Amine-based gas-sweetening processes prove economically more viable than the Benfield HiPure process

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ABSTRACT

The removal of CO_2 and H_2S from natural gas streams is a crucial step for the Liquefied Natural Gas (LNG) industry. Since its inception, ADGAS (Abu Dhabi Gas Liquefaction limited) has been employing the Benfield HiPure process in its LNG plant in Das Island to purify sour natural gas to ultra-sweet specifications before it is compressed to produce LNG.

This work compares the technical and economic performance of some simple amine-based processes to the Benfield HiPure process used in ADGAS LNG units. These process alternatives are evaluated and assessed based on the product purity, energy consumption and overall economic performance.

Simulation results using the simulator ProMax clearly underpin the possibility of replacing the Benfield HiPure process with amine-based processes that achieve the same level of gas purity but with a better economics.

Processes using mixed amines such as MDEA/DGA and MDEA/DEA prove more economically viable with a 50% reduction in capital costs and 20% and 48% savings on the stripping cost and the annual power needed for the solvent circulation respectively.

Therefore, the prospect of shutting down the potassium carbonate section and retrofitting the DEA section to MDEA/DEA or MDEA/DGA appears very promising, allowing ADGAS to decrease operating costs and possibly increase capacity.

1. Introduction

There are many processes available for removal of contaminants from natural gas feeds. The selection of these processes is based on economic feasibility and cleanup ability. Chemical and physical solvent processes, or combinations of both, have been used extensively in many existing LNG facilities [8]. The removal of both H_2S and CO_2 from natural gas before liquefaction is done primarily to meet the LNG product specifications, prevent corrosion of process equipments, and meet environmental performance standards. Thus, purified natural gas streams prepared for LNG production should typically contain no more than 1 ppmv of H_2S and 50 ppmv of CO_2 [4,8].

Since its inception, ADGAS (Abu Dhabi Gas Liquefaction limited) has been employing the Benfield HiPure process in its LNG plant in Das Island to purify sour natural gas to the above mentioned ultra-sweet specifications before it is sent to the liquefaction step. This process that was first described by Benson and Parrish[1] uses two independent but compatible circulating solutions in series to achieve the required gas purity. The process consists of an amine (diethanolamine) promoted hot potassium carbonate section followed by an amine (diethanolamine) section. The choice of diethanolamine (DEA) as promoter has been underpinned by various studies that have proven DEA as the most effective promoter for potassium carbonate over others amines such as monoethanolamine (MEA), diisopropylamine (DIPA), diglycolamine (DGA), or Piperazine (PZ) [9,12,23]

Beside its relatively high capital and operating costs and high stripping energy, major process concerns of corrosion, erosion and column instability are frequently reported for units using the Benfield HiPure process in particular for the hot potassium carbonate absorber [11,22]. Recent studies on methyl diethanolamine (MDEA)-based processes have shown their commercial advantages over the Benfield HiPure Process in that MDEA is less corrosive to carbon steel, the solution is stable, and it is not as susceptible to degradation[10,13,15,19,21]. Since MDEA is not very corrosive, higher concentrations of up to 50% can be used without any significant effects on the process equipments [7,14]. Also, MDEA-based processes are simple with lower capital and maintenance costs than the Benfield HiPure process. However, MDEA is known for its low ability to absorb COS and Mercaptans. This disadvantage can be overcome by mixing MDEA with secondary amines such as DEA or DIPA, or an isomeric primary amine, such as DGA, known for increasing the mixture reaction rates with these sulfur and Mercaptan compounds [7,8,14]

In this work, the technical and economic performances of the Benfield HiPure process used to sweeten gas streams in ADGAS plant in Das Island are assessed along with some proposed MDEA-based alternatives. The study is carried out using the process simulator ProMax® V3.2.[2] The economic analysis of the different processes is based on the equipment data published in the literature [5,6,16,18,20]. Blended amines are greatly expected to bring a significant improvement in the absorption capacity, absorption rate, and also savings in solvent regeneration energy requirements. These process alternatives are expected to significantly reduce capital and operating costs while providing more flexibility in achieving specific purity requirements.

2. Process Description

2.1. ADGAS Train#3 Plant

The integrated schematic of the Benfield HiPure process used in ADGAS plant in Das Island is shown in **Figure 1**. The hot potassium carbonate absorption system comprises a split flow absorber and a regenerator with no side draws. The carbonate absorber and regenerator are both vertical packed bed columns. The treated gas from the carbonate absorber is fed directly into the amine absorber. The DEA amine system comprises an absorber and a stripper. All the columns are vertical and made of a stack structured packing. After absorbing the acid gases, the rich solution from the absorber is pumped to the DEA regenerator that has no condenser. Then the overhead gas is fed to the middle of the carbonate regenerator which does have a condenser. Liquid from the carbonate regenerator condenser is fed to the top of the DEA regenerator as reflux. Sweet gas exiting the DEA absorber will undertake further processing before being sent to the cryogenic section to produce LNG. The stripped acid gases from both the carbonate and DEA regenerators proceed to a sulfur recovery unit (SRU), where they are processed to produce molten liquid sulfur. **Figure 2** shows the conventional amine-based flow scheme process that is used as a base case to evaluate all the amine mixtures alternatives.

2.2. ADGAS Operating Data and Absorber Specifications

The operating data and absorber specifications of the Benfield HiPure columns are given in Tables 1 and 2. The equipment specifications of the MDEA-based alternatives are obtained by sizing the equipment in ProMax®[2] based on the operating data provided in Table 1. Some of the DEA equipment sizes are taken the same for the alternatives. This applies to flash vessels, coolers and heat exchangers.

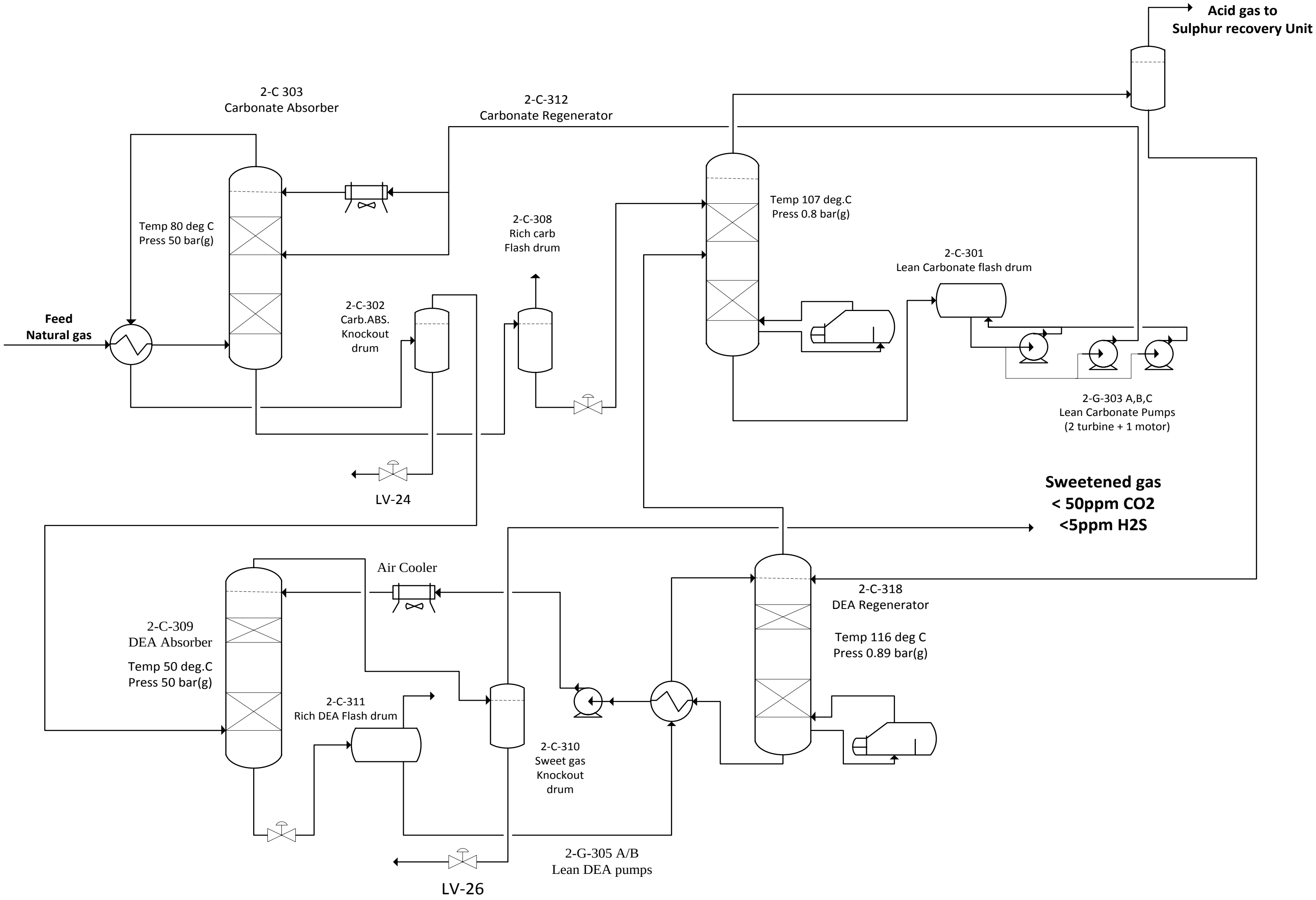

Figure 1: ADGAS' Train#3 gas treating plant – Schematic of the Benfield HiPure process

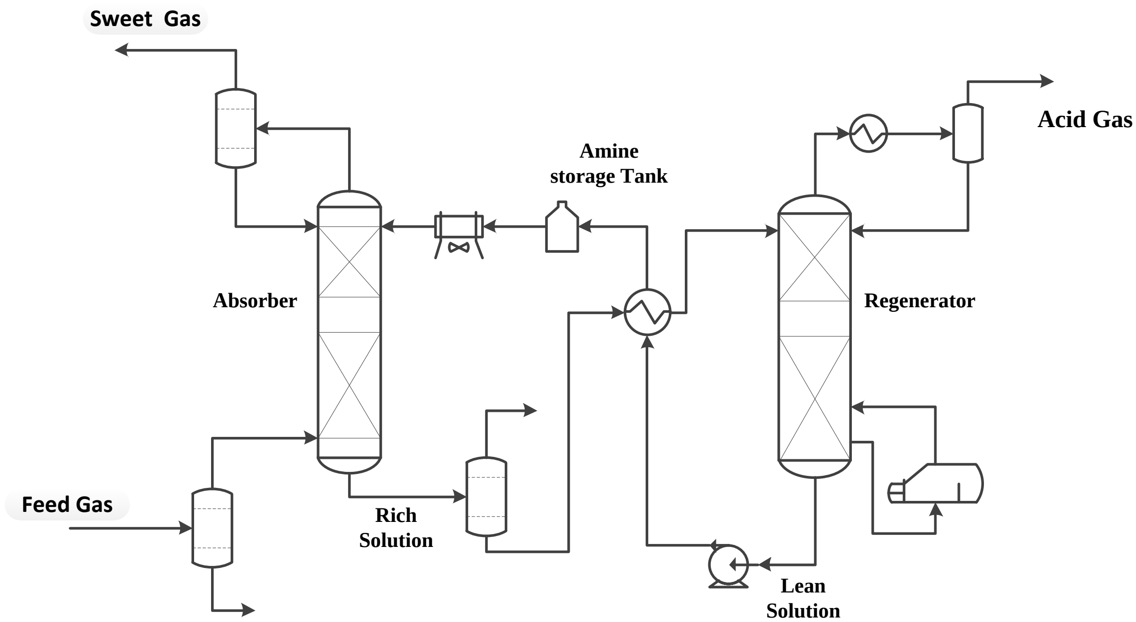

Figure 2: Schematic Process Flow Diagram for an Amine gas treating plant

Table 1: Typical ADGAS operating data

Parameter		Value
Feed Gas Flow Rate (MMSCFD)		476.93
Feed Gas Temperature (ºC)		25.03
Feed Gas Pressure (barg)		52.08
H_2S Feed Gas Composition (%)		4.67
CO_2 Feed Gas Composition (%)		2.11
Hot Potassium Carbonate Unit		
Circulation Rate (m^3/hr)	Main:	343.50
	Split:	1292.20
Lean Solvent Temperature (ºC)	Main:	81.84
	Split:	117
Lean Solvent Pressure (barg)		51.4
K_2CO_3 Concentration (wt %)		30
Promoter Concentration (DEA) (wt %)		3
Amine Unit		
Circulation Rate (m^3/hr)		109.8
Lean Solvent Temperature (ºC)		49.94
Lean Solvent Pressure (barg)		53.71
DEA Concentration (wt %)		20

Table 2: Absorber specifications for ADGAS Train#3 Plant

Potassium carbonate Absorber	
Top section	
Column Diameter (m)	3.581
Packing Height (m)	9.144
Bed 1 Packing Type	#2.5 S.S Mini Rings
Bottom Section	
Column Diameter (m)	4.724
Packing Height (m)	9.144
Bed 2 Packing Type	#3 S.S Mini Rings
Amine Absorber	
Column Diameter (m)	2.972
Packing Height (m)	15.24
Packing Type	#3 S.S Mini Rings

3. Simulation Case

3.1. Process Tool

The process calculations were completed using ProMax®V3.2 (Bryan Research and Engineering 2010)[2]. The electrolytic property package was used to predict the H_2S and CO_2 absorption in both potassium carbonate and amine units of the Benfield HiPure process. The TSWEET kinetics model in ProMax was used to predict the relatively slow CO_2-amine/carbonate kinetic reactions taking place in all absorbers.

3.2. Simulation Results

Simulation outputs reasonably matched ADGAS' operating data as shown in **Table 3**. Subsequently, comparison of ADGAS' Benfield HiPure process to the proposed MDEA-based alternatives is given in **Table 4**. The concentrations of the solvents used in this study are 50 wt. % MDEA and 3% wt. of the additive. However, due to fear of corrosion caused by high MDEA concentration, optimized concentrations are considered for each alternative, as shown in **Table 5**. It can be noted that the optimized cases have no bigger differences from the base cases. Therefore, for clarity and better comparison, the optimized case was considered for the economic analysis.

The results show that a conventional single amine MDEA system does not meet the CO_2 requirement; however, with addition of a promoter both acid gas specs are met. Simulations also show that promoted MDEA systems are able to run at about 50% of the solvent circulation rate required by the Benfield HiPure. Optimizing the alternatives shows that replacing ADGAS' Benfield HiPure with an MDEA/DGA system will reduce the solvents' regeneration energy by about 15% and 9% when MDEA/DEA is employed. Apart from the 50% reduction in circulation rate, replacing the Benfield HiPure process with MDEA/DIPA system gives no savings in regeneration energy. Therefore, to make an appropriate decision on MDEA/DIPA system, an overall economic evaluation needs to be considered. However, as shown by Table 5, MDEA/DGA and MDEA/DEA are technically judged as the best alternatives to replace the Benfield HiPure.

Table 3: Comparison of ADGAS operating data to the simulation results for the Benfield HiPure Process

Components	Sour gas	Feed Gas to DEA (ppmv)		Sweet Gas (ppmv)	
	(Mole %)	Simulation	Plant data	Simulation	Plant data
CO_2	4.7	570	574	25	19
H_2S	2.1	683	707	0.40	0.41

Table 4: Preliminary Evaluation of MDEA-based Alternatives

Process design	Solvent (wt. %)	Circulation rate (m^3/hr.)	Reboiler Duty (Gcal/hr)	H_2S (ppmv)	CO_2 (ppmv)
	30% K_2CO_3 + 3% DEA	1635.7			
Benfield HiPure	20% DEA	109.8	59	0.45	23
MDEA	50% MDEA	800	72	0.98	5786
MDEA/DEA	50% MDEA + 3% DEA	800	53	0.94	33.28
MDEA/DGA	50% MDEA+3% DGA	800	56	0.80	17.44
MDEA/DIPA	50% MDEA + 3% DIPA	800	63	0.95	4.81

Table 5: Optimized Solvent Concentration on the Alternatives

Process design	Solvent (wt. %)	Circulation rate (m^3/hr.)	Reboiler Duty (Gcal/hr)	H_2S (ppmv)	CO_2 (ppmv)
	30% K_2CO_3 + 3% DEA	1635.7			
Benfield HiPure	20% DEA	109.8	59	0.45	23
MDEA	50% MDEA	800	72	0.98	5786
DEA/DEA	30% MDEA +20% DEA	800	54	0.79	9.74
MDEA/DGA	30% MDEA + 20% DGA	800	50	0.75	7.47
MDEA/DIPA	45% MDEA +10% DIPA	800	59	0.76	4.03

4. Economic Analysis

In process economics, the total expenses on capital (CAPEX) and operation (OPEX) of a plant are directly influenced by the design and operating parameters of the process [18,20]. The circulation rate is considered to be the most important factor influencing the economics of gas treating with chemical solvents. Solvent circulation rate influence the size of pumps, lines, heat exchangers and regeneration tower, thus has a large effect on the capital cost of gas treating plants. Circulations rates also influence the energy requirement for solvent regeneration because the reboiler heat duty is associated directly with the liquid rate. Another factor that plays an important role in gas treating economics is solution corrosivity, which determines the material of construction of units due to the high temperatures and solution acidity. Estimates of the cost of equipment and other cost related to the capital investment play a crucial role in selecting among design alternatives. Capital cost estimates combined with process operating cost and other expenses are vital factors that need to be given full consideration since the viability of a proposed change to an existing gas treating unit depends on them.

4.1.Economic Assumptions and methodology

The capital and operating costs are estimated for the Benfield HiPure process and its MDEA-based alternatives within the operating and economic environments of ADGAS gas plants. The total capital cost includes all key process equipment shown in the process flow diagrams, plus a general facilities cost. The design parameters of equipment like absorber and regenerator columns, flash vessels and pumps are obtained from the plant design data sheets. The rest of the equipment is sized in ProMax® to provide information on equipments' costs as required by the calculations. The operating cost includes fixed general maintenance costs comprising labor, local taxes and general insurance. The process operating expenses are estimated through the unit cost of utilities and consumables with reference to the techno-economic performance parameters. To enable complete investigation of the process, information on both equipment and operating cost is obtained from a number of sources including vendor and public sources [5,6,17,18,20,24]. The assumptions used in carrying out the economic analysis and evaluation are given in **Table 6.** Economic gains for the existing process and its alternatives are compared using the economic potential and net present value/worth.

Table 6: Overview of the Economic Assumptions

Economic assumptions	
Project life(yrs.)	22
Equipment salvage Value	0
Construction period(yrs.)	3
Plant operating time (hr. /yr.)	7920
Interest rate (%)	5
K_2CO_3 cost ($/kg)	3.0
DEA cost ($/kg)	3.8
MDEA cost ($/kg)	2.6
DGA cost ($/kg)	4.06
DIPA cost ($/kg)	3.0
Natural gas Price ($/MMBTU)	3.5
Tax rate (%)	20

4.1.1. Capital Costs

The total capital expense (CAPEX) is divided into two main components: The fixed capital investment and the working capital investment. All the cost estimates are represented in constant million US dollars using the 2000 cost index [6,18,20]. **Table 7** gives the cost factors used to estimate all the costs involved in the construction of the plant basing on the purchase cost of each individual equipment. **Table 8** gives an overview of the equipment cost estimates for the Benfield HiPure process of ADGAS in million US dollars (M$). The most expensive equipments are the absorber and the regenerator, which are responsible for about 45% of the fixed cost on investment. Cost factors for the effects of material of construction, temperature and pressure are also included in the capital cost estimation of the equipment. Equipment such as valves, steam jet ejectors and pipes were neglected in this study.

Table 7: Typical factors for capital cost based on delivered equipment

Item[20]	Factors
Direct Costs	
Equipment delivered	1
Equipment Erecting	0.4
Piping	0.7
Instrumentation and control	0.2
Electrical	0.1
Utilities	0.5
Off-sites	0.2
Building	0.2
Site Preparation	0.1
Total Capital Cost of Installed Equipment	3.4
Indirect Costs	
Design,Engineeing & Construction	1
Contingency	0.4
Total Fixed Capital cost	**5.8**

Table 8: Overview of the equipment cost estimates for the Benfield HiPure Plant of ADGAS

Potassium Carbonate Section[a,b]	Material	Cost(M$)
Absorber	SS	18.92
Regenerator	SS	33.70
Feed/Sweet gas HEX	CS	0.55
Lean Carbonate Filter	CS	1.48
Lean Carbonate Pumps	CS	0.63
Acid gas Condenser	CS	0.26
Lean Carbonate Vessel	CS	0.14
Reflux Drum	CS	0.06
Lean Carbonate Cooler	CS	0.31
Kettle Reboiler	CS	1.42
Rich Carbonate vessel	CS	0.11
Fixed Costs		**117.65**
Diethanolamine Section[a,b]	**Material**	**Cost(M$)**
Absorber	SS	10.11
Regenerator	SS	5.74
Kettle Reboiler	CS	0.48
Lean/Rich HEX	CS	0.60
Lean Amine Pump	CS	0.15
Lean Amine Cooler	CS	0.29
Rich Amine Flash Vessel	CS	0.05
Lean Amine solution Filter	CS	0.14
Gas Condensation Drum	CS	0.07
Pure gas Flash	CS	0.08
Process Gas Cooler	CS	0.47
Lean Amine Vessel	CS	0.10
Fixed Cost		**38.37**

Fixed Capital Investment(FCI)		**156.02**
	% of FCI	
Working Capital Investment[a]	28%	43.49
Start-up + Solvent Cost[c]	10%	15.53
Total Capital Expenses (CAPEX)		**214.53**

[a] Estimated from [18]

[b]Estimated from [20]

[c]Estimated from[17]

4.1.2. Operating costs

The total operating expense is divided into five sections: the charges, direct production costs, plant overhead costs, general expenses and utility costs. **Table 9** gives an overview of these expenses in million US dollars (M$). 40% of the operating costs is for the direct production costs and only 11% is spent on utilities basically steam and electricity.

Table 9: Typical Overview of Annual Expenditure on the Benfield HiPure Process of ADGAS

	Range	**Used Value**	**Cost(M$/yr)**
Fixed Charge[a]	10-20% OPEX	20	8.99
Local Taxes[a]	1-10% FCI	9	1.18
Insurance[a]	0.2-1% FCI	1	0.12
Direct Production Costs			
Raw Materials[a]	10-60% OPEX	10	4.50
Operating labor(OL)[a]	10-20% OPEX	20	8.99
K_2CO_3 makeup (98wt %)	20.1 kg/hr		0.48
DEA makeup	5.47868 kg/hr		0.165
Activated Carbon Filter[b]			0.08
Maintenance and Repairs(M)[a]	7-11% FCI	8	0.95
Supervision & Clerical[a] Assistance(S)[a]	15 % of OL	15	1.35
Operating Supplies[a]	15% M	15	0.14

Laboratory Charges[a]	10-20% OL	20	0.27
Patents & Royalties[a]	0-6% OPEX	3	1.35
Plant Overhead Cost	50-70% (M+OL+S)	60	6.05
General Expenses			
Administrative Cost[a]	2-5% OPEX	5	2.25
Distribution and Marketing[a]	2-20% OPEX	2	0.90
R&D cost[a]	1-5% OPEX	1	0.45
Utilities			
Steam[b]		4.40$ /1000kg	4.25
Electricity[b]		0.0245 $/KW-hr	0.69
Total Operating Cost (OPEX)			**45.27**

[a] Estimated from [18]

[b] Estimated from [20]

4.2. Economic Potential

The economic potential or net annual cash income is the revenue from annual sales of product minus the total annual cost of expense required to produce and sell the product excluding any annual provision for plant depreciation but including tax.

$$EP = (R - OPEX)(1 - t) \tag{1}$$

The estimated economic potential is 16.43 million US Dollars per year with all the assumptions considered in the calculations. The product capacity is predicted from the ADGAS model developed in ProMax®. However, due to the uncertainties in both market price and production capacity, the economic potential will most likely shift to another value. Therefore economic uncertainty and sensitivity analysis should be carried out to identify the limits of such estimation.

4.2.1. Risk and Uncertainty

Due to the dynamic change in market price of natural gas, the gross income of ADGAS plant was estimated for a range of market prices published by the International Energy Agency [5] and a probability normal distribution curve is developed to cater for the uncertainties expected in market price of the purified gas. An average price of 3.5 US Dollars per million BTU is considered as the base case for this study.

The probability distribution curve integrated between limits to determine the probability that a random price value lies between the selected limits is shown in **Figure 3**. This distribution curve gives an idea on how close an estimated market price of natural gas can be to the average value (US$ 5/MMBTU). The relative likelihood is greatest near the center and diminishes near the endpoints.

In the same way, random values of the respective annual revenue can be obtained and then used to plot the cumulative–probability curve shown in **Figure 4.** This shows that the annual revenue for ADGAS will fall in the range 96 – 98 million US Dollars. Due to the unexpected change in market price, there is only 27% chance that the annual revenue of ADGAS will fall below the predicted value. This probability distribution is used to predict ADGAS profits under sales price uncertainty.

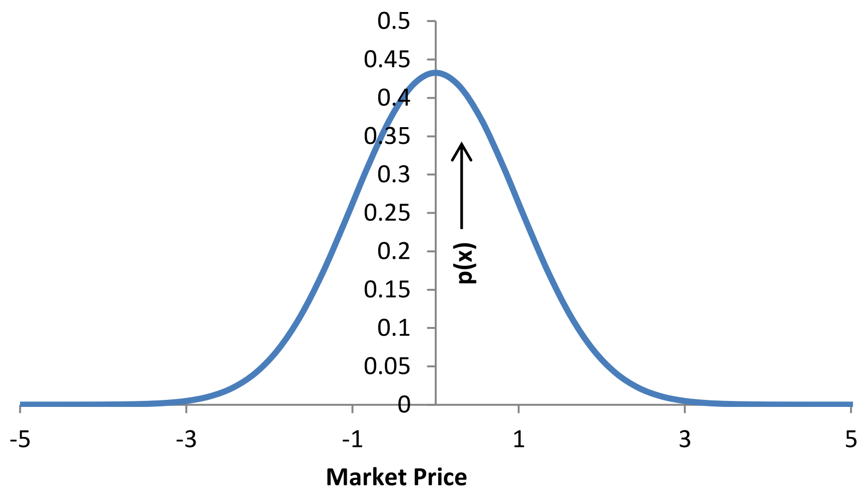

Figure 3: Distribution of Market Price

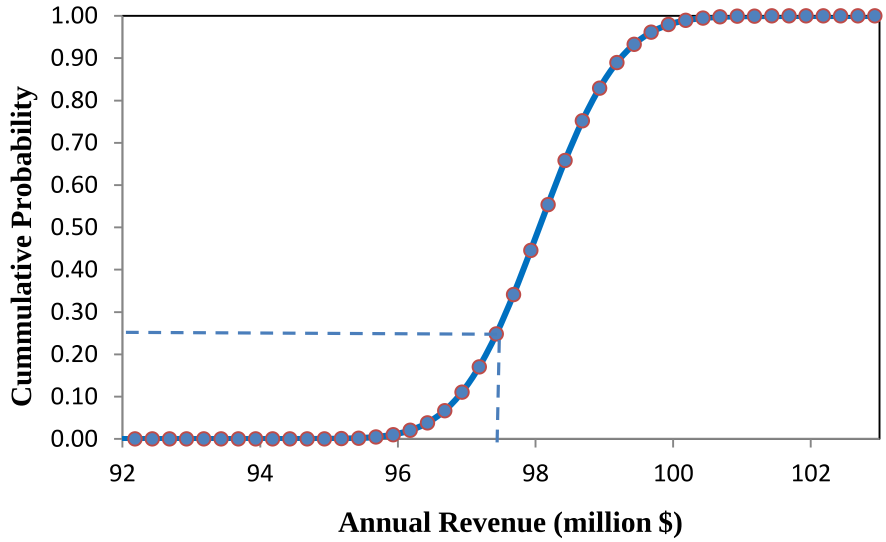

Figure 4: Cumulative Probability for the Expected Annual Revenue

4.2.2. Sensitivity Analysis

Sensitivity analysis can be made by pinpointing areas susceptible to economic changes while economic analysis can only be used to predict the relationships and magnitude of the estimated economic parameters rather than the exact cost of the plant[6] (. The major purpose of sensitivity analysis is to determine factors which are most sensitive to the profitability of a project. This is always carried out to observe the effect of departures from the estimated values.

Figures 5 and **6** show that the effect of the fixed capital cost of ADGAS on the OPEX and EP is a linear relationship with a gradient of 0.5 and 1.0 respectively. A 10% increase in equipment cost (ΔFCI) will increase the operating cost of a process by 5% of its initial value. The increase in fixed cost investment is most expected when retrofitting designs or improving process technology. Due to the fact that the operating cost of the process is affected, the economic potential is also directly affected. **Figure 6** shows that, a 10% increase fixed capital investment will reduce the economic potential by 10%. Therefore, for any additional process equipment, an additional cost of operation should be carefully estimated to avoid losses which may come along with an increase in fixed capital investment.

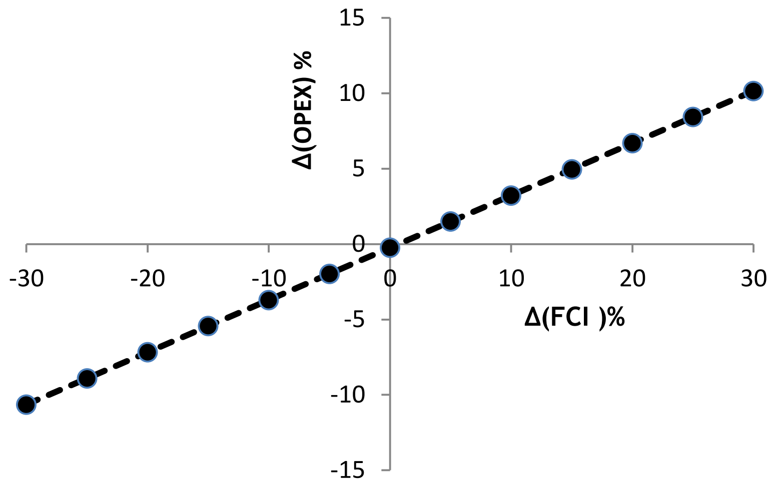

Figure 5: Effect of change in Fixed Cost Investment to Operating costs of ADGAS plant

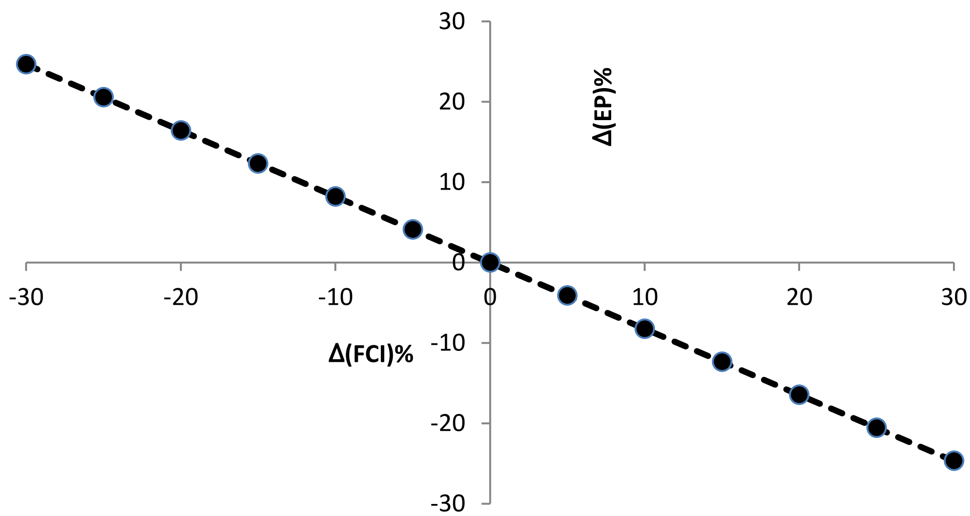

Figure 6: Effect of change in Fixed Cost Investment on the economic potential of ADGAS plant.

4.3. Net Present Value

The net present value is defined as the total present worth of cash flows minus the present worth of all capital investments as defined by equation (2).

$$NPV = \sum_{i=1}^{i=n} \{ f_D \left[(R_i - OPEX_i)(1-t) + FCI * d_f * t \right] \} - \sum_{j=-3}^{j=n} f_D * C_j \qquad (2)$$

Where

NPV net present value (profit)

n number of years

f_D discount factor

$(R_i - OPEX_I)$ savings in operating cost compared to the base case

d_f depreciation factor (straight-line depreciation considered)

FCI fixed Capital Investment

t tax rate

C_j Cash flow on capital Investment

Figure 7 shows the cash flow pattern for ADGAS plant considering at a discount rate of 5%. The first 3 years of the cash flow profile represents the construction stage of the plant. The capital cost is assumed to be released in installments throughout the whole period of construction. At low discount rates the net annual cash flow is higher and increasing the discount rate may lead to a negative net annual cash flow after some time. At high discount rate, a bigger percentage of the capital cost cash flow will increase (more negative) and more income will need to be spent to cover that gap. It can be seen from **Figure 7** that the cash flow at 30% discount rate drops faster towards the zero axes than the ones at lower discount rate. Therefore, for more profitable operation, lower discount rates should be preferred.

The cumulative annual cash flow for the net annual discounted cash flow is presented by **Figure 8**. This shows that at a discount rate of about 20%, the breakeven point will not be met up with a period of 22 years and that of 15% crosses the breakeven point after 14 years. The highest profit is obtained if the discount rate is as low as 2%. **Figure 9** shows that the estimated discounted cash flow rate of return on the ADGAS plant is about 12-15 % and this gives a payback period of 10 years.

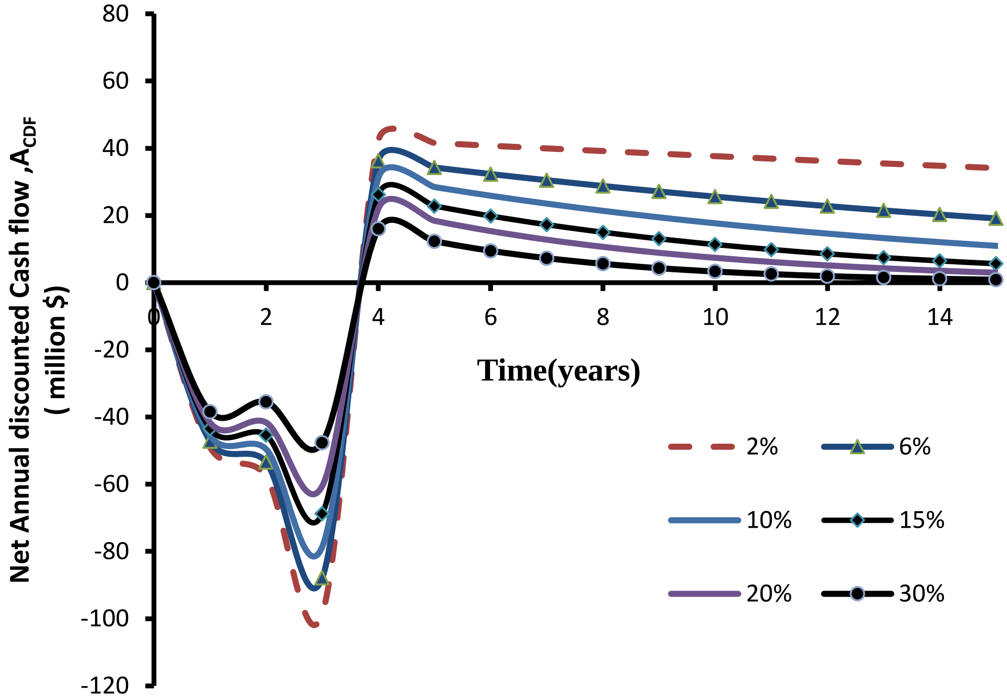

Figure 7: Net Annual discounted cash flow for the ADGAS Plant at different discount rates.

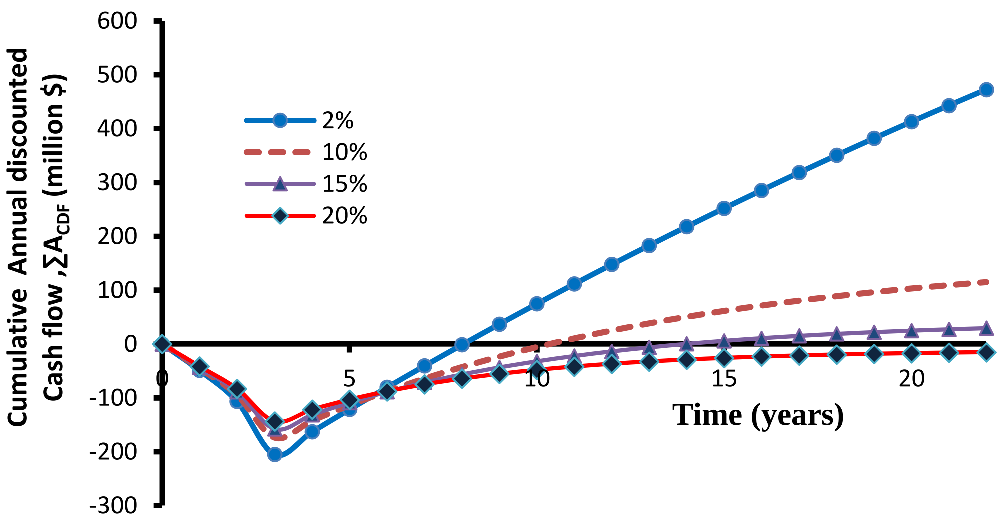

Figure 8: Cumulative Annual discounted cash flow for the ADGAS at different discount rates.

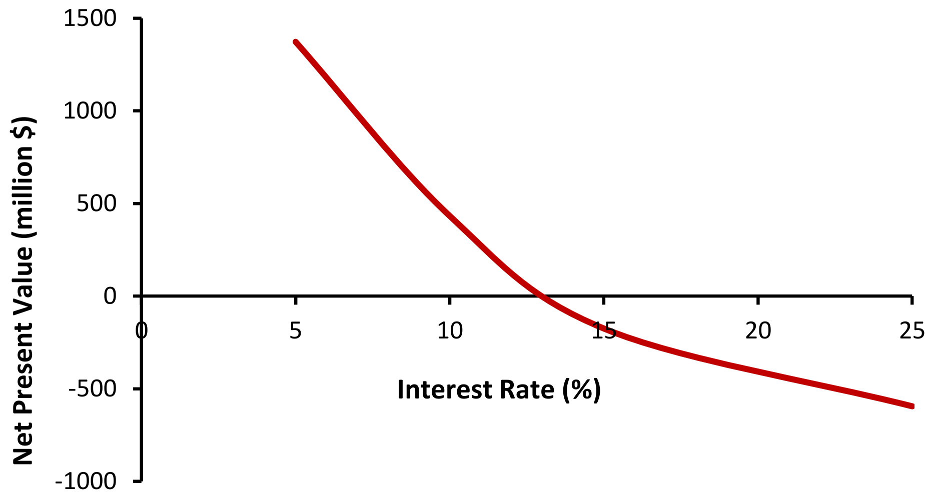

Figure 9: Estimating the DCFRR of ADGAS Plant

4.4. Economic Analysis of the Alternatives

4.4.1. Capital Costs

The capital cost of the Benfield HiPure process of ADGAS was compared with three MDEA-based alternatives as shown in **Figure 10**. From this figure, it can be noticed that considering process alternatives like MDEA/DEA, MDEA/DIPA and MDEA/DGA will reduce the capital costs by 50%. For a given economic analysis, choosing a process with low initial installation cost might not be the only best option since the operating cost may be high, making the breakeven point unattainable [3].Therefore operating cost is another factor which will need to be analyzed before a final decision can be made on the alternatives.

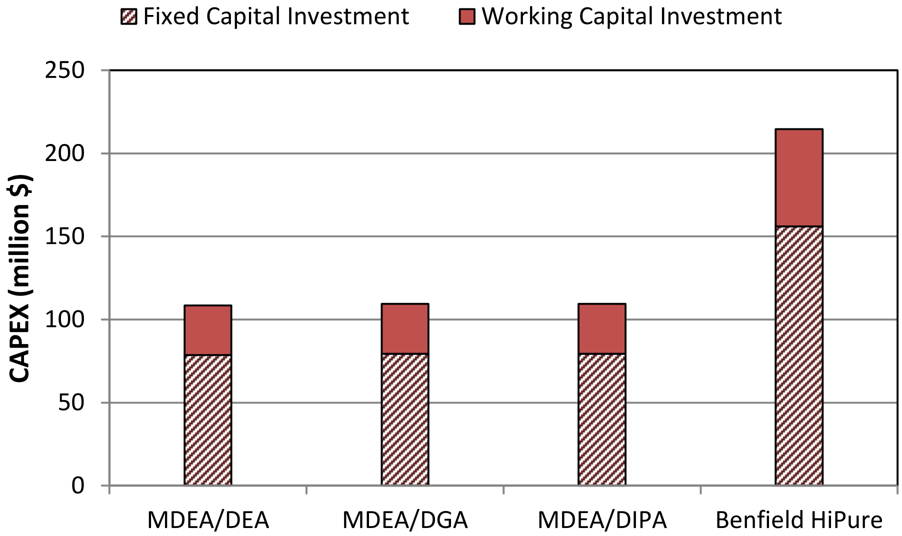

Figure 10: Comparison of the CAPEX of the Alternatives to the Benfield HiPure Process

4.4.2. Operating Costs

The operating cost of a process is directly estimated from the fixed cost of investment. Therefore, using the OPEX to compare these alternatives would not be the best option since the capital cost of the Benfield HiPure is already reported as the highest. However, more systematic means of considering the cost of stripping per unit mass of acid gas and the annual expenditure on power consumption, gives a better evaluation criteria for these alternatives.

Figure 11 shows that MDEA/DIPA has the lowest economic efficiency among all the MDEA-based alternatives with more extra costs incurred on stripping a unit mass of acid gas compared to the Benfield HiPure process.

Figure 12 shows a comparison of the estimated annual cost of power for all the alternatives. It can be noticed that MDEA-based alternatives have considerable lower annual costs spent on power consumption. Replacing the Benfield HiPure with any of the alternatives will save up to 48% of the annual expenditure on power. However, the MDEA/DIPA system would not be considered a good alternative due its high costs of stripping. Therefore, making a decision depending on the operating cost will consider only MDEA/DGA or MDEA/DEA as a better alternative the Benfield HiPure process of ADGAS.

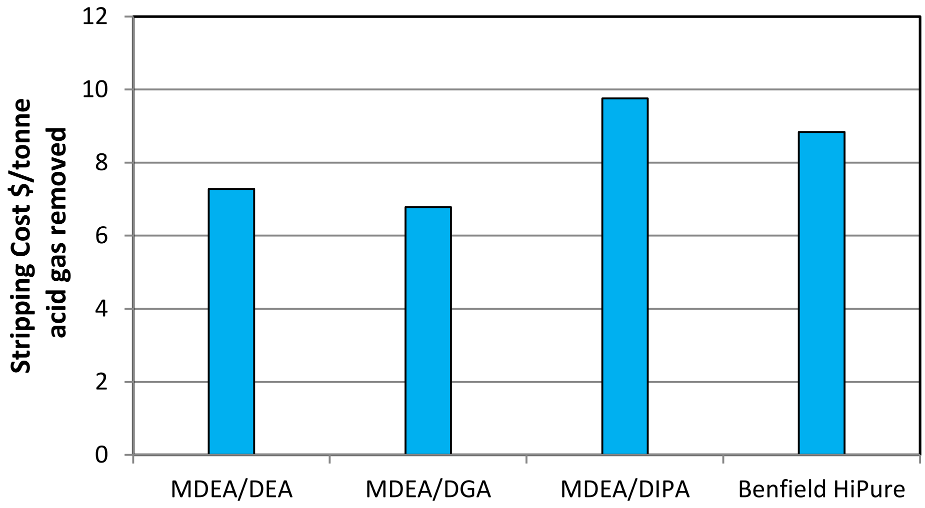

Figure 11: Comparison of the Economic Efficiency on Acid gas Stripping

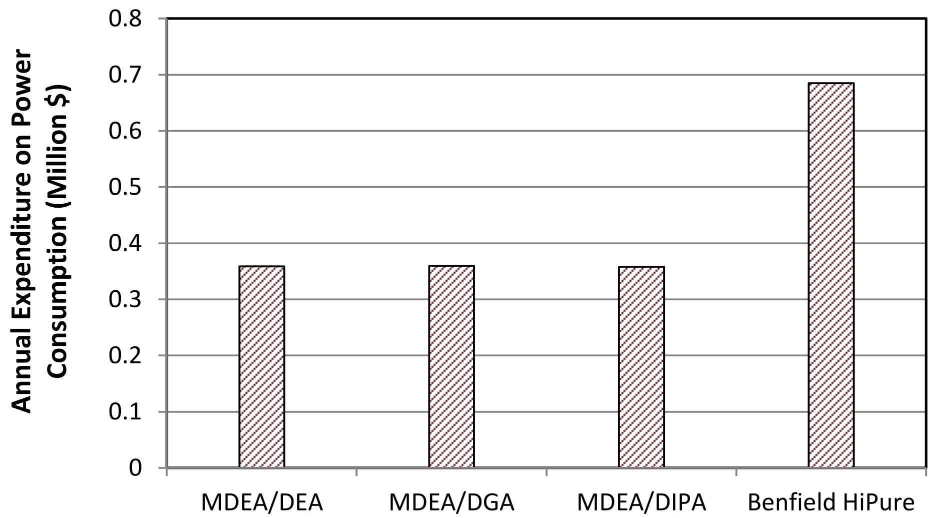

Figure 12: Comparing the Annual Expenditure on Power Consumption

4.4.3. Economic Potential

Figure 13 shows that using any of the proposed alternatives can increase the economic potential of ADGAS by averagely 37% (approximately 7 million USD per year) or 16.7%(approximately 4.4 million USD/year) on the net profit. The difference between the economic potential and net profit is that the later includes depreciation (straight line method) on the fixed capital costs.

On comparing the alternatives, the economic potential of the Benfield HiPure process is lower because of its high annualized cost. When a manufacturing enterprise invests money, it expects to receive a return during the time the money is being used. The amount of return demanded usually depends on the degree of risk that is assumed. This risk can be quantified in form of the amount of capital invested, therefore since the Benfield HiPure process has the highest capital investment; it is expected to have the highest minimum acceptable rate of return irrespective of the interest rate assumed. The higher annualized capital cost will significantly reduce the economic potential of the project. However, introducing the effect of depreciation will increase the profit of the Benfield HiPure by a greater magnitude due to its high fixed capital investment. As shown in **Figure 10**, the Benfield HiPure process has the highest expenses on capital therefore its minimum rate of return is expected to be higher than that of all the alternatives irrespective of the amount of income earned.

A decision on the choice among the alternative could not be made from these results due to the fact that their economic potentials were so indifferent.

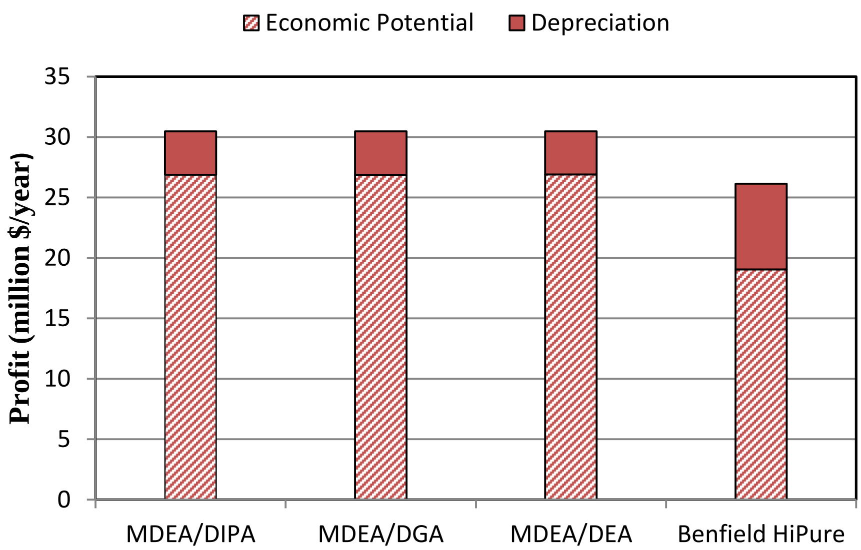

Figure 13: Comparison of the potential profits with the alternatives

4.4.4. Cash Flow

Figure 14 shows a comparison of the annual cash flow of the alternatives discounted at a rate of 15 %. Due to the fact that the Benfield HiPure process has a higher capital cost and its construction has to be completed in the same period as the alternatives, a greater minimum will be produced. A comparison of the cumulative net discounted cash flow is shown in **Figure 15**, and the PBP of the Benfield HiPure process is longer due to the higher investment and operating costs involved. However less consideration should be put on the OPEX, because this is directly estimated using the fixed investment cost. Using the alternatives will give a PBP of about 6-8 years which makes them highly more profitable than the HiPure design.

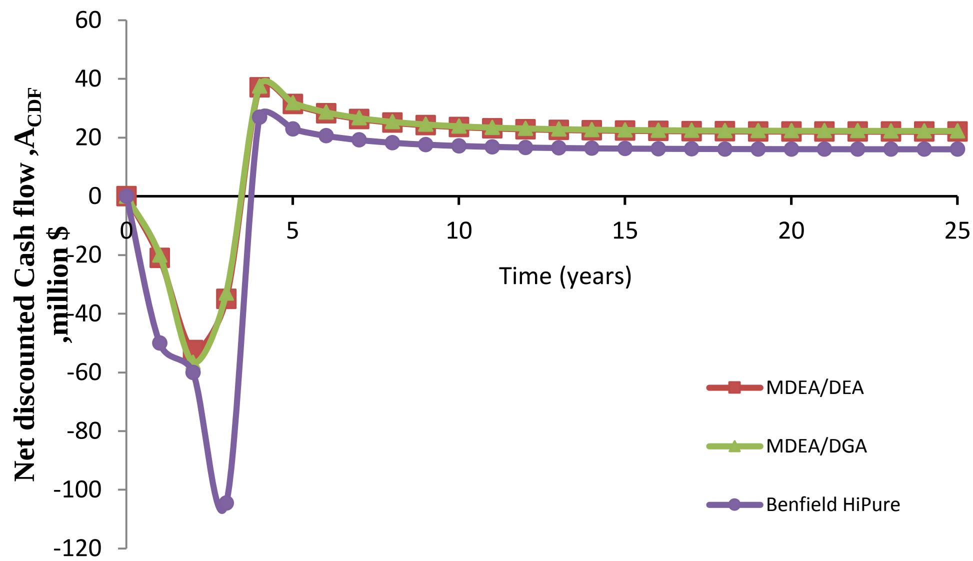

Figure 14: Evaluating Net Annual discounted cash flow at different discount rates.

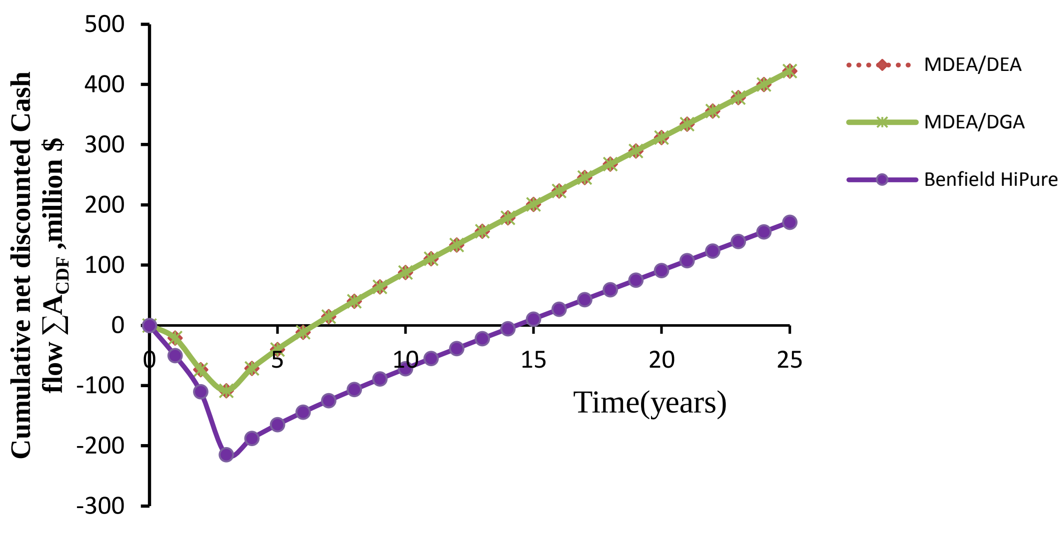

Figure 15: Evaluating the Cumulative the Annual discounted cash flow of the Alternatives at a discount rate of 15%

5. Conclusions

This study explored the potential possibilities for replacing the two-stage Benfield HiPure process currently in use in ADGAS gas plants with a single-stage amine process based on a mixture of MDEA and secondary amines. The latter has proved to be techno-economically attractive in terms of product quality and capital and operating costs. A single amine MDEA process, though economically more efficient than the Benfield HiPure process, is not able to purify the sweet gas to the required specifications. Despite a considerable reduction in solvent circulation rate, the optimized mixture of MDEA/DIPA required the same amount of regeneration energy as the Benfield HiPure process. MDEA/DGA and MDEA/DEA process alternative show a considerable reduction in solvent circulation rate with significant savings on stripping costs. The economic analysis demonstrates that MDEA/DGA or MDEA/DEA are potentially the more economically viable alternatives with approximately 50% reduction in capital costs and 48% savings on the annual power consumed. The economic study of the Benfield HiPure process also shows that more than 75% of the total fixed capital cost is taken by the hot potassium carbonate process. Therefore, all additional works may incorporate feasibility studies focusing on the prospect of shutting down the potassium carbonate section completely, and retrofitting the DEA process to an MDEA/DEA or MDEA/DGA system. If that scenario is successful, ADGAS will see a notable decrease in operating costs and, possibly, additional capacity.

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References

1. Benson, H. E. and R. W. Parrish (1974). Hydrocarbon Process **53**: 2.
2. Bryan Research and Engineering, I. (2010). ProMax ProMax V3.2. Bryan, Texas, Bryan Research and Engineering,Inc.
3. Bullin, A. and R. Hall (2000). "Optimization of Natural Gas processing plants including Bussiness Aspects." Proceedings of the Seventy-ninth GPA Annual Convention.
4. Ebenezer, S. A. and J. S. Gudmundsson (2005). "Removal of Carbondioxide from Natural Gas for LNG Production." NTNU Semester Project
5. EC-IEA (2010). Medium-Term Oil and Gas Markets I. E. Agency.
6. Holland, F. A. and J. K. Wilkinson (1997). Process Economics. Perry's Chemical Engineer's Handbook. New York, NY, Mc Graw Hill.
7. Kidney, A. J. and W. R. Parrish (2006). "Fundamentals of Natural gas processing." Taylor and Francis Group,USA.
8. Kohl, A. and R. Nielsen (1997). Gas Purification.
9. Laddha, S. S. and P. V. Danckwerts (1982). "The Absorption of CO_2 by Amine-Potash Solutions." Chemical Engineering Science **37**(No.5): 3.
10. Lallemand, F. and A. Minkkinen (2009). Highly Sour Gas Processing in More Sustainable World. Sustainable Industrial Processes. G. C. F.Gavani, S.Perathoner and F.Trifiro. Weinheim, WILEY-VCH Verlag GmbH and Co.KGaA.
11. Maddox , R. N. (1977). "Gas and Liquid sweetening." Campbell Petroleum Series **Second Edition**: 85.
12. Mahajani, V. and P. V. Danckwerts (1983). "The stripping of CO2 from Amine-Promoted Potash solutions at 100^{o}C." Chemical Engineering Science **38**: 321.
13. Mandal, B. and S. B. Shyamalendu (2006). "Simultaneous Absorption of CO_2 and H_2S Into Aqueous Blends of N-Methyldiethanolamine and Diethanolamine." Environ. Sci. Technol **40**: 9.
14. Manning, F. S. and R. E. Thompson (1991). Oilfield Processing of Petroleum : Volume one: Natural Gas. Tulsa,Oklahoma, Penn Well
15. Meissner, R. and U. Wagner (1983). "Low-energy process recovers CO_2." Oil and Gas Journal: 4.
16. Mohammad, R., L. Schneiders, et al. (2007). "CO_2 Capture from power plants: Part I : A Parametric Study of the technical Performance based on monoethanolamine." International Journal of Greenhouse Gas Control **1**: 10.
17. Mohammad, R. M. A. Z. (2009). Carbondioxide Capture from Flue Gas. Ph.D, University of Twente-Netherlands.
18. Peters, M., K. Timmerhaus, et al. (2003). Plant Design and Economics for Chemical Engineers. New York,NY, Mc Graw Hill.
19. Polasek, J. C. and G. A. Iglesias (1992). Using Mixed Amines Solutions for Gas Sweetening. Proceedings of the 71st Annual Gas Processors Association Convention, Tulsa,Ok.
20. Smith, R. (2005). Chemical Process Design and Integration. England, John Wiley and Sons.

21. Sohbi, B., M. Meakaff, et al. (2007). "The Using of Mixed Amines in an Industrial Gas Sweetening Plant." World Academy of Science,Engineering and Technology **31**: 5.
22. Tamadher, Al Bishr. (2012). "Train-3 Acid Gas Removal unit performance." SOGAT Proceedings of 8th International Conference,Abu Dhabi ,U.A.E: 21.
23. Tondinca, T., C. Tanasie, et al. (2007). "Absorption with chemical reaction: evaluation of rate promoters effect on CO_2 absorption in hot potassium carbonate." 17th European Symposium on Computer Aided Process Engineering-ESCAPE17.
24. Wells, G. (1973). Process Engineering with Economic Objective London, Leonard Hills Books.