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Optimization of Heat Recovery and Dehydration Unit in the Gas Project

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Abstract

Many of the process used in the gas handling production facilities require the transfer of heat. This is necessary for heating and cooling the gas , as well as for regeneration the various substances used in the gas treating and processing. Natural gas dehydration is the process of removing water vapor from the gas stream to achieve the required export gas specifications and avoid corrosion & hydrate problems. The allowable water vapor content in the export gas ranges from 4 to 7 pounds per MMSCF. Removing water vapor prevents hydrate formation, corrosion and maximizes the pipelines efficiency.It has been noticed that cooling the gas from the CO_2 absorber prior to feeding the dehydration unit renders both CAPEX and OPEX benefits. Exchanging heat with the cold gas from the cold gas separator in two stages, one before and the other after the dehydration unit, optimizes the heat recovery in the gas train by reducing the size of TEG contactor , minimizing the TEG regeneration unit capacity, reducing (and possibly eliminating) the stripping gas requirements, this minimizes the flaring of the gas leaving the glycol still column or the costly alternative requirement of a vapor recovery system to recover the gas into the fuel gas system. Its highly recommended cooling the gas to 30°C prior to the dehydration unit in order to optimize the heat recovery in the gas treatment section of the CPF. This will then satisfy the company philosophy & standards for reduction in fuel gas use, greenhouse gas emissions and improving the process efficiency.

***Keywords**: PROMAX; HYSYS; TEG, CAPEX; OPEX.*

1. Introduction

Raw natural gas is produced from gas wells or extracted at the surface from the fluids produced from oil wells can't be delivered directly to residential, commercial and industrial consumers [1-2]. 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+}) [3-4]. In addition to hydrocarbon components there is gas impurities as carbon dioxide, nitrogen, Hydrogen sulphide and water vapor. 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 gas specifications [5-6].

Natural gas has powerful importance according to its economic and environmental benefits. It is one of the major sources of electricity among energy sources of coal, nuclear and petroleum [7]. Natural gas also burns cleaner than coal or petroleum products, and as more governments begin implementing national or regional plans to reduce carbon dioxide emissions, they may encourage the use of natural gas to displace more carbon-intensive coal and liquid fuels [8].

Raw natural gas produced from underground gas fields or extracted at the surface from the fluids produced from oil wells can't be delivered directly to the industrial consumers [9]. To make it suitable and environmentally safe to use, it is crucial to purify it from all contaminates that can affect its utilization and optimal energy capacity [10]. These contaminants can also cause problems such as corrosion, freezing, plugging, erosion, health, and environmental hazards [11]. Stewart and Arnold noted that the gas contracts regulation restrict H_2S content about 4 ppm and CO_2 about 2% in natural gas stream [12].

Consumption of natural gas worldwide is projected to increase from 120 trillion cubic feet (TCF) in 2012 to 203 TCF in 2040 in the International Energy Outlook 2016 [13]. By energy source, natural gas accounts for the largest increase in world primary energy consumption. Abundant natural gas resources and robust production contribute to the strong competitive position of natural gas among other resources [14].

World consumption of natural gas for industrial uses increases by an average of 1.7% / year, and natural gas consumption in the electric power sector increases by 2.2% / year, from 2012 to 2040. The industrial and electric power sectors together account for 73% of the total increase in world natural gas consumption, and they account for about 74% of total natural gas consumption through 2040 [15-17].

Heat transfer is playing an essential role in the gas processing because heat exchangers are used extensively. A heat exchanger is a system used to transfer heat between two or more fluids [18]. Heat exchangers are used in both cooling and heating processes [19]. Heating medium is any solid or fluid (such as water, steam, air, or flue gas) which is used to convey heat from a heat source (such as an electric immersion heater) to a process or space being heated. Electric heat is often used as a temporary or permanent solution to heat the medium which is then used in various types of heat exchangers throughout the gas plant. Several natural gas processing plants utilize thermal fluid heating systems to offer temperature control and precise [20-21].

The main objective of this study is to investigate the heat recovery options in the gas processing train to optimize the dehydration and hydrocarbon dew pointing units. Also, to review the suitable inlet feed gas temperature to the dehydration unit and ways to achieving it, options to reduce the TEG purity and stripping gas requirements and avoid the need for a vapor recovery unit to maintain zero flaring policy.

2. Gas project description

The gas project will include gas wellhead flowlines and the main process facilities to achieve the required gas and condensate export specifications.

2.1. Central process facilities (CPF)

Figure 1 reveals the major processing units which make up the central processing facility CPF while Figure 2 displays the gas and condensate process facilities in the gas project. The gas project consists of eight wells, a gathering system and CPF, where the production stream from the various fields will be separated into condensate and dew pointed gas products for export. The gas processing involves inlet facilities for liquid separation, mercury removal unit, CO_2 removal unit, dehydration unit, and a hydrocarbon dew-pointing unit to meet the export gas specifications. The condensate separated from the gas in the inlet facilities is stabilized to meet the RVP specification for export condensate. The gas will be exported via export gas pipeline and treated in a dedicated liquefied petroleum gas (LPG) extraction facility to commercial specification required for end user consumption. The condensate will be exported via export pipeline to the oil terminal.

Mercury has been detected up to 70 ng/Sm^3 in some well samples. Well samples are reported to contain no elemental Sulphur, no wax, and no paraffin. Also, the H_2S content of the wells is zero.

2.2. Gas wellhead flowlines

Eight producing wells are initially considered for the gas project. Figure 3 demonstrates the wells and the length for each wellhead flow line. A wellhead pressure of 267 bara, wellhead temperature of 50°C and the flowline pressure of 56 bar at the design flow rate of 0.425 MSCMD (15 MMSCFD) shall be used. The all eight gas wells have the same design flowrate which is 0.425 MSCMD. A range of compositions of different condensate gas ratios (CGRs) can be delivered by each well depending on the layer being produced.

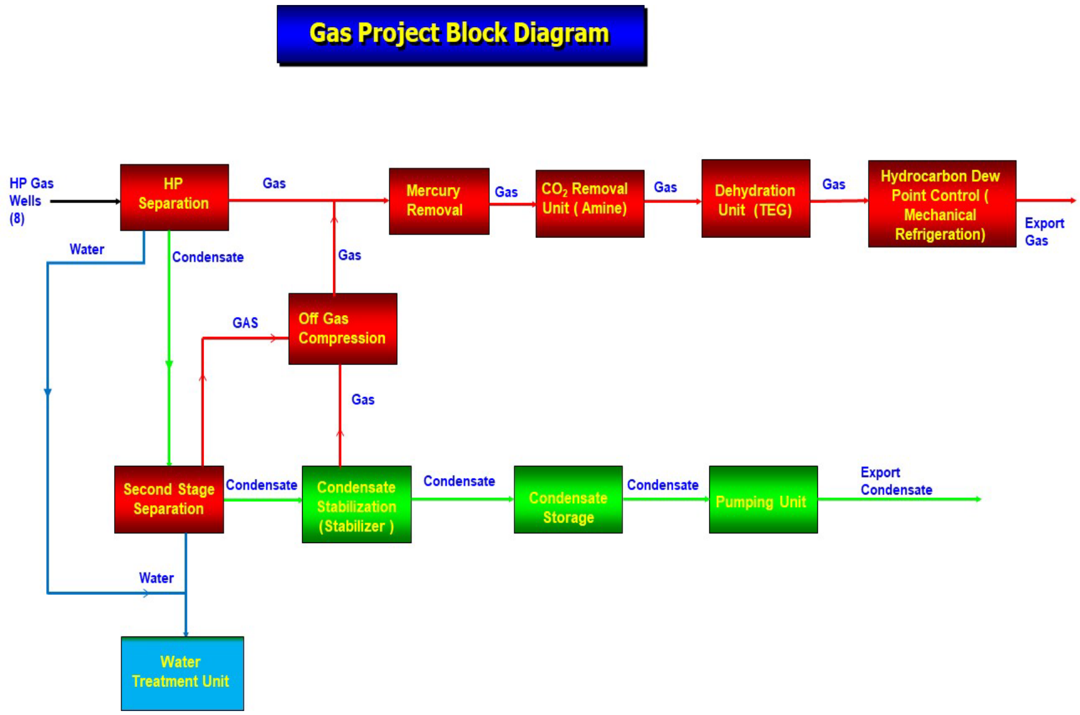

Figure 1. Schematic of the CPF process units

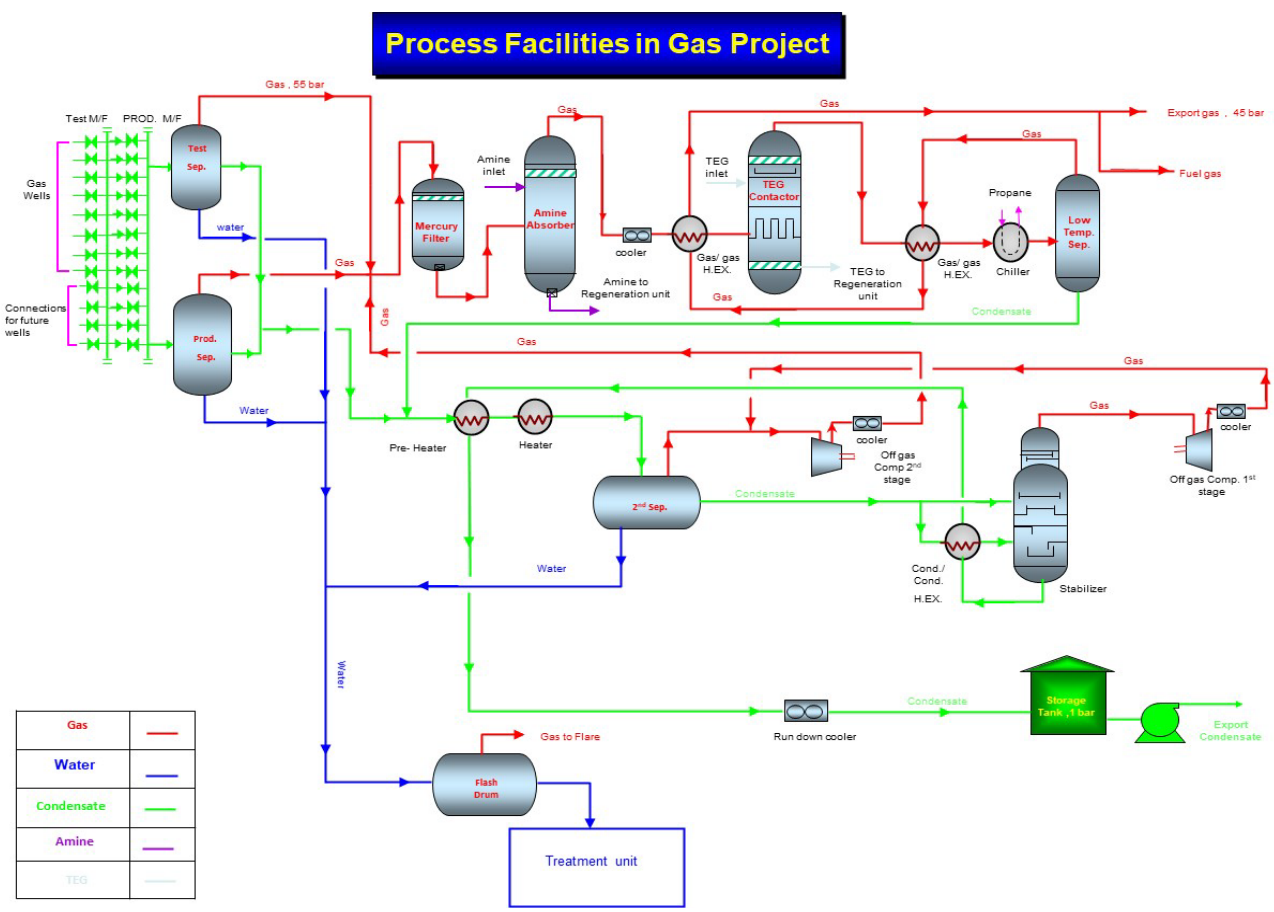

Figure 2. Process Facilities in the Gas Project

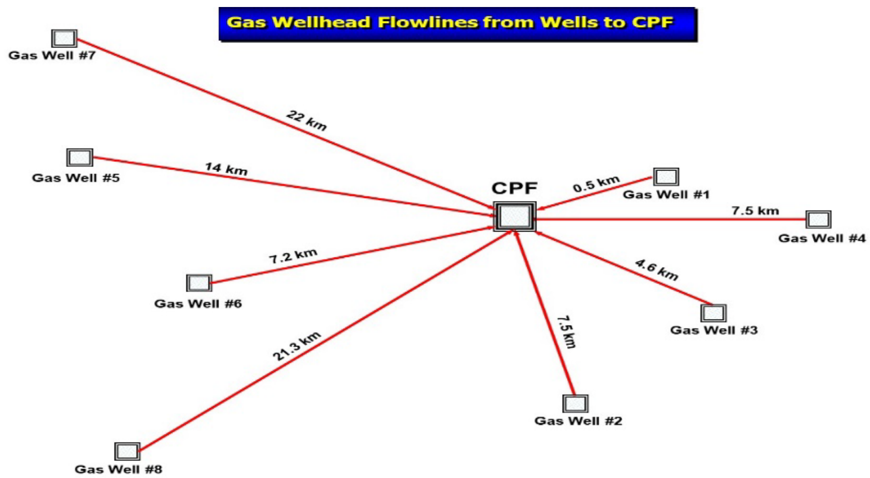

Figure 3. Gas Wellhead Flowlines from Wells to CPF

3. Gas project design capacities

The gas project is designed for a production of 2.7 MSCMD export gas and 10,000 STB/day export condensate. Table 1 displays the design flowrates for the for the process facilities of the gas project.

Table 1. Flowrates design production

Design capacity	Unit	Value
Production from wells	MSCMD	2.9 (lean gas) 3.3 (rich gas)
Gas export (for gas pipeline design)	MSCMD	2.7
Condensate export, maximum	STB/day	10,000
Water production, water-cut	% vol.	10

3.1. Export gas specifications

The export gas specifications for gas are: water dew point-12°C; hydrocarbon dew point at 35 barg +10°C, and CO_2 content < 2.0 mole %. The pressure of the export gas is 44 barg.

3.2. Export condensate specifications

The condensate export specification requires removal of water and light hydrocarbons to meet the BS&W (< 1 %v/v) and RVP (< 0.8 bar & 37.8 °C) specifications.

4. Methodology

4.1. Simulation basis

The gas composition entering the dehydration unit has been obtained from the simulation of the amine sweetening unit in ProMax. The gas flow rate and composition are expected to vary for the following 4 cases:

- Lean gas in winter condition; Rich gas in winter condition.
- Lean gas in summer condition; Rich gas in summer condition.

At a fixed temperature, say 60°C (which can be attained out of a gas cooler downstream of the CO_2 absorber at a maximum ambient air temperature of 50°C) and a fixed mass flow rate, the volumetric flow rate of the gas and the saturation water in the hydrocarbons are observed to be the highest for the lean gas in summer case. Accordingly, this study has been carried out for the lean summer case.

With the addition of the amine sweetening unit there is a possibility of the gas temperature being higher than 60°C, in which case it is proposed to cool the gas to 60°C by an air cooler, and if required for process optimization, further cooling by process heat exchange and refrigerant before feeding the dehydration unit.

The TEG contactor is expected to operate at a higher pressure of around 52 barg. The higher operating pressure is expected to slightly decrease the water content of the gas, but this is over-shadowed by the significant increase due to the higher temperature. Accordingly, the following basis has been considered for this study for the gas at the inlet to the dehydration unit: glow rate 2.85 MSCMD, pressure 51.7 barg, and temperature 60°C (alternative check with 30, 40 & 50°C).

The minimum temperature approaches used for heat exchangers in the study are: shell and tube exchangers 7°C; air coolers 10°C; chiller kettle type 3°C.

4.2. Simulation basis

The all feed gas composition was simulated by using Aspen HYSYS simulation software to choose the best composition for the design of the central process facilities for the gas project based on the total condensate production of 10,000 BOPD and total gas export of 2.7 MSCMD.

Four steady state simulation cases have been developed for the selected feed compositions. These are for a winter case and a summer case with a lean and rich composition each. The varying inlet temperatures and compositions create four unique simulations.

Units are based upon the metric system, with pressures quoted in barg. Standard conditions are defined as 15.6°C (60°F) and 1.01323 bara (1 atm). TVP is calculated at 37.78°C (100°F). All liquid production rates are shown as standard barrels per day (SBPD).

4.3. Simulation software

4.3.1. Aspen HYSYS software

The main process simulations in the gas plant have been carried out with Aspen HYSYS software version 12 [22]. The selected physical property package for the aspen HYSYS model developed for the gas project is the Peng Robinson equation of state with modified interaction parameters fluid package. The binary coefficients in aspen HYSYS have been selected as recommended by the software.

4.3.2. ProMax software

Bryan Research and Engineering "ProMax" software version 5, ProMax was used to simulate the amine (MDEA) and TEG systems [23]. This software is a specialist for amine and TEG packages and give more accurate predictions of performance of CO_2 removal systems. ProMax software is used to provide input to Aspen HYSYS regarding the process outlet temperature from both columns, as well as estimating the amine circulation rate, TEG circulation rate and duties for heat exchangers in the respective regeneration packages.

ProMax software uses SRK and SRK equations of state for vapor phase properties of the amine package and the electrolytic ELR and SRK models for liquid phase properties of the amine and TEG packages. The binary coefficients in ProMax software have been selected as recommended by the software

5. Results and discussions

The export gas is required to be dehydrated and chilled to meet the water and hydrocarbon dew point of the export gas. The subsequent addition of a gas sweetening unit upstream of the dehydration unit will increase the gas temperature at inlet to the dehydration unit. Simulations showed that temperatures as high as 70°C can be seen at the outlet from the CO_2 absorber. This results in the requirement to cool the gas upstream of the dehydration unit in order to optimize the process efficiency, reduction of water loading in feed gas and lower amine losses. The higher the feed gas temperature, the higher the water content; more equilibrium stages are required to achieve the water dew point and a higher TEG purity. Higher TEG purity increases the requirement of stripping gas for TEG regeneration.

A step-by-step procedure as listed below has been adopted to establish the optimum design for the dehydration unit. Ascertain the governing case for the gas feed to the TEG contactor for this study, its composition, flow rate and water dewpoint specifications.

- Review the water concentration at inlet and outlet of TEG contactor, and its water removal efficiency.
- Review the impact of feed gas temperature, TEG contactor stages and the TEG concentration on the TEG flow rates, reboiler duty and the requirement of stripping gas.
- Evaluate the CAPEX and OPEX for the cases.
- Establishing optimum design case.

5.1. Effect of feed gas temperature

The water removal efficiency is defined as [24]: $WRE = \frac{W_{in}-W_{out}}{W_{in}}$.

Table 2 displays the W_{in} and W_{out} calculations based on ProMax software

Table 2. Water content in gas

Parameter	Units	TEG contactor inlet @ 60°C	TEG contactor outlet
Gas flow rate	Sm^3/h	118637	117967
Water dew point	°C	60	-15
Water flow rate	kg/h	405.6	4.5
Water conc.	mg/Sm^3	3419	38.2

Based on calculations mentioned in Table 2 , the required water removal efficiency is calculated as 0.989. Cooling the gas upstream of the dehydration unit condenses the water out of the gas, which reduces the required water removal efficiency and thus reduces the cost of dehydration.

The cooling of the gas from the CO_2 removal unit is achieved first in an air cooler, followed by further cooling in a gas-gas exchanger upstream of the dehydration unit as shown in Figure 4. However, care should be taken to operate this gas-gas exchanger above the hydrate formation temperature. The impact of feed temperature on the TEG concentration and TEG circulation rates are illustrated in Table 3 considering 3 equilibrium stages for the TEG Contactor using ProMax software. It is noted from table 3 that by cooling the feed gas to 30°C, not only is the lean TEG concentration reduced from 99.84 %w to 99.06 %w, but also the lean TEG circulation rate is reduced from 16 Sm^3/h to 2.1 Sm^3/h.

Table 3 shows the effect of feed gas temperature on TEG concentration and TEG circulation rates. The TEG concentration and flow rates are estimated by ProMax software based on 3 number of equilibrium stages.

Table 3 Effect of feed gas temperature

Affected parameters		Gas inlet temperature	
		60°C	30°C
Water at inlet	mg/Sm^3	3419	742
Water at outlet	mg/Sm^3	38.2	37.5
Water removed	Kg/h	401.1	83.1
Water removal efficiency	-	0.989	0.949
TEG concentration required	%w	99.84	99.06
TEG circulation rates	liters TEG / kg water	40.0	24.0
Lean TEG flow	Sm^3/h	16.0	2.1

5.2. Effect of number of stages

It is noted in Table 3 that the TEG concentration and circulation rates are very high for the inlet gas temperature of 60°C. Calculations were conducted by ProMax Software to reduce these by increasing the number of equilibrium stages in TEG contactor, as shown in Table 4.

Table 4 reveals the effect of number of equilibrium stages in TEG contactor based on feed gas temperature of 60°C. The TEG concentration and flow rates are estimated by ProMax software

Table 4. Effect of number of equilibrium stages

Affected parameters		Number of stages 3	5
TEG concentration required	%w	99.84	99.81
TEG circulation rates	liters TEG/kg water	40.0	27.5
Lean TEG flow	Sm^3/h	16.0	11.0

It is observed that although there is a reduction in the lean TEG flow rate, the reduction is not as significant as that achieved by having a cooler feed gas. More importantly, the reduction in TEG concentration is insignificant.

5.3.Equipment capacities and utility requirements

The cases analyzed in Tables 3 and 4 are summarized and their impact on the equipment/unit capacities are listed in Tables 5 and 6. Table 5 illustrate the impact of gas feed temperature, number of stages in the TEG contactor on the TEG concentration , size of TEG contactor , capacity of TEG regeneration unit , stripping gas and reboiler duty.

Table 5. Summary of the cases for CAPEX/OPEX study

Study case	Case description	TEG concentration, %w	Contactor size, Dia. x height, m	TEG regeneration unit capacity, m^3/h	Reboiler duty, kW	Stripping gas, Sm^3/h
1	Gas feed at 60° C; No. of stages = 3	99.84	1.3 x 8.5	18.0	469	335
2	Gas feed at 60° C; No. of stages = 5	99.81	1.3 x 12.5	12.5	412	275
3	Gas feed at 30° C; No. of stages = 3	99.06	1.2 x 8.5	3.0	127	0 (*)

From Table 5 , it can be noticed that when the feed gas temperature the TEG concertation decreases while the TEG contactor size decreases , TEG regeneration unit decreases , reboiler duty decrease and finally the stripping gas will be not required in this case .

Table 6 reveals the impact of gas feed temperature, number of stages in the TEG contactor on duty of the gas-gas exchanger and chiller. From table 6 , it can be noticed that number of gas -gas exchanger will be increased when the gas feed temperature become 30 C . Also, the chiller duty is almost the same at different feed gas temperature and finally, the total power required will decrease with decreasing the feed gas temperature and the number of stages in TEG contactor.

Table 6. Summary of the cases for CAPEX/OPEX study

Study case	Case description	Number of Exchangers: Air cooler	Gas-gas exchanger	Chiller	Gas-gas exchanger process duty, kW	Combined chiller process duty, kW	Total power, kW
1	Gas feed at 60° C; No. of stages = 3	1	1	1	2793	976	1135
2	Gas feed at 60° C; No. of stages = 5	1	1	1	2793	976	1078
3	Gas feed at 30° C; No. of stages = 3	1	2	1	2238+796	972	790

Since the gas needs to be cooled to approximately 5.8°C for hydrocarbon dew pointing, the total cooling requirements for the gas is not changed significantly, i.e., a reduction in the gas temperature to 30°C prior to the dehydration unit does not increase the refrigeration unit capacity significantly. Exchanging heat with the cold gas from the cold gas separator in two stages, one before and the other after the dehydration unit as shown in Figure 4 , optimizes the heat recovery in the gas train. This is evident by minimizing the TEG regeneration unit capacity and the stripping gas requirements, as seen in Tables 5 and 6. Figure 4 shows the optimization of heat recovery in the gas project

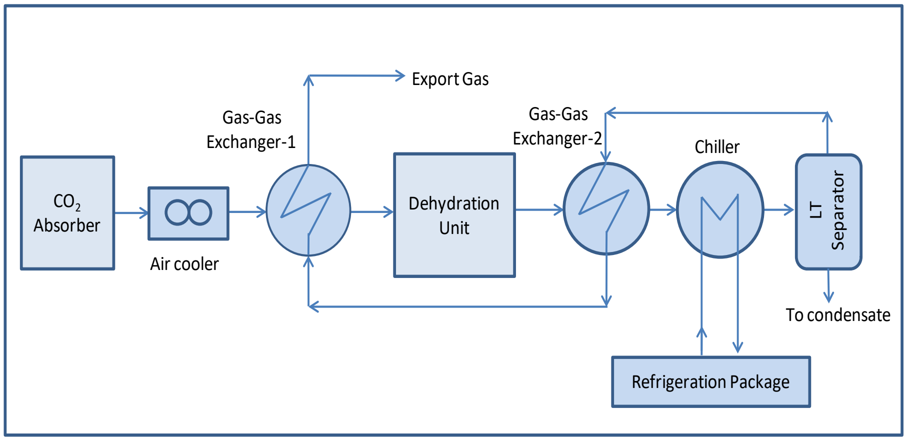

Figure 4. Optimization of heat recovery

The hydrate formation temperature for the gas from CO_2 removal unit is approximately 15.5°C. The cold streams entering the gas-gas exchanger upstream of the TEG Contactor should be maintained above this to avoid risk of hydrates forming on cold spots in the exchangers. The governing scenario for design of the gas-gas exchangers and the chiller is the rich summer case as can be noted from Table 7.

Although an additional exchanger is required for cooling the gas to 30°C upstream of the TEG contactor as compared to the 60°C feed case, the reduction in the total chilling duty is only marginal for the governing "rich gas in summer" case. Accordingly, there is no significant impact on the refrigeration package duty and power generation requirements by cooling the gas upstream of the TEG dehydration unit. Table 7 indicate the duty of the heat exchangers for lean and rich gas compositions at different feed gas temperature

Table 7. Heat exchangers duty comparison for lean & rich gas cases

Study case	TEG absorber feed	Sizing case	Gas-gas exchanger-1 (Upstream of TEG absorber)	(Downstream of TEG absorber)	
				Gas-gas exchanger -2	Chiller
1 & 2	60°C	Lean summer	Not required	2793	976
1 & 2	60°C	Rich summer	Not required	3005	1172
3	30°C	Lean summer	2238	796	972
3	30°C	Rich summer	2370	843	1137

5.4. CAPEX vs OPEX comparison for heat recovery and dehydration study

As noted in the Table 7, the CAPEX comparison for the cases do not include the refrigeration package and power generation unit since the governing case for this equipment is the "rich summer" case and these are not significantly different. However, for OPEX comparison of the "lean gas in summer" case in order to optimize the dehydration unit gas feed temperature, the refrigeration and power generation duty has been considered. Table 8 lists the major equipment required and compares the CAPEX estimates for cases analyzed. Any equipment common for all the cases have not been listed.

Table 8. Equipment List and CAPEX Comparison

Case	Equipment	Equipment US$ x 10^6	Total cost US$ x 10^6
Case 1 Gas feed at 60°C No. of stages = 3	TEG Contactor size: 1.3 m Dia., 8.5 m height	0.708	3.187
	TEG regeneration unit: 18 m3/h	6.940	25.88
	Gas-gas exchanger: 3005 kW (x 1)	0.141	0.63
	Power generation: 1135 kW	1.507	6.78
	Total		36.49
Case 2 Gas feed at 60°C No. of stages = 5	TEG contactor size: 1.3 m Dia., 12.5 m height	1.018	4.581
	TEG regeneration unit: 12.5 m3/h	5.576	20.80
	Gas-gas exchanger: 3005 kW (x 1)	0.141	0.63
	Power generation: 1078 kW	1.457	6.56
	Total		32.57
Case 3 Gas feed at 30°C No. of stages = 3	Contactor size: 1.2 m Dia., 8.5 m height	0.650	2.926
	TEG regeneration unit: 3.0 m3/h	2.368	8.83
	Gas-gas exchanger: 2370 kW (x 1) + 843 kW (x 1)	0.209	0.94
	Power generation: 790 kW	1.190	5.36
	Total		18.06

It can be seen that cooling the gas upstream of the TEG contactor to 30°C reduces the capacity of the TEG regeneration unit from 18 m^3/h to 3 m^3/h TEG flow rate resulting in CAPEX savings of approximately US$ 18.4 million.

It should be noted that for the cases with higher feed temperature to TEG contactor, there is a significant flaring requirement of the gases leaving the glycol still column resulting from higher stripping gas requirement. Avoiding the flaring by compressing and use in the fuel gas system will increase the CAPEX and OPEX further.

Table 9 compares the OPEX estimates for the cases analyzed. The only major utility for the comparison of these options is the fuel gas required to generate power and for stripping gas. It can be seen that by cooling the feed to the TEG contactor, there is a sharp decrease in the TEG reboiler duty and the stripping gas could be eliminated, by cooling the gas to 30°C.

Table 9. Utility Requirements and OPEX Comparison

Case	Fuel gas requirement, Sm^3/h			Cost US$ x 10^6 / year
	For power generation for refrigeration package & TEG regeneration reboiler	For stripping gas	Total fuel gas	
Case 1: Gas feed @ 60°C No. of stages = 3	391	335	726	2.02
Case 2: Gas feed @ 60°C No. of stages = 5	371	275	646	1.03
Case 3: Gas feed @ 30°C No. of stages = 3	272	0	272	0.76

5.5. Summary of heat recovery and dehydration unit optimization study

Figure 5 displays the relative CAPEX requirements at different feed gas temperature and different number of stages in the TEG contactor.

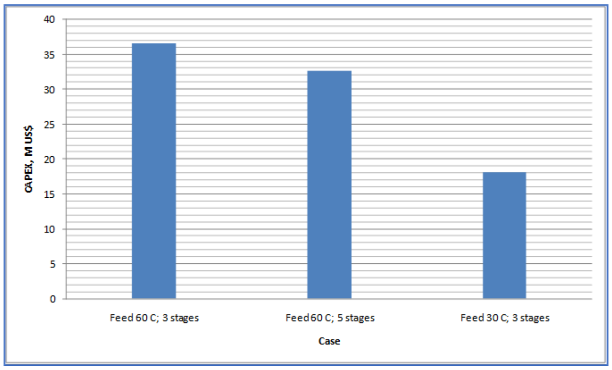

Figure 5. CAPEX comparison for heat recovery and dehydration unit study

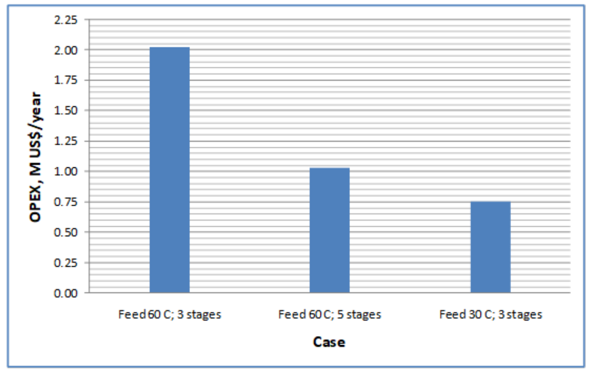

Figure 6. OPEX comparison for heat recovery and dehydration unit study

From Figure 5 , it can be observed that the total CAPEX of the equipment in the gas project are increasing with the increase of the feed gas temperature and number of stages in the TEG contactor. cooling the gas upstream of the TEG contactor to 30°C reduces the capacity of the TEG regeneration unit from 18 m^3/h to 3 m^3/h TEG flow rate resulting in CAPEX savings of approximately US$ 18.4 million.

Figure 6 shows the relative OPEX requirements at different feed gas temperature and different number of stages in the TEG contactor. From Figure 6 , it can be noticed that the total OPEX of the equipment in the gas project are increasing with the increase of the feed gas temperature and number of stages in the TEG contactor. Increasing the feed gas temperature will increase the fuel gas required to generate power and for stripping gas.

6. Conclusions

Based on the economic and technical comparison, the following can be concluded:

- Cooling the gas from CO_2 absorber prior to feeding to the dehydration unit provides both CAPEX and OPEX benefits.
- Since the gas needs to be cooled to approximately 5.8°C for hydrocarbon dew pointing, the total cooling requirement for the gas is not changed significantly, i.e., a reduction in the gas temperature to 30°C prior to the dehydration unit does not increase the refrigeration unit capacity significantly.
- By exchanging heat with the cold gas from the cold gas separator in two stages, one before and the other after the dehydration unit, optimizes the heat recovery in the gas train by reducing the size and weight of TEG contactor and regeneration skid package, minimizing the TEG regeneration unit duty, reducing (and possibly eliminating) the stripping gas requirements . This minimizes the flaring of the gas leaving the glycol still column (note that being very low pressure, this gas cannot be routed to LP flare) or the costly alternative requirement of a vapor recovery system to recover the gas into the fuel gas system.

7. Recommendations

Its highly recommended cooling the gas to 30°C prior to the dehydration unit in order to optimize the heat recovery in the gas treatment section of the CPF. This will then satisfy the company philosophy & standards for reduction in fuel gas use, greenhouse gas emissions and improving process efficiency.

Nomenclature

bpd	*Barrels Per Day.*
BS&W	*Basic Sediment and Water*
CAPEX	*Capital Expenditure.*
CGR	*Condensate Gas Ratio.*
CPF	*Central Processing Facility.*

Dia.	*Diameter.*
ELR	*Extended Long Range.*
K.O	*Knock Out Drum.*
kW	*Kilowatt.*
LP	*Low Pressure.*
LPG	*Liquefied Petroleum Gas.*
MDEA	*Methyl diethanolamine.*
MW	*Megawatt.*
MMSCF	*Million Standard Cubic Feet.*
MMSCFD	*Million Standard Cubic Feet Per Day.*
MSCMD	*Million Standard Cubic Meter per Day.*
NGL	*Natural Gas Liquid.*
OGC	*Off Gas Compressor.*
OPEX	*Operating Expenditure.*
PR	*Peng-Robinson*
RVP	*Reid Vapor Pressure.*
SBPD	*Standard Barrels Per Day.*
SRK	*Soave-Redlich-Kwong.*
TCF	*Trillion Cubic Feet.*
TEG	*Tri ethylene glycol.*
TVP	*True Vapor Pressure.*
Vol.	*Volume.*
W_{in}	*Water content in the gas at the inlet to TEG contactor, mg/Sm^3.*
W_{out}	*Water content in the gas at the outlet of TEG contactor, mg/Sm^3.*

References

[1] Kidnay AJ, Parrish WR, McCartney DG. Fundamentals of natural gas processing. 3rd edition, CRC Press, Boca Raton, New York, USA, 2019.

[2] Speight JG. Natural gas: a basic handbook. Gulf Professional Publishing , 2019.

[3] Perry, R.H. , Green, D.W., Perry's Chemical Engineers' Handbook , 9th edition, McGraw Hill, 2018.

[4] Lyons WC, Plisga GJ, Lorenz M. Standard Handbook of Petroleum and Natural Gas Engineering. Second edition, Gulf professional, Imprint of Elsevier, Burlington, MA, USA, 2016.

[5] Abdel-Aal H, Aggour M, Fahim L. Petroleum and Gas Field Processing. 2nd edition , New York: CRC Press,2016.

[7] Noaman AA. Modelling and Optimization of the Gas and Condensate Process Facilities in the Gas Plant. Australian Journal of Basic and Applied Sciences, 2020;14(12): 26-45.

[8] Noaman AA, Ebrahiem EE. Comparison of Natural Gas Hydrocarbon Dewpointing Control Methods. Journal of Advanced Engineering Trends,2021; 40(2): 99-116.

[9] Ashour I, Abdelaziz AA, Ebrahim IE, Nassar MM, Farag T. Comparison of CO_2 and H_2S capture methods from natural gas. The 4th Int. Conf. on Advanced Sciences (ICAS4), Hurghada, Egypt, Nov. 7 – 10, 2017.

[10] Ebrahiem EE, Ashour IA, Nassar MM, Abdel Aziz A. A Comparison of Natural Gas Dehydration Methods. Minia Journal of Engineering & Technology (MJET), 2018; 37(2)..

[11] Ebrahiem EE, Noaman AA, Aboulfotouh TM, Mansour MI, Hakim YA. CO_2 Emissions, Green House Gas Calculations and Controlling in the Gas Plant. Egyptian Journal of Chemistry, 2021; 64(3): 1421-1431.

[12] Stewart M, Arnold K. Gas Sweetening and Processing Field Manual , Gulf Professional Publishing, Waltham, 2011.

[13] DOE/EIA., International Energy Outlook, Energy Information Administration, Department of Energy, Washington DC,2020.

[14] British Petroleum (BP), BP Statistical Review of World Energy, 68th edition, 2019.

[15] International Gas Union, World LNG Report , 2019, Barcelona, Spain,2019.

[16] U.S. Energy Information Administration, Monthly Natural Gas Trades, New York: EIA Press, 2019.

[17] International Energy Agency (IEA), Energy Efficiency 2017, Paris: OECD/ IEA, 2017.

[18] Green DW Southard MZ. Perry's Chemical Engineers' Handbook, 9th Edition, McGraw-Hill , 2018.

[19] Al-Sammarraie AT, Vafai K. Heat transfer augmentation through convergence angles in a pipe, Numerical Heat Transfer, Part A: Applications.,2017; 72 (3): 197–214.
[20] Polley GT, Athie R, Gough M. Use of Heat Transfer Enhancement in Process Integration. Heat Recovery System, Chp. ,2013; 12(3): 191.
[21] Kianfar F, Mahdavi Moghadam SR, Kianfar E. Energy Optimization of Ilam Gas Refinery Unit 100 by using HYSYS Refinery Software, Indian Journal of Science and Technology, 2015; 8(S9): 439–444.
[22] Aspen Tech., Advanced Process Modeling Using HYSYS, 2019.
[23] BR&E ProMax® Version 5, Bryan Research & Engineering. Inc. P.O. Box 4747 Bryan, Texas, 2017.
[24] GPSA Engineering Data Book, 14th edition, Gas Processors Suppliers Association, Tulsa, Oklahoma,2017.

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