Optimum ethane recovery in conventional turboexpander process

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Ab s t r a c t

Ethane recovery in a conventional turboexpander process is optimized considering different demethanizer pressures and different feeds: a lean gas and a rich one. The design variables are varied, while meeting process constraints, in order to find the optimum conditions achieving the maximum profit. The analysis covers the whole process including the refrigeration part, and the entire typical demethanizer pressure range. The optimum ethane recovery is compared with the maximum possible recovery for each value of the demethanizer pressure .Recommendations are given regarding the selection of the level of ethane recovery, along with the demethanizer pressure, and refrigeration recovery system..

Keywords: Natural gas liquids (NGL); Ethane; Recovery; Optimization; Turboexpander; Simulation

1. Introduction

Natural gas liquids are valuable components in natural gas. Several extraction processes were proposed. Reviews about processes for NGL recovery can be found in Manning and Thompson (1991), Arnold and Stewart (1999), GPSA (2004) and Kidnay and Parrish (2006). A recent overview is given by Chebbi et al. (2008), and briefly reviewed here. Different options include JT (Joule Thompson) valve expansion, external refrigeration (Russel, 1977) using propane as a refrigerant, and turbo expansion. Typically, a turboexpander is used in combination with JT expansion and propane refrigeration. Other methods include cascade refrigeration which is complex and requires high compression cost (Manning and Thompson, 1991). Mixed refrigerant (MacKenzie and Donnelly, 1985;ManningandThompson,1991)is commonly used in LNG processes, but much less in NGL recovery. Refrigerated lean oil absorption is expensive in terms of equipment and energy requirements and is hard to operate(ArnoldandStewart,1999; GPSA, 2004).

The turboexpander process is dominating ethane recovery processes. Different processes have been proposed. The evolution in design for the "old" generation was summarized by McKee (1977). The simple plant consists of turbo expansion.
The other options include the use of side reboiler, refrigeration, and two stages of expansion (McKee, 1977).

The next generation of ethane recovery processes includes the residue recycle (RR) process, the gas subcooled process (GSP) (Pitman et al., 1998; GPSA, 2004). In the GSP, the gas stream leaving the cold separator is split into two streams, one of them feeding the turboexpander, and the second one is subcooled by the demethanizer overhead stream, flashed in a valve, and then sent to the demethanizer as a reflux. The next generation also includes the CRR process. It has one addition to the GSP and is reported to achieve very high ethane recovery (Pitmanetal.,1998);but with acryogenic compressor, cost may be prohibitive (Leeetal.,1999). The IPSI is another modification of the GSP (Lee et al., 1999; GPSA, 2004). Combining GSP with liquid subcooled process (LSP) (Jibril et al., 2006) was found to yield higher recoveries than GSP and LSP used individually using process simulation for eight different feeds. RSV (recycle split-vapor) process (Pitman et al., 1998) is another modification of the GSP, and RSVE (recycle split-vapor with enrichment) process is itself a modification of the RSV (Pitman et al., 1998) found to lower capital cost compared to RSV.

Studies have been carried out in order to optimize ethane recovery. Wang considered different feeds with low CO_2 content and found that a combination of turboexpansion
and external refrigeration achieves minimum energy requirements (Bandoni et al., 1989). Bandoni et al. (1989) divide the process into two sectors. The first one includes compression, recompression and atmospheric heat exchange while the second sector includes separation, expansion refrigeration, and heat exchange below ambient temperature. A general superstructure embedding different features was considered. The main optimization variables they used are: cold tank temperature and pressure in addition to demethanizer column pressure. The energy balance over sector 2 was found to provide guidelines for the optimum ethane recovery process selection, and the economic objective function was found to increase with demethanizer pressure due to less energy requirement for gas recompression. Fernandez et al. (1991) extended the analysis to CO_2-

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https://doi.org/10.1016/j.cherd.2009.11.003

containing mixtures. One constraint was added: prevention of carbon dioxide precipitation in the demethanizer. The effect of increasing the demethanizer pressure was also found to enhance the net annual benefit(calculated as sales minus operating cost minus annual investment cost).

A cost-revenue comparison was performed for GSP, SFR (split-flow reflux), ISS (typical industry-standard stage plant) a typical turboexpander process, and CRR processes by Wilkinson and Hudson (1992) for one gas feed containing 6.76% C_2+, a flow rate of 100MMscfd, and 1050psig nominal pipeline pressure. The SFR process is based on GSP, with the subcooled stream used to reflux the demethanizer indirectly. The GSP was found to be the best choice if minimum capital cost is needed, and CRR process was found to be the process of choice if ethane market is expected to be favorable. SFR process was found to be the best option if ethane price projections are low.

LSP, BTEP (basic turboexpander process), GSP, and 2-sDP (two-stage demethanization process) were compared by Diaz et al. (1997) for different feed gas mixtures with % C_2+ ranging from 6% to 25%. The GSP was found to give the maximum profit for most mixtures, especially the lighter ones (lower C_2+ content). Optimal design and operating conditions were found using a superstructure embedding different expansion alternatives together with mixed-integer nonlinear programming (MINLP). The impact of CO_2 content was found to decrease ethane recovery.

Konukman and Akman(2005)analyzed flexibility and operability of a HEN-integrated natural gas expander plant, using HYSYS process simulation. They showed that HEN synthesis design and flexibility analysis, based on nominal values, would provide incorrect results due to significant changes of physical properties. Konukman and Akman concluded the difficulty or impossibility of automated HEN synthesis for a highly energy-integrated plant like the turboexpander process using process flowsheet simulators.

Mehrpooya et al. (2006) simulated an existing NGL recovery unit using HYSYS. Two modifications were considered suitable for optimization: turboexpander, and turboexpander exchanger configurations to find the best revamping alternative. A genetic algorithm that used for optimization to ensure the maximum profit achieved is global and not only local.

Five different turboexpander ethane recovery processes were compared by Chebbi et al. (2004), and it was shown that more complex processing scheme may yield the same or less ethane recovery. Two very common processes were considered in Chebbi et al. (2008): the conventional turboexpander process and the GSP which adds reflux to the conventional process. Several other processes include modifications of GSP as we have mentioned above. Considering four different gas feeds and four different demethanizer pressures, and varying all available design variables, Chebbi et al. (2008) showed that the conventional turboexpander process yields larger maximum ethane recovery in most cases except when both the feed is lean and the demethanizer pressure is low.

In the present work, we consider the effect of demethanizer pressure on optimal ethane recovery level selection to achieve the maximum profit in a conventional turboexpander process. Optimization is performed by changing all the design variables for each value of the demethanizer pressure, while satisfying process constraints. In contrast to the works in Bandoni et al. (1989) and Diaz et al. (1997), the process simulation does not start at the cold tank, and does include the whole NGL recovery unit. Recommendations are given based on the present results.

2. Simulation and optimization

As described in Chebbietal.(2008),in the conventional process considered (Fig. 1 in Chebbi et al. (2008)), the feed is first cooled while providing the reboiler duty, then cooled further by heat exchange with the residue gas, followed by heat exchange with propane in a chiller to reach a temperature of −31◦F. After separation from the liquid, the gas leaving the separator is cooled by heat exchange with the overhead stream leaving the demethanizer, then separated from the liquid in the cold separator, expanded in a turboexpander and sent to the demethanizer. The liquids leaving the two separators are expanded through JT expansion and sent to the demethanizer at lower levels. The turboexpander provides part of the power needed to recompress the residue gas. The other portion is provided by a recompressor to raise the residue gas pressure to 882psia.

The temperatures after cooling by heat exchange with the residue gas leaving the demethanizer are design variables that are changed in the optimization process to maximize ethane recovery for different demethanizer pressures. The constraints are conditions preventing temperature cross in these two heat exchangers.

Fig. 1 also shows the refrigeration loop. The chiller is common to both the turboexpander process and the refrigeration unit. A propane economizer cycle is selected (Manning and Thompson, 1991). The compression cost is smaller compared to propane refrigeration simple cycle since part of the vapor is obtained at an intermediate pressure, called economizer pressure after the first valve expansion (Manning and Thompson, 1991); therefore the economizer cycle requires less power of recompression.

The optimum economizer pressure achieving minimum total compression power selected in our process is given by Fraser (Manning and Thompson, 1991):

$$P_{econ} = P_{ch}\left(\frac{P_{cond.out}}{P_{ch}}\right)^{0.614} \qquad (1)$$

in which $P_{cond,out}$ and P_{ch} are the pressures of propane after condensation (achieved by air cooler in our case) and inside the chiller.

We consider two different feeds: A and D (Bandoni et al., 1989; Diaz et al., 1997). Their compositions in terms of mole fractions are given in Table 1. Feed A is a lean gas with 6% C_2+ content, and feed D is a rich gas with 30% C_2+ content. The four demethanizer pressures considered are 100, 215, 335,

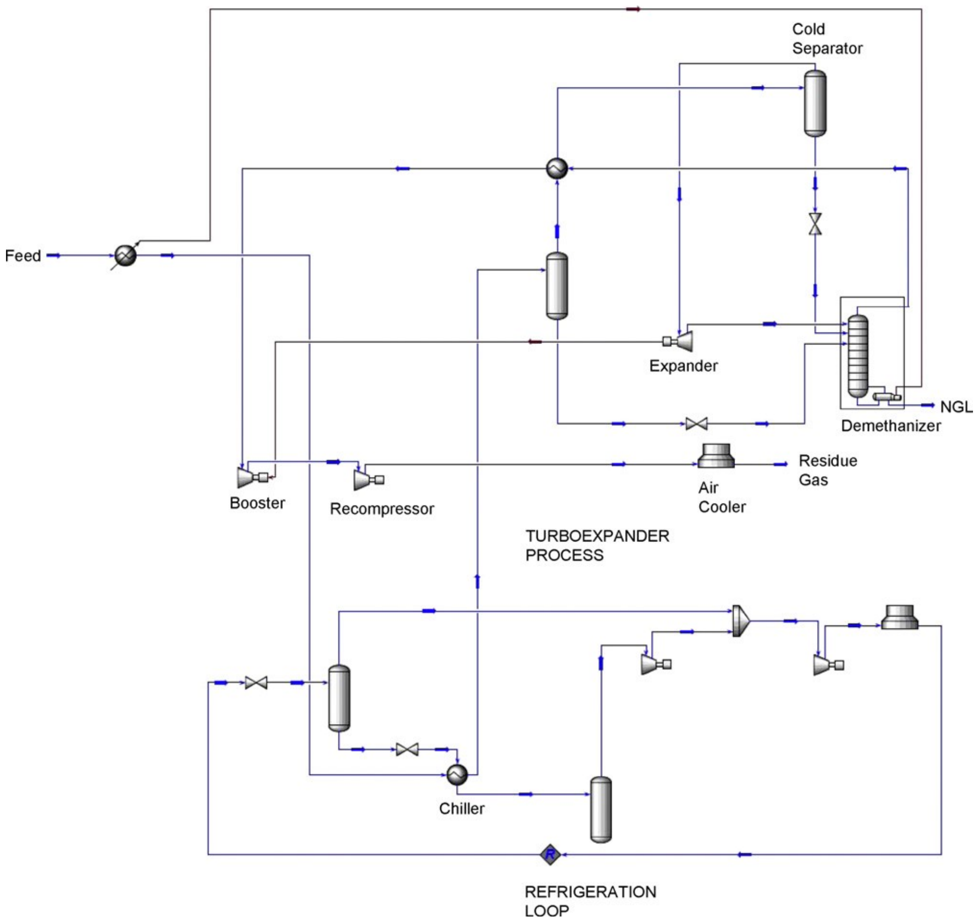

Fig. 1 – Conventional ethane recovery process with the propane refrigeration loop at low demethanizer pressure.

and 450psia as in Chebbi et al. (2008). In all cases, the feed gas is at 100◦F and 882psia, the residue gas is recompressed to 882psia, and in the NGL stream, the molar ratio of C_1 to C_2 is set at 0.02 (Manning and Thompson, 1991). Also in all cases the feed gas rate is 10,980lbmol/h as in Chebbi et al. (2008).

2.1. *Low demethanizer pressure*

The demethanizer pressure considered here is 100psia. The outlet temperature from the first gas-to-gas heat exchanger is a design variable that affects the cooling partition between the first gas-to-gas heat exchange and refrigeration.

As explained in the next section, the simulation and optimization study for this case shows that the profit is enhanced when the first gas-to-gas heat exchanger is discarded, with refrigeration providing the full cooling duty required to lower the feed gas temperature to −31◦F. The process flowsheet shown in Fig. 1 reflects the abovementioned process change.

Table 1 – Feed gas composition.

Component	A	D
Nitrogen	0.01	0.01
Methane	0.93	0.69
Ethane	0.03	0.15
Propane	0.015	0.075
Butanes	0.009	0.045
Pentanes	0.003	0.015
Hexanes	0.003	0.015
% C_2+	6	30

2.2. *Intermediate demethanizer pressure*

The pressure considered as intermediate is 215psia. As will be mentioned in the next section, it is more economical in this case to keep the first gas-to-gas heat exchanger to cool the feed gas before refrigeration to recover refrigeration from the residue gas leaving the demethanizer, after it exits the second gas-to-gas heat exchanger operating at low temperature. The process at intermediate demethanizer pressure corresponds tothetypicalflowsheetinwhichtwogas-to-gasheatexchangers are used (Fig. 1 in Chebbi et al. (2008)).

2.3. *High demethanizer pressures*

The pressures considered here are 335 and 450psia. As for the intermediate pressure case, the results show that a first gasto-gas heat exchanger is also economically justified and yields a higher profit. Another modification, dictated by thermodynamics, is required at high demethanizer pressures. As the demethanizer pressure increases, the temperature profile in the column is shifted up. The simulation results show that, at 335 and 450psia, the feed gas temperature is not high enough to provide heat in the reboiler. Thus, an external heat source has to be used in this case. The process flowsheet shown in Fig. 1 was modified to reflect the two abovementioned process changes.

3. Results and discussion

HYSYS simulation was performed with the Peng–Robinson equation selected as the thermodynamic model. The design variables are varied to maximize the profit P calculated in \$/year as

$$P = SG + SNGL - C_{RM} - COM_d \quad (2)$$

in which SG and SNGL represent residue gas and NGL sales, respectively, C_{RM} is the cost of feed gas (raw material), and COM_d is the cost of processing without depreciation. This cost is given for a typical chemical process as (Turton et al., 2003)

$$COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM}) \quad (3)$$

where FCI is the fixed capital investment, C_{OL} is the cost of operating labor, C_{UT} is the cost of utilities, C_{WT} is the cost of water treatment, and C_{RM} is the cost of raw materials. We define our objective function as

$$B = SG + SNGL - 0.180FCI - 1.23C_{UT} \quad (4)$$

The operating labor and feed gas costs, C_{OL}, and C_{RM}, are constant; therefore $P–B$ is constant, and in order to maximize the profit, we can simply maximize B.

The fixed capital cost is obtained using the Lang factor technique (Turton et al., 2003), and the purchase cost of the major equipment is obtained using the cost correlations in Turton et al. (2003).

3.1. High NGL to gas price ratio

To determine residue gas and NGL sales, the output from HYSYS is used along with the following assumed prices: 8\$/MMBtu and 1.53\$/US gallon. The ratio in this case is high: 0.191(\$/USgallon)/(\$/MMBtu).The utility is assumed to be provided by the residue gas. If more ethane is extracted from the feed gas, it is clear that this will result in less residue gas sales, and more NGL sales. The basis used for updating the costs is the chemical engineering plant cost index (CEPCI) for January 2008.

For the low demethanizer pressure case (100psia), optimization shows that the first gas-to-gas heat exchanger between the feed gas and the overhead stream from the demethanizer should be discarded in order to maximize profit, which means that the refrigeration duty needs be higher to achieve the same final temperature, −31°F, after chilling.

Table 2 – Results from optimization for feed D (high NGL to gas price ratio).

P (psia)	C_{UT} (MM\$)		FCI (MM\$)		Total FCI (MM\$)	Objective function (MM\$)	Ethane recovery (%)
	Main process unit	Refrigeration cycle	Main process unit	Refrigeration cycle			
100	4.62	3.61	18.2	13.4	31.6	657.1	91
215	3.35	2.11	10.7	9.62	20.4	653.9	85.3
335	3.9	3.43	9.17	13	22.1	637.3	76.2
450	3.41	3.39	7.08	12.9	20	619.1	64.3

3.1.1. First gas-to-gas heat exchange versus additional refrigeration load

The optimization results show that for the 100psia demethanizer pressure case, the first gas-to-gas heat exchanger needs to be discarded in order to maximize the profit.

Although this seems to be contradictory because of less refrigeration recovery and more load on the refrigeration unit in this case, the overall effect can be determined by considering the impact on the main process side as well. In order to explain the apparent contradiction, we consider as an example two cases in which the overall duty (needed to lower the temperature of the feed gas to the same temperature of −31°F) is shared equally between the first gas-to-gas heat exchanger and the chiller (refrigeration unit) in the first case, and shared as 40% for the refrigeration part in the second case, with more refrigeration recovery (60% of the load) provided by the residue gas in the first gas-to-gas heat exchanger. Tables A1–A3 in the appendix show the effect of changing the partition from 50–50% (case 1) to 60–40% (case 2). Also included in the appendix are the equations that illustrate how most CBM values for case 2 can be obtained simply from those calculated for case 1. Although the partition change results in a reduction of the refrigeration cost as expected, the process side is affected adversely: more gas-to-gas heat exchange yields a higher residue gas temperature at the inlet to the booster, and therefore a higher cost of compression (in the booster) and air cooling. At intermediate and high demethanizer pressures, the compression cost (process side) is less (compared to the 100psia pressure case) and the simulation study shows that, for a given intermediate (215psia) or high demethanizer pressure (335 or 450psia), higher profit is achieved by having a gas-to-gas heat exchanger for refrigeration recovery before the chiller in addition to the low temperature gas-to-gas heat exchanger shown in Fig. 1.

3.1.2. Maximum profit

Table 2 clearly shows that selecting lower demethanizer pressures yields higher profit for feed D, which is rich. The higher NGL recovery offsets the additional cost of compression at low pressure. For feed A, which is lean, there is an optimum demethanizer pressure at which the objective function, and therefore the profit, is at a maximum (see Table 3). Compared to feed A, the profits are substantially higher in the case of feed D

Table 3 – Results from optimization for feed A (high NGL to gas price ratio).

P (psia)	C_{UT} (MM\$)		FCI (MM\$)		Total FCI (MM\$)	Objective function (MM\$)	Ethane recovery (%)
	Main process unit	Refrigeration cycle	Main process unit	Refrigeration cycle			
100	4.91	1.63	24.1	7.25	31.3	353.2	92.6
215	3.77	0.7	14.4	4.18	18.6	354.9	82.4
335	3.03	1.24	10.8	6.1	16.9	349.9	65.4
450	2.03	1.23	8.2	6.05	14.3	344.7	45.9

(due to the substantially higher NGL content) with comparable capital costs at low and intermediate (100 and 215psia) demethanizer pressures.

3.1.3. Overall capital cost

For feed A, as the demethanizer pressure decreases, the overall capital cost increases, which is expected due to higher recompression cost needed to raise the residue gas pressure for the purpose of transportation. For feed D, the overall capital cost increases as pressure increases from 215 to 335psia. Although the recompression cost is lower at higher demethanizer pressure, the refrigeration load is higher due to a lower potential of cooling for the residue gas which leaves the demethanizer at a higher temperature. Another reason for the increase of the capital cost at higher demethanizer pressures is the necessity of an external heat source.

3.1.4. Refrigeration costs

The results in Table 3 clearly show that refrigeration is less costly (both in terms of capital cost and utility) in the 215psia case compared to both 335 and 450psia cases, which is expected since the residue gas reaches the first gas-to-gas heat exchanger at a lower temperature, and has a larger potential to cool the feed gas, and as a result, the load on the refrigeration unit is reduced. We would expect the refrigeration cost to be lower in the 335psia case compared to the 450psia one for the same reason; never the less another factor should be taken into account. At these high demethanizer pressures, the temperatures of the residue gas leaving the demethanizer become closer, making the difference in potential for cooling less; however, the heat capacity of the residue gas, is larger for the 450psia case due to the fact that more ethane and heavier components are left in the residue gas, and this tends to give more potential for cooling. Simulations show that the second effect (higher heat capacity) slightly surpasses the first effect (lower residue gas temperature) at these high demethanizer pressures (335 and 450psia).

The refrigeration cost(both in terms of capital cost and utility) is the highest in the 100psia case, due to the fact that the first gas-to-gas heat exchanger is eliminated, which adds the corresponding cooling load to the refrigeration unit through the chiller.

3.1.5. Main process capital and utility costs

For both feeds A and D, the main process capital cost decreases as the demethanizer pressure increases due to lower recompression cost (see Tables 2 and 3). The same trend is noticed for the utilities cost for all demethanizer pressures in the case of feed A and for low and intermediate demethanizer pressures (100 and 215psia) in the case of feed D. At high pressures (335 and 450psia), the utility cost needs to include the additional

Table 4 – Results for maximum ethane recovery in the case of feed D (high NGL to gas price ratio).

	Maximum ethane recovery (%)	Objective function (MM$
91.0	657.1 85.3 653.9 76.3 637.3	
64.3		619.1

external heating utility cost, which is by far larger in the case of the rich feed D. For this reason the utility cost rises to a higher value instead of declining as for feed A when the demethanizer pressure increases from 215 to 335psia.

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Table 5 – Results for maximum ethane recovery in the case of feed A (high NGL to gas price ratio).

	Maximum ethane recovery (%)	Objective function (MM$)
94.1	353.1 82.4 354.9 65.4 349.9	
45.9		344.7

3.1.6. Comparison between optimum and maximum recovery cases

Using the design variables values obtained to get the maximum ethane recovery for each demethanizer pressure in Chebbi et al. (2008) yields the profit at maximum recovery. Results are shown in Tables 4 and 5 for feeds A and D, respectively.

Comparing results from Tables 2 and 4 clearly shows that the optimum occurs at the maximum ethane recovery for feed D, which is rich. For feed A, which is lean, the same conclusion is reached at intermediate and high demethanizer pressures (see Tables 3 and 5); however, even at low demethanizer pressure, 100psia, the difference is not significant; and operating at the highest ethane recovery level would give a profit nearly equal to the optimum, along with a higher ethane recovery (1.5% difference at the lowest demethanizer pressure 100psia).

The trends are better illustrated in Figs. 2–5, in which the difference in profit and ethane recovery are plotted versus demethanizer pressure.

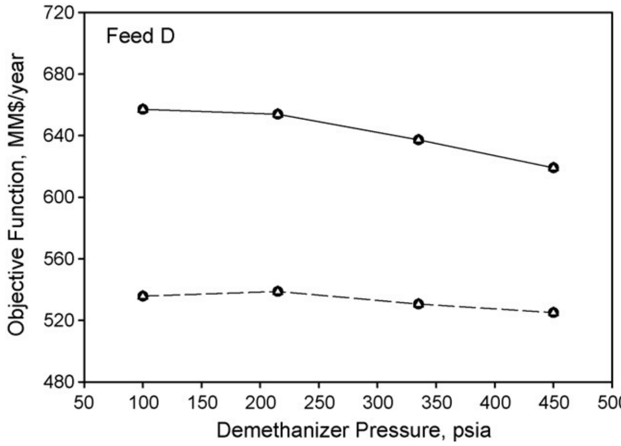

Fig. 2 – Objective function as a function of demethanizer pressure for feed D (circle: optimum value, triangle: value corresponding to maximum ethane recovery, continuous curve: high NGL to gas price ratio, dashed curve: low NGL to gas price ratio).

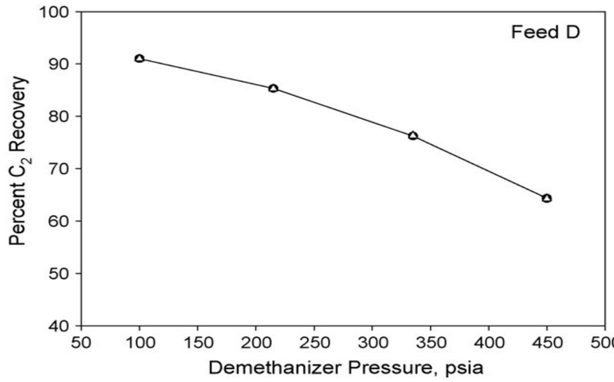

Fig. 3 – Objective function as a function of demethanizer pressure for feed A (circle: optimum value, triangle: value corresponding to maximum ethane recovery, continuous curve: high NGL to gas price ratio, dashed curve: low NGL to gas price ratio).

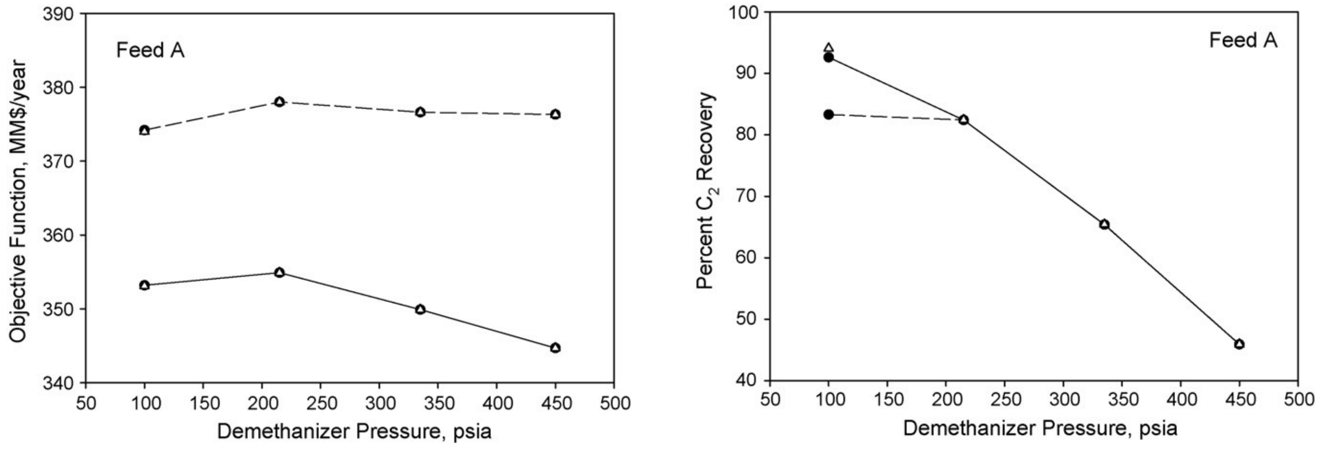

Fig. 4 – Percent C_2 recovery as a function of demethanizer pressure for feed D (circle: optimum value, triangle: maximum ethane recovery, continuous curve: high NGL to gas price ratio, dashed curve: low NGL to gas price ratio. The dashed curve coincides with the continuous one).

3.2. *Low NGL to gas price ratio*

The simulations were repeated with the following prices: 9.60$/MMBtu and 1.00$/US gallon. The ratio in this case is low: 0.104($/US gallon)/($/MMBtu). Only differences or important similarities with respect to the previous case (high NGL to gas price ratio) will be discussed in this part. The compared results can be seen in Figs. 2–5 introduced previously.

3.2.1. *Profit*

The results obtained are shown in Tables 6 and 7. In all cases, the objective function is seen to show a peak, with an optimum pressure at which the profit is maximum. This was not thecaseinpart1(high NGL to gas price ratio)for rich feed D, for which producing the maximum NGL, by operating at the lowest pressure 100psia, yields the highest profit, a likely result at high NGL prices for a rich gas. At low NGL to gas price ratio, the values of the objective function, and therefore the profit, are smaller in the case of feed D, which is expected since the gas is rich, and NGL price is low. On the contrary, the profits are higher in the case of feed A, which is also anticipated since the feed is lean (with a significantly smaller NGL production rate), and gas prices are high. Considering feeds A and D separately, the objective function values are relatively close for the different demethanizer pressures at low NGL to gas price ratio.

3.2.2. *Main process and refrigeration utilities costs*

Compared to the high NGL to sales gas price ratio case, the utilities costs are higher in all cases due to the higher gas price except for feed A at low demethanizer pressure (100psia), for which the refrigeration utility cost increases but the main process utility cost declines. Optimization yields an increase in the outlet temperature of the feed gas exiting the cold gas to-gas heat exchanger shown in Fig. 1 (only one such heat exchanger is included at 100psia, for the reasons mentioned above), which enhances the power of the turboexpander, and reduces the load on the booster, and therefore the utility cost on the process side.

Table 6 – Results from optimization for feed D (low NGL to gas price ratio).

P (psia)	C_{UT} (MM$)		FCI (MM$)		Total FCI (MM$)	Objective function (MM$)	Ethane recovery (%
	Main process	Refrigeration unit cycle	Main process	Refrigeration unit cycle			
100	5.55	4.32	18.2	13.4	31.6	535.7	91
215	4.02	2.54	10.7	9.62	20.4	538.7	85.3
335	4.69	4.11	9.17	13	22.1	530.6	76.2
450	4.1	4.07	7.08	12.9	20	525	64.3

Table 7 – Results from optimization for feed A (low NGL to gas price ratio).

P (psia)	C_{UT} (MM$)		FCI (MM$)		Total FCI (MM$)	Objective function (MM$)	Ethane recovery (%)
	Main process unit	Refrigeration cycle	Main process unit	Refrigeration cycle			
100	4.16	2.43	23.3	8.31	31.6	374.2	83.3
215	4.52	8.4	14.4	4.18	18.6	378	82.4
335	3.64	1.49	10.8	6.1	16.9	376.6	65.4
450	2.44	1.47	8.2	6.05	14.3	376.3	45.9

Table 8 – Results for maximum ethane recovery in the case of feed D (low NGL to gas price ratio).

Maximum ethane recovery (%)	Objective function (MM$)
91.0	535.7
85.3	538.7
76.2	530.6
64.3	525.0

Table 9 – Results for maximum ethane recovery in the case of feed A (low NGL to gas price ratio).

Maximum ethane recovery (%)	Objective function (MM$)
94.1	374.0
82.4	378.0
65.4	376.6
45.9	376.3

3.2.3. Capital costs

The main process and refrigeration unit capital costs are the same as for the high NGL to sales gas price ratio case, except for feed A at demethanizer pressure equal to100psia.Thisspecial case is the only case in which the design variable (temperature of the feed gas exiting the cold gas-to-gas heat exchanger) at optimum conditions does not correspond to the limiting case where the design variable is set just to avoid temperature cross in the gas-to-gas heat exchanger. In all other cases, equipment sizing is the same, and equipment prices are considered not affected by the relative costs of NGL and sales gas. As indicated in the previous part, there is less load on the booster in this special case, and this yields not only a decrease of the utility on the process side, but also a drop in the capital cost compared to the high NGL to sales gas price ratio case.

3.2.4. Ethane recovery

The results are shown in Tables 6–9. As for the high NGL to sales gas price case, for a given demethanizer pressure, the optimum is found to occur at maximum ethane recovery except for the lean feed A at low demethanizer pressure, 100psia. The deviation from maximum ethane recovery is significantly higher for the low NGL to sales gas price case (see Tables 7 and 9). This result is likely since it yields more ethane left in the residue gas, at optimum condition, in the case of higher sales gas price. However for both prices ratios, the maximum profit (optimum) is only slightly higher than the profit at maximum recovery.

4. Conclusions

The present work shows that in all studied cases except one, there is an optimum demethanizer pressure at which the profit is maximum. The exception corresponds to rich gas D and high NGL to gas price ratio. The optimum is reached in this special case at the lowest demethanizer pressure, 100psia. In contrast to the works in Bandoni et al. (1989) and Diaz et al. (1997) the process simulation does not start at the cold tank, and does include the whole NGL recovery unit in our analysis. In addition, the demethanizer pressure range is wider in our work and does cover the typical range indicated in Manning and Thompson (1991). Recovery is found to be adversely affected at higher demethanizer pressures as expected. For a given demethanizer pressure, the recovery at optimum conditions is found to be equal to the maximum recovery at the specified pressure, except for feed A at low demethanizer pressure (100psia). In addition, the maximum profit (optimum) is found equal to the profit at maximum ethane recovery except for the special case of feed A at low demethanizer pressure for which the optimum profit and value at maximum ethane recovery are found not equal but very close.

Optimization shows that the use of a gas-to-gas heat exchanger before the chiller to recover refrigeration yields lower profit at low demethanizer pressure (100psia), and thus should be discarded at low pressure. At high demethanizer pressures (335 and 450psia), the feed gas cannot be used to provide the reboiler duty. Therefore, less cooling (refrigeration recovery) can be achieved, and an external heat source is required. At intermediate pressure (215psia), the main process is a conventional turboexpander unit with two gas-to-gas heat exchangers, and where the demethanizer reboiler duty is provided by the feed gas, which results in further cooling before refrigeration. The costing structure is relatively complex and accounts for the overall cost including the interaction between the main process part and the refrigeration unit through the chiller. Details are given to explain the costing structure. Due to fluctuations in the prices of NGL and sales gas, it is essential, for the purpose of optimization, to have a control system capable of adapting to changes needed in the design variables (including the demethanizer pressure) for optimization, along with a flexible process that encompasses possible required changes in terms of refrigeration recovery as the demethanizer pressure changes as discussed in the present paper.

Acknowledgement

This work was supported by the American University of Sharjah under grant no. FRG07-040.

Appendix A. Appendix

The relevant bare module equipment costs (CBM) (Turton et al., 2003), utility costs, and total annual costs are compared in Tables A1–A3. The CBM values for case 2 can be obtained directly as for case 1; however most of those values can be obtained by simply using the equations given below.

The utilities in Table A1 are proportional to the compressor duties, therefore we have

The mass flow rate of the refrigerant, propane, is obtained as

$$\dot{m}_2 = \frac{0.4}{0.5}\dot{m}_1 \qquad \text{(A2)}$$

The intensive properties inside the propane loop remain the same. The utilities cost on the refrigeration side can be simply obtained as

$$\text{Utility}_2 = \frac{0.4}{0.5}\text{Utility}_1 \quad \text{(refrigerationloop)} \qquad \text{(A3)}$$

The sizes of the compressors and their drives in the refrigeration loop can be found as

$$S_2 = \frac{0.4}{0.5}S_1 \qquad \text{(A4)}$$

Table A1 – Compared annual utility cost estimates for two different cooling partitions between gas-to-gas heat exchange and refrigeration.

Equipment	Refrigeration part: 50%		Refrigeration part: 40%	
	Duty (Btu/h)	Duty ($/year)	Duty (Btu/h)	Duty ($/year)
Refrigeration				
First compressor	7,635,000	535,100	6,108,000	428,100
Second compressor	7,298,000	511,400	5,838,000	409,100
Total (refrigeration)	14,933,000	1,046,500	11,946,000	837,200
Main Process (equipments affected by the partition change)				
Compressor (booster) (2)	70,437,000	4,936,200	75,224,000	5,272,000
Total	85,370,000	5,982,700	87,170,000	6,109,000

Table A2 – CBM estimates compared for two different cooling partitions between gas-to-gas heat exchange and refrigeration (the only equipment included in the total cost are those affected by the partition change).

Equipment (number of units if more than 1)	Refrigeration part: 50%		Refrigeration part: 40%	
	Size per unit	CBM ($)	Size per unit	CBM ($)
Refrigeration				
First separator	20.0ft^3	13,500	14.3ft^3	11,800
Second separator	250.6ft^3	95,500	179.4ft^3	77,800
Chiller (2)	820ft^2	1,461,400	750ft^2	1,342,500
First compressor	1,606,000Btu/h	1,454,400	1,285,000Btu/h	1,210,800
Drive for the first compressor	2,294,000Btu/h	343,700	1,835,000Btu/h	287,300
Second compressor	1,535,000Btu/h	680,900	1,228,000Btu/h	566,400
Drive for the second compressor	2,192,000Btu/h	331,400	1,754,000Btu/h	277,000
Air cooler	3189ft^2	241,300	2552ft^2	217,100
Total CBM (refrigeration)		4,622,100		3,990,700
Main process (equipment affected by the partition change)				
First gas-to-gas heat exchanger	786ft^2	183,200	1054ft^2	212,800
Compressor (booster) (2)	7,936,000Btu/h	4,610,900	8,475,000Btu/h	4,817,400
Drive for the booster (2)	10,581,000Btu/h	2,319,100	11,300,000Btu/h	2,442,600
Air cooler	3278ft^2	261,300	3525ft^2	270,700
Total CBM (process equipment affected by the partition change)		7,374,500		7,743,500
Total		11,996,600		11,734,200

Table A3 – Compared annual cost for two different cooling partitions between gas-to-gas heat exchange and refrigeration (the only equipment included in the total cost are those affected by the partition change).

Cost	Refrigeration part: 50%	Refrigeration part: 40%
0.18FCI+1.23C_{UT} ($/year)	9,906,800	10,006,400

The volumes of the two separators in the refrigeration loop can be obtained using

$$\frac{V_2}{V_1} = \frac{A_2 L_2}{A_1 L_1} = \frac{A_2 D_2}{A_1 D_1} = \left(\frac{A_2}{A_1}\right)^{3/2} = \left(\frac{Q_2/u_a}{Q_1/u_a}\right)^{3/2} = \left(\frac{\dot{m}_2}{\dot{m}_1}\right)^{3/2} = \left(\frac{0.4}{0.5}\right)^{3/2} \tag{A5}$$

where *V*, *A*, *D* and *L* represent the volume, the cross-sectional area, diameter and length of the vessel, respectively, and u_a the maximum allowable vapor velocity.

The sizes of the booster and the first gas-to-gas heat exchanger cannot be calculated directly from the values obtained for case 1 since there are changes in relevant intensive properties. The same is also valid for the chiller since it has one stream on the process side.

The size of the equipment being found, and the bare module factors being the same for cases 1 and 2, the bare module equipment cost, CBM_2, can be obtained using

$$CBM_2 = CBM_1 \frac{10^{K_1+K_2\ \log_{10}(S_2)+K_3[\log_{10}(S_2)]^2}}{10^{K_1+K_2\ \log_{10}(S_1)+K_3[\log_{10}(S_1)]^2}} \tag{A6}$$

where the *K* values, for the corresponding equipment, are given in Turton et al. (2003).

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