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Recovery of Valuable Fuel Fractions from Oil Fields Waste Gas Streams

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Abstract

Large amounts of gases recovered from oil and gas extraction operations are flared because they will not be used. One of the top challenges in the energy and environmental sectors is gas flaring which not only wastes valuable fuel, but causes severe environmental damages at the same time. To meet the increasing global energy demands and reduce negative environmental impacts, a more efficient use of gas needs to be found. In this study, a simulation investigation of the recovery and emission reduction of the technical and economic flare gas systems was performed. The simulations were done using HYSYS Version 11 where 10.6 MMSCF of gas were recovered from 5 different oil fields which would have otherwise been flared daily. The economic value of the products that could be extracted from this gas was estimated to be $15.65 million annually which shows how valuable the findings of this study are both, environmentally and economically. The profit came from recovering 7.413 MMSCFD of sales gas (light end product), 70.28 Ton/Day of LPG, and 200 BPD of condensate (C_5^+).

Keywords: *Waste gas recovery; Environmental emissions; Simulation system; Hydrate inhibition; Sweeting; Refrigeration; Economic evaluation.*

1. Introduction

The process of burning-off associated, waste/unwanted, or excess generated gases are referred to as gas flaring. Gas flaring is performed in industrial processes such as hydrocarbon production, coal bed methane and landfill gas extraction, wastewater treatment, and petrochemicals manufacturing where excess gas is generated during normal or unplanned over-pressuring operations. Despite flaring being a controversial method, it is very popular. More than 150 billion cubic meters (bcm) of gas is flared annually from the oil and gas production industry alone [1]. Russia topped the list of gas flaring countries in 2020 when it was found to contribute 15% of global flaring on its own [2].

The chemical compositions of flare gas differ depending on its generating source and industrial processes. For instance, natural gas is mostly composed of methane, some ethane, and varying quantities of other hydrocarbons and gases. Emam reported that the evaluation of the flare gas heat source is very difficult because of measurement challenges due to the noticeable composition changes of the gas at high flow velocities [3].

Gas flaring contributes to the overall burden of global warming by producing a significant amount of greenhouse gases (GHGs) which is considered a major environmental concern, globally. The main sources of GHGs are the emission of smoke and carbon dioxide (CO_2) [4]. Flaring is considered the second in overall emissions with 8 and 32% of CH_4 and GHG, respectively [5]. Moreover, flaring releases about 400 million tons of CO_2 into the atmosphere annually [1]. Also, flaring leads to several types of harmful compounds because of the incomplete combustion of hydrocarbons which impacts both human populations and agriculture by acid rain and heavy metals [6].

Eliminating routine gas flaring can provide much required energy, especially to developing countries such as Egypt, Mexico, and Nigeria. Globally, the revenues lost through gas flaring is estimated at about US$ 25 billion per year at $5.00 per MMBTU [1]. So, the broad description of flaring is a multibillion-dollar waste and a catastrophe for the environment [7]. As a counter of the harmful effects of gas flaring, flare gas recovery systems (FGRS) can be utilized to minimize, or even eliminate, these effects. The World Bank global gas flaring reduction (GGFR) partnership and the global methane initiative have an international direction to decrease gas flaring and venting [8]. FGRS reduce operation and maintenance costs, thermal radiation, gas emission, noise, and air pollution. In addition, FGRS reduce the consumption of fuel gas and steam [8]. Flaring gas can be eliminated, minimized and/or recovered by alternative techniques, including: redistribution of the natural gas networks, using pipelines for transportation, developing cleaner power generation programs, reinjection in enhanced oil recovery, feedstock for petrochemical manufacturing, and liquefied and compressed natural gas [9].

The installation of most FGRS is based mainly on economics, where the design includes the payback on the equipment in a short enough timespan to justify the investment [8]. Different recovery technologies for electricity production or alternative efficient applications can be used. Furthermore, method of vapour recovery can be controlled the hydrocarbons damage where recaptured resources can be reprocessed in the plant [10]. The recovered flared gas can be used in the generation of electricity instead of the adopted method of burning fossil fuels in conventional gas turbines [11]. This method has a potential of reducing the emissions of CO_2 about 50 million tons annually as per today's production rates [10]. Hajizadeh *et al.* [12] used different flare gas recovery methods such as liquefaction, LPG (liquefied petroleum gas) production, and gas compression unit. Their economic analysis showed that the rate of return for LPG and liquefaction methods is above 200% [12]. Flare recovery units depend on sub cooling by through external refrigeration packages. These units have result in high recovery of sales gas and LPG, good economics reflected in high net present value (NPV), high internal rate of return (IRR), and shorter payback periods [13]. Three methods for flare gas recovery and reuse were studied through simulation and economic evaluation: gas-to-liquid production, gas turbine electricity generation, and compression and injection into the refinery pipelines [14]. Economically, the gas compression technique was found to be the best approach in gas refinery with a medium capital investment. This is because of the lower capital investment costs as well as the higher return on investment rate [14]. Comodi *et al.* [15] reported that, the use of a liquid ring compressor to treat a flow rate of 400 kg/h flare gas and reuse it was economically profitable with an interesting payback time estimated at less than 3 years.

The purpose of this paper is to study the design scheme of a flare gas recovery system to conserve or utilize the waste gas streams onsite and at other nearby facilities. This design scheme is used to recover valuable fuel fractions such as: sales gas, LPG and condensate products for market delivery. The simulation models are built using HYSYS steady state simulation software (Version 11) which has been proved as a reliable tool to approach the industrial results. The recovery of NGLs from the feed in the flare recovery scheme depends on sub cooling by means of an external refrigeration package. Additionally, an economic analysis is performed using Aspen Process Economic Analyzer (Version 11).

2. Methodology

To conserve energy, the general strategy followed was to utilize the onsite gas as well as that of nearby facilities. This is particularly important in reducing the reliance on purchasing fuel for the energy supply.

2.1. Case study

The proposed flare gas recovery unit is to be built close to 5 Egyptian oil fields in the Egyptian Eastern Desert with a total lifetime of 20 years of the designed unit. About 10.6 MMSCFD of flare gas is reported from these fields which makes it one of the largest flaring sites in Egypt. The inlet feed temperature and pressure will be 20°C and 260 psig, respectively and the feed composition to the flare recovery unit is presented in Table 1. Mono ethylene glycol (MEG) with 80 wt% MEG and 20 wt% H_2O is used for hydrate inhibition. The expected products of the recovery unit are to be sales gas, LPG, and condensate. The flare gas recovery units are to be built as skid mounted units to facilitate the relocation to other production sites for future projects with an on-stream factor of 340 days/year.

Table 1. Flare gas composition

Component	Mole %	Component	Mole %
CO_2	2.87	i-C_4	1.32
N_2	0.49	N-C_4	3.1
C_1	59.26	C_{5+}	2.86
C_2	12.23	H_2	0
C_3	9.13	H_2S	8.65

The operation of the flare gas recovery unit as well as the heaters of the de-ethanizer and de-butanizer towers requires large amounts of fuel gas. This fuel gas is supplied from the condensated sales gas (after reaching the dew point) as wells as the light components recovered from the de-ethanizer.

2.2. Unit simulation

Figure 1 displays the overall designing process of the flare gas recovery unit in addition to the steady state simulation model of the recovery process. This study recovers NGLs from the flare gas. This is followed by the fractionation of the NGLs to produce condensate and LPG. This case study was based on the flare gas composition and amount from different oil fields.

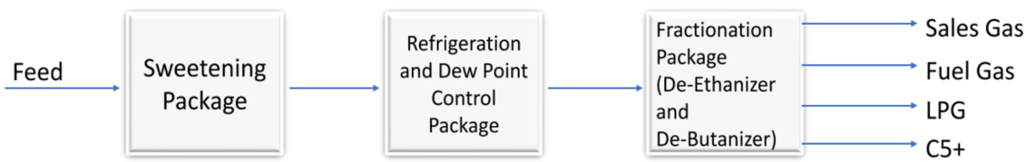

Figure 1. Flare gas recovery schematic

All simulation models presented in this work were built using HYSYS Steady State simulation software (Version 11), which has been proven as a reliable tool in the industry [13]. The two software packages used inside HYSYS were CPA and Acid-Gas Chemical Solvent Fluid.

3. Results and discussions

This section describes the proposed process scheme applied in this study for flare gas recovery. Moreover, the main steps of producing saleable products (i.e. sales gas, LPG, and condensate) were reported and discussed in addition to the results and the economic evaluation.

3.1. Feed

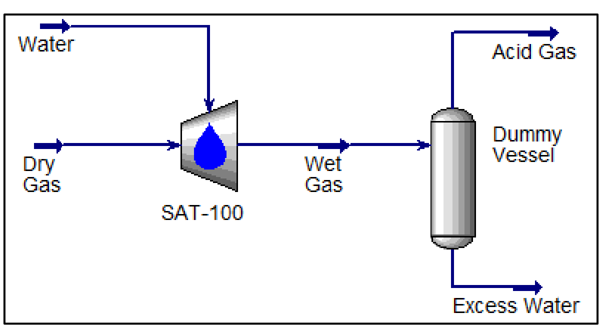

Figure 2. Feed - Simulation model

A dummy vessel is used to saturate the dry gas of the feed with water through the combination of the two streams (Figure 2). Any excess water is drained from the dummy vessel to ensure the saturation of the dry feed gas. This is done by the saturation tool which sets the water flow rate to 12.97 kg/h at the input valve

3.2. Sweeting package

The purpose of the sweetening process is to remove any sulfur in the feed in the form of H_2S. The removal of the high H_2S concentration in the feed (8.65%) is critical to prevent corrosion, pipe deterioration, and any health related issues. An amine treatment unit is used for the sweetening process. It consists of an absorber, a two-phase separator, and an amine regulator (Figure 3). The amine is regenerated and recycled back into the process through steam stripping. The heat from the regenerator's lean amine is exchanged with the rich amine feed to increase the energy consumption efficiency.

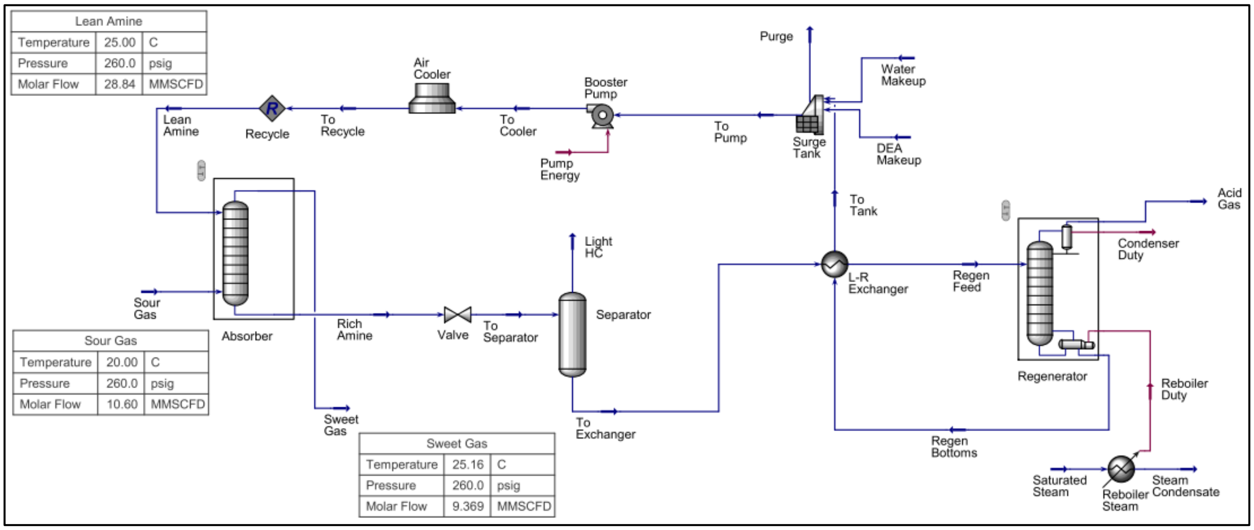

Figure 3. Sweetening package

3.3. Refrigeration and dew point control package

The refrigeration and dew point control package is shown in Figure 4. It consists of the MEG injection package, the feed heat exchangers, propane refrigeration unit, and low temperature separator. MEG is injected to the recovery unit to inhibit the formation of hydrates in the sweet gas stream and the chiller feed stream in the sections operating at low temperatures (below 0℃). The temperature of MEG at the injection time is set at 50℃ and is injected at 50 kg/h into the sweet gas stream to maintain a 10℃ safety margin between the operating temperature and the hydrate formation temperature.

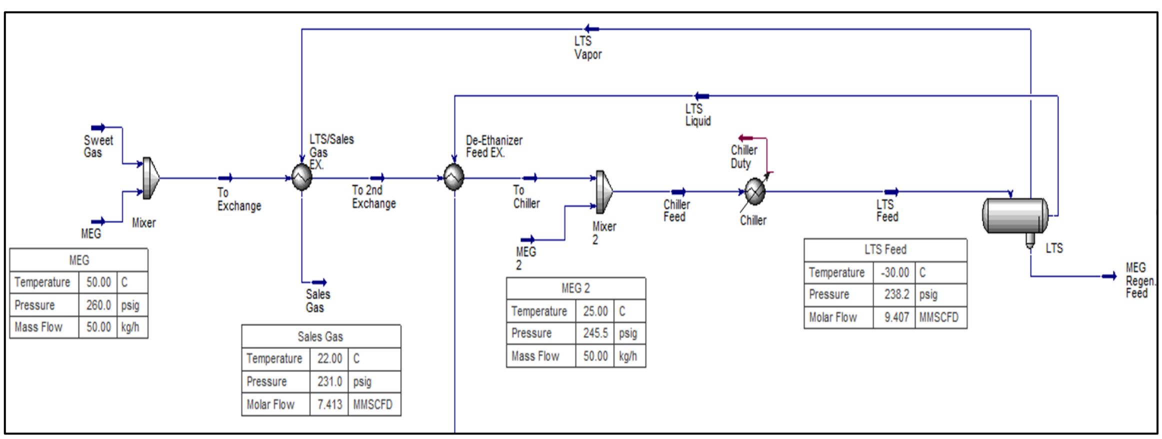

Figure 4. MEG injection, refrigeration and dew point control package - Simulation model

To reduce the cooling requirement from external sources (i.e. refrigeration package), heat integration is done. This is achieved by cooling the dehydrated feed gas in two successive heat exchangers (Figure 5) against low temperature separators. The feed temperature is reduced from 26.05℃ to 0.00℃ through the exchange with the cooled products of the recovery units such as LTS/sales gas. After that, the chiller feed is directed to the propane refrigeration package which is simulated as a simple cooler with -30℃ outlet temperature. Propane refrigeration is necessary for sales gas and liquid hydrocarbon recovery as heat integration by itself is not enough to achieve these low temperatures. After the propane refrigeration, the feed is directed to the low temperature separator to be separated into gas and liquid. The separated gas is then heated against the hot feed gas of the heat exchanger.

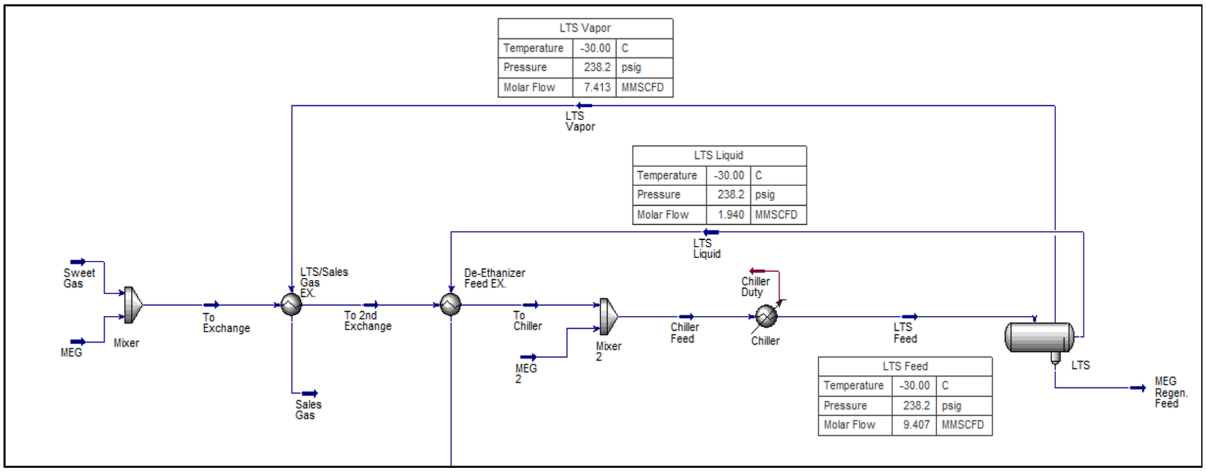

Figure 5. Reduction of feed temperature - Simulation model

3.4. Fractionation package

Figure 6-a displays the fractionation package which consists of two systems: the de-ethanizer, and the de-butanizer. The fractionation is used to separate the hydrocarbon recovered from the feed to the required commercial products, i.e. LPG and condensate. Any feed components lighter than LPG are recovered from the top of the de-ethanizer column and circulated back to the fuel system. As for the de-butanizer, the incoming liquid stream from the de-ethanizer is fractionated into LPG and condensate.

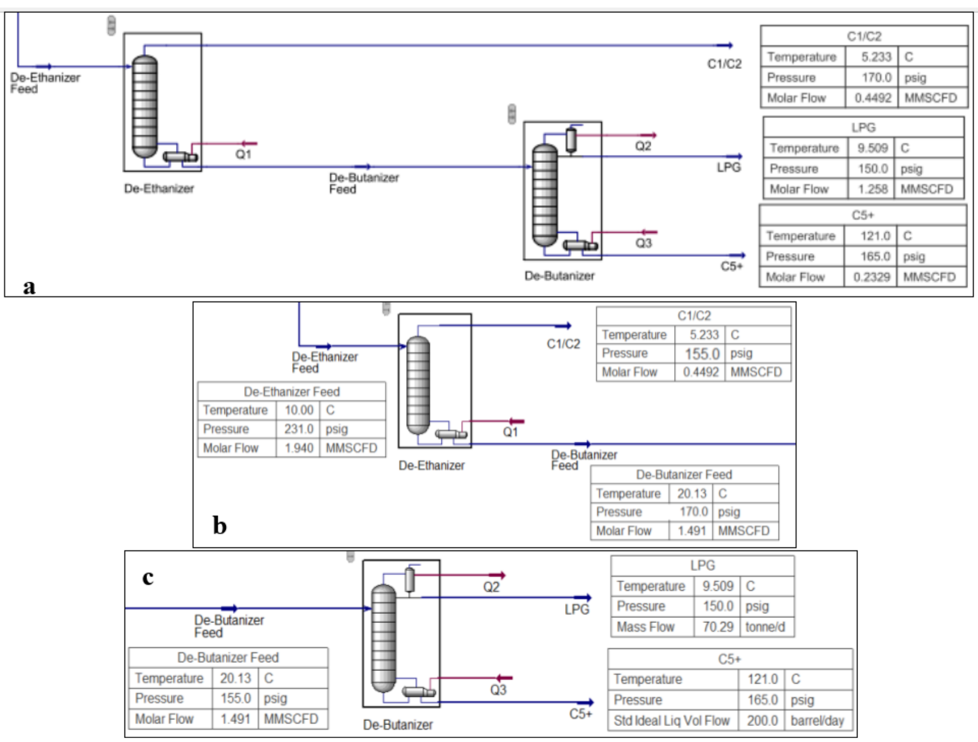

Figure 6. a) Fractionation towers, b) de-ethanizer, c) de-butanizer - simulation model

Table 2. De-ethanizer tray profile

	Stage	Pressure (psig)	Temperature (C)	Net Liquid (MMSCFD)	Net Vapor (MMSCFD)
Main Tower (1)	0	170.0	5.233	1.569	0.449200
Main Tower (2)	1	168.3	5.178	1.567	0.007779
Main Tower (3)	2	166.7	5.119	1.566	0.007619
Main Tower (4)	3	165.0	5.058	1.564	0.007445
Main Tower (5)	4	163.3	4.995	1.562	0.007287
Main Tower (6)	5	161.7	4.932	1.561	0.007118
Main Tower (7)	6	160.0	4.877	1.559	0.006949
Main Tower (8)	7	158.3	4.880	1.558	0.006791
Main Tower (9)	8	156.7	5.212	1.561	0.006698
Main Tower (10)	9	155.0	7.451	1.581	0.006958
Re-boiler	10	155.0	20.13	1.491	0.008967

3.4.1. De-ethanizer

The de-ethanizer was designed to be a re-boiled absorber with 10 trays with the feed coming in to the first tray. The top outgoing stream (mainly methane and ethane at 155 psig) is recycled back to the fuel system for energy saving while the bottom one (which is set at 20.1°C and 170 psig) flows into the de-butanizer for separating LPG from the heavy condensate. Figure 6-b and Table 2 show the de-ethanizer system and its tray profile, respectively.

3.4.2. De-butanizer

The de-butanizer is designed as a 10-tray distillation column with the inlet stream connected to the second tray (Figure 6-c). The pressures of the condenser and re-boiler are set at 150 and 165 psig, respectively (Table 3). The condensate is flown to its storage tanks at a rate of 200 bbd while the separated LPG is flown at 70.29 ton/day to the LPG storage bullets.

Table 3: De-butanizer tray profile

	Pressure (psig)	Tempera-ture (C)	Net Liquid (MMSCFD)	Net Vapor (MMSCFD)	Net Feed (MMSCFD)	Net Draws (MMSCFD)	Duty (kcal/h)
Condenser	150.0	9.509	0.314571			1.25830	3.778E+05
Main Tower (1)	150.0	56.500	0.319895	1.57286			
Main Tower (2)	151.7	62.770	1.952640	1.57819	1.4911		
Main Tower (3)	153.3	80.080	2.031680	1.71983			
Main Tower (4)	155.0	90.080	2.071340	1.79881			
Main Tower (5)	156.7	96.990	2.091230	1.83853			
Main Tower (6)	158.3	102.600	2.103550	1.85842			
Main Tower (7)	160.0	107.500	2.115910	1.87074			
Main Tower (8)	161.7	111.900	2.131580	1.88310			
Main Tower (9)	163.3	115.700	2.150450	1.89877			
Main Tower (10)	165.0	118.900	2.176150	1.91764			
Re-boiler	165.0	121.000		1.94334		0.23281	4.104E+05

3.5. Flare gas recovery scheme - simulation results

Table 4 displays the simulation results. The obtained products of the recovery unit were sales gas, LPG, and condensate at the rates of 7.413 MMSCFD, 70.28 T/D, and 200 bbl/D, respectively. These products are free of CO_2 and H_2S. From an environmental perspective, the flare gas recovery unit will result in a reduction of GHG emissions from the refinery. Also, normal waste gas flaring will be eliminated by recovering valuable and clean fuel fractions. Ultimately, by reducing the emissions of the flare, this recovery unit will eliminate the impacts of the refinery plant on the surrounding environment.

Table 4. Steady state simulation results for flare gas recovery scheme

Mole %								
Component	Dry Feed	Wet Feed	MEG	DE amine	Sales Gas	LPG	C5+	Fuel Gas
CO_2	2.87	2.87	0.00	0.00	0.00	0.00	0.00	0.00
N_2	0.49	0.49	0.00	0.00	0.73	0.00	0.00	0.12
C_1	59.26	59.18	0.00	0.00	81.29	2.12	0.00	47.35
C_2	12.32	12.30	0.00	0.00	12.64	18.38	0.00	29.77
C_3	9.13	9.12	0.00	0.00	4.50	43.82	0.01	17.90
$I\text{-}C_4$	1.32	1.32	0.00	0.00	0.26	8.97	0.470	1.36
$N\text{-}C_4$	3.10	3.10	0.00	0.00	0.44	20.99	8.60	2.51
C_{5+}	2.86	2.86	0.00	0.00	0.14	5.71	90.68	0.99
H_2	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
H_2S	8.65	8.46	0.00	0.00	0.00	0.00	0.00	0.00
H_2O	0.00	0.14	20.00	75.00	0.00	0.00	0.00	0.00
MEG	0.00	0.00	80.00	25.00	0.00	0.00	0.00	0.00
DE amine	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Conditions								
Vapor fraction	0.9944	1.00	0.00	0.00	1.00	0.00	0.00	1.00
Temperature (°C)	20	20	50	25	22	9.509	121	5.233
Pressure (psig)	260	260	260	260	230	150	165	170
Molar flow (MMSCFD)	10.59	10.6	0.0188	28.84	7.413	1.258	0.2328	0.4492
Mass flow (Ton/D)	331.8	332.1	1.2	791.8	173.2	70.28	19.72	14.74
Liquid flow (bbd)	5120	5122	6.847	4870	3304	866.8	200	240.5

3.6. Economic analysis

The economic analysis for the flare gas recovery scheme is discussed in this section which was done using Aspen Process Economic Analyzer (Version 11). Aspen Process Economic Analyzer allows process engineers to: i. use initial process data to perform economic feasibility decisions, ii. estimate the capital and operating costs for a process, iii. evaluate alternative processes to compare relative profitability. Profitability for each scheme is obtained by presenting capital and utilities costs, products sales, NPV, IRR, and payback period.

To achieve this economic analysis, the following assumptions were made:

- The unit's economic lifetime is 20 years.
- Discount rate and taxes are 17.25% and 40%, respectively.
- A straight-line depreciation model and 5% working capital..
- Operating charges are 25% with 50% plant overhead.

The budgetary prices for the 10.6 MMSCFD flared gas recovery unit are shown in Table 5. Additionally, assumed product prices and total annual sales are presented in Table 6.

Table 5. Budgetary price for 10.6 MMSCFD flared gas recovery unit

Compression package	1.9 M US Dollars
Amine gas sweetening package	3.95 M US Dollars
Dew point control package	0.89 M US Dollars
Fractionation package	1.7 M US Dollars
Refrigeration package	1.9 M US Dollars
Meg regeneration	0.89 M US Dollars
Heat medium	0.83 M US Dollars
Instrument air	0.265 M US Dollars
Piperack and Interconnecting Piping	1.86 M US Dollars
Modularized MCC and control room	1.135 M US Dollars
Total	15.32 M US Dollars

Table 6. Assumed product prices & total annual sales

Product	Unit Price US $	Quantities / yr.	Total Price $
Sales gas	4750 $ / MMSCF	2520	11,970,000
LPG	950 $ /T	23895	22,700,250
Condensate	65 $ / Barrel	68000	4,420,000
Total Sales $/yr.			39,090,250

The flare gas recovery scheme results are as follows:
Total Capital Cost= $15,320,000; Total approximate Product Cost= $13,000,000 /Yr
Profit= $26,090,250 /Yr; Net Profit= $15,654,150 /Yr
Payback period= $\frac{Total\ Fixed\ Cost}{Net\ Profit} \approx 1\ year$

These results show that the recovery unit is profitable through high net profits. Also, the feasibility study indicates positive results meaning that the current payback period of investments may be adequate for many investors.

4. Conclusions

A flare gas recovery scheme is designed, simulated, and evaluated based on factors such as feed analysis, feed capacity, feed conditions, products, and the lifetime of the unit. Hydrate inhibition is achieved through MEG injection which is necessary for the sub cooling required for the recovery of NGLs in the flare recovery unit by means of an external refrigeration package. Furthermore, the flare recovery unit includes two heat exchangers steps of the dehydrated feed gas to achieve the required cooling from 26.05℃ to 0.00℃. Following that, the chiller feed is directed to the propane refrigeration package with -30℃ outlet temperature. CO_2 and H_2S are separated in the recovery unit from the obtained products i.e., sales gas & LPG, and condensate. Therefore, the optimum process scheme for flare gas recovery is a unit depending on sub cooling through external refrigeration which reduces GHG emissions and normal waste gas flaring from the plants by recovering valuable and clean fuel fractions. Additionally, this scheme has better economics reflected in higher net profit and less payback period.

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