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Supersonic separation in onshore natural gas dew point plant

Priscilla B. Machado [a], Juliana G.M. Monteiro [a], Jose L. Medeiros [a], Hugh D. Epsom [b], Ofelia Q.F. Araujo [a,*]

[a] Escola de Química, Universidade Federal do Rio de Janeiro, Avenida Horácio Macedo, 2030-Ilha do Fundão, Rio de Janeiro-RJ 21941-909, Brazil
[b] Twister BV, Einsteinlaan 10, 2289 CC RIJSWIJK, The Netherlands

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ABSTRACT

Conditioning of natural gas (NG) for sales mainly involves meeting water- and hydrocarbon-dew points (WDP and HCDP, respectively) while assuring high heating value (HHV) specifications achievable through minimal extraction of C_{5+} components (NGL). This paper compares technically and economically a supersonic separator technology – Twister®, which can promote simultaneously WDP, HCDP and enhanced NGL extraction – to a conventional gas treating technology, consisting of an onshore natural gas dew pointing plant with TEG Dehydration unit coupled to a Joule-Thomson/Low Temperature Separation unit (TEG + JT/LTS). In Twister® technology, water and hydrocarbon dew pointing normally requires pressure drop, which results in more NGL recovery than necessary to meet the established product specifications. An economic scenario was evaluated with NG and crude oil prices of US$ 4.22/GJ and US$ 50/bbl, respectively. The economic performance of each process alternative is evaluated in terms of the net present value (NPV) after 20 years of operation, with an assumed discounted rate of 10%. Twister® based process outperformed conventional TEG + JT/LTS process as the additional revenue from the increased NGL production compensates for the lower revenue from the NG sale resulting from decreased flow rate and lower NG HHV.

1. Introduction

Processing of natural gas is the largest industrial gas separation application. The U.S. consumption of natural gas is higher than 22 trillion scf/year (0.62 trillion m^3/a), and the total worldwide consumption surpasses 95 trillion scf/year. This consumption drives a worldwide market for new natural gas separation equipment of more than $5 billion per year (Baker and Lokhandwala, 2008). Natural gas contains many contaminants, water being the most common undesirable component. Most natural gases will be nearly water-saturated at the temperature and pressure of production. Dehydration of natural gas is hence a critical step of the natural gas conditioning process as it reduces the potential for corrosion, hydrate formation and freezing in the pipeline. Water is also removed to meet a water dew point requirement of sales specifications, normally ranging from 32.8 to 117 kg/10^6sm^3 (Gandhidasan et al., 2001).

A conventional method for dehydration in the natural gas industry is the use of a liquid desiccant contactor-regeneration process. In this process, the wet gas is contacted with a lean solvent (containing only a small amount of water). The water in the gas is absorbed by the lean solvent, producing a rich solvent stream and a dry gas (Carrol, 2009). The solvent is regenerated in a second column and is then returned to the first column for water removal from feed gas. Glycols have proved to be the most effective liquid desiccants in current use since they have high hygroscopy, low vapor pressure, high boiling points and low solubility in natural gas (Gandhidasan et al., 2001). TEG has gained nearly universal acceptance as the most cost effective of the glycols due to superior dew point depression, operating cost, and operation reliability. However, there are several operating problems with glycol dehydrators. Suspended foreign matter may contaminate glycol solutions and overheating of the solutions may produce decomposition products. Foaming of solution may also occur with resultant carry-over of liquid. Last, there are environmental issues associated to fugitive emissions and efforts for reducing these emissions are being sought (Gandhidasan et al., 2001).

Besides water contamination, natural gas contains liquids that should be commonly removed to meet hydrocarbon dew point specification. In most instances, natural gas liquids (NGLs) have higher values as separate products, and cryogenic processing, although a costly alternative is the preferred technology for this purpose. It is worth noting that hydrocarbon dew point in natural

* Corresponding author. Tel.: +55 21 2562 7637; fax: +55 21 2562 7535.
E-mail addresses: machadopriscilla@ig.com.br (P.B. Machado), julianamoretzsohn@yahoo.com.br (J.G.M. Monteiro), jlm@eq.ufrj.br (J.L. Medeiros), Hugh.Epsom@twisterbv.com (H.D. Epsom), ofelia@eq.ufrj.br (O.Q.F. Araujo).

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gas is operationally important and HCDP is a quality parameter for gas sale (Herring, 2010). An undesirable result of extracting NGL is a lower heating value of the gas product which can reduce its market value. HCDP specification usually is met through Low Temperature Separation (LTS).

In this work, an alternative technology – the Twister® Supersonic Separator – is evaluated and compared to a conventional TEG + JT/LTS technology. Twister® can achieve both water- and hydrocarbon dew pointing in one unit. The supersonic separation equipment has thermodynamics similar to a turbo expander, combining cyclone gas/liquid separation, and re-compression in a compact, tubular device (Brouwer and Epsom, 2003). According to Brouwer and Epsom (Brouwer and Epsom, 2003), whereas turbo-expander transforms pressure drop into shaft power, Twister® achieves a similar pressure drop by converting pressure to kinetic energy. Table 1 displays the advantages and disadvantages of Twister® and TEG + JT/LTS for achieving HCDP.

2. Process simulation

To assess the technical feasibility of TEG + JT/LTS and Twister® gas processing alternatives, material and energy balances were performed using process simulator UniSim® Design (Honeywell). Equipment sizing was conducted according to Campbell (Campbell, 2004). Capital and operational expenditures (CAPEX and OPEX), revenue and cash flow were evaluated according to Turton et al. (2009). The composition and flow rate of the gas to be treated in the present study is reported in Table 2, in a water free base. The gas was taken as saturated in water at process inlet conditions. As process design premises, it was assumed specifications of WDP of −45 °C@1 atm and HCDP of 0 °C@45 bar, which are consistent with the Brazilian market regulations.

2.1. TEG + JT/LTS

Fig. 1 shows the flowsheet used for the conventional scheme. Water absorption by TEG is used to achieve the WDP (−45 °C, 1 atm). Subsequently, HCDP (0 °C, 45 bar) is adjusted in the Joule-Thomson expansion system. The saturated feed gas enters to the bottom of the absorber and flows in countercurrent to the TEG solution to absorb water. The water lean TEG solution is fed to the top of the absorber, whilst a water rich TEG solution is recovered at the bottom. The dry gas leaving the top of the absorber has the specified water content. The column operates at 70 bar and the dry gas stream temperature is 26.4 °C. This gas is cooled down to 11.2 °C in a heat integration exchanger before entering the Joule–Thomson valve. The isenthalpic expansion drops the temperature to −3.3 °C and the heavier hydrocarbons condensate. The condensed hydrocarbon liquid is recovered in a vessel (V-102), while the gas passes through the gas-gas heat exchanger and leaves the plant at 13.7 °C and 39.5 bar. The HCDP@45 bar is −3 °C (specification: 0 °C max) and the WDP@1 atm is −54 °C (specification: −45 °C max).

Table 1
Comparison of natural gas HC dew point technologies (Mokhatab and Meyer, 2009).

Process	Advantages	Disadvantages
TEG + JT/LTS	- Simple and compact process - Ease in operation - Low capital cost - Low maintenance cost	- Hydrocarbon dew point control is directly related to the pressure reduction cross Joule–Thomson (JT) valve - Off-spec gas during start up - Sensitive to feed gas composition
Twister®	- Can achieve dehydration and dew point control simultaneously - Remove more hydrocarbons than JT valve for the same pressure drop - Compact module design - Ease of installation and operation - Low maintenance cost	- High compression horsepower - Limited commercial test experience and performance relies on proprietary information - Limited turndown without operator involvement

Table 2
Inlet gas flow rate and composition.

Flow rate ($MMsm^3/d$)	6.00
Component	% mol
CO_2	0.1394
C1	95.7814
C2	2.3263
C3	0.7095
iC4	0.2139
nC4	0.1808
iC5	0.1026
nC5	0.0550
nC6	0.0724
nC7	0.1232
nC8	0.0886
nC9	0.0094
N_2	0.1975

The rich TEG solution that leaves the absorber contains dissolved gas, which is released through depressurization to 4 bar. The expanded stream flows into a coil that passes through the top of the TEG regenerator column and is consequently heated up to 85 °C. The solution is fed to a vessel, where the gas is separated and sent to a flare. The equilibrium calculations predict a mass flow of 2.2 kg/h for this flare gas. The rich TEG is heated in a TEG-TEG heat integration exchanger and is fed to the top of the TEG regeneration column at 140 °C. The regenerator re-boiler operates at 195 °C, and receives a stripping gas flow of 201 sm^3/h to improve the concentration of TEG in the bottom stream. As a result, the lean TEG solution has a purity of 99.3% w/w. The lean TEG coil helps to lower the top temperature to about 110 °C, avoiding massive TEG losses. The top outlet shows ca 0.4% TEG, which is equivalent to 1.93 kg/h. The high purity lean TEG is cooled down to 140 °C at the TEG-TEG exchanger and pumped to a 70 bar pressure at which it enters an auxiliary heat exchanger that uses water to cool the TEG down to 40 °C. At this temperature and pressure, the lean TEG is fed to the top of the absorber.

The plant has direct emissions as the top product of the regenerator, consisting of a mixture of water (41.77%), hydrocarbons (53.52%), CO_2 (1.92%), N_2 (2.73%) and TEG (0.06%). These values are given by UniSim® Design, based on thermodynamic equilibrium calculations (Peng Robinson EOS), but practice indicates that carry over effects can lead to TEG losses of about 1%. This more conservative value was used for OPEX and emissions calculations.

2.2. Twister® scheme description

Fig. 2 shows the flowsheet used for the Twister® scheme. The saturated production gas is compressed (K-100) from 70 bar to 82 bar, leading to a discharge temperature of 40 °C. The gas exchanges heat with the export gas (E-100), reaching −21 °C. Knock-out vessel V-101 separates the liquid phase that results from both compressing and cooling the gas. Saturated gas at 81 bar and −21 °C enters the Twister® tubes. At the primary outlet, the temperature reaches −34 °C and the pressure drops to 54.2 bar. At the secondary outlet, the temperature reaches −40 °C and the

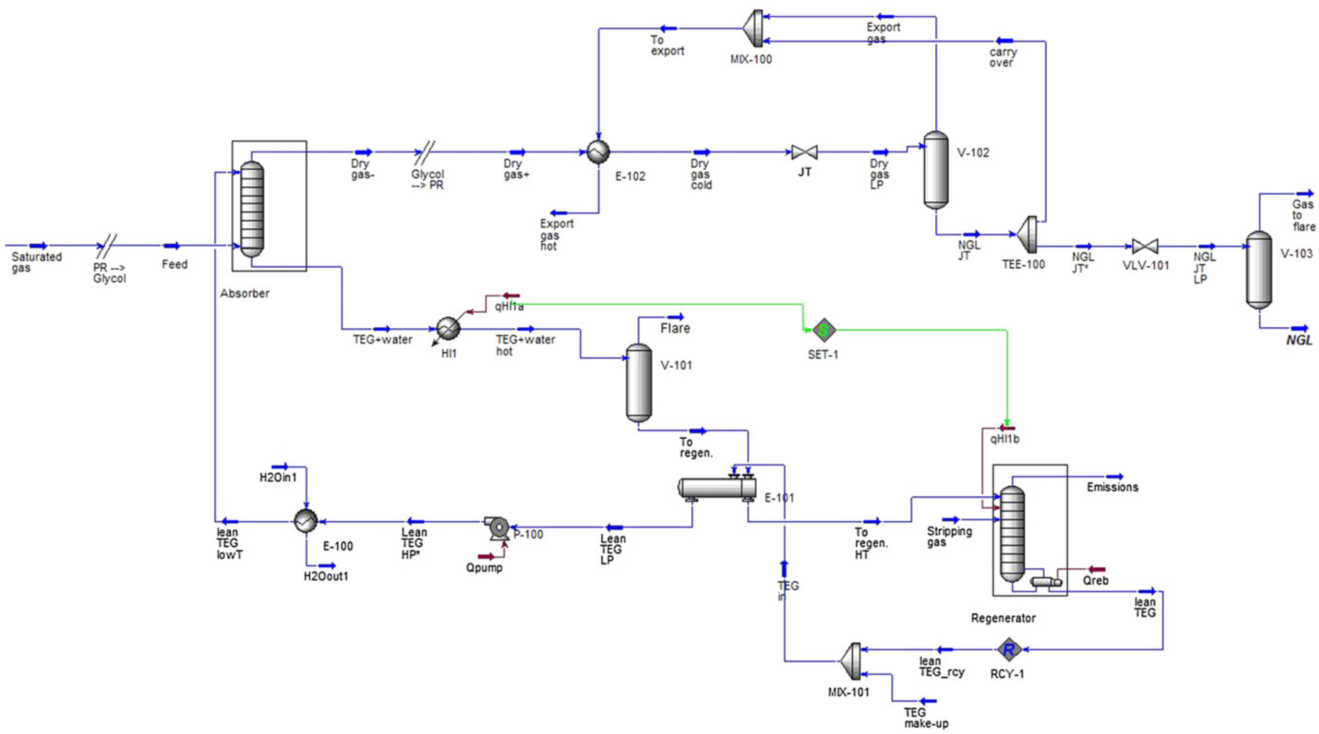

Fig. 1. TEG + JT/LTS process flowsheet for natural gas dew point control.

pressure drops to 52.8 bar (35% of the inlet pressure). This stream contains mostly water and heavy components, but also contains gas. This gas is recovered at the LTX vessel. The gas outlet of the LTX joins the primary outlet of the Twister® tubes, and the resulting NG stream exchanges heat with the inlet gas. The export gas is at 35.5 °C and 51.9 bar which is 12.4 bar higher than for the TEG + JT/LTS case. The HCDP@45 bar is −39 °C (specification: 0 °C max) and the WDP@1 atm is −70 °C (specification: −45 °C max). Details on equipment sizing are given in section 3.

It is worth stressing out that supersonic velocity is the key to separation and condensation of natural gas liquids and water in the Twister® technology. While a turbo-expander transforms pressure to shaft-power, Twister® achieves a similar temperature drop by transforming pressure to kinetic energy (i.e., supersonic velocity). Hence, a superior pressure at the front end of the equipment is required. Since for comparative purpose the gas is fed to both processes at 70 bar, Twister® technology, as opposed to the TEG + JT/LTS alternative, requires a compressor at the equipment's entrance. However, natural gas at the exit point of the Twister® process (stream "Gas to Export", in Fig. 2) is at 51.9 bar while the exit point of the TEG + JT/LTS (stream "Exported Gas hot", in Fig. 1) is at 39.5 bar, therefore representing a benefit of the technology as compression power to reach gas pipeline pressure will be reduced.

The gas outlet from the NGL stabilization drum is used as fuel in the gas turbine power generation system.

3. Equipment sizing

Process equipment sizing was performed according to principles outlined in the literature (Campbell, 2004; Turton et al., 2009) and is summarize in Tables 3 and 4.

4. Economic analysis

TEG + JT/LTS and Twister® schemes were evaluated based on capital expenditures (CAPEX), operational expenditures (OPEX), revenue and profitability estimations.

4.1. CAPEX estimation

Although capital cost estimation study performed from a process flow diagram (PFD) has expected accuracy between +40% and −25%, this type of preliminary feasibility estimate can be used

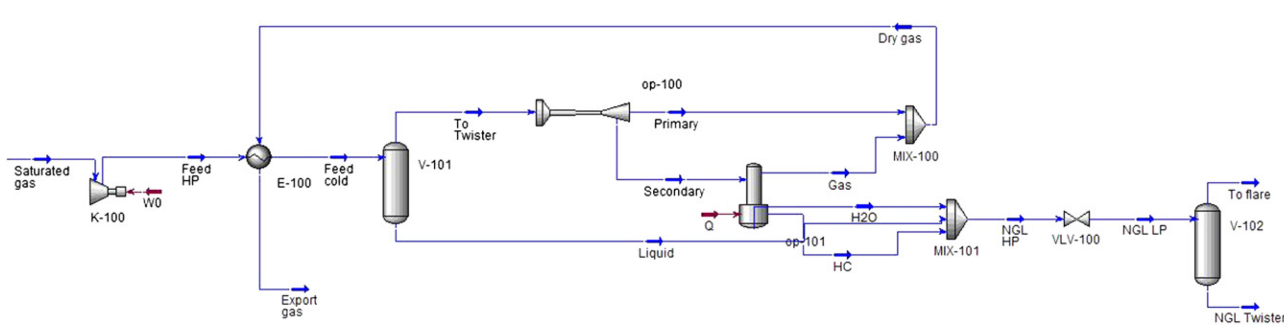

Fig. 2. Twister® process flowsheet for natural gas dew point control.

Table 3
Equipment summary for TEG + JT/LTS processing scheme.

Vessels			
Equipment tag name	**V-101**	**V-102**	**V-103**
Material of construction	Carbon steel	Carbon steel	Carbon steel
Diameter (m)	0.60	2.47	0.60
Height (m)	1.80	9.88	1.80
Orientation	Vertical	Vertical	Vertical
Bare Module Cost (USD)	12,739	1,061,959	12,739
Pumps			
Equipment tag name	**P-100**		
Material of construction	Carbon steel		
Power (shaft) (kW)	9.0		
Efficiency	75%		
Type/drive	Centrifugal/electric		
Pressure in (bar and Pa)	1.00 bar (1.0×10^5 Pa)		
Pressure out (bar and Pa)	70.00 (7.0×10^6 Pa)		
Bare Module Cost (USD)	200,379		
Heat exchangers			
Equipment tag name	**E-100**	**E-101**	**E-102**
Type	Fixed TS	Fixed TS	Fixed TS
U (kW/m^2 °C)	0.37	0.40	0.367
Area (m^2)	17.72	7.50	433.52
Shell side			
Max. temp (°C)	35.00	194.90	13.74
Pressure (bar and Pa)	4.00 bar (1.0×10^5 Pa)	1.05 bar (1.05×10^5 Pa)	40.00 (4.0×10^6 Pa)
Material of construction	Carbon steel	Carbon steel	Carbon steel
Tube side			
Max. temp (°C)	143.60	140.00	26.43
Pressure (bar and Pa)	70.00 bar (7.0×10^6 Pa)	4.00 bar (4.0×10^5 Pa)	70.00 bar (7.0×10^6 Pa)
Material of construction	Carbon steel	Carbon steel	Carbon steel
Bare Module Cost (USD)	71,373	66,501	228,199
Towers and reboiler			
Equipment tag name	**Absorber**	**Regenerator**	**Fired heater**
Material of Construction	Carbon Steel	Carbon Steel	
Diameter (m)	2.50	0.40	
Height (m)	13.58	4.00	
Orientation	Vertical	Vertical	
Internals	11.70 m, 304 SS	2.12 m, 304 SS	
Pressure (bar and Pa)	70.00 bar (7.0×10^6 Pa)	2.00 bar (2.0×10^5 Pa)	
Duty (kW)			390
Bare Module Cost (USD)	2,310,646	13,407	642,874
JT valve			
Equipment tag name	**JT**		
Pressure drop (bar)	29.50		
Bare Module Cost (USD)	115,000		

for comparing process alternatives (Turton et al., 2009). The estimation includes the main process equipments: pumps, compressors and turbines, columns and vessels, fired heaters and heat exchangers. To consider the effect of time on purchased equipment cost, the Chemical Engineering Plant Cost Index (CEPCI) was employed, with a value of 524 for CEPCI, corresponding to the year 2009 (http://www.che.com/pcitrial/).

The capital cost of chemical plants was estimated using the bare module cost (BMC), which considers both direct and indirect costs for each unit. For a given piece of equipment, the BMC corresponds to the purchased cost in base conditions, multiplied by a correction factor that accounts for material of construction and operating pressure effects. The calculation technique is explained in detail by Turton et al. (2009), and can be calculated with CAPCOST, an MS Excel file provided by the authors (CAPCOST_2008.xls). CAPCOST was used in the present work to determine the BMCs of the plants. The total bare module cost (TBMC) is obtained by simply adding up all the plant's BMCs. Finally, the total module cost (TMC) is obtained by multiplying TBMC by a factor of 1.18, which accounts for contingency and fee costs (Turton et al., 2009). As an approximation, the capital expenditure (CAPEX) is herein calculated by multiplying TMC by a factor of 1.5. Hence, CAPEX equals 1.77 times TBMC.

Table 4
Equipment summary for Twister® processing scheme.

Vessels		
Equipment tag name	**V-101**	**V-102**
Material of construction	Carbon steel	Carbon steel
Diameter (m)	2.00	0.60
Height (m)	8.05	1.80
Orientation	Vertical	Vertical
Bare Module Cost (USD)	1,033,015	12,739
Compressor and drive		
Equipment tag name	**K-100**	**D-100**
Material of construction	Carbon steel	Carbon steel
Power (shaft) (kW)	1338	1338
Type	Centrifugal	Electric
Bare Module Cost (USD)	2,499,121	481,328
Heat exchanger		
Equipment tag Name	**E-100**	
Type	Fixed TS	
U (kW/m^2 °C)	0.367	
Area (m^2)	2972.00	
Shell side		
Max. temp (°C)	35.58	
Pressure (bar and Pa)	52.76 (5.276×10^6 Pa)	
Material of construction	Carbon steel	
Tube side		
Max. temp (°C)	40.39	
Pressure (bar g and Pa g)	82.00 (8.2×10^6 Pa)	
Material of construction	Carbon steel	
Bare Module Cost (USD)	1,160,607	
Twister®		
Equipment tag name	**Op-100**	
Inlet pressure (bar and Pa)	80.0 (8.0×10^6 Pa)	
Outlet pressure (bar and Pa)	52.7 (5.27×10^6 Pa)	
Inlet temperature (°C)	−21.0	
Max. outlet temperature (°C)	−40.0	
Bare Module Cost (USD)	3,430,000	
Hydrate separator		
Equipment tag name	**Op-101**	
Bottom outlet temperature (°C)	20.0	
Max. gas outlet temperature (°C)	−37.6	
Bare Module Cost (USD)	1,430,000	

It should be mentioned that the methodology adopted for CAPEX evaluation (Turton et al., 2009) in the present work considers exclusively the impact of the number of equipments based on its cost (type, size and material of construction). The weight of the equipments and the penalty for an excess in the number of items of equipments is not embodied in the adopted approach, although their contribution to the total investment cost in offshore facilities is recognized.

Furthermore, it is reasonable to expect that a smaller equipment count should be advantageous, reducing the negative impact of items that dominate CAPEX in the Twister® case.

4.2. OPEX estimation

Operation costs are a function of labor, maintenance, utilities and raw material costs, as given by Eq. (1) (Turton et al., 2009).

$$OPEX = F1{*}CAPEX + F2{*}COL + F3{*}(RM + U) \quad (1)$$

where $F1$, $F2$ and $F3$ are cost factors, COL represents the costs of operation labor, RM represents the raw materials costs, and U

Table 5
Values for the cost factors.

Cost factors	TEG + JT/LTS	Twister®
*F*1	0.17	0.10
*F*2	2.64	2.49
*F*3	1.19	1.19

represents the utilities costs. The cost factors values for each scheme are presented in Table 5.

The lower cost factors used for Twister® Scheme is based on the fact that Twister® requires less maintenance since it has no rotating parts and does not require supply of chemicals.

4.2.1. Cost of operation labor (COL)

The COL is estimated according to the number of equipment pieces, which determines the number of operators required. The base salary considered was US$ 45,300/year, which is consistent with the Brazilian market. The number of operators (NOP) per shift is given by Eq. (2) (Turton et al., 2009).

$$\mathrm{NOP} = 4.5*(6.29 + 0.23*\mathrm{Neq})^{0.5} \qquad (2)$$

where Neq is the number of equipment in the plant (considers only the main equipment, such as compressors, heat exchangers, vessels, towers and reactors). Notice that the Turton et al. (2009) procedure foresees a minimum number of 12 operators (when Neq equals 0). This Neq, however, is not realistic for the Twister® Scheme, which was estimated based on similar plants in operation, set as 8 as the Scheme is less intensive in operators due to its simpler design.

4.2.2. Utilities costs

The considered utilities are:

- Cooling water: the required flow was calculated with UniSim® Design simulations. Cooling water price was adopted as US$ 0.0148/t (Turton et al., 2009);
- Fuel: fired heater is considered to burn natural gas, at a fuel price of US$ 2.11/GJ.

The value published by Turton et al. (2009) should reflect the US reality. However, in the present study, they were considered adjusted to comply with Brazilian values.

4.2.3. Raw materials costs

Chemicals are only required in the TEG + JT/LTS Scheme. TEG make-up flow rate was calculated with UniSim® Design simulations, and its price was adopted as US$ 1450/t TEG (http://www.alibaba.com/product-free/106947801/TEG_DEG_MEG.html). For the crude natural gas (before dew-pointing), a realistic cost value is given by the NG "supply cost", defined by the Canadian National Energy Board as "the present value of producing a gigajoule of natural gas over the life of a well" (NEB National Energy Board, September, 2008). The difference between the supply costs and the NG selling prices, referred to as NG price margin, determines the economic viability of the gas well in question. The non-associated NG production scenario was investigated, in which the supply cost is estimated at US$ 2.11/GJ, i.e., half the selling price of the NG (taken as US$ 4.22/GJ).

Table 6
Capital Expenditures (CAPEX) for Dew Point Control Technologies.

	Dew point control technology	
	TEG + JT/LTS	Twister®
Bare module costs		
Compressors, pumps and turbines	200,379	2,499,121
Compressors drivers	0	481,328
Vessels	1,087,437	1,045,754
Heat exchangers	366,073	1,160,607
Towers	2,324,053	0
Fired heater	642,874	0
Twister®	0	3,430,000
Hydrate separation vessel	0	1,430,000
JT valve	115,000	0
Total Bare Module Cost (US$)	4,735,816	10,046,810
CAPEX (US$)	8,382,394	17,782,854

Table 7
Operational expenditures (OPEX) and revenue for dew point control technologies.

	OPEX (US$/year)	Revenue (US$/year)	NPV (Millions US$/year)
TEG + JT/LTS	202,813,299	321,450,838	437
Twister®	210,370,247	336,557,008	460

NPV = Net Present Value.

4.2.4. Availability

The conventional plant is considered to operate continuously, with a yearly two-week shut down for maintenance. This is equivalent to 8424 hours/year operation. Availability for the Twister® case is estimated based on actual Twister® operating data. The Twister® technology has achieved over 99% availability with the Twister® system operating offshore with Sarawak Shell/Petronas in East Malaysia during over six years of operation, and onshore with Shell/NNPC in Nigeria, which has been operating continuously since April 2009.

4.3. Revenue estimation

The natural gas price is a function of its high heating value (HHV). The adopted price is US$ 4.22/GJ. Ethane and liquid products prices are given as a percentage of the crude oil price. The adopted values are: ethane, 30%; propane, 55%; butanes; 75%; pentanes and higher, 95% (OPIS, 2010). The conservative value of US$ 50/bbl was adopted.

4.4. Profitability evaluation

A realistic profitability evaluation requires, besides capital and operational costs, information on revenue, working capital, depreciation, taxation rate, discount rate, etc. Turton et al. (2009) indicate that, when comparing mutually exclusive investment alternatives, the alternative with the greatest positive net present value (NPV) should be chosen. NPV evaluation requires that the plant's cumulative discounted cash flows are calculated. The discounted cash flows are calculated for the first 20 years of a plant's project life, and the construction period was set as 2 years. The total fixed capital investment is represented by the grass roots cost (GRC), and is considered to be distributed during the construction period (60% at year 1 and 40% at year 2). Cost of land is not taken into consideration. Last, depreciation calculation is based on modified accelerated cost recovery system (MARCS), using a half-year convention and considering a 5-year recovery period (Turton et al., 2009). The discount and taxation annual rates were considered as 10% and 45%, respectively. The resulting NPVs are obviously affected by the adopted assumptions and simplifications. Nevertheless, the difference between the net present values (NPV) of the considered alternatives is expected to be appropriate for the comparison purposes of the present study.

To simplify the comparison between alternative technologies, the results are presented in an incremental basis, deducting the

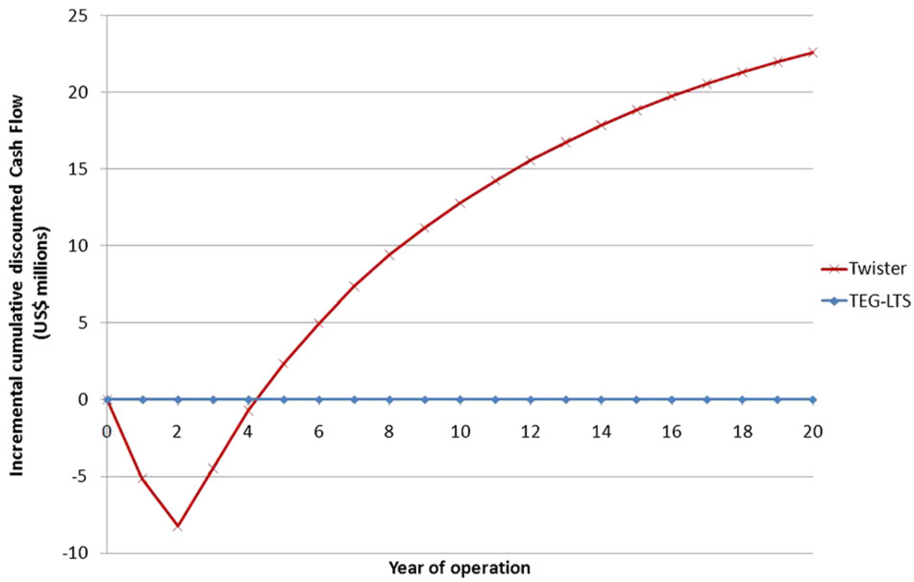

Fig. 3. Economic performance of process alternatives. Incremental cumulative discounted cash flow of Twister® compared to TEG + JT/LTS technologies for natural gas dew point control.

base case (the TEG + JT/LTS conventional technology) after tax discounted cash flow from the Twister® case result.

5. Results and discussion

The NPV and cash flow results are presented for non-associated NG production (NAP): NG supply cost is half the NG selling price (US$ 2.11/GJ).

Table 6 indicates that Twister® Scheme has higher CAPEX when compared to the TEG + JT/LTS scheme.

Table 7 shows that Twister® scheme OPEX is higher because it has more availability (NG supply costs per year are therefore higher). Concerning the calculated revenue, Twister® Scheme outperforms TEG + JT-LTS process due to its higher availability and superior liquid recovery of NGL, improving the economic results. For NPV results, Twister® scheme has a higher value than the TEG + JT/LTS process.

Fig. 3 displays the incremental cumulative discounted cash flow.

6. Conclusions

The present work has as main objective to compare two natural gas dew point technologies: Twister® technology and a conventional technology using TEG + JT/LTS, regarding technical and economic aspects. The first technical aspect observed was that the Twister® technology considerably reduces the number of major equipments (heat exchangers, vessels, towers, etc.) required in the NG processing plant. Since the Twister® scheme avoids the use of glycols (TEG), it eliminates the need for glycol regeneration units. It is worth noting that the higher CAPEX of the Twister® system is mainly due to the fact that Twister's® compression duty offers a 12.5 bar higher export pressure than the TEG + JT/LTS case. However, an increased production of NGL attributes economic advantage to the supersonic technology when NPV is considered (Fig. 3).

The value added associated with CAPEX depends on the volumes and the prices of the products sold. Twister® technology produces more NGL, by-products of natural gas production, which can increase revenue from the produced gas stream – NGLs typically compete with crude-oil-derived products, so their prices tend to follow crude-oil prices. Therefore increased NGL production presents added revenue, which allows early payback of the investment.

Furthermore, a key conclusion of the comparative study presented is that, within the bounds of uncertainty of the economic analysis, it is not possible to distinguish which of the two options is best. On the other hand, it does clearly set out that Twister® in such an application merits further consideration and indeed with a sensitivity assessment on the CAPEX (which is often the barrier to initial investment) may have a clearer advantage.

Last, another factor crucial to investment is depreciation. If a machine still produces well and without increasing maintenance costs after depreciation period, then continued production is still economically reasonable. Opportunity costs (the cost of not investing in new machinery) also must be considered. Due to the fact that Twister® demonstrated 99% uptime in over six years of commercial operation, near zero maintenance costs and inspection maximum once every six years, it is plausible to expect that this technology poses lower opportunity costs compared to traditional TEG + JT/LTS technology.

A final note concerns the glycol regeneration process, which includes the use of a stripping gas to obtain high purity lean TEG. The wet stripping gas (with a water content of 55% mol) is released to the atmosphere. Considering that the stripping gas may have BTX (Benzene–Toluene–Xylene) components, this emission issue is of relevancy when comparing natural gas processing alternatives.

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