# PRACTICAL TROUBLESHOOTING

# TECHNIQUES

# FOR CRYOGENIC GAS PLANTS


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## ABSTRACT

Almost any cryogenic plant performance problem can be modeled and conclusively identified, since modern process simulators can accurately predict actual plant performance. This paper describes a systematic approach to plant troubleshooting based on process simulations. The proposed troubleshooting logic should help operators and engineers minimize the time and expense necessary for identifying cryogenic plant problems and deciding what to do about them.

The recommended troubleshooting approach includes the following steps: (1) model the original plant design; (2) simulate the expected plant performance for the current inlet conditions and pressure profile; (3) develop a model matching the actual field data; (4) analyze any process equipment problems; (5) simulate the target plant performance assuming no pressure profile problems; (6) analyze any pressure profile problems; and (7) make changes in the plant to achieve acceptable plant performance and verify the performance change with the models.

A key feature of this approach is that process equipment problems are first distinguished from pressure profile problems and then either solved or accepted before analyzing any pressure profile problems.

When evaluating the economics of each potential modification, the plant models allow the engineer to quantify the corresponding incremental increase in product recovery. The goal of any plant troubleshooting effort is to achieve actual plant performance (product recovery) as close as possible to the target plant performance, within the constraints of good economic sense. The techniques presented here should help in achieving this goal as quickly and inexpensively as possible.

# PRACTICAL TROUBLESHOOTING TECHNIQUES FOR CRYOGENIC GAS PLANTS

## INTRODUCTION

Ortloff Engineers, Ltd. is routinely involved in troubleshooting cryogenic gas plants in support of its process retrofit and plant upgrade business. Before the performance benefits of a process retrofit can be identified, we must evaluate overall plant performance and the performance of individual equipment items such as heat exchangers and columns. These efforts have led Ortloff to develop a standardized approach to plant performance evaluation.

Ortloff is also routinely involved in helping plant owners with troubleshooting and improving existing plant performance, even when a process retrofit is not under consideration. The pitfalls encountered in these efforts are very similar to those found in the pre-retrofit evaluation. The systematic overall strategy for approaching any cryogenic plant processing problem described in this paper was developed from our experience in these two areas.

### Plant Performance Problems

For the purposes of this paper, a plant performance problem means a low product recovery problem. Operational stability problems are not addressed. It is assumed that the recovery level is stable and repeatable but less than expected, or that the recovery has suddenly decreased over a short period of time (i.e., a day or a week).

Perceived plant performance problems are included in the scope of this paper. Many times we find that the feed composition, flow, inlet pressures, and residue pressures have changed significantly from the original design, and that no one knows for sure how the plant should be performing — there is just a feeling that the performance should be better than it is. The troubleshooting steps presented here will help separate real from perceived performance problems.

### Applicable Process Designs

The examples presented herein are taken from standard single-stage expander plants (Figure 1) and Ortloff's Gas Subcooled Process (GSP) design plants (Figure 2), but the troubleshooting steps are the same for any expander plant. The focus of the paper is the cryogenic process equipment downstream of the dehydration system through the residue compression system and through the bottoms liquid exchanger. Upstream gas treating, downstream liquid treating, and fractionation are excluded from the scope of this paper. The techniques and logic can be applied to plants that have external refrigeration systems, a pre-boost expander/compressor, or inlet compression.

The examples given are for ethane recovery plants. The troubleshooting logic, however, can be applied to propane recovery plants as well.

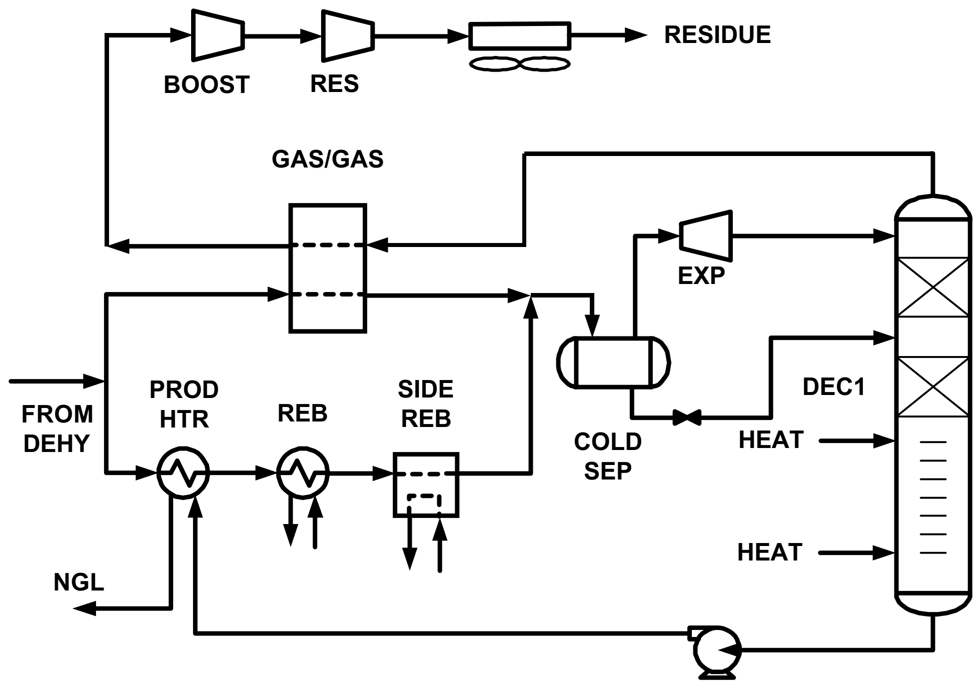


Figure 1 – Single-Stage Expander Process

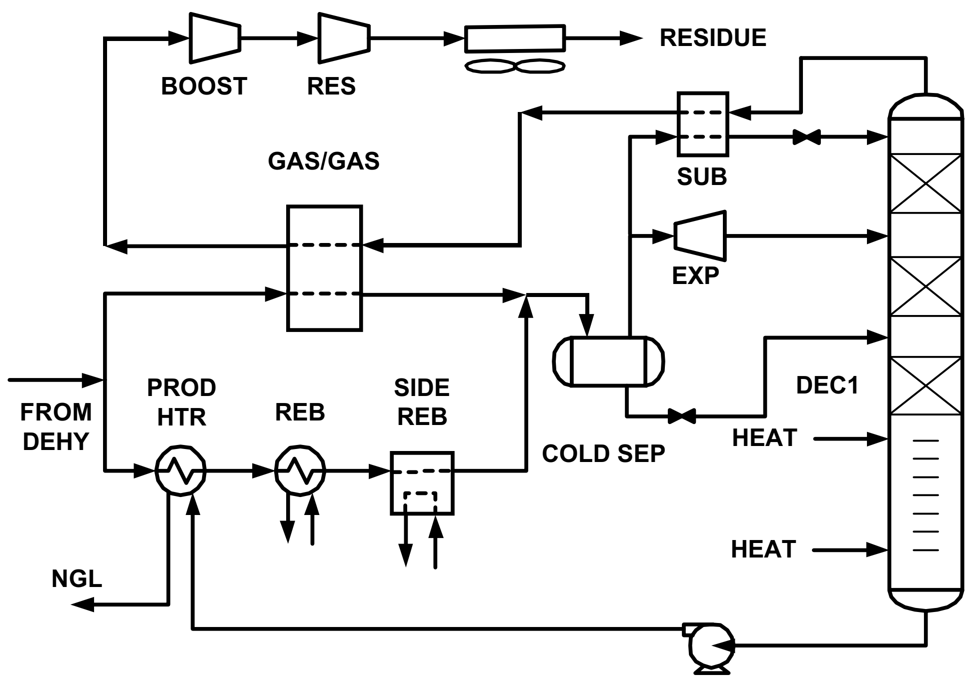


Figure 2 – Ortloff's GSP Design

## Minimum Requirements for the Troubleshooting Effort

You (the engineer involved in troubleshooting) must have access to and knowledge of process simulation and plant modeling. Initially, you must have access to the original process flow diagrams and material balance information. You also need a set of current mechanical flow diagrams (piping and instrument diagrams). The plant personnel must be able to supply accurate gas chromatographic (GC) analyses of the plant inlet gas stream, the residue sales stream, and the demethanizer bottom liquid stream. Accurate process temperature, pressure, and flow data must also be available. You will eventually need access to most plant engineering data and the mechanical job books.

## TROUBLESHOOTING STEPS

When a plant troubleshooting project is first initiated, the tendency is to go immediately to the plant site. If at all possible, perform some preliminary work first. Once the preliminary work is completed, you can schedule an effective plant trip. Plan the activities for that trip once you know much more about what to expect. This will become clear in the discussion that follows.

Plant performance problems fall into two broad categories: (1) process equipment problems, and (2) pressure profile problems. Process equipment problems include, for example, exchanger fouling and poor column performance. Pressure profile problems include high pressure drop, as well as compressor and compressor driver performance problems. We have found it very helpful to identify which category of problem we are dealing with as early as possible during a troubleshooting assignment. Another benefit of this approach is to quantify changes in product recovery that can be attributed to feed condition changes alone, thus distinguishing the perceived problems from the real ones.

**Summary of Troubleshooting Steps**

1. Model the original plant design (Case 1).
2. Determine expected performance at current feed conditions and pressure profile (Case 2).
3. Determine and quantify actual plant performance by matching field data (Case 3).
4. Compare Case 2 and Case 3 results, then identify, analyze and address any process equipment problems.
5. Determine target plant performance at maximum pressure ratio (Case 4).
6. Compare Case 4 and Case 3 results, then identify, analyze and address any pressure profile problems.
7. Make changes and verify performance change by updating Cases 2, 3, and 4, and comparing results again.
8. Achieve an economically acceptable plant performance level and stop.

These steps are shown in the logic diagram below (Figure 3) and are discussed in detail in the sections that follow.

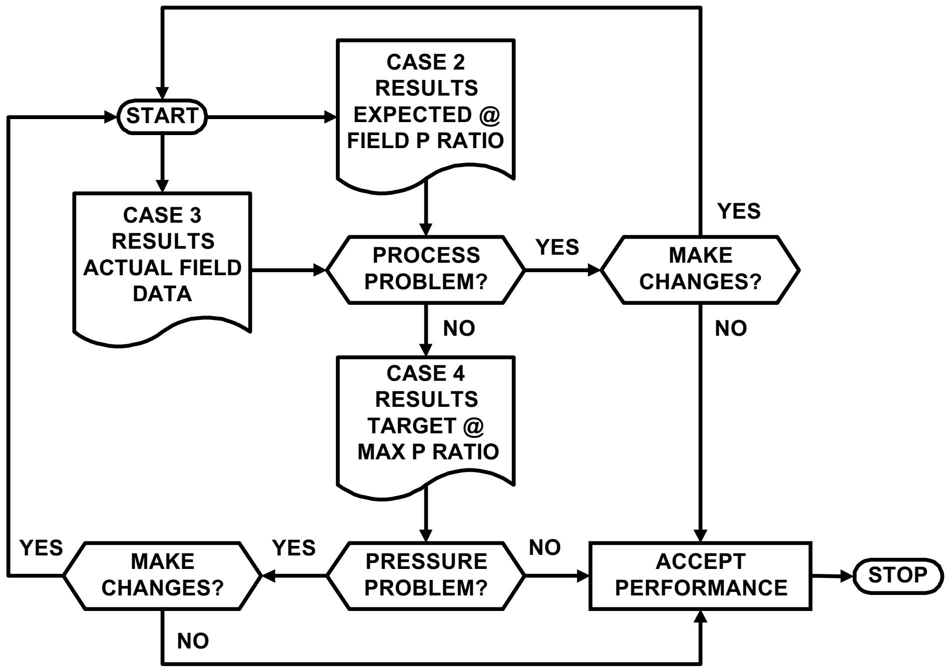


Figure 3 – Troubleshooting Logic Diagram

## MODEL ORIGINAL PLANT DESIGN (CASE 1)

**Requirements**

The first step is modeling the original plant design, which requires the original process design flow diagrams and material balance. The purpose of building this model first is to have access to the original design basis for the process equipment and determine if the original calculations are viable. You can determine the original design exchanger UAs (overall U-value times the exchanger surface area) even if the manufacturer's data sheets are not available. Examine the original feed composition and inlet conditions of flow, pressure, and temperature, as well as the column bottoms specification. In addition, examine the phase envelope margin at the cold separator and the physical properties at each stage in the demethanizer for any potential problem areas. Also, determine the compression horsepower and compressor throughput design basis.

Include all the available heat exchanger rating data in the model and check it against the original design at this time. Embedding this information in the original design simulation file wherever possible will speed up the performance evaluation in subsequent simulations.

Complete the model of the original design before going to the plant. The original design simulation will always be needed at some later time regardless of how far the current operation is off from the original design conditions. The exchanger UAs, column stages, expander and compressor efficiencies, and pressure profile are used for reference and comparison later.

**Simulation Guidelines for Case 1**

The original design recovery and temperature can usually be matched with current process simulators by using the same physical properties package as was used originally. Good agreement may not be achieved with different physical properties packages. We usually start with Peng-Robinson properties and equation of state. Since the original design simulation is used to determine the design basis for the equipment, it is more important to match the original design temperatures for this simulation rather than the original design recovery. Use the properties package which best matches the original design temperatures and duties. However, regardless of what was used to match the original design, use the Peng-Robinson or Soave–Redlich-Kwong properties package for all subsequent runs (Cases 2, 3, and 4 below). (Based on our experience, the Peng-Robinson equation of state is the best available for modeling cryogenic natural gas plants.)

Begin by assuming 82% adiabatic efficiency for the expander and 75% polytropic efficiency for the booster compressor. For small plants (<50 MMSCFD) using smaller frame size expanders, use one percentage point less; for larger plants (>300 MMSCFD), use one percentage point higher. Assume 1-2% bearing loss in calculating the horsepower available to the booster compressor from the expander. Manufacturer's data, if available, can be used, but the guaranteed efficiencies may have been used in the original design, not the calculated efficiencies.

If no curves are available for the compression equipment, assume 75% polytropic efficiency for a centrifugal compressor and 80% for a reciprocating compressor. Use higher numbers for very large machines and lower numbers for smaller or high-speed machines.

Start with an assumption of 60% tray efficiency in setting the number of stages in the model for a trayed demethanizer column. Use an HETP of 1.5 to 3.0 feet for packing, depending upon the packing size.

Make note of the exchanger UAs, horsepowers, efficiencies, pressure drops, and recovery. Check the $CO_2$ freeze margin, the cold separator phase envelope margin, and the physical properties of the stage liquids and vapors in the column. Cold separators operating within about 10°F of the full condensation temperature will not be stable. Density differences less than about 20 lbs/ft$^3$ between the stage liquid and stage vapor in the column require special tray spacing, as do surface tensions less than 2.0 dyne/cm$^2$.

## EXPECTED PERFORMANCE AT FIELD CONDITIONS AND PRESSURE RATIO (CASE 2)

### Simulation Guidelines for Case 2

The next step in the troubleshooting sequence is to use the field inlet composition, flow, temperature, and pressure to determine what the plant performance should be for the current field pressure profile. Not all of the field data is needed at this time to develop this simulation. Try to complete this run before going to the plant if at all possible.

Start with the original design simulation, Case 1, and change to the field inlet composition, flow, temperature, and pressure. Also input as a minimum the cold separator pressure, the column pressure, and the column bottom specification. Do not change the column stages and feed locations at this time.

Adjust the temperatures and flow splits to converge on all the exchanger UAs or ratings. When possible, rigorously re-rate shell and tube exchangers and try to get within 5% of their design rating. For plate-fins, if you do not have a way to rigorously re-rate them and the flow rates are not within about 10% of the original design values, then correct the target UAs by the factor (actual flow rate / original design flow rate) raised to the 0.7 power. The expander and booster compressor efficiencies should also be corrected for flow rate. The objective is to determine what the plant recovery and temperature profile should be for the current inlet composition and pressure profile.

**Evaluate Case 2 Results**

After this run is complete, you should have a good idea of exactly what field temperature profile and recovery to expect **IF** there are no equipment problems such as fouled exchangers, damaged column internals, or low expander efficiency.

Check the expected temperature profile against the original design for encroachment on the minimum design temperatures of the materials used in the plant equipment. This can be a problem when the inlet gas has "leaned up" substantially from the original design. The column temperature profile can become much colder at the original operating pressures when this occurs. As a minimum, check the side reboiler draw and return piping, the cold separator, and the expander inlet piping. Check the reboiler and side reboiler hydraulics, too.

Make note of the exchanger UAs, horsepowers, efficiencies, pressure drops, and recovery. Check the $CO_2$ freeze margin, the cold separator phase envelope margin, and the physical properties of the stage liquids and vapors in the column. Identify potential phase envelope, physical properties, and $CO_2$ freeze problems from this simulation just as you did for the original design simulation. A comparison of the Case 2 and Case 3 vapor and liquid traffic in the column will also help identify stages in the column where significant deviations from the original design are expected — but watch for lower as well as higher than design numbers.

If time permits, re-rate the column internals for the Case 2 conditions, make any necessary adjustments in theoretical stages, and converge on the exchangers one more time.

This simulation allows you to identify any overly optimistic expectations the plant owner may have for the plant with its current feed composition and pressure profile. Many plants have never been rigorously re-rated for the current feed composition and pressure profile. Preparation of this run may save everyone a lot of time and money by providing a reasonable expectation of plant performance for comparison with the actual performance. It is quite common to underestimate the effect of a reduction in inlet pressure or an increase in the richness of the inlet gas on product recovery.

**Development Logic for Case 2**

The logic diagram for generating the expected performance simulation is shown in Figure 4.

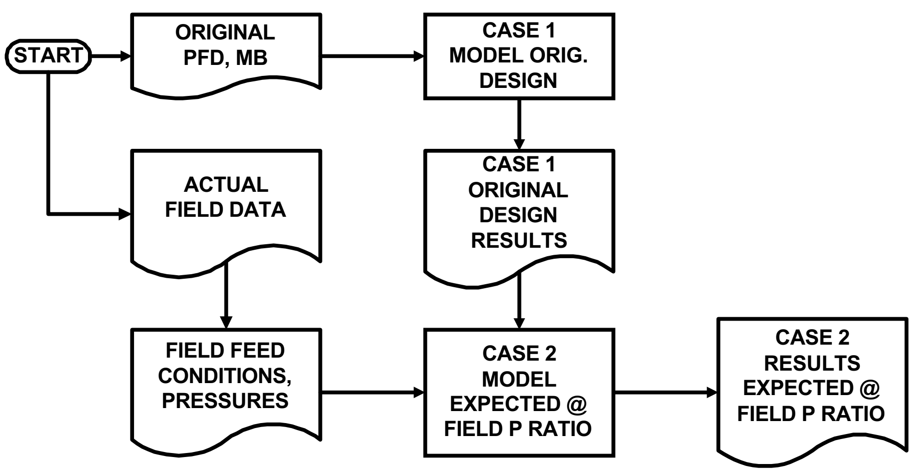


Figure 4 – Case 2 Logic Diagram

This run provides the expected performance data needed to evaluate the actual operation of the plant as described in the next section. You are now prepared to analyze the actual plant performance.

# DETERMINE ACTUAL PLANT PERFORMANCE (CASE 3)

**Recommendations for Obtaining and Evaluating Field Data**

Use steady-state snapshot data, not 24-hour averages. The plant should have been steady for several hours before the snapshot is taken. Ask that the GC samples be taken at the same time that the process data snapshot is taken. Obtain DCS printouts for the snapshot if possible. Ask that the demethanizer bottoms sample be taken upstream of any surge tanks or product storage vessels, preferably directly off the column bottoms pipe. If the product comes from a kettle reboiler, take the sample off of the liquid outlet from the kettle. Check the data when it is first received for obvious problems, such as temperature crosses and pressure increases where decreases are expected. Many field data problems can be caught with a cursory examination of the data before a simulation is attempted. Do not waste your time trying to model questionable data.

**Simulation Guidelines for Case 3**

Start with the Case 2 simulation. Remove the constraints on the exchanger UAs and, instead, specify as many of the temperatures and flows per the actual field data as possible without causing inconsistency errors in the simulation. Leave the pressures alone, since they were used to develop Case 2 anyway. Obtain as close a match as possible with the field data. Then compare the residue gas composition from the simulation to the GC analysis reported from the plant. Also compare the demethanizer bottoms compositions.

A good match between the field data and the simulation exists when the column temperatures agree within 1°F, all the exchanger temperatures agree within 2°F, and the GC analyses agree within 2% on a component flow basis.

Good agreement is rarely achieved between the simulation and the field data on the first attempt. Usually there is some bad field data and occasionally there are errors or omissions in the model. It is not uncommon to accept poor agreement simply out of frustration. Keep in mind that modern simulators are extremely accurate and that good agreement with accurate field data is not only possible but also necessary to properly identify and solve the plant performance problem. When good agreement is not achieved, there are only two possible reasons: bad data or a bad model. Each possibility is addressed below.

***Bad Data?***

A repeatable but inaccurate plant instrument is still very useful to plant operations, but it will cause problems when modeling the plant. Invariably plant personnel report that all their instruments are good and that they can be used without concern for their accuracy. Instrument repeatability is not sufficient for plant testing and troubleshooting, so some problems with instrument accuracy may surface during the troubleshooting analysis and modeling that were not discovered in the daily operation of the plant.

GC data can be checked independently of other data. The sum of the products should equal the plant inlet. A simple component splitter block can be inserted into the simulation that will force a fixed percentage amount of each component to the NGL product with the remainder to the residue. Use the resulting overall estimated product flow rates to calculate the sum of the products from the GC analyses. Then compare the calculated product compositions to the reported numbers to see which GC analyses, if any, are in error.

Generally, the vapor phase analyses are more reliable than the liquid analyses. For plant troubleshooting, the feed composition and the residue composition are most often used for comparison to the models and the liquid product is calculated by difference.

Another check of the GC data can be performed if the column overhead temperature and pressure and the column bottoms temperature and pressure are known. Input the GC data compositions into the simulator and specify the stream pressures as measured in the plant. Then calculate the dew point temperature of the overhead and the bubble point temperature for the bottoms product. These temperatures should agree very closely with the actual column overhead temperature and bottoms temperature. If they don't, then the analysis is not correct, or there is a carry-over problem in the column (example later).

Since the simulator forces a heat and material balance, many instrument discrepancies will become apparent when the field data is entered into the model. Some combinations of temperatures and flow rates simply are not possible. When this happens, you must decide which data is accurate and which instruments must be checked in the field. Some of the most common problems with field data include the following:

1. Calibration errors for GCs and transmitters.
2. DCS configuration errors for flows, i.e., no square root extraction or two square root extractions.
3. Instrument installed in different location from that shown on DCS graphics and operator screen.
4. Poor location of element in piping to read the desired process variable, such as a column pressure transmitter at the residue line on the gas/gas exchanger rather than on the column bottoms.
5. Temperature elements installed at mixing tees rather than on the cold separator.
6. Incorrect meter factors for non-compensated flow transmitters.

The quickest way to resolve inconsistencies in the data is to analyze the heat and material balance across individual pieces of equipment. Usually several temperature indicators and transmitters have to be calibrated before a consistent set of field data is obtained. Discrepancies in the instrument locations are also identified at this time. Obviously, modeling the plant data can become an iterative process. The tendency is to give up too soon and accept the model as the best one can do. Giving up too soon is counterproductive, especially if the plant has a mechanical problem and the data is accurate.

***Bad Model?***

The problem with accepting the model too soon is that there may be a mechanical problem in the plant that must be included in the model to properly represent what is happening in the field. The model initially represents what we expect, and often needs to be modified to represent what we observe. This is demonstrated in the following examples for column carry-over, a reboiler or side reboiler leak, and poor column performance.

- **Column Carry-over**

The symptoms of column carry-over are low ethane recovery, with larger than normal amounts of propane and small amounts of butane in the column overhead. The residue gas temperature leaving the gas/gas exchanger will be cooler than expected. The cold separator may be cooler than normal. It will be impossible to match the residue gas temperature profile because the simulator assumes the column overhead is pure vapor.

The key is to check the residue gas sample dew point. The only way the dew point can be warmer than the column overhead is if liquids are being carried out of the column. The liquids are then vaporized in the subcooler and gas/gas exchangers. The heat of vaporization of the liquids is what is missing in the model.

The model can easily be corrected to include the effects of the carry-over. All you need to do is add a side draw to the column operation at the top stage and then mix it with the overhead upstream of the first exchanger, as shown in Figure 5. Specify the rate as some portion of the total first stage liquid rate and let the model run. Adjust the side draw rate until the calculated residue composition agrees with the field data. The temperatures should then show agreement with the field data.

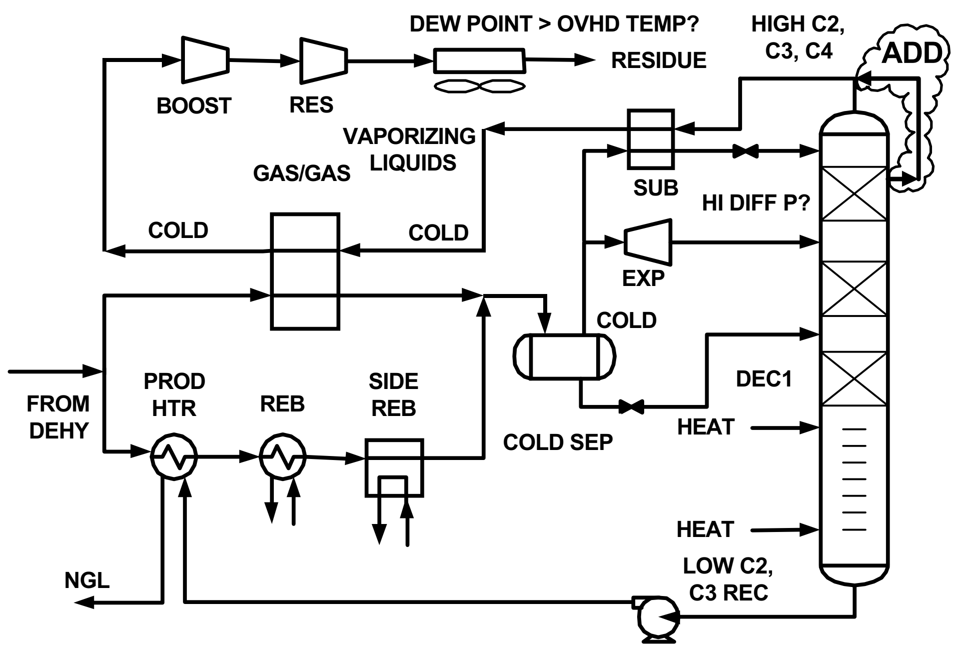


Figure 5 – Modify Model for Column Carry-over

Once the model and field data are in agreement, a little more detective work is needed to determine why the carry-over is occurring. The mist pad may be overloaded or damaged, or there may be hydrates or $CO_2$ obstructing a significant portion of the column cross-sectional area. The column differential pressure across the affected section will be higher than normal if there is an obstruction.

Plants that have front-end treating for $CO_2$ may exhibit carry-over after repeated treating system upsets. This is because high $CO_2$ content in the cryogenic plant feed, even for only short periods of time, can result in the $CO_2$ freezing and building up in the column. The buildup will continue until carry-over problems develop. If this is the case, then (after the treater problems are solved) the plant can be warmed up enough to allow the $CO_2$ to dissipate and the obstruction will disappear if there are no hydrates mixed with it. It may be necessary to shut down the expander for a few hours to warm up enough to quickly dissipate the $CO_2$ solids. A shutdown should not be necessary. Determine the success of the thaw by monitoring the propane and heavier content of the residue gas. The quantities of the heavier components in the residue gas should return to normal when the obstruction is gone.

Hydrates are much more stubborn. Only shutting down the plant, warming up all the cryogenic equipment to 100°F, and going through a complete dryout can successfully eliminate them. Cutting corners will result in either not removing all of the hydrate or moving the water (via methanol) to another troublesome location in the plant equipment and piping.

- **Reboiler / Side Reboiler Leak**

The symptoms of a reboiler or side reboiler leak are poor ethane recovery, warm temperatures throughout the expander plant, problems with making the methane specification at the bottom of the demethanizer and high heat input requirement at the column bottoms. The column or residue-side low-pressure relief valve may also relieve on a plant ESD when all the shutdown valves are working properly. Matching the field data will not be possible without changing the model to include the problem.

The key evidence here is the difficulty in making the methane specification in the bottom of the column. Typically, a plant experiencing these symptoms has a plate-fin or brazed aluminum side reboiler and bottom reboiler exchanger combined into one core. Many of these exchangers have failed due to thermal stresses, allowing high pressure inlet gas to pass through the exchanger return piping directly to the column. The introduction of this high methane content gas directly to the lower section of the column makes it difficult to maintain the methane specification in the bottom. When the leak first develops, it may be small enough so that the methane specification can be met by an increase in reboiler heat. As the leak worsens, additional heat may not be sufficient. The product recovery starts dropping as the reboiler heat is increased.

Again, the model can be modified to approximate the effects of the leak as shown in Figure 6, below. All that is needed is to add a tee in the inlet gas reboiler loop and route a portion of the flow through a valve and then on to the column as a separate feed at the same stage as the side reboiler or reboiler return line. Some trial-and-error iterations are required to match what is observed from the plant data, but, once again, good agreement with all the plant temperatures is possible after the model is modified to include the problem.

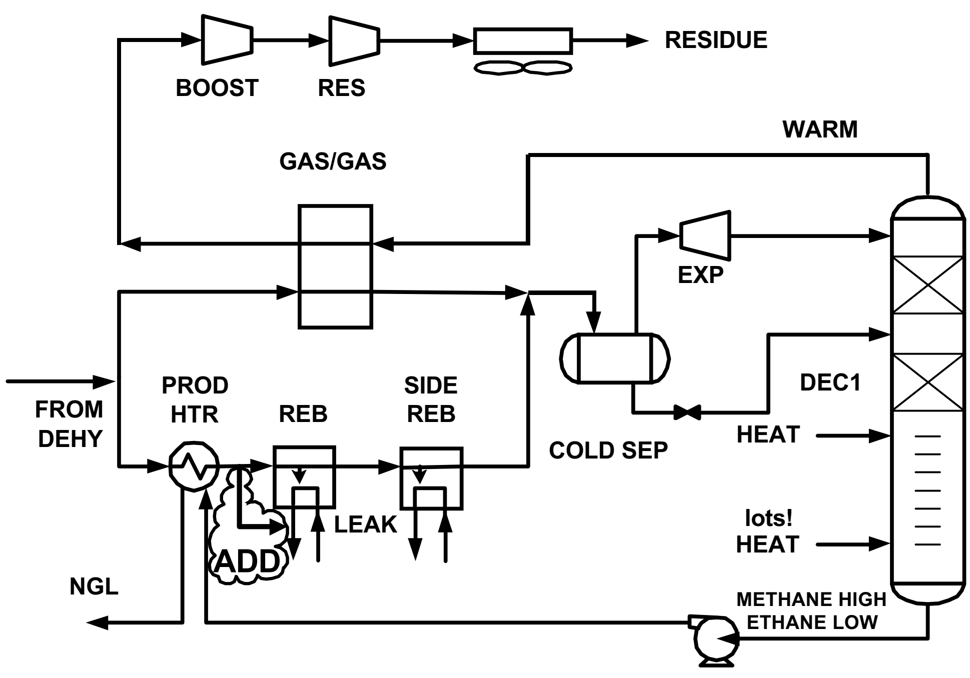


Figure 6 – Modify Model for Exchanger Leak

The solution to the plant problem is to replace the leaking heat exchanger with a newer design that is not as susceptible to the thermal stress failure. In extreme cases, it may be necessary to separate the two services into two cores, especially for plants in which the inlet gas stream has become much leaner than the original design. If this is the case, check the minimum design temperature, metallurgy, and pipe stresses for the side reboiler piping and the column.

- **Poor Column Performance**

The symptoms of poor column performance are poor recoveries and warm temperatures. The model will not match the field data with the number of stages in the column that were used in the original design. The side reboiler draw and return temperatures cannot be matched.

Reduce the number of stages in the column operation and iterate on the split between the side reboiler and reboiler duties until a match is obtained. Then compare the actual number of stages in the column with those in the original design model (Case 1). The differences can be due to poor turndown operation or a mechanical problem with the distributors or trays. Additional analysis using field column scans can be used to identify the details of the mechanical problem inside the column. Once the simulation matches the field data, however, the confidence level in improving the plant performance increases significantly.

Your Case 2 run can be used to re-rate the existing column internals to determine if a design change is necessary for the current operation. You may find that the column is working as would be predicted for the actual field conditions in Case 3. This may also be helpful in determining what steps to take next.

**Development Logic for Case 3**

This approach to modeling the actual plant operation is shown below in Figure 7. Start with the Case 2 model based on the actual inlet conditions and pressure profile and change the temperature profile and flows to match the field data. Then test the resulting simulation against all the field data. Until good agreement is achieved, either the field data or the simulation is modified and the iterations continue. Only when the model and the field data agree can you go to the next step, which is evaluating the actual plant performance.

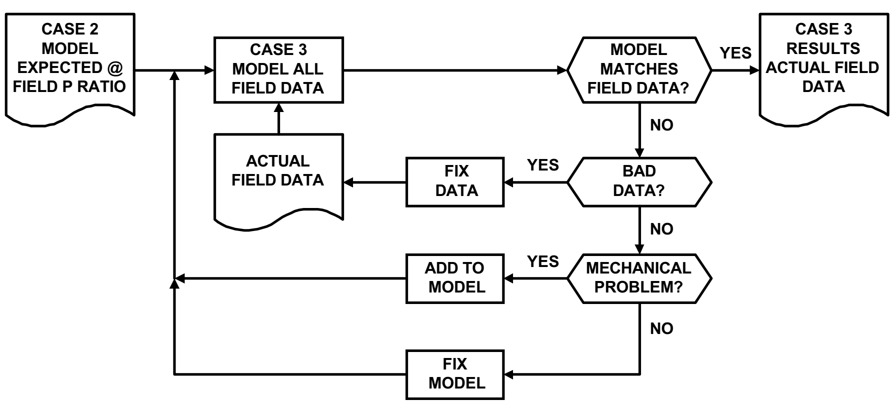


Figure 7 – Case 3 Logic Diagram

It is often very difficult to distinguish between data problems and model problems. The tendency is to accept a "close" model before it is sufficiently accurate to be used for analyzing plant performance. It is very important to achieve a good match of the field data with the Case 3 model before proceeding.

# PROCESS PROBLEM?

**Compare Case 2 and Case 3 Results**

Once you have the Case 2 run (the expected performance) and the Case 3 run (the actual performance match) completed, the simulation results can be compared and the less obvious problems identified. Tabulate the key performance data for Case 2 and Case 3 on one list, showing exchanger UAs, equipment efficiencies, column stages, and product recovery. Address the differences.

For example, the UA on the gas/gas exchanger may be only 50% of the value expected from the original design. The actual cold separator temperature in Case 3 will be warmer than expected from Case 2 and the recovery will be correspondingly lower. The column stages and expander efficiency may agree closely with the expected values, however. It will be obvious from this comparison that the exchanger has become fouled.

The fouling may have resulted from mole sieve dust, lube oil, or hydrates. Additional study of plant records and some testing may be required to determine the cause of the fouling before committing to a shutdown. "Puffing" the exchanger, chemically cleaning the exchanger, or warming up and drying out the plant may be required. But fixing the exchanger is not sufficient. The source of the fouling must also be identified and eliminated during the shutdown or the exchanger will become fouled again.

**Decide What to Do**

You (the troubleshooter) have two very valuable tools in deciding what to do once the problem has been identified. The Case 2 model represents what the plant should do if there were no process equipment problems at the current pressures and feed conditions. The Case 3 model matches what the plant is actually doing. The potential incremental product recovery improvement is the difference between the product streams in these two runs. Compare the product difference and the estimated cost to fix the plant (including any downtime) in an economic analysis to decide whether the incremental recovery is worth the cost. If the repair is too expensive, then the decision is made to accept the plant performance as it is. If the costs are reasonable, then make the changes to the plant and take a new round of field data. Update Cases 2 and 3 and repeat the process until no further improvement is economical. This logic is shown in Figure 8 below.

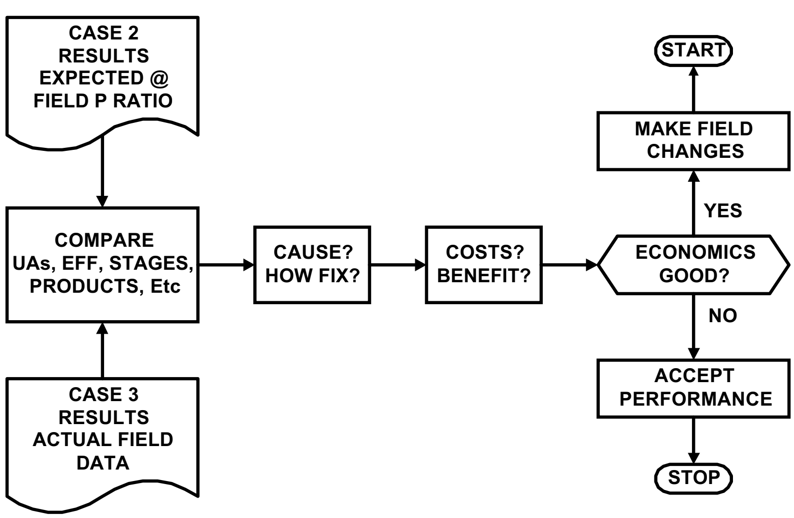


Figure 8 – Process Problem Logic Diagram

When the actual plant performance (Case 3) agrees with the expected performance at the current pressure ratio (Case 2), then at least the exchanger and column performance are acceptable for the current conditions. In other words, the product recovery is as good as it is going to be with the present pressure profile. The only way to increase the product recovery further is to increase the plant expansion ratio.

## PRESSURE PROBLEM?

**Simulation Guidelines for Case 4**

Analyze the current pressure profile next. Pressure profile problems include pressure drop problems, poor expander/booster compressor performance, and poor residue compressor or driver performance. The next step is to develop a new case, Case 4, for the target plant performance at the maximum achievable pressure ratio. The purpose of this case is to determine what the plant performance would be if the pressure drops were reasonable and the compression equipment performance is consistent with its performance curves and design horsepower capabilities. You need the compressor curves and driver data, as well as enough piping design drawings to do a point-to-point pressure drop study of any piping section suspected to have excessive pressure drop.

For a meaningful pressure drop study, analyze process piping pressure drops separately from equipment pressure drops (and rises). For example, calculate booster compressor pressures at the booster compressor flanges. Calculate the drops across an upstream orifice plate or suction screen separately and add them to the line loss from the gas / gas exchanger. You must include the pressure drops for interstage cooling on high-ratio residue compressors in the model. As another example, the pressure drop between the booster compressor discharge flange and the residue compressor suction flange could actually be as high as 15 PSI, and this can significantly affect the actual horsepower requirement of the residue compressor driver. The diagnosis can change from "sick compressor driver" to "restrictive piping design" based on the level of detail addressed in this pressure drop study.

Start at the plant fence and back-calculate the residue gas compressor discharge pressure for the flow rate in Case 3. Then calculate the compressor suction pressure corresponding to the residue compressor flow rate and target driver HP and speed. Add the line drop between the residue compressor suction and the booster compressor discharge to get the booster compressor discharge pressure. Input all the expected drops from the column overhead to the booster compressor suction. The booster compressor discharge pressure is calculated within the simulator by energy balance with the expander using the specified efficiency and the suction conditions.

On the inlet side of the plant, start at the dehydrator outlet and calculate the pressure drops to the cold separator, and then from the cold separator to the expander, and add them to the model. Allow some drop for the static head from the expander outlet to the column feed point and add that to the model.

Now, you must iterate the simulation to load up the exchanger UAs in the same way as you did to develop Case 2. The difference for Case 4 is that you also adjust the column pressure to hold the residue compressor suction pressure constant as the cold separator temperature and reflux split (for a GSP design) are optimized. The result is a Case 4 simulation that represents what the plant target performance should be for the assumed residue horsepower availability and the calculated pressure drops.

**Compare Case 3 and Case 4**

Now, compare the Case 3 actual plant performance and pressure profile to the Case 4 target performance and pressure profile. The difference in recovery and horsepower should be attributed to the pressure profile differences, either due to pressure drops or compressor performance.

First compare the residue compressor suction pressure and performance. What are the actual speed, pressure ratio, and driver HP compared to the target compressor performance? Is the compressor running at the expected speed? If not, why not? If it is, but the actual pressure ratio is low, what is the measured efficiency? Are both the efficiency and the pressure ratio low? Are the interstage labyrinths worn out? If the compressor efficiency is okay but the speed is low, is the gas turbine driver performance poor? Does the gas turbine need cleaning or re-blading? Is the maximum firing temperature control setpoint lower than design?

If the actual compressor operating point appears to be further out on the curve than expected, check for some recycle source through the plant, like a surge control valve open or leaking, or a leaking pressure control station.

If the residue compressor performance is acceptable, then move upstream to the booster compressor. Compare the target and actual pressure ratio and speeds. If the actual pressure ratio is less than the target and the temperature rise across the machine is greater than expected, the machine efficiency can be low due to worn wheel seals, or the surge control valve could be leaking. If the machine speed is low, the expander efficiency or horsepower input to the booster compressor may be low.

By far the most common pressure profile problem is poor performance of the gas turbine drivers on the residue compressors. These machines deteriorate slowly but surely over time, and the plant owners may not realize that the driver performance has deteriorated because the poor performance was attributed to other changes in plant operation.

The second most common cause of pressure problems is poor booster compressor performance due to off-design operation, mechanical wheel damage, or open bypass surge control valves. These are typically low-pressure ratio machines, making it difficult to monitor performance accurately using the local control panel gauges. For example, if a suction gauge reads 5 PSI low and the discharge gauge reads 5 PSI high, poor compressor performance may not be detected. Booster compressor performance is best determined with differential pressure transmitter readings, digital thermometers, and carefully calibrated suction flow measurement. The only way expander efficiency can be determined conclusively is by a horsepower balance with the booster compressor (plus the bearing loss). Matching outlet temperatures alone is not sufficient if the expander performance is really in question. (Determining expander performance is a topic for another paper.)

**Decide What to Do**

In analyzing the pressure profile you have completed two runs from which the incremental recovery can be calculated. Determining the cost of the repair is a separate problem. But the costs of each improvement in pressure ratio can be compared to the incremental increase in product revenue. Calculating the ethane recovery improvement due to a 20 PSI drop in column pressure or a 5% increase in gas turbine horsepower capability is not difficult once these models are completed.

The objective is to make all the improvements in the pressure ratio that can be justified economically. Eventually you will obtain an economically acceptable recovery level that is as close as practical to the target recovery level determined in Case 4. The logic is identical to determining if there was a process equipment problem, with only the cases being compared different, as shown in Figure 9, below.

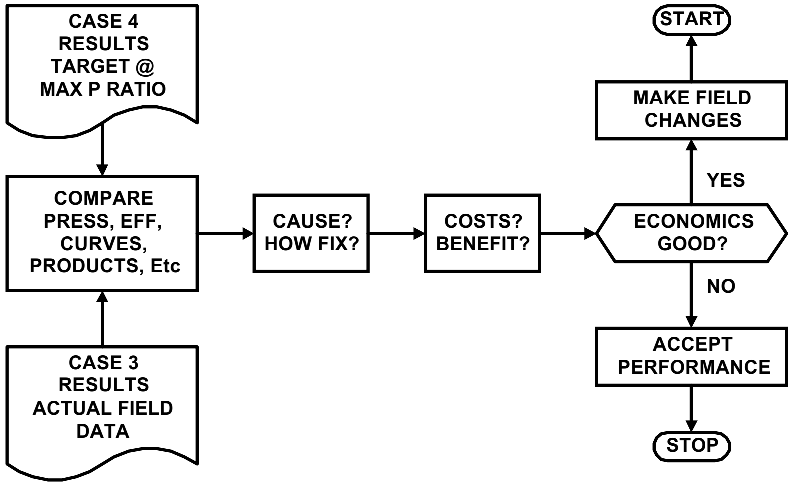


Figure 9 – Pressure Problem Logic Diagram

## RECOMMENDATIONS / CONCLUSIONS

We hope this systematic and logical approach to plant troubleshooting will help operators and engineers minimize the time and expense necessary for identifying problems and deciding what to do about them. The basic premise in our approach is that modern process simulators are sufficiently accurate in predicting actual plant performance that almost any performance problem can be successfully modeled and conclusively identified. When there is a discrepancy between the computer model and the plant data, either the data is bad or the model is bad. Bad data will always be a solvable problem. However, perseverance in fixing the model many times leads to identifying the real problem with the plant. You (the troubleshooter) must remain open-minded and completely logical in your approach to simulating the actual plant data if you are to be successful in solving the plant problem. Pre-judging the plant data or the simulation results will only delay solving the problem.

The approach presented here, in which a model is developed to determine the expected plant performance at the field pressure ratio, helps reduce troubleshooting time by separating the column and exchanger problems from the pressure drop and compressor problems. This approach immediately shifts the focus away from comparisons with the original design to a much more useful comparison to what the plant performance and temperature profile should be with the current feed composition and pressures. Development of an accurate model of the plant data becomes a critical task in the troubleshooting effort, especially if mechanical problems are involved. Only after the actual performance is compared to the expected performance at the same pressures does the focus then shift to an evaluation of the compressors and pressure drops. The pressure problems are then isolated and analyzed separately from the process equipment problems. These troubleshooting techniques can be successfully applied to almost any cryogenic expander plant problem.