Maximization of Condensate Production by Revamping of Gas-Oil Separation Plant in Gulf of Suez

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The present work is a retrofit of the gas processing train of a gas-oil separation plant (GOSP) located in Gulf of Suez in Egypt to produce the NGL. Three modification schemes based on self-refrigeration and external refrigeration of raw natural gas are proposed. Applying either of the mentioned methods causes the heavy components in the raw gas (C_{4+}) to condense and then get separated. Steady-state simulation of the GOSP is conducted using HYSYS software based on the actual plant data. The first and second revamps are based on "self-refrigeration" while the third one relies on "external refrigeration". The expected outcome of the retrofit schemes is to increase the condensate recovery of GOSP by approximately 10 %. Also adjust the Reid vapor pressure (RVP) of the GOSP condensate product according to export specifications i.e. maximum RVP = 82.74 kPa. The three schemes succeeded to maximize the condensate production and yielded about 9.7 % increase in GOSP condensate product. The RVP issue is managed by addition of re-boiled absorber tower in all schemes (RVP = 65.5 kPa).

1. Introduction

Raw natural gas composition is typically a high mole percentage of methane (CH_4), ethane, propane & butane (Liquefied petroleum gases) and heavier hydrocarbons (C_{5+}) also known as condensates and all hydrocarbon components other than methane are known as natural gas liquids (NGL, C_{2+}). In addition to hydrocarbon components there is gas impurities as carbon dioxide, nitrogen, Hydrogen sulphide and water vapour. The raw natural gas produced from wells is routed to a gas processing facility to remove the aforementioned impurities and be dehydrated from water vapors in order to meet export pipeline specifications. It is well known that NGL as a separate product yields higher revenues than pure natural gas product (Mokhatab et al., 2006). Hence, purified natural gas undergoes further NGL recovery process. In industry, the richness of natural gas mixture is determined as cubic meter of liquid recoverable per 1,000 standard cubic meter (m^3/km^3) and it requires the knowledge of the gas composition and the cubic meter of liquid per gmole of each component (Kidnay et al., 2006). Various NGL recovery applications are available in industry and selection of certain process depends on the natural gas liquid content, gas inlet pressure and economics (Kidnay et al., 2006). NGL recovery is achieved either by applying refrigeration, lean oil absorption, adsorption or membranes. Refrigeration is accomplished by self-refrigeration or external refrigeration. Self-refrigeration is managed by pressure drop of the gas stream across a Joule-Thompson valve (usually from 3,447 to 6,895 kPa) causing the stream temperature decrease at constant enthalpy (Mokhatab et al., 2006). The condensed liquid is then separated in one or more separators to meet vapour pressure and composition specifications. While external refrigeration technique depends on external refrigeration cycle such as propane cycle to condense the NGL and then separated in low temperature separator (Ghorbani et al., 2016). The both techniques may be preceded by glycol contactor or hydrate inhibitor injection to prevent hydrate formation at the low pressures. Heat exchange of inlet warm gas with cold gas stream from low temperature separator is preferable (minimum temperature approach 5 °C) before refrigeration to minimize energy consumption.

2. AMAPETCO Gas-Oil Separation Plant (GOSP) Process Description

As shown in Figure 1, the feed fluids to GOSP facilities are directed to the slug catcher D-2050 were the separation between vapors and liquids takes place. Combined gases from D-2050 & D-2010 are divided into two parts, the first part flow to the high pressure scrubber D-2280 where the separated gases are directed to the glycol contactor D-2230 to adjust the water content/water dew point specifications of the export gases.
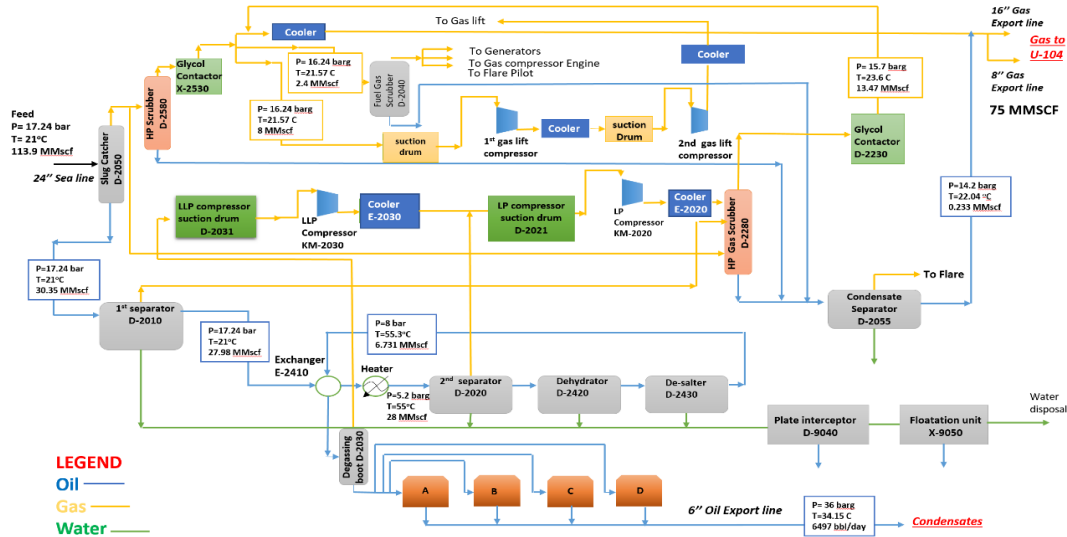

Figure 1: Process-flow diagram of the GOSP before modification (Elsheemy, 2017)

The estimated flow of this part is ranging between 498.1 & 647.5 kgmole/h. The second part of gases with an estimated flow rate of 3,686 kgmole/h are directed to the high pressure scrubber D-2580 followed by the glycol contactor D-2531 where the water content of the gases reaches a value of 150 ppm under the action of back pressure controller. The dehydrated gases from the existing TEG units are combined together and flow through the metering skid X-2535 where the flow of the produced gases is measured. The combined gases are split again based on the hydraulics of the pipeline and the operating conditions, the small part (498.1 to 647.5 kgmole/h) is sent to U-104 through the existing 0.2032 m export pipeline, while the remaining part (3,686 kgmole/h) is sent to U-104 through the 0.4064 m export pipeline. Two slip streams are taken up stream the back pressure controller of the two existing glycol contactors (to increase flexibility. of operation), one of the two slip streams will supply the fuel gas to the fuel gas scrubber D-2040, and a new line up stream D-2040 will be installed to supply the gas turbine generators with their requirement of fuel gas, the gas turbine generators shall be equipped with a small compressor to increase the fuel gas pressure to the required level. The estimated consumption of fuel gas is ranging between 74.7 & 99.6 kgmole/h (ENNPI, 2010).

3. Process simulation

The GOSP is simulated on HYSYS software using GOSP actual data (PFD, streams data, production data, and equipment design data), see Table 1 and Table 2 (ENNPI, 2010). The Ping–Robinson equation of state is suggested for this purpose. The simulation model of the GOSP is validated by comparing between actual plant data and the software model data as recommended by (Lynch et al., 2007), see Table 3. After the matching between the real and simulated data, four revamping schemes are applied on the validated simulation model. The refrigeration schemes are applied after the glycol contactor unit (X-2530) excluding the glycol separator unit as there is no need for it in the presence of the glycol contactor (X-2530). The following conservative design parameters are considered upon simulation which is within the design range recommended by literature and best practice recommendation by the process engineering designers: For compression stage, the gas pressure is raised to 7601 kPa which corresponds to 149.3 °C (T < 150 ºC) and does not exceed 4.4:1 compression ratio (Kidnay et al., 2006). Air coolers is set to cool down the gas to 55 °C which is the maximum temperature achieved by the existing GOSP air coolers and pressure drop to 151.3 kPa. For heat exchangers, the pressure drop across the shell-side is set 130.7 kPa, 151.3 kPa across the tube-side and the minimum temperature approach is 8 ºC (Gpsa, 2004). For the Joule-Thompson throttling valves, the pressure drop is set to 5,821 kPa which corresponds to -30 ºC (Abdel-Aal, 2003). The pressure drop across the stabilizer is set to 131.3 kPa and "specs"

set to vapor pressure equals 65.5 kPa which is the max RVP at which no gases formed at storage tanks conditions.

Table 1: Feed stream to GOSP in mole fraction (ENNPI, 2010)

Component	Mole fraction	Component	Mole fraction	Component	Mole fraction	Component	Mole fraction
H_2O	0.197	n-Hexane	0.0115	n-Decane	0.0027	n-C_{20}	0.0002
Nitrogen	0.0034	Benzene	0.0003	n-C_{11}	0.0012	n-C_{21}	0.0001
Carbon dioxide	0.0016	n-Heptane	0.008	n-C_{12}	0.0006	n-C_{22}	0.0001
Methane	0.6047	Toluene	0.001	n-C_{13}	0.0005	n-C_{23}	0.0001
Ethane	0.0731	n-Octane	0.0078	n-C_{14}	0.0004	n-C_{24}	0.0001
Propane	0.041	E-Benzene	0.0003	n-C_{15}	0.0003	n-C_{25}	0.0001
i-Butane	0.0074	O-Xylene	0.0005	n-C_{16}	0.0003	n-C_{26}	0.0001
n-Butane	0.0162	m-C_8H_{10}	0.0007	n-C_{17}	0.0004	n-C_{27}	0.0000
i-Pentane	0.0073	p-C_8H_{10}	0.0003	n-C_{18}	0.0003	n-C_{28}	0.0000
n-Pentane	0.0072	n-Nonane	0.0029	n-C_{19}	0.0002	n-C_{29}	0.0000
						C_{30+}	0.0001
						Total	1.0000

Table 2: Feed Stream Properties and conditions

Properties	
Vapor/Phase Fraction	0.7336
Temperature, ^oC	21
Pressure, kPa	1,825
Molar Flow, kgmole/h	5,673
Mass Flow, kg/h	140,057.95
Standard Liquid Volume Flow, m^3/h	133,290.82
Molecular Weight	24.68
Average Mass Density, kg/m^3	26.55

Table 3: Actual gas product vs. software gas product

Properties	GOSP terminal gas	Software terminal gas
Temperature, ^oC	18.83	19
Pressure, kPa	1,468	1,468
Flowrate, kgmol/hr	3,191	3,746
Molecular weight	21.01	20.98
Minimum LHV, $Kcal/Sm^3$	9,975	10,253
Viscosity [cP]	18.83	19

3.1 Low Temperature Separation Using Self-Refrigeration LTSUSR-1 – first revamp

As shown in Figure 2, the gas feed (streams 24 and 34A) from the glycol contactors X-2530 and D-2230 are divided into two streams; stream 25 goes to fuel gas scrubber (D-2040) with 1,725 kPa, 19.7 ^oC and 119.5 kgmole/h while stream 35 is directed to LTSUSR-1 process with rate 4,059 kgmole/h where its pressure is raised to 7,601 kPa in one stage compression (K-100) with a compression ratio 4.5. The compressed gas (stream 45) is cooled from 147.3 ^oC to 55 ^oC by air cooler (E-102). The outlet stream 49 from Air Cooler (E-102) is divided into two streams; stream 27 goes to gas lift train with 7,551 kPa, 55 ^oC and rate 398.5 kgmole/h. This leads to get rid of the gas lift compressor (KM-2080A) and use one compression stage only in the Gas lift train to approach 10,301.33 kPa as in the GOSP base case. While stream 50 is directed to the revamp one process with 7,551 kPa, 55 ^oC and rate 3,661 kgmole/h. The gas is cooled using gas-gas heat exchanger E-103, condensate-gas Heat Exchanger (E-104), and finally J-T throttling (VLV-110) to -28.07 ^oC by applying pressure drop 7,621 kPa (outlet pressure 1,731 kPa). The condensate is separated from the gas by entering low-temperature (three phase) separator. The outlet condensate (stream 76) is used to cool the gas entering the heat exchanger E-104 which is further expanded using J-T throttling to 133.3 kPa. The condensate is separated from the gas by low-temperature separator. The pressure of this stream is about 163.3 kPa exiting the stabilizer's reboiler and its pressure drops about 151.3 kPa in the air cooler (E-108), so when the condensate (stream 104) reaches the storage tanks it is nearly atmospheric about 113.3 kPa.
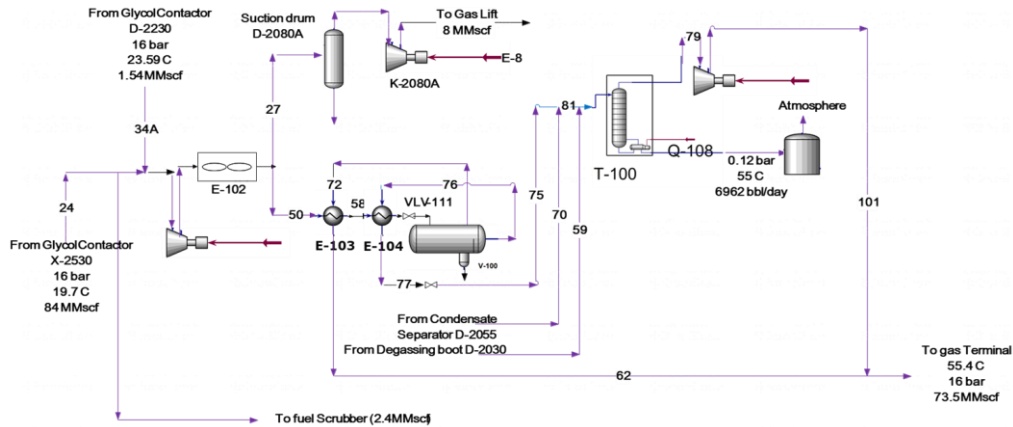
Figure 2: Process flow diagram of LTSUSR-1 (Elsheemy, 2017)

3.2 Low Temperature Separation Using Self-Refrigeration LTSUSR-2 – second revamp

As shown in Figure 3, two stage compression (K-100 and K-101) instead of one stage as in LTSUSR-1 to raise the gas pressure to 10,301.33 kPa. In this case, there is no need for the gas lift compression stages and stream 33 is branched with rate 398.5 kgmole/h from the main compressed gas stream and is sent directly for gas lift with 10,301.33 kPa as in the GOSP base case. The rest of compressed gas (stream 50) is cooled using gas-gas heat exchanger E-103, condensate-gas heat exchanger E-104, and finally J-T throttling (VLV-110) to - 42.47 ᵒC by applying pressure drop 8,321 kPa (outlet pressure is 1,731 kPa). The condensate is separated from the gas by entering low-temperature separator (V-100). The outlet condensate (stream 67) is used to cool the gas (stream 58) entering the heat exchanger E-104 and then further expanded using J-T throttling valve (VLV-106) to 133.3 kPa. The cooled condensate stream 77 is then combined with the condensate stream 59 coming from the Degassing boot (D-2030) and the condensate separator (D-2055) (stream 76) by MIX-108 and enters (stream 81) the near atmospheric stabilizer (T-100) at 133.3 kPa and - 4.8 ᵒC to achieve the required RVP (65.5 kPa) which is the maximum RVP at which no gases are formed at the storage tanks conditions. The condensate from the stabilizer (stream 78) exits at 163.3 kPa and 69.7 ᵒC is cooled to 55 ᵒC using air cooler (E-108) and directed (stream 22) to the fixed roof storage tanks at conditions 113.3 kPa and 55 ᵒC. While the gas (stream 62) from the Low-temperature separator and stabilizer (stream 79) are compressed to 1,701 kPa and directed to the sales gas pipeline.
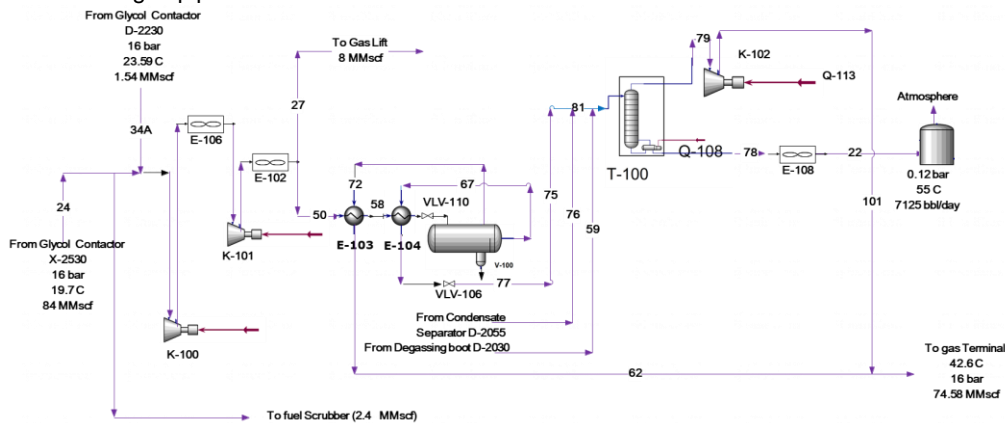

Figure 3: Process flow diagram of LTSUSR-2 (Elsheemy, 2017)

3.3 Low temperature separation using external refrigeration LTSUER – third revamp

As shown in Figure 4, no compression stages are conducted on the gas coming from the glycol contactors (X-2530 and D-2230) and an external refrigeration "Propane cycle" is applied instead of cooling using J-T expansion. The gas (stream 50) is cooled from 19.85 ᵒC to -15.3 ᵒC using gas-gas heat exchanger (E-103), condensate-gas heat exchanger E-104 and finally, refrigerated to -36.9 ᵒC in chiller (E-109) using propane cycle with approach 5 ᵒC. Condensed NGL recovered in separator at 1,601 kPa and -36.9 ᵒC, heated in E-104 and

combined with the condensate (stream 59) from degassing boot (D-2030) and from the condensate separator (D-2055) at pressure 163.3 kPa. The combined condensate is heated in heat exchanger (E-106) against the hotter condensate stream from the stabilizer's re-boiler (reboiler@T-100) from 3.2 °C to 18.9 °C. The heated condensate stream then enters the top stage inlet of stabilizer (T-100) where the condensate RVP is adjusted to 75.84 kPa (which is the maximum RVP at which no gases are formed at the storage tanks conditions). The outlet condensate stream (stream 78) from re-boiler is cooled to 11.2 °C in heat exchanger (E-106) and then routed to the storage tanks. The separated gases from low-temperature separator (V-100) and stabilizer (T-100) are compressed and sent to the gas export pipeline (stream 41). Table 4 shows the recovered condensate and gas products' composition, volumetric flow rate per day, properties and conditions. Comparing the condensate products compositions of the proposed schemes, they have nearly the same trend. LTSUSR-2 recovered more of butanes and pentanes than the other schemes.
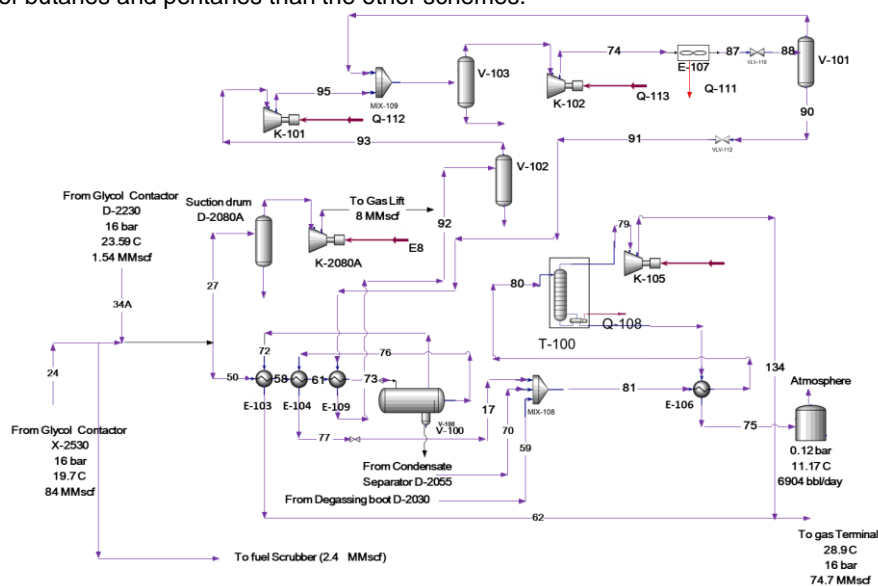

Figure 4: Process flow diagram of LTSUER (Elsheemy, 2017)

Table 4: Properties and conditions of Condensate & Gas products (Elsheemy, 2017)

Condensate product properties				Gas product properties			
	LTSUSR-1	LTSUSR-2	LTSUER		LTSUSR-1	LTSUSR-2	LTSUER
C_4 mol%	9.75	9.89	10.95	C_1 mol%	80.95	81.47	81.21
C_5 mol%	18.7	19.35	16.45	C_2 mol%	9.8	9.9	9.87
C_{6+} mol%	71.02	70.3	70.2	C_3 mol%	5.53	5.62	5.6
Temperature, °C	55	55	11.17	Temperature, °C	55.36	42.6	28.89
Pressure, kPa	113.3	113.3	113.3	Pressure, kPa	1,701	1,701	1,468
Molecular Weight	104.5	104.00	104.6	Molecular Weight	20.48	20.42	20.64
Mass Density, kg/m^3	669.6	669.6	708.6	Mass Density, kg/m^3	13.7	13.7	12.7
Condensate flow, m^3/h.	46.32	47.2	45.73	Gas flow, kgmole/h	3,661	3,715	3,721
RVP at 37.8 °C, kPa	65.5	65.5	75.84	LHV Vol. Basis, MJ/m^3	41.85	41.85	42.27

4. Cost estimation

Preliminary economic analysis is conducted on the schemes using the "Factorial method" mentioned by (Sinnott, 1999) as an economic indicator of the feasibility of this study. The equipment purchase cost is estimated based on the charts mentioned mid-1998 by (Sinnott, 1999). The estimated purchase cost is escalated to year 2016. After obtaining the estimated total investment, their profitability is determined based on the pay-back time equation (2). Total investment cost is calculated using equation (1) (Guthrie, 1969):

$$Total\ investment\ cost\ = Fixed\ capital\ cost\ +\ working\ capital\ cost \tag{1}$$

A simple method for estimating the payback time is to divide the total capital investment by the average annual cash flow:

$$Simple\ pay-back\ time = {Total\ Investment}/{Average\ Annual\ Cash\ flow} \tag{2}$$

It assumes that all the investment is made in year 0 and revenues begin immediately. The simple payback time also neglects taxes and depreciation. Table 5 shows the economic evaluation of the proposed schemes.

Table 5: Economic evaluation for the revamp schemes

	LTSUSR-1	LTSUSR-2	LTSUER
Added Condensate, m^3/y	24,393.60	33,264.00	21,384.00
Equipment Cost, US$	$ 5,433,754.60	$ 8,061,775.17	$ 7,939,185.42
Total investment Cost, US$	$ 20,083,157	$ 29,796,322	$ 29,343,230
Annual production cost (D), US$/y	$ 5,521,753	$ 7,660,867	$ 5,562,634
Added Condensate Recovery Sales income ($503.2/$m^3$) (E), US$/y	$ 12,274,859.52	$ 16,738,444.8	$ 10,760,428.8
Average annual cash flow (E-D), US$/y	$ 6,753,106.52	$ 9,077,577.8	$ 5,197,794.8
Payback time, y	2.97	3.28	5.65

5. Conclusion

A novel NGL recovery configurations was introduced and accomplished either by self-refrigeration or external refrigeration of raw gas. Applying the mentioned methods condenses the heavy components in the raw gas (C_{4+}) which is separated by low temperature separator. Three modifications to the validated model are applied to increased condensate recovery. The revamps' results and recommendations can be summarized as follows: (1) LTSUSR-2 yields the highest condensate production 47.2 m^3/h, LTSUSR-1 yields 46.32 m^3/h, and LTSUER achieves the lowest recovery which is 45.73 m^3/h. (2) As a measure of the schemes profitability, LTSUSR-1 has the lowest payback period (2.97 years), LTSUSR-2 recovers in (3.3 years) and LTSUER recovers in 5.6 years. (3) RVP is adjusted below 82.74 kPa successfully in all the recommended retrofit solutions. This is attained by the usage of re-boiled absorber tower that operates at near atmospheric conditions. The maximum RVP for LTSUSR-1 and -2 is 65.5 kPa while LTSUER max RVP is 75.84 kPa. (4) LTSUSR-2 is the recommended scheme for this GOSP as it leads to highest recovery (47.2 m^3/h) which resembles 9.7 % increase in GOSP condensate production and reasonable payback period (3.28 years).

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