Optimization of Natural Gas Processing Plants Including Business Aspects

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ABSTRACT

A new method to determine the optimum performance of natural gas processing plants has been developed. This methodology reduces the overall plant material balance equations into a linear form using the volatility of components and product specifications. Simulator response modeling relates key process variables to plant performance satisfying the remaining unknown information from the material balance equations. Rigorous economics are subsequently applied to the process model. This technique adequately combines contractual terms, product prices, and process information to calculate the optimum set of operating conditions for the plant offline. It is also a valuable tool to analyze the economic impact of processing additional streams and investigating new potential contract scenarios.

Proceedings of the Seventy-Ninth GPA Annual Convention. Atlanta, GA: Gas Processors Association, 2000.

INTRODUCTION

For most industrial processes a single company buys raw material and processes the material to convert it into a more valuable product(s). In the case of gas plants, the plant operating company typically does not own all of the raw feed stock. As a result, gas plant economics have unique characteristics that make it inherently more difficult to operate.

Gas plants have several distinctive features that cause economic complexity. One unique characteristic is the fact that several gas streams with different ownership can be combined at the plant inlet and processed as a group while the customer retains ownership in the same form throughout the process. Another interesting characteristic is that the plant enters into a contract with each customer for the terms and conditions under which the plant processes the raw gas. These contracts may be entirely different for each customer depending upon the age of the contract, the composition and amount of gas, plant recoveries, and the contractual preferences of the customer.

From a processing point of view, raw inlet gas may originate from nearby gas fields, be a product of another type of process *i.e.* refinery off-gas, or collected as associated gas from oil fields. As a result, the composition and volume of each plant inlet stream may be drastically different. Upon completion of processing, the products are allocated to each inlet gas owner based upon the composition and amount of gas contributed minus the contractual plant processing fees subject to any other contractual penalties for poor performance by either party.

The plant receives a processing fee *per* the terms and conditions of the gas processing contracts with each inlet owner. The plant processing fee is usually determined in one of three ways: a fixed processing fee, a 'keep-whole' contract, or retention of a portion of the produced liquids. A fixed processing fee agreement pays the plant a flat fee based upon the volume of inlet gas. A 'keep-whole' contract allows the plant to remove liquids from the gas and pay the supply company based upon the BTU value of the fuel and shrinkage.

In addition to processing fees, the plant may receive income from compression, marketing, or pipeline transmission fees. The situation may be complicated further depending upon the terms for fuel allocation and shrinkage. Contractual penalties may also exist for low recovery, insufficient inlet gas flow, low plant inlet suction pressure, high field pressure, high levels of impurities in the inlet or product, and lean inlet gas.

The goal of gas plant economic optimization is to utilize all available plant information to determine the set of economically optimum operating conditions. Over time, the typical gas plant finds itself processing gas under such a wide variety of contracts that it is difficult, if not impossible, to recognize intuitively the process conditions for the plant economic optimum. The optimum is tangled further by the frequency of product price changes and fluctuations in inlet stream flowrates. As a result, the economic optimum set of plant process conditions may be different for each gas supply company while the plant may have yet another optimum.

New Method

A new method has been developed to determine the optimum performance of gas processing facilities including business aspects. The new technique incorporates all contractual terms, product prices, and process information to calculate the optimum set of operating conditions for the plant. The major benefit to the new methodology is that it is tailored specifically to natural gas plants and therefore can utilize information about their structure to simplify process modeling. The result is a new methodology with an accurate reduced order process model combined with rigorous economic modeling. Presentation of the method is divided into three parts: process flow modeling, process energy modeling, and economic modeling. Each portion is described in the following sections. For more information on the details of the methodology see Bullin[1], 1999.

Process Flow Modeling

Gas plants commonly fractionate the raw gas into residue gas, ethane, propane, butane, and natural gasoline products based upon boiling point differences. During the separation process, some of the components are not affected by normal variations in process conditions. In the demethanizer, for example, the major separation is between methane and ethane. Carbon dioxide and propane may also appear in significant quantities in either the vapor or liquid product stream. Butane and heavier components, however, possess a low enough boiling point that 99.5+% of these components always are in the demethanized liquid product stream. Similar approximations exist for the deethanizer, depropanizer, and debutanizer. In addition, the boiling points of some components are so high relative to the others, that they no longer exist in the system during some separations. A summary of the simplifying assumptions on this basis appears in Table 1.

Additionally, the product purity constraints can be used as an approximation to the composition of the light components in the vapor products. The plant always tries to maximize the amount of impurities in a product stream. Utilizing these assumptions, the component material balance equations for an entire plant may be reduced significantly and combined into a linear form. This set of linear equations is very powerful because it relates each product stream composition and flowrate to the residue composition. As a result, the solution hinges on the composition of only a few components in the residue gas namely CO_2, C_2, and C_3. The next step is to develop a simple relationship between selected key plant control variables to the demethanizer overhead product.

The performance of the demethanizer column is sensitive to several feed parameters including changes in composition, flowrate, temperature, and feed location. These feed parameter fluctuations result from changes in upstream process variables. Common process variables impacting the demethanizer consist of plant inlet fluctuations, refrigeration load, and inlet pressure. It is desirable to correlate the performance of the demethanizer in terms of several upstream variables without introducing the difficulties associated with an equation-based approach. It is also desirable to develop a methodology to retain the accuracy of a rigorous simulation but without its complexity.

Table 1. Gas Plant Fractionation Simplifying Assumptions

Column	Component in Vapor Product Only	Component in Liquid Product Only	Component Not Present in Liquid
Demethanizer	N_2	C_4, C_5, C_{6+}	None
Deethanizer	C_1	C_5, C_{6+}	N_2, CO_2
Depropanizer	C_2	C_{6+}	N_2, CO_2, C_1
Debutanizer	C_3	None	N_2, CO_2, C_1, C_2

A compromise is to tailor rigorous simulations of the gas plant to generate a correlation of the residue stream composition as a function of a few influential control variables. A statistical response surface modeling design is used to select the minimum number of simulations to generate a given order model. A correlation is then developed for each unknown component composition in the residue stream as a function of all the key process variables. The composition and flowrate of the plant product streams may can now be calculated by a set of linear equations which are a function of the plant control variables.

Process Energy Modeling

Shortcut distillation[2] is applied to each column in the fractionation system to determine the energy required for separation. Additional plant energy uses are either calculated directly based upon stream flowrate for compressors, pumps, and heaters or correlated with process parameters using a rigorous steady state simulator. The process correlations are determined as a function of the key process parameters in the same manner as used for the residue stream. Each energy cost may be subsequently allocated to the appropriate party.

Economic Modeling

Because of the complexity of natural gas plant processing arrangements, a rigorous approach must be taken to plant economics. Each portion of gas plant economics must be incorporated into the economic model including: product allocation, fuel, shrink, fees, and contract terms. The economic calculations depend upon the product flowrates, plant energy requirements, product prices, and contract terms. The calculational methods for each portion of the plant economics are specified in the processing contract and have slight variations.

The ownership of the product streams is allocated based on the amount of each component fed by each inlet owner. Fuel costs are issued based upon the value of the residue gas consumed as fuel. Shrinkage is distributed back to the inlet owner based upon the heating value of the demethanized products. All additional economic terms are determined based upon the individual contract arrangements. Once all of the economic factors are calculated, the terms are combined to determine the gross plant revenue.

The new methodology creates a process model based upon tailored rigorous simulations of the gas plant over a wide variation of operating conditions and directly incorporates economics. Due to the specific application of the method to gas processing plants, only a small amount of information is required to predict the behavior of gas plants. Once the methodology is applied to a particular gas plant, the economic analysis possibilities

extend far beyond the calculation of gross plant revenue. Revenue to the plant from any portion of the contract *i.e.* fees or product retention, may be computed with variations in the contract arrangements, inlet flowrates, inlet composition, adjustable process variables, product prices, or contract scenarios. The process model is also in a form to easily generate the set of optimum plant operating conditions to maximize plant revenue. The flexibility of this method is demonstrated with the following two examples.

Example 1: Joule-Thompson (JT) Expansion Plant

A JT plant schematic diagram appears in Figure 1. This plant cannot achieve the high recovery of turboexpander designs, but it attains adequate recoveries without the large capital expenditure of a turboexpander. Three key process variables are available for manipulation: the chiller temperature, the inlet compressor discharge pressure, and the C_1/C_2 bottoms product specification. The goal is to use the new methodology to investigate the economic trade-off among variations in these key variables while investigating the relationships between product prices and contractual scenarios on the gross plant revenue.

Process Description

Two rich inlet gas streams enter the plant at 120 °F, 300 psig and 40 MMSCFD. The flowrate of each stream is 40 MMSCFD. The inlet gas composition for both streams appears in Table 2. The demethanized liquid product is sold without further processing as a Y-Grade product.

The plant processes stream A under a retention of liquids arrangement. The plant keeps 15% of each of the liquid products and does not pay for fuel or shrink in return for processing the gas. The plant processes stream B on a keep whole basis. As a result, the plant pays for the fuel and shrinkage allocated back to stream B in return for retaining all the liquids. These contract terms are summarized in Table 3. The product prices for this example appear in Table 4. Normal operating ranges for the key process variables appear in Table 5.

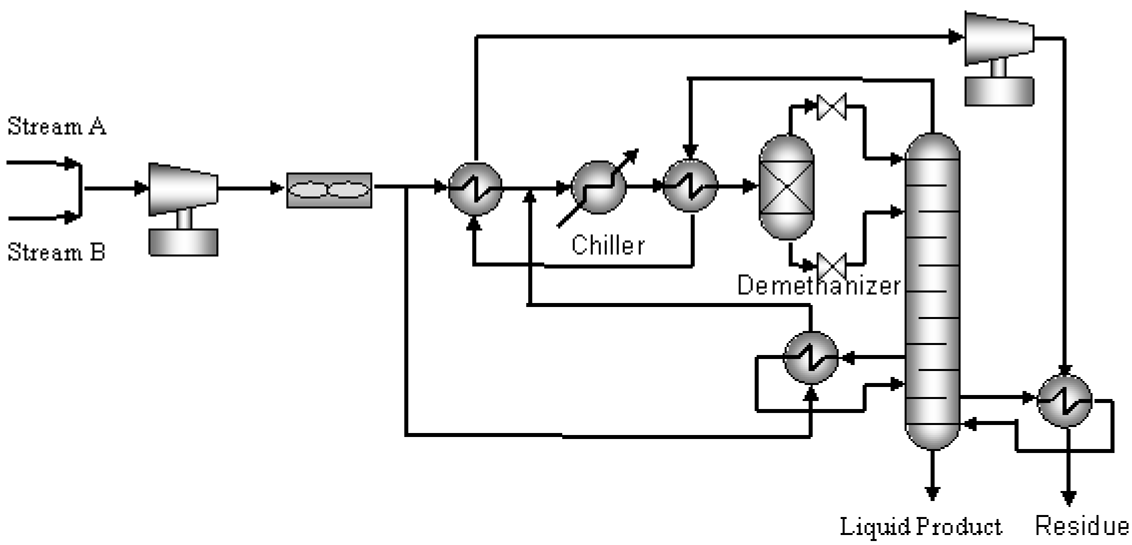

Figure 1. JT Plant Schematic Diagram

Table 2. Inlet Composition for Example 1

Component	Stream A (mol %)	Stream B (mol %)
N_2	4.20	1.85
CO_2	0.53	0.75
C_1	84.24	79.00
C_2	5.30	9.00
C_3	3.20	4.50
$i\text{-}C_4$	0.63	1.50

n-C_4	0.53	1.20
i-C_5	0.42	0.80
n-C_5	0.42	0.75
C_{6+}	0.53	0.65

Table 3. Summary of Contract Terms for Example 1

Plant Portion	Stream A	Stream B
Retention of Liq.	15%	100%
Fuel	0%	100%
Shrink	0%	100%

Table 4. Summary of Product Prices for Example 1

Product	Price
Lean Gas	$1.7500 / MMBTU
C_2	$0.1050 / gal
C_3	$0.1587 / gal
i-C_4	$0.2000 / gal
C_{5+}	$0.2200 / gal

Table 5. Normal Operating Ranges for Key Variables

Key Variable	Normal Operating Range
Chiller Outlet Temperature	-35 to 10 °F
Inlet Compressor Pressure	500 to 900 psig
Demethanizer Bottoms C_1/C_2	0.005 to 0.026

Process Model

A quadratic model of the process is created using PROSIM® for the rigorous simulation calculations and a central composite response surface modeling design[3]. The original process model required 30 rigorous simulations of the process to create a quadratic model.

Price Variations and Optimal Operation

Rigorous process simulations of the JT plant demonstrate that the maximum ethane recovery of the process is approximately 60% and this occurs at the process constraints. These process constraints are the coldest chiller outlet temperature (-35 °F), the highest inlet pressure (900 psig), and the highest C_1/C_2 ratio

(0.026). Operation at these constraints achieves the highest plant recovery, but with the highest fuel consumption. With high liquids prices, the economic optimum operating conditions are those that maximize liquids recovery. At a certain point, however, it is desirable for the plant to move away from these constraints and suffer a loss in recovery to reduce fuel consumption.

Figure 2 demonstrates this concept by fluctuating ethane price while fixing the other product prices at their current value. At each ethane price, the optimum set of key variables are determined to maximize the gross revenue to the plant under the given contract scenario. While these product prices may not reflect current market prices, this analysis is beneficial to understand how economic factors affect plant operation. For high relative ethane prices, the plant keeps the key variables at their constraints to maximize recovery. When the ethane price falls to about $0.10 per gallon, the economic optimum no longer occurs at the constraints on the key process variables, but how should the plant optimally reject ethane?

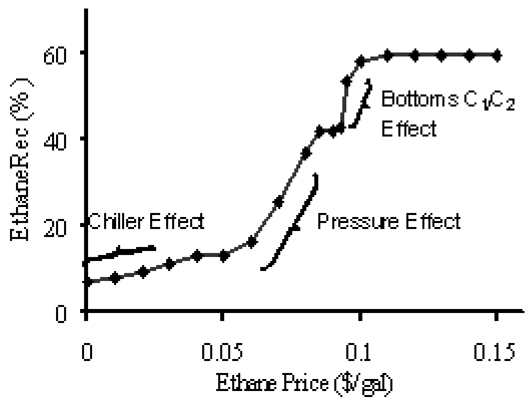

Figure 2. Ethane Recovery at Optimal Gross Plant Revenue vs. Ethane Price.

Figure 2 shows that the first process variable adjustment is the demethanizer column bottoms C_1/C_2 ratio. The C_1/C_2 ratio is the first key variable to be manipulated because it lowers the recovery of ethane while maintaining the recovery of the propane and heavier components.

As the ethane price falls further, a second process variable adjustment reduces the inlet compressor discharge pressure. This effect reduces the recovery of ethane by increasing the demethanizer inlet temperature and significantly reducing plant fuel consumption. The reduction in the inlet compressor discharge pressure begins at an ethane price below $0.085 per gallon. When the ethane price falls to $0.05 per gallon, the pressure is lowered to its minimum constraint of 500 psig.

Chiller adjustments begin to occur at prices lower than $0.04 per gallon. The chiller temperature is the last key process variable to be adjusted because it has the most dramatic effect on the propane recovery.

Contractual variations also have a direct influence on the gross plant revenue as well as the optimal point of operation. Figure 3 is a comparison of the gross plant revenue for different contract scenarios. As the figure demonstrates, the most conservative approach for the plant is to process gas for a fixed fee. Fixed fee processing allows the plant to operate independent of the current product prices. Fixed fee process optimization strategy is to increase revenue for the inlet owner to maintain the processing contract.

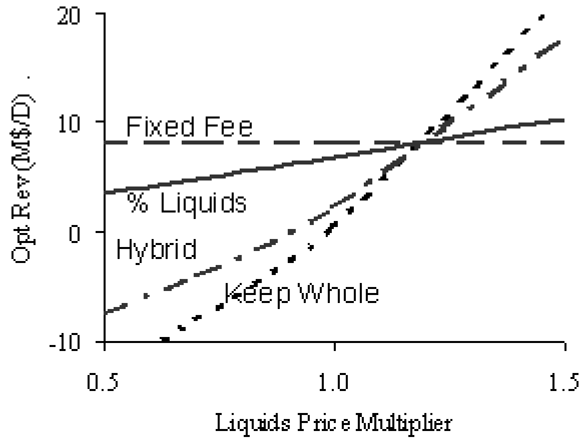

Figure 3. Optimal Revenue vs. Liquids Price Multiplier.

Another conservative approach is for the plant to operate in return for a percentage of the liquids captured. Retention of liquids contracts allow the plant to benefit while prices are high, but take a reduction in gross revenue when prices fall. Plant optimization in this mode requires the plant to generate as much of the liquid products as possible regardless of the product prices if the plant does not pay for fuel.

The most risky approach is for the plant to process the gas on a keep-whole basis. During periods of high prices, the plant receives more income than with other types of contracts. During low price periods, however, the plant loses significantly more money. Plant optimization for keep-whole contracts considers revenue from products as well as all processing costs.

Hybrid contracts offer the plant the opportunity to dampen the disastrous effect of low product prices at the expense of reduced income during periods of high prices. Figure 3 demonstrates an example of a hybrid contract scenario in which the plant receives 15% of the liquids from stream A while processing stream B on a keep-whole basis. The gross plant revenue for the hybrid scenario falls between the keep-whole and retention of liquids contracts.

The inlet owner's perspective generally is not included in the plant optimization process because the inlet owner does not have authority to run the plant. The inlet owner's portion of the gross revenue may also be determined from Figure 3. The amount of gross revenue the inlet owners receive is the amount of gross revenue in a keep-whole contract minus the amount paid to the plant.

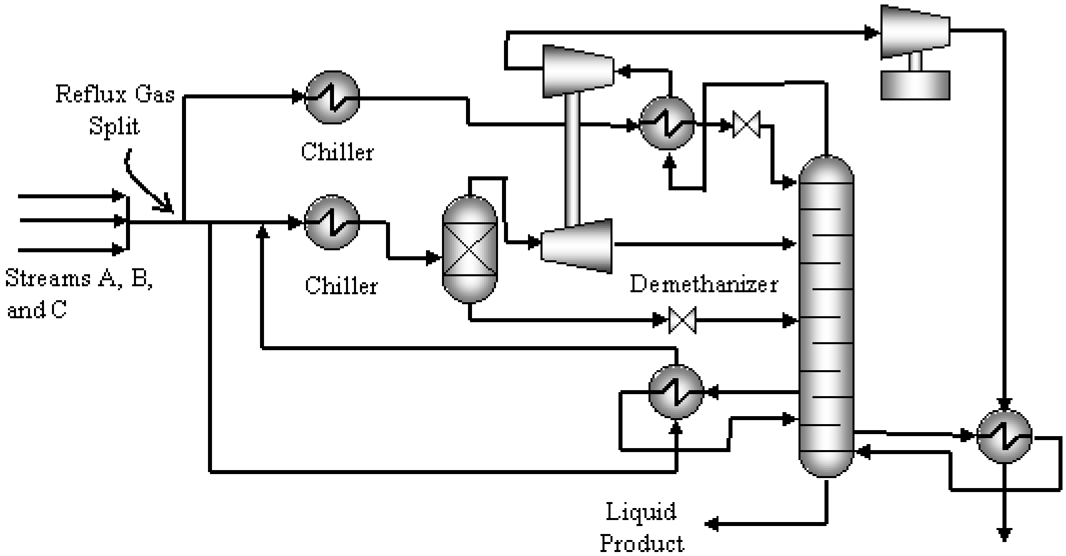

Figure 4. Refluxed Demethanizer Process.

Example 2: Gas Plant With A Refluxed Demethanizer

The refluxed demethanizer process is similar to the typical expander plant except a portion of the raw inlet gas is diverted through a chiller, cross-exchanged with the demethanized vapor product, flashed to column pressure, and injected at the top of the column. Refluxing the column in this manner injects cold rich liquid into the top of the column and tends to stabilize and wash ethane and heavier components from the rising vapor phase.

Two rich inlet gas streams (streams A and B) currently enter the plant at 120 °F and 900 psig. The plant has the opportunity to process a third stream (stream C) with a drastically different composition. Before accepting a new contract, the new methodology is applied to determine the effect of the new stream on the plant economics.

Process Description

The composition for each of the inlet streams appears in Table 6. The current flowrate of stream A is 70 MMSCFD and stream B is 7.5 MMSCFD. A schematic diagram of the process appears in Figure 4. The residue gas is partially recompressed using the shaft work generated by the expander and then compressed to 700 psig for pipeline injection. The demethanized liquid product is further fractionated.

The contract terms are summarized in Table 7. The product prices for this example appear in Table 8. The plant pipeline and marketing fees are shown in Table 9. The normal operating ranges appear in Table 10.

Table 6. Inlet Composition for Example 2

Component	Stream A (mol %)	Stream B (mol %)	Stream C (mol %)
N_2	0.16	0.4	11.543
CO_2	0.43	0.96	0.000
C_1	90.11	90.41	24.737
C_2	4.90	4.47	45.028
C_3	2.17	1.95	15.956
$i\text{-}C_4$	0.63	0.44	1.368
$n\text{-}C_4$	0.61	0.52	1.247

i-C_5	0.33	0.17	0.121
n-C_5	0.33	0.20	0.000
C_{6+}	0.33	0.48	0.000

Table 7. Summary of Contract Terms

Plant Portion	Stream A	Stream B	Stream C
Liq. Retention	100%	0%	5%
Fuel	100%	0%	0%
Shrink	100%	0%	0%
Fixed Fee	---	$0.12/MCF	---

Table 8. Summary of Product Prices

Product	Price
Lean Gas	$2.0000 / MMBTU
C_2	$0.1875 / gal
C_3	$0.2235 / gal
i-C_4	$0.2975 / gal
C_{5+}	$0.3200 / gal

Table 9. Plant Pipeline and Marketing Fees

Stream	Fee	Ethane	Propane	Butane +
A	Marketing			
	Pipeline			
B	Marketing	$0.005/gal	$0.005/gal	$0.005/gal
	Pipeline		$0.01/gal	$0.02/gal
C	Marketing	$0.005/gal	$0.005/gal	$0.005/gal
	Pipeline		$0.01/gal	$0.02/gal

Table 10. Normal Operating Ranges for Key Variables

Key Variable	Normal Operating Range
Chiller Outlet Temperature	-35 to –20 °F
Reflux Split	25 to 35%
Demethanizer C_1/C_2	0.005 to 0.026
Flowrate of Stream C	0 to 15 MMSCFD

Process Model

A quadratic model of the process is created using PROSIMÒ for the rigorous simulation calculations and a Box-Behnken[3] response surface modeling design.

The gross plant revenue increases with the flowrate of the new stream (stream C) at an almost constant rate of $250 per day per MMSCFD as shown in Figure 5. This increase in gross revenue is surprising because the ethane recovery drops dramatically with increases in the flowrate of stream C. Figure 6 shows the dramatic drop in recovery from about 97% to 83% for 15 MMSCFD of the new stream.

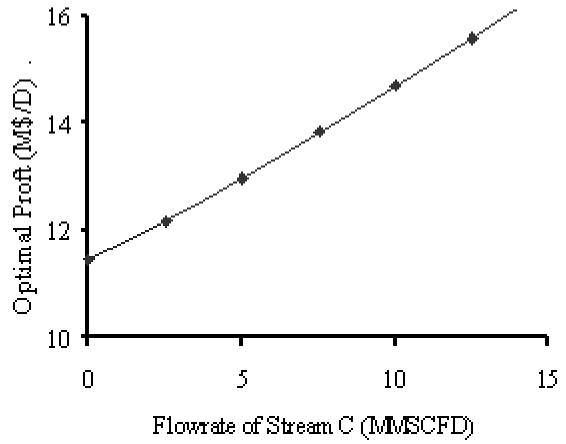

Figure 5. Optimal Plant Profit with New Stream

Given the decline in recovery with each increase in the flowrate of stream C the question naturally arises: Why does the gross plant revenue increase so dramatically when the recovery falls? The answer to this question stems from an understanding of the relationship of the economics.

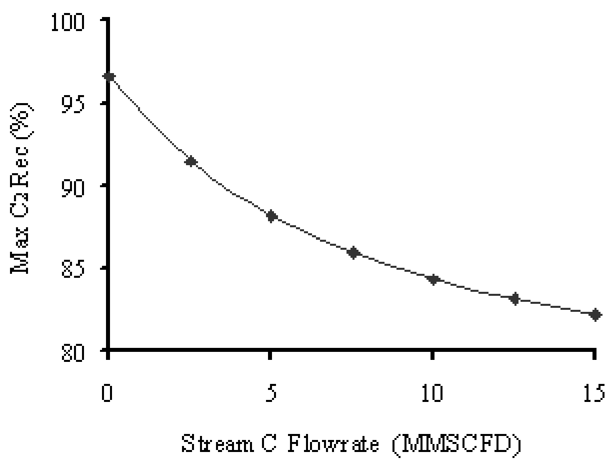

Figure 6. Maximum C_2 Recovery vs. Stream C Flowrate.

Figure 7 displays the relation between the plant revenue generated from stream A and the flowrate of stream C. The revenue from processing stream A is directly related to the recovery of the process.

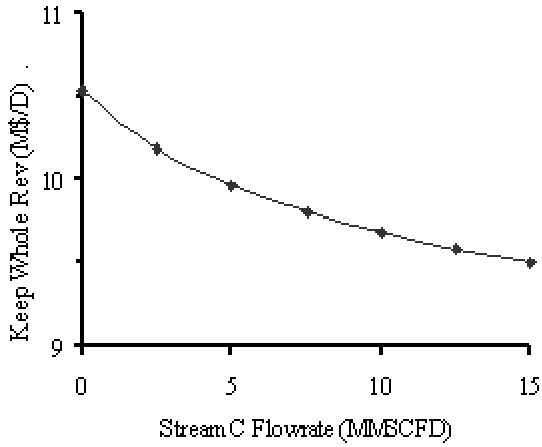

Figure 7. Revenue from Keep Whole Stream vs. New Stream Flowrate

Because the plant processes stream B on a fixed fee basis, the processing fee is constantly $900 per day. If the plant elects to process stream C, the revenue from processing stream B will not decrease. This is one of the advantages of a fixed fee contract.

The plant revenue attributable to the new stream is shown in Figure 8. This diagram illustrates that the revenue generated from stream C is not significantly dampened by the drop in recovery. The increase in revenue caused by stream C overpowers the loss in revenue from stream A due to the drop in ethane recovery. If the flowrate of stream C is 15 MMSCFD, then plant gross revenue increases about $6,000 per day. A substantial portion of the revenue from stream C, approximately $4,000 per day, comes from product fees. In fact, the plant derives more money from allowing stream C to use its pipeline and market its products than it does retaining 5% of the liquids.

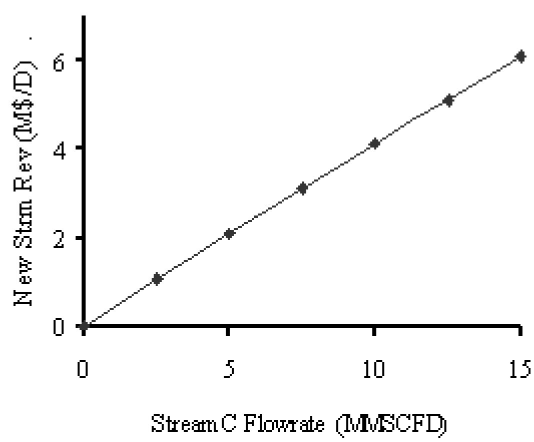

Figure 8. Revenue from New Stream vs. New Stream Flowrate

The results from the application of the new methodology show that the plant gross revenue increases significantly with the addition of the new stream. A decline in ethane recovery is expected with the addition of the new stream, but the economic benefits significantly outweigh the loss in ethane recovery.

This example also demonstrates the importance of investigating each processing alternative and its economic consequences. Based upon the information presented above, the addition of stream C should boost plant profits. One item has been overlooked in the solution to this example. What are the economic effects of adding the new stream on the owners of stream B? Figure 9 shows the economic impact.

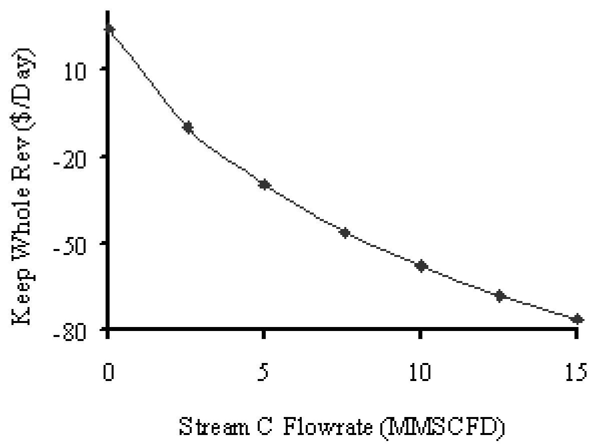

Figure 9. Gross Revenue for the Owner of Stream B vs. Stream C Flowrate

At current liquids prices the owners of stream B are making $22 per day. This is not a significant amount, but if liquids prices rise, they could make substantially more money. With the addition of stream C, the process achieves a lower ethane recovery and the revenue for the owners of stream B begins to fall. If the plant decides to process 15 MMSCFD of stream C gas, the owners of stream B will lose $78 per day. The loss in revenue to stream B from the addition of stream C is $100 per day. Under these circumstances, the owners of stream B would probably not renew their contract with the plant. They would also reduce the flowrate to the plant as much as they are able contractually.

The revenue from stream A decreases by $1,000 per day while the liquids revenue from stream C is only $1,700 per day. Thus, in looking at the desirability of stream C from a plant view point, the increase in liquids value is only about $700 per day. Should the owners of stream B terminate their contract with the plant, the cancellation would result in a loss of $900 per day. The real value in stream C comes from the ability to charge the pipeline and marketing fees.

As a result, the plant must decide which action to take to determine the business economic optimum for the plant. The application of this new methodology is a valuable tool to help find it.

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