DESIGN AND SIMULATION OF NATURAL GAS PROCESSING

BY

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This thesis, written by **Aban Abdullah Sakheta** under the direction of his thesis advisor and approved by his thesis committee, has been presented and accepted by the Dean of Graduate Studies, in partial fulfillment of the requirements for the degree of **MASTER OF SCIENCE IN CHEMICAL ENGINEERING**.

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Dedicated to my beloved parents

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LIST OF ABBREVIATIONS

CO_2	:	Carbon Dioxide
AGR	:	Acid Gas Removal
NGL	:	Natural Gas Liquids
SGP	:	Shedgum Gas Plant
KGP	:	Khursaniyah Gas Plant
H_2S	:	Hydrogen Sulphide
DWC	:	Dividing Wall Column
CDWC	:	Conventional Dividing Wall Column
DGA	:	Diglycolamine
MDEA	:	Monodiethanolamine
DEA	:	Diethanolamine
MEA	:	Monoethanolamine
TEG	:	Triethyleneglycol
deC_1	:	Demethanizer
deC_2	:	Deethanizer
deC3	:	Depropanizer

deC_4 : Debutanizer

deiC_4 : Deisobutanizer

C_1 : Methane

C_2 : Ethane

C_3 : Propane

nC_4 : n-butane

iC_4 : i-butane

ABSTRACT

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Recently, the amount of greenhouse gas emissions such as carbon dioxide in the atmosphere increases continuously. This tendency leads to several issues harming the environment such as global warming. Using natural gas as a fuel source instead of other fusel fuels emits lower amount of carbon dioxide in the atmosphere. Several treatment steps must be done before start producing the natural gas namely acid gas removal, dehydration and NGL fractionation. The objective of this thesis is to study and model these treatment steps in details and filling some gaps in this area. Two models employing two different solvents were analyzed in the acid gas removal section namely, DGA and MDEA. The results revealed that the plant employed MDEA requires less energy compared to DGA one. Energy saving opportunities were investigated using different design alternatives. Mechanical Vapor Recompression configuration was found to be superior to other designs and more energy efficient while maintaining the product purity.

The analogy between natural gas and carbon capture dehydration using two different design approaches namely, conventional and stripping gas was investigated in the dehydration part. Results showed that the stripping gas configuration requires less amount of circulation rate and energy maintaining the same water removal level as the conventional scheme for both cases. NGL fractionation process consists of a connecting series of distillation columns to separate the NGL components. Dividing wall column sequence was selected as an alternative design for this part. Results show that using dividing wall column arrangement can save 24% of the energy requirement compared to the conventional sequence.

ملخص الرسالة

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كمية انبعاثات الغازات الدفيئة في الغلاف الجوي مثل ثاني أكسيد الكربون في ازياد مستمر، هذه الزيادة تؤدي إلى مشاكل بيئية جمة مثل الاحتباس الحراري، استخدام الغاز الطبيعي كمصدر وقود صناعي يؤدي إلى انبعاثات أقل من الغازات الدفيئة إلى الغلاف الجوي، الغاز الطبيعي يحتاج إلى المعالجة قبل البدء باستخدامه أو تصديره ومنها إزالة الغازات الحمضية ونزع الماء وتقطيع وفصل مكوناته، تهدف رسالة الماجستير هذه لدراسة ومحاكاة عمليات المعالجة المذكورة بالتفصيل، تمت دراسة نموذجان في جزء نزع الغازات الحمضية، الأول يستخدم المذيب (DGA) والآخر يعالج باستخدام (MDEA)، أشارت نتائج هذه الدراسة أن الطاقة المطلوبة في المصنع الذي يستخدم (MDEA) أقل من نظيرتها في المصنع الذي يستخدم (DGA)، وشمل هذا الجزء من الدراسة محاكاة تصاميم مصنعية بديلة لبحث فرص توفير الطاقة اللازمة لعمل المصنع، وتوضح النتائج بأن أفضل تصميم بديل هو ضغط البخار الميكانيكي القادر على توفير أكبر قدر ممكن من الطاقة.

تمت في جزء نزع الماء دراسة التشابه السلوكي بين عمليتا نزع الماء من الغاز الطبيعي ونزع الماء من الكربون الملتقط باستخدام التصميم المتعارف عليه وتصميم الغاز النازع، نتائج هذا الجزء تشير إلى أن استخدام طريقة الغاز النازع تتطلب كمية أقل من الطاقة والمذيب مع مراعاة المساواة في كفاءة نزع الماء بين التصميمين، هذه النتيجة المذكورة تشمل عمليتا نزع الماء من الغاز الطبيعي ومن الكربون الملتقط، عملية تقطيع الغاز الطبيعي السائل تتكون من سلسلة متتالية من أعمدة التقطير لفصل مكونات الغاز الطبيعي السائل عن بعضها البعض، ولقد تم اختيار تصميم أعمدة التقطير بحائط قاسم كتصميم بديل للتصميم المتعارف عليه لبحث سبل توفير الطاقة، نتائج هذا الجانب من هذه الدراسة تشير إلى أن استخدام هذا التصميم البديل يساعد على توفير 24% من الطاقة المطلوبة مقارنة بالتصميم المتعارف عليه.

CHAPTER 1

INTRODUCTION

1.1 Overview Background

The continuous increase in the amount of carbon dioxide (CO_2) contributes negatively to climate change and cause global warming. Recently, CO_2 emission level has reached 399 ppm and is expected to increase continuously if no preventive measure is taken (Anon 2016a). Among other strategies to limit the emissions, use of clean fuel have been one of the main approaches to mitigate the impact of climate change. Natural gas is considered as a clean energy source and produce 71% and 56% less emissions compared to oil and coal, respectively (Lim et al. 2013). Conventionally, natural gas is produced from the underground reservoirs together with or without crude oil. The natural gas recovered along with oil is referred as associated gas while the gas produced without much oil phase is referred to as non-associated gas. Usually, the associated gas contains lower methane and higher amount of high molecular weight hydrocarbons than the non-associated gas. Before the natural gas can be used for any domestic and industrial application, it must be processed to purify it from the undesired impurities. Also, the conditioning of natural gas is important for its transmission, which if not done, can be disastrous to pipelines and personnel health. There are several stages in the natural gas processing as shown in figure 1. First, the gas is pre-processed to remove any free water or associated liquid phases. It is then sent to the acid gas removal (AGR) section for the removal of acid gases before entering the dehydration plant. The purpose of dehydration unit is to remove water contents below certain level so that gas hydrate formation can be prevented. Finally, the gas is either sent as sales gas or further processed in fractionation units to separate C_1 -C_5 hydrocarbons

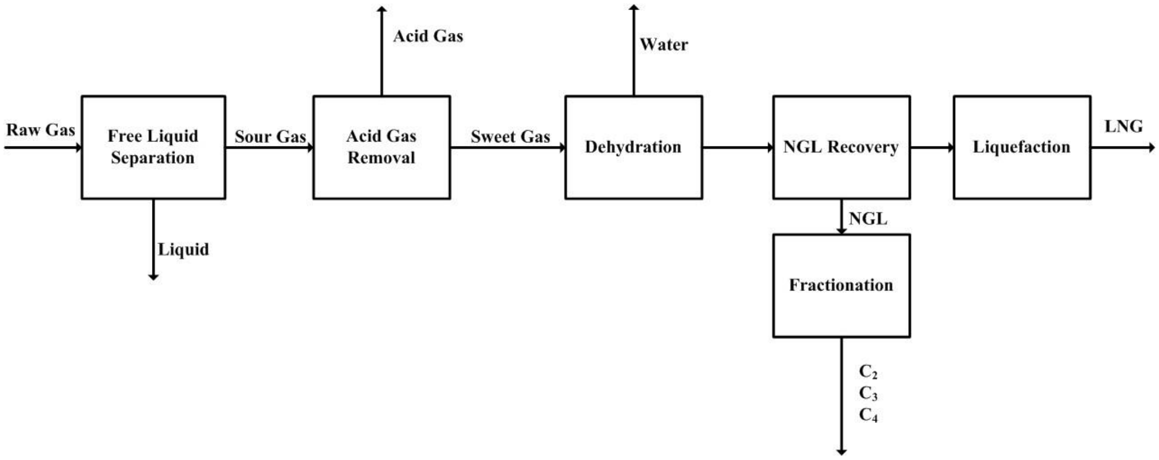

Figure 1 Stages Involved in the Natural Gas Processing

1.1 Objectives

The objectives sought for this project are as following:

Acid Gas Removal

1- Process model development for the AGR
2- Model validation with the literature
3- Improve the process performance maintaining the product purity.
4- Development of several alternative designs to look for energy savings opportunity

Dehydration

1- Process model development for the dehydration for natural gas and carbon capture cases.
2- Model validation with the literature
3- Improve the process performance maintaining the product purity.
4- Comparison the analogy between the natural gas and carbon capture dehydration

NGL Fractionation

1- Process model development for the NGL fractionation columns
2- Parametric study on some of the important design variables
3- Alternative design development such as dividing wall column for process improvement

1.2 Layout of Thesis

This thesis mainly focuses on the three natural gas treatment stages namely, AGR, dehydration and natural gas liquids (NGL) fractionantion, sections. Two representative acid gas removal plants namely Shedgum and Khursaniyah were selected based on the source of natural gas. Shedgum gas plant (SGP) receives the low pressure associated natural gas, while Khursaniyah gas plant (KGP) processes the high pressure non-associated gas. These two plants employed two different amine solvents to absorb acid gas primarily, CO_2 and H_2S. The aim of this section is to study and analyze the sweetening and the energy performance of the two cases. The model used is validated against plant data to check the reliability of the model. Finally, parametric analysis is performed to investigate the effect of some key parameters on the removal and energy performance. The sweetened gas stream from the KGP is selected to be the feed to the dehydration plant where water is removed at this stage. The dehydration model is validated against plant data as well. The analogy between the natural gas and captured CO_2 dehydration is investigated by performing sensitivity analysis for both scenarios. The dried gas stream is then sent to the fractionation unit where C_1 – C_5 hydrocarbons are separated. In this section, energy saving opportunities using dividing wall column (DWC) designs has been investigated.

CHAPTER 2

LITERATURE REVIEW

2.1 Acid Gas Removal

This study will mainly focus on the use of amine solvents for the gas treatment. Chemical amines can be mainly classified into three main categories, namely primary, secondary and tertiary amines. Monoethanol amine (MEA) and diglycolamine amine (DGA) are examples of primary amines, whereas diethanolamine (DEA) and monodiethanolamone (MDEA) are representative examples of secondary and tertiary amines, respectively. Furthermore, more recently sterically hindered amines such as 2-amino-2 methyl-1-propanol (AMP), 1,8-p-methanediamine (MDA) and 2-piperidiene ethanol (PE) got attention for their capability to efficiently absorb acid gases. However, these amines are still at the development stage and deemed too expensive for commercial use. Each of these above mentioned amines have their own characteristics and limitations, and have been used in various industries.

In the past, many studies investigated the performance of DGA and MDEA solvents. However, not many studies have analyzed the application of these solvents in the actual industrial plants. Specially, open literature on the performance and application of DGA is rarely reported. Pellegrini et al. (Pellegrini et al. 2010) compared various amines that can be employed for the CO_2 capture. The authors investigated the performance of MEA, DGA and ammonia in terms of their capture capability and found that the NH_3 is superior to other solvents. However, when using ammonia solvent, salt formation can occur and require another separation unit to remove solid salts from the streams. Erfani et al. (Erfani et al. 2015) also compared between various amines including MEA, DGA, MDEA, DEA and DIPA. Their results showed that the mixture of MEA-DGA amine can reduce the reboiler duty compared to the MEA base case. Kazemi et al. (Kazemi et al. 2014) investigated the effect of CO_2 content in the lean amine for MEA, DGA and MEA-MDEA mixture on the acid gas removal process. They reported that the use of DGA solvent to be more economical compared to the other processes. More recently, Zahid et al, (Zahid et al. 2017) performed an energy analysis of a gas sweetening unit employing DGA solvent and estimated the reboiler duty to be 5.57 GJ/ton CO_2 removed. However, this study considered only CO_2 in the sour feed gas and did not investigate the effect of H_2S presence in the feed.

Similarly, many researchers have reported the application of MDEA in the gas sweetening industry. For example, Borhani et al.,(Borhani et al. 2016) used rate-based model to describe the absorption and reactive system between MDEA and acid gases. They investigated the impact of some key parameters and found that decreasing the surface area of packing increases the removal efficiency. Also, their results showed that increasing the liquid flow rate in the column can enhance the overall mass transfer coefficient leading to a better absorption. Aliabad and Mirzaei (Aliabad & Mirzaei 2009) studied the MDEA and DEA solvents and found that increasing the circulation rate and concentration of the solvents increase the removal of acid gases. Jassim (Jassim 2016) performed a parametric analysis and reported that the MDEA concentration and lean amine flow rate are the most important parameters which influence the energy performance and removal efficiency. Mofarahi et al., (Mofarahi et al. 2008) compared the performance of various amines for the acid gas removal application and found that MDEA is a good solvent for selective removal of H_2S. Fouad and Berrouk (Fouad & Berrouk 2013) compared blends of various amines and found the combination of MDEA/TEA to have the least regeneration energy compared to the other amines. Al-qahtani and Garland (Al-Qahtani & Garland 2013) investigated the foaming problem in Khursaniyah plant which employs MDEA as the chemical solvent. The authors recommended increasing the sour gas temperature along with the use of antifoaming solvent to avoid foaming in the absorber.

2.2 Dehydration

Several dehydration methods are available depending on the downstream requirement including condensation by cooling, adsorption, absorption and membranes (L. L. Faulkner 2006) as shown in figure 2. The simplest dehydration method is condensation by cooling which involves cooling the gas in order to condense the water content which can then be separated. However, water removal by cooling can only achieve water contents down to 600 ppmv in the dried gas which in most of the applications is higher than the required limits. Another drawback of this process is the formation of methane hydrates which require hydrate inhibitors. Adsorption can be done by a variety of desiccants such as silica gel, activated alumina and molecular sieves. The amount of water removed mainly depends on the type of desiccant used and the feed gas thermodynamic conditions. Usually, the moisture removal efficiency by adsorption increases with an increase in the pressure and decrease in the temperature. Various desiccants have their own characteristics and are typically employed for adsorption in a cycle of two beds with one bed under adsorption while the other regenerates the solid adsorbent. Regeneration can be performed either by temperature swing adsorption (TSA) or pressure swing adsorption (PSA). However, the TSA process is more common employed than that of the PSA process. Adsorption can achieve very low water concentration in the dried gas with dew point as low as < -50 ^{o}C. However, adsorption offers high capital and operating cost compared to the other technologies. Studies suggest that CAPEX for adsorption can be 2 -3 times higher than that of the absorption process (Netusil & Ditl 2011; Kinigoma & Ani 2016). More recently, membranes received significant attention for the dehydration of natural gas because of their less weight, large turn down ratio and ability to operate at medium to high pressures (50-70 barg). However, throughput scale-up and a pretreatment stage requirement for solids and droplets (> 3micrcons) removal pose a technical challenge (L. L. Faulkner 2006). Absorption is the most widely used method employed for the purpose of dehydration. Liquid solvent usually a glycol is used to remove the moisture content from the feed gas. The common glycols include diethylene glycol (DEG), triethylene glycol (TEG), ethylene glycol (EG) and tetraethylene glycol (TREG). The glycols are preferred choice of solvent because of their ability to selectively absorb water with no significant loss of natural gas, low viscosity and volatility at absorption conditions, and thermally stable with low corrosion rate. In this study, absorption process using TEG solvent has been selected for the investigation of water removal from the natural gas and CO2 streams. In industrial dehydration processes, TEG is a preferred solvent because of its superior dew point depression characteristics, low cost and high operational reliability compared to the other glycols.

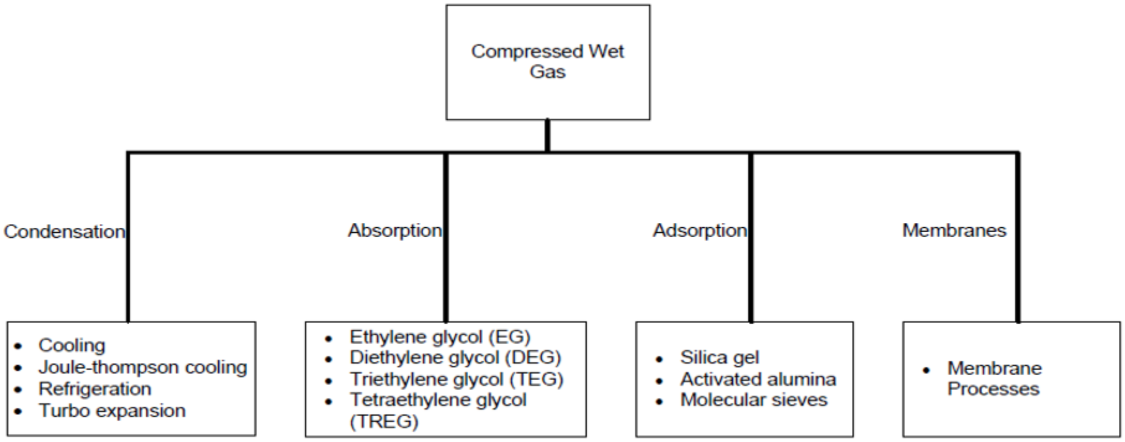

Figure 2 Gas Dehydration Methods

2.2.1 Natural Gas Dehydration

Dehydration has been employed industrially for many decades and significant literature is available on various dehydration technologies. For example, Netusil and Ditl (Netusil & Ditl 2011) compared absorption, adsorption and condensation technologies over a wide operational range in terms of energy consumption. Their results showed that the adsorption is the highest energy consuming process compared to the other technologies. They also showed that the absorption is a better choice for low pressure operation while the condensation is suitable for high pressure process. Similar results were reported by Kinigoma and Ani (Kinigoma & Ani 2016).Their results showed that the condensation can only remove water content up to certain level which limits the process performance. Nemati Rouzbahani et al., (Nemati Rouzbahani et al. 2014) simulated the natural gas dehydration process employing DEG as the solvent and reported that the volatile organic compounds emissions have a great influence on enhancing the purity of DEG in the stripper. Neagu and Cursaru (Neagu & Cursaru 2017) simulated two design configurations of natural gas dehydration and showed that the TEG concentration of the regenerated glycol can be increased from 98.74 % to 99.85 % when stripping gas configuration is used without substantial increment in the capital investment. Anyadiegwu et al (Anyadiegwu et al. 2014) studied the effect of TEG flow rate on the water removal performance. Their study results showed that, at 92 bar and 30 ^{o}C absorption conditions, 3.5 m^3/h is sufficient to remove the water up to 6.8 Ib/MMSCF from the gas stream flowing at 10 MMSCFD. Similar work was done by Collins et al,. (Collins et al. 2015) at relatively lower pressure operation i.e. 62 bar. Simulation results showed that 10 MMSCFD gas flow rate required 53 to 70 L/h solvent rate to obtain 6 to 7

Ib/MMSCFD water concentration in the gas stream, respectively. Mawgoud et al (El Mawgoud et al. 2015), discussed the opportunities to revamp an existing natural gas dehydration plant. Their results revealed that decreasing the lean TEG temperature before the absorption leads to higher water removal efficiency. Rahimpour et al (Rahimpour et al. 2013), investigated the influence of several parameters on the dehydration performance. They reported that increasing the pressure operation, glycol concentration, glycol flow rate can enhance the water removal process and cost saving. Felicia and Evboumwan (Felicia & Evbuomwan 2015) performed an optimization study of natural gas dehydration plant that employed TEG. Their analysis implied that by varying the circulation rate of TEG, an optimum point of the water purity in the dry gas stream exists.

2.2.2 Carbon Capture Dehydration

Several studies were done in carbon capture dehydration process as well. Abbas et al, (Abbas et al. 2013) compared refrigeration, adsorption and absorption in terms of water removal performance. Their analysis showed absorption by EG can be applied at wide range of circulation rate and the water can be removed below 50 ppmv. However, the refrigeration is the most advantageous among other technologies for CO_2 dehydration application because it requires lower cost than absorption. Oi and Fazlagic (Øi & Fazlagic 2014), simulated several retrofit designs for CO_2 dehydration process using two different equilibrium models. Their results showed that injecting the stripping gas to an additional stripping column that is connected to the main regenerator can reduce the water content in the dry gas more than injecting the stripping gas to the reboiler of the stripper. Wise and Chapoy (Wise & Chapoy 2016), analyzed the carbon dioxide solubility in TEG over different operating conditions. Their model results are agreed with the experimental findings. According to their results, as pressure increases, the solubility of CO_2 in TEG increases as well for a wide range of temperatures. Also, the presence of water can reduce the solubility of CO_2 in TEG substantially.

2.3 NGL Fractionation

NGL recovery is done in the demethanizer where methane is separated from NGL and leave the column from the top while NGL is recovered in the bottom of the column. The most concerning issue regarding demethanizer is that it requires low temperature operation to partially condense it. Many retrofit designs are used to reduce the plant capital and operating cost and save the energy for cooling the feed gas such as utilizing the low temperature from the overhead product, using some external refrigeration cycles such as propane refrigeration and turboexpansion of feed gas which will consequently results in low overhead temperature.

The NGL fractionation process is done by a connected series of distillation columns. Conventionally, the number of distillation columns required to get the pure NGL components depends primarily on the number of products desired. Separating ternary mixtures can be done by different sequences or configurations and it requires at least two columns to perform the separation. The simplest methods are direct splitting and indirect splitting. The direct sequencing involves separating the lightest component in the first column and the heavier components are leaving at the reboiler stage of the first column and then isolated to their pure components in the second column. The indirect splitting sequence separates the heaviest component at the first column while the other components enter the second column to obtain their pure products. Another possible configuration is using three columns where the first column performs sloppy separation. The top product of the first column is the light component and some portion of the second component and it is fed to the second column while the bottom product consists of the other portion of the second component and the heavy key product and it enters the third column. The separation to the pure component is continued in the second and third column. However, the disadvantage of the mentioned columns, aside from high energy requirement, is lack of ability to perform reversible split. To overcome these drawbacks, a thermally linked column was proposed, which simply combining the second and third column of the sloppy configuration in one shell and the intermediate component is withdrawn from the middle side of the second column. another improvement reported is to construct one main column along with side rectifier or side stripper. The light and heavy components are withdrawn from the main column while the middle component leaves from the side column. One reboiler or condenser can be omitted using these two configurations. Another energy efficient column design is the fully thermally coupled column, namely Petlyuk column, which consists of prefractionator and main columns as shown in figure 3. The overhead, bottom and side products are withdrawn from the main column. This arrangement can reduce the capital cost and save energy since a single condenser and a single reboiler are used to supply the whole sequence with heat. Combining the prefractionator with the main column into a single shell with a welded wall in the middle section is the recent configuration proposed and it is known as dividing wall column (DWC) as shown in figure 4. Moreover, the idea of top divided wall column and bottom divided wall column arises from integrating the two columns into one single shell in the side rectifier and side stripper, respectively, as well (Asprion & Kaibel 2010; Dejanovic et al. 2010).

Many simulation and optimization studies were done in NGL fractionation to compare the direct split and more advanced arrangements. Long and Lee (Long & Lee 2012) investigated several design alternatives to improve the performance of the deethanizer and depropanizer. The results revealed that using DWC arrangement reduces the total annual cost (TAC) up to 33.63%. The same authors in another study. Uwitonze et al, (Uwitonze et al. 2016) proposed new process schemes which involve a heat and energy integration between LNG production and NGL recovery processes. Their results showed the heat integration between dual mixed refrigerant (DMR) and fully thermally coupled distillation column configuration leads to lower total reboiler duty than DMR-conventional and DMR-heat integrated distillation sequences. For total condenser duty, DMR- heat integrated distillation sequence has the lowest value among all cases. Factorial design method was used to design and optimize dividing wall in the work of Long and Lee (van Duc Long & Lee 2012). After optimizing DWC which separates i-butane, n-butane and pentane components, energy and cost analysis showed that using DWC can reduce the reboiler duty, investment cost and TAC by 25.6, 13.7 and 18.9%, respectively. Chew et al., (Chew et al. 2014) investigated several design approaches to enhance the energy performance of the DWC in various well-known processes including depropanizer and debutanizer. Their results show that vapor recompression, closed cycle compression, absorption heat pump, absorption heat transformer, organic Rankine cycle and Kalina cycle configurations are thermodynamically inefficient and costly. Long et al.,(Long et al. 2016) studied various retrofit designs to reduce energy consumption in

NGL fractionation. Optimization results show that using side reboiler and heat pump-assisted, thermally coupled distillation to integrate deethanizer and depropanizer minimizes the operating cost. For debutanizer and deisobutanizer, vapor remporession heat pump system is used for each column and both top products are used to supply heat to the reboilers of the deisobutanizer. This retrofit design can save 52.8 and 68.80% of operating cost and reboiler energy, respectively. Ching et al., (Ching et al. 2016) conduct a research to investigate the feasibility of distinct retrofit designs for depropanizer column using commercial simulation software. their analysis showed that mechanical vapor recompression and vapor compression save more than 65% of energy cost. Another study (Van Duc Long & Lee 2012) investigated the possibilities of separating quaternary mixtures that contains propane, i-butane, n-butane and propane+ using DWC configurations. After comparing the results with the conventional sequence, double prefractionator arrangement and double dividing wall column are the most efficient configurations which lowers the energy consumption substantially.

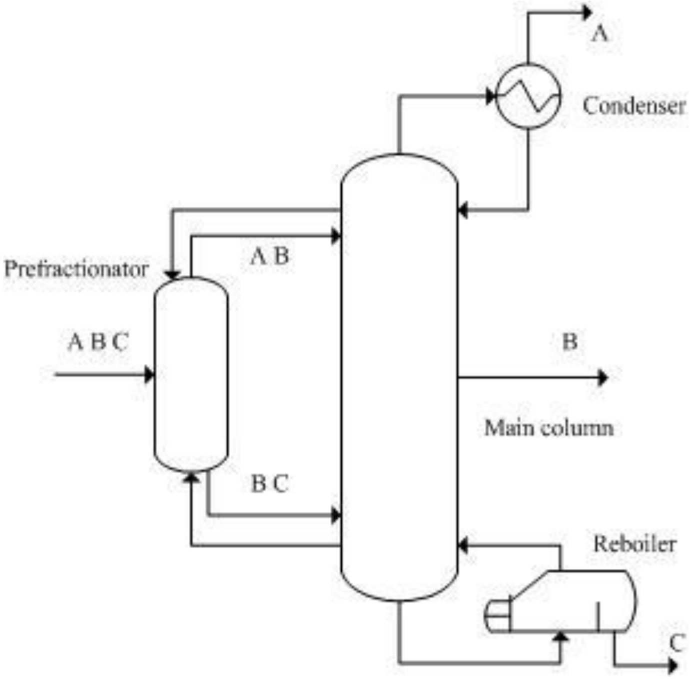

Figure 3 Petlyuk Column

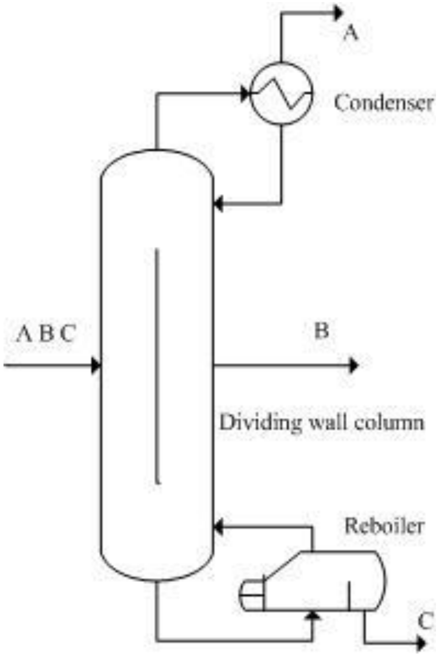

Figure 4 Conventional Dividing Wall Column

CHAPTER 3

ACID GAS REMOVAL

3.1 Introduction

The scope of this chapter is limited to the AGR section where the acid gases, mainly H_2S and CO_2, are removed from the gas stream. Depending on the feed conditions, composition and operating environment, a number of methods can be employed for the removal of acid gases. The common industrial methods for the removal of acid gases include treatment using chemical amines, physical solvents and mixtures of physical and chemical solvents (L. L. Faulkner 2006). The natural gas produced in the Saudi Arabia comes from both associated and non-associated sources. In 2015, the non-associated gas accounted to 80% of natural gas production while the rest 20% was associated gas (Anon 2016b). Inherently, the associated gas is available at low pressure compared to the non-associated gas which is mostly available at a high pressure. The natural gas produced in the Saudi Arabia is mainly directed to six main gas plants, namely Haradh, Hawiyah, Uthmaniyah, Khursaniyah, Berri and Shedgum. Figure 5 shows the source of natural gas that is being treated at each of the processing facility along with the solvent employed for gas treatment (McMurray 2011). It is interesting to see that four of the gas plants employ diglycolamine (DGA) as the chemical solvent for the gas cleaning while the Khursaniyah and Haradh employ monodiethanolamine (MDEA) for gas treatment. Although acid gas removal processes have been used for many decades, however, the details on how an industrial plant performance can be manipulated is not extensively reported in open literature. The goal of this study is to investigate the dynamics of natural gas sweetening process especially in the Middle East region. For this reason, this study has selected two representative natural gas processing plants from the Saudi Arabia, namely Shedgum and Khursaniyah gas plants. Shedgum gas plant (SGP) receives low pressure associated natural gas and DGA solvent is used for the acid gas removal. On the other hand, Khursaniyah gas plant (KGP) which receives high pressure non-associated natural gas and employs MDEA as the chemical solvent. It is well understood that the two solvents belong to different class of amines and a direct performance comparison cannot be drawn. However, the goal of this chapter is to investigate the performance of two different cases from associated and non-associated sources available at different pressures. In the presence of both H_2S and CO_2 in the feed gas, MDEA is mainly utilized for the selective removal of H_2S, while DGA is used for the removal of both acid gases. In this study, first a steady-state model is simulated using a commercial software for each of the gas plants. The models are then validated against the available actual plant data in order to ensure the model reliability. Finally, a sensitivity analysis for both the models has been done to observe the impact of key process parameters such as circulation rate, solvent strength, stripper pressure and lean amine temperature on the energy performance and product purities.

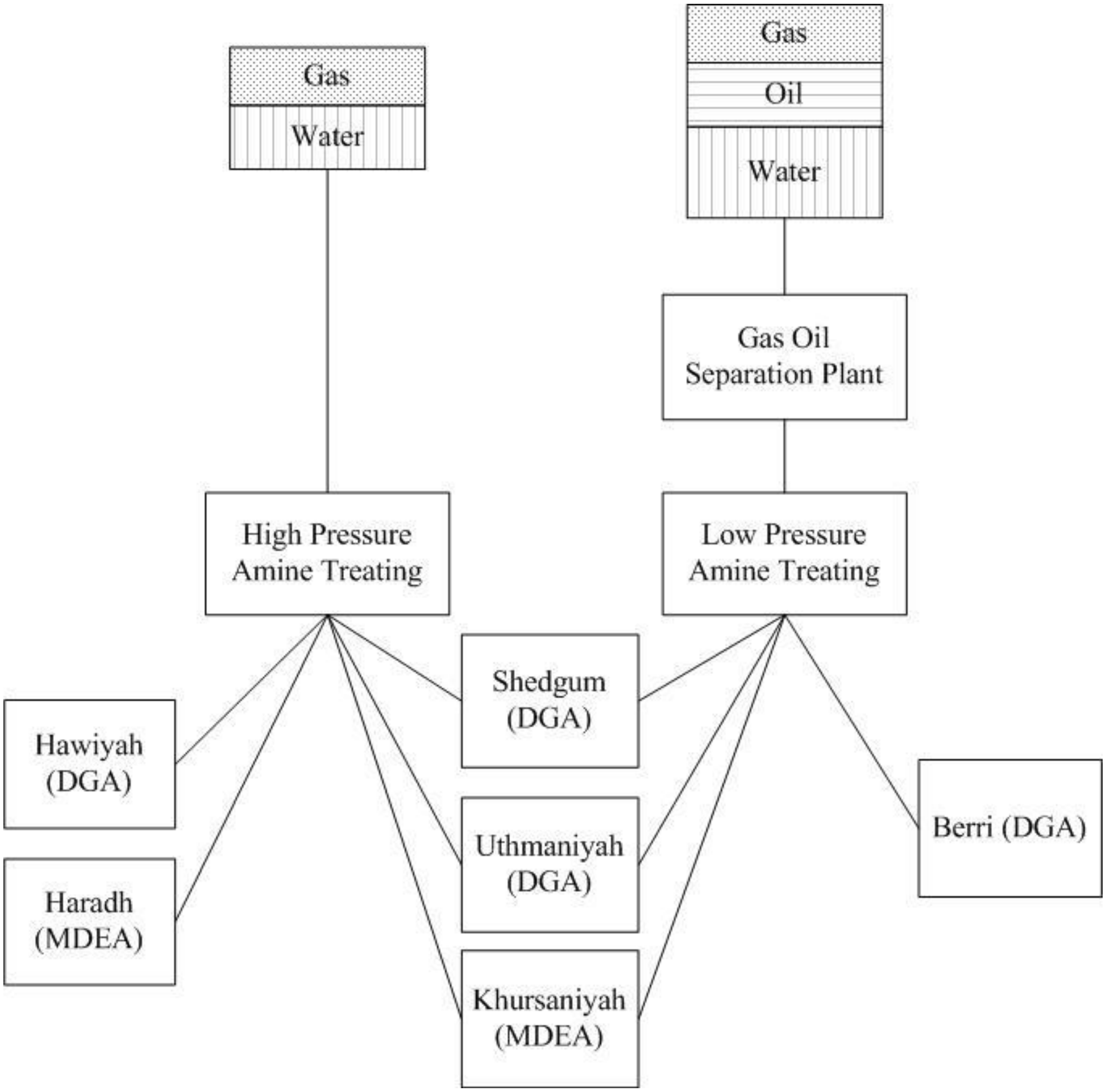

Figure 5 Natural Gas Processing Plants in the Saudi Arabia Employing Amines as the Solvent for Gas Sweetening

3.2 Design Basis

This study considered two representative base cases namely Shegum and Khursaniyah Gas plant. The base case flowsheet structure is same for both the designs. However, the feed specification and solvent employed is different. The base case configuration for the Shedgum gas plant considers the associated feed gas available at a low pressure. The feed composition along with other specification for the SGP is shown in table 1.

Table 1 Feed Gas Specifications for Shedgum Gas Plant

Parameters	Value
Inlet gas flow rate (MMSCFD)	249
Inlet gas temperature (^oC)	38
Inlet gas pressure (bar)	12.76
C_1 in sour gas (mol%)	56
C_2 in sour gas (mol%)	12.7
C3 in sour gas (mol%)	9
C_4+ in sour gas (mol%)	10.5
H_2O in sour gas (mol%)	0.9913
BTEX in sour gas (mol%)	0.0311
H_2S in sour gas (mol%)	2.9
CO_2 in sour gas (mol%)	7.9

The feed has relatively high amount of higher hydrocarbons because the feed is coming from an associated source. The acid gas content is also high compared to the other regional feeds containing H2S and CO_2 mole percent of 2.9 and 7.9 % respectively. Sour feed gas at SGP is treated using DGA solvent. The allowable DGA concentration in the solution is usually in the range of 35 – 70 wt. % which can lead to lower solvent circulation rates compared to that of the MEA solvent (Mofarahi et al., 2008). However, the degradation of DGA solvent in the presence of CO_2 adds up to the solvent makeup cost. Also, when both H_2S and CO_2 are present in the natural gas, DGA is not suitable for the selective removal of H_2S. In this study, 50 wt % DGA solution is employed at the SGP with a circulation rate of 2862 GPM for a sour gas feed flow rate of 249 MMSCFD as shown in table 1.

The base case for the KGP receives the sour feed from a non-associated high pressure source. KGP employs MDEA solvent for the sweetening process. Since the non-associated feed is usually available at high pressure, there are some benefits associated with the high pressure gas sweetening process. High pressure operation tends to enhance the removal performance that can lead to a decrease in required circulation rate and consequently decrease in the reboiler duty. However, operating absorber at a high pressure requires more pumping energy. Therefore, there is a trade-off between pumping energy requirement and reboiler duty reduction at high feed gas pressure that must be optimized for required product purity. KGP is one of the largest gas processing plants having three processing trains of identical capacity with total processing capacity of 1860 MMSCFD (Al-Qahtani & Garland 2013). This study simulated one of those identical process trains. Table 2 shows the feed gas composition along with the feed specifications for the KGP. The feed to KGP comes from Karan gas field which has around 2 mol% H2S and high CO_2 content of around 8 mol%. In this study, 45 wt% MDEA solvent is used to selectively absorb the H2S from the sour feed gas.

Table 2 Feed Gas Specifications for Khursaniyah Gas Plant

Parameters	
Inlet gas flow rate (MMSCFD)	620
Inlet gas temperature (^oC)	36
Inlet gas pressure (bar)	61
C_1 in sour gas (mol%)	82.47
C_2 in sour gas (mol%)	1.62
C3 in sour gas (mol%)	0.33
C_4+ in sour gas (mol%)	0.39
H_2S in sour gas (mol%)	2.06
CO_2 in sour gas (mol%)	8.3
N_2 in sour gas (mol%)	4.83

MDEA is a tertiary amine with typical concentration of 20-50 wt% in the gas sweetening applications. The weight concentration of MDEA used is proportional to the operating pressure i.e. lower pressure operations requires low concentration of MDEA to remove H2S efficiently. In order to ensure that the amine is not degraded at a high temperature, the operational temperature range should be between 25-127 ^oC. MDEA has unique characteristics among other amines, for example low heat of reaction, low vapor pressure and low degradability. Also, MDEA causes less corrosion compared to other amines provided it does not react with oxygen to form corrosive acids.

The design requires the sweet gas to have less than 4 ppm of H_2S. For CO_2 the acceptable range for DGA process is to have less than 100 ppm in the sweet gas. The required product purities must be achieved by operating the plant within reasonable range of rich and lean loadings that result in the minimum energy requirement ensuring solvent is not thermally degraded. This section aims to briefly discuss the thermodynamic model used to perform the simulation. Moreover, Process flowsheet for the base case scenario for SGP and KGP and alternative designs for SGP will be described along with their advantages.

3.2.1 Thermodynamic Background

Alkanolamine solutions are polar and highly complex due to strong interaction forces between their molecules. There are some empirical thermodynamic models that were used in the past for describing the behavior of alkanolamine systems such as Gabrielsen model (Gabrielsen et al. 2005). However, since these models do not handle the non-ideal behavior with significant accuracy, alternative rigorous models should be utilized to describe the thermodynamic properties of alkanolamines-CO_2-H_2S systems. Activity coefficient models based on excess Gibbs free energy can be used for the liquid phase with association of equation of states in the vapor phase. This study employed the electrolyte nonrandom two liquids (eNRTL) activity coefficient model for liquid phase associated with Peng Robinson (PR) equation of state for vapor phase. The reactions taking place in the absorber between the amine solvents and acid gases are exothermic in nature. The regeneration process is opposite to that of the absorption and involves endothermic reactions.

The essential feature of using MDEA is selective absorption of H_2S and is usually used when sulfur recovery is required. The key reactions taking place between alkanolamine solutions with CO_2 and H_2S are shown below:

Ionization of water:

(2.1)

Ionization of dissolved H_2S

(2.2)

Hydrolysis and ionization of dissolved CO_2:

(2.3)

Protonation of amine

(2.4)

Carbamate formation

(2.5)

All of the above reactions occur for the primary amines including DGA. However, in case of MDEA, the reaction (5) of carbamate formation is absent. Since Henry’s law can be applied to determine the equilibrium concentration of CO_2 and H_2S in the solution, the forward pathway of second, third and fifth reaction can be enhanced by increasing the partial pressure of CO_2 and H_2S. Alkalinity of amines is the prime reason for their effective performance towards absorption of acid gases (Kohl & Nielsen 1997).

3.2.2 Process Flowsheet

3.2.2.1 Base Case Scheme

The gas sweetening process consists of two steps. First, the lean amine and sour gas contact counter currently in the absorber where the acid gases are absorbed in the solvent. The rich amine from the bottom of the absorber is then sent to the regenerator where the acid gases are stripped from the solvent by the addition of energy to the reboiler. The typical absorption-stripping configuration is shown in figure 6. The lean amine with known circulation rate and concentration contacts with the sour gas to chemisorb acid gases (CO_2 and H_2S). The rich solvent is then throttled and flashed using a flash drum to separate the light hydrocarbons. The rich amine stream is then heated in the rich/lean heat exchanger before feeding it to the full reflux stripper column. The degree of freedom for the stripper is two, therefore two more specifications are required to solve the column stage by stage. The acid gases leave at the top while the lean amine leave at the bottom of the regenerator. Finally, the lean amine is mixed with a makeup stream to maintain its flow rate and amine concentration. The lean amine is then pumped and cooled by a trim cooler, before being recycled back to the absorber column. The key input specifications for the simulation of the two plants namely SGP and KGP are listed in the table 3. All the stage numbering sequence followed in this study is top to bottom.

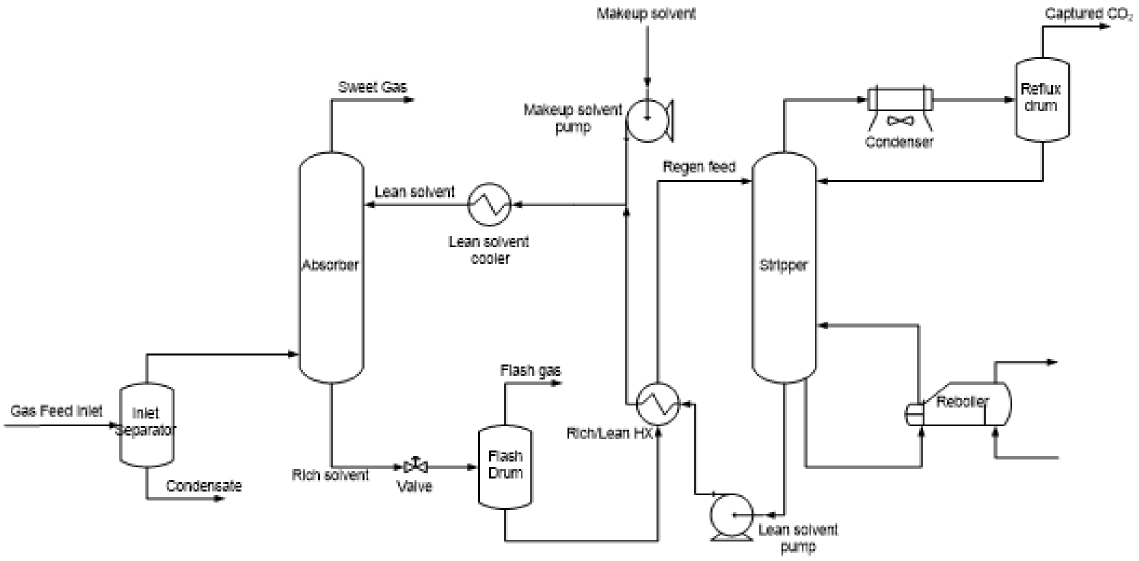

Figure 6 Process Flow Diagram for Acid Gas Removal Process

Table 3 Design Specifications Used in the Simulation for SGP and KGP

Parameters	SGP	KGP
Number of absorber trays	23	20
Number of stripper trays	20	20
Stripper top temperature (^{o}C)	70	70
Stripper bottom Temperature (^{o}C)	130	121
Absorber top pressure (bar)	12.76	51
Absorber bottom pressure (bar)	12.80	52
Stripper pressure (bar)	2.2	1.85
Stripper inlet temperature (^{o}C)	105	80
Solvent weight percent	50	45
Solvent Circulation rate (USGPM)	2862	3082

3.2.2.2 Absorber Intercooling

Several approaches can be applied to enhance the absorption capability. Absorption of acid gases by amines is favored by low temperature since the reaction is exothermic. In order to provide the desired low temperature to the absorber, an intercooling cycle has been inserted to the absorber where small liquid portion is withdrawn from the column, cooled and entered back to another tray as shown in figure 7. Since the efficiency of the absorber is improved, lower circulation rate than the conventional design can be used to achieve the required product purity. The selection of the withdrawn stage and the inlet cooled stream is based on product purity obtained.

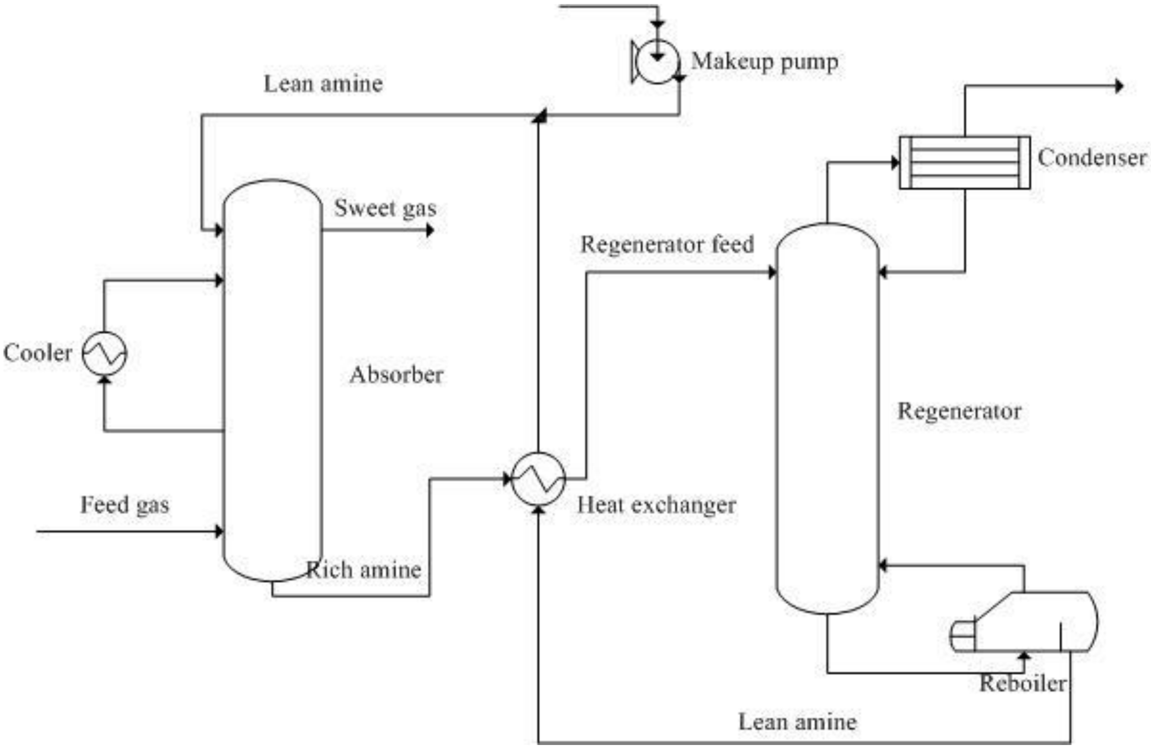

Figure 7 Absorber Intercooling Configuration

3.2.2.3 Split Lean Solvent Flow Scheme

This scheme helps to reduce the regeneration energy required in the reboiler by withdrawing some of the liquid flow rate from the middle stages in the regenerator. This withdrawn stream, known as semi lean solvent, is recycled back to the middle stages of the contactor. This design requires also an additional heat exchanger where the heat energy from the rich solvent can be utilized to precool the semi lean amine stream as shown in figure 8. The selection of withdrawn stage location from the stripper and the inlet feed stage location of the absorber depends mainly on the product purity and reboiler duty.

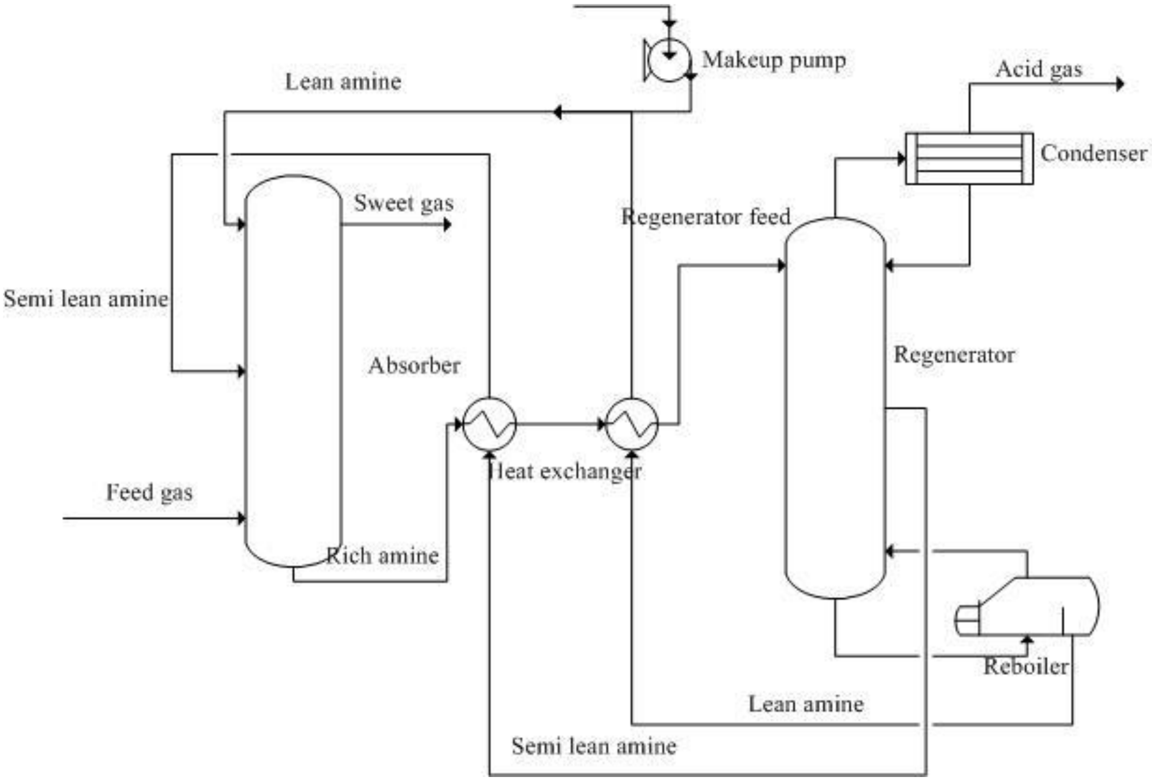

Figure 8 Split Lean Solvent Flow Process Flow Diagram

3.2.2.4 Split Lean and Rich Solvent Flow Scheme

Figure 9 shows a process flow diagram of a split solvent and rich amine scheme. This scheme is considered more complex than There are two main differences between split flow and split solvent and flow. Firstly, the rich amine that exits the contactor is splitted into two streams. These two streams are used to precool the lean amine and semi lean amine streams that leave the regenerator. Secondly, the semi lean amine stream is mixed with a makeup solvent to maintain the same concentration as the lean amine. The side draw stage location from the stripper, inlet side tray location in the absorber and the semi lean amine circulation rate are the key parameters that must be selected carefully in the split flow configuration.

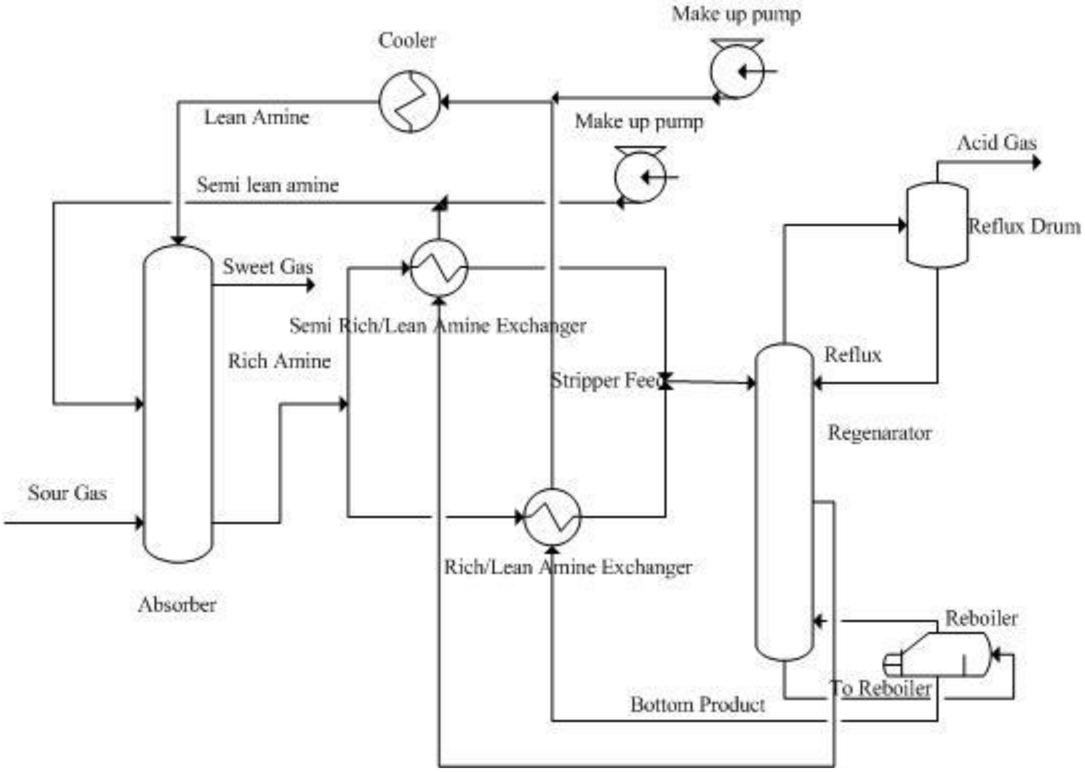

Figure 9 Split Lean and Rich Solvent Flow Process Flow Diagram

3.2.2.5 Mechanical Vapor Recompression

This type of configuration takes care mainly with the thermodynamic efficiency of the stripper. In conventional strippers, the thermodynamics efficiency of such system is considered low because around similar amount of energy is consumed and released in the reboiler and condenser, respectively (Andika et al. 2017). The condenser in the regenerator is replaced by a compressor. The compressed stream which has high temperature is heat exchanged with the bottom product of the column. Then this stream is cooled and flashed so the acid gases leaves the flash drum from the top while some portion of the bottom product is recycled back to the column as depicted in figure 10.

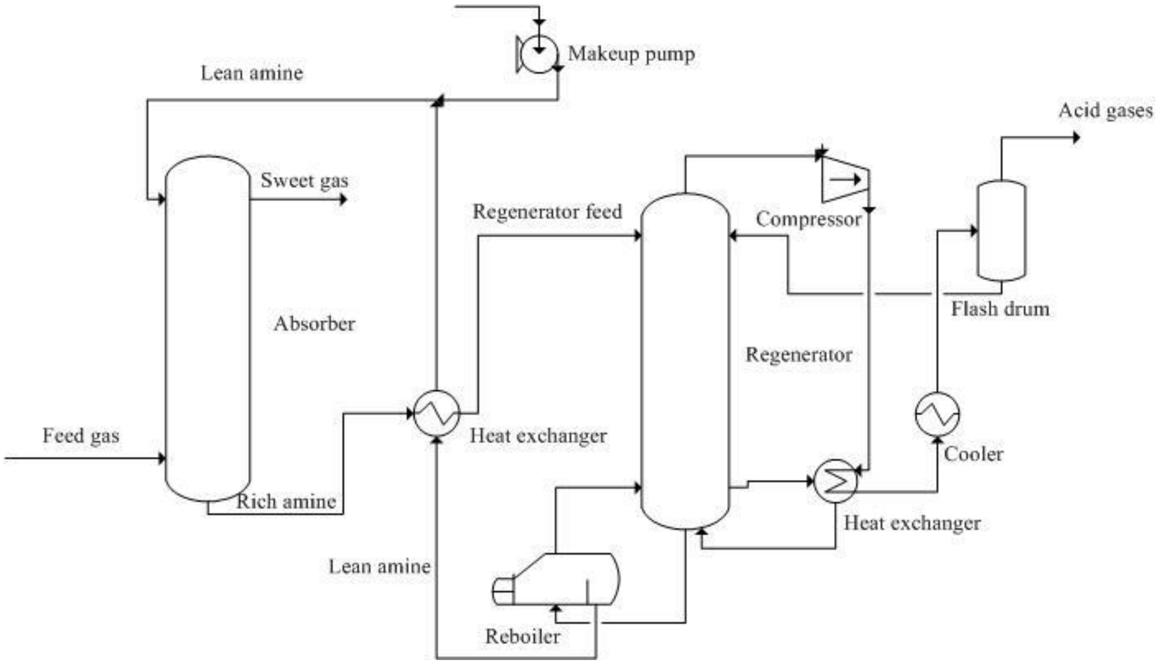

Figure 10 Process Flow Diagram for Mechanical Vapor Recompression Scheme

3.3 Results and Discussion

3.3.1 Model Validation

Model validation is an important step for ensuring the reliability and accuracy of the simulation model. The results for SGP have been validated against the available plant data and results show that there is a good agreement between the developed model and the plant data as shown in table 3. When using the DGA solvent, the acid gas concentration in the sweet gas should be less than 4 ppm and 100 ppm of H_2S and CO_2 respectively. The results show that the H_2S and CO_2 concentration in the sweet gas is 3 and 63.04 ppm respectively. According to SGP design, the H2S concentration in the acid gas stream leaving at the top of the stripper should be in the range of 17-35 mol%. The simulation model results show the H_2S concentration of 35 mol% in the acid gas. The solvent to feed mole ratio of 1.75 is used in this study which is within the operating range of design data. The inlet temperature of sour gas and lean amine entering the absorber is 38 ^oC and 54 ^oC respectively. Figure 11 shows the temperature profile of the DGA absorber. It can be seen from the figure 11 that most of the absorption is taking place near the bottom of the absorber. The rich stream leaving the bottom of the absorber is at 78.89 ^oC which matches well with this study's findings. Figure 12 shows the regenerator temperature profile for the DGA case. The results show that the condenser temperature is around 70 ^oC, while the temperature at the reboiler is 130 ^oC.

Table 4 Comparison of Simulation Results and Plant Data for the SGP

Parameter	This study	Plant data

H_2S in acid gas stream	35 mol%	17-35%
Rich amine stream temperature	78.89 ^{o}C	< 85 ^{o}C
Lean amine temperature (^{o}C)	54	54
Inlet gas temperature (^{o}C)	38	38
Solvent to gas ratio (mole basis)	1.75	1.66 – 1.94
CO_2 concentration in sweet gas (ppm)	63	<100
H_2S concentration in sweet gas (ppm)	3	4

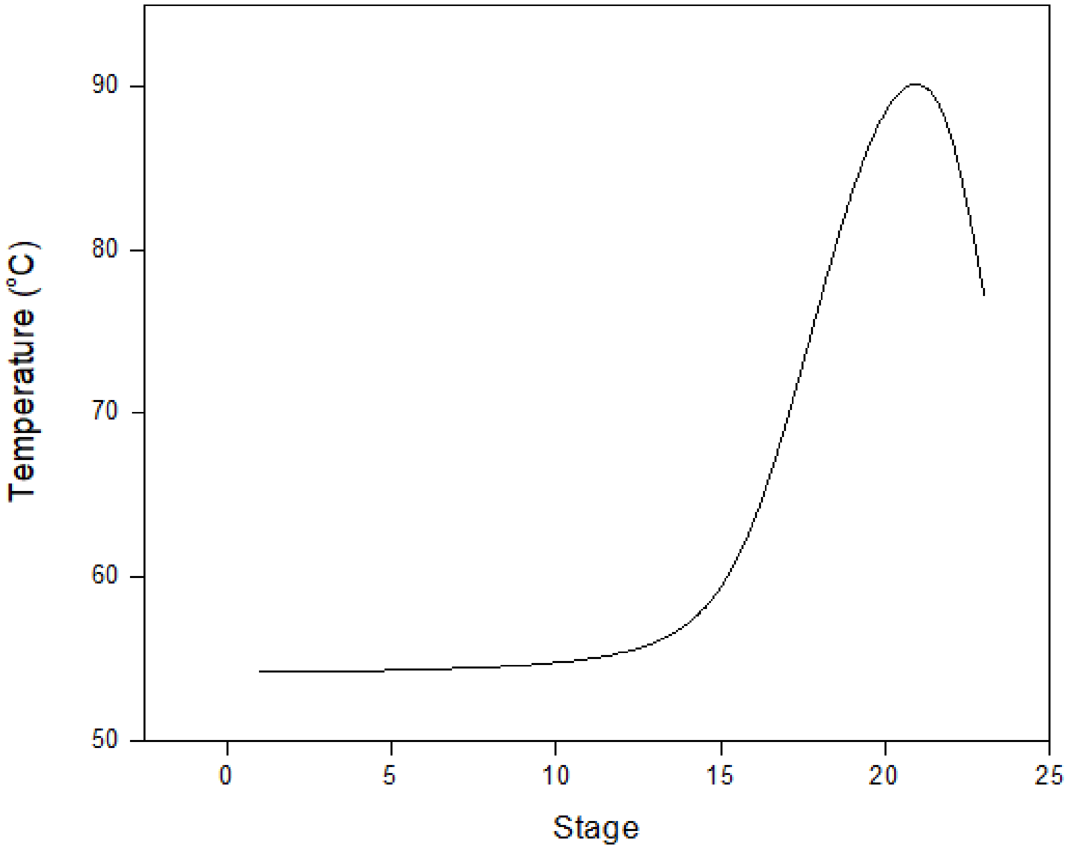

Figure 11 Absorber Temperature Profile for DGA Solvent Employed at SGP

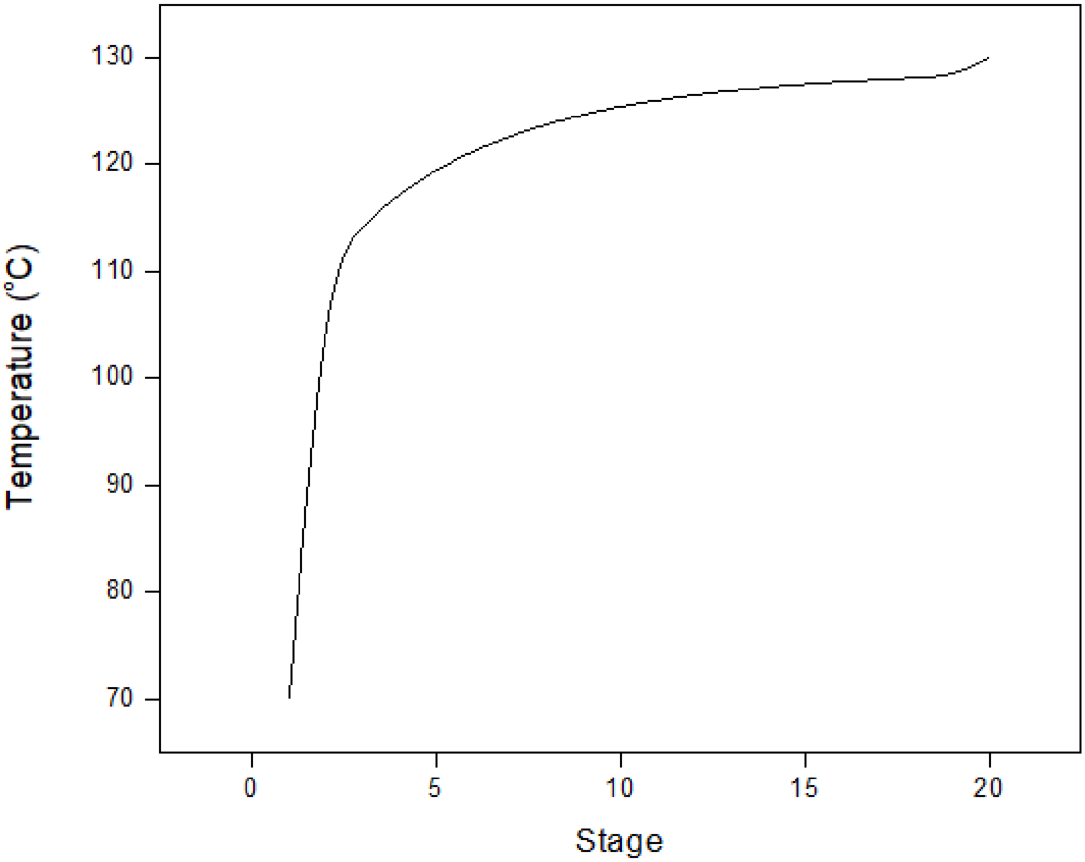

Figure 12 Stripper Temperature Profile for DGA Solvent Employed at SGP

KGP has been simulated using the plant feed as shown in table 2. However, an extensive validation results cannot be reported due to data unavailability. To check the reliability of KGP results, the result trends of MDEA model are compared with the literature as presented in the next section. The results show that MDEA selectively absorbs H2S while CO_2 mainly leaves with the sweet gas stream. The model results show that the H2S concentration in the sweet gas stream is less than 4 ppm as per the design requirements. Figure 13 and 14 show the temperature profile of the absorber and stripper respectively with the MDEA solvent. As can be seen from the comparison of DGA and MDEA temperature profiles, the temperature peak of DGA solvent is higher because of large heat of reaction between the solvent and acid gases compared to that of the MDEA solvent.

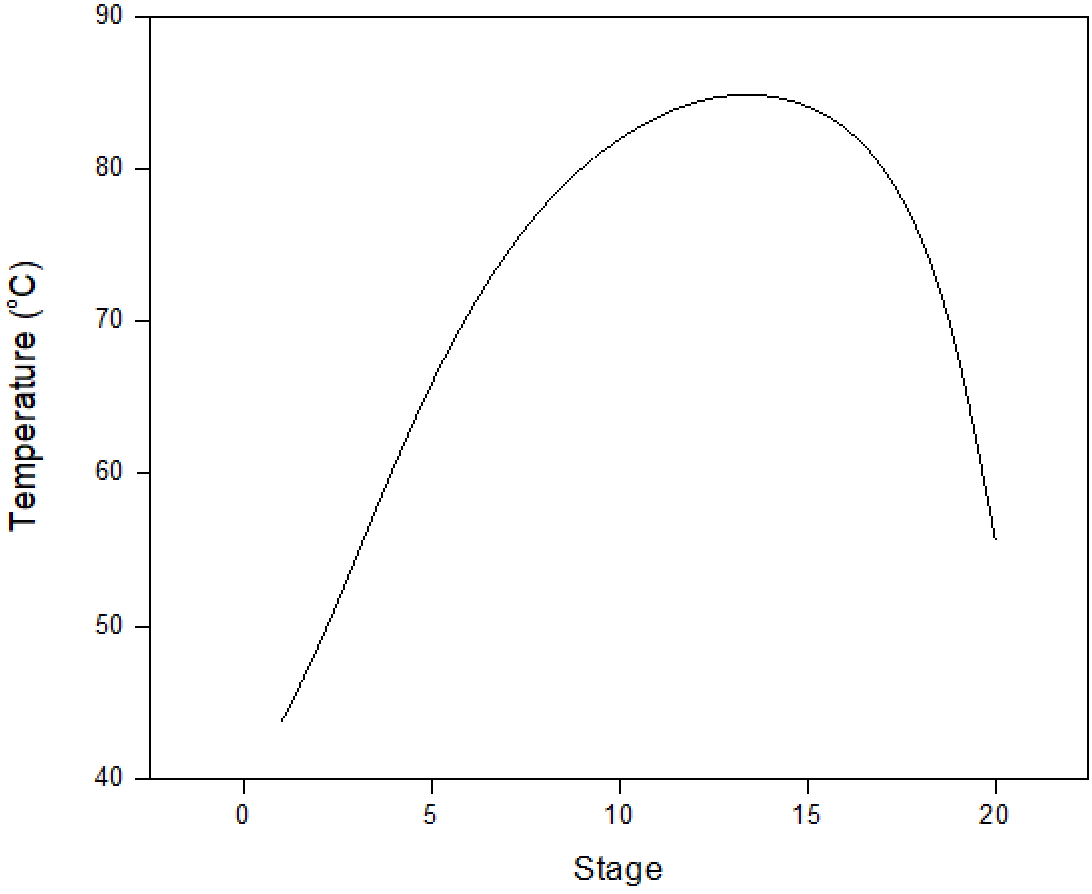

Figure 13 Absorber Temperature for MDEA Solvent Employed at KGP

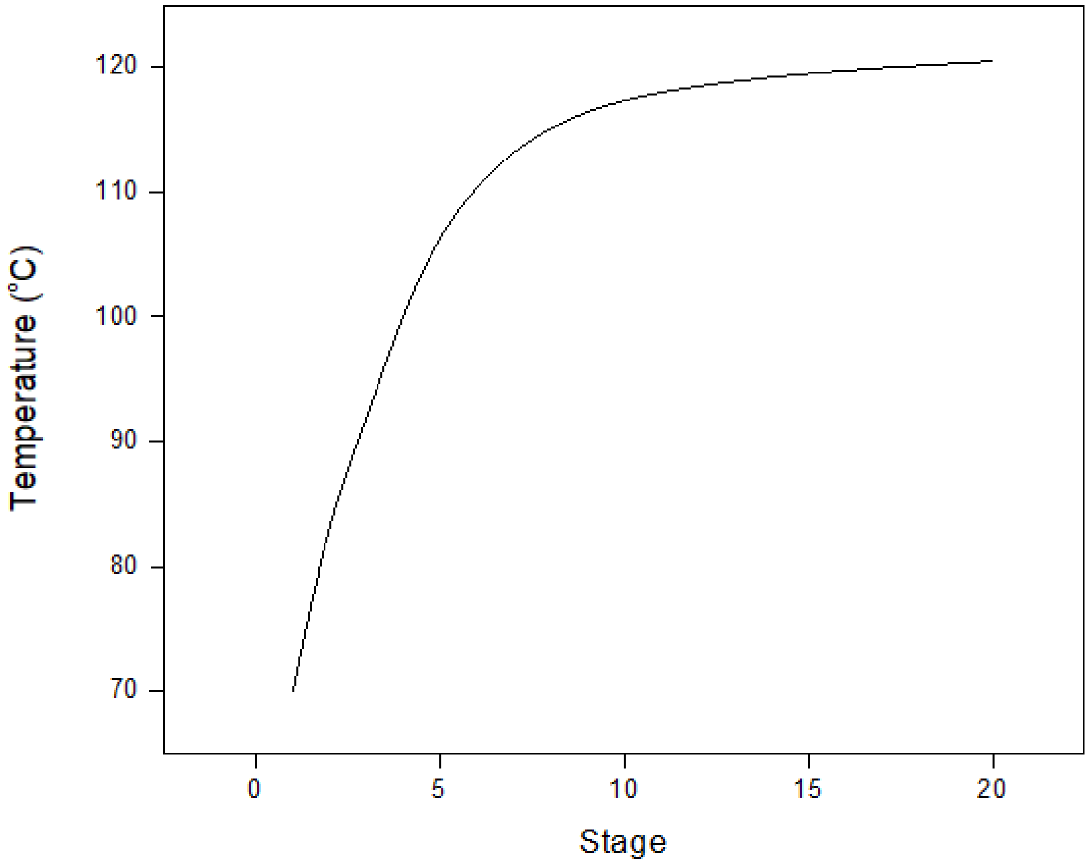

Figure 14 Stripper Temperature Profile for MDEA Solvent Employed at KGP

3.3.2 Energy Performance

It has been reported that approximately 20% of the total natural gas processing cost is attributed to the pre-processing stages, of which AGR is the main section (Lim et al. 2013). Therefore, energy analyses have been done to investigate the power consumption of the acid gas removal plant. The results show that the most energy intensive section in the acid gas cleaning process is the stripper duty. Figure 15 shows the comparison of energy contribution for the major equipment involved in the process. In case of DGA solvent, around 52.1% of the total energy requirement is attributed to the reboiler duty whereas, in case of MDEA solvent, the reboiler duty has an approximate share of 52.3% in the energy consumption. The lean solvent cooler is the second most energy consuming equipment. This also presents an opportunity to investigate feasible heat integration possibilities within the plant. Table 5 shows the breakdown of duty requirements for the two plants. The reboiler duty requirement for the SGP and KGP is calculated as 179 and 185 GJ/h, respectively. However, since both the plants have different feed conditions and compositions, a direct comparison between the performance cannot be made. Therefore, a standard method employed in the literature is to calculate the energy requirement per unit ton of acid gas removed. The results in table 5 show that the energy requirement for the removal of CO_2 and H_2S in case of DGA solvent is higher compared to the MDEA solvent. The lower energy requirement for MDEA solvent is mainly because of its selective absorption capabilities. Another reason for lower energy requirement in case of MDEA solvent is the feed pressure. In case of MDEA solvent, the absorber at KGP is operated at around 52 bar compared to DGA absorber pressure of 12.8 bar at the SGP. Salkuyeh and Mofarahi (Salkuyeh & Mofarahi 2012) reported that significant reboiler duty can be saved when the absorber is operated at a higher pressure. The results also show that the solvent losses in case of MDEA is around 7.5 times higher than that of the DGA solvent which adds to the solvent makeup cost. The results and trends of energy requirement is similar to what have been reported in the literature (Ghanbarabadi & Khoshandam 2015; Banat et al. 2014; Gutierrez et al. 2017; Abdulrahman & Sebastine 2013; Aliabad & Mirzaei 2009; Jassim 2016; Slagle 2013; Sharif Dashti et al. 2015).

Table 5 Comparison of Energy Performance for DGA and MDEA Solvent

Parameter	DGA	MDEA
Reboiler Duty (GJ/h)	179	185
Reboiler duty/ton CO_2 (GJ/ton)	8.199	6.158
Reboiler duty/ton H_2S (GJ/ton)	14.593	8.593
Reboiler duty/ton acid gas (GJ/ton)	5.25	3.59
Reboiler and condenser duty/ton acid gas (GJ/ton)	6.632	3.779
Reboiler and condenser duty/ton feed gas (GJ/ton)	0.674	0.318
Solvent Loss (m^3/h)	0.002	0.0149

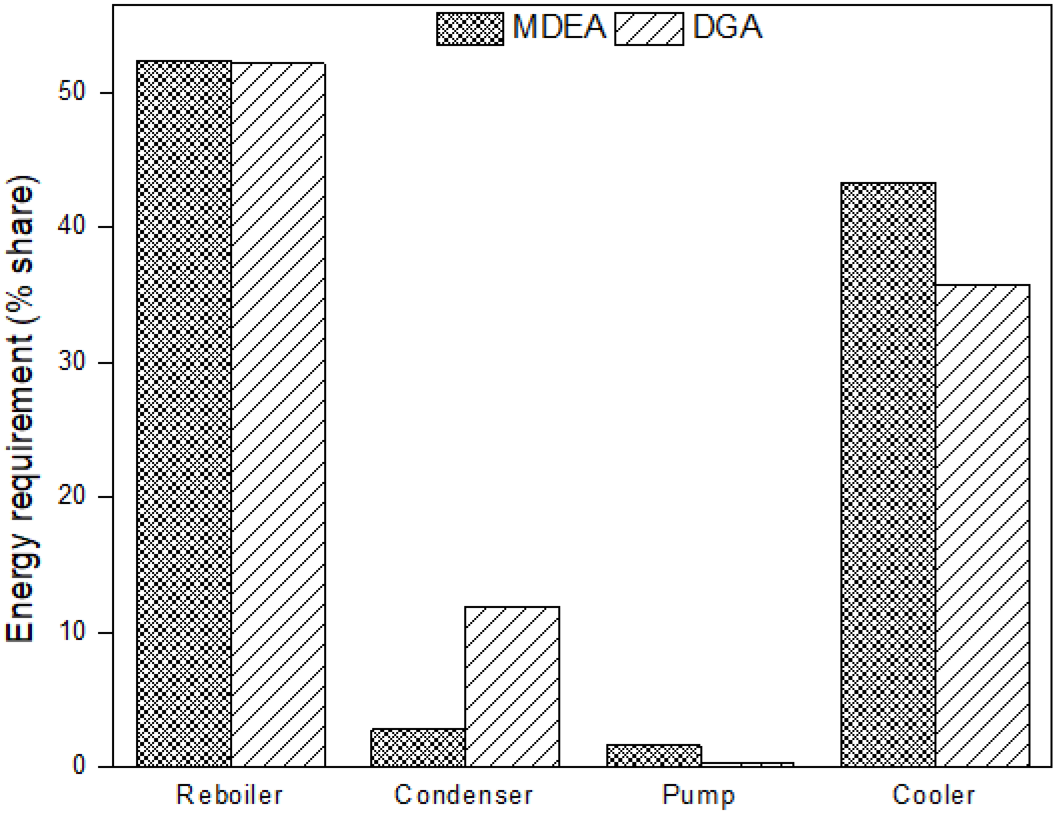

Figure 15 Energy Requirement of Various Units in the AGR Plant for DGA and MDEA Solvent

3.3.3 Design Alternatives Results

Various design configurations were investigated in the present study to look for energy saving opportunities compared to base case scheme. SGP that employed DGA was selected in this comparison. Some of the design specifications are used based on optimization while the other parameters are selected from the recent literature. Table 6 summarizes the results of these alternative designs and the comparison between these configurations and base case in terms of sweetening performance energy consumption. The product purity is maintained for fair comparison between the suggested designs except for the split lean and rich solvent flow scheme. This trend is in good agreement with the literature (Bae et al. 2011). All the design alternatives requires lower DGA circulation rate than conventional scheme except for the mechanical vapor recompression scheme. However, the highest reboiler duty and total energy saving can be obtained using mechanical vapor recompression. The calculated energy saving values are 42% and 21% for reboiler duty and total energy, respectively.

Table 6 Comparison between conventional and design alternatives results

Configuration	Conventional	Split lean solvent flow	Split lean and rich solvent flow	Absorber intercooling	Mechanical vapor recompression
Circulation rate (m^3/h)	650	625	600	636	650
CO_2 in sweet gas (ppm)	63	63	77	63	63
H_2S in sweet gas (ppm)	3	2.5	2	3	4
Reboiler duty/ acid gas removed (GJ/t)	5.294	5.09	4.83	5.148	3.079
Total energy/ acid gas removed (GJ/t)	9.96	9.67	9.72	9.73	7.84

3.3.4 Sensitivity Analysis

Many parameters can be tuned to improve the overall performance of gas sweetening process. Some of the parameter variations are attributed to the feed conditions, while other parameters can be manipulated to achieve the product purities. Therefore, parametric analyses have been performed for some of the important process variables that can affect the performance and cost of the process. Since removing acid gases from the sour gas with minimum energy is the goal of any acid gas removal plant, therefore, all the results presented here are in terms of sweet gas purity and required reboiler duties.

3.3.4.1 Effect of Circulation Rate

Solvent circulation rate is one of the most important variable that can be used directly to achieve the product purity. For both DGA and MDEA absorbers, the acid gas content in the sweet gas can be controlled by an increase in the amine circulation rate. An increase in the solvent will allow a longer contact time between sour gas and solvent leading to an improved solubility. The results show that the rich amine loading decrease as the recirculation rate increases for both the amines as shown in figure 16. The results show that with an increase in the DGA recirculation rate, the acid gas content in the sweet gas decreases sharply up to 640 m3/h and then almost remains at the same level as shown in figure 17. In case of MDEA, as the recirculation rate is increased, there is a continuous sharp drop in the H_2S content of the sweet gas as presented in figure 18. However, the reduction in CO_2 content is relatively small because MDEA preferentially absorbs H2S in the presence of both H_2S and CO_2 gases. The results also indicate that as the recirculation rate increase, there is a linear increase in the reboiler duty because of an increased flow to the stripper as shown in figure 16.

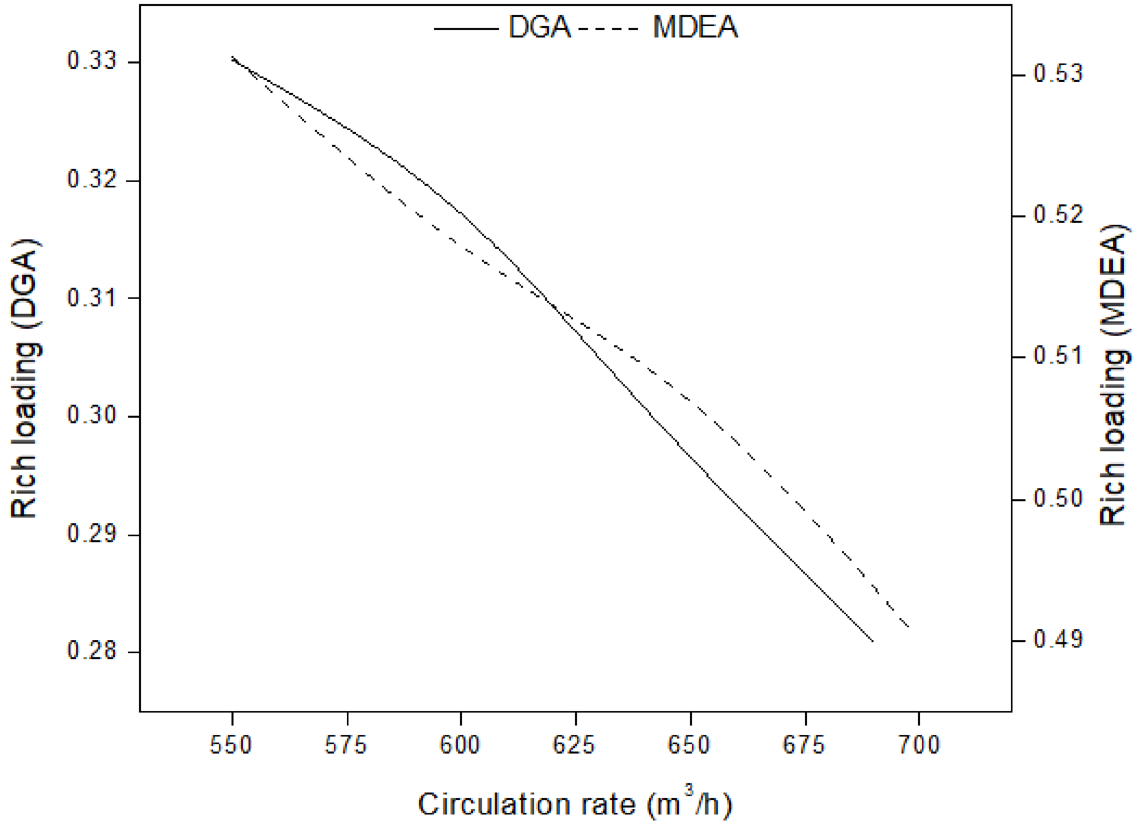

Figure 16 Effect of Circulation Rate on Rich Loading for DGA and MDEA Solvents

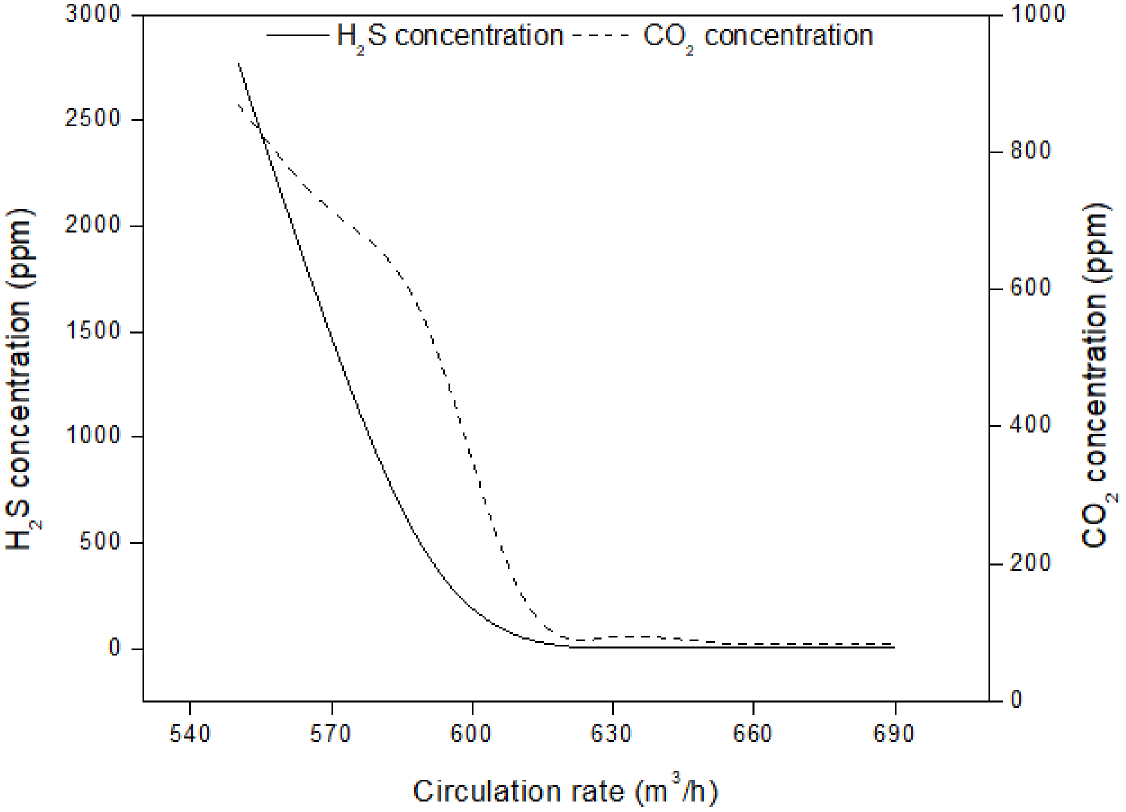

Figure 17 Effect of Circulation Rate on the Product Purity Using DGA Solvent

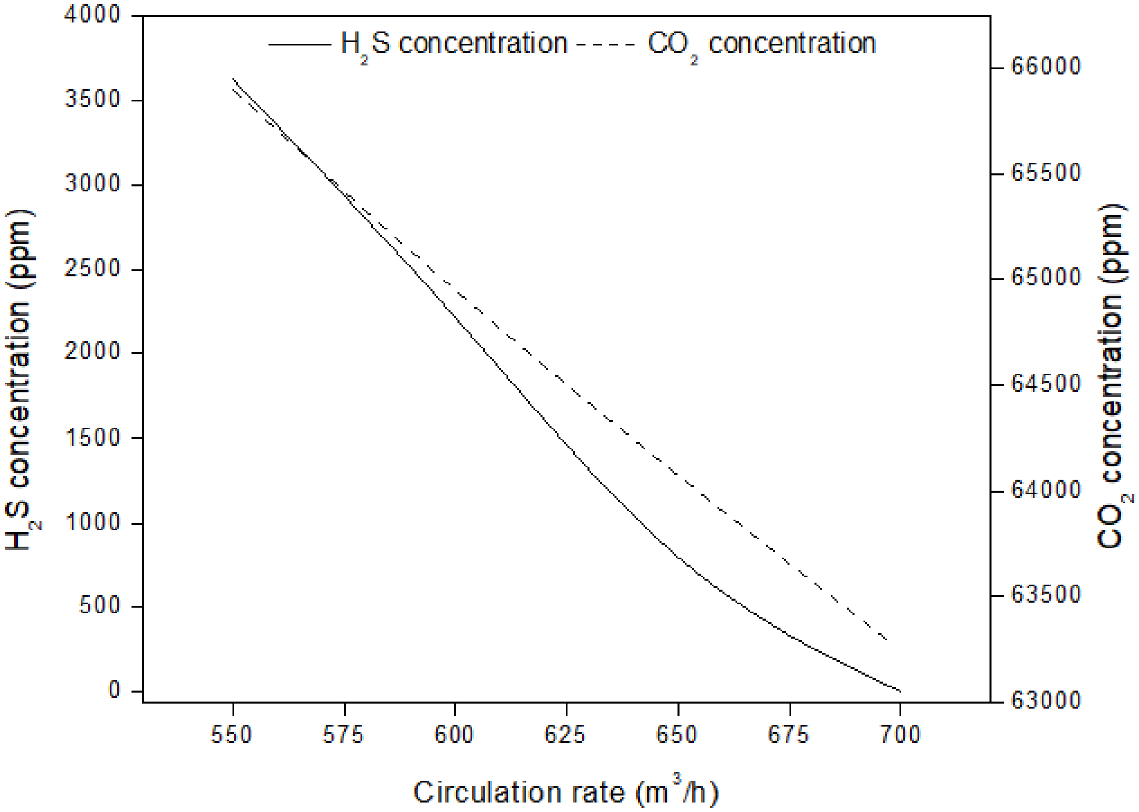

Figure 18 Effect of Solvent Circulation Rate on the Removal of Acid Gases for MDEA Solvent System

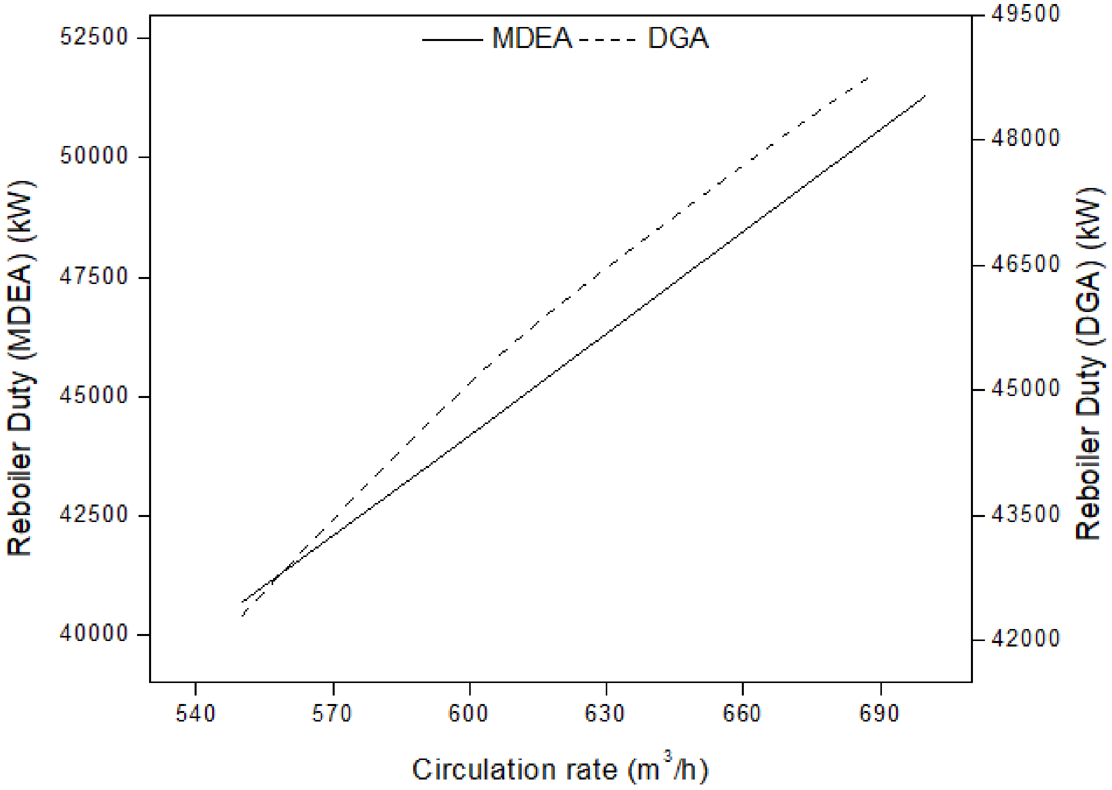

Figure 19 Effect of Solvent Circulation Rate on the Reboiler Duty for DGA and MDEA Solvent

3.3.4.2 Effect of Lean Solvent Temperature

The reaction taking place in the absorber is exothermic, therefore, the absorber temperature and its reactions kinetics can be manipulated by changing the feed gas temperature and amine temperature. Usually, the absorption performance can be enhanced by lowering the operating temperature of the absorber. However, due to large flowrate of the sour feed gas, it is usually not easy to manipulate. Therefore, the only variable to control the temperature profile of the contactor is the lean amine temperature. Parametric analyses have been done to analyze the impact of changing lean solvent temperature on the product purities and process performance. It can be observed that as the lean amine temperature increases, the acid gas content in the sweet gas reduces till it reaches a minimum point. Any further increase in the lean temperature beyond that minimum ppm level decreases the solubility of the acid gases causing an increase of its content in the sweet gas. Figure 20 shows that in case of DGA solvent, the minimum acid gas content can be achieved when the lean amine temperature is about 50 ^{o}C. On the other hand, in case of MDEA solvent, any increase in solvent temperature beyond 40 ^{o}C rises the acid gas content in the sweet gas as shown in figure 21. Also, figure 19 shows the effect of lean temperature on the reboiler duty for both solvent cases. The reboiler duty for DGA solvent increases with an increase in the lean temperature up to 50 ^{o}C and then decrease as the lean amine temperature increases. In case of MDEA solvent, the reboiler duty reduces consistently as the lean amine temperature is increased as shown in figure 22. It is important to mention that the lean solvent temperature also directly effects the cooler duty. As shown in figure 15, lean solvent cooler is the second largest energy consuming equipment in the acid gas removal plant. Therefore, a detailed analysis must be done in order to optimize the lean amine temperature.

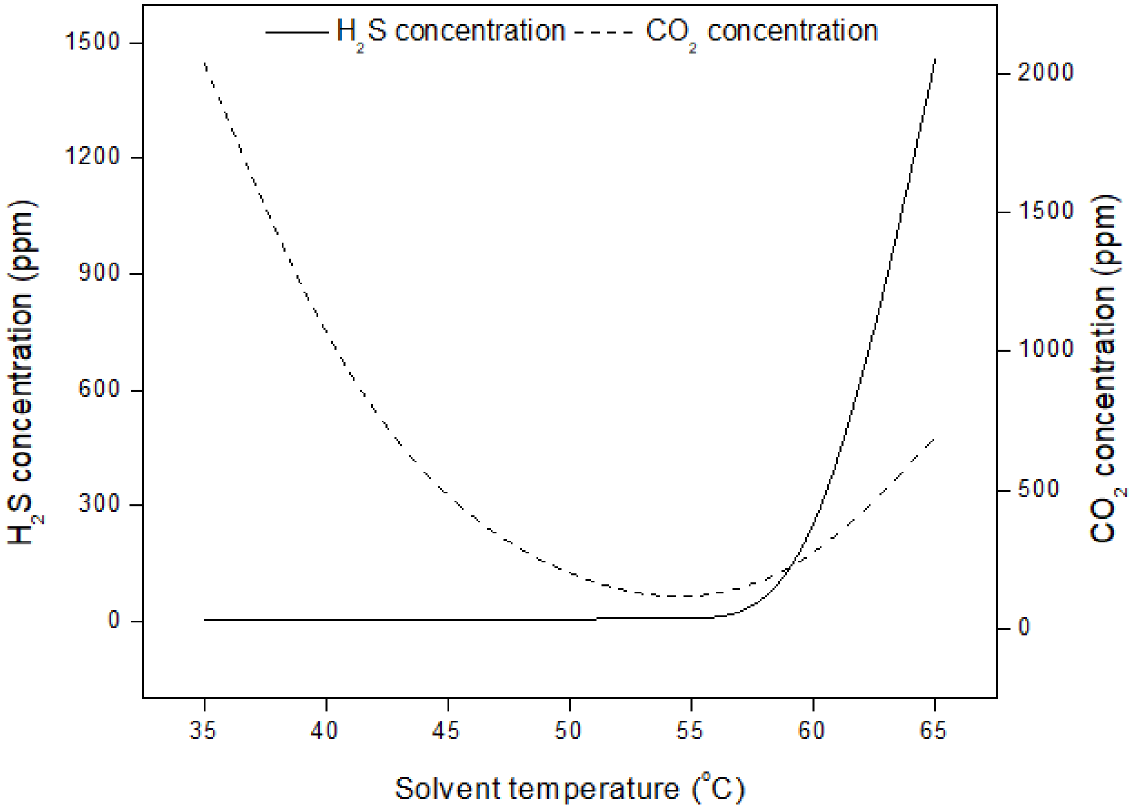

Figure 20 Effect of Lean Solvent Temperature on the Product Purity Using DGA Solvent

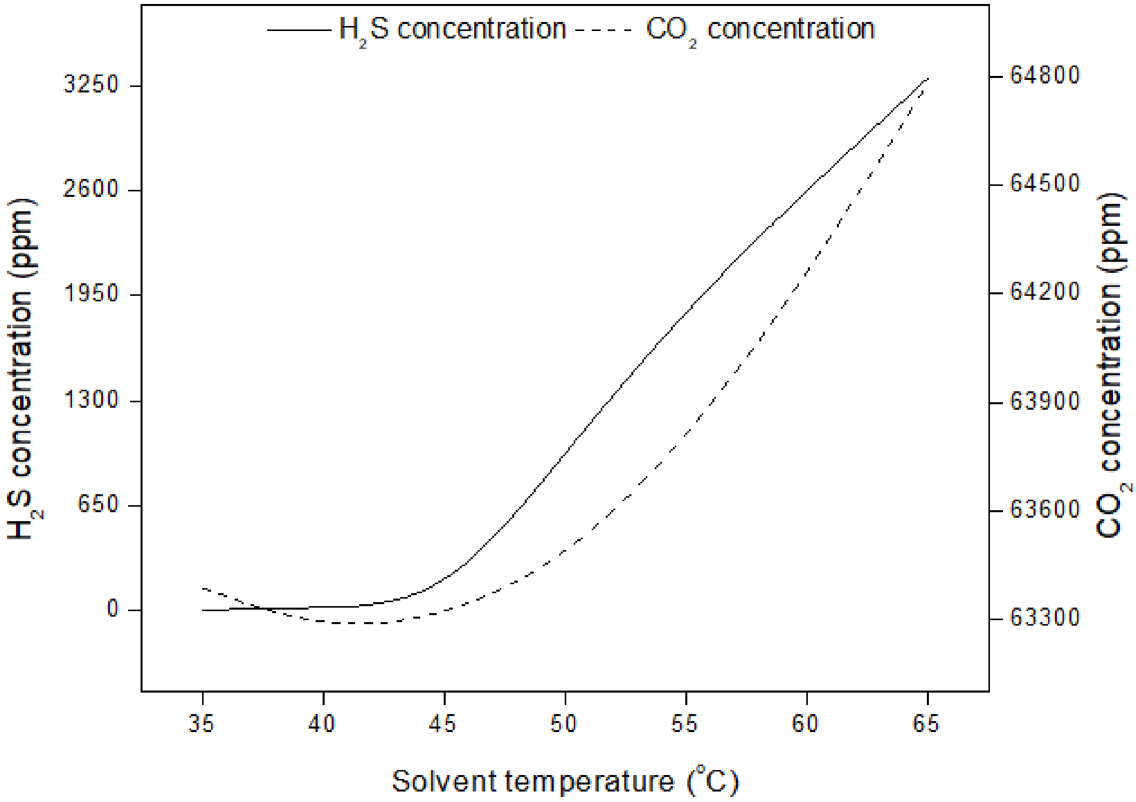

Figure 21 Effect of Lean Solvent Temperature on the Product Purity Using MDEA Solvent

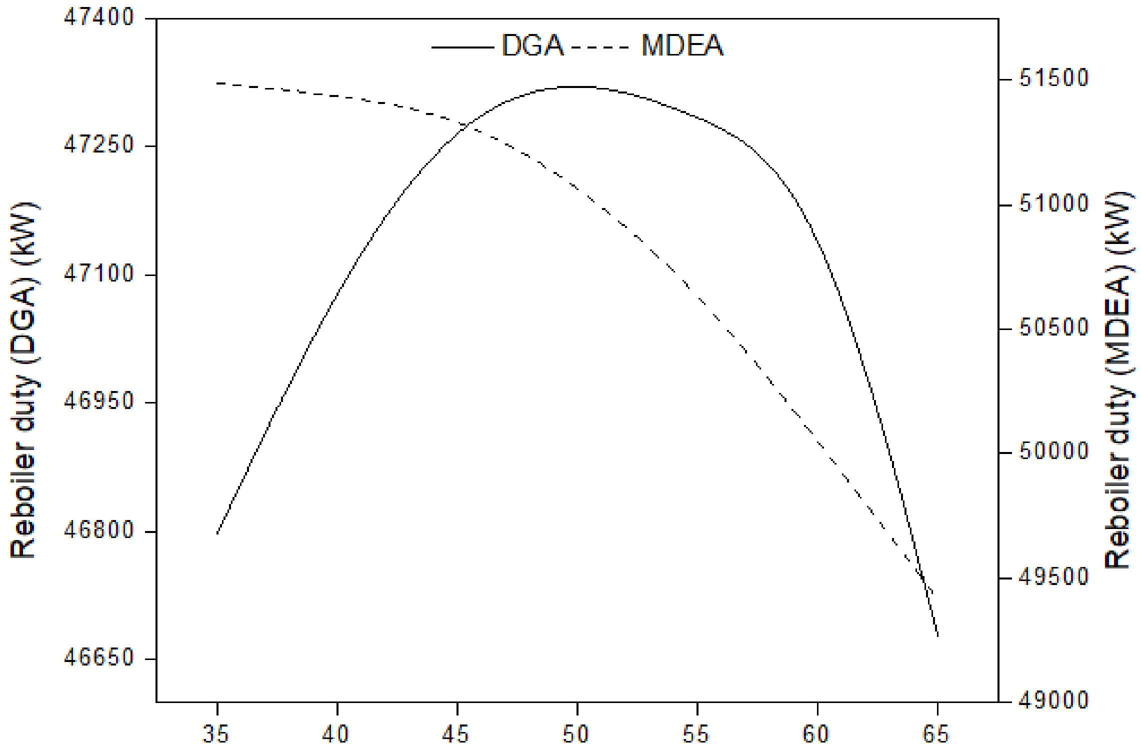

Figure 22 Effect of Lean Solvent Temperature on the Reboiler Duty for DGA and MDEA Solvent

3.3.4.3 Effect of Solvent Strength

Solvent weight concentration is another variable that can be manipulated to investigate its impact on the process performance. The base case considered 50 % DGA and 45 % MDEA for the two plants respectively. The results of the sensitivity show that in case of DGA solvent, the acid gas concentration in the sweet gas decrease sharply as the solvent strength is increased from 45 wt% to 50 wt% as shown in figure 23. Any further increase in the solvent weight percent reduces the acid gas content linearly. The results show that the sweet gas specifications can be achieved at 50 wt% DGA solvent. On the other hand, in case of MDEA solvent, a minimum H_2S concentration was achieved in the sweet gas with 40 wt% MDEA solvent as shown in figure 24. Any increase in the MDEA solvent concentration beyond 40 wt% increases the H_2S in the sweet gas. The results obtained in this work are similar to what is reported in the literature (Jassim 2016). The effect of solvent concentration was also investigated on the reboiler duty and the results (figure 25) show that increasing the lean solvent concentration leads to a decrease in the reboiler duty which is consistent with industrial findings (Fouad et al. 2011; Satyadileep et al. 2015)

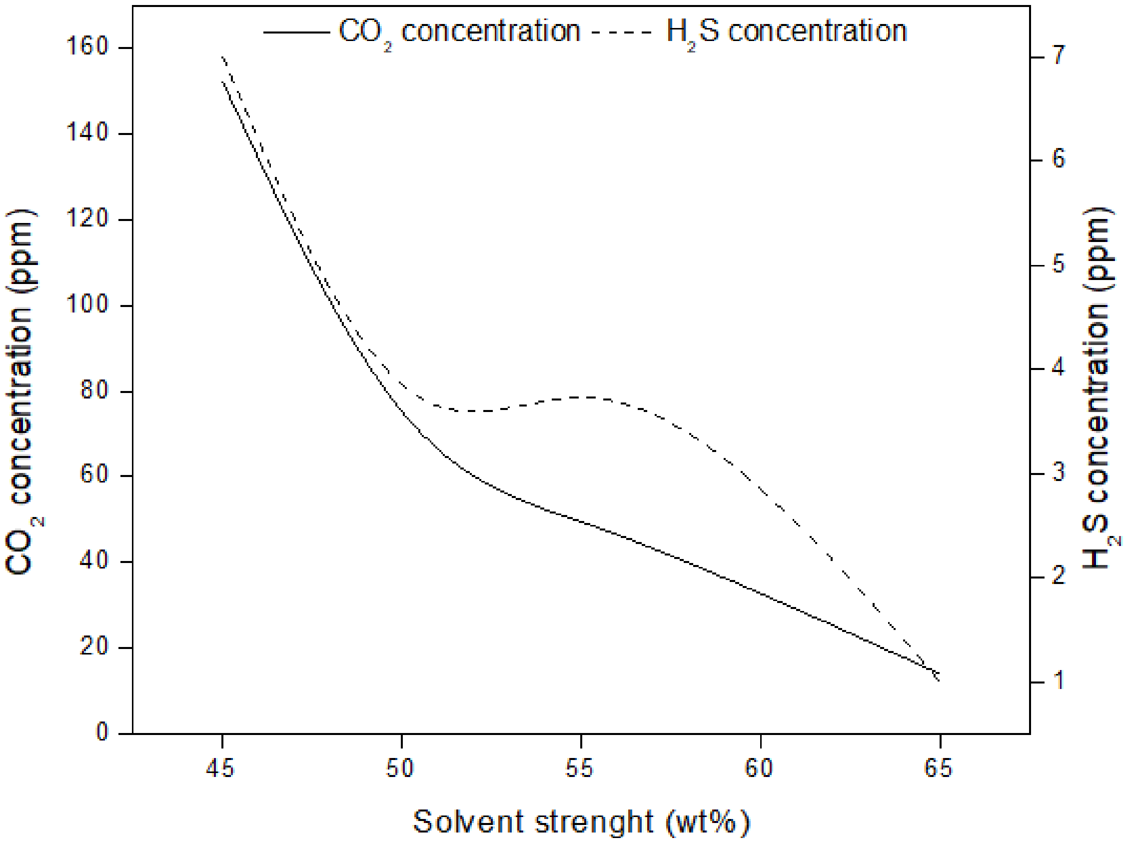

Figure 23 Effect of Solvent Concentration on the Product Purity Using DGA Solvent

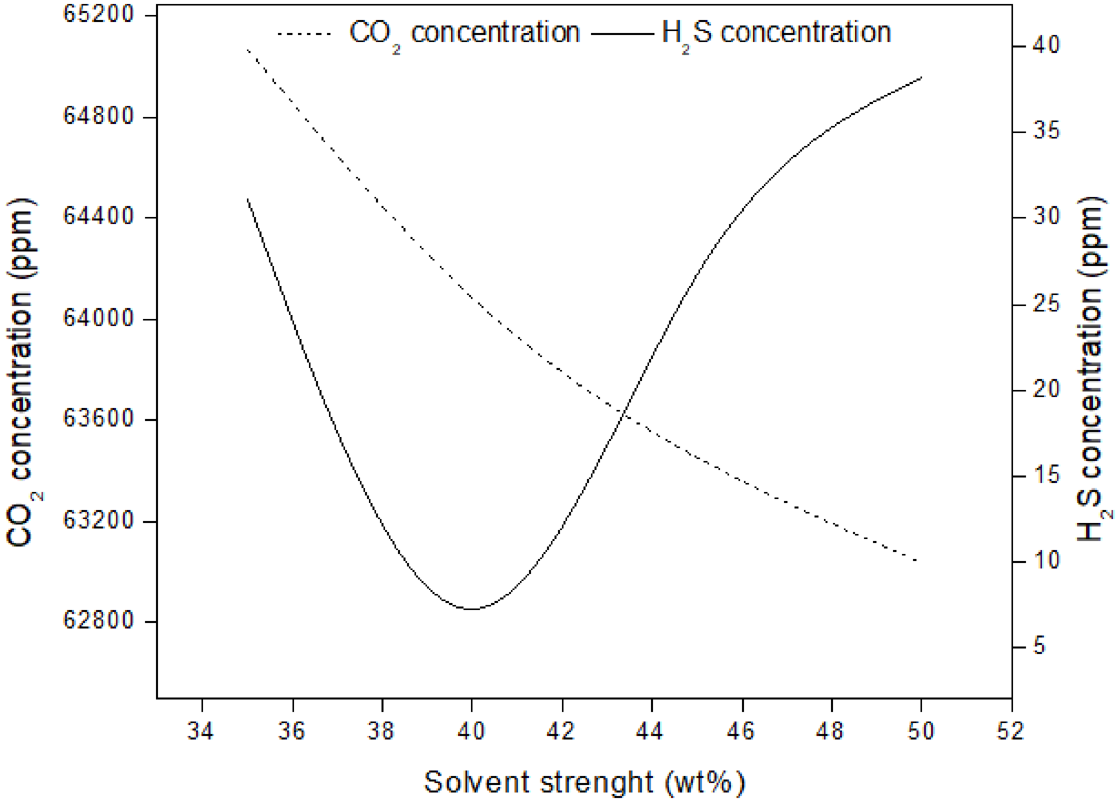

Figure 24 Effect of Solvent Concentration on the Product Purity Using MDEA Solvent

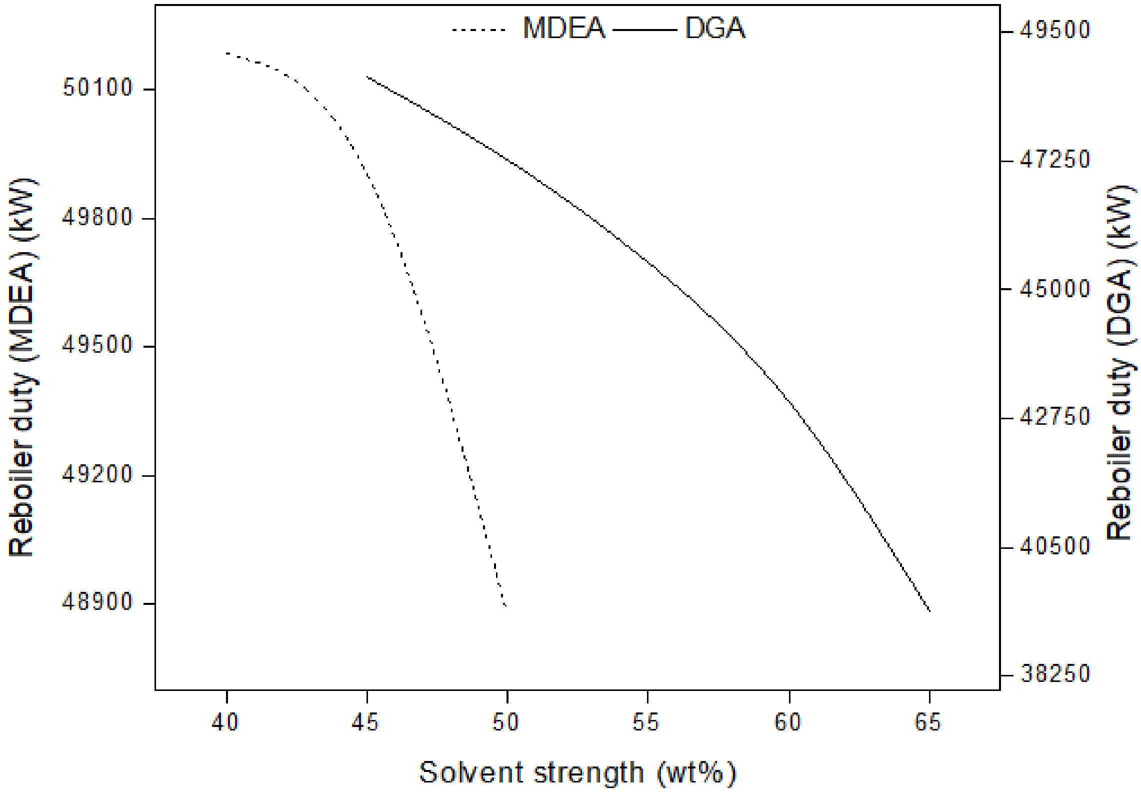

Figure 25 Effect of Solvent Concentration on the Reboiler Duty for DGA and MDEA Solvent

3.3.4.4 Stripper Inlet Temperature

The rich amine temperature feed to the regenerator can be varied to influence the reboiler duty. Therefore, analyses have been done in order to study the influence of changing stripper inlet temperature on the process performance. Literature has widely reported stripper inlet temperature of around 99 ^{o}C with slight variations (Addington & Ness 2010). The base case in the study assumed the stripper inlet temperature of 105 ^{o}C in case of DGA solvent, while 80 ^{o}C in case of MDEA solvent. The results show that with an increase in the stripper inlet temperature, the reboiler duty decreases for both the DGA and MDEA solvent as shown in figure 26. However, the maximum stripper inlet temperature is also limited from an operational point of view, since increasing the rich amine temperature beyond its saturation temperature will produce vapors in the stream which can cause corrosion problems in the downstream equipment.

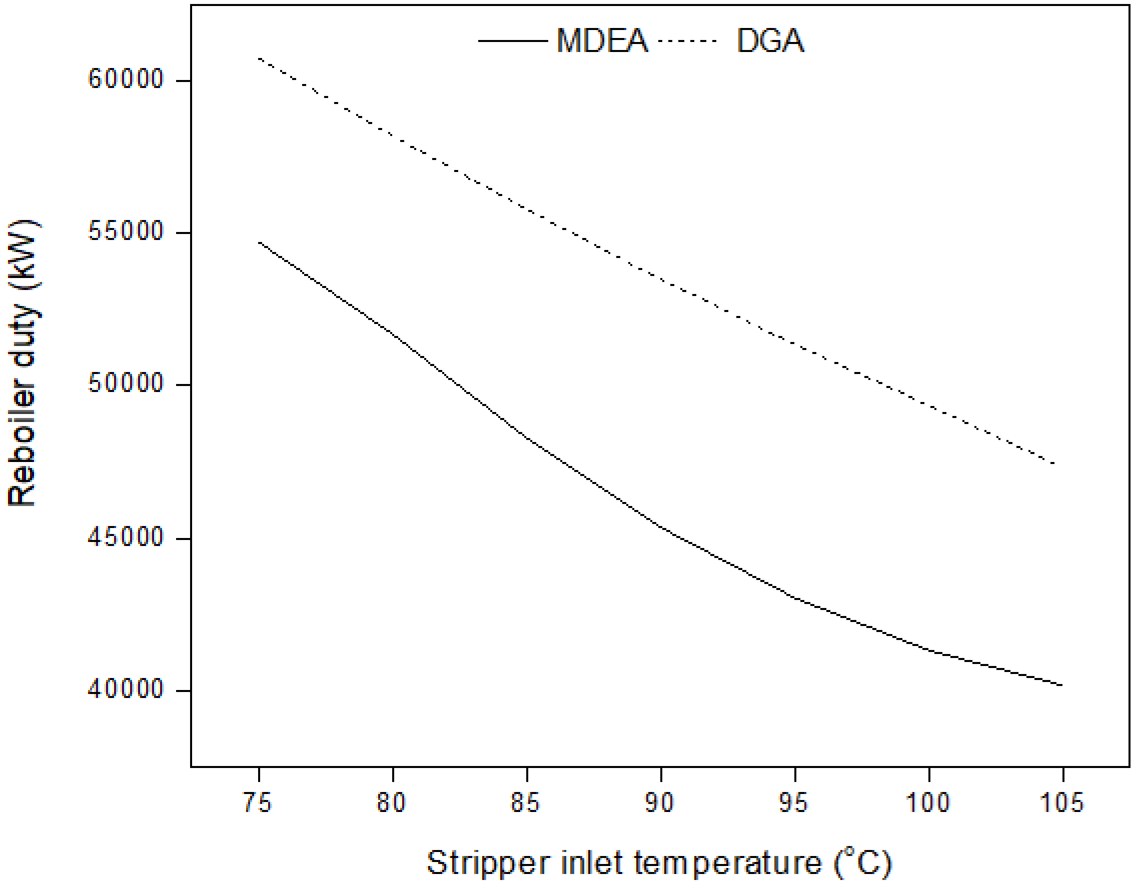

Figure 26 Effect of Stripper Inlet Temperature on the Reboiler Duty for DGA and MDEA Solvent

3.3.4.5 Effect of Stripper Pressure

Stripper operating pressure is an important variable which is directly connected with the reboiler temperature. Usually, it is recommended to have a stripper pressure of approximately 2.1 – 2.2 bar, since higher stripper pressure will cause an increase in the reboiler temperature. Parametric analysis has been done in order to analyze the influence of stripper pressure on the reboiler temperature and hence its duty. The results show that as the stripper pressure is increased, the reboiler temperature increases linearly. In case of DGA solvent, any increase in the stripper pressure reduces the reboiler duty as shown in figure 27. This result is in line with the findings of Addington and Ness, (Addington & Ness 2010). In case of MDEA solvent, the results show that the reboiler duty increases with an increase in the stripper pressure as shown in figure 28. It is evident from the results that when the sour feed has both CO_2 and H_2S, increasing stripper pressure in case of DGA solvent will reduce the process duty while any increase in stripper pressure will hurt the performance in case of MDEA solvent. Another important aspect related to stripper pressure is the reboiler temperature. Increasing the stripper pressure increases the reboiler temperature, but amines start to thermally degrade above 132 ^oC. Therefore, care should be exercised while increasing the reboiler pressure which impacts the reboiler temperature.

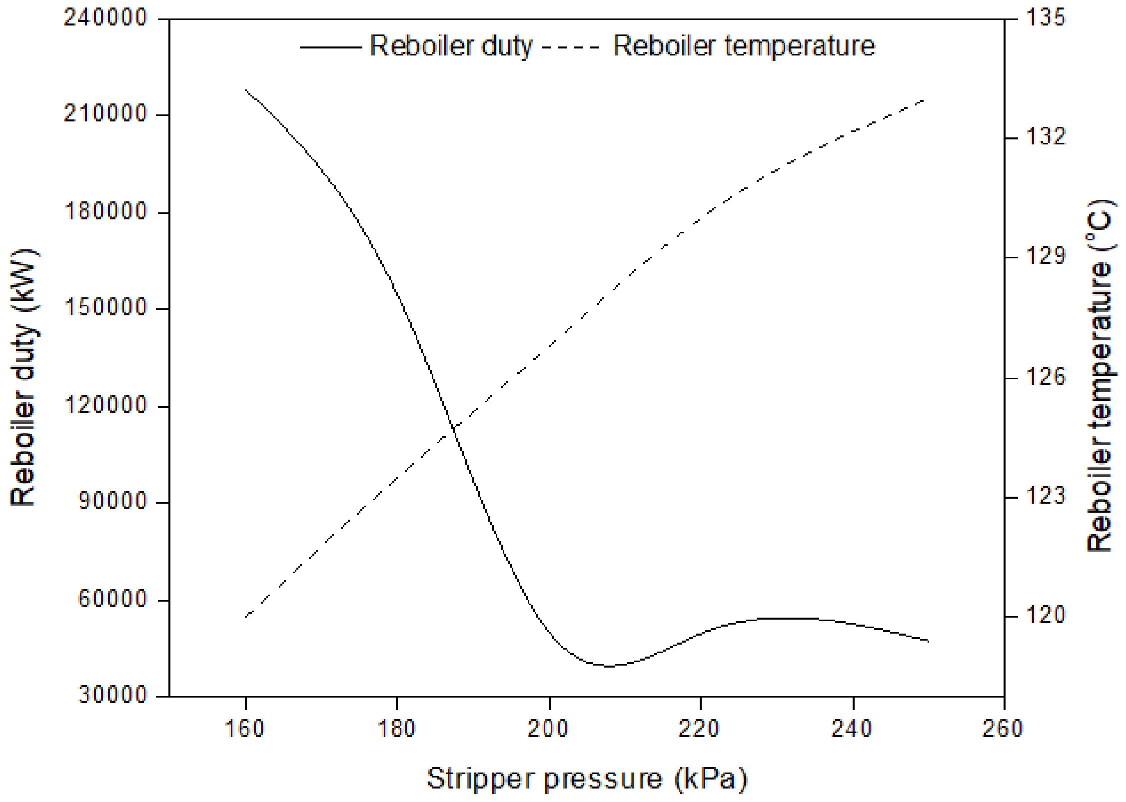

Figure 27 Effect of Stripper Pressure on the Reboiler Temperature and Duty for DGA Solvent

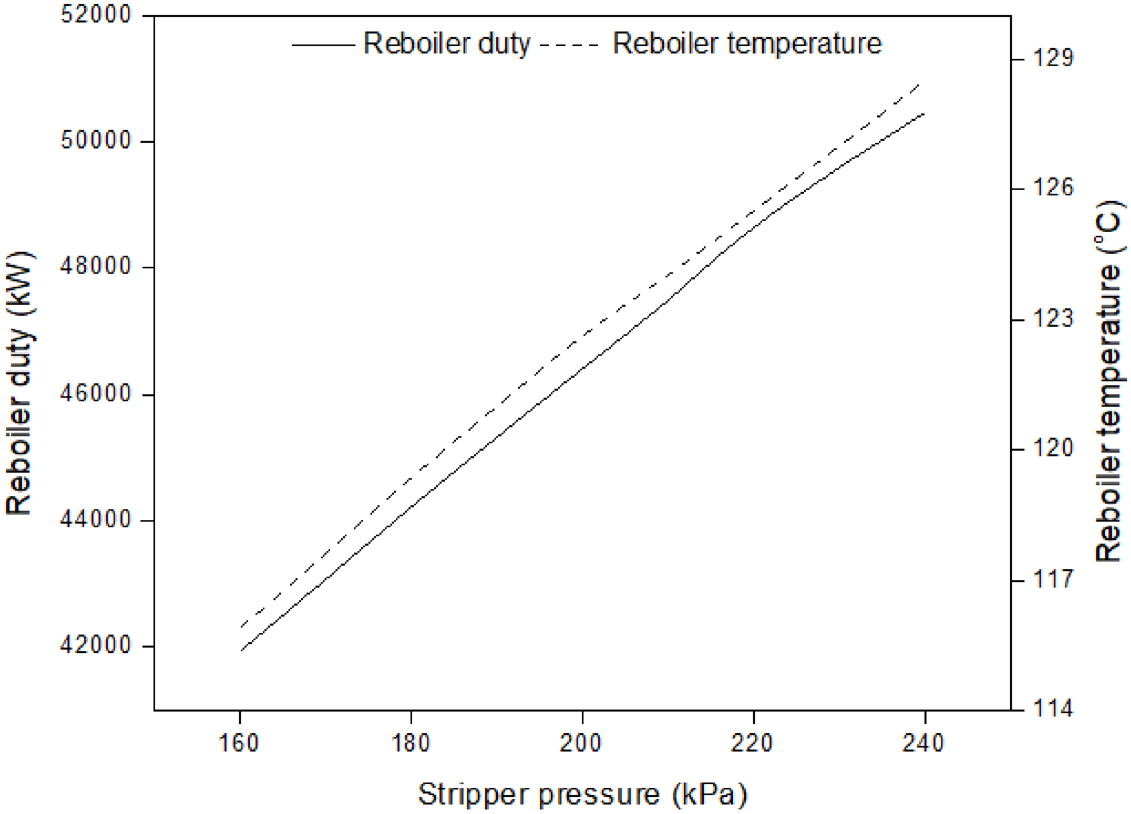

Figure 28 Effect of Stripper Pressure on the Reboiler Temperature and Duty for MDEA Solvent

CHAPTER 4

NATURAL GAS DEHYDRATION

4.1 Introduction

The removal of water is a common process employed for both the natural gas processing and CO_2 conditioning known as dehydration. The life of a pipeline for natural gas or CO_2 transmission depends on its corrosion rate which is directly associated with the amount of water present in the process stream. Therefore, it is critical to reduce the water content in the feed before its transportation in order to avoid pipeline corrosion and prevent hydrate formation. The water dew point temperature must be lower than the minimum pipeline temperature to avoid any condensation.

Although many studies were conducted on the natural gas dehydration and recently few studies also investigated the water removal for the CCS application. However, most of these studies are done without any validation with the actual plant data. Also, no substantial research is performed to compare the analogy between the natural gas dehydration and water removal from the wet CO_2 stream. The purpose of this part is to investigate the glycol dehydration process for the application of natural gas processing and CCS. The study aims to model and simulate a glycol dehydration process. The developed model results are then validated against the actual plant to ascertain the accuracy of the simulation results. The results of this study are also compared with that of the literature to compare the performance trends. Alternative design using stripping gas has also been investigated to compare the process performance with that of the base case model. Finally, a parametric analysis has been performed to study the impact of key process parameters on the performance of the dehydration process.

4.2 Process Description

The dehydration process itself consists of two parts, gas dehydration and solvent regeneration. In gas dehydration, water is removed from the wet gas stream using the solvent in the absorber. The rich solvent is then sent to the solvent regeneration section to strip water out of the solvent. Figure 29 shows the process flow diagram of the TEG dehydration process. The wet CO_2 stream saturated with water is first sent to a flash drum where free water is separated. Removing free water in the flash drum reduces the absorption load resulting in smaller column and lower solvent circulation requirement. The wet gas enters the bottom of the absorber where it counter currently contact the lean TEG solvent fed at the top of the column. Water is absorbed from the gas stream into the TEG. The dried gas stream leaves at the top of the absorber, while the rich glycol stream leaving at the bottom of the absorber is passed through an expansion valve to reduce the pressure. Rich TEG is pre-heated through the reflux condenser in the top of the regeneration column. Rich TEG is further heated in the lean/rich heat exchanger before it is fed to the regeneration column. In the regeneration column, water and CO_2 are stripped off from the solvent and leaves at the top of the regenerator column. Whereas, the lean TEG leaving at the bottom of the regenerator is cooled in the lean/rich exchanger with the incoming rich TEG. The lean TEG is further cooled using a cooler before being pumped back to the top of the absorber. Glycol make-up is added to the recycled lean TEG stream to ensure the required glycol recirculation rate. The following section will investigate two cases for the dehydration application, namely natural gas processing and carbon capture and storage.

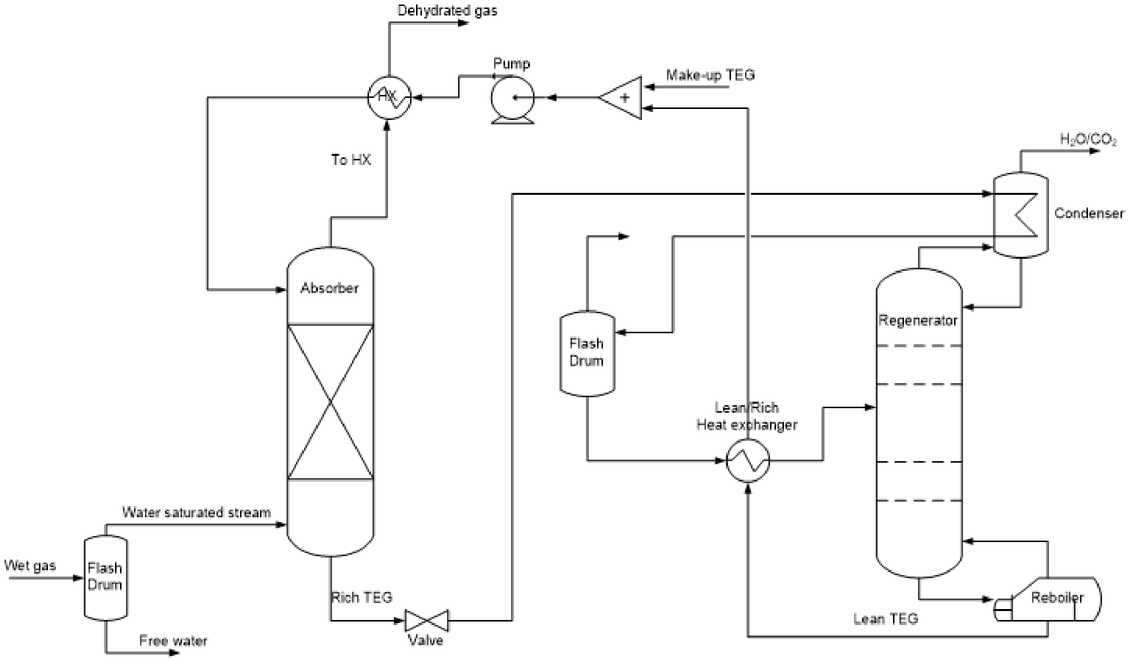

Figure 29 Conventional Gas Dehydration Process

4.3 Design Basis

The simulation has been performed using one of the commercial simulator, Aspen HYSYS which is a widely used software in chemical process industry. The Glycol package available in the Aspen HYSYS is used as the thermodynamic fluid package. The Glycol property package contains the TST (Twu-Sim-Tassone) equation of state to determine the phase behavior accurately for the TEG-water mixture. The Glycol property package can predict the vapor-liquid equilibrium (VLE) over the range of temperatures, pressures, and component concentration encountered in a typical TEG-water dehydration system. The Glycol package can reasonably predict the properties between 15 °C to 50 °C and between 10 atm to 100 atm for the gas dehydrator, and between 202 °C to 206 °C and 1.2 atm for the glycol regenerator. Glycol package uses three adjustable parameters to correlate VLE mixture data. The Glycol property package uses the Cavett model for enthalpy and entropy calculations. In this study, two cases have been investigated for the application of dehydration process, namely, natural gas and captured CO_2 dehydration.

Since the goal of dehydration process is to remove water, typically, the acceptable water content in the natural gas transmission pipelines is 7 lb/MMSCF which corresponds to approximately water content of 147 ppm. Therefore, the design basis for natural gas dehydration has been set to achieve the product purity with no more than 100 ppm of water content in the dried gas. In case of CO_2 stream dehydration, the allowable water content in the dried CO_2 depends on the mode of transmission. Currently, the captured

CO_2 is mainly transported in supercritical phase through pipelines for the application of enhanced oil recovery (EOR). Various countries have defined the regional policy regarding the allowable water content in the CO_2 stream. For example, US and Canada limits the water content in the dehydrated CO_2 up to a maximum limit of 0.064 %, which corresponds to 640 ppmv. The EU Dynamics project set a limit of less than 200 ppmv of water impurity in the CO_2 stream. However, the allowed limit for the CO_2 transport in the liquid phase is more stringent. In this study, the water content limit of 200 ppmv in the dehydrated CO_2 stream is assumed considering supercritical phase pipeline transport.

4.3.1 Natural Gas Dehydration

In order to illustrate the application of dehydration in natural gas processing, the feed specification is taken from one of the gas processing plant located in the Middle East. The feed to the dehydration unit is coming from the acid gas removal unit as shown in table 7. The high pressure wet feed enters the absorber from the bottom side at approximately 40 ˚C and contacts the TEG counter currently in the column. The glycol absorbs the moisture content in the feed gas and leaves at the bottom of the absorber, while the dried gas leaves at the top of the column. The inlet temperature of the lean glycol to the absorber is usually kept 3 – 10 ˚C higher than that of the feed gas temperature. The base case considered the lean TEG inlet temperature of 45 ^{O}C. The pressure of the rich TEG stream is throttled down from 51 bar to 5 bar through a valve. The temperature of the rich glycol is then increased to 44 ˚C by exchanging heat with the reflux condenser in the top of regenerator. It is then fed to a flash drum to separate the hydrocarbons absorbed in the rich TEG stream. If flash drum is not placed before the regenerator, then the hydrocarbon emissions will leave at the top of regenerator with the water content leading to another environmental issue. Therefore, the flashed hydrocarbons can be taken from the flash drum for utilization as a fuel or process gas in the plant. The temperature of the rich TEG stream is further increased to 145 ˚C by passing it through the lean-rich heat exchanger before feeding it into the regenerator. The water is stripped in the regeneration column and leaves at the top, while the lean TEG (98.9 wt. %) leaves at the bottom of the column and is pumped back to the absorber. The simulation input specifications for the natural gas dehydration are shown in table 8. The number of absorber and regenerator stages are specified as one and three respectively.

Table 7 Feed Gas Specifications and Compositions for Natural Gas Dehydration

Feed conditions	
Temperature (^{O}C)	40
Pressure (bar)	51.7
Flow Rate (m^3/h)	1236
Components (mole fraction)	
Nitrogen	0.0504
CO_2	0.0633
H_2S	0.0000
C_1	0.8603
C_2	0.0169
C_3	0.0034
iC_4	0.0006
nC_4	0.0013
iC_5	0.0004
nC_5	0.0004
nC_6	0.0002
nC_8	0.0002

Table 8 Design Specifications for the Natural Gas and Carbon Capture cases

Parameter	Natural gas		Carbon capture	
	Conventional	Stripping gas	Conventional	Stripping gas

Feed gas temperature (^{o}C)	40	40	35	35
Lean TEG temperature (^{o}C)	45	45	45	45
Absorber Pressure (bar)	51.3	51.3	59.8	59.8
Absorber stages	3	3	3	3
Circulation rate (kgmole/h)	26	10	15	3.8
Stripping gas rate (kgmole/h)	-	0.1178	-	1.019
Stripper pressure (bar)	1.05	1.05	1.05	1.05
Condenser temperature (^{o}C)	100	100	100	100
Reboiler temperature (^{o}C)	200	200	187.8	187.8
Stripper stages	1	3	1	3

4.3.2 CO_2 Dehydration

The second application of dehydration investigated in this study is for the CCS. The feed to the dehydration plant is coming from a CO_2 capture plant as shown in table 9. The feed mainly consists of CO_2 rich stream with some amount of water. The water content present in the CO_2 stream should be removed before it is transported through pipelines or liquefied to prevent hydrate formation and inhibit corrosion. The base case design for CO_2 dehydration is same as explained earlier for the natural gas dehydration application. The input simulation specifications for this case are shown in table 8. The absorber has three equilibrium stages and operates at a pressure of 59.9 bar. The wet CO_2 stream enters the absorber from the bottom at 35 ^{O}C which contacts the lean TEG entering from the top of the column at 45 ^{O}C. The dried CO_2 stream leaving at the top of the absorber is set to have water content of no more than 200 ppmv. The rich TEG is sent to the regenerator where lean solvent is produced having concentration of 98.4 wt. %.

Table 9 Feed Gas Specification and Composition for CO_2 Capture Dehydration Case

Feed condition	
Temperature (^{O}C)	35

Pressure (bar)	59.9
Flow rate (m^3/h)	522.7
Components (mole fraction)	
CO_2	0.989922
H_2O	0.005008
C_1	0.004189
C_2	0.000508
C_3	0.000149
Benzene	0.000118
Toluene	4.05E-05

4.3.3 Stripping Gas Configuration

The base case designs (natural gas and CO_2 dehydration) can be modified in order to improve the performance of the dehydration process. The common method for enhancing the process efficiency is by adding stripping gas to the reboiler or the bottom of the regeneration column. The main effect of adding stripping gas is that the partial pressure of water in the gas phase is reduced so that more water will evaporate and increase the wt. % concentration of lean TEG (Øi & Fazlagic 2014; Neagu & Cursaru 2017). The stripping gas can come from an external source (usually N_2) or CO_2 from the dehydrated gas can be used. The vent of the flash drum located after the absorber column can also be used for the purpose of stripping in the regenerator. In this study, the flash gas available after the absorber is used as a stripping source in the regenerator as shown in figure 30.

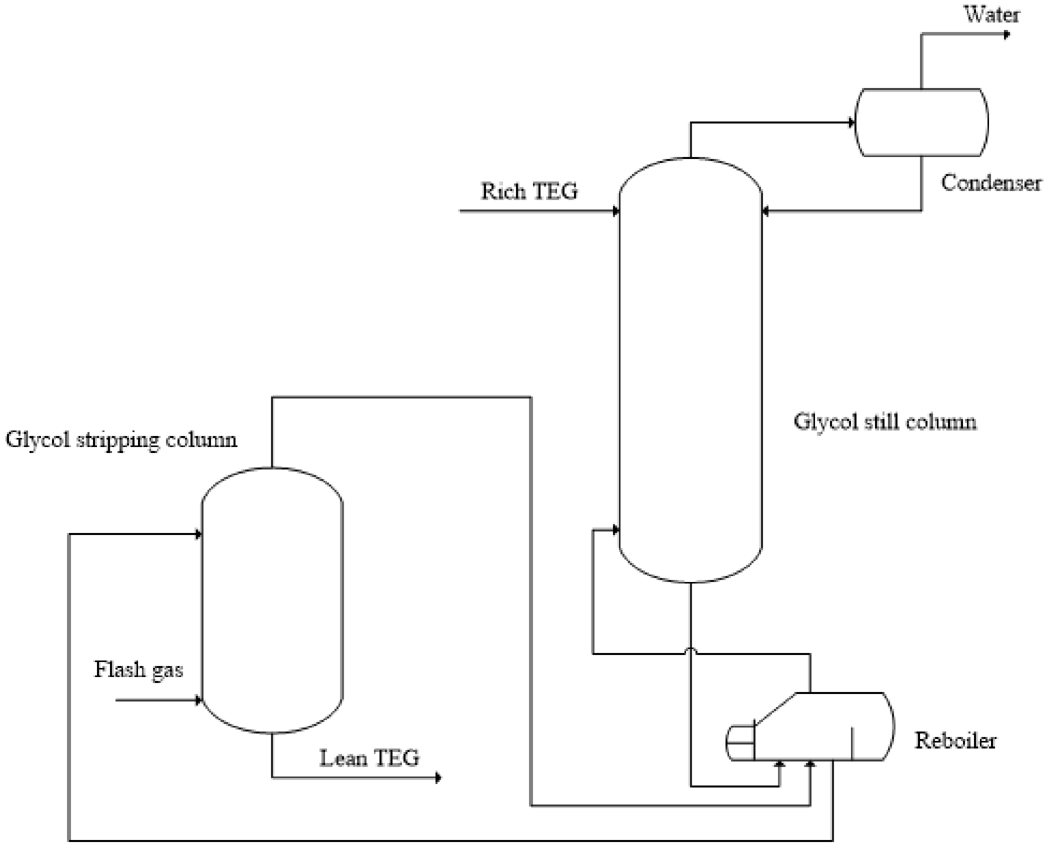

Figure 30 Regeneration System in the Stripping Gas Configuration

Both the cases, natural gas and CO_2 dehydration have been simulated to investigate any process improvement with stripping gas configuration. Table 8 shows the input specification for the stripping gas design along with the conventional design. The number of absorber stages for all the cases is set to three. The feed in both the cases, natural gas and CO_2 stream, is available at high pressure of 51.7 and 59.9 bar respectively. It has been assumed in this study that all the vapor flow from the flash drum is directed to the regenerator reboiler. The lean TEG circulation rate has been calculated based on the desired dehydrated product purity which is set as 100 and 200 ppmv for natural gas and CO_2 transport respectively. At higher temperatures especially in the regenerator, TEG can react with water and decompose to Monoethylene glycol (MEG). Various sources have reported the decomposition temperature of TEG to be in the range of 180 – 207 ^{O}C. In this study, the reboiler temperature for all the cases have been set not to exceed 200 ^{O}C.

4.3.4 Model Validation

Before any performance analyses and design comparisons are made, the foremost step in this work is to validate the simulation model with the plant data. Model validation is an important step for accurately predicting the process behavior over wide range of operating conditions. The model validation has been done for the CO_2 dehydration unit with stripping gas injection and the results show that there is a fair agreement between the model and plant operational data. Table 10 shows the comparison of simulation results with the actual data for some of the key parameters. The results show that the absorber and regenerator temperatures are in good agreement with the plant design data. Similarly, the simulation model predicted a reasonable rich and lean TEG molar concentration compared to that of the plant data. The dehydrated gas purity shows some variation from the plant data, however, that can be explained with the minor variations of reboiler and condenser duties. Overall, the simulation model predicted reasonably accurate results in comparison to the plant data.

Table 10 Dehydration Model Validation with Plant Data

	Unit	Plant Data	Simulation
Dehydrated gas purity	ppmv	132	101
Water removed	%	97.55	98.90
Absorber top temperature	^{O}C	45.5	44.6
Absorber bottom temperature	^{O}C	42.2	42.8
Condenser temperature	^{O}C	100.0	100.0
Reboiler temperature	^{O}C	187.7	187.8
Reboiler duty	GJ/h	0.74	0.83
Condenser duty	GJ/h	0.026	0.039
Lean TEG	mol. fraction	0.977	0.970
Rich TEG	mol. fraction	0.581	0.597

4.4 Results and Discussion

4.4.1 Natural Gas Dehydration

Operating conditions for the dehydration units are governed by the degree of dehydration required. However, there is a limit on the level of dehydration that can be achieved with the standard operational procedures. The base case results show that the target water content of 100 ppmv is achieved in the dehydrated gas with a TEG circulation molar rate of 26 kmol/h. Lean TEG leaving at the bottom of the regeneration column has the concentration of 98.9 wt. %. However, in the case of stripping gas addition to the regenerator, the target water content in the dried gas can be achieved with the TEG circulation rate of 10 kmol/h with all the other variables being held constant. The lean TEG concentration increased to 99.3 wt. % at the bottom of the regeneration column which can be further increased with an increase in the stripping gas flowrate to the reboiler. The results show that the glycol requirement reduced by 61.5 % in the case of stripping gas design maintaining dehydrated stream purity same as that of the conventional design. Figure 31 shows the process performance comparison for the two designs of natural gas dehydration. The results show that the highest energy consuming equipment in the flowsheet is the regenerator reboiler. The calculated reboiler energy requirement for the conventional and stripping gas design came out to be 0.83 and 0.41 GJ/h respectively. Similarly, the total energy requirement of the dehydration process for the two designs is calculated as 1.51 and 0.64 GJ/h respectively. The calculated energy requirement of the plant (conventional and stripping gas) corresponds to 21.8 and 9.2 GJ per ton of water removed respectively. The results show that the stripping gas design requires approximately 57.7 % less energy compared to the conventional dehydration design. Another positive impact of stripping gas configuration on the overall performance is the reduction in the TEG losses as shown in figure 31.

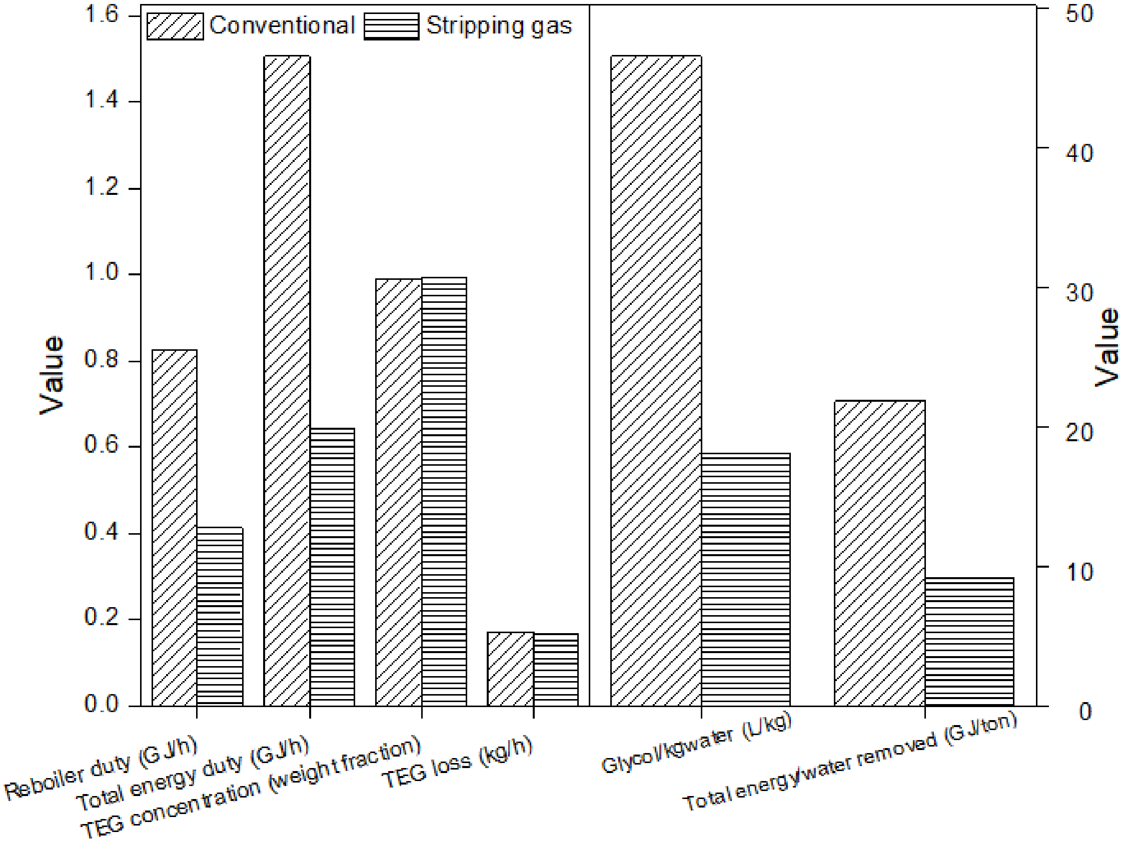

Figure 31 Comparison between Conventional and Stripping Gas Configuration Results for Natural Gas Case

4.4.2 CO_2 Dehydration

The performance of CO_2 dehydration process has been analyzed in the similar fashion to that of the natural gas dehydration. In case of CO_2 dehydration, the target dried stream purity is set to have no more than 200 ppmv of water content. The conventional design results show that the target water content in the dehydrated CO_2 was achieved with a glycol molar flowrate of 15 kmol/h. Lean TEG entering the top of the contactor column has a concentration of 98.43 wt. %. The results show that approximately 20.5 liters of glycol is required per kg of water removed from the wet CO_2 stream. In the case of stripping gas design, the required purity was achieved with a TEG flowrate of 3.8 kmol/h. The lean TEG concentration leaving at the bottom of the regenerator increased to 99.75 wt %. The results show that there is a significant decrease in the glycol requirement for the stripping gas configuration compared to the conventional design for the same level of dehydrated stream purity. Around 9.43 liters of glycol is required per kg of water removed in case of stripping gas design which is 72.6 % less compared to the conventional design. Figure 32 shows the comparison between the conventional and stripping gas design for CO_2 dehydration. The reboiler duty is the largest energy consumption in case of CO_2 dehydration which is consistent with natural gas dehydration. In case of CO_2 dehydration, the calculated reboiler duty for the conventional and stripping gas design is 0.88 and 0.49 GJ/h respectively. Similarly, the total energy requirement of the dehydration process for the two designs is calculated as 1.52 and 0.57 GJ/h respectively. The energy requirement per unit ton of water removed for the two designs (conventional and stripping gas) is 8.9 and 3.3 GJ respectively. The results show that the stripping gas configuration requires approximately 63 % less energy compared to that of the conventional CO_2 dehydration design.

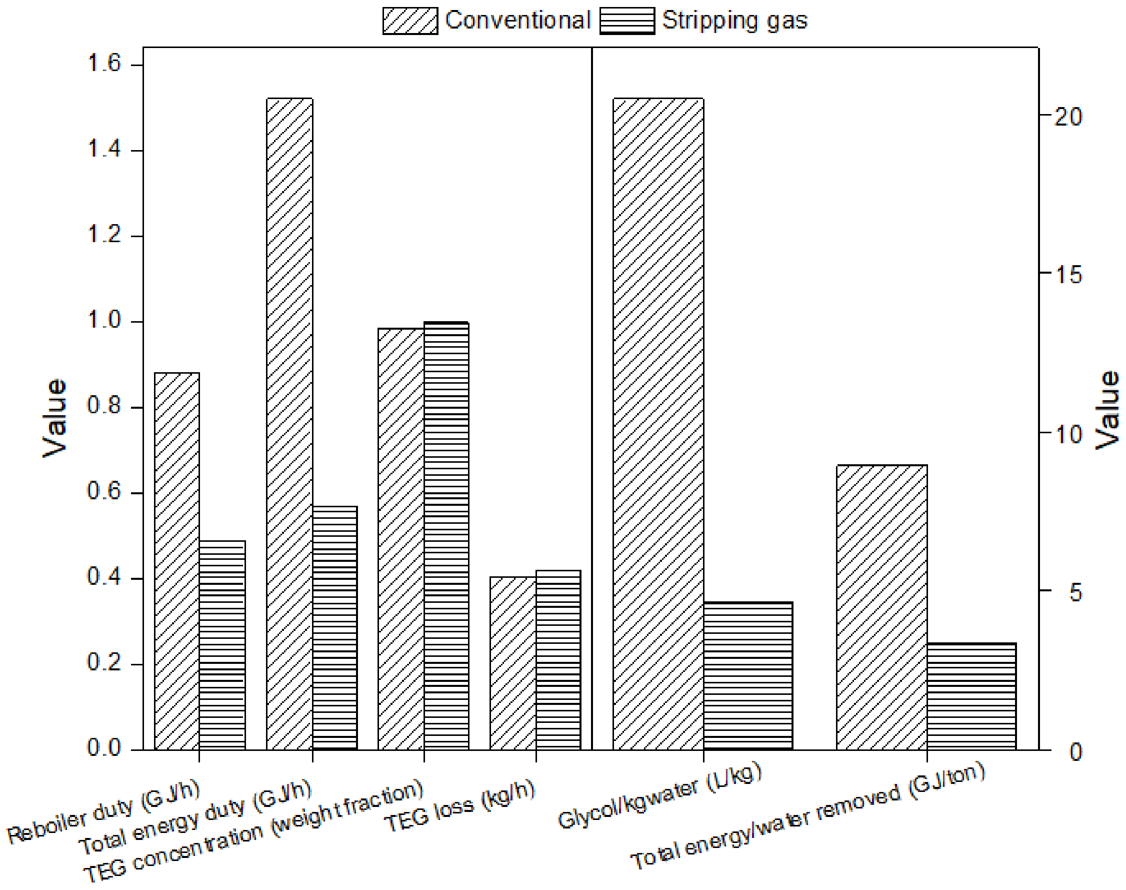

Figure 32 Comparison between Conventional and Stripping Gas Configuration Results for Natural Gas Case

4.4.3 Comparison between natural gas and CO_2 dehydration process

Dehydration process has been employed for removing the water from natural gas for many years. However, the application of TEG solvent for dehydration of wet CO_2 streams is relatively new and no significant literature has been reported on it. In case of CO_2 dehydration, TEG solvent has high affinity for CO_2 gas compared to the methane gas which is the main component of the natural gas stream. Therefore, TEG not only removes the water content from the CO_2 stream but also carries some of the CO_2 which is released at the top of the regenerator. Another potential issue in case of CO_2 dehydration is the loss of TEG when the feed CO_2 stream is at a high pressure. GPSA handbook (Russell et al. 2004) reported that the glycol losses will be significant for the CO_2 dehydration particularly at pressure above 62 bar. At high pressures, the TEG solubility in the CO_2 stream increases leading to high glycol losses. One possible option to reduce TEG losses is to reduce the CO_2 pressure before feeding it to the contactor. However, this solution may not result in an energy efficient process since the dried CO_2 has to be recompressed again at a high pressure for pipeline transportation purposes. The direct process performance comparison between natural gas and CO_2 dehydration is not straight forward because of different feed conditions and dried stream purity requirements. Therefore, in order to have a fair comparison, all the parameters for the two dehydration processes are made similar including feed temperature, lean solvent temperature, regenerator inlet temperature, condenser temperature and reboiler temperature. The dehydrated product stream for both the cases have been set to contain no more than 100 ppmv of water content. Table 11 shows the comparative results for the natural gas and CO2 dehydration. The results show that for the same dehydration level, natural gas requires 0.2 liters of more glycol per kg of water removed compared to that of the CO_2 dehydration. This results in a decrease of the reboiler duty and hence total energy requirement for the CO_2 dehydration case. However, the solvent losses in case of CO_2 dehydration are 4.21 times more than that of the natural gas process. The high TEG losses in case of CO2 dehydration are mainly attributed to the high pressure feed stream leading to an increased TEG solubility in the CO_2 stream.

Table 11 Performance comparison of natural gas and CO2 dehydration with stripping gas configuration

	Natural gas feed	CO2 feed
Water content in dehydrated stream (ppmv)	100	100
Glycol consumption (L/kg H_2O removed)	18.37	18.17
Reboiler duty (GJ/t of H_2O removed)	5.907	4.89
Total energy (GJ/t of H_2O removed)	9.20	7.37
TEG loss (kg/h)	0.168	0.7065

4.4.4 Sensitivity Analysis

The comparative results showed that the stripping gas configuration offers better performance and lower energy demand compared to the conventional design. Therefore, some of the key variables affecting the process performance have been tested with the stripping gas design. The recorded variables are water content in the dehydrated stream and energy requirement of the process.

4.4.4.1 Effect of Circulation Rate

Glycol circulation rate can directly affect the absorption performance and consequently the water removal performance. The recommended TEG recirculation rate is 17- 42 L per kg of H_2O removed (L. L. Faulkner 2006). Therefore, a sensitivity has been performed to analyze the effect of glycol circulation rate on the process performance keeping reboiler temperature constant. Figure 33 and 34 shows the water content in the dried gas stream with varying TEG recirculation flow rate for the natural gas and CO_2 capture cases, respectively. The results show that any increase in the TEG recirculation rate reduces the water content in the dried stream and hence lower dew point temperatures could be achieved. Results show that the dehydrated gas purity up to 55 and 20 ppmv can be achieved for the natural gas and CO_2 respectively using the stripping gas configuration. However, as the TEG recirculation rate increases, the reboiler duty also increases linearly due to an increased flow to the regenerator. The results show that the total energy requirement increases with an increase in the glycol circulation rate because of an increase in the pump power, cooler duty and condenser and reboiler duties.

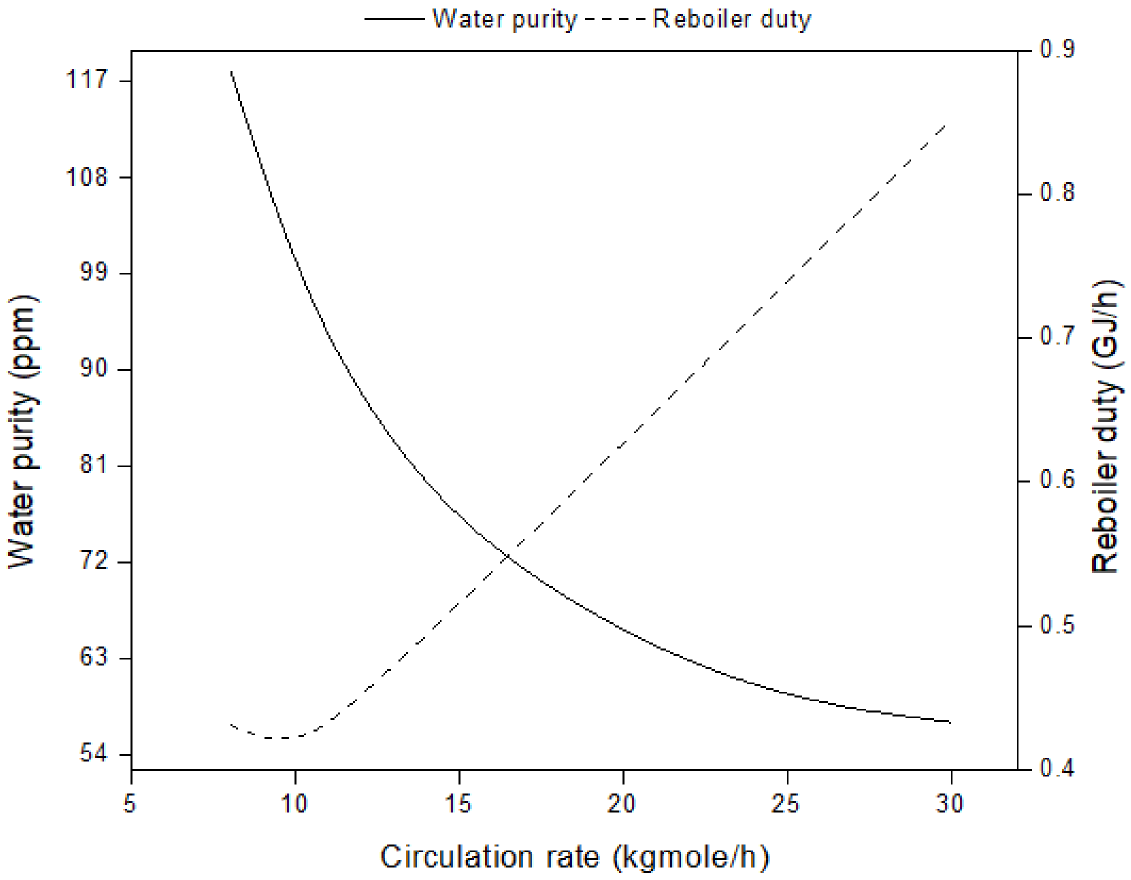

Figure 33 Effect of Circulation Rate on the Water Removal and Reboiler Duty in the Natural Gas Case

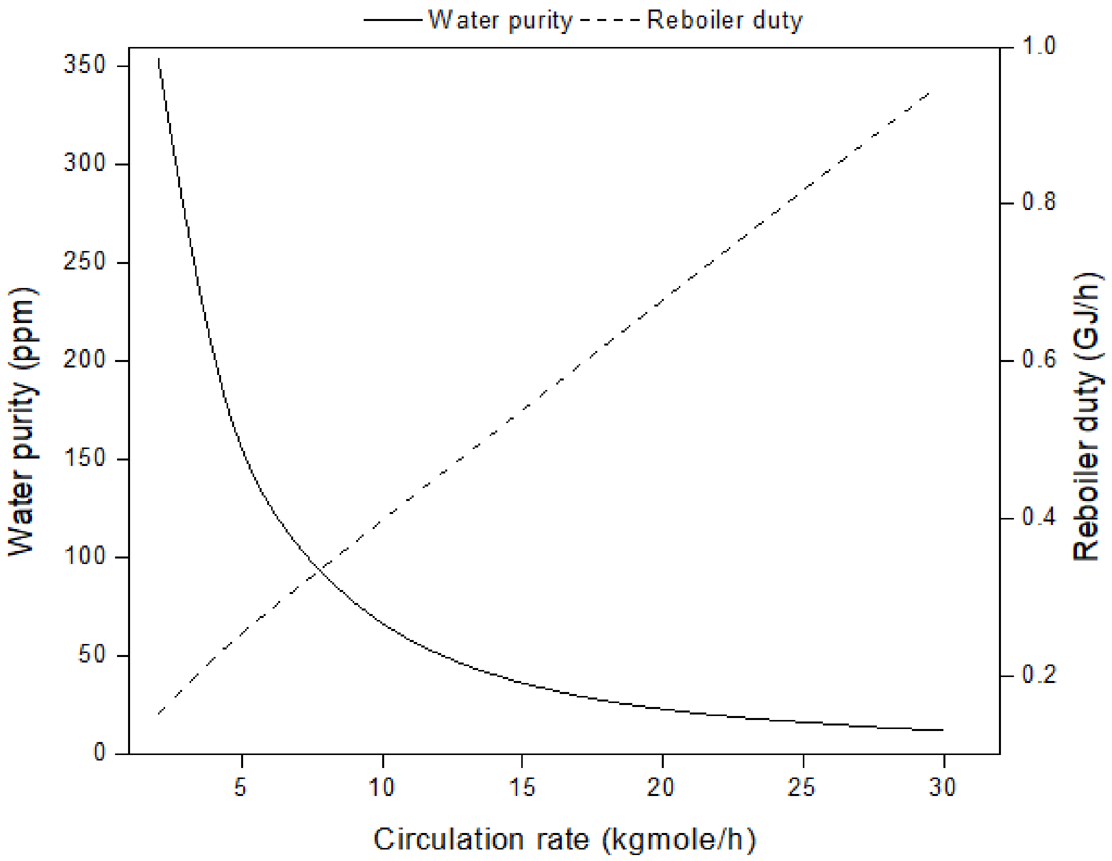

Figure 34 Effect of Circulation Rate on the Water Removal and Reboiler Duty in the Carbon Capture Case

4.4.4.2 Effect of Inlet Gas Temperature

Wet feed gas temperature is an important parameter that can affect the dehydration process performance significantly. In this study, a sensitivity has been performed for the wet gas temperature by maintaining a ΔT of 5 ^{O}C between feed gas and lean solvent. Figure 33 and 34 shows the effect of varying feed temperature on the process performance for the natural gas and CO_2 dehydration designs. The results show that as the contactor temperature decreases, percent removal of water increases significantly. As the feed gas temperature decreases, more water content is removed in the flash separation before entering into the contactor which leads to an improved performance. Secondly, the decrease in the feed temperature increases the solubility of gaseous components in the glycol leading to a decrease in the water content of the dehydrated stream. The results show that as the inlet feed gas temperature decreases, the reboiler duty also decreases because of reduction in the load. While performing this analysis, the temperature difference between the feed gas and lean glycol is maintained at 5 ^{O}C, however, the main impact on the performance is caused because of the gas temperature and lean glycol temperature has a minor effect. This is because of the high feed gas flowrate compared to the lean glycol circulation rate which determines the overall contactor temperature profile.

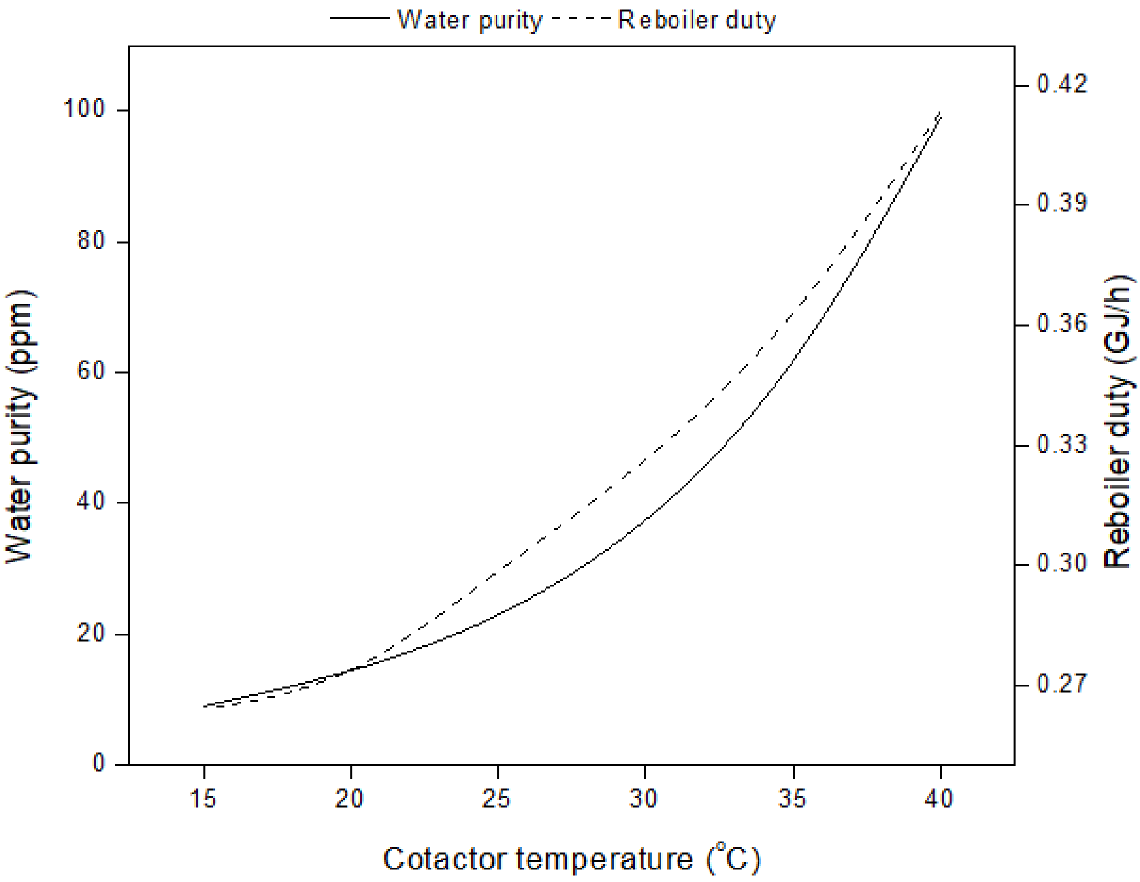

Figure 35 Effect of Contactor Temperature on the Water Removal and Reboiler Duty in Natural Gas Case

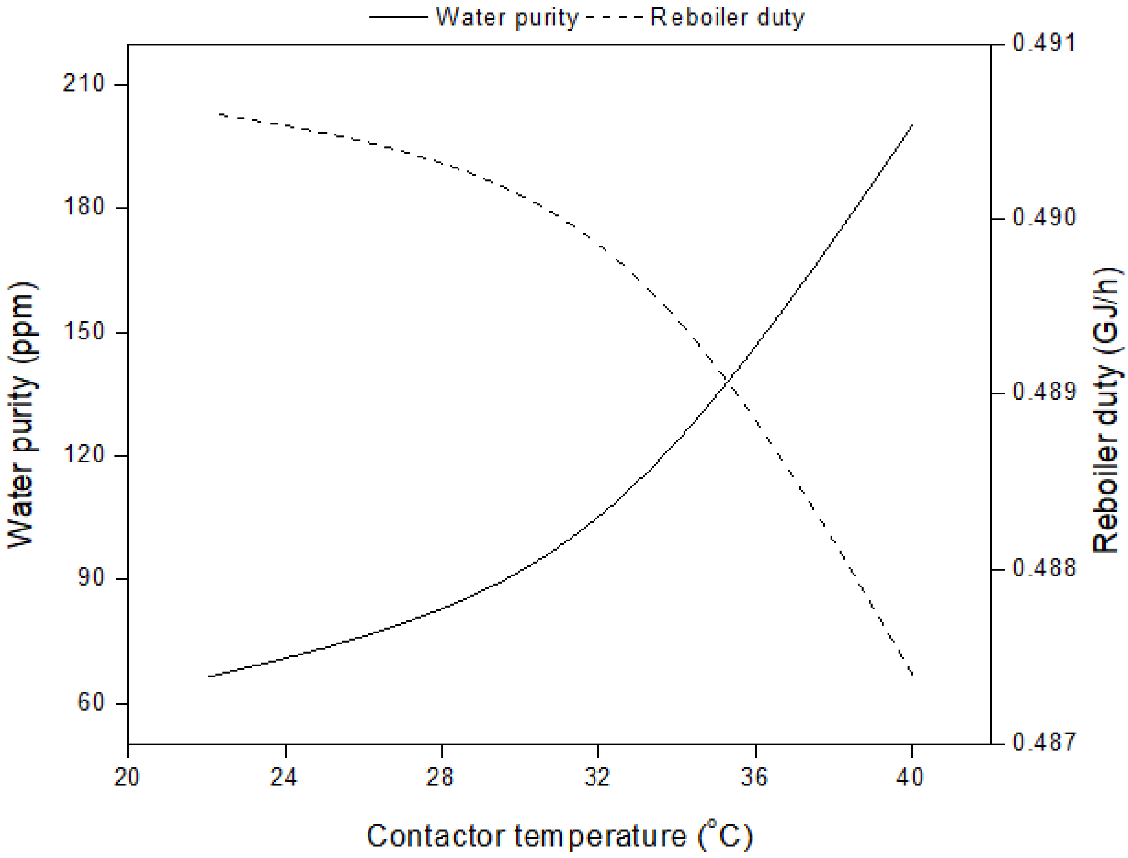

Figure 36 Effect of Contactor Temperature on the Water Removal and Reboiler Duty in Carbon Capture Case

4.4.4.3 Effect of Stripping Gas Flow Rate

The use of stripping gas in the glycol dehydration is a commonly employed technique in order to improve the process performance. The stripping gas can be added directly to the regenerator's reboiler or can be added to the stripping column as shown in the figure 30. Lean glycol concentrations of 99.9 wt. % can be achieved by the stripping gas configuration. The addition of stripping reduces the water partial pressure while the total pressure is maintained constant. A sensitivity analysis has been performed to investigate the effect of stripping gas rate on the process performance as shown in figures 37 and 38. The results show that as the stripping gas rate increases, the lean TEG concentration increases leading to a lower water content in the dehydrated stream. The stripping gas addition to the regenerator also has a positive impact on the reboiler duty as shown in the figures 37 and 38. The results show a reduction in the reboiler duty as the stripping gas flowrate is increased. Many studies in the past have only emphasized on the performance improvement by using the stripping gas, however, there is a maximum allowable stripping gas rate that can be fed to the reboiler in order to keep TEG losses to the minimum. Different equipment vendors have their recommendations for the maximum allowable stripping gas rate ranging between 35 – 60 SCM per m^3 of glycol. Also, the results show the effect of stripping gas on the TEG losses. As evident from the results, TEG losses are proportional to the stripping gas rate.

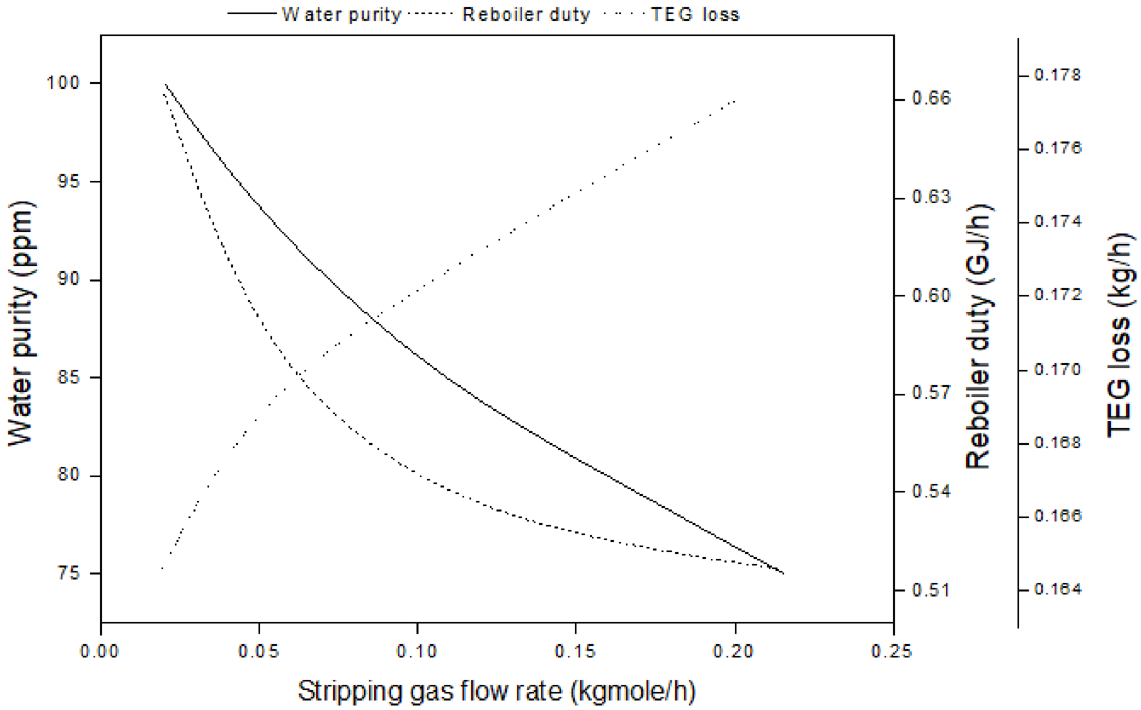

Figure 37 Effect of Stripping Gas Flow on the Water Removal, Reboiler Duty and TEG Loses in the Natural Gas Case

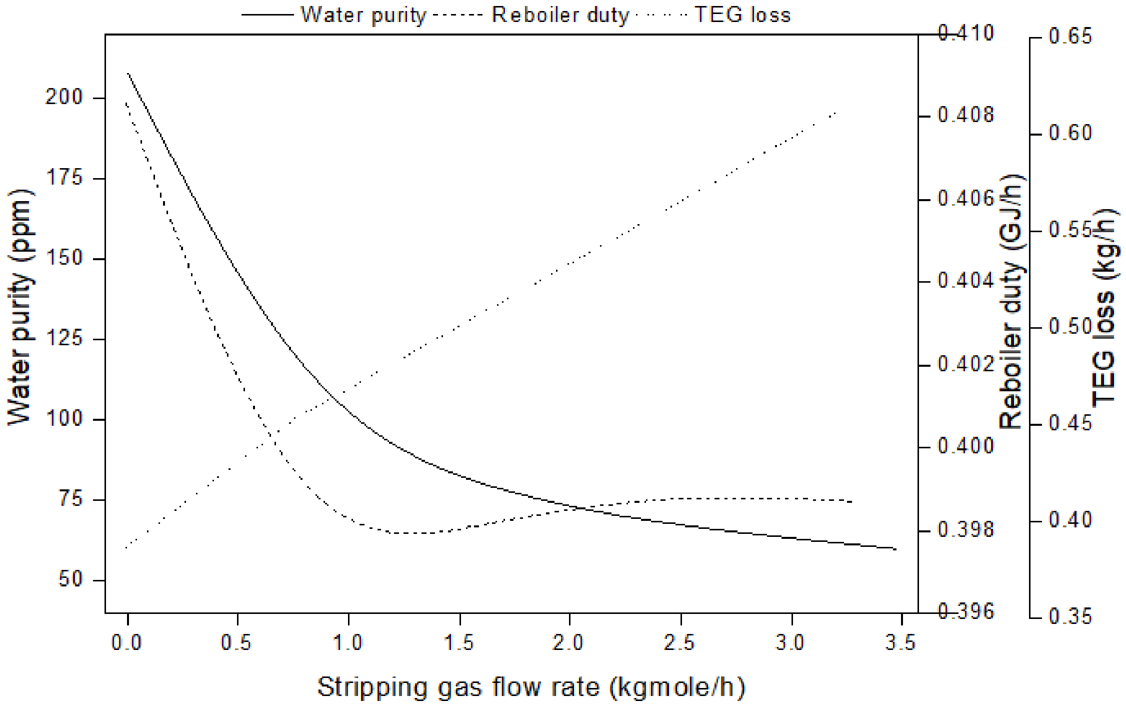

Figure 38 Effect of Stripping Gas Flow on the Water Removal, Reboiler Duty and TEG Loses in the Carbon Capture Case

4.4.4.4 Effect of Reboiler Temperature

Reboiler temperature can be changed to provide additional heat and is directly connected with the reboiler duty. TEG can thermally decompose at temperatures higher than 204 ^{O}C and hence the maximum reboiler temperature should be always kept below this limit (Russell et al. 2004). Figures 39 and 40 shows the effect of reboiler temperature on the dehydration process performance for natural gas and CO_2 stream, respectively. The results show that as the reboiler temperature is increased, leaner TEG can be produced leading to a lower water content in the dried gas. However, an increase in the reboiler temperature results in high reboiler duty as shown in figures 39 and 40.

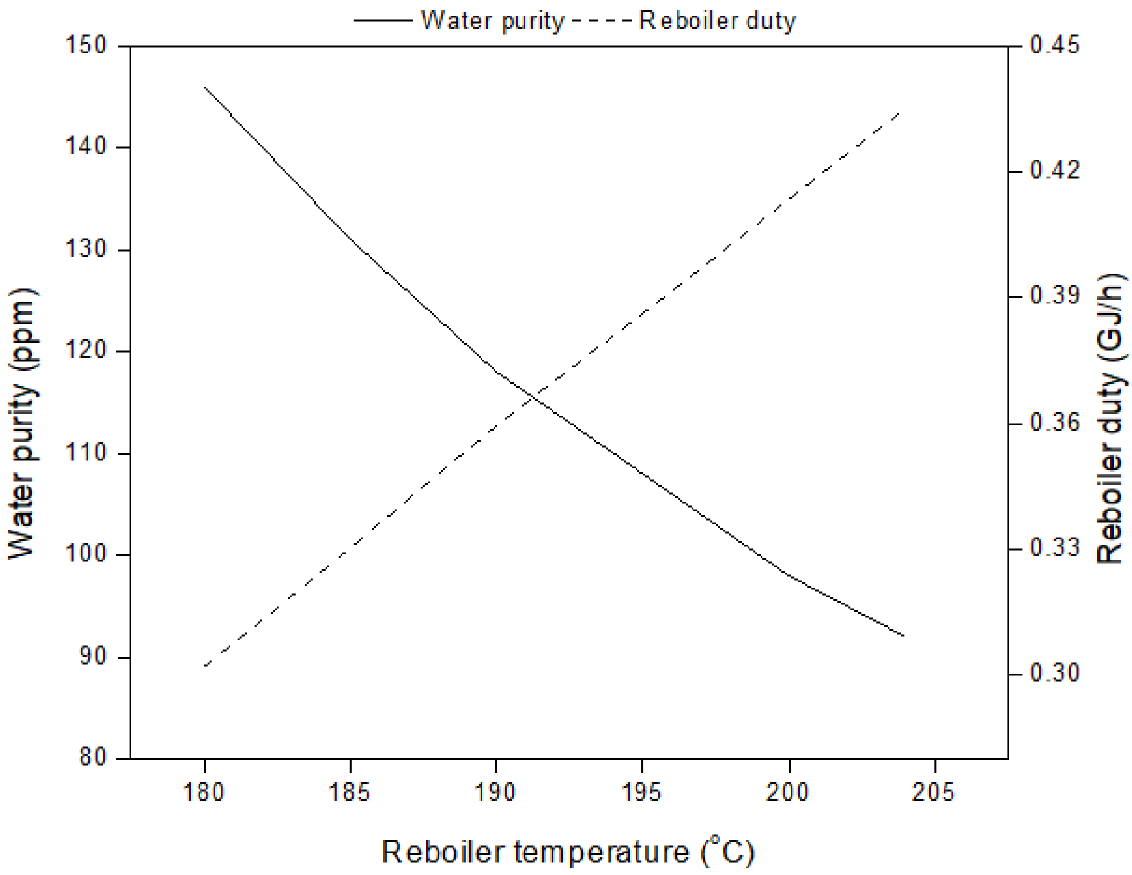

Figure 39 Effect of Reboiler Temperature on the Water Removal and Reboiler Duty in Natural Gas Case

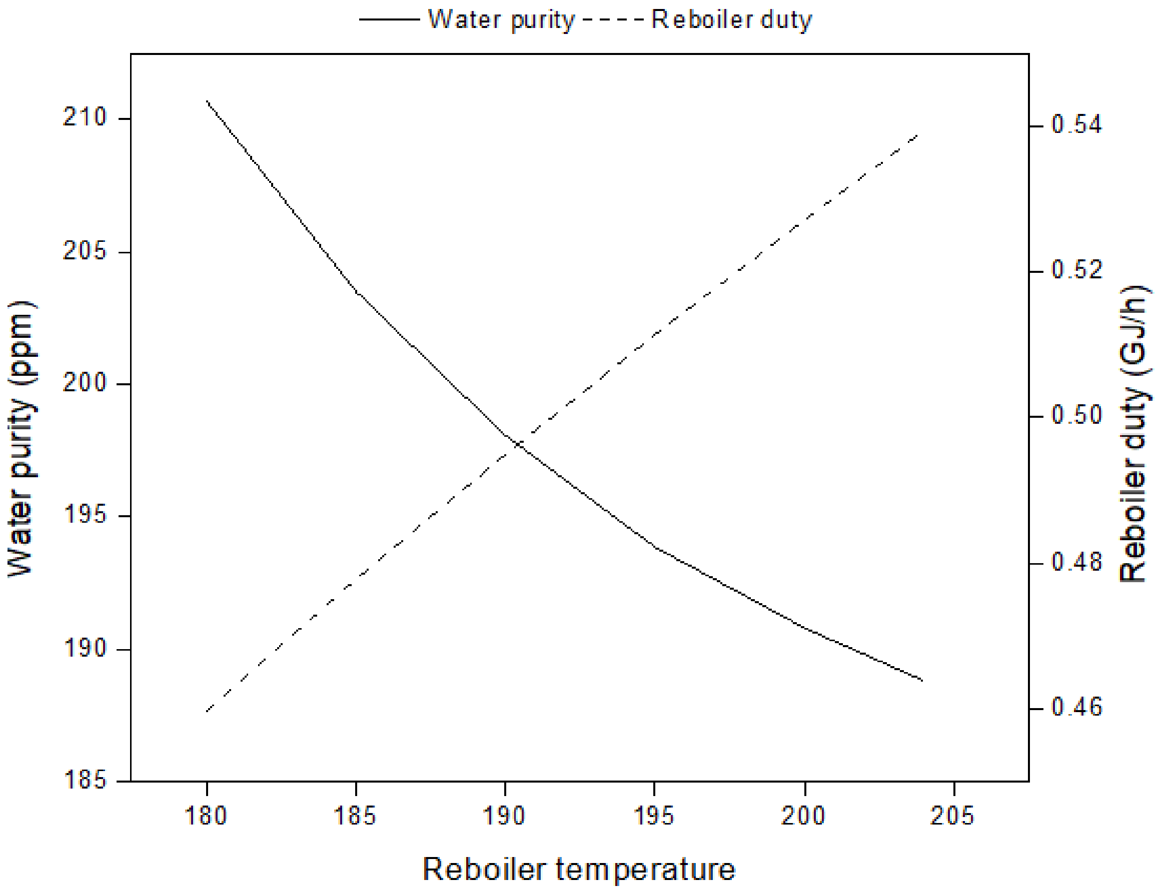

Figure 40 Effect of Reboiler Temperature on the Water Removal and Reboiler Duty for Carbon Capture Case

4.4.4.5 Effect of Absorber Stages

The base case assumed three absorber stages for all the design configurations. Parametric analysis has been done to analyze the impact of number of contactor stages on the dehydration performance. The results show that as the number of absorber stage increases, lower water contents can be achieved in the dehydrated stream. However, the water content in the dry gas becomes steady after certain number of stages as shown in figures 41 and 42. The reboiler duty varies directly with the water load in the rich TEG stream. The results show that as the number of absorber stages, quantity of water absorbed increases in the rich TEG leading to a higher reboiler duty.

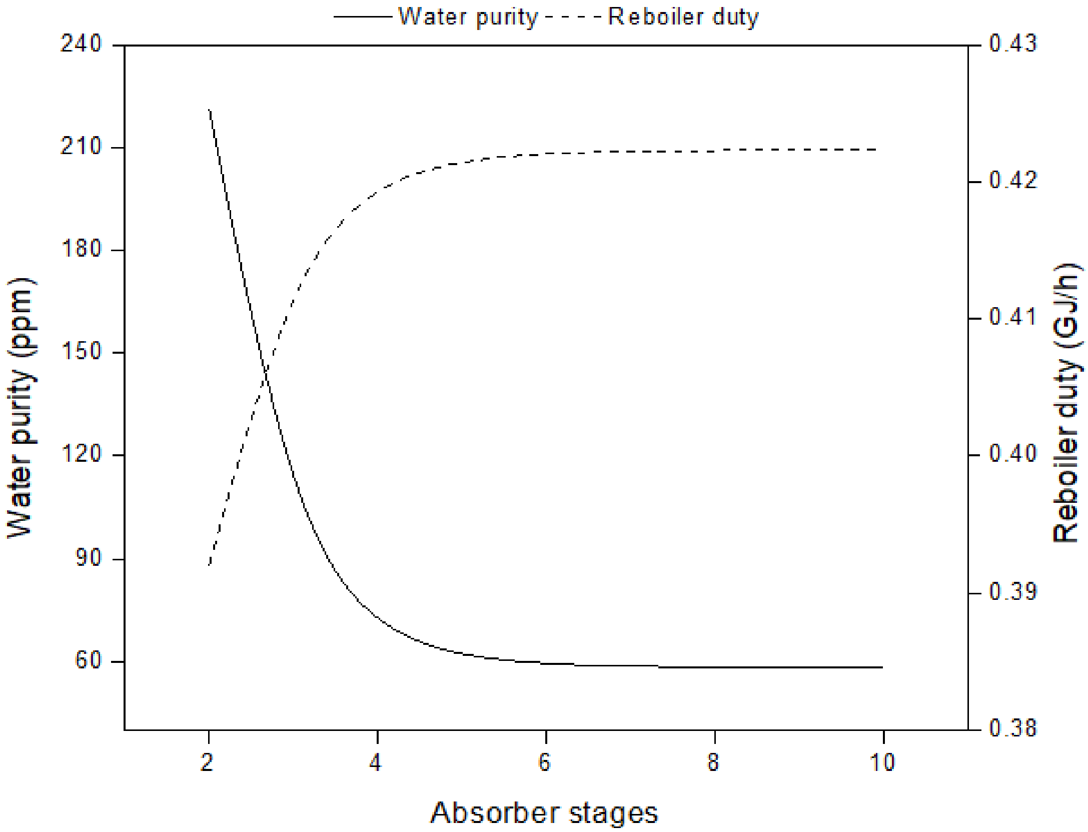

Figure 41 Effect of Absorber Stages on the Water Removal and Reboiler Duty in Natural Gas Case

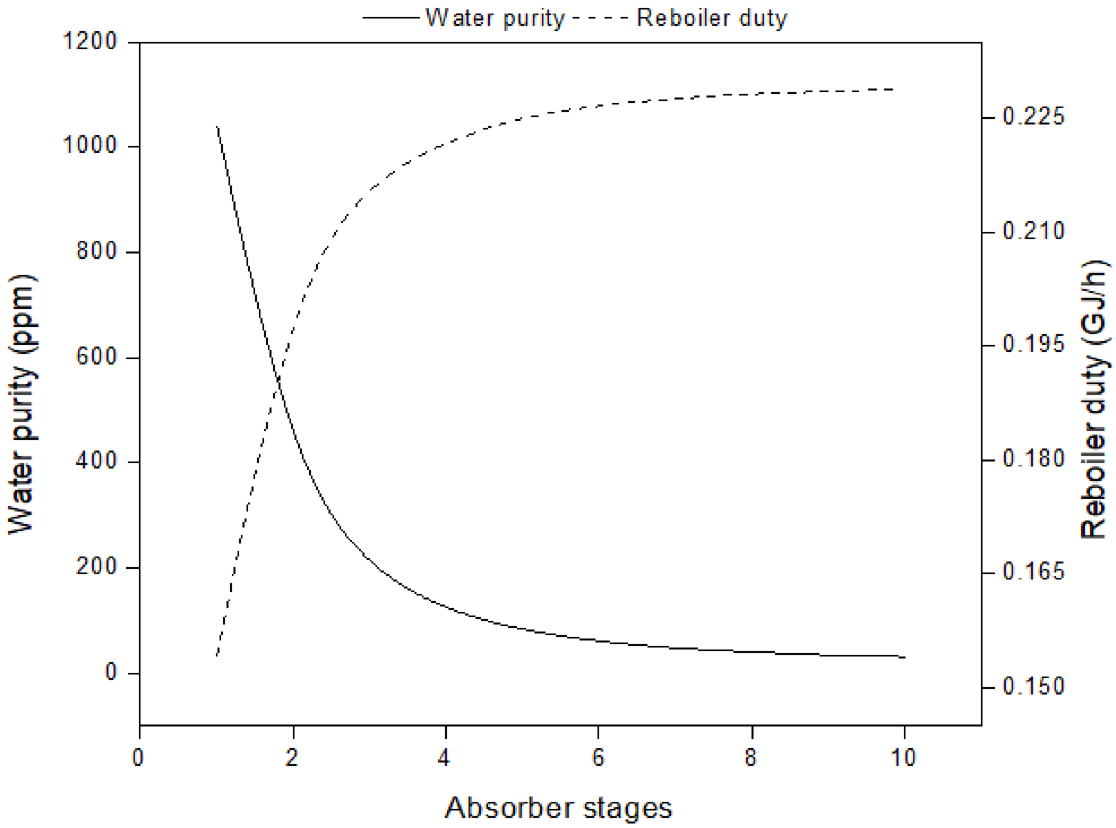

Figure 42 Effect of Absorber Stages on the Water Removal and Reboiler Duty for Carbon Capture Case

CHAPTER 5

NGL FRACTIONATION

5.1 Introduction

The focus in the present study will be on NGL recovery and fractionation. This process aims to separate the NG components, mainly C_1-C_{5+} into their pure products. Distillation is the main purification technology used for NGL fractionation process although Distillation columns are highly energy intensive equipment. Most of separation and purifications in industrial scale are done using distillation (Humphrey Austin, TX (United States)] 1995). Moreover, the largest portion of plant operating cost in most of the industrial plants is the distillation section. Other design alternatives and methods for distillation column are tested to reduce the operating cost required for the plant and pollutant emissions which consequently help saving the environment. Binary or multicomponent mixtures can be separated at high purities using distillation technology.

Many studies were done in NGL recovery and fractionation areas. Most of these studies are mainly focused on optimizing various retrofit designs and comparing them with the direct sequencing in terms of cost and energy savings. However, there are no recent studies that investigated parametric analysis for turbo-expander and dividing wall column configurations for NGL systems.

The present study aims to understand the attitude of NGL recovery and fractionation processes. One of the Middle East feed specifications has been selected to perform the analysis and the feed specifications are listed in table 12. The feed is the product from the acid gas removal and dehydration plants so the risk of forming gas hydrates is relatively low. A conventional direct sequence that includes turboexpansion demethanizer, deethanizer, depropanizer, deisobutanizer and debutanizer will be simulated first using commercial simulation software. Two consecutive CDWC are the proposed alternatives for the deC$_2$-C$_3$ and deiC$_4$-nC$_4$, respectively. Since the distillation technology is highly energy intensive, the energy performance will be analyzed and compared between the conventional and the alternative designs for each component individually. The proposed configurations will be connected to determine the energy savings using the proposed design compared to the conventional scheme. Finally, sensitivity analysis for the conventional and proposed schemes will be done to investigate the impact of key parameters namely, feed temperature, operating pressure, feed stage location, split ratio, number of column stages, product purity and the withdrawn side stage to the energy performance of the separation process.

Feed specifications	
Feed flow (kgmole/h)	2732
Feed temperature (^{o}C)	42
Feed pressure (bar)	51
C_1 (mole fraction)	0.9258
C_2 (mole fraction)	0.0169
C_3 (mole fraction)	0.0034
iC_4 (mole fraction)	0.0006
nC_4 (mole fraction)	0.0013
iC_5 (mole fraction)	0.0004
nC_5 (mole fraction)	0.0004
nC_6 (mole fraction)	0.0002
nC_8 (mole fraction)	0.0002
N_2 (mole fraction)	0.0506
CO_2 (mole fraction)	0.0001
H_2O (mole fraction	0.0001

5.2 Simulation Basis

Many models can be used to estimate the thermodynamic properties of the light hydrocarbons. Peng Robinson equation of state has been selected to perform this study because of its reasonable accurate estimations over wide ranges of temperature and pressure.

5.2.1 Turbo-expander Configuration

Separating methane from NGL components has some challenges in terms of maintaining the operating conditions. Turbo-expander has been selected for the design of demethanizer due to its capability to purify methane with low energy requirements. The feed gas that represents the product from the dehydration unit is splitted two times to cool the stream. For the purpose of cooling, two pump arounds streams that are withdrawn from the demethanizer and reboiler duty were used as shown in figure 43. The feed gas is mixed again and flashed. The bottom product which consists of lower flow rate enters the demethanizer where the top product is expanded. The expander duty is used as an electrical source for the compressor to pressurize the methane product. The outlet stream from the expander is fed to a secondary flash drum where the flash gas which consists mainly of methane is mixed with the overhead of the demethanizer. The bottom product of the flash at -85 ^{o}C is entered to the demethanizer which has 11 trays.

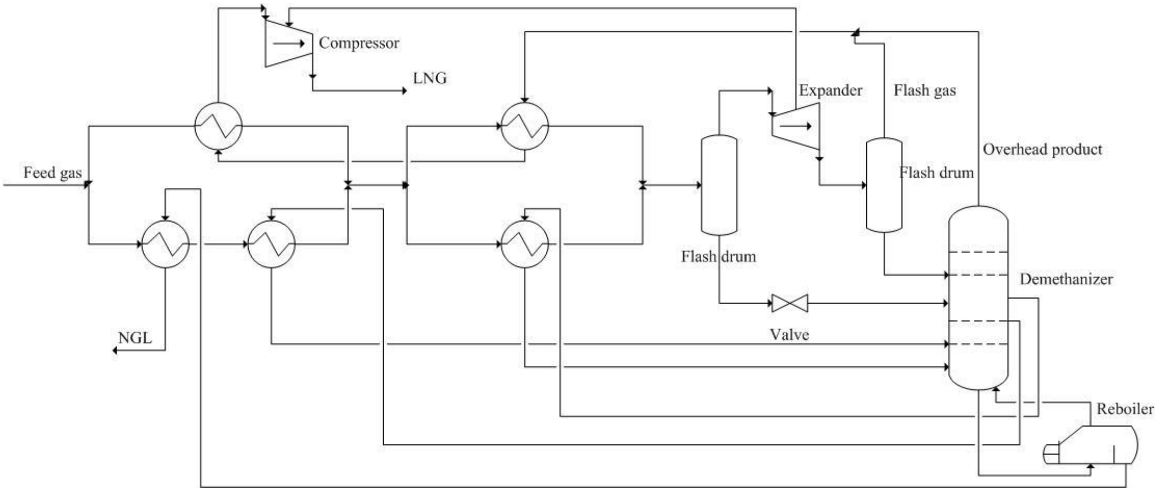

Figure 43 Turbo-Expander Configuration

5.2.2 Conventional Distillation Sequence

The components targeted to be purely produced are the C_2-C_{5+}. For this purpose, connected series of distillation columns are used namely deethanizer, depropanizer, debutanizer and deisobutanizer. All the targeted products are withdrawn from the top of the columns except for the n-butane where it leaves from the bottom of the deisbutanizer. All columns have a reboiler and a condenser to supply the heat required to perform the separation as shown in figure 44. The deethanizer is operated at low temperature and require refrigerated condenser while cooling water is sufficient to condense propane, butane and isobutanee in the depropanizer, debutanizer and deisobutanizer, respectively. To overcome the excessive cost of the refrigeration, the deethanizer are operated at high pressures. Table 12 shows the design specifications that are used for all the columns. Since the boiling point of the i-butane and n-butane is close to each other, the number of trays required to perform the separation is relatively high i.e. >70. Shortcut distillation columns and GPSA Engineering Data Book (Russell et al. 2004) were used to select the number of stage, operating pressure and assure that the reflux ratio lies within the acceptable range of operation.

Table 12 Design Specifications for Conventional Columns Series

	Deethanizer	Depropanizer	Debutanizer	Deisbutanizer
Operating pressure (bar)	31	18	5	6.5
Number of trays	34	33	35	80
Feed stage location	25	7	4	47
Reflux ratio	1.92	3.32	1.80	12.47

Top temperature	10	50	42	44
Bottom temperature	109	123	103	64

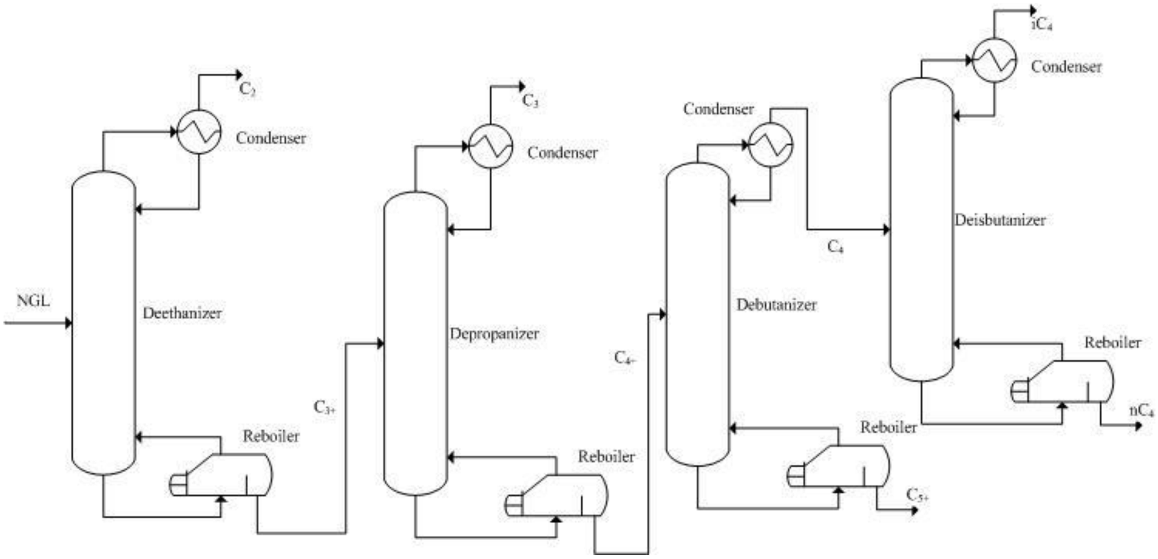

Figure 44 Conventional Column Sequence

5.2.3 Conventional Dividing Wall Column Sequence

CDWC is selected as an alternative configuration to the conventional sequence for the NGL fractionation due to its capability to save energy and installation space. There are two CDWC are inserted in this flowsheet as shown in figure 45. The first column substitutes the deethanizer and depropanizer where propane is withdrawn from the middle right section of the column. Similarly, to the conventional configuration, the first CDWC is operated at high pressure (18 bar) to compromise between the capital cost, refrigeration cost and reboiler duty required. The second CDWC is used to separate i-butane, n-butane and pentane and higher molecular weight components where the high purity (95%) n-butane leaves the column from the middle right section. For both CDWCs, the feed stage enters the columns from the middle left section, namely prefractionator section. The total number of stages and the distribution of stages are selected based on the heuristics given in the Advanced Distillation Technologies book (Kiss 2013). As a base case, number of stages of CDWC is selected as 80% of the total number of stages of its corresponding conventional columns. Moreover, the distribution of stages is assumed to be equal i.e. 33% for each of top, middle and bottom sections. Table 13 shows the design specifications of the first and second CDWC. Since the design of CDWC is more complex and the degree of freedom for this column is five, five parameters were specified to solve the CDWC stage by stage. The parameters selected for both CDWC are vapor split ratio, liquid split ratio, overhead product purity, middle product purity and the flow rate of the withdrawn side stream based on the results got from conventional scheme.

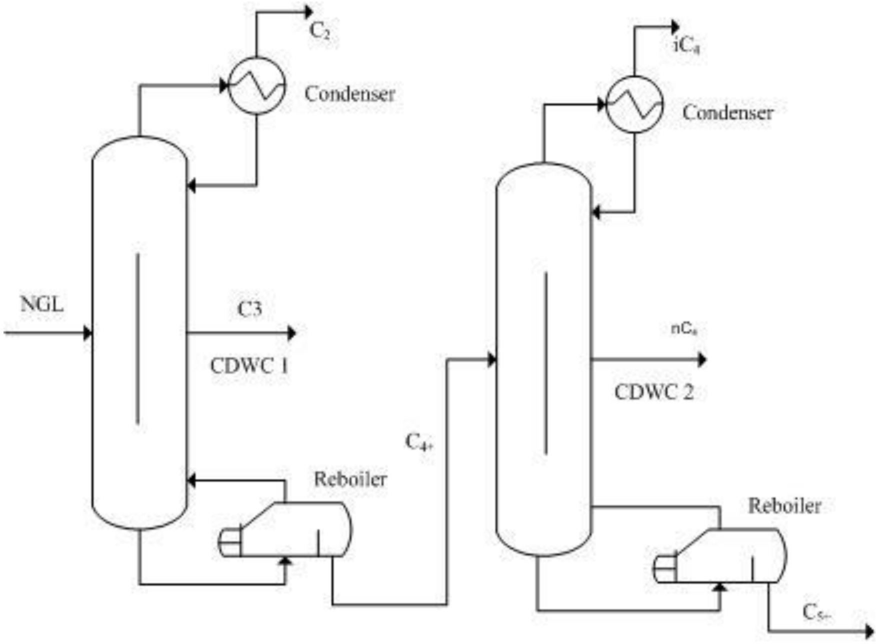

Figure 45 CDWC series

Table 13 Dividing Wall Column Design Specifications

	CDWC 1	CDWC 2
Operating pressure (bar)	18	5
Number of trays	54	92
Feed stage location	2 (prefractionator section)	7 (prefractionator section)
Reflux ratio	2.3	14.7
Top temperature	-12	30
Bot temperature	123	107
Distribution of stages	18 in each section	31 in the top and middle sections and 30 in the bottom section
Vapor split ratio	0.5	0.5
Liquid Split ratio	0.5	0.5

5.3 Results and Discussion

Because the NGL recovery and fractionation process are mainly done by distillation technology, these units are highly energy intensive. This section will shed a light on the energy consumption of both conventional and the proposed design sequence and the energy savings can be obtained in each sub-unit and in entire flowsheet.

5.3.1 Shortcut Distillation

The shortcut method, based on Fenske-Underwood and Gililand models, was used to estimate minimum number of stages, actual number of stages, feed stage location, minimum reflux ratio, condenser and reboiler duties and overhead and bottom temperatures. The parameters must be specified is the product purity in overhead and bottom streams along with the operating pressure. Table 14 summarizes the results of the shortcut distillation. Most of these results are in good agreement with the GPSA Engineering Data Book values which makes our model reliable.

Table 14 Shortcut Distillation Results

	Deethanizer	Depropanizer	Debutanizer	Deisbutanizer
Number of trays	34	29	45	85
Feed stage location	25	9	5	50
Reflux ratio	1.74	2.87	1.92	16.24
Top temperature	9	50	42	44
Bot temperature	109	123	103	64

5.3.2 Turbo Expander Configuration

Turbo-Expander configuration has been selected to simulate the demethanizer in order to obtain high purity methane utilizing as low energy as possible since this process requires low temperature operation. About 94.4% purity of methane is obtained at the overhead of the demethanizer and it leaves the column at -82 ^{o}C. In terms of energy performance, the heat integration in this configuration is sufficient to omit all the external energy sources. The pump around streams, overhead product and NGL provide the cooling duty required for the feed stream while the expander duty can be used as an electrical energy source for the compressor.

5.3.3 Comparison between Conventional and CDWC Configuration

Since one centered divided wall column with a single reboiler and condenser can substitute two conventional columns, the number of energy units i.e. reboilers and condensers are reduced by half. This can lead to lower investment cost and space saving. To easily compare between conventional scheme and CDWC configuration, the desired purities for ethane, propane, isobutane and n-butane are fixed for both schemes. Moreover, the side product from the middle right section of the CDWC flow rate value is selected the same as the result of the conventional design. Table 15 showed the desired purities and the reboiler and condenser duty for the conventional and CDWC configurations. The sum of the reboiler duties of the deethanizer and depropanizer is evaluated as 1090 kW. As expected and cited in the literature, the CDWC design in the present study has the capacity to save 18% of the energy utilized in the conventional flowsheet. Similar trend is found in the second CDWC where the energy savings reaches 40 % compared to the conventional deisobutanizer and debutanizer.

Table 15 Product Purities and the Energy Duties required for the Conventional and Dividing Wall Column Sequence

Product purities	
C_1 (mole fraction %)	0.9444
C_2 (mole fraction %)	0.99
C_3 (mole fraction %)	0.973
iC_4 (mole fraction %)	0.99
nC_4 (mole fraction %)	0.95
Energy parameters	

Deethanizer reboiler duty (kW)	808
Deethanizer condenser duty (kW)	416
Depropanizer reboiler duty (kW)	282
Depropanizer condenser duty (kW)	323
Debutanizer reboiler duty (kW)	138
Debutanizer condenser duty (kW)	190
Deisobutanizer reboiler duty (kW)	234
Deisobutanizer condenser duty (kW)	226
First CDWC reboiler duty (kW)	828
First CDWC condenser duty (kW)	675
First CDWC reboiler duty saving (%)	24
First CDWC condenser duty saving (%)	9
Second CDWC reboiler duty (kW)	214
Second CDWC condenser duty (kW)	262
Second CDWC reboiler duty saving (%)	43
Second CDWC condenser duty saving (%)	37

5.4 Sensitivity Analysis

Many parameters are tuned in both conventional and CDWCs to investigate the impact of these parameters on the reboiler and condenser duties per column. These parameters include operating conditions such as feed temperature and column pressure and column design aspects such as inlet feed location and side draw stage.

5.4.1 Conventional Columns Sensitivity Analysis

5.4.1.1 Effect of Feed Temperature

Feed temperature for the deethanizer, depropanizer, debutanizer and deisobutanizer is manipulated to study its effect on the reboiler and condenser duties. The range of feed temperature selected is between the condenser and reboiler temperatures. Figures 46, 47, 48 and 49 show the influence of feed temperature on the energy performance of deethanizer, depropanizer, debutanizer and deisbutanizer, respectively. In all columns, rising the feed temperature leads to an increase in the condenser duty. Deethanizer, depropanizer and desiobutanizer reboiler duties tend to decline as the temperature increases. All columns share the same behavior where after a certain point the decrease in the reboiler duty becomes sharper. However, for the debutanizer, reboiler duty increases while the feed temperature increases.

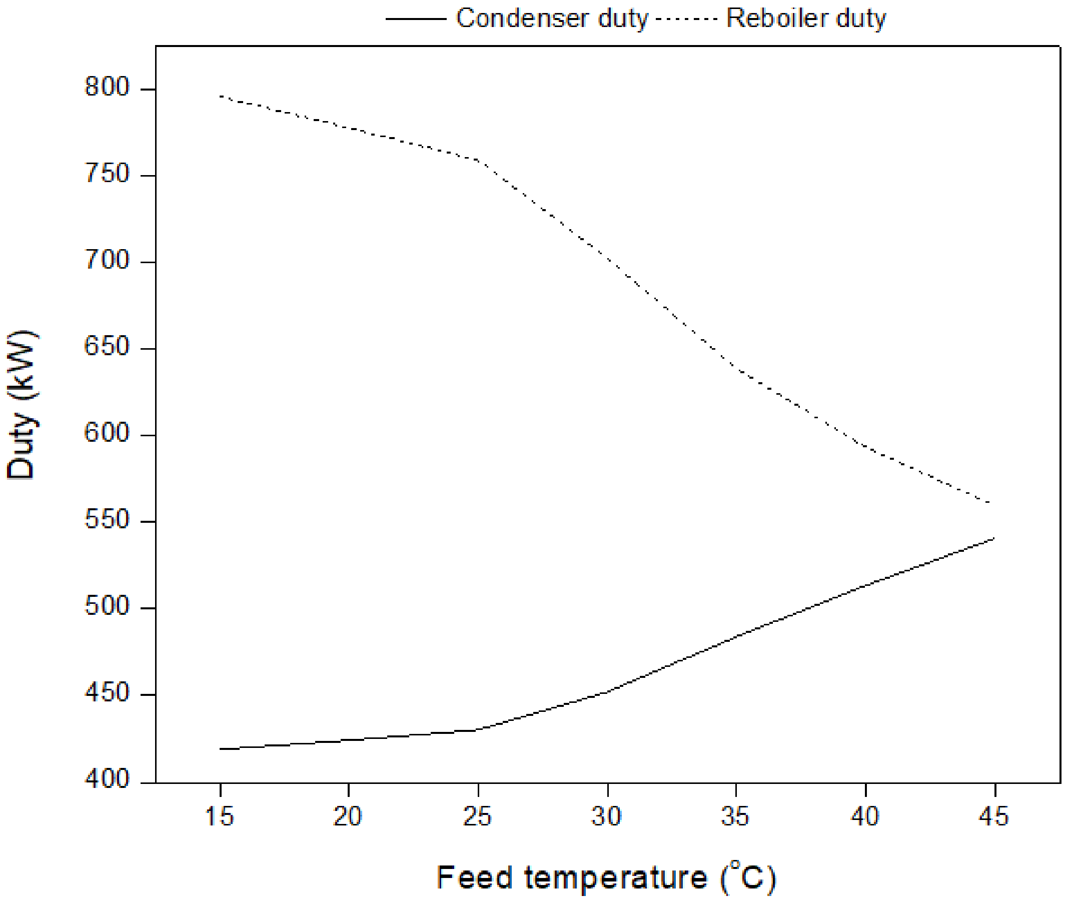

Figure 46 Effect of Feed Temperature on the Reboiler and Condenser Duties in the Deethanizer

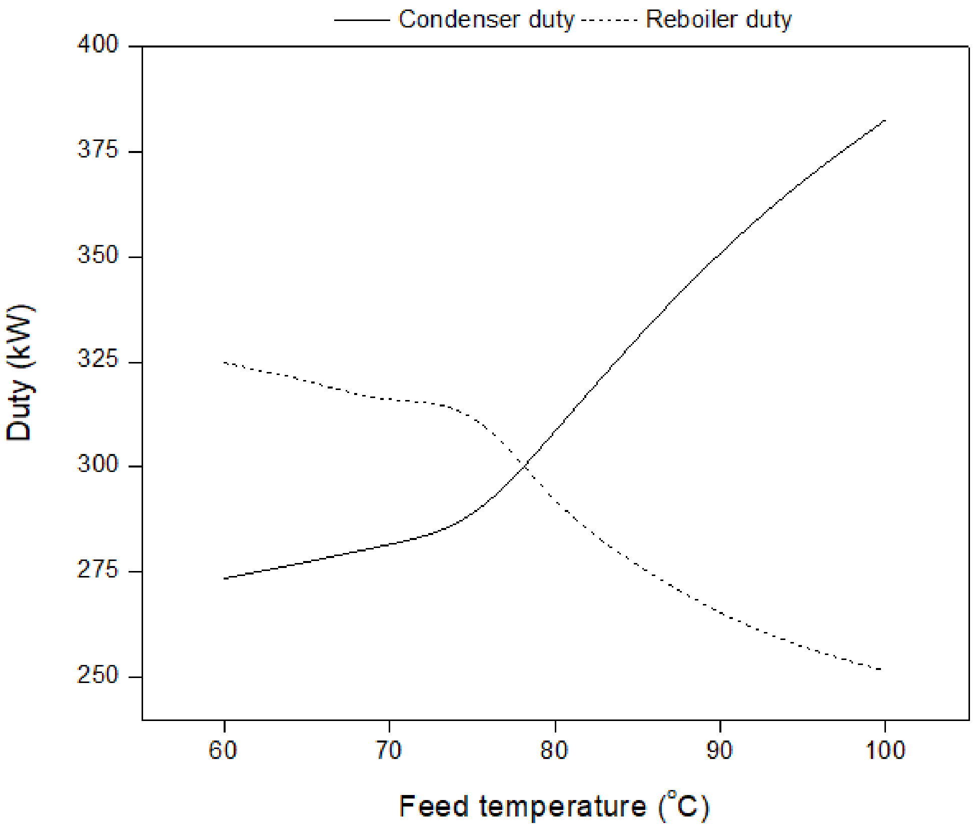

Figure 47 Effect of Feed Temperature on the Reboiler and Condenser Duties of the Depropanizer

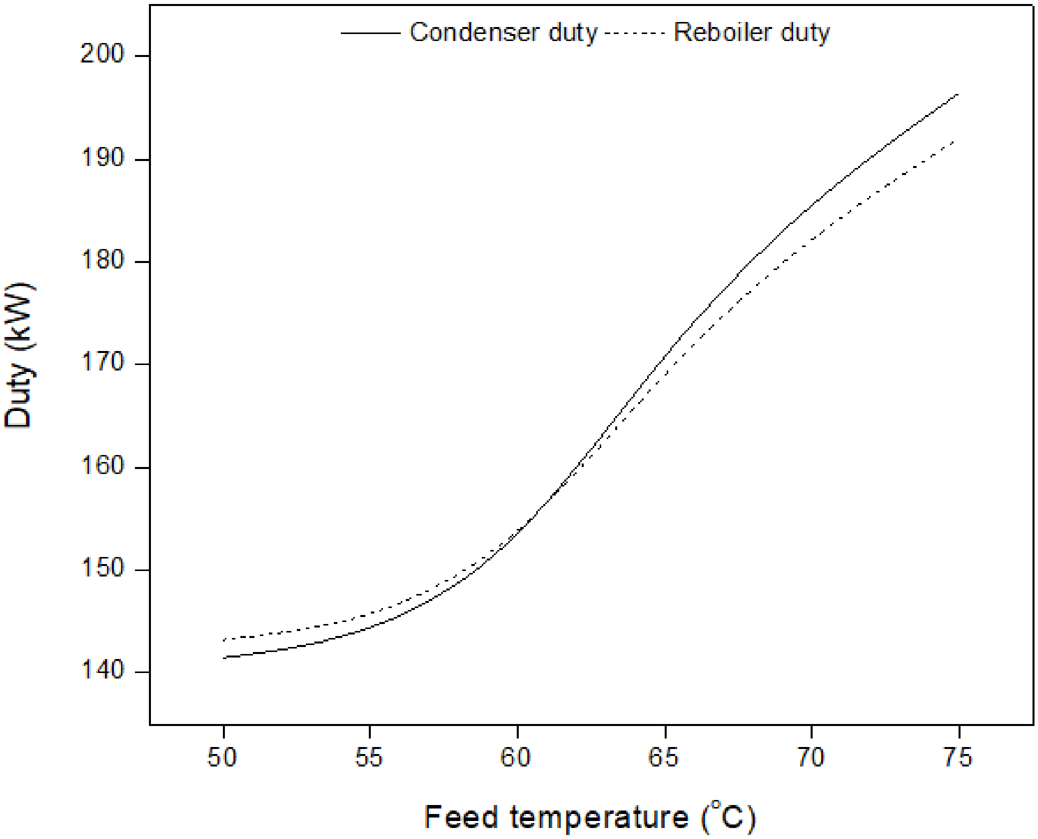

Figure 48 Effect of the Feed Temperature on the Reboiler and Condenser Duties in the Debutanizer

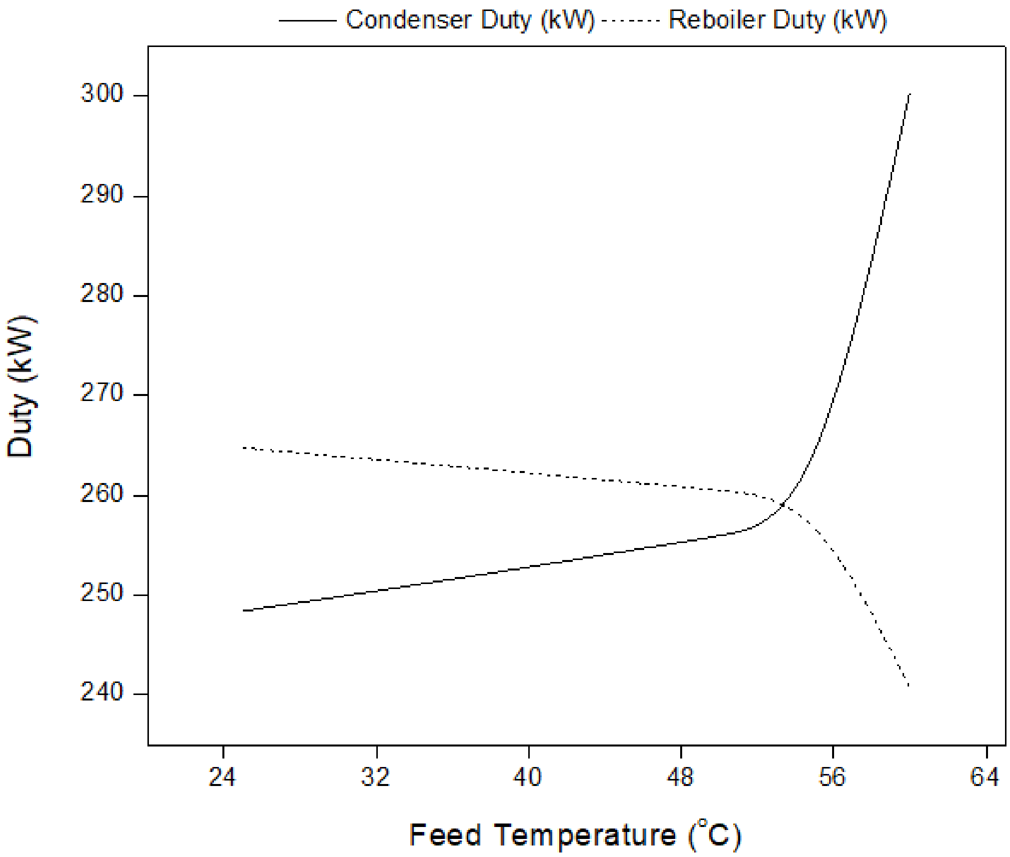

Figure 49 Effect of the Feed Temperature on the Reboiler and Condenser Duties in the Deisobutanizer

5.4.1.2 Effect of Column Pressure

Column operating pressure has significant impact on the energy and operating cost of the process. The effect of tuning the pressure of fractionation columns has been tested against the reboiler and condenser duties in the present study. Column pressure typically associated with the reboiler and condenser temperatures as shown in figures 50-57. These pressure values are selected according to the typical ranges suggested in the GPSA Engineering Data Book (Russell et al. 2004). Results show that increasing the column pressure within this range increases the corresponding saturation temperature of the condenser and reboiler. Moreover, as agreed with literature (Long & Lee 2012), the results show that increasing the column pressure will lead to an increase in the reboiler and condenser duties. One exceptional case in the debutanizer where condenser duty decreases as the column pressure increases.

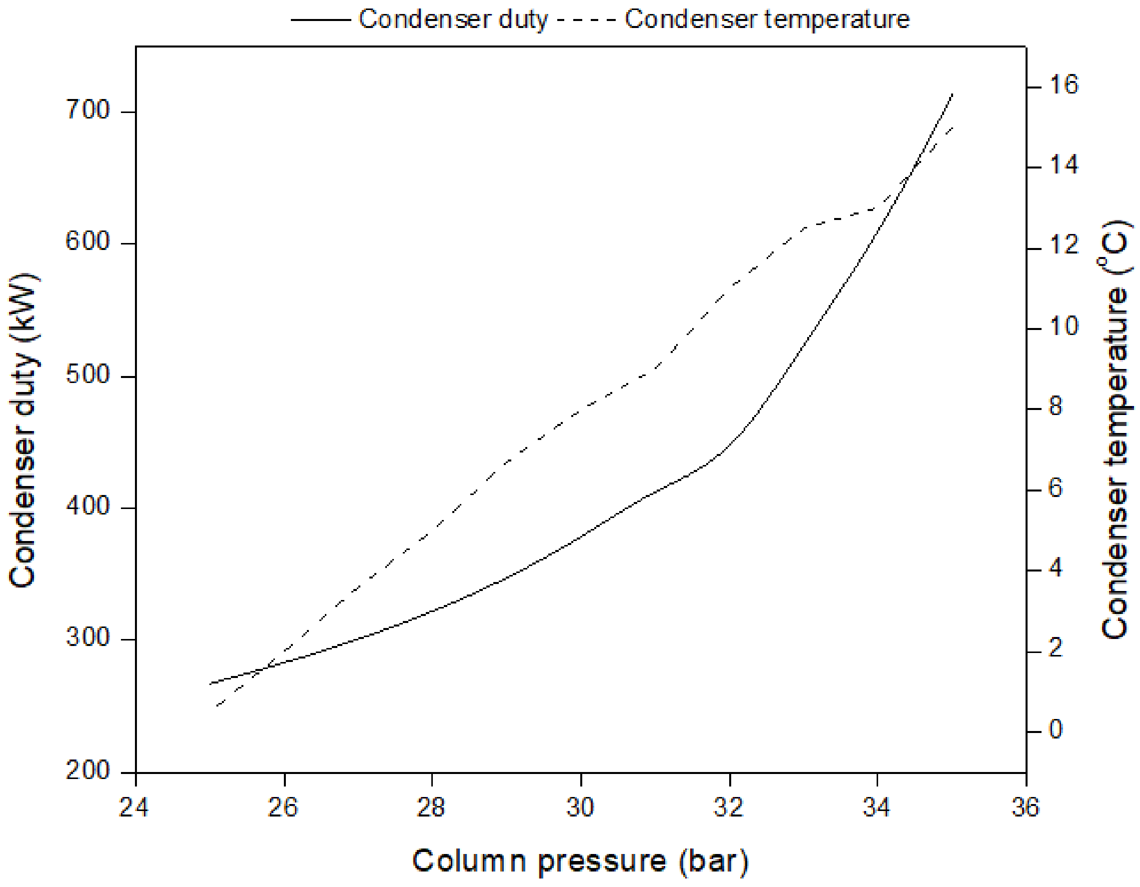

Figure 50 Effect of the Column Pressure on the Condenser Duty and Temperature in the Deethanizer

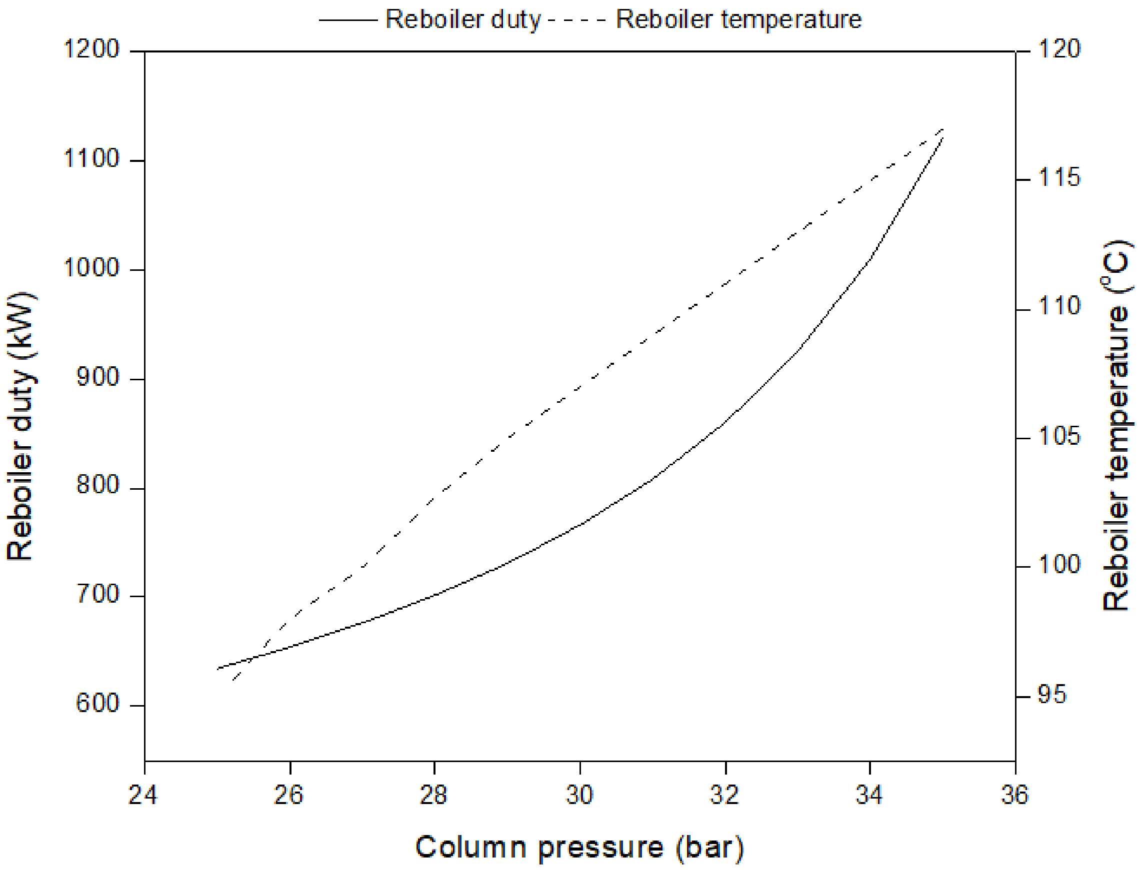

Figure 51 Effect of the Column Pressure on the Reboiler Duty and Temperature in the Deethanizer

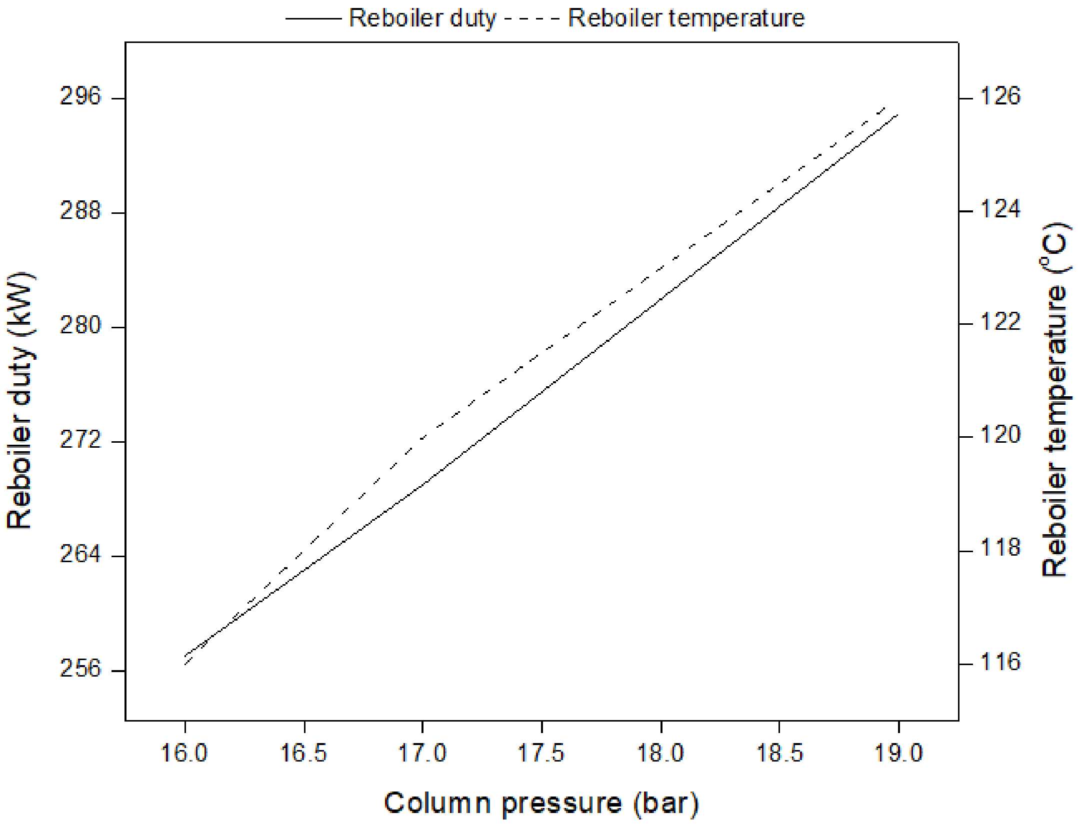

Figure 52 Effect of the Column Pressure on the Reboiler Duty and Temperature in the Depropanizer

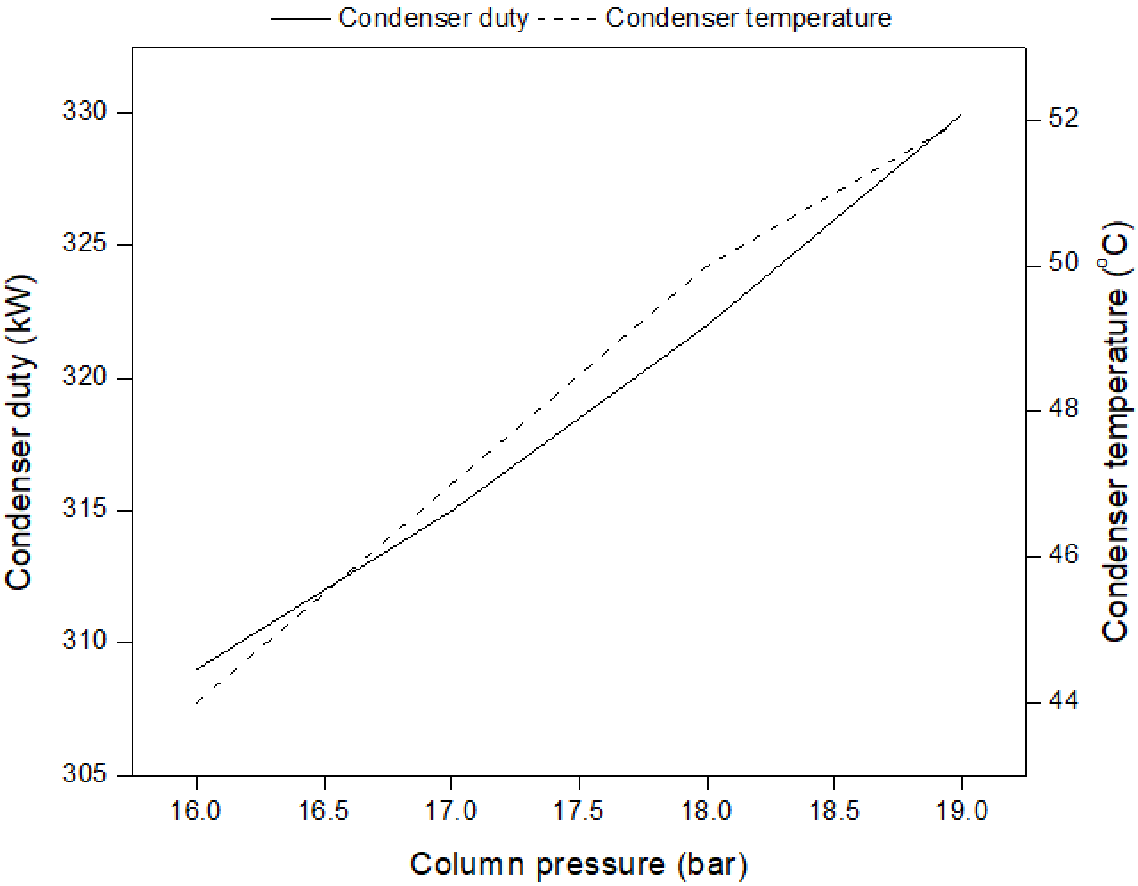

Figure 53 Effect of the Column Pressure on the Condenser Duty and Temperature in the Depropanizer

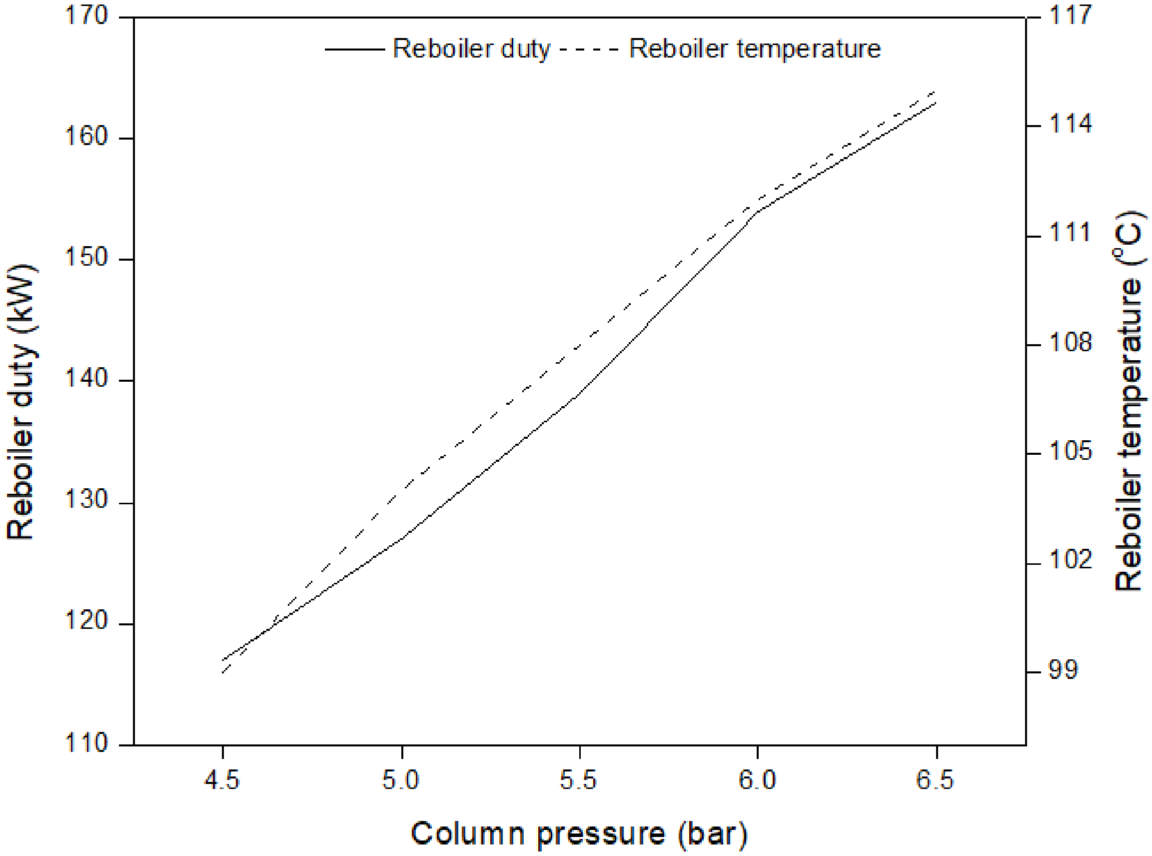

Figure 54 Effect of the Column Pressure on the Reboiler Duty and Temperature in the Debutanizer

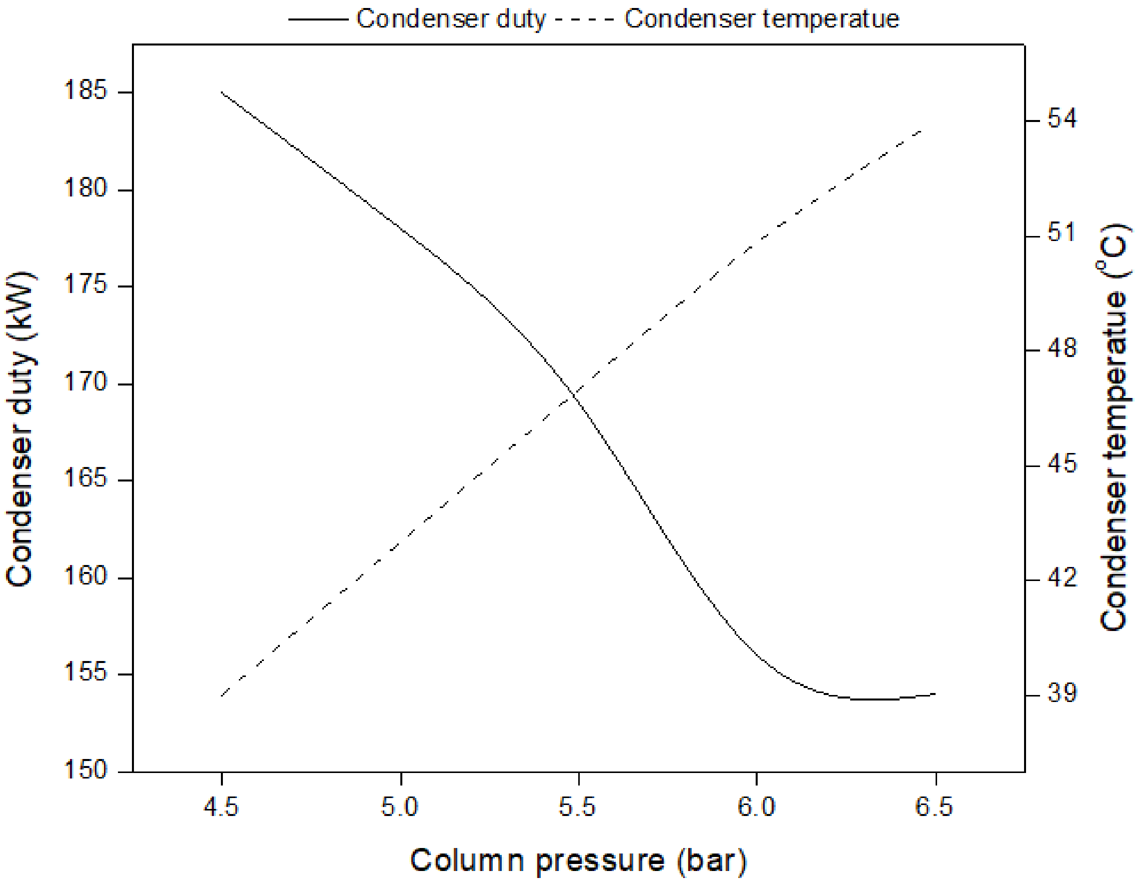

Figure 55 Effect of the Column Pressure on the Condenser Duty and Temperature in the Debutanizer

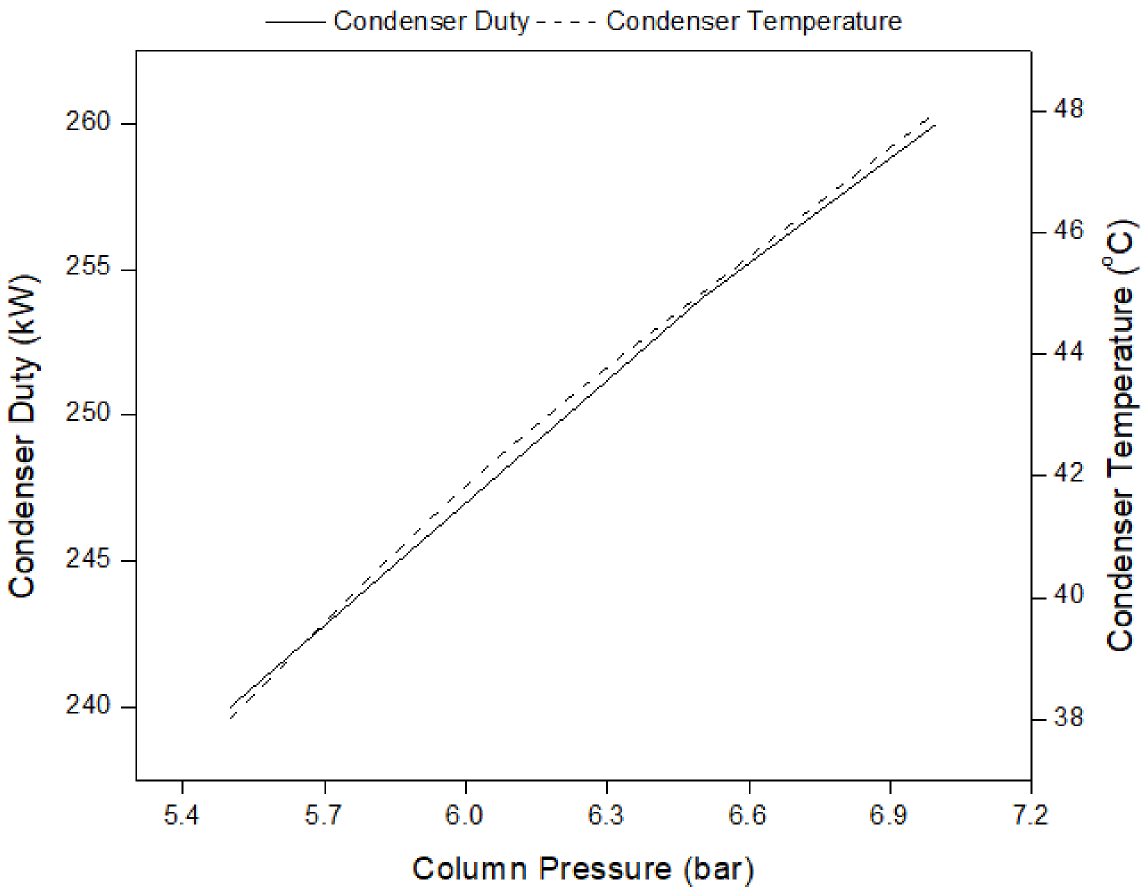

Figure 56 Effect of the Column Pressure on the Condenser Duty and Temperature in the Deisobutanizer

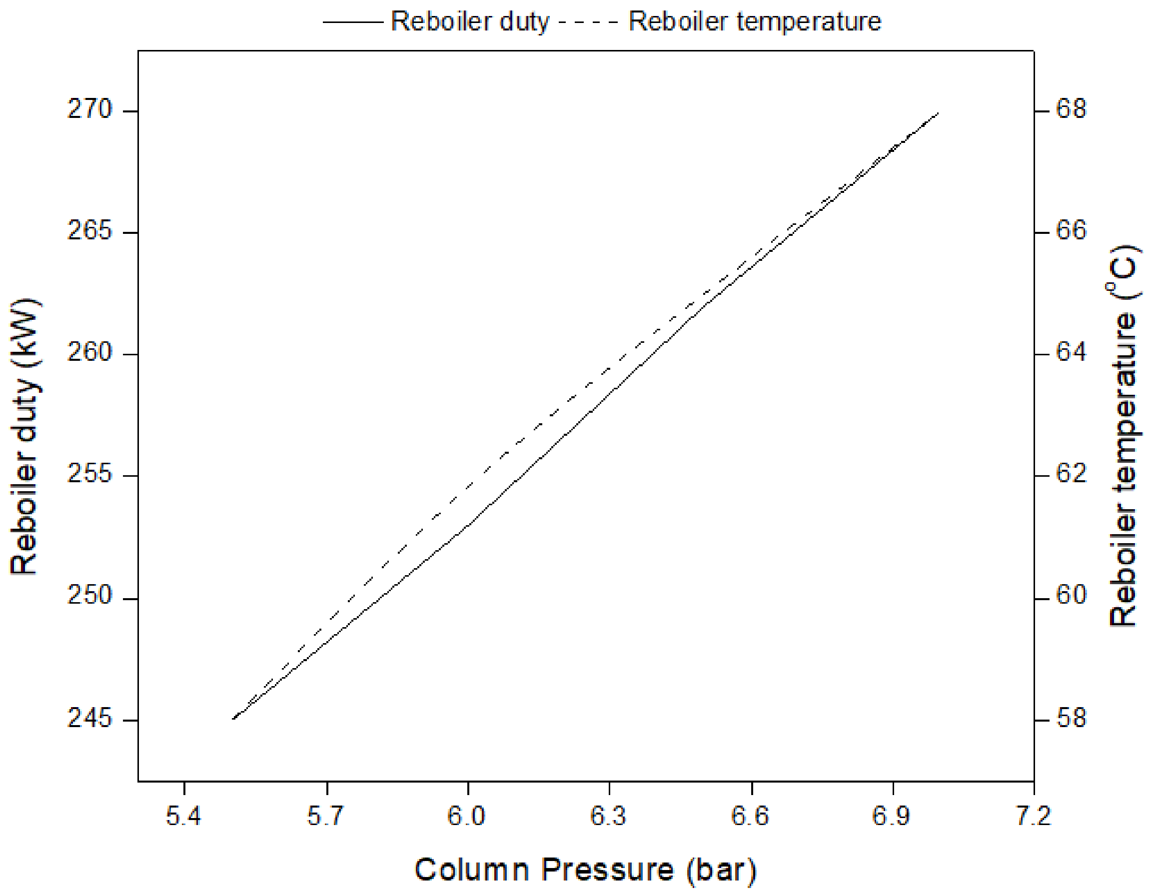

Figure 57 Effect of the Column Pressure on the Reboiler Duty and Temperature in the Deisobutanizer

5.4.1.3 Effect of Number of Column Trays

Number of column trays is another variable that can be manipulated to investigate its effect on the energy performance of the process. Fractionation columns stages values are selected within the range of typical values in the industry (Russell et al. 2004). As expected, in all columns, increasing the number of trays reduces both reboiler and condenser duties required to achieve the purities required as depicted in figures 58-61. However, in the deethanizer, about 80%-90% of the energy required can be reduced by increasing the number of trays from 28 to 35 as shown in figure 58 while lower amount of energy can be saved for the other columns. Another behavior that is observed during performing this analysis is that the reflux ratio decreases as the number of trays for each column increases as shown in figures 58-61.

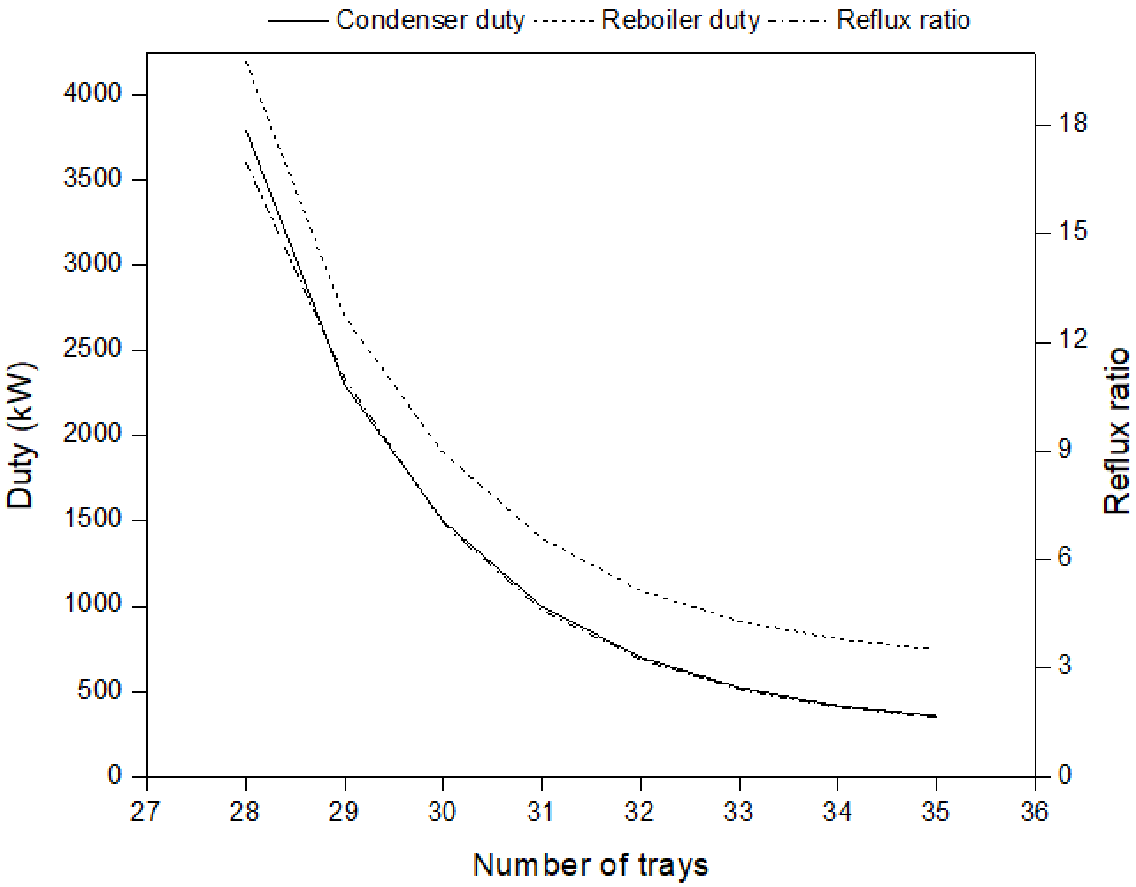

Figure 58 Effect of the Number of Stages on the Column Duties and Reflux Ratio in the Deethanizer

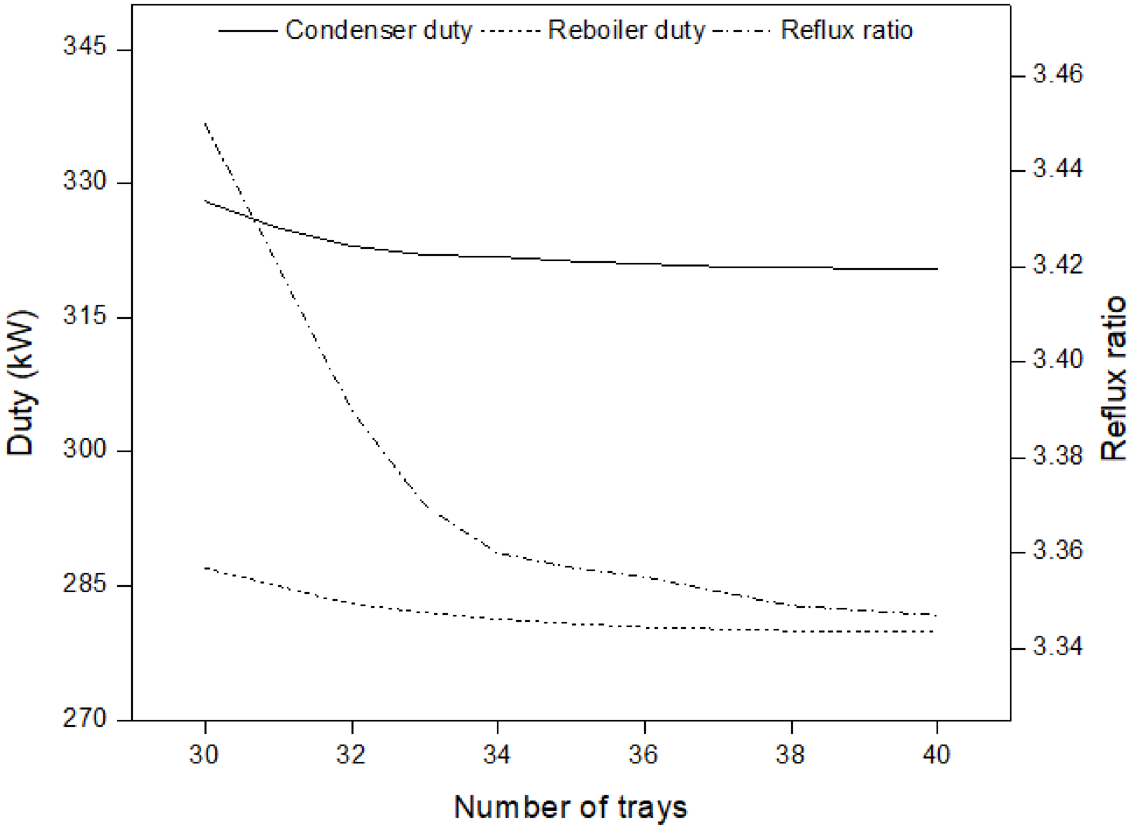

Figure 59 Effect of the Number of Trays on the Column Duty and Reflux Ratio in the Depropanizer

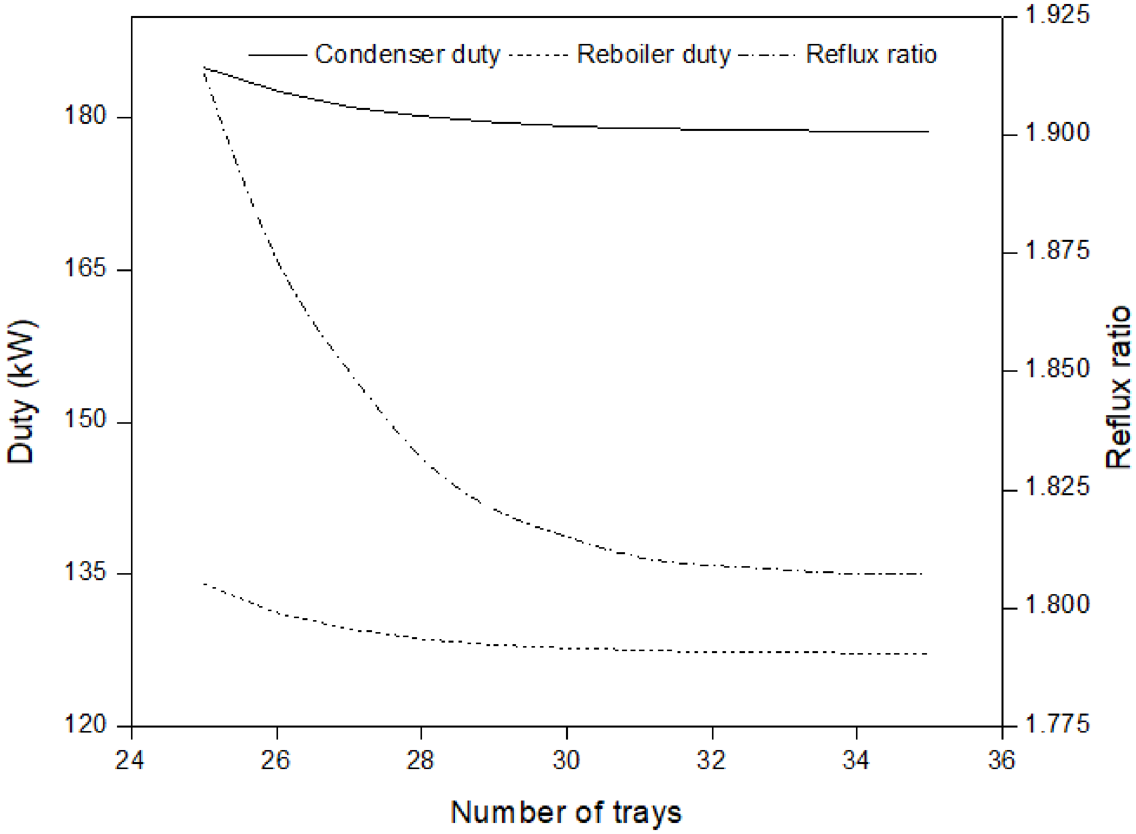

Figure 60 Effect of the Number of Trays on the Column Duty and Reflux Ratio in the Debutanizer

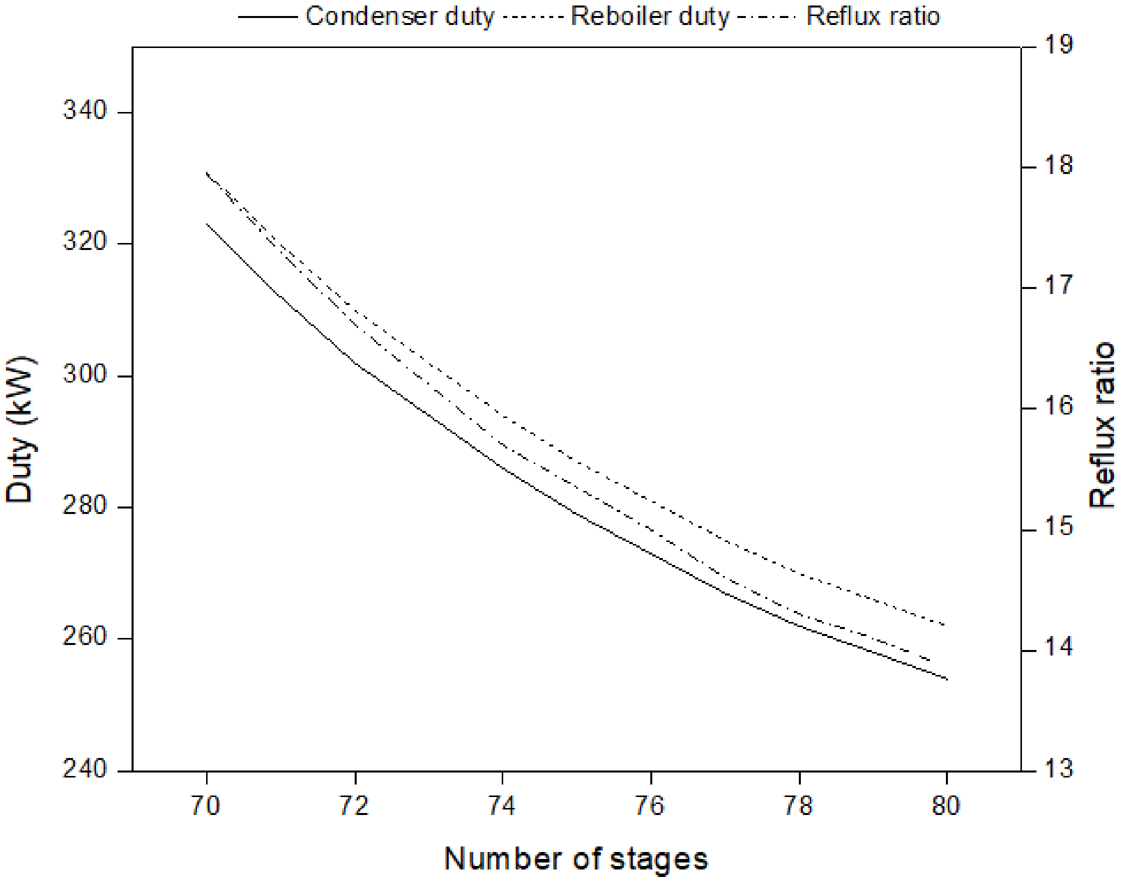

Figure 61 Effect of the Number of Trays of the Column Duty and Reflux Ratio in the Deisobutanizer

5.4.1.4 Effect of Product Purity

Product purity is the most crucial factor influencing the reboiler and condenser duties required to perform the separation. The purities of ethane and propane were manipulated in the deethanizer and deporpanizer, respectively while the purities of i-butane and n-butane are changed in the deisobutanizer. As expected, to achieve high purity of each component, higher energy must be supplied from the reboiler and the condenser as shown in figures 62-65.

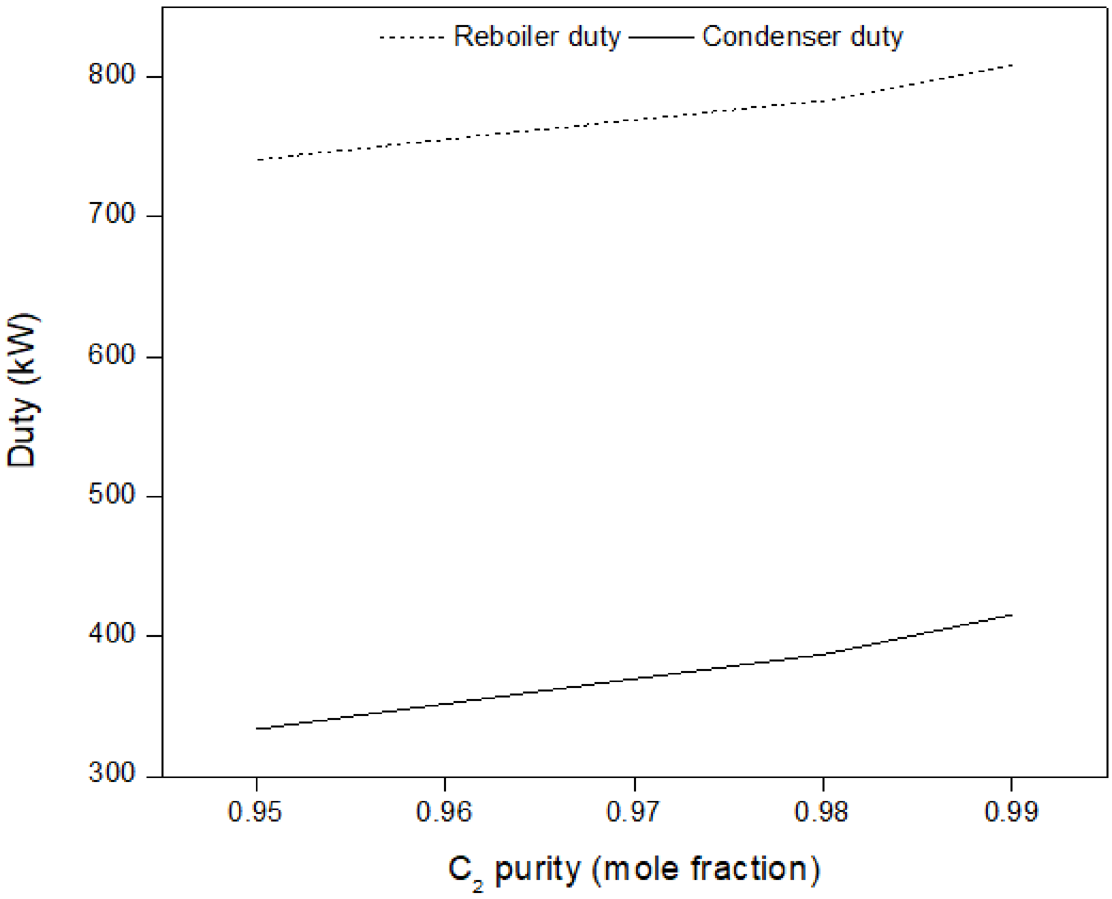

Figure 62 Effect of C_2 purity on the Condenser and Reboiler Duties in the Deethanizer

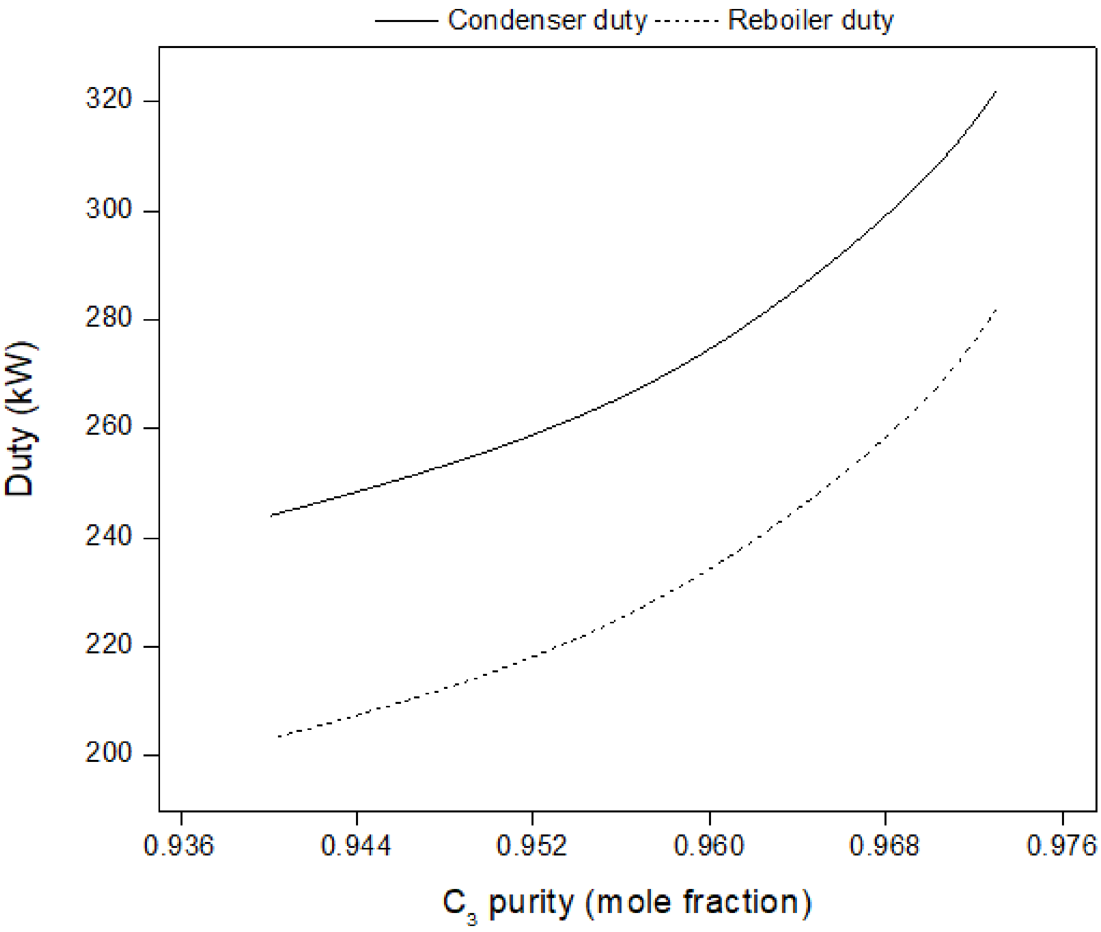

Figure 63 Effect of the C3 Purity on the Reboiler and Condenser Duties in the Depropanizer

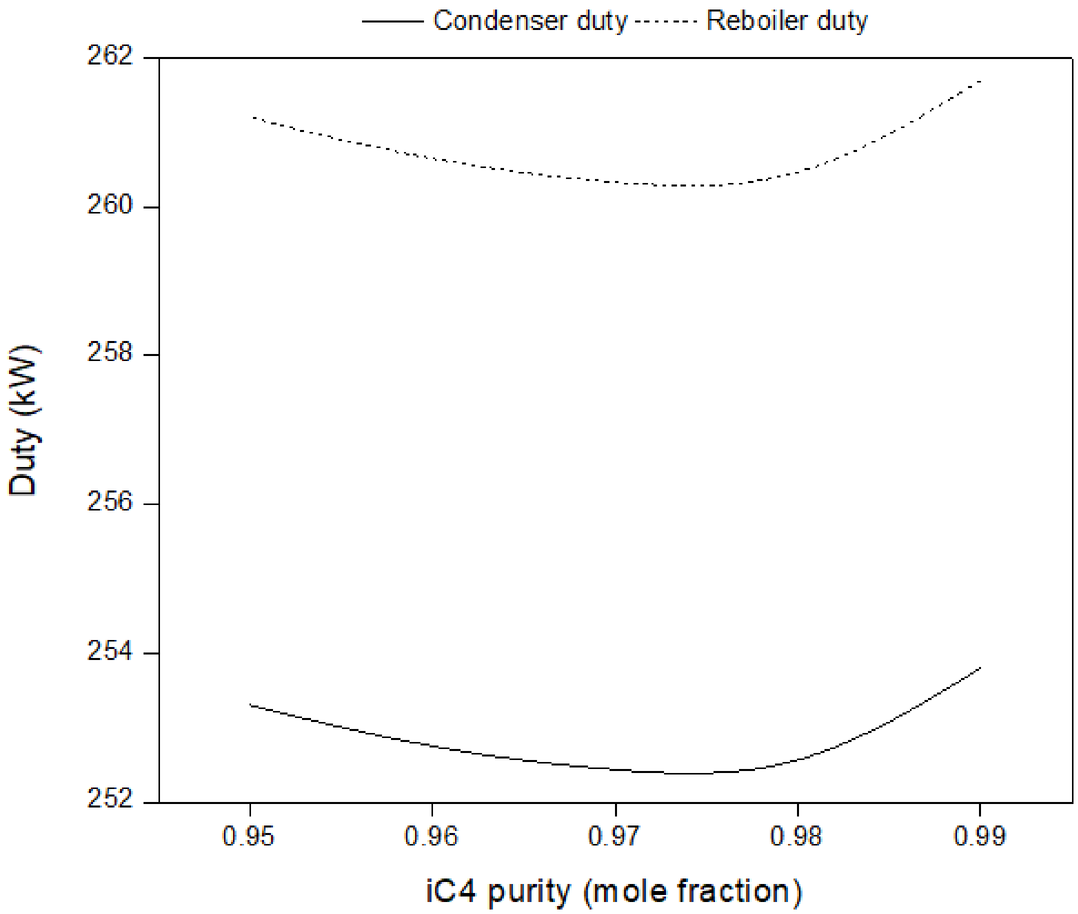

Figure 64 Effect of the iC4 Purity on the Reboiler and Condenser Duties in the Deisobutanizer

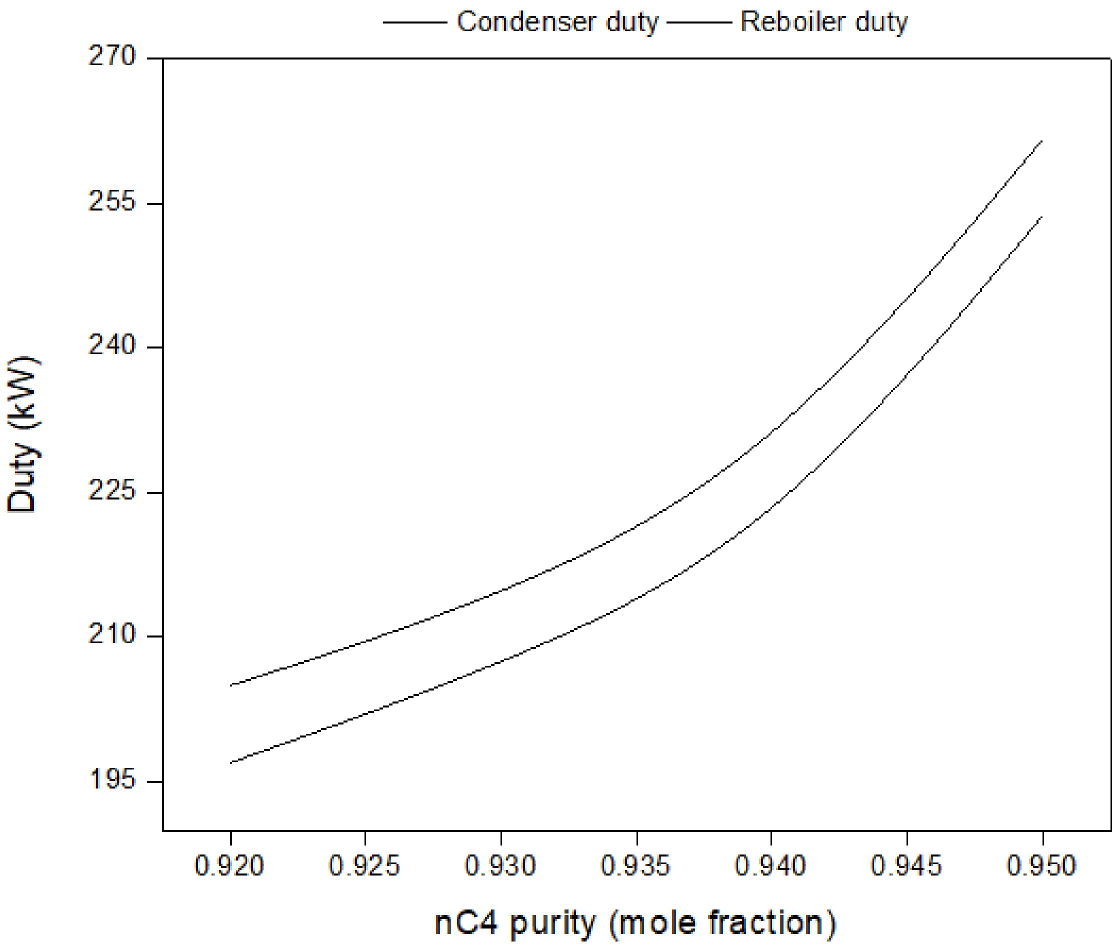

Figure 65 Effect of the nC4 Purity on the Reboiler and Condenser Duties in the Deisobutanizer

5.4.2 CDWC Sensitivity Analysis

5.4.2.1 Effect of Feed Temperature

The effect of feed temperature is studied as well in the dividing wall columns. The same criteria are used for this analysis to select the temperature range. However, different trend can be observed than the feed temperature study in the conventional columns. In both CDWC, an optimum temperature which gives the lowest reboiler and condenser duties is found as shown in figures 66 and 67. These values are about 30 ^{o}C and 65 ^{o}C in the first and second CDWC, respectively. Increasing the feed temperature beyond these values leads to an increase in the reboiler and condenser duties.

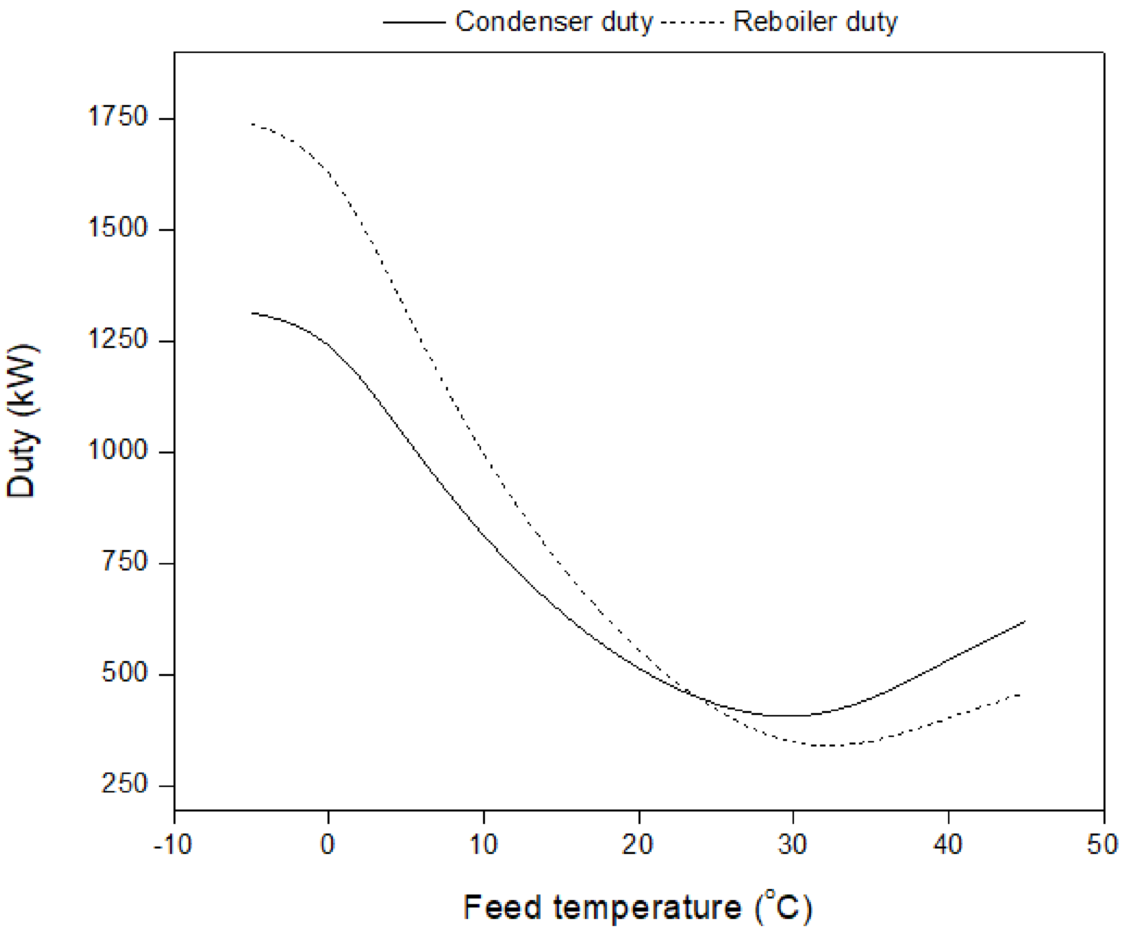

Figure 66 Effect of the Feed Temperature on the Reboiler and Condenser Duties in the First CDWC

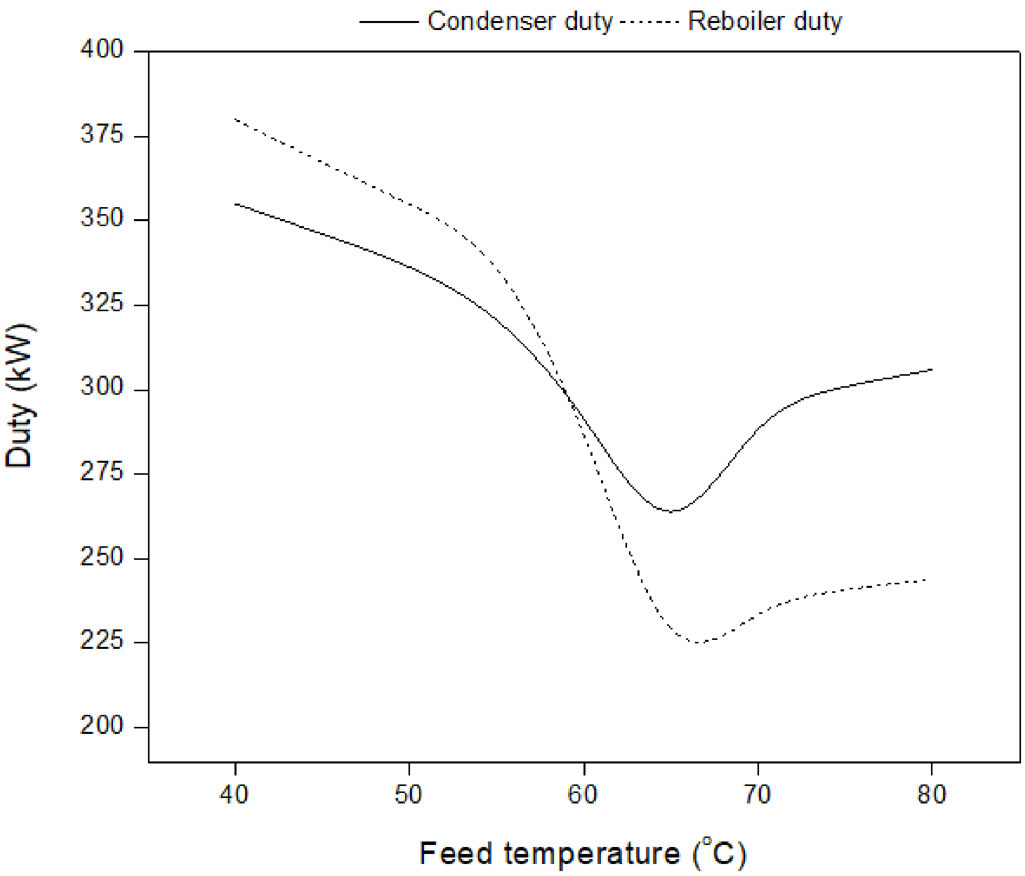

Figure 67 Effect of Feed Temperature on the Reboiler and Condenser Duties in the Second CDWC

5.4.2.2 Effect of Column Pressure

Operating pressure must be selected properly in the CDWC. High pressure operation in the CDWC will result in a high capital cost requirement since the high pressure operation needs thicker material of construction and the size of CDWC is larger than the conventional columns. Moreover, as approved in many studies and in the conventional sensitivity analysis section, Column pressure has a major influence on the reboiler and condenser duties. As shown in figures 68-71, the saturation reboiler and condenser temperatures increase as a result of increasing the column pressure. For the first CDWC, condenser duty increases linearly as the column pressure increases. Reboiler duty for the first CDWC shares the same trend with the reboiler and condenser duties in the second column where there is a minimum duty point can be found by increasing the operating pressure. These values are approximately 16 bar for the first column and 5 bar for the second CDWC.

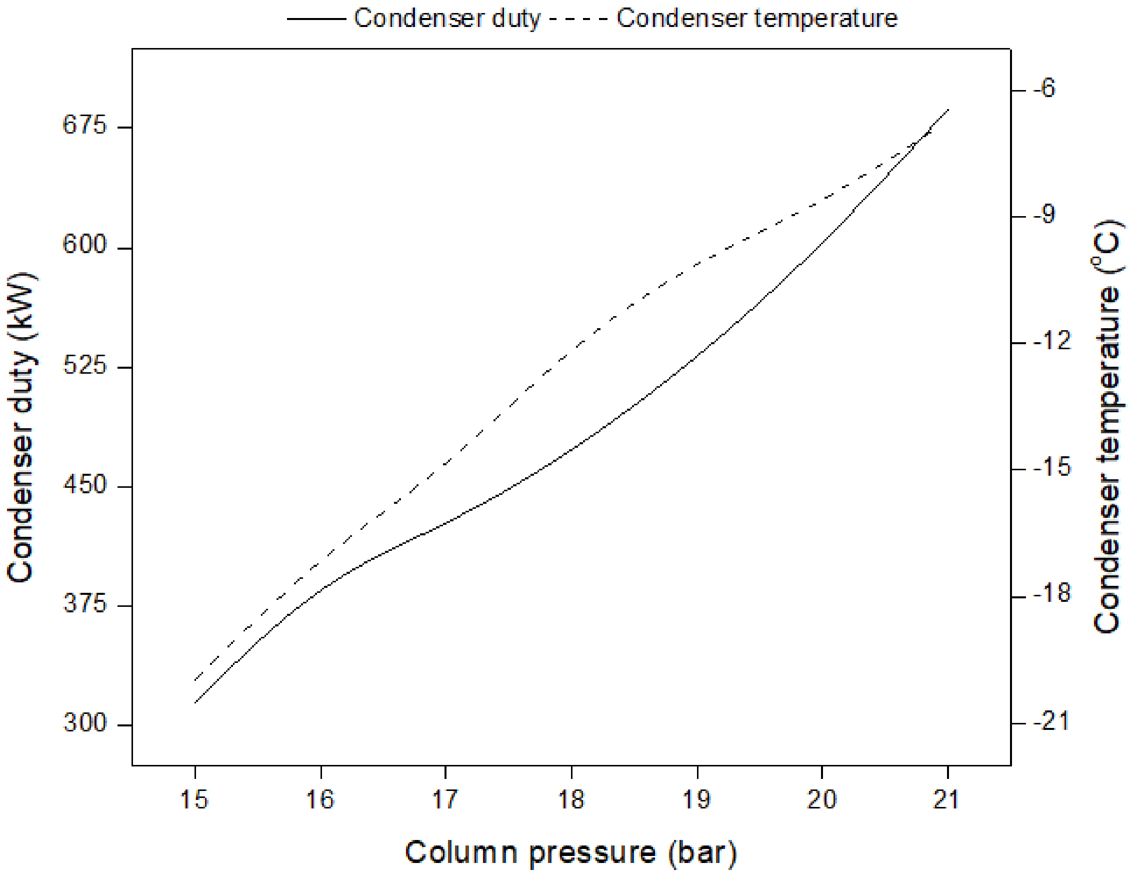

Figure 68 Effect of Column Pressure on the Condenser Temperature and Duty in the First CDWC

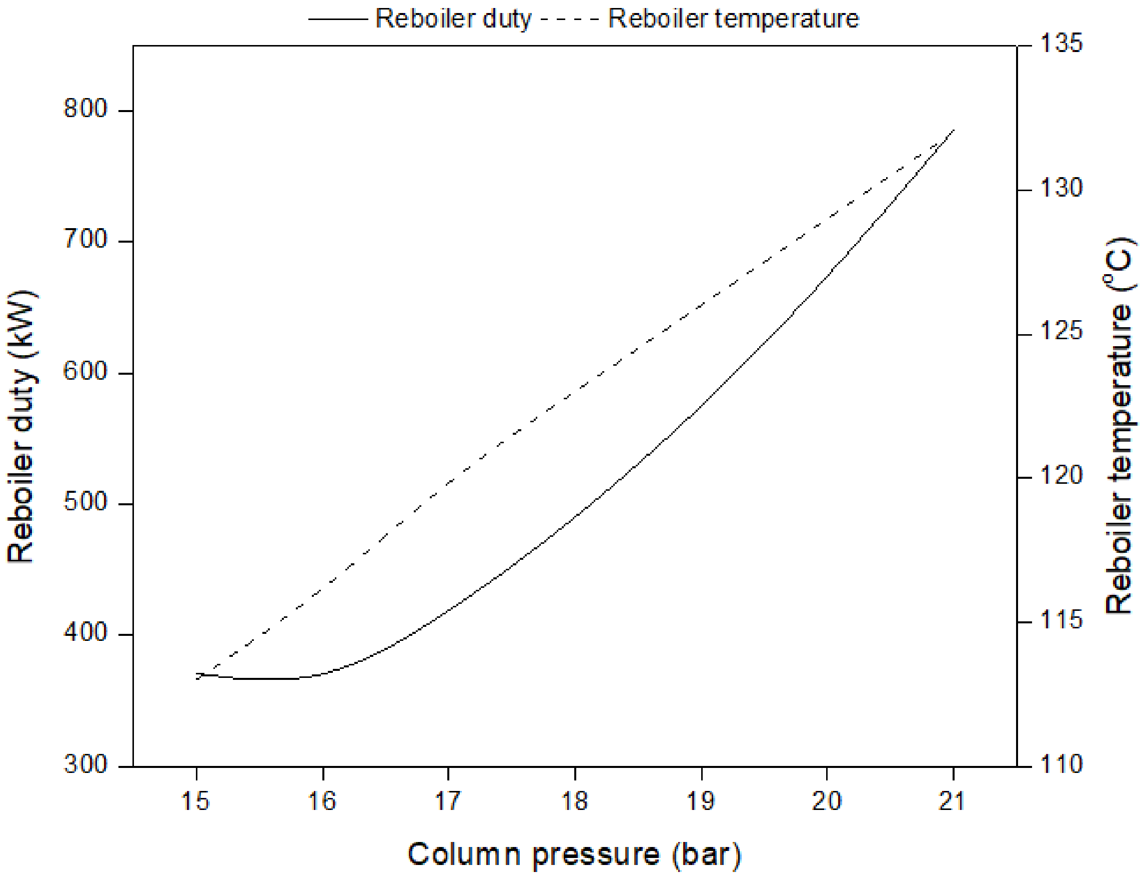

Figure 69 Effect of Column Pressure on the Reboiler Temperature and Duty in the First CDWC

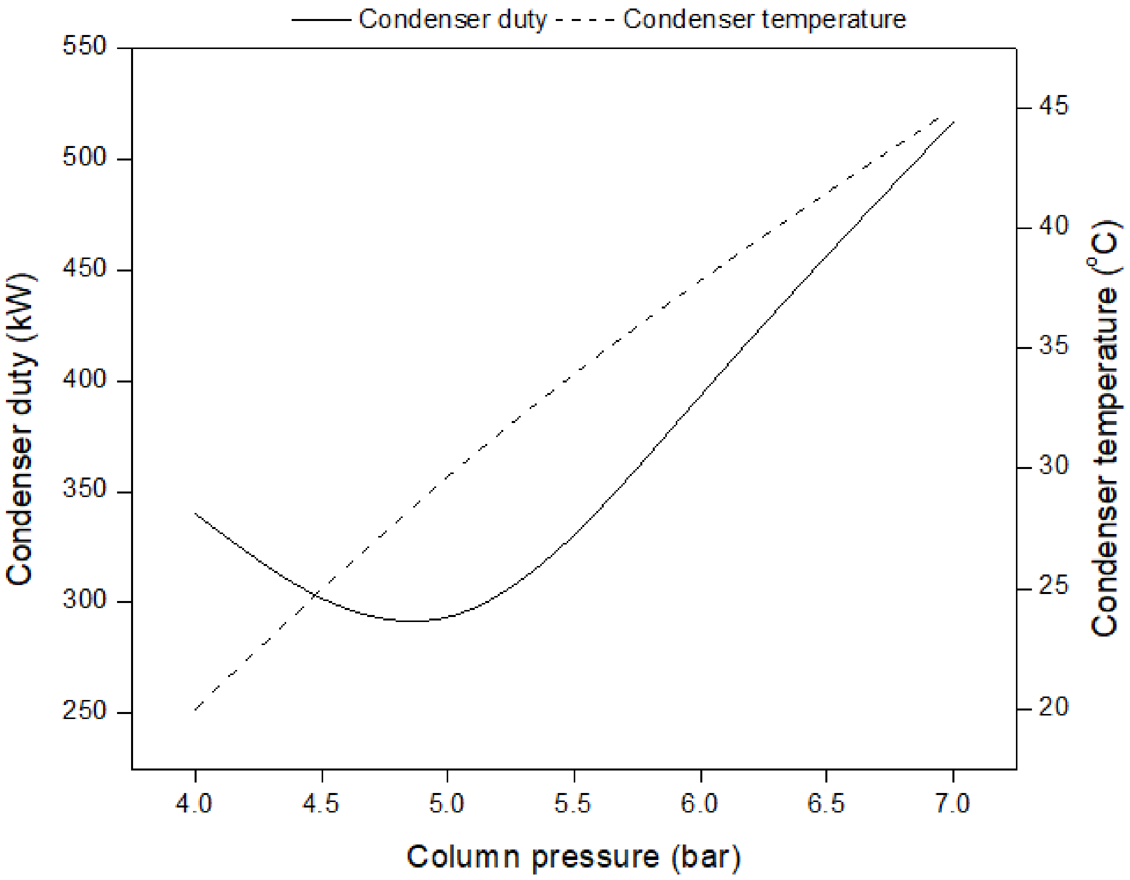

Figure 70 Effect of the Column Pressure on the Condenser Temperature and Duty in the Second CDWC

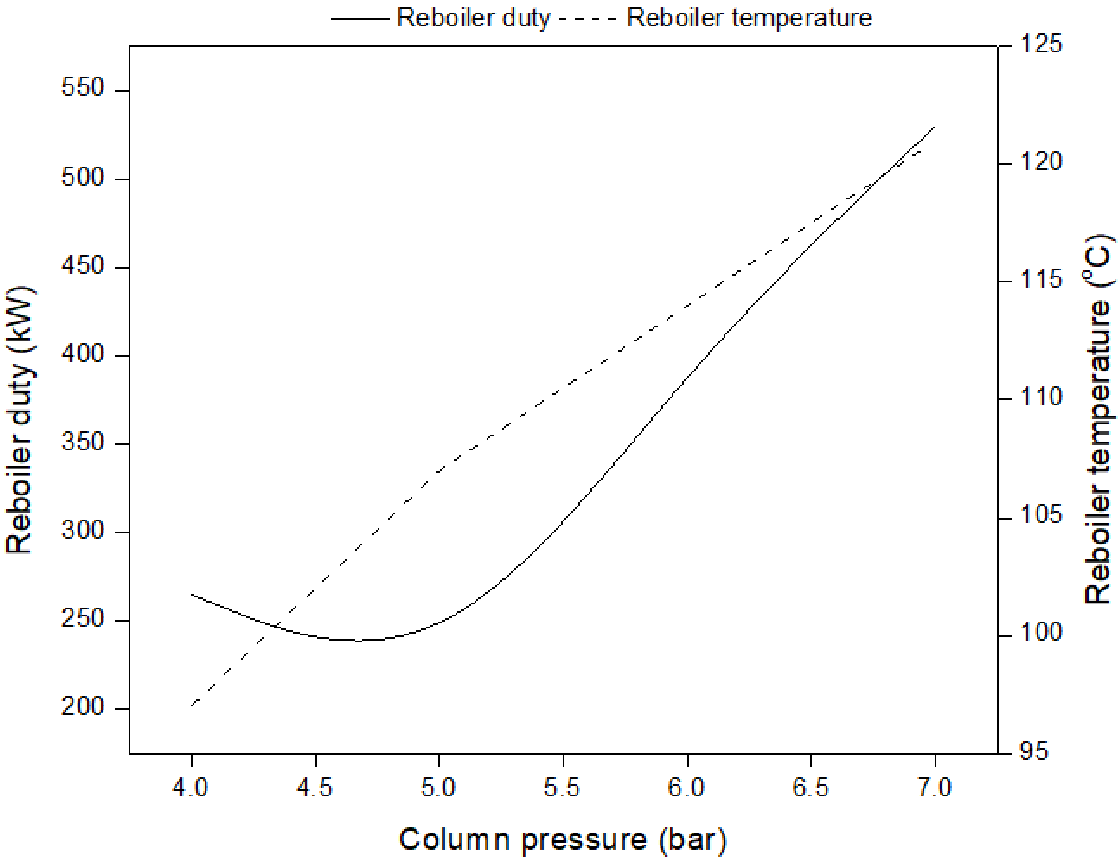

Figure 71 Effect of the Column Pressure on the Reboiler Duty and Temperature in the Second CDWC

5.4.2.3 Effect of Feed Stage Location

The feed is inserted in the prefractionator section (middle left) of the CDWC. The basis to select feed tray location is done usually by optimization. In the present study, sensitivity analysis was done to investigate the effect of changing feed tray location in the CDWC. Figures 72 and 73 showed the reboiler and condenser duties become minimum at the 7th tray for the first CDWC while 8th tray is the optimum location for the second CDWC. The reason for this behavior is the difference between the feed composition and the concentration profile inside the columns where lower differences can minimize the duties required to perform the separation.

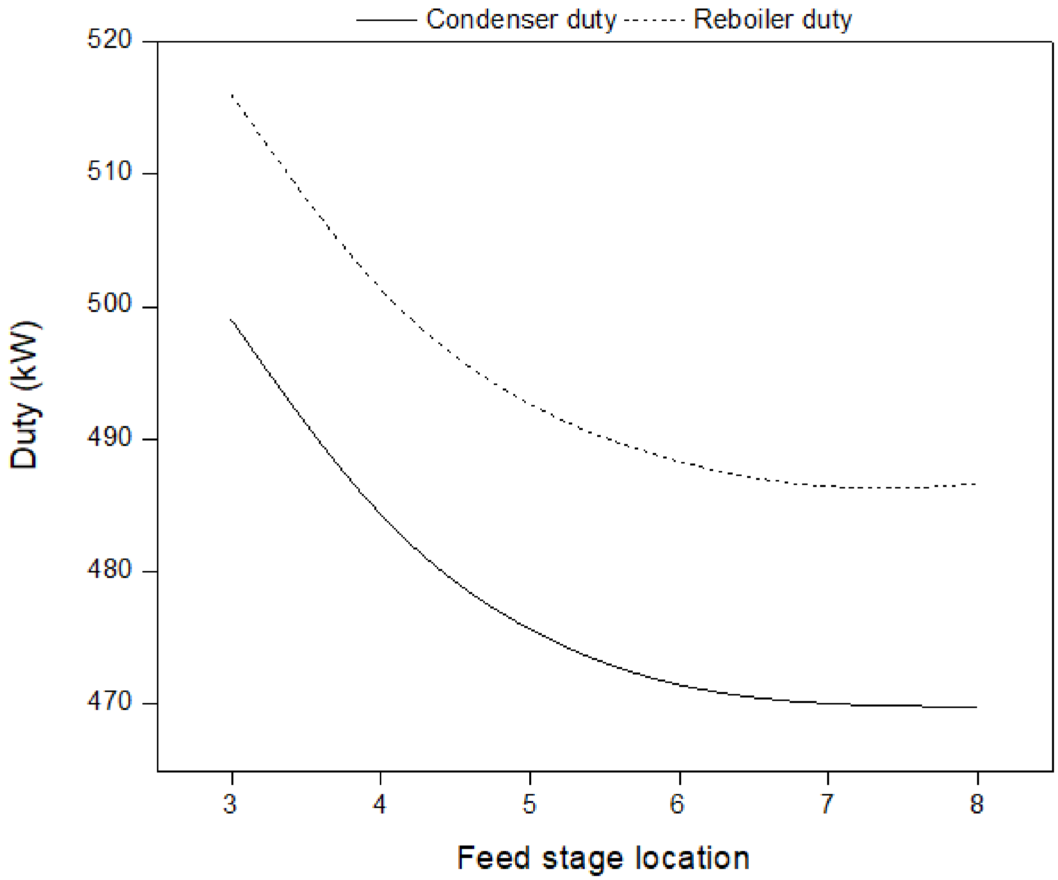

Figure 72 Effect of the Feed Stage Location on the Reboiler and Condenser Duty in the First CDWC

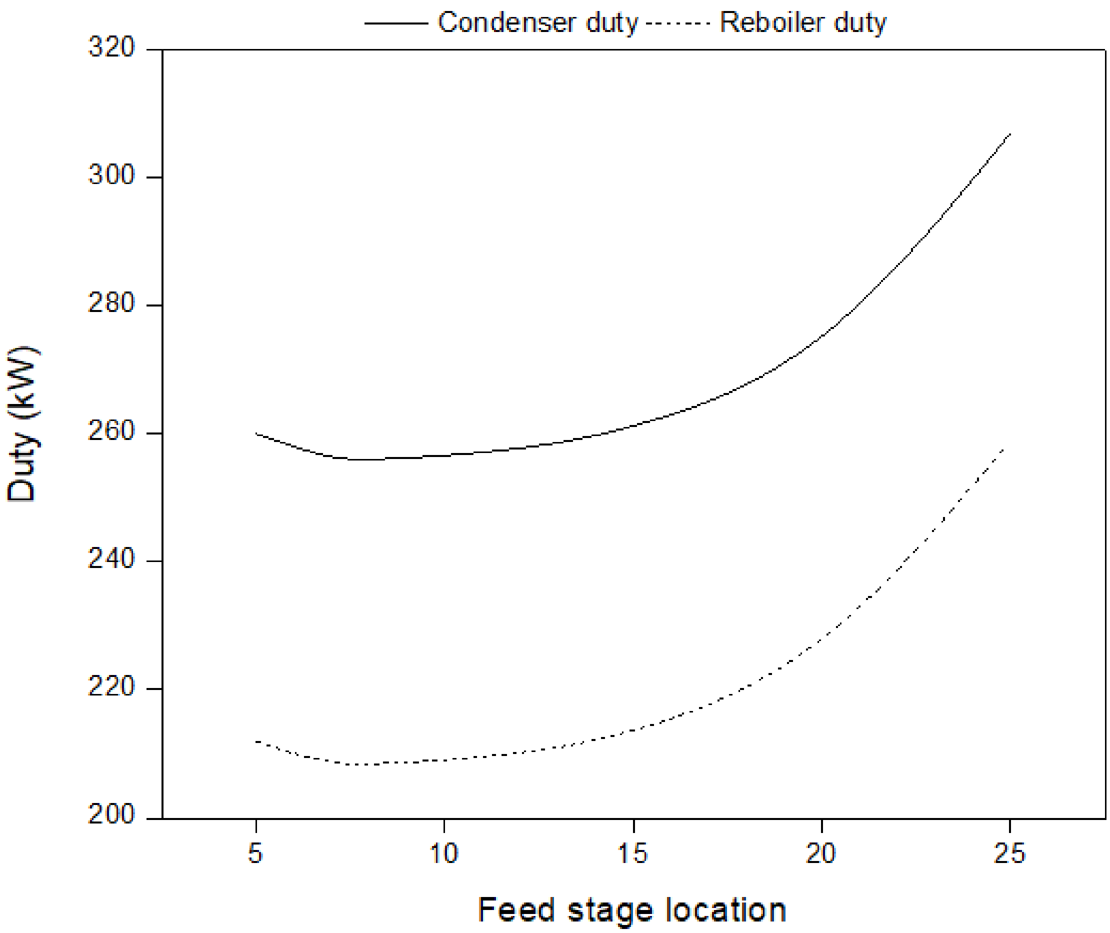

Figure 73 Effect of the Feed Stage Location on the Reboiler and Condenser Duties in the Second CDWC

5.4.2.4 Effect of Side Draw Location

Side draws location which the second lowest relative volatility product exits the CDWC is another variable that has been tested to observe its effect on the energy performance of both CDWC. The analysis results showed that withdrawing the side product from the 8th stage gives the lowest reboiler and condenser duties for the first column. The second CDWC gives better energy performance when the n-butane is withdrawn from stages closer to the bottom section of the column as shown in figures 74 and 75.

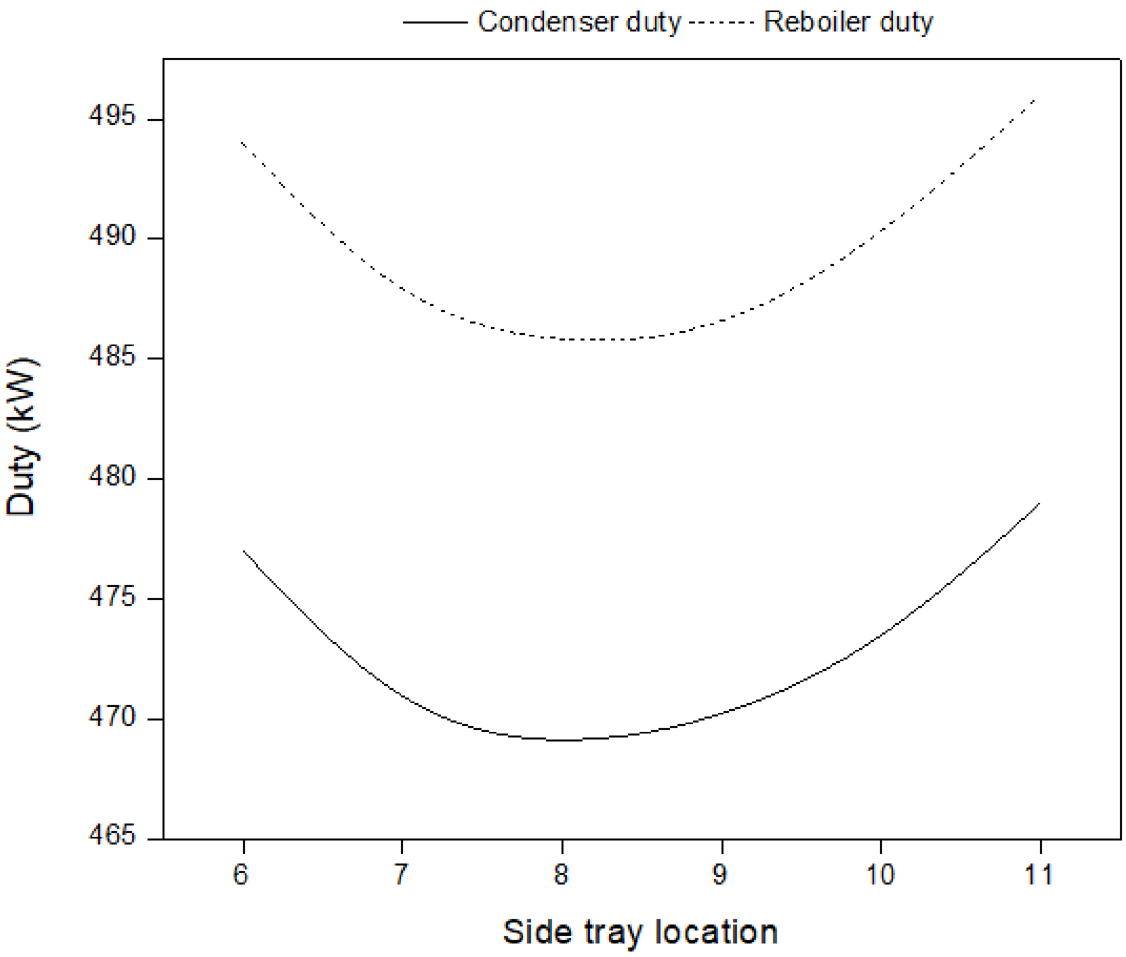

Figure 74 Effect of Side Tray Location on the Reboiler and Condenser Duties in the First CDWC

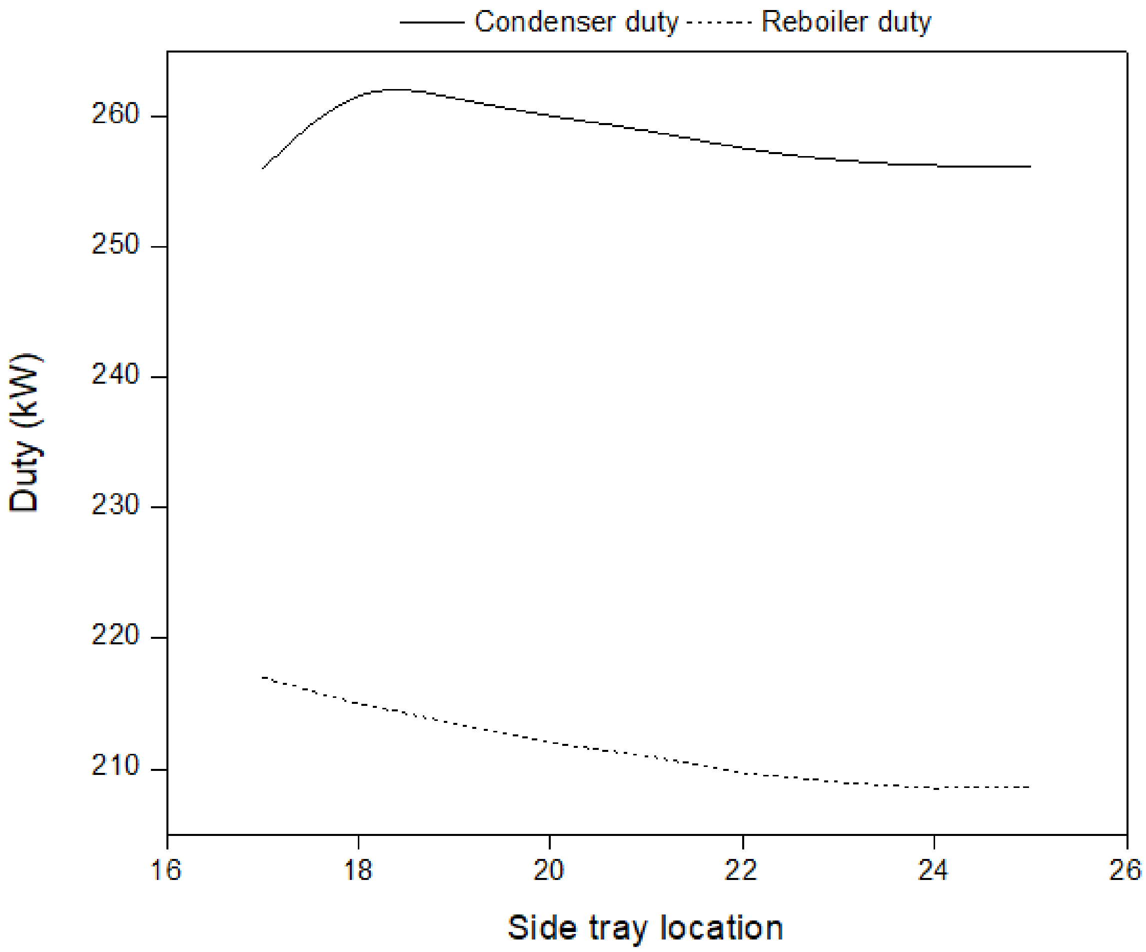

Figure 75 Effect of the Side Draw Location on the Reboiler and Condenser Duty in the Second CDWC

5.4.2.5 Effect of Split Ratio

Vapor split ratio that is coming from the bottom side of the column and liquid split ratio which is distributed from the top section of the CDWC are other unique parameters that can influence the energy performance of the column. Both split ratios are manipulated simultaneously to study their effect on the reboiler and condenser duties in the present study. Using equally split ratios from 0.3 – 0.7 requires lower duties than different ratios for the second CDWC as shown in figured 76-79. For the first CDWC, slightly higher amount of the vapor split ratio to the left side of the column along with equally distributed liquid split ratio leads to the lowest reboiler and condenser duties as depicted in figures 76 and 77.

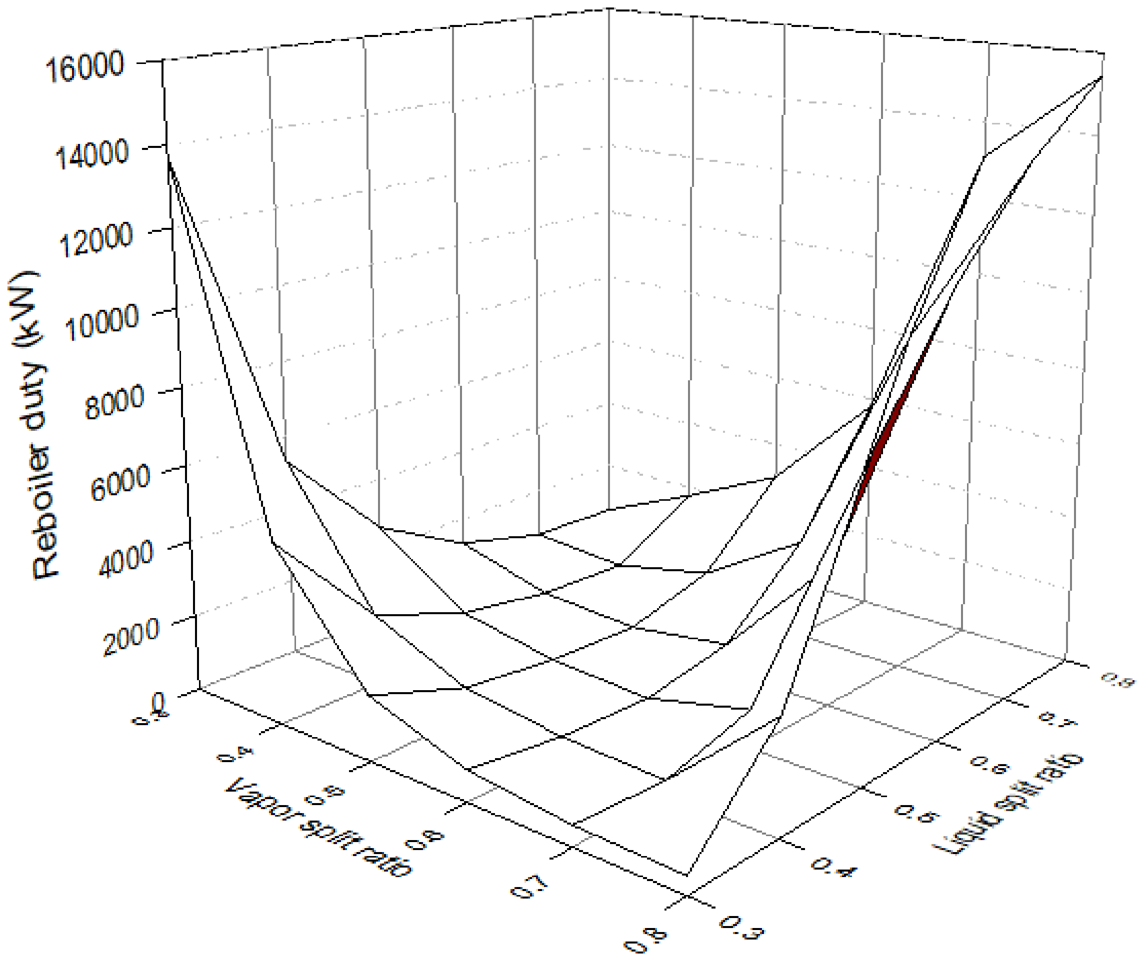

Figure 76 Effect of Split Ratios on the Reboiler Duty in the first CDWC

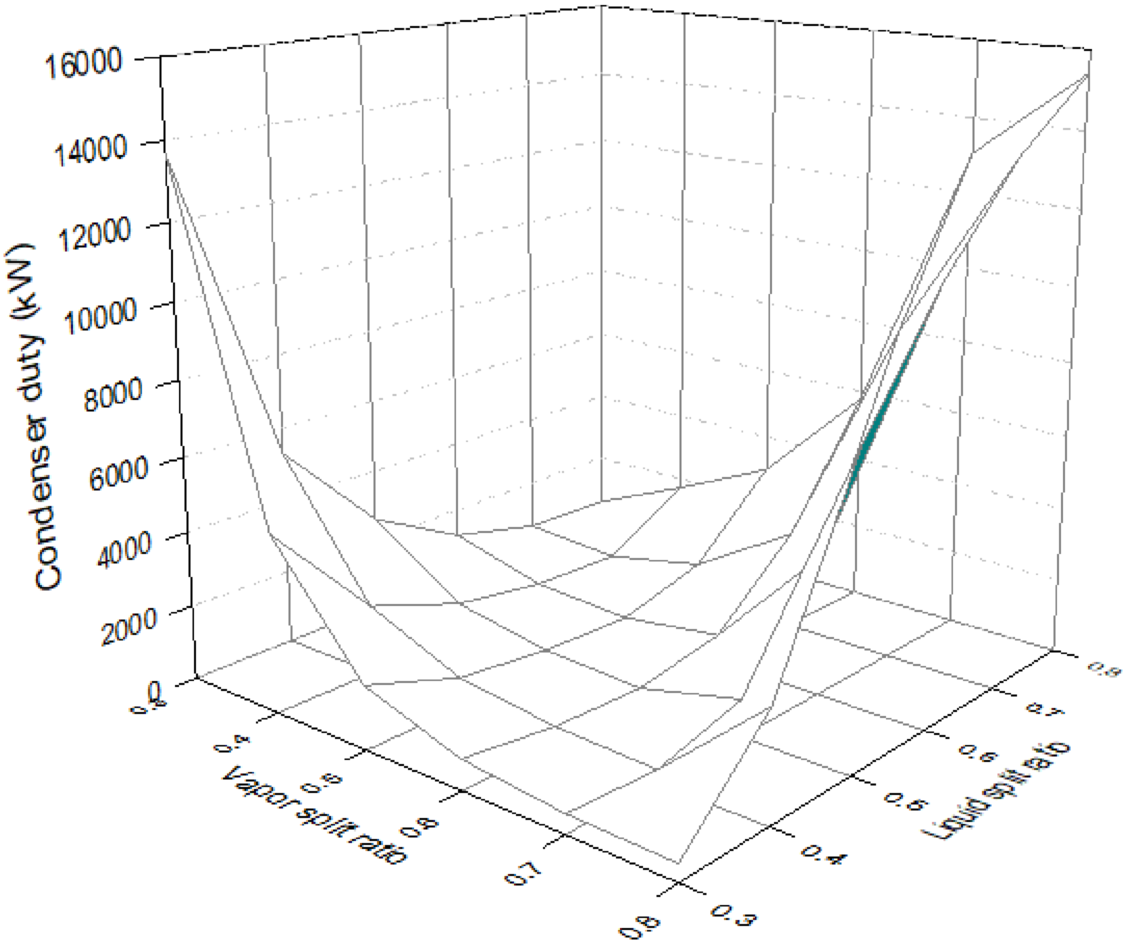

Figure 77 Effect of Split Ratios on the Condenser Duty in the First CDWC

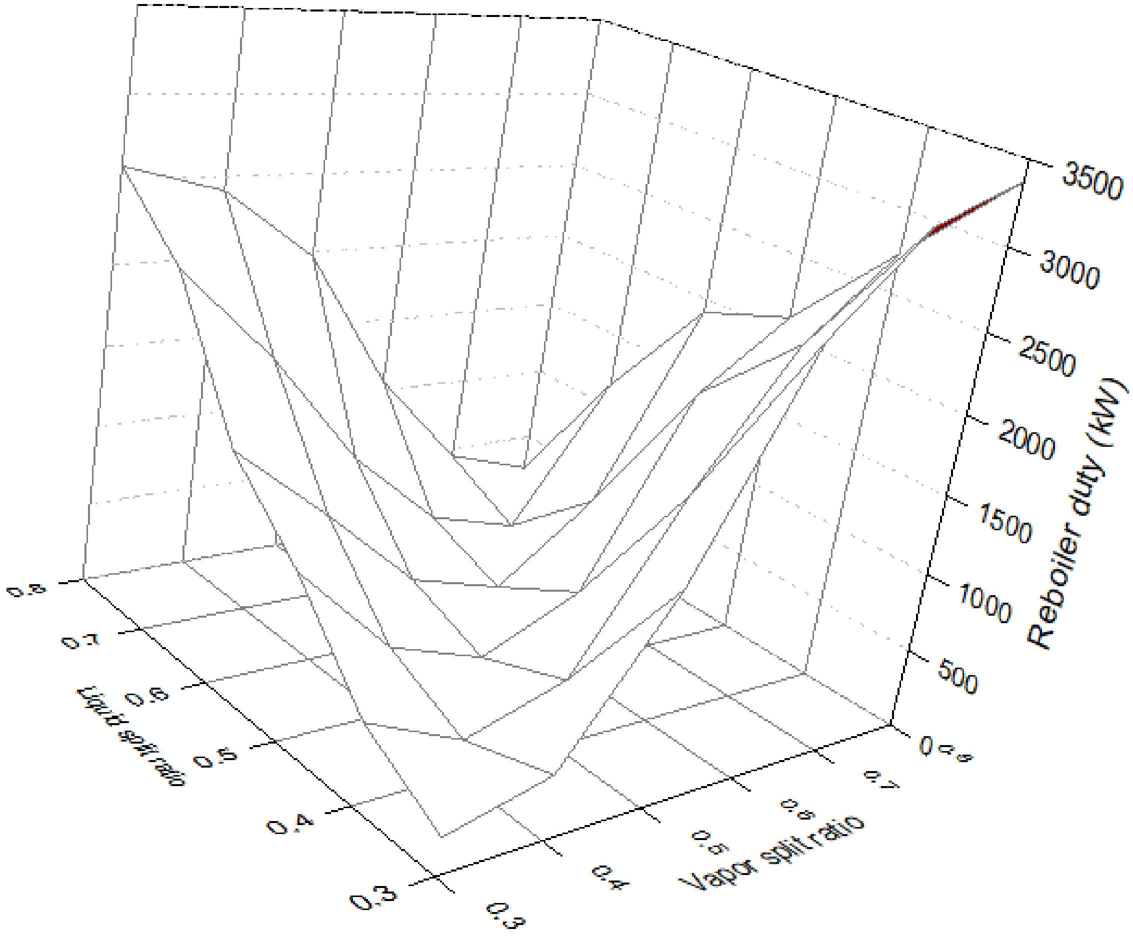

Figure 78 Effect of Split Ratios on the Reboiler Duty in the Second CDWC

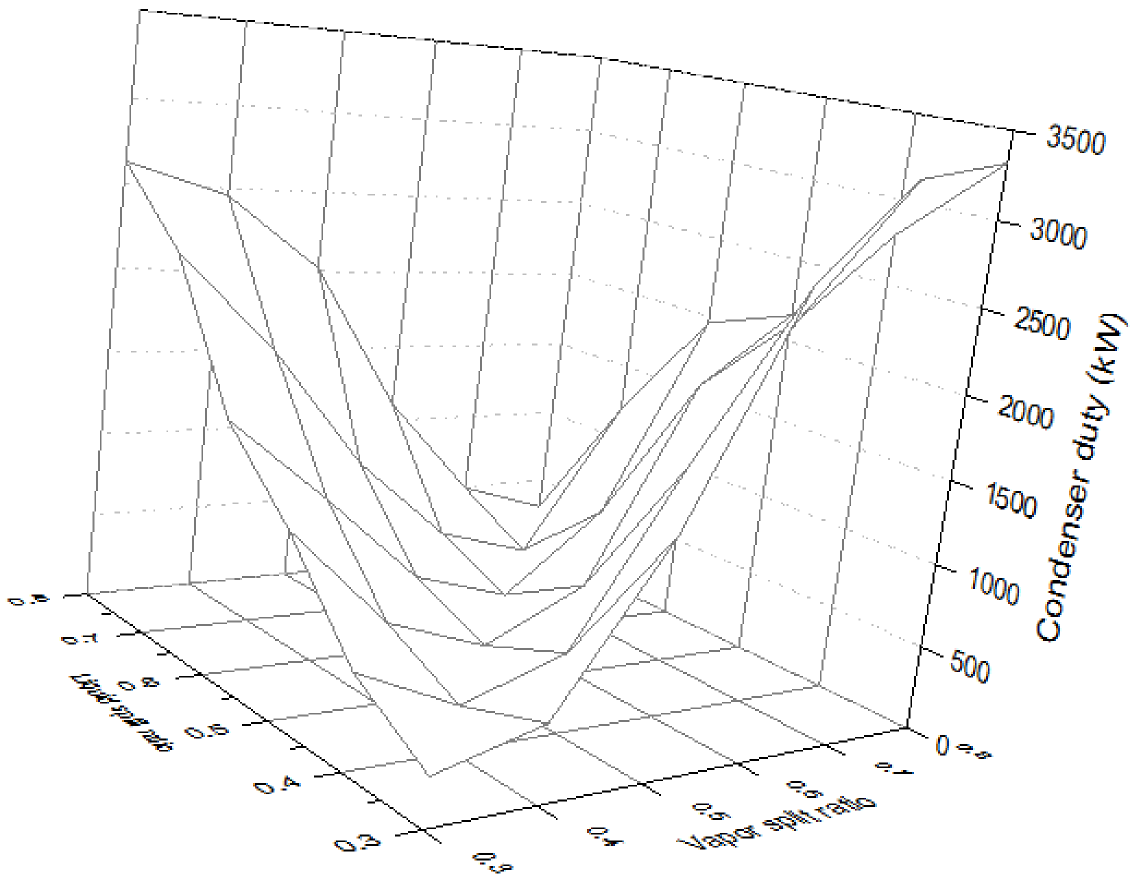

Figure 79 Effect of Split Ratios on the Condenser Duty in the Second CDWC

CHAPTER 6

CONCLUSIONS AND RECOMMENDATIONS

6.1 Conclusions

Three natural gas treatment processes were discussed and analyzed in the present work namely, acid gas removal, dehydration and NGL fractionation. Two plants that employed DGA and MDEA solvents were simulated in the acid gas removal section. The model results are in good agreement with the plant data so the model is assured to be reliable. MDEA tends to selectively absorb the H_2S while DGA can absorb both acid gases. Results showed that MDEA case required lower energy than DGA case. Also, several design alternatives had been investigated as well in order to check the possibilities to save the energy. As evident from the results, the MVR scheme has the highest capability to save energy while maintaining the same level of sweetening performance compared to conventional scheme. Key process parameters affecting the energy and removal efficiency, directly or indirectly, such as solvent circulation rate, solvent strength, inlet solvent temperature and stripper pressure can be tuned to improve the process performance.

Dehydration was the second part in this thesis and the goal of this part is to remove the water from the gas stream. The analogy between carbon capture and natural gas dehydration is investigated in two configurations, namely conventional and stripping gas. It was found that circulation rate and energy requirements can reduced by using stripping gas configuration for both cases. Moreover, TEG loss tends to be higher in the carbon capture case although the glycol flow rate required per water removed is lower in the carbon capture dehydration case. Sensitivity was also performed in this part to study impact of key parameters on the water removal and energy performance circulation rate, contactor temperature, stripping gas rate and reboiler temperature.

NGL fractionation is an energy intensive section in natural gas processing since it is done by a connected series of distillation columns. In this work, a conventional series had been simulated and dividing wall column sequence had been selected as a retrofit design to look for energy saving opportunities. According to our findings, 24% of the reboilers and condensers duties required in the conventional sequence can be saved using CDWC sequence. Parametric analysis was performed for the conventional and CDWC sequence to investigate the possibilities to reduce the energy requirements. The parameters are feed temperature, column pressure, product purities and number of trays and other unique parameters for CDWC such as split ratios and side tray location.

6.2 Recommendations

Further improvements can be done in this study and several future works can be investigated as well. The recommended aspects to be studied are the following:

1- In acid gas removal section, extending the investigation of the design alternatives to the other solvents such as MDEA and DEA can give better understanding for these alternatives.
2- Looking for methods to minimize the TEG lose in the dehydration unit especially for the carbon capture case.

3- Performing parametric analyses for different column configuration to detect more approaches to save energy and maximize the product purity.

4- Gathering more plant data to ensure the reliability of the models used in future studies.

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