OPTIMIZED DESIGN OF SHALE GAS PROCESSING AND NGL RECOVERY PLANT UNDER UNCERTAIN FEED CONDITIONS

A Thesis

by

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ABSTRACT

Shale gas is an increasingly booming resource and it has been predicted to increase from 1% in 2000 to 40% in 2035 of the total US domestic gas produced. Since shale gas is both industrially economical and environmentally clean compared to oil or coal as a resource, many studies are focused on developing technologies to monetize shale gas. However, one of the key challenges in utilizing shale gas is its fluctuating flow rate and compositional behavior. The flow rate of a shale gas well dwindles over a period of time and the composition of shale gas differs from well to well in the same shale play. This provides a challenge in designing a plant of optimum size for a shale gas processing and NGL recovery plant.

In this study, this uncertainty in shale gas feed flow rate and composition is addressed while designing a shale gas processing and NGL recovery plant. First, different shale gas flow rates are chosen over a period of shale gas well life based on the average shale gas rate declination curve of a shale play. Second, two different process flow sheets are developed (i) using conventional technology and (ii) using novel technology. In the novel technology, the NGL recovery section of the conventional technology is modified to accommodate novel changes such as using a divided wall column or pre-fractionated sequence to separate methane, ethane, and propane. Later, these process flow sheets are simulated in Aspen plus for comparing the economics of different plant sizes. Furthermore, heat integration and optimization of individual units of the process flow sheets are carried out using pinch and sensitivity analyses, respectively. Lastly, the

economic analysis of a plant of optimum size with constant feed flow rate over its plant life is evaluated. In this case, shale gas from different wells is collected in a header and adjusted such that the shale gas flow rate is constant to the plant. Environmental impact of the process is also observed.

From the economic analysis of various cases for conventional and novel technology, it is observed that case-3 provides the optimum plant design with highest ROI percentage compared to other cases and for case-3, novel technology ROI is 4.17% more compared to conventional technology. Finally, constant production rate case, at the flow rate of case-3, the ROI percentage is observed to be more than minimum requirement implying that this processing plant is economically viable to implement.

DEDICATION

To My Parents, Brother and Grandparents

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All work for the thesis was completed by the student, under the advisement of Professor Mahmoud EI-Halwagi of the Department of Chemical Engineering.

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CHAPTER I

INTRODUCTION AND LITERATURE REVIEW

1.1 Introduction

Shale gas is natural gas trapped within shale formation, and it is derived from underground shale deposits that are broken up by hydraulic fracturing and horizontal drilling. Hydraulic fracturing is a process in which pressurized liquid is injected into a wellbore to create cracks in deep rock formations through which deposits like shale gas, natural gas, and petroleum will flow and, Horizontal drilling technique employs slant type of well boring.

Shale gas is one of the promising sources for producing Natural Gas (NG) in the United States. Production of the shale gas has increased from less than 1% of domestic gas in the United States in 2000 to over 20% by 2010 and it is predicted that shale gas will account for 46% of United States gas supply by 2035 (Stevens, 2012). Life-cycle emissions for shale gas is 6% lower than conventional natural gas, 23% lower than gasoline, and 33% lower than coal. In addition, the energy available per unit of shale gas is more compared to oil or coal. Hence, Shale gas is an industrially economical and environmentally clean compared to other resources.

Shale gas feed contains Methane and Natural Gas Liquids (C_2, C_3, C_4 and C_5+) compounds among other impurities such as H_2O, H_2S, CO_2, N_2, and inert gases. The composition of shale gas varies between different shale gas reservoirs also; two wells in the same field may yield gaseous products that are different in composition. Table 1 shows average shale gas composition for different shale gas reservoirs. Table 2 and Table 3 show how composition varies between different wells in Barnett and Marcellus shale plays respectively.

Table 1: Average shale gas compositions from wells of different shale plays (Bullin & Krouskop, 2009)

Reservoir	C1	C2	C3	CO2	N2
Barnett	86.8	6.7	2.0	1.7	2.9
Marcellus	85.2	11.3	2.9	0.4	0.3
Fayetteville	97.3	1.0	0.0	1.0	0.7
New Albany	89.9	1.1	1.1	7.9	0.0
Atrium	62.0	4.2	1.1	3.8	29.0
Haynesville	95.0	0.1	0.0	4.8	0.1

Table 2: Barnett shale gas composition (Bullin & Krouskop, 2009)

Well	C1	C2	C3	CO2	N2
1	80.3	8.1	2.3	14	7.9
2	81.2	11.8	5.2	0.3	1.5
3	91.8	4.4	0.4	2.3	1.1
4	93.7	2.6	0.0	2.7	1

Table 3: Marcellus shale gas composition (Bullin & Krouskop, 2009)

Well	C1	C2	C3	CO_2	N_2
1	79.4	16.1	4.0	0.1	0.4
2	82.1	14.0	3.5	0.1	0.3
3	83.8	12.0	3.0	0.9	0.3
4	95.5	3.0	1.0	0.3	0.2

Shale gas processing generally involves removal of oil and dirt or sand, water, elements such as sulfur, helium and carbon dioxide, and recovering natural gas liquids (NGL). There are many chemical processing technologies available to refine shale gas. Depending upon various process variables, the choice of a process technology to be employed is decided. While making this decision several factors must be considered such as (Speight, 2013)

- The types and concentrations of contaminants in the gas
- The degree of contaminant removal desired
- The selectivity of acid gas removal required
- The process conditions such as temperature and pressure
- The composition and volume of the gas that is to be processed
- The desirability of sulfur recovery due to process economics or environmental issues

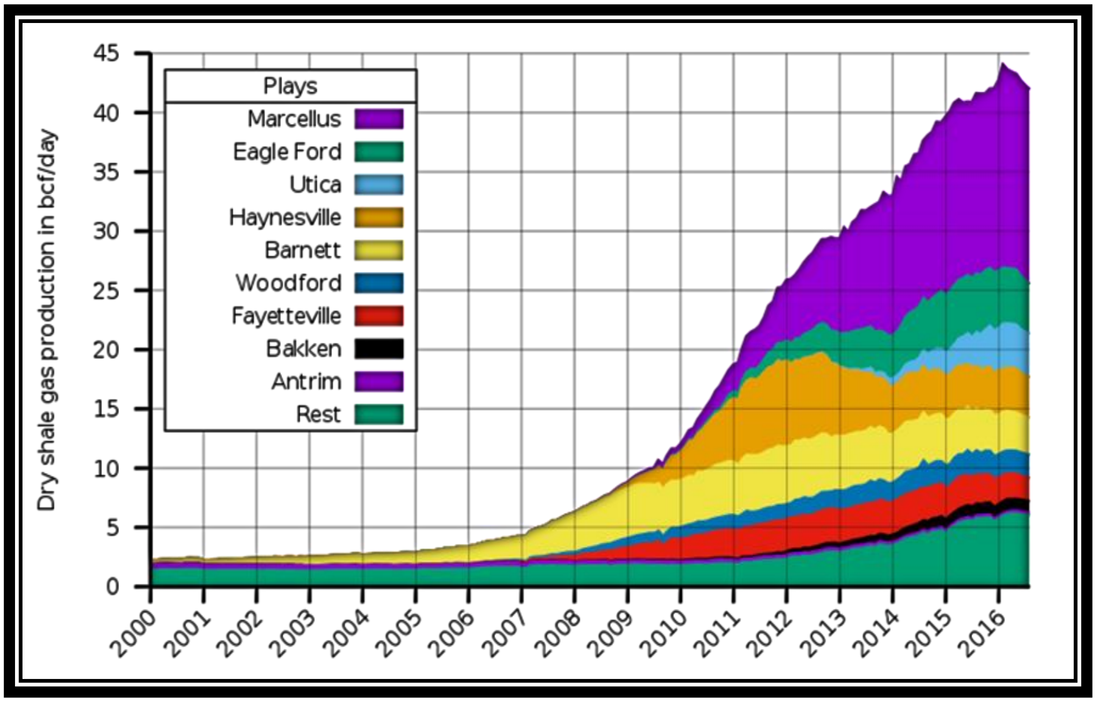

Figure 1: Shale gas production by geological formation

From Figure 1, it can be inferred that Barnett is one of the longest shale gas producing reserve and Marcellus shale gas reserve is the largest producer of the shale gas in the US. Hence, in this study, one of these two shale plays data will be considered.

The percentage of impurities specification in the shale gas feed varies based on the feedstock consumer. If the consumer is a power plant or a plant that uses raw material in a gas form generally require pipeline quality gas and, the cryogenic gas plants or the liquefaction gas plants require more stringent feed gas specifications. Table 4 gives the specification of feed gas for pipeline quality and feed to cryogenic gas plant. The major products and by-products from the process must also adhere to the requirements of the user. Table 5 shows the specification of products and by-products used in this work.

Table 4: : Pipeline gas quality and feed to cryogenic gas plant specifications (Al-Sobhi & Elkamel, 2015; Mokhatab & Poe, 2012; Rufford et al., 2012)

Feed Criteria	CO_2 content	H_2S content	Nitrogen content	Water content
Pipeline gas quality	2 mole%	0.25–1.0 grain/100 scf	< 3 mole%	4–7 lbm H_2O/ MMscf of gas
Feed to cryogenic gas plant	50 ppmv	5 ppmv	1 mole%	1 ppmw

Table 5: Specification of products and by-products for plant design (Al-Sobhi & Elkamel, 2015)(Johnson)

Component	Methane	Ethane	Propane	Butane
Methane	98 mole%	2 mole%	1 ppmw	0
Ethane		96 mole%	2 mole%	1 ppmw
Propane		2 mole%	95 mole%	2 mole%
Butanes		0	3 mole%	97 mole%

1.2 Literature Review

There are many studies concentrating on shale gas monetization in recent years. Ehlinger et al. developed a process for the production of methanol from shale gas (Ehlinger, Gabriel, Noureldin, & El-Halwagi, 2013). Andrea et al. designed two different processes for converting shale gas to ethylene (Ortiz-Espinoza, Noureldin, El-Halwagi, &

Jiménez-Gutiérrez, 2017), which mainly concentrates on designing economically viable alternatives to convert raw shale gas to salable products.

However, there are limited studies addressing the fluctuating behavior of shale gas. Getu et al. (Getu, Mahadzir, Long, & Lee, 2013; Wang & Xu, 2014) studied the effect of feed gas composition variation in deciding optimum NGL recovery process scheme. In their work, the feed composition varies and the flow rate is assumed constant for all different feed scenarios. All the feed scenarios are categorized into either lean feed or rich feed based on C_2 and C_3 composition.The profitability analysis comparison between all different process schemes is carried out by taking average values calculated for lean and rich feed. In a paper by Wang et al. (Wang & Xu, 2014) uncertainty in shale gas feed rate is addressed by assuming that shale gas feed flow rate variation follows a standard normal distribution. In this paper, five discrete feed rate cases are selected based on probability. The optimum operating conditions are determined using an objective function that represents an operating cost, based on the sum product of probability with shale flow rate at five discreet feed rates. In this study, plant size is constant and variation is only seen in the operating cost. As a result, the final optimal plant design is not expected to handle the feedstocks in all the scenarios. Gong et al. designed an intensified shale gas processing and NGL recovery system assuming composition of each well site raw shale gas as constant and feedstock composition uncertainty is caused by unstable flowrates of shale gas from different well sites (Gong, Yang, & You, 2017). In this work, the capacity of each equipment of intensified design is selected as the largest value among the operating levels of the same unit in process design. Designing for maximum capacity during uncertain feed scenario may not guarantee the optimum plant design.

Given the drastic drop in shale gas wells flowrate with time, as shown in Figure 2, it is imperative to consider varying plant size while designing against uncertain feed conditions for an economically optimum plant design. In my work, this problem will be addressed by selecting five different discrete flow rate cases along the life of shale gas well, based on the drop in flow rate and important milestones, to determine the optimum plant size.

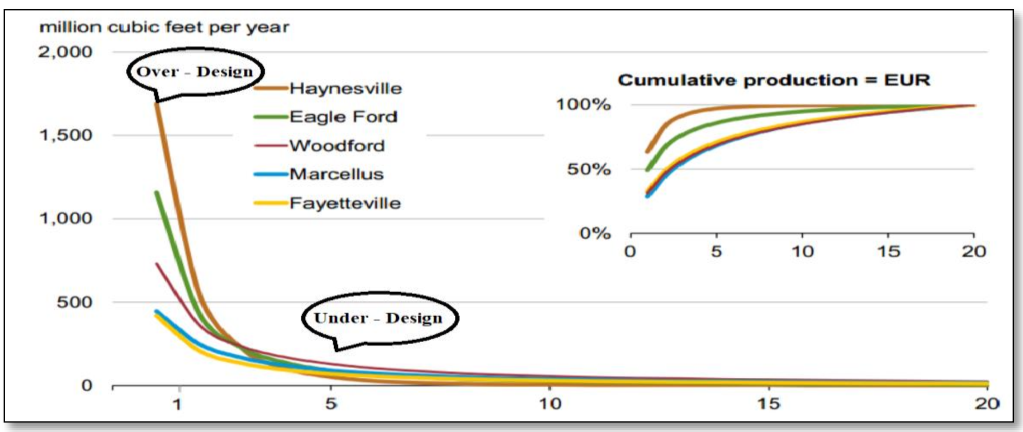

Figure 2: Average well decline rate curve for various shale plays

CHAPTER II

PROCESS TECHNOLOGIES

2.1 Process Overview

Shale gas is mainly composed of Methane. It also contains a minor amount of impurities such as CO_2, H_2S, N_2, H_2O and other natural gas liquids. Methane and other NGLs such as ethane and propane are required to maintain certain pipeline specification for transmission through interstate and intrastate pipelines and to use as products respectively. Moreover, Methane, which is sent to the cryogenic unit is required to maintain very stringent CO_2 and dew point specifications compared to pipeline quality gas, since these impurities lead to icing and large energy consumption in Natural gas liquefaction or cryogenic gas plants. Hence, typical shale gas processing and NGL recovery plants contain four sections. As shown in Figure 3, the shale gas is first sent through an acid gas removal unit where the impurities like carbon dioxide and hydrogen sulfide are removed from raw shale gas to avoid freezing and corrosion of downstream units. Generally, a dehydration unit follows the acid removal unit where water is removed to meet specification required by downstream shale gas processing units to avoid icing in cryogenic units. The first cryogenic column is the de-methanizer, which is designed to remove methane from the higher boiling hydrocarbons. Later methane is sent to the nitrogen gas rejection unit to separate nitrogen and other lighter gas to meet methane product purity. Heavies from the de-methanizer column bottom are sent to the fractionator train unit that consists a series of cryogenic distillation columns in order to recover the natural gas liquids (NGLs). The second cryogenic column is the de-ethanizer and separates ethane from higher boiling hydrocarbons such as propane and butane followed by the de-propanizer to separate propane from butane (Ehlinger et al., 2013).

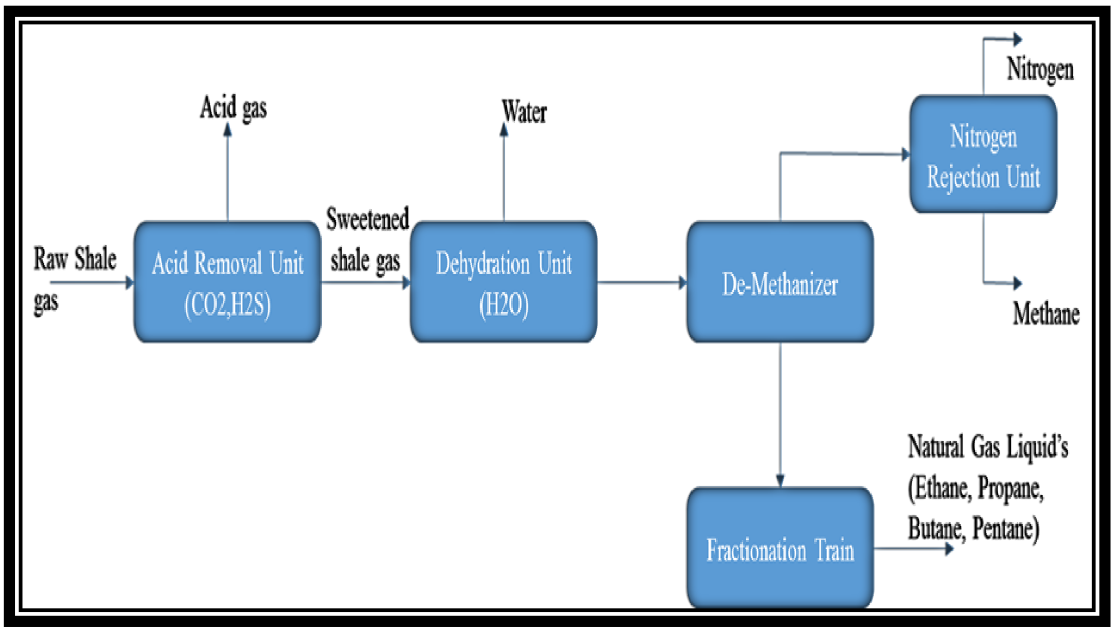

Figure 3: Block flow diagram of shale gas processing and NGL recovery

2.2 Shale Gas Processing Technologies

Figure 4 lists the available technologies for major units in shale gas processing and NGL recovery plants (Al-Sobhi & Elkamel, 2015; Bahadori, 2014; Mokhatab & Poe, 2012; L. Peters, A. Hussain, M. Follmann, T. Melin, & M.-B. Hägg, 2011). These technologies are further discussed in the sections below.

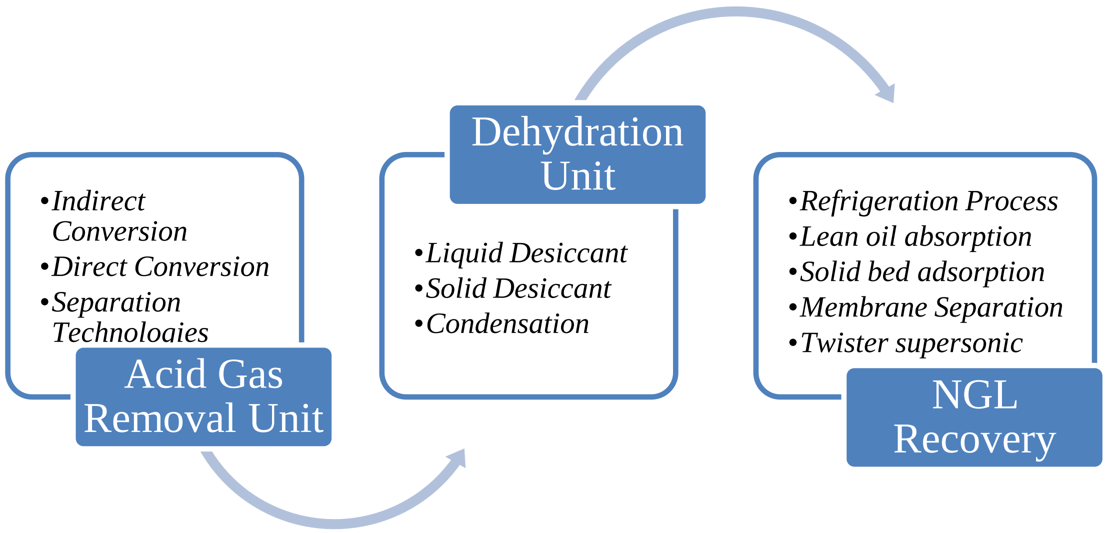

Figure 4: Available technologies for major process units

2.2.1 Acid Gas Removal Unit

The acid gas removal unit is designed to remove the acidic components to meet sales gas CO_2 and Sulphur specifications. Following are the various technologies available for acid gas removal,

❖ *Indirect Conversion*

The indirect conversion uses either adsorption or absorption process. Adsorption is a physical-chemical phenomenon in which impurities from the gas are removed by adhering physically or chemically to the surface of a selective solid. Whereas absorption is achieved by a physical phenomenon called dissolution, which uses a solvent to physically absorb impurities, or by a chemical phenomenon, that uses a reaction of solvent with the acid gases impurities.

- *Direct Conversion*

Direct conversion is used when a sulfur recovery is an option. In these processes, H_2S is absorbed in an alkaline solution containing a chelating agent and this is oxidized with air to form elemental sulfur.

- *Separation Technologies*

  - Membrane Systems

The membrane-based gas separation process largely depends on the gas components, membrane material, and the process conditions. Typically membrane unit is divided into two sections by a membrane, the stream leaving from the side where the feed is entering is called retentate and the stream leaving from another side of the membrane is called as permeate.

In a typical single stage membrane process, without any recycle streams. The crude shale gas flows over the feed side of the membrane.CO_2 in the gas stream permeates through the membrane to the permeate side. The retentate leaves the membrane with nearly the same pressure as the feed and on the permeate side, a permeate stream enriched with CO_2 leaves the membrane.

- *Cryogenic Fractionation*

Cryogenic separation involves cooling the acid gases to a very low temperature so that the acid gases such as CO_2 can be liquefied and separated from the feed gas stream. This technology requires pretreatment and dehydration of the feed gas to remove components that would result in hydrate formation and CO_2 freezing in the cold section of the fractionation equipment. Moreover, it requires substantial power to operate the refrigeration unit.

2.2.2 Dehydration

Gas from the acid gas removal unit is fed to the gas dehydration unit to meet the water dew point specification for pipeline transmission. In colder climate areas, the water dew point specification can be as low as -40^{o}F in order to avoid hydrate formation in the pipeline. Different types of dehydration methods are available depending on the plant capacity and extent of drying,

❖ *Liquid Desiccant*

This uses the absorption process where the liquid desiccant absorbs moisture from the feed gas. The liquid which is more desirable to use for commercial dehydration purpose should have properties such as high absorption efficiency, easy and economic regeneration, non-corrosive and non-toxic, no operational problems when used in high concentrations, no interaction with the hydrocarbon portion of the gas, and no contamination by gases. Some examples of liquid desiccants are glycols, particularly EG, DEG, TEG and tetra ethylene glycol.

❖ *Solid Desiccant*

This uses the adsorption process where water is adsorbed onto a solid surface from shale gas. This desiccant system generally has two or more towers. In a typical two tower system, one tower is on-stream adsorbing water from gas while the other tower is being regenerated and cooled. The towers are switched before the on-stream tower becomes saturated with water. Examples of typical desiccants are activated alumina and a granular silica gel material.

❖ *Condensation*

This method employs gas cooling to turn water molecules in the feed gas into a liquid phase and then removing this liquid water from the feed stream using flash column.

2.2.3 Nitrogen Rejection

Nitrogen Rejection can be done in two methods,

❖ *Cryogenic Processes*

This is the most common nitrogen rejection technology to separate nitrogen from shale gas. This technology uses Joule-Thomson (JT) cooling of the high-pressure gas and the difference in boiling points between nitrogen and methane for separation in a distillation column. Typically cryogenic processes are used to treat shale gas containing more than 10 mole percentage nitrogen and are economically viable in processing gas flow rate ranging from 30 MMscfd to 900 MMscfd.

❖ *Non-Cryogenic Processes*

▪ *Pressure Swing Adsorption (PSA)*

PSA selectively separates nitrogen from methane in a cyclic process using a zeolite adsorbent. In this process, nitrogen is separated at a pressure by adsorption of methane at high pressure. The adsorbent can be regenerated with a combination of pressure and thermal changes. PSA technology is limited in capacity and can typically handle 2–15 MMscfd gas flow rate.

▪ *Membrane Separation*

In this process, the feed gas is compressed and passed across the membrane surface which in turn separates the hydrocarbon permeate. Permeate is later compressed back to the pipeline pressure while non-permeate, nitrogen-rich waste gas can be used as fuel. Similar to the PSA system, membranes also can handle only small flow rates varying from 0.5 to 25 MMscfd.

2.2.4 NGL Recovery

NGL recovery process selection must be evaluated based on meeting the NGL recovery levels, feed gas pressure, temperature, gas compositions, product specifications and NGL recovery flexibility. The NGL process must be designed to handle both a rich gas case and a lean gas case. Rich gas has higher NGL content than lean gas. NGL designs that are optimized today may not work well in the future when feed gas compositions change. An optimized gas plant design must be flexible and be suitable for revamp to meet future requirements and project economics while preserving most of the key equipment. Following are the processes available for NGL recovery,

- *Refrigeration Process*

- *Cascade Refrigeration Process*

The cascade system generally uses two or more refrigeration loops in which the expanded refrigerant from one stage is used to condense the compressed refrigerant in the next stage. Cascade refrigeration with two refrigerants consists of two refrigeration circuits that are thermally connected to a cascade condenser, which in turn acts as a condenser for the low-temperature circuit and as the evaporator for the high-temperature circuit. Generally, this process is used for temperature levels below –90 F, when light hydrocarbon gases or other low boiling gases and vapors are being cooled. To obtain the highest overall efficiency of the system, the refrigerants for the two superimposed systems should be different.

- *Joule-Thompson (JT) Expansion Process*

Joule –Thompson expansion process a gas is sent from one pressure to a lower pressure while keeping the system isolated i.e., no heat transfer to the surrounding. During this process, gas is cooled by isenthalpic expansion.

In typical process unit working with JT valve principle, the inlet gas is pre-cooled against the treated gas via a gas-to-gas exchanger, and subsequently further cooled by isenthalpic expansion (i.e., Joule–Thomson expansion) through a valve resulting in heavy hydrocarbons and water to condense.

- *Turbo-Expansion Process*

This technology utilizes cryogenic low-temperature distillation process which involves expansion of the gas through a turbo-expander followed by distillation in a de-methanizer fractionating column. Turbo-expander is a machine which has an expander/compressor as a single unit. It consists of two primary components integrated as a single assembly, one component is radial inflow expansion turbine and the other is a centrifugal compressor. The expansion turbine is the power unit and the compressor is the driven unit. In cryogenic NGL recovery processes, the turbo-expander achieves two different but complementary functions. The main function is to generate refrigeration to cool the gas stream. This is done by the expansion turbine end that expands the gas isentropically by extracting the enthalpy from the gas stream, causing it to cool. The other function is the use of the extracted energy to rotate the shaft to drive the compressor end of the turbo-expander, which recompresses the residue gas stream.

❖ *Lean Oil Absorption*

The lean oil absorption process was developed in the early 1910s and was used exclusively until the 1970s. The absorption unit uses a lean oil to absorb the C3+ components followed by a de-ethanizer and a rich oil still to regenerate the rich oil.

❖ *Solid Bed Adsorption*

Solid-bed adsorption process can be designed to selectively remove specific hydrocarbons. The adsorbent can be silica gel (i.e., Sorbead) that can be designed to remove most of the C_6+ hydrocarbons. In a typical two-bed adsorption process, regeneration is accomplished by passing heated recycle gas through the bed. The heavy hydrocarbon is recovered from the regeneration gas by cooling, condensation, and separation.

❖ *Membrane Separation*

In this process, a slipstream of the pipeline gas is processed in the membrane unit, which removes the heavy hydrocarbons, producing a lean gas. The hydrocarbons can be recycled back to the compressor suction and recompressed back to the pipeline system. The hydrocarbon contents can also be recovered by condensation and chilling to produce a liquid by-product.

❖ *Twister Supersonic*

Twister separation technology uses a supersonic mechanism (a combination of aerodynamics, thermodynamics, and fluid dynamics) to condense and remove water and heavy hydrocarbons from natural gas. This technology is based on the concept that condensation and separation at supersonic velocity reduce the residence time to milliseconds, allowing no time to form hydrates. This separation technology can potentially offer significant cost and environmental benefits for offshore operation.

2.3 Selected Process Technology – Process Overview

Process technology for major units from all the available options discussed in the section above is selected based on the current most widely used technologies in the industry.

2.3.1 Acid Gas Removal Unit

For acid gas removal unit, process technology selected is amine absorption process. Amine absorption process mainly comprises of two sections: Absorption section and Regeneration section.

In absorption section, an amine solvent is used to remove impurities such as CO_2 and H_2S from feed gas in the absorption column. The amine solvent that is free of impurities is also called as a Lean solvent. Lean solvent and feed gas enter absorption column in a counter-current fashion. In this work, di-ethanol amine (DEA) is used as the solvent. The sweet gas, gas free of impurities, is sent to a downstream unit, and the rich solvent, solvent rich with impurities, flows over to regenerator section. Regenerator is typically a stripper column with a reboiler. In stripper column, the rich solvent is regenerated by stripping out acid impurities from the rich solvent and venting it to flare. Later, lean out from stripper column bottom is recycled back to the absorption column for reuse. Figure 5 represents the schematic of the acid gas removal process.

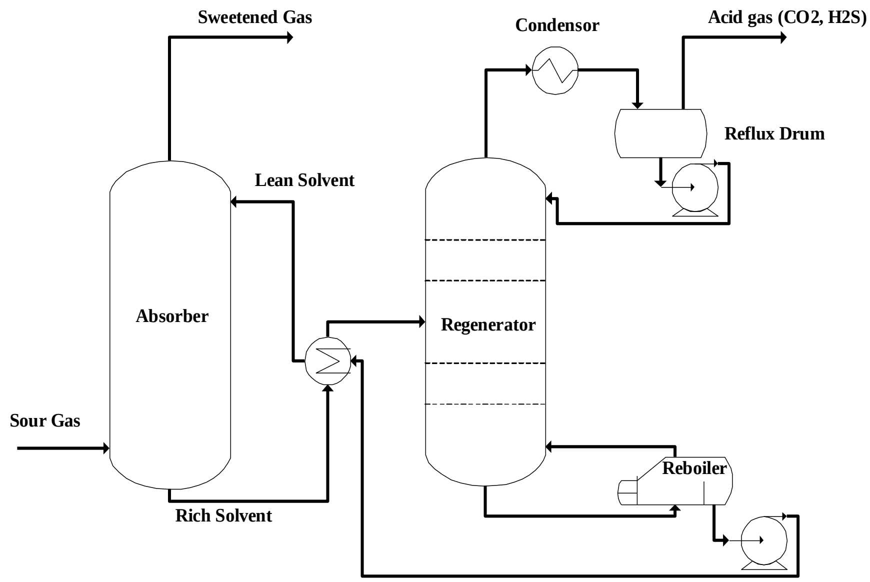

Figure 5: Amine process schematic flow sheet

2.3.2 Dehydration Unit

In this study, TEG absorption process is selected for gas dehydration unit. Similar to acid gas removal unit, this process also has two sections: absorption section and stripping section. In absorption section, water is stripped out of sweet gas entering into absorption column using lean solvent. Tri-ethylene glycol, TEG, is used as the solvent in this process. Solvent now enriched with water known as rich solvent flows over to stripping column from the absorber. In stripping section, TEG solvent is regenerated in a stripper column by removing moister. The regenerated TEG is recycled back to absorber for reuse. Figure 6 represents the schematic of TEG dehydration process flowsheet.

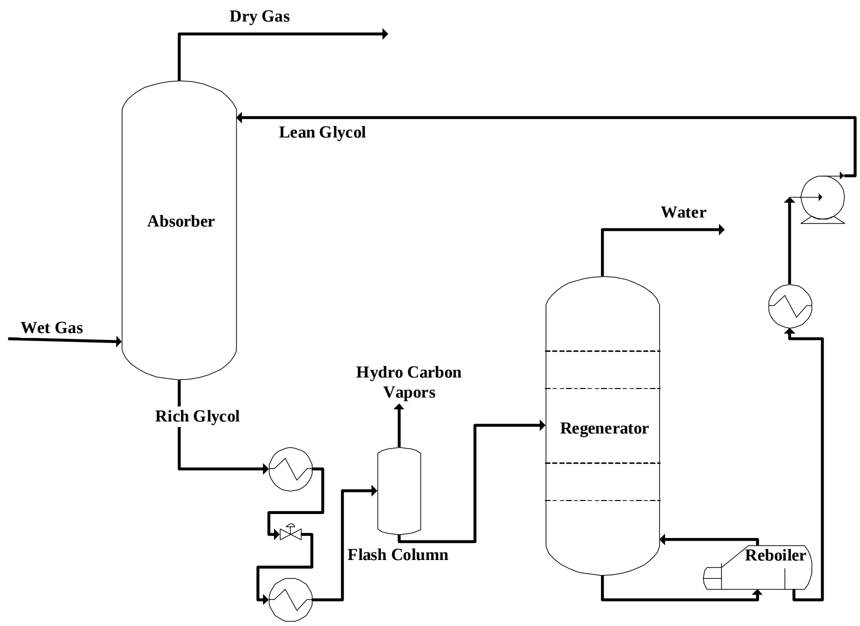

Figure 6: Gas dehydration process flow sheet schematic

2.3.3 NGL Recovery Unit

NGL recovery plant has mainly two sections one is turbo-expander and other is a de-methanizer column. The feed gas is first cooled in a heat exchanger and then flashed in a flash column. The liquid stream from a flash column is expanded through a JT valve and gas stream is expanded through turbo-expander. This liquid and gas streams enters de-methanizer at different tray locations. Figure 7 represents schematic of typical turbo-expander NGL recovery process.

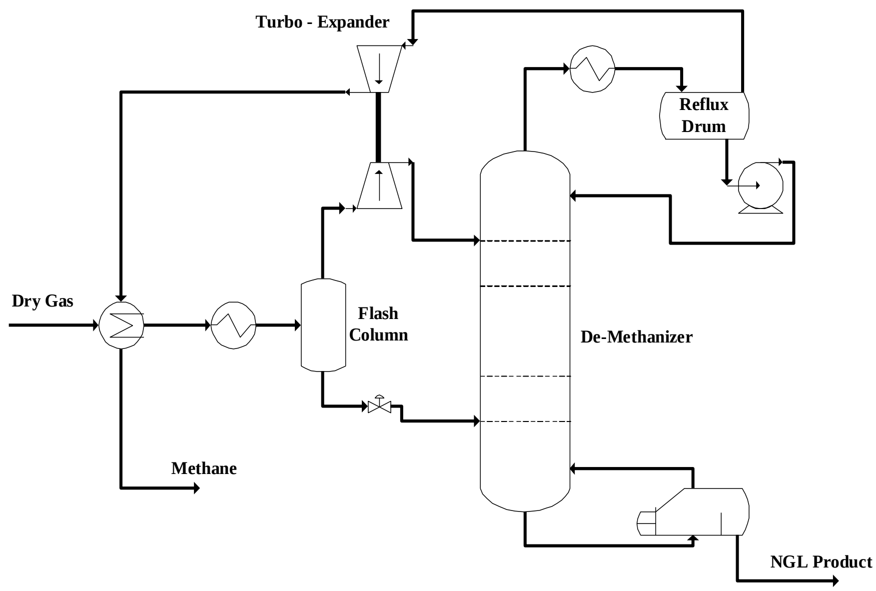

Figure 7: Turbo-expander process for NGL recovery

Fractionation train can be either a direct sequence or an indirect sequence for simple column configuration. The economically profitable sequence is selected based on the energy consumption which intern is directly proportional to the vapor load. The sequence with least vapor load will be more economically profitable.

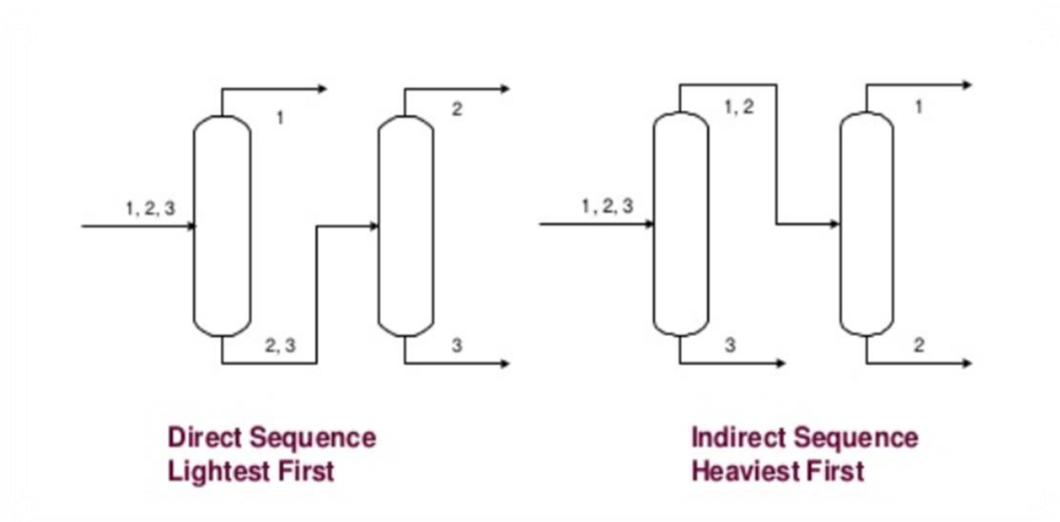

Figure 8: Classical distillation schemes for tertiary mixture

The following formula is used for distillation column sequence selection (Smith & Smith, 2005)

Vapor Load, $V = (F_A + F_B + \cdots + F_{LK}) + (F_A + F_B + \cdots + F_{LK} + F_{HK} + \cdots + F_{NC}) * \frac{R_F}{\alpha - 1}$

Where FA, FB etc. are the molar flow rates of component A, component B etc., RF is ratio of Reflux Ratio and Minimum Reflux Ratio, LK – Light key, HK – Heavy Key, NC- Non-Condensable

Using the formula above it is found that vapor load in the direct sequence is lesser than that of the indirect sequence. Hence, the direct sequence is used for NGL fractionation section.

CHAPTER III

PROBLEM STATEMENT AND APPROACH

3.1 Problem Statement

The percentage of methane, natural gas liquids, and impurities in shale gas varies among different wells in the same shale play. In addition, the flow rate of shale gas from each well declines with time. If a gas plant is designed, to process shale gas with a fixed flow rate, then the plant will either be under-designed or over-designed for a particular period of its life. For an instance, designing a plant for maximum flow rate will cause a plant to be over-designed for the most part of its life. It is possible that this case would not make an optimum plant size.

In this study, this uncertainty in shale gas feed gas flow rate is considered while designing shale gas processing and NGL recovery plant. The aim is to minimize total annual cost, by optimizing shale gas processing plant size and the amount of excess gas in the process when the plant is under-designed. The detailed techno-economic analysis is carried out to prove the viability of this approach.

3.2 Solution Approach

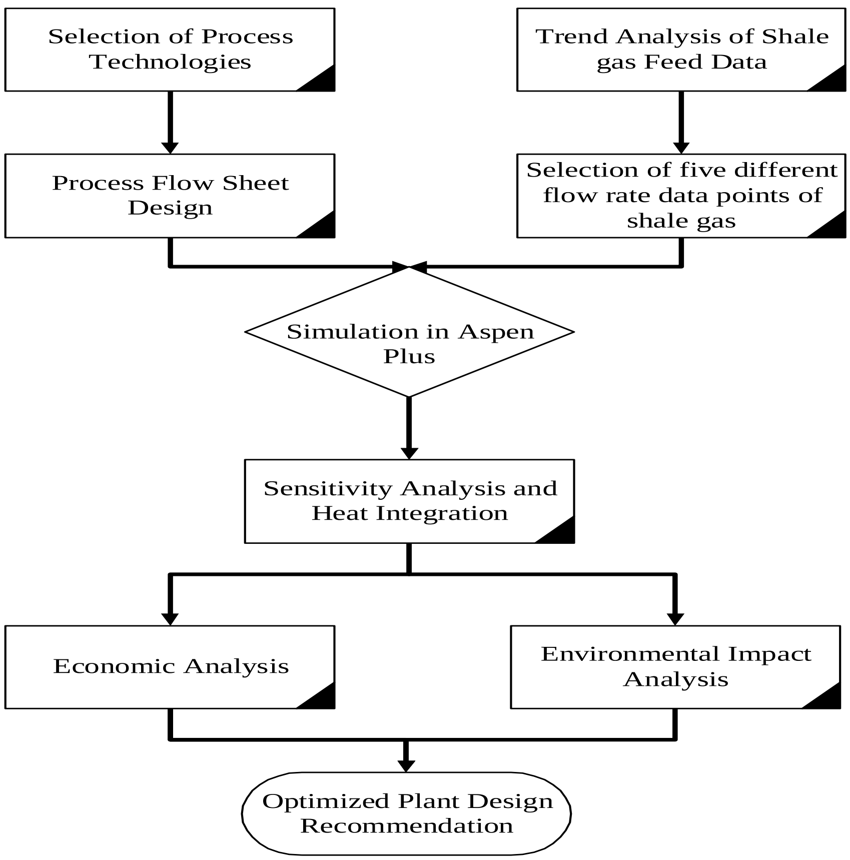

Figure 9: Solution approach

3.3 Scope of the Work

Scope of this project from the solution approach flow diagram shown in Figure 9,

- Shale gas flow rate from a particular well declines over a period of time. In this study, five discrete shale gas flow rates are selected at different time intervals over 3 years of shale gas well life.
- Two shale gas processing and NGL recovery process technologies are chosen to perform case studies. One of which is a conventional process and other is novel technology.
- Conventional technology is designed and simulated in Aspen plus based on the five different sets of shale gas flow rate selected, as mentioned before.
- Sensitivity analysis is carried out in Aspen Plus to choose optimum operating conditions such as reflux ratio in a distillation column.
- Heat integration of the process is explored
- Economic Analysis is carried out for all the case studies to recommend an optimum plant size and design.
- Economics with constant production rate of optimum design case is done to apply this work to real life scenario.
- Environmental impact is observed

CHAPTER IV

SIMULATION AND HEAT INTEGRATION

4.1 Selection of Shale Gas Data

Marcellus shale play is considered for this study. In these case studies, feed composition to the shale gas processing and NGL recovery unit is considered constant and the feed flow rate fluctuation is taken into account by the multi-period approach. For feed composition, the average of the four different well data compositions is used, and feed flow rate is accounted by selecting five discrete flow rates. Table 6 shows typical data for Marcellus shale gas for four different wells.

Figure 10 shows the declination curve for the average well production rate for Marcellus shale play. From this graph, it can be inferred that shale gas flow rate from a given well declines drastically with time. By examining this graph, it is observed that 75% of the peak production is dropped within 1 year of shale play well life and almost 90% of the production flow is declined by end of the third year. Hence, discrete flowrate points are chosen between zero to three years of a shale play well life. Further, data for these flow rate cases are selected based on the amount of drop in flow rate and important milestones.

Table 6: Marcellus shale play well data

Component	Well 1	Well 2	Well 3	Well 4	Average Composition
CO_2	0.1	0.1	0.9	0.3	0.35
N_2	0.4	0.3	0.3	0.2	0.3
C1	79.4	82.1	83.8	95.5	85.2
C2	16.1	14	12	3	11.275
C3	4	3.5	3	1	2.875

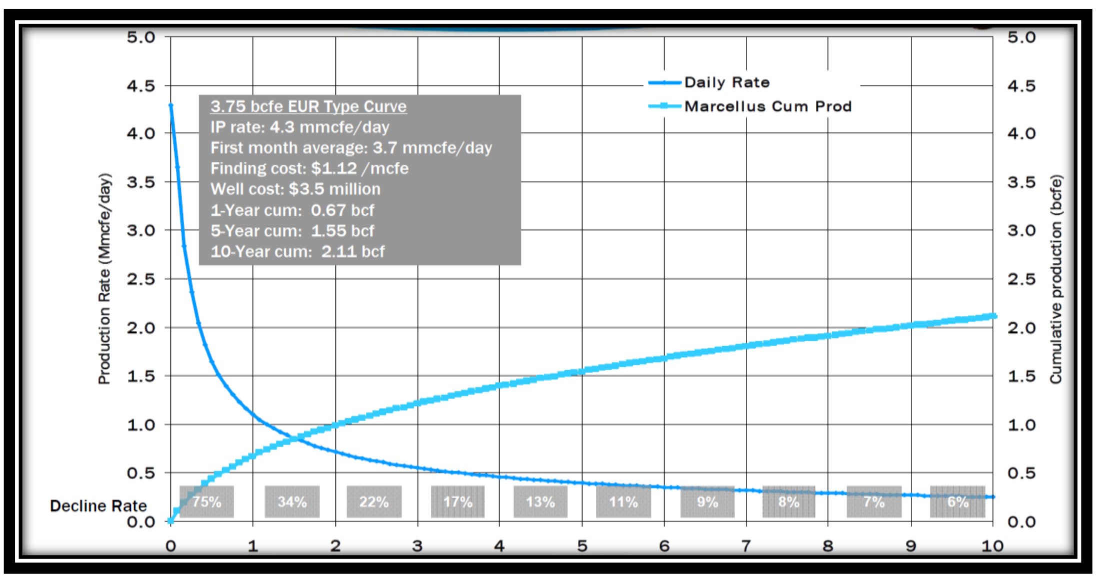

Figure 10: Average Marcellus shale gas well production rate declination curve

In this work, as mentioned above, five different production flow rates are chosen to perform five different process simulation cases of the given technology respectively. The process technology given in this study will be elaborated in later sections. Case-1 corresponds to the peak production; Case-2 corresponds to 40% of the peak production and so forth as mentioned in Table 7. The production rate, given in the declination curve in volumetric flow rate is assumed to be available at international standard conditions i.e., at 60°F and 288.15 K. Production rate is converted from volumetric flow to molar flow using formula below,

$$Molar\ Flow\left(\frac{kmole}{hr}\right) = \frac{Volumetic\ flow(1000 * CFD)}{379.5\left(\frac{ft^3}{m^3}\right) * 2.2\left(\frac{m^3}{kmole}\right) * 24(\frac{hours}{day})}$$

Total input flow rate to the process simulations is taken as the product of the production rate multiplied by a number of wells i.e., 4 for this study.

Table 7: Shale gas flow rate data from decline curve for various cases

Case #	Flow (CFD*1000) per well	Flow (Kmol/hr) per well	Flow (Kmol/hr) for four wells	Time line on declination curve
Case 1	4300	214.60	858.39	Beginning of 1st year
Case 2	2580	128.76	515.03	40% of 1st year
Case 3	1075	53.65	214.60	End of 1st year
Case 4	710	35.41	141.63	End of 2nd year
Case 5	553	27.62	110.47	End of 3rd year

4.2 Simulation

Process simulations for the selected technologies are carried out in Aspen plus. The scope of process simulation in this study includes acid gas removal, dehydration, and NGL recovery units.

4.2.1 Acid Gas Removal Unit

Figure 11 represents the simulated acid gas removal unit process flow sheet in Aspen plus. In the simulation, feed gas entering at standard conditions is sent via series of compressors and inter-coolers, M-Comp, to attain feed conditions required for the absorption column. In this absorption column, ABSORBER, a lean di-ethyl amine (DEA) solvent, DEA, absorbs impurities from entering feed gas stream. The DEA solvent used in this simulation is of 29-weight percentage concentrated. Treated gas, GASOUT, is sent to a downstream dehydration unit and, the solvent rich with impurities from the absorption column, RICHOUT, is expanded through a pressure reducer, VALVE-1, and then heated in an exchanger, HX-2.This heated rich stream, RICHIN, is stripped-off of acid gas impurities in a stripping column, STRIPPER. The top stream from this stripping column, CO2OUT, is vented to flare. Whereas, the bottom stream from this stripping column, LEANOUT, is cooled in the exchanger, HX-3, and mixed with makeup DEA in the mixer, MX-1 before recycling it back to absorption column through the pump, PUMP-1, for reuse.

In Aspen plus, ELECNRTL property method is used to simulate the acid gas removal unit. Absorption and stripping columns calculations are performed by choosing rate based calculation type in aspen plus. Compression, expansion, and pumping are assumed to be carried out adiabatically. Table 8 shows important equipment operating conditions. Process conditions for this unit are taken from aspen plus reference examples. Further, it is assumed that throughout all cases process conditions including L/G ratio are constant. Additionally, column diameter for various cases is calculated using the correlation below

$$D^2 \alpha V$$

Where ‘D’ is column diameter and ‘V’ is vapor rate in the column.

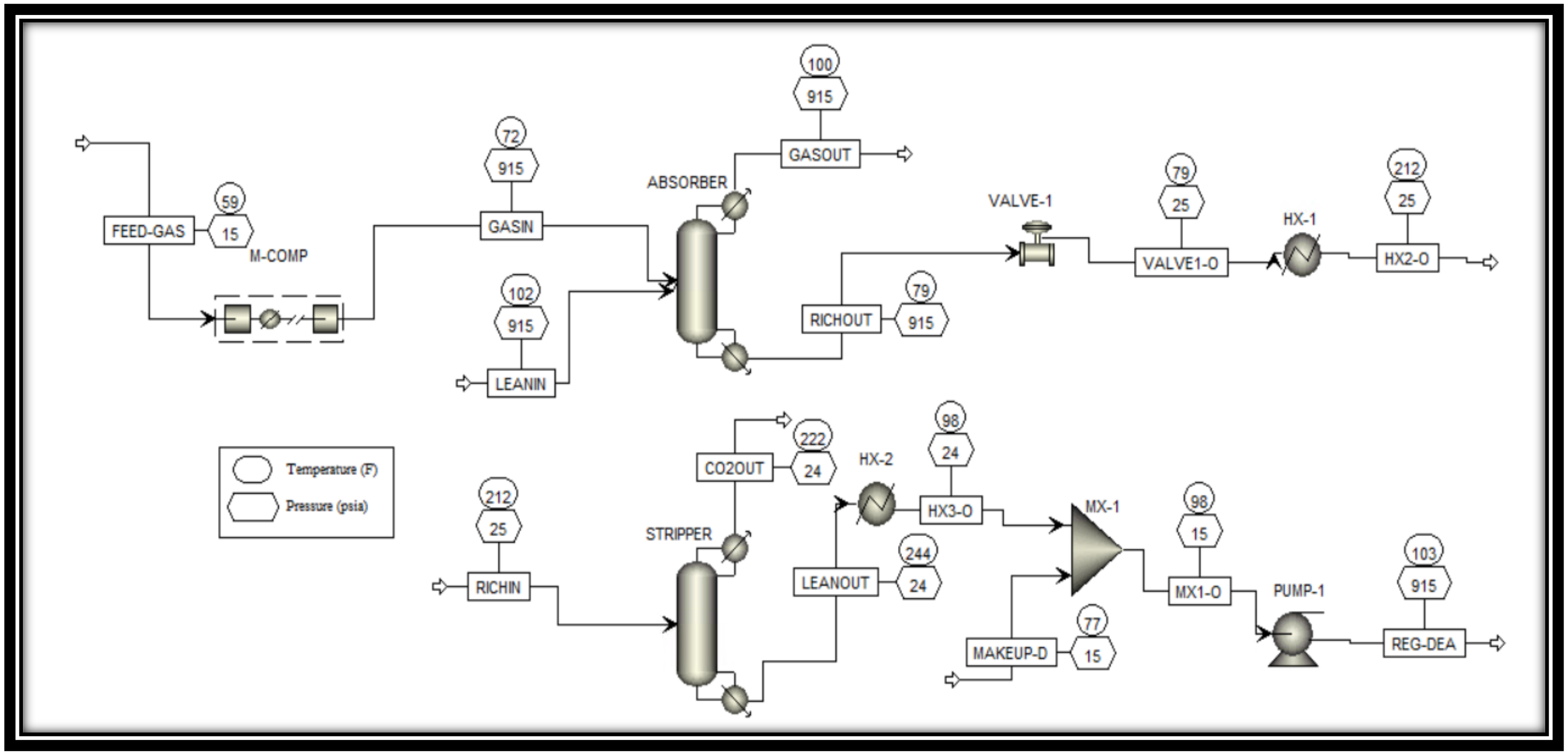

Figure 11: Acid gas removal unit is Aspen plus

4.2.2 Gas Dehydration

Figure 12 represents process flow sheet of the dehydration unit in Aspen plus simulation interface. In this, feed gas coming from the acid gas removal unit is expanded through the turbine, TURB-1, and then sent through the exchanger, HX-3, to achieve proper feed conditions required for absorption column, ABS-TEG. In ABS-TEG, moisture from feed gas is physically absorbed by lean tri-ethylene glycol (TEG). The purity of TEG is assumed to be greater than 99%. Dry gas from the absorber is sent to the downstream unit. Whereas glycol solvent rich with water is flashed in a flash column, FLASH, and expanded through the valve, VALVE-2. This rich solvent is further heated in the exchanger, HX-4, before sending it through stripping column, TEG-REG. In this stripper, moisture is removed from the rich solvent and vapor distillate is sent to flare and, bottom stream from the stripper is cooled in an exchanger, HX-5, and mixed with make-up TEG in a mixer, MX-2, before finally recycling it to absorber as lean in via pump, PUMP-2.

In Aspen plus, SR-POLAR property method is used for gas dehydration unit. Absorber and stripper columns use equilibrium calculation method. Turbine, valve, and pump are assumed to operate adiabatically. Table 8 shows important equipment operating conditions. Process conditions are taken from aspen plus reference examples. In addition, L/G ratio of the absorber is assumed to be constant for all cases.

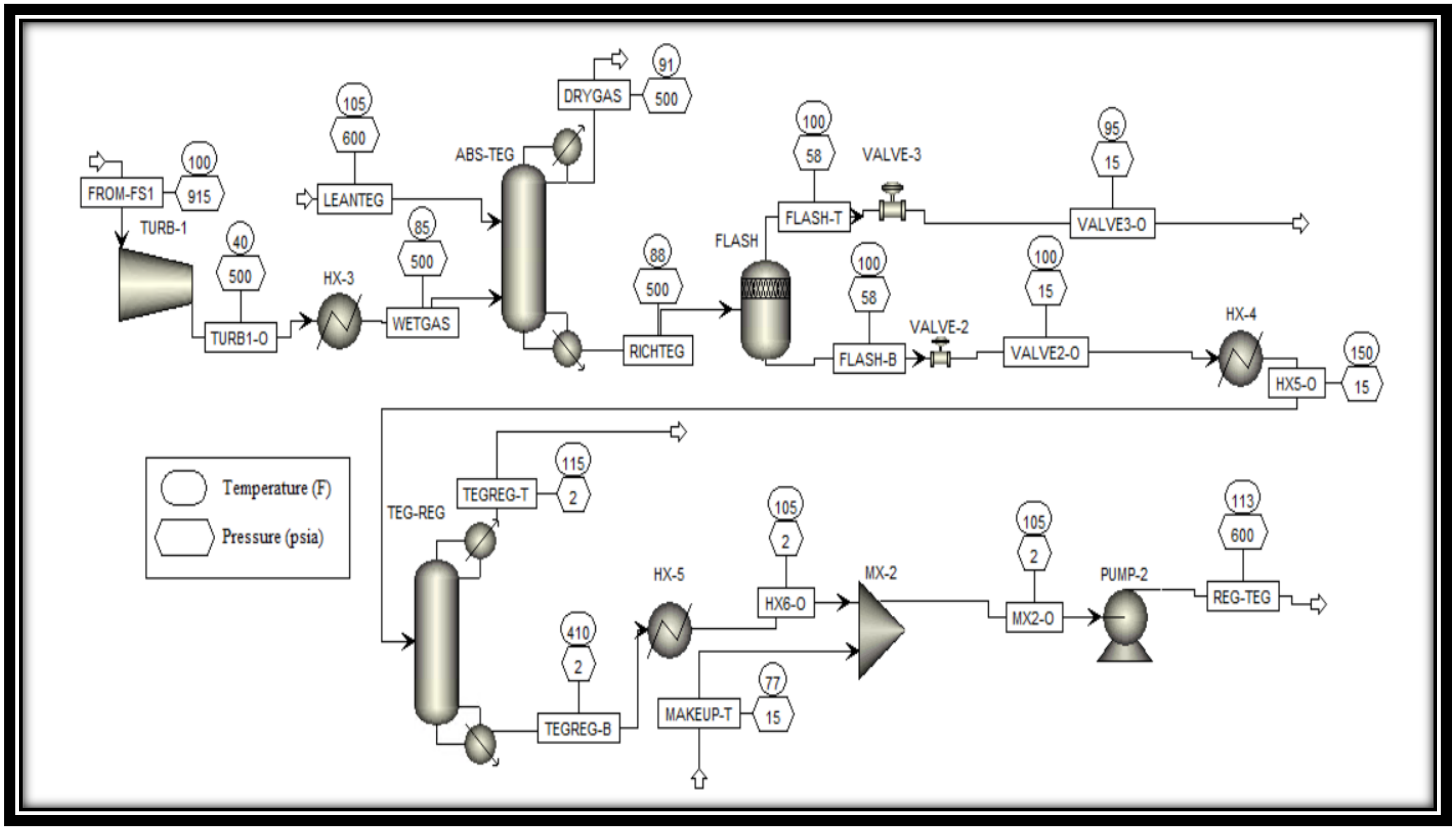

Figure 12: Gas dehydration unit process flowsheet in Aspen plus

4.2.3 NGL Recovery

Figure 13 shows simulated NGL process flow sheet in aspen plus. In this process, Feed from Dehydration unit is expanded adiabatically through the turbine, TURBINE-2, before sending it to the de-methanizer column, DE-METH, where methane is separated from Natural gas liquids (NGL). Later NGL is processed through a de-ethanizer column, DE-ETH, where propane is separated from butane.

For simulation, Peng-Robinson property method is used for NGL recovery and fractionation process. De-methanizer and De-ethanizer columns use equilibrium calculation mode in Aspen plus. Table 8 shows important equipment operating conditions. Process conditions for this unit are taken from literature.

Table 8: Main equipment input data

Equipment ID	Pressure (atm)	No. of Stages	Pressure Ratio
ABSORBER	62.24	20	-
STRIPPER	1.67	33	-
M-Compressor (3-Stage)	-	-	3.96
ABS-TEG	34.02	3	-
TEG-REG	0.12	5	-
DE-METH	25.00	10	-
DE-ETH	22.50	18	-

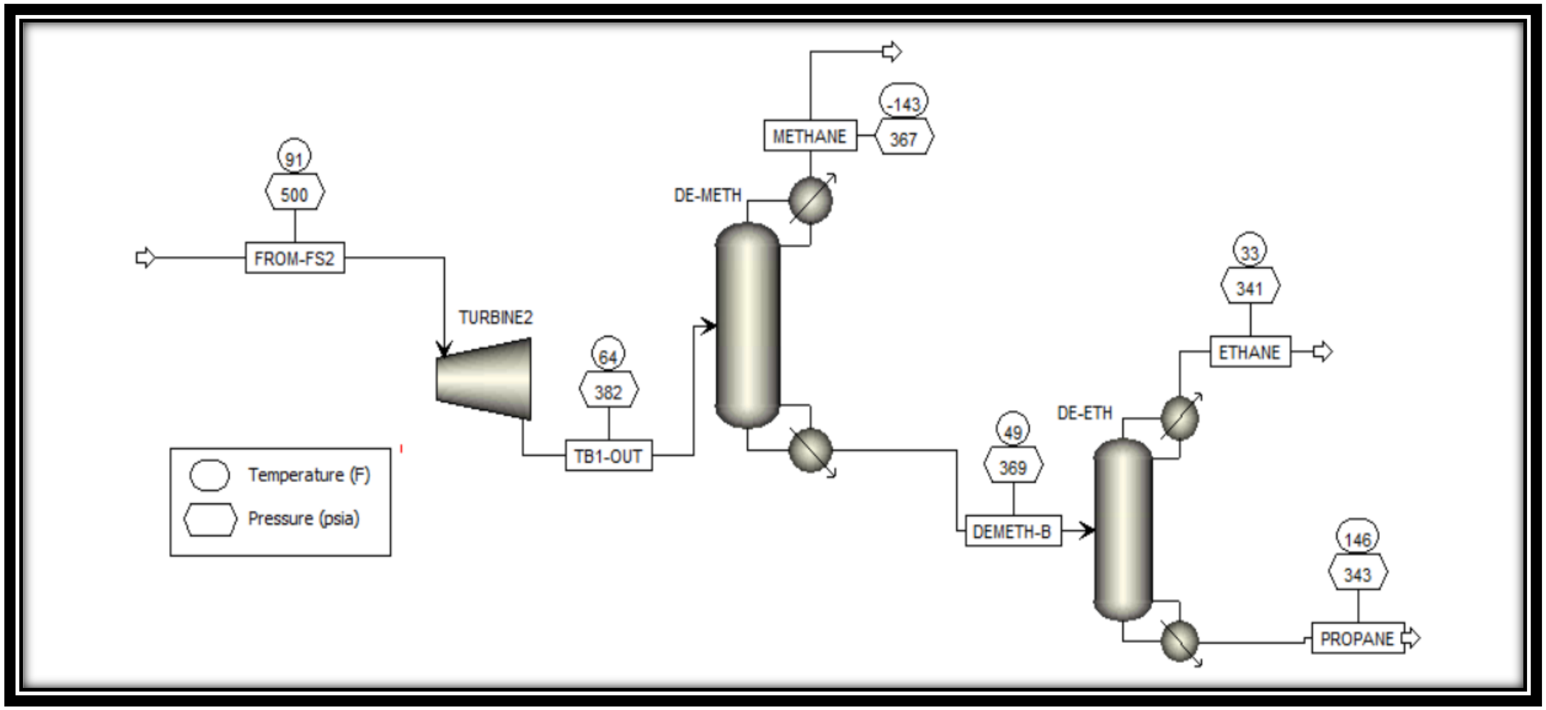

Figure 13: NGL recovery unit in Aspen plus

4.3 Assumptions

- Liquid to gas ratio's for amine absorber and TEG absorption columns are assumed to be constant throughout all the cases
- Operating conditions of all unit operations are same in all cases
- Pressure drop via heat exchangers is negligible
- Pressure drop across column is assumed to be negligible
- Delta temperature approach for pinch analysis is 5 Kelvin
- Plant life is taken as 10 years for economic analysis
- Tax-rate is assumed to be 30%

4.4 Case Studies

Five cases with five different feed flow rates respectively are simulated in aspen plus. Since operating conditions of all unit operations are constant for different cases and Aspen plus model varies equipment size linearly with feed flow rate, it is implied that the production rate of different streams varies linearly with feed flow rate. Specific flow value per unit feed flow rate for important streams is given in Table 9. In addition, the calculated cumulative flows for a period of three years for all five cases is also mention in Table 9.

Cumulative flowrates are calculated to make flowrates independent of the time factor. In each given case, plant size is designed based on the input flow rate. In the event of case-1 where the flow rate is at peak, from shale gas well production rate declination curve, it is observed that plant will be over-designed for throughout its life. And in case-

2, the plant will be under-designed for starting 40% of the year and it will be over-designed for remaining the time of plant life. Similarly, for case-3, case-4 and case-5 plant will be under-designed for first one, two and three years respectively. During this under-design period in case-2, case-3, case-4, and case-5, there will be excess shale gas feed flow available which cannot be processed given the limited plant size. The viability of co-generation is explored for utilizing this excess gas. Therefore, because of this loss in feed flow, net feed processed in a particular case is calculated by subtracting this excess gas flow from the actual cumulative flow available as shown by the violet curve in Figure 10.

Table 9: Simulation results for conventional technology

Stream	Specific Values (Per Kmole of Feed Gas IN)	Cummulative Feed Rate (kmole) for 3 years					
		Case-1	Case-2	Case-3	Case-4	Case-5	Constant Prod Rate Case
Feed Gas IN	1.00	5988741.17	5877015.21	4658641.75	3679171.16	2903264.82	5150317.40
Sweet Gas	1.00	5975958.49	5864471.01	4648698.11	3671318.14	2897067.94	5139324.30
Dry Gas	1.00	5961521.96	5850303.81	4637467.93	3662449.09	2890069.30	5126908.89
Methane	0.87	5191220.85	5094373.43	4038250.70	3189216.15	2516637.21	4464449.93
Ethane	0.10	609320.09	597952.61	473990.10	374334.58	295390.56	524015.28
Propane	0.03	160984.41	157981.09	125229.78	98900.45	78043.18	138446.60
Make up DEA	0.01	57224.74	56157.16	44515.13	35155.91	27741.82	49213.28
Make up TEG	0.00	1.55	1.52	1.20	0.95	0.75	1.33
Acid gas removal Vent	0.01	87485.62	85853.49	68055.06	53746.62	42411.91	75237.63
Dehydration Vent-1	0.00	1679.17	1647.84	1306.23	1031.59	814.04	1444.09
Dehydration Vent-2	0.00	12759.00	12520.97	9925.23	7838.47	6185.40	10972.74
Co-generation Fuel flow		0.00	111725.96	1330099.41	2309570.01	3085476.34	NA

4.5 Heat Integration

The conventional process technology used in this work is further energy integrated by performing

- Energy analysis of major equipment's
- Pinch analysis
- Co-generation

After performing an energy analysis of major equipment's, the unit operation, which is utilizing maximum energy, is modified and this modified process flowsheet is termed as 'Novel Technology'. Next, pinch analysis of the conventional technology is carried out for utility targeting. Finally, the viability of incorporating co-generation option for utilizing excess feed gas from the system is explored. The application of these concepts is further elaborated in sections below.

4.5.1 Novel Technology In-corporation

Pie chart with major equipment's energy consumption in shale gas processing and NGL recovery plant is shown in Figure 14. From this pie chart, it is clear that costs of hot and cold utility in de-methanizer and de-ethanizer columns amounts to around 85% of the total hot and cold utility costs in the process. Hence, these column configurations are modified in novel technology to reduce operating costs. In the conventional technology, a simple two-column system is used for NGL fractionator section and in the novel technology, this system is replaced with a complex configuration.

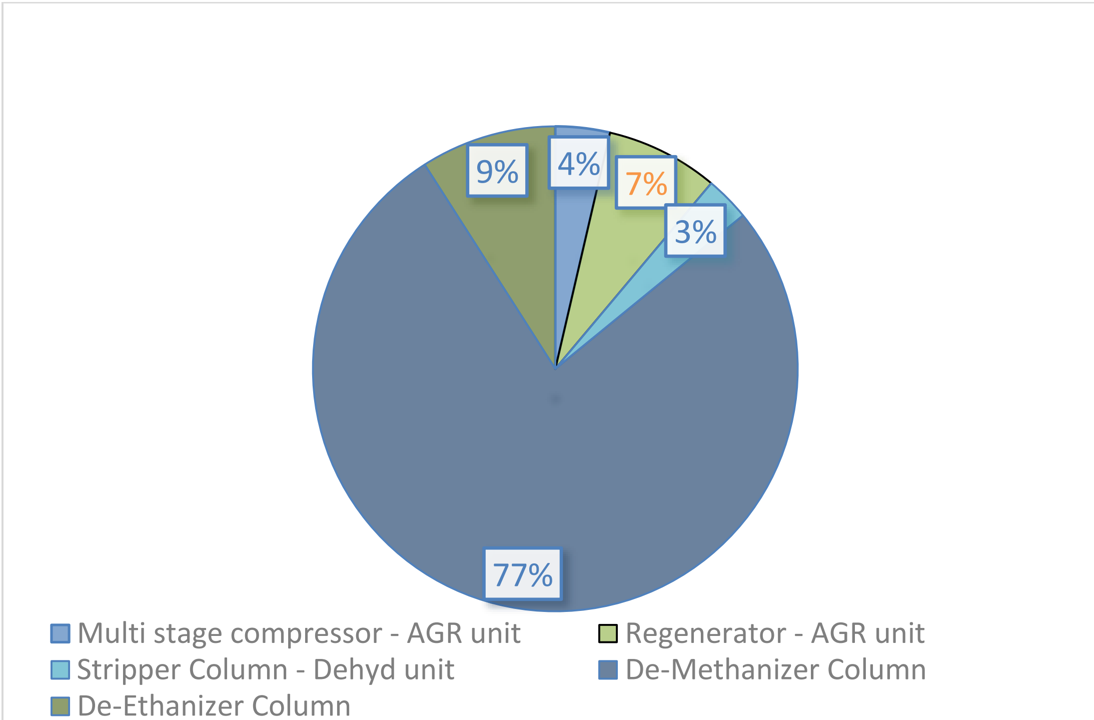

Figure 14: Percentage utility costs for main units

The simple column configuration as stated in conventional technology should have a reboiler and a total condenser. This restriction curtails the use of materials flow to provide some of the necessary heat transfer by direct contact. This transfer of heat via direct contact is known as thermal coupling (Smith & Smith, 2005). Two famous technologies designed using this concept and now commercially available are Petlyuk Column and Dividing-wall column. Figure 15 shows schematic of the Petlyuk column and the Dividing-wall column.

Petlyuk column is also known as an equivalent thermally coupled pre-fractionator arrangement. Petlyuk column arrangement has two columns first is pre-fractionator and second one is the main column. In this arrangement, feed entering to pre-fractionator is split into vapor and liquid fractions utilizing the pump around streams from the main column as reboiler and condenser. The vapor and liquid fractions from pre-fractionator enter main column at two different feed locations, near top and bottom respectively. Since vapor and liquid are entering at top and bottom of the main column respectively, this avoids remixing of vapor and liquid streams that generally happens in the traditional column system. Typically, thermocouple arrangement requires 30% less energy compared to a conventional column due to reduced heat loss from remixing (Smith & Smith, 2005).

Dividing Wall Column uses a single shell with a vertical baffle dividing the central section of the shell into two parts. The feed side section in this arrangement acts as pre-fractionator. This arrangement works on the same principle as Petlyuk column and it utilizes the same amount of energy as Petlyuk column. In addition, Dividing-wall column typically require 30 percent less capital cost than traditional two-column arrangements of simple columns (Smith & Smith, 2005) as it has only one column instead of two.

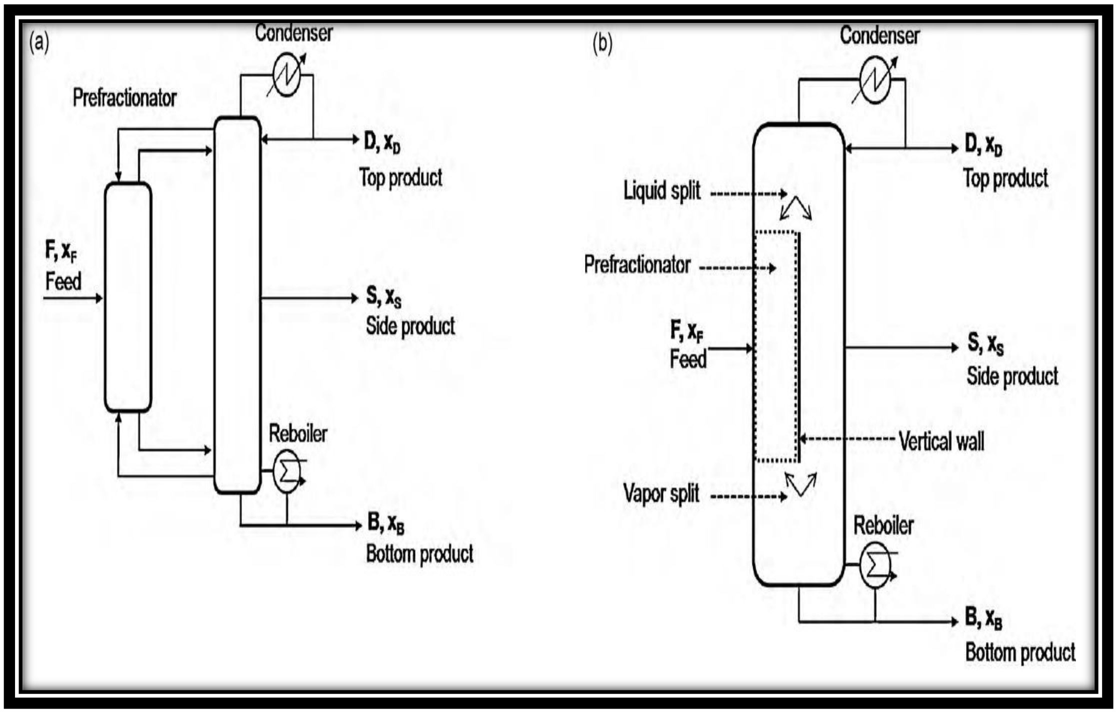

Figure 15: Schematic diagram of (a) Petlyuk column and (b) Dividing-wall Column

4.5.2 Pinch Analysis

To accomplish the minimum usage of heating and cooling utilities for a given process technology, it is necessary to maximize the heat exchange among process streams. Using this theory, there is a graphical technique known as the "thermal pinch diagram" originally developed by Linnhoff and Hindmarsh in 1983, and a simplified algebraic method, which uses “hot and cold temperature interval diagram”, available for utility targeting. Thermal pinch diagram is an enthalpy Vs temperature graph of cold and hot stream utilities. In this graph, cold stream utility graph is displaced horizontally to touch the hot utility curve at a particular point. This point is known as a Pinch point; it is a location that gives minimum hot and cold utility consumption targets achievable for a particular process system. In the algebraic method, amount of cold and hot utility consumption for each temperature interval is calculated. The net difference between hot and cold utility available in an interval is carried on to the next interval as a hot utility. The inlet utility to the first interval in a temperature interval diagram which is zero initially is adjusted to a value such that the net difference of hot and cold utility between one of the intervals becomes zero. The interval at which this zero occurs it is called as Pinch point or location.

In this study, pinch analysis algebraic method is used to calculate the targeted values for minimum hot and cold utility. The delta approach temperature is assumed, 5-degree Kelvin. This pinch analysis is carried out on conventional technology, the data for which is available from Aspen plus simulations. Table 10 below shows the utility consumption for various heat transfer equipment before pinch analysis and Table 11 shows the minimum utility consumption after pinch targeting. Moreover, Figure 16 and Figure 17 shows percentage energy savings achieved due to pinch analysis. It is observed that the utility consumption of hot and cold utility is reduced by 92% and 31% respectively after performing pinch analysis.

Table 10: Data from simulations for pinch analysis

Stream ID	Equipment	Ts (K)	Tt (K)	Case - 1		Case - 2		Case - 3		Case - 4		Case - 5	
				Heat Duty (KW)	FCp (KW/K)	Heat Duty (KW)	FCp (KW/K)	Heat Duty (KW)	FCp (KW/K)	Heat Duty (KW)	FCp (KW/K)	Heat Duty (KW)	FCp (KW/K)
H1	M-comp-1	415.2	350.0	684.6	10.5	410.7	6.3	171.1	2.6	113.0	1.7	88.1	1.4
H2	M-comp-2	488.1	350.0	1552.6	11.2	931.5	6.7	388.1	2.8	256.2	1.9	199.8	1.4
H3	M-comp-3	489.6	295.4	2288.8	11.8	1353.3	7.0	572.2	2.9	377.7	1.9	294.6	1.5
H4	Amine Stripper Cond	378.7	377.7	190.7	190.7	114.3	114.3	47.7	47.7	31.5	31.5	24.5	24.5
H5	HX-3	390.7	310.0	822.4	10.2	493.2	6.1	205.7	2.5	135.8	1.7	105.9	1.3
H6	TEG Stripper Cond	319.1	318.1	23.6	23.6	14.1	14.1	5.9	5.9	3.9	3.9	3.0	3.0
H7	HX-6	483.1	313.7	199.3	1.2	119.3	0.7	49.7	0.3	32.8	0.2	25.6	0.2
H8	De-Meth Cond	176.1	175.1	1596.4	1596.4	957.8	957.8	399.0	399.0	263.4	263.4	205.4	205.4
H9	De-Eth Cond	273.8	272.8	179.2	179.2	107.4	107.4	44.8	44.8	29.5	29.5	23.1	23.1
C1	HX-2	299.1	373.2	776.1	10.5	464.5	6.3	193.7	2.6	127.8	1.7	99.7	1.3
C2	Amine Stripper Reb	390.7	391.7	534.7	534.7	320.6	320.6	133.7	133.7	88.3	88.3	68.8	68.8
C3	HX-4	277.4	302.6	280.1	11.1	168.0	6.7	70.0	2.8	46.2	1.8	36.0	1.4
C4	Flash Column in Dehyd unit	310.9	311.9	7.3	7.3	4.4	4.4	1.8	1.8	1.2	1.2	0.9	0.9
C5	HX-5	310.7	338.7	34.0	1.2	20.3	0.7	8.5	0.3	5.6	0.2	4.4	0.2
C6	TEG Stripper Reb	483.1	484.1	215.0	215.0	128.6	128.6	53.6	53.6	35.4	35.4	27.6	27.6
C7	De-Meth Reb	282.7	283.7	273.4	273.4	164.1	164.1	68.3	68.3	45.1	45.1	35.2	35.2

Table 11: Pinch analysis results

Case #	Hot Utility (KW)	Cold Utility (KW)			
	IP Steam	Refrigerant	Brine	Chilled Water	CW
Case 1	202.81	1596.44	179.16	172.34	3241.18
Case 2	121.38	957.79	107.44	101.90	1927.06
Case 3	50.54	399.04	44.83	43.09	810.57
Case 4	33.35	263.39	29.55	28.44	535.01
Case 5	26.02	205.43	23.07	22.18	417.24

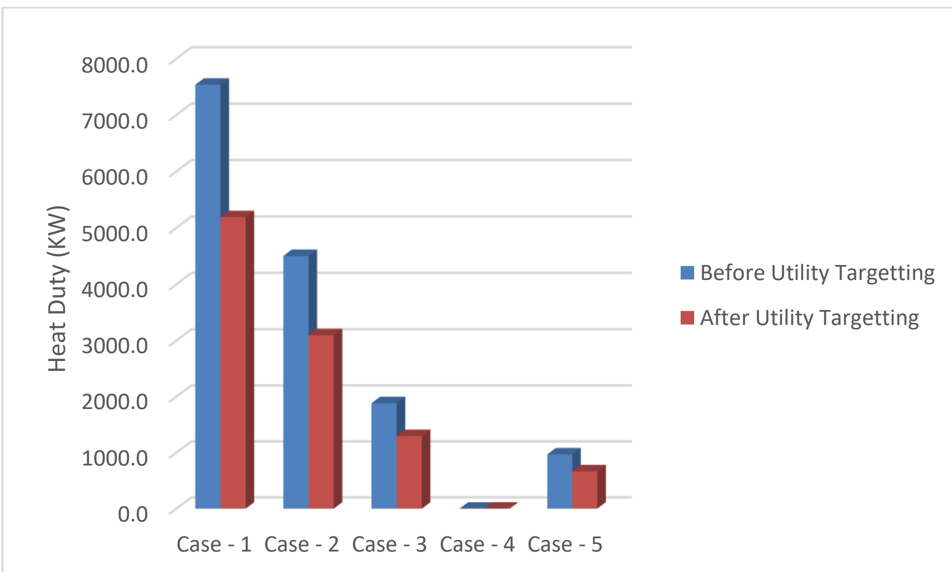

Figure 16: Cold utility before and after energy targeting using pinch analysis

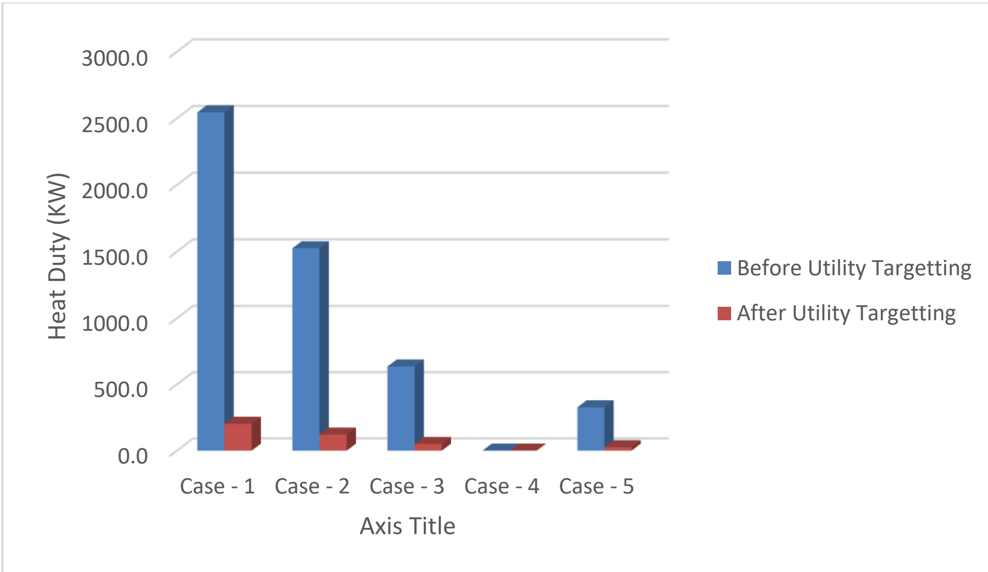

Figure 17: Hot utility before and after energy targeting using pinch analysis

4.5.3 Co-generation

The power plant with a steam turbine includes a boiler that corresponds to the high-temperature heat source where a fuel is burned to transfer heat to water, thereby producing high-pressure steam. The high-pressure steam is directed toward a turbine where the pressure energy of the steam is converted into rotational energy in the form of a shaft work. The shaft work can be converted to electric energy through an electric generator. The steam leaving the turbine is condensed, the condensate is fed to a pump that increases the pressure of the condensate and returns the water back to the boiler to be heated, transformed into steam, and the cycle continues. Typically, the generated work from the turbine is significantly higher than the work used in the pump. Therefore, the net effect of the system is that the heat from the boiler is converted into work while the discharged heat (with the steam exiting the turbine) is transferred to the condenser that serves as the low-temperature heat sink (El-Halwagi, 2017). Figure 18 represents the schematic of a steam turbine power plant.

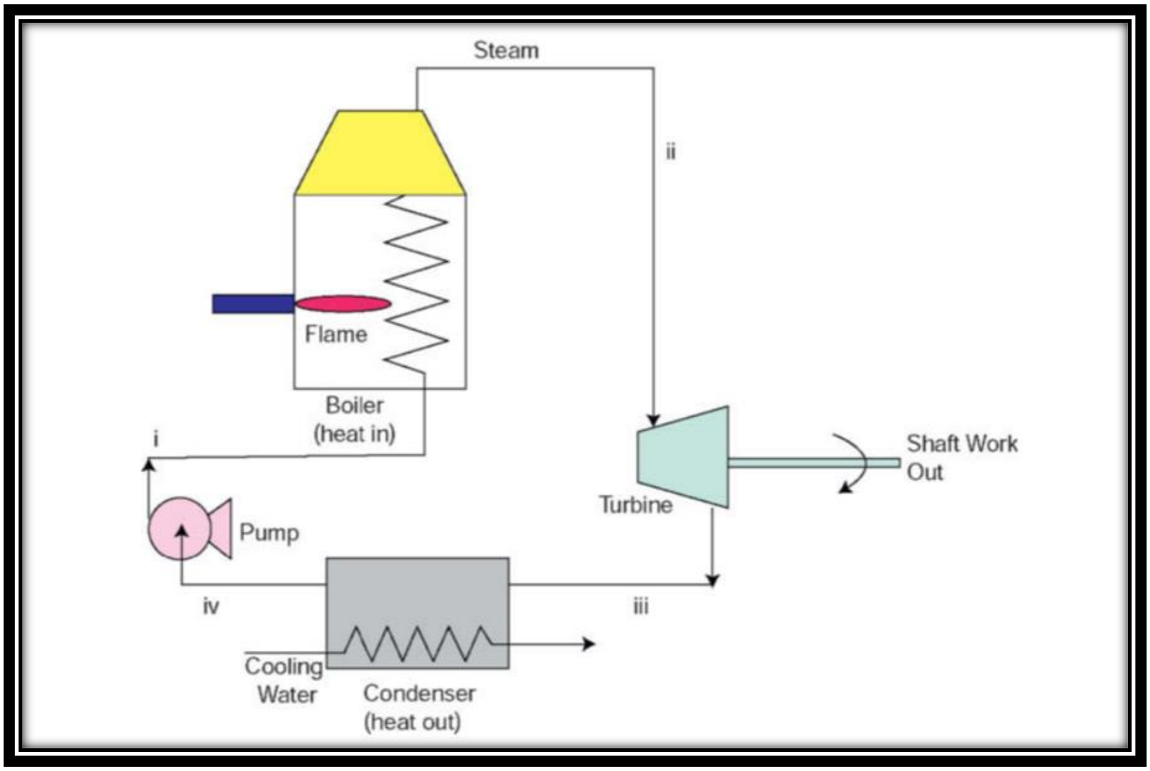

Figure 18: A schematic representation of a steam power plant

As discussed in before sections, shale gas processing and NGL recovery plant are simulated with five different flow rate scenarios. The excess shale gas flow available for co-generation is given in Table 9. In here, the amount of excess shale gas available in each case is cumulated over the period of 3 years to make the flow desensitized to time.

Operating conditions for co-generation are taken from 'Sustainable design through process integration' book by Mahmoud EI- Halwagi (El-Halwagi, 2017). The operating conditions of various unit operations used in steam turbine process are assumed constant in all cases. The energy available from burning excess shale gas in steam reboiler is used to calculate heat duty available from a boiler (Q boiler). The steam flow rate is calculated by assuming the enthalpy change across pump section is negligible. Enthalpy values are calculated using co-relations developed by Al-Azri N et al. (El-Halwagi, 2017). Finally, electric power generated in the turbine is calculated from enthalpy balance and steam flow rate. Table 12 below shows the electricity generated and utility consumption for different cases.

Table 12: Co-generation calculations for various cases

Description	UOM	Case-2	Case-3	Case-4	Case-5
Q Boiler	MM Btu	57395	683292	1186461	1585056
Steam Flow	Kilo Lb	59735	711151	1234835	1649681
Q Condenser	MM Btu	54461	648360	1125805	1504022
W pump	MM Btu	125	1487	2581	3448
Turbine (Elec Produced)	MM Btu	3024	35999	62508	83508

CHAPTER V

ANALYSIS AND CONCLUSION

5.1 Economic Analysis

In these case studies, calculations are carried out by assuming cumulative production rate of products, by-products, and utilities over a period of 3 years, except for the constant production flow rate case. Production rate is cumulated over 3 years because the five different shale gas flow rate data points considered for case studies are distributed over 3 years of the shale play well life. Table 13 gives the cumulative feed flows and the excess gas cumulative flow available for various cases that are used in this economic analysis. For easy understanding economic analysis is divided into four sections,

- Conventional technology economics
- Novel technology economics
- Excess shale gas economics
- Constant production rate economics

Conventional technology is nothing but the traditional process flow sheet used in this study and data for this is generated using simulation. Novel technology accommodates the modification of NGL fractionator part of the conventional technology. In excess shale gas section, the economics of co-generation are compared with selling the excess gas as the product. Finally, constant production rate section extrapolates the applicability of this study to a practical situation. Further detail analysis of these sections is carried out which is given below.

Table 13: Cumulative production rate processed through plant in 3 years

Case #	Flow (kmole/hr)	Time Period (year)	Cummulative Flow (bcfe @ STP) per well	Cummulative Flow (kmole) for 4 wells	Actual Feed Processed (kmole)	Loss in Feed rate (kmole)
Case 1	858.39	0.00	0.00	0.00	0.00	0.00
Case 2	515.03	0.40	0.40	1916397.17	1804671.22	111725.96
Case 3	214.60	1.00	0.67	3209965.27	1879865.85	1330099.41
Case 4	141.63	2.00	1.00	4790992.93	2481422.92	2309570.01
Case 5	110.47	3.00	1.25	5988741.17	2903264.82	3085476.34

5.1.1 Conventional Technology Economics

Plant economic analysis mainly comprises of three parts, one is capital cost estimation, the other is operating cost estimation and the last one is profit evaluation.

Capital cost is further sub-divided into fixed capital cost investment, working capital investment and maintenance costs. Similarly, operating cost comprises of utility cost, raw material cost and labor charges in this study. Finally, profit is evaluated assuming a tax rate, salvage value and time period

- Capital Cost Estimation

In this work, capital cost is estimated using data from a reference paper and scaling it to current case plant size with the help of six-tenths rule. The formula for six-tenth rule is given by,

$$C_B = C_A * \left(\frac{S_B}{S_A}\right)^{0.6}$$

Where C_A and C_B are the respective costs for plants A and B; S_A and S_B are the respective plant sizes for plants A and B

The inflation rate due to the time difference between reference paper and these studies is compensated using chemical engineering process cost index (CEPCI). CEPCI is available from the literature. Formula for capital cost to accommodate for inflation rate is,

$$C_B = C_A * \left(\frac{S_B}{S_A}\right)^{0.6} * \left(\frac{CEPCI_B}{CEPCI_A}\right)$$

Table 14 shows reference paper data for various process units. The capital cost calculated from reference papers is the fixed capital investment (FCI), which includes the cost of the land, equipment, piping, and instrument. The costs such as solvent flow rate and inventory come under working capital, which is assumed to be (15/85) fraction of FCI. Total capital cost investment (TCI) is calculated using below equation,

$$TCI = FCI + WCI$$

Annualized capital cost investment (ACI) is calculated using formula,

$$ACI = \frac{TCI - S}{n}$$

Where S is salvage value and n is the lifetime of a plant.

Depreciated Fixed capital investment (Dep. FCI) is calculated using formula,

$$Dep.FCI = \frac{FCI - FCI_s}{n}$$

Where FCI_S is FCI salvage value and n is the lifetime of a plant.

In these case studies, salvage value for calculating both ACI and FCI is assumed zero. Plant life is assumed to be 10 years. Table 15 below shows capital cost investment for different cases studied in this report.

Table 14: Reference paper data for capital cost estimation

Process Unit	Capital Cost (Million USD)	year	CEPCI	Flow Rate (kmole/hr)	Reference Paper
Acid Gas Removal	8.26	2008	575.4	18560	(L. Peters, A. Hussain, M. Follmann, T. Melin, & M. B. Hägg, 2011)
Dehydration	11.60	2002	395.6	2788	(Binci, Ciarapica, & Giacchetta, 2002)
NGL Recovery	7.38	2013	567.3	5000	(Getu et al., 2013)

Table 15: Capital cost for conventional technology

Case #	Flow Rate (kmole/hr)	FCI (MM $)	WCI (15/85*FCI)	TCI (MM $)	ACI (MM $)
Case-1	858.39	12.05	2.13	14.18	1.42
Case-2	515.03	8.87	1.57	10.44	1.04
Case-3	214.60	5.25	0.93	6.17	0.62
Case-4	141.63	4.09	0.72	4.81	0.48
Case-5	110.47	3.52	0.62	4.14	0.41

- Operating Cost Estimation

Operating cost comprises mainly of the cost of raw material, the cost for utility, labor expenses, maintenance cost and cost of solvent make-up. In this study, the cost for the make-up solvent is assumed negligible compared to raw material cost. For equipment integrity, maintenance cost (MC) per annum is assumed as 5% of the FCI. Prices used for raw material, products and by-products are given in Table 16 and prices for different utilities are mentioned in Table 17.

Total operating cost is calculated by summation of raw material cost, utility cost, maintenance cost, and labor charges. Eight thousand working hours per year is assumed for evaluating annualized operating cost (AOC). Labor cost is calculated by assuming two operators, a supervisor, an engineer and two staff with annual salaries of 50000, 75000, 75000, 40000 dollars per annum respectively with additional benefits of 45%. Operating cost for different cases is mentioned in Table 13.

Total operating cost is given by,

$$\boxed{TOC = Raw\ material\ cost + Utility\ cost + MC + Labor\ charges}$$

Table 16: Raw material and product prices

Description	Cost ($/MM Btu)	Cost ($/Kmole)
Shale gas	3	2.50
Methane	4.2	3.19
Ethane	4.5	6.12
Propane	4.5	8.91

Table 17: Utility prices

Utility Type	Cost ($/GJ)
Cooling Water	0.35
Electricity	16.8
Steam (IP)	6.08
Refrigeration cost	20
Chilled Water	0.35
Brine	7.89

- Profit Evaluation

For this study, the tax rate is assumed 30 percentage and plant life is assumed 10 years. Net Present Value (NPV), Payback period and Return On Investment (ROI) can be to determine the optimum plant size. For the project to be economically profitable NPV should be positive, Payback period should be as less as possible typically less than 3-4 years and ROI should be as high as possible typical minimum ROI is 10%.

Payback period (P) can be evaluated using the formula,

$$P = \frac{Depreciable\ FCI}{Annual\ net\ profit\ (after - tax)}$$

ROI is calculated using the formula,

$$ROI = \frac{Annual\ net\ profit\ (after - tax)}{TCI} * 100$$

NPV takes into account the time value of money. An interest rate (i) of 10% is assumed to accommodate the time value while calculating net profit at the end of first, second and third years. All the working capital investment can be recovered at the end of the plant-operated life and part of the fixed capital investment can be recovered based on assumed salvage value. For this study equation for NPV is given by,

$$NPV = -FCI - WCI + \frac{AATP(1)}{1+i} + \frac{AATP(2)}{(1+i)^2} + \frac{(AATP(3) + 0.7\,FCI + WCI)}{(1+i)^3}$$

Where, AATP (1), AATP (2), AATP (3) is annual after-tax profit at the end of first, second and third years respectively. Annual after-tax profit is calculated by subtracting the total operating cost from total revenue and deducting tax rate from it and dividing this amount by 3 for evenly distributing profit over three years.

Annual net profit after tax is estimated using the formula,

$$\begin{aligned} &Annual\ net\ profit\ (after - tax) \\ &\quad = (Annual\ gross\ profit - Dep.FCI) * (1 - tax\ rate) + Dep.FCI \end{aligned}$$

Annual gross profit in this study is determined by dividing cumulative gross profit calculated over the period of three years with number three. Gross profit is calculated using formula below,

$$Gross\ Profit = Total\ operating\ cost - Total\ revenue$$

Figure 19 shows the ROI for various cases in the conventional technology. From this figure, it can be inferred that case-3 provides optimum plant size case.

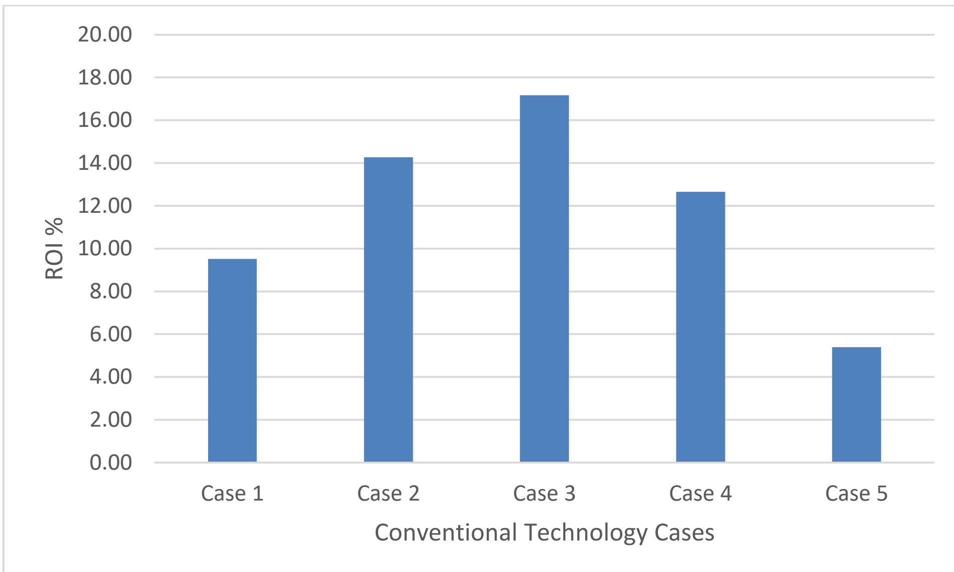

Figure 19: Economic analysis of conventional technology

5.1.2 Novel Technology Economics

When direct sequence distillation columns are replaced by a dividing wall column, there would be a 30% decrease in both capital and operating costs of the column (Smith & Smith, 2005). The economics for novel technology is carried out similar to conventional technology but with reduced operating cost and capital cost for NGL recovery unit. Utility consumed in de-methanizer and de-ethanizer columns is reduced by 30%.Also, the capital cost of NGL recovery unit is reduced by 30%, assuming that major capital cost is because of the distillation columns in NGL recovery unit. Figure 20 shows that the ROI for various cases of the novel and conventional technologies. From this figure, it is observed that case-3 is the optimum case and case-3 ROI percentage for novel technology is 21.34 %, which is 4.17 % more than conventional technology.

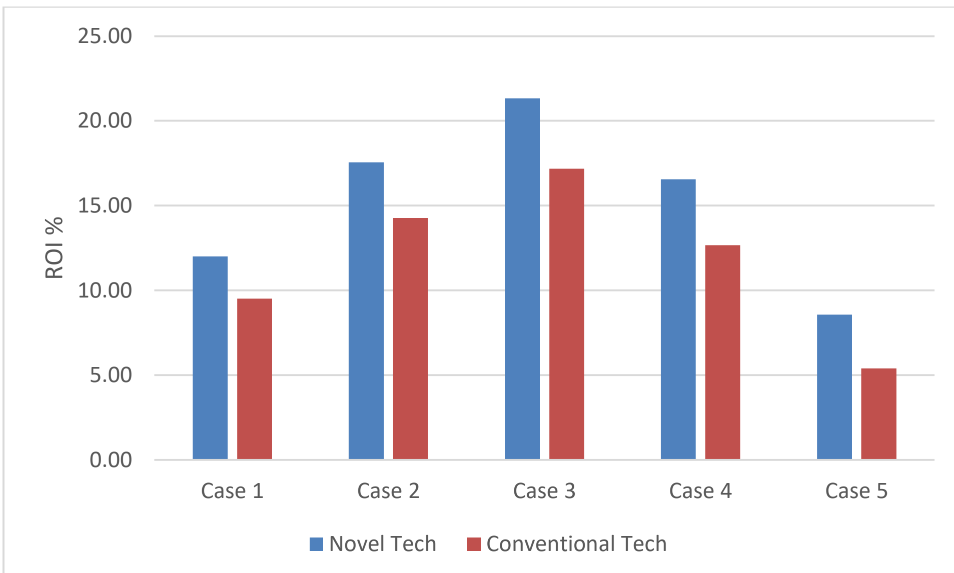

Figure 20: Comparison between conventional and novel technology economics

5.1.3 Excess Shale gas Economics

Excess shale gas from different cases can either be utilized as fuel to a steam boiler, and steam from this boiler is in turn used in a steam turbine to produce electricity, or else this feed gas can be further processed to be sold as salable pipeline quality gas.

❖ *Co-generation: Assuming Zero Fuel/Raw Material Cost*

Purchased cost of the steam turbine is evaluated using Figure 21 as a reference. Capital cost for the steam generation power plant is calculated by multiplying lang factor for fluids processing with steam turbine purchased cost.

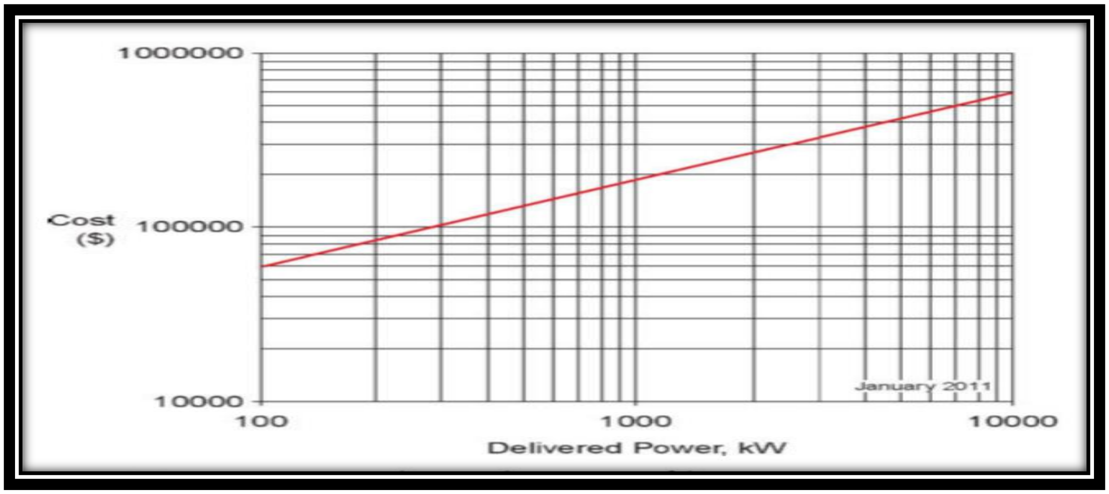

Figure 21: Purchased cost of steam turbine

Table 18: Co-generation economics with zero fuel cost

Description	UOM	Case-2	Case-3	Case-4	Case-5
Q Condenser Required	$	18067	215084	373470	498938
Net Electricity Produced	MM Btu	2899	34513	59927	80060
Net Electricity Produced sales	$	46162	549555	954241	1274821
Fixed Capital Investment	$	1770000	2478000	2655000	2743500
Depreciated Fixed Capital Investment	$	177000	247800	265500	274350
Dep. FCI (for given processing time)	$	70800	247800	531000	823050
Net profit after tax	MM $	0.0315	0.4882	0.8285	1.0853

Net electricity generated by the steam turbine power plant is evaluated by subtracting power utilized by condensate pump from electricity generated in a turbine. In addition, cooling water is used as a cold utility to provide heat duty to the condenser. Fuel cost to the boiler is assumed as zero. Profitability analysis of cogeneration opportunity is as shown in Table 18.

❖ *Excess Shale Feed Gas as Pipeline Gas Product*

Another alternative of utilizing the excess feed shale gas when the plant is under-designed is by processing it to pipeline gas quality and selling it as a product. For Marcellus shale play, the average impurities present in the feed is very less and thereby can be sold as pipeline quality natural gas without processing it through acid gas removal unit and dehydration unit. This feed gas can directly be sent to NGL recovery section to separate products. From Table 19, it is observed that net profit from operating co-generation process is very less compared to the gas revenue when sold as the product. Hence, it is advised that co-generation is not a viable option in this case, except for when there is no market for selling the excess gas as the product.

Table 19: Revenue from excess gas as pipeline gas Vs net profit from cogeneration

Case #	Net Profit (After-tax) Cogen (MM $)	Excess Gas Revenue (MM $)
Case-2	0.032	0.28
Case-3	0.49	3.33
Case-4	0.83	5.77
Case-5	1.09	7.71

5.1.4 Constant Production Rate Economics

In the economic analysis of previous cases, the capital and operating costs were calculated on cumulative production basis over a period of three years. However, a typical chemical engineering plant is designed for a plant life of about 10-20 years. Therefore, in this scenario, the economics of plant is performed assuming the plant is operating with constant throughput over its entire lifetime. For this scenario, the case-3 production rate is chosen, as it is the optimum plant size compared to other cases from above economic analysis. Constant feed flow rate condition is maintained by assuming the drop in flow rate is made up by flow from a grid. The flow to the grid is in turn maintained by drilling new wells as existing ones get dwindled. Table 20 shows the ROI and payback period for constant flow rate scenario for conventional and novel technology. From this table, it can be inferred that the ROI calculated for constant throughput case is higher than the minimum ROI percentage requirement, which is 10%, making it an economically viable project to implement.

Table 20: Economic analysis with constant production rate of case-3

Description	UOM	Conventional Technology	Novel Technology
Annualized Capital Cost	MM $	0.62	0.58
Raw material Flow per Annum	kmole	1716772.47	1716772.47
Raw material cost	MM $	4.30	4.30
Utility Cost	MM $	0.26	0.17
Product Revenue	MM $	6.57	6.57
Depreciated FCI	MM $	0.52	0.49
Net profit (After-tax)	MM $	0.45	0.49
ROI	%	24.18	28.23

5.2 Environmental Analysis

Environmental analysis for process technology is done by calculating the amount of CO_2 emitted from combustion of process vents and electricity consumed in the process. For cogeneration case, The CO_2 emissions due to combustion of the fuel gas are evaluated. Then net CO_2 emission is calculated by subtracting the CO2 utilized by selling electricity from total CO_2 emitted from fuel gas combustion. For simplicity purpose, 100% combustion of carbon is assumed, and from EPA website it is taken that 7.44*10^-4 Metric tons of CO_2/kWh of electricity consumed (EPA, 2016).Table 21 shows the specific metric ton of CO_2 emissions per ton of feed processed for conventional/Novel technology and for co-generation.

Table 21: Specific CO_2 emissions

Process Name	Specific CO_2 Emission (Tonne CO_2/ Tonne of Feed Processed)
Conventional/Novel Technology Cases	0.0087
Co-generation	0.8625

5.3 Results & Recommendations

❖ *The important results can be summarized as*

- The plant size designed for case-3 will provide optimum plant design with a ROI percentage of 17.17% for conventional technology and 21.34% for Novel technology
- Pinch analysis for utility targeting has saved the energy consumption by 31% of cold utility and 92% of hot utility.
- Co-generation option for the excess gas is profitable only when fuel gas rate is neglected. Selling excess gas as a natural gas with pipeline gas quality is observed to be more profitable than cogeneration.
- When assumed constant feed flow rate with optimum plant size i.e., case-3 for plant life of 10 years. The loss in feed rate with time is made up by drilling new wells. The ROI for conventional technology comes to be 24.2% and for novel technology, it is 28.2%.

- *The following recommendations are suggested for further work*
    - Exploring use of vent streams as a fuel
    - Further reduction in refrigeration consumption by employing technologies like IPSI-1 for NGL recovery section
    - Determining optimum cases based on Multi-Criteria analysis which includes sustainability, safety, and economics
    - Performing reservoir simulations for better statistical data estimates

5.4 Conclusion

This study presents the solution approach to design an optimum plant size for a given average well production rate declination curve. In this, a simulation study of shale gas processing and NGL recovery plant at five discrete shale gas feed flow rates (five cases) has been carried out, to determine the optimum plant size under uncertain feed conditions. Furthermore, pinch analysis for utility targeting, process modification of NGL recovery unit and cogeneration options are evaluated to optimize the energy consumption.

After performing economic analysis, it is found that case-3 provides the optimum plant design with better ROI percentage compared to other cases. To extend the solution approach from this study to a practical scenario, a constant production rate case is considered, in which flow rate is maintained constant by making up the loss in shale gas from a well with time from a grid system and flow rate to this grid is maintained by drilling new wells. ROI percentage for this constant production rate case is found to be more than minimum ROI requirement i.e., 10%, thereby making it economically viable implementation.

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