Design of a Natural Gas Processing Plant

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Chemical Engineering Capstone Project (CP303)

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Abstract

Natural gas is a valuable resource that is extracted from underground and undersea reserves and used as a source of energy worldwide. However, raw natural gas contains impurities which must be removed before the gas can be used efficiently. The design of a natural gas processing plant plays a crucial role in purifying and separating the components of raw natural gas, making it safe for transportation and use, and increasing its value. Proper design ensures that the processing plant operates safely, efficiently, and with minimal environmental impact. In this regard, this report provides a detailed overview of the plant design, including the equipment and processes used to achieve the desired separation and purification of the natural gas. The design process involves the use of DWSIM, an open-source process simulation software to ensure that the plant meets necessary performance standards, followed by design and costing of equipment via well established procedures. A break-even analysis has been conducted to establish the plant's economic viability, with the break-even period coming out to be about 4 months. The robustness of the plant's design under uncertainties in feed conditions has also been demonstrated via a sensitivity analysis. Lastly, a Hazard and Operability (HAZOP) study has been conducted for the propane chiller, a key equipment, which is followed by recommendations to augment safe operation.

Keywords: natural gas, impurities, plant design, DWSIM, simulation, break-even analysis, uncertainties, sensitivity analysis, HAZOP, propane chiller

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1. Introduction

Natural gas is a valuable energy resource that is produced from oil, gas and condensate wells. It contributes to about 22% of the global energy mix [1]. It is a cleaner source of energy than other fossil fuels and will play an important role during the transition to renewables, and is hence also known as 'transition fuel'. It is primarily composed of methane, a colourless, odourless and highly flammable gas. It also contains other hydrocarbons such as ethane, propane, and butane, as well as impurities such as water, carbon dioxide, and sulphur compounds. These impurities must be removed to ensure that natural gas can be safely transported and used as a fuel source. Natural gas processing plants play a critical role in the production of natural gas by purifying the raw gas and separating it into its constituent components, making the safe and efficient design of such plants a necessity.

The design of a natural gas processing plant involves multiple factors that must be taken into account. These factors include the characteristics of the raw gas, the desired composition of the processed gas, industry standards, environmental and safety considerations, and the economic feasibility of the project. Proper design is essential to ensure that the plant operates safely, efficiently, and sustainably. The first step involves characterizing the composition, temperature and pressure of raw natural gas as they influence gas behaviour during processing. Then, the desired composition of processed gas is determined based on industry standards which dictates the required processing steps such as dehydration, desulphurization and the removal of heavier hydrocarbons. Typical equipment used include separators, absorbers, distillation columns, heat exchangers, etc. Next, based on the processing plant's capacity, the equipment is adequately sized based on procedures available in the literature.

From the point of view of process economics, it is necessary to estimate the cost of a plant design to evaluate design options, optimize the process and determine profitability. For this, the equipment can be costed based on their size using historical data, indices, etc. However, the capital investment is much more than the mere cost of equipment and must include investments in utilities, instrumentation, off-site investments, working capital, etc. which must be accounted for [2]. Together with the operating expenditures, it can give us an idea of the project's feasibility and can be used to perform a break-even analysis to determine profitability.

Well-designed plants are generally robust to deviations and upsets in upstream conditions. They can offset deviations to ensure product quality is within control limits and that the process operates safely. A robust plant is desirable to ensure high efficiency, economic viability, reliability, scalability, low maintenance and downtime, and most importantly, safety. In the past, there have been several incidents at natural gas processing plants that highlight the importance of safe design and operation. In June 2016, a heat exchanger like allowed natural gas to leak and ignite, causing an explosion that killed one worker and injured three at a plant in Pascagoula, Mississippi. In April 2013, an explosion occurred at a plant in Opal, Wyoming operated by William Partners LP, resulting in the evacuation of nearby residences and a fire that lasted for several days. A heat exchanger leak was again responsible for an explosion at ExxonMobil's Torrance Refinery in February 2015 [3]. These incidents elucidate the importance of regular maintenance, inherently safe designs, proper safety procedures, hazard identification and risk assessment.

Thus, this project covers the important phases of designing of a natural gas processing plant in detail such as process simulations using DWSIM, followed by a sensitivity analysis to evaluate the plant's robustness under uncertainty. Next, the plant equipment is sized and costed and the break-even period is determined. Finally, hazards in the design and operation of a key component (propane chiller) have been identified along with the possible causes and consequences, and requisite safeguards have been recommended to enhance safety in a HAZOP study.

2. Literature Survey

Based on its chemical composition, particularly sulphur content, natural gas can be classified as sweet or sour. If there is little to no hydrogen sulphide present, the gas is termed as sweet while if the concentration is more than 5 mg per normal m^3, it is termed as sour [4]. Natural gas liquids, which contain ethane and higher hydrocarbons can be separated by cooling, oil absorption, adsorption or membrane processes and then fractionated. Water is removed as it reduces the heating value of the gas, condensation in pipes can cause slug flow, corrosion and erosion in conjunction with acid gases, and even hydrate formation. The corrosive acid gases, namely carbon dioxide and hydrogen sulphide can be removed via a number of processes. Chemical absorption processes include the amine process which uses alkanol amines, potassium carbonate washing and sodium hydroxide washing, while physical absorption processes include the rectisol, selexol and purisol processes. Adsorption by metal organic frameworks for example and membrane processes can also be used. Speaking of the amine process which is the most common for acid gas removal, the reactivity of monethanol amine is the greatest, followed by diethanolamine and diglycol amine. Hence, the solution strengths in water used are 15-20%, 30% and 60% by wt. respectively. Unlike MEA, DEA is not degraded by mercaptans, carbon disulphide and carbonyl sulphide and it is also less corrosive, making it the most favoured solvent [5]. Recent research has focused on the economic optimization of natural gas plants, development of advanced membrane processes, synthetic natural gas production, etc. There is also the need to reduce emissions using carbon capture storage and utilization, heat recovery, sulphur recovery, use of low NOx burners, etc. One important indicator of natural gas quality is the hydrocarbon dew point, which is the temperature at which heavy hydrocarbon components begin to condense out of the gaseous phase when the gas is cooled at constant pressure. A higher dew point indicates higher heavy hydrocarbon proportions and a greater risk of condensate formation in the pipeline.

3. Methodology

3.1. Sour Gas Processing

Raw natural gas enters a vertical gas-liquid separator to remove free water, and is then fed to an amine contactor for removal of acid gases. The absorbing medium is diethanolamine (DEA) in water at a 28 wt. % concentration. There are 20 real stages in the contactor. Before it reaches the lean/rich amine exchanger, the rich amine is flashed from the contactor pressure of 1000 psia to 90 psia to liberate the majority of the absorbed hydrocarbon gas. The rich amine is heated to a regenerator feed temperature of 200°F in the lean/rich exchanger. The regenerator has 20 real stages. Lean amine is regenerated at about 255°F, while acid gas is rejected from the regenerator at 120°F. After cooling in the lean/rich exchanger, the lean amine is recycled back to the contactor. The process simulation flowsheet has been created in DWSIM and is shown in Fig. 1.

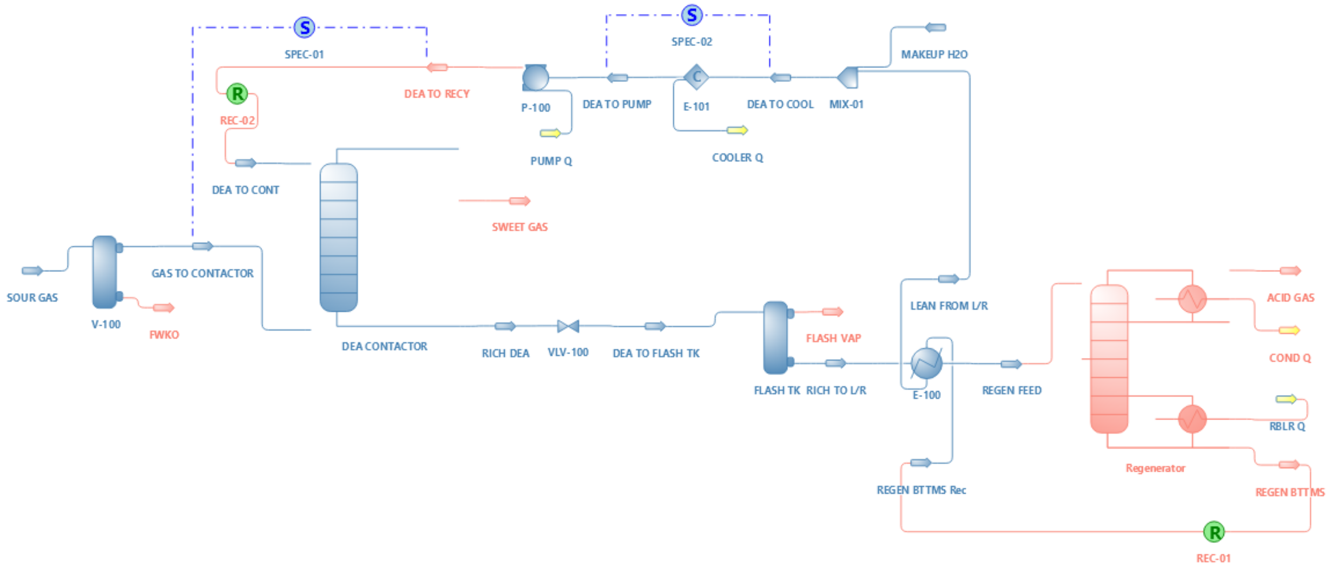


Fig. 1: Sour Gas Processing Flowsheet in DWSIM

The parts highlighted in red show errors in simulating the flowsheet. Unfortunately, DWSIM does not have the thermodynamic package required to adequately simulate the amine process and neither does gPROMS, the only licensed software available with the department. Also, not enough information to create a simulation is available in literature on other processes such as rectisol (specifically with regards to natural gas sweetening, unlike the generic acid gas removal from syngas). Thus, the simulation and subsequent sensitivity analysis, equipment sizing, costing, break-even and pinch analyses and emission reduction objectives had to be junked and efforts redirected towards the next processing step, i.e., processing of the sweetened gas.

3.2. Sweet Gas Processing

Two sweetened natural gas streams at 60°F and 600 psi (compositions shown in Tables 1 and 2) are mixed together and sent to a vertical gas-liquid separator to remove heavier hydrocarbons. The total molar flow of the two feed streams is 10 MMSCFD, which will henceforth be referred to as the plant capacity. The gaseous stream from the separator is cooled in a gas/gas heat exchanger and then further refrigerated in a propane chiller to 0°F. The incoming propane refrigerant is assumed to be a saturated liquid at -22°F. The motive is to condense the heavier organics so that they can be separated from the lighter ones in the low temperature separator. The gaseous stream from this separator is recycled back to the heat exchanger where it is heated. This *SalesGas* stream is then transported through pipes. A typical dew point specification for natural gas streams is 15°F at 800 psia. The liquid streams from both the separators are mixed and sent to a depropanizer with 12 real stages to obtain a bottoms product that is rich in butanes and lean in propane. The DWSIM flowsheet is shown in Fig. 2.

Tables 1 and 2: Species Mole Fractions in Feed1 and Feed2

Compound	Amount
Nitrogen	0.01
Carbon dioxide	0.01
Methane	0.6
Ethane	0.2
Propane	0.1
N-butane	0.04
Isobutane	0.04

Compound	Amount
Nitrogen	0.017879149
Carbon dioxide	0
Methane	0.62441266
Ethane	0.16656859
Propane	0.11358401
N-butane	0.034469153
Isobutane	0.043086441

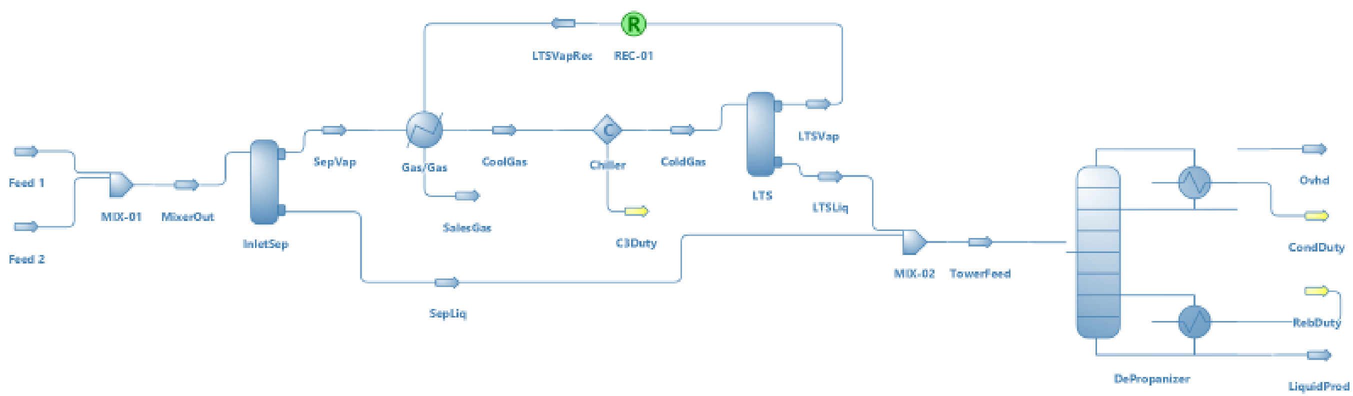


Fig. 2: Sweet Gas Processing Flowsheet in DWSIM

3.3. Sensitivity Analysis

The molar flow rate, temperature, pressure and methane/propane mole fraction of both feeds were varied one at a time and the variation of the *SalesGas* dew point at 800 psi and the mole fraction of methane in it were analysed to ascertain whether the design can handle uncertainties and fluctuations in upstream conditions and is robust enough.

3.4. Equipment Sizing

The main equipment in the flowsheet includes gas-liquid separators, a gas/gas shell and tube heat exchanger, a propane chiller and a distillation column along with its reboiler and condenser. Since the condenser operates at low temperatures, a propane chiller can be used for the purpose. Propane chillers further consist of a shell and tube heat exchanger working as the evaporator, a compressor and an air-cooler for condensing the propane refrigerant.

The superficial velocity of gas in the separator vessel can be determined from

$$v = K\sqrt{\frac{\rho_L - \rho_g}{\rho_g}} \qquad (1)$$

where ρ_L and ρ_g are the liquid and gas densities respectively in lb/ft^3 and K = 0.18 or 0.265 for a vessel height (H) of 5 ft or 10 ft. Dividing the gas volumetric flow rate by its superficial velocity gives the vessel cross-sectional area and consequently its diameter (D). H/D should be between 2 and 5 [6]. Finally, the vessel volume can be calculated from its dimensions.

Distillation column design involves selecting a trial plate spacing, estimating column diameter based on flooding considerations, deciding the flow arrangement and plate layout, and checking weeping, entrainment, plate pressure drop and downcomer backup. Column thickness is calculated based on pressure considerations, after which the column shell's volume and consequently mass is calculated using its dimensions. The detailed procedure will not be stated here in the interest of brevity and can instead be found in [7].

For the propane chiller, in kW, the duty of the air-cooler (Q_{cond}) and the compressor power (Q_{comp}) per MW of refrigeration can be found using the following correlation [8]:

$$Q_{cond} \text{ and } Q_{comp} \ (per\ MW\ refrigeration) = \sum_{i=0}^{3} \left(\sum_{j=0}^{3} a_{ij} T_{cond}^{j} \right) T_{evp}^{i} \qquad (2)$$

where T_{cond} and T_{evp} (in K) are the refrigerant temperatures in the air-cooled condenser and the evaporator respectively. The list of a_{ij} is available in [8]. T_{evp} is taken as 308.15 K for both the gas chiller as well as the depropanizer's condenser, while T_{cond} is taken as 243.15 K and 263.15 K respectively.

Next, the overall heat transfer coefficient (U) for the shell and tube heat exchangers and air-coolers can be estimated from Fig. 3. Having calculated the log mean temperature difference (ΔT_{LM}) and knowing the exchanger duty (Q), the heat transfer area can be calculated as:

$$A = \frac{Q}{U \times \Delta T_{LM}} \qquad (3)$$

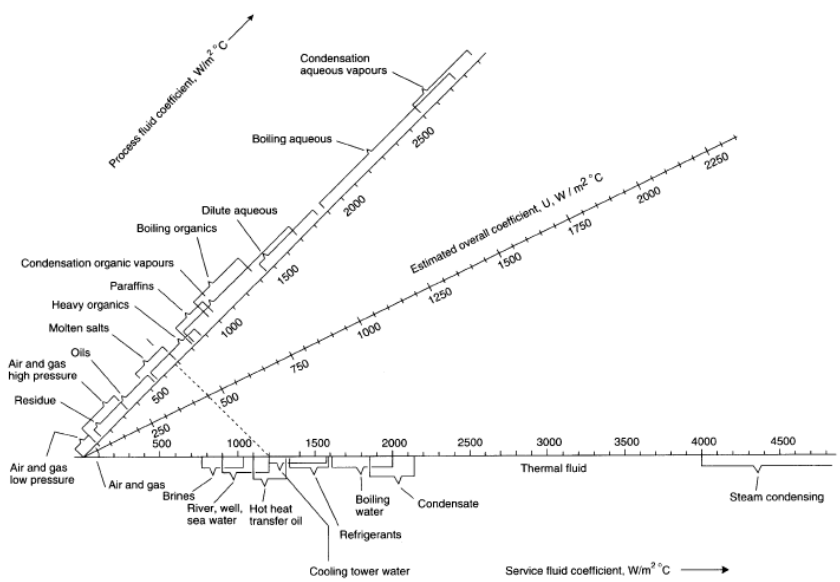


Fig. 3: Estimation of Overall Heat Transfer Coefficient, Source: [7]

3.5. Plant Costing

The cost of each equipment (C_E) assuming carbon steel as the material of construction and moderate temperature and pressure is calculated using:

$$C_E = C_B \left(\frac{Q}{Q_B}\right)^M \frac{INDEX_2}{INDEX_1} \qquad (4)$$

where Q is the size or capacity of the equipment, Q_B is the size or capacity of the base equipment, C_B is the cost of the base equipment and M is a constant depending on equipment type. Data on C_B and M can be found in [2], [9]. The index values are used to adjust for the time value of money. The costs are adjusted to USD_{2022} using the Chemical Engineering Plant Cost Indices (CEPCI). Finally, the fixed capital investment is calculated as:

$$C_F = \sum_i [f_M f_P f_T (1 + f_{PIP})]_i C_{E_i} + (f_{ER} + f_{INST} + f_{ELEC} + f_{UTIL} + f_{OS} + f_{BUILD} + f_{SP} + f_{DEC} + f_{CONT} + f_{WC}) \sum_i C_{E_i} \qquad (5)$$

The actual equipment cost and piping cost depend on the material of construction and the design temperature and pressure. The costs for erection, installation, electricals, utilities, offsites, buildings, site preparation, design engineering construction, contingency and working capital will be largely unchanged and hence calculated using the unadjusted values.

3.6. Break-Even Analysis

The plant's capital cost has been estimated, and we assume the yearly operating expenditures as 10% of the capital cost. At the end of the break-even period (n years), the sum of operating expenditures incurred during this period and the capital cost would equal the revenue generated during the period. We assume the sale of *LiquidProd* (butanes) and *SalesGas* (natural gas) as the sources of revenue. For natural gas, the price is USD_{2022} 6.45/MMBtu [10], with 1 MMSCFD being equivalent to 1037 MMBtu [11]. For butanes, the price is USD_{2022} 13.75/MMBtu (after adjustment from USD_{2021} 12.45/MMBtu, which is the average for n-butane and isobutane [12], using US Producer Price Index), with 1 MMSCFD being equivalent to 3225 MMBtu [13].

Fixed capital investment + n*Yearly operating expenses = n*Revenue (7)

3.7. Hazard and Operability (HAZOP) Study

A HAZOP study is a formal and effective procedure to identify hazards in a chemical process facility. The process flowsheet is broken down into elements, and nodes related to each element are analysed in the study. A process parameter relevant to the node is picked and a relevant guide word is applied to suggest possible deviations. Possible deviation causes and consequences are identified, and actions recommended to reduce the likelihood of occurrence and the severity of consequences. A component of the propane chiller shown in Fig. 2 is a shell and tube heat exchanger working as the evaporator in which propane would evaporate while cooling the gas stream (Fig. 4). The propane enters the heat exchanger as a saturated liquid at -22°F and leaves as a saturated vapour at the same temperature. In this project, the chiller's evaporator/shell and tube heat exchanger is chosen as the study element, and the nodes are the streams *CoolGas*, *ColdGas*, *Propane In* and *PropaneOut*, the heat exchanger itself and the control system. Safeguards and design features are then recommended as a consequence of the HAZOP study.

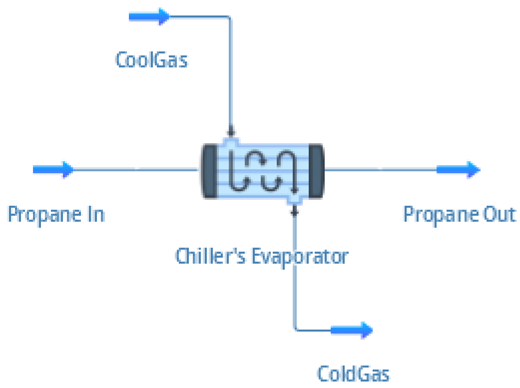


Fig. 4: Shell and Tube Heat Exchanger working as Propane Chiller's Evaporator

4. Results and Discussion

Now that the process design and analysis methodology have been described, their performance and results must be discussed.

Firstly, the sweet gas processing flowsheet created in DWSIM is run. The compositions of the solved *SalesGas, Ovhd* and *LiquidProd* streams are shown in Tables 3, 4 and 5.

Tables 3, 4 and 5: Species Mole Fractions in SalesGas, Ovhd and LiquidProd

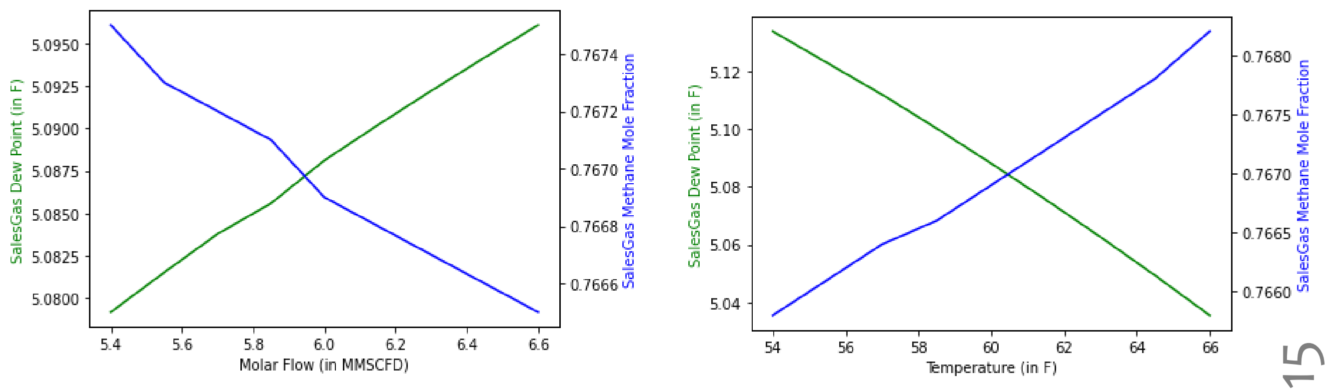
Compound	Amount
Nitrogen	0.017855327
Carbon dioxide	0.0065852936
Methane	0.76692577
Ethane	0.15614152
Propane	0.04135932
N-butane	0.0044882365
Isobutane	0.0066445305

Compound	Amount
Nitrogen	0.0026576718
Carbon dioxide	0.006106781
Methane	0.31427879
Ethane	0.34265314
Propane	0.3340397
N-butane	2.0022649E-05
Isobutane	0.00024389178

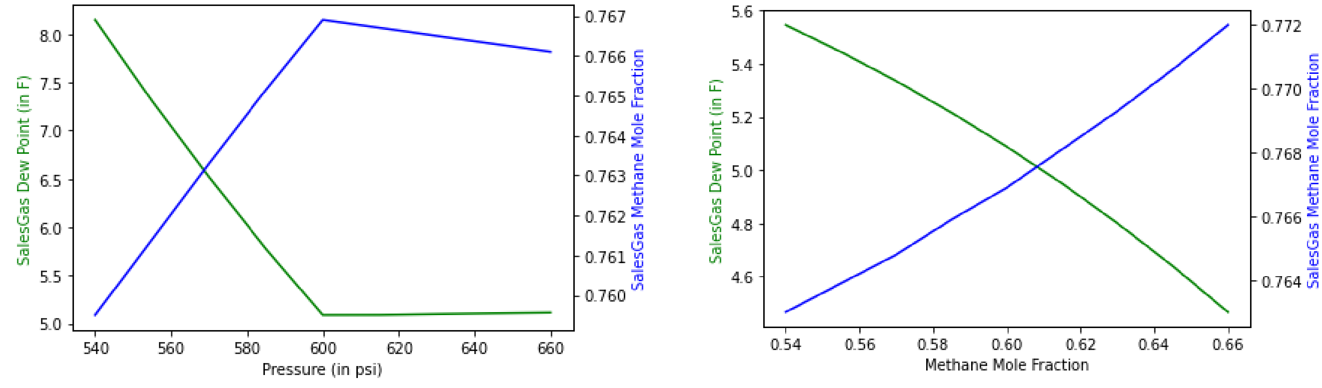
Compound	Amount
Nitrogen	4.3302336E-20
Carbon dioxide	2.1602149E-12
Methane	5.7153222E-14
Ethane	2.350263E-07
Propane	0.019998648
N-butane	0.47704749
Isobutane	0.50295362

The *SalesGas* stream is very rich in methane with a mole fraction of 0.7669. The mole fraction of heavier hydrocarbons is very less. Further, the dew point of the stream at 800 psi is 5.08812°F, which is much less than the maximum limit of 15°F and is an indicator of the high quality of the stream and its suitability for pipeline transport. Thus, the design achieves the goal of producing natural gas that meets industry standards. Next, as expected, the mole fraction of propane in the bottoms product obtained from the depropanizer is very low. The stream is rich in n-butane and isobutane which can be sold. The overhead product contains nearly one-third of each of methane, ethane and propane which can be further separated or used as internal fuel for the plant.

Having demonstrated the design's performance, its robustness to handling fluctuations and uncertainties in feed conditions must be determined. The results of the sensitivity analysis are shown in Figs. 5 and 6.

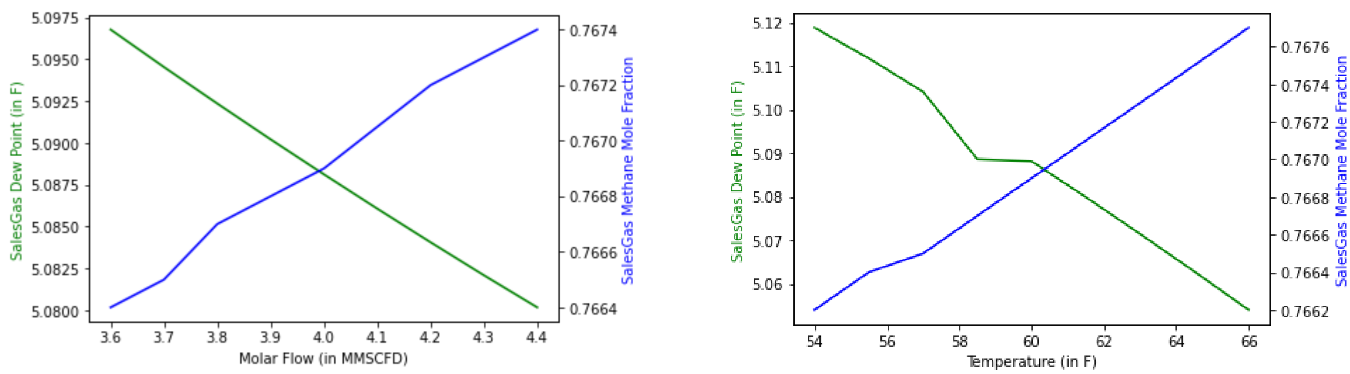


(a) (b)

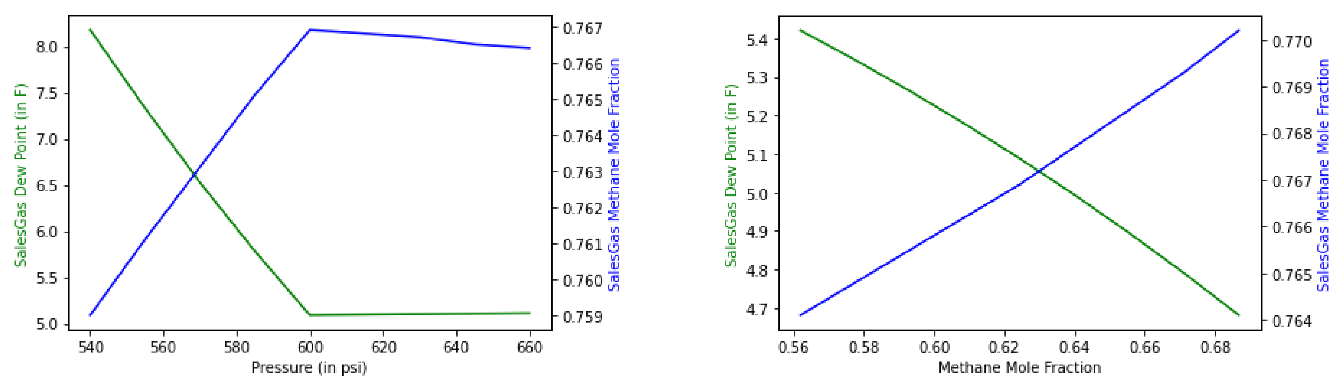


(c) **(d)**

Fig. 5: SalesGas Dew Point (at 800 psi) and Methane Mole Fraction vs Feed 1's a) Molar Flow, b) Temperature, c) Pressure, d) methane Mole Fraction



(a) **(b)**



(c) **(d)**

Fig. 6: SalesGas Dew Point (at 800 psi) and Methane Mole Fraction vs Feed 2's a) Molar Flow, b) Temperature, c) Pressure, d) methane Mole Fraction

The mole fraction of methane is always in the range of 0.76. The dew point is also in the range of 4-6°F, except when operating at lower feed pressures where it is in the range of 8°F. Nevertheless, the dew point is always less than the maximum value of 15°F. Clearly, the design is robust under uncertainty. When methane content increases in the final product, the dew point decreases due to a decrease in the concentration of heavier hydrocarbons. The data generated in this sensitivity analysis may be used to create a predictive model for predicting SalesGas properties based on feed conditions.

The next step is the sizing of equipment and plant costing. The size/capacity of equipment and their costs are summarized in Table 6.

Table 6: Size/Capacity and Cost of Process Equipment

Equipment	Attribute	Value	C_E (in USD_{2022})
InletSep	Volume	0.483 m^3	9700.58
LTS	Volume	0.303 m^3	6811.05
Chiller's evaporator	Area	26.975 m^2	2476.74
Chiller's compressor	Power	149.042 kW	1.96×10^5
Chiller's air-cooler	Area	393.486 m^2	5.905×10^5
Depropanizer	Shell Mass	957.7 kg	37151.16
10 Sieve Trays	Diameter	0.892 m	22750.72
Reboiler	Area	47.808 m^2	2683.14
Condenser's evaporator	Area	45.269 m^2	2641.86
Condenser's compressor	Power	82.856 kW	1.14×10^5
Condenser's air-cooler	Area	318.771 m^2	4.889×10^5

The C_E values are used in equation (5) to get the fixed capital investment which comes out to be USD_{2022} 9.09 million.

Based on the assumption that the yearly operating expenditures are 10% of the capital investment, the former comes out to be USD_{2022} 0.909 million/year. Equating the sum of capital investment and operational expenses incurred during the break-even period to the revenue generated during the same period yields a break-even period of 0.3245 years or 3.9 months. This indicates that all investments are recovered at the end of 3.9 months and the operation of the plant becomes profitable very quickly. However, a drawback of the break-even analysis is that it does not take into account the time-value of money.

Table 7 presents the HAZOP study conducted on the propane chiller's heat exchanger. Action items have been suggested to prevent or combat the causes and consequences of deviations in process parameters.

Table 7: HAZOP Study for Evaporator of Propane Chiller

Item	Study Node	Process Parameter	Guideword (Deviation)	Possible Causes	Possible Consequences	Action Required
1A	*Cool Gas*	Flow	No	1. Gas supply failure	No gas, downstream processes upset	Install flow meter and low flow alarm
				2. Valve closure	Same as above	Same as above, add valve lock
				3. Line plugging	Same as above	Same as 1A.1, regular maintenance
				4. Line rupture	Same as above, gas leakage, flammable mixture formation risk	Same as above, install automatic shutdown system, bypass line, adequate ventilation and fire protection system
1B			More	1. Increased supply pressure	High gas flow, downstream process upset	Install flow meter, high flow alarm, relief valve and ventilation, pressure sensor and high-pressure alarm
1C			Less	1. Partial gas supply failure	Low gas flow, downstream process upset	Same as 1A.1
				2. Low supply pressure	Same as above	Same as 1A.1, install pressure sensor and low-pressure alarm
				3. Partial valve closure	Same as above	Same as 1A.2
				4. Partial line plugging	Same as above	Same as 1A.3
				5. Line rupture	Same as above, gas leakage, flammable mixture formation risk	Same as 1A.4
1D/E/F			As well as/Part of/Other than	1. Supply contamination	Contamination, downstream process upset	Quality control
1G			Reverse	1. High downstream pressure	Backflow into supply	Install check valve, pressure sensor, high-pressure alarm, relief valve and ventilation, check downstream process
1H/I			Early/Late	1. Early/late valve opening	Downstream process upset	Check downstream process
1J/K		Temperature	High/Low	1. Temperature control failure	Line and shell damage under stress, gas leakage and combustion, downstream process upset	Install temperature sensor, alarm, redundant control system and power backup,

						fire protection system regular maintenance
1L		Pressure	Low	1. Control valve failure	Reduced cooling, downstream process upset	Regular maintenance, install pressure sensor and alarm
				2. Pump failure	Same as above	Same as above, install backup pump and power
				3. Blockages	Same as above	Same as 1L.1
				4. Line rupture	Same as 1C.5	Same as in 1A4
1M			High	1. Control valve failure	Line and shell damage under stress, gas leakage and combustion, downstream process upset	Same as 1L.1, install relief valve, fire protection system and ventilation
				2. Outlet blockage	Same as above	Same as above
2A	*Propane In*	Flow	No	1. Compressor failure	No propane, no cooling of gas, downstream processes upset	Install flow meter and low flow alarm, regular maintenance, install backup compressor and power
				2. Valve closure	Same as above	Install flow meter and low flow alarm, add valve lock
				3. Line plugging	Same as above	Install flow meter and low flow alarm, regular maintenance
				4. Line rupture	Same as above, propane leakage, flammable mixture formation risk	Same as above, install automatic shutdown system, bypass line, fire protection system and adequate ventilation
2B			More	1. Increased supply pressure	High propane flow, unnecessary extra gas cooling, propane and compressor power wastage, downstream process upset	Install flow meter, high flow alarm, relief valve and ventilation, pressure sensor and high-pressure alarm
2C			Less	1. Compressor malfunction	Low propane flow, reduced gas cooling, downstream process upset	Same as 2A.1
				2. Low supply pressure	Same as above	Install flow meter and low flow alarm, pressure sensor and low-pressure alarm
				3. Partial valve closure	Same as above	Same as 2A.2
				4. Partial line plugging	Same as above	Same as 2A.3
				5. Line rupture	Same as above, propane leakage, flammable mixture formation risk	Same as 2A.4
2D/E/F			As well as/Part of/Other than	1. Supply contamination	Contamination, cooling upset, downstream process upset	Quality control
2G			Reverse	1. High downstream pressure	Backflow into supply	Install check valve, pressure sensor, high-pressure alarm, relief valve and ventilation, check downstream process

2H/I			Early/Late	1. Early/late valve opening	Wasteful chiller operation, downstream process upset	Check downstream process
2J/K		Temperature	High/Low	1. Temperature control failure	Line and tube damage under stress, propane leakage and combustion, reduced/excess gas cooling, downstream process upset	Install temperature sensor, alarm, redundant control system and power backup, fire protection system, regular maintenance
2L		Pressure	Low	1. Control valve failure	Reduced gas cooling, downstream process upset	Regular maintenance, install pressure sensor and alarm
				2. Compressor failure	Same as above	Same as above, install backup compressor and power
				3. Blockages	Same as above	Same as 1A.1
				4. Line rupture	Same as 2C.5	
2M			High	1. Control valve failure	Line and tube damage under stress, propane leakage and combustion, downstream process upset	Same as 2L.1, install relief valve, fire protection system and ventilation
				2. Outlet blockage	Same as above	Same as above
3A	*Cold Gas*	Flow	No	1. Same as 1A		
				2. Shell damage in exchanger	No gas, downstream process upset, gas leakage, flammable mixture formation risk	Install flow meter, low flow alarm, automatic shutdown system, adequate ventilation, low flow alarm in exchanger, fire protection system
3B			More	1. Same as 1B		
3C			Less	1. Same as 1C		
				2. Shell damage in exchanger	Same as 3A.2	
3D/E/F			As well as/Part of/Other than	1. Supply contamination	Contamination, downstream process upset	Quality control
				2. Tube damage in exchanger	Propane leakage causes contamination, downstream process upset	Regular maintenance, install high flow alarm in exchanger and line
3G			Reverse	1. Same as 1G		
3H/I			Early/Late	1. Same as 1H/I		
3J/K		Temperature	High/Low	1. Same as 1 J/K		
				2. Inadequate propane flow and specifications	Same as 2	
				3. Interaction with external environment	Same as 1J/K	Install and maintain insulation
3L		Pressure	Low	1. Same as 1L		
3M			High	1. Same as 1M		
4A	*Propane Out*	Flow	No	1. Same as 2A		
				2. Tube damage in exchanger	No propane, downstream process upset, propane leakage and mixing with gas, flammable mixture formation risk	Install flow meter, low flow alarm, automatic shutdown system, adequate ventilation, high flow alarm in exchanger, fire protection system

4B			More	1. Same as 2B		
4C			Less	1. Same as 2C		
				2. Tube damage in exchanger	Same as 4A.2	
4D/E/F			As well as/Part of/Other than	1. Supply contamination	Contamination, downstream process upset	Quality control
				2. Tube damage in exchanger	Propane leakage causes contamination, downstream process upset	Regular maintenance, install high flow alarm in exchanger and line
4G			Reverse	1. Same as 2G		
4H/I			Early/Late	1. Same as 2H/I		
4J/K		Temperature	High/Low	1. Same as 2 J/K		
				2. Inadequate gas flow and specifications	Same as 1	
				3. Interaction with external environment	Same as 2J/K	Install and maintain insulation
4L		Pressure	Low	1. Same as 2L		
4M			High	2. Same as 2M		
5A/B	*Heat Exchanger*	Temperature	Low/High	1. Same as 1/2/3/4J/K		
5C/D		Pressure	Low/High	1. Same as 1/2/3/4L/M		
5E/F		Normal Functioning	Fouling/Corrosion	1. Supply contamination	Leakage, reduced heat transfer, downstream process upset	Regular maintenance, supply quality control
				2. Chemical deposition	Same as above	Same as above
5G			Damage	1. Same as 1/2/3/4J/K/M		
6A	*Control Systems*	Normal Functioning	Failure	1. Faulty sensors	Incorrect readings, upsets in heat transfer and downstream processes, damage to equipment	Regular maintenance, install redundant control systems
				2. Circuital problems	Same as above	Same as above
				3. Power loss	Same as above	Same as above, install backup power

The chief recommendations from this study would be the installation of temperature, pressure and flow sensors, alarms, backup rotating equipment and power lines, bypass lines, automatic shutdown systems, fire protection systems and adequate ventilation, and performing regular maintenance. In addition, it would be necessary to add control valves to all nodes along with redundant control systems. These measures would augment the equipment's safety and mitigate risks.

5. Conclusions

In this project, the design of a natural gas processing plant has been covered comprehensively. The focus was on the design and simulation of the plant in DWSIM software. The simulation was used to evaluate the plant's performance and yielded good results in terms of the natural gas composition, its dew point specification and the butane rich bottoms product from the depropanizer. A sensitivity analysis was carried out to determine how the process would perform under different feed conditions, and the results showed that the plant was robust and could handle variations in feed flow rate, temperature, pressure and composition. The equipment was sized based on the design specifications, and the cost of the plant was estimated. The break-even analysis showed that the plant would be economically viable and profitable within nearly 4 months of operation. Finally, a HAZOP study was conducted on the propane chiller's evaporator to identify potential hazards and to recommend risk mitigation measures, a critical requirement to ensure safe and reliable operation.

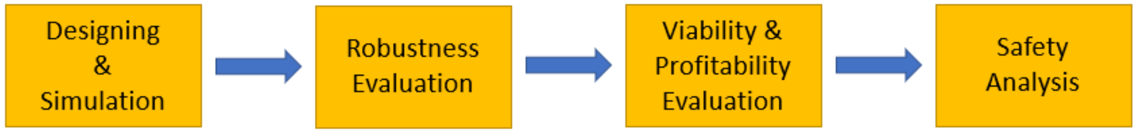


Fig. 7: Project Workflow

Overall, the project demonstrates the importance of a systematic and rigorous approach to the design and operation of natural gas processing plants. Although the analysis would have been more complete in the presence of a fully simulated sour gas processing flowsheet, satisfactory and reliable results and insights fulfilling project objectives have been obtained despite its absence.

6. Future Scope

This study has focused on analysing sweet gas processing in detail. A more in-depth safety analysis can be performed by undertaking a HAZOP study for all equipment in the flowsheet. A vital step towards a complete and holistic natural gas processing plant design would be the simulation of the sour gas sweetening process in a commercial simulation software such as ASPEN, UniSim or PRO\II which have the requisite thermodynamic package and capabilities. This would unlock the doors to equipment sizing and plant costing whose results could be used in a break-even analysis of the entire plant. The two flowsheets could then be integrated into one, and the availability of more streams for a pinch analysis means that they can be paired to reduce the consumption of energy and utilities, making the process sustainable, greener and more economically viable. Further, the data generated in the sensitivity analysis could be used to create a predictive model. Lastly, the possibility of undertaking facility siting based on hazard identification and accident modelling could be explored.

7. References

[1] "IEA: Global short-term gas demand growth comes to a halt | Oil & Gas Journal." https://www.ogj.com/general-interest/economics-markets/article/14283648/iea-global-shortterm-gas-demand-growth-comes-to-a-halt (accessed Mar. 18, 2023).

[2] Robin Smith, *Chemical Process Design and Integration- 2nd Edition*, 2nd Edition. John Wiley & Sons, 2016.

[3] "List of natural gas and oil production accidents in the United States - Wikipedia." https://en.wikipedia.org/wiki/List_of_natural_gas_and_oil_production_accidents_in_the_United_States (accessed May 10, 2023).

[4] S. Faramawy, T. Zaki, and A. A. E. Sakr, "Natural gas origin, composition, and processing: A review," *J Nat Gas Sci Eng*, vol. 34, pp. 34–54, Aug. 2016, doi: 10.1016/J.JNGSE.2016.06.030.

[5] D. S. J. Jones and P. R. Pujadó, *Handbook of petroleum processing*. Springer, 2006.

[6] C. RICHARD SIVALLS, "OIL AND GAS SEPARATION DESIGN MANUAL," 2010. Accessed: Mar. 18, 2023. [Online]. Available: https://pacs.ou.edu/media/filer_public/c9/4a/c94a97ac-9609-4262-ab06-

b7b2dda1c4fa/3_oil_and_gas_separation_design_manual_by_c_richard_sivalls.pdf

[7] RK Sinnott, *Coulson & Richardson's Chemical Engineering*, vol. 6. 2003.

[8] A. Bahadori and H. Vuthaluru, "Estimating compressor power and condenser duty in a refrigerant system A simple-to-use predictive tool calculates compressor power and condenser duty per refrigeration duty in a three-stage propane refrigerant system," 2010, Accessed: Mar. 18, 2023. [Online]. Available: www.digitalrefining.com/article/1000520

[9] Max S. Peters, Klaus D. Timmerhaus, and Ronald E. West, *Plant Design and Economics for Chemical Engineers*, Fifth Edition. McGraw-Hill Higher Education, 2003.

[10] "U.S. Energy Information Administration - EIA - Independent Statistics and Analysis." https://www.eia.gov/todayinenergy/detail.php?id=55119 (accessed May 10, 2023).

[11] "Frequently Asked Questions (FAQs) - U.S. Energy Information Administration (EIA)." https://www.eia.gov/tools/faqs/faq.php?id=45&t=8 (accessed May 10, 2023).

[12] "Prices for hydrocarbon gas liquids - U.S. Energy Information Administration (EIA)." https://www.eia.gov/energyexplained/hydrocarbon-gas-liquids/prices-for-hydrocarbon-gas-liquids.php (accessed May 10, 2023).

[13] "Fuel Gases - Heating Values." https://www.engineeringtoolbox.com/heating-values-fuel-gases-d_823.html (accessed May 10, 2023).