



Matthew Efe Takerhi

ORCID: 0000-0002-9098-1177
AGH University of Science and Technology in Krakow

SEPARATION OF NATURAL GAS LIQUIDS AND WATER FROM GAS CONDENSATE

Date of submission:
21.05.2021

Date of acceptance:
14.07.2021

Date of publication:
31.07.2021

© 2021 Author. This is an open access publication, which can be used, distributed, and reproduced in any medium according to the Creative Commons CC-BY 4.0 License

<https://journals.agh.edu.pl/jge>

Abstract: Unlike dry gas reservoirs, condensate gas reservoirs contain a considerable amount of natural gas liquids which should be extracted to maximize energy usage. This paper uses Bryan ProMax to set up the processing units for the recovery of natural gas liquids and removal of water. The parameters for the simulation were a gas composition which assumes a sweet gas content. The outcome of the simulation includes reduction of water content below 7 lbm/MMscf, recovery of methane and recovery of propane and isobutane only. The glycol dehydration unit minimized water impurity, while cooling with Joule–Thomson valve and heat exchangers help in methane recovery and separation from natural gas liquids. The results show that natural gas liquid recovery which depends on gas composition can be recovered by controlling the conditions of several units, namely heat exchangers, flash vessels, cold separators, fractionators, stabilizers with their reboilers and condensers.

Keywords: gas condensate, Natural Gas Liquids (NGL), fractionations

1. Introduction

In 1910, a displeased motorist complained to Dr Walter Snelling of the U.S Bureau of Mines that half of his gas tank had disappeared when he got home. Snelling examined a sample and discovered that the disappearing liquid was evaporating gases. He made his distillation column using an old water heater, some copper tubing, and wire-wrapped glass bottles. This allowed him to collect the gases as liquid and to patent the first process for capturing natural gas liquids. By so doing Snelling was able to separate the fraction of disappearing liquid into a separate product using fractional distillation [1]. In the middle of the 20th century, the natural gas business was not clearly noticed, it was an obscure business in production and exploration companies. The growth in the petrochemical and transportation industry led to more demand for natural gas liquids. More recently, the increasing supply of natural gas liquids has put the product on the business agenda [1].

Natural gas liquids (NGL) are produced from natural gas wells or from associated gas oil wells. In gas wells, natural gas liquids also called condensate which is separated at the separator and further removal is carried out in the gas plant. Gas plants clean up contaminants in the gas and NGL, separates NGL from the natural gas, and then several fractionators separate the NGL from each other. The various NGL, each meeting its own specification, move into several shared markets such as refining and petrochemical feedstocks, burner tips and internal combustion engine [1].

Natural gas mainly contains methane and natural gas liquids, which include ethane, LPG (propane and butane), gas-condensate. Methane is commonly known as fuel gas used for heating and power plants. Additionally, NGL are also valuable as separate products.

Ethane is very valuable in the petrochemical industry but as the most volatile NGL it is used almost exclusively as a petrochemical feedstock; in the petrochemical plants, ethane is converted to ethylene, and changed to polyethylene or ethyl alcohol, or various ethylene-based petrochemicals.

Propane is too volatile to be placed in gasoline and too valuable and heavy to be left with natural gas, it is used as a fuel for heating, cooking and sometimes lighting; as a petrochemical feedstock to make both ethylene and propylene as co-products; as a motor fuel; petrochemical companies use propylene to make polypropylene and isopropyl alcohol and other products.

Butane is less volatile than propane but is still shipped and handled under pressure as liquids. Normal butane is used in the fuel market for heating and cooking, though not as much as propane. It is used as a gasoline-blending component to add little volatility. It is also

used in the petrochemical plants to produce ethylene, propylene, butylene and butadiene, while LPG are used for cooking and heating.

Natural gasoline exist as a liquid at room temperature and pressure, has a low octane number, although refineries can clean up natural gasoline and even upgrade it and blend it with better components to make it suitable for sale at gas stations. Natural gasoline is also used as diluents for very heavy crude oils, and it reduces the viscosity of crude oils so it can be moved in a pipeline [1]. It is therefore profitable to separate the ethane, propane and butane and gasoline from the natural gas.

1.1. Description of NGL extraction

The NGL are first extracted from the natural gas and later separated into different components. Let us focus on optimization of the gas-condensate exploitation processes in the recovery of LPG and ethane [1]. The order of processes in natural gas processing depends on its composition. The first process is separation, which consists of one or more separators used to set condensation and remove the free water, two-phase gas liquid separators are used for separation of two-phase gas liquids, while three-phase separators are used for three-phase component containing oil, gas, and water.

The well stream from the gas condensate reservoir undergoes staged separation to separate the gas from the condensate. Stage separation is the process by which gaseous and liquid hydrocarbons are flashed into vapor and liquid phases by two or more separators. The aim of separation is to achieve maximum liquid recovery and stabilization of the gas products [2]. A large pressure reduction in a single separator will cause flash vaporization and is a safety hazard. In the first stage of separation, water content is usually reduced to 5%; there must be a minimum pressure difference between each stage of the separation to ensure satisfactory performance in pressure and level control loops.

As gas flows into the separator, the inlet diverter causes a sudden change in momentum which results in settling of the liquids droplets, other liquids droplets fall to the bottom of the separator due to gravity and are separated into oil and water phase. Vapor rises upwards and passes through the demister, which have tiny perforations that condense the tiny vapor in the gas into the liquids, which fall back to the bottom of the separator. The second separator is also known as the intermediate separator, its pressure is reduced for optimum recovery of methane gas and natural gas liquids, usually to the level of atmospheric pressure where temperature should be below 100°C, and water content will be reduced to about 2%. The pressure of this separator should be modified to ensure that the optimum amount

of natural gas liquids is recovered from the gas stream. The condensate stream is flashed into the low-pressure separator. Processing facilities would often have either two or three stage separators. In two stage separators, only one intermediate separator is possible while a three-stage separator allows for either one operating at second or third stage suction pressure [3].

The last separator is usually at standard conditions of 15°C and 14.7 psi. The gases from the last stage separator have lost so much pressure that they must be recompressed for transportation [2: 43] and should be cooled and recompressed before transfer to the natural gas pipeline. Natural gas liquids from the bottom of the separator are dehydrated. Table 1 shows stage separation guidelines [3: 45]. The condensates are sent to the refinery while raw natural gas is transported to a gas processing plant. The natural gas is treated to remove acid gases. For fractionation of the natural gas liquids, Abdel Aal et al., suggests two essential requirements for designing a fractionating unit: understanding the process used for separation; the guidelines to determine the sequence of separation [4].

Table 1. Stage separation guidelines [3]

Initial separation pressure		
[psig]	[kPa]	number of stages
25–125	170–860	1
125–300	860–2100	1–2
300–500	2100–3400	2
500–700	3400–4800	2–3

1.2. Gas dehydration process

The next unit is the dehydration unit which uses either regenerable absorption or a regenerable adsorption unit for removing water vapor from gas. In an absorption process, a special solvent is used as the absorbent, which selectively absorbs water vapor from the rising gas vapor, as it moves down the column. Well known absorbents used include glycols, among which triethylene glycol (TEG) is most used and which works with gas temperatures as high as 120°C. The other glycols include ethylene glycol (EG), diethylene glycol (DEG), tetraethylene glycol (TREG) and propylene. For the adsorption process, activated charcoal or molecular sieves are used as the adsorbent [4: 44–50]. Table 2 contains the operational parameters of a glycol dehydration unit [5: 142], Table 3 presented the fractionator operating conditions.

Removal of water from gas helps to prevent hydrate formation when temperature drops in transmission and distribution system and prevents corrosion problems because of the condensation of water vapor due to high pressure drops or cold temperatures.

Table 2. Operational parameters of a glycol dehydration unit [5: 234]

Parameters	Values
Pressure in inlet separator and glycol contactor	Feed gas pressure
Feed gas pressure	600–1200 psia
Lean glycol temperature	10°F hotter than the inlet gas temperature
Reboiler temperature	375°F to 390°F
Minimum glycol circulation rate	2 gallons of glycol per one pound of water
The regenerator, typical reboiler temperature	350°F (176.6°C)

Table 3. Typical fractionator operating conditions [4]

Fractionating column	Operating pressure [psig]	Number of trays	Tray efficiency
Deethanizer	375–450	25–35	60–80
Depropanizer	240–270	30–40	80–90
Debutanizer	70–100	25–35	85–95
Butane Splitter	80–100	60–80	90–100

2. Research problem and the method of solution

The data for the natural gas condensate well stream which is used for the simulation is shown in Table 4 which contains some water and hydrocarbons from methane to decane at 100°F, 970 psia with a standard volumetric flow of 290 MMscf. The challenge is to recover the hydrocarbons and meet the product specifications. For the natural gas product (fuel gas), some specification require at least 70% mole of methane with a maximum water content not exceeding 7 lbm/MMscf, and its dewpoint temperature should be sufficiently low. This is to ensure that at any temperature during operations, liquids are not formed because this may cause hydrates or corrosion. The problem also concerns fractionation of natural gas liquids (C2 to C10+) to obtain ethane, propane, and butane.

In the simulation used for obtaining the result, the sequence of operation is in the following order: dehydration of natural gas stream, demethanizer operation and fractionation of natural gas liquids. Bryan ProMax is used to estimate the content of water in the input stream from the well, and that of the output stream from the glycol contactor, in order to determine if dehydration has been fulfilled sufficiently.

From the literature review, the operating parameters for the separator, glycol dehydration units, and

the fractionators were summarized in Tables 1–3, and can be used as a reference for simulation. Then a review is carried out to find the optimal technologies to separate those valuable components – methane, ethane, LPG, gas-condensate. Finally, the methodology for processing this well is suggested as follows: at first, water is removed from the stream by dehydration unit using triethylene glycol; then, methane is recovered as a fuel gas stream, propane and butanes are recovered and finally the remaining heavier hydrocarbons, C5+ (hydrocarbons). To achieve the optimal conditions for these gas processes, a simulation of a gas plant was created using ProMax version 4.0 software with some facilities like: propane refrigeration loop, the Joule–Thompson valve, the turbo expander included.

3. Results

The gas plant has been simulated by 3 main parts: dehydration unit, demethanizer processing unit and fractionation of natural gas liquid which was restricted to only the depolarizer column. The various processing units represented in Brayan Pro-Max are dehydration unit, demethanizer unit and depropanizer units respectively as shown in Figures 1, 4, and 7.

3.1. Operation of the dehydration unit

The dehydration unit is shown in Figure 1, where the input stream contains hydrocarbon components from methane to decane as shown in Table 4, Figure 2 and 3, at 100°F of and 950 psia, with standard vapor volumetric flow of 153.5 MMscfd and standard liquid volumetric flow of 2799.78 sgpm as shown in Figure 2. Analysis of the streams with the glycol contactor indicate water content of 55.37 lbm/MMscf at the input stream (wet gas feed), and 6.96 lbm/MMscf at the output stream (stream 4), with the triethylene glycol flow of 43.07 sgpm (std liquid volumetric flow), as shown in Figure 2. Water dew temperature is 100.02°F at the input stream (wet gas stream) and 35.91°F at the output stream, where water dew point temperature represents the temperature where water begins to form. To meet the specification of water which is below 7 lbm/MMscf, the specifier of the standard liquid volumetric flow of triethylene glycol was adjusted from water-fed multiplied by 3 to water-fed multiplied by 5.9 which resulted in meeting the water content value at the output stream as shown in Figure 2.

The dehydration process is such that triethylene glycol was pumped at a rate of the water-fed multiplied by a constant value, which resulted in triethylene glycol

flow of 43.07 sgpm of glycol in the glycol maker shown in Figure 2. The triethylene glycol produced in the triethylene glycol makeup unit is transported by the glycol pump which increases the temperature of the flowing stream, which is subsequently cooled by the gas/glycol exchanger. Next the stream is passed through a recycle block to the glycol contactor, a recycle block is a unit where the initial guesses are specified – they include: temperature, pressure, composition (100% triethylene glycol) and molar flow [lbmol/hr]. Those parameters are shown in Figure 3. The temperature of triethylene glycol should be below 110 to prevent loss in the glycol contactor. Initially, triethylene glycol is passed into the top of the contactor resulting in downward flow to the bottom of the contactor simultaneously absorbing water from the rising natural gas stream on contact.

Table 4. Gas composition used for the simulation

Component	Mole percent [% mole]	Mole fraction [Yi]
N ₂	23.83599	0.23836
CH ₄	54.32819	0.543282
C ₂ H ₆	5.079988	0.0508
C ₃ H ₈	2.678428	0.026784
i-C4	0.424063	0.004241
n-C4	0.764013	0.00764
i-C5	0.233935	0.002339
n-C5	0.176109	0.001761
n-C6pent	0.118333	0.001183
2-methylpentane	0.174827	0.001748
3-methylpentane	0.963353	0.009634
n-C7	0.462133	0.004621
2-methylhexane	0.250883	0.002509
C8	0.352549	0.003525
2-methylheptane	0.500026	0.005
3-methylheptane	0.688129	0.006881
C9	0.530791	0.005308
2-methyloctane	0.544235	0.005442
3-methyloctane	0.401946	0.004019
C10	0.328732	0.003287
2-methylnonane	0.332425	0.003324
3-methylnonane	0.307783	0.003078
C10+	6.523137	0.065231
H ₂ O	0.07636	0.0007636

Subsequently after flowing out of the glycol contactor, rich triethylene glycol passes through a reflux coil and Joule–Thompson valve, the reflux coil helps to add some heat, and the Joule–Thompson valve reduces the pressure of the stream by 890 psia which also reduces the temperature by 8.78°F, and subsequently passed to the rich flash separator to produce a flash gas, and

a condensate liquid which is a mixture of triethylene glycol and water. The triethylene glycol and water mixture are passed into the glycol regenerator for recovery, where mostly triethylene glycol is produced from the bottom stream and water from the top stream. The stream from the top of the glycol regenerator is passed

into the condenser for more recovery of triethylene glycol. From the condenser, water vapour is recovered and heat is transferred from the condenser to an exchanger used to warm the glycol contactor end products with the QRecycle block. The QRecycle block is used to provide initial estimates of recycled energy.

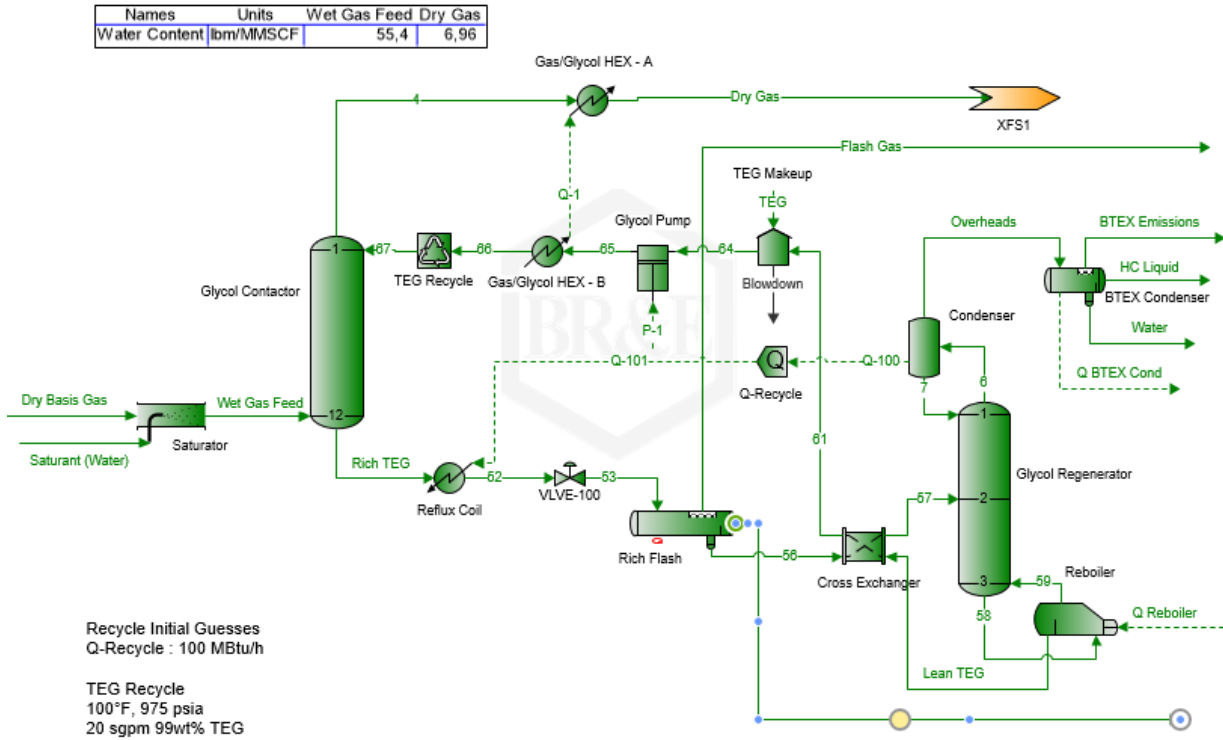


Fig. 1. Dehydration unit for water removal

3.2. Conditions for running dehydration

The conditions for running the glycol dehydration are summarized in Table 5. It shows that the glycol contactor inlet stream with 12 stages and a pressure of 970 psia which drives the pressure of the glycol contactor. Since there is no reboiler for the glycol contactor, the temperature and the pressure are driven by the property of the input stream (wet gas feed), which is the only input to the glycol contactor. The gas/glycol HEX-A and JT valve (vlve-100) are used to cool the stream with a temperature drop of 89.43°F and 8.82°F respectively. The Reflux Coil was used to add heat to the stream with a temperature increase of 12.81°F, while the glycol pump operation result pressure drop of 965.3 psia, indicating high flow and temperature increases of 1.79°F. Across the glycol regenerator there was small pressure drop and heat was added by the reboiler resulting in an increase

in temperature of 86°F to remove the water content, while the condenser aids in the recovery of triethylene glycol.

3.3. Demethanizer processing unit

The conditions of the dehydrated stream (inlet stream) are 955 psia, 113°F and at 153.66 mmscfd (standard vapor volumetric flow) as shown in Figure 5, with 63% mole of methane with other hydrocarbon components. There are two output streams: the first output stream product is fuel gas, predominantly methane, with about 63.5% mole of methane at 293 psia and 83°F with 153.5 mmscfd (standard vapor volumetric flow); the other output product stream is a liquid stream – unlike the first output product, it contains almost no methane of about 5.525e-06% mole and also serves as the input to the fractionation processing unit, the preceding values are shown in Figure 6.

Name: Wet Gas Feed

Properties Composition Analyses Notes

		Total	Vapor	Light Liquid	H ^
Temperature	°F	100	100	100	
Pressure	psia	970	970	970	
Mole Fraction Vapor	%	81,3561	100	0	
Mole Fraction Light Liquid	%	18,6439	0	100	
Mole Fraction Heavy Liquid	%	0	0	0	
Molecular Weight	lb/lbmol	34,539	21,2375	92,5829	
Mass Density	lb/ft^3	6,80696	3,70555	41,9622	
Molar Flow	lbmol/h	20870	16979	3890,97	
Mass Flow	lb/h	720828	360590	360238	
Vapor Volumetric Flow	ft^3/h	105896	97311	8584,82	
Liquid Volumetric Flow	gpm	13202,6	12132,3	1070,32	
Std Vapor Volumetric Flow	MMSCFD	190,076	154,638	35,4375	
Std Liquid Volumetric Flow	sgpm	2799,78	1724,45	1075,34	
Compressibility		0,819471	0,925607	0,356327	
Specific Gravity			0,733275	0,672808	
API Gravity				72,4247	
Enthalpy	Btu/h	-7,5088e+08	-4,05516e+08	-3,45364e+08	
Mass Enthalpy	Btu/lb	-1041,69	-1124,59	-958,71	

Name: TEG Makeup

Execute

Connections Process Data Streams Notes

Streams	Stream	TEG	64
Properties	From Block		
Composition	To Block	TEG Makeup	TEG Makeup Glycol Pump
Temperature	°F	100	199,341
Pressure	psia	25	14,7
Mole Fraction Vapor	%	0	0
Mole Fraction Light Liquid	%	100	100
Mole Fraction Heavy Liquid	%	0	0
Molecular Weight	lb/lbmol	149,079	140,816
Mass Density	lb/ft^3	69,4613	65,7416
Molar Flow	lbmol/h	1,357	172,629
Mass Flow	lb/h	202,301	24308,9
Vapor Volumetric Flow	ft^3/h	2,91243	369,764
Liquid Volumetric Flow	gpm	0,363108	46,1004
Std Vapor Volumetric Flow	MMSCFD	0,0123591	1,57224
Std Liquid Volumetric Flow	sgpm	0,358064	43,0709
Compressibility		0,00893344	0,00445218

Name: Glycol Flow Control_x1

Simple Specifier Notes

Specified Variable

Value: 43.0708553981289 sgpm

Replace: WaterFed * 5.9

Independent Variables

Variable	Value	Units
WaterFed	7.30014498273371	lb/min

Fig. 2. Conditions of the inlet stream to the glycol contactor with conditions at the triethylene glycol and its specifier

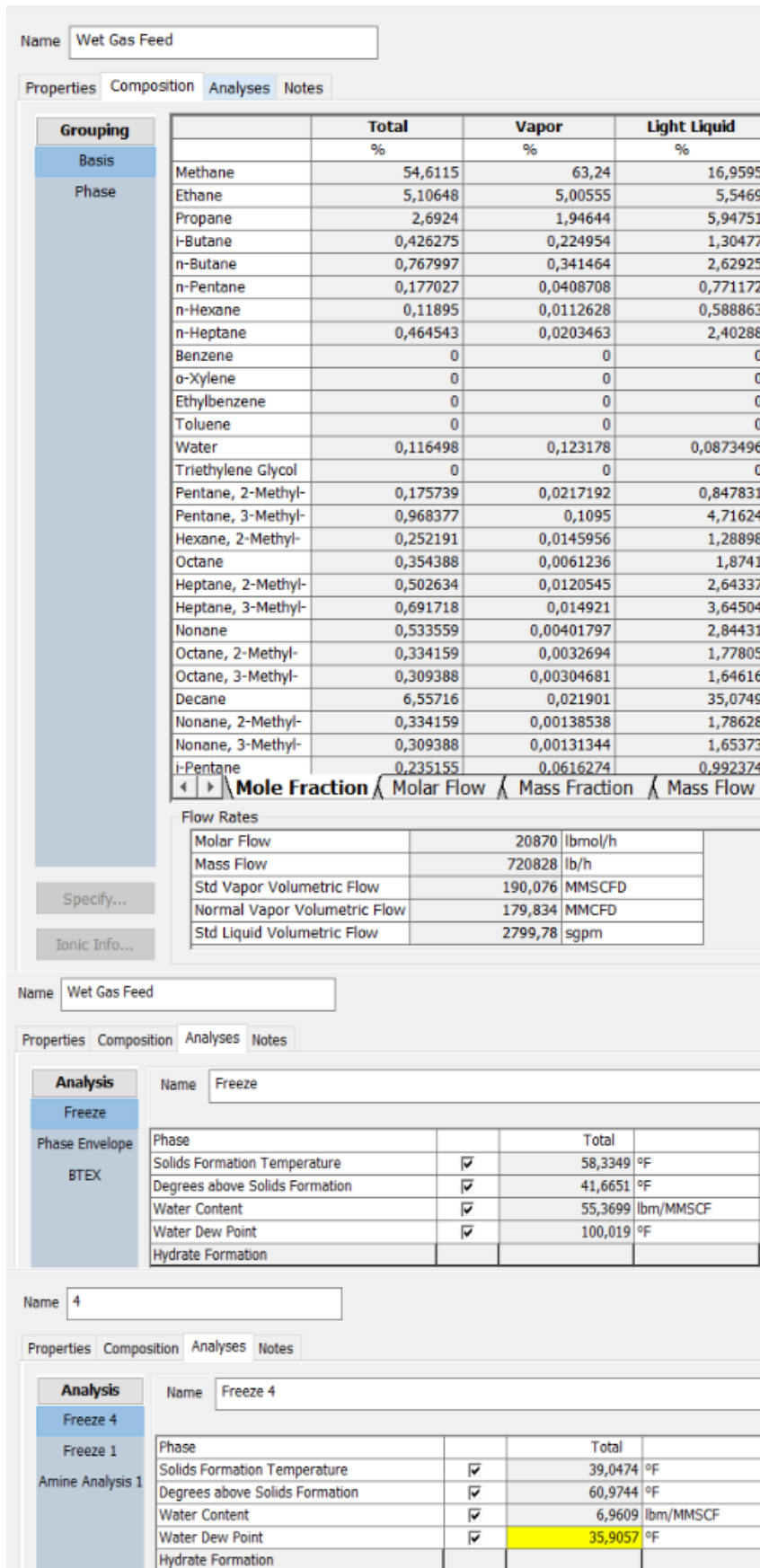


Fig. 3. Input stream to the dehydration unit

Table 5. The conditions for running dehydration part

Units	Inlet pressure [psia]	Pressure drop [psia]	Inlet temperature [°F]	Temperature change [°F]	Other properties
Glycol contactor	970	10	92.7–91.4	6.7	12 stages
Gas/glycol HEX-A	980	5	199.87	89.43	Heat duty: $-1.8e+06$ btu/h
Reflux coil	970	5	79.09	-12.81	Heat duty: $-563,106$ kJ/h
Glycol pump	14.7	965.3	199.34°F	-1.79	Power: 52.33 hp Overall efficiency: 68%
JT valve: VLVE-100	965	890	103.23	8.78	-
Rich flash	75	0	94.41	0	-
Glycol regenerator	70	0.5	98.86	192.69	Reflux ratio target: 1 Reflux ratio value: 0.1 Boilup ratio target: 0.2289 Boilup ratio value: 0.1969 Reboiler temperature target: 400°F Reboiler temperature value: 400°F
Condenser for glycol	14.7	0	218.90	11.58	Heat duty: -64869.8 btu/h
Glycol regenerator reboiler	15.2	0	314.50	-86	Heat duty: $286e+06$ btu/h
Cross exchanger	-	-	-	-	Pressure drops across lean TEG to blowdown: 0.5 psia. And temperature drops by 201.07°F Pressure drop across rich flash and stream to inlet to glycol regenerator: 5 psia. And temperature increased by -205.88°F

JT Plant with Stabilizer

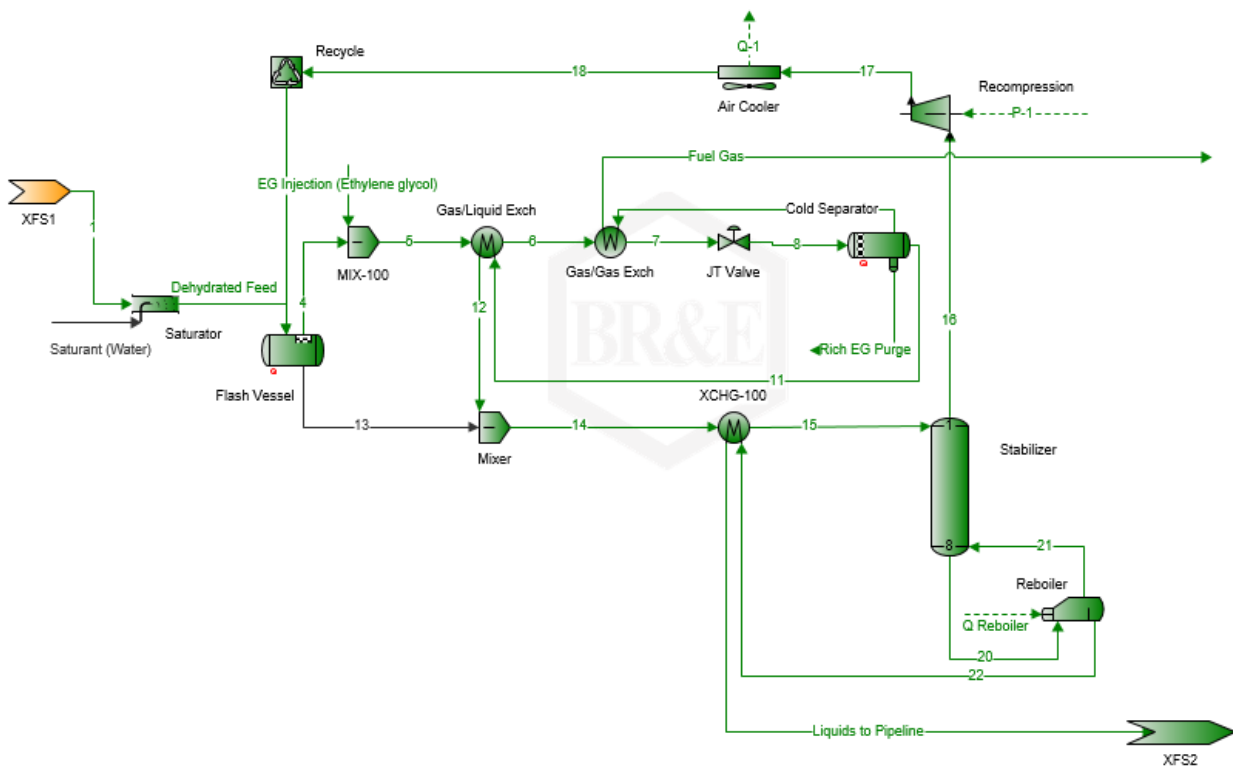


Fig. 4. Methane and nitrogen removal units to separate methane and nitrogen as sales gas

Name: Dehydrated Feed		Total	Vapor	Light Liquid	H ^
Temperature	°F	113,372	113,372		
Pressure	psia	955	955		
Mole Fraction Vapor	%	100	100		
Mole Fraction Light Liquid	%	0	0		
Mole Fraction Heavy Liquid	%	0	0		
Molecular Weight	lb/lbmol	21,1727	21,1727		
Mass Density	lb/ft ³	3,54926	3,54926		
Molar Flow	lbmol/h	16871,2	16871,2		
Mass Flow	lb/h	357208	357208		
Vapor Volumetric Flow	ft ³ /h	100643	100643		
Liquid Volumetric Flow	gpm	12547,7	12547,7		
Std Vapor Volumetric Flow	MMSCFD	153,656	153,656		
Std Liquid Volumetric Flow	sgpm	1710,19	1710,19		
Compressibility		0,926383	0,926383		
Specific Gravity		0,731037	0,731037		
API Gravity					
Enthalpy	Btu/h	-3,98447e+08	-3,98447e+08		
Mass Enthalpy	Rtu/lb	-1115,45	-1115,45		

Name: Fuel Gas		Total	Vapor	Light Liquid	H ^
Temperature	°F	83,3503	83,3503		
Pressure	psia	293	293		
Mole Fraction Vapor	%	100	100		
Mole Fraction Light Liquid	%	0	0		
Mole Fraction Heavy Liquid	%	0	0		
Molecular Weight	lb/lbmol	21,0762	21,0762		
Mass Density	lb/ft ³	1,09655	1,09655		
Molar Flow	lbmol/h	16850	16850		
Mass Flow	lb/h	355133	355133		
Vapor Volumetric Flow	ft ³ /h	323864	323864		
Liquid Volumetric Flow	gpm	40377,8	40377,8		
Std Vapor Volumetric Flow	MMSCFD	153,463	153,463		
Std Liquid Volumetric Flow	sgpm	1704,15	1704,15		
Compressibility		0,966387	0,966387		
Specific Gravity		0,727706	0,727706		
API Gravity					
Enthalpy	Btu/h	-3,96326e+08	-3,96326e+08		
Mass Enthalpy	Rtu/lb	-1115,99	-1115,99		

Name: Liquids to Pipeline		Total	Vapor	Light Liquid	H ^
Temperature	°F	376,353		376,353	
Pressure	psia	285		285	
Mole Fraction Vapor	%	0		0	
Mole Fraction Light Liquid	%	100		100	
Mole Fraction Heavy Liquid	%	0		0	
Molecular Weight	lb/lbmol	100,379		100,379	
Mass Density	lb/ft ³	31,5139		31,5139	
Molar Flow	lbmol/h	20,5558		20,5558	
Mass Flow	lb/h	2063,38		2063,38	
Vapor Volumetric Flow	ft ³ /h	65,4752		65,4752	
Liquid Volumetric Flow	gpm	8,16314		8,16314	
Std Vapor Volumetric Flow	MMSCFD	0,187214		0,187214	
Std Liquid Volumetric Flow	sgpm	6,01288		6,01288	
Compressibility		0,101183		0,101183	
Specific Gravity		0,505283		0,505283	
API Gravity		72,7701		72,7701	
Enthalpy	Btu/h	-1,61816e+06		-1,61816e+06	
Mass Enthalpy	Rtu/lb	-784,277		-784,277	

Fig. 5. Conditions at the inlet stream to the flash vessel (dehydrated stream), fuel gas and natural gas liquid gas

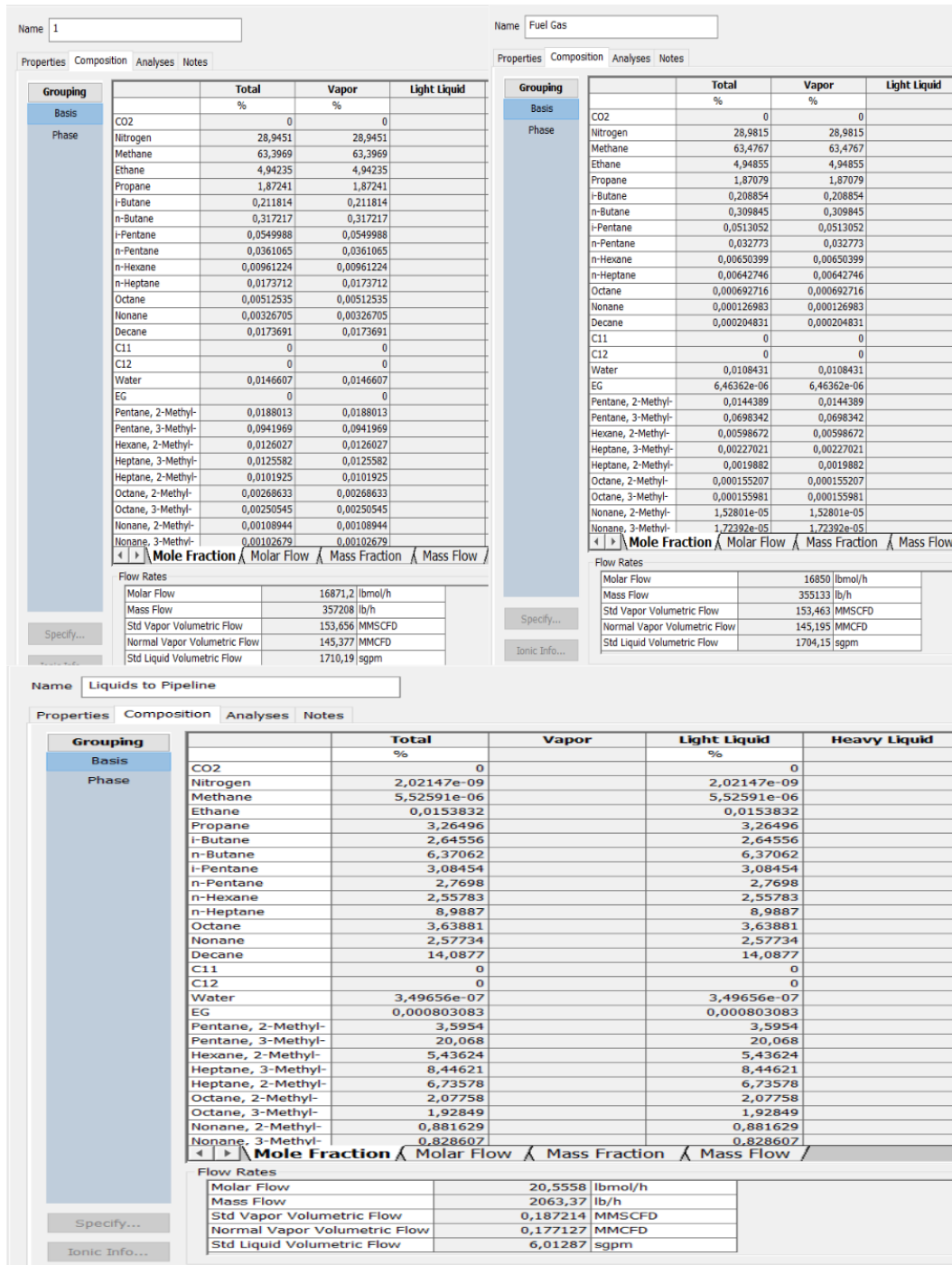


Fig. 6. Data of two outlet streams from the demethanizer and output stream for fuel gas and natural gas liquid

3.4. Conditions for running demethanizer operations

The conditions for running the demethanizer operation is summarized in Table 6, where the flow from the input stream which have been dehydrated is saturated with water before being sent with compressed vapor from the recycle stream into the flash vessel, where a pressure drop

of 20 psia occurs across the flash vessel. From the flash vessels the light components are mixed with ethylene glycol and cooled by the gas/liquid and gas/gas exchangers, and subsequently by the Joule-Thompson valve before it is sent to the cold separator. The gas/gas exchanger contributes the greatest drop in temperature, cooling by 59.6°F, followed by the Joule-Thompson valve with 34.9°F and gas/liquid exchanger with 0.5°F.

Table 6. The conditions for running demethanizer operations

Units	Inlet temperature [°F]	Temperature change [°F]	Inlet pressure [psia]	Pressure drop [psia]	Other properties
Flash vessel	Dehydrated feed: 110.27 Recycle feed: 120	temperature change = 2.5°F	Dehydrated feed: 955 Recycle feed: 900	22	–
MIX-100	107.9	0	898	0	–
Gas/liquid exchanger	107.8	0.6 (cooling)	898	5	Heat duty: 60,592.99 btu/h
Gas/gas exchanger	Stream 6: 107.2 Stream 9: 14.94	Stream 6: 59.6°F (cooling) Stream 9: –68.4 (heating)	Stream 6: 893 Stream 9: 298	Stream 6: 5 Stream 9: 5	Heat duty: 1.016e+07 btu/h
JT valve	50	34.9°F (cooling)	888	588	Joule–Thomson coefficient: 0.06095°F/psi
Mixer	Stream 12: 60	0	Stream 12: 293	0	–
XCHG-100 liquid exchanger	Stream 56: 94 Lean TEG: 400	Stream 56: –206 (heating) Lean TEG: 201.07 (cooling)	Stream 56: 75 Lean TEG: 15.2	Stream 56: 5 Lean TEG: 0.5	–85121.997 Btu/h
Cold separator	14.2	0.15	300	2	–
Air cooler	289.7	169.7	905	5	–
Recompression	158.38	–129.25 (heating)	288	–617 (increased pressure)	Performance curve type: isentropic (adiabatic) efficiency: 85%
Stabilizer	120	Inlet: –38.38 262.84° across all stages	290	2	Number of stages: 8
Stabilizer reboiler	357.21	–62.14 (heating)	290	0	–

Cold separator with a pressure drop of 2 psi and temperature drop on 0.15°F, aids the separation of gas from the natural gas liquids. Natural gas from the cold separator cools the incoming stream to the gas/gas exchanger, simultaneously warming itself, before being sent out as the natural gas product containing mainly methane. Natural gas liquids from the cold separator are used as coolant for the gas/liquid exchanger, before moving to the mixer, then to the XCHG-100 (vapor/liquid exchanger), where heating by the liquid stream from the stabilizer reboiler takes place. This causes a temperature rise of 60°F in the stream to XCHG-100 exchanger. The stabilizer unit recovers more light components through its stages in addition, through heating the bottom components in its reboiler. Across the stabilizer, from its input to the top, where the lighter components pass out, temperature increases by 38.38°F while temperature change along all its stages of the stabilizer is about 262.84°F with a pressure drop of 2 psi and the reboiler provides heat with temperature increase of 63.39°F.

The product of the reboiler includes natural gas liquids and vapor. The natural liquids are sent to cool the input of the stabilizer in the XCHG-100 (vapor/liquid exchanger) before it is sent as an input to the fractiona-

tion unit to obtain various natural gas liquid products. The vapor stream produced from the stabilizer is sent for recompression which increases pressure by 617 psia resulting in a temperature rise of 129°F. Next, the pressurized stream is sent to the air cooler which reduces the temperature by 169°F, which appears to compensate for the drastic increase in temperature due to recompression. The cooled vapor from the air cooler becomes the recycle value which is sent to the flash vessel.

3.5. Depropanizer processing unit

The liquid stream from the demethanizer unit serves as the input to the fractionation unit (Fig. 7). The simulation was not feasible for the demethanizer unit. The input stream was removed from the demethanizer column and connected to the depropanizer column which resulted in a feasible simulation and with a propane product containing 14.62% mole i-butane and 30.40% mole n-butane while product of the bottom stream contained 0.37% mole i-butane and 1.8% mole n-butane as shown in Figure 8. The simulation was infeasible to separate i-butane from n-butane and other heavier hydrocarbons accounted for the rest of the mole percent.

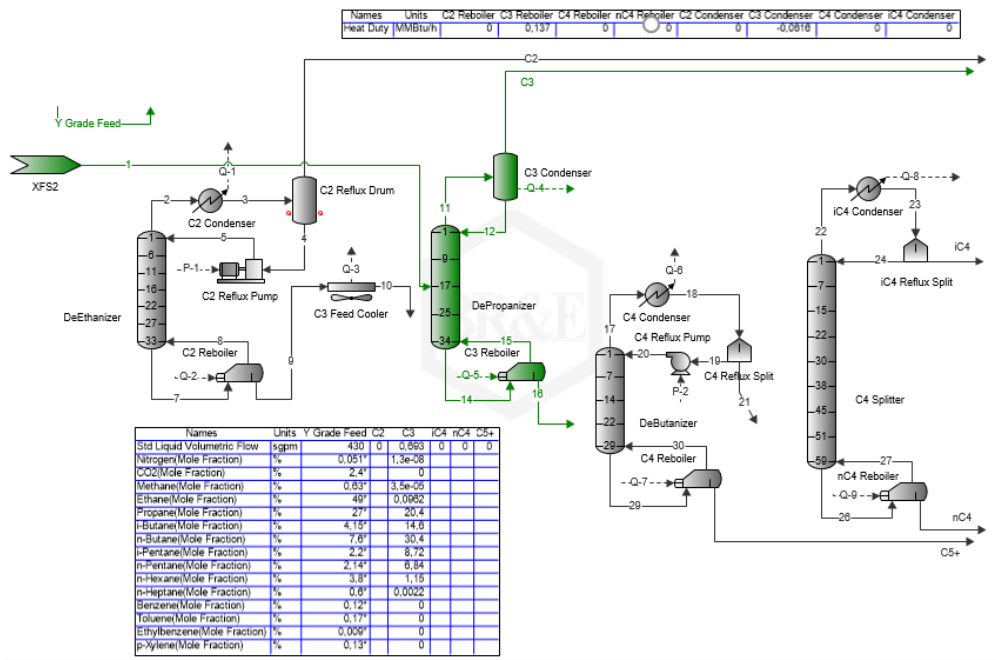


Fig. 7. Fractionation to separate C5+, i-C4, n-C4, C3, C2 components

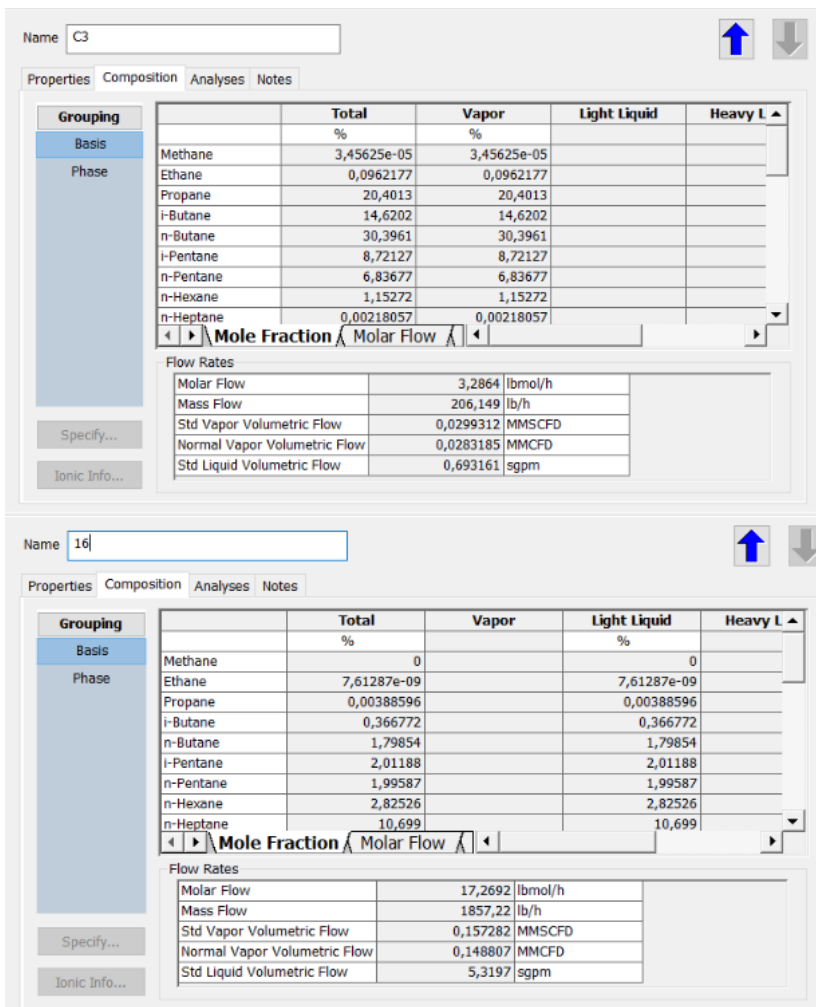


Fig. 8. Propane products from the depropanizer in stream C3 and the heavier hydrocarbon products in stream 16

3.6. Conditions for running fractionation

The natural gas liquid stream from the demethanizer was first sent to the deethanizer column, but results in an infeasible solution, and was repeated for other columns. The only feasible solution was passing natural liquid to the depropanizer column which was isolated from the debutanizer column where input stream conditions

were 376.54°F, 285 psia and 0.1872 MMscfd (standard vapor volumetric flow), and 6.31 MMscfd (standard liquid volumetric flow) as shown in Figure 8. The operation is summarized in Table 7 with 34 stages, and as shown in Figure 9, the specification target includes 99.9% recovery rate for propane while the actual value in the simulation is 99.9001% and the bottom product specification target of isobutane was 98% while the actual value was 11.64%. The reflux ratio target value is 2, as well as the value used.

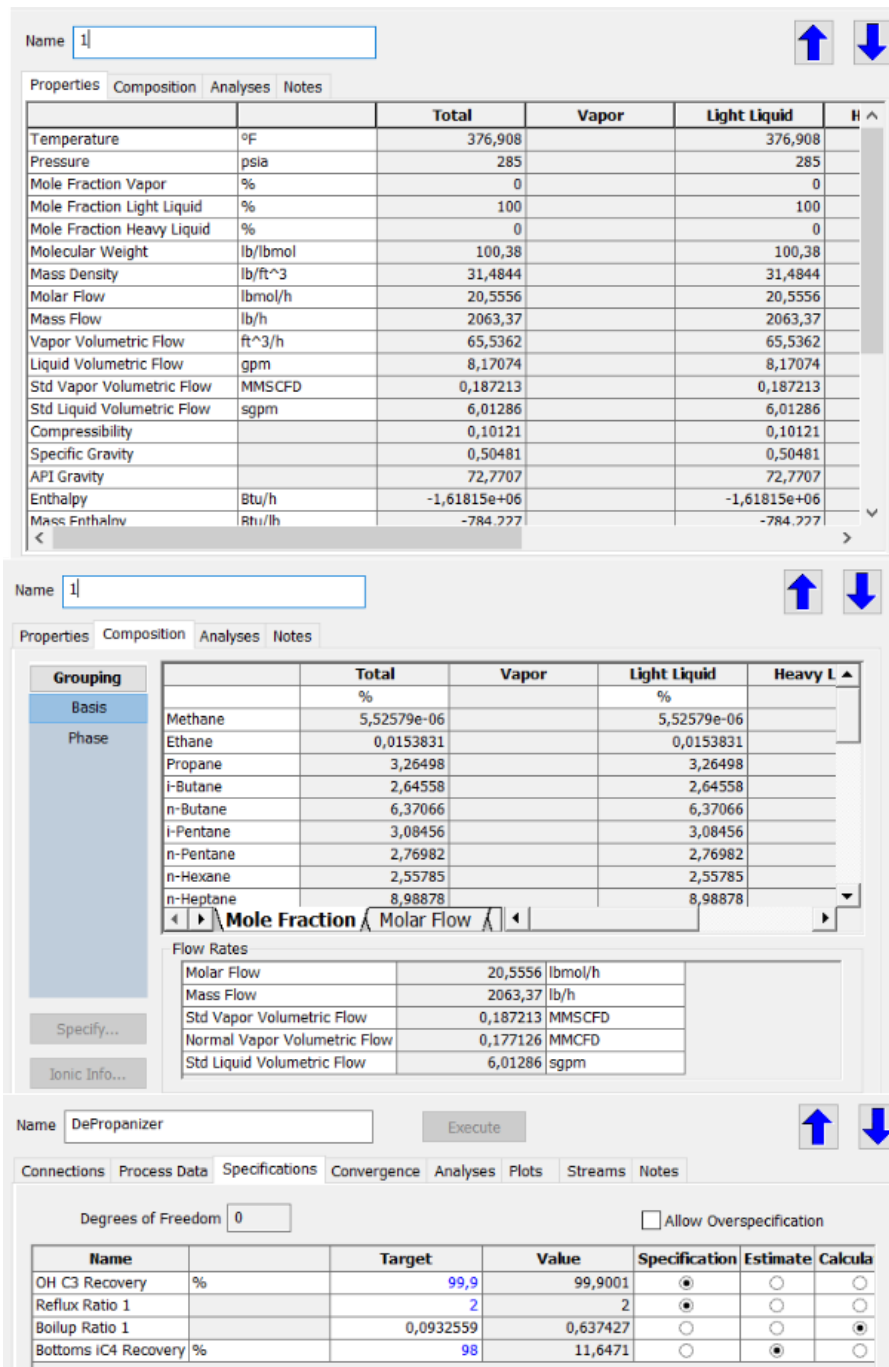


Fig. 9. Conditions of the inlet stream to the depropanizer with its composition and the specification of the depropanizer

Table 7. The conditions for running fractionation

Units	Inlet temperature [°F]	Temperature change [°F]	Inlet pressure [psia]	Pressure [psia]	Other properties
Depropanizer	376.54	Temperature drop across all stages: 178.6	285	Pressure drop across all stages: 2	Number of stages: 34 OH C3 recovery: 99.9% Reflux ratio target and value: 2 Boilup ratio target: 0.0932559 Boilup ratio value: 0.637047 Bottoms i-C4 i-C4 recovery: 98% i-C4 recovery value: 11.67
C3 condenser	280.43	31.01	205	0	Heat duty: -62970.85 BTU/h Mole fraction vapor: 33.33% Mole fraction light liquid: 66.67%
C3 reboiler	405.29	22.65	207	0	Heat duty: 139586.31 BTU/h Mole fraction vapor: 38.91% Mole fraction light liquid 61.09%

4. Discussion

The glycol contactor operating conditions are driven by the input stream which is at 100°F and 970 psia with a pressure drop of 10 psi across the contactor and outlet temperature at the top of 100.02°F. The outcome of the simulation showed that water content was 6.96 lbm/MMscf which is less than the maximum required of 7 lbm/MMscf. The triethylene glycol from the contactor contains water and other hydrocarbon which are separated to obtain a more concentrated glycol for reuse. First, it is warmed by a reflux coil by 12.81°F and is greatly depressurized and cooled by the JT Valve (VLVE-100), by a pressure drop of 890 psia and a 8.78°F temperature drop. It then passes into the rich flash, which is a three-phase separator which separates the gas from the gas liquids from the water to produce three streams of natural gases, natural gas liquid and a mixture of triethylene glycol and absorbed water. From the rich flash, the triethylene glycol and water mixture passes through a cross exchanger where it was warmed by the heated liquid stream from the glycol reboiler and simultaneously cools the warming stream, resulting in 205.88°F rise in temperature of the triethylene and water stream, which moves into the glycol contactor where water is driven through the top outlet while mostly glycol falls to the bottom, where it passes into the reboiler for more heating. From the reboiler, mainly water and hydrocarbons are sent back to the glycol regenerator for recovery. Also, more concentrated triethylene glycol flows to the triethylene makeup from the reboiler, but first loses heat at the cross exchanger where it is cooled by 201.07°F as shown in Table 5, which also shows the glycol pump operates with a pressure drop of 965.3 psia due to flow and a 1.79°F rise in temperature with an efficiency of 68% and 38.17 hp.

The temperature, pressure and flow of the input stream conditions are 113.38°F and 955 psia and at 153.66 MMscfd (standard vapor volumetric flow), while there are two output streams. The first output product stream is fuel gas, predominantly methane, with about 63.5% mole of methane at 293 psia and 83°F with 153.46 MMscfd (standard vapor volumetric flow) while the other output product stream is a liquid stream. Unlike the first output product, it contains almost no methane of about 5.525e-06% mole and also serves as the input to the fractionation processing unit.

Table 6 shows that cooling occurred at the gas/liquid exchanger, gas/gas Exchanger and JT Valve by a temperature drop of 0.6°F, 59.6°F and 34.9°F respectively with pressure drop of 5 psia, 5 psia and 588 psia respectively, while cooling and warming simultaneously due to energy exchanges between the two streams, where heating occurs from stream 56 and cooling occurs from Lean TEG stream. Separations of gas from natural gas liquids occurred in the cold separator and stabilizer, although separation is also possible in the flash vessel, which did not occur. The cold separator and stabilizer temperature and pressure drop are 0.2°F, 40°F and 2 psia, 2 psia, respectively. The reboiler supplies heat to the stabilizer with a temperature increase of 62.14°F which produces the final natural gas liquid stream, while the vapor from the stabilizer is recompressed by 288 psia and cooled by the air cooler by 169.7°F. Natural gas consisting mainly of methane was produced from the cold separator and warmed at the gas/gas exchanger by 68.4, after which is sent as fuel gas.

The natural gas liquid stream from the demethanizer was first sent to the deethanizer column. The feasible solution is when natural liquid from the demethanizer served as input to the depropanizer column which was isolated from the debutanizer column. The operation

of the depropanizer column is summarized in Table 7 with 34 stages, where input stream conditions are 376.91°F, 285 psia and 0.187 MMscfd (standard vapor flow). Across the depropanizer stages, temperature and pressure drops are 178.58°F and 2 psi respectively. The specification target includes a 99.9% recovery rate for propane which was fulfilled as indicated by the actual value of 99.9001%, while for the bottom product, the target for isobutane was 98% – it was not fulfilled as indicated by the actual recovery of 11.65%.

5. Conclusions

The project has been done successfully, with the water content reduced from 55.37 to 6.62 lbm/MMscf at the

end of the dehydration processes, where the glycol liquid flow rate is 59.10 sgpm and temperature is 110.10°F and the conditions of the feed gas glycol contactor was 100°F, 970 psia with vapor flow of 190.07 MMscfd as shown in Figure 7. Recommendations for operations were fulfilled, lean glycol from the glycol pump followed the recommendation that its temperature should be 10°F greater than the input stream of the glycol contactors, and the glycol flow rate also exceeds the minimum recommended glycol flow of 2 sgpm.

In addition, from the depropanizer fractionation process, recovery are as follows C3: 99.9%; i-C4: 11.67%. The recovery of natural gas liquids depends on the composition of the natural gas, and the application of cooling by the exchangers, recycling of stream and conditions in the flash vessels, cold separators stabilizers and its reboilers and condensers.

References

- [1] Leffler W.: *Natural gas liquids: A nontechnical guide*. PennWell Corporation, Oklahoma 2014.
- [2] Devold H.: *Oil and gas production handbook: An introduction to oil and gas production transport, refining and petrochemical industry*. Edition 3.0, ABB Oil and Gas, Oslo, August 2013.
- [3] Stewart M., Arnold K.: *Gas-liquids and liquid-liquid separators*. Elsevier, United Kingdom, 2008.
- [4] Abdel-Aal H.K., Aggour M.A., Fahim M.A.: *Petroleum and Gas Field Processing: Recovery, Separation and Fractionation of Natural Gas Liquids*. Marcel Dekker, New York 2003.
- [5] Kidnay A.J., Parrish W.R.: *Fundamentals of natural gas processing*. Taylor and Francis Group, London 2006.

