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RETROFIT OF THE AMERADA HESS SEA ROBIN PLANT FOR VERY HIGH ETHANE RECOVERY

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ABSTRACT

The Sea Robin cryogenic gas processing plant located near Henry, Louisiana was originally built in 1972 to process 900 MMSCFD of lean offshore production gas from the Sea Robin Pipeline at an ethane recovery level of 44%. The inlet flow rate and composition, inlet and delivery pressures, and plant operators have changed over the years. After Amerada Hess Corporation (AHC) became operator in 2000, a plan was conceived for increasing the plant profitability by retrofitting the plant to increase the ethane recovery to 98% at a design flow rate of 600 MMSCFD. The process design used in the retrofit along with the unique process equipment and compression changes which were made to the plant are discussed.

The original plant was a "first generation" cryogenic turboexpansion process design, using two parallel shell and tube exchanger and expander trains and a single demethanizer column. The residue gas compression was provided by a gas turbine driven compressor and an identical parallel steam turbine driven compressor which used the steam generated by a waste heat boiler installed on the gas turbine.

Two Ortloff "third generation" process designs were implemented as part of the retrofit. The Recycle Split Vapor (RSV) design was combined with the Carbon Dioxide Control (CDC) technology to allow AHC to achieve the desired ethane recovery levels without CO_2 freezing and without exceeding the NGL specification for CO_2 content. Nearly all of the existing equipment was reused in the retrofit process design. The modified plant is a success and the ethane recovery is limited only by the CO_2 content of the inlet gas.

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INTRODUCTION

The Sea Robin Plant was originally built in 1972 to process gas produced into the Sea Robin Pipeline System near Henry, Louisiana. The original design basis was for 900 MMSCFD (15.025 PSIA and 60°F) of 1.209 GPM gas at an inlet pressure of 650 PSIG. The ethane recovery (limited by residue gas minimum heating value) was 44% yielding 18,246 barrels per day (BPD) of ethane plus mix in an ethane recovery mode and 11,849 BPD of propane plus in an ethane rejection mode. The process design was a first generation, single stage expander design using parallel exchanger and expansion trains. A reflux exchanger and reflux pumps were included in the original design for ethane rejection operation. The residue compression system consisted of two parallel 50% centrifugal compressors, one gas turbine driven and one steam turbine driven using steam generated from waste heat from the gas turbine. A schematic of the original process design is shown in Figure 1.

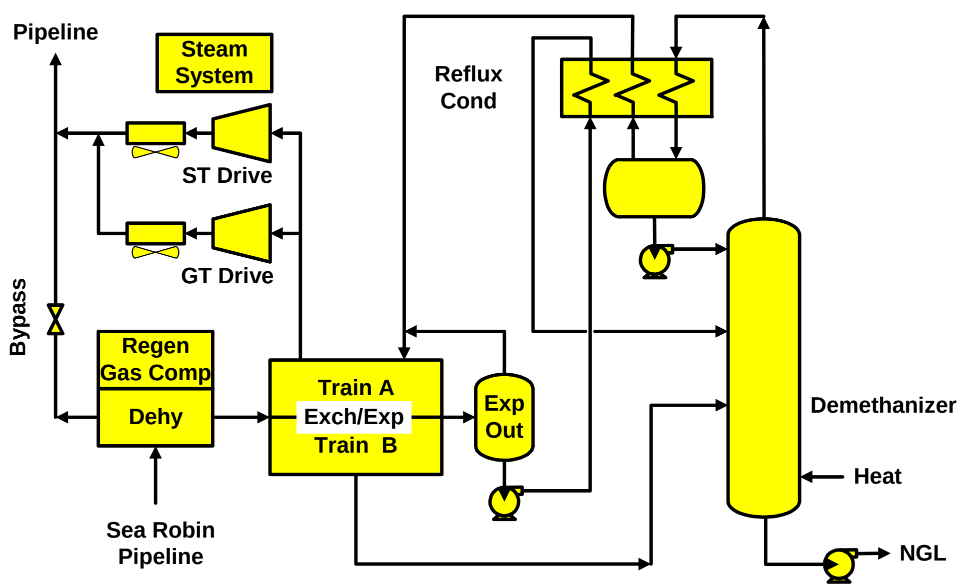

Figure 1 – Sea Robin Original Process Design

Note the two parallel exchanger/expander trains in the flow diagram. Each train consists of gas/gas exchangers, a liquid feed preheater exchanger, a cold separator, and an expander/compressor. This arrangement was used in the original design because of the available expander/compressor frame sizes and the multiple shells required for the large gas/gas exchangers. Figure 2 on the next page, shows the arrangement within each of the original exchanger/expander trains.

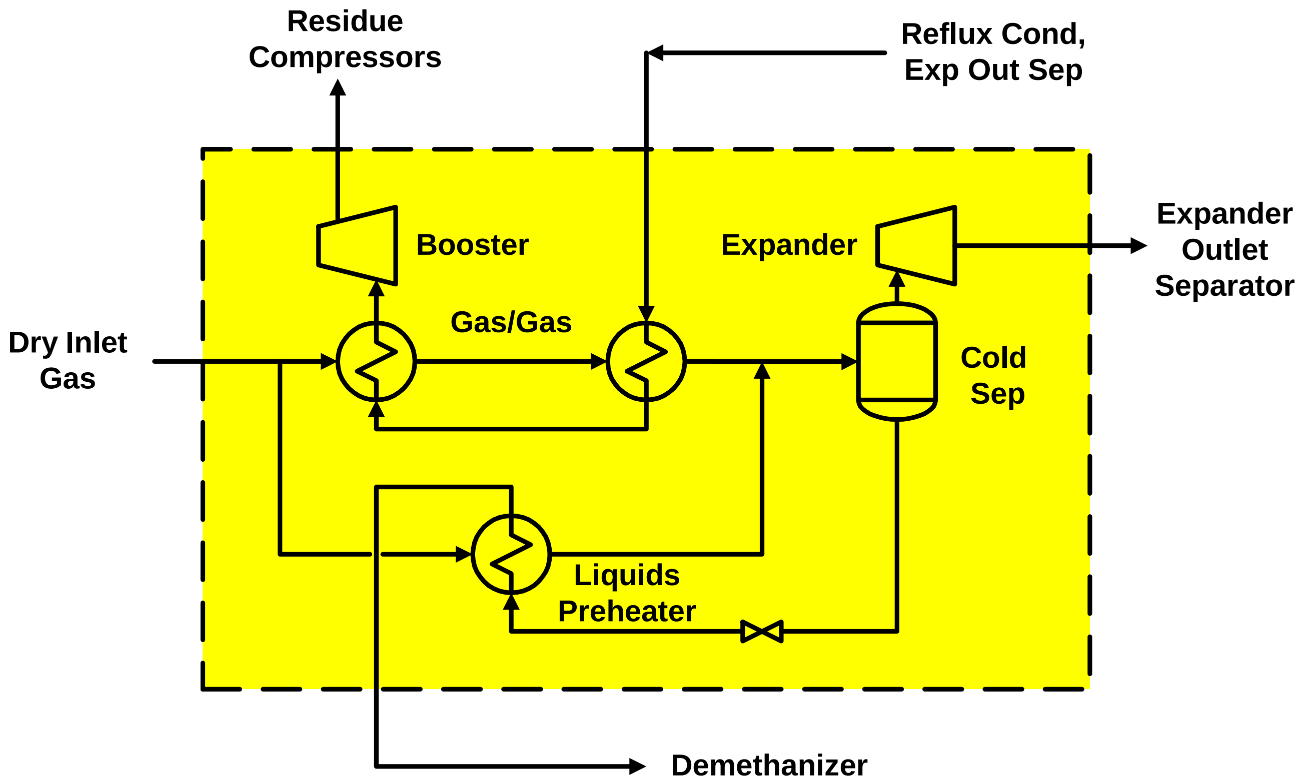

Figure 2 – Exchanger/Expander Train

The expander outlet streams from the two trains are routed to a common expander outlet separator with the liquids pumped to the reflux condenser and the vapor combined with the reflux separator vapor before entering the gas/gas exchangers.

PLANT OPERATION IN EARLY 2000

When Amerada Hess became operator in 2000, the plant was processing 425 MMSCFD of 1.6-1.8 GPM gas at an inlet pressure of 800 PSIG, achieving an ethane recovery level of 60%. The operating philosophy prior to year 2000 was to operate as cheaply as possible. This was achieved by shutting down almost half of the equipment to operate the plant at half design throughput and bypassing any gas the plant could not handle. About 100 MMSCFD of gas was bypassing the plant. The process design was unchanged from the original 1972 design.

Several major pieces of equipment had been shut down. The regeneration gas compressor was out of service. One of the two exchanger/expander trains and the steam turbine driven residue compressor had been shut down. The associated steam system had been decommissioned and was in need of many repairs. There were chronic reliability problems with the aging gas turbine driven residue compressor train which resulted in considerable plant downtime.

IMPROVEMENT GOALS FOR THE SEA ROBIN PLANT

Amerada Hess embarked on an ambitious plan to improve the plant beginning in April of 2000. The immediate short term goals were to:

- Process all the gas in the Sea Robin Pipeline and not bypass any gas around the plant when product pricing was favorable.
- Improve the plant on-line time.
- Reduce plant fuel consumption.

Improvements to the mechanical equipment, especially to the gas turbine driven residue compressor, were key to achieving these short term goals. The long term goals were to improve the plant operating flexibility and liquids recovery, specifically to:

- Increase ethane recovery when operating in ethane recovery mode.
- Increase propane recovery when operating in ethane rejection mode.

A modern process design would be required to achieve these recovery goals. The constraints were to minimize the cost of any process design changes and to use as much of the original equipment as possible while minimizing plant downtime to make the changes.

IMPROVEMENTS TO MECHANICAL EQUIPMENT

Most of the improvements to the aging mechanical equipment were completed in the years 2000-2002, before the process retrofit project was started.

Expander/Compressor Improvements

The single expander/compressor in operation at the 425 MMSCFD plant inlet rate had been operating at an efficiency level between 50% and 60%. In 2000, Texas Turbine, Inc. was charged with performing a complete redesign of the two parallel units. The initial evaluation of the expander/compressors indicated that the maximum compressor end throughput at the year 2000 conditions was equivalent to about 600 MMSCFD of plant inlet gas with the original process design. Good efficiencies and performance for the redesigned machines could be expected at the 500-600 MMSCFD plant throughput range with both machines in service. The redesign resulted in expander efficiency of 80+% and an increase in recompressor boost pressure of approximately 20 PSI. This project was completed in April of 2001. Obviously, successful rework of the original expanders was crucial to achieving the highest possible ethane recovery and in reducing the load and the fuel requirement of the residue compressors.

The expander inlet pressure was increased by placing both exchanger trains in operation thereby reducing the pressure drop from the dehydrators to the expander inlet nozzles. The increase in inlet pressure to the expanders resulted in an increase in the expander power. Operating both exchanger trains also reduced the pressure drops on the residue gas side of the plant and resulted in higher pressure at the suction to the booster compressors.

Regeneration Gas Compressor

The electric motor driven regeneration gas compressor had been out of service for nearly twenty years. Instead of operating the compressor, the discharge of the residue gas compressor was throttled by about 40 PSI to provide the pressure drop needed to force regeneration gas flow through the heater, dehydration bed, cooler, separator, and then on to the residue gas sales line. The 40 PSI additional residue compressor discharge pressure resulted in additional residue compressor fuel consumption over what was needed when the regeneration gas compressor was in service.

The operating pressures were higher now than the original design conditions. Texas Turbine, Inc. was hired to overhaul the unit and to redesign it for the current process conditions. The modified regeneration gas compressor was put back into service in March 2001. The result was a reduction in back-pressure on the residue compressor that reduced fuel usage by 200 MSCFD.

Gas Turbine Driven Residue Compressor Train

With the reduction in the required discharge pressure of the residue gas compressor due to placing the regeneration gas compressor back in service and the increase in suction pressure due to the improvement in expander/compressor efficiencies, the operating conditions for the residue compressors were greatly improved. However, there were still many opportunities for improvement in both residue compressor trains.

The gas turbine driver had a history of shutdowns due to high vibration levels. After extensive investigation, the auxiliary drive jackshaft was found to be out of balance. The vibration levels were reduced significantly when the jackshaft was re-balanced.

Problems with the turbine disc cavity cooling, turbine blade ring, turbine blades, and vane sections were analyzed and repaired during three shutdowns in 2000. The compressor was overhauled and the balance piston re-designed for the higher current operating pressures. These changes reduced bearing temperatures and thrust loads. The compressor case O-rings were also re-designed for the new pressure conditions.

Steam Turbine Driven Residue Compressor Train

The steam system, boiler, and economizer associated with the steam turbine driver were all in poor mechanical condition due to oxygen corrosion in the steam system. The system had not been operated since 1998. The repairs required several years to complete and the steam turbine driven compressor was re-started in the summer of 2004, nearly a year after the process retrofit described below was completed.

Control System Changes

Modern digital compressor surge control systems were added to each residue compressor. The original design did not protect the machines during startup and shutdown and was no help in controlling parallel operation of the two machines. A DCS process control system was installed in 2000. The new DCS allowed new plant control strategies to be implemented, and this capability became significant for the process design improvements which followed.

PROCESS DESIGN IMPROVEMENTS

In late 2001, the historical operating data and forecasts indicated that a process retrofit of the plant for higher recovery should be designed for an inlet gas CO_2 concentration of 0.5% and a liquids content of about 1.91 GPM. The 600 MMSCFD expander/compressor throughput limitation set the throughput design rate for any process retrofit.

The plant operating requirements versus original design offered some unusual opportunities for a retrofit design. There was plenty of residue compression power and throughput capability available if the steam system and steam turbine driven compressor were placed back in operation. There was ample surface area available in the original gas/gas exchangers, although the exchanger arrangement was not optimum for high ethane recovery, i.e., no side heater or product heater was included in the original design because of the low ethane recovery requirement. The original demethanizer reboiler used hot residue compressor discharge gas as a heat source and was sized to provide enough heat for ethane rejection operation with a lean inlet gas stream.

Simulations at the 600 MMSCFD inlet rate and 0.5% CO_2 content confirmed that very high ethane recovery levels (98%) could be achieved simply by running both residue compressors and retrofitting the plant to an Ortloff Recycle Split Vapor (RSV) process design [1,2].

The RSV retrofit design includes a new absorber column, a reflux condenser, and a pair of absorber bottoms pumps. In addition, revisions to the existing demethanizer internals were required. Note, however, that all of the original ethane recovery exchangers, separators, pumps, and reboiler were reused in the retrofit design. Only the small reflux condenser, accumulator, and pumps used in the original design's propane recovery mode were abandoned. The RSV retrofit equipment items were all additions to, not replacements of, the original plant equipment.

One result of operating a plant at this very high ethane recovery level, however, is that the CO_2 recovery is also very high. For the Sea Robin design, the resulting CO_2 concentration in the NGL product exceeded the allowable pipeline specification. The CO_2 recovery was about 65% in the RSV configuration, resulting in an NGL product CO_2/C_2 liquid volume ratio of nearly 3%, or about twice the allowable 1.5 liquid volume percent. This specification could easily be met by lowering the ethane recovery level or by adding product treating. However, neither approach provided acceptable project economics. A better solution for meeting the CO_2 specification had to be found if the project economics were to remain attractive.

An Ortloff Carbon Dioxide Control (CDC) [3] feature was then added to the RSV design to reduce the CO_2 content of the NGL product while keeping the project economics favorable. The CDC design feature is used to strip some of the CO_2 from the column liquids several fractionation stages above the reboiler using a methane-rich stream from the absorber bottoms liquids. Stripping some CO_2 results in increasing the CO_2 concentration in the upper stages of the column and in the absorber column, so CO_2 freeze margins are narrowed as more and more CO_2 is rejected from the NGL product. Also, adding the CDC stripping vapors normally results in a small increase in the compression power requirement to maintain high ethane recovery while rejecting more CO_2.

The Sea Robin requirements were a perfect fit for the CDC technology. The product CO_2 specification was met with little increase in the residue compression requirement while keeping the ethane recovery at 98%. With the addition of the CDC feature, the process retrofit design was finalized as shown in Figure 3 on the next page.

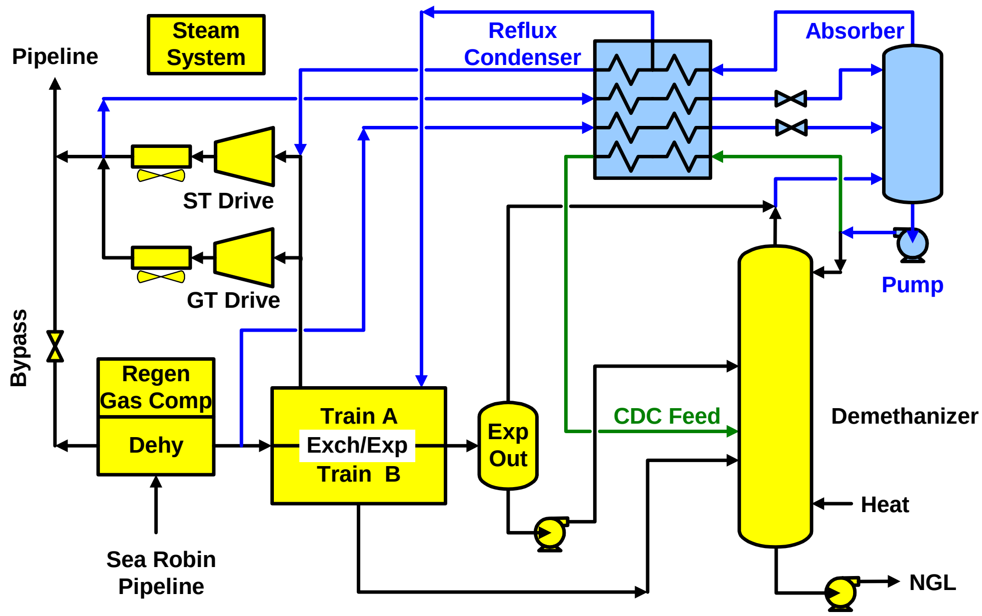

Figure 3 – Sea Robin RSV+CDC Retrofit Process Design

Revisions to the demethanizer column internals were required to accommodate the turndown conditions resulting from operating the column at a lower rate than the original design. The RSV+CDC retrofit design further unloads the demethanizer column internals by adding fractionation stages via the new absorber column. The new absorber column becomes a process extension of the existing demethanizer, with the demethanizer overhead routed to the bottom of the absorber column and the absorber bottoms pumped to the top of the demethanizer.

The RSV design uses the very lean residue gas as a source of reflux to the column above a reflux feed which originates from the plant inlet. The top reflux stream is a recycle stream routed through the residue compressors. At Sea Robin, the required reflux stream flow rate moves the compressor operating point enough to keep the surge control valves closed at 500 MMSCFD plant inlet rate with both compressors in service. The unusual combination of compression equipment capacity versus residue gas flow rate including the reflux stream requirement actually favored the use of the RSV design at the very high ethane recovery level. The horsepower required to provide the residue reflux stream is largely cancelled out by moving to a higher efficiency operating point on the compressor map. Running the steam turbine driven compressor adds very little incremental fuel cost since it is driven primarily from steam generated by the waste heat boiler on the gas turbine driven compressor.

The expander/compressor operating point also changed with the retrofit process design, with the expander end ACFM staying fairly constant at lower flow rate but with a warmer inlet temperature. The compressor end operating point also changed, but not enough to warrant another redesign. These units were not modified from the 2001 configuration.

The CDC feature is very effective at reducing the CO_2 concentration in the bottom product. This reduction is achieved by forcing enough CO_2 overhead to allow the bottom product specification to be met. Since the CO_2 concentration in the tray liquids in the cold upper section of the absorber column is increased, CO_2 freeze margin is reduced. At the design 0.5% CO_2 concentration in the feed gas, CO_2 freeze was not a problem, even at 98% ethane recovery. However, when the CO_2

concentration in the feed gas to the plant is higher than design, the ethane recovery must be controlled at the highest possible recovery level without freezing while also meeting the bottoms CO_2 specification.

With the RSV+CDC process design finalized in early 2002, the retrofit project was contracted to Optimized Process Design (OPD) in Katy, Texas, for detailed engineering, procurement, and construction. The process retrofit was completed in June of 2003 and has been in service for nearly two years.

SEA ROBIN PIPELINE GAS RATE AND COMPOSITION

Figure 4 below shows the Sea Robin Pipeline gas volume available for processing over the last five years. The gas available to process peaked in mid-2000 and has remained within the 400-500 MMSCFD range since early 2001.

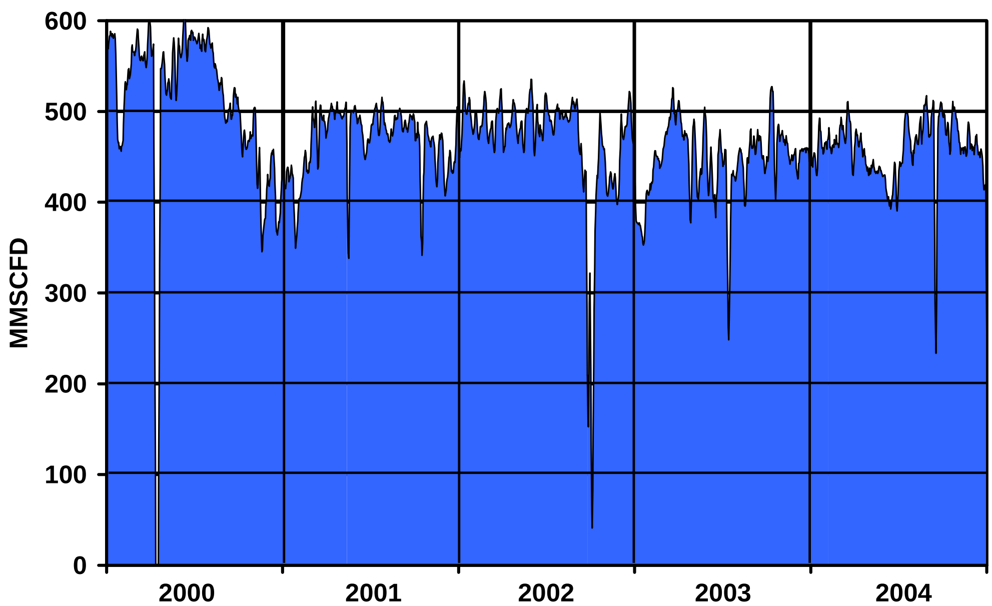

Figure 4 – Sea Robin Pipeline Gas Rate

The dips in September 2002, July 2003, and September 2004 were all weather related. There is some data missing in early 2000, and the remaining dips were primarily from short interruptions during pigging operations.

Figure 5 on the next page, shows the variation in liquids content on a gallons per MSCF basis (GPM) and the CO_2 content on a mole percent basis. The liquids content of the gas available for processing varied from 1.5 to 2.0 GPM, resulting in a significant impact on the NGL production, regardless of the processing efficiency.

The most significant change in inlet composition has been in the CO_2 content of the feed gas since April of 2003. With the original process design operating at 60% ethane recovery, the CO_2 content was not a factor in operation of the plant. With the RSV+CDC retrofit designed to operate at

98+% recovery, the CO_2 content is much more of a concern because of its limiting effect on the ethane recovery due to CO_2 freezing. The CO_2 content trend was generally downward from 2001 into 2003, but turned up and continued to rise through 2002 to almost 1.0%, nearly twice the design value of 0.5% for the retrofit. It has remained well above 0.5% and volatile since mid-2003.

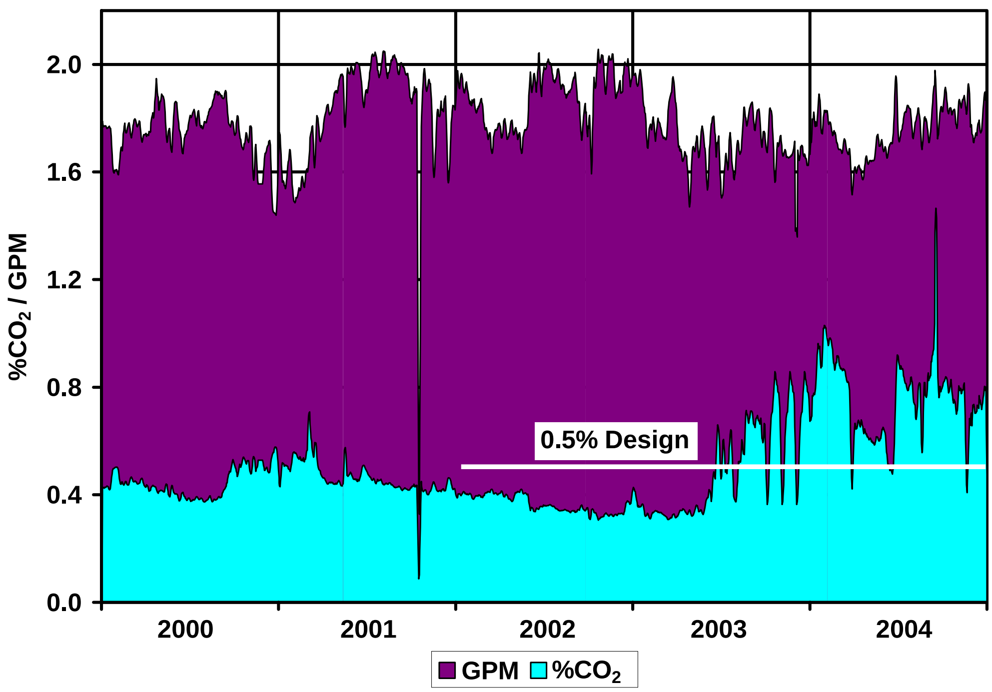

Figure 5 – Sea Robin Inlet Liquids and CO_2 Content

OPERATING RESULTS

Operating results for 2000-2004 are shown on several graphs which follow. Not all plant constraints were technical, however. There were significant swings in processing margins over these years, and there were several weather related interruptions which add some variability to the graphs.

Available and Bypassed Daily Volume

The daily volume of Sea Robin gas available to the plant compared to the daily volume bypassed is shown in Figure 6 on the next page. Figure 6 is the same as Figure 4 with the bypassed volume rate added. The gas volume processed through the plant is the difference between the two rates on any given day.

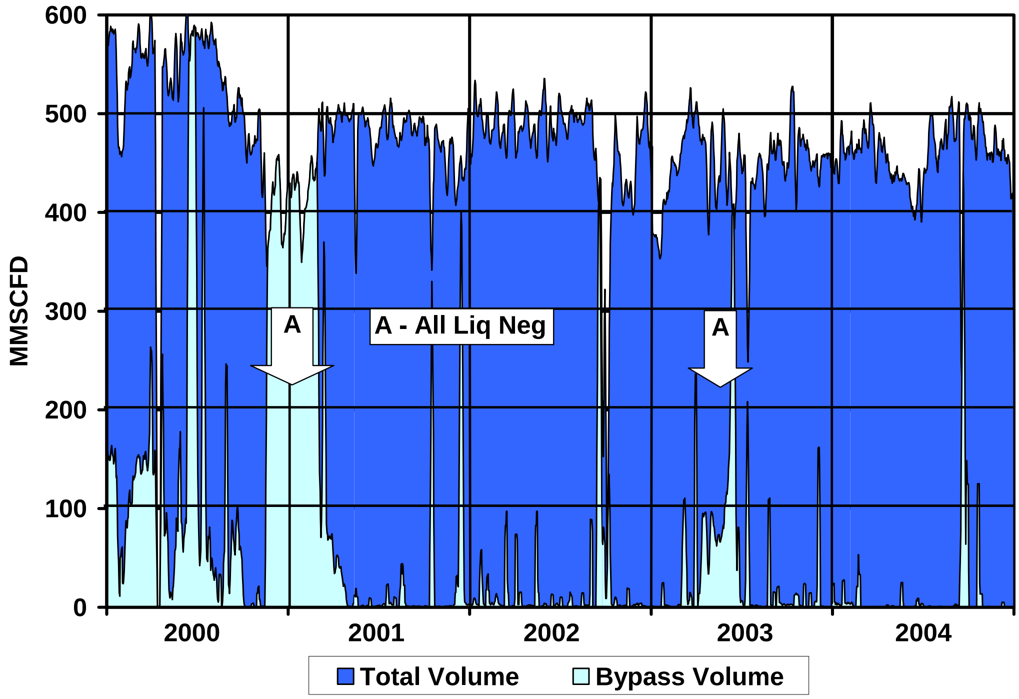

Figure 6 – Available and Bypassed Daily Volume

In 2000, the bypass rate is variable but averages about 100 MMSCFD. By mid-2000, sufficient changes had been made to increase the plant throughput to over 500 MMSCFD so the bypass was closed off. The bypass remained closed for two months before market conditions warranted changes in the plant operation.

November 2000 to March 2001. All liquids were negative in value. The plant was shut down during this time. The time was used to make major repairs on the gas turbine at the plant. The expander/compressors were redesigned for the current operating pressures and flows.

Many liquids extraction plants were shut down during this period due to the negative liquids value, resulting in problems with liquids drop out in the transmission lines. Since that time, most gas plants are required to recover sufficient liquids to avoid dewpoint problems in transmission lines, regardless of the liquids pricing relative to gas.

April 2003 to May 2003. All liquids were once again negative but propane was the least negative product. Some liquids had to be extracted to meet the dewpoint specification. During this time, some gas was bypassed and the plant was operated in an ethane rejection mode to allow the plant to recover the least amount of propane possible and still meet the dewpoint requirement. The plant bypass was used as a means of butane+ rejection. The unloaded plant operated at 5% ethane recovery and 90% propane recovery. The timing was perfect for completing the tie-ins for the RSV+CDC retrofit.

Note that the plant bypass was normally closed after mid-2001 and has not been used much since, other than for the market-driven adjustment in 2003.

Ethane and Propane Recovery

Both the original plant design and the RSV+CDC retrofit could be adjusted to operate in either ethane rejection or ethane recovery mode. Figure 7 below shows the ethane and propane recovery results as a percentage of those components in the cryogenic plant inlet before and after the process retrofit to the Ortloff RSV+CDC process design.

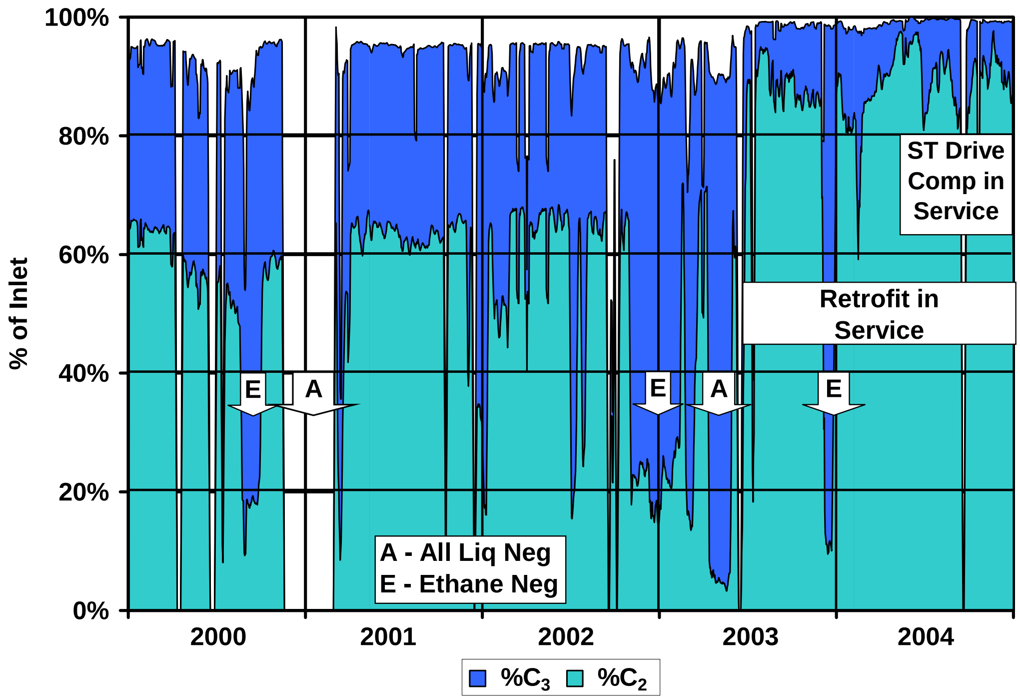

Figure 7 – Ethane and Propane Recovery as Percent of Inlet

After the expander rework in April 2001, ethane recovery averaged 60-67% with the propane recovery at 95%. These recovery levels were limited by the original process design. Starting in June, 2003, after the RSV+CDC retrofit was completed, both the ethane recovery and the propane recovery increased dramatically even without running the second residue compressor. The second (steam turbine driven) residue compressor was started up in April, 2004. After that compressor was started, propane recovery approached 100%, limited only by a small leak in one of the original shell and tube gas/gas exchangers. Ethane recovery also increased but the results were not consistent, for reasons described later. The graph indicates that there were several extended time periods when the plant was operated in ethane rejection mode.

August 2000 to September 2000. Ethane liquid value was negative and the propane liquid value was positive. Ethane recovery was dropped to 18% and the propane recovery dropped along with the ethane recovery to around 91%. Rejecting any more ethane would have resulted in higher propane losses.

November 2002 to February 2003. Ethane liquid value was negative and the propane liquid value was positive. The plant was operated in an ethane rejection mode using the original process design with 22% ethane recovery. Propane recovery dropped from 96% to 90%.

December 2003. Ethane liquid value was negative and the propane liquid value was positive. The RSV+CDC retrofit was in operation with only one residue compressor in service. Ethane recovery was 11% and propane recovery remained very high at 98%. The ethane rejection level was limited by the original reboiler size in ethane rejection mode. The operating flexibility of the new process design was clearly demonstrated during this time.

Effect of CO_2 Content on Ethane Recovery

Figure 5 on page 8 showed how the CO_2 concentration in the Sea Robin Plant inlet gas changed over the last five years. The trend was slightly downward in early 2002 when the design basis for the retrofit design was frozen at 0.5%. However, in mid-2003, shortly after the retrofit was started up, new off-shore wells and the associated treating system resulted in significant unforeseen swings in the CO_2 concentration. When the CO_2 concentration exceeded 0.5%, the ethane recovery had to be reduced to avoid CO_2 freeze conditions in the new absorber column.

The result is that the plant is currently controlled by adjusting recovery level to avoid freezing while making the NGL product CO_2 content specification. For the high CO_2 concentrations, the plant is controlled to operate on the verge of CO_2 freeze. When the CO_2 content in the feed drops down close to 0.5%, the ethane recovery is very high, about 99%. When the CO_2 content rises, the ethane recovery must be reduced several percentage points.

The operators monitor pressure drop across the absorber column trays to detect the onset of freezing. In normal operation the top differential pressure is 6.5 to 7.5 inches of water and the bottom differential pressure is 12 to 15 inches of water. When either the top differential pressure increases to 8 inches of water or the bottom differential pressure increases to 18 inches water, column alarms alert the operators. Then either the column pressure is increased or the reflux rates decreased to warm up the column temperature profile. These changes also result in a decrease in ethane recovery. In this fashion, the ethane recovery is constantly adjusted as the CO_2 content of the feed changes. Currently the adjustments are done manually. The results are reflected in the larger variations in the ethane recovery levels for the time periods where the CO_2 content variations are the most severe, such as in the last half of 2004.

A plot of inlet CO_2 has been added to the ethane and propane recovery graph for years 2003 and 2004 in Figure 8 on the next page. Note how the ethane recovery was reduced as the CO2 content of the inlet gas increased. When the CO_2 content came back down, the ethane recovery was increased. Note also that the propane recovery remained very high regardless of the ethane recovery as long as some recycle reflux flow is maintained.

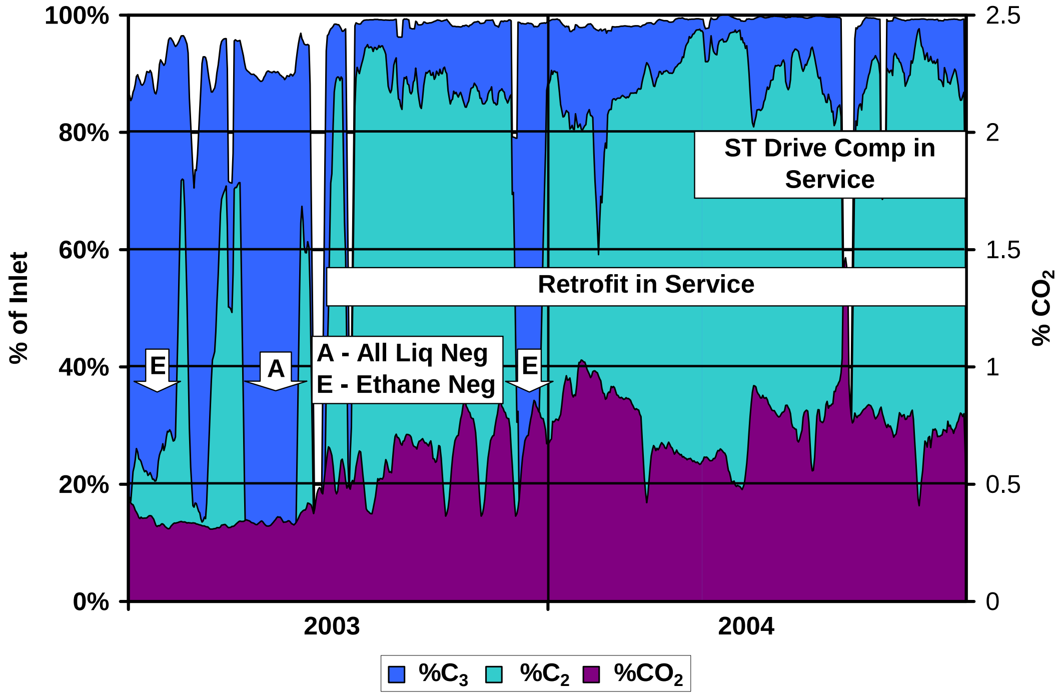

Figure 8 – Recoveries and CO_2 Content for 2003-2004

NGL Production

Figure 9 on the next page shows the daily NGL production for the five year period 2000-2004. The liquid production volume was a result of the liquid product processing economics, the gas rate available for processing, the liquids content of the gas, the CO_2 content if it is high enough to limit ethane recovery, and plant on-line time.

In 2000, simply taking more pressure drop across the plant and closing the plant bypass increased liquids production from about 13,500 BPD to 15,500 BPD. After the second expander train was placed in operation in April of 2001, essentially all long term use of the plant bypass ceased and liquids recovery climbed to about 18,000 BPD.

When processing economics were favorable, the NGL production increased from 13,500 BPD average in early 2000 to nearly 20,000 BPD near the end of 2004 even with a higher CO_2 content and lower gas availability rate. It is now possible to maintain nearly 100% propane recovery regardless of the ethane recovery economics. The plant's operating flexibility and recovery capability have been improved significantly as a result of the process retrofit and improvements made to the plant equipment.

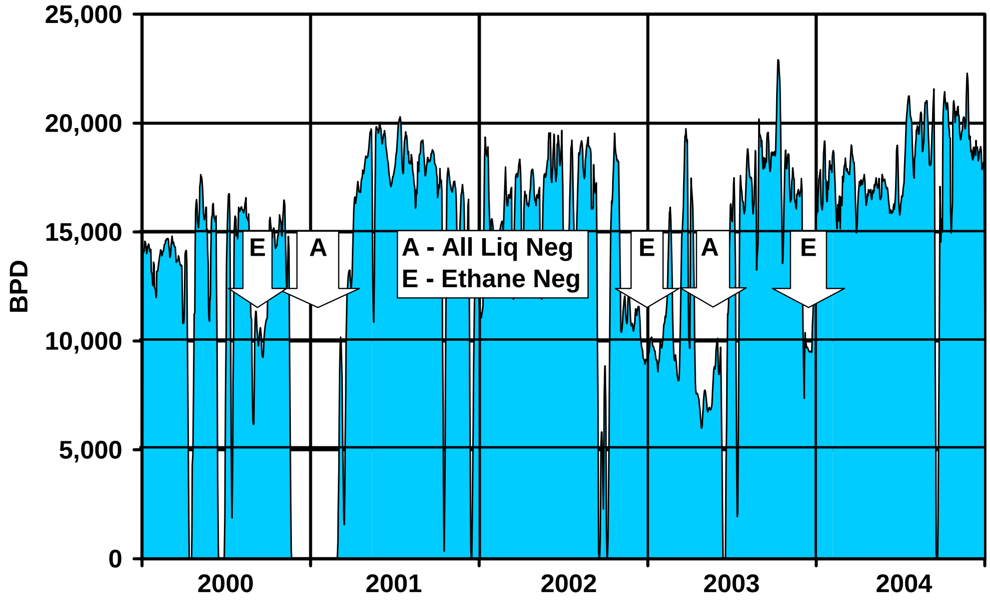

Figure 9 – Daily Total NGL Production

FUTURE IMPROVEMENTS

There are additional projects planned for the next two years. A stand-alone fired boiler will be installed to allow full load operation of the steam turbine driven compressor if the gas turbine driven compressor and its waste heat boiler are down for scheduled maintenance. Without the stand-alone boiler, the plant is down when the gas turbine driven compressor train is down.

The residue compressor gas turbine will undergo a major overhaul in 2006, after the stand-alone fired boiler is installed. Only stop-gap work has been done on the hot gas path since 2000. The compression systems should be very reliable for many years after this major overhaul.

The gas turbine driven compressor speed controls are an outdated analog design. These will be replaced with digital controls in 2005. Currently the compressor is run on fixed speed control because of the antiquated control system. It should be possible to vary the compressor speed to better match throughput changes without disturbing all of the plant pressures after the turbine controls are replaced.

The tube leak in the gas/gas exchanger will be repaired when the plant is down long enough to make the repairs.

The original demethanizer reboiler used hot residue compressor discharge gas for a heat source. Piping has been added to allow steam to be used as a heat source for ethane rejection mode operation. Ethane recovery of 2% should be possible with the retrofit design in an ethane rejection mode while maintaining very high propane recovery.

A continuous inlet gas CO_2 analyzer will be installed. This addition will allow some feed forward control of the ethane recovery as the CO_2 changes.

It should be possible to further automate some of the ethane recovery process variables after the turbine controls are improved. For example, it should be possible to reset or bias the column pressure control using the differential pressure measurement across the absorber column. If so, control on the verge of CO_2 freeze can be automated rather than controlled manually. Automatic control should allow tighter control of the ethane recovery versus CO_2 freeze without constant manual intervention by the board operator.

CONCLUSIONS

This project is a good example of what can be done to an older first generation facility to bring it up to date in terms of process design technology, efficiency, and reliability. The modifications have resulted in greatly increased liquids recovery, reduced unscheduled downtime, and reduced operating cost per barrel of liquid product. Use of the latest liquids recovery technology provided very high product recovery levels without adding a product treater.

Improvements in the plant have been achieved by maximizing the use of the original equipment. Upgrading the process design to accommodate the changes in plant inlet conditions of rate, pressure, and composition has resulted in a very flexible and predictable plant design, which can be operated to maximize income as the liquid market pricing changes.

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