ETHANE PRODUCTION PLANT FOR BETTER ENERGY INTEGRATION AND COST REDUCTION IN JORDAN

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ABSTRACT

The recovery of ethane from natural gas is considered to be more economically feasible due to shift in ethane prices and advances in process modifications. Process design and retrofitting of ethane recovery technologies are reviewed. A viable process design for the production of ethane using Qatari natural gas as feedstock in Jordan is proposed. The plant capacity is determined to be 8.83*10⁹ scf per year at purity of 98.6 percent. Process design and energy integration are performed using CHEMCAD and HENSAD simulation software. The optimal process flow diagram (in terms of energy integration) consists of three major sections: Pretreatment section, liquefaction and cryogenic separation of methane and ethane. By using parametric studies, the major operating and design variables are optimized. The simulation results show that the ethane recovery plant proved be able to produce ethane and methane as a byproduct at a purity up to 98.6 and 99.7, respectively. Moreover, preliminary economic evaluation (profitability analysis) based on discounted and not discounted methods is performed. It results in an average ROROI of 16 percent value, and a nondiscounted payback period of 4.5 years.

Keywords: ethane recovery, natural gas, process synthesis, process integration, computer-aided flowsheeting, economic evaluation.

INTRODUCTION

Jordan has very limited energy resources of fossil fuels. Iraq supplies Jordan with 10,000 barrels of crude oil a day, but still not enough. Jordan’s government forecasts that domestic fuel consumption will increase to 10 million metric tons per year of oil equivalent by 2020, up one third from today’s 7.5 million tons. The government expects electricity consumption to almost double within 10 years to 5 200 megawatts. Natural gas is considered another viable energy option and there are two available sources; (a) Al-Risha well near the Jordan-Iraq common border, with ethane content of only 1 %, so the low capacity of this well, results in excluding this source, (b) the Egyptian natural gas, which contains 4% ethane and provides Jordan with most of the energy
it needs for the last couple of years. The Arab Gas Pipeline is a natural gas pipeline in the Middle East and delivers Egyptian natural gas to Jordan, Syria, and Lebanon. The pipeline has a total length of 1200 km. But due to the present Arab Spring which started in December 2010 and the Egyptian revolution, this natural gas pipeline in the Sinai Peninsula was bombed more than fifteen times which affected the stability of natural gas supplies to Jordan. These actions had increased Jordan’s energy bill which hit the economy badly. Therefore serious thinking of replacing this option to a more stable energy source and explore other long-term alternatives is an obvious need. The Qatari natural gas seems to be the best alternative for Jordan as a profitable project since it is rich in ethane with a content of about 13 %, more over the negotiations and studies to import the gas from Qatar are being seriously taken into account. The high ethane percentage in the natural gas is a crucial factor in any plant feasibility analysis. Even with an acceptable margin of profit, the production of Ethane in Jordan will have direct impact on the country’s economy through ensuring stable energy policies for the industry and on both the strategic and political levels where the plant will contribute towards minimizing the high rate of unemployment.

Natural gas releases a high amount of energy upon combustion with few emissions. Therefore, natural gas is an indispensable source of energy for most industries, and electricity generation, and has played a major role in the industrial revolution and in raising the life standards. The composition of natural gas varies depending on the geology of the ground, the location of the well and its depth. The Qatari gas has the composition shown in Table 1.

Recently software simulators have played an important role in synthesis and design of natural gas plants. Also, they are used in technically assessing various overall design scenarios and optimizing operating conditions to reduce the overall cost [2-5]. Konukman and Akkaya [6], minimized the exergy destruction for a turboexpander plant by integrating Aspen Plus process simulator and the optimization-related features of MATLAB. Whereas, Tirandazi et al. [7], simulated an ethane recovery unit with its refrigeration cycle and the exergetic efficiency of the refrigeration cycle is determined to be 43.45 %, indicating a great potential for improvements. Most of the ongoing research is concentrated on efficient utilization of Liquefied Natural Gas (LNG) in a cryogenic process and to separate light hydrocarbons (C2+) from LNG with low power consumption [8]. The capital cost and potential revenue from LNG extraction can have a significant impact on the overall economics of an LNG project. An economy based model uses several online parameters from upstream and downstream plants to minimize C2+ loses and recover C3+ NGL Plants are described [9].

The main aims of this work are to review the available ethane extraction technologies and address various process design scenarios for an ethane production plant from natural gas. An optimized design would be selected which will be located in the Hashemite Kingdom of Jordan. The selected process design should satisfy the targets of maximum heat integration, low capital cost with the country’s needs and suitability. This is carried out through process synthesis and analysis, computer-aided flow sheeting, parametric studies of the basic flowsheet and feasibility studies. The use of simulation

<table>
<thead>
<tr>
<th>Component</th>
<th>Formula</th>
<th>Composition (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Methane</td>
<td>CH₄</td>
<td>76.6</td>
</tr>
<tr>
<td>Ethane</td>
<td>C₂H₆</td>
<td>12.59</td>
</tr>
<tr>
<td>Propane</td>
<td>C₃H₈</td>
<td>2.38</td>
</tr>
<tr>
<td>i-Butane</td>
<td>C₄H₁₀</td>
<td>0.11</td>
</tr>
<tr>
<td>n-Butane</td>
<td>C₄H₁₀</td>
<td>0.21</td>
</tr>
<tr>
<td>Pentane</td>
<td>C₅H₁₂</td>
<td>0.02</td>
</tr>
<tr>
<td>Nitrogen</td>
<td>N₂</td>
<td>0.24</td>
</tr>
<tr>
<td>Hydrogen Sulfide</td>
<td>H₂S</td>
<td>1.02</td>
</tr>
<tr>
<td>Water</td>
<td>H₂O</td>
<td>6.83</td>
</tr>
</tbody>
</table>

Table 1. Qatari Natural Gas Compositions [1].
software to assess the design of the ethane recovery process is highlighted. The Qatari gas, which is rich in ethane 13 % is chosen as a feed stock for the proposed plant capacity of $8.83 \times 10^9$ scf per year ethane with purity about 98.6 %.

**ETHANE EXTRACTION TECHNOLOGIES**

Ethane was typically not separated from the methane component of natural gas, but simply burnt along with the methane as a fuel. With technology advancement [10], a widespread industrial production of ethane, which is used as a chemical feed stock to produce ethylene, created a high and continuous demand on ethane as a raw material.

The common ethane recovery technologies are summarized as follows: (a) Steam cracking is one of the well-established technologies to produce ethane from naphtha and heavy fuel. It consists of a pyrolysis furnace, followed by fractionation and compression stages and finally - the product recovery stage [11]. (b) The adsorption process offers another available alternative. In this process, ethane is extracted from a natural gas mixture at $25^\circ C$ and 101.3 kPa using a packed bed adsorber. Ethane is completely separated from the mixture, however, after a prolonged period. The adsorbent must be large-pored and much larger than the kinetic diameters of ethane (4.44 Å). The advantage of using adsorption to recover ethane from natural gas is that the separation conditions are likely to be not extreme [12]. The main disadvantage is the low recovery of ethane (50 %) if carbon is used as the adsorbent. (c) Membrane technology is applied to separation of ethane on the basis of the difference in its rate of permeation through the membrane. The recovered ethane (the permeate) is never 100 % pure because of finite partial pressure difference. So ethane can be further separated by multiple stages. The membrane permeation has many advantages, which include low capital investment, ease of operation, good weight and space efficiency, no moving parts flexibility, low environmental impact, ease of installation and reliability. The principle disadvantage is that a clean feed is required, leading to the use of filtration to remove particles down to one micron in size [13]. (d) Cryogenic processes generally involve cooling a natural gas stream to temperatures near -85°C. These low temperature requirements have high associated energy consumption costs [14]. Ethane is most efficiently separated from methane by liquefying at cryogenic temperatures. The most economical process presently in wide use employs turbo expansion which can recover over 90 % of the ethane in natural gas [15-17]. There are certain guidelines that should be taken into consideration when choosing between cryogenics or absorption gas processing systems [18]. A pilot plant technology based on the adiabatic cooling of swirling gas flow in a supersonic nozzle which is effective in separating and processing natural gas components was developed [19]. To be competitive with liquid based crackers, ethane must be extracted at the lowest marginal cost. Barthe, and Gahier [20] presented a discussion, which covers the new processes that are being applied in different gas and LNG projects.

Recently, retrofitting an existing NLG plant to increase ethane recovery is considered [21-25]. Retrofitting typically requires modifications to the process equipment and expansion of the existing refrigeration system [26]. The operating conditions and the effect of column pressures, minimum approach temperature in the LNG exchanger, and split fraction have direct impact on maximizing ethane yield [27].

**SYNTHESIS OF ETHANE RECOVERY PROCESS**

Process synthesis and design can be categorized into two groups: Optimization based and knowledge-based methods. In optimization based methods an explicit or implicit flowsheet superstructure is required. On the other hand, knowledge-based (heuristic) methods concentrate on the representation and knowledge organization of the design problem. In the synthesis of ethane recovery process, we used an heuristic-based approach rather than rigorous structural optimization. Therefore, we followed the same lines of the Onion model [28,29]. In ethane recovery from natural gas, the synthesis of the basic flowsheet consists of three major stages: (a) The pretreatment “purification” step in which the gas pretreatment involves the removal of hydrogen sulfide (H$_2$S), carbon dioxide (CO$_2$), water, nitrogen and some other possible impurities to sufficiently low levels to meet the specifications and permit additional processing in the plant without any problems, (b) The liquefaction step-natural gas liquefaction is required whenever cryogenic separation is applied. Separation
of hydrocarbons generally occurs by the creation of a second phase. Since hydrocarbons (in natural gas) are gases at normal conditions and their boiling points are sufficiently low, a second phase (liquid phase) is created by liquefying part of the gas. Cryogenic separation relies on the fact that heavier components tend to be in the liquid phase more than lighter ones and so liquefy more rapidly, and finally (c) the separation step—the separation can be achieved by different techniques, such as lean oil absorption, and cryogenic fractionation. In natural gas industry, lean oil absorption is not widely employed, while cryogenic fractionation is the most common technique because of its high separation efficiency especially when high recovery rates are required, as in the present case [14]. Cryogenic fractionation simply employs a series of distillation columns that operate at very low temperatures to achieve the desired separation. The synthesis of the required process is based on the nature of the feed stock, which is the Qatari natural gas that contains the highest ethane content among other natural gases.

As mentioned above, a basic process flow diagram is developed using the onion model procedure. At each layer synthesis decisions may require analysis on the individual unit level and/or the flowsheet as a whole. These analyses are performed using a process simulation tool. The simulation software “CHEMCAD” was used in the synthesis and analysis of the ethane recovery process. It helped in optimizing the flowsheet configuration, by eliminating infeasible options. The “HENSAD” Software [28] was used to improve the flowsheet economics through implementing heat exchangers network synthesis, analysis and design and heat integration based on pinch analysis.

### Computer-Aided Flowsheeting and Sensitivity Analysis

Sensitivity analysis is used to estimate optimum process conditions, which result in optimum design in terms of desired product yield, and by minimizing utilities (steam & cooling water) consumption. The sensitivity analysis is performed by changing critical process parameters (input conditions), which have major impact on process output variables. With the use of the computer-aided flow sheeting CHEMCAD package, the possibility to test different process configurations, different thermodynamics packages and equipment

<table>
<thead>
<tr>
<th>Stream No.</th>
<th>Mass Flow (kg/s)</th>
<th>Cp (kJ/kgK)</th>
<th>Temp In (K)</th>
<th>Temp Out (K)</th>
<th>Stream Enthalpy (kW)</th>
<th>Film Transfer Coefficient (W/m²/K)</th>
</tr>
</thead>
<tbody>
<tr>
<td>22</td>
<td>19.75</td>
<td>4.0</td>
<td>473.0</td>
<td>313.1</td>
<td>12628</td>
<td>850.0</td>
</tr>
<tr>
<td>24</td>
<td>48.3</td>
<td>2.60</td>
<td>321.0</td>
<td>297.0</td>
<td>3013</td>
<td>30.0</td>
</tr>
<tr>
<td>27</td>
<td>48.3</td>
<td>2.75</td>
<td>297.0</td>
<td>244.1</td>
<td>7019</td>
<td>30.0</td>
</tr>
<tr>
<td>31</td>
<td>48.3</td>
<td>4.54</td>
<td>248.1</td>
<td>222.1</td>
<td>5701</td>
<td>30.0</td>
</tr>
<tr>
<td>34</td>
<td>34.25</td>
<td>2.49</td>
<td>349.0</td>
<td>340.0</td>
<td>767.5</td>
<td>30.0</td>
</tr>
</tbody>
</table>

Available Cumulative Hot Stream Energy = 31062.7 kW

<table>
<thead>
<tr>
<th>Stream No.</th>
<th>Mass Flow (kg/s)</th>
<th>Cp (kJ/kgK)</th>
<th>Temp In (K)</th>
<th>Temp Out (K)</th>
<th>Stream Enthalpy (kW)</th>
<th>Film Transfer Coefficient (W/m²/K)</th>
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</thead>
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<tr>
<td>45</td>
<td>34.24</td>
<td>4.8</td>
<td>151.0</td>
<td>161.0</td>
<td>-1643</td>
<td>850.0</td>
</tr>
<tr>
<td>42</td>
<td>33.27</td>
<td>3.0</td>
<td>207.0</td>
<td>222.7</td>
<td>-1996</td>
<td>1140</td>
</tr>
<tr>
<td>54</td>
<td>4.94</td>
<td>3.17</td>
<td>312.1</td>
<td>322.1</td>
<td>-156.5</td>
<td>1140</td>
</tr>
</tbody>
</table>

Available Cumulative Cold Stream Energy = -30780.7 kW

### Table 2. Hot streams extracted from the ethane recovery process (base case).

### Table 3. Cold streams extracted from the ethane recovery process (base case).
models can be easily performed. In the case of recovery of ethane from natural gas, the final ethane purity, which is produced from the Deethanizer, is optimized. The amine package is the default thermodynamics model, which is used in the simulation of the recovery of ethane process. This is because the present process contains a gas sweetening unit in which diethanolamine (DEA) is used to absorb H₂S gas. As an alternative to the amine thermodynamics package, the UNIFAC package and the SRK equation of state for the estimation of the k-values were employed for the enthalpy model. The SRK is known to predict accurately enthalpy changes and can quantitatively describe the phase equilibrium especially for light hydrocarbons [28].

The flowsheet is simulated following the process design hierarchy, where the recycle loops are first cut and initialized to recover the sequential modular approach. Once all the recycle calculations converged, sensitivity analyses were carried out on the flowsheet level. The results from the integrated process flowsheet are shown in Fig. 1. The energy integrated flowsheet was generated following the heat exchanger network design and analysis using the HENSAD software [28].

Firstly, the De-ethanizer is analyzed by investigating the effects of feed pressure and temperature on the ethane recovery. The first case is studied by varying the feed temperature of the Deethanizer at a given feed pressure. It is clear that as the feed temperature increases, the ethane purity increases too. At very low cryogenic temperatures, the purity of ethane is not affected at feed pressures in the range of 10 to 14 bar. On the other hand, at high feed pressures (18, 22, 26, 30 bars), the purity of ethane is found more sensitive to temperature changes. The purity of ethane in the overhead product increases by decreasing the feed pressure at constant temperature. On the other hand, the ethane purity decreases when the temperature is decreased at constant pressure. This can be attributed to the increase of the vapor fraction of the feed, where at some pressure and temperature, the feed composition is equal to that at the feed tray. Therefore, at moderate feed pressure and temperature an optimum ethane purity can be obtained. Accordingly, the best operating conditions used in the final simulation flowsheet, shown in Fig. 1, are 14 bar and −17°C. The temperature −17 °C is chosen for further heating the process stream in E-105 that is imposed by the demethanizer bottom stream temperature, which is −30°C. On the other hand, the pressure of the feed is found to affect the condenser duty, where the latter has a strong impact on the fixed capital cost of the plant (which increases the area of the condenser), and utility costs of the process. This is because the condenser needs a refrigeration cycle which is known to be a very expensive utility [28]. It is found that the minimum condenser duty is achieved at 14 bar. This results in a minimum fixed capital and operating costs, when compared to other operating pressures. The use of the amine fluid package has a substantial impact on the simulated results. This behavior is reported in many simulation case studies, where using different fluid packages can lead to different results [28].

Synthesis of the Heat Exchanger Network

The objective of the heat integration is to avoid the expensive refrigeration cycles that are required in the feed preparation of the separation section (liquefaction section), and the bottom stream of the DEA regeneration distillation column. Moreover, heat integration can help in the elimination of hot utility (steam) by replacing the usage of the low pressure steam in the deethanizer reboiler. This is done through the pinch analysis, which is an integral part of the overall strategy for process development, design and optimization (process synthesis) [29-32].

The heat integration and analysis were carried out using "HENSED" software [28]. The heat integration of the ethane production process is done systematically. Firstly, the hot and cold streams in the process are identified. Table 2 shows the hot streams in the process along with the necessary information that should be provided to the HENSAD software. Table 3 shows the cold streams in the process that is used to eliminate the cooling utilities (refrigerants, cooling water). A phase change occurs in streams (45 and 42 of Fig. 1), where a (10°C) of sensible heat to take into account the latent heat of the phase change in the streams, is used [28].

As an initial guess, the minimum approach temperature can be taken as 10°C for fluids, and 5°C for refrigerants [28]. In the present case, we selected ΔTmin = 10°C, as a compromise between fixed capital investment and utility costs [28]. The composite curve for the ethane recovery process is given in Fig. 2. It shows that the integrated ethane extraction process (Fig. 1) is a threshold problem for the hot utility side. At the design level, it was necessary to screen and eliminate some streams from
Fig. 1. Energy integrated process flow diagram for the ethane recovery process.
the pinch analysis of the process. Some of the major streams that are eliminated from the pinch analysis are:

Stream (19) of the integrated PFD due to its low flow rate, and its high content of H$_2$S, which is considered to be a highly corrosive material.

Stream (23) is also eliminated from the pinch analysis, where the stream temperature has to be reduced from 106 to 48°C. In order to achieve this temperature reduction, this stream needs about 7376 kW. The optimum way to do this is to use a high flow rate of cooling water with inlet temperature of 30°C and return temperature of 40°C, rather than cooling this stream with the available cold streams. This is because these streams are used for cooling other process streams, where cooling water is not possible.

The synthesis of the Heat Exchanger Network (HEN) based on the temperature interval diagram is performed using the HENSAD software. The results on Fig. 3 shows that the heat integrated process has only five heat exchangers (E-104, E-105, E-106, E-107, E-110) using a minimum approach temperature of 10°C.

The minimum approach temperature has a direct effect on the operating and capital cost of the heat exchanger network. An optimal minimum approach temperature is expected to minimize the total annualized cost. This is represented by the total Equivalent Annualized Operating Cost (EAOC) as function of the minimum approach temperature, which is shown in Fig. 4. It is clear that the EAOC curve as function of the minimum approach temperature is flat around the optimal value, which is around 25°C. At these optimal conditions, the operating cost of the heat exchanger network is found equal to 10.86 $ million/year, and the total operating cost of the process is 42.14 $ million/year (see the section for cost estimation). The heat integration achieved about 26% savings with respect to the total operating cost of the plant, which is a very important step in the phase of process synthesis and design.

The energy recovery is considered a crucial issue, which is achieved through heat integration, because of the rapid increase in the fuel costs, and the energy crisis. The heat integration results in a reduction of energy requirements by about 319.4 MW or 967980 GJ/year. This high energy comes from burning of fuels to generate steam for heating, and in the refrigeration cycles that consumes a huge amount of electricity, generated in gas turbines to drive the compressors of
the refrigeration cycles. On the other hand, the heat integration made the process more complex, since it created a more interactions between the process streams and major pieces of equipment. This complicates the control strategy and process start up and shutdown [29].

**Process Description**

Fig. 1 shows the energy integrated PFD for the ethane production process. The process shows that the sour natural gas in stream <1> is fed to the packed absorption tower (T-101), which contains Pall Rings as packing. The gas enters from the bottom at 30.7 bar and 35°C. Hydrogen sulfide (H\(_2\)S) is absorbed using a 30 mass % (6.8 mol %) aqueous solution of Di-Ethanol Amine (DEA). The solvent, stream <2>, flows at of 3,257 kmol/h. The solvent passes through the flash drum (V-102) where hydrocarbon traces are recovered in stream <13> and mixed with the methane product in stream <26>. The solvent stream <15> is regenerated in the stripping distillation column (T-102). The regenerated solvent becomes concentrated due to water losses, and hence a makeup water in stream <45> is used.

The sweet natural gas in stream <10> is dehydrated in the adsorption towers (T-103 A/B), which contain molecular sieves of average pore size equal to 4Å. After that, the dehydrated sweet gas in stream <21> enters the separation feed preparation stage where it is compressed, using the compressor (C-101) to 65 bar. This stream is then cooled in a heat exchangers network consisting of four heat exchangers in series (E-103, E-104, E-105, and E-106) to -51°C. The cooled gas in stream <35> enters the low-temperature separator (V-104), where it is split into two streams: the overhead vapor stream <36> and the liquid (condensed gas) stream <37>. The vapor stream is expanded in a turbo-expander (D-101) where its pressure is reduced to 14 bar. The liquid stream pressure is also reduced to 14 bars using a throttling valve. The two cold streams, <38> and <40>, are then sent to the Demethanizer (T-104) with two feeds, introduced at trays 4 and 15, respectively. Liquid methane is recovered as an overhead product in stream <45> and used as refrigerant for cooling the process gas in (E-106). This stream is then used to cool the regenerated solvent in (E-107), and the process gas in (E-104), where it is mixed with the hydrocarbon recovered in (V-102) and compressed to 30 bar using the compressor (C-102). This stream is then cooled to 76°C using the heat exchanger (E-110), before it is sent to the final storage. The remaining hydrocarbons (C2+), which leave the bottom of the Demethanizer in stream <42> are partially reboiled in (E-105) and recycled back for flashing in V-106. The vaporized portion returns to the Demethanizer at the last bottom tray, and the liquid portion is pumped to the Deethanizer (T-105) in stream <47>. Ethane (the major product) is recovered in the Deethanizer overhead stream <52> at purity of 98.6 % and then sent storage tanks. The remaining hydrocarbons (C3+) leave the bottom of the Deethanizer in stream <49> and are partially reboiled in (E-110). This stream is then flashed in vessel (V-108).

**ECONOMIC & PROFITABILITY ANALYSES**

In process synthesis and design, economic analysis is essential to decide whether the project is feasible or not. The feasibility study is carried out using the preliminary estimation method that has an uncertainty of ± 20 % [28].

**Economic Evaluation**

The total Fixed Capital Investment (FCI) estimate is based on the bare module cost algorithm, presented in Turton et al. [28]. The estimated capital investment is found to be $125,942,528, which includes all direct costs, indirect costs, contingency and fee, and auxiliary facilities. The Working Capital is estimated as 15 % of the FCI, and is found to be $18,891,379. Therefore, the total Capital Investment for the ethane recovery process is $144,833,907. The costs of manufacturing, or the operating costs, are estimated from the following cost items [30]:

- **Fixed Capital Investment (FCI = $125,942,528).**
- **Cost of Operating Labor (C_{OL})**

The number of operators required per shift depends on the number of shifts per day and the number of hours per shift. The labor costs are estimated based on the specific labor costs in the region and the number of operators required.

![Fig. 5. Distribution of utilities costs for the ethane recovery process.](image)
on the major equipment in the plant. Then the number of hired operators can be taken as 4.5 times the required number to cover all shifts [28]. Based on the method of Turton et al. [28], the average payment for each employee is estimated to be 46 800 $/year in 1996 and 71 800 $/year in 2010. The number of operators required in the ethane recovery plant per shift is 3.55 operators. Therefore, 16 operators are to be employed, with total annual operating labor cost equals to $ 1 148 500 per year.

Cost of Utilities (C_{UT})

The utilities requirements for the integrated ethane recovery process and their costs are based on a stream factor of 0.95, which amounts to annual costs of 30 199 053. The distribution of the utility costs is shown in Fig. 5. It is clear that the steam contribution is around 50 % of the total utilities costs, followed by refrigeration, which is about 32 %. On the other hand, the cooling water has the lowest share that is about 0.6 %. This means that the energy integration has reduced the annual utility cost by 28.4 %.

Cost of Waste Treatment (C_{WT})

The only stream that may be considered to contain waste is the overhead stream coming from the stripper, which contains hydrogen sulfide and water in the vapor phase. This stream is assumed to go directly to the flaring system. Therefore, no costs are associated with treating any type of wastes in this plant.

Cost of Raw Materials (C_{RM})

The main raw material for the ethane recovery process is the natural gas. Recent prices of natural gas were obtained through the Chemical Market Associates, Inc. (CMAI). The price of natural gas is 4.26 $/MMBtu. The amount of natural gas flowing into the plant is 418 325 lb/h. The lower heating value (LHV) of natural gas is 19 800 Btu/lb. And thus, the total cost of raw material (based on stream factor 0.95) is calculated based on the LHV of natural gas, which is $ 293 640 747 per year. Based on these estimated values, the Cost Of Manufacturing (COM) is estimated using the following formula [28]:

\[
COM = 0.304FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})
\]

(1)

In the above equation a depreciation of 10 % is assumed. On the other hand, the COM without depreciation (COMd) is:

\[
COM_d = 0.180FCI + 2.73C_{OL} + 1.23(C_{UT} + C_{WT} + C_{RM})
\]

(2)

Therefore, based on equations (1) and (2), the total costs of manufacturing with capital depreciation is COM = 439 739 952 $/year and that without depreciation is COM_d = 424 123 079 $/year. To find the Equivalent Annualized Operating Cost (EAOC), the FCI is annualized using the capital recovery factor (A/P, i, n):

\[
EAOC = FCI \left[ \frac{(1+i)^n}{(1+i)^n - 1} \right] + COM
\]

(3)

Assuming an interest rate of 8 % and a project life of 10 years, the EAOC is found to be 458 509 103 $/year.

The distribution of the EAOC is shown in Fig. 6, where it is clear that the raw materials contribute to about 79 % of the total EAOC. This value is an expected result,
since the cost of raw materials has a major impact on the total production cost, which ranges from 30 to 85 % [28]. The plant is expected to make revenue from selling three products; Ethane, Methane, and C3+ (containing 87 % propane). Recent prices for these three products were obtained through CMAI. The revenues are shown in Table 4.

Note that the methane revenue is calculated based on the LHV of 21 537 Btu/lb of methane. On the other hand, ethane and propane revenues are calculated based on the price per volume ($/gallon). Accordingly, the total revenue is equal to $ 474 400 780 per year.

**Profitability Analysis**

Profitability analysis is performed for the ethane recovery process based on three criteria: time, cash, and interest rate. For each of these bases, discounted and non-discounted techniques are used. In order to develop a cash flow diagram, the following assumptions are made [30]:

- Straight line method of depreciation over the project life (10 years).
- Since industrial plants are not constructed to be sold at the end of the project life, a zero salvage value is assumed (only land price is retrieved).
- Income taxes: 10 %.
- A typical interest rate of 8 % is assumed.
- Land costs 1.5 % of the total capital investment.

Based on the above assumptions, a discrete after tax cash flow diagram is shown in Fig. 7. A period of two years for constructions and a project life of ten years are assumed. Fig. 7 shows that the FCI is spent evenly during the construction period (-2 to 0). The plant start-up is assumed after year (0) and the net cash flow is reduced in the first year due to the deduction of the working capital. A constant net cash flow of $ 32 454 MM/yr from years (2 to 9) is achieved before it jumps in the last year when the working capital and the land price are retrieved. The non-discounted time criterion gives a good indication for the project’s profitability. The term used for this criterion is the payback period (PBP) and it is found to be 4.5 years. On the other hand, the discounted PBP can also be used in time profitability analysis. It is estimated by converting all non-discounted cash flows to the discounted form, then reproducing the cumulative cash flow diagram based on the new discounted values. The discounted PBP for the ethane recovery process is found equal to 6.0 years after start-up. The Net Present Value (NPV) for the ethane recovery plant is calculated with the aid of the NPV built-in function in Microsoft Excel at (i = 8%) and found to be $ 60 738 824. Whereas, Cumulative Cash Position (CCP), is simply the cumulative non-discounted cash flow at the end of the project life. For this project, the CCP is equal to $198 594 000. The Discounted Cash Flow Rate of Return (DCFROR) represents the highest after tax interest rate, at which the project can just break even. This rate is also called the Internal Rate of Return (IRR), and when computed, is compared with the Minimum Attractive Rate of Return (MARR). IRR is calculated by trial and error after setting the NPV equal to zero. The IRR for the ethane recovery project is calculated also using the IRR built-in function in Microsoft Excel and found to be 17 %. When it is compared to the common MARR of 10-12 % for chemical industries, it is considered very satisfying. An industrial project with such an IRR value is considered to be a low
risk investment [29]. Also, the Rate of Return on Investment (ROROI) is used for non-discounted interest rate profitability analysis. It represents the non-discounted rate at which money is made from the fixed capital investment. The ROROI is calculated using equation (4) and found to be equal to 16%.

\[
\text{RORI} = \frac{\text{average annual net profit}}{\text{FCI}} \frac{1}{n}
\]

The major sources of uncertainty, which will have an effect on the cost economics and savings, come directly from the fluctuating cost price of the raw material and indirectly from the equipment design equations. The uncertainty may reach up to 15% error [28, 29].

**PLANT LOCATION**

The plant site should be ideally located where the cost of production and distribution can be at a minimum level, with a good scope for plant expansion, suitable environment and safe living conditions for easy plant operation. Therefore, the Aqaba Special Economic Zone (ASEZ) in the southern part of Jordan is selected. It is directly adjacent to the site where the new Aqaba seaport will be built over the coming five to seven years [31]. The South Industrial Zone (SIZ) is a principal investment opportunity within ASEZ.

The natural gas and ethane storage tanks are located outside the battery limits of the plant. A thirty day capacity for the raw materials and products storage tanks is recommended [28]. The raw material flowrate is 173,983 m³/day, and hence the volume of natural gas needed is 5,219,500 m³. This volume should be divided into 131 floating head tanks with each 40,000 m³. This is because the maximum volume of storage tanks available from vendors is 40,000 m³ with a diameter of 42 m. The main product which is ethane has a flowrate of 2,060 m³/day, and the storage capacity needed is 61,870 m³. This volume can be divided into 2 huge cylindrical tanks of 20,623 m³ volume, 5 m length and 17 m diameter.

**CONCLUSIONS**

Natural Qatari gas, with 13 percent ethane content is considered a feasible feed stock for the production of ethane in Jordan. In spite of cryogenic conditions, the proposed process design is found to be feasible with an efficient process at high recovery rates. As expected, the cryogenic liquefaction section consumes most of the energy required in the ethane recovery process. The feed preparation through dehydration of natural gas is a critical step, and is found to affect the performance of down processing steps, due to freezing of water under severe cryogenic conditions prevailing in the separation section. The desulfurization of natural gas is crucial, since sulfur compounds are poisonous and corrosive by their nature, and high precautions should be taken into account while dealing with them. Accordingly, H₂S is removed early in the process in the feed preparation section.

With the aid of computer-aided flowsheeting, the sensitivity analysis shows that an operating pressure of 14 bar for the demethanizer yields minimum losses of ethane in the overhead product. Pinch analysis is used to design the optimal heat exchangers network, and the use of liquid methane stream as a refrigerant, has led to significant annual money savings; however at the expense of flowsheet complexity structure. The Equivalent Annualized Operating Cost (EAOC) is found to be around 78%. The profitability analysis shows that the ethane recovery process has an Internal Rate of Return (IRR) of 17%. Thus, such an investment is considered to be a safe investment.

To further assess the feasibility of ethane recovery from natural gas, it is recommended to consider other natural gas feed stocks with different ethane contents such as the Jordanian or Egyptian natural gas.

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