OPTIMUM DESIGN OF TURBO-EXPANDER ETHANE RECOVERY PROCESS

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Abstract—*This paper explores methods for determining the optimum design of turbo-expander ethane (C_2) recovery processes, focusing on constrained maximum recovery (C-MAR), a new methodology. C-MAR—successor to the system intrinsic maximum recovery (SIMAR) methodology introduced recently—uses a set of curves developed to benchmark C_2 recovery applications based on the popular gas sub-cooled process (GSP) and external propane (C_3) refrigeration (–35 °C). Using the C-MAR curves, a process engineer can quickly determine the optimum design and estimate the performance and cost of various C_2 recovery opportunities without performing time-consuming simulations. Moreover, the C-MAR curves enable alternative process configurations to be compared against GSP performance.*

Keywords—*C-MAR, compressor, ethane, expander, refrigeration, SIMAR, turbo-expander*

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INTRODUCTION

Since its acceptance by the industry in the 1970s, the expander-based process has become the mainstay technology in ethane (C_2) recovery applications. [1] Despite the great technical and commercial success of this technology, a systematic methodology for determining the optimal system design has remained elusive until recently. Design optimization was approached as an art to be mastered; to this end, a new process engineer would typically spend several years gaining experience and acquiring the necessary expertise. The steep and frustrating learning curve was not conducive to extending this art beyond the province of process specialists to general engineers.

Recently, a methodology called SIMAR, which stands for system intrinsic maximum recovery, was described in papers presented at a key technical conference. [2, 3] These works and subsequent follow-up papers [4, 5] identified a systematic approach to arrive at the optimal design for a given feed stream.

Although SIMAR greatly facilitates the design procedures by reducing a two-dimensional (2-D) search to a single dimension, its reference case is a hypothetical scenario in which infinite amounts of refrigeration are available to the system. In many real cases, the refrigeration supply is limited and costly. Therefore, it is necessary to move to sub-SIMAR operations, and additional steps are required.

This paper presents a new approach to eliminate the aforementioned shortcomings of SIMAR. The new method is called C-MAR, which stands for constrained maximum recovery. C-MAR redefines the reference case by adopting the gas sub-cooled process (GSP), a well-known industrial design [6], as the benchmark case and by incorporating a fixed refrigeration temperature of –35 °C, the practical lower bound of propane (C_3) refrigeration circuits. Since this new reference case is a realistic industrial design, its results are more readily transferable to industrial applications (for example, cost estimates).

The technical background for the development of C-MAR is described in some detail in this paper. SIMAR methodology is discussed and illustrated. C-MAR's usefulness and applications are demonstrated in real cases using the Enhanced Natural Gas Liquid (NGL) Recovery Process[SM] (ENRP) [1, 7, and 8] (employing a stripping gas system) and the lean reflux process. [9]

TECHNICAL BACKGROUND FOR DEVELOPMENT OF C-MAR

Following a general categorization and discussion of expander-based C_2 recovery processes, SIMAR methodology is explored in this section. A scenario in which liquefied

ABBREVIATIONS, ACRONYMS, AND TERMS

1-D	one-dimensional
2-D	two-dimensional
C_1	methane
C_2	ethane
C_3	propane
C-MAR	constrained maximum recovery
DeCl	demethanizer column
ENRP	Enhanced NGL Recovery Process[SM]
GPA	Gas Processors Association
GPM	gallons per Mscf
GSP	gas sub-cooled process
JT	Joule-Thomson
LNG	liquefied natural gas
LRP	lean reflux process
MMscfd	million scf per day
Mscf	thousand scf
NG	natural gas
NGL	natural gas liquid
SB	side reboiler
scf	standard cubic feet
SIMAR	system intrinsic maximum recovery
VF	vapor fraction
XPDR	expander; expressed as XPDR in conjunction with XPDR 1, 2, or 3 process configuration categories

natural gas (LNG) is used as feed is described. Since LNG contains abundant refrigeration, the SIMAR reference case can be approximated well. SIMAR curves are compared for the expander (XPDR) 1, XPDR 2, and XPDR 3 process configuration categories. The discussion then examines typical results when the feed is shifted from LNG to natural gas (NG), based on the XPDR 3 category.

C_2 Recovery Processes

Figure 1 shows a generalized scheme for C_2 recovery based on expander technology. The process is intended to strip the inlet NG of its heavier components. The residue gas is recompressed and returned to the pipeline. Sweet, dry inlet NG flows through an inlet chilling section, where the gas is chilled to a

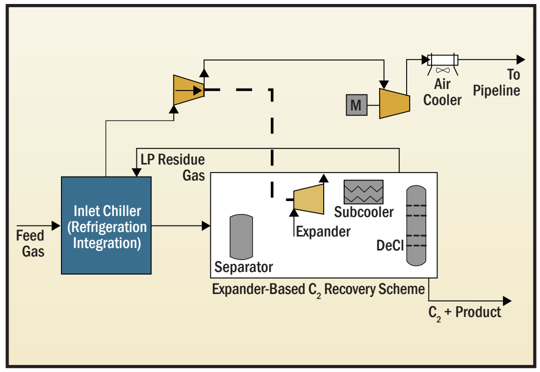

Figure 1. Generalized Gas Processing Scheme for C_2 Recovery

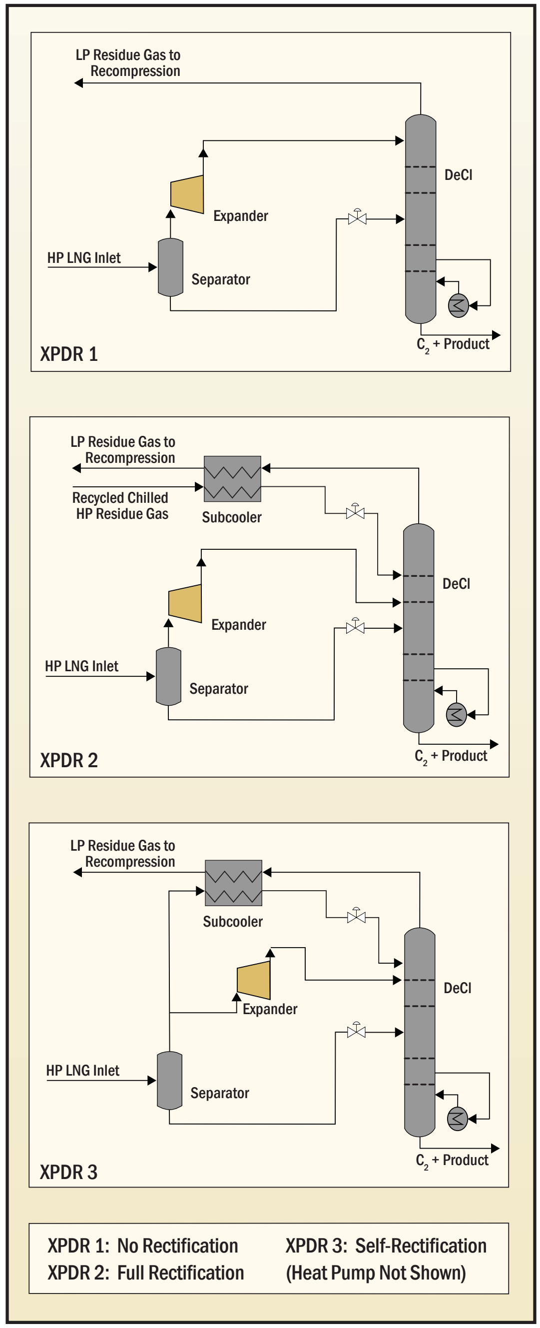

Figure 2. Categorizing Expander-Based Schemes

suitable level before entering the heart of the plant: an expander-based C_2 recovery system. The refrigeration of the inlet chilling section is mainly provided by the returning residue gas and is supplemented by side-draws from the demethanizer column (DeCl) (side-draws not shown in Figure 1). Depending on actual requirements, external refrigeration may be required (not shown in Figure 1).

The main components of the expander section include a separator, an expander, and a DeCl. A subcooler is usually provided for improved refrigeration integration in the low temperature regions. The exact design of this section is an art.

Figure 2 provides further details on the expander section. Following earlier practice, the expander schemes are grouped into three process configuration categories: separator top not rectified (XPDR 1), separator top fully rectified (XPDR 2), and separator top self-rectified (XPDR 3). A fourth category, heat pumps, is not shown but is described shortly. The XPDR 3 is the well-known GSP in the industry, which uses a small portion of the non-condensed vapor as the top reflux to the demethanizer, after substantial condensation and sub-cooling. The main portion, typically in the range of 65%–70%, is subjected to turbo expansion as usual.

Configurations with heat pumps are discussed separately because: (1) a heat pump can be an enhancement to any of the aforementioned three categories, and (2) a heat pump moves heat from low to high temperature and changes the temperature distribution in its base configuration. Its working principle is different from that of the three categories. A heat pump design can be recognized by the use of a compressor; a cooler for rejecting heat to a high temperature sink, a Joule-Thomson (JT) valve, or a second expander; and, optionally, a second exchanger to take heat from the low temperature source. **Figure 3** depicts the ENRP as an example.

In the Figure 3 configuration, a side-draw liquid stream from the bottom of the demethanizer is expanded to generate refrigeration. This stream is then heated by indirect heat exchange with inlet gas to generate a two-phase stream. The two-phase stream is flashed in a separator. The flashed vapor is compressed and recycled to the demethanizer as a stripping gas. The flashed liquid stream can be mixed with other NGL product streams or returned to the column. This heat pump effectively moves heat from the inlet stream to the bottom of the column.

The main features of this novel design center on the fact that the stripping gas (1) enhances relative volatility ratios and NGL recovery levels, and (2) lowers the column temperature profile and makes heat integration easier.

The lean reflux process, which belongs to the XPDR 3 category, was developed to achieve high recovery levels of C_2 in an NG feed without

The XPDR 3 process configuration category is the well-known gas sub-cooled process in the industry.

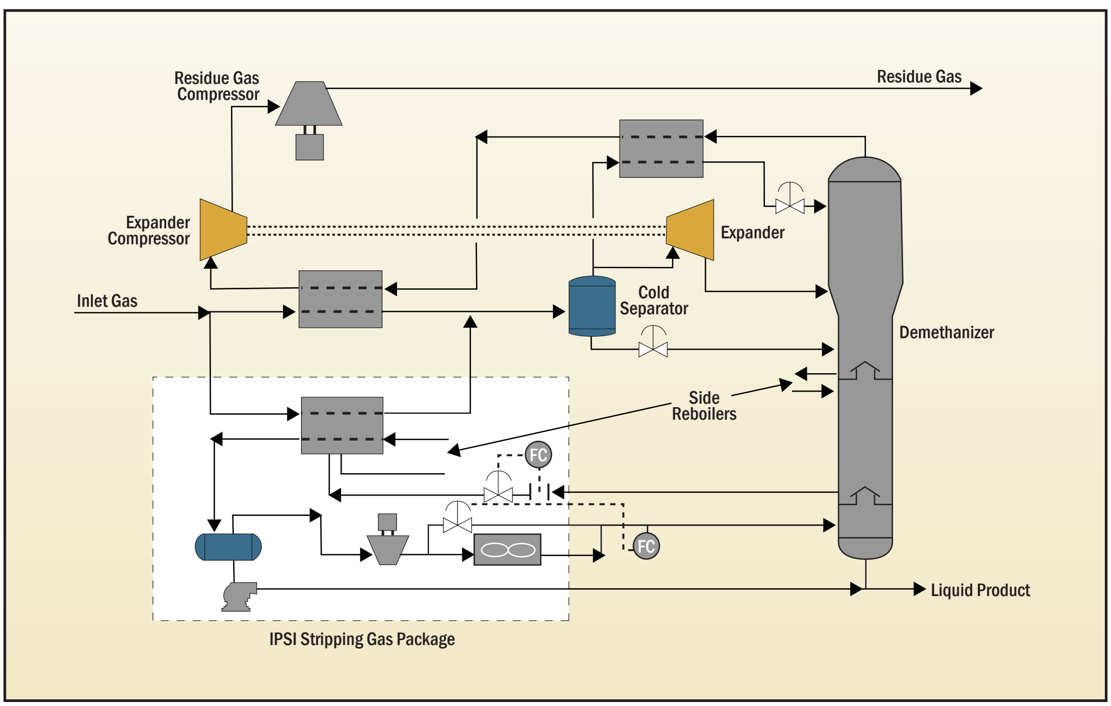

Figure 3. Enhanced NGL Recovery Process

The entire inlet pre-chilling section can be eliminated when the NG feed is replaced by LNG.

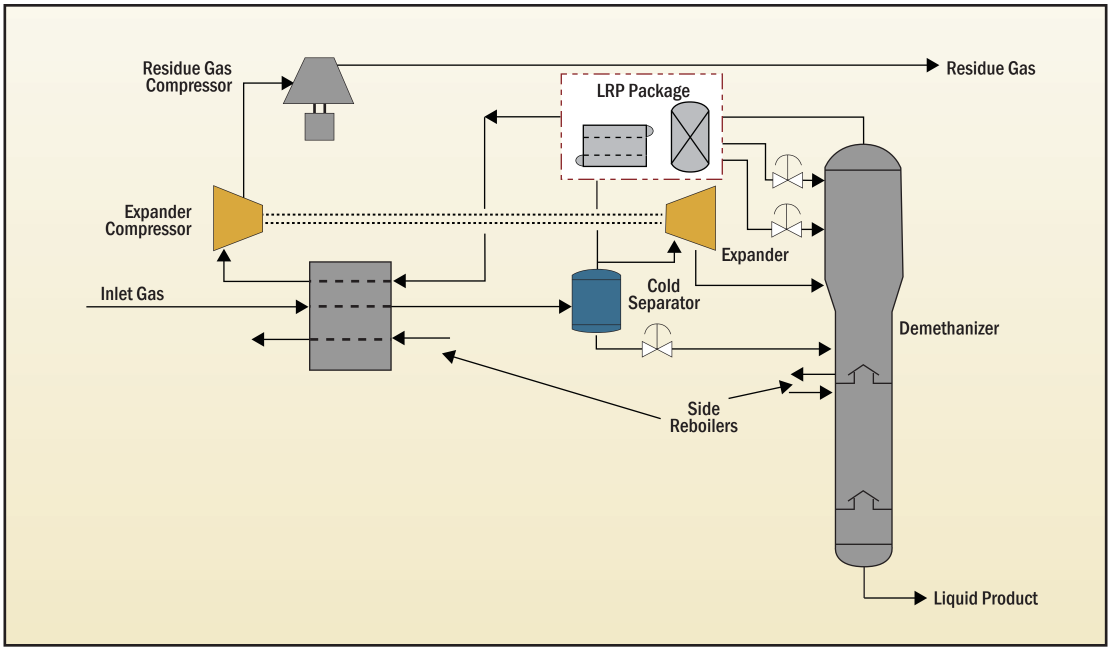

Figure 4. Lean Reflux Process

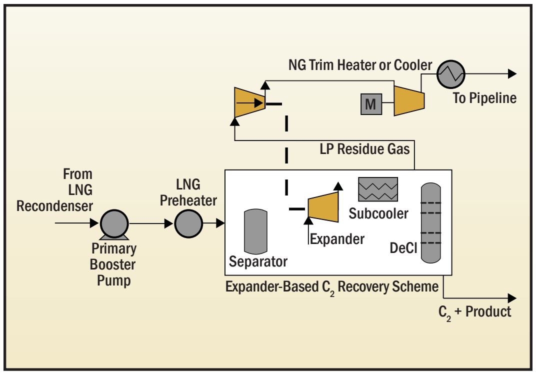

Figure 5. Generalized Processing Scheme for C_2 Recovery with LNG as Feed

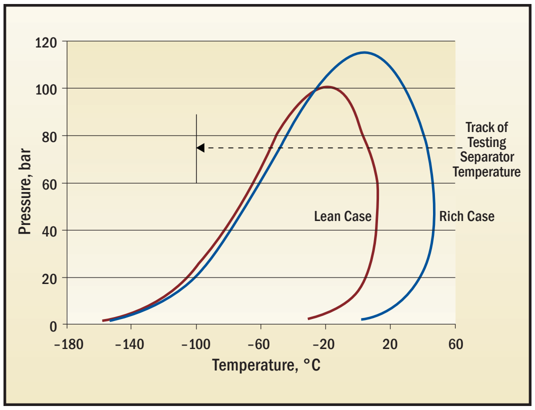

Figure 6. Phase Envelope of Inlet Gas

adding substantial amounts of recompression and/or external refrigeration power. This process uses a slipstream from the cold separator or feed gas to generate an essentially C_2-free stream as a lean reflux to the demethanizer (see **Figure 4**).

Introducing a lean reflux considerably reduces equilibrium loss, thereby leading to high C_2 recovery while maintaining the demethanizer at a relatively high operating pressure. The process overcomes deficiencies in the commonly used gas sub-cooled reflux process in which C_2 recovery levels are ultimately restricted to approximately 90% due to equilibrium loss, or otherwise demand a lower demethanizer pressure and a higher recompression and/or refrigeration horsepower.

SIMAR Methodology

C_2 Recovery with LNG as Feed

For the sake of easy visualization, **Figure 5** depicts a scenario wherein the NG feed shown in Figure 1 is replaced by LNG. The major difference resulting from this change is the fact that the entire inlet pre-chilling section can be eliminated when LNG is used as the feed. The refrigeration in the residue gas can be retained, thus dramatically reducing the recompression power.

A SIMAR curve can be constructed following a few simple steps. The process starts from a relatively high temperature at a reasonable pressure level, as shown in **Figure 6**. The track of testing temperatures penetrates through the two-phase region and ends at an arbitrarily chosen level of –100 °C. The fluid remains liquid at and below this temperature level. Once the temperature reaches a certain point, the column's operating limits are exceeded and the column no longer converges.

Figure 7 plots the C_2 recovery level against the test separator temperature for the three categories of process configurations. The DeCl operating pressure is 22 bar. Both XPDR 1 and XPDR 2 shows a monotonic trend of improvement as the temperature decreases. This trend continues even after the inlet gas is totally liquefied, when the expander is replaced by a JT valve and no expansion work is recovered. XPDR 3 shows a different trend, however. As the separator temperature decreases, the C_2 recovery reaches a maximum value and decreases. In other words, too much refrigeration at the separator may hurt the C_2 recovery.

For XPDR 3, the SIMAR is defined as the maximum of the curve. For XPDR 1 and XPDR 2, the SIMAR is defined at the temperature level where the separator fluid has 30% vapor fraction (VF). The choice of 30% is based on a practical consideration that no expanders would be installed if the gas flow is below this level.

A SIMAR curve is the collection of all the SIMAR conditions that cover the entire range of DeCl operations of interest. **Figure 8** depicts typical SIMAR curves corresponding to the three XPDR categories. The characteristics of the three categories can be observed. The reflux stream makes XPDR 2 more efficient than XPDR 1 throughout the entire range, which corresponds to the operating pressures of DeCl from 17 to 42 bar. The efficiencies of XPDR 2 and XPDR 3 are comparable, while each has its advantages over a certain span.

C_2 Recovery with NG as Feed

Figure 9 shows typical results, based on the XPDR 3 category, when the feed is shifted from LNG to NG. Since the refrigeration in the residue gas must be recovered to cool the inlet gas, the recompression power increases significantly by this shift in feed. A big gap is apparent between the two thin curves on the left.

In addition to the recovered refrigeration in the residue gas, external refrigeration may also be needed. When this is true, the compression power required in the external refrigeration power should be added to the aforementioned recompression power to form the total compression power. The two thick curves on the right represent the recompression duties by using one or two side reboilers (SBs). The gap between the curves of total compression and the recompression curves on the left in Figure 9 represents the external refrigeration. Using two SBs reduces the external refrigeration because of improved refrigeration integration in the pre-chilling section.

The gap narrows in Figure 9 when the DeCl operating pressure decreases, indicating the decreased demands for external refrigeration. As can be observed in Figure 8 as well, when the DeCl operating pressure decreases or C_2 recovery level increases, the need for external refrigeration also decreases. The expander provides more refrigeration for process needs at lower DeCl operating pressures.

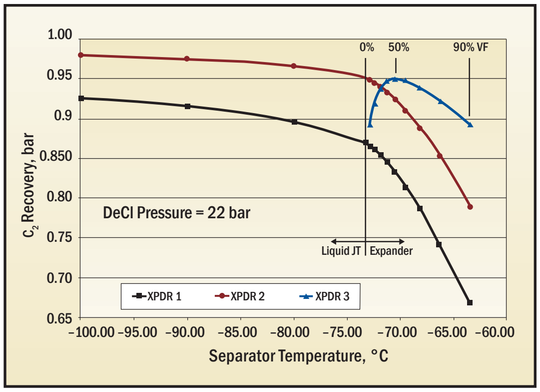

Figure 7. Two Types of Behavior for Different Categories

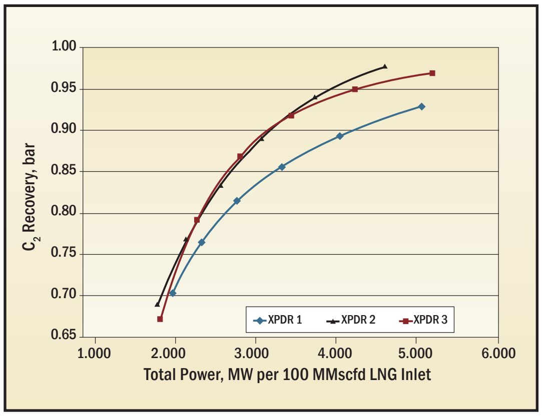

Figure 8. Comparing SIMAR Curves for Three XPDR Categories

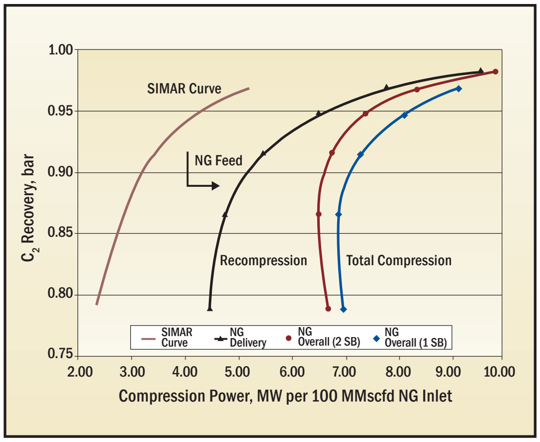

Figure 9. Comparing Compression Duties and Impact of Side Reboilers Based on XPDR 3

In addition to the recovered refrigeration in the residue gas, external refrigeration may also be needed.

FEED GAS COMPOSITION AND SIMULATION PARAMETERS

Table 1 lists two feed gas compositions used in this paper, rich case and lean case. They represent different richness in C_2+ components. The richness of a gas sample is reflected in its C_2+ or C_3+ components, expressed in gallons per Mscf (GPM). The GPM value for the rich case is 5.71 and for the lean case is 2.87. The phase envelopes corresponding to the two compositions are shown in Figure 6. The richer the gas, the wider its envelope becomes. The raw gas supply is 300 MMscfd (dry basis).

The two feed gas compositions used in this paper, rich case and lean case, represent different richness in C_2+ components.

All simulations in this paper are performed using Aspen HYSYS® 3.2. **Table 2** lists pertinent parameters. The delivery pressure to the pipeline is similar to the inlet pressure. Two temperature levels of heat sinks are defined: The high temperature represents air coolers, and the low temperature represents the external refrigeration temperature supplied by two-stage C_3 compressor loops.

Table 1. Feed Gas Compositions

Components	Rich Case, mole %	Lean Case, mole %
Nitrogen	0.315	0.750
CO_2	0.020	0.217
Methane	79.550	88.910
Ethane	10.600	4.950
Propane	5.470	3.090
i-Butane	0.926	0.442
n-Butane	1.690	0.894
i-Pentane	0.468	0.224
n-Pentane	0.478	0.221
n-Hexane	0.295	0.300
n-Heptane	0.132	0.000
n-Octane	0.060	0.000
n-Nonane	0.020	0.000
GPM for C_2+	5.710	2.870

Table 2. Simulation Parameters Used in This Paper

Parameter	Value
Inlet Temperature, °C	27
Inlet Pressure, bar	69 or 55
Send-Out Residual Gas Temperature, °C	38
Send-Out Residual Gas Pressure, bar	74 or 55
Number of Trays in DeCl	16
DeCl Operating Pressure, bar	17 to 37
Composition Ratio of C_1 to C_2 in DeCl Bottom Product	0.015
High-Temperature Sink, °C	38
Low-Temperature Sink, °C	–35

C-MAR METHODOLOGY

Principal Elements and Assumptions

C-MAR methodology includes two major elements:

- XPDR 3 process configuration as the benchmark model (the GSP, which is well-known in the industry)
- Fixed refrigeration temperature of –35 °C

For purposes of conceptual discussions, the pre-chiller is simulated using one integrated exchanger, which handles all streams including inlet gas, returning residue, SBs, and external refrigeration. Only the minimum amount of refrigeration is added to satisfy the refrigeration balances. The intent is to minimize the additional compression work. Unless specified otherwise, two SBs in an integrated exchanger are assumed. External refrigeration implies closed-loop designs of C_3 circuits.

The results of C-MAR methodology, including the characteristics of resultant curves and their relation to SIMAR, are examined below. The paper concludes with a discussion of the C-MAR curve in relation to the ENRP and the lean reflux process.

C-MAR Methodology Results, Curves, and Relation to SIMAR

Figure 10 shows the C_2 recovery versus separator temperature for the lean case. As the separator temperature decreases, the C_2 recovery shows a maximum at about –57 °C for all curves.

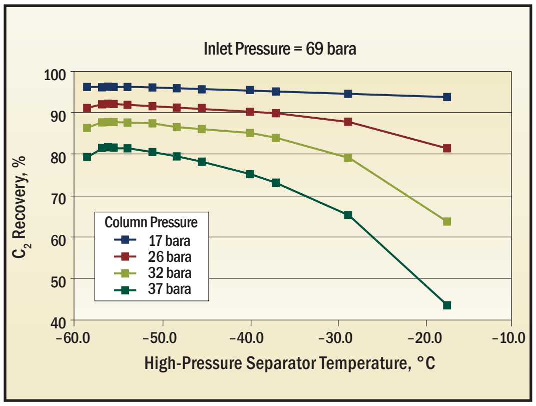

Figure 10. Determining Maximum C_2 Recovery Using C-MAR Methodology (Lean Case)

This pattern bears similarities to the XPDR 3 curve in Figure 7, indicating the existence of the maximum behavior for the GSP configuration under different constraints, e.g., DeCl pressure and refrigeration availability. Physically, separator temperatures that are too cold result in C_1 condensation. The DeCl reboiler would input extra heat to prevent excessive C_1 loss from the bottom. The net result is the increased C_2 loss in the residue gas. To calculate the power requirement for the external refrigeration, this discussion assumes that the external refrigeration is from two-stage C_3 compressor circuits with evaporator temperature at –41 °C and refrigerant condensing temperature at 49.5 °C. From the Gas Processors Association (GPA) data book [9], the value for the power can be obtained.

It should be noted that the maximum C_2 recovery using C-MAR occurs at a higher temperature (–57 °C) than that of SIMAR (about –70 °C). Using C-MAR, the constraint in refrigeration prevents the separator temperature from decreasing further. Using SIMAR, the constraint is imposed last by forcing the selection into the sub-SIMAR region. Either approach would lead to similar results.

Figure 11 shows trends for the rich case similar to those described above. With the separator temperature further decreasing below some point, the C_2 recovery decreases due to C_1 condensation.

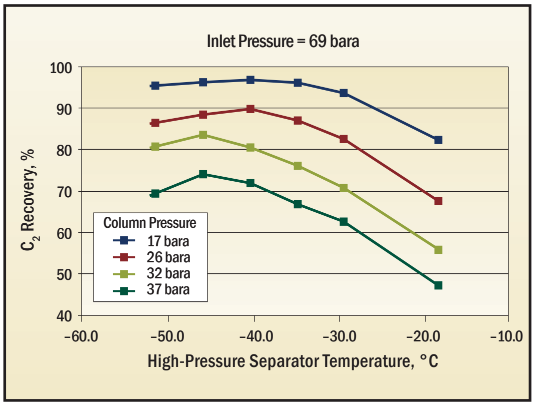

Figure 11. Determining Maximum C_2 Recovery Using C-MAR Methodology (Rich Case)

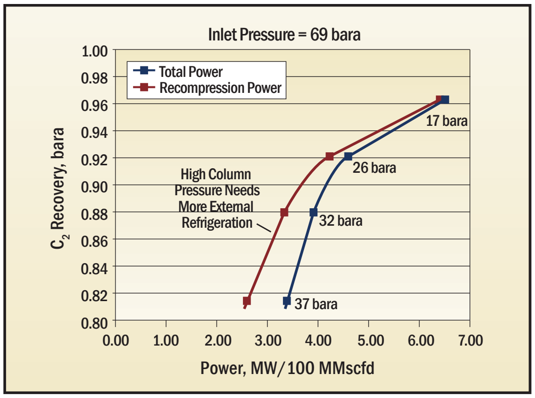

Figure 12. Operation Curves Determined by C-MAR Methodology (Lean Case)

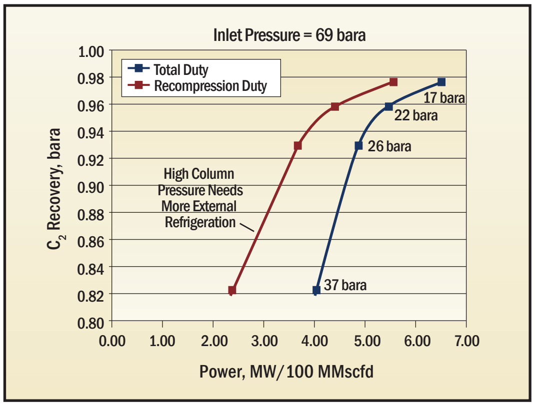

Figure 13. Operation Curves Determined by C-MAR Methodology (Rich Case)

Figures 12 and 13 depict operation curves determined by C-MAR methodologies for lean and rich cases. As anticipated, the trends are similar to those of SIMAR shown in Figure 9. The gap between recompression and total power represents the external refrigeration. Again, as anticipated, the rich case demands more refrigeration duties than the lean case at the same inlet pressure. And with the decrease of DeCl pressure, C_2 recovery increases and less external refrigeration is needed because the relative volatility is greater at lower column pressure.

Figures 14 and 15 show C-MAR curves at two inlet pressure levels and two inlet gas GPM values. Figure 14 is for total power and Figure 15 is for recompression power only. In Figure 14, rich feed gas needs more total power (or more power than lean feed gas) because more external refrigeration is needed to condense heavy components in rich feed gas in the DeCl into liquid product. But the total power (or the power of lean and rich feed gas) will be about the same, or the lean case can even require more power than the rich case, at high C_2 recovery level. The reason is that more recompression power is needed for lean feed gas to handle the larger residual gas flow.

As can be seen from Figure 15, for both 69 bara and 55 bara inlet pressure cases, rich feed gas requires more recompression power at low C_2 recovery level than lean feed gas. But at high C_2 recovery level, lean feed gas needs more

It should be noted that the maximum C_2 recovery using C-MAR occurs at a higher temperature than that of SIMAR.

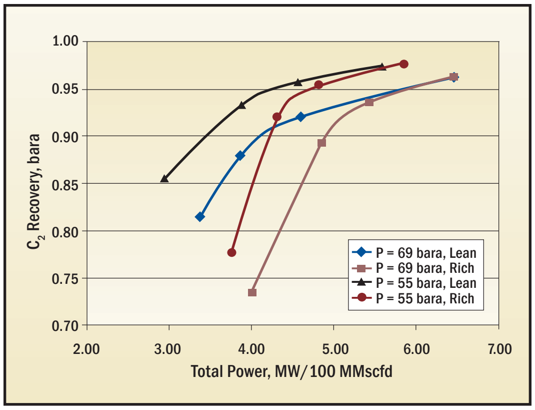

Figure 14. Impact of Inlet Pressure and Richness of Feed Gas on C-MAR Curves (Total Power)

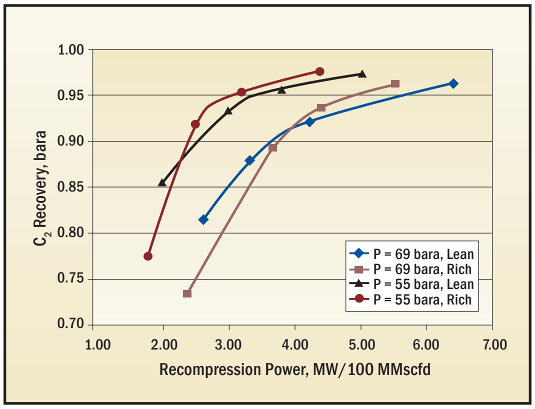

Figure 15. Impact of Inlet Pressure and Richness of Feed Gas on C-MAR Curves (Recompression Power)

At a given feed gas pressure and inlet gas richness, it is possible to develop general correlations to interpolate required duties for different feed gases.

recompression power than rich feed gas. At low C_2 recovery level or high DeCl pressure, to obtain the same C_2 recovery, rich feed gas needs lower DeCl pressure to create higher relative volatility, which leads to a higher recompression power requirement. But at high C_2 recovery or low DeCl pressure, either lean or rich feed gas has high relative volatility, while lean feed gas requires a greater flow rate to achieve the same C_2 recovery. This explains the larger recompression power requirement of the lean case at high C_2 recovery. It is easy to understand that high pressure feed gas (69 bara) needs more recompression power than low pressure feed gas (55 bara) because of the assumption that the inlet pressure is the same as the delivery pressure. As mentioned earlier, the external refrigeration requirement can be deduced from the curves in Figures 14 and 15 because it is simply the difference between the total power and the recompression power.

Examining the regularities between the rich and lean cases in Figure 14 leads to an important conclusion. At a given feed gas pressure and a given richness of inlet gas (i.e., GPM value), it is possible to develop general correlations to interpolate required duties for different feed gases. Since the curves in Figure 14 represent the maximum C_2 recoveries achievable by the GSP with realistic refrigeration supplies, the interpolated results provide expedient estimates in feasibility investigations.

In addition, since the GSP has practically become a benchmark configuration in this field, the curves in Figure 14 acquired by C-MAR methodology can be used to evaluate different process configurations. The following subsection provides an illustration using the ENRP and the lean reflux process as examples.

C-MAR Curve and the ENRP and Lean Reflux Process

In **Figure 16**, the ENRP and lean reflux process are compared with the C-MAR curve. Either of the two processes or both in combination can achieve higher C_2 recovery at lower power than the C-MAR curve or the highest recovery by the GSP. Improvement can be expected from the two processes. Obviously, the ENRP can expend less power to achieve higher C_2 recovery than the GSP, and the lean reflux process can achieve high C_2 recovery with less power than the GSP. Combining the ENRP and lean reflux process is better because the combination can achieve high C_2 recovery with less power.

In **Figure 17**, the C-MAR curve is compared with the recovery and power for the Pascagoula NGL plant, which uses the GSP. The point plotted for Pascagoula falls on the right side of the C-MAR curve and is quite close to it. This shows that the design of this plant can achieve a C_2 recovery close to the maximum achievable by the GSP.

Another example shown in **Figure 18** is the Neptune II NGL plant, which uses the ENRP in its design. For comparison, the point for the GSP without refrigeration is also marked. The

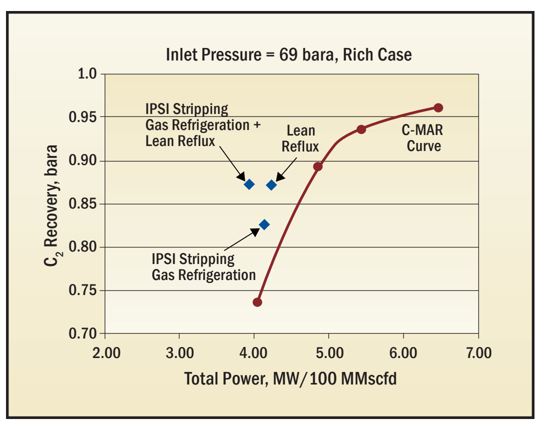

Figure 16. Comparison of ENRP and Lean Reflux Process with C-MAR Curve

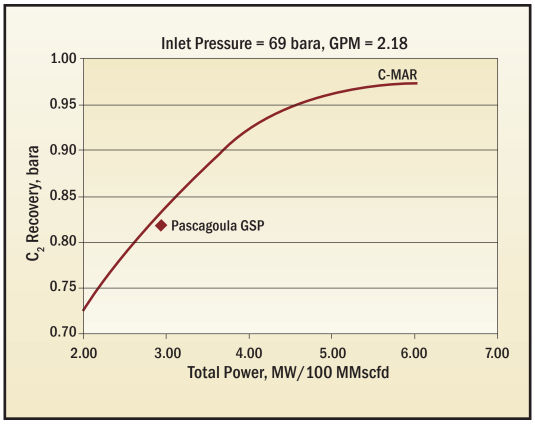

Figure 17. C-MAR with Pascagoula GSP

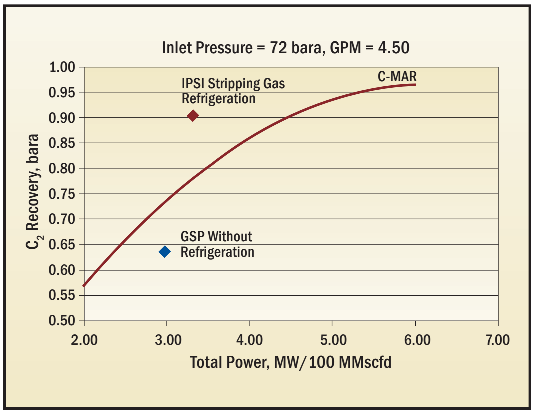

Figure 18. C-MAR with Neptune II

point for the ENRP is on the left side of the C-MAR curve and shows the improvement realized from use of the ENRP over the GSP. GSP without refrigeration is some distance away from the C-MAR curve on the right side; the C_2 recovery is limited because no external refrigeration is supplied.

CONCLUSIONS

In this paper, the authors explored the technical background for developing design optimization methodologies for turbo-expander C_2 recovery processes. Methods and processes to optimize design and improve system performance were examined and illustrations presented. The discussion and data support the following conclusions, briefly summarized below:

- C-MAR is a valuable tool and a new approach that eliminates the shortcomings of the SIMAR methodology by expediently determining the maximum C_2 recovery and compression power based on use of the well-known GSP.
- Using the C-MAR curves enables optimum design to be determined and initial cost estimates to be prepared for project scoping, avoiding the need to perform intricate simulations.
- Separately, use of the stripping gas process (ENRP) and the lean reflux process can significantly improve the system performance.
- A combination of the aforementioned two processes further improves the system performance. ■

C-MAR is a valuable tool and a new approach that eliminates the shortcomings of SIMAR methodology and enables optimum design to be determined.

TRADEMARKS

Aspen HYSYS is a registered trademark of Aspen Technology, Inc.

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BIOGRAPHIES

Wei Yan has more than 10 years of experience in the oil and gas industry. He joined IPSI LLC[1] as a senior process engineer in 2006 to work on design and technology development for LNG and natural gas processing projects.

Before joining IPSI LLC, Dr. Yan worked at Tyco Flow Control Co. as an application engineer focused on new flow-control product development. He also served as a process engineer for China Huanqiu Chemical Engineering Corp., where he worked on the process design of petrochemical projects. Previously, as a research assistant at Rice University, Dr. Yan focused on the foam-aided alkaline-surfactant-enhanced oil recovery process.

Dr. Yan is a member of the Society of Petroleum Engineers and the American Institute of Chemical Engineers.

Dr. Yan holds a PhD from Rice University, Houston, Texas, and a Bachelor's degree from Tianjin University, China, both in Chemical Engineering.

Lily Bai, a senior process engineer with IPSI LLC, has more than 10 years of experience in research, process design, and development in chemicals, petrochemicals, gas processing, and LNG. Dr. Bai works on the Wheatstone LNG project and is responsible for process simulation. The Wheatstone facility, to be located on the northwest coast of mainland Australia, will have initial capacity of at least one 5 million-ton-per-annum LNG production train.

Before her current assignment, Dr. Bai worked on projects such as Angola LNG, Santos Gladstone LNG, and Atlantic LNG (Train 4) reliability. Her responsibilities included process simulation and preparation of process flow diagrams and equipment datasheets.

Dr. Bai holds a PhD from Rice University, Houston, Texas, and MS and BS degrees from Tianjin University, China, all in Chemical Engineering.

Jame Yao has 28 years of experience in the development of gas processing and LNG technologies. As vice president of IPSI LLC, Dr. Yao is responsible for all IPSI/Bechtel process design/simulation and development in cryogenic gas processing, nitrogen rejection, and LNG technology. He holds several patents in the field.

Dr. Yao joined International Process Services, Inc., the predecessor to IPSI LLC, in 1986 as a senior process engineer. During his tenure with IPSI, he has co-invented several processes for the cryogenic separation and liquefaction of N_2, He, LNG (methane), and other light hydrocarbons. Previously, Dr. Yao worked as a member of the worldwide Technology Center for Gas Processing of DM International (Davy McKee) in Houston, Texas.

Dr. Yao performed graduate study/research at Purdue University related to the measurement and prediction of thermodynamic properties of cryogenic gas mixtures. This work enabled him to co-invent several processes for the separation and processing of natural gas. He also contributed to the design of gas plants in Australia, New Zealand, Venezuela, the UK, North Sea, Norway, and the United States.

Dr. Yao is the author of more than 20 technical publications and holds more than 15 patents. He is a member of the American Institute of Chemical Engineers.

[1] Bechtel affiliate IPSI LLC, based in Houston, Texas, was formed in 1986 to develop technology and provide conceptual/front-end design services for oil and gas production and processing facilities as well as for engineering, procurement, and construction companies.

Dr. Yao holds PhD and MS degrees from Purdue University, West Lafayette, Indiana, and a BS degree from National Taiwan University, Taipei, all in Chemical Engineering.

Roger Chen has more than 30 years of experience in research, process design, and development in gas processing and in oil and gas production facilities. As a senior vice president of IPSI LLC, he is responsible for process design and development for gas processing facilities.

Dr. Chen designed the Enterprise Neptune II natural gas plant in Louisiana, constructed to match the capacity of Neptune I. He used an IPSI patent process in the design. Dr. Chen also has served as the technical auditor for several LNG projects, including Darwin in Australia, Zaire Province in Angola, and BG Egyptian in Idku, Egypt.

Previously, Dr. Chen was senior process engineer for IPSI. In this role, he initiated the process design for BG's Hannibal gas processing plant located near Sfax, Tunisia. Dr. Chen also has served as a chief process engineer for IPSI, with a focus on the BG Pascagoula liquid recovery facility, part of the 1.5-billion-cubic-feet-per-day Pascagoula natural gas processing plant in Mississippi. His activities included process design and startup assistance.

Dr. Chen has been a member of the American Institute of Chemical Engineers and the American Chemical Society for more than 40 years, and the Gas Processors Association Research Steering Committee for 8 years. He holds 10 patents and has authored more than 30 technical publications.

Dr. Chen holds PhD and MS degrees from Rice University, Houston, Texas, and a BS degree from National Taiwan University, Taipei, all in Chemical Engineering.

Doug Elliot, a Bechtel Fellow and a fellow of the American Institute of Chemical Engineers, has more than 40 years of experience in the oil and gas business, devoted to the design, technology development, and direction of industrial research. He is president, chief operations officer, and co-founder (with Bechtel Corporation) of IPSI LLC.

Before helping establish IPSI, Dr. Elliot was vice president of Oil and Gas for DM International (Davy McKee). He started his career with McDermott Hudson Engineering in 1971 following a post-doctoral research assignment under Professor Riki Kobayashi at Rice University, where he developed an interest in oil and gas thermophysical properties research and its application.

Dr. Elliot has authored or co-authored more than 65 technical publications and holds 12 patents. He served on the Gas Processors Association Research Steering Committee from 1972 to 2001 and as chairman of the Gas Processors Suppliers Association Data Book Committee on Physical Properties. Dr. Elliot also served as chairman of the South Texas Section and director of the Fuels and Petrochemical Division of the American Institute of Chemical Engineers and is currently a member of the PETEX Advisory Board.

Dr. Elliot holds PhD and MS degrees from the University of Houston, Texas, and a BS degree from Oregon State University, Corvallis, all in Chemical Engineering.

Stanley Huang is a staff LNG process engineer with Chevron Energy Technology Company in Houston, Texas. His specialty is cryogenics, particularly as applied to LNG and gas processing. Since 1996, Dr. Huang has worked on many LNG baseload plants and receiving terminals. He has also fostered process and technology improvements by contributing more than 20 publications and corporate reports.

Before joining Chevron, Dr. Huang worked for IPSI LLC and for KBR, a global engineering, construction, and services company supporting the energy, petrochemicals, government services, and civil infrastructure sectors.

By training, Dr. Huang is an expert in thermodynamics, in which he still maintains a keen interest. After leaving school, he worked for Exxon Research and Engineering Company as a post-doctorate research associate. Dr. Huang then worked for DB Robinson and Associates Ltd. in Alberta, Canada, a company that provides phase behavior and fluid property technology to the petroleum and petrochemical industries. He contributed more than 30 papers and corporate reports before 1996, including one on a molecularly based equation of state called SAFT, which is still popular in polymer applications today.

Dr. Huang holds PhD and MS degrees in Chemical Engineering and an MS in Physics, all from Purdue University, West Lafayette, Indiana, and a BS in Chemical Engineering from National Taiwan University, Taipei, Taiwan. He is a registered professional engineer in the state of Texas.