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Cost Comparisons of Some Separation Processes

S. O. Momoh[a]
[a] Department of Chemical Engineering, Obafemi Awolowo University, Ile-Ife, Oyo State, Nigeria

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COST COMPARISONS OF SOME SEPARATION PROCESSES

S.O. Momoh, Ph.D
Department of Chemical Engineering,
Obafemi Awolowo University,
Ile-Ife, Oyo State, Nigeria.

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1.0 INTRODUCTION

There are many separation methods in use today in the chemical industries[1]. In the last 50 years or there about, the common and dominant separation processes have been the

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conventional processes of distillation, liquid-liquid extraction, gas absorption, leaching, crystallization, evaporation, e.t.c. This is so because of advantages of using these processes among which are the technical know-how on the design, construction, control and operation of these processes. Naturally, manufacturers are usually reluctant to venture into new areas. But in recent years, attention is being focussed on newer separation processes which are sometimes referred to as the "Novel Separation Processes (NSP)". These are usually seen as alternatives to conventional processes. Example of such novel processes include the membrane processes (like the Ultrafiltration (UF), Electrodialysis (ED), Reverse Osmosis (RO), etc.), adsorption, ion exchange and gas or liquid chromatography.

For a particular separation problem, it is technically possible to achieve a desired separation by more than one method of separation. Then a need arises for evaluation of the relative competitiveness of some of the available separation methods. There may be several factors that could be used as basis for achieving this, but the processing cost is usually the most important factor. It is therefore the objective of this paper to present a review of selected papers (most of which are recent) on the cost comparisons of some of the novel separation processes with conventional separation processes for solving particular separation problems. This paper is not intended to review literature on costs of separation processes per say. But the aim is to identify and review relevant practical examples in the literature (where author(s) solved a particular separation problem using two or more separation methods) to illustrate the relative competitiveness of some separation processes on cost basis. By so doing, the reader will be made aware of the economic viability of competing separation processes. First, a word on cost estimation and cost comparison of processes.

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2.0 COST ESTIMATION AND COST COMPARISON

To be able to predict whether capital should be invested in a particular project, the design engineer must carry out a cost estimation of the plant or process. There are different types of costs involved in the manufacturing process. The total annual cost of the process usually include the depreciated total capital investment which consists of "fixed capital investment for the physical equipments and facilities in the plant plus the working capital which must be made available to pay salaries, keep raw materials and products on hand, and handle other special items requiring direct cost outlay"[2] ; and the operating costs which consist mainly of the cost of heating, cooling, and utilities (e.g. supply of steam, water, power, compressed air and fuel, waste disposal, fire protection etc.).

Although there are many factors[3-7] on which the comparisons of separation processes may be based, processing cost is the most important of them all in considering the viability of any separation process. A particular separation may be achievable technically but uneconomical to carry out. In such a case, the final decision is made having examined the cost comparison of the process with the other competitive alternatives. Therefore, processing cost is usually considered as the final arbiter on any conflict over which method of separation is to be used. Cost comparisons among various processes are always complex and usually application dependent. This complexity is due to some reasons. First, different cost factors are usually considered by different authors in costing a particular process. Second, the basis for most calculations are rarely given. Third, the different methods of accounting for depreciation, cost indices that are used, type of cost estimate made and the year to which the cost is applied often cloud the situation. All the same, the

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cost comparison approach can be most useful in deciding which process is cost effective if the essential parameters are taken into account in estimating the cost of the processes.

3.0 REVIEWING THE RELEVANT LITERATURE

The following sections attempt to review some published articles relating to the cost comparisons of competing separation processes. Some of these articles deal with comparison on the basis of energy cost or utility cost alone while the majority have been on the basis of the total annual processing cost (i.e the sum of capital and operating costs). The published articles reviewed are categorized into applications in liquid hydrocarbon processing, gas processing, solid de-watering, carbon dioxide enhanced oil recovery (CO_2-EOR), and air separation.

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3.1 Liquid Hydrocarbon Processing.

The petroleum and the petrochemical industries produce great varieties of hydrocarbon mixtures which may require further separation for specific uses. The mixtures can be used as solvents and as raw materials for other chemical industries. The separation of these mixtures into relatively pure product forms may be easy or difficult depending to a large extent on the relative volatilities of the mixture under consideration. The conventional methods of separation had been distillation, absorption, liquid-liquid extraction,etc. Today, new methods of separation are being considered which may have the advantages of lower cost and purity of products.

Bonmati et al[8] gave an appraisal of the Production Scale Gas Chromatography (PSGC) developed by the Societe Natinale Elf Aquitane and Societe Recerches Techniques et Industrielle for

TABLE I

Cost comparison of the separation of 2/3 Bromothiophene Mixture

	Feed, %	B.pt ^0C	Product Purity, %
2 bromothiophene	69.5	150	99.1
3 bromothiophene	30.5	158	99.5

Efficiency

	Distillation	PSGC
Theoretical plates	50	350
Reflux ratio	15.4	-

Energy requirement (per kg of purified product)

	Distillation	PSGC
Feed preheating	71	-
Reboiler	599	-
H2 Compressor	-	231
H2 preheating	-	122
Feed vaporization	-	82
TOTAL	670kcal	435kcal

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the separation of 2/3 bromothiophene mixture. With a feed consisting of 69.5 percent of 2-bromothiophene and 30.5 percent of 3-bromothiophene, the energy requirement per kg of purified product is 670kcal for a conventional vacuum distillation process and 435kcal for the PSGC process (Table I). This shows that less energy is required in PSGC than in the distillation process. Besides the economic advantage of the PSGC method, products in the PSGC method are rarely subjected to thermal degradation due to the low residence time associated with gas chromatography. Moreover, faster throughput results in increased productivity and eliminates the need for large inventories of expensive materials. The gain in residence time is however accompanied by the need for a much larger number of plates.

This is a close boiling mixture that expectedly requires a high boiling rate. Can we still expect this high percentage energy savings if non-close boiling mixtures are involved? This is not likely. For a mixture with a relative volatility of 1.20 and above, the ordinary distillation method will be preferred. It is very unlikely too that a high percentage energy savings will be gained if a higher feed rate is required. The energy contributed by the hydrogen compressor may be high thereby reducing the energy saving proportion in the use of the PSGC method.

Rush[9] investigated ways of replacing a conventional distillation column separating dimethylformamide (DMF) from aqueous solution by alternative methods of extraction, freeze crystallization, vapor recompression, and multiple-effect distillation, each of which hopefully results in a saving of energy. The basis was a 10,000kg/hr feed of 12.5 wt % DMF, and the purity of 0.1 wt % water in the recovered DMF

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TABLE II

Economic Comparison of Alternative Separation Processes[9]

	Extraction	Freeze Crystal-ization	Vapor Recomp-ression	Multiple effect distillation.
Net savings, M S/year	120	75	180	88
Net money required, MS	3200	1600	2000	1400
Net return on new money, %	3.8	4.7	9.8	6.3
Investor's return on new money, %	16.6	18.1	32.1	25.0

Table III compares the net dollar savings of energy (thousand of dollar) for each alternative to the conventional distillation. On the basis of the investor's return (IR) (i.e. cash flow calculations) on the new money, only the vapor recompression process gives an IR which looks promising. There are cases in other processes, however, when some of these alternatives are justifiable on the ordinary net return basis and are being used. For example, we use liquid-liquid extraction in preference to distillation in recovery process containing inorganic chloride that are extremely corrosive. The low temperature extraction process permits the use of less exotic and hence much less expensive materials of construction in the recovery area.

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The approach of the IR calculations requires assumptions on escalating costs of energy sources and probable life of the project. The calculations include such items as maintenance, depreciation, and solvent losses as well as energy. A 13-year life project and a 7% annual increase in energy were used to determine the IR figures. However, the author concluded that "the greatest energy reduction in the immediate future could be accomplished by operating the existing distillation systems more efficiently". This can be fully realized when energy integration of distillation into the total process is considered; that is by designing the process so that more of the heat rejected in one part of the process is used in another part of the process.

Null[10] in 1980 estimated energy requirements of various separation processes based on the work value of the energy flow and presented criteria that will guide the process engineer in choosing the most energy efficient process. The processes considered were distillation, extraction, crystallization, RO, adsorption and ion-exchange. Based on some assumptions and

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simplifications, the author developed equivalent work loss expressions for each of the processes considered. The expressions obtained for the conventional distillation process was compared with those of other processes. This was done so as to determine the relevant analytical criteria that would make any of the processes more energy-efficient than the conventional distillation process. In the derivation of the expressions, Null noted that the true measure of the energy economy of any process is the quality of fuel that must be burned to supply the required energy input to the process. The work, W, available from a heat source, q, at absolute temperature T with surrounding at T_o is given by

$$W = q.e_p.(T-T_o)/T \qquad \text{----------}(1)$$

When $T < T_o$, W becomes negative, implying that work must be supplied to remove heat. Distillation in its simplex form, receives heat at the bottom (reboiler) and rejects heat at the top (condenser). The quantities of heat (energy) flowing are usually approximately equal as:-

$$q_B = q_o = D\Delta H_v\ (R_D + 1) \qquad \text{----------}(2)$$

For the case of distillation column heated by steam and cooled by cooling water (this is the usual case)

$$W_{eq} = D\Delta H_v\ (R_D + 1)\ e_p\ (T_S - T_o)/T_S \quad \text{----}(3)$$

where W_{eq} is the equivalent work loss for the distillation process.

Similarly, he showed that for an extraction process, equivalent work becomes:-

$$W_{eq} = (D + B)\,\Delta H_v\, e_p\, (T_{SE} - T_o)/T_{SE} \qquad \text{— (4)}$$

In order for the extraction process to be more attractive than the distillation process from an energy point of view, Null suggested that W_{eq} calculated by equation 4 must be less than W_{eq} calculated by equation 3. This criterion leads to

$$R_D > (1 + B/D)(T_S/T_E)(T_{SE} - T_o)/(T_S - T_o) - 1 \qquad \text{—(5)}$$

In general, $T_{SE} \geqslant T_S$, and usually $T_{SE} > T_S$. If $T_{SE} = T_S$,

$$R_D > B/D \qquad \text{——(6)}$$

Thus equation (6) implies that "when all reboiler temperatures are equal and the extraction solvent is nonvolatile, extraction is more energy efficient than distillation when the distillation reflux ratio exceeds the ratio of bottom product to overhead product."[10]

If the solvent is less volatile than the product, but that a reflux ratio R_E will be required for the extract separation and R_E for the raffinate, the equivalent work loss is given by

$$W_{eq} = [D(1+R_E)(T_{SE}-T_o)/T_{SE} + N_B(1+R_E)(T_{SE}-T_o)/T_{SE}]\,e_p\,\Delta H_v \qquad \text{—(7)}$$

Equation (7) compared with equation (3) leads to the criteria that extraction is more energy efficient when

$$R_D > R_E + (1 + R'_E)B/D \qquad \text{————(8)}$$

$(\text{i.e } T_{SE} = T'_{SE}\ ;\ T_{SE} = T_S)$

If the extraction solvent becomes more volatile, and with same reasoning as above with $T_{SE} = T_{SE} = T_S$,

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$R_D > [(1+R_E)N_S/D + (1+R_E')N_S'/B]\ (\Delta H_{VS}/\Delta H_V) - 1$ —(9)

Similarly, melt crystallization will be considered more attractive than distillation from energy point of view when
$R_D > (R_C + 1)\ (\Delta H_F/\Delta H_V) - 1$ ---- (10).

As for adsorption and ion exchange, they would be preferred to distillation on energy ground only if

$R_D > [\ N_{ads}/D)(\ell/H_V)(T_S/T_{RG})\{(T_{RG} - T_O)/(T_S - T_O)\}] - 1$ — (11)

This equation (11) is for the temperature swing gas adsorption process. The author's graphical presentation shows that adsorption is favored mainly for the small quantities of the less volatile component (distillation bottom product) and for very difficult distillation operations.

For Reverse Osmosis (RO) process, the work required is pressure, P, across the membrane multiplied by the volumetric flow rate. If pump mechanical efficiencies are neglected, the net energy required is given

$W_{RO} = -F.P/\pi'.R.T_{RO}\ Ln(\gamma x_{rej})$ ---- (12)

Comparing RO at 25°C and 3445kpa (500 psi) excess of pressure with single-stage evaporator using 149°C steam, we get

$W_{RO}/(W_{eq})_{Evap} = F/4737\ N_{RO}\ [26.69-1066\ Ln\{(Fx_F - N_{RO})/(F-N_{RO})\}]$
------ (13)

A graphical representation was presented for each of the above cases. On the basis of the criteria developed, distillation replacement will be difficult to justify in a situation where the

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heat can be supplied by low-pressure steam, and refrigeration is not required to condense the overhead. This is the common distillation operation case found in practice. The author well noted that the equivalent work loss concept does not apply to plants in which steam is generated sorely for heating purposes. However, such plants are inherently wasteful of energy.It can also be observed that even with the assumptions and simplifications made, most of the analyses are far from being simple. Moreover, energy while a very significant factor, is only one but a major component of the total annual processing cost of a separation process. The author did realize this too. Capital and capital-related expenses which often increase as energy costs decrease may play an important role in an economic assessment of the processes. The consideration of capital cost factor (especially for the difficult separations) in addition to the energy criteria proposed here may completely offset some of the conclusions arrived at as to the preferred use of one separation process to another. However, the proper use of the equations and the criteria presented can lead to an initial screening of possible alternatives so as to be able to eliminate those which are obviously wasteful of energy from further consideration.

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Sander et al[11] compared the utility consumption for the dehydration of ethanol of 94 wt % to 99.8 wt % by entrainer distillation and by pervaporation. The pervaporator system consisted of three LURGI pervaporator units arranged in series with respect to the feed flow. Each unit comprised three pervaporation stages which are connected on the permeate side to the integrated condenser by an internal collector. The feed entered the first pervaporation stage at a preheated temperature of 100 ^{0}C. The cost comparisons are summarized in Table III

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TABLE III

Utility Consumption and Cost for Dehydration of Ethanol of 94 wt % to 99.8 wt % by Entrainer Distillation and Pervaporation.[11]

Utilities		Entrainer Distillation		Pervaporation	
LP Steam 50DM/tD	t	1.0-1.5		0.125	
	DM		50-75		6.25
Cooling Water, 0.10DM/m^2	m^3	75		20	
	DM		7.50		2.00
Electric power, 0.15DM/kwh	kwh	15		38	
	DM		2.25		5.70
Entrainer, 1.50DM/l cyclohexane	l	1.6-3.0		-	
	DM		2.40-4.50		-
Membrane Replacement, 4-2 years.		-	-	-	8-16
Total Utility Cost	DM/t		62-89		23-30
	DM/l		0.05-0.07		0.02-0.03

(DM =Deutche Mark; tD = ton per day; l=litres;m=metres)

There are significantly lower utility costs for the pervaporation process and this amounts to one-third of those for entrainer distillation. This allows for substantial cost savings despite the claimed higher capital cost of pervaporation. The lower utility cost for the pervaporation process are mainly due to the comparative low energy requirement since in pervaporation only the heat of evaporation for the small amount of separated water has to be supplied. Apart from the utility cost advantage, there are few other practical advantages of pervaporation over the

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entrainer distillation. It is easily "adjustable and highly flexible with respect to changing feed concentration and varying throughput while ensuring a simple and safe product quality control"; and thereby producing a product that is free from all traces of entrainer. However, the frequent replacement of membrane in the pervaporation process may crucially influence the economics of the process.

Liquid Membrane Permeation (LMP) process schemes for the separation of toluene from n-heptane were compared with liquid-liquid extraction by Cahn et al[12]. The LMP schemes considered were the Light-Solvent Scheme (LSS), the Heavy Solvent Scheme (HSS) and the No-Solvent Scheme (NSS). With product rates of 23,000 ton/year and 100,000 ton/year only the cost of NSS appeared to be of the same magnitude as that of the extraction process. The costs of other processes were much higher. For example, for a product rate of 23,000 ton/year (TABLE IV), the sum total of investment and purification costs (in million dollars) were 2.5 - 3.5, 7.7, 6.3 and 3.8 respectively for extraction, LSS, HSS and NSS processes. (The investment cost includes on-site and offsite investments, while the operating costs include utilities, depreciation, and return on investment.) The membrane replacement cost is one other critical factor which could make the LMP process more uneconomically viable. It is not clear whether the cost of membrane replacement is built into the costs shown on the table. The result of the study should however be considered not only on the basis of the merits of the LMP versus solvent extraction for the separation and purification of toluene, but also as an evaluation of LMP as a novel separation tool versus a well established conventional process.

Nippon Kokan[13] designed a system for solvent vapor recovery for gasoline vapor application. The system was designed for

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TABLE IV

Economic of IMP Schemes for Toluene and Recovery[12]

		IMP Schemes		
	Solvent Extraction	Light Solvent	Heavy Solvent	No Solvent
Product Rate, 23,000 ton/year				
Total investment (S * 10^6)	1.5-2.0	3.8	3.1	1.8
Purification cost (S * 10^6)	1.0-1.5	3.9	3.2	2.0
Total (S * 10^6)	2.5-3.5	7.7	6.3	3.8
Product Rate, 100,000 ton/year				
Total investment (S * 10^6)	3.0-4.0	8.9	7.1	4.3
Purification cost (S * 10^6)	0.6-1.0	2.2	1.7	1.2
Total (S *10^6)	3.6-5.0	11.1	8.8	5.5

TABLE V

Comparisons of processes for Gasoline Vapor Recovery.[13]

	Membrane Separation	Ambient Temp Adsorption	Ambient Temp.and Atm. Pressure adsorption. vacuum regime	Refrigerated Adsorption	Chilling condensation	Adsoption
Performance	00	00	00	000	000	000
Operability	000	00	00	0	0	0
Safety	000	000	000	0	00	00
Required Space	000	00	00	0	0	00
Construction Cost	000	000	00	0	0	00
Operating Cost	000	00	000	0	0	00
Points	17	14	14	8	9	12

higher level of vapor in the feed (15-50 vol. %) and including a recovery column for gasoline recovery. The nonquantitative comparison of the options used is presented in Table V. The more the circles the better the option.

3.2 Gas Processing

In this section, we review papers dealing with the natural gas processing and hydrogen separation and recovery. Some natural gas streams contain impurities such as carbondioxide, moisture and hydrogen sulfide that require removal prior to delivery to a pipeline. The most common process for removing carbondioxide and hydrogen sulfide from natural gas is the amine absorption. Amine plant are large, heavy and need tall structure. These limitations invite the introduction of new technology like the membrane processing. Spillman[14] in a recent paper titled "Economics of Gas Separation Membranes" presented a review of some commercial applications for gas separation membranes and highlighted some of the economic advantages and constraints in these applications.

Hydrogen can be recovered from the ammonia and refinery industries. The purge gas from the conversion of hydrogen from natural gas and nitrogen from air to ammonia contains valuable hydrogen that needs to be recovered. Hydrogen is also recovered from offgases from existing refinery processes that generate hydrogen as a by-product. Conventional hydrogen recovery processes include cryogenic, catalytic and pressure-swing adsorption (PSA), oil scrubbing etc. Today, membrane processes compete favorably with some of these processes.

Grace Membrane System[15] field-tested membranes on a variety of gas streams for both natural gas processing and hydrocarbon recovery; and presented cost comparison data for the typical case

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of a packaged amine plant and a two-stage membrane processing plant (the type was not mentioned). The overall capital cost for the membrane process was found to be about 25% less than the amine plant and the operating costs about 60% less. This was on a basis of 84,900 m^3/day natural gas containing 8% carbondioxide. The company similarly presented installed capital and operating costs for a 30% DEA plant versus a two-stage membrane facility. Capital cost advantages of the membrane over the conventional method were found to increase substantially as the carbondioxide content of gas increases; and the operating cost savings increase even more as the carbondioxide content increases. It thus shows from this study that the higher the carbondioxide content the more the cost savings. This was illustrated by the work of Spillman and Dethloff[21] reviewed below. Membrane are very efficient at higher concentrations of carbondioxide and in fact it was shown[14] that the amount of membrane area actually decreases after passing the 15% carbondioxide level. On the other hand, it is not likely to have an increase in cost savings in the use of the membrane process for a higher feed rate of gas. The cost of compression in the membrane process would have increased tremendously. Moreover, the economics of scale favors amine treatment process, since processing costs decrease rapidly as the size of the gas flow rises[21].

Weber and Bowman[16] presented some economic results of comparing membrane processes with a conventional separation process in the processing of (i) a gas separation scheme (hydrogen recovery process, HDS) and (ii) a groundwater remediation system, GRS. For the HDS process, the results given for a particular case of low purity (60%) HDS purge showed that a membrane system can reduce hydrogen recovery cost by as much as 25% compared with the established technology used. The comparison would have been more interesting if a case of higher purity were considered. It would have been interesting to know if there would still be any savings in cost if, for example, about 99% purity is required.

For the GRS process, in which the objective was to remove 98% of the total groundwater volatile components (VOC) and to reduce each VOC constituents to no greater than 50 ppb, the results showed that the capital cost and the operating cost of the hybrid membrane concentrator were 2.0 to 2.5 million dollars and 600,000 dollars/year respectively, while for the conventional process of stream stripping, the costs were 2.5 million dollar and 500,000 dollars/year respectively. On the basis of this result, the membrane process is only barely competitive with the conventional process being 20% cheaper on the basis of operating cost. (The treatment capacity was 490 kl/day and the raw water flowing into the system was estimated to be in excess of 45 kg/day.). But with increased knowledge in the design and operation of membranes processes in recent years, the result may be different.

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Air products[17] made a comparison of all three processes for upgrading a catalytic reformer off-gas stream containing 75-85% hydrogen at 1.72MPa. Their results are presented in Table VI below. The membrane system was judged to have higher operating cost than the PSA system; and its capital cost, however, was the lowest of the three options. Details on each process were not given however, and it is impossible to judge process conditions and bases for costs.

In a similar study, Schendel et al[18] made a comparative cost of the use of membrane separation and PSA in the reclaiming of hydrogen from hydrotreator purge gas. A 210,000m^3/day purge stream was available at 5.5MPa and 37^{0}C containing 72% hydrogen.

TABLE VI

Product Purity, Recovery and Relative Operating, Capital, and Product Costs to remove hydrogen from a catalytic reformer off-gas[17]

Process	Purity %	Recovery %	Relative capital cost	Relative operating cost	Relative product cost
Cryogenic	97.5	96.0	1.44	1.22	1.06
Membrane	96.9	89.4	1.00	1.17	1.09
PSA	99.9	86.0	1.40	1.00	1.00

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The membrane operated with a permeate pressure of 1.7MPa. The PSA process consisted of vessels packed with selective modular sieves that adsorb hydrocarbons and other impurities to produce a 99% pure hydrogen product. Two PSA design alternatives were studied. One operates at 3.1MPa with a desorption pressure of 483kPa to feed a 413kPa refinery fuel system. The second case operates at 1.72MPa with a low desorption pressure feeding waste gas to a low pressure burner at 34.5kPa. The cost and performance factors for the three processes are summarized in Table VII. The installed capital cost for the membrane system is 51.4 and 61.5% of the two PSA processes respectively; and the operating costs are relatively the same for the three processes. The PSA processes have the advantage of higher hydrogen purity which could impact hydrotreator economics. There is no significant advantage of one PSA process over another in terms of cost. However, the second PSA is preferred of the two since it leads to a higher hydrogen recovery. It is observed that, in the overall, the recovery capacity (in this case the hydrogen recovered) is lower for the PSA processes of 99.5% hydrogen purity than the membrane process of 93% hydrogen purity. Maintaining a high recovery capacity and

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TABLE VII

Cost Comparison of Membrane and PSA for hydrotreator hydrogen recovery[18]

	Membrane	PSA No.1	PSA No.2
Waste gas pressure (psig)	770	60	5
Hydrogen recovery (MMSCFD)	4.10	3.02	4.05
Hydrogen purity (%)	93	99.5	99.5
Installed capital cost (1000$)	630	1225	1025
Operating cost ($/MSCF)	0.11	0.05	0.09
Total cost ($/MSCF)	0.20	0.29	0.24

TABLE VIII

Comparison of Membranes with Adsorption and cryogenic processing (hydrogen recovery from refinery off gas)[20]

	Membrane Process		Adsorption	Cryogenic
	80°C	120°C		
Hydrogen recovery (%)	87	91	73	90
Recovery Hydrogen purity (%)	97	96	98	96
Product gas flow rate (MMSCFD)	2.76	2.86	2.24	2.86
Power (kW)	220	220	370	390
Steam (kg/hr)	230	400	-	60
Cooling water (t/hr)	38	38	64	79
Investment ($*$10^6$)	1.12	0.91	2.03	2.66
Installation Area (ft^2)	86	52	651	1292

a high percentage purity will attract additional facilities and thus increases the investment cost of the process. A reasonable balance has to be struck between the percentage purity and the recovery capacity if cost saving is to be achieved. A similar observation can be made for results displayed below in Table VIII.

Nakamura and his co-workers[19,20] compared membranes with adsorption and cryogenic methods for hydrogen recovery process. Their comparison showed that the membrane process has between 50-70% of the capital cost of PSA or cryogenic processes. Table VIII displayed the cost-and-performance comparison of a specific case of hydrogen recovery from a refinery off gas using polymide membranes. Beside the cost advantages, polymide membranes have the advantage of high temperature operation for applications requiring such performance. Relative ease of operation and versatility are added advantages for the membrane process.

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Spillman and Dethloff[21] presented a cost comparison of membrane and amine separation for natural gas treating where carbondioxide and hydrogen sulfide are to be removed from the natural gas. The Table IX below provides a breakdown of DEA amine and membrane costs for a flow rate of 1,110,000 m^3 of gas per day at 5.00 MPa. As observed from the table

(i) Membrane processing showed economic advantage over the DEA amine system over a wide range of feed gas compositions.

(ii) Membrane processes were particularly competitive at the lower flow rates (because of their modularity) or for high carbon dioxide concentrations.

TABLE IX

Amine Versus Membrane Processing of Natural Gas[26]

Carbondioxide Content of Feed:	5%	10%	15%	20%	30%	40%
Amine						
Capital (millions of $)	3.35	4.54	5.45	6.21	7.50	8.56
Expenses (millions of $/yr)	1.22	1.81	2.33	2.82	3.73	4.58
Loss Product (" " ")	0.02	0.04	0.07	0.09	0.14	0.19
Capital charge (" ")	0.91	1.23	1.48	1.68	2.03	2.32
Processing cost ($/MSCF feed)	0.17	0.24	0.30	0.36	0.46	0.55
Membrane (Multistage Process)						
Capital millions of $)	1.86	3.33	3.87	3.69	3.37	3.32
Expenses millions of $/yr)	0.53	0.85	0.97	1.00	0.98	0.96
Lost Product (" ")	0.43	0.69	0.93	1.24	1.54	1.49
Capital charge (" ")	0.51	0.90	1.05	1.00	0.91	0.90
Processing cost ($/MSCF Feed)	0.11	0.19	0.23	0.25	0.27	0.26

(iii) Lower capital costs were observed for a wide range of compositions.

(iv) Membrane processing showed advantage in all key categories except lost product value.

The authors explained that the lost product value results from the loss of hydrocarbon into the permeate due to incomplete membrane separation; but this can be minimized through proper process design.

Meyer et al[22] presented the results of economic studies on the *Higee* centrifugal gas-liquid contactor for (i) selective hydrogen sulfide removal and (ii) a carbon molecular sieve in PSA process being developed for nitrogen removal. The Higee unit consists of a rotor containing packing in an annulus and a stationary casing. Gas enters the casing and is forced inward

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through the rotating packing where it comes into intimate contact with the liquid which enters the 'eye' of the packing annulus and is forced outward by the centrifugal force. For the economic comparison, two case problems of hydrogen sulfide removal were simulated and compared with a conventional bubble cap trayed towers. The two cases are (i) a conservative case in which the relative selectivity for the hydrogen sulfide removal by reducing the carbondioxide tray efficiency in half; and (ii) an optimistic case where the carbondioxide tray efficiency was reduced by an additional ten-fold. The feed was 1,500,000m^3/day of raw gas containing 1.7% hydrogen sulfide and 18% carbondioxide. The overall results summarized in Tables X and XI show that the Higee cases have a saving of 15-37% processing cost compared to conventional processing; and these cost savings were distributed as 22% in capital related expenses and 78% in operating expenses. The operating saving mainly resulted from a decrease in imported steam and electricity requirements due to lower circulation rates achieved by the more selective Higee unit. The results suggest that the Higee process could compete favorably with established technology.

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For the nitrogen removal process, three methods were evaluated; the cryogenic distillation, an advanced PSA process using carbon molecular sieve (CMS), and a selective membrane process. Trace concentration of water vapor and carbondioxide were initially removed in a molecular sieve adsorption bed, the purified gas cooled before an expansion cooling in the distillation column. The product stream was finally compressed from 1.1MPa to 7.0MPa. The PSA process made use of the CMS which has a higher selectivity towards adsorbing oxygen, nitrogen and carbondioxide than hydrocarbons. By adsorbing the hydrocarbon gases on the CMS and allowing the nitrogen and oxygen to pass through, nitrogen can be rejected from the natural gas. In the

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TABLE X

Estimated Costs of Selective Hydrogen sulfide Removal and Sulfur Recovery.

($1000 U.S., Mid-1987)	Conventional Tower	Higee Conservative	Higee Optimistic
Total plant Investment	13,077	12,219	10,933
Variable Operating & Maintenance Costs, First Year			
- Imported Steam	518	336	172
- Imported Electricity	392	387	280
- Catalysts & Chemicals	21	17	9
- Operating Labor.,etc	(84)	(124)	(185)
Net operating Costs	847	616	276

1,500,000m^3/day, Cost of 65psi steam=$2.44/1000lbs, electrical power = 5.24cent/kWh, sulfur credit = $100/long ton, labor rate = $10.30/hr, total maintenance cost = 3%, Indirect cost/direct cost = 1.30.

TABLE XI

Processing Costs for Selective Hydrogen Sulfide Removal and Sulfur Recovery.

(cent/MM Btu, U.S. Mid-1987)	Conventional Tower	Higee Conservative	Higee Optimistic
Capital Related Costs	7.8	7.3	6.5
Levelized Net operating costs	7.1	5.3	2.8
Levelized working capital	0.4	0.4	0.4
Levelized constant dollar Cost-of-Gas-Processing	15.3	13.0	9.7

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TABLE XII

Estimated cost for Nitrogen Removal from Natural Gas

			Membrane (Selectivity)		
	Cryogenic	PSA	10	20	35
Total Facility construction Investment ($MM)	14.3	14.7	28.9	16.7	11.2
Annual Operating Cost ($M)	(167.2)	(149.5)	210.3	(105.8)	(248.)

$300,000 m^3$/day product gas, 1987 dollars

TABLE XIII

Processing Costs for Nitrogen Removal from Natural Gas

			Membrane (Selectivity)		
	Cryogenic	PSA	10	20	35
Capital Related cost	1.12	1.16	2.28	1.32	0.89
Operating & Maintenance	(0.05)	(0.05)	0.07	(0.03)	(0.08)
Lost Gas	0.09	0.11	0.07	0.07	0.07
Processing Cost	1.16	1.22	2.42	1.36	0.88

membrane nitrogen removal system, a two-stage membrane with one-stage recycle was used. The economic comparison of the different methods is summarized in Tables XII and XIII. The capital costs for the PSA were 3% higher than the cryogenic plant cost, whereas membrane can only be competitive at selectivity greater than 20. The high cost of PSA could result from the utilization of the wide pore process which requires more adsorbent and more power to produce than the narrow pore process. For the membrane with a selectivity of 10, the higher capital cost led to maintenance costs that exceeded the electrical credit and resulted in a negative cash flow for this system. In Table XIII, it is observed that the costs for lost gas were a direct reflection of the thermal efficiency of each process. The results suggest that PSA may offer an attractive alternative to cryogenic operation in smaller capacity applications; however the potential for further improvement to PSA method is promising.

Three different processes for the separation of hydrogen gas from different feedstocks were studied by Meindersma[23]. These processes were the cryogenic distillation (purge gas from ammonia synthesis), PSA (hydrogen rich gas stream from naphtha cracker) and a membrane permeation (purge gas from ammonia synthesis). Each of these processes is fully described in the paper. The process route for the cryogenic recovery consists of pretreatment and a separation in the so-called cold box. The PSA utilizes zeolite to selectively adsorb the required gas. In the membrane unit the purged gas was delivered to the membrane at a pressure of 130 to 140 bar. Table XIV summarizes the process conditions for the feed and the products obtained, while Table XV showed the comparison of PSA and membrane permeation for the hydrogen purification. The advantages of the cryogenic process is the additional possibility of producing high purity argon and nitrogen. A disadvantage is that the hydrogen purity is relatively low (89-92%). The main advantage of the PSA is the very high purity of the hydrogen product gas, but the inerts and a considerable amount of hydrogen are desorbed at low pressure. For a membrane unit it is possible to produce hydrogen gas with a purity of 98 to 99% and with recoveries varying from 64 to 95%. From Table XV it is observed that the best choice for hydrogen purification is membrane permeation with a precompression of 50bar and a hydrogen purity of 98%. The calculated costs included the cost of precompression and product compression. In summary, it can be concluded from the study that cryogenic recovery is

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TABLE XIV

Process Conditions of hydrogen separation

	Cryogenic	PSA	Membranes			
			1983		1988	
Feed, Nm^3/hr	20000	1100	8600		7290	
H_2, mole %	59	90	61.3		62.3	
Pressure, bar	76	13	140		136	
Main inert gas	N_2	CH_4	N_2		N_2	
Product, Nm^3/hr	12000	500	3320	2290	1900	2120
H_2, mole %	89-92	99.9995	90	84.6	91.7	90.0
Pressure, bar	70	12.5	69	26.5	69	25.8
Temperature, °C	33	25	38	38	32	33
H_2 Recovery, %	90-94	50	93		80	

more suitable for capacities above $8000Nm^3/hr$ feed gas. Below this capacity value membrane permeation is likely more suitable. For a very high purity hydrogen gas, PSA is far more the best. However, with the nitrogen present in large amounts PSA is usually not economically feasible for hydrogen separation. High purity gas can also be produced by means of membranes if the hydrogen concentration in the feed is relatively high, or if the recovery is lowered. Thus hybrid system will probably play a role in hydrogen separation processes in future as this promise to combine a higher hydrogen purity with lower overall costs.

Methane gas contaminated with carbondioxide (approximately 50%) is usually generated by biogas of sewage treatment plants and landfills. The gas can be upgraded to natural gas quality with product purity of up to 99 volume percent of methane. In this application, membrane process competes with water scrubbing, amine treatment and PSA methods. Rautenbach and Ehresmann[24]

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TABLE XV

Comparison of PSA and membrane permeation for hydrogen purification (product gas is compressed to 100 bar).

	Feed			Product			Tail			Rec	Relative Costs/kg		
	Q_{H2}	H_2	P	Q_{H2}	H_2	P	Q_{H2}	H_2	P		Fix.	Var.	Tot.
	kg/hr	mole%	bar	kg/hr	mole%	bar	kg/hr	mole%	bar	%	-	-	-
A.													
PSA	224	90	20	188	99.9	19	36	59.1	1.3	84	0.49	1.03	1.52
	211	90	20	188	99.9	19	23	49.9	1.3	89	0.54	0.98	1.52
Membr	230	90	20	189	98.1	10.3	41	65.1	17	82	0.43	1.06	1.49
	295	90	20	189	99.0	9.0	106	77.6	17	64	0.42	1.31	1.73
	197	90	50	188	98.3	20	9	32.7	48	95	0.43	0.94	1.37
	244	90	50	188	99.0	20	56	69.2	48	77	0.44	1.13	1.57
B.													
PSA	225	92.3	20	180	99.6	19	45	71.4	1.3	80	0.56	0.84	1.40
Membr	161	92.3	70	149	98.2	32.4	12	52.1	68.5	93	0.27	0.73	1

A: Feed is naphtha cracker gas, B: Feed is cryogenic product gas.

reported the production of natural gas from landfill by a membrane process. The specific operation was the enrichment of methane on a landfill site to about 90-96 volume percentage. After discussions on material transport in gas permeation and mathematical modelling of the gas permeation process, the authors presented a comparison graph of a specific product cost between gas permeation and alternative processes of water absorption, HEA absorption and an adsorption process. Although no details are given of other processes, it is observed from the graph that the membrane cost are the lowest of the alternatives for the wide range of flow rates considered.

3.3 Solid De-watering Processes

These are processes involving the removal of a fluid from a solid-fluid mixture thus concentrating the mixture. The conventional methods of carrying out this include the methods of evaporation, freeze crystallization,etc. Newer methods of separation include RO, Ultrafiltration and ion exchange. These newer methods have the advantage of low residence time, product purity and low cost of operation.

Wagner[25] carried out an analysis in which RO was used as an alternative process to evaporation in the processing of effluents from a Norwegian paper mill. The analysis was based on cost comparison. On having some problems with their evaporation method, the company decided to install hyperfiltration plant working at an average pressure of 4.0MPa and a temperature of 30^0C. The solids in the feed are 6% on average and in the concentrate are 12%. The water removal capacity was about 7500 litres/hour. The savings in oil of 1100t/yr over the conventional process of evaporation was worth 56,000 pound sterling/yr.

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A graphical comparison of operating and capital cost of RO,freeze crystallization and evaporation processes for the desalination of water was produced by Porter et al[26]. The result (obtained using cost data from the literature) showed that membrane processes are most attractive at small throughput but economics of scale favor the use of evaporation at large throughput. Freeze crystallization was considered to probably be the least attractive. Furthermore, it was observed that membrane processes are attractive up to a water throughput of about 100 tons/day, and if suitable membranes were developed it may be that membrane processes (RO and Ultrafiltration) will be used to remove water in large applications of biotechnology, at least as an initial step.

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Bemberis and Neely[27] compared UF with vacuum filtration (VF) and spray drier(SD) for the processes of kaolin de-watering, fermentation broth clarification, apple juice clarification and egg white. The kaolin clay de-watering cost for UF and SD were based on the filtration surface area required in 1152m^2 of membranes versus 426m^2 vacuum filter. The internal rate of return (IRR) for the rotary drum vacuum filter was 33% with a payback period of 3.5 years. The lower capital cost and operating costs of the UF system· increase the IRR to 59.2%. For the filtration area of 1152m^2 of membrane against 1135m^2 of vacuum filter area, the result showed that 'the high capital cost for the VF reduced the IRR to 8.5%. However, the UF system is extremely attractive with IRR of 74.2% . For these cases the major differences in operating costs are high chemical and labor requirements for the vacuum filter. The high cost component for the UF system is the membrane replacement. The economic comparison of a UF system using diafiltration and a precoat filter with cake washing was similarly considered. The capital cost of the UF was four times greater than that of the precoat filter, but the operating costs

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TABLE XVI

Ultrafiltration Compared with Conventional Clarification Technique for Juice Clarification[28]

Basis: 70,000 gal/20hr day: 150 days/year

Ultrafiltration		Conventional Clarification	
Pretreatment		Pretreatment	
Enzymes	$150.00	Enzymes	$600.00
Membrane Replacement	207.00	Diatomaceous earth	700.00
Cleaning Chemicals	16.00	Filter Pada	160.00
Maintenance	12.00	Maintenance	25.00
Power (32hp)	32.00	Power (75hp)	75.00
Labor	50.00	Labor	600.00
Total/day	$467.00		$2,160.00
Total/year	$70,000.00		$324,000.00
Total yearly Savings is $253,950.00			

were only 44% of the precoat operating cost. However, it would be observed that a UF system can significantly outperform precoat filter on slimy fermentation broths that produce poor filtrate rates. A juice clarification by Plate and Frame presses and UF showed that, using UF would lead to a saving of about $\$8.84/m^3$ based on operating costs and with a resultant IRR of 81%. Also, the operating costs to concentrate to 20% total solids using UF was found to be one-fourth those of a spray drier removing the equivalent amount of water. This difference results because of energy savings since UF de-watering does not require a phase change to remove the water.

Blanck and Eykamp[28] illustrated a comparison of UF with a conventional clarification processing technique with respect to major production costs for a juice clarification process. A total yearly savings of 253,950 dollars in favor of UF was obtained

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TABLE XVII

Cost comparison of RO process with Distillation for the Production of Purified Water (Cost in US dollars)[29]

		RO	Distillation
I	Capital Costs		
	System	115,500	207,000
	Pretreatment	48,000	43,000
	Total purchase price	163,500	250,000
II	Operating Costs		
	Energy cost	38	862
	Water cost	60	32
	Membrane cost	123	14
	Regeneration chemicals	14	14
	Total	255	910
III	Total per day	381	1,102
IV	RO saving per year	137,300	
V	Payback period	1.2	
VI	5-year saving	686,500	

(Table XVI). The single unit operation of the ultrafilter increases juice yield, eliminates diatomaceous earth filtration, and reduces operating costs when compared with the clarification process incorporating centrifugation and filtration. These were the added advantages of UF over the conventional process of clarification.

In 1986, Parekh et al[29] considered a cost comparison of a Reverse Osmosis (RO) process and a comparable size distillation process in the production of purified water. The result, summarized in Table XVII showed that the capital cost of the distillation process is 52% higher than the RO operation. The operating cost data also indicated that the RO process costs \$4.70/1,000 litres while the distillation process costs \$20.9/1,000 litres. On the whole, RO process costs are approximately 25% of the distillation costs. This is a useful and important comparison since most of the major costs are accounted for.

3.4 Carbon Dioxide Enhanced Oil Recovery (EOR)

The carbondioxide which is pumped into drying oil reservoirs to extend the life of an oil field can over time begin to contaminate the natural gas associated with the well. The purpose of gas separation is to claim both the natural gas and the carbondioxide. The claimed carbondioxide can be utilized again. The most common method of separation was the amine absorption treatment process. In recent years membrane systems are quite useful for this application because of their effectiveness at the high carbondioxide content.

The separation of carbondioxide from hydrocarbon stream containing a large percentage of carbondioxide using (i) extractive distillation and (ii) distillation with membrane was studied by Schendel[30]. The typical case was a mixture of methane(0.005), ethane (0.218), carbondioxide (0.475), propane (0.188), i-butane(0.022), n-butane (0.058), i-pentane (0.014), n-pentane (0.012), hexane (0.008) and hydrogen sulfide (0.0093). The concentrations are in mole percentages. The operating pressure is 41.4 MPa at a temperature of 289K and volume of $0.3 * 10^6$ m^3/day. The utility consumption and the installed capital cost are as shown in Table XVIII with an assumed life of three years.

The membrane process is competitive for this particular case with a savings of 24% in utility cost and 37% in capital cost. Thus the savings in both operating and capital cost are claimed to be

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TABLE XVIII

Utility and Installed Capital Cost for Carbondioxide Separation[30]

Process	Power kW	Steam kW	Membrane Replacement	Total Annual Cost,$
Extractive distillation	484	7228	-	$1.34 * 10^6$
Membrane (including Hydrogen Sulfide removal)	888	3442	98,800	$1.08 * 10^6$

and the installed capital cost

Extractive distillation	$\$5.65 * 10^6$
Distil-membrane	$\$4.12 * 10^6$

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sufficient to pay for the hydrogen sulfide removal at the front end of the process. It is to be noted that membranes are most effective at high concentrations of carbondioxide because membranes use partial pressure as their driving force. They are however, uneconomical for high feed rates of the mixture. The horsepower requirement of membrane which include compressors, pumps, refrigeration and air cooled fans may be very high for high feed rates of the mixtures. This will surely lead to a reduction in energy savings against the extractive distillation process at higher feed rates of the mixture.

Table XIX represents the results of comparing amine and membrane processes for natural gas treating for a specific gas stream of 300,000 m^3/day, 12.2% carbondioxide stream[31]. In this case, the use of a single-stage membrane reduced the capital cost significantly. The increase in the operating cost is due to methane losses in the permeate. The hybrid system reduced the

TABLE XIX

Comparison of amine and membrane processing for natural gas treating[31]

	DEA Amine	Membranes	Membrane/ DEA Hybrid
Relative Capital Cost	1.0	0.26	0.72
Relative Operating Cost	1.0	1.51	1.14
Relative net present cost at 15%	1.0	0.76	0.89

TABLE XX

Process comparison for natural gas treating[14]

Option Description	Overall operating	Overall capital	Overall space	Overall weight
1. Single-stage membrane	1	1	1	1
2. Membrane plus DEA	0.44	4.6	1.19	5.39
3. Two-stage membrane plus compression	0.85	1.43	1.65	1.47
4. Selexol	0.13	6.18	1.74	7.79
5. DEA	0.087	6.36	1.96	8.10
6. Act. MDEA	0.068	6.25	1.58	7.91

methane loss due to the use of membrane for bulk removal and DEA system for final clean-up. Table XX similarly shows the relative merits of membranes for a specific gas treating application. In natural gas processing for offshore platforms in particular, weight and space savings translate rapidly into money saved. As observed from the table, the weight and space savings are quite significant for the membrane operations.

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TABLE XXI

Gas separation process comparison for Carbondioxide EOR[32]

	DEA Amine	Cryogenic	TEA/DEA Amine	Membrane/ DEA
Capital investment (millions of $)	103.6	73.5	65.0	47.0
Operating cost ($*10^6/yr)				
Ex. Utilities	10.8	7.8	6.9	8.8
Utilities	24.9	18.0	15.9	10.7
Product losses	0.1	2.6	0.1	0.9
Capital charge ($*10^6/year)	28.0	19.9	17.6	12.7
Total cost ($ * 10^6/yr)	68.2	48.2	40.4	33.1

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A detailed case study was performed by Amoco Production Company[32] involving 4,440,000 m^3/day, 90% carbondioxide enhanced oil recovery stream, at 1.86MPa pressure. DEA amine, cryogenic, combined TEA/DEA amine, and membrane/DEA processes were compared (Table XXI). In the analysis, membranes were used as "topping" process for bulk removal of carbondioxide prior to a "polishing" process with a conventional DEA unit. The hydrocarbon gas product and carbondioxide stream have pressure requirements of 4.39 and 3.00 MPa respectively.The analysis summarized investment costs and operating expenses for treating the gas at several stages in the project. For example, the carbondioxide content and flow rate increase over time from 1.9% carbondioxide/444,000m^3/day at start to 90% carbondioxide/4,440,000m^3/day at maturity. The membrane/DEA option shows the most favorable economics.

Table XXII shows a similar analysis comparing membrane with hot potassium carbonate and cryogenic separation process for a

TABLE XXII

Process Comparison for Carbondioxide EOR[33]

	Hot Potassium Carbonate	Cryogenic DEA	Membrane/ DEA
Capital Investment (millions of Dollars)			
CO_2 Removal Unit	21.1	24.2	16.1
DEA Treating	-	4.0	4.9
Compression	19.9	17.3	18.1
Other	4.0	-	4.0
Total	45.0	46.4	43.1
Operating Expense ($MM/yr)	11.8	8.5	6.5
Capital Charge at 27% ($MM/yr)	12.2	12.5	11.6
Total ($MM/yr)	24.0	21.0	18.1

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3,000,000 m^3/day, 80% carbondioxide stream[33]. The feed gas was available at pressure of 0.17MPa and was compressed to 1.72MPa for amine and membrane systems and to 3.44MPa for the hot potassium carbonate processes. This cost of compression according to Spillman[14] tends to obscure the capital costs benefits of the membrane approach. Similarly, Schendel et al[34] in 1984 showed the cost comparison (expressed in ratios) of a membrane process with membrane/hot potassium carbonate and straight amine treating of 3,120,000m^3/day, 40% carbondioxide stream. Although details of their cost calculations were not available, the results are presented in Table XXIII. A straight TEA amine system provided the best economics for this case. The large volume of gas to be processed contributed to the high investment cost of membrane for this application. Spillman[14] observed that one of the contributing factor to the high cost of membrane operation is the fact that a 40% carbondioxide content falls nearly on top of the

TABLE XXIII

Process Comparison for Carbondioxde EOR[34]

	TEA Amine	Membrane/ Hot Potassium Carbonate	Membrane only
Capital cost	1.0	1.0	1.6
Operating cost	1.0	1.3	1.9
Evaluated cost	1.0	1.1	1.7

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cost curve for membrane separation. With increased knowledge of design and operation of membrane processes in the last few years, the results would probably be different if carried out now.

3.5 Air Separation Processes

Oxygen and Nitrogen are the primary products of air separation. For example, nitrogen production is utilized in inert gas blanketing while the major potential market for oxygen is that for enhanced combustion. Due to membrane performance limitations, majority of membrane applications are for the production of nitrogen. Membrane design for air separation generally consists only of a single-stage membrane. Recycle is technically feasible but very expensive.

Monsanto[35] published some cost comparison calculations on their Prism Alpha air separation method between the gas cylinder delivered liquid nitrogen gas and a nitrogen gas produced via membrane process. The liquid nitrogen gas cost was \$0.40 per 3 m^3 and the tank rental was \$370/month. The powder metal company using the nitrogen required 8,580 m^3 of nitrogen gas. This amounts to a total cost of \$0.53/3 m^3 of nitrogen to the user. A membrane process producing the same volume of 99% nitrogen gas is calculated to have a compressor power requirement of 7.05kW and a monthly capital and compressor maintenance charge of \$475/month. Assuming a power cost of \$0.07/kWh, the membrane approach provides nitrogen at a cost of \$0.26/3 m^3. This represents more than 50% saving over the delivered liquid nitrogen. The cost saving for the membrane unit would increase for lower volume nitrogen requirement and purity.

The energy requirements, operating costs and capital costs for an A/G Technology membrane system was compared with PSA and DOW membrane by Gollan and Kleper[36]. All the major cost components were taken into account in this analysis. The systems investigated were the generation of 10 tons/day of 35% Oxygen Enriched Air (OEA) for combustion air and 3 ton/day of 95% Nitrogen Enriched Air (NEA) for blanketing applications. As observed from Table XXIV the capital cost advantage for the A/G membrane over PSA is about \$26,000 ton/year (representing 92% capital cost savings); and the operating cost advantage is about \$15,000 ton/year (representing 51% savings). If higher purity is required, membrane may not be suitable. For the NEA, the cost comparison indicated in Table XXV showed that A/G membrane was less expensive (33% less) in terms of the capital cost than the PSA. The operating costs for the A/G membrane and the PSA system are similar, about \$35/ton and \$37/ton respectively. The DOW system is about \$57/ton. As implied by the authors, these results are generally not true for higher concentration and purity requirements. In such a case, the PSA system is generally preferred to membrane systems on a purely cost basis. It is noted by Spillman[14] that current membrane technology cannot economically produce oxygen concentration above 40-50%. Higher oxygen concentration are possible by reducing the membrane area

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TABLE XXIV

Cost Comparison for 35% OEA Generation (10 tons/day available oxygen)[36]

	A/G Membrane	PSA
Installed capital cost	$287,500	$552,000
Membrane replacement	38.11	0.00
Power cost	85.50	130.50
Operator labor cost	8.60	8.60
Maintenance & taxes	9.91	19.02
Capital charges	105.18	201.95
Depreciation	33.03	63.41
$/day	280.33	423.49
$/ton	28.03	42.35

TABLE XXV

Cost Comparison for 95% NEA Generation (3 tons/day available Nitrogen)[36]

	A/G Membrane	PSA	DOW Membrane
Installed capital cost	$90,000	$120,000	$132,000
Membrane replacement	16.01	0.00	40.24
Power cost	35.10	40.50	38.61
Cooling water cost	0.43	0.43	0.39
Operator labor cost	4.30	4.30	4.30
Maintenance & taxes	3.89	5.18	4.02
Capital charges	32.93	43.90	48.29
Depreciation	12.96	17.28	13.41
$/day	105.61	111.59	149.27
$/ton	35.20	37.20	57.41
$/100SCF	0.13	0.13	0.21

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contact time and by membrane recycle process in which the permeate stream is recompressed and passed over a second membrane unit. Although this is technically feasible such an approach is relatively expensive.

4.0 CONCLUSION AND FUTURE TREND

All the available literature have not been mentioned in this review. The paper, however, covers some of the relevant publications on the cost competitiveness of some key separation processes. Most of the cost comparisons analyzed suggest that under certain conditions there is an economic advantage in the use of novel separation processes over the conventional ones. This assertion may not be generalized for all feed rates, feed and product conditions and compositions, only on the basis of the cost comparisons available in the literature. Most of the cases found in the literature are for specific feed rates, feed and product conditions. Cost of processes are strongly affected by the feed and product conditions and compositions. For some of the cases, not all the major cost components are taken into account by the respective authors; and it is also likely that in many cases the costs developed are out of date. However, the paper achieves its objective of providing an overview of the economic viability of some separation processes in relation to others.

Quantitative suggestions can be made from the review.

(1) Membranes are particularly good for gas separation processing. They are particularly competitive for lower flow rates or lower purity natural gas treatment. Higher feed flow rates result in higher product flows but lower purities for the same amount of membrane area. For example, (i) membrane technology has brought down the cost of air separation to levels

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where it can be competitive in some applications with on-site deliveries of cryogenically produced gases; (ii) membranes utilization in the recovery . of hydrogen is now a wide spread in the ammonium and refinery industries and, to a lesser extent in the petrochemical industry. They compete favorably with older processes of cryogenic, catalytic, pressure swing adsorption, etc,; (iii) the membrane processes can be useful in carbondioxide enhanced oil recovery processing. The objective of the separation is to claim both the natural gas and the carbondioxide (for reinjection) in an oil field. They are very competitive at lower flow rates or for high carbondioxide concentrations; (iv) contaminated methane generated from the landfill can be free of carbondioxide by the use of membrane process instead of the common processes of water scrubbing, amine treatment, etc,.

(2) Use pervaporation method for ethanol dehydration. It ensures a lower cost of operation and a product that is free from impurities.

(3) Use a reverse osmosis method for desalination of water up to a water throughput of about 100 tons/day.

(4) Gas chromatography are useful for the separation of hydrocarbon mixtures with a relative volatility of less or equal to 1.10. Above this value, ordinary distillation operation may be more economical.

What will be the trend in the future? Granted that the cost comparisons reviewed in this paper suggest economic advantages in the use of novel separation processes over the conventional ones, would it mean that, in the long run, the latter processes would become less and less attractive? This is not likely in the immediate future. The conventional processes are well

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established; their feasibility in use, process control, method of design and construction, etc., are well known. These advantages would still continue to make these processes attractive. However, as the novel or newer processes are being developed and used successfully on more occasions and with the design procedures more established, they would become a more important component of the spectrum of separation processes to be considered for new plants.

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NOMENCLATURE

B = distillation bottom product, Kmol/hr

D = distillation overhead product rate, Kmol/kr

F = Feed in Kmol/hr

ΔH_F = heat of fusion of the component at the melting end of the crystallizer.

ΔH_v = latent heat of vaporization of overhead product

ΔH_{vs} = latent heat of vaporization of the solvent in extraction

N_{ads} = average rate of adsorption, Kmol/hr

N_B = extraction molar rate of bottom product, Kmol/hr

N_S = molar rate of solvent for extract separation, Kmol/hr.

N_S' = molar rate of solvent for raffinate separation, Kmol/hr

P = Pressure, N/m^2

q = heat source quantity.

e_p = the power cycle efficiency (relative to a Carnot Cycle)

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q_B = rate of heat absorption at the bottom of the distillation column.

q_o = rate of heat rejection at the top of the distillation column.

R = gas constant

R_C = the crystallization reflux ratio

R_D = external distillation reflux ratio.

R_E = reflux ratio required for the extract separation.

R'_E = reflux ratio required for the raffinate separation

T = heat source temperature, K

T_O = surrounding temperature, K

T_{RG} = supply temperature of the regenerated gas in the adsorption, K

T_{RO} = osmotic absolute temperature, K

T_S = absolute temperature of the steam used to supply the heat to the distillation column.

T_{SE} = absolute temperature of the heat source in the solvent separation step of the extraction.

T'_{SE} = heat source temperature of the raffinate solvent separation step.

x_{rej} = mole fraction of the solvent in the concentrated solution.

1 SCF = standard cubic feet = $0.03m^3$

1 CSCF = hundred standard cubic feet = $3m^3$

1 MSCF = thousand standard cubic feet = $30m^3$

1 MMSCF = million standard cubic feet = $30{,}000m^3$

1 MM = million.

Greek Symbols

γ = activity coefficient of solvent, usually assumed equal to 1.0 for water.

ζ = total regeneration heat per Kmol adsorbed.

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π = osmotic pressure in N/m^2

= $\rho_m RT \ln(\gamma x_{rej})$

ρ_m = molar density of the solvent.

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