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Why Cryogenic Processing (Investigating the Feasibility of a Cryogenic Turbo-Expander Plant)

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INTRODUCTION

As is well known, attractive product prices from the late 1960's to date have been providing the natural gas processing industry increasing incentives to recover greater percentages of available gas liquid products. Particular emphasis has been placed on recovery of ethane. As industry sought these higher recoveries, the turbo-expander plant evolved as the most attractive processing system.

The purpose of this paper is to present a method to quickly investigate the economic feasibility of a new plant or modification of an existing plant to achieve high product recoveries. Curves for this purpose were developed from engineering studies and are presented as valid (within their noted limitations) for a wide range of applications.

ECONOMIC INCENTIVES

As product prices have soared, cryogenic processing of natural gas for high liquids recovery has become an item of major interest. While it is anticipated that natural gas prices are certain to increase, it is also anticipated that an adequate spread between BTU values as a gas and BTU values as a liquid will be maintained to justify liquid recovery. Not only are existing plants being modified to recover a greater percentage of the available liquid products, but new plants are also being built to process gas volumes previously considered uneconomical. Today cryogenic plants are designed to recover as much as 80-90% of the ethane and to process inlet gas volumes as low as 5-10 MMscfd.

THE TURBO-EXPANDER PLANT

Several processes such as Turbo-Expansion, Joule-Thomson (J-T) Expansion, Cascade Refrigeration, and Multicomponent Refrigeration (MCR) comprise those commonly referred to as "cryogenic" in the gas processing industry. Because of operating and economic factors, the Turbo-Expander process has evolved as the most attractive. It is the most practical cryogenic process available to the industry for high ethane recovery, particularly at inlet gas pressures above 500 psi.

Figure 1 is a simplified flow diagram of a typical Turbo-Expander gas liquids extraction plant. Turbo-expansion utilizes the energy of a gas stream when reducing its pressure to produce recoverable energy and obtain cooling. Gas expanded across a valve (J-T) has a certain refrigerating effect, but by expanding the gas across a turbine wheel (turbo-expansion) work is removed from the stream providing a significantly lower final temperature for the same expansion ratio. This lower temperature both shows a thermodynamic efficiency increase and provides usable energy to replace or supplement mechanical drivers that are fuel consumers in a gas plant.

The expander plant has only one major utility cost — the residue gas recompressor or, in some cases, an inlet gas precompressor. An aerial cooler for cooling the compressed gas, a product pump, and some source of heat for regeneration of the dehydrator are also used, but these energy consumers are comparatively small. In the alternate design using propane refrigeration with the expander (dotted lines on Figure 1), propane compressors are fuel consumers as well, but this fuel consumption is an efficient trade-off with a reduction in the horsepower required for recompression and a subsequent reduction in fuel consumption.

Figure 8 is a picture of a scale model of a large capacity, high ethane recovery expander plant. The plant area shown occupies a plot size of 80 ft. x 120 ft., including a deethanizer with its associated equipment and product surge tank. Since the expander plant requires few pieces of equipment and little space, it is an ideal candidate for some form of packaging. This may be a true skid unit in the smaller sizes with graduation toward pre-fabrication of major portions or block mounted but precut and match-marked structures. Any of these forms of preassembly combined with inherent compactness provide portability and high salvage value. Portability, likewise, offers high versatility of location.

METHOD FOR RAPID INVESTIGATION OF PROJECT ECONOMICS

The natural gas processor often has need for a quick method of investigating the approximate economics of a potential project. Using six curves, Figures 2 through 7, he should be able to:

1. Determine the inlet gas richness (GPM C_2+) from the composition (Mol %) of the gas (Figure 2);
2. Estimate the maximum ethane recovery possible as limited by a residue gas heating value of either 1000 BTU/cf or 950 BTU/cf (Figure 3A or 3B);

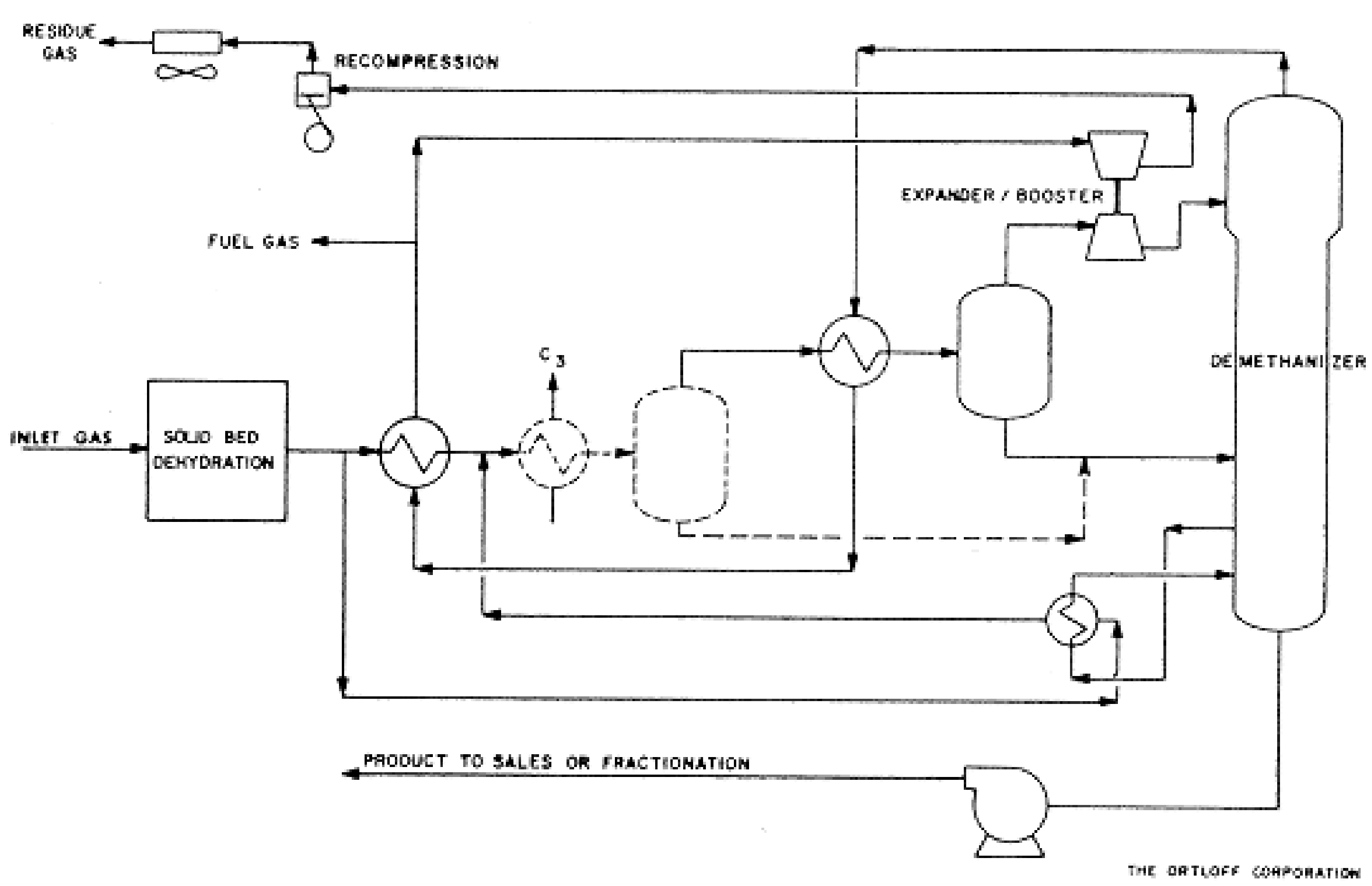

Figure 1
Simplified Process Flow Diagram
Turbo-Expander Gas Liquids Extraction Plant

3. Approximate the propane, butane, and heavier recoveries associated with the targeted ethane recovery level (Figure 4);
4. Calculate the net value of the products recovered and the resulting gross income before operating costs, depreciation, and taxes (Figure 5);
5. Determine the order-of-magnitude plant cost (Figure 6) and resulting payout before operating costs, depreciation, and taxes; and
6. Approximate the required compression horsepower (Figure 7) for calculating the fuel consumption and maintenance costs.

From the above, the feasibility of a project can be determined. A detailed study may then be started to determine the economics of the plant more accurately.

Conversion Curves For Liquefiable Fractions

Figure 2 provides curves for directly converting mol percent to gallons per Mscf (GPM) for each of the major liquefiable fractions (ethane, propane, butane, and pentanes plus). These curves are based on values published in the 1972 GPSA Data Book. The mixed butanes value used in plotting the curve is based on an assumed average composition of 25% isobutane and 75% n-butane. The value used for pentanes plus is assumed to be that of normal hexane.

Example: Butane Composition(total of isobutane & n-butane) = 5 Mol %

Butane Content per Figure 2 = 1.6 GPM

Note: For compositions with Mol % less than 1.0, divide the values on each axis by 10; e.g., for .5 Mol%, GPM = 0.16.

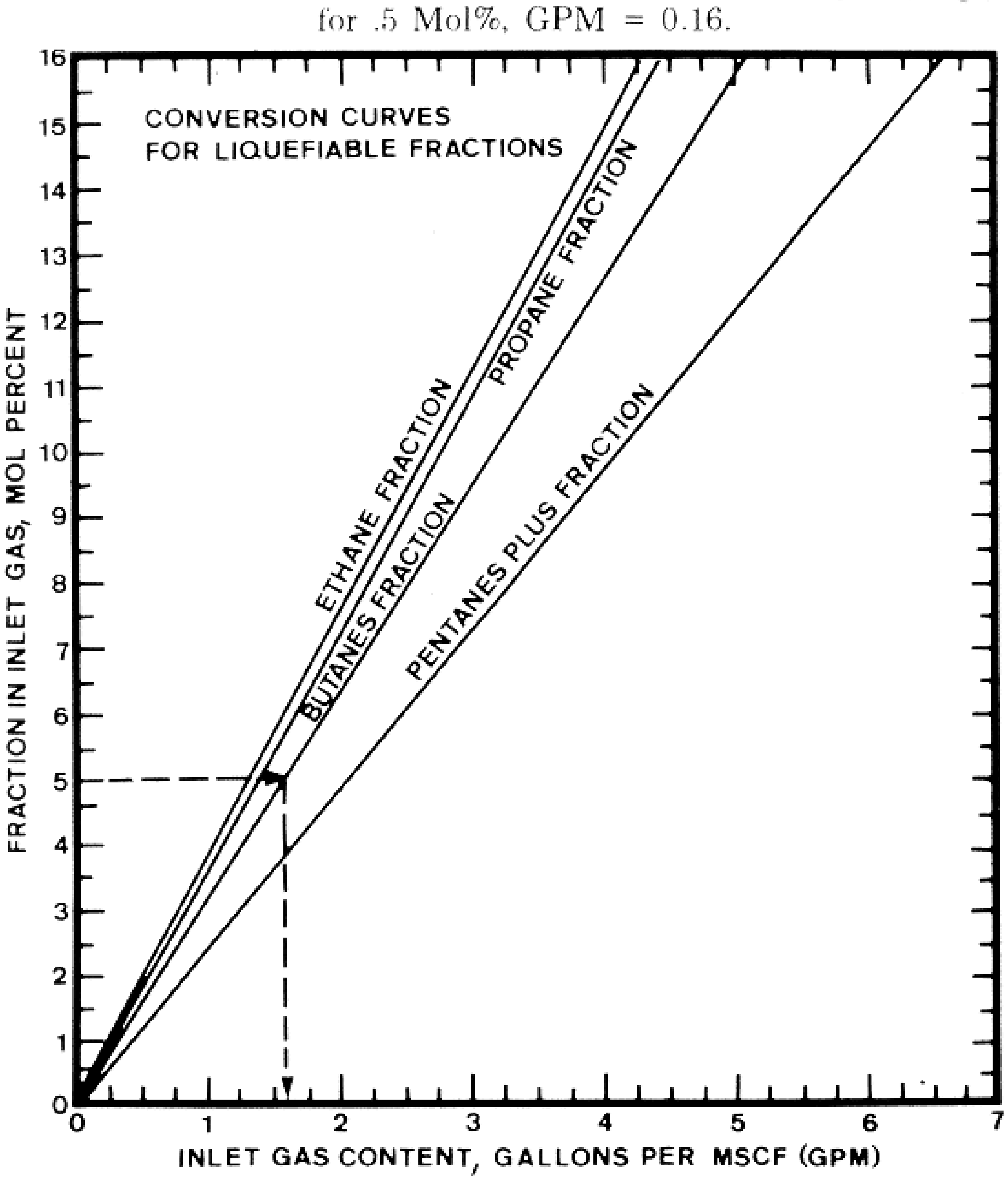

Figure 2
Conversion Curves for Liquefiable Fractions

Maximum Ethane Recovery Correlated With Inlet Gas Nonhydrocarbon Fractions

Limits to liquid hydrocarbon recovery are imposed by the need to maintain a certain heating value of the residue gas. Since heating values of 1000 BTU/cf and 950 BTU/cf (real gas, saturated at 30 in. Hg, 60°F) are common contract minimums, these values were selected as the bases for the plotted curves. Figures 3A and 3B plot such limits as a function of inlet gas richness (GPM C_2+) and nonhydrocarbon content (Mol %). These curves were developed from typical natural gas compositions. Estimated maximum percent ethane recovery should generally be accurate within ± 2% ethane. Larger deviations could occur if the relative concentration of each component varies from that used as a representative sample in developing these two curves. Table I presents the general range of relative component concentrations for three categories of richness which were used in developing Figures 3A and 3B.

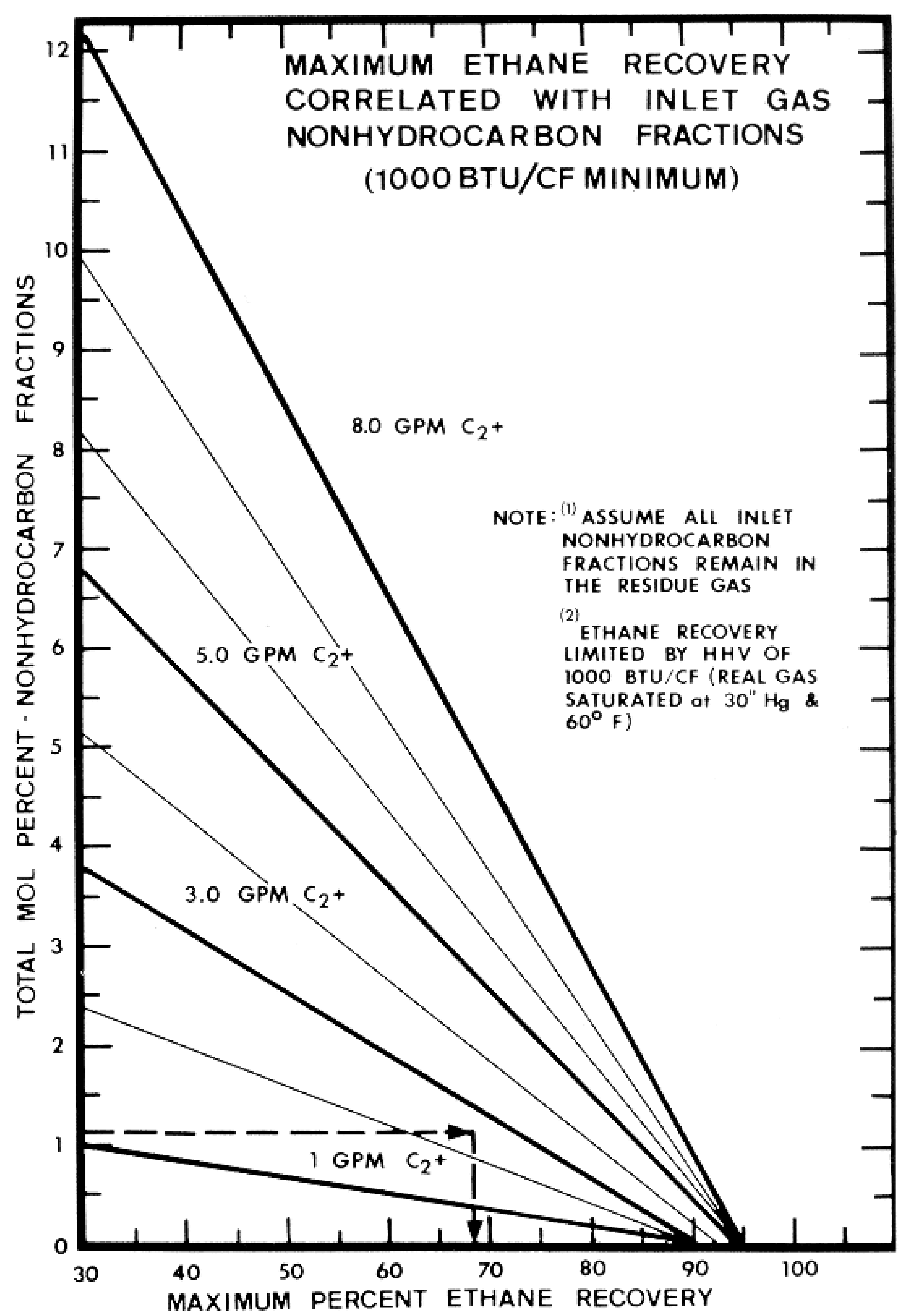

Figure 3A
Maximum Ethane Recovery Correlated with Inlet Gas Nonhydrocarbon Fractions
(1000 BTU/CF Minimum)

TABLE I

Concentration of Components

Component	Percent of Total Ethane Plus Fraction		
	Lean gas	Moderately Rich Gas	Very Rich Gas
	(C_2+GPM < 2.5)	(C_2+GPM=2.5-5.0)	(C_2+GPM > 5.0)
Ethane	45-60	50-65	35-50
Propane	15-20	20-30	30-35
Butanes	10-15	10-15	10-20
Pentanes Plus	10-15	5-10	5-10

Should the inlet gas (after any acid gas pretreatment) contain large quantities of carbon dioxide, a portion of this amount will be absorbed into the liquid product thereby reducing the dilution effect of carbon dioxide on the residue gas heating value. This will permit higher ethane recovery than initially indicated by the curves. Therefore, to adjust for carbon dioxide contained in the liquid product, the following equation should be used:

$$\text{Resultant \% Nonhydrocarbon} = \text{Inlet \% Nonhydrocarbon} - \frac{\text{Inlet \% Carbon Dioxide}}{3.5}$$

Since each process design might result in a different amount of carbon dioxide extraction, this equation is merely an approximation to improve the accuracy of the curves.

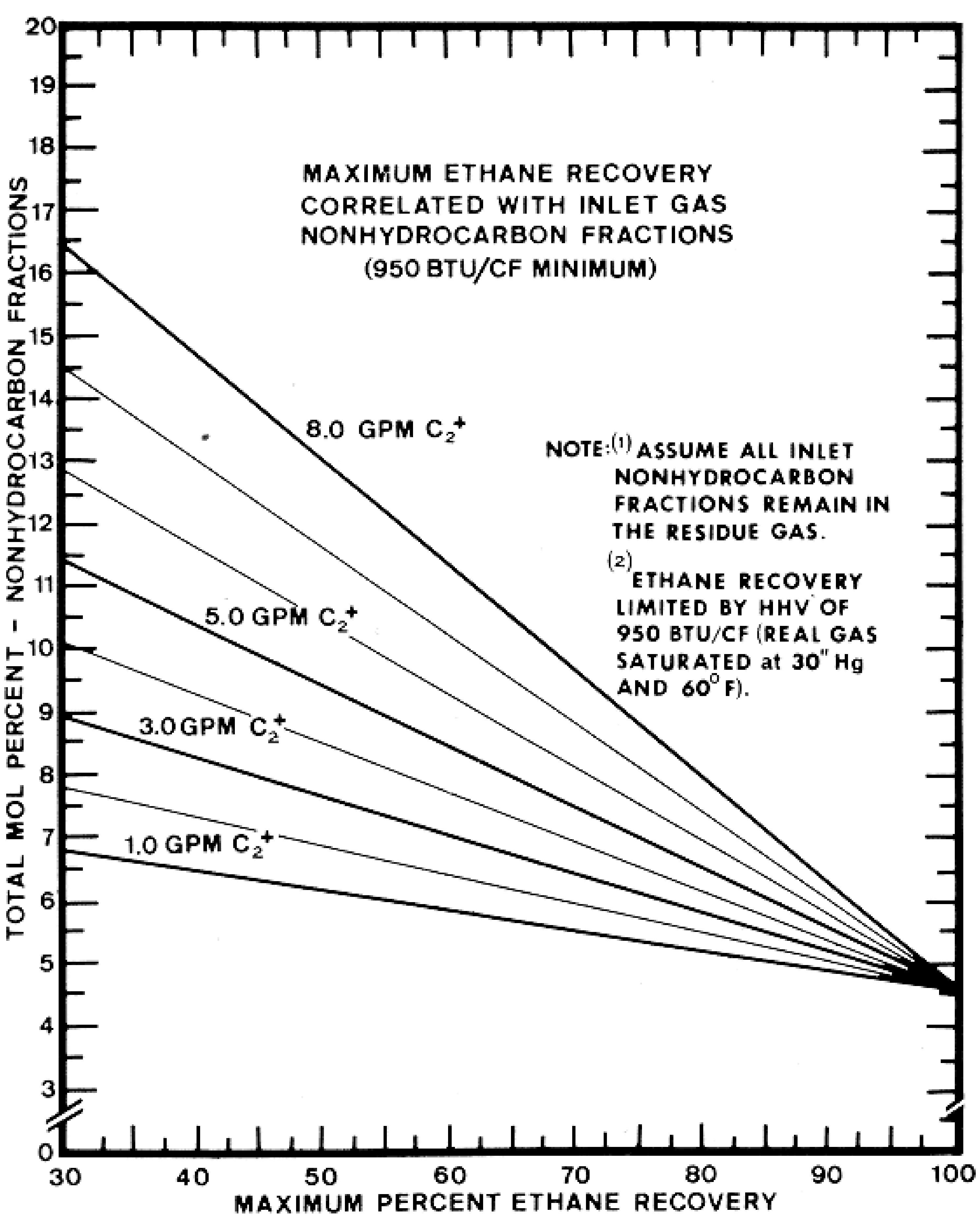

Figure 3B
Maximum Ethane Recovery Correlated with Inlet Gas Nonhydrocarbon Fractions (950 BTU/CF Minimum)

TABLE II

Inlet Gas Analysis

Component	MOL %	GPM
Nitrogen	0.93	—
Carbon Dioxide	0.16	—
Methane	89.99	—
Ethane	5.08	1.36
Propane	2.10	0.58
I-Butane	0.56	0.18
N-Butane	0.54	0.17
I-Pentane	0.19	0.07
N-Pentane	0.14	0.05
Hexane	0.15	0.06
Heptane Plus	0.16	0.08
	100.00	2.55

Inlet Gas Richness = 2.55 GPM C_2+
Nonhydrocarbon Fractions = 0.93 + 0.16 = 1.09%
Maximum Ethane Recovery Possible Due to Residue Gas HHV of 1000 BTU/cf: Rigorous Design Calc. 67.8%
Estimated (Figure 3A) 68.0%

Relative Recovery Curves

Although propane and butane recoveries are not independent of the specific process design, they are primarily related to the percentage of ethane extracted from the inlet gas. The curves shown in Figure 4 were developed from rigorous expander plant designs for various inlet gas compositions and process conditions. Recoveries shown should generally be accurate within ± 0.5% butane. For ethane recoveries above 80%, the indicated propane recoveries should be accurate within ± 0.75% propane. At lower ethane recoveries, propane recoveries should be accurate within ± 1.5%. Larger deviations may occur should gas compositions vary significantly from those used in design of the curve. (See tabulation in the Figure 3A and 3B discussion for representive compositions.)

TABLE III

Product Recovery

Component	Recoveries, % of Inlet	
	Rigorous	Estimated
Ethane	67.8	68.0
Propane	95.3	95.5
Butanes	99.4	99.3
Pentanes Plus	100.0	100.0

Based on inlet gas already defined.

Equivalent Liquid Value of Hydrocarbon Fuels

Curves plotted in Figure 5 provide the information necessary for determination of the added or incremental value of a given product. The heating value of 3310 BTU/cf for butanes is based on an assumed average composition of 25% isobutane and 75% n-butane. The heating value of 5086 BTU/cf for the pentane and heavier fraction is based on the equivalent value of hexane.

Conversion from the gross price of an extracted liquid to the added or incremental value requires one step. From the gross price of a gallon of liquid hydrocarbon, subtract the equivalent value of the extracted fraction as related to the price such a gallon would bring if sold in the vapor phase by millions of BTU's. This is the incremental or added value of the product, accounting for sales gas shrinkage.

Example: Gas Value - 80¢/MMBTU
From Figure 5, equivalent liquid value of propane is 7.3¢/gal.
If propane is selling for 14¢/gal., the added value is: 14 − 7.3 = 6.7¢/gal.

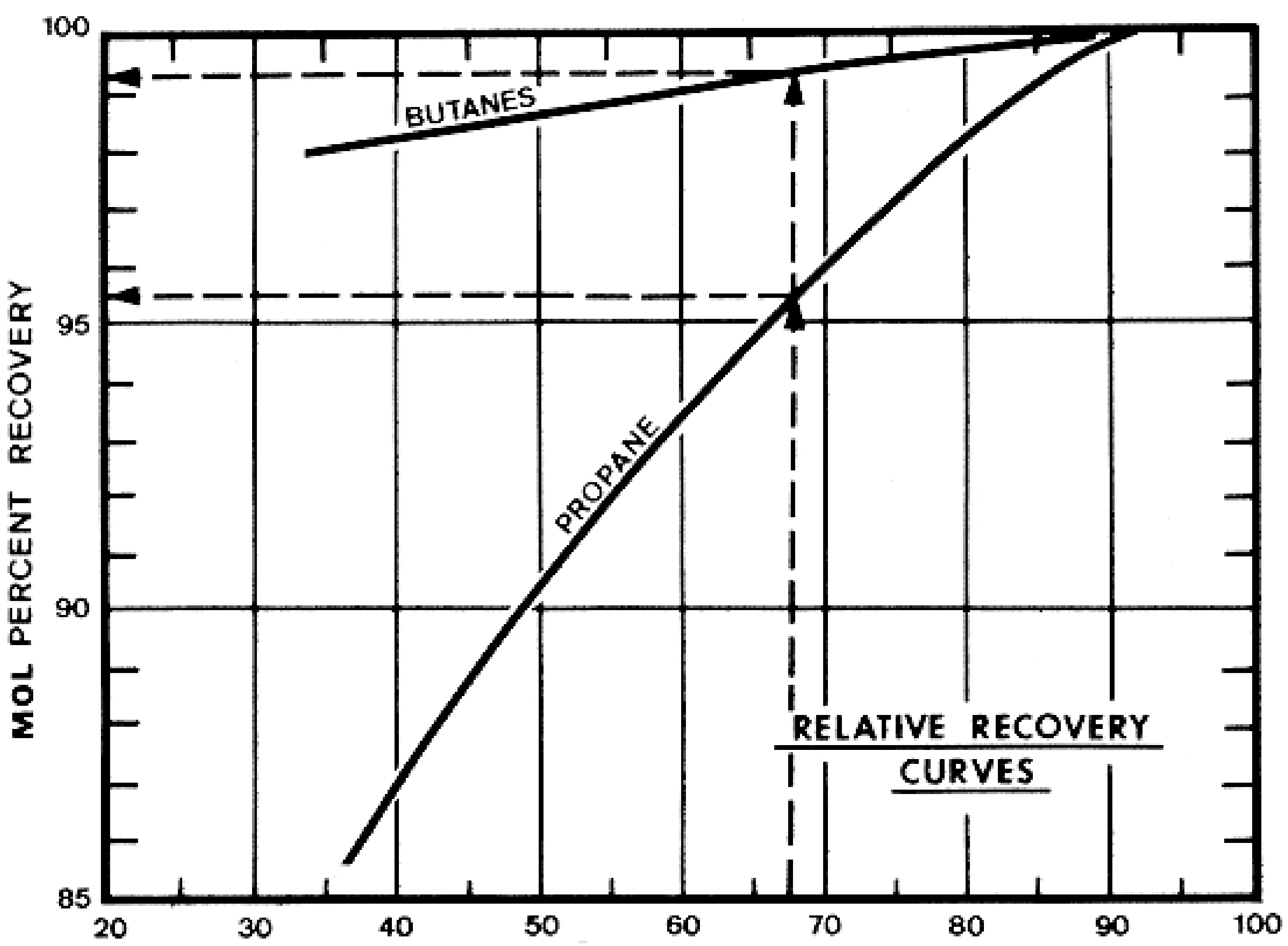

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Figure 4
Relative Recovery Curves

Estimated Installed Cost of Cryogenic Expander Plants

The curves in Figure 6 were developed from detailed expander plant design. While each plant was based on different processing requirements, each was designed for maximum ethane recovery limited only by a residue gas heating value of 950-1000 BTU/cf (saturated at 30 in. Hg, 60°F) or a maximum recovery level of 85%. Obviously, for ethane recoveries below the maximum, a particular plant will cost less than that indicated by Figure 6.

The battery limits cost of each plant used in the development of these curves was escalated from the original estimate date to the first quarter of 1975. Costs were adjusted to account for inclusion of nominal demethanized product surge capacity and compression for full pressure restoration. These costs include the use of either reciprocating or centrifugal compressors as determined by the most economical selection for the specific horsepower requirements. They were also adjusted to exclude all fractionation other than demethanization and all product treating or storage. Based on these data, it was possible to correlate the installed cost of a cryogenic expander plant with the inlet gas rate (MMscfd) and richness GPM C_2+). A check of the correlation against several actual plant designs indicates an order-of-magnitude accuracy of ± 20% can generally be expected.

Approximate Compressor Horsepower Requirements For Expander Plants

The largest single operating and maintenance cost item in a cryogenic expander plant is compression. Figure 7 is presented to assist the natural gas processor in estimating this cost. Using this curve, it is possible to approximate the total compression requirements, including all recompression, precompression, and propane refrigeration compression horsepower. Applying the specific fuel consumption and unit maintenance costs (supplied by the user), it is then possible to estimate the total horsepower related expenses.

This curve is presented to provide an order-of-magnitude cost only. The actual compression requirements for any plant will vary considerably depending on the specific design and product recovery requirements. In general, it is assumed that the plant is designed for relatively high ethane recovery as limited by a minimum residue gas heating value of 950-1000 BTU/cf but not exceeding approximately 85%. No allowance is made for additional compression that might be required for deethanization of the demethanized product.

Example: Inlet gas richness = 4.0 GPM C_2+
Inlet gas flow rate = 100 MMscfd
Adiabatic horsepower from Figure 7 = 49 Ahp/MMscfd
Assuming that centrifugal compressors will be used, overall efficiency = 72%.

Thus,
Total plant compression horsepower =

$$\frac{49 \text{ Ahp/MMscfd}}{0.72} \times 100 \text{ MMscfd} = 6806 \text{ bhp}$$

Compressor fuel consumption and maintenance costs can now be estimated.

SAMPLE ECONOMICS

The following case is presented to illustrate the economic incentives for high ethane recovery plants. It is based on an actual plant situation.

Case Statement

An owner-operator desires to determine the approximate economics of replacing an existing refrigeration type plant recovering propane and heavier components only with a turbo-expander plant capable of high ethane recovery.

Inlet gas flow rate to the plant is 20 MMscfd, and reserves are adequate to maintain this rate for at least five years.

Component recoveries of the present plant are: C_3 = 70%, C_4 = 90%, C_5^+ = 100%.

The first step in determining the economic feasibility is to calculate the increase in recovery for each component using the following equation:

Increase in Component Recovery (Gals/Day)	=	Incremental Component Recovery (%/100)	x	GPM Content for Component	x	Inlet Gas Volume (Mscfd)

Once the liquid recoveries are calculated, the added value of each fraction is determined and the annual gross income figured. Simple payout is calculated by dividing this income into the installed plant cost as shown in Figure 6.

Fuel Gas Value: 50¢/MMBTU

Liquid Product Values: (¢/Gal)

Product	Gross Market Value	Equivalent Liquid Value (Fig. 5)	Added Value
Ethane	11	3.2	7.8
Propane	16	4.6	11.4
Butanes	17	5.2	11.8

Gas Composition:

Component	Mol %	GPM (Fig. 2)
Methane & Non-hydrocarbons	74.5	—
Ethane	9.4	2.51
Propane	8.9	2.45
Butanes	4.8	1.52
Pentanes Plus	2.4	0.95
	100.0	7.43

Summary

Liquid Component	Ethane	Propane	Butanes	Pentanes Plus	Total
Percent Now Recovered	0.	70.	90.	100.	
Percent Proposed to Recover	82.	99.	100.	100.	
Percent Incremental Recovery	82.	29.	10.	0.	
Liquid Content, GPM	2.51	2.45	1.52	0.95	7.43
Increase in Recovery Gal/Day	41,160.	14,210.	3,040.	0.	58,410
Added Value, $M/Yr.	1,172.	591.	131.	0.	1,894

Annual Income Before Operating Costs, Depreciation, and Taxes	*$1,894,000*
Estimated Plant Investment (Figure 6)	*3,000,000*
Payout in Years	*1.58*

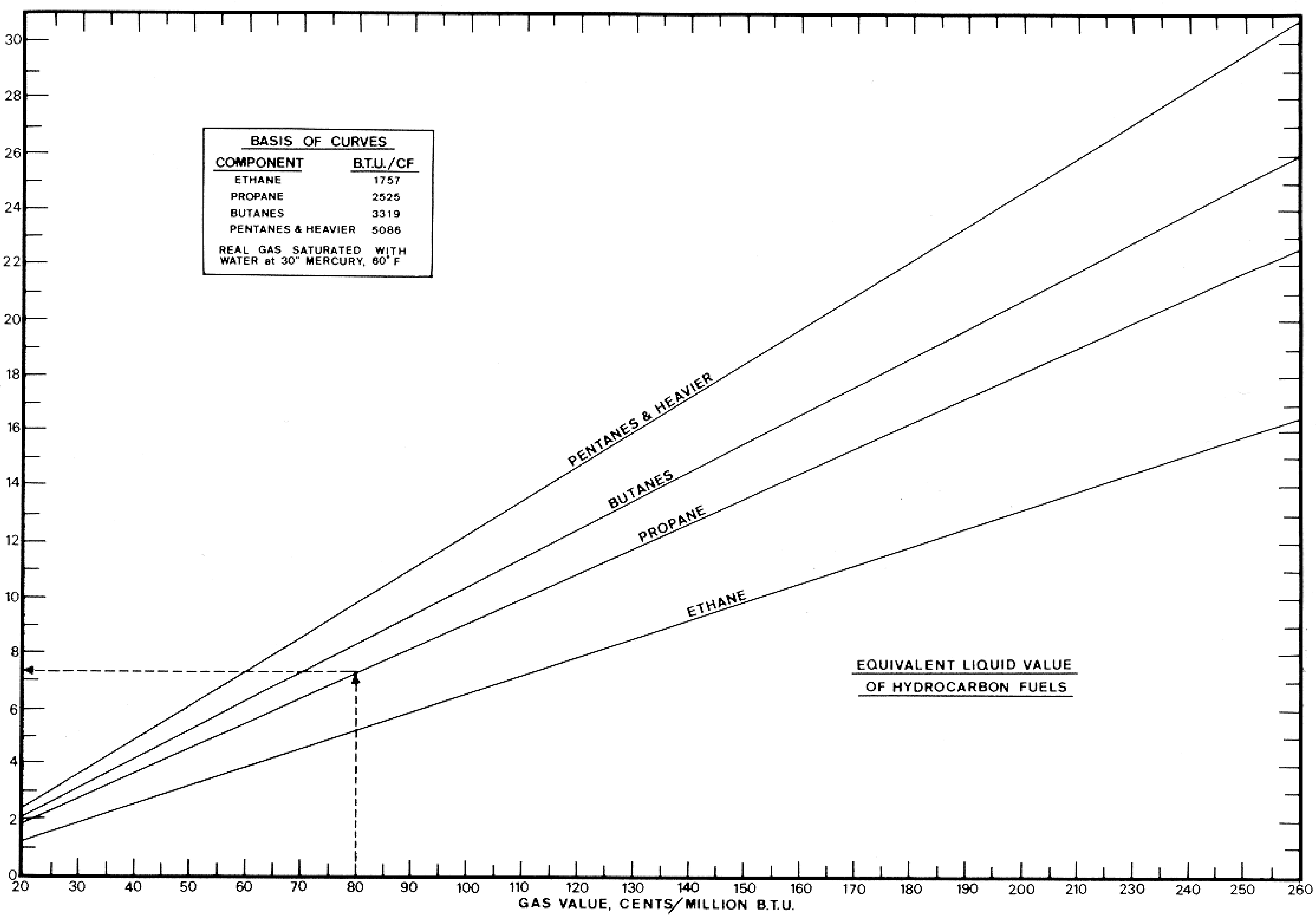

Figure 5
Equivalent Liquid Value of Hydrocarbon Fuels

If desired, the above calculation could be further refined for the effect of incremental compressor fuel consumption. Propane refrigeration in the replaced plant used 700 bhp. Approximate compressor horsepower needed in the new cryogenic plant may be determined from Figure 7. Using the horsepowers so defined, incremental compressor fuel cost can be estimated.

Given: Current Compressor bhp = 700

Estimated horsepower in the new facility (assuming use of reciprocating compressors)

$$= \frac{78 \text{ Ahp/MMscfd}}{0.81} \times 20 \text{ MMscfd} = 1926 \text{ bhp}$$

Incremental bhp = 1926 — 700 = 1226

Incremental compressor fuel cost (for 8000 BTU/bhp-Hr consumption)

$$= 1226 \text{ bhp} \times \frac{8000 \text{ BTU}}{\text{bhp-Hr}} \times \frac{24 \text{ Hr}}{\text{Day}}$$

$$\times \frac{\$.50}{1{,}000{,}000 \text{ BTU}} = \$118/\text{day} \quad \text{or} \quad \$42{,}960/\text{Yr}.$$

SUMMARY

In conclusion, the data presented should provide an effective tool for those wishing to make a quick analysis of expander plant investment and economics.

Our approach to economic appraisal is carried only through the steps of estimated capital investment and calculation of gross incremental product income. This is adequate to arrive at a first-glance payout before taxes and plant expenses leading to the decision for a further and more detailed analysis of the project.

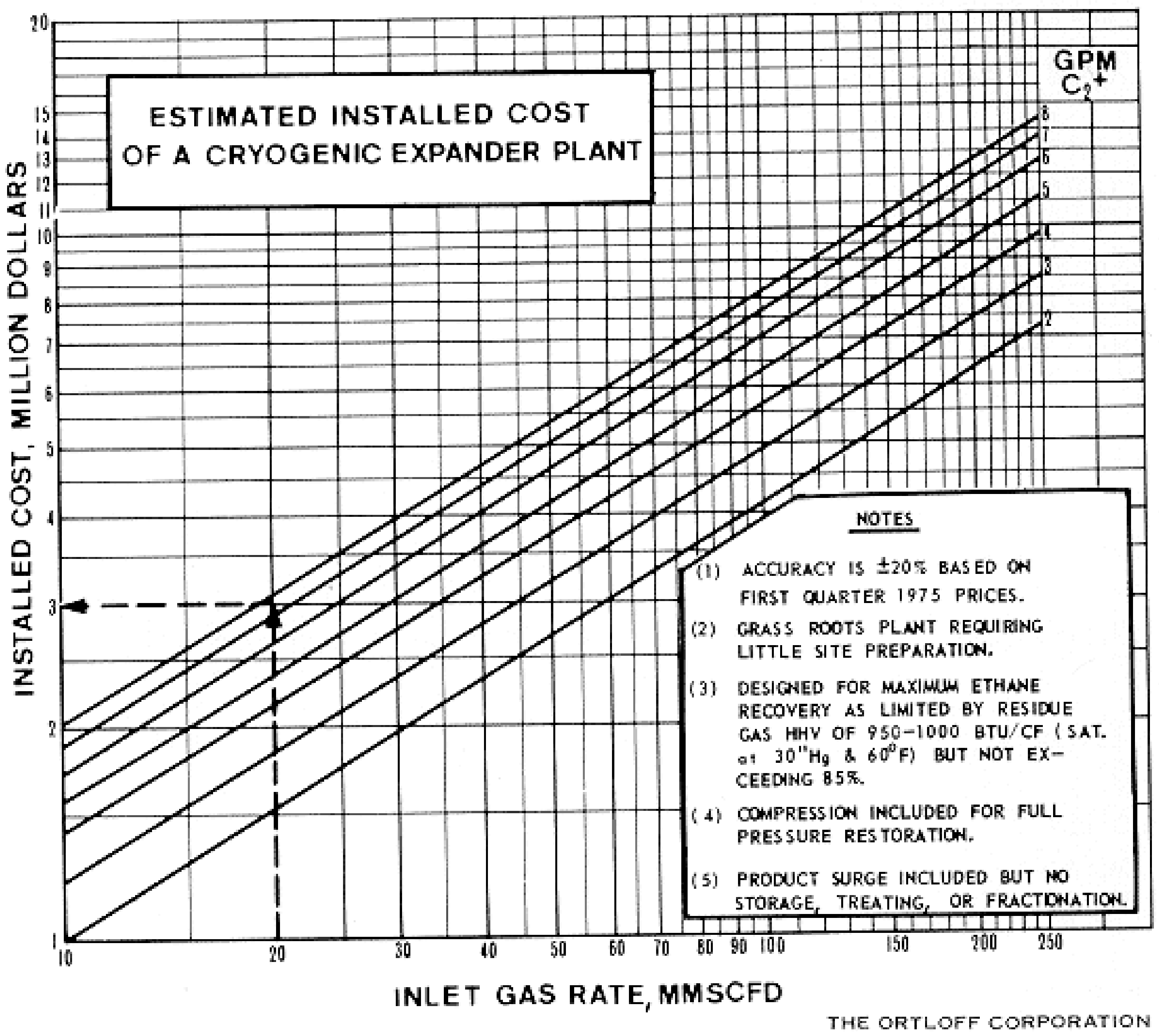

Figure 6
Estimated Installed Cost of a Cryogenic Expander Plant

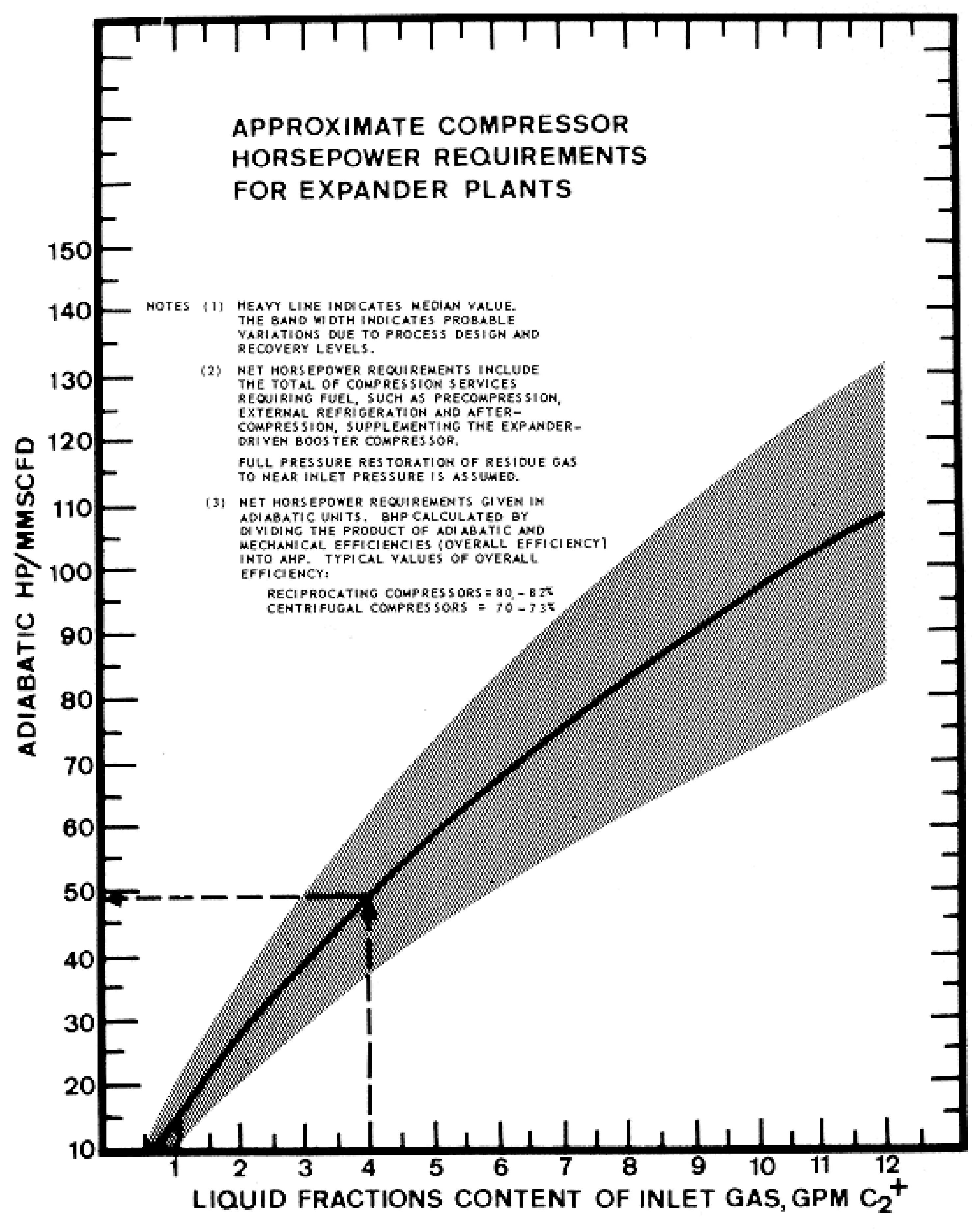

Figure 7
Approximate Compressor Horsepower Requirements for Expander Plants

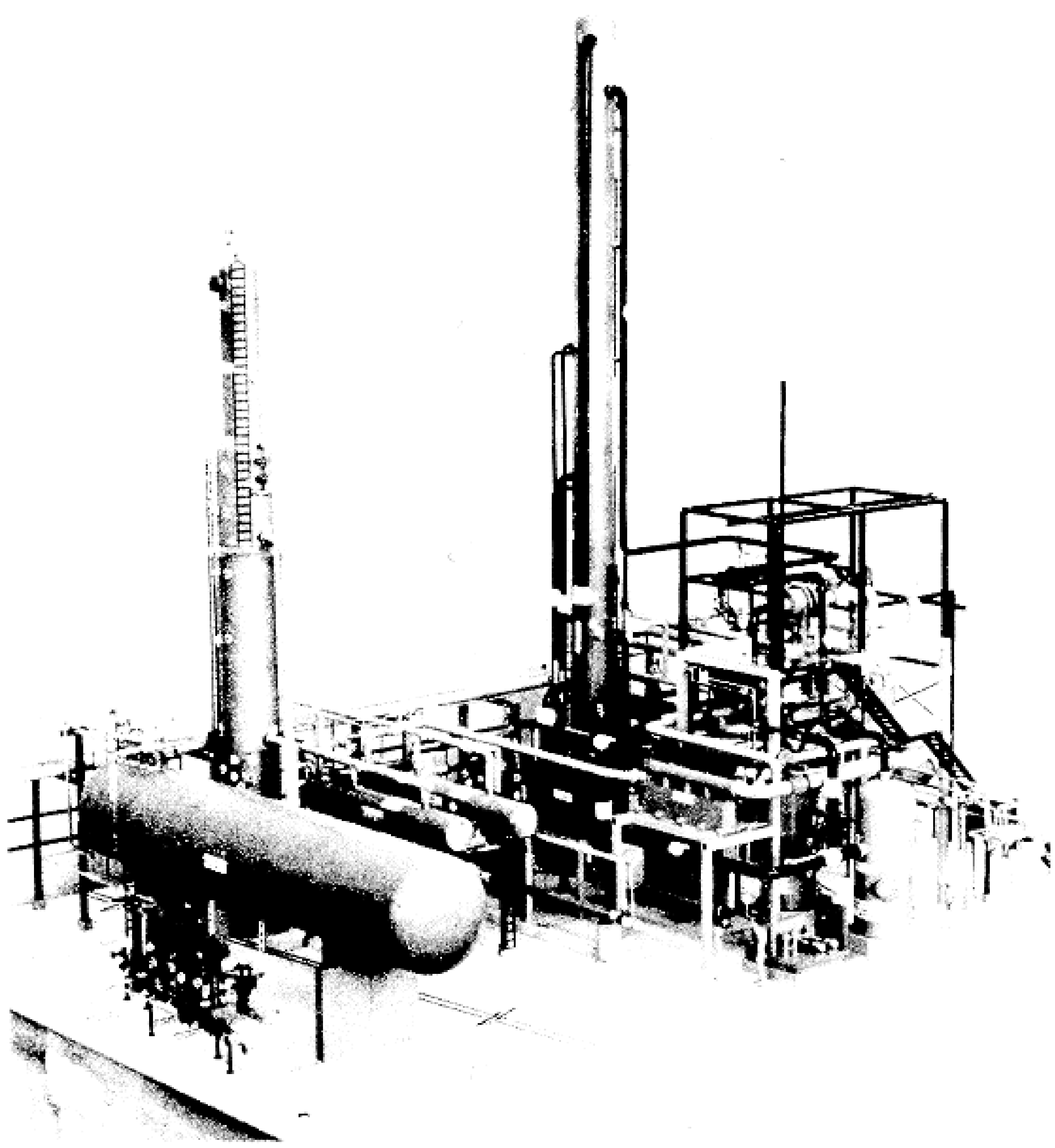

Figure 8
Scale Model of Large Capacity High Ethane Recovery Expander Plant