UNIQUE DESIGN CHALLENGES IN THE AUX SABLE NGL RECOVERY PLANT

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Joe T. Lynch, P.E.
Ortloff Engineers, Ltd.
Midland, Texas, USA

Chuck Cairo
Aux Sable Liquids Products
Channahon, Illinois, USA

Hank M. Hudson, P.E., John D. Wilkinson, P.E.,
Richard N. Pitman, and Michael C. Pierce
Ortloff Engineers, Ltd.
Midland, Texas, USA

ABSTRACT

One of the largest gas processing plants in the world, the Aux Sable NGL Recovery Plant in Channahon, Illinois, was commissioned in December 2000. This plant has an ultimate inlet gas processing capacity of 2.1 BCFD and a design NGL production rate of 123,000 BPD in two trains. Its size and its location at the end of the Alliance Pipeline gas transmission line presented unique design challenges, including:

- Supercritical (dense phase) inlet gas pressure
- Process selection for a broad range of inlet rate and product recovery requirements
- Control of ethane product specification
- Startup and expander-down operating modes

Plant operations since commissioning have shown the overall plant design philosophy to be quite satisfactory, including successful operation at off-design conditions. This paper discusses each of these design challenges, the possible solutions considered for each, and the solutions chosen for the final plant design.

INTRODUCTION

The history and the economic drivers for the Aux Sable NGL Recovery Plant project were covered in an article published last year.[1] The plant was designed in 1997-99, built in 1999-2000, and started up in December of 2000. The facility is located near the termination point of the Alliance Pipeline, about 50 miles southwest of Chicago, Illinois, USA. The facility consists of two parallel NGL extraction trains, a single fractionation train with product treating, and the associated compression, product storage, and utility systems as shown in Figures 1 and 2.

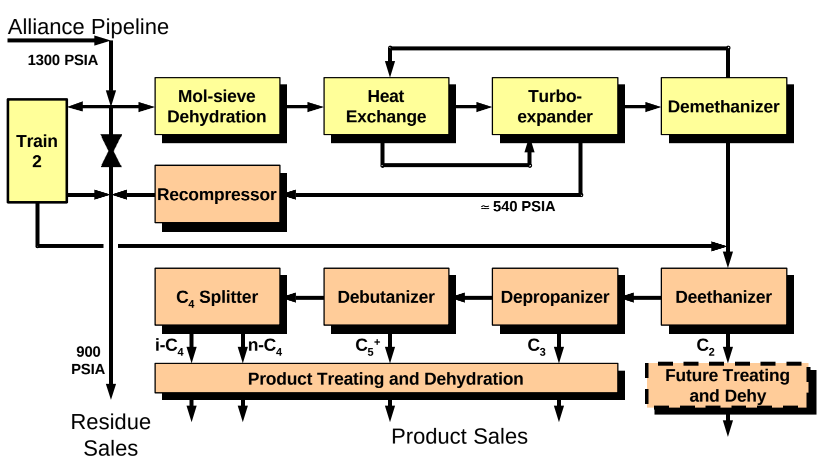

Figure 1 – Aux Sable Block Diagram

Figure 2 – Aux Sable NGL Recovery Plant

This paper describes several unique technical challenges encountered while designing the NGL recovery trains. These challenges resulted from the high volume, high pressure, dense phase inlet gas conditions, as well as the broad range of operating requirements for the facility.

INLET CONDITIONS

The Alliance Pipeline was designed to deliver 1.6 BCFD of gas initially, with expansion to 2.1 BCFD after adding thirteen more pipeline booster compressor stations. Aux Sable's design requirements therefore included the ability to process the entire range of potential flow rates.

The plant inlet pressure was fixed at 1300 PSIA for the base case. An off-design case in which one pipeline compressor is down and the inlet gas pressure drops below the cricondenbar to 950 PSIA was also included. A phase envelope for a typical inlet gas composition with the base and off-design inlet pressures identified is shown in Figure 3.

The exact inlet composition was not known at the time the plant was being designed. Lean and rich inlet gas compositions were established to cover the expected range of inlet composition, with the rich case composition containing 50% more ethane and heavier components than the lean case. In addition, a large number of off-design cases were run with varying amounts of ethane versus heavier components to quantify the plant performance for the possible variations in inlet composition.

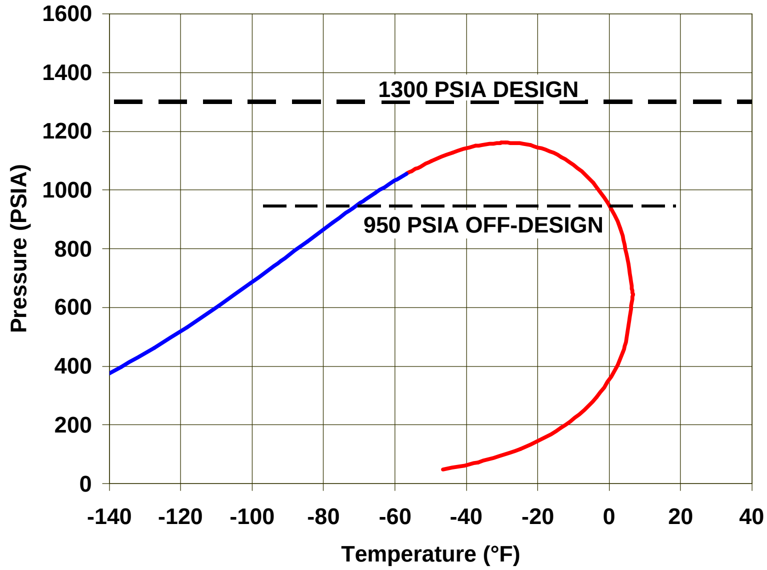

Figure 3 – Phase Envelope

The range of inlet compositions used for the design basis is shown in Table 1. The CO_2 composition was specified at a maximum of 0.5 mole percent, even though the hydrocarbon content was expected to vary considerably depending upon market conditions in Canada and the U.S.

Table 1 – Inlet Gas Composition

Composition mole %	Lean Case	Rich Case
Ethane	6.50	9.75
Propane	1.90	2.90
Butane +	0.70	1.10
CO_2	0.50	0.50

PRODUCT REQUIREMENTS

During the design phase, several possible markets for the residue gas were identified. It was predicted that these customers would require delivery pressures (up to thirteen miles from the plant) ranging from 685 to 837 PSIA. This meant that the plant residue compressors would need to be able to accommodate the wide discharge pressure range, and that the selected NGL recovery process would need to provide the highest liquids recovery possible over this range of residue delivery pressures.

An additional constraint was the maximum higher heating value of the residue gas. The plant would need to be able to remove enough liquids from the gas to keep the higher heating value below a 1050 BTU/SCF limit.

The plant location provides access to several hydrocarbon liquid product markets, with one ethane product customer located within one mile of the plant. The liquid ethane product specification allows for a maximum CO_2 concentration of 1000 ppm(w).

MECHANICAL DESIGN AND OPERATIONAL CONSTRAINTS

Several mechanical constraints affected the process design and the NGL recovery train size. Obviously, the initial and ultimate capacity of the Alliance Pipeline required a facility with large size equipment. Preliminary column sizing indicated that a minimum of two trains would be required if the demethanizer columns were to be shop fabricated, which was a project requirement.

The two train concept would also allow using a single 15,000 HP expander per train. At the time the facility was being designed, the 15,000 HP level was a reasonable upper limit for the available machines. The available expander case sizes could easily meet the volumetric flow capacity requirements for the Aux Sable trains because of the high stream pressures.

The original conceptual design for the facility was based on electric motor driven residue gas compressors, but the design had to be changed to gas turbine drivers due to difficulty in obtaining a satisfactory electric utility contract. Instead, the process design was optimized for the capabilities of the same GE LM-2500 gas turbine drivers purchased for the Alliance pipeline booster compressor station services. The compressors could not be the same, however, as the plant requirements dictated less flow and higher head than the pipeline machines.

Yet another design requirement was to be able to operate with an upstream pipeline compressor station out of service. This would result in a 950 PSIA plant inlet pressure, below the dense phase pressure, so two-phase conditions could exist after inlet stream cooling.

The initial ethane product market demand could be met at a relatively low ethane recovery level. This resulted in another design constraint, the ability to meet the 1000 ppm(w) CO_2 concentration limit in the ethane product without treating. This capability would allow the installation of the ethane product treating and dehydration systems to be deferred. The selected process would need to be able to recover as much ethane as possible while reboiling the column to meet the CO_2 specification. An ethane product treater would be added later as ethane product demand increased, so operation with and without the ethane product treater was included in the design.

Finally, the NGL recovery trains needed to be capable of full ethane rejection operation with minimal loss of propane. This mode of operation would be used when the market conditions or maintenance requirements were such that the deethanizer system needed to be shut down.

PROCESS DESIGN SELECTION

The inlet conditions, product requirements, mechanical constraints, and operational requirements described above became the design basis for evaluating several process design options. Various process schemes using inlet gas and/or residue gas to reflux the tower were evaluated over the range of operating conditions to determine the performance when operating for high ethane recovery, moderate ethane recovery, and full ethane rejection. Ortloff's Recycle Split-Vapor (RSV) process provided excellent performance over the broad range of conditions in the design basis.[2] All of the other process designs considered had significant limitations at one end or the other of the range of operating conditions.

With RSV, the available horsepower would allow 94-98% ethane recovery when the residue delivery pressure and inlet rate were at the low end of the design basis conditions. As the inlet rate and residue pressure increased, the RSV design could maintain the ethane recovery in the 80% range while holding very high propane recovery. Product recovery could be maximized by adjusting the RSV process reflux rates to maximize use of the residue compressor horsepower for any combination of inlet rate and residue delivery pressure.

The two-train RSV design was optimized to meet all the mechanical constraints while providing excellent performance, resulting in the following final equipment configuration:

1. Single 20' diameter, 400 ton demethanizer column (500 PSIG design pressure) in each 1.05 BCFD train.
2. Single 15,000 HP expander/compressor in each train.
3. Single 31,200 HP (ISO) LM-2500 gas turbine driven centrifugal compressor package in each train.
4. No refrigeration system needed for the NGL recovery trains.
5. No ethane product treating for removal of CO_2 initially.

The recovery performance for the RSV process design with this combination of equipment ranged from 75% ethane recovery when reboiling to a CO_2 specification to as high as 98% when operating at low residue pressure at the 1.6 BCFD initial inlet rate. The process could be operated at the high ethane recovery level without CO_2 freezing with the inlet CO_2 concentration at the design value of 0.5 mole percent. A simplified process flow schematic of the final RSV process design is shown in Figure 4.

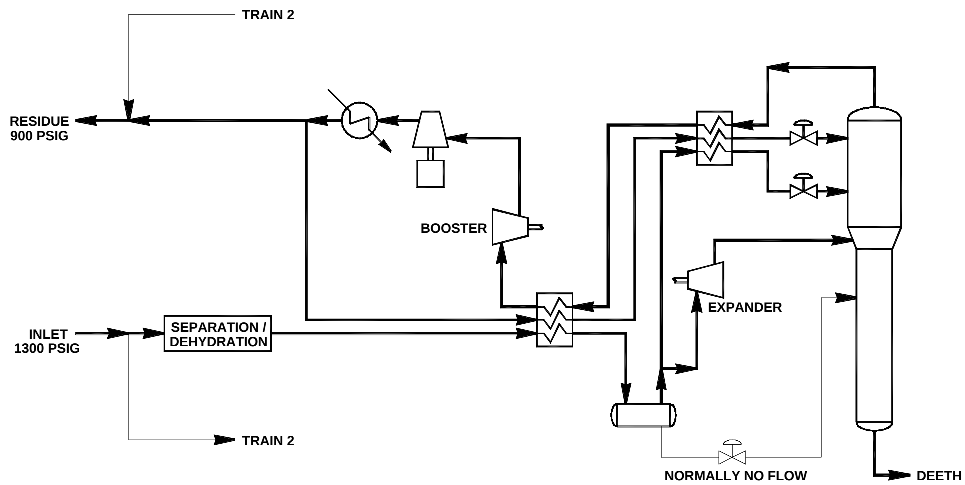

Figure 4 – RSV Process Flow Schematic

DENSE PHASE INLET GAS PRESSURE

The normal inlet pressure for this facility is 1300 PSIA. This is above the cricondenbar, so the pipeline fluid density is between that of a gas and a liquid. Unlike a typical cryogenic process design, there is no condensation of liquids as the high pressure inlet gas stream is cooled, since no liquid phase can exist at the 1300 PSIA inlet pressure. Rather than forming two distinguishable phases, the stream density simply increases as the stream is cooled.

The maximum acceptable demethanizer operating pressure is limited by phase separation considerations to no more than 500 PSIA, so the minimum expansion ratio is essentially fixed at 2.6 by the 1300 PSIA inlet pressure. It should be noted that when the RSV process is operated for high ethane recovery, the optimum configuration typically coincides with the maximum allowable demethanizer operating pressure. This then allows the residue compressors to recycle the maximum amount of residue gas, providing a larger top reflux stream. For the original Aux Sable design, the maximum operating pressure was set at 460 PSIA, since the pressure at which phase separation difficulties begin was uncertain. (Field experience at Aux Sable has now shown that reasonable mass transfer in the demethanizer is possible at 470 PSIA.)

When the inlet flow rate and/or residue delivery pressure increase, the ethane recovery using the available horsepower is maximized by first reducing the top reflux feed provided by recycled residue gas so that the tower pressure does not have to increase. In the extreme case, shutting off the recycle flow entirely may offer the best recovery by allowing the tower pressure to be lower.

Because the minimum expansion ratio is so high, mechanical refrigeration using an external propane refrigeration system is of limited value with these inlet conditions. Instead, the ethane recovery in the demethanizer column is largely a function of the work extraction and cooling that can be achieved using the expander. In some of the design cases, a warmer expander inlet temperature resulted in better recovery than a colder expander inlet temperature for a given residue compressor suction pressure. There are two reasons for this behavior.

First, a warmer expander inlet temperature results in more power recovery, increasing the power supplied to the booster compressor upstream of the residue compressor. With more power available, the ratio across the booster compressor can be increased, allowing a lower demethanizer operating pressure. The improved recovery at the lower column pressure more than offsets the recovery lost due to the warmer cold separator.

Second, with this high inlet pressure it is possible to condense more methane than desired at the expander outlet. Additional methane condensation without a corresponding increase in ethane condensation results in an increase in reboiler duty to meet the column bottoms specification, and reduces the ethane recovery as the extra methane is stripped from the tower liquids.

The plant must also be capable of operating under the inlet pressure conditions which exist when a pipeline booster compressor is out of service. At 950 PSIA, the cooled inlet stream may be two-phase, so a cold separator is required upstream of the expander. A special vortex-type separator (see Figure 5) was installed, which is much smaller than a traditional cold separator but meets the process requirements. This high efficiency vessel operates dry as long as the inlet pressure is above the cricondenbar.

Figure 5 – Aux Sable Cold Separator

DEMETHANIZER COLUMN CONTROL

Design cases were run at demethanizer column pressures ranging from 330 PSIA to 460 PSIA. The column pressures were calculated based on loading the residue compressor at the various residue delivery conditions and adjusting the reflux rates to optimize the recovery while using all the available expander horsepower. Numerous off-design cases were also run.

The CO_2 content of the feed was specified at 0.5 mole percent, with a significant amount of the CO_2 being recovered with the ethane if the bottoms temperature is adjusted to control the methane content rather than the CO_2 content of the liquid product. The column heat input needed to limit the CO_2 in the bottoms product to 1000 ppm(w) is higher than that required to control the methane content. Some ethane is lost along with the CO_2 rejected to the residue if the CO_2 content is controlled by adjusting the column reboiler heat input.

The case studies showed that the CO_2 concentration is very sensitive to temperature around the 1000 ppm(w) control limit. Figure 6 shows how the recovered amounts of CO_2 and ethane change when reboiling a demethanizer at constant pressure to meet different product specifications.

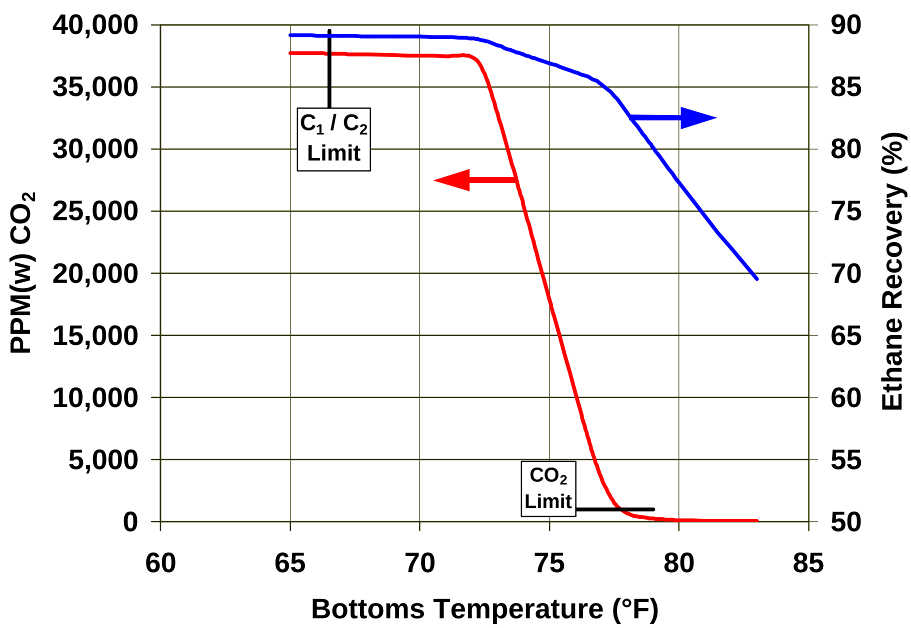

Figure 6 – Bottoms Composition vs. Bottoms Temperature

Once ethane product treating is installed at the Aux Sable facility, the demethanizer will be reboiled to a methane:ethane product specification. This corresponds to a much cooler bottoms temperature than when reboiling to the stringent 1000 ppm(w) CO_2 specification for the ethane in the product. At cooler bottoms temperatures, essentially all the ethane (and CO_2) flowing down the column is recovered, and the stripping gas generated by the reboiler is mostly methane. However, when the bottoms temperature is increased to reduce the CO_2 concentration, no methane reaches the reboiler to be vaporized to provide the necessary stripping vapors. Instead, ethane must be vaporized to strip the CO_2 from the product, and ethane recovery begins to suffer.

Note particularly how quickly the amount of CO_2 in the bottom product changes with very small changes in bottoms temperature. When the Aux Sable plant is operating without ethane product treating and the tower is reboiled to meet the 1000 ppm(w) CO_2 specification, the operating point is very near the steep portion of the curve. This could cause large changes in CO_2 concentration as plant conditions change slightly, so the plant operators are forced to set the bottoms temperature to a safe level (i.e., a few degrees warmer than the optimum) to provide some margin for operating variations without exceeding the specification.

The conditions for the Figure 6 graph correspond to a tower pressure of 400 PSIA, but the sensitivity is not unique to that pressure. Figure 7 shows a very small section of the CO_2 concentration curve at three different tower pressures. All three curves illustrate the same large changes in CO_2 content with small changes in bottoms temperature.

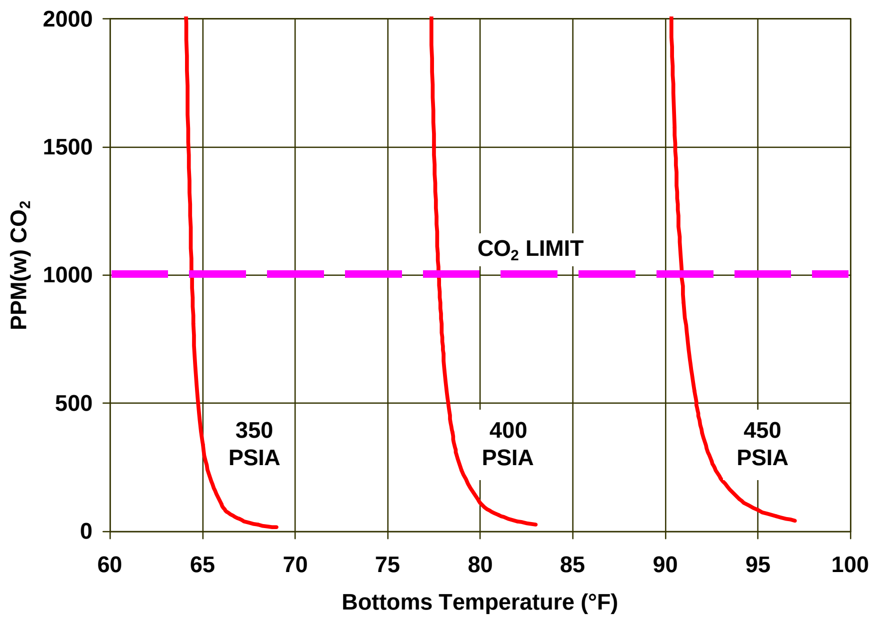

Figure 7 – CO_2 Concentration in Bottoms vs. Bottoms Temperature

This demonstrates a situation where Ortloff's Carbon Dioxide Control (CDC) technology would be applicable.[3] By redistributing more methane lower in the tower, less ethane must be vaporized in the reboiler to act as stripping vapor for the CO_2. Unfortunately, CDC was conceived at about the same time as construction on the Aux Sable plant was commencing.

STARTUP AND EXPANDER-DOWN OPERATING MODES

The dense phase, high pressure inlet stream generates high expander horsepower, which generates a correspondingly high booster compressor pressure rise. For example, when the residue compressor is delivering to the residue line at 900 PSIG and the residue compressor suction pressure is 540 PSIG, the pressure rise across the booster is 120 PSI.

In the typical post-boost cryogenic expander plant with a much lower inlet pressure than found at Aux Sable, the column pressure is allowed to rise when the expander is out of service to make up for the loss of the expander booster compressor. The column pressure must rise to equal the original residue compressor suction pressure (plus the pressure drop from the column to the residue compressor). The mechanical design pressure of the column is then specified high enough to allow for operation in this Joule-Thomson (J-T) mode with the expander out of service.

For the Aux Sable plant with its much higher than typical boost, however, the column pressure would have to rise to 550 PSIG or higher to operate in J-T mode. Fractionation is poor at this column pressure, and the mechanical design pressure of the column would have to be at least 600 PSIG to cover operation at the high pipeline delivery pressure. This requirement would have added significantly to the cost of the 20' diameter columns over that required for the 500 PSIG design pressure set by operation with the expander/compressor in service.

Also, as stated earlier, the ethane recovery for this plant is very dependent upon the horsepower extracted from the gas at the expander. With the expander down, the recovery decreases substantially. Since the effect of having the expander down is so dramatic, a different approach was taken for the expander-down operation of the Aux Sable plant. This approach was then applied to the startup and shutdown sequence of each train as well.

Each parallel NGL recovery train was piped so that it can be operated independently as shown in Figure 8. The residue compressor piping is completely separate for each train, with the residue compressor inside the train isolation valves. If an expander is out of service, that train is taken out of service. Since the residue compressor suction pressure is limited by the 500 PSIG design pressure of the column, the residue compressor cannot reach the residue pipeline delivery pressure if it is at the high end of the design basis range without the booster compressor in service. Without the expander in service, the residue compressor discharge check valve would close and the residue compressor would begin to operate in a closed recycle loop via its surge control valve, halting inlet gas flow through the train. Thus, it makes sense to shut down the train if the expander goes off-line.

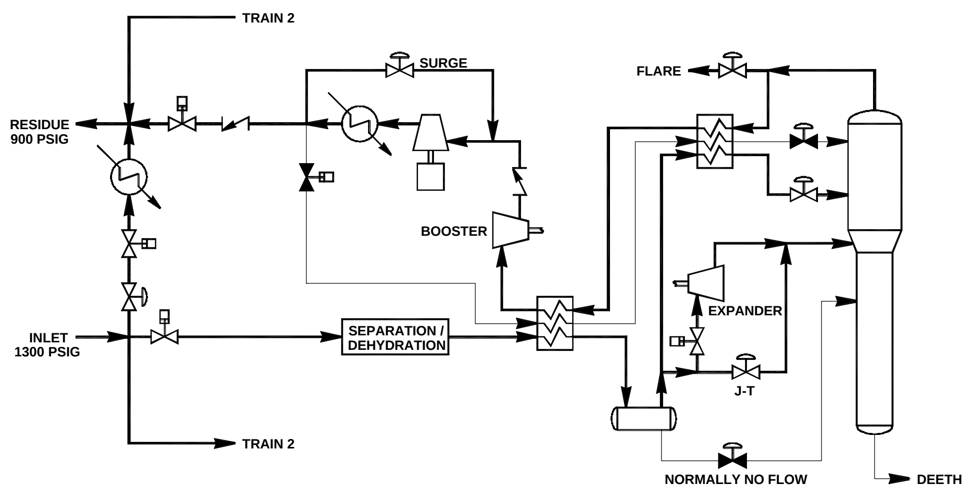

Figure 8 – Startup Flow Schematic

The following procedure is used to bring a train back on-line. Once its residue gas compressor is on-line (operating in total recycle mode via its surge control spill-back line), the expander for that train is started up by routing inlet gas through the expander and sending the demethanizer overhead vapor to the flare on pressure control. As the expander comes up to speed, the booster compressor differential pressure and residue compressor suction pressure will increase until the residue compressor discharge check valve opens. As soon as the check valve opens and gas starts leaving the system, the demethanizer pressure drops below the flare valve setpoint and the flare valve closes on pressure control. The plant is then back on-line, with the column pressure controlled by the expander inlet vanes.

The procedure given above is used at Aux Sable because the high inlet pressure prevents operating in a closed loop to "bootstrap" up the expander. The train inlet pressure is always 400-600 PSI higher than the residue delivery pressure. It would be possible to operate the plant in a closed loop and start up the expander without flaring if the piping and controls were modified. The existing dry-out loop piping, several additional pressure control loops, and a very large control valve would be needed to control the train inlet pressure and to introduce gas simultaneously with the residue compressor check valve opening.

CURRENT OPERATION

The Aux Sable NGL Recovery Plant is currently operating at conditions which are slightly different from the design cases. The inlet rate is lower than design at 1.5 BCFD and the current residue delivery pressure is 900 PSIG. The inlet composition is leaner in ethane than the lean case design basis and the CO_2 concentration is above design at 0.75 mole percent. One result of this is that the potential ethane recovery while reboiling to a CO_2 specification will be lower than the design cases because the CO_2:C_2 ratio in the inlet is significantly higher.

Average ethane recovery is currently less than expected for the current conditions due, at least in part, to a continuous buildup of hydrates over time in the cryogenic heat exchangers. Higher than design pressure drops and lower than design heat transfer rates result in the current ethane recovery of around 62% while reboiling the column to meet the CO_2 specification. One source of water causing the hydrates has been found and piping changes related to the expander seal gas and regeneration gas lines in each train will be made in 2002 to eliminate this known problem. Additional investigation will follow, if needed. After these changes, plant performance should then match the numbers predicted for current operation.

CONCLUSIONS

The most challenging consideration for the design of this plant turned out to be the high inlet pressure, but not because of any unexpected phase behavior. Rather, it was because of the high pressure itself and the corresponding high demethanizer pressure, and the fact that the horsepower generated by the expander has such a significant impact on the ethane recovery. Operationally, the plant is easier to run than most because of its simplicity and operating flexibility.

Enough flexibility was included in the original design to allow efficient operation over a wide range of product requirements and flow rates. At the same time, the capital cost of the plant was minimized by using the simplest equipment arrangement compatible with operation of the plant at the high inlet pressure.

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