



Study of Energy Consumption for CO₂ Removal Using Gas Twister Systems in Natural Gas Production and Related Attributes

Prepared by:
Muhamad Hidayat Kamalzaman
14397

Supervisor: AP Dr Usama Mohamed Nour El Demerdash

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Universiti Teknologi PETRONAS

Bandar Seri Iskandar

31750 Tronoh,

Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

**STUDY OF ENERGY CONSUMPTION FOR CO₂ REMOVAL
USING GAS TWISTER SYSTEMS IN NATURAL GAS
PRODUCTION AND RELATED ATTRIBUTES**

By

Muhamad Hidayat bin Kamalzaman
14397

A project dissertation submitted to the

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in partial fulfilment of the requirement for the
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Approved by,

AP Dr Usama Mohamed Nour El Demerdash
Universiti Teknologi PETRONAS
Bandar Seri Iskandar
31750 Tronoh,
Perak Darul Ridzuan

CERTIFICATION OF APPROVAL

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

Muhamad Hidayat bin Kamalzaman

Abstract

Out of the total world natural gas reserves, more than 40% of it are estimated to be sour that is high in hydrogen sulfide and carbon dioxide. Sour natural gas poses bigger challenges in harvesting and processing it. In this study, steady state simulations of the natural gas production facility that uses supersonic gas separation technology of gas twister are developed using Aspen HYSYS. The simulations are used to study the total energy consumption of the topside facility that uses the Gas Twister system. The amount of hydrocarbon lost from the whole system is also being paid attention to. The total energy consumption is studied by varying the CO₂ loading of the natural gas and the arrival conditions of natural gas. Four case studies are conducted in order to achieve the objectives of the research. The results of this research show some notable trends that are crucial in the removal of CO₂ from sour natural gas. The production facility uses less energy for a high CO₂ concentration natural gas and the arrival temperature and arrival pressure of the natural gas affects the work produced by the turbo expander.

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Chapter 1: Introduction

1.1 Background

According to Society of Petroleum Engineers (2014) more than 40% of world gas reserves are estimated to be in reservoirs that have high content of hydrogen sulfide, H₂S and carbon dioxide, CO₂. The presence of the two components could be as high as 70% and their presence poses challenges for drilling, cementing, completion, production and intervention.

Let alone in Malaysia, it is approximated that up to 13 trillion cubic feet of high CO₂ – Natural Gas fields remain undeveloped in Malaysia. The development of CO₂ recovery from natural gas technologies will enable monetization of high CO₂ natural gas. The ability of natural gas to offer energy saving benefits instead of oil and coal has made natural gas the most energy-efficient fuel source. Besides being used primarily as fuel, natural gas is a hydrocarbons source for petrochemical feedstock. It is also a major source of elemental sulphur. The popularity of natural gas is increasing drastically since it can achieve two significant energy goals – ability to provide sustainable energy supplies and ability to reduce environmental effects. (Mokhatab & Poe, 2012a)

The composition of raw natural gas varies from well to well (Baker & Lokhandwala, 2008). Generally, 30% - 90% of natural gas is made up of methane with other light hydrocarbons i.e. ethane and propane and other heavier hydrocarbons. Apart from hydrocarbons, natural gas also contains carbon dioxide, hydrogen sulfide and nitrogen with varying composition (Scholes, Stevens, & Kentish, 2012). Traces amount of metallic substances – arsenic, selenium, mercury and uranium are also known to exist in raw natural gas. (Mokhatab & Poe, 2012a)

The composition of CO₂ in natural gas varies from 4% to 50%. Natural gas from CO₂ gas re-injected EOR (enhanced oil recovery) well can contain up to 90% CO₂. Before transporting the natural gas, the gas must be treated to meet the typical export gas specification of 2%-5% CO₂ (Datta & Sen, 2006). Today, many natural gas reserves are left undeveloped due to their low production rate and low quality (high CO₂ content).

This situation has led to the needs of developing efficient process of CO₂ separation from natural gas (Ahmad, Lau, Shariff, & Murshid, 2012).

The main challenge in natural gas treatment processes is to remove the acid gas as much as possible so that it is consistent with emission regulation and to treat the natural gas as economical as possible. Acid gas such as CO₂ can cause severe damage to the pipelines if not handle properly. A number of methods are available and being practiced in industry for the removal of carbon dioxide (Mokhatab & Poe, 2012c).

The processing of high CO₂ content gas requires advancement of technology in pretreatment, separation, utilization, transportation and storage. Indirectly, this increases the cost of obtaining the natural gas.

1.2 Problem Statement

PETRONAS is currently evaluating the possibilities of extracting gas from several high CO₂ gas field developments in both Peninsular and East Malaysia. It is estimated that the gas content for those fields ranges from 20 mol% to 70 mol%.

In order to monetize the gas fields, the carbon dioxide from the natural gases must be removed. Several methods have been used widely in the industry in order to combat the high CO₂ content in the natural gases. Some of the methods are being explained in Chapter 2. PETRONAS is considering using Gas Twister system on its natural gas production facilities to remove the carbon dioxide. Gas Twister (a supersonic separation method) is a proprietary technology owned by TwisterBV.

The Gas Twister technology is known for its low energy consumption and compact design. This research is aimed to investigate the effect of the variation in process parameters on the total energy consumption of the offshore gas production facilities. Hence, this research will identify and highlight the factors that influence the total duty or power consumption of the production facility that uses Gas Twister system.

Energy consumption or total duty is an important aspect in a process operation. Higher duty requirement means that bigger equipment (pump or compressor) are needed for increasing pressure of a liquid or a gas stream. Equipment with high duty will cost more and the operation cost are expected to be higher since more energy is required. In an offshore production facility energy consumption is a big concern.

Amongst the challenges that is expected in this research is the lack of data on the real natural gas composition which is most of the time are confidential information. The biggest constraint in this study is the unavailability of proprietary unit – Gas Twister, Liquid Twister and etc in the process simulator, Aspen HYSYS. They have to be modelled using typical equipment in Aspen HYSYS in order to represent them. Another constraint to this study is the limited literature available on the study on Gas Twister system.

1.3 Objective

The main aim of this study is to develop a steady-state simulation for the gas production facility that uses the Gas Twister system. The process simulations are then used to study the following objectives;

- i. To study the effect of CO₂ loading on total energy consumption.
- ii. To study the effect of different natural gas composition on total energy consumption
- iii. To study the effect of inlet condition in terms of temperature and pressure on the total energy consumption.

1.4 Scope of Study

This research covers the whole topside natural gas production facility. It is based on a real high CO₂ gas field offshore of Sarawak. The Aspen HYSYS simulations of the production facility are developed subject to the following.

Process Boundary:

This research will be focusing the offshore topside facilities of a natural gas production system. The boundary of the system will be from the well head to the export gas stream. However, the utility systems (propane cooling system, heating system and other basic utilities) are not being considered in the study.

Variables

The CO₂ loading, concentration of C1(methane), C2(ethane), C3(propane) and C4(butane), arrival conditions of the natural gas in terms of pressure and temperature.

Chapter 2: Literature Review

2.1 Natural Gas Processing

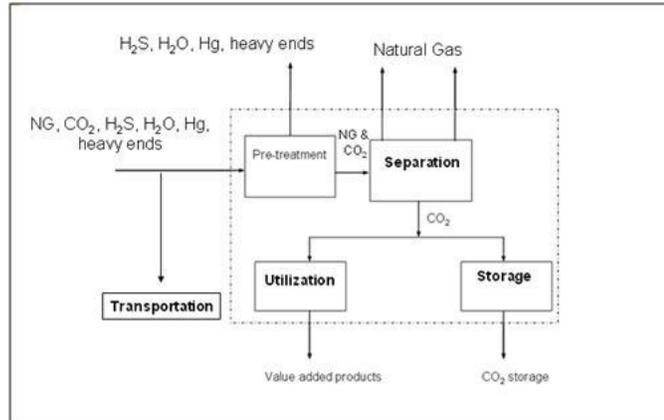


Figure 2.1-1 Carbon Dioxide - Natural Gas Management. (Manning & Thompson, 1991)

Manning and Thompson (1991) define oilfield processing as the processing of oil and/or gas for safe and economical storage and/or transport by pipeline, tanker or truck. Water treatment, whether produced waters for disposal and/or reinjection is a part oilfield processing as well.

Three main objectives of natural gas processing are (Manning & Thompson, 1991):

- To produce salable gas stream that meets the typical export specifications
- To maximize natural gas liquid recovery
- To deliver commercial gas

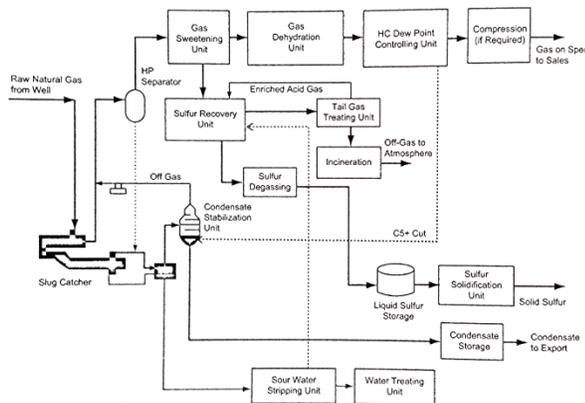


Figure 2.1-2 Typical setup of gas processing plant producing sales gas (Mokhatab & Poe, 2012b).

Gas production facilities or topside facilities are varying in shape and in size – from a basic gas receiving and letdown terminal to a complex processing facility with natural gas liquid (NGL) and sulfur recovery packages. Gas processing facilities and topside facilities are used interchangeably in this report. Figure 2.1-2 shows a typical process flow scheme for a gas processing facility. The figure shows several processing modules which could be made up of one single equipment or a group of equipment. All the modules shown are not necessary will be present in every gas plant. The actual process needed by a gas processing facilities depends on the feed gas composition and sales specifications for both the gas and liquid products (Mokhatab & Poe, 2012b).

For the sake of simplicity a typical gas processing facility will have the following processing operation modules (Manning & Thompson, 1991):

Table 2.1-1 Gas Processing Plant Operation Module (Manning & Thompson, 1991).

Operation Module	Function
Cleaning	Removing liquids and solids such as sand, pipeline dirt, reservoir fines, corrosion products and inhibitors, liquid or free water, salt and drilling mud
Treating	Sweetening or removal of acid gasses (H ₂ S and CO ₂)
Dehydration	Drying or removing water vapor or controlling H ₂ O dew point
Hydrocarbon Dew Point Control	Recovery of ethane and heavier hydrocarbons as condensate

2.2 Carbon Dioxide Recovery (Gas Treating)

Sour gas – carbon dioxide CO₂ and hydrogen sulfide, H₂S can lead to extensive damage to pipeline if not properly processed. Presence of CO₂ in fuel gas is undesirable because it is nonflammable and will reduce fuel gas heating value. With water CO₂ will form a weak, corrosive acid. Hence, these are among the reasons why CO₂ removal from natural gas is mandatory (Mokhatab & Poe, 2012c). In this section different method of CO₂ recovery is being compared highlighting Gas Twister system advantages over the others.

2.2.1 Absorption processes

Absorption is one of the most common CO₂ recovery methods in natural gas processing. In normal practice, two towers are used. In the first tower, the feed gas will enter at high pressure from the bottom of the tower. Absorbent liquids will be flown counter current of the natural gas. The absorbent liquids will absorb CO₂, water and heavy hydrocarbon are collected at the bottom of the first tower. The collected liquid is heated and sent to a low pressure stripper for absorbent liquids regeneration. (Baker & Lokhandwala, 2008)

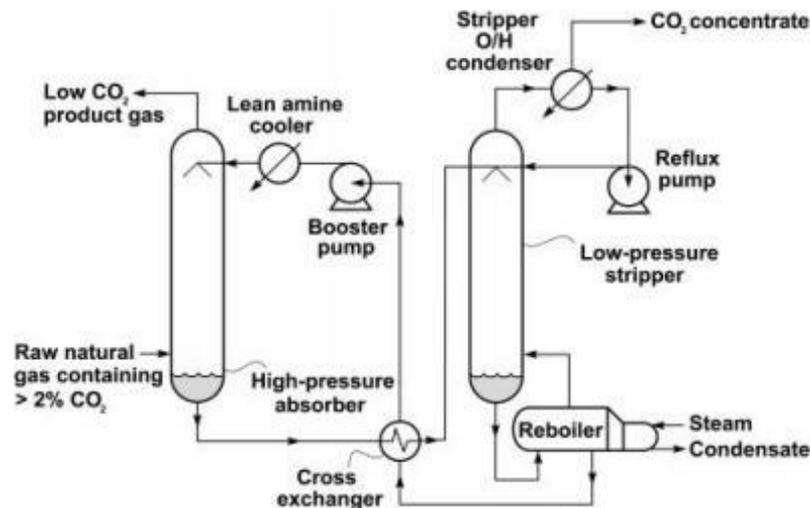


Figure 2.2-1 Typical natural gas absorber-stripper treatment process using amine absorbents (Baker & Lokhandwala, 2008)

The needs of using high pressure tower is expensive since the tower is large, thick walled and heavy vessel. The size of the tower will increase with large natural gas flowrate. 10%-20% of hydrocarbon is expected to be lost during the absorption of carbon dioxide. The stripper column is also a high operating and maintaining cost unit

Typically there are three types of membrane for CO₂ removal: cellulose acetate, polyimides and perfluoropolymers. Removal of CO₂ using polymeric membrane has become a major focus due to the easiness of manufacturing it (Scholes et al., 2012).

Major advantages of membrane technology is its smaller footprint. The separation using membrane is cleaner, require less maintenance, low energy consumption and environmental friendly. However, the main drawback of membrane technology is the hydrocarbon loss – the natural gas permeated into the membrane together with CO₂. To resolve this issue, two stage system are being practice instead of single stage membrane separation (Mokhatab & Poe, 2012c).

Another type of physical separation is cryogenic fractionation. The practice of using low temperature distillation to remove CO₂ has been long practiced. The distillation process are able to liquefy and purify CO₂ to up to 90% of its source. This method require the gas to be cooled to a very low temperature so that the CO₂ can be liquefied and separated. Pretreatment of natural gas is also required to prevent freezing of lines and blockages due to the components with freezing point lower than operating condition of distillation column. Another problem with the distillation of natural gas at cryogenic condition is that the difficulties of separating CO₂ with methane, CO₂ with ethane and CO₂ with H₂S (Mokhatab & Poe, 2012c).

2.2.4 Supersonic Separation

The separation of CO₂ from natural gas using supersonic separation technology is developed by TwisterBV. This technology shares similar benefits and simplicity, robustness and ease of operation as the JT Valve (LTS) (Mokhatab & Poe, 2012d).

a. Twister® Supersonic Separator

Twister® Supersonic Separator or for the remainder of this report will be called the Gas Twister is a one of the proprietary devices developed by the TwisterBV. This device separate CO₂ from natural gas at supersonic velocity. Gas twister has thermodynamics similar to a turbo expander in which it combines the three process steps in a compact tubular device (TwisterBV, 2011):

- Expansion
- Cyclonic gas/liquid separation
- Recompression

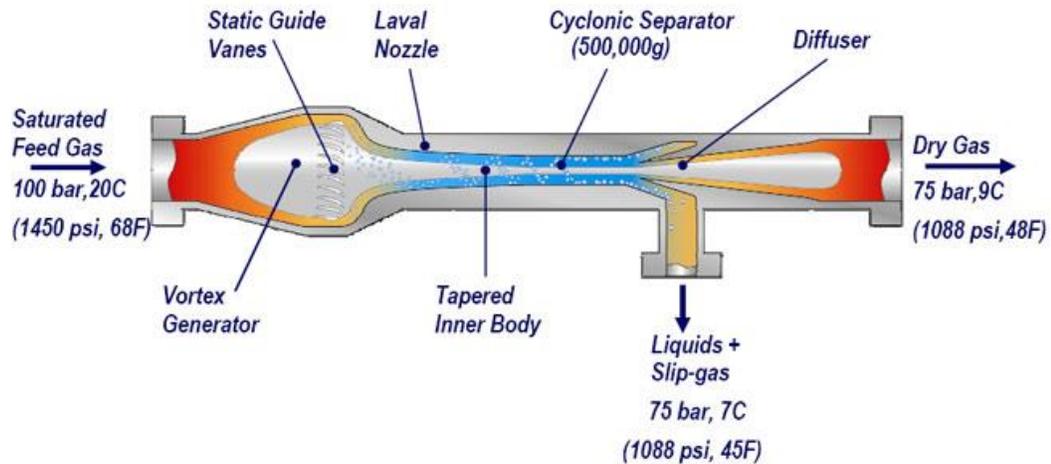


Figure 2.2-3 Cross-section view of a Gas Twister (TwisterBV, 2011).

In the expansion stage, the feed gas enters the gas twister. As the gas passes through the Laval nozzle the gas is accelerated to supersonic velocity. The increase in velocity results a pressure and a temperature drop. This then causes heavier hydrocarbon to form condensates (TwisterBV, 2011).

After the expansion, the fluid inside the gas twister is separated by the swirling effect generated by the vortex generator. The swirling effect makes the Gas Twister to act like a cyclone separator which then results in gas-condesates separation. The gas leaves the gas twister through the gas outlet while liquid leaves the Gas Twister through the liquid outlet (TwisterBV, 2011).

After the separation, the gas leaves the Gas Twister, the gas enters a diffuser region in order for recompression. Recompression is necessary to recover the pressure loss which occurs when the gas is accelerated to supersonic velocity. Typically there will be about 25% of pressure loss across the Gas Twister (TwisterBV, 2011).

b. Supersonic Cyclonic Oil Degasser™

Supersonic Cyclonic Oil Degasser™ or for the remainder of the report will be called Liquid Twister is a similar proprietary device as the Gas Twister. However instead of having gas inlet, Liquid Twister has liquid as inlet. In this device, Liquid is accelerated once it enters the vortex generator. The accelerated liquid transform into vapor as the pressure dropped. As with the Gas Twister, static guide vanes are used to introduce a swirl and the vortex produce will separate the two phase mixture into liquid and gas outlet stream (TwisterBV, 2011).

c. Crystallization Vessel

The separation of CO₂ is further enhance by connecting the CO₂ rich stream from both twisters – Liquid and Gas to a CO₂ crystallization vessel. In the crystallization vessel, the recovery of pure CO₂ is assisted by the formation of solid CO₂ operation, the feed of the vessel comes from either a Liquid or Gas twister or both has a temperature below the CO₂ freeing line. CO₂ solidfies as it enter the vessel. The tangential inlet creates swirling flow as such in a cyclone separator which separates the condensed phase and the gas phase. Usually the separation has an efficiency of greater than 99% (TwisterBV, 2011).

Solid CO₂ is collected at the bottom of the vessel and is free from hydrocarbons. The CO₂ is melted by supplying heat input at the bottom of the vessel. The heat supplied is to prevent the accumulation of solid CO₂ at the bottom of the vessel. The crystallization vessel concept is based on the TwisterBV's hydrate separator.

Gas Twister technology removes the needs of chemicals usage for the in the recovery of CO₂. The system also uses less major equipment. (Machado, Monteiro, Medeiros, Epsom, & Araujo, 2012) The compact design of Gas Twister minimize the footprint of the system especially for offshore production facility. (Karimi & Abdi, 2009)

2.3 Gas Hydration

In most cases natural gas is in supercritical dense phase and it is very likely that water and hydrocarbon component to condense and form a liquid phase. The heating value of natural gas decreases with the presence of water and if condensed the water will cause operational problem such as corrosion, excessive pressure drop, hydrate formation and slug flow. (Karimi & Abdi, 2009)

The major reasons for dehydrating or removing water from natural gas are (Manning & Thompson, 1991):

- a. When natural gas combine with liquid or free water hydrates will form and it can plug valves, fitting or even pipelines.
- b. With the presence of high concentration of CO₂ and/or H₂S, water will form acidic solution which is corrosive.
- c. Condensed water in pipeline will caused slug flow.
- d. Presence of water vapor will increase volume and decrease heating value of gas
- e. To meet export gas water content specification usually 7 lb H₂O per MMSCFD
- f. To prevent freeze up in cryogenic operation.

Hydrate is formed when pressure is suitably high. At this point, water will form a complex solid structure at a temperature higher than normal freezing point. Hydrate is inherently unstable since it is made up a large network of open cavities. As cooling continues the normally compact and stable ice structure will eventually form unless some foreign molecule enters the structure and supports the cavity. In natural gas processing, the most abundant guest molecule is methane. The stable solid compound formed between water and guest molecule is called hydrate. C₂, C₃, i-C₄ and CO₂ are also capable of forming hydrates. Hydrates typically will formed around 15°C – 25°C (in natural gas processing) depending on gas composition and gas pressure. Once hydrates are formed will block the flow of gas in the system hence disrupting the overall process (GPSA, 2004).

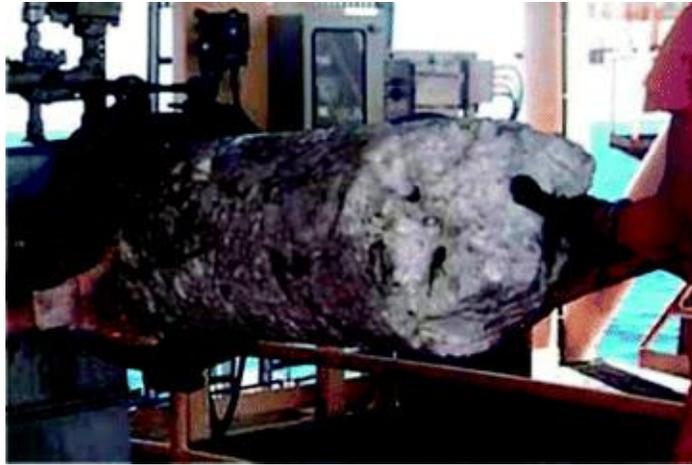


Figure 2.3-1 Hydrate formed in an equipment inside a natural gas process facilities (TwisterBV, 2011).

Listed are different methods of natural gas dehydration (Karimi & Abdi, 2009)

- i. Absorption using liquid desiccants
- ii. Adsorption using solid desiccants
- iii. Dehydration using calcium chloride
- iv. Dehydration using membrane permeation
- v. Supersonic dehydration

2.3.1 Hydrate Separator™

The gas hydration techniques are not being compared in this study. However, the study will be using another TwisterBV's proprietary technology - Hydrate Separator™ to remove water in order to prevent hydrate formation. This type of hydrate separator eliminates the needs of using chemicals such as ethylene glycols and methanol in order to separate hydrates and liquids from natural gas (TwisterBV, 2011).



Figure 2.3-2 Installed Hydrate Separator™ for JT-LTX system (TwisterBV, 2011).

Hydrate Separator™ is based on the conventional Low Temperature Extraction (LTX) technology which has been used as early as 1950s. The operational principle is the same as the Low Temperature Separator (LTS) with heating coils in the liquid section to melt the hydrates. This enables the operation below hydrate temperature without the use of any chemicals (TwisterBV, 2011). The topside of the Hydrate Separator™ is a mono-cyclone separator which could achieve a high separation greater than 99% of liquid and hydrates from the gas flow. On the other hand, the bottom part of the Hydrate Separator™ is made up of a water bath heater to melt the hydrate layer rendering a water phase and hydrocarbon liquid phase (TwisterBV, 2011).

2.4 Factors Influencing Natural Gas Processing

Processing plants of oil and gas are designed to handle peak production rates for both gas and oil. Therefore they will suffer from variation of operating conditions starting from the start of field life to the end of field life. (Nguyen, Fülöp, Breuhaus, & Elmegaard, 2014)

Separation oil gas and water is influenced by the temperature and pressure. Low pressure and high temperature will promotes the separation of the oil and gas. (Nguyen et al., 2013)

Nguyen et al stressed that in production facilities highest energy demand will be the compression of natural gas, pumping and injection systems and management of this are crucial. Typically, recompression of gas will be utilizing the most energy out of the total energy consumption. (Nguyen et al., 2013)

Low pressure will promotes the separation of natural gas at cryogenic condition however at low pressure the total duty needed for the recompression of will be very high and uneconomic. (Nawaz & Jobson, 2010)

In term of energy demand, condensation method (low temperature separation) is the most appropriate natural gas dehydration method. At low pressure, this method requires more energy as the JT effect can't be used. In low pressure cases, absorption and adsorption method is favorable. (Netusil & Dittl, 2011)

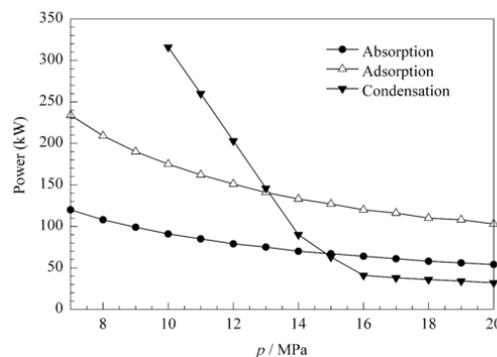


Figure 2.4-1 Power Demand vs Pressure of Natural Gas (Netusil & Dittl, 2011)

Graph in Figure 2.4-1 Power Demand vs Pressure of Natural Gas (Netusil & Dittl, 2011), depicted the energy demand at different pressure for the different natural gas dehydration method. Processing of high pressure gas requires less energy compare to low pressure.

Low temperature separation requires the process to be in cryogenic conditions. Therefore, 4 issues are highlighted in optimizing a low temperature system. The four issues are as follows (Tahouni, Hassan Panjeshahi, & Ataei, 2011):

- a. Options of sequences.
- b. Options of simple and complex separation devices.
- c. Options of conditions such as the pressure and feed condition.

d. Integration of refrigeration cycles.

Selection of refrigerant used to cool the process stream to a sub-ambient temperature is also an important considerations since sometimes one refrigerant is not enough to cover the entire span of temperature range (Tahouni et al., 2011).

The major advantage of Gas Twister system is that it requires considerably less number of major equipment such as the vessels, heat exchangers, tower and etc. With the use of Gas Twister it eliminates the use of glycols in which it eliminates the need for glycol regeneration tower which is big and heavy. Twister technology produces more natural gas liquids (NGL), by products of natural gas production, in return it can increase the revenue from the produce gas stream. (Machado et al., 2012)

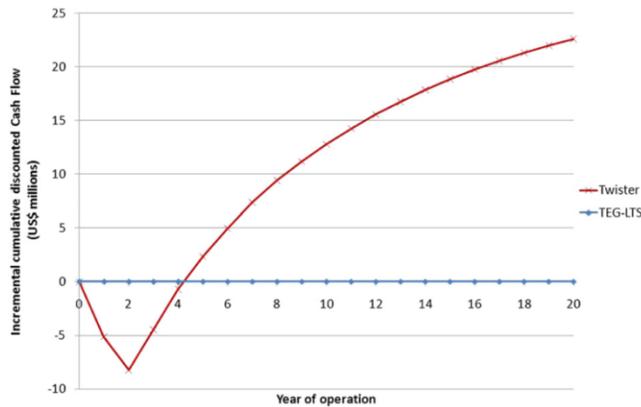


Figure 2.4-2 Economic Performance of process alternatives (Twister vs TEG + JT-LTS) (Machado et al., 2012)

Karimi and Abdi (2009) manipulated several parameters in order to optimize the supersonic separation using supersonic nozzle. By manipulating parameters as such the inlet pressure, inlet temperature and flow rate on the behavior of the fluids exhibits significant changes which is represented in a phase diagram.

The stream pressure was manipulated to 10, 30, 50 and 70000 kPa. Water capacity in the stream increases as the inlet pressure increases. Besides, the inlet flow capacity increases with the increase of the inlet pressure. As the result of increased inlet pressure, the temperature-pressure variation is more likely to stay in the supercritical phase of the phase envelope.

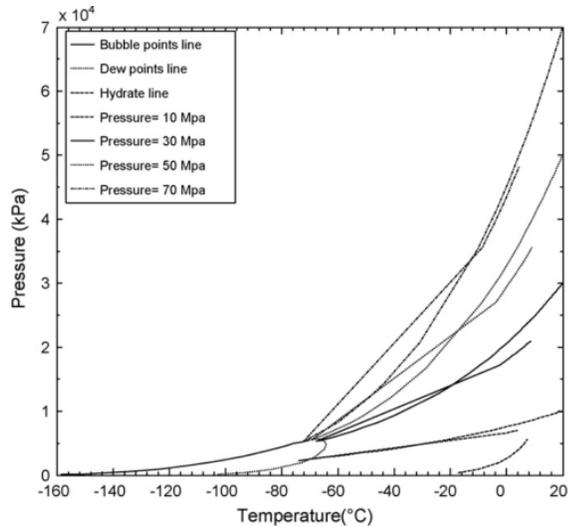


Figure 2.4-3 Phase envelope and pressure temperature distribution with the manipulated inlet stream pressure (Karimi & Abdi, 2009).

The inlet stream was manipulated by Karimi and Abdi (2009) with different temperatures of 1, 20, 40 and 60°C. With the increase in temperature, the water capacity of gas to hold water increases as well. Based from the phase diagram (Figure 2.2-6), as the temperature increases the amount of the liquid phase decreases.

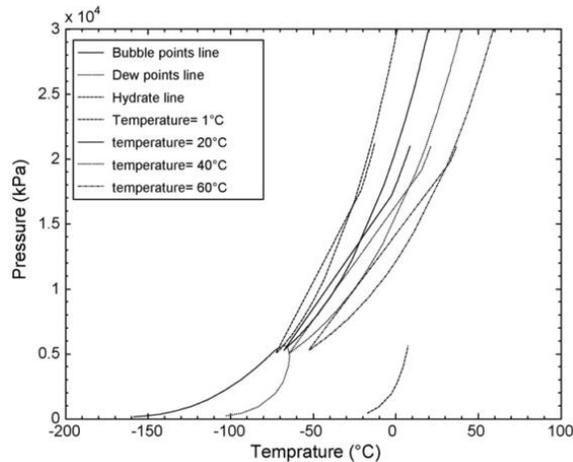


Figure 2.4-4 Phase envelope and pressure temperature distribution with the manipulated inlet stream temperature (Karimi & Abdi, 2009).

Besides temperature and pressure Karimi and Abdi (2009) also manipulated the flow rate of the stream (1000, 2000, 3000, 4500 and 5000 kmol/h). From the research, flow rate should be kept the same with the design pressure in order for the flow to achieve its Mach number or else the flow will never be choked and separation will not occurred.

The behavior of the fluid stream of the supersonic separator also depends on the nozzle's back pressure (pressure at the nozzle exit). After choking occurs at the throat, two possibilities might happen – pressure recovery or pressure decrease. Lowering the pressure recovery (higher pressure drop across the nozzle), gas will expand more and pressure and temperature at the exit will be lower. Increasing the backpressure might reduce the separation efficiency of the nozzle (Karimi & Abdi, 2009).

Chapter 3: Methodology

3.1 Project Flowchart

The figure below shows the overall project flowchart which begins with the problem defining and ends with making the conclusion.

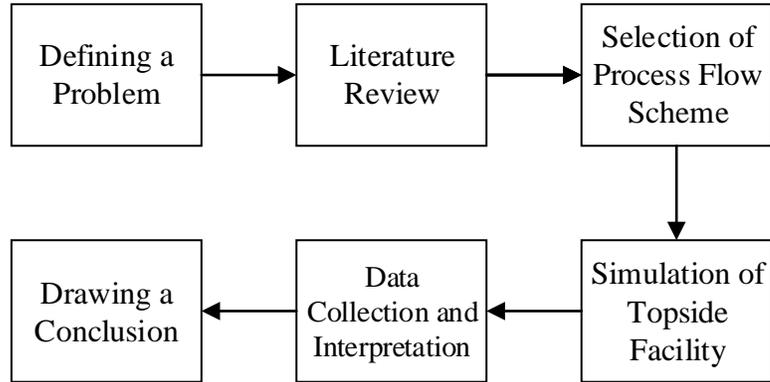


Figure 3.1-1 Process flow diagram of the study.

a) Defining the Problem

The first step in scientific research is defining the problem. In this research the problem has been defined in the previous section – 1.2 Problem Statement i.e to study the effect of varying the process parameters on the power consumption of the topside facilities. At this stage as well the scope of the studies is set so that the planned project work will be feasible. By defining the problem, it would lead the research to analysis stages and hence it will lead the research to achieve its objectives by solving the defined problem.

b) Literature Review

Literature review is an important stage in a research. Journals, articles, books and project documents are some examples of literature that are being used in this research. By referring to these sources, it enables the author to get a clear view of how the proprietary technology works hence enabling the author to identify possible factors that influence the energy consumption of the natural gas production facilities. In this literature review stage, the author looks also into the production of natural gas before starting specific research on the gas twister technology.

c) Selection of Process Flow Schemes

A natural gas production facility that uses Gas Twister system for its removal of CO₂ is identified. A production facility for an offshore platform in Sarawak is selected for this research. In this research the gas field is named as Field 'X'. The process flow scheme and operating condition of each equipment and streams for acquired for the purpose of the next step in the research i.e the simulation of the topside production facility.

d) Simulation of Topside Facility

The steady state simulation of the topside facilities with Gas Twister system is done using Aspen HYSYS 7.3. Simulation using Aspen HYSYS removes the needs of conducting the real experiments which is costly at large scale. This stage will begin with developing the base case simulation. Once the base case is verified, several process parameters are manipulated in order to detect changes in the total energy consumption of the production facilities.

e) Data Collection and Interpretation

Three important data that needs to be collected are the composition of CO₂ in export gas, total power or duty required by the process flow schemes and hydrocarbon loss in the recovery process. These three important parameters are main indicator for the performance of the topside facilities. Data collected are tabulated and graphs are made to identify in the trends as changes are introduced. The trends shown are analyzed and interpreted.

f) Drawing a Conclusion

Conclusion is made once the data collected has been analyzed and interpreted. Recommendations are made for the production system to improve its operability and for future work continuation.

3.2 Topside Facility Process Flow Schemes

The process flow scheme selected for this study is developed by TwisterBV for PETRONAS for the development of an oil field in offshore of Sarawak. For the purpose of this study, field is named as Field X. In this process flow schemes, several proprietary technology owned by TwisterBV are being used that includes – Gas Twister, Liquid Twister, Hydrate SeparatorTM and CO₂ crystallizer.

The process flow schemes of the production facility is shown in the appendix section – Appendix 1.

3.3 HYSYS Simulation Methodology

Aspen HYSYS 7.3 is used for simulating the process flow schemes for all the cases that are being studied in this research. In order to model the process flow schemes the following steps were followed:

- i. First, the thermodynamics properties for components involved in the simulation are determined by selecting the fluid packages. In this research, Peng-Robinson is chosen as the fluid package. At the same time, the components of natural gas are also selected and a component package Field X is created.
- ii. Once, the components and fluid package are determined the simulation environment is entered and the modelling of the process flow schemes is started.
- iii. Common equipment such as 2-phase separator, pumps, compressor and heat exchangers are simulated using equipment available on the object palette. For, the TwisterBV's proprietary technology equipment, they are not available in HYSYS. Therefore, for each equipment – Gas Twister, Liquid Twister, Hydrate Separator and CO₂ Crystallizer, they are modelled using the process template option. By using the process template option, a few equipment are used to represent them in order for it to have the same working principle as the original equipment. The modeling of the equipment is made in a sub-flowsheet environment – refer to Appendix 3 to Appendix 6.

- iv. Before the simulation is run. The operating conditions for each equipment are defined. For case 2, the component package is redefined for each simulation.
- v. Once all the unit operations are converged, necessary data are taken.
- vi. Steps iv and v are repeated for other case studies. Four case studies are prepared for this research i.e:
 - a. Case 1: Effects of Varying CO₂ Composition
 - b. Case 2: Effects of Different Natural Gas Composition
 - c. Case 3: Effects of Variation in Arrival Temperature of Natural Gas
 - d. Case 4: Effects of Variation in Arrival Pressure of Natural Gas
- vii. Data obtained are gathered and analyzed. Discussion are made from the analyzed data.
- viii. Finally, conclusion of the project is made based on the result and discussion.

3.4 Base Case Feed Gas Composition and Other Process Parameters

The base case is based on a high CO₂ gas field in offshore Sarawak. Again for this research purpose the gas field is named as Field X. The gas field has a peak gas rate of 2000 MMSCFD. Therefore, the simulation will be run at maximum production rate of 2000 MMSCFD of natural gas.

Table 3.4-1 Gas Composition of Field X.

Component	Mole %
N2	0.02927
CO2	0.68954
C1	0.25852
C2	0.00925
C3	0.00350
i-C4	0.00071
n-C4	0.00061
i-C5	0.00032
n-C5	0.00022
n-C6	0.00035
methylcyclopentane	0.00012
benzene	0.00028
cyclohexane	0.00012
n-C7	0.00018
methylcyclohexane	0.00016

toluene	0.00020
n-C8	0.00035
E-benzene	0.00002
m-Xylene	0.00000
o-Xylene	0.00001
n-nonane	0.00035
124-Mbenzene	0.00003
n-Decane	0.00013
n-C11	0.00009
C12+	0.00159
H2O	0.00407

Table 3.4-2 Process operating condition for natural gas feed from Field X.

Parameter	Value
Pressure	138.9 bar
Temperature	40.0°C
Vapor fraction	0.996

The feed composition of the natural gas from Field X is shown in Table 3.4-1 Gas Composition of Field X and the operating condition of the feed gas stream is shown in Table 3.4-2 Process operating condition for natural gas feed from Field X.

For other cases, the data the composition of gas are manipulated based on the base case composition or if other composition is used it will be mentioned. Other gas composition are attached in the Appendix II.

3.5 Calculations of Total Duty and Hydrocarbon Loss

Total duty is defined as the sum of the shaft duties measured in kW required for the chiller unit, cascade cooling, booster compression, gas recycling, export compression, heating medium pump and CO₂ re-injection pumps.

$$Total\ Duty(kW) = Total\ Duty\ of\ Pumps + Total\ Duty\ Compressors$$

Total hydrocarbon loss is defined on molar basis as shown in the formula below:

$$\text{Hydrocarbon Loss} = \frac{HC_{\text{Stream100}} - HC_{\text{Export Gas Stream}}}{HC_{\text{Stream100}}} \times 100$$

Where HC = hydrocarbon (C1 to C12+) in kgmole/h and Stream 100 is the wellhead acid gas with saturated water. Hydrocarbon loss includes the losses in the liquid outlet streams from the inlet separator and hydrate separator.

3.6 Gantt Chart and Key Milestones

This final year project covers for the whole final two semesters in the campus. It began at the 1st week of the final year first semester in FYP 1 and will end at the final week of the final semester in FYP2. Through the two semesters, a lot of project works will be going on. For, FYP 1 the project works are illustrated in detailed in the Gantt Chart and Key Milestones for FYP 1 Table 3.6-1 Gantt Chart and Milestone for Final Year Project 1..

On the other hand the proposed project works for FYP 2 is shown in the Table 3.6-2 Gantt Chart and Milestones for Final Year Project 2.. For this table the Gantt Chart and Milestone might be changing in the next semester.

Table 3.6-1 Gantt Chart and Milestone for Final Year Project 1.

Tasks	1	2	3	4	5	6	7	8	9	10	11	12	13	14
1. Selection of Project Topic.											HARI RAYA HOLIDAY			
2. First Meeting and Discussion with the Supervisor.														
3. Reviewing the documents provided by TwisterBV.														
4. Literature reviews: collecting information on previous similar research.														
5. Modelling the proposed design of the topside facilities for Base Case.														
6. Reviewing the Base case simulation with the Supervisor														
7. Preparation and submission of the extended proposal.							●							
8. Proposal Defense presentation with Supervisor and External Examiner.								●						
9. Editing the simulation based on comments from Supervisor														
10. Reviewing edited simulation with the Supervisor														
11. Initial analysis of the simulations.														
12. Preparation and Submission of interim report.														



Process Work



Suggested Milestone

Table 3.6-2 Gantt Chart and Milestones for Final Year Project 2.

		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
1.	Running the simulation (manipulating the composition of natural gas)	■	■	■	■											
2.	Reviewing simulation result with Supervisor			■	■	■	■									
3.	Submission of Progress Report							●								
4.	Running the simulation (manipulating feed composition)								■	■						
5.	Reviewing obtained result with supervisor									■	■					
6.	Analysis of obtained results										■	■				
7.	Pre-Sedex											●				
8.	Preparation of Final Report												■			
9.	Submission of Final Report (draft)													●		
10.	Submission of Technical Paper													●		
11.	Viva														●	
12.	Submission of Project Dissertation															●

■ Process Work ● Suggested Milestone

Chapter 4: Result and Discussion

For the time being, 4 simulation cases are being simulated:

- a. Case 1: Effects of Varying CO₂ Composition
- b. Case 2: Effects of Different Natural Gas Composition
- c. Case 3: Effects of Variation in Arrival Temperature of Natural Gas
- d. Case 4: Effects of Variation in Arrival Pressure of Natural Gas

The simulation cases are run on Aspen HYSYS 7.3 and the data obtained from the simulation are tabulated, analyzed and interpreted. All the simulation are based on a base case which is being explained in the following section.

4.1 Base Case Simulation.

A simulation model is developed for a topside facility for natural gas production from an offshore field to produce an export gas stream with CO₂ content of 8% mol fraction.

The feed operating condition is based on Table 3.4-1 Gas Composition of Field X. and Table 3.4-2 Process operating condition for natural gas feed from Field X.. The process is described in Section 4.1.1-Process Description of Base Case.

4.1.1 Process Description of Base Case

2000 MMSCFD of wellhead gas at 138.9 bar and 40°C enters the Inlet Separator, V-100. V-100 operates at 138.9 bar and 40°C in which the vessel is responsible in removing liquids in the wellhead natural gas. The removed liquids hydrocarbon is sent to the liquid handling system while the gas hydrocarbon is then expanded by expander K-100 to 24.9 bar. The pressure of the gas is further expanded using a pressure control valve to 11.8 bar prior entering the Hydrate Separator, V-101. The hydrate separator removes up to 76 mol% of water from the gas hydrocarbon. The liquids removed from the Hydrate Separator, V-101 is sent to the liquid handling system. The rest of the water in the gas is removed by the molecular sieve, T-100.

The gas hydrocarbon is then cooled by a series of Cold Boxes to achieve a cryogenic condition just before entering the Liquid Twister, S-200. The gas hydrocarbon is first cooled to -4.70°C by Cold Box Hex 1, E-200. It is then further cooled to -20.0°C

by C3 Chiller HEX, E-201. The gas stream is then split into two streams in the ratio of 80:20. The majority of the gas is cooled to -47.0°C while the rest is used as the refrigerant for the cold box HEXs. The refrigerant is then recompressed and cooled in by the refrigeration recompressing system. The recompressed refrigerant is then recycled back to the upstream of Cold Box HEX 1, E-200.

The gas is then cooled to -50.5°C just before entering the Liquid Twister, S-200. The liquid twister removes CO_2 to 24mol%. The CO_2 rich outlet of the Liquid Twister is connected to the CO_2 Crystallizer, V-301. The methane rich outlet is connected to a CO_2 Separator, V-200 to further maximize the separation of gas hydrocarbon and liquid CO_2 . Due to the pressure loss across the Liquid Twister the gas hydrocarbon is recompressed in a booster compressor, K-201 to 69.0 bar prior to entering the Gas Twister, S-300. Before entering the Gas Twister, S-300 the stream is cooled to -47.8°C using Cold Box HEX, E-300.

The Gas Twister, S-300 removes CO_2 further to 7.3 mol%. The gas outlet of the gas twister is connected to the export compressor trains while the liquid outlet is connected to a Low Temperature Separator, V-300 to recover methane from the liquid CO_2 . The gas outlet of the Low Temperature Separator is reconnected to the export compressor trains while the liquid outlet is sent to the CO_2 Crystallizer, V-301. The export gas, is heated by the Cold Box HEX, E-300 before entering a double compression system of the gas export compressor trains. The export gas with the flow rate of 623.3 MMSCFD is at 138.6 bar, 40°C with 7.3 mol% CO_2 .

The CO_2 rich liquid outlet of Liquid Twister and Gas Twister is sent to the CO_2 crystallizer, V-301. The crystallizer separates CO_2 from methane by crystallizing the CO_2 . Once the CO_2 is melted to liquid, the methane will separates as gas from the CO_2 . The methane is recompressed by compressor, K-300 and then cooled to 40°C by cooler, E-304 before being recycled back to the upstream of the Cold Box HEX 1. The liquid CO_2 from the CO_2 crystallizer is heated by CO_2 heater, E-303 before being mixed with the liquid CO_2 from the CO_2 separator, V-200. The mixed CO_2 liquids is pumped to 60.0 bar and heated up again by Cold Box HEX 1, E-200 to 15°C . The heated CO_2 is then pumped to 173.4 bar and then re-injected to the well.

4.1.2 Process Flow Schemes Base Case

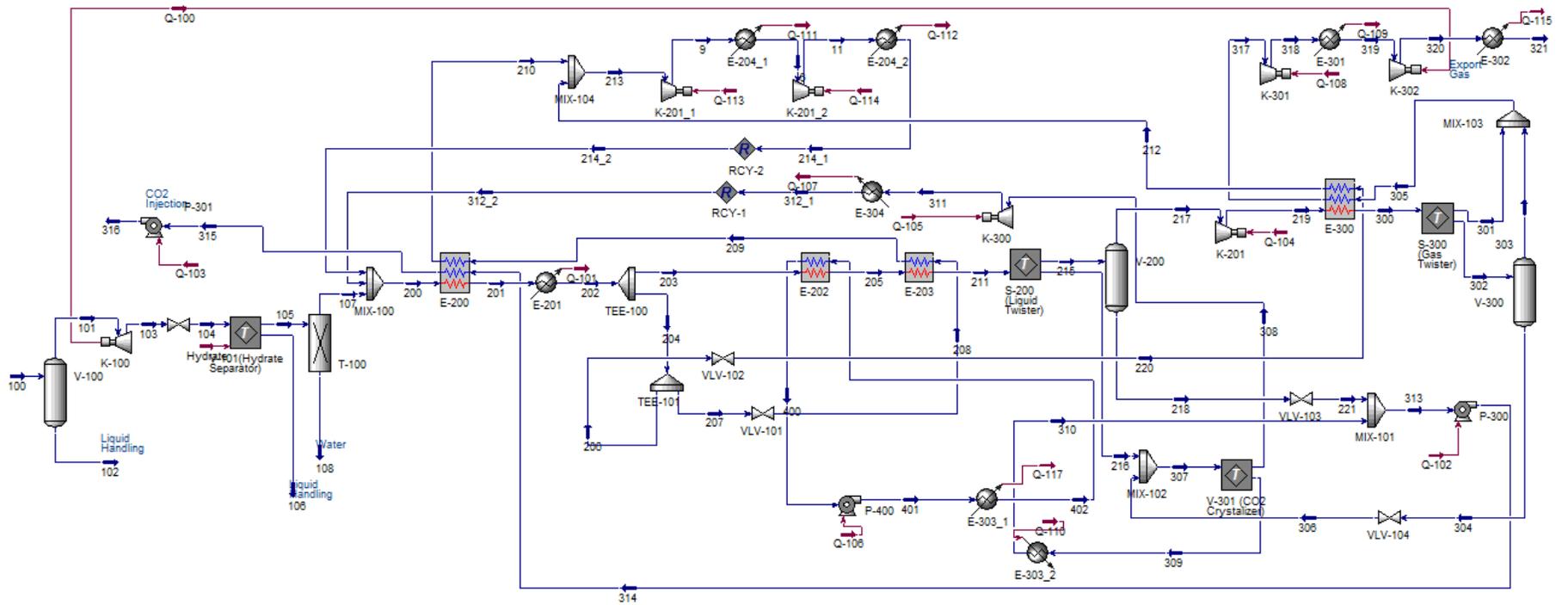


Figure 4.1-1 Process Flow Schemes of Base Case from Aspen HYSYS

4.2 Validation of Base Case Simulation.

In order to validate the base case simulation, the result of the simulation is compared with an existing simulation that have been simulated by TwisterBV using the similar process flow schemes and operating condition. The heat and mas balance of the two simulation is in Appendix 8 and Appendix 9.

The two simulation results are shown in the following table.

Table 4.2-1 Comparison between Simulation of Base Case with Simulation by TwisterBV.

	Existing	Base Case Simulation	Differences, %
CO ₂ in Export Gas	7.90 %	7.65 %	3.10
Total Duty (MW)	166.40	166.69	0.17
Total Hydrocarbon Loss	9.86 %	9.91 %	0.51
Export Gas Flow (MMSCFD)	605	603.8	0.20

The main cause for the differences are due to the fact that the developed simulation are not using the actual unit in the HYSYS. Due to unavailability of Gas Twister, Liquid Twister, Hydrate SeparatorTM and CO₂ crystallizer, these units are modeled in the sub-flow sheet environment using typical equipment that are available in HYSYS. This has also been aforementioned in the problem statement as one of the main constraint of this study. Hence, it is concluded that this will be the main source of error in the simulation. However, based on data obtained from the based case the error from the simulation is considerably low. The modeled equipment will have thermodynamic properties the same as the original equipment.

Another possible causes for the error is that the hypothetical component, C12+ is being represented as C20 in the in the HYSYS component manager. This is because the information on C12+ component for the field X is not available and it can be introduced in HYSYS theoretical component manager. C12+ component is separated in the Inlet Separator, V-100. In order to choose the best component to represent the C12+ hypothetical component several heavy hydrocarbon were tested. C20 gave the closest separation result for the separation of heavy component in the Inlet Separator, V-100.

4.3 Case 1: Effects of Varying CO₂ Concentration

The first case study is to investigate how the variation of CO₂ composition will affect the processing of the natural gas in the topside production facilities. Several simulation cases were simulated by varying the methane composition to reduce the CO₂ to 20% CO₂, 40% CO₂, 50% CO₂ and 70% CO₂.

In these simulations, the base case simulation using the Field X composition is used as the max case - 70% CO₂. The other simulation cases (20% CO₂, 40% CO₂, 50% CO₂) the natural gas composition are produced by changing the composition of CH₄ and CO₂ of the Field X natural gas composition.

The results of the simulation cases are tabulated in the following table.

Table 4.3-1 Summary of Simulation of the Varying CO₂ Composition based on 70% Base Case.

Composition of CO ₂ , %	Total Duty, kW	Export Gas Specifications			Hydrocarbon Loss, %
		Molar Flow, MMSCFD	CO ₂ , %	CH ₄ , %	
20	264376.78	1621	5.25	90.38	2.73
40	226132.14	1191	6.22	88.08	4.30
50	204236.22	973.9	6.86	86.30	5.58
70	166693.06	603.8	7.65	81.71	9.91

The data obtained are depicted on the following graphs in order to see the trends in the result.

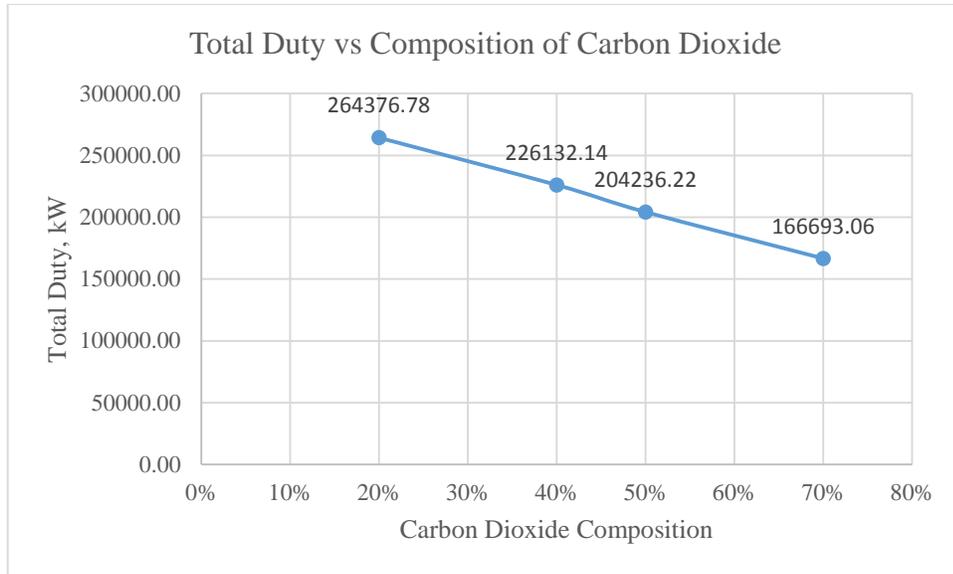


Figure 4.3-1 Graph of Total Duty vs Composition of Carbon Dioxide-Case 1.

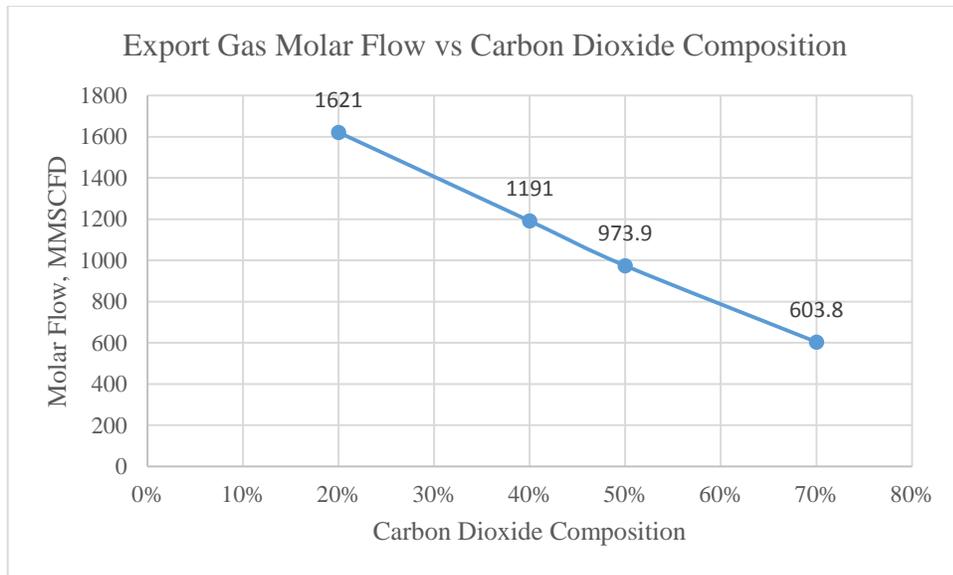


Figure 4.3-2 Graph of Export Molar Flow vs Composition of Carbon Dioxide - Case 1.

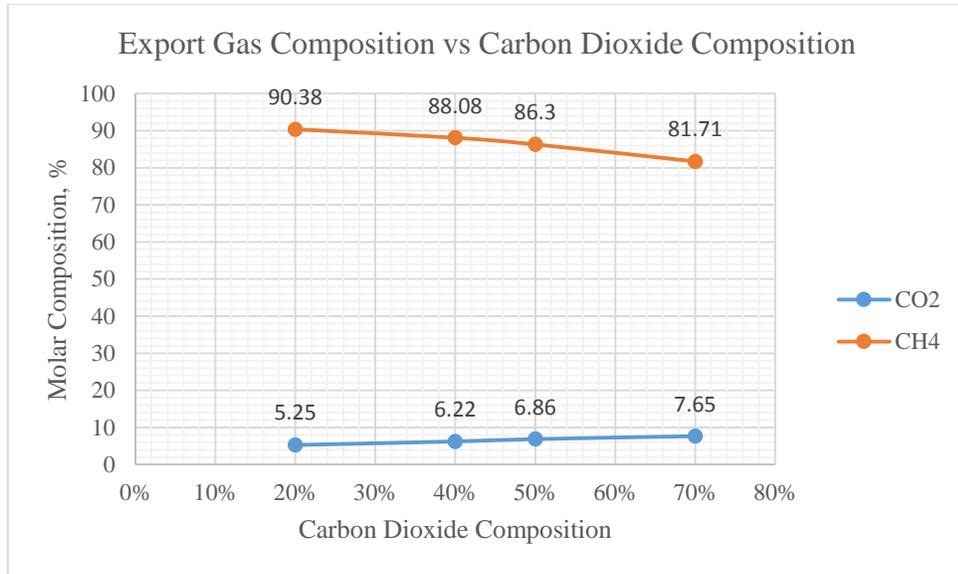


Figure 4.3-3 Graph of Methane and Carbon Dioxide Composition in Export Gas vs Composition of Carbon Dioxide-Case 1.

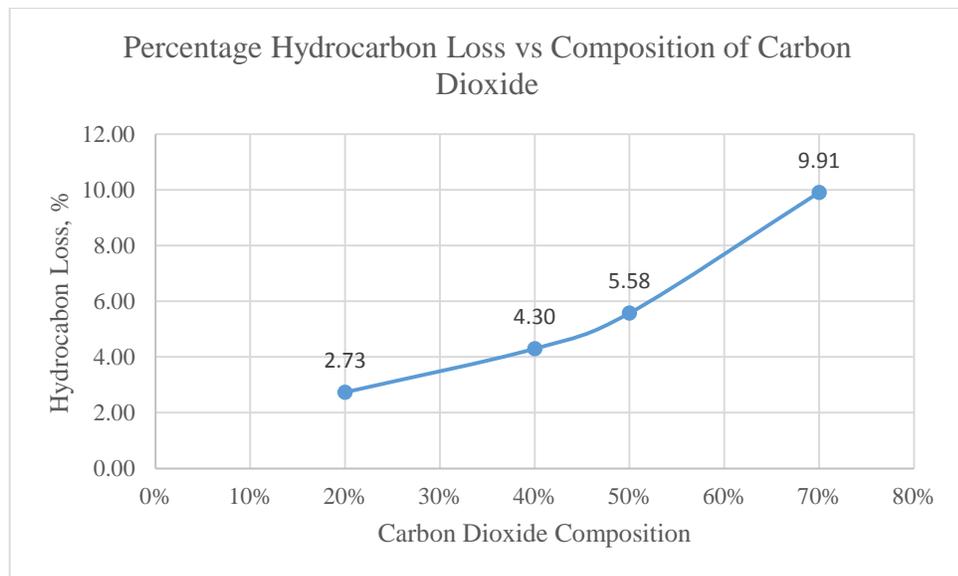


Figure 4.3-4 Graph of Percentage Hydrocarbon Loss vs Composition of Carbon Dioxide-Case 1.

The results of the simulation clearly show that as the composition of CO₂ increases the total duty or the power consumption of the production facilities decreases. This is due to the fact that with the increase of CO₂ the concentration of CH₄ in the natural gas decreases. With the decrease of the CH₄, the total recovery of export gas also decreases. Lower recovery of export gas means less energy consumption for the recompression of export gas. From the data of power consumption obtained, at 20% CO₂ 52% of the total duty (138 MW from 264MW) are for the recompression of the export gas to the export pressure of 138.6 bar. This agrees with Nguyen et al. 2012 that most of energy consumption of production facility is from the recompression of export gas. This shows that the amount of natural gas recovered from the system will have high impact on the total energy consumption.

With the increase of the CO₂ concentration the amount of hydrocarbon loss in the system increases as well. Hydrocarbon are lost mainly in the first two separation module – the inlet separator, V-100 and the hydrate separator, V-101. In this two units, heavy components are being removed from the natural gas. In high CO₂ natural gas, the percentage of heavy component are much smaller as compared to the low CO₂ natural gas. When the heavy component are being removed from the natural gas, natural gas with high CO₂ will experienced greater hydrocarbon loss in both of the separation module. Hydrocarbon losses are also expected in the Liquid Twister, S-200, Gas Twister, S-300 and CO₂ Crystallizer, V-301 in which CO₂ recovery is taken place.

However with different CO₂ concentration the topside facility with Gas Twister system are still able to reduce the CO₂ in the export gas well below the 8% requirement. With lower CO₂ in the inlet, it is expected and proved that the export gas will has lower CO₂ concentration.

The topside production facility is designed to handle 70% CO₂ natural gas. In order for the facility to work better with lower CO₂ natural gas the modifications must be made to design so that the facility can operate at an optimum conditions.

4.4 Case 2: Effects of Different Natural Gas Composition

In the second case study, instead of manipulating the composition of the CO₂ and CH₄, a set of several totally different natural gas composition are being used. In this case, natural gas composition with 10% CO₂, 20% CO₂, 40% CO₂, 50% CO₂ and 70% CO₂ are being used. These natural gas are from different sources (gas fields). The composition are attached in the appendix section. Again in this case study the base case simulation is labeled as the 70% CO₂.

The results of the simulation cases are tabulated in the following table.

Table 4.4-1 Simulation Results for the Different Natural Gas Composition.

Composition of CO ₂ , %	Total Duty, kW	Export Gas Specifications			Hydrocarbon Loss, %
		Molar Flow, MMSCFD	CO ₂ , %	CH ₄ , %	
10	263037.47	1473	4.02	94.10	21.37
20	265442.57	1621	5.25	90.38	19.15
40	233122.62	1191	6.22	88.08	7.630
50	209627.22	973.9	6.86	86.30	21.95
70	166693.06	603.8	7.65	81.71	9.910

The data obtained are depicted on the following graphs in order to see the trends in the result.

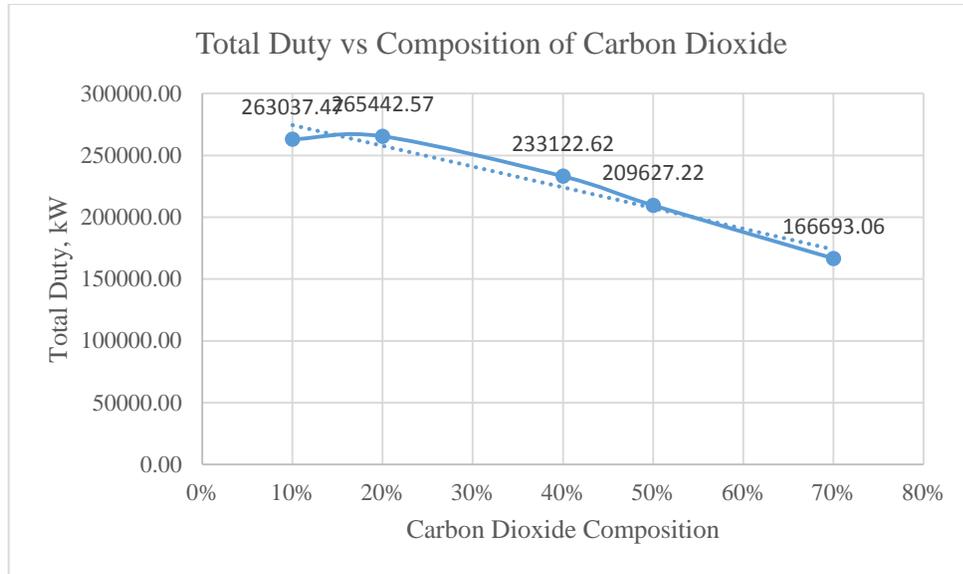


Figure 4.4-1 Graph of Total Duty vs Composition of Carbon Dioxide-Case 2.

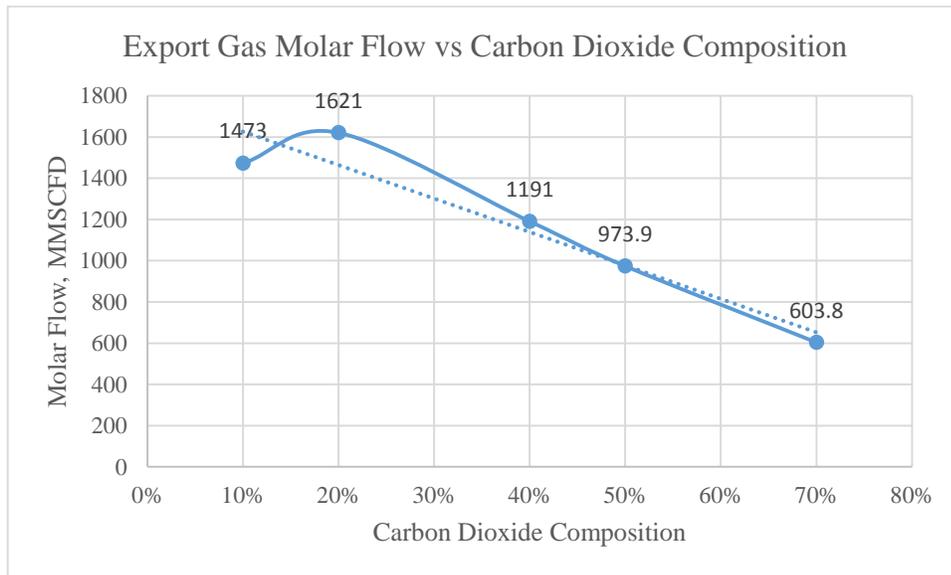


Figure 4.4-2 Graph of Export Molar Flow vs Composition of Carbon Dioxide - Case 2.

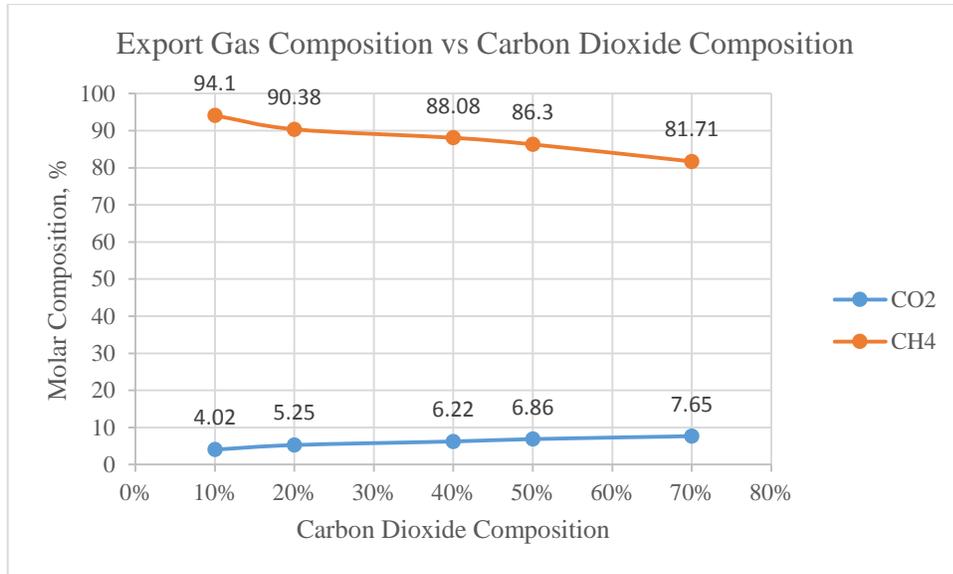


Figure 4.4-3 Graph of Methane and Carbon Dioxide Composition in Export Gas vs Composition of Carbon Dioxide-Case 2.

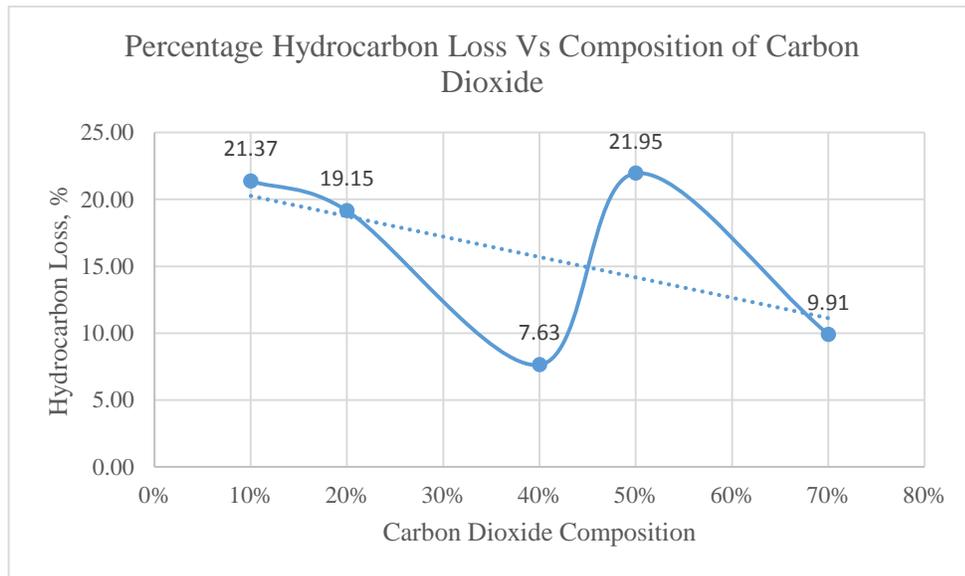


Figure 4.4-4 Graph of Percentage Hydrocarbon Loss vs Composition of Carbon Dioxide-Case 2.

In the second case study, the results obtained are pretty much similar to the previous case study. As the composition of CO₂ increases the total duty or the power consumption of the production facilities decreases. The reason behind this is also the same as in Case Study 1 and this proves the Case Study 1 is true. The CH₄ concentration decreases as the CO₂ concentration increases. With the decrease of the CH₄, the total recovery of export gas also decreases. Lower recovery of export gas means less energy consumption for the recompression of export gas. . However from this case study, at 10% CO₂ 45% of the total duty (118 MW from 263MW) are for the recompression of the export gas to the export pressure of 138.6 bar. Again this agrees with Nguyen et al. 2013 statement on production facility energy consumption. This shows that the amount of natural gas recovered from the system will have high impact on the total energy consumption. Besides, this clearly shows that most of energy required in natural gas processing is the recompression of the natural gas.

With different natural gas composition from different sources, the percentage of hydrocarbon loss does not shows a significant trend. This is due to the fact that different natural gas composition has different percentage of heavy component. Therefore it is quite difficult to analyses the hydrocarbon loss. As being mentioned before the hydrocarbon loss occurs mainly in the inlet separator, V-100 and the hydrate separator V-101. Hydrocarbon losses are also expected in the Liquid Twister, S-200, Gas Twister, S-300 and CO₂ Crystallizer, V-301 in which CO₂ recovery is taken place.

However with different natural gas composition from different natural gas sources the production facility is still capable of reducing the CO₂ composition in the export gas below 8%. Again with lower CO₂ in the inlet, it is expected and proved that the export gas will has lower CO₂ concentration.

The topside production facility is designed to handle 70% CO₂ natural gas from field X. In order for the production facility to handle the high hydrocarbon loss from other natural gas sources modification must be made. This to ensure that, the production facility is able to recover valuable product and not let it wasted.

4.5 Case 3: Effects of Variation in Arrival Temperature

Variation of temperature in Malaysia usually from 23°C to 40°C that is lowest during the monsoon season and hottest during the dry season. For this case, considering the production facilities to be in offshore Malaysia, the arrival temperature of the natural gas will be influenced by the variation of temperature. The simulation cases are run with the temperature varying from 23°C to 40°C with two different CO₂ composition i.e 70% and 40%.

The results of the simulation cases are tabulated in the following table.

Table 4.5-1 Simulation Results for Different Arrival Temperature of Natural Gas.

	Temperature, °C	Total Duty, kW	Export Gas Specifications			Hydrocarbon Loss, %
			Molar Flow, MMSCFD	CO ₂ , %	CH ₄ , %	
40% CO ₂	23	236822.30	1150	5.72	89.38	8.35
	25	236526.17	1150	5.72	89.38	8.35
	27	236295.85	1150	5.72	89.38	8.35
	29	235941.49	1150	5.72	89.38	8.35
	31	235653.38	1150	5.72	89.38	8.35
	33	235368.28	1150	5.72	89.38	8.35
	35	235082.68	1150	5.72	89.38	8.35
	37	234806.98	1150	5.72	89.38	8.35
	39	234531.21	1150	5.72	89.38	8.35
	40	234394.63	1150	5.72	89.38	8.35
70% CO ₂	23	167914.76	601.8	7.73	81.63	10.27
	25	168091.85	601.8	7.73	81.63	10.27
	27	167799.23	601.8	7.73	81.63	10.27
	29	167459.66	601.8	7.73	81.63	10.27
	31	167138.45	601.8	7.73	81.63	10.27
	33	167283.12	602.8	7.73	81.65	10.12
	35	166962.27	602.8	7.73	81.65	10.11
	37	166646.69	602.7	7.73	81.64	10.13
	39	166311.75	602.7	7.73	81.64	10.13
	40	166693.06	603.8	7.65	81.71	9.91

The variation of the temperature has a significant effect on the duty of the turbo expander, K-100. The following table summarizes the duty of the turbo expander, K-100.

Table 4.5-2 Turbo expander -K100 duty at different arrival temperature.

CO ₂	Turbo Expander - K-100, Duty, kW									
	23°C	25°C	27°C	29°C	31°C	33°C	34°C	37°C	39°C	40°C
40%	8088	8288	8486	8684	8879	9074	9266	9457	9647	9741
70%	4911	5558	5978	2159	6447	6627	6860	7093	7341	7442

The data obtained are depicted on the following graphs in order to see the trends in the result.

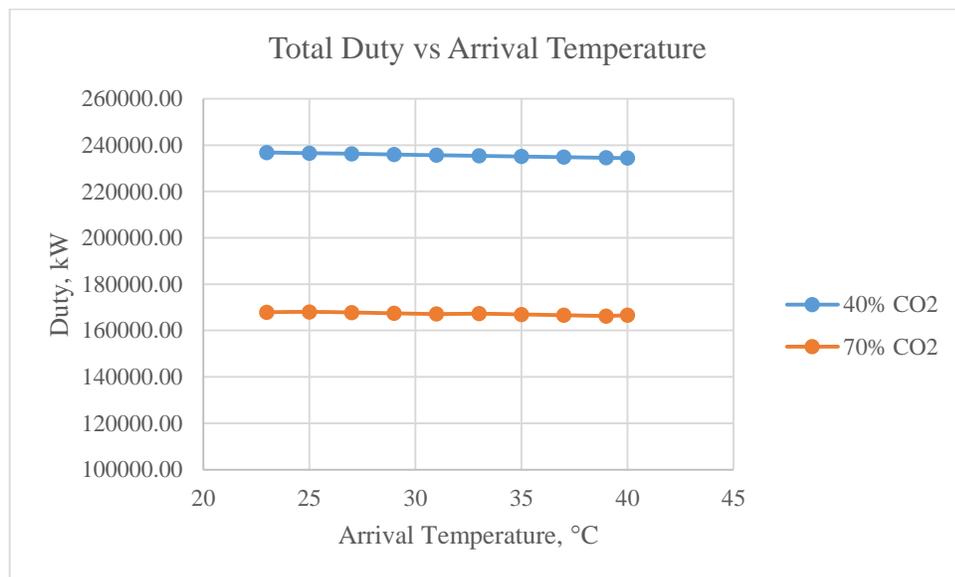


Figure 4.5-1 Graph of Total Duty vs Arrival Temperature.

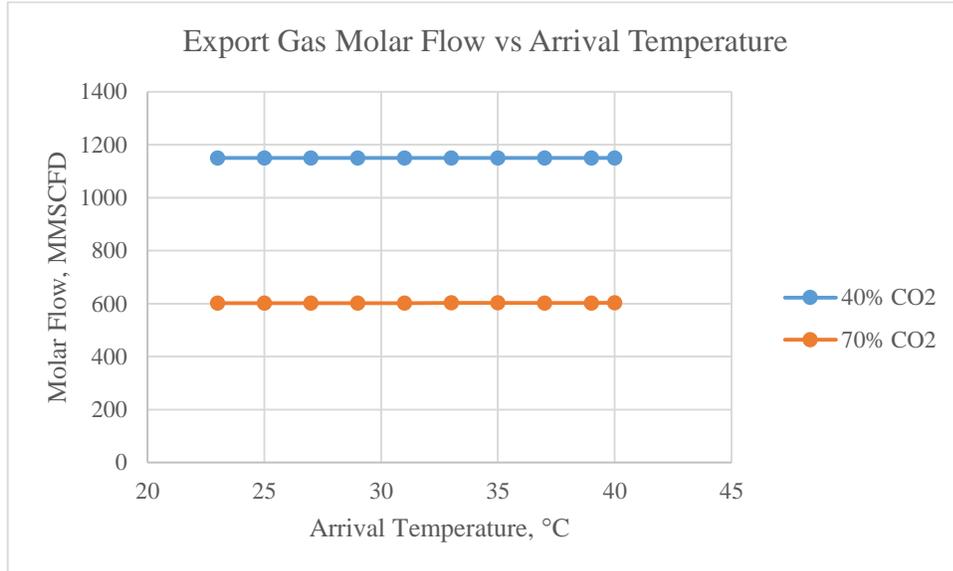


Figure 4.5-2 Graph of Export Gas Molar Flow vs Arrival Temperature.

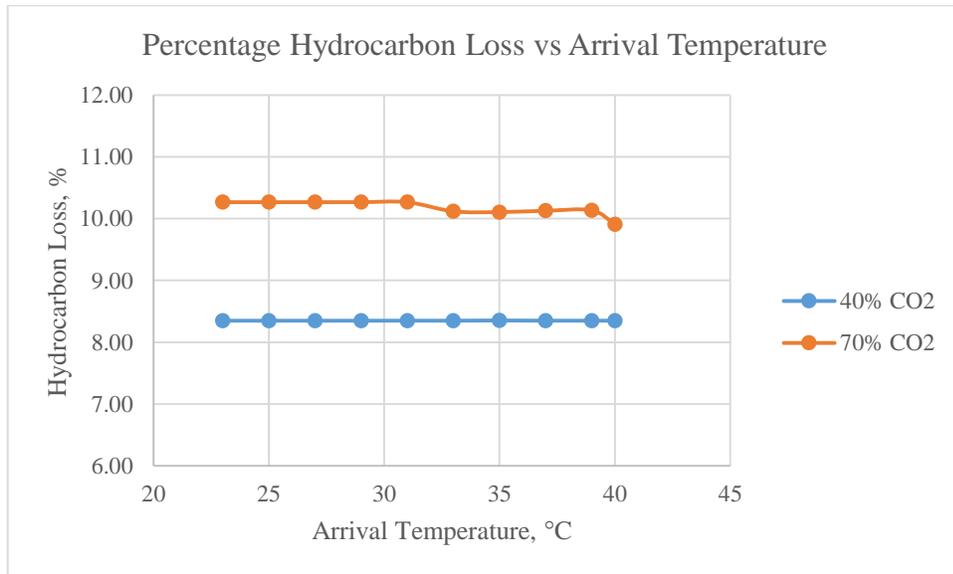


Figure 4.5-3 Graph of Percentage Hydrocarbon Loss vs Arrival Temperature.

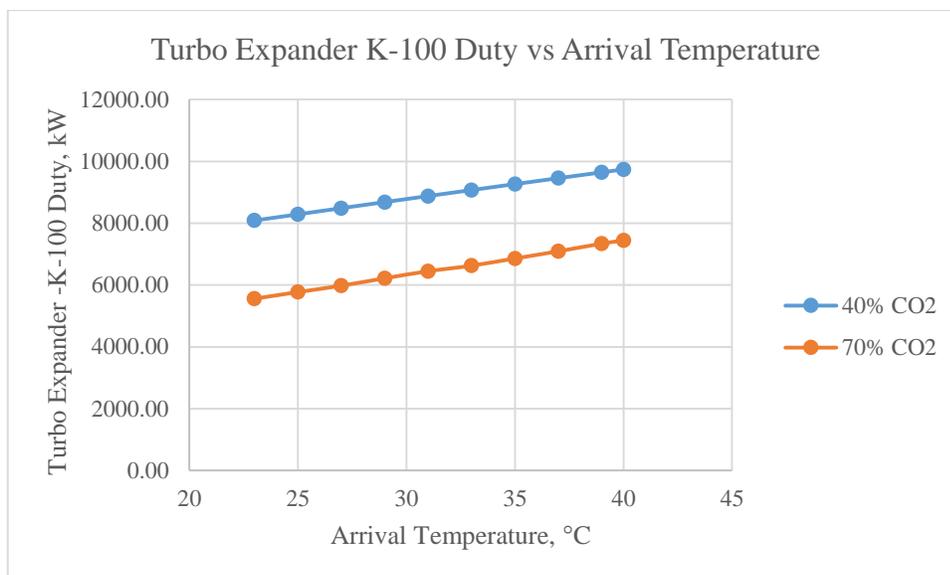


Figure 4.5-4 Graph of Turbo Expander K-100 Duty vs Arrival Temperature.

The arrival temperature of natural gas does not significantly affecting the total duty required for the production of the natural gas. From Figure 4.5-1 Graph of Total Duty vs Arrival Temperature., it can be clearly seen that for both cases 40% and 70% the line remains almost horizontal with respect to small variation. Yet, still most of the total energy required are from the recompression of natural gas which agrees with Nguyen et al. 2013. From the simulation it can be said that the variation of the arrival temperature will not have significant effect on the process of the carbon dioxide recovery. This is mainly because the variation of the arrival temperature does not intervene the operating condition of the 3 main units for carbon dioxide recovery i.e the liquid twister, gas twister and CO₂ crystallizer.

Other parameters such as the export gas flow rate, hydrocarbon losses and composition of CO₂ and CH₄ also does not significantly vary with the variation of the temperature. This is because of the same reason as the effect on the total duty where the arrival temperature does not affecting CO₂ recovery process. However, the variation of the temperature will affect the heating and cooling system of the plant. But, the heating and cooling system are not part of this study.

Although the arrival temperature does not affect the total duty of the process facility significantly, it is worth to point that the Turbo Expander duty, K-100 is quite

significantly affected by the variation of arrival temperature. This can be explained with the concept of work done by the isentropic expansion. At higher temperature, the gas will have higher enthalpy. When expanded, the gas will cool down and the enthalpy will be reduced. The different in enthalpy is the work done by the expansion. With an expansion to a constant pressure, higher temperature of the arrival natural gas will mean that higher work will be produced from the isentropic expansion of the natural gas.

In places as such Malaysia which temperature variation is not that big this wouldn't be a big issues. But for places that have a really low temperature during the winter and high temperature in the summer, the variation of the arrival temperature will have significant effect on the turbo expander, K-100 duty hence will affect the total duty of the production facility. A need of a heater or a cooler upstream of Inlet Separator, V-100 might be required to be address if the temperature variation is big.

4.6 Case 4: Effects of Variation in Arrival Pressure

The fourth case study investigate the effects of different arrival pressure of the natural gas. The value of arrival pressure is manipulated from 105 bar to 138.9 bar the base case. However, the restriction of the arrival is not to be 105 bar and below since the usage of the turbo expander K-100. In this production facilities, turbo expander, K-100 will expand the natural gas from the arrival temperature 138.9 bar (base case) to 105 bar. The turbo expander will not producing any work if the pressure is less than 105 bar.

The results of the simulation cases are tabulated in the following table.

Table 4.6-1 Simulation Results for Different Arrival Pressure of Natural Gas.

	Pressure, bar	Total Duty, kW	Export Gas Specifications			Hydrocarbon Loss, %
			Molar Flow, MMSCFD	CO ₂ , %	CH ₄ , %	
40% CO ₂	105	Simulation Does Not Converge				
	110	246602.11	1149	5.72	89.38	8.41
	115	243941.94	1149	5.72	89.38	8.41
	120	241555.88	1149	5.72	89.38	8.40
	125	239400.27	1149	5.72	89.38	8.39
	130	235672.20	1149	5.72	89.38	8.36
	138.9	234394.63	1149	5.72	89.38	8.35
70% CO ₂	105	Simulation Does Not Converge				
	110	175678.01	603.8	7.65	81.7	9.91
	115	173596.99	603.6	7.65	81.7	9.93
	120	171780.12	603.6	7.65	81.7	9.94
	125	170205.35	603.6	7.65	81.7	9.94
	130	168813.27	603.6	7.65	81.7	9.93
	138.9	166693.06	603.8	7.65	81.71	9.91

Same with the previous case, the variation of the arrival pressure has a significant effect on the duty of the turbo expander, K-100. The following table summarizes the duty of the turbo expander, K-100.

Table 4.6-2 Turbo expander -K100 duty at different arrival pressure.

CO ₂	Turbo Expander - K-100, Duty, kW					
	110 bar	115 bar	120 bar	125 bar	130 bar	138.9 bar
40%	1791	3437	4953	6353	8856	9741
70%	1464	2768	3935	4986	5939	7442

The data obtained are depicted on the following graphs in order to see the trends in the result.

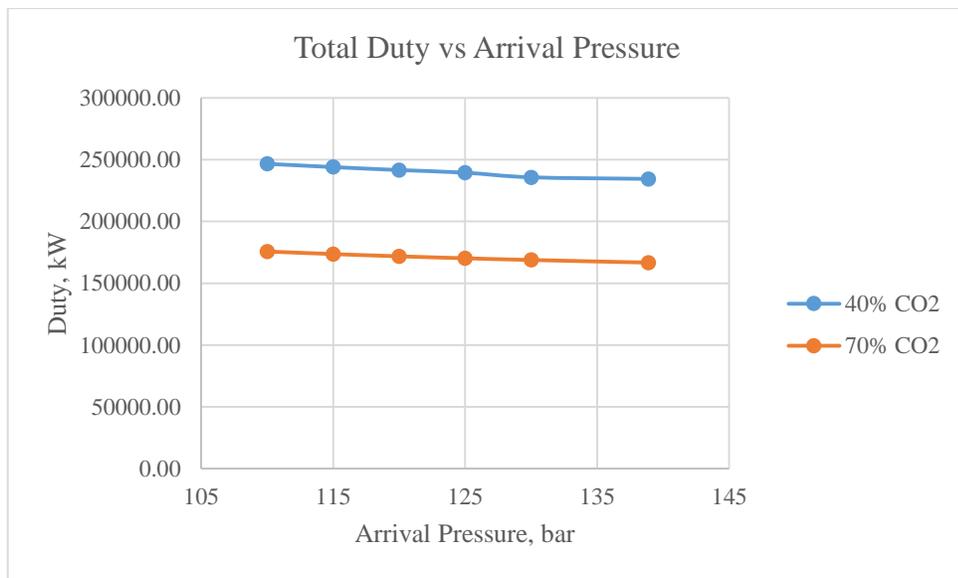


Figure 4.6-1 Graph of Total Duty vs Arrival Pressure.

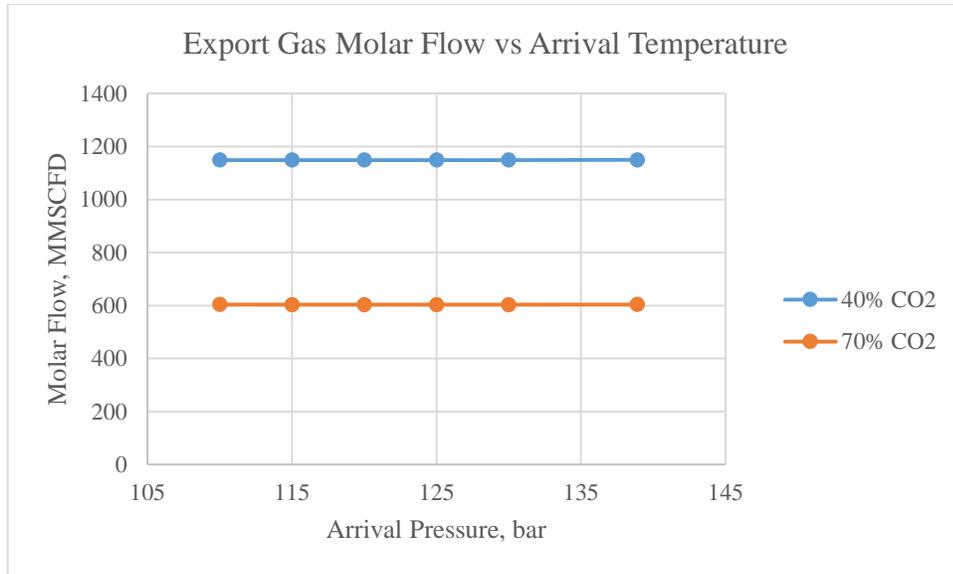


Figure 4.6-2 Graph of Export Gas Molar Flow vs Arrival Pressure.

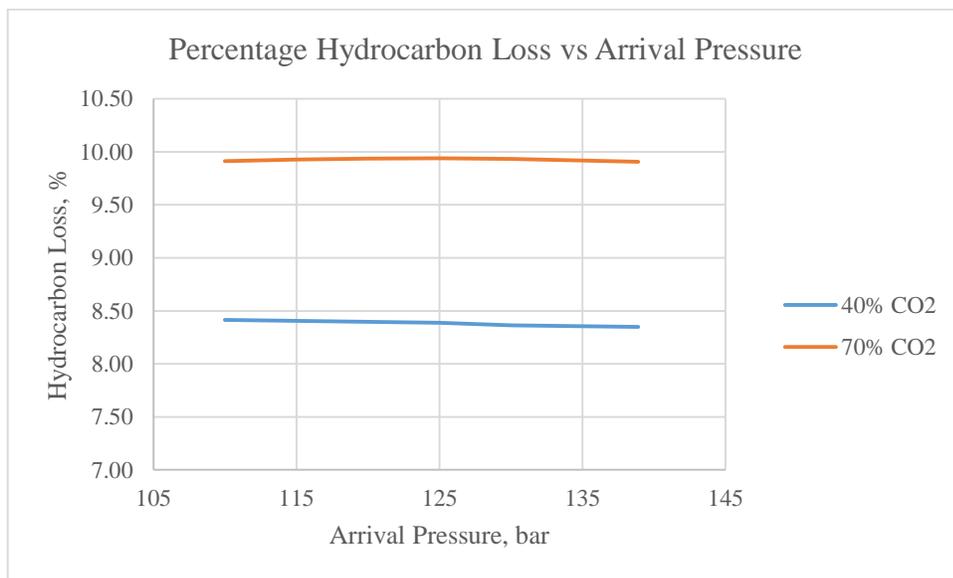


Figure 4.6-3 Graph of Percentage Hydrocarbon Loss vs Arrival Pressure.

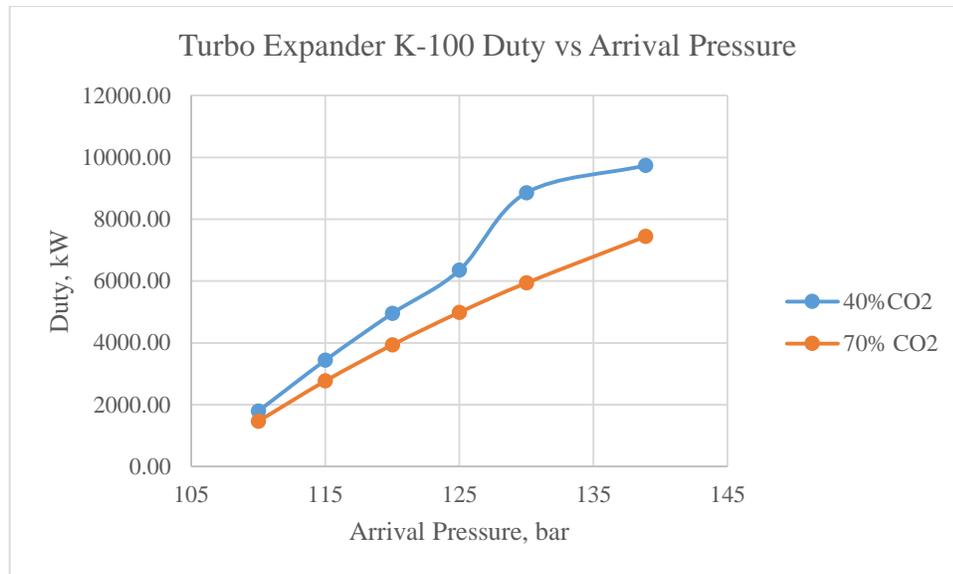


Figure 4.6-4 Graph of Turbo Expander K-100 Duty vs Arrival Pressure.

From Figure 4.6-1 Graph of Total Duty vs Arrival Pressure. it can be clearly seen that at low pressure the total duty required by the production facilities are greater than at higher pressure. This can be relate to the effect of the arrival temperature variation to the turbo expander, K-100 duty. In which at lower pressure the duty of the turbo expander is very low. As the turbo expander produces low duty, the production facility will require more power for the recompression of the export gas hence increasing the total energy consumption of the system.

The effect of arrival temperature on the duty of the turbo expander, K-100 can be explained using the work done by the isentropic expansion as well. At higher pressure greater work will be produced as the natural gas will have higher enthalpy. So, with the expansion to the same final pressure it is expected that higher pressure will produce greater work as compared to lower pressure.

Same with the variation of pressure, variation of arrival pressure does not affect the export gas molar flow rate and the concentration of CH₄ and CO₂ in the export gas stream. This is due to the same reason of the previous case, in which in this case the arrival pressure of the natural gas does not affect the process of the carbon dioxide recovery in the three major unit in which the recovery of CO₂ takes place. The reasons

are also true for the percent hydrocarbon loss in the system which are not really affected by the variation in arrival temperature.

Chapter 5: Conclusion

The aim of this research is to study the energy consumption for carbon dioxide removal of Gas Twister system in natural gas production. In order to achieve the project objectives, steady state simulation of a natural gas production facility that uses Gas Twister system is developed. Four case studies are then conducted to study the effect of variation in process parameters on the total energy consumption.

In the first case study, the CO₂ loading are manipulated and with increase in CO₂ content in the natural gas, energy consumption of the process facility decreases as less export gas are being recovered. The same condition is also true for the second case study. Instead of only manipulating only the CO₂ and methane composition, the whole composition of the natural gas is manipulated in the second case study. The same results are obtained as in the first case study.

The third and the fourth case study are aimed to investigate the effect of variation in arrival temperature and arrival pressure of the natural gas respectively. In both cases, the variation in both parameters doesn't affect much on the total energy consumption of the production facility which may be due to small variation. But it is worthy to note that, these two variations has an impact on the total duty produced by the turbo expander, K-100. With the drop of pressure and temperature, the work done by the compressor will be reduced and with big variation in the two parameters this will cause a huge problem. Variation in pressure can be avoided since the production facility is equipped with CO₂ reinjection system. A need of a heater or a cooler upstream of Inlet Separator, V-100 might be required to be address if the temperature variation is big.

For future recommendations, the error of the simulation can be reduced by simulating the proprietary technologies using the actual unit in Aspen HYSYS instead using other equipment in a sub-flow sheet to represent them. In order to get a more precise energy consumption, the process boundary of this project can be enlarge to include the utilities systems that are ignored. In future work as well, the cooling and heating duties of the system can also be considered with respect of the same variation in the parameters. This study also does not address the amount of fuel gas needed to drive the compressor and pumps.

This study proves that the Gas Twister system is a reliable carbon dioxide removal method especially for a high CO₂ gas fields. Compact design, smaller footprint (less equipment), no liquid or solid handling made Gas Twister system a step ahead amongst other technology. With this study it is hope that the monetization of high CO₂ gas fields both in Peninsular Malaysia and Offshore Sabah and Sarawak will be possible as the profit from those gas fields will bring benefits to the country – Malaysia.

Chapter 6: References

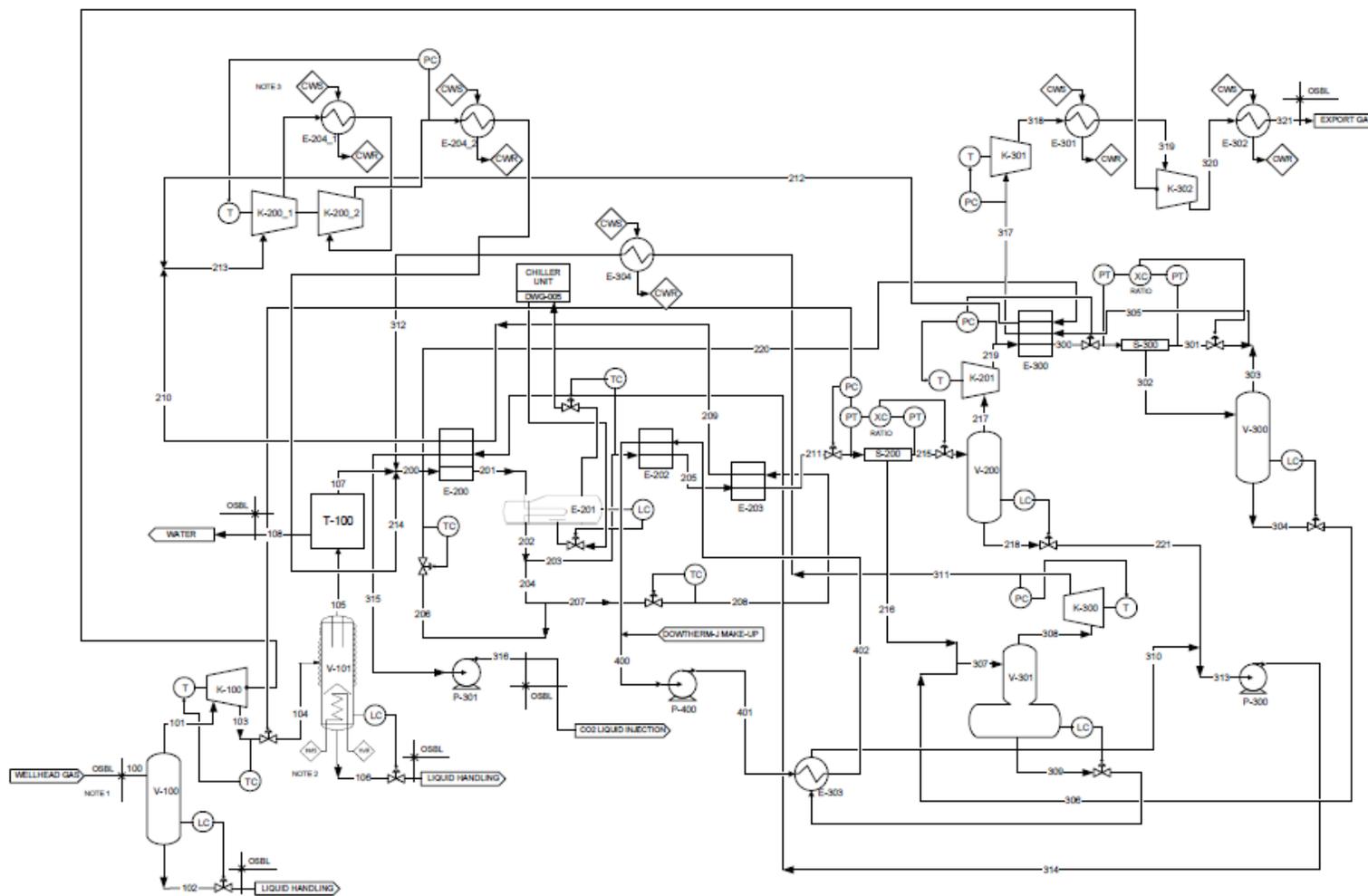
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Chapter 7: Appendix

- Appendix 1: Process Flow Diagram of the Gas Twister System
- Appendix 2: Natural Gas Composition
- Appendix 3: HYSYS subflowsheet for Hydrate SeparatorTM, V-101
- Appendix 4: HYSYS subflowsheet for Liquid Twister, S-200
- Appendix 5: HYSYS subflowsheet for Gas Twister, S-300
- Appendix 6: HYSYS subflowsheet for CO₂ Crystallizer, V-301
- Appendix 7: Heat and Material Balance of Simulation by TwisterBV
- Appendix 8: Heat and Material Balance of Base Case Simulation

Process Flow Diagram of a Topside Natural Gas Production Facility Using Gas Twister System.



Natural Gas Composition

For Case 2: Effects of Different Natural Gas Composition

10% Case

Component	Mole Fraction
N2	0.00028
CO2	0.09998
H2S	0.00000
C1	0.76889
C2	0.07103
C3	0.03404
i-C4	0.00544
n-C4	0.00721
i-C5	0.00349
n-C5	0.00207
C6	0.00125
C7	0.00506

20% Case

Component	Mole Fraction
N2	0.00025
CO2	0.20001
H2S	0.00000
C1	0.68166
C2	0.06316
C3	0.03043
i-C4	0.00491
n-C4	0.00654
i-C5	0.00220
n-C5	0.00193
C6	0.00123
C7	0.00528

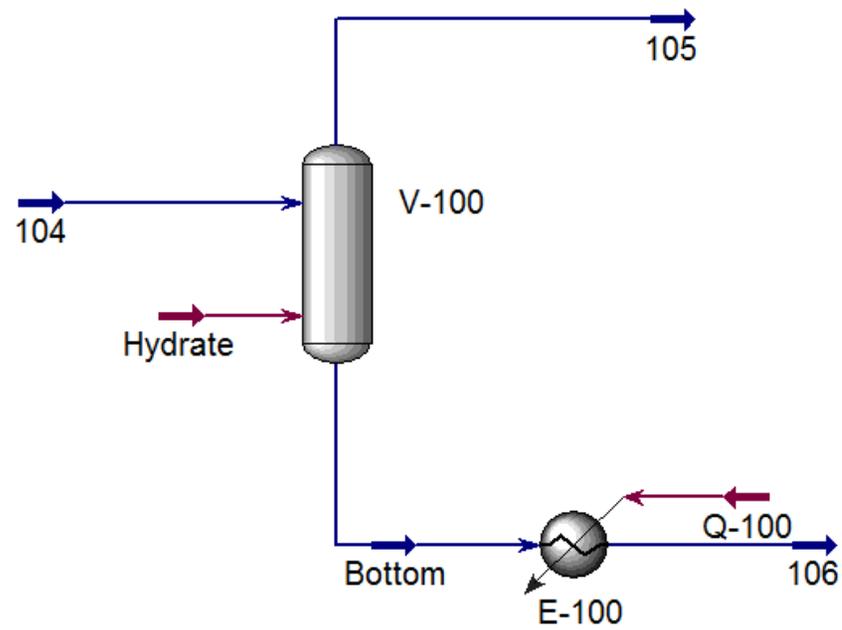
40% Case

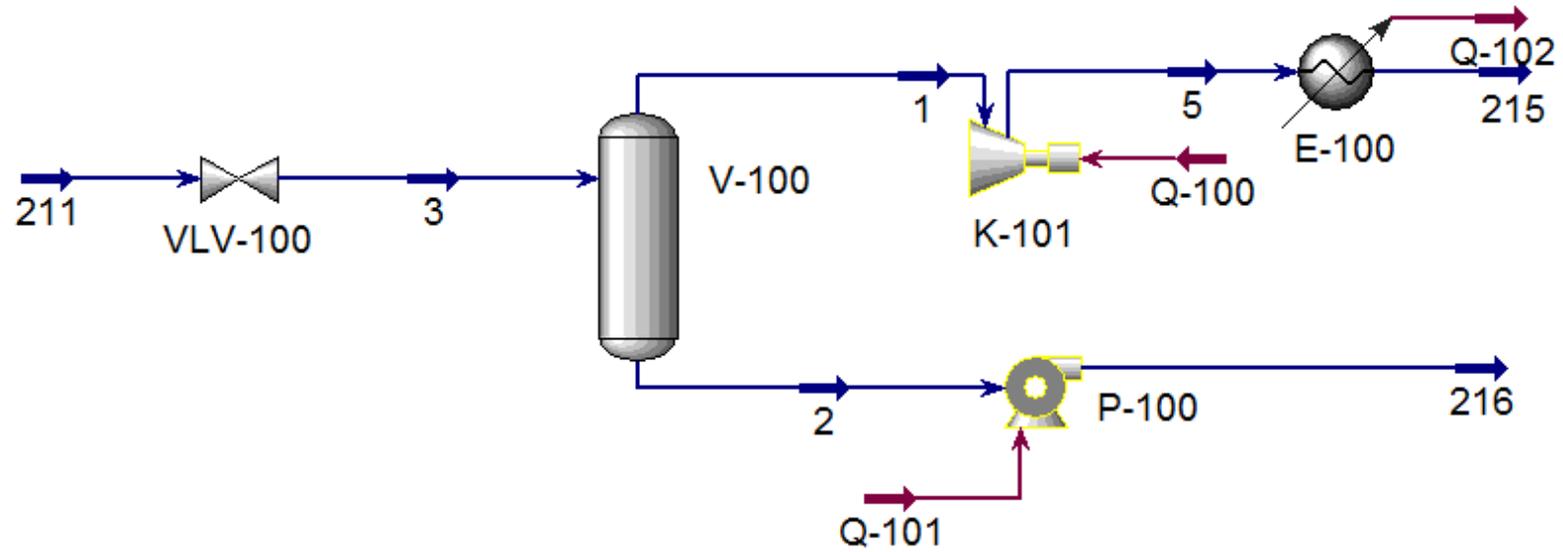
Component	Mole Fraction
N2	0.01994
CO2	0.40782
C1	0.52847
C2	0.02493
C3	0.00760
i-C4	0.00270
n-C4	0.00171
i-C5	0.00100
n-C5	0.00060
n-C6	0.00070
methylcyclopentane	0.00000
benzene	0.00010
cyclohexane	0.00000
n-C7	0.00090
methylcyclohexane	0.00000
toluene	0.00002
n-C8	0.00040
E-benzene	0.00000
m-Xylene	0.00003
o-Xylene	0.00002
n-nonane	0.00020
1,2,4-Methylbenzene	0.00000
n-Decane	0.00010
n-C11	0.00003
C12+	0.00006
H2O	0.00267

50% Case

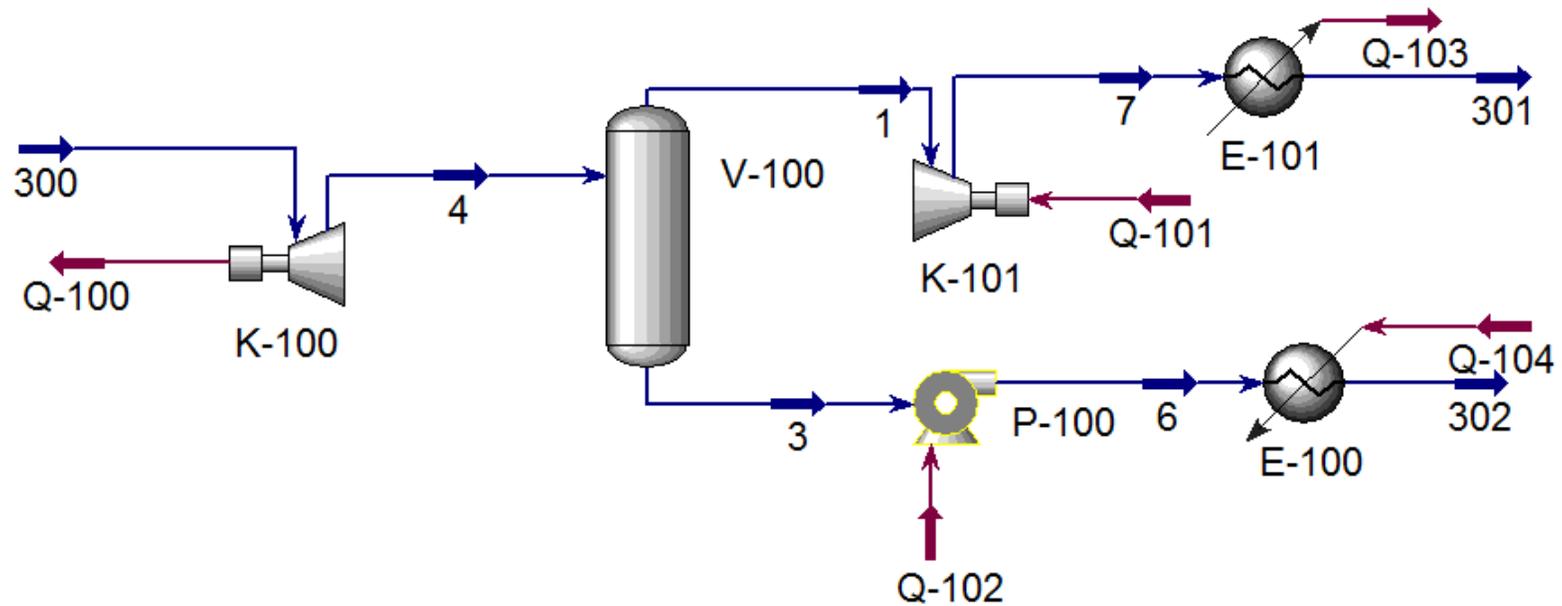
Component	Mole Fraction
N2	0.0050
CO2	0.5060
H2S	0.0000
C1	0.3970
C2	0.0350
C3	0.0240
i-C4	0.0090
n-C4	0.0090
i-C5	0.0060
n-C5	0.0060
C6	0.0020
C7	0.00000

HYSYS subflowsheet of Hydrate Separator, V-100

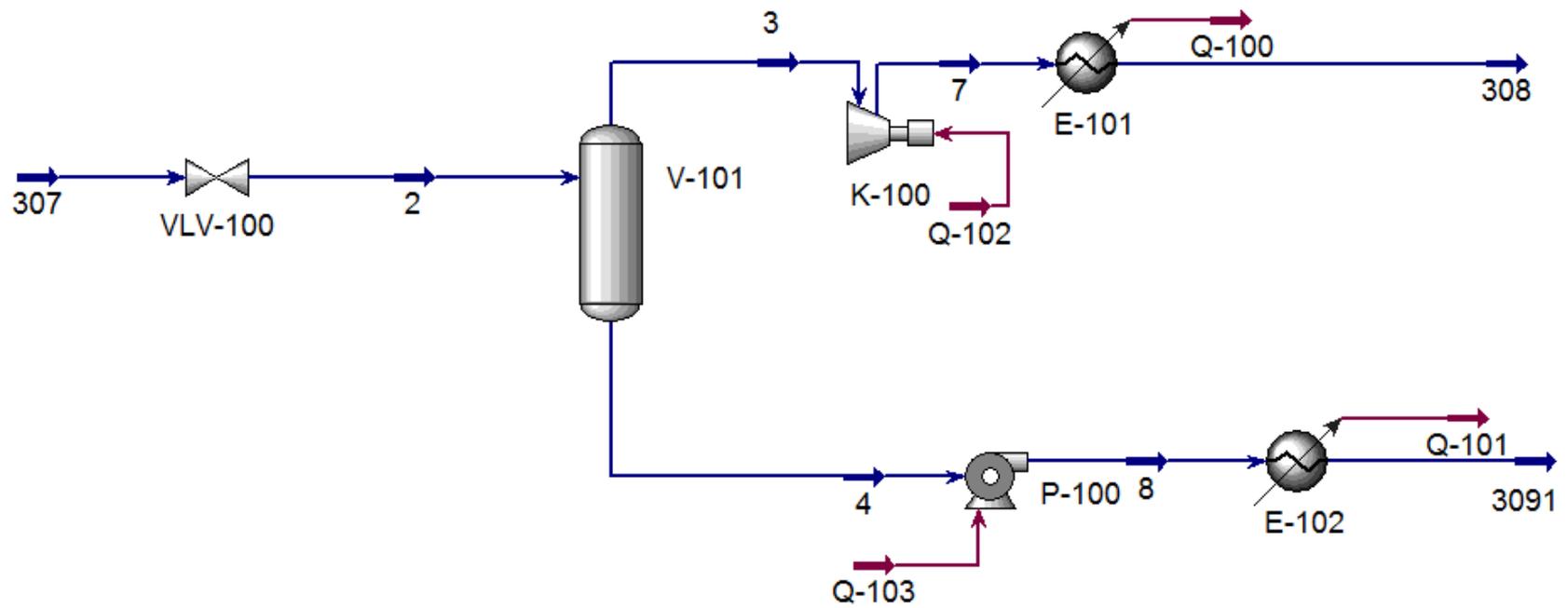


HYSYS subflowsheet of Liquid Twister, S-200

Both compressor, K-101 and pump, P-100 are not converged. This is because of the gas and liquid carry over introduced in the separator V-100. This can be ignored since the two equipment are being used to represent the recompression in Liquid Twister.

HYSYS subflowsheet of Gas Twister, S-300

Both compressor, K-101 and pump, P-100 are not converged. This is because of the gas and liquid carry over introduced in the separator V-100. This can be ignored since the two equipment are being used to represent the recompression in Gas Twister.

HYSYS subflowsheet of CO₂ Crystallizer, V-301

Heat and Material Balance for the Simulation Developed TwisterBV

HEAT & MATERIAL BALANCE - CASE 1: FIELD X 8% Liquid & Gas Twister																														
Property / Stream Number	100	101	102	103	104	105	106	107	108	200	201	202	203	204	205	206	207	208	209	210	211	212	213	214	215	216	217	218		
Vapour Fraction	0.996	1.000	0.000	0.998	0.947	1.000	0.000	1.000	0.000	1.000	0.405	0.022	0.022	0.022	0.000	0.022	0.022	0.501	0.620	1.000	0.000	1.000	1.000	1.000	0.950	0.087	1.000	0.000		
Pressure [bar]	138.9	138.9	138.9	105.0	81.0	81.0	81.0	81.0	81.0	80.0	79.7	79.3	79.3	79.3	79.0	79.3	79.3	14.5	14.2	13.8	78.6	14.2	13.8	80.0	21.4	10.0	21.4	21.4		
Temperature [C]	40.0	40.0	40.0	24.9	11.8	13.7	19.3	13.9	13.9	18.2	-4.8	-20.0	-20.0	-20.0	-47.0	-20.0	-20.0	-55.4	-50.1	15.0	-50.5	17.0	18.6	40.0	-62.5	-76.4	-62.5	-62.5		
Molar Flow [MMSCFD]	200.0	1952.0	8.0	1952.0	1992.0	1980.1	11.9	1978.2	1.9	2834.3	2834.3	2324.1	510.2	2324.1	155.2	355.0	355.0	355.0	355.0	355.0	2324.1	155.2	510.2	510.5	769.6	1554.6	731.0	38.0		
Mass Flow [kg/h]	3650831	3606171	44660	3606171	3609171	3575809	51082	3573387	1701	5002794	5002794	4102291	900503	4102291	273896	626607	626607	626607	626607	626607	4102291	273896	900503	901262	921794	3180487	842051	79743		
Heat Flow [kW]	8291193	-8247012	-44180	-8254464	-8254464	-8188798	34065	-8191234	-7563	-11318758	-11486796	-11589105	-9503066	-2086039	-9590802	-834488	-1451151	-1451151	-1442018	-1402742	-9600335	-612114	-2014857	-2027431	-1619229	-7981108	-1420354	-198876		
Master Comp Mole Frac (Nitrogen)	0.02927	0.02935	0.00927	0.02935	0.02935	0.02951	0.00214	0.02954	0.00000	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	0.02974	
Master Comp Mole Frac (CO2)	0.88954	0.89021	0.52028	0.89021	0.89021	0.89206	0.28494	0.89332	0.00000	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	0.89372	
Master Comp Mole Frac (Methane)	0.25852	0.25902	0.13377	0.25902	0.25902	0.26032	0.04375	0.26057	0.00000	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	0.26824	
Master Comp Mole Frac (Ethane)	0.00925	0.00925	0.00926	0.00925	0.00925	0.00925	0.00928	0.00929	0.00000	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	0.01158	
Master Comp Mole Frac (Propane)	0.00350	0.00349	0.00572	0.00349	0.00349	0.00349	0.00393	0.00349	0.00000	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	0.00341	
Master Comp Mole Frac (n-Butane)	0.00071	0.00071	0.00160	0.00071	0.00071	0.00070	0.00133	0.00070	0.00000	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064	0.00064
Master Comp Mole Frac (i-Butane)	0.00061	0.00061	0.00160	0.00061	0.00061	0.00060	0.00144	0.00061	0.00000	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054	0.00054
Master Comp Mole Frac (n-Pentane)	0.00032	0.00032	0.00114	0.00032	0.00032	0.00031	0.00121	0.00031	0.00000	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027	0.00027
Master Comp Mole Frac (i-Pentane)	0.00022	0.00021	0.00087	0.00021	0.00021	0.00021	0.00099	0.00021	0.00000	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018	0.00018
Master Comp Mole Frac (n-Hexane)	0.00035	0.00034	0.00218	0.00034	0.00034	0.00033	0.00093	0.00034	0.00000	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028	0.00028
Master Comp Mole Frac (Methylpentane)	0.00012	0.00012	0.00077	0.00012	0.00012	0.00011	0.00115	0.00011	0.00000	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010
Master Comp Mole Frac (Benzene)	0.00028	0.00027	0.00190	0.00027	0.00027	0.00025	0.00209	0.00026	0.00000	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022	0.00022
Master Comp Mole Frac (Cyclohexane)	0.00012	0.00012	0.00088	0.00012	0.00012	0.00011	0.00137	0.00011	0.00000	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010	0.00010
Master Comp Mole Frac (n-Heptane)	0.00018	0.00017	0.00161	0.00017	0.00017	0.00016	0.00269	0.00016	0.00000	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014	0.00014
Master Comp Mole Frac (Toluene)	0.00020	0.00019	0.00225	0.00019	0.00019	0.00017	0.00403	0.00017	0.00000	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015	0.00015
Master Comp Mole Frac (n-Octane)	0.00035	0.00033	0.00458	0.00033	0.00033	0.00028	0.00093	0.00028	0.00000	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024	0.00024
Master Comp Mole Frac (E-Benzene)	0.00002	0.00002	0.00027	0.00002	0.00002	0.00001	0.00055	0.00001	0.00000	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001
Master Comp Mole Frac (m-Xylene)	0.00001	0.00001	0.00015	0.00001	0.00001	0.00001	0.00031	0.00001	0.00000	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001
Master Comp Mole Frac (p-Xylene)	0.00000	0.00000	0.00007	0.00000	0.00000	0.00000	0.00015	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Master Comp Mole Frac (n-Nonane)	0.00035	0.00032	0.00613	0.00032	0.00032	0.00024	0.01388	0.00024	0.00000	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021	0.00021
Master Comp Mole Frac (12-M-Benzene)	0.00003	0.00002	0.00071	0.00002	0.00002	0.00001	0.00152	0.00001	0.00000	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001
Master Comp Mole Frac (n-Decane)	0.00013	0.00012	0.00357	0.00012	0.00012	0.00008	0.00755	0.00008	0.00000	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007	0.00007
Master Comp Mole Frac (n-C11)	0.00009	0.00008	0.00301	0.00008	0.00008	0.00004	0.00655	0.00004	0.00000	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003	0.00003
Master Comp Mole Frac (C12+)	0.000159	0.00047	0.28354	0.00047	0.00047	0.00001	0.07597	0.00001	0.00000	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001	0.00001
Master Comp Mole Frac (H2O)	0.00007	0.00008	0.00148	0.00008	0.00008	0.00008	0.00096	0.00008	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Master Comp Mole Frac (DTRM-J)	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000

HEAT & MATERIAL BALANCE - CASE 1: FIELD X 8% Liquid & Gas Twister																											
Property / Stream Number	219	220	221	300	301	302	303	304	305	306	307	308	3														

