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# ALTERNATIVE CONCEPTS FOR SUPPLYING CARBON DIOXIDE FOR ENHANCED OIL RECOVERY PROJECTS

FINAL REPORT

Date Published—July 1980

Work Performed for the Department of Energy
Under Contract No. EW-78-C-21-8333

Science Applications, Inc.
Golden, Colorado

Bartlesville Project Office
U. S. DEPARTMENT OF ENERGY
Bartlesville, Oklahoma

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# ALTERNATIVE CONCEPTS FOR SUPPLYING CARBON DIOXIDE FOR ENHANCED OIL RECOVERY PROJECTS

**Final Report**

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U.S. DEPARTMENT OF ENERGY

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# PREFACE

A study of alternative concepts for supplying carbon dioxide for enhanced oil recovery projects was undertaken by Science Applications, Inc. The work was funded under the System Support Services Contract with the Morgantown Energy Technology Center of the Department of Energy (Contract No. EW-78-C-21-8333, Task - 4). This report describes the conduct and the findings of this study. The purposes of the study were to:

- Define the $CO_2$ requirements for candidate EOR reservoirs.
- Generate alternative concepts for collecting, transporting, storing and processing $CO_2$ from naturally occurring sources.
- Compare the alternatives on the basis of cost effectiveness.
- Reestimate the magnitude of the $CO_2$ EOR reserve base on the results of the above findings.
- Recommend future DOE program activities.

The results of the study reported herein fulfills the above stated objectives. The results, opinions, and recommendations expressed in this report are those of Science Applications, Inc. and not necessarily those of the U.S. Department of Energy.

FIGURES



TABLES

# I. INTRODUCTION

The United States has reached a critical point in its energy consumption patterns. The reliance on liquid petroleum for nearly fifty percent of its energy supply and the enormous growth in energy consumption over the past two decades have outstripped the ability to produce from domestic sources. The shortfalls in domestic petroleum production have been overcome by the importation of crude oil from foreign countries, to the point where the U.S. currently imports more crude and refined products than it produces. The 1974 oil embargo by the Arab states and the 1979 supply interruptions caused by the political unrest in Iran have left their marks on the national economic and social consciousness. The high rate of crude importation has caused an unfavorable balance of payments and an unprecedented inflation rate in the U.S., and has placed the national security in serious jeopardy.

There has been a concerted effort to arrest the decline of national production. Following the 1974 Arab oil embargo, the oil industry embarked on an intensive drilling program in the lower 48 states. The active rotary rig count has averaged more than 2,000 units over that period of time, and has recently reached a 30 year record of over 2,500 units. In other words, since 1974, approximately 70% of the <u>world's</u> drilling resources have been devoted to improving the petroleum reserve and production picture of the continental United States. Unfortunately, an improved picture has not evolved. In 1978 the nation's drawdown of proven reserves was seven times as great as the establishment of new reserves. Texas and other critical producing states experienced record declines in production rates.

In summary, the nation is faced with three elements of a perplexing and critical energy problem:

- Our energy utilization infrastructure is heavily dependent on products derived from liquid petroleum. Significant alterations of this infrastructure (such as massive shifts to coal utilization or nuclear energy) would require huge investments in capital and elapsed time.

- Our declining petroleum production and domestic reserve situation are falling short of meeting the demand for petroleum products by a factor of two.

- The shortfall between domestic demand and supply is being accomodated by the importation of petroleum produced in foreign countries, resulting in intolerable consequences to the economic and strategic well being of the United States.

Coupled with these facts is the realization that less than one-third of the original oil-in-place is produced by presently used primary and secondary recovery techniques. It is evident that improvements in the percentage recovery of the oil-in-place could significantly improve the U.S. energy supply picture.

The production techniques that are applied after completion of secondary recovery operations or to heavy oil reservoirs is referred to as enhanced oil recovery (EOR) methods. The methods are normally characterized as thermal, chemical, or gas miscible recovery processes.

The specific resource target for EOR is 250 to 300 billion barrels of light and heavy crude left in reservoirs after the utilization of conventional production methods. This is one of the few hydrocarbon resources that can be utilized without causing massive perturbations to the energy utilization infrastructure. With successful technology development, as much as 50 billion of the 300 billion barrels could be recovered. (REF.1) As much as 21 billion of the 50 billion barrels could be recovered by the gas miscible process using carbon dioxide as the injected gas.

Despite EOR's potential, its development and implementation have been slow to materialize. Only 400,000 bbl/day of the nations 8,400,000 bbl/day production can be attributed to the implementation of EOR processes. The vast majority of this production results from thermal recovery methods applied to heavy oil reservoirs which have experienced little to no primary production. The reasons for the slow progress of EOR implementation are multifaceted. In some cases, the technology is not sufficiently mature, in other cases the economics are unattractive at the present risk level. The implementation is also hindered by institutional constraints.

For example, in highly integrated oil companies, EOR projects must compete for investment funds with projects that vary from the acquisition of coal and uranium properties to the acquisition of electric motor manufacturing operations. Until the preceived economic risk of EOR relative to other investment options is substantially reduced, EOR's potential for reducing the nation's energy dependence on foreign sources will never be realized.

The principal risks that are associated with the $CO_2$ miscible process are:

- Incomplete understanding of the behavior of $CO_2$ after it is injected into representative reservoir fluids.
- Inability to contact all of the reservoir due to viscous fingering, poor mobility control, and gravity segregation.
- Insufficient quantities of low-cost carbon dioxide near the candidate reservoirs.

It is the last stated risk that was addressed in the study reported herein.

## $CO_2$ Requirements - Identified Sources

Several attempts have been made since the 1974 Arab oil embargo to estimate the potential for production of petroleum in the United States by utilizing enhanced oil recovery techniques. (REF. 1-3) The most recent estimate by the Office of Technology Assessment of the U.S. Congress (REF. 1) is that under the most favorable sets of assumptions up to 51 billion barrels of additional oil could be produced using EOR techniques. The most favorable assumptions were that:

- EOR technology would improve considerably over current levels.
- The prices would be virtually unconstrained.
- The injected fluids would be available in the quantities needed at "reasonable" prices.

The study projected that up to 21.1 billion of the 51 billion barrels of oil produced would be recovered by the $CO_2$ miscible EOR technique. The specific technology assumption that accompanied that projection was that the oil recovery efficiency of $CO_2$ would be improved from the current value (approximately 10 to 7 MCF/barrel of oil recovered) to 5.1 MCF/barrel. To achieve the oil production estimate of 21.1 billion barrels a total of 107.6 trillion SCF would be required. Since approximately 25% of the $CO_2$ requirements could be met by reinjecting some of the $CO_2$ that is co-produced with the tertiary oil (also assumed in REF. 1) a net of 80.7 TSCF of $CO_2$ would be required for the initial injection into tertiary oil reservoirs.

The cost and availability of $CO_2$ from various sources within the United States have been estimated in the previously conducted studies. (REF. 4-6) On the basis of the reserves stated in these references or assuming a twenty year production of the rates reported in these references the following conclusions were drawn by the writers:

SUMMARY OF $CO_2$ RESERVES

| | |
|---|---|
| Natural sources containing high concentrations of $CO_2$ ($>80\%$) | 30 TSCF |
| Process vents containing high concentrations of $CO_2$ | 5 TSCF |
| Flue gas stacks containing low concentrations of $CO_2$ ($5 < CO_2 < 17\%$) | 182 TSCF |

On the surface of the figures shown in the above table one can conclude there are sufficient quantities of $CO_2$ available to meet the needs of future EOR projects. Unfortunately, the most voluminous source is, for the most cases, very costly. The costliness stems from the very high cost associated with separating the low concentrations of $CO_2$ from the products of combustion and the high cost of transporting the $CO_2$ from large fossil fuel burning installations to candidate petroleum reservoirs. The second cost contributor results from the fact that most large flue gas sources are electrical generating facilities and that these facilities are located near densely populated regions of the U.S. and not near oil producing regions.

Although estimates of the delivered cost vary widely the OTA study indicates that the cost of recovering $CO_2$ from flue gas is more than twice that required to recover $CO_2$ from natural reservoirs. It is reasonable to conclude that the availability of only 35 TSCF from less costly sources could conceivably reduce the amount of oil recovered by the $CO_2$ miscible process by over a factor of two. It was this observation that prompted the initiation of the study reported herein. The question asked was "Is there another source of $CO_2$ available that is more attractive than removing $CO_2$ from flue gases?". The answer to the question was an assessment of the quantity of the $CO_2$ that could be recovered from the production and processing of natural gas.

Looking at natural gas production as a potential source of $CO_2$ was appealing for the following reasons:

- Even at low $CO_2$ concentrations the sheer magnitude of natural gas production could result in a large source of $CO_2$. (Example: If the known reserves of natural gas (approximately 200 TSCF) contained 5% $CO_2$ the results would be 10 TSCF of $CO_2$.)
- The cost of separating the $CO_2$ from the natural gas could be partially borne by the revenue generated from the sale of natural gas.
- Since natural gas and petroleum are normally produced in the same general regions the transportation distance between $CO_2$ sources and tertiary oil reservoirs would in most cases be less than the distance from the reservoirs to large electrical generating plants.
- The advent of significantly higher natural gas prices and the depletion of older reservoirs of "sweet" natural gas have made production from reservoirs containing higher concentrations of $CO_2$ and $H_2S$ more attractive than in the past (witness recent development in the Western Overthrust Belt and Louisiana's Tuscaloosa Trend). Estimates that over 200 TSCF of natural gas could be produced from the Western Overthrust Belt which extends from

the Canadian to the Mexican border. The $CO_2$ content has been relatively high but there is insufficient production data to confirm the average $CO_2$ content. The probable reserve forecast for the Tuscaloosa Trend is over 80 TSCF. Several wells have produced in excess of 10% $CO_2$ at very high total flow rates. Additional drilling and production will be required to firm up both the extent of the natural gas reserves and the content of the $CO_2$ in the gas. Any $CO_2$ that would be separated from either of the plays could find application to reservoirs in relatively close proximity. They also share the advantage of new, high capacity reservoir systems whose productive lives coincide with the need for $CO_2$ for EOR projects.

Status of $CO_2$ Supply Projects

The $CO_2$ supply projects discussed in the following pages are those that are associated with large scale EOR projects. There have been numerous $CO_2$ pilots and mini tests which have been serviced with temporary $CO_2$ supply systems. There are nine $CO_2$ supply projects that have been announced in open literature. The status of these projects range from fully operational to the preliminary planning stages (Table 1). It is useful to analyze the projects to test the validity of the assumptions that have been made in previous supply studies. The first pipeline, SACROC, is the only one that is in operation. It has been in service since the early 1970's transporting a 90% $CO_2$, 10% $CH_4$ mixture from the Val Verde basin in Southerwestern Texas to the Kelly-Snyder Field (Canyon Reef Reservoir) in Scurry County, Texas and the North Crossed Field in Upton County, Texas. Virtually all of the operational experience that exists has been gained on the SACROC project. (Reference 7)

The next four projects are transporting $CO_2$ from large natural reservoirs in Southern Colorado, Northeastern New Mexico and North Central Mississippi to large reservoirs in West Texas and Central Mississippi. It is interesting to note that these $CO_2$ producing provinces were the ones identified in Reference 5 as having the most attractive near-term utility to large $CO_2$ EOR projects. The producing provinces included:

- McElmo Dome - Doe Canyon, Colorado (Shell)
- Sheep Mountain - Dike Mountain, Colorado (ARCO)
- Brovo Dome Unit - New Mexico (Amoco)
- Jackson Dome - Mississippi (Shell)

The next two projects are pipelines that connect $CO_2$ from process vents (anhydrous ammonia plants) to smaller reservoirs. The chracterisitcs of these lines are lower in flow rates (most anhydrous ammonia plants yield less than 50 MMSCFD) and shorter transport distances than those of the four western lines.

The eighth and ninth supply projects listed derive their $CO_2$ from the co-production of natural gas and $CO_2$ where the $CO_2$ concentration ranges from 15 to 66% by volume.

The trend that is developing with the announcement of these projects is probably indicative of the near future and includes the following characteristics:

- The coupling of the least expensive naturally occurring sources having high deliverability to the large (in some cases giant) reservoirs such as Canyon Reef, Wasson, North Cowden, etc.

- The utilization of high concentration process vent sources of $CO_2$ to flood reservoirs of more modest sizes that are located relatively close to the source of the $CO_2$.

- The exclusive use of pipelines as the transport mode.

In all cases for which detailed information has been published, the pipelines have been designed to operate in the super-critical dense-phase-fluid thermodynamic state (i.e. the pressure is greater than the critical pressure, and the temperature is at ground ambient which is usually less than the critical temperature (88$^{o}$F).

## Status of Processing Natural Gas/$CO_2$ Streams

There are over fifty well known gas purification processes to date and new ones are being continuously developed. The numerous choices available make the selection of a process applicable to specific situations more difficult. There are many factors that need to be considered when selecting a process for a given sweetening application. Some of these factors are:

1. The types of impurities to be removed from the gas stream
2. The concentration level of these impurities and the degree of removal desired
3. The acid gas selectivity required, if any
4. The volume of gas to be processed and the temperature-pressure conditions at which gas is available
5. The feasibility and desirability of sulfur recovery
6. Relative economic factors of the suitable pressures
7. Other factors such as, track record, ease of operation, depending on location, etc.

Each of the above factors, except #7, is further described in Ref. 15 . Depending on the site specific criteria, one or two of the above factors will be critical and affect the selection of a process much more than the others. The factors mentioned in #7 are involved in the final decision making step. If these factors could be digitized and some dollar value put on it, it can be combined into the "Relative Economic Factors" (#6).

The commercially available gas purification processes can be divided into the following general categories:

A - Processes which involve a reversible chemical reaction between the impurities and the solution used to extract it.

B - Processes using solvents which physically absorb the impurities.

C - Solid bed and adsorption processes, where the impurities are either selectively adsorbed on the surface area of the adsorbent, or engage in a chemical reaction by coming in contact with the reagent in the solid bed.

D - Processes involving separation based on boiling point differences, etc.

Some of the processes which fall into the "A" category are:

| | |
|---|---|
| Amine Processes: | MEA, DEA, TEA, MDEA, DIPA |
| New Amine Type Processes: | Diglycolamine, Alkanolamine, SNPA-DEA, Alkazid, ADIP, Sulfiban |
| Ammonia Processes: | Collin, Diamox, and Aqueous Ammonia |
| Carbonate Processes: | Seaboard, Vacuum Carbonate, Tripotassium Phosphate, Sodium Phenolate, Alkacid, Hot Carbonate, Catcarb, Benfield (Modified and Hipure), Giammarco-Vetrocoke |

Typical processes which fall into the "B" category are:

Water Absorption
Fluor Solvent, Selexol, Rectisol, Purisol, Estrasolvan

A number of other physical solvents whose list and physical properties are readily available have been investigated by Woertz (15 ).

There are processes which can be considered as a hybrid between the categories "A" and "B". The most commonly known process which utilizes a solvent which is a combination of chemical reactant as well as a physical absorbent is sulfinol. Other processes, such as AMISOL can also be named.

There are processes such as: Burkheiser, Ferrox, Gluud, Manchester, Thylox, Fischer, Cataban, Konox, Perox, Lo-Cat, and others which are selective for removal of hydrogen sulfide only.

Processes which fall into the "C" category are:

Iron Oxide (sponge), Activated-Carbon, Cold Bed Adsorption (CBA), EFCO processes, and Appleby-Frodingham

In category "D" type processes, refrigeration at predetermined pressures will take place in order to separate the hydrocarbons from their impurities. Depending on the available upstream pressure, the use of turboexpansion machinery to provide the required refrigeration can be considered.

In general, the natural gases containing $CO_2$ can be categorized on the basis of their $CO_2$ content. The natural gas is divided into three categories: 1) Natural gases containing less than 5 mole percent $CO_2$; 2) Natural gases containing 5-30 mole percent $CO_2$; and 3) Natural gases containing greater than 30 mole percent $CO_2$. In the section entitled "Analysis of Process Alternatives," the basis for the above categories is described in detail.

A comparison of the most common commercial gas purification processes, in as far as their limitations on the feed and product stream conditions are concerned, is given in Table 2 . Depending on specific feed stream conditions and the desired quality of the product stream, a process can be selected using Table 2, especially if the $CO_2$ concentration is between 5 and 30 mole percent.

## II. APPROACH

The objective of this study was to explore alternative means of supplying $CO_2$ to enhanced oil recovery projects. The study emphasis was placed on:

- Analyzing processing alternatives for separating $CO_2$ from natural gas streams of varying compositions.
- Analyzing alternatives to pipeline only transport modes.
- Projecting the impact that any of the evolving concepts might have on the availability of low-cost $CO_2$ and hence the quantity of oil potentially recoverable by $CO_2$ EOR methods.

From the outset it appeared both possible and prudent to separate the "processing" and "transportation" portions of the study from each other and thus allow those areas to be worked in parallel. The transportation analysis address the transport of nearly pure (98% $CO_2$) carbon dioxide irrespective of its source.

### Approach to Analysis of Processing Alternatives

In order to analyze processing alternatives for natural gases containing $CO_2$, the natural gases were divided into the following three general categories:

1 - Natural gas streams containing less than 5 mole % $CO_2$
2 - Natural gas streams containing greater than 5 but less than 30 mole % $CO_2$
3 - Natural gas streams containing greater than 30 mole % $CO_2$

The reasoning behind the above division is based on the heating value and the Wobbe Index.

The Wobbe Index is used as the most convenient index of commercial gas quality, combining measurements of calorific value and specific gravity, and indicating the range of gas qualities that can be satisfactorily handled by a predetermined orifice size.

$$\text{Wobbe Number} = \frac{\text{Calorific Value}}{\sqrt{\text{Specific Gravity}}}$$

Calorific Values, Specific Gravities and Wobbe Nos. for Typical Fuel Gases are (22):

| Type of Gas | Average Calorific Value Btu/Cuft | Specific Gravity | Wobbe Number |
|---|---|---|---|
| 1. Coal gas | | | |
| - horizontal retorts | 560 | 0.472 | 878 |
| - vertical retorts w/steaming | 475 | 0.4744 | 690 |
| - low temp. process | 889 | 0.5351 | 1215 |
| 2. Coke oven gas | 525 | 0.3922 | 838 |
| 3. Water gas | | | |
| - "blue" | 295 | 0.5500 | 398 |
| - "carburetted" | 500 | 0.6457 | 622 |
| 4. Producer gas | | | |
| - from coke | 132 | 0.8940 | 140 |
| - from coal | 163 | 0.8669 | 175 |
| 5. Blast furnace gas | 92 | 1.0100 | 92 |
| 6. "Oil gas" I.C.I. "500" process | 500 | 0.4714 | 728 |
| C.R.G. Process | 800 - 975 | 0.5614 | 1170 |
| 7. Natural gas | | | |
| Average "Commercial" gas | 1,000 - 1,100 | 0.65 | 1302 |
| Methane | 1,100 | 0.5517 | 1481 |
| Ethane | 1,800 | 1.0345 | 1770 |
| Propane | 2,300 | 1.5172 | 1867 |
| Butane | 2,940 | 2.0000 | 2079 |

It is important to define Natural Gas on a commercial basis. Some twelve gas samples amongst 200 or more samples were chosen from different states and counties. The heating value for these ranged from 198 to 2323 Btu/scf. The methane, $CO_2$, heavier hydrocarbons, Nitrogen content of these gases are given in Table 3. The specific gravity and the Wobbe Number were calculated for these gases. For a typical unprocessed natural gas the Wobbe Number will range from 800 to 1500.

Table K shows how, on a commercial basis, natural gases are sometimes classified in the U.S. into "high inert," "high methane" and high "Btu" groups. The specific gravity, heating value, and the Wobbe for the commercial gases are given in this table.

TABLE K

Classification of Commercial Natural Gases

| | S. Gr. | Btu/Cuft | Wobbe No. |
|---|---|---|---|
| 1. High inert | .66 - .708 | 958 - 1051 | 1179 - 1249 |
| 2. High methane | .59 - .614 | 1008 - 1071 | 1312 - 1367 |
| 3. High Btu | .62 - .719 | 1071 - 1124 | 1360 - 1326 |

To relate the above discussions to the reasoning behind categorizing the natural gas into the a) $CO_2 < 5$ m%, b) $5 < CO_2 < 30$ m%, and c) $CO_2 > 30$ m% , it is assumed that the natural gas is made up of $CH_4$ and $CO_2$. It is important to know the limits or the $CO_2$ content in order to classify the gas as follows:

1 - Natural gas containing $CO_2$

2 - Natural $CO_2$ source containing other impurities

Table K-1 gives Gross Heating Value, Specific Gravity, and Wobbe Number for methane-carbon dioxide mixture.

TABLE K-1

GHV, S.Gr. and Wobbe No. for $CH_4$ - $CO_2$ Mixtures

| $CH_4$ - $CO_2$ Mixutre | S. Gr. | GHV Btu/SCF | Wobbe No. |
|---|---|---|---|
| 95 - 5 | 0.6 | 997.5 | 1288 |
| 90 - 10 | 0.65 | 945.0 | 1174 |
| 80 - 20 | 0.74 | 840.0 | 973 |
| 70 - 30 | 0.84 | 735.0 | 801 |
| 60 - 40 | 0.94 | 650.0 | 651 |
| 50 - 50 | 1.03 | 525.0 | 516 |

From the results given in the above table it can be seen that the Wobbe Index reaches the mininum limit as shown in Table 3. As a result, the cut-off in the $CO_2$ content of a natural gas, as far as the processing is concerned, is set at 30 mole percent $CO_2$.

Another reason for the 30 mole % $CO_2$ content as a cut-off point in categorizing natural gas is that most physical absorption processes (as defined in Section I-C) are most effective economically up to about 30 mole % acid gas loading ($CO_2$ in this case).

Choice of solvent for gas treating is important because solvent circulation rate usually determines equipment sizes, the overall cost of the unit, and therefore the cost of the investment. Circulation rate of the solvent also determines the amounts of cooling or heating needed and sets the required horsepower for the moving machinery.

Solvent selection should be based on composition, temperature, and pressure of the feed gas plus the desired conditions of the product gas. These parameters usually determine whether a chemical or physical solvent is more economical. For a case where all parameters except the acid gas loading is the same, the circulation rate becomes directly proportional to the acid gas content of the feed. Chemical solvents can absorb acid gases without a high sensitivity to pressure, whereas the physical solvents can absorb acid gases in proportion to their partial pressures.

Chemical solvents usually consist of an aqueous solution of a reagent. Substantial amounts of energy (heat) are needed to regenerate the rich solution. Physical solvents have low heats of solution and in most cases do not require heat for regeneration.

As an example, with acid gas partial pressure at 75 psig (90 psia), capacity of physical solvents in commercial use far surpasses that of chemical solvents.

Figure 1 shows the range of treating for various processes. Assuming natural gas containing only $CO_2$ at an operating pressure of 1000 psig, the amine processes are economically preferred up to a partial pressure of 10 - 15 psi (1 to 1.5 $CO_2$ mole percent). Other chemical processes such as activated hot potassium carbonate can be considered up to a partial pressure of 50 - 75 psi (5 - 7.5 mole percent) while producing a comparable product purity. This can be considered as a main reason for drawing the line at 5 mole percent $CO_2$ in categorizing the natural gas containing carbon dioxide.

Natural gas streams containing greater than 30 mole % $CO_2$ must be purified in order to meet the Wobbe Index or other gas quality specifications, even if the product is not restricted to a certain $CO_2$ purity level. These types of streams require special processing. A typical schematic flow sheet is shown in Figure 2. Considering natural gases containing $CO_2$ as $CO_2$ resource for the Enhanced Oil Recovery processes, the natural gas containing more than 30% $CO_2$ can be considered more as a $CO_2$ resource where upon processing will yield a very valuable by-product namely, methane and other hydrocarbons.

Processing alternatives for natural gas can be summarized as follows:

1) Streams containing less than 5% mole $CO_2$ which can be treated using chemical solvents.

2) Streams containing greater than 5% mole and less than 30% $CO_2$ are excellent candidates for the physical solvents.

3) Streams containing greater than 30% $CO_2$ reqiure special processing. A typical schematic flow sheet is shown in Figure 2.

## Approach to Analysis of Transportation Alternatives

The transportation modes considered in previously conducted global $CO_2$ supply studies (References 4 and 6) were restricted to pipelines only. While it was obvious that pipelines were clearly the most efficient method of moving large quantities of $CO_2$ over a long period of time, it was not obvious that it was the best means of supplying small quantities for short periods of time. Specifically, the crossover point between truck, rail, and pipeline transport had never been reported in a generalized sense. (The study performed by Lawrence-Allison & Assoc. did address this issue for the Los Angeles Basin on a restrictive basis). (Reference 8) There remained the open issue; could a mixed mode transportation system consisting of a large diameter trunk line and combinations of lateral pipelines and trucks better serve the needs of more modest sized EOR projects.

This question arose from the fact that large pipelines favored long term (20 year +) investment recovery economics, and that most $CO_2$ injection projects required on the order of five years to complete.

The second area of concern with the results of past studies was the assumption made regarding the thermodynamic state of $CO_2$ for both transport and pressurization. The past studies concluded that the supercritical state was the thermodynamic state which provided the most efficient pipeline transport. The definition "super critical" merely means that the pressure of the fluid is higher than the critical pressure. At pressures over the critical pressure there can be no abrupt phase change thus ruling out the problems associated with two-phase flow. Although there are no distinct phase changes in this regime, there is a significant difference in the fluid's compressibility depending on the temperature assumed. The previous studies (Reference 6) recommended that the fluid be heated to $130^{\circ}$ F prior to compression resulting in very high compressibility and hence, high compressor work. The energy requirements and economics of pressurizing in the high compressibility regime were compared to those which assumed a lower temperature prior to compression.

The approach followed to access the impact of novel methods of processing and supplying $CO_2$ on the amount of EOR oil was to determine the bounds of additional $CO_2$ that was available from natural sources (specifically natural gas $CO_2$ mixtures) if the novel methods were applied. Unfortunately, the accessment is tied to both the amount of new gas found and its $CO_2$ concentration.

The details of the methodology followed in the analysis are discussed in the next section of this report.

## Approach to Analysis of Changes in $CO_2$ Reserves

The changes in $CO_2$ reserves will occur if the newly discovered natural gas sources containing 5 to 10% $CO_2$ are added to the $CO_2$ reserves. As discussed in the "Approach to Processing Alternatives," natural gases containing up to 10% $CO_2$ can be purified (taking the $CO_2$ out) using chemical or physical solvents. The economic incentives for taking small amounts of $CO_2$ out of the natural gas are discussed in the "Results" section. As long as the amount of $CO_2$ produced is in the range of 5-10 MMSCFD and the processing plant is within a 100 miles of the oil reservoir where the $CO_2$ can be utilized, separating the $CO_2$ from the hydrocarbon portion of the natural gas will have distinct economic incentives. The 5-10 MMSCFD $CO_2$ can be supplied from various combinations of amounts of natural gas and its $CO_2$ concentration. A 200 MMSCFD natural gas stream containing only 2.5 mole % $CO_2$ will be as attractive as a 100 MMSCFD natural gas with 5 mole % $CO_2$ in as far as total $CO_2$ content is concerned. When sufficient amounts of $CO_2$ can be produced, the natural gas containing only a few mole percent $CO_2$ can be considered a $CO_2$ reserve.

## III. RESULTS

The results of the analysis are divided into the categories discussed in the previous section:

- Processing Alternatives
- Transportation Alternatives
- Changes in Availability of $CO_2$ from Natural Sources

The discussions of the results that appear in this section are divided into the same above mentioned categories.

## Processing Alternatives

As described earlier, the processing of natural gas containing $CO_2$ was divided into three general categories.

1 - Natural gas streams containing less than 5 mole % $CO_2$

2 - Natural gas streams containing greater than 5 but less than 30 mole % $CO_2$

3 - Natural gas streams containing greater than 30 mole % $CO_2$

The results from each of the above will be discussed separately.

### A. Processes for Natural Gas Streams Containing Less Than 5 Mole % $CO_2$

Economically, chemical type purification processes are best suited to natural gases containing less than 5 mole percent $CO_2$. Depending on feed composition, temperature and especially pressure, the chemical processes may have an economic edge over physical processes even up to about 10 mole percent $CO_2$ content in the feed.

In as far as the purpose of this study is concerned, the natural gases containing 5 - 10 mole percent $CO_2$ were examined as an additional $CO_2$ resource.

### Cost of Processes Used for 5 to 10 Mole % $CO_2$

Natural gas streams containing 5 to 10 mole % $CO_2$ will need to be purified to meet pipeline quality standards. In the process of sweetening the natural gas, a high purity $CO_2$ mixed with a small amount of hydrocarbons will come out of the regenerator.

The pressure and temperature of the $CO_2$ coming out of a regenerator will depend on the type (chemical or physical) and the specific process used. Ranges of outlet pressure and temperature will be:

| | |
|---|---|
| Outlet Pressure | Vacuum to 30 psig |
| Outlet Temperature | 200 to 450°F |

The processing cost of $CO_2$ will be mainly dependent upon the mode and condition of transportation. The $CO_2$ delivery conditions considered in this report were:

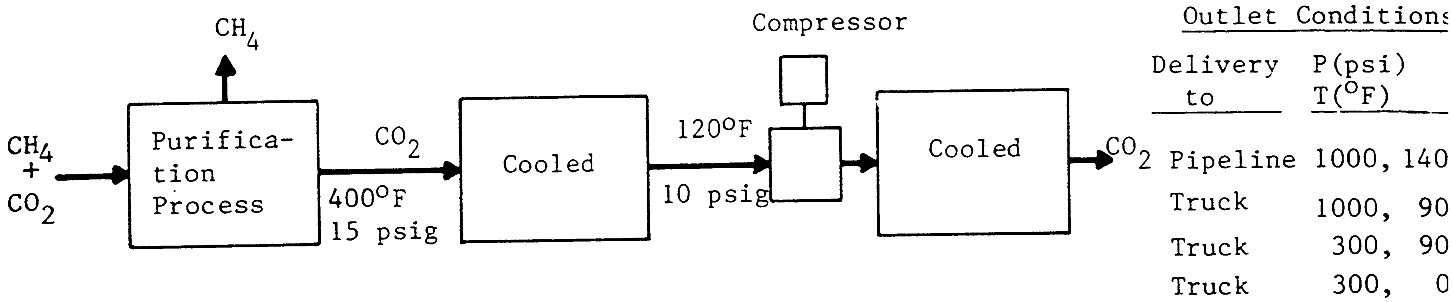
$CO_2$ Delivery Conditions

| Delivery To | Pressure (psi) | Temperature ($^oF$) |
|---|---|---|
| Pipeline | 2000 | 140 |
| Truck | 1000 | 90 |
| Truck | 300 | 90 |
| Truck | 300 | 0 |

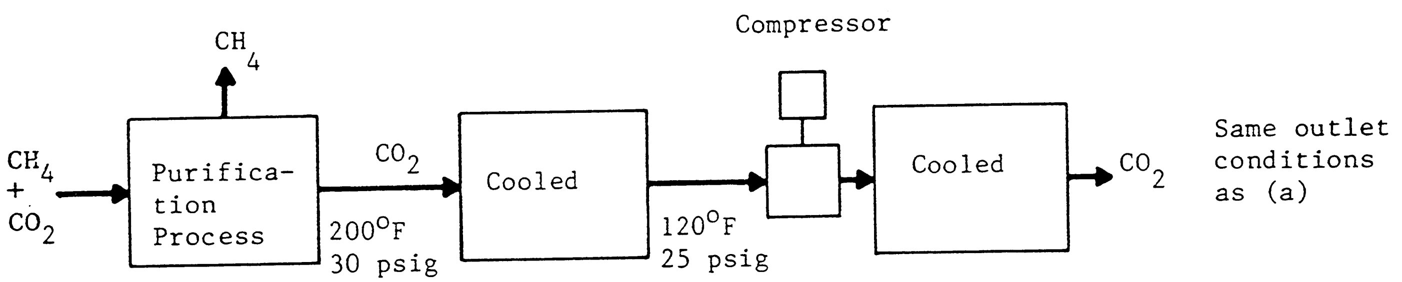
The general processing scheme required is given in Figure R-1.

Numerous costing alternatives can be considered. The following table describes the alternatives and the elements involved.

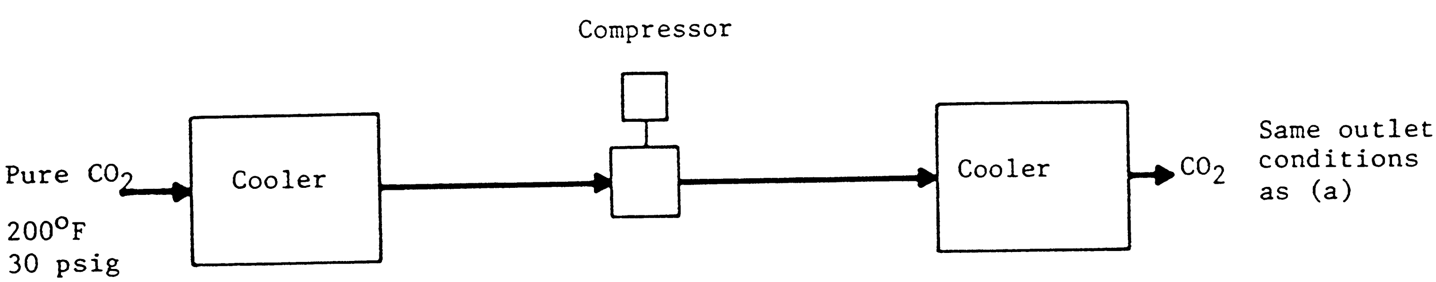
| | Purification Cost | | | | |
|---|---|---|---|---|---|
| Alternative | Included | Adjusted | Ignored | Type of Process | Type of Delivery Pressure, Temp. |
| A-1 | yes | -- | -- | Chemical | Pipeline (2000, 140) |
| A-2 | yes | -- | -- | Chemical | Truck (1000, 90) |
| A-3 | yes | -- | -- | Chemical | Truck (300, 90) |
| A-4 | yes | -- | -- | Chemical | Truck (300, 0) |
| B-1 | yes | -- | -- | Physical | Pipeline (2000, 140) |
| B-2 | yes | -- | -- | Physical | Truck (1000, 90) |
| B-3 | yes | -- | -- | Physical | Truck (300, 90) |
| B-4 | yes | -- | -- | Physical | Truck (300, 0) |




a) Processing reqiured using a chemical processing




b) Processing required using a physical processing




c) Processing required for pure $CO_2$ out of a sweetening plant

Figure R-1 General Processing Schemes for Delivering $CO_2$ to Various Transporting Conditions

As seen from the above table, there are eight alternatives when the total purification cost is included in the final cost of $CO_2$. Obviously this is unfair since the natural gas will be upgraded as far as its heating value is concerned. Assuming the price of natural gas depends on its Btu content, i.e., so many dollars per million Btu, then the purification cost can be adjusted such that the $CO_2$ product does not bear the entire purification cost. It is also possible to completely ignore the purification cost in costing the $CO_2$ product. This is assuming that the natural gas had to be purified for other reasons, and its $CO_2$ content was to be released into the atmosphere.

To determine the various cost elements the following example will serve as a basis:

| Natural Gas Stream | Conditions |
|---|---|
| Amount | 100 MMSCF/D |
| Methane | 92.5 mole percent |
| Carbon Dioxide | 7.5 mole percent |
| Pressure | 1000 psig |
| Temperature | 100°F |
| Other Impurities | Nil |

The example is chosen such that either a chemical or physical solvent can be used.

To calculate the total cost of processing the above gas and delivering it in accordance to the transportation conditions required, the costing will be broken down as follows:

- Cost of Purification
- Cost of Additional Processing Required by the Transportation

## Cost of Purification

For the given example, it was assumed that the natural gas product stream will be virtually free of $CO_2$ and the product stream will not contain any hydrocarbons. The conditions of the product streams will be:

## Product Streams

| | Natural Gas | Carbon Dioxide |
|---|---|---|
| Amount (MMSCF/D) | 92.5 | 7.5 |
| Pressure (psig) | 985 | vacuum - 30 |
| Temperature ($^{O}F$) | 110 | 200 - 450 |
| Impurities | Nil | Nil |

As stated earlier, the natural gas in this example can be processed using either a chemical or a physical solvent. Knowing that the actual costs depend on the specific chemical or physical process, however, for simplicity, in this section the cost of a typical chemical process is compared against that of a typical physical process.

## Purification Costs for Chemical and Physical Processes

| | Typical Chemical Process | Typical Physical Process |
|---|---|---|
| Capital Investment $/MSDCF Acid Gas | 100 | 130 |
| Total Operating Costs* $/MSCFD Acid Gas | 18.0 | 12.0 |

*Includes Operating Labor, Maintenance, Chemical Losses, Utilities, and Indirect Operaitng Costs

## Cost of Additional Processing Required Before Transportation

This cost must be divided into the following:

- Cost of cooling before compression
- Cost of compression
- Cost of cooling after compression
- Cost of refrigeration when necessary

For comparison purposes, all cooling was assumed to be done by air except when cooling to lower than 140$^{O}$F is required. The same cost of compression horsepower is used. The basis for cost calculations are:

- $600/Installed Horsepower
- 9000 Btu/HP-HR
- 7.5 HP/ton of refrigeration
- 340 Days/Year
- 1979 dollars
- $2.5/MMBtu

Basic engineering principles as outlined in references (9, 18) when used to determine heat exchanger area required, compression horsepower or refrigeration loads.

Cost of Cooling Before Compression

The cost of cooling before compression will depend on whether a chemical or physical process was used. Since the pure $CO_2$ gas coming out of a chemical process is at a higher temperature than that coming out of a physical process, the amount of heat to be removed will be higher for a chemical process than the physical one. The following table shows a comparison between cooling requirements and costs for a typical chemical against a typical physical process.

Comparison of Cooling Cost Before Compression

| | Typical Chemical Process | Typical Physical Process |
|---|---|---|
| Amount $CO_2$ (lbs/yr) | 36,278 | 36,278 |
| Temperature before cooling ($^oF$) | 450 | 200 |
| Temperature after cooling ($^oF$) | 140 | 140 |
| Heat Removed (MMBtu/hr) | 4.53 | 0.848 |
| HX surface area (sq.ft.) | 1,140 | 488 |
| Motors (Horsepower) | 30 | 10 |
| Capital Cost (1979 dollars) | 25,800 | 11,800 |
| Operating Cost ($/year) | 5,500 | 1,840 |

## Cost of Compression, Intercooling, and After-cooling

Depending on the mode and specific conditions of transportation, the compression costs (capital and operating costs), and Intercooling and After-cooling costs (capital and operating costs) are given in Tables 4-7 for the following;

- Truck Transport at 300 psi and 90°F (Table 4 )
- Truck Transport at 300 psi and 0°F (Table 5 )
- Truck Transport at 1000 psi and 90°F (Table 6 )
- Pipeline Transport at 2000 psi and 140°F (Table 7 )

## Total Capital and Operating Costs

The total capital and operating cost includes:

- Purification Cost (Chemical or Physical)
- Cooling Cost Before Compression (Chemical or Physical)
- Compression Cost, Intercooling, and After-cooling, with the following Transportation Options:
    - Truck 300 psi, 90°F
    - Truck 300 psi, 0°
    - Truck 1000 psi, 90°F
    - Pipeline 2000 Psi, 140°F

As stated earlier, other options where the purification costs are either ignored or adjusted should also be included. Table 8 provides the total capital and operating cost matrix for the various options:

For the "Adjusted Option" the purification costs were adjusted on the following basis:

- 100 MMSCFD gas inlet to the sweetening plant
- 92.5 MMSCFD sweet gas out if processed
- Distance to be transported is 100 miles
- Btu value of the gas when it reaches the consumer will be the same whether 100 MMSCFD of sour gas or 92.5 MMSCFD of sweet gas is put through a trunkline
- Same trunkline and distance to market is assumed for the sweet and the sour gas

Trunkline transportation requirements are that the gas must be dried to less than 7 lbs/MMSCF. In both cases (sour or sweet gas) the gas must be dried to the above specification. Incentives for sweetening a natural gas containing $CO_2$ are:

- Less gas to be dried (92.5 vs. 100 MMSCFD) while delivering the same heating value to the consumer.
- Less gas to be transported through a trunk line

The following table shows the reductions in drying and transportation costs:

| Amount of Gas MMSCF/D | Nature of Gas | Dehydration Cost | | Trunkline Transportation Co |
|---|---|---|---|---|
| | | Capital Cost ($1000) | Operating Cost ($1000/yr) | ($1000/yr) |
| 925 | sweet | 475 | 157.3 | 3145 |
| 100 | sour | 525 | 186.2 | 3400 |
| Savings | | $50 | 28.9 | 255 |

Therefore, for the "Adjusted Option" the total capital cost of the purification process (chemical or physical) will be reduced by $50,000, while the annual operating cost is reduced by 255,000 + 28,900 = $283,900.

In the "Ignored Option" it is assumed that the $CO_2$ product will not be burdened with the purification costs. However, the capital and operating costs for the necessary pre-compression cooling will still be added.

Cost of $CO_2$ at the Delivery Point to the Consumer

The cost of $CO_2$ per MSCF at the delivery point to the consumer will include the following:

- Cost of purification
- Cost of pre-compression cooling
- Cost of compression including intercooling costs
- Cost of after-compression cooling or refrigeration (when necessary)
- Cost of transportation from the process plant to the point of consumption

The same economic criteria were used for determining the purification cost and other processing costs as those used for the transportation.

Table 9 shows the cost of $CO_2$ at the point of consumption for the various options given in Table 8 . The total cost was determined assuming the distance between the process plant and point of consumption to be 100 miles.

## B. Process for Natural Gas Streams Containing 5-30 Mole% $CO_2$

For natural gas streams falling in this category the choice amongst processes will become narrower. Almost without question, physical processes will have a clear advantage. The only shortcoming some of the physical processes will have is the inability to meet strict purity requirements in the processed gas. In most cases like a "Hipure" Benfield Process, this problem is solved by adding a DEA unit to scrub the gas to the desired purity requirement.

Some of the processes which are suitable in the range of $CO_2$ concentration are:

Benfield (Modified or Hipure, if necessary)
Selexol
SNPA-DEA
Sulfinol

The final choice will depend on the economics and energy efficiency factor of the above processes.

Natural gases containing more than 10 mole % $CO_2$ are the best candidates for physical type sweetening processes. The economics compared to the economics of sweetening natural gas streams containing less than 10 mole % will be improved because both the capital and annual operating costs per MSCF of $CO_2$ processed will improve as the $CO_2$ concentration in the natural gas stream increases.

The increase in the $CO_2$ content of the feed gas will also improve the economics of the other processing required prior to transportation and the economics of transportation.

## C. Processes for Natural Gas Streams Containing Greater Than 30 Mole % $CO_2$.

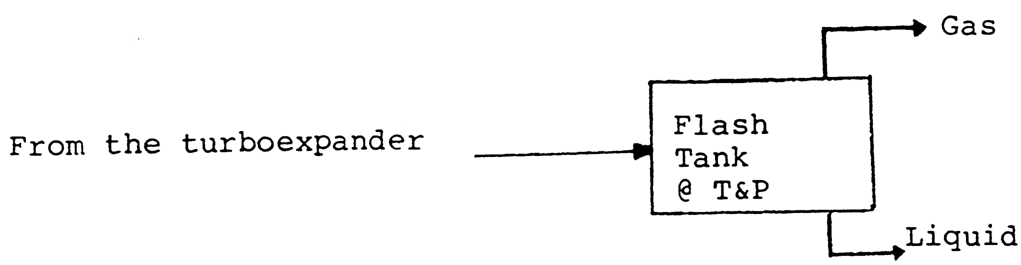
As described earlier, these types of natural gas streams should be categorized as natural $CO_2$ sources. The hydrocarbon content should be looked at as a welcomed by-product and a valuable one. The general processing scheme for these gases are given in Figure 2.

As shown in this figure, the gas stream must be cooled to a desired temperature in order to optimize liquid $CO_2$ recovery. The cooling process can be achieved a) by expanding the gas (if it is available at high enough pressure), and b) by using outside refrigeration. Economizers will be used to exchange heat between the hot streams (such as the feed) to be cooled and the cold streams which need to be heated.

It was mentioned in Part B (above) that physical sweetening processes show an economic advantage as the $CO_2$ concentration increases. However, at some point, like 30 mole %, the capital and operating cost of the sweetening process will be proportional to the circulation rate of the physical solvent and the circulation rate will be directly proportional to the $CO_2$ concentration in the sour fuel gas.

As shown in the schematic flow sheet, use of turboexpanders (when a high wellhead pressure is available) or other refrigeration means (when the wellhead pressure decreases) will produce a liquid stream with a high $CO_2$ content and a gas stream with a $CO_2$ content in an optimized range for a physical solvent sweetening process.

Assuming the natural gas to be a binary mixture of $CO_2$-$CH_4$, flash calculations can be run at different pressure and temperatures for different mixture compositions.




Moles of $CO_2$ in the Liquid Phase = $X_{CO_2}L$

$$X_{CO_2}L = \frac{Z_{CO_2} K_{CH_4} + K_{CO_2}(1 - K_{CH_4} - Z_{CO_2})}{(K_{CH_4} - K_{CO_2})\,(1 - K_{CO_2})} \qquad (1)$$

Where: $X_{CO_2}$ = Mole fraction of $CO_2$ in the liquid

$L$ = Moles of liquid

$Z_{CO_2}$ = Mole fraction of $CO_2$ in the feed

$K_{CH_4}$ = Vapor-liquid equilibrium ratio for methane

$K_{CO_2}$ = Vapor-liquid equilibrium ratio for carbon dioxide

As seen in Equation (1) which is a result of material balance and vapor liquid equilibrium, the moles of $CO_2$ in the liquid phase are related to the moles of $CO_2$ in the feed, the equilibrium constants for $CH_4$ and $CO_2$.

The temperature and the pressure in the flash tank following the turboexpander (or other means of refrigeration) will determine:

- Amount of liquid stream containing higher $CO_2$ concentration
- Amount of gas stream containing lower $CO_2$ concentration

As mentioned earlier, a simple computer program was written to determine the temperature and pressure (for specific feed composition) at which the amount of $CO_2$ in the liquid phase will be a maximum. Tables 10 to 14 show the temperature and pressure results for feed composition of 30 to 50 % $CO_2$. The above tables were prepared only for temperature, pressure, and composition for which some experimental data were available.

The proper choice of temperature and pressure will depend on the well head pressure and the percent $CO_2$ in the fluid. The high pressure streams can be expanded through a turboexpander to drop the temperature of the fluid.

The cost of the gas processing plant will depend on the flow rate and the pressure of the feed gas. Most sweetening processes are temperature sensitive. The MEA process for example will not be effective at all at inlet gas temperatures above 120$^{o}$F. Regardless of the type of sweetening process (chemical, physical, or solid bed) the cooler the inlet gas is (within limits, i.e., prevent gas hydrates, etc.), the more efficient the separation process will be. The inlet gas pressure has the opposite effect as the inlet gas temperature. The higher the pressure (within practical limitation), the more effective the separation will be. The inlet gas pressure has a twofold effect: 1) more efficient separation process, especially physical and solid bed type processes, will be at higher pressures, 2) smaller processing equipment required (such as contact towers) to process the same amount of gas at low pressures. However, higher pressure means thicker vessels, more steel and greater cost. An optimum contact tower pressure can be determined based on vessel size, wall thickness and weight, pump horsepower for returning the solvent into the tower and other sundry equipment, such as heat exchanges, etc. The contact tower pressure will also be determined by inlet pressure required for the pipeline transporting the processed gas. Therefore, it is definitely uneconomical to process the gas at much higher pressure than required as the inlet pressure to the gas pipeline plus 10-20 psi for pressure drop in contact tower.

Natural gas reservoirs containing sufficient $CO_2$ will be more interesting if the wellhead pressure is higher than the pressure required for the contact tower. The higher wellhead pressure can be utilized to cool the gas using turboexpansion.

Use of Turboexpansion in Recovering $CO_2$ from Natural Gas

In order to carry out any sort of process calculations on a mixture of $CH_4$-$CO_2$, a reliable range of vapor-liquid equilibria (K) are required. The K data for the mixutre of $CH_4$-$CO_2$ are especially critical in the design of a Turboexpander and the following separation equipment. As will be explained in the ensuing discussion, it is very difficult to find any consistent data.

Inconsistency in the Vapor-Liquid Equilibrium (K) Data for $CH_4$-$CO_2$ Mixtures

The K values for any component in a mixture are a function of temperature, pressure, and composition.

$$K = K\ (T,P\ X_i) \qquad (2)$$

considering the phase rule Equation:

$$P + F = C + 2 \qquad (3)$$

for a two component mixture; i.e., $CH_4$-$CO_2$, C = 2. Hence, in a two phase region (vapor-liquid equilibrium), p = 2. Substituting the values for C and P, the degree of freedom (F) from Equation (3) is calculated to be 2. Therefore, if temperature and pressure are specific (two degrees of freedom), the K values for the $CH_4$-$CO_2$ mixutre must be the same for all various compositions.

The K values for $CH_4$ in a 50-50 mixture of $CH_4$-$CO_2$ from Reference 19 are shown in Figures 3-5. The K values computed using the NGSPA program for a 40-60 ($CH_4$-$CO_2$) mixture are shown in Figure 6. As can be seen from these figures, the K data for $CH_4$ at the same T and P are not the same for the (50-50) and the (40-60) mixture.

The K data for methane at different convergence pressures of 800, 1000, 1500, 2000, 3000, and 5000 psia were obtained at temperatures of -50, 0, and 50$^{O}$F and pressures of 100, 250, 500, 750, 1000, and 1500 psia. A plot of $K_{CH_4}$ against convergence pressure is given for the applicable pressure range for each of the above temperatures (-50, 0, 50$^{O}$F). These plots are shown in Figures 7,8, and 9. As can be seen from these plots, the $K_{CH_4}$ values are very much a function of the convergence pressure when C.P. is below 2000 psia. The convergence pressure is a way to determine the effect of composition on the K value of a component in a mixture.

An attempt was made to determine whether there was any consistency between the $K_{CH_4}$ data in a 50-50 $CH_4$-$CO_2$ mixture experimental data, 40-60 mixture calculated values, and values obtained from the $K_{CH_4}$ vs. convergence pressure plots.

Unfortunately, no consistency or meaningful correlation can be found. The $K_{CH_4}$ was also calculated using a correlation given by Canfield (23). In this reference $K_{CH_4}$ is given as a function of T, P, and convergence pressure. The result of $K_{CH_4}$ determined from various sources at a given T, and P are compared in Table 15 As it can be seen from the results given in Table 15 , the $K_{CH_4}$ values differ by a wide enough margin that none can be trusted.

This leads one to think that there is a very dire need for reliable and consistent vapor-liquid equilibrium data for $CH_4$-$CO_2$ mixtures. Such data should be determined experimentally at wide composition, temperature, and pressure ranges such that the two phase (vapor-liquid) envelope can be plotted for each mixture composition. A plot of critical point locus as a function of composition for the $CH_4$-$CO_2$ mixture can also be prepared.

Most $CH_4$-$CO_2$ mixtures transported by pipeline or other means will have a small amount of other impurities, such as higher Hydrocarbons (ethane, propane), Nitrogen, and other gaseous elements. It will not be possible to determine all necessary K values experimentally due to infinite numbers of combinations that can exist. However, an attempt can be made to determine the effect of presences of a third element (which can be a single component or combination of two or more components) in the $CH_4$-$CO_2$ mixture. The presence of a third element will provide extra degrees of freedom in accordance to Equation 2 and hence, make the $K_{CH_4}$, and $K_{CO_2}$ a function of composition as well as temperature and pressure.

Lack of Other Physical Property Data for $CO_2$ - $CO_4$ Mixtures

- Critical Temperature and Pressure Data

A plot of critical pressure against critical temperature for different $CH_4$ - $CO_2$ mixture compositions was prepared using available experimental data references (19, 20, and 21). This plot is shown in Figure 10. As it can be seen from this figure, $CH_4$ - $CO_2$ mixtures behave very nonideally. The deviation of the critical pressure of the mixture at some points is as much as 100 percent from those which can be predicted using various mixing rules.

The critical property of the mixture is very important in determining phase behavior.

- Heat Capacity Data

No reliable data for heat capacities of the $CH_4$-$CO_2$ mixtures could be found. The heat capacities are needed in performing heat transfer calculations. The ratio of $C_p/C_v$ is very important in calculating the outlet temperature out of the turbo-expander.

The heat capacities for the mixture at a wide range of temperatures were calculated using reduced temperature ($T_r$) and reduced pressure ($P_r$) and generalized enthalpy departure charts $(H-H^*)/T_C$. Since $T_C$, and $P_C$ data for a few mixtures were available hence, Figure 10 was prepared and then used to determine $T_C$ and $P_C$ for other mixtures.

The values for $1/m = k/(k-1)$ were also calculated using $C_p$ values as described above. The values for various $CH_4$-$CO_2$ mixtures at different pressures and temperatures are given in Tables 16 through 21. The 1/m values are extremely critical in calculating outlet temperatures out of an expander of a compressor. The 1/m values calculated using experimental $T_C$ + $P_C$ values as explained varied as much as 200-300 percent from those obtained from a molecular weight correlation given in Reference (9).

- <u>Compressibility Factor, Viscosity, and Thermal Conductivity Data</u>

Compressibility, viscosity, and thermal conductivity data for $CO_2$ containing small amounts of methane (5-8%) play an important role in performing pipeline, compression, and other transportation costs.

In order to show that the type of processing shown in Figure 2 is economically more cost effective than just a physical solvent type purification process, it is important to use reliable physical properties for $CH_4$- $CO_2$ mixtures. In preparing the table given in Figure 2 it was realized that variation in equilibrium vapor-liquid data or other physical property data causes a great variation in the amount of liquid or vapor produced and their concentrations. The amounts and compositions of liquid or vapors produced are the main determining factors in the size, costs, and as a result, the optimized conditions of the process.

- <u>Calculation of Upstream Pressure and Work Out of the Turboexpander</u>

The procedure for calculating the upstream pressure necessary to provide the proper outlet temperature when expanded is given in Appendix B.

The amount of work which can be expected out of the turboexpander is also given in Appendix B.

## TRANSPORTATION ALTERNATIVES

In this section we evaluate alternative means for transporting carbon dioxide from a source to an EOR project. Large, natural $CO_2$ resources appear to be inadequate to supply projected U. S. EOR demand for carbon dioxide (reference 1 ), but published studies of $CO_2$ transportation seem directed principally to high-rate supply from large, naturally-occuring resources. Furthermore, though rapid gains are being made, the U. S. petroleum industry has had very little experience building and operating carbon dioxide pipelines, and our brief review (reference 10 ) indicated that serious uncertainties surround the thermodynamic properties of both pure $CO_2$ and also $CO_2$-hydrocarbon mixtures. Consequently, this section examines the following basic questions:

> Are pipelines being designed for the transmission of $CO_2$ properly accommodating the unique thermodynamic properties of $CO_2$?
>
> Is the exclusive use of pipelines the optimum transport mode for all $CO_2$ EOR projects, regardless of injection rates and project durations?

Cost-saving innovations in $CO_2$ transport may permit economical utilization of smaller amounts of $CO_2$ than are currently judged practical, so that an expanded number of the lower-purity, natural $CO_2$ reservoirs may be economically extracted, and thus contribute to increased U. S. $CO_2$ reserves.

## Thermodynamic Properties Affecting Pipeline Design

The critical design parameter for economical $CO_2$ pipeline design is pressure drop, which is largely dependent upon the thermodynamic properties of the transported gas. Existing $CO_2$ pipelines and previous studies of $CO_2$ transport have placed emphasis upon operation in the supercritical gas phase (References 4, 6, 8). A pressure-enthalpy (Mollier) diagram for $CO_2$ is shown in Figure 11. The four regions identified in the figure are the liquid two-phase, gas, and the "dense-phase-fluid," which is the description of the condition of carbon dioxide above the critical pressure of 1,071 psia. The supercritical gas condition refers to the part of the dense-phase-fluid (DPF) region in Figure 11 labeled "high compressibility." Operation in the supercritical gas phase is economically attractive because higher densities can be achieved than those possible below the critical pressure (the top of the "dome" in Figure 11), and because abrupt pressure and temperature drops will not necessarily induce two-phase flow, which is damaging to pressurization equipment (Reference 7).

Within the DPF region of $CO_2$ there are operating conditions of pressure and temperature that are economically more attractive than others for $CO_2$ pipeline design. In such a region, which is identified as "low compressibility" in Figure 11, the lines of constant-entropy (S - constant) become more vertical as one progresses from the higher temperature to the lower temperature of the DPF region. In intuitive terms, $CO_2$ in the DPF behaves more as a "liquid" than as a "gas" as the temperature is decreased at constant pressure. This characteristic is normally expressed in the equation of state by a reduction in the compressibility factor (hence, the labeling of Figure 11). The work required to compress the DPF in the low-compressibility region is significantly less than that required in the high-compressibility region, so that one would expect the energy (and thus the cost) required to transport $CO_2$ as a low-compressibility DPF would be less than as a high-compressibility DPF.

The economic incentive for $CO_2$ pipeline operation in the low-compressibility-DPF region is a function of the following comparative design requirements for low-compressibility and high-compressibility schemes to transport a given quantity of gas:

Costs of pressurization at compressor stations
Costs of compressor repressurization stations
Separation between compressor stations
Diameter of pipeline

The optimal selection of these pipeline system design parameters depends on the quantity of $CO_2$ to be transported, the cost of pipe as a function of diameter, the cost of compressors as a function of horsepower (and, precisely but to a limited extent, flow rate), the cost of energy to pressurize the $CO_2$, and the physical properties of carbon dioxide at the various pressures and temperatures of the pipeline (e. g., at compressor inlets and outlets). In specific applications, terrain and routing also are key factors in pipeline design; for example, at the base of a steep mountain ridge, a <u>de</u>pressurization station may be necessary to assure that static head pressure in the pipeline does not exceed the allowable stress for the pipe. While a specific design must consider such real-world constraints as terrain, to characterize the potential economic advantages of low-compressibility operation, we assumed a "flat," unconstrained pipeline route, and evaluated alternative low- and high-compressibility paths for $CO_2$ pipeline operation at various delivery rates.

The candidate paths for the initial pressurization of the carbon dioxide are illustrated in Figure 12 . The following assumptions are common to both cases:

200 psia and 140°F wellhead conditions

First-stage compression from wellhead conditions to 1000 psia and 80°F, preceded by cooling

Single-stage pressure ratio limited to 3:1 for high isentropic efficiency

In order to access the low-compressibility region without the aid of refrigeration, it is necessary to compress $CO_2$ in the gas phase to a pressure which has a reasonably high saturation temperature; this first compression stage is common for both low- and high-compressibility paths. The gas was initially cooled to reduce the compressor work (1 to 2), and compressed in the first stage to 1000 psia and 80°F (2 to 3). The gas was then intercooled (3 to 4), and compressed in the second compressor stage. In the second stage, the two paths depart. The high-compressibility path continues in the near-isentropic manner to point 5. The low-compressibility path from the departure point 4' to point 6 consists of heat transfer and condensation of the carbon dioxide. The remainder of the pressurization for this path is performed with $CO_2$ in the liquid and low-compressibility DPF states. The departure reduces the pressurization work for the low-compressibility path; the advantage appears in Figure 12 as the difference between $\Delta h_2'$ and that portion of $\Delta h_2$ above the point of path departure (4' to 5). As previously stated, the reason for the departure at 1000 psia is that this pressure represents the lowest at which condensation can be reasonably acheived without the aid of refrigeration.

The compressor horsepower requirements associated with the two thermodynamic paths are shown in Figure 13 , which is a plot of compressor horsepower per MSCFD vs $CO_2$ wellhead pressure. Of course, in an actual application $CO_2$ reservoir pressure should be utilized whenever possible to reduce compressor horsepower requirements. For the range of hypothetical pressures of 400 to 1200 psia in the example, however, the Figure indicates that substantial reductions in compressor horsepower requirements can be achieved by utilizing the low-compressibility path. At low wellhead pressures, these differences become smaller, however, for the relatively longer length of the commonly shared legs decreases the difference in total work between the two paths. Nevertheless, even though the percentage energy savings for the low-compressibility path are less at lower wellhead pressure, the difference over the range of 400 to 800 psia is still 20 to 50 hp per MSCFD.

Repressurization is required for long pipelines to overcome the frictional losses encountered by the gas. The total energy requirement of repressurization is the sum of the energy required by the compression, plus any energy required for special thermal conditioning. Assuming that the pipeline can be buried sufficiently deep to eliminate temperature changes due to variations in ambient air (which, for normal soils, requires about 4-ft burial), we assumed that the ground temperature never exceeds $70^{\circ}F$. (Most of the Midcontinent region has temperatures significantly lower than $70^{\circ}F$.) As cited in reference 6 , we also assume that $CO_2$ must be heated to $130^{\circ}F$ (at 1400 psia pipeline pressure) to provide a sufficiently high compressibility factor to meet gas compressor manufacturer's warranty conditions. Thus, compliance with manufacturer's warranty requirements on the compressors forces additional energy to be added to the carbon dioxide for thermal conditioning prior to repressurization. However, this thermal conditioning is only needed in the high-compressibility path.

Figure 14 shows the relative temperature changes experienced by low- and high- compressibility pipelines for 70$^{\circ}$F ground temperature and minimum pressure of 1400 psia. The temperature of the $CO_2$ must be reheated to meet the compressibility requirements prior to entry into the compressor.

If pressurization is accomplished in the low-compressibility area, then the temperature of the $CO_2$ never departs significantly from ground temperature. For pipelines of equal diameter and capacity, the pressure loss per unit length will be lower for the lower temperature, more dense fluid.

These two approaches to repressurization can be analyzed by plotting the paths on a Mollier diagram as shown in Figure 15. The pump work required in the low-compressibility path is indicated by the $\Delta h_p$. The energy required to condition the $CO_2$ (i.e., heat it to 130$^{\circ}$F) will depend upon the temperature the $CO_2$ had reached just prior to the compressor station. One can see that in cases where the $CO_2$ temperature approaches the ground temperature, substantial amounts of energy are required to reheat the fluid to the required compressor inlet conditions.

To determine the impact of the above mentioned phenomena, the operating characteristics were calculated for $CO_2$ pipelines of four different capacities. In all cases a maximum operating pressure of 2000 psia and a minimum operating pressure of 1400 psia were assumed. In one set of calculations, the initial conditions of 2000 psia and 80°F were assumed (low-compressibility path). In the second set of calculations, the initial conditions were assumed to be 2000 psia and 185°F (high-compressibility path). Modified Panhandle (Panhandle B) equations (REF. 9 ) were used to compute the pressure drop of the high-compressibility path. Both the Modified Panhandle equations and methods more suitable for product pipelines were used for the low compressibility path. The heat transfer from the $CO_2$ was limited by the heat transfer coefficient at the outer wall of the pipe, which was assumed to be 0.6 BTU/ft-°F-hr, a value calculated for average soil conductivities. The pipeline diameter was held fixed for each line capacity.

The results of the pressure loss and heat transfer calculations are shown in Table 22. The pressure loss is represented by the distance between compressor stations (i.e., the line length at which the low-pressure limit had been reached - $\Delta p$=600 psi). As a result of the higher density, the low-compressibility path always resulted in longer distances between compressor stations (i.e., lower unit pressure losses) than the high-compressibility path. The differences in distances ranged from 17% for the 8-in diameter pipeline to 150% for the 20-in line. This variation in differences is caused by the fact that the smaller-diameter pipeline transfers heat more rapidly than the large-diameter pipe because of the higher surface-to-volume ratio of the smaller pipe. Additional evidence of the decreasing unit heat transfer rate with increasing pipe diameter is the higher final fluid temperatures of the larger-diameter pipes.

The compressor power and thermal conditioning requirements for the four pipelines are listed in Table 23. The significant differences in compressor power (hp/mi) results from both the closer compressor spacing and the higher power required to compress the higher temperature fluid for the high-compressibility path. The differences in compressor requirements for the two paths are substantial, ranging from a factor of 2 for the 8-in line to almost a factor of 8 for the 20-in line. The thermal conditioning requirements for the smaller pipelines are significant and tend to negate the beneficial effects that transferring heat to the ground had in reducing the pressure loss. Though thermal conditioning and compression requirements are superior for the low-compressibility designs, these differences are not representative of pipelines whose designs would be optimized for the respective fluid conditions. The higher capacity lines designed for the high-compressibility conditions would definitely require larger diameters to arrive at more optimum operating conditions.

The differences in heat and power requirements for alternative optimized designs may be more or less than the examples in Table 23, but the differences reported there are illustrative of the improvements possible with a given line diameter operating in a low-compressibility path.

As one would expect from the decreased energy requirements of the low-compressibility path, the estimated $CO_2$ transportation costs are lower for the low-compressibility pipeline than for the high-compressibility case. Table 24 summarizes the investment costs for two cases. The comparison is based on the standard pipeline diameter most appropriate to low-compressibility fluid transport at each of the four line capacities of 50,/125,/250 and 500 MMSCFD. Although actual high-compressibility fluid pipelines would probably be different from the diameters chosen, the analysis is illustrative of the relative economic efficiency of the two transport alternatives.

The investment costs include pipeline and repressurization facilities. Pipeline investment was assumed to be \$20,000/in mile for both low- and high- compressibility cases. Repressurization costs were assumed to be \$540/hp for both cases. As seen in Table 24, at the lower two capacities the investment costs per mile are roughly the same. The investment for the high-compressibility transport mode increases more rapidly with capacity than does the low-compressibility investment. At 500 MMSCFD, the investment for the high-compressibility system is 27% greater than the investment for the low-compressibility system with a 20-in pipeline diameter. The difference in investments is due principally to the closer spacing of compressor stations for the high-compressibility pipeline than that required for the low-compressiblity pipeline.

The difference in operating costs between the two alternatives is much more pronounced than the difference in investment. Operating costs include repressurization;

station labor and supervision; and fuel for both repressurization and revaporization (needed only for the high-compressibility transport of 50 to 125 MMSCFD). Table 24 shows that the high-compressibility system has significantly higher operating costs than the low-compressibility system. At 50 MMSCFD, the high-compressibility operating costs are 52% greater than the low-compressibility costs; and at 500 MMSCFD, the high-compressibility operating costs are 147% greater than the low-compressibility costs.

The cost analysis suggests that both investment and operating costs are lower for a dense-phase-fluid $CO_2$ pipeline than for a pipeline designed for a high-compressibility path. These results are only qualitative, however, since the same diameter pipe was assumed for both low- and high-compressibility cases. Though beyond the scope of this study, a more accurate economic comparison would involve selection of the cost-minimizing combination of pipe diameter, pressures, temperatures and pump stations. The analysis summarized in Table 24 is nonetheless useful as an indication of the potential economic benefits to be gained from operation of a $CO_2$ pipeline in an energy-conserving, low-compressibility mode. Of additional interest in this respect are the fuel costs of the two cases. Fuel cost accounts for from 29% to 43% of the operating cost for the high-compressibility pipeline, but fuel cost is only 3% to 16% of the operating cost for the dense-phase-fluid systems. Hence, the cost of the high-compressibility $CO_2$ transport system is more sensitive to rising energy cost than is the low-compressibility transport cost.

In conclusion, substantial reductions in compressor work and thermal conditioning can be achieved by maintaining $CO_2$ in the lower-temperature, dense-phase-fluid region (above the critical pressure) in pipelines.

The reductions in compressor work and thermal conditioning can translate into substantial reductions in investment and/or operating costs. These results are strongly dependent on assumptions regarding compressor operating characteristics. Specifically, the stated need to provide the compressor a fluid having a compressibility factor of 0.5 ($T = 130^{o}F$ at 1400 psia) or higher, places a severe thermodynamic penalty on $CO_2$ pressurization schemes and pipelines. The advantages of dense-phase-fluid pressurization and transport should provide the motivation to explore means of pressurizing $CO_2$ in the low-compressibility, more liquid-like area of the dense-phase-fluid thermodynamic region.

Two large diameter $CO_2$ pipeline projects have been announced since this phase of the study was conducted and reported (REF.10). The first by Shell is a pipeline that transports $CO_2$ from Montezuma County, New Mexico to the Denver Unit of the Wasson Field in Yoakum County, Texas (REF. 11). The line was sized at 36 in nominal diameter to transport 400 MMSCFD. The net elevation change between the origin and the terminus is approximately 3600 feet or a positive head equivalent to 1250 psia. Shell is providing for only one recompression station (at the 150-mile point) for the entire 478 mile length of the pipeline.

The second pipeline announced by ARCO transports $CO_2$ from Huerfano County, Colorado to the Willard Unit of the Wasson Field and the Seminole Field in Yoakum County and Gaines County, Texas, respectively (REF.12). This line will have a nominal diameter of 20 inches and will transport 300 MMSCFD. The net elevation change from the highest elevation of the line to the terminus is 4800 feet, with a positive head equivalent to 1680 psi. There will be no repressurization stations along the 420 mile-length of the pipeline.

Both of the above mentioned lines are conservatively sized (i.e., low average velocity). The flow rate per unit flow area is 23% (ARCO) and 67% (Shell) lower than the low-compressibility path cases previously discussed (which is representative of the SACROC pipeline design). Apparently, constantly rising energy costs and the difficulty of providing energy to remote compressor locations is causing operators to adopt a more conservative (i.e., more capital-intensive) design philosophy. This observation will be discussed in greater detail in the next section.

The approach of making use of the low-compressibility thermodymanic regions of $CO_2$ was followed by ARCO in the design of it's pipeline. ARCO's thermodynamic design assumptions are discussed in detail in REF.12 . There were insufficient details presented in REF.12 to determine the thermodynamic design assumptions made by Shell in the design of the McElmo Dome pipeline.

## Transportation Mode Optimization

The relatively higher unit investment for small-capacity pipelines suggests that exclusive use of pipeline might not be optimum for the transport of $CO_2$ to smaller EOR reservoirs. Studies have assumed a twenty-five-year economic life for a pipeline; also, flow rates less than 25 MMSCFD have not been considered. Reference 1 suggests that $CO_2$ project lives of 5 to 7 years should provide a planning basis for $CO_2$ transport systems.

The assumption of a twenty-five year pipeline life is reasonable for long-distance, high-capacity trunk lines, such as those being planned by ARCO, Shell, and Amoco. In these cases the pipeline life coincides with expected productive life of the $CO_2$ natural reservoirs. It is unrealistic, however, to assume that pipelines that distribute $CO_2$ from the trunkline to a given field would be in use for twenty-five years. The average injection period reported (1) indicates that five years is a more reasonable distribution system life. The injection system was projected to operate at full capacity for only two years due to the increased use of reinjected $CO_2$ in the latter phases of the project. A lateral pipeline or alternative transportation system that would be dedicated to a particular field would have to be selected on the basis of a five-year, rather than a twenty-year useful life.

## Pipeline Transport - Trunkline

Two sets of pipelines were designed to allow the comparison of alternative transport modes. The first set was felt to be representative of trunk lines, the second set represented lateral pipelines (i.e., lines whose sole function was to supply a given project). The lines were sized to provide very low average velocities with the attendant low pressure losses. This design approach is consistent with the pipeline projects previously discussed and is founded on the following:

- The rapid rate of increase of energy costs and the large uncertainty that this causes the economics of long-term projects.
- Pipeline operation and maintenance costs are almost exclusively a function of the number of compression stations along the line.
- Since the $CO_2$ being transported cannot be used as a fuel for the compressor's prime mover (as in the case of a natural gas pipeline), additional compressor stations mean long distance electrical transmission lines are required.

The results of the pipeline-sizing calculations are shown in Table 25.

The cost of transporting $CO_2$ through the trunk lines sized as those in Table 25 were almost exclusively due to the fixed cost associated with the capital required for the construction of the pipelines. The cash flow assumptions used in computing the cost for the trunk line delivered $CO_2$ were as shown below:

- The line would be constructed in two years and the investments would be made at the ends of years -1 and 0 (where time "0" designates the start of the productive life of the project).

- The line would operate the first year at an average of 80% of design capacity, the second year at 93% of design capacity, and at full capacity for the remainder of its twenty-year life. This start-up profile is consistent with that planned for the Shell McElmo line. The less-than-capacity start-up is due to the time required to complete the development of the $CO_2$ fields.

- The line would have a zero salvage value at the end of its twenty-year productive life.

Other assumptions made in the calculation of the cost of transporting $CO_2$ through the trunk line were:

- Investment tax credit of 10% of the total investment.

- Investment costs of \$15,000/inch·mile and \$20,000/inch·mile for the entire pipeline system.

- Discounted cash flow rate of return (after taxes) of 15%.

- Total (state and federal) income tax rate of 50%.

- Straight-line depreciation over 20 years.

- Yearly operating cost equal to 1% of the total investment cost.

The operating cost assumption was checked for several of the lines and found to be a conservative (i.e., generous allowance) estimate of operating cost. The resulting cost of transporting $CO_2$ is shown in Table 26. The delivery costs for the two different pipeline construction cost assumptions bound the pipeline cost reported in Reference 6. The costs are totally dependent on the pipeline construction cost assumptions. The assumptions used represented the range of average cost of pipelines that were reported in the Oil and Gas Journal's Pipeline Cost Index adjusted by the second quarter 1979 composite cost index (Reference 13).

The long distance, high-capacity pipeline is by far the most efficient means of transporting carbon dioxide. The only issue is to provide sufficient time to recover the initial investment. For the twenty-year project life that was assumed for the trunk line application, the pipeline has no competition.

## Pipeline Transport - Lateral

The principal difference in the analysis of the lateral and trunk pipeline cases is the assumed length of the project. In the OTA Study (Reference 1) and in the earlier Lewin Study (Reference 2), a universal $CO_2$ injection profile was developed that was felt to be representative of most candidate EOR reservoirs. The profile accounted for the reinjection of $CO_2$ that would be produced with the recovered oil in the latter phase of the projects. The injection profile that was developed in the two referenced studies is shown in the following table:

| Year | Purchased $CO_2$ % of total | Recycled $CO_2$ % of total |
|---|---|---|
| 1 | 20 | 0 |
| 2 | 20 | 0 |
| 3 | 16 | 4 |
| 4 | 13 | 7 |
| 5 | 6 | 14 |

The relatively short project life and the sub-capacity operation for three of the five years motivated comparison of truck transport cost and lateral pipeline cost.

The cash flow assumptions used in computing the cost of the $CO_2$ delivered via lateral pipelines were as follows:

- The line would be constructed in one year and the investment would be made at the end of year -1 or the start of year 0 (where time "0" designates the start of the productive life of the project).
- The line would operate the first and second year at design capacity and at 84%, 65%, and 30% of design capacity for the third, fourth and fifth years respectively.
- The line would have a zero salvage value at the end of its five-year life.

Other assumptions made in the calculation of cost of transporting $CO_2$ through the lateral pipeline were

- Investment tax credit of 6.66% of the total investment (owing to the 5-year production life).
- Investment costs of $10,000/inch·mile, $15,000/inch·mile, and $20,000/inch·mile (depending on size) for the entire system.

- Discounted cash flow rate of return (after taxes) of 15%
- Total (State and Federal) income tax rate of 50%
- Straight-line depreciation over 5 years.
- Yearly operating cost equal to 1% of the total investment cost.

The resulting cost of delivering $CO_2$ is shown in Table 27. For the cases where the lateral pipeline capacity overlapped the trunk line, the cost of the lateral delivered $CO_2$ was 48% higher than that of the trunk line. The difference in the costs is due to both the shorter assumed project life and the fact that the lateral does not flow at design capacity for three of the five years. To separate these two effects, the cost calculations were performed assuming that the the line flowed at full capacity for all of the five years. The change in flow assumption reduced the cost of the delivered $CO_2$ by 18%. The difference between the trunk line delivered and a full capacity lateral-line cost was reduced to a 19% difference. This discrepancy is totally attributable to the difference between the project life assumptions of 20 years and 5 years. It should be recalled that in time-value-of-money problems the cash flows in the earlier years have a much larger impact on cost than those in later years (see Reference 14).

## Truck Transport

Truck-trailer distribution of $CO_2$ was investigated as an alternative to pipeline laterals. Lubbock Manufacturing Company of Lubbock, TX, provided specifications and performance data on its line of insulated steel $CO_2$ trailers. The largest Lubbock trailer, having 5,9000-gallon capacity, was selected for the subsequent cost analysis. Transport of pure $CO_2$ at 300 psia and 0°F was assumed. The pressure is limited by the weight associated with pressure vessel tankage. Trailer capacity in standard cubic feet is as follows:

Trailer capacity =

5,900 gal x 0.13368 cu ft/gal x 4.29 lb/cu ft

(@ 300 psia + 0°F) ÷ 0.11 lb/cu ft (@ std cond.)

= 0.10246 MMSCF (@ std. conditions)

* This density is the maximum that could be achieved by turboexpanding (isentropic expansion) from 1400 psia and 70°F.

If Q is the daily $CO_2$ injection requirement in MMSCFD, then Q/0.10246 is the number of daily trailer loads required. The total requirement for trailers, however, depends upon the total trailer cycle time, which was assumed to consist of the following elements:

| | |
|---|---|
| Fill time (from manufacturer): | 0.75 hr |
| Trailer hitch and turnaround: | 0.15 hr |
| Travel time (D=one way haul,miles): | 2 x D/(30 mph)hr |
| Drain time (Q=MMSCFD $CO_2$ requirements): | 0.10246/Qx24 hr. |

TOTAL TRAILER CYCLE TIME: (1 + D/15 + 2.459/Q)hr

The above formula is based upon assumption of an average travel speed of 30 mph, irrespective of haul distance, and assumption of drain rate being a function of $CO_2$ demand rate only (i.e. trailers can be operated to drain rapidly enough). If a sufficient number of tractors are available to pull the tanker-trailers, and the average availability of trailers is 95%, then the total number of trailers required is as follows:

$$T = \text{Total Trailers} = \frac{CO_2 \text{ requirement per hour} \div \text{Trailer availability}}{\text{Trailer capacity} \div \text{Trailer cycle time, hours}}$$

$$= \frac{Q/24/0.95}{0.10246/(1 + D/15 + 2.459/Q)}$$

$$T = Q(0.4281 + D/35.0416) + 1.0526 \qquad \text{(Equation 1)}$$

Because the truck tractors need not loiter during trailer filling and draining times, the tractor cycle time is shorter than the trailer cycle time.

| | |
|---|---|
| Trailer hitch and turnaround: | 2 x 0.25 hr |
| Travel time: | 2 x D/(30 mph) hr |

TOTAL TRACTOR CYCLE TIME: (0.5 + D/15) hours

According to various trucking companies we contacted, tractor availability cannot, on the average, exceed 64% (due to maintenance, breakdowns, etc.) so that the total number of tractors required to distribute QMMSCFD is

$$S = \text{Total Tractors} = \frac{CO_2 \text{ requirement per hour} \div \text{Tractor availability}}{\text{Trailer capacity} \div \text{Tractor cycle time, hours}}$$

$$= \frac{Q/24/0.64}{0.10246/(0.5 + D/15)}$$

$$S = (D/23{,}607 + 0.3177)Q \qquad \text{(Equation 2)}$$

Equation 2 does not explicitly include tractor units needed to avoid trailer waiting time that could occur because fewer tractors than trailers are required. Turnaround and travel times were deliberately hedged to provide an extra tractor capacity "cushion" to simplify the calculations without explicit consideration of waiting-time problems. For large daily $CO_2$ requirements, the trailer waiting time should be negligible.

Table 28 presents estimates of tractor and trailer requirements for various $CO_2$ distribution requirements and distances from the trunk pipeline to the consumer oil fields. Equations 1 and 2 were used to estimate unit requirements under the assumption that tractor and trailer usage could be prorated among EOR projects of identical size and haul distance. (Put another way, "fractional tankers" and "fractional tractors" are charged to each $CO_2$ project over the haul distances and at the demands shown in Table 28.) For the 10-mile haul at a 0.2-MMSCFD demand, only about one and a third trailers are required, and these can be serviced by one tractor dedicated only 26.5% of its time. At the other extreme of haul and demand, a 10-MMSCFD $CO_2$ demand 100 miles from the trunk pipeline requires service by about 34 trailers and 46 tractors. Note that at haul distances and $CO_2$ demands beyond 25 miles and 2 MMSCFD, respectively, the tractor unit requirements exceed the trailer requirements. (These cases are enclosed by the polygon on Table 28.) This situation is due to the low availability (about 64%) of tractors relative to trailers (95%).

Based upon a brief survey of highway tanker companies and manufacturers' data, we estimate distribution costs to include the following components:

Tractor Purchase Cost: $50,000
Trailer Purchase Cost: $45,000

Driver Labor
(including fringe benefits): $30,000 per year

Tractor Operation
(fuel, tires, maintenance, etc.): $0.35/haul mile

Trailer Operation
(tires, wheels, maintenance,etc.): $0.10/haul mile

Tractor Fixed Cost
(dispatching, taxes, insurance,etc.): $6,000/year

Trailer Fixed Cost: $1,000/year

The ICC permits 7-year, straight line depreciation to be taken on 90% of the investment in a tractor and trailer. Tables 29 and 30 display details of the development of annualized tractor and trailer charges as functions of one-way haul distance and daily $CO_2$ requirements, and for 15% discount rate and 50% effective tax rate. For a given demand ($Q$) and haul distance (D), the product of the total tractor requirement times the annualized cost per tractor, plus the product of the total trailer requirement times the annualized cost per trailer, is the total $CO_2$ distribution cost by truck. (A 340-day operating year is assumed.)

Table 31 summarizes the truck-and-trailer transport costs for truck-delivered $CO_2$ (Complete cost tables are presented in Appendix A.) For trailers loaded with saturated-vapor $CO_2$ at $0^{\circ}F$ and 300 psia, Table 31 shows that the delivered-gas cost varies from $0.304/MCF to haul 10 MMSCFD over 5 miles, to $3.086/MCF to haul 0.5 MMSCFD over 100 miles. Additional Costs for truck-delivered $CO_2$ at the transport conditions of Table 31 include turboexpansion from 1400 psia (minimum trunk pipeline pressure) and $70^{\circ}F$ to 300 psia and $260^{\circ}F$ for injection at the trailer delivery point. Costs of turboexpansion (less recovered work) and recompression total about $0.07/MCF. Total truck-delivered $CO_2$ cost thus ranges from $0.374/MCF to $3.156/MCF over hauls of 5 to 100 miles and for demands of 0.5 MMSCFD to 10 MMSCFD.

Different trailer load conditions of temperature and pressure change the $CO_2$ costs. The 5,900-gal Lubbock trailer is rated to carry 44,000lb of commodity load. If the trailer carries saturated liquid at $0^{o}F$ and 300 psia (approximately 7lb/gal), then each trailer can deliver 375,480 SCFD, a 266% increase over the saturated-vapor transport case. For 0.5 MMSCFD $CO_2$ and a 100-mi haul, the transport cost is $1.066/MCF, while at 10 MMCFD and a 5-mi haul, the transport cost is $0.100/MCF. To these costs, however, must be added the cost of refrigerating the saturated vapor at 300 psia and $0^{o}F$ to saturated liquid, a change of 121 Btu/lb. The cost of refrigeration is about $0.34/MCF, so that the total cost of transporting a saturated liquid ranges from $0.420/MCF to $1.486/MCF.

We also considered a $CO_2$ distribution by a 5,900-gal trailer rated for 1000 psia and $67^{o}F$. Such a trailer could be constructed using state-of-the-art technology in pressure vessel design, as with Kevlar wrapping on a steel shell, the design of such a trailer is beyond the scope of the present study. Assuming that the trailer cost would double ($90,000 trailer), we estimate that truck-delivered $CO_2$ cost would vary between $0.948/MCF (0.5 MMCFD and 100 miles) and $0.096/MCF (10 MMCFD and 5 miles). Additional costs for turboexpansion and recompression we estimate at less than 5¢/MCF.

## Truck vs. Lateral Pipeline

Figure 16 compares lateral-pipeline, 300-psi-truck; (saturated vapor), and 1000-psi-truck transport system costs. The saturated-vapor, 300-psi-truck system is economically inferior to the lateral pipeline at all $CO_2$ rates and distances. However, the 1000-psi (saturated vapor) truck is economically preferred for delivering $CO_2$ at distances beyond 15 miles (assuming, of course, that such a trailer could be built for $90,000 or less).

## Impact on $CO_2$ Reserves & Recoverable Oil

The prospects for generating substantial quantities of naturally occurring $CO_2$ by the co-production of natural gas-$CO_2$ mixtures were discussed in Section I of the report. The prospects hinged on the existence of development of efficient means of separating relatively low concentrations (5% $CO_2$) from natural gas streams, the development of new natural gas production regions whose productive lives would coincide with the time frame of high $CO_2$ demand and efficient means of transporting small quantities of $CO_2$ over reasonably short distances.

The processing techniques discussed in the previous sections provide a large array of methods for separating $CO_2$ from natural gas mixtures of varying concentrations. In most cases, the natural gas will have to be separated from the $CO_2$ in order to meet the natural gas market specifications. The principal cost of the $CO_2$ processing is merely the cost of pressurizing the $CO_2$ to the required transport associated with high-purity process-vent sources, such as, anhydrous ammonia plants. The only additional cost associated with co-produced $CO_2$ could be the cost of gathering several relatively small quantities of $CO_2$ from the natural gas processing site. The practice that seems to be evolving in the Overthrust Belt is to build gas processing plants in modules of 250 MMSCFD capacity. If the feed gas $CO_2$ concentration were 5%, then a 250 MMSCFD processing plant would provide a $CO_2$ source of 12.5 MMSCFD, a source equivalent to a small anhydrous ammonia plant. The need to gather many of these sources would depend on the proximity of petroleum reservoirs.

If the estimate of ultimate natural gas recovery from the Western Overthrust Belt and the Tuscaloosa Trend are realized, then 14 TSCF of additional low-cost $CO_2$ could be available for $CO_2$-flooding projects. This could result in the production of an additional 2.7 billion barrels of recovered oil under the assumptions presented in Reference 1. While the above-mentioned source does not completely resolve the $CO_2$ shortfall problem, it does present a source that is approximately three times the size of currently available high-purity process-vents.

## IV. Conclusions and Recommendations

### A. Conclusions

The following objectives were accomplished:

- Defining the $CO_2$ requirements for candidate EOR reservoirs.
- Generating alternative concepts for collecting, transporting, storing and processing $CO_2$ from naturally occurring sources and comparing them on the basis of cost effectiveness.
- Re-estimating the magnitude of the $CO_2$ EOR reserves based on the findings of this study.

The details for each of the above are given in the "Results" section. The conclusions derived from the study are divided into the following:

- Conclusions from the processing alternatives
- Conclusions from the transportation alternatives
- Overall effects of conclusions from processing and transportation on $CO_2$ - EOR reserves.

#### Conclusions from the Processing Alternatives

After studying about 50 different natural gas purification processes for separating $CO_2$, the following conclusions were reached:

- If the amount of $CO_2$ separated out of a natural gas exceeds 5 MMSCFD, the natural gas containing $CO_2$ should be sweetened even if it contains only 2-3 mole % of $CO_2$.
- As a rule of thumb, in separating $CO_2$ from natural gas, chemical solvents should be used when the $CO_2$ content is less than 4-5 mole %, physical solvents when the $CO_2$ content is 5-30 mole %, and a combination of refrigeration (using turboexpansion when high wellhead pressures are available) and physical solvents when the $CO_2$ content is greater than 30 mole percent.

- Cost of $CO_2$ prior to transportation depends on
  - mode of transportation (.30-.65 $/MMSCF)
  - type of purification (.29-.30 $/MSCF)
  - whether or not it includes all, adjusted capital and operating cost, or none of tne purification cost (.03, 0.18, 0.21 $/MSCF)
- Cost of processing $CO_2$ prior to transportation depends on the mode and conditions required at the loading point. The cost of processing compared to the cost of transportation ranges from about 10 to 75 percent, even when all of the purification costs are burdened on to the $CO_2$ costs.

## Conclusions From The Transportation Alternatives

Various modes of transportations and combinations were studied. Conclusions are:

- Pipelines and truck-trailers at about 1000 psi have a distinct advantage over trucks at 300 psi for all distances considered.
- Trucks at 1000 psi and ordinary temperatures ($90^{o}F$) have an advantage over lateral pipelines up to distances of 100 miles.
- Refrigerated trucks at low pressure (300 psi, $0^{o}F$) have an advantage over unrefrigerated low pressure trucks (300 psi, $90^{o}F$) due to a great change in density of $CO_2$ as a result of a phase change. Total cost of processing plus transport (100 miles) for an unrefrigerated low pressure truck is the highest (see table 9).

## Conclusions on $CO_2$-EOR Reserves

If natural gases containing less than 10 mole percent $CO_2$ are processed, and trucks/trailers at 1000 psi and $90^{o}F$ are used to transport it (a distance of 100 miles), the $CO_2$ can be made available at the point of consumption for about $1.15/MSCF. The above cost includes all of the purification cost for 100 MMSCFD of natural gas containing 7.5 mole % $CO_2$.

- Considering only the natural gases from the Western Overthrust Belt and the Tuscaloosa Trend, the $CO_2$ reserves could be increased about 14 trillion standard cubic feet.

B. Recommendations

It is recommended that DOE include the following in their future programs:

- Experimental work determining vapor-liquid equilibrium ratios for mixtures of methane-carbon dioxide at temperatures and pressures encountered in processing the natural gas.
- Experimental work determining other physical properties, such as heat capacity, conductivity, compressibility, viscosity, etc. for $CH_4$-$CO_2$ mixtures containing other impurities, such as $H_2S$ or others which exist in natural gases in the United States.
- High pressure truck/trailers should be studied in greater detail to determine an optimum pressure considering practical construction materials plus transportation rules and regulations.

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Fig. 1 GAS TREATING FOR $CO_2$ REMOVAL

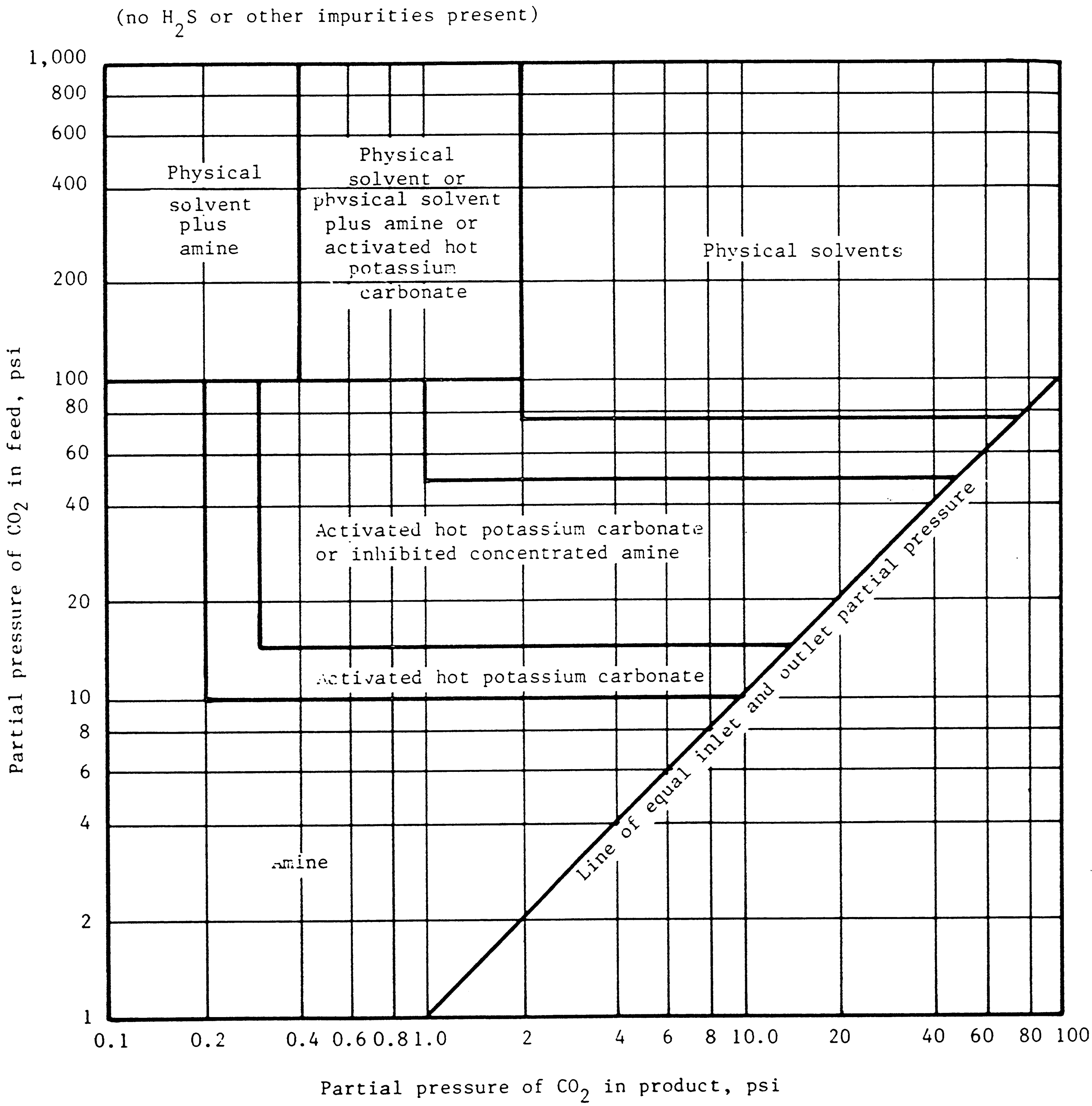
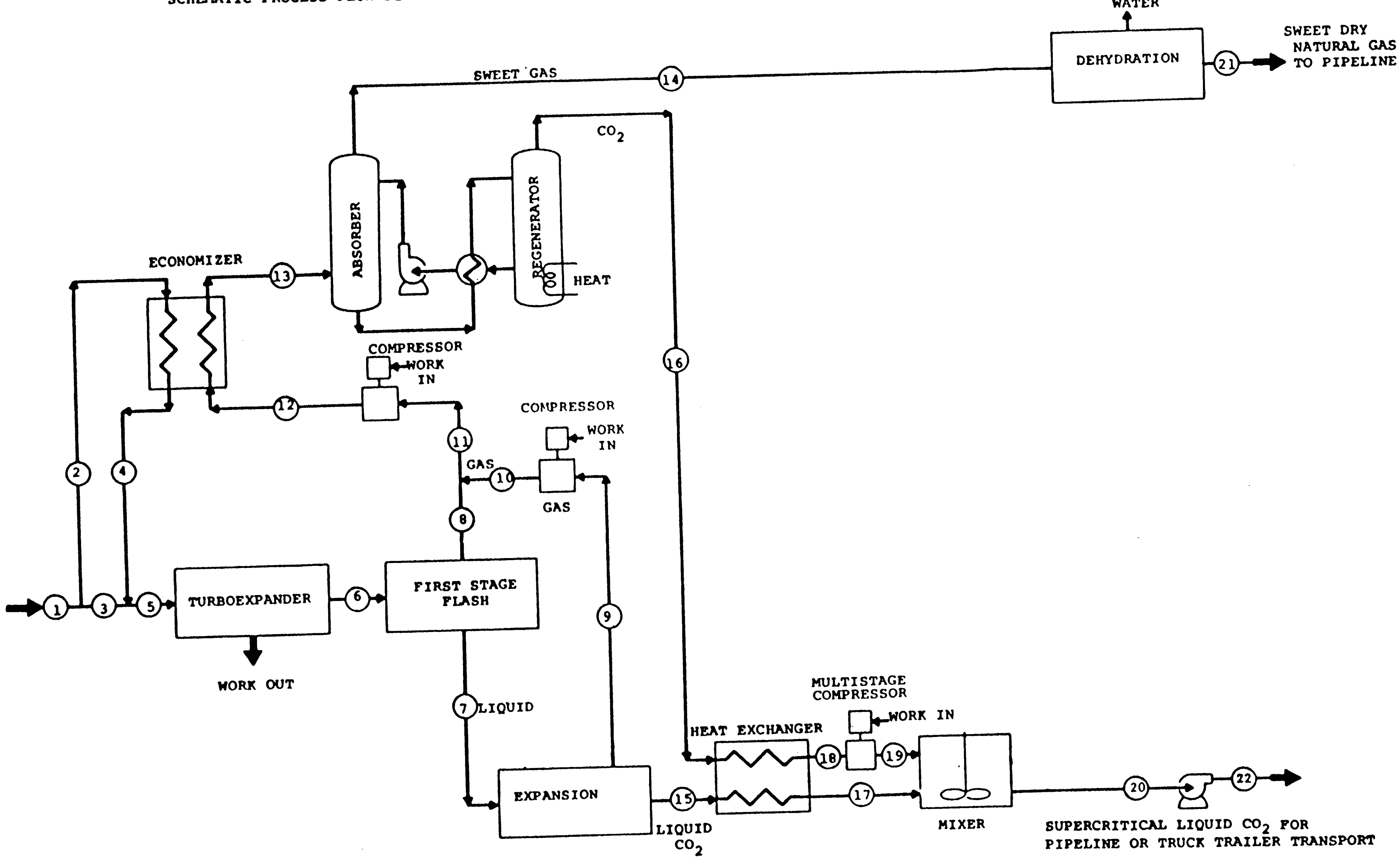


Fig. 2

SCHEMATIC PROCESS FLOW DIAGRAM FOR NATURAL GAS STREAMS CONTAINING GREATER THAN 30 MOLE PERCENT $CO_2$




| Description / Stream No. | | | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 | 13 | 14 | 15 | 16 | 17 | 18 | 19 | 20 | 21 | 22 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Amount $CH_4$ | Gas | MMSCF/D | 50 | 50 | 0 | 50 | 50 | 50 | 0 | 19 | 24 | 24 | 43 | 43 | 43 | 43 | 0 | 0 | 0 | 0 | 0 | 0 | 43 | 7 |
| | Liq | Gallons/Hour | 0 | 0 | 0 | 0 | 0 | 0 | 1601 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 509 | 0 | 652 | 0 | 0 | 652 | 0 | 493 |
| Amount $CO_2$ | Gas | MMSCF/D | 50 | 50 | 0 | 50 | 50 | 50 | 0 | 8.7 | 12 | 12 | 20.7 | 20.7 | 20.7 | 0 | 0 | 20.7 | 0 | 20.7 | 0 | 0 | 0 | 0 |
| | Liq | Gallons/Hour | 0 | 0 | 0 | 0 | 0 | 0 | 33780 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 19863 | 0 | 20318 | 0 | 14333 | 34651 | 0 | 34552 |
| Temperature, °F | | | 110 | 110 | 0 | 90 | 90 | -30 | -30 | -30 | -30 | 11 | -10 | 5 | 52 | 100 | -30 | 400 | 55 | 45 | 55 | 55 | 110 | 65 |
| Pressure, Psi | | | 2093 | 2093 | 0 | 2088 | 2088 | 1000 | 1000 | 1000 | 750 | 1000 | 1000 | 1110 | 1100 | 1090 | 750 | 15 | 745 | 15 | 750 | 750 | 1080 | 1000 |

Fig. 3 $50\%\ CH_4 - 50\%\ CO_2$

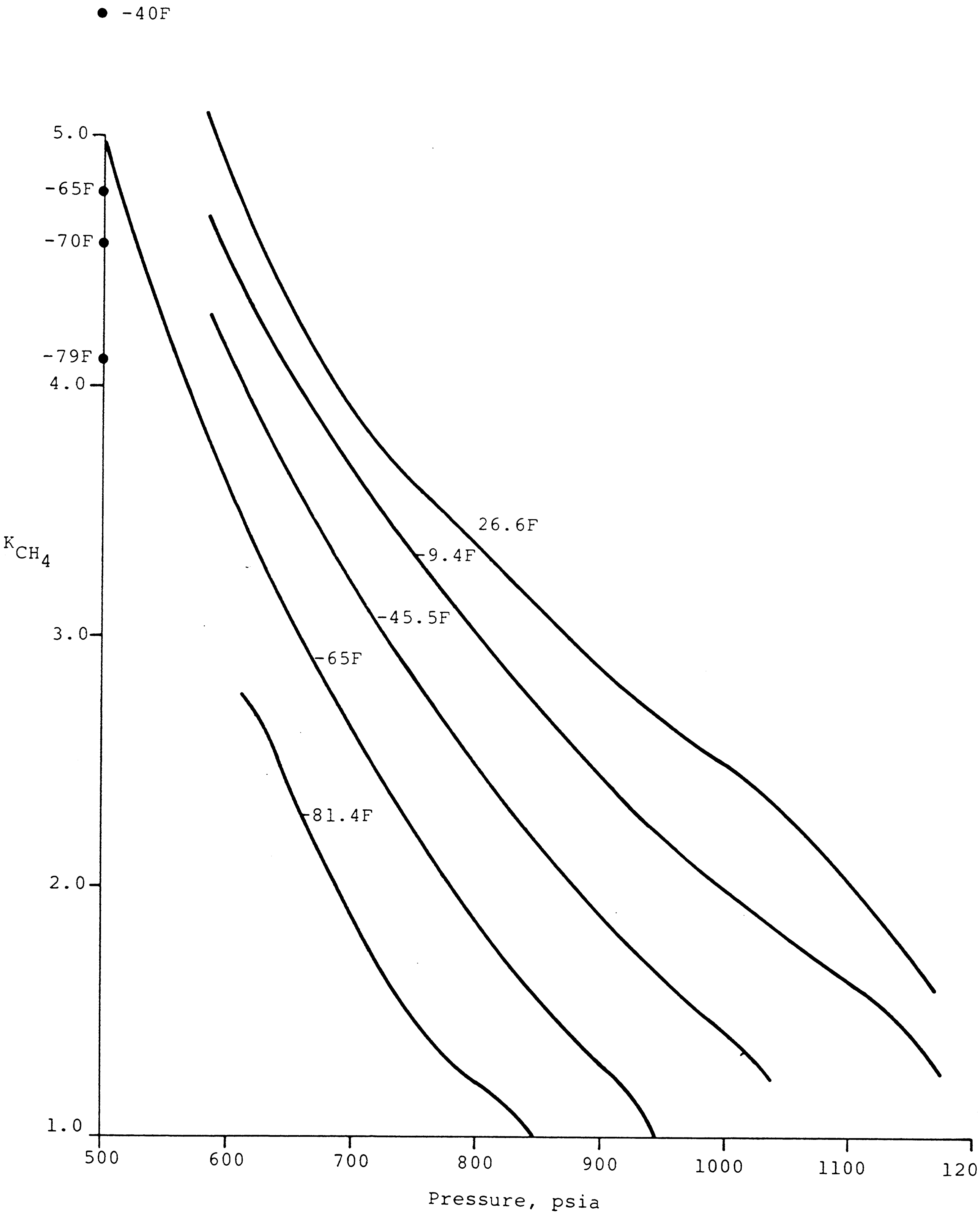
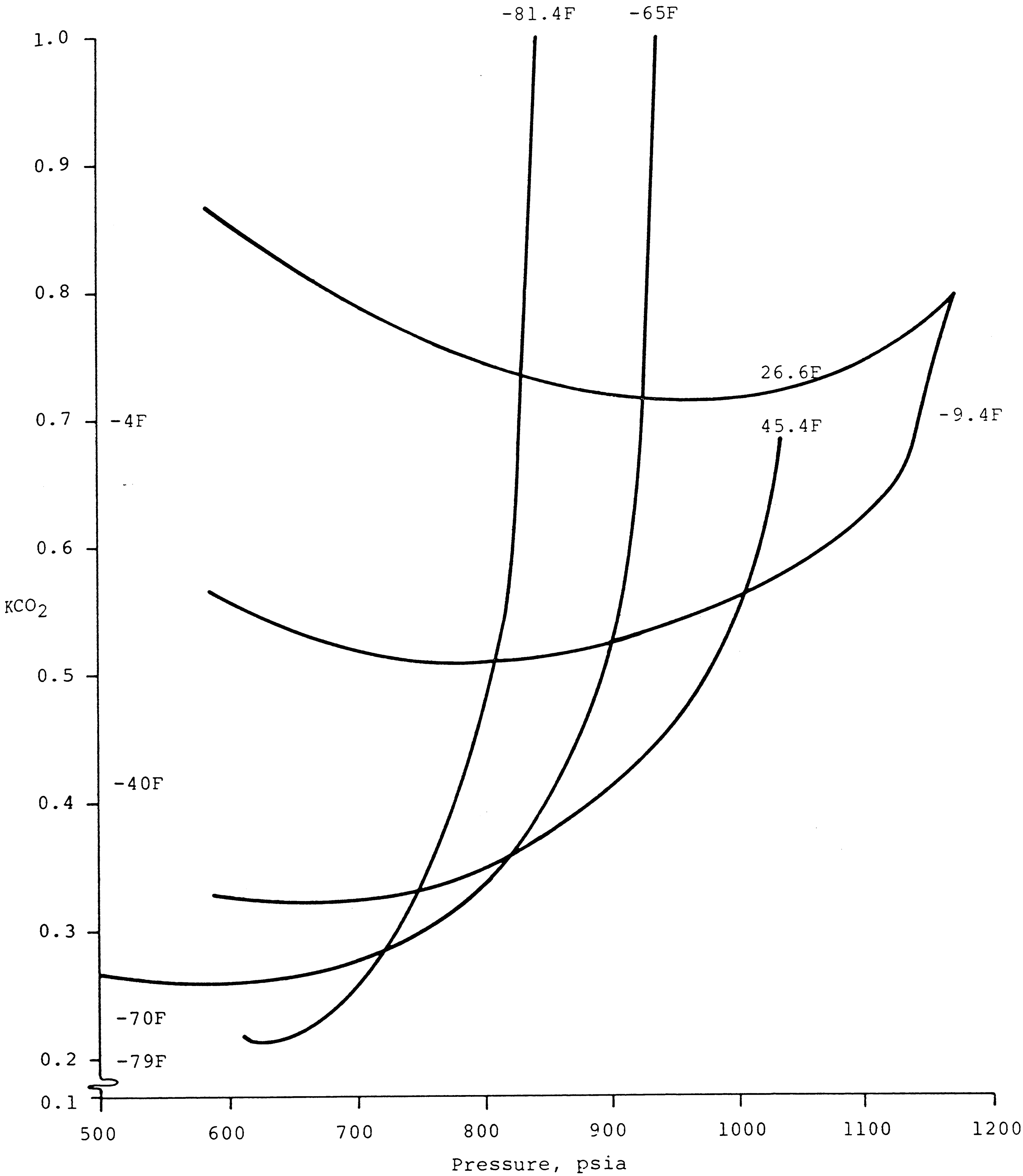


Fig. 4 50% $CH_4$ - 50% $CO_2$



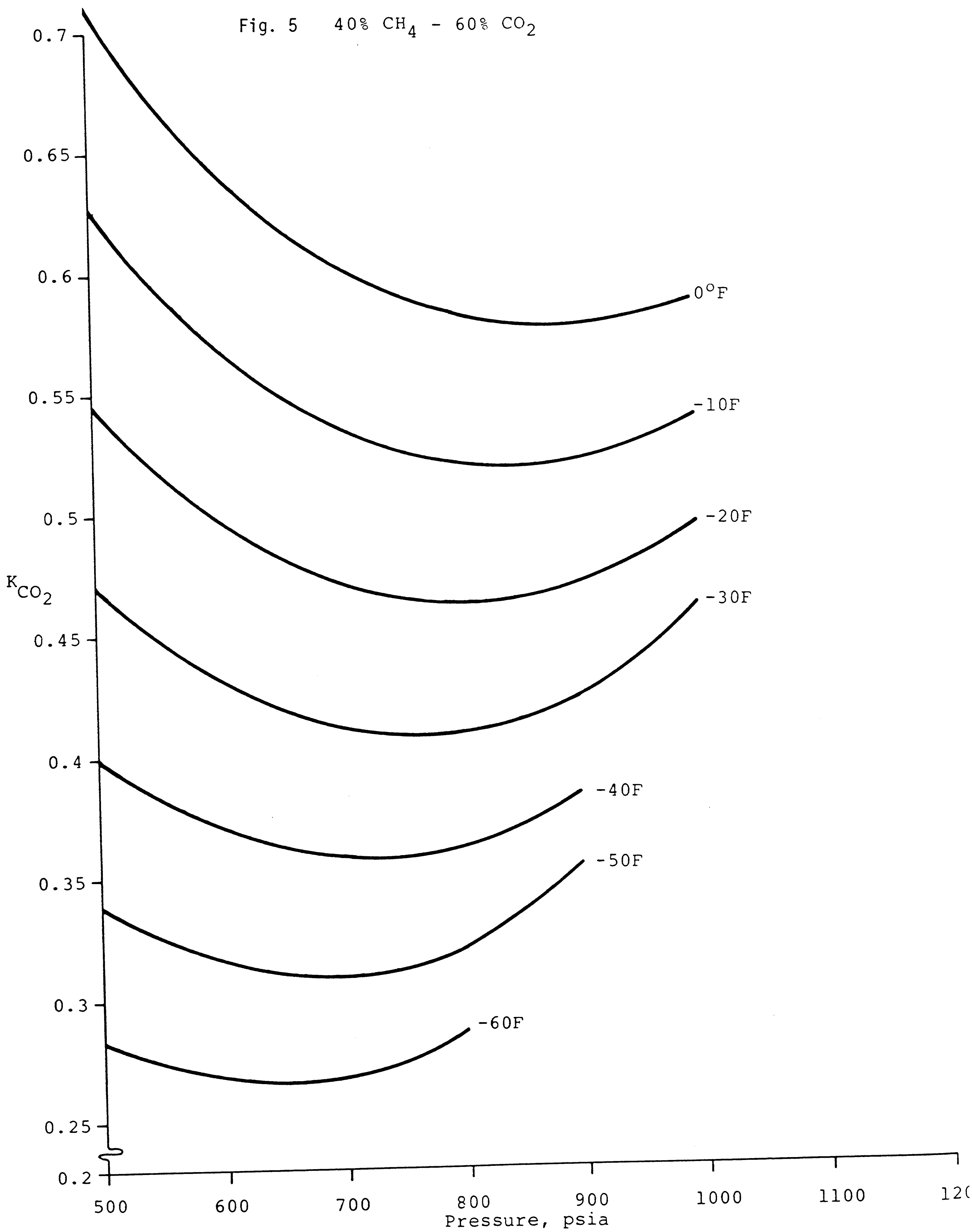


Fig. 5 40% $CH_4$ - 60% $CO_2$
$K_{CO_2}$
0.7
0.65
0.6
0.55
0.5
0.45
0.4
0.35
0.3
0.25
0.2
0°F
-10F
-20F
-30F
-40F
-50F
-60F
500
600
700
800
900
1000
1100
Pressure, psia

Fig. 6 40% $CH_4$ - 60% $CO_2$

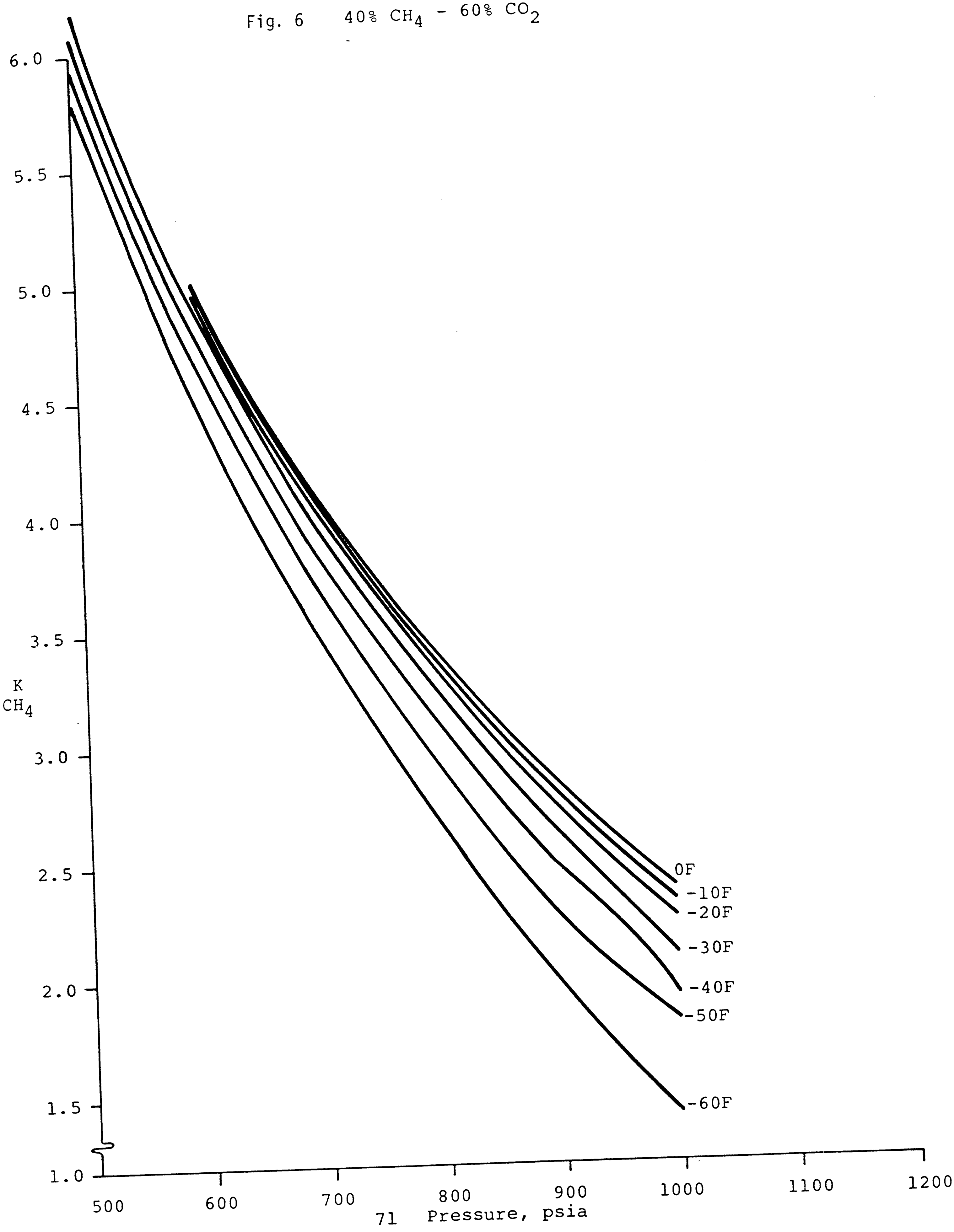


Fig. 7 $-50^{o}F$ for $CH_4$

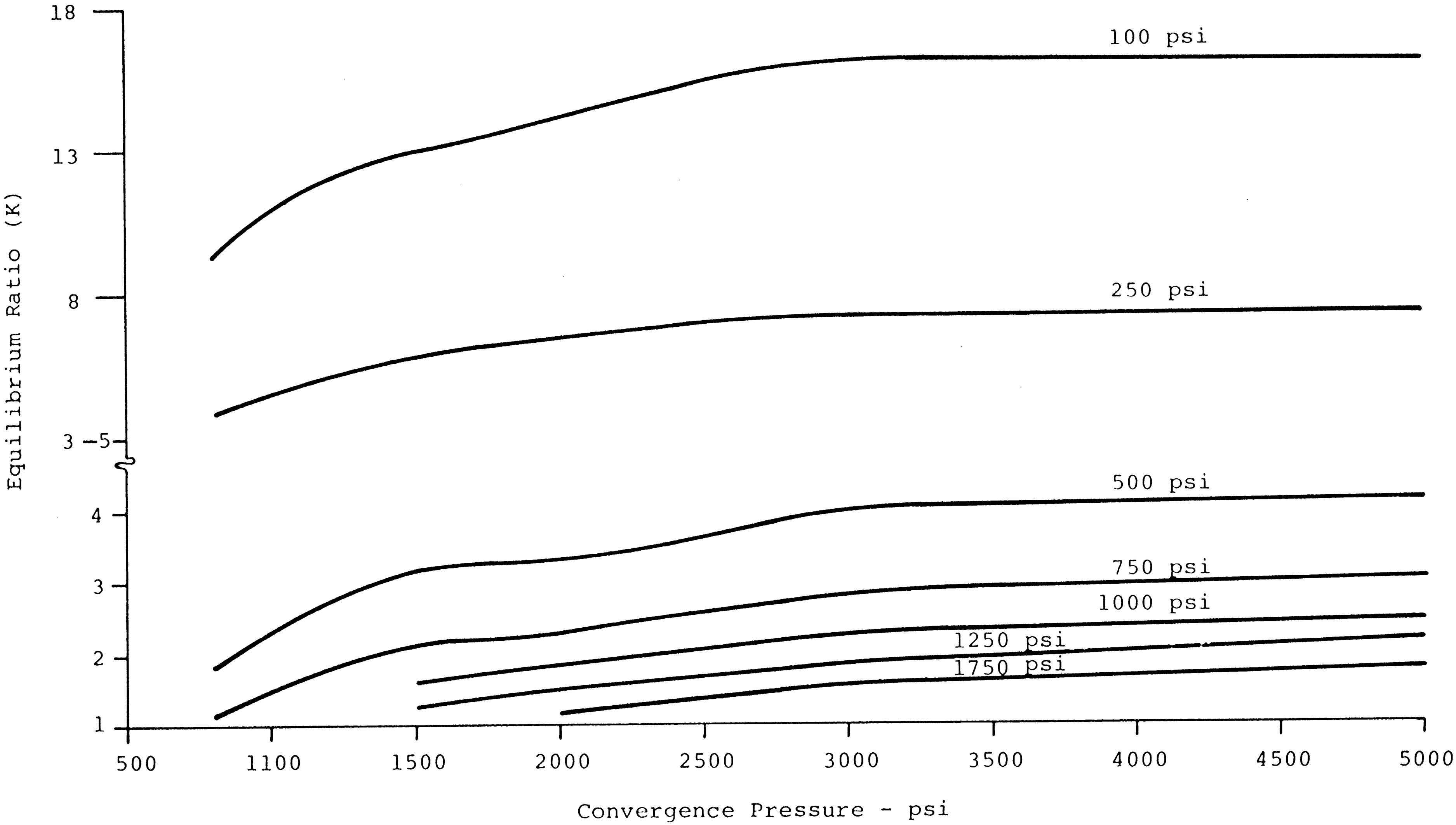


Fig. 8 $0^{o}F$ - K for $CH_4$

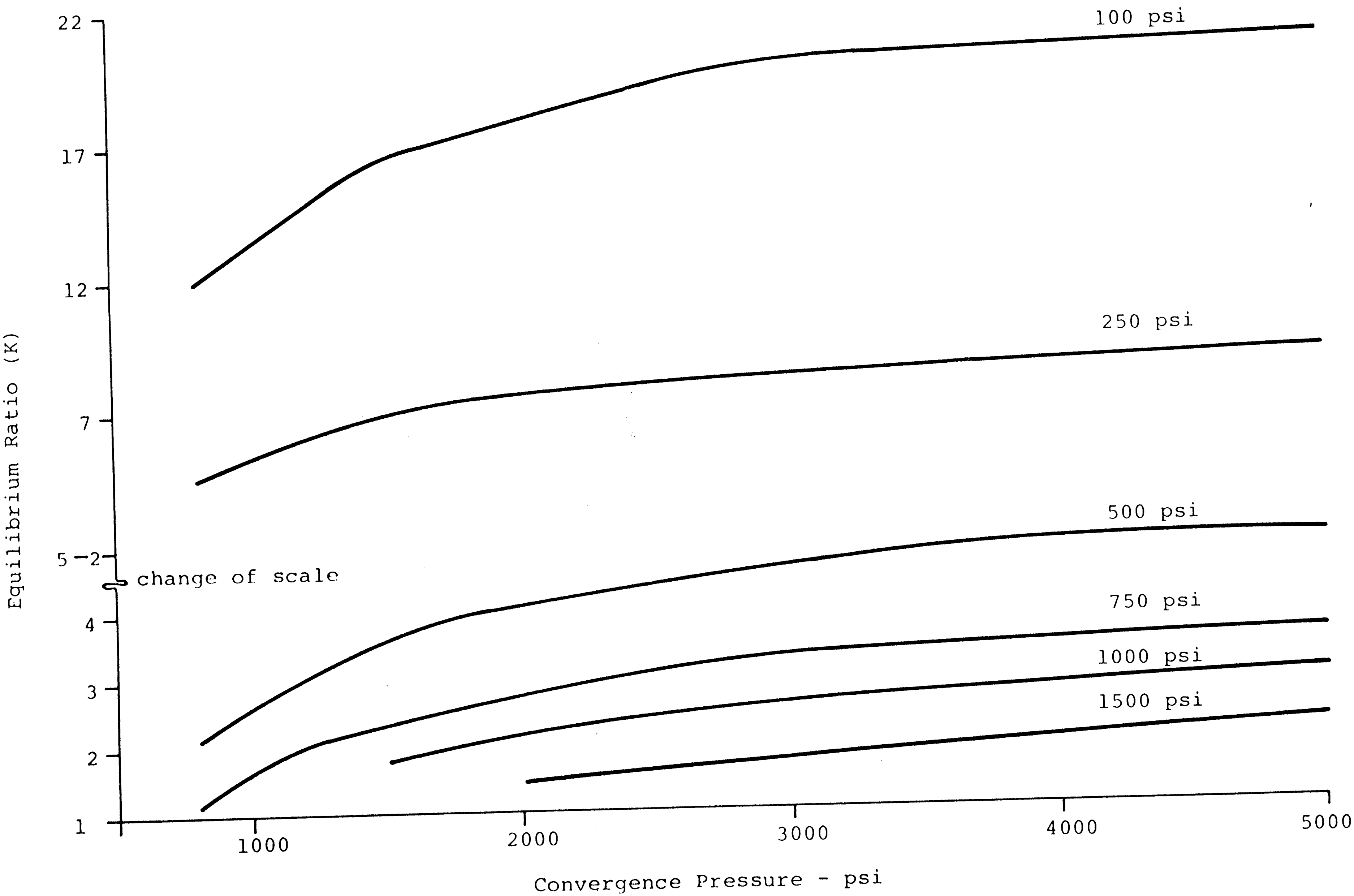


Fig. 9 +50F for $CH_4$

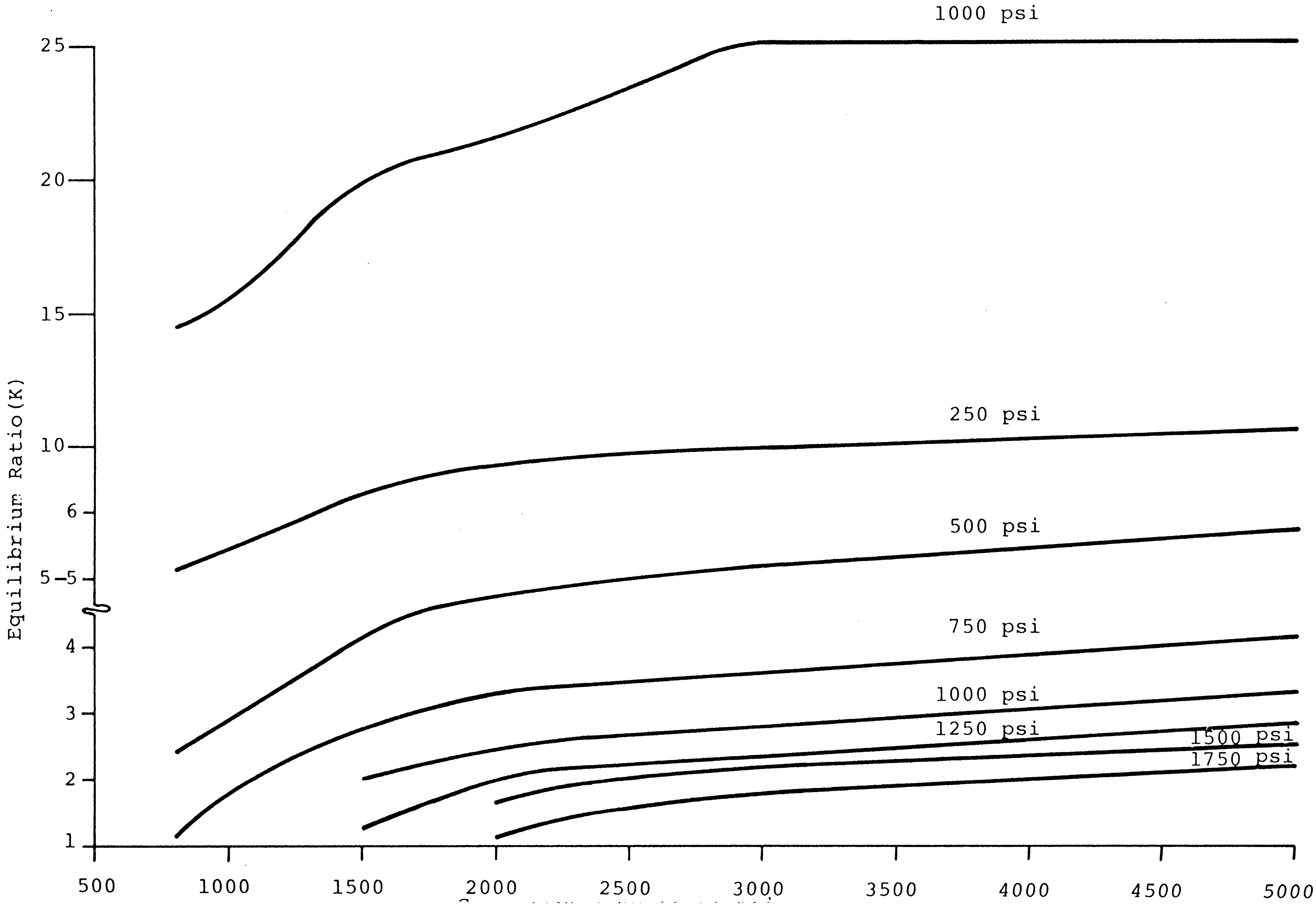


Fig. 10 Critical Pressure vs. Critical Temperature for Mixtures of Methane and Carbon Dioxide

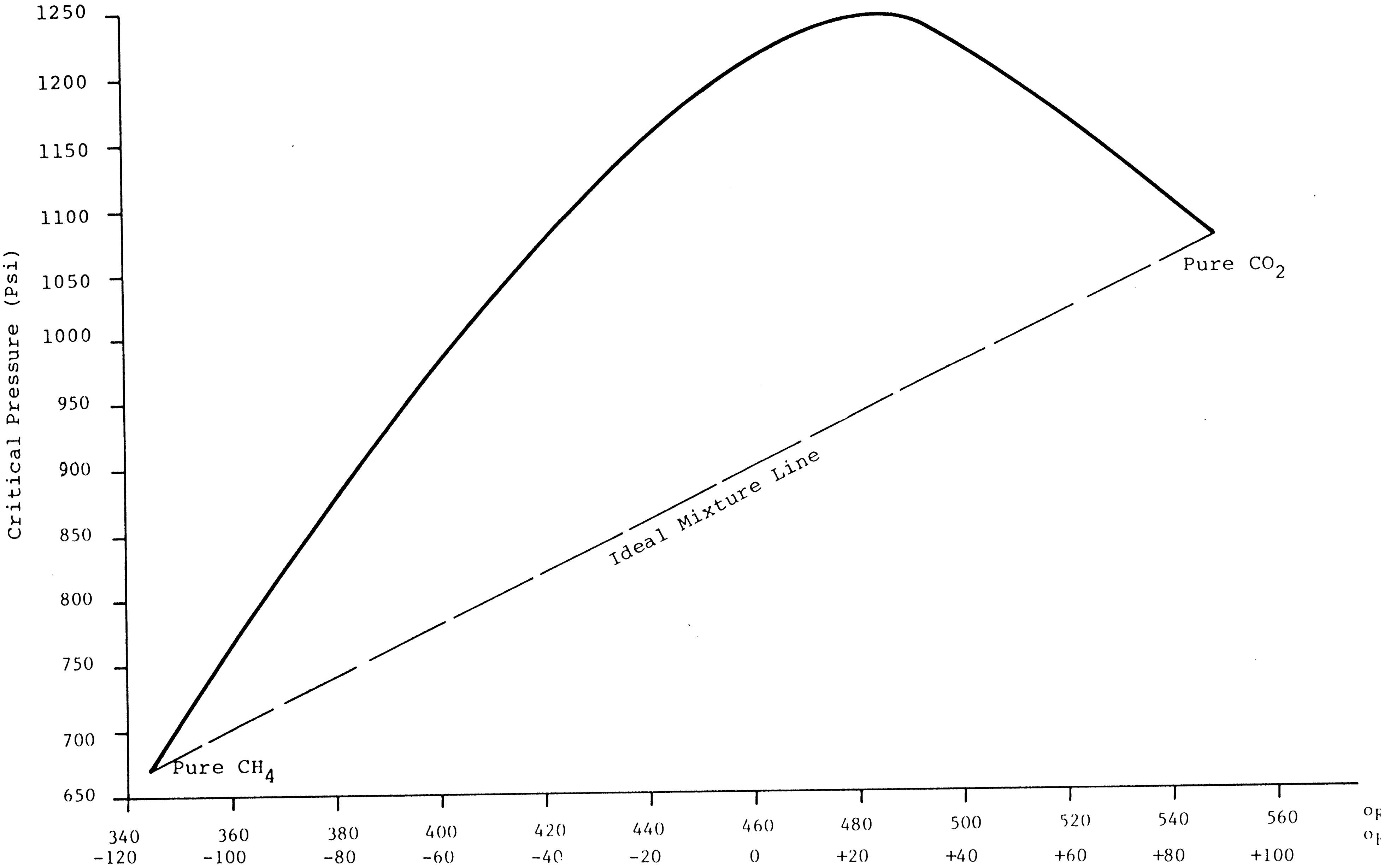


Fig. 11 CARBON DIOXIDE THERMODYNAMIC PROPERTIES PRESSURE-ENTHALPY (MOLLIER) DIAGRAM

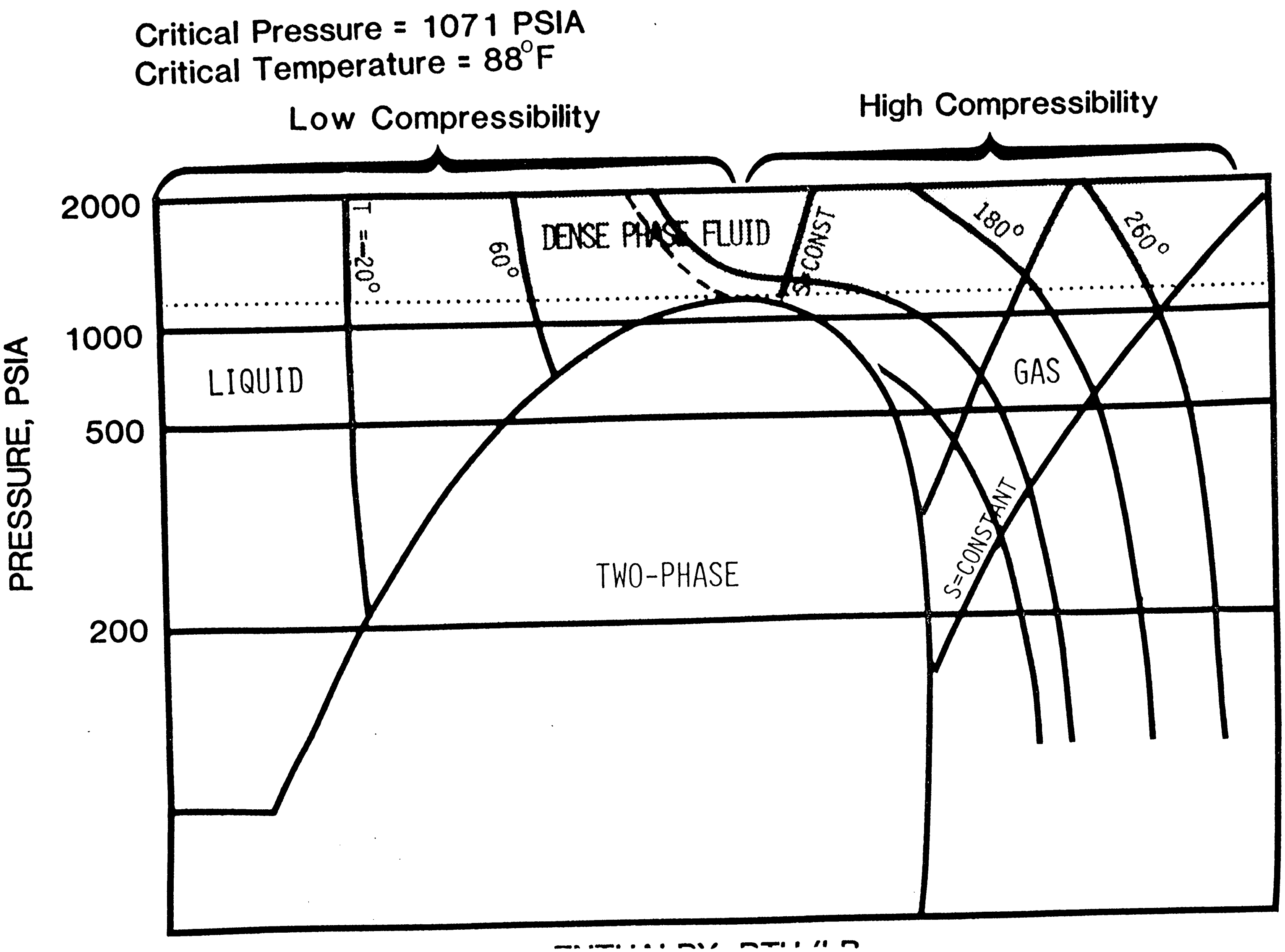


Fig. 12 INITIAL PRESSURIZATION OF $CO_2$

Thermodynamic Comparison

Compressor/Pump Work = $\Delta h_1 + \Delta h_2$

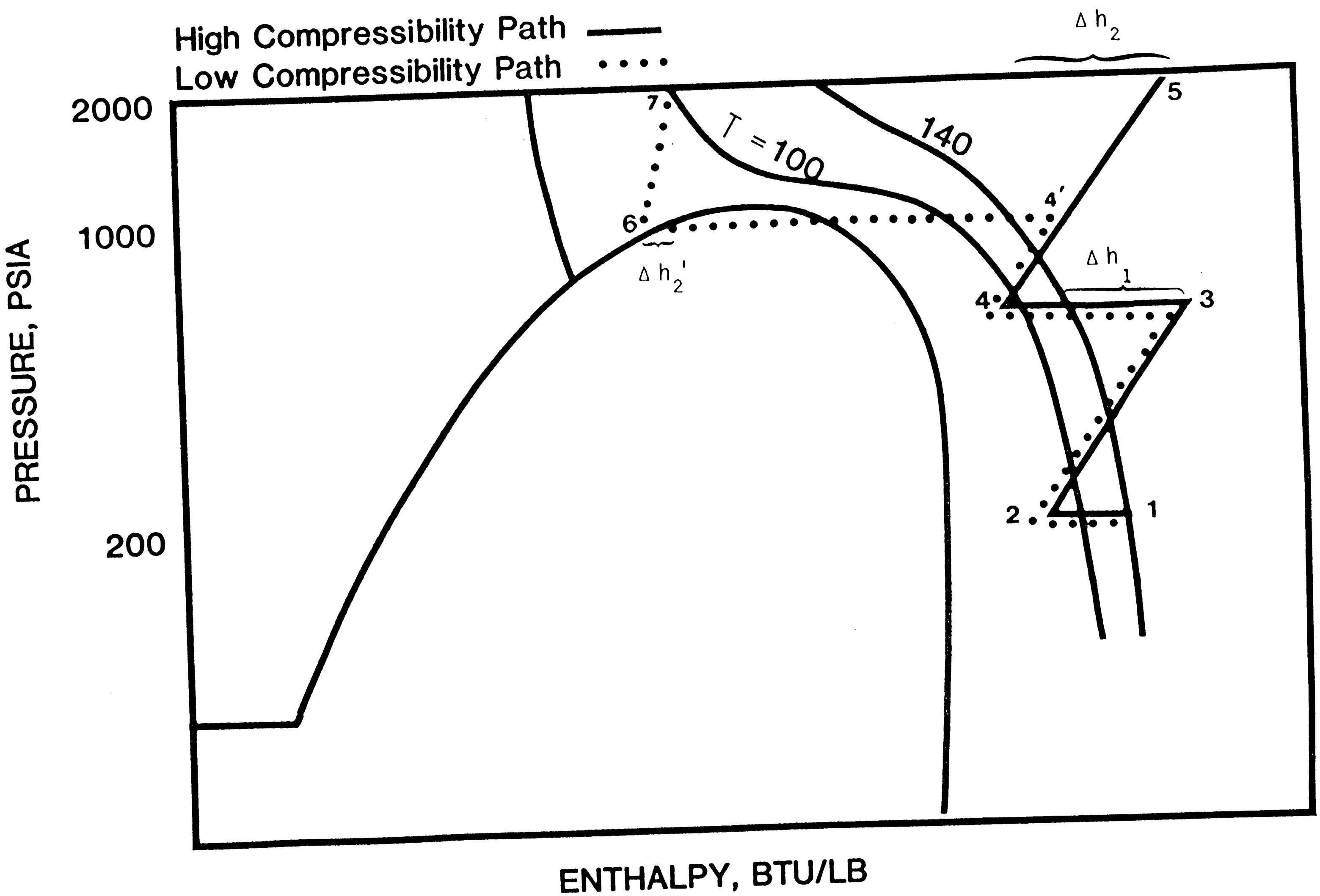


Fig. 13 INITIAL PRESSURIZATION OF $CO_2$

COMPRESSOR HORSEPOWER COMPARISON

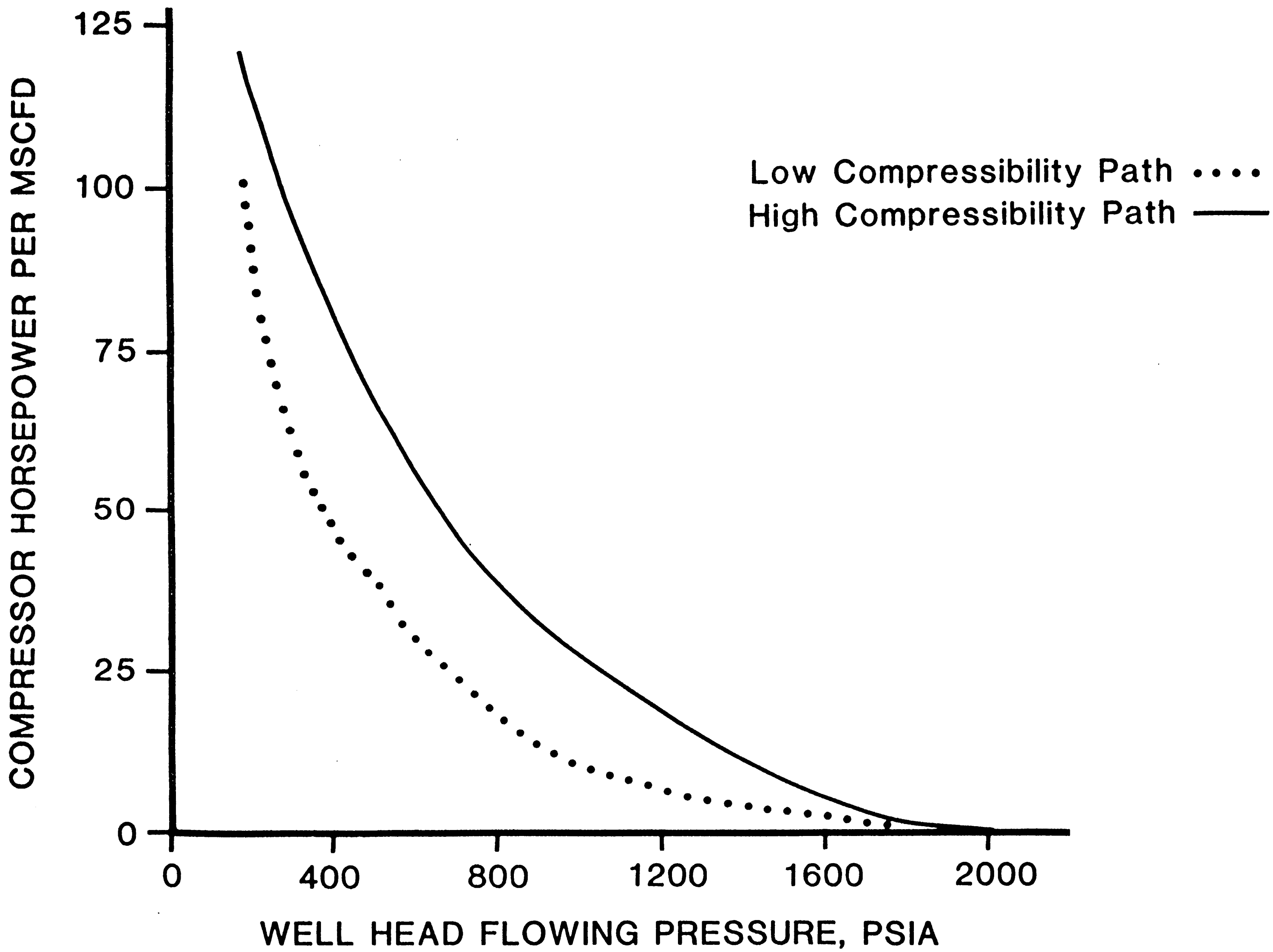


Fig. 14 TEMPERATURE DISTRIBUTION IN 50 MMSCFD $CO_2$ PIPELINES

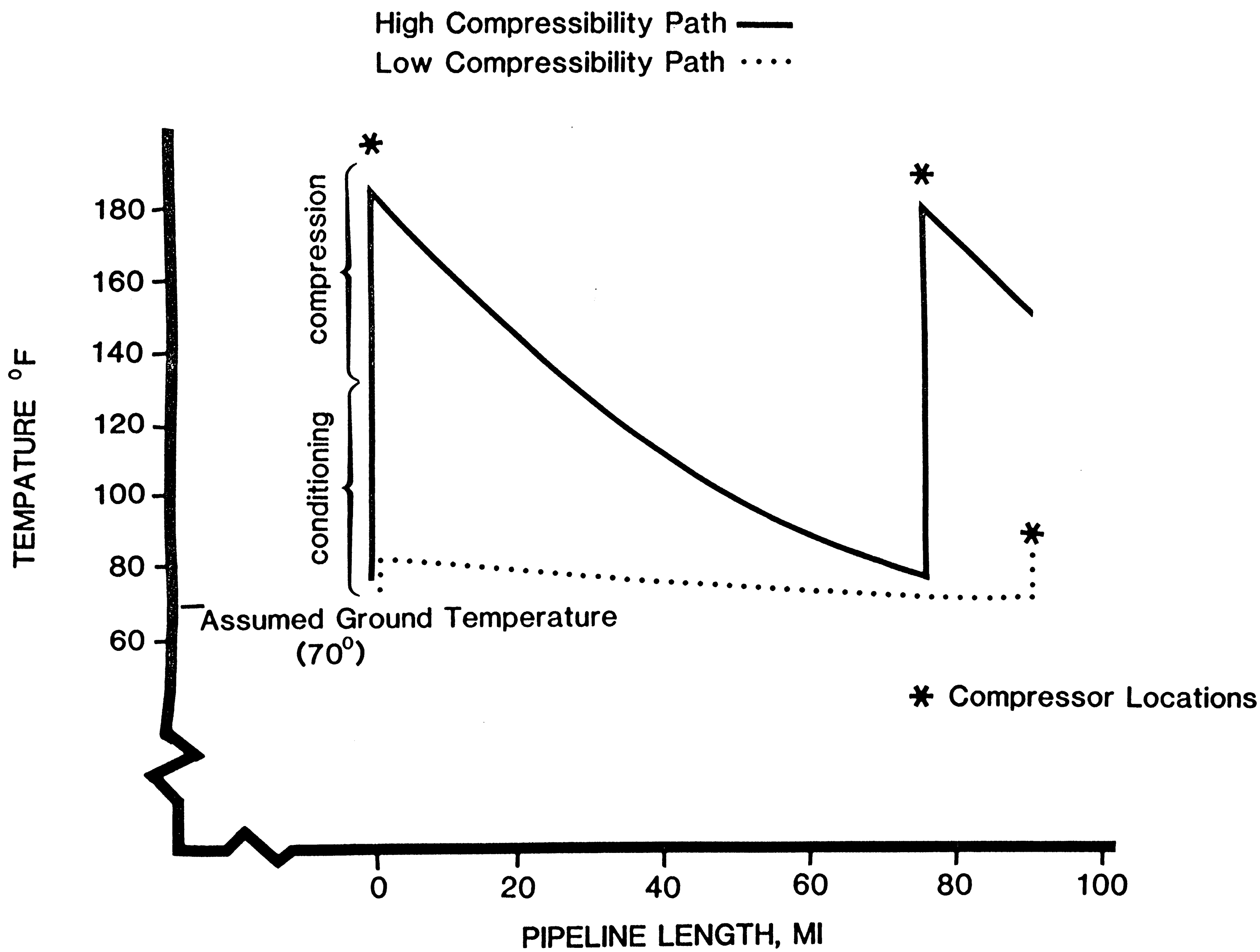


Fig. 15 **PIPELINE REPRESSURIZATION & THERMAL CONDITIONING OF $CO_2$**

**Thermodynamic Comparison**

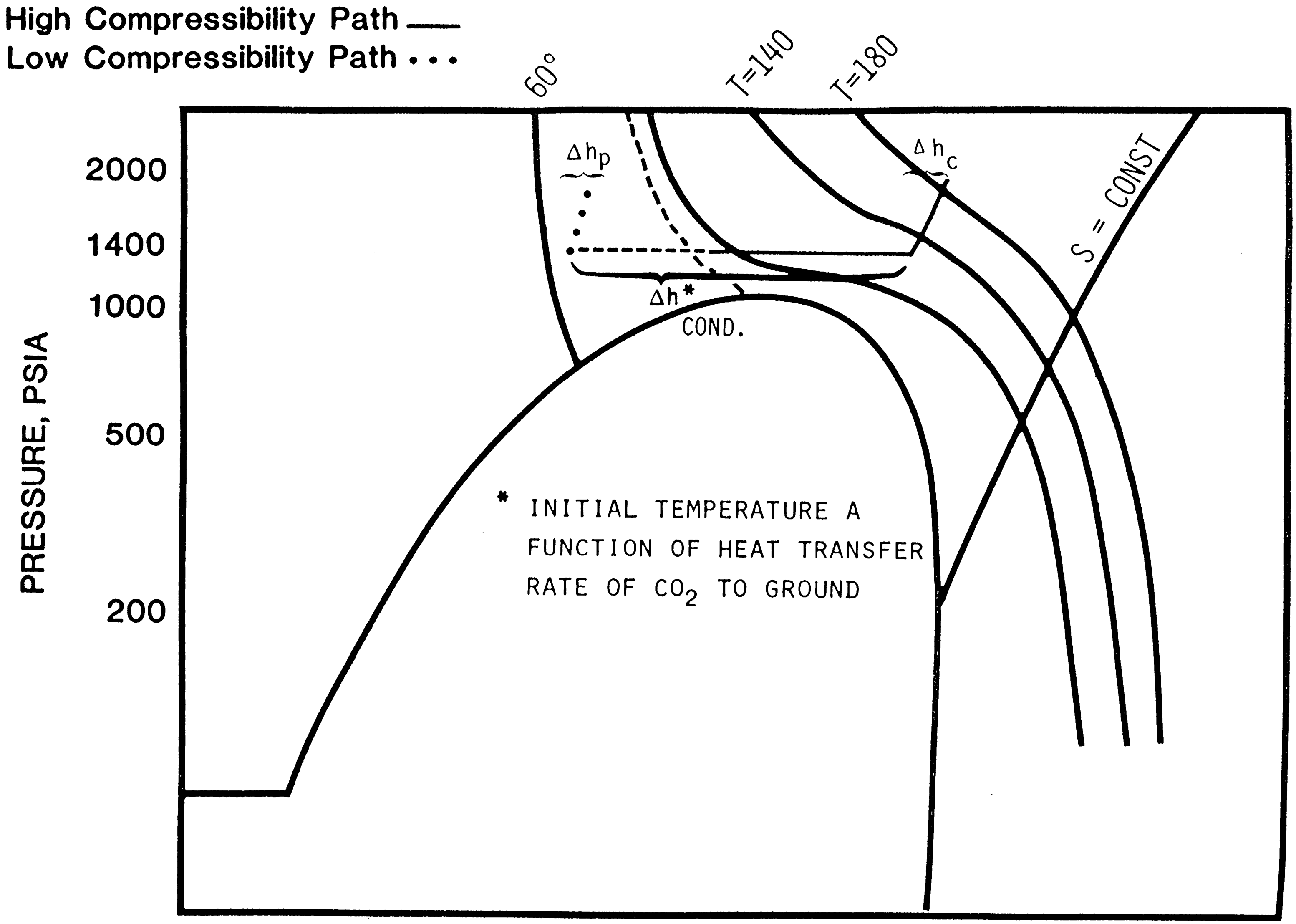
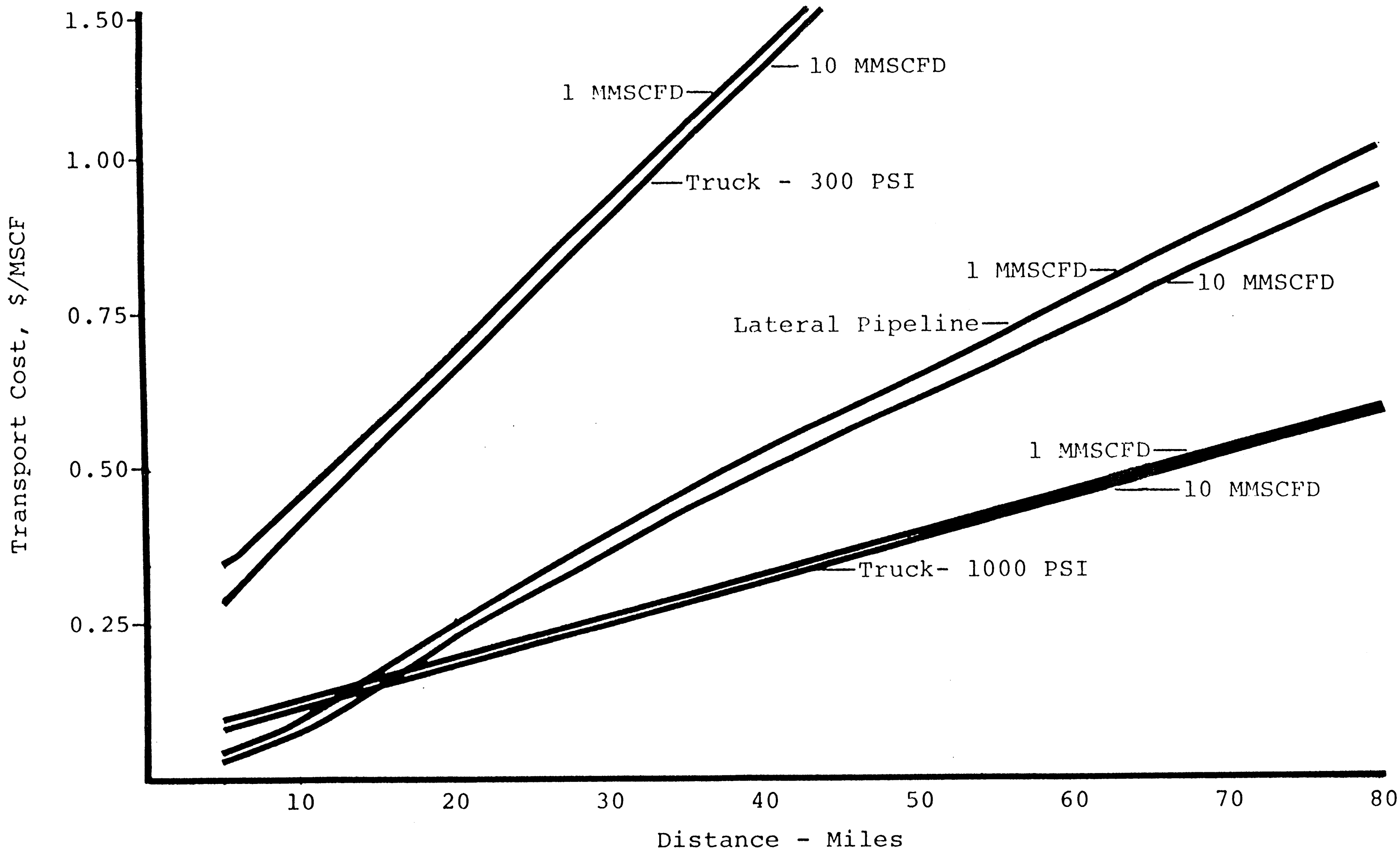


Fig. 16 COMPARISON OF TRUCK AND LATERAL PIPELINE
$CO_2$ TRANSPORT COST



SUMMARY OF $CO_2$ SUPPLY PROJECTS

| | Operator | Origin | $CO_2$ Source | Capacity (MMSCFD) | Terminus | Field Served |
|---|---|---|---|---|---|---|
| | SACROC | Terrell & Pecos Counties, TX | Val Verde Basin Natural Gas (18 to 53% $CO_2$) | 240 | McCamey (Upton Ct) & Snyder (Scurry Ct.), TX | North Crosset & Kelly-Snyder |
| 2 | Shell | Dolores & Montezuma Ct, CO | McElmo Dome & Doe Canyon $CO_2$ Fields (96-99% $CO_2$) | 400 | Denver City (Yoakum Ct.) TX | Wasson/Denver Unit |
| 3 | ARCO | Huerfano Ct, CO | Sheep Mt & Dike Mt $CO_2$ Fields (97-99% $CO_2$) | 300 | Denver City (Yoakum Ct) & Seminole (Gains Ct) TX | Wasson/Willard Unit & Seminole |
| 4 | AMACO | Union & Harding Counties, N.M. | Brovo Dome Unit $CO_2$ Fields (99-100% $CO_2$) | ? | West Texas | Probably: North Cowden Slaughter and Leveland |
| 5 | SHELL | Rankin Ct, MISS | Jackson Dome $CO_2$ Field (98-99% $CO_2$) | 200 | Lincoln & Pike Cts, MISS (&beyond) | Little Creek |
| 6 | TRANS-PETRO | Borger (Hutchinson Ct), TX | Cominco, Inc. Anhydrous Ammonia Plant Vent | 27 | Ochiltree & Hansford Cts., TX | Marmaton |
| 7 | CITY SERV. ARCO | Garfield Ct, OK | Farmlands Ind Anhydrous Ammonia Plant Vent | 35 | Garvin Ct., OK | E.Velma & N.E.Purdy |
| 8 | COLUMBIA GAS | Kanawa Ct., W VA | Indian Creek Natural Gas/$CO_2$ Field (66% $CO_2$) | 30 | ? | ? |
| 9 | CONOCO | Pecos Ct, TX | Val Verde Basin Natural Gas | 20 | Orla (Reeves & Culberson Cts.) TX | Ft. Geraldine |

TABLE 1

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Table 2 Comparison of Some Commonly Used Gas Purification Processes Considering Limitations on the Feed and the Product Streams

| Process/Type | Allowable Ranges of Acid Gas Concentrations in the Feed | Feed Conditions Allowable | Acid Gas Concentrations in Product | Product Conditions | Acid Gas Conditions | Special Considerations |
|---|---|---|---|---|---|---|
| Fluor Econamine Process, Diglycolamine (DGA) Sorbent | Usually 2 - 5% total acid gas - been used to 11.41% $CO_2$ with nil $H_2S$ and 31.7% $H_2S$ with 3.9 $CO_2$ | < 160°F with no limit on pressure with previous design flow rates running from 3 to 550 MMSCFD | 0.01 mole % $CO_2$, 0.25 grains/100 SCF $H_2S$ capable of 4 ppmv $H_2S$ and 50 ppmv $CO_2$ | 90 - 220°F pressure dependent on feed, etc. | Near atmospheric pressure ambient temp. | Will remove other organic sulfurs |
| Fluor Solvent Process, Propylene Carbonate Sorbent | Acid gas partial pressure > 75 psi | Feed gas pressure whatever, ambient or below temperatures | < .25 grains $H_2S$/100SCF<br>< 2% $CO_2$ | Subambient temperatures and atmospheric pressures | ~25°F, Atmospheric pressure | Can be used for selective removal $H_2S$ from $CO_2$, $H_2S$; also will remove other organic sulfur.<br>Inert gas is injected in the stripper diluting acid gas |
| Benfield Processes - Hot Potassium Carbonate & High Pure Modification | 5 - 50% acid gas max to date 12% $H_2S$ + 42% $CO_2$ | Moderate to high pressures, (10 - 225 MMSCFD, 100 - 2000 psig, ambient to 400°F. Highest acid concentrations | Hot Carbonate<br>$CO_2$ < 500 ppm<br>$H_2S$ < 1 ppm<br>Hi Pure<br>$CO_2$ < 20 ppm<br>$H_2S$ < 1 ppm | 100-200 psig pressure & temperature - ambient to 400°F | Ambient Conditions | Will handle substantial $H_2O$ & heavy hydrocarbons selective $H_2S$ absorption capability |
| Union Carbide - Molecular Sieve | 0.1 - 50% $CO_2$<br>< 1 - 400 grains $H_2S$/100SCF | 110 - 115°F + 200 - 260 psig or greater | < 50 ppmv $CO_2$<br>< ¼ grain $H_2S$/100 SCF | 1 - 2 psi less than inlets and same temperature | Low pressure, probability but not necessarily atmospheric Acid gas at elevated temps. | - Regeneration gas is required so $CO_2$ is not recovered in a pure stream<br>- Molecular sieves can remove water & heavy hydrocarbons as well |
| Shell ADIP Diisopropanolamine (DIPA) | High concentrations of acid gas,<br>< 1 to > 50% $H_2S$<br>< 1 to > 50% $CO_2$ | 100 - 140°F usually <100°F psia | < 50 ppmv $H_2S$ to <1 ppmv $H_2S$ as desired variable $CO_2$ recovery generally proportionally small | 40 psi, 110°F | Ambient Conditions | Selective absorption of $H_2S$ even in very high concentrations of $CO_2$ |
| Shell Sulfinol Sulfolane (tetrahydrothiophenedioxide) & alkonolamine | < 1 to > 50% $H_2S$<br>< 1 to > 50% $CO_2$ | ≈ 110°F & 1000 psia typically pressure can vary from 22 to >1330 psia | < 1 ppmv $H_2S$+COS<br>< 50 ppmv $CO_2$ usually < 2-3% achievable | 1000 psia & 113°F | Ambient Conditons | Can be used for selective removal of $H_2S$ |
| Selexol - Allied Chemical, dimethyl ether of polyethelene | Typical case 30% $CO_2$ 20% $H_2S$ concentrations not critical | 1000 psig, ambient or lower temperature | < 2 ppm $CO_2$<br>< ¼ gr/100 scf | Ambient temp. 1000 psig | Ambient Conditions | Energy efficient, can be somewhat selective |

Table 3 Composition, Heating Value, Specific Gravity, Wobbe Number of Natual Gas Samples from Various States, Counties, and Fields in U.S.A.

| Component | Gas Sample Number | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | 1 | 2 | 3 | 4 | 5 | 6 | 7 | 8 | 9 | 10 | 11 | 12 |
| Methane | 18.9 | 30.3 | 40.5 | 49.7 | 79.4 | 93.6 | 83.8 | 73.1 | 66.7 | 68.4 | 60.2 | 26.2 |
| Ethane | 0.2 | 1.8 | 2.4 | 3.3 | 0.8 | 2.5 | 8.9 | 7.7 | 12.5 | 7.9 | 15.8 | 15.8 |
| Propane | 0.1 | 0.9 | 1.2 | 1.1 | 0.2 | 1.2 | 2.8 | 6.6 | 8.5 | 10.4 | 12.0 | 26.0 |
| N-Butane | -- | 0.3 | 0.4 | 0.2 | T | 0.3 | 0.6 | 2.6 | 4.3 | 4.0 | 3.7 | 14.2 |
| Isobutane | -- | 0.2 | 0.3 | 0.3 | 0.1 | 0.4 | 0.5 | 1.1 | 1.8 | 3.0 | 2.8 | 5.5 |
| N-Pentane | -- | T | T | T | T | 0.1 | 0.1 | 0.7 | 1.5 | 1.2 | 0.9 | 4.5 |
| Isopentane | -- | 0.1 | 0.2 | 0.1 | T | 0.1 | 0.3 | 1.2 | 1.4 | 1.6 | 1.8 | 3.4 |
| Cyclopentane | -- | T | T | T | T | 0.1 | T | 0.5 | 0.1 | 0.4 | 0.2 | 0.4 |
| Hexanes plus | -- | T | T | T | T | 0.2 | 0.2 | 0.8 | 0.5 | 1.0 | 1.4 | 1.6 |
| Nitrogen | 3.0 | 26.2 | 25.7 | 42.3 | 9.4 | 0.7 | 1.1 | 4.3 | 1.0 | 0.4 | 0.3 | 2.1 |
| Oxygen | 0.4 | -- | -- | -- | -- | -- | -- | 0.4 | 0.1 | 0.1 | T | 0.3 |
| Argon | -- | 0.1 | 0.1 | 0.2 | 0.1 | -- | -- | T | T | -- | -- | -- |
| Hydrogen | -- | 0.4 | 0.3 | -- | T | -- | -- | -- | T | -- | -- | -- |
| Carbon Dioxide | 77.3 | 0.1 | 0.5 | T | 9.7 | 0.9 | 1.7 | 0.9 | 1.6 | 1.8 | 0.9 | T |
| Helium | -- | 39.5 | 28.4 | 2.8 | 0.2 | T | -- | -- | -- | T | T | T |
| Heating Value Btu/Scf | 198 | 385 | 518 | 612 | 827 | 1070 | 1145 | 1315 | 1479 | 1526 | 1618 | 2323 |
| Spec. Gravity | 1.319 | 0.521 | 0.582 | 0.784 | 0.692 | 0.589 | 0.674 | 1.037 | 0.867 | 0.879 | 0.937 | 1.378 |
| Wobbe Number | 172 | 533 | 679 | 691 | 994 | 1394 | 1395 | 1291 | 1588 | 1628 | 1672 | 1979 |
| State | Alaska | Texas | Texas | Kansas | Colorado | Louisiana | Colorado | Colorado | Texas | Texas | Oklahoma | Oklahoma |
| County | Kemai Pen Burough | Potter | Potter | Kiowa | Mesa | Plaque-nines Parish | Morgan | Morgan | Colorado | Colorado | Custer | Lincoln |
| Field | Not Given | Cliff-side | Cliff-side | Nullin-vilie | Wildcat | Bastian Bay SE | Wildcat | Renegade | Third Creek | Name Le | Weather-ford | Fallis |

Table 4 TRUCK LOADING @ 300 PSI AND 90$^{o}$F

| | | Compression | | | Cooling | | |
|---|---|---|---|---|---|---|---|
| Stage Number | Description | Power (HP-hr) | Capital Cost ($1000) | Operating Cost ($1000/yr) | Heat Removed MMBtu/hr | Capital Cost $1000 | Operating Cost ($1000/yr) |
| 1 | Boosting press. to 50 psi | 957 | 574 | 175.6 | -- | -- | -- |
| 2 | Cooling down to 140$^{o}$F | -- | -- | -- | 1.85 | 10.75 | 1.38 |
| 3 | Boosting press. to 150 psi | 694 | 416 | 127.3 | -- | -- | -- |
| 4 | Cooling down to 140$^{o}$F | -- | -- | -- | 1.45 | 10.75 | 1.38 |
| 5 | Boosting press. to 300 psi | 465 | 280 | 85.4 | -- | -- | -- |
| 6 | Cooling down to 90$^{o}$F. | -- | -- | -- | 1.38 | 9.10 | 1.20 |
| | TOTAL | 2116 | 1270 | 388.3 | 4.68 | 30.6 | 3.96 |

TOTAL (Compression + Cooling) Capital Cost ($1,301,000) Annual Operating Cost ($392,300)

NOTE: Operating cost consists of fuel cost based on 340 days of operation per year

Table 5 TRUCK LOADING @ 300 PSI AND $0^{o}F$

| | | Compression | | | Cooling | | |
|---|---|---|---|---|---|---|---|
| Stage Number | Description | Power (HP/hr) | Capital Cost ($1000) | Operating Cost ($1000/yr) | Heat Removed MMBtu/hr | Capital Cost $1000 | Operating Cost ($1000/yr) |
| 1 | Boosting press. to 50 psi | 957 | 574 | 175.6 | -- | -- | -- |
| 2 | Cooling down to $140^{o}F$ | -- | -- | -- | 1.85 | 10.75 | 1.38 |
| 3 | Boosting press. to 150 psi | 694 | 416 | 127.3 | -- | -- | -- |
| 4 | Cooling down to $140^{o}F$ | -- | -- | -- | 1.45 | 10.75 | 1.38 |
| 5 | Boosting Press. to 300 psi | 465 | 280 | 85.4 | -- | -- | -- |
| 6 | Cooling down to $90^{o}F$ | -- | -- | -- | 1.38 | 9.10 | 1.20 |
| 7 | Refrigeration to $0^{o}F$ | -- | -- | -- | 5.45 | 2040 | 624 |
| | TOTAL | 2116 | 1270 | 388.3 | 10.13 | 2070.6 | 627.96 |
| | TOTAL (Compression + Cooling) | Capital Cost ($3,341,000) | | | Annual Operating Cost ($1,061,260) | | |

NOTE: Operating cost consists of fuel cost based on 340 days of operation per year

Table 6 TRUCK LOADING @ 1000 PSI AND 90°F

| | | Compression | | | Cooling | | |
|---|---|---|---|---|---|---|---|
| **Stage Number** | **Description** | Power HP-hr | Capital Cost ($1000) | Operating Cost ($1000/yr) | Heat Removed MMBtu/hr | Capital Cost ($1000) | Operating Cost ($1000/yr) |
| 1 | Boosting press. to 50 psi | 957 | 574 | 175.6 | -- | -- | -- |
| 2 | Cooling down to 140°F | -- | -- | -- | 1.85 | 10.75 | 1.38 |
| 3 | Boosting press. to 150 psi | 694 | 416 | 127.3 | -- | -- | -- |
| 4 | Cooling down to 140°F | -- | -- | -- | 1.45 | 10.75 | 1.38 |
| 5 | Boosting press. to 500 psi | 732 | 440 | 134.4 | -- | -- | -- |
| 6 | Cooling down to 140°F | -- | -- | -- | 1.67 | 9.10 | 0.92 |
| 7 | Boosting press. to 1000 psi | 465 | 280 | 85.4 | -- | -- | -- |
| 8 | Cooling down to 90°F | -- | -- | -- | 2.07 | 12.20 | 2.20 |
| | TOTAL | 2848 | 1710 | 522.7 | 7.04 | 42.8 | 5.88 |

TOTAL (Compression + Cooling) Capital Cost ($1,753,000), Annual Operating Cost ($528,600)

NOTE: Operating cost consists of fuel cost based on 340 days of operation per year

Table 7 PIPELINE @ 2000 PSI AND 140°F

| Stage Number | Description | Compression | | | Cooling | | |
|---|---|---|---|---|---|---|---|
| | | Power (HP-hr) | Capital Cost ($1000) | Operating Cost ($1000/yr) | Heat Removed MMBtu/hr | Capital Cost $1000 | Operating Cost ($100/yr) |
| 1 | Boosting press. to 50 psi | 957 | 574 | 175.6 | -- | -- | -- |
| 2 | Cooling down to 140°F | -- | -- | -- | 1.85 | 10.75 | 1.38 |
| 3 | Boosting press. to 150 psi | 694 | 416 | 127.3 | -- | -- | -- |
| 4 | Cooling down to 140°F | -- | -- | -- | 1.45 | 10.75 | 1.38 |
| 5 | Boosting press. to 500 psi | 732 | 440 | 134.4 | -- | -- | -- |
| 6 | Cooling down to 140°F | -- | -- | -- | 1.67 | 9.10 | .92 |
| 7 | Boosting press. to 1000 psi | 465 | 280 | 85.4 | -- | -- | -- |
| 8 | Cooling down to 140°F | -- | -- | -- | 1.35 | 10.75 | 1.38 |
| 9 | Boosting press. to 2000 psi | 380 | 230 | 69.8 | -- | -- | -- |
| 10 | Cooling down to 140°F | -- | -- | -- | 2.29 | 12.2 | 1.84 |
| | TOTAL | 3228 | 1940 | 592.5 | 8.61 | 53.55 | 6.90 |

TOTAL (Compression + Cooling) Capital Cost ($1,994,000) Annual Operating Cost ($599,400)

NOTE: Operating cost consists of fuel cost based on 340 days of operation per year

TABLE 8 TOTAL CAPITAL AND OPERATING COSTS FOR VARIOUS OPTIONS (NATURAL GAS STREAMS CONTAINING 5-10% $CO_2$)

| Options | | | | | Total Capital Cost ($1000) | Total Operating Cost ($1000/year) |
|---|---|---|---|---|---|---|
| Purification Cost | Type of Purification | Type of Transportation | Pressure psi | Temp. °F | | |
| Included | Chemical | • Truck | 300 | 90 | 2076.80 | 532.80 |
| | | • Truck | 300 | 0 | 4116.80 | 1201.76 |
| | | • Truck | 1000 | 90 | 2528.80 | 669.10 |
| | | • Pipeline | 2000 | 140 | 2769.80 | 739.90 |
| | Physical | • Truck | 300 | 90 | 2287.80 | 484.14 |
| | | • Truck | 300 | 0 | 4327.80 | 1153.10 |
| | | • Truck | 1000 | 90 | 2739.80 | 620.44 |
| | | • Pipeline | 2000 | 140 | 2980.80 | 691.24 |
| Adjusted | Chemical | • Truck | 300 | 90 | 2001.00 | 243.40 |
| | | • Truck | 300 | 0 | 4041.00 | 912.36 |
| | | • Truck | 1000 | 90 | 2453.00 | 379.70 |
| | | • Pipeline | 2000 | 140 | 2694.00 | 450.50 |
| | Physical | • Truck | 300 | 90 | 2226.00 | 198.40 |
| | | • Truck | 300 | 0 | 4266.00 | 867.36 |
| | | • Truck | 1000 | 90 | 2678.00 | 334.70 |
| | | • Pipeline | 2000 | 140 | 2919.00 | 405.50 |
| Ignored | Chemical | • Truck | 300 | 90 | 1326.80 | 397.80 |
| | | • Truck | 300 | 0 | 3366.80 | 1066.76 |
| | | • Truck | 1000 | 90 | 1778.80 | 534.10 |
| | | • Pipeline | 2000 | 140 | 2019.80 | 604.90 |
| | Physical | • Truck | 300 | 90 | 1312.80 | 394.14 |
| | | • Truck | 300 | 0 | 3352.80 | 1063.10 |
| | | • Truck | 1000 | 90 | 1764.80 | 530.44 |
| | | • Pipeline | 2000 | 140 | 2005.80 | 601.24 |

Table 9 COST OF $CO_2$ AT THE POINT OF CONSUMPTION
BASIS: DISTANCE OF 100 MILES BETWEEN PROCESS PLANT AND POINT OF CONUSMPTION

| Options: Purification Cost | Options: Type of Purification | Options: Type of Transportation | Pressure psi | Temp. $^oF$ | Cost of $CO_2$ Prior to Transport | Cost of Transport ($/MSCF) | Cost of $CO_2$ @ Delivery Point ($/MSCF) |
|---|---|---|---|---|---|---|---|
| Included | Chemical | • Truck | 300 | 90 | 0.30 | 2.65 | 2.95 |
| | | • Truck | 300 | 0 | 0.65 | 0.87 | 1.52 |
| | | • Truck | 1000 | 90 | 0.37 | 0.78 | 1.15 |
| | | • Pipeline | 2000 | 140 | 0.41 | 1.25 | 1.66 |
| | Physical | • Truck | 300 | 90 | 0.29 | 2.65 | 2.94 |
| | | • Truck | 300 | 0 | 0.64 | 0.87 | 1.51 |
| | | • Truck | 1000 | 90 | 0.36 | 0.78 | 1.14 |
| | | • Pipeline | 2000 | 140 | 0.40 | 1.25 | 1.65 |
| Adjusted | Chemical | • Truck | 300 | 90 | 0.18 | 2.65 | 2.83 |
| | | • Truck | 300 | 0 | 0.53 | 0.87 | 1.40 |
| | | • Truck | 1000 | 90 | 0.25 | 0.78 | 1.03 |
| | | • Pipeline | 2000 | 140 | 0.29 | 1.25 | 1.54 |
| | Physical | • Truck | 300 | 90 | 0.17 | 2.65 | 2.82 |
| | | • Truck | 300 | 0 | 0.52 | 0.87 | 1.39 |
| | | • Truck | 1000 | 90 | 0.25 | 0.78 | 1.03 |
| | | • Pipeline | 2000 | 140 | 0.28 | 1.25 | 1.53 |
| Ignored | Chemical | • Truck | 300 | 90 | 0.21 | 2.65 | 2.86 |
| | | • Truck | 300 | 0 | 0.56 | 0.87 | 1.43 |
| | | • Truck | 1000 | 90 | 0.29 | 0.78 | 1.07 |
| | | • Pipeline | 2000 | 140 | 0.32 | 1.25 | 1.57 |
| | Physical | • Truck | 300 | 90 | 0.21 | 2.65 | 2.86 |
| | | • Truck | 300 | 0 | 0.56 | 0.87 | 1.43 |
| | | • Truck | 1000 | 90 | 0.28 | 0.78 | 1.06 |
| | | • Pipeline | 2000 | 140 | 0.32 | 1.25 | 1.57 |

Table 10

MOLE PERCENT CO2 IN FEED XCO2= 0.30

TEMPERATURE= -81.40 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 612. | 2.770 | 0.219 | 0.306 | 0.694 | 0.848 | 0.152 | 0.273 | 0.727 | 0.1897 |
| 625. | 2.699 | 0.210 | 0.317 | 0.683 | 0.857 | 0.143 | 0.291 | 0.709 | 0.1984 |
| 650. | 2.383 | 0.221 | 0.360 | 0.640 | 0.859 | 0.141 | 0.318 | 0.682 | 0.2037 |
| 675. | 2.119 | 0.236 | 0.406 | 0.594 | 0.860 | 0.140 | 0.352 | 0.648 | 0.2090 |
| 700. | 1.888 | 0.257 | 0.456 | 0.544 | 0.860 | 0.140 | 0.396 | 0.604 | 0.2156 |
| 725. | 1.654 | 0.290 | 0.520 | 0.480 | 0.861 | 0.139 | 0.473 | 0.527 | 0.2266 |
| 750. | 1.464 | 0.337 | 0.588 | 0.412 | 0.861 | 0.139 | 0.590 | 0.410 | 0.2430 |
| 775. | 1.327 | 0.396 | 0.649 | 0.351 | 0.861 | 0.139 | 0.758 | 0.242 | 0.2664**** |

MOLE PERCENT CO2 IN FEED XCO2= 0.30

TEMPERATURE= -65.00 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 500. | 5.042 | 0.267 | 0.154 | 0.846 | 0.774 | 0.226 | 0.119 | 0.881 | 0.1009 |
| 600. | 3.651 | 0.261 | 0.218 | 0.782 | 0.796 | 0.204 | 0.166 | 0.834 | 0.1299 |
| 700. | 2.643 | 0.277 | 0.306 | 0.694 | 0.808 | 0.192 | 0.215 | 0.785 | 0.1491 |
| 800. | 1.858 | 0.335 | 0.437 | 0.563 | 0.811 | 0.189 | 0.297 | 0.703 | 0.1676 |
| 900. | 1.294 | 0.520 | 0.620 | 0.380 | 0.803 | 0.197 | 0.563 | 0.437 | 0.2137**** |

Table 11

MOLE PERCENT CO2 IN FEED XCO2= 0.35

TEMPERATURE= -81.40 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 612. | 2.770 | 0.219 | 0.306 | 0.694 | 0.848 | 0.152 | 0.366 | 0.634 | 0.2537 |
| 625. | 2.699 | 0.210 | 0.317 | 0.683 | 0.857 | 0.143 | 0.383 | 0.617 | 0.2617 |
| 650. | 2.383 | 0.221 | 0.360 | 0.640 | 0.859 | 0.141 | 0.419 | 0.581 | 0.2679 |
| 675. | 2.119 | 0.236 | 0.406 | 0.594 | 0.860 | 0.140 | 0.462 | 0.538 | 0.2744 |
| 700. | 1.888 | 0.257 | 0.456 | 0.544 | 0.860 | 0.140 | 0.520 | 0.480 | 0.2829 |
| 725. | 1.654 | 0.290 | 0.520 | 0.480 | 0.861 | 0.139 | 0.620 | 0.380 | 0.2971 |
| 750. | 1.464 | 0.337 | 0.588 | 0.412 | 0.861 | 0.139 | 0.773 | 0.227 | 0.3185 |
| 775. | 1.327 | 0.396 | 0.649 | 0.351 | 0.861 | 0.139 | 0.994 | 0.006 | 0.3492**** |

MOLE PERCENT CO2 IN FEED XCO2= 0.35

TEMPERATURE= -65.00 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 500. | 5.042 | 0.267 | 0.154 | 0.846 | 0.774 | 0.226 | 0.200 | 0.800 | 0.1691 |
| 600. | 3.651 | 0.261 | 0.218 | 0.782 | 0.796 | 0.204 | 0.253 | 0.747 | 0.1975 |
| 700. | 2.643 | 0.277 | 0.306 | 0.694 | 0.808 | 0.192 | 0.314 | 0.686 | 0.2182 |
| 800. | 1.858 | 0.335 | 0.437 | 0.563 | 0.811 | 0.189 | 0.431 | 0.569 | 0.2427 |
| 900. | 1.294 | 0.520 | 0.620 | 0.380 | 0.803 | 0.197 | 0.837 | 0.163 | 0.3178**** |

Table 12

MOLE PERCENT CO2 IN FEED XCO2= 0.40
TEMPERATURE= -81.40 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 612. | 2.770 | 0.219 | 0.306 | 0.694 | 0.848 | 0.152 | 0.458 | 0.542 | 0.3177 |
| 625. | 2.699 | 0.210 | 0.317 | 0.683 | 0.857 | 0.143 | 0.476 | 0.524 | 0.3250 |
| 650. | 2.383 | 0.221 | 0.360 | 0.640 | 0.859 | 0.141 | 0.519 | 0.481 | 0.3320 |
| 675. | 2.119 | 0.236 | 0.406 | 0.594 | 0.860 | 0.140 | 0.572 | 0.428 | 0.3399 |
| 700. | 1.888 | 0.257 | 0.456 | 0.544 | 0.860 | 0.140 | 0.643 | 0.357 | 0.3501 |
| 725. | 1.654 | 0.290 | 0.520 | 0.480 | 0.861 | 0.139 | 0.766 | 0.234 | 0.3675 |
| 750. | 1.464 | 0.337 | 0.588 | 0.412 | 0.861 | 0.139 | 0.956 | 0.044 | 0.3939**** |

MOLE PERCENT CO2 IN FEED XCO2= 0.40
TEMPERATURE= -65.00 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 500. | 5.042 | 0.267 | 0.154 | 0.846 | 0.774 | 0.226 | 0.280 | 0.720 | 0.2374 |
| 600. | 3.651 | 0.261 | 0.218 | 0.782 | 0.796 | 0.204 | 0.339 | 0.661 | 0.2652 |
| 700. | 2.643 | 0.277 | 0.306 | 0.694 | 0.808 | 0.192 | 0.414 | 0.586 | 0.2873 |
| 800. | 1.858 | 0.335 | 0.437 | 0.563 | 0.811 | 0.189 | 0.564 | 0.436 | 0.3179**** |

Table 13

MOLE PERCENT CO2 IN FEED XCO2= 0.45
TEMPERATURE= -81.40 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 612. | 2.770 | 0.219 | 0.306 | 0.694 | 0.848 | 0.152 | 0.550 | 0.450 | 0.3817 |
| 625. | 2.699 | 0.210 | 0.317 | 0.683 | 0.857 | 0.143 | 0.569 | 0.431 | 0.3883 |
| 650. | 2.383 | 0.221 | 0.360 | 0.640 | 0.859 | 0.141 | 0.619 | 0.381 | 0.3962 |
| 675. | 2.119 | 0.236 | 0.406 | 0.594 | 0.860 | 0.140 | 0.682 | 0.318 | 0.4054 |
| 700. | 1.888 | 0.257 | 0.456 | 0.544 | 0.860 | 0.140 | 0.767 | 0.233 | 0.4174 |
| 725. | 1.654 | 0.290 | 0.520 | 0.480 | 0.861 | 0.139 | 0.913 | 0.087 | 0.4379**** |

MOLE PERCENT CO2 IN FEED XCO2= 0.45
TEMPERATURE= -65.00 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 500. | 5.042 | 0.267 | 0.154 | 0.846 | 0.774 | 0.226 | 0.361 | 0.639 | 0.3056 |
| 600. | 3.651 | 0.261 | 0.218 | 0.782 | 0.796 | 0.204 | 0.426 | 0.574 | 0.3328 |
| 700. | 2.643 | 0.277 | 0.306 | 0.694 | 0.808 | 0.192 | 0.513 | 0.487 | 0.3565 |
| 800. | 1.858 | 0.335 | 0.437 | 0.563 | 0.811 | 0.189 | 0.698 | 0.302 | 0.3930**** |

Table 14

MOLE PERCENT CO2 IN FEED XCO2= 0.50
TEMPERATURE= -81.40 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 612. | 2.770 | 0.219 | 0.306 | 0.694 | 0.848 | 0.152 | 0.642 | 0.358 | 0.4457 |
| 625. | 2.699 | 0.210 | 0.317 | 0.683 | 0.857 | 0.143 | 0.662 | 0.338 | 0.4515 |
| 650. | 2.383 | 0.221 | 0.360 | 0.640 | 0.859 | 0.141 | 0.720 | 0.280 | 0.4604 |
| 675. | 2.119 | 0.236 | 0.406 | 0.594 | 0.860 | 0.140 | 0.792 | 0.208 | 0.4708 |
| 700. | 1.888 | 0.257 | 0.456 | 0.544 | 0.860 | 0.140 | 0.890 | 0.110 | 0.4847**** |

MOLE PERCENT CO2 IN FEED XCO2= 0.50
TEMPERATURE= -65.00 F

| PRESS. PSIA | K CH4 | K CO2 | X CH4 | X CO2 | Y CH4 | Y CO2 | MOLES LIQUID | MOLES GAS | MOLES CO2 IN LIQUID |
|---|---|---|---|---|---|---|---|---|---|
| 500. | 5.042 | 0.267 | 0.154 | 0.846 | 0.774 | 0.226 | 0.442 | 0.558 | 0.3738 |
| 600. | 3.651 | 0.261 | 0.218 | 0.782 | 0.796 | 0.204 | 0.512 | 0.488 | 0.4005 |
| 700. | 2.643 | 0.277 | 0.306 | 0.694 | 0.808 | 0.192 | 0.613 | 0.387 | 0.4256 |
| 800. | 1.858 | 0.335 | 0.437 | 0.563 | 0.811 | 0.189 | 0.831 | 0.169 | 0.4682**** |

Table 15 Comparison of Vapor-Liquid Equilibrium Ratio For Mehtane ($KCH_4$) From Various Sources

| P (PSIA) | T = -50°F (1) | T = -50°F (2) | T = -50°F (3) CP=1000 | T = -50°F (3) CP=2000 | T = -50°F (3) CP=5000 | T = 0°F (1) | T = 0°F (2) | T = 0°F (3) CP=1000 | T = 0°F (3) CP=2000 | T = 0°F (3) CP=3000 | T = +50°F (1) | T = +50°F (2) | T = +50°F (3) CP=1000 | T = +50°F (3) CP=2000 | T = +50°F (3) CP=5000 |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| 500 | NA | 5.97 | 2.25 | 3.3 | 4.1 | NA | 5.93 | 2.6 | 4.0 | 5.0 | NA | NA | 2.0 | 4.7 | 5.7 |
| 750 | 2.70 | 3.26 | 1.5 | 2.3 | 3.0 | 3.43 | 3.75 | 1.7 | 2.8 | 3.6 | NA | NA | 1.8 | 3.3 | 4.1 |
| 1000 | 1.25 | 1.39 | NA | 1.85 | 2.45 | 2.15 | 2.37 | NA | 2.15 | 3.0 | NA | NA | NA | 2.45 | 3.3 |
| 1500 | NA | 0.86 | NA | 1.36 | 1.9 | NA | 1.47 | NA | 1.45 | 2.25 | NA | NA | NA | 1.65 | 2.5 |

(1) Experimental Data (50% $CH_4$ - 50% $CO_2$)

(2) Calculated Using NGPA Program (40% $CH_4$ - 60% $CO_2$)

(3) CP = Convergence Pressure

NA Not Available

Table 16

COMPOSITION: 100. %CH4 0. %CO2
CRITICAL TEMPERATURE = 344. RANKINE
CRITICAL PRESSURE = 673. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 |
| | (H*-H)/TC | 2.80 | 3.25 | 3.65 | 3.95 | 4.30 | 4.70 |
| | CP* | 7.490 | 7.490 | 7.490 | 7.490 | 7.490 | 7.490 |
| | CP | 5.082 | 4.695 | 4.351 | 4.093 | 3.792 | 3.448 |
| | 1/M | 2.55 | 2.36 | 2.19 | 2.06 | 1.91 | 1.73 |
| 420. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.22 | 1.22 | 1.22 | 1.22 | 1.22 | 1.22 |
| | (H*-H)/TC | 2.20 | 2.75 | 3.10 | 3.40 | 3.75 | 4.05 |
| | CP* | 7.629 | 7.629 | 7.629 | 7.629 | 7.629 | 7.629 |
| | CP | 5.827 | 5.377 | 5.090 | 4.844 | 4.558 | 4.312 |
| | 1/M | 2.93 | 2.70 | 2.56 | 2.43 | 2.29 | 2.17 |
| 440. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.28 | 1.28 | 1.28 | 1.28 | 1.28 | 1.28 |
| | (H*-H)/TC | 2.05 | 2.30 | 2.55 | 2.80 | 3.05 | 3.30 |
| | CP* | 7.770 | 7.770 | 7.770 | 7.770 | 7.770 | 7.770 |
| | CP | 6.167 | 5.972 | 5.776 | 5.581 | 5.385 | 5.190 |
| | 1/M | 3.10 | 3.00 | 2.90 | 2.80 | 2.71 | 2.61 |
| 460. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.34 | 1.34 | 1.34 | 1.34 | 1.34 | 1.34 |
| | (H*-H)/TC | 1.75 | 2.00 | 2.25 | 2.40 | 2.65 | 2.95 |
| | CP* | 7.911 | 7.911 | 7.911 | 7.911 | 7.911 | 7.911 |
| | CP | 6.602 | 6.415 | 6.228 | 6.116 | 5.929 | 5.705 |
| | 1/M | 3.32 | 3.22 | 3.13 | 3.07 | 2.98 | 2.87 |
| 480. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.40 | 1.40 | 1.40 | 1.40 | 1.40 | 1.40 |
| | (H*-H)/TC | 1.55 | 1.75 | 1.95 | 2.10 | 2.30 | 2.50 |
| | CP* | 8.054 | 8.054 | 8.054 | 8.054 | 8.054 | 8.054 |
| | CP | 6.943 | 6.800 | 6.656 | 6.549 | 6.406 | 6.262 |
| | 1/M | 3.49 | 3.42 | 3.34 | 3.29 | 3.22 | 3.15 |
| 500. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.45 | 1.45 | 1.45 | 1.45 | 1.45 | 1.45 |
| | (H*-H)/TC | 1.35 | 1.55 | 1.75 | 1.85 | 2.10 | 2.25 |
| | CP* | 8.198 | 8.198 | 8.198 | 8.198 | 8.198 | 8.198 |
| | CP | 7.269 | 7.132 | 6.994 | 6.925 | 6.753 | 6.650 |
| | 1/M | 3.65 | 3.58 | 3.51 | 3.48 | 3.39 | 3.34 |

Table 16 Continued

COMPOSITION: 100. %CH4 0. %CO2
CRITICAL TEMPERATURE = 344. RANKINE
CRITICAL PRESSURE = 673. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.51 | 1.51 | 1.51 | 1.51 | 1.51 | 1.51 |
| | (H*-H)/TC | 1.25 | 1.40 | 1.55 | 1.65 | 1.85 | 2.00 |
| | CP* | 8.554 | 8.554 | 8.554 | 8.554 | 8.554 | 8.554 |
| | CP | 7.727 | 7.628 | 7.529 | 7.462 | 7.330 | 7.231 |
| | 1/M | 3.88 | 3.83 | 3.78 | 3.75 | 3.68 | 3.63 |
| 540. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.57 | 1.57 | 1.57 | 1.57 | 1.57 | 1.57 |
| | (H*-H)/TC | 1.15 | 1.30 | 1.40 | 1.50 | 1.60 | 1.85 |
| | CP* | 8.639 | 8.639 | 8.639 | 8.639 | 8.639 | 8.639 |
| | CP | 7.906 | 7.811 | 7.747 | 7.683 | 7.620 | 7.460 |
| | 1/M | 3.97 | 3.93 | 3.89 | 3.86 | 3.83 | 3.75 |
| 560. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.63 | 1.63 | 1.63 | 1.63 | 1.63 | 1.63 |
| | (H*-H)/TC | 1.05 | 1.15 | 1.25 | 1.35 | 1.50 | 1.65 |
| | CP* | 8.722 | 8.722 | 8.722 | 8.722 | 8.722 | 8.722 |
| | CP | 8.077 | 8.016 | 7.954 | 7.893 | 7.801 | 7.708 |
| | 1/M | 4.06 | 4.03 | 4.00 | 3.97 | 3.92 | 3.87 |
| 580. R | PR | 1.34 | 1.49 | 1.63 | 1.78 | 1.93 | 2.08 |
| | TR | 1.69 | 1.69 | 1.69 | 1.69 | 1.69 | 1.69 |
| | (H*-H)/TC | 0.95 | 1.00 | 1.10 | 1.20 | 1.35 | 1.50 |
| | CP* | 8.817 | 8.817 | 8.817 | 8.817 | 8.817 | 8.817 |
| | CP | 8.254 | 8.224 | 8.165 | 8.105 | 8.016 | 7.927 |
| | 1/M | 4.15 | 4.13 | 4.10 | 4.07 | 4.03 | 3.98 |

Table 17

COMPOSITION: 90. %CH4 10. %CO2
CRITICAL TEMPERATURE = 362. RANKINE
CRITICAL PRESSURE = 760. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.10 | 1.10 | 1.10 | 1.10 | 1.10 | 1.10 |
| | (H*-H)/TC | 3.15 | 3.55 | 4.00 | 4.45 | 4.85 | 5.30 |
| | CP* | 7.524 | 7.524 | 7.524 | 7.524 | 7.524 | 7.524 |
| | CP | 4.674 | 4.312 | 3.904 | 3.497 | 3.135 | 2.728 |
| | 1/M | 2.35 | 2.17 | 1.96 | 1.76 | 1.58 | 1.37 |
| 420. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 |
| | (H*-H)/TC | 2.50 | 2.80 | 3.20 | 3.50 | 3.80 | 4.20 |
| | CP* | 7.666 | 7.666 | 7.666 | 7.666 | 7.666 | 7.666 |
| | CP | 5.511 | 5.252 | 4.908 | 4.649 | 4.390 | 4.046 |
| | 1/M | 2.77 | 2.64 | 2.47 | 2.34 | 2.21 | 2.03 |
| 440. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.22 | 1.22 | 1.22 | 1.22 | 1.22 | 1.22 |
| | (H*-H)/TC | 2.20 | 2.40 | 2.65 | 2.90 | 3.20 | 3.50 |
| | CP* | 7.808 | 7.808 | 7.808 | 7.808 | 7.808 | 7.808 |
| | CP | 5.998 | 5.834 | 5.628 | 5.422 | 5.176 | 4.929 |
| | 1/M | 3.01 | 2.93 | 2.83 | 2.72 | 2.60 | 2.48 |
| 460. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.27 | 1.27 | 1.27 | 1.27 | 1.27 | 1.27 |
| | (H*-H)/TC | 1.85 | 2.10 | 2.30 | 2.50 | 2.70 | 3.00 |
| | CP* | 7.950 | 7.950 | 7.950 | 7.950 | 7.950 | 7.950 |
| | CP | 6.494 | 6.298 | 6.140 | 5.983 | 5.825 | 5.589 |
| | 1/M | 3.26 | 3.16 | 3.09 | 3.01 | 2.93 | 2.81 |
| 480. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.33 | 1.33 | 1.33 | 1.33 | 1.33 | 1.33 |
| | (H*-H)/TC | 1.65 | 1.80 | 2.00 | 2.20 | 2.45 | 2.70 |
| | CP* | 8.093 | 8.093 | 8.093 | 8.093 | 8.093 | 8.093 |
| | CP | 6.849 | 6.736 | 6.585 | 6.434 | 6.246 | 6.057 |
| | 1/M | 3.44 | 3.38 | 3.31 | 3.23 | 3.14 | 3.04 |
| 500. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.38 | 1.38 | 1.38 | 1.38 | 1.38 | 1.38 |
| | (H*-H)/TC | 1.45 | 1.65 | 1.80 | 1.95 | 2.15 | 2.30 |
| | CP* | 8.237 | 8.237 | 8.237 | 8.237 | 8.237 | 8.237 |
| | CP | 7.187 | 7.042 | 6.934 | 6.825 | 6.680 | 6.572 |
| | 1/M | 3.61 | 3.54 | 3.48 | 3.43 | 3.36 | 3.30 |

Table 17 Continued

COMPOSITION: 90. %CH4 10. %CO2
CRITICAL TEMPERATURE = 362. RANKINE
CRITICAL PRESSURE = 760. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.44 | 1.44 | 1.44 | 1.44 | 1.44 | 1.44 |
| | (H*-H)/TC | 1.50 | 1.60 | 1.75 | 1.90 | 2.10 | 2.30 |
| | CP* | 8.571 | 8.571 | 8.571 | 8.571 | 8.571 | 8.571 |
| | CP | 7.527 | 7.457 | 7.353 | 7.248 | 7.109 | 6.970 |
| | 1/M | 3.78 | 3.75 | 3.69 | 3.64 | 3.57 | 3.50 |
| 540. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.49 | 1.49 | 1.49 | 1.49 | 1.49 | 1.49 |
| | (H*-H)/TC | 1.40 | 1.50 | 1.60 | 1.80 | 1.95 | 2.05 |
| | CP* | 8.661 | 8.661 | 8.661 | 8.661 | 8.661 | 8.661 |
| | CP | 7.722 | 7.655 | 7.588 | 7.454 | 7.353 | 7.286 |
| | 1/M | 3.88 | 3.85 | 3.81 | 3.75 | 3.70 | 3.66 |
| 560. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.55 | 1.55 | 1.55 | 1.55 | 1.55 | 1.55 |
| | (H*-H)/TC | 1.30 | 1.40 | 1.50 | 1.65 | 1.80 | 1.90 |
| | CP* | 8.748 | 8.748 | 8.748 | 8.748 | 8.748 | 8.748 |
| | CP | 7.908 | 7.843 | 7.778 | 7.681 | 7.584 | 7.520 |
| | 1/M | 3.97 | 3.94 | 3.91 | 3.86 | 3.81 | 3.78 |
| 580. R | PR | 1.18 | 1.32 | 1.45 | 1.58 | 1.71 | 1.84 |
| | TR | 1.60 | 1.60 | 1.60 | 1.60 | 1.60 | 1.60 |
| | (H*-H)/TC | 1.20 | 1.30 | 1.40 | 1.55 | 1.65 | 1.80 |
| | CP* | 8.846 | 8.846 | 8.846 | 8.846 | 8.846 | 8.846 |
| | CP | 8.097 | 8.034 | 7.972 | 7.878 | 7.816 | 7.722 |
| | 1/M | 4.07 | 4.04 | 4.01 | 3.96 | 3.93 | 3.88 |

Table 18

COMPOSITION: 80. %CH4 20. %CO2
CRITICAL TEMPERATURE = 381. RANKINE
CRITICAL PRESSURE = 890. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.05 | 1.05 | 1.05 | 1.05 | 1.05 | 1.05 |
| | (H*-H)/TC | 3.15 | 3.75 | 4.50 | 5.25 | 5.95 | 6.55 |
| | CP* | 7.559 | 7.559 | 7.559 | 7.559 | 7.559 | 7.559 |
| | CP | 4.558 | 3.987 | 3.273 | 2.558 | 1.891 | 1.320 |
| | 1/M | 2.29 | 2.00 | 1.64 | 1.29 | 0.95 | 0.66 |
| 420. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.10 | 1.10 | 1.10 | 1.10 | 1.10 | 1.10 |
| | (H*-H)/TC | 2.50 | 2.90 | 3.30 | 3.60 | 3.95 | 4.40 |
| | CP* | 7.702 | 7.702 | 7.702 | 7.702 | 7.702 | 7.702 |
| | CP | 5.435 | 5.072 | 4.709 | 4.437 | 4.119 | 3.711 |
| | 1/M | 2.73 | 2.55 | 2.37 | 2.23 | 2.07 | 1.86 |
| 440. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.15 | 1.15 | 1.15 | 1.15 | 1.15 | 1.15 |
| | (H*-H)/TC | 2.10 | 2.35 | 2.65 | 2.95 | 3.20 | 3.65 |
| | CP* | 7.847 | 7.847 | 7.847 | 7.847 | 7.847 | 7.847 |
| | CP | 6.028 | 5.812 | 5.552 | 5.292 | 5.076 | 4.686 |
| | 1/M | 3.03 | 2.92 | 2.79 | 2.66 | 2.55 | 2.35 |
| 460. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.21 | 1.21 | 1.21 | 1.21 | 1.21 | 1.21 |
| | (H*-H)/TC | 1.70 | 1.95 | 2.25 | 2.55 | 2.70 | 3.00 |
| | CP* | 7.989 | 7.989 | 7.989 | 7.989 | 7.989 | 7.989 |
| | CP | 6.581 | 6.374 | 6.126 | 5.877 | 5.753 | 5.505 |
| | 1/M | 3.31 | 3.20 | 3.08 | 2.95 | 2.89 | 2.77 |
| 480. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.26 | 1.26 | 1.26 | 1.26 | 1.26 | 1.26 |
| | (H*-H)/TC | 1.55 | 1.80 | 2.00 | 2.25 | 2.45 | 2.65 |
| | CP* | 8.133 | 8.133 | 8.133 | 8.133 | 8.133 | 8.133 |
| | CP | 6.902 | 6.704 | 6.545 | 6.347 | 6.188 | 6.029 |
| | 1/M | 3.47 | 3.37 | 3.29 | 3.19 | 3.11 | 3.03 |
| 500. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.31 | 1.31 | 1.31 | 1.31 | 1.31 | 1.31 |
| | (H*-H)/TC | 1.35 | 1.50 | 1.75 | 1.90 | 2.15 | 2.25 |
| | CP* | 8.276 | 8.276 | 8.276 | 8.276 | 8.276 | 8.276 |
| | CP | 7.247 | 7.133 | 6.943 | 6.828 | 6.638 | 6.562 |
| | 1/M | 3.64 | 3.58 | 3.49 | 3.43 | 3.34 | 3.30 |

Table 18 Continued

COMPOSITION: 80. %CH4 20. %CO2
CRITICAL TEMPERATURE = 381. RANKINE
CRITICAL PRESSURE = 890. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.36 | 1.36 | 1.36 | 1.36 | 1.36 | 1.36 |
| | (H*-H)/TC | 1.45 | 1.65 | 1.80 | 1.90 | 2.10 | 2.30 |
| | CP* | 8.588 | 8.588 | 8.588 | 8.588 | 8.588 | 8.588 |
| | CP | 7.526 | 7.379 | 7.269 | 7.196 | 7.049 | 6.903 |
| | 1/M | 3.78 | 3.71 | 3.65 | 3.62 | 3.54 | 3.47 |
| 540. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.42 | 1.42 | 1.42 | 1.42 | 1.42 | 1.42 |
| | (H*-H)/TC | 1.30 | 1.45 | 1.60 | 1.70 | 1.85 | 2.00 |
| | CP* | 8.682 | 8.682 | 8.682 | 8.682 | 8.682 | 8.682 |
| | CP | 7.765 | 7.659 | 7.553 | 7.483 | 7.377 | 7.271 |
| | 1/M | 3.90 | 3.85 | 3.80 | 3.76 | 3.71 | 3.65 |
| 560. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.47 | 1.47 | 1.47 | 1.47 | 1.47 | 1.47 |
| | (H*-H)/TC | 1.20 | 1.35 | 1.50 | 1.60 | 1.70 | 1.85 |
| | CP* | 8.774 | 8.774 | 8.774 | 8.774 | 8.774 | 8.774 |
| | CP | 7.958 | 7.856 | 7.753 | 7.685 | 7.617 | 7.515 |
| | 1/M | 4.00 | 3.95 | 3.90 | 3.86 | 3.83 | 3.78 |
| 580. R | PR | 1.01 | 1.12 | 1.24 | 1.35 | 1.46 | 1.57 |
| | TR | 1.52 | 1.52 | 1.52 | 1.52 | 1.52 | 1.52 |
| | (H*-H)/TC | 1.10 | 1.25 | 1.40 | 1.50 | 1.60 | 1.75 |
| | CP* | 8.875 | 8.875 | 8.875 | 8.875 | 8.875 | 8.875 |
| | CP | 8.152 | 8.053 | 7.955 | 7.889 | 7.824 | 7.725 |
| | 1/M | 4.10 | 4.05 | 4.00 | 3.96 | 3.93 | 3.88 |

Table 19

COMPOSITION: 63. %CH4 37. %CO2

CRITICAL TEMPERATURE = 414. RANKINE

CRITICAL PRESSURE = 1039. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 0.97 | 0.97 | 0.97 | 0.97 | 0.97 | 0.97 |
| | (H*-H)/TC | 8.20 | 8.25 | 8.35 | 8.50 | 8.60 | 8.75 |
| | CP* | 20.463 | 20.463 | 20.463 | 20.463 | 20.463 | 20.463 |
| | CP | 11.976 | 11.925 | 11.821 | 11.666 | 11.562 | 11.407 |
| | 1/M | 6.02* | 5.99* | 5.94* | 5.86* | 5.81* | 5.73* |
| 420. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.01 | 1.01 | 1.01 | 1.01 | 1.01 | 1.01 |
| | (H*-H)/TC | 2.85 | 3.70 | 6.00 | 7.00 | 7.45 | 8.70 |
| | CP* | 7.765 | 7.765 | 21.191 | 21.191 | 21.191 | 21.191 |
| | CP | 4.956 | 4.118 | 15.277 | 14.291 | 13.848 | 12.615 |
| | 1/M | 2.49 | 2.07 | 7.68* | 7.18* | 6.96* | 6.34* |
| 440. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.06 | 1.06 | 1.06 | 1.06 | 1.06 | 1.06 |
| | (H*-H)/TC | 2.30 | 2.75 | 3.20 | 3.60 | 4.20 | 4.75 |
| | CP* | 7.912 | 7.912 | 7.912 | 7.912 | 7.912 | 7.912 |
| | CP | 5.748 | 5.324 | 4.901 | 4.524 | 3.960 | 3.442 |
| | 1/M | 2.89 | 2.68 | 2.46 | 2.27 | 1.99 | 1.73 |
| 460. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 |
| | (H*-H)/TC | 2.00 | 2.30 | 2.60 | 2.90 | 3.15 | 3.50 |
| | CP* | 8.056 | 8.056 | 8.056 | 8.056 | 8.056 | 8.056 |
| | CP | 6.256 | 5.986 | 5.716 | 5.446 | 5.221 | 4.906 |
| | 1/M | 3.14 | 3.01 | 2.87 | 2.74 | 2.62 | 2.47 |
| 480. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 |
| | (H*-H)/TC | 1.65 | 1.85 | 2.20 | 2.35 | 2.60 | 2.85 |
| | CP* | 8.200 | 8.200 | 8.200 | 8.200 | 8.200 | 8.200 |
| | CP | 6.777 | 6.604 | 6.302 | 6.173 | 5.957 | 5.742 |
| | 1/M | 3.41 | 3.32 | 3.17 | 3.10 | 2.99 | 2.89 |
| 500. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.21 | 1.21 | 1.21 | 1.21 | 1.21 | 1.21 |
| | (H*-H)/TC | 1.40 | 1.60 | 1.80 | 2.00 | 2.25 | 2.45 |
| | CP* | 8.343 | 8.343 | 8.343 | 8.343 | 8.343 | 8.343 |
| | CP | 7.183 | 7.018 | 6.852 | 6.687 | 6.480 | 6.314 |
| | 1/M | 3.61 | 3.53 | 3.44 | 3.36 | 3.26 | 3.17 |

Table 19 Continued

COMPOSITION: 63. %CH4 37. %CO2
CRITICAL TEMPERATURE = 414. RANKINE
CRITICAL PRESSURE = 1039. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.26 | 1.26 | 1.26 | 1.26 | 1.26 | 1.26 |
| | (H*-H)/TC | 1.50 | 1.65 | 1.85 | 2.05 | 2.20 | 2.40 |
| | CP* | 8.617 | 8.617 | 8.617 | 8.617 | 8.617 | 8.617 |
| | CP | 7.423 | 7.303 | 7.144 | 6.985 | 6.865 | 6.706 |
| | 1/M | 3.73 | 3.67 | 3.59 | 3.51 | 3.45 | 3.37 |
| 540. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.30 | 1.30 | 1.30 | 1.30 | 1.30 | 1.30 |
| | (H*-H)/TC | 1.40 | 1.50 | 1.75 | 1.90 | 2.00 | 2.20 |
| | CP* | 8.719 | 8.719 | 8.719 | 8.719 | 8.719 | 8.719 |
| | CP | 7.646 | 7.569 | 7.377 | 7.262 | 7.186 | 7.032 |
| | 1/M | 3.84 | 3.80 | 3.71 | 3.65 | 3.61 | 3.53 |
| 560. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.35 | 1.35 | 1.35 | 1.35 | 1.35 | 1.35 |
| | (H*-H)/TC | 1.30 | 1.40 | 1.60 | 1.70 | 1.80 | 1.95 |
| | CP* | 8.818 | 8.818 | 8.818 | 8.818 | 8.818 | 8.818 |
| | CP | 7.857 | 7.783 | 7.635 | 7.561 | 7.487 | 7.377 |
| | 1/M | 3.95 | 3.91 | 3.84 | 3.80 | 3.76 | 3.71 |
| 580. R | PR | 0.87 | 0.96 | 1.06 | 1.15 | 1.25 | 1.35 |
| | TR | 1.40 | 1.40 | 1.40 | 1.40 | 1.40 | 1.40 |
| | (H*-H)/TC | 1.15 | 1.30 | 1.45 | 1.55 | 1.60 | 1.80 |
| | CP* | 8.924 | 8.924 | 8.924 | 8.924 | 8.924 | 8.924 |
| | CP | 8.103 | 7.996 | 7.889 | 7.817 | 7.781 | 7.639 |
| | 1/M | 4.07 | 4.02 | 3.96 | 3.93 | 3.91 | 3.84 |

Table 20

COMPOSITION: 45. %CH4 55. %CO2
CRITICAL TEMPERATURE = 450. RANKINE
CRITICAL PRESSURE = 1184. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 0.89 | 0.89 | 0.89 | 0.89 | 0.89 | 0.89 |
| | (H*-H)/TC | 10.00 | 10.00 | 10.05 | 10.10 | 10.15 | 10.20 |
| | CP* | 20.202 | 20.202 | 20.202 | 20.202 | 20.202 | 20.202 |
| | CP | 8.952 | 8.952 | 8.896 | 8.840 | 8.784 | 8.727 |
| | 1/M | 4.50* | 4.50* | 4.47* | 4.44* | 4.41* | 4.39* |
| 420. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 0.93 | 0.93 | 0.93 | 0.93 | 0.93 | 0.93 |
| | (H*-H)/TC | 9.30 | 9.40 | 9.40 | 9.45 | 9.45 | 9.50 |
| | CP* | 20.788 | 20.788 | 20.788 | 20.788 | 20.788 | 20.788 |
| | CP | 10.824 | 10.717 | 10.717 | 10.663 | 10.663 | 10.609 |
| | 1/M | 5.44* | 5.39* | 5.39* | 5.36* | 5.36* | 5.33* |
| 440. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 0.98 | 0.98 | 0.98 | 0.98 | 0.98 | 0.98 |
| | (H*-H)/TC | 2.50 | 3.35 | 7.85 | 8.00 | 8.15 | 8.30 |
| | CP* | 7.981 | 7.981 | 21.406 | 21.406 | 21.406 | 21.406 |
| | CP | 5.424 | 4.555 | 13.378 | 13.225 | 13.071 | 12.918 |
| | 1/M | 2.73 | 2.29 | 6.72* | 6.65* | 6.57* | 6.49* |
| 460. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.02 | 1.02 | 1.02 | 1.02 | 1.02 | 1.02 |
| | (H*-H)/TC | 2.15 | 2.55 | 3.15 | 4.25 | 5.50 | 6.25 |
| | CP* | 8.127 | 8.127 | 8.127 | 8.127 | 8.127 | 22.359 |
| | CP | 6.023 | 5.632 | 5.045 | 3.969 | 2.746 | 16.245 |
| | 1/M | 3.03 | 2.83 | 2.54 | 1.99 | 1.38 | 8.16* |
| 480. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.07 | 1.07 | 1.07 | 1.07 | 1.07 | 1.07 |
| | (H*-H)/TC | 1.70 | 2.10 | 2.40 | 2.80 | 3.25 | 3.65 |
| | CP* | 8.271 | 8.271 | 8.271 | 8.271 | 8.271 | 8.271 |
| | CP | 6.677 | 6.302 | 6.021 | 5.646 | 5.224 | 4.849 |
| | 1/M | 3.36 | 3.17 | 3.03 | 2.84 | 2.63 | 2.44 |
| 500. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 |
| | (H*-H)/TC | 1.60 | 1.80 | 2.05 | 2.40 | 2.65 | 3.00 |
| | CP* | 8.413 | 8.413 | 8.413 | 8.413 | 8.413 | 8.413 |
| | CP | 6.973 | 6.793 | 6.568 | 6.253 | 6.028 | 5.713 |
| | 1/M | 3.50 | 3.41 | 3.30 | 3.14 | 3.03 | 2.87 |

Table 20 Continued

COMPOSITION: 45. %CH4 55. %CO2
CRITICAL TEMPERATURE = 450. RANKINE
CRITICAL PRESSURE = 1184. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 | 1.16 |
| | (H*-H)/TC | 1.35 | 1.50 | 1.75 | 2.00 | 2.25 | 2.40 |
| | CP* | 8.647 | 8.647 | 8.647 | 8.647 | 8.647 | 8.647 |
| | CP | 7.479 | 7.349 | 7.133 | 6.917 | 6.700 | 6.571 |
| | 1/M | 3.76 | 3.69 | 3.58 | 3.48 | 3.37 | 3.30 |
| 540. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.20 | 1.20 | 1.20 | 1.20 | 1.20 | 1.20 |
| | (H*-H)/TC | 1.25 | 1.35 | 1.55 | 1.75 | 2.00 | 2.15 |
| | CP* | 8.758 | 8.758 | 8.758 | 8.758 | 8.758 | 8.758 |
| | CP | 7.716 | 7.633 | 7.466 | 7.299 | 7.091 | 6.966 |
| | 1/M | 3.88 | 3.84 | 3.75 | 3.67 | 3.56 | 3.50 |
| 560. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.24 | 1.24 | 1.24 | 1.24 | 1.24 | 1.24 |
| | (H*-H)/TC | 1.15 | 1.25 | 1.45 | 1.60 | 1.80 | 1.95 |
| | CP* | 8.865 | 8.865 | 8.865 | 8.865 | 8.865 | 8.865 |
| | CP | 7.941 | 7.861 | 7.700 | 7.579 | 7.419 | 7.298 |
| | 1/M | 3.99 | 3.95 | 3.87 | 3.81 | 3.73 | 3.67 |
| 580. R | PR | 0.76 | 0.84 | 0.93 | 1.01 | 1.10 | 1.18 |
| | TR | 1.29 | 1.29 | 1.29 | 1.29 | 1.29 | 1.29 |
| | (H*-H)/TC | 1.10 | 1.15 | 1.25 | 1.40 | 1.55 | 1.75 |
| | CP* | 8.975 | 8.975 | 8.975 | 8.975 | 8.975 | 8.975 |
| | CP | 8.122 | 8.083 | 8.006 | 7.889 | 7.773 | 7.618 |
| | 1/M | 4.08 | 4.06 | 4.02 | 3.96 | 3.91 | 3.83 |

Table 21

COMPOSITION: 28. %CH4 72. %CO2
CRITICAL TEMPERATURE = 486. RANKINE
CRITICAL PRESSURE = 1245. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 400. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 0.82 | 0.82 | 0.82 | 0.82 | 0.82 | 0.82 |
| | (H*-H)/TC | 11.20 | 11.25 | 11.25 | 11.30 | 11.30 | 11.30 |
| | CP* | 19.956 | 19.956 | 19.956 | 19.956 | 19.956 | 19.956 |
| | CP | 6.348 | 6.287 | 6.287 | 6.226 | 6.226 | 6.226 |
| | 1/M | 3.19* | 3.16* | 3.16* | 3.13* | 3.13* | 3.13* |
| 420. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 0.86 | 0.86 | 0.86 | 0.86 | 0.86 | 0.86 |
| | (H*-H)/TC | 10.65 | 10.70 | 10.70 | 10.70 | 10.75 | 10.75 |
| | CP* | 20.407 | 20.407 | 20.407 | 20.407 | 20.407 | 20.407 |
| | CP | 8.084 | 8.026 | 8.026 | 8.026 | 7.968 | 7.968 |
| | 1/M | 4.06* | 4.03* | 4.03* | 4.03* | 4.00* | 4.00* |
| 440. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 0.91 | 0.91 | 0.91 | 0.91 | 0.91 | 0.91 |
| | (H*-H)/TC | 9.85 | 9.85 | 9.90 | 9.95 | 9.95 | 10.00 |
| | CP* | 20.902 | 20.902 | 20.902 | 20.902 | 20.902 | 20.902 |
| | CP | 10.022 | 10.022 | 9.967 | 9.911 | 9.911 | 9.856 |
| | 1/M | 5.04* | 5.04* | 5.01* | 4.98* | 4.98* | 4.95* |
| 460. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 0.95 | 0.95 | 0.95 | 0.95 | 0.95 | 0.95 |
| | (H*-H)/TC | 8.75 | 8.80 | 8.85 | 8.85 | 8.90 | 9.00 |
| | CP* | 21.846 | 21.846 | 21.846 | 21.846 | 21.846 | 21.846 |
| | CP | 12.601 | 12.548 | 12.495 | 12.495 | 12.443 | 12.337 |
| | 1/M | 6.33* | 6.31* | 6.28* | 6.28* | 6.25* | 6.20* |
| 480. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 0.99 | 0.99 | 0.99 | 0.99 | 0.99 | 0.99 |
| | (H*-H)/TC | 2.15 | 2.60 | 3.50 | 7.25 | 7.55 | 7.75 |
| | CP* | 8.338 | 8.338 | 8.338 | 23.638 | 23.638 | 23.638 |
| | CP | 6.161 | 5.705 | 4.794 | 16.297 | 15.994 | 15.791 |
| | 1/M | 3.10 | 2.87 | 2.41 | 8.19* | 8.04* | 7.94* |
| 500. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 1.03 | 1.03 | 1.03 | 1.03 | 1.03 | 1.03 |
| | (H*-H)/TC | 1.85 | 2.25 | 2.65 | 3.20 | 4.00 | 4.65 |
| | CP* | 8.480 | 8.480 | 8.480 | 8.480 | 8.480 | 8.480 |
| | CP | 6.681 | 6.293 | 5.904 | 5.369 | 4.592 | 3.960 |
| | 1/M | 3.36 | 3.16 | 2.97 | 2.70 | 2.31 | 1.99 |

Table 21 Continued

COMPOSITION: 28. %CH4 72. %CO2
CRITICAL TEMPERATURE = 486. RANKINE
CRITICAL PRESSURE = 1245. PSIA

| | | 900. PSIA | 1000. PSIA | 1100. PSIA | 1200. PSIA | 1300. PSIA | 1400. PSIA |
|---|---|---|---|---|---|---|---|
| 520. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 1.07 | 1.07 | 1.07 | 1.07 | 1.07 | 1.07 |
| | (H*-H)/TC | 1.80 | 2.10 | 2.40 | 2.80 | 3.30 | 3.80 |
| | CP* | 8.676 | 8.676 | 8.676 | 8.676 | 8.676 | 8.676 |
| | CP | 6.994 | 6.714 | 6.433 | 6.059 | 5.592 | 5.125 |
| | 1/M | 3.51 | 3.37 | 3.23 | 3.04 | 2.81 | 2.58 |
| 540. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 | 1.11 |
| | (H*-H)/TC | 1.60 | 1.80 | 2.00 | 2.30 | 2.60 | 3.00 |
| | CP* | 8.795 | 8.795 | 8.795 | 8.795 | 8.795 | 8.795 |
| | CP | 7.355 | 7.175 | 6.995 | 6.725 | 6.455 | 6.095 |
| | 1/M | 3.70 | 3.61 | 3.51 | 3.38 | 3.24 | 3.06 |
| 560. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 1.15 | 1.15 | 1.15 | 1.15 | 1.15 | 1.15 |
| | (H*-H)/TC | 1.50 | 1.70 | 1.90 | 2.10 | 2.40 | 2.70 |
| | CP* | 8.909 | 8.909 | 8.909 | 8.909 | 8.909 | 8.909 |
| | CP | 7.607 | 7.434 | 7.260 | 7.087 | 6.826 | 6.566 |
| | 1/M | 3.82 | 3.74 | 3.65 | 3.56 | 3.43 | 3.30 |
| 580. R | PR | 0.72 | 0.80 | 0.88 | 0.96 | 1.04 | 1.12 |
| | TR | 1.19 | 1.19 | 1.19 | 1.19 | 1.19 | 1.19 |
| | (H*-H)/TC | 1.35 | 1.50 | 1.70 | 1.90 | 2.10 | 2.40 |
| | CP* | 9.024 | 9.024 | 9.024 | 9.024 | 9.024 | 9.024 |
| | CP | 7.893 | 7.767 | 7.600 | 7.432 | 7.265 | 7.013 |
| | 1/M | 3.97 | 3.90 | 3.82 | 3.73 | 3.65 | 3.52 |

## COMPARISON OF $CO_2$ PIPELINE

## REPRESSURIZATION AND THERMAL CONDITIONING REQUIREMENTS

| | | Low Compressibility Path | | | High Compressibility Path | | |
|---|---|---|---|---|---|---|---|
| Line Capacity (MMSCFD) | Nominal Diameter (in) | Distance Between Compressor Stations (mi) | Initial Temp ( F) | Final Temp ( F) | Distance Between Compressor Stations (mi) | Initial Temp ( F) | Final Temp ( F) |
| 50 | 8 | 88 | 80 | 75 | 75 | 185 | 77 |
| 125 | 12 | 107 | 80 | 75 | 61 | 185 | 101 |
| 250 | 16 | 87 | 80 | 75 | 38 | 185 | 147 |
| 500 | 20 | 70 | 80 | 75 | 28 | 185 | 167 |

NOTE: Compressor stations are located where pressure has declined from 2000 psia to 1400 psia

TABLE 22

## COMPARISON OF $CO_2$

## PIPELINE REPRESSURIZATION POWER AND THERMAL CONDITIONING

| | | Low Compressibility Path | | | High Compressibility Path | | |
|---|---|---|---|---|---|---|---|
| Line Capacity (MMSCFD) | Nominal Diameter (in) | Distance Between Compressor Stations (mi) | Compressor Power (hp/mi) | Thermal Conditioning (BTU/mi/day) | Distance Between Compressor Stations (mi) | Compressor Power(hp/mi) | Thermal Conditioning (BTU/mi/day) |
| 50 | 8 | 88 | 2.9 | 0 | 75 | 9.1 | $6.2 \times 10^6$ |
| 125 | 12 | 107 | 5.9 | 0 | 61 | 28.0 | $9.6 \times 10^6$ |
| 250 | 16 | 87 | 14.5 | 0 | 38 | 89.9 | 0 |
| 500 | 20 | 70 | 36.0 | 0 | 28 | 244.1 | 0 |

Note: Compressor stations are located where pressure has declined from 2000 psia to 1400 psia

# SUMMARY OF TRANSPORTATION COSTS

| Line Capacity (MMSCFD) | Nominal Diameter (in) | Investment Cost Per Mile: Low Compressibility | Investment Cost Per Mile: High Compressibility | Annual Operating Cost Per Mile: Low Compressibility | Annual Operating Cost Per Mile: High Compressibility |
|---|---|---|---|---|---|
| 50 | 8 | $162,000 | $165,000 | $11,700 | $17,800 |
| 125 | 12 | 243,000 | 255,000 | 13,900 | 26,900 |
| 250 | 16 | 308,000 | 349,000 | 18,200 | 35,600 |
| 500 | 20 | 419,000 | 532,000 | 26,000 | 64,100 |

TABLE 24

## SUMMARY OF $CO_2$ PIPELINE DESIGNS

<u>Trunk Lines</u> (20 Year Life)

| | | | | | |
|---|---|---|---|---|---|
| Flow Rate (MMSCFD) | 20 | 50 | 125 | 250 | 500 |
| Diameter (inches) | 8 | 10 | 16 | 22 | 30 |
| Length (miles) | 50 | 100 | 200 | 400 | 500 |

<u>Laterals</u> (5 Year Life)

| | | | | | | |
|---|---|---|---|---|---|---|
| Flow Rates (MMSCFD) | 1 | 2 | 5 | 10 | 20 | 50 |
| Diameter (inches) | 2 | 3 | 4 | 5 | 8 | 10 |
| Length (miles) | 5 | 10 | 20 | 40 | 80 | 100 |

Table 25

TRUNK PIPELINE CALCULATIONS
(SUMMARY TABLE)

Flow Rate = 500 MMSCFD, D = 30 inches

| Length (Miles) | Cost $15,000/in·mile ($/MSCF) | Cost $20,000/in·mile ($/MSCF) |
|---|---|---|
| 50 | 0.0394 | 0.0526 |
| 100 | 0.0789 | 0.1052 |
| 200 | 0.1578 | 0.2103 |
| 400 | 0.3155 | 0.4207 |
| 500 | 0.3944 | 0.5259 |

Flow Rate = 250 MMSCFD, D = 22 inches

| | | |
|---|---|---|
| 50 | 0.0578 | 0.0771 |
| 100 | 0.1157 | 0.1543 |
| 200 | 0.2314 | 0.3085 |
| 400 | 0.4628 | 0.6170 |
| 500 | 0.5784 | 0.7713 |

Flow Rate = 125 MMSCFD, D = 16 inches

| | | |
|---|---|---|
| 50 | 0.0841 | 0.1122 |
| 100 | 0.1683 | 0.2244 |
| 200 | 0.3365 | 0.4487 |
| 400 | 0.6731 | 0.8975 |
| 500 | 0.8414 | 1.1218 |

Flow Rate = 50 MMSCFD, D = 10 inches

| | | |
|---|---|---|
| 50 | 0.1315 | 0.1753 |
| 100 | 0.2629 | 0.3506 |
| 200 | 0.5259 | 0.7011 |
| 400 | 1.0517 | 1.4023 |
| 500 | 1.3146 | 1.7528 |

Flow Rate = 20 MMSCFD, D = 8 inches

| | | |
|---|---|---|
| 50 | 0.2629 | 0.3506 |
| 100 | 0.5259 | 0.7011 |
| 200 | 1.0517 | 1.4023 |
| 400 | 2.1034 | 2.8046 |
| 500 | 2.6293 | 3.5057 |

Table 26

LATERAL PIPELINE CALCULATIONS
(SUMMARY TABLE)

Flow Rate = 50 MMSCFD, D = 10 inches

| Length (Miles) | Cost $15,000/in·mile ($/MSCF) | Cost $20,000/in·mile ($/MSCF) |
|---|---|---|
| 5 | 0.0194 | 0.0259 |
| 10 | 0.0388 | 0.0517 |
| 20 | 0.0776 | 0.1035 |
| 40 | 0.1556 | 0.2070 |
| 80 | 0.3105 | 0.4140 |
| 100 | 0.3881 | 0.5175 |

Flow Rate = 20 MMSCFD, D = 8 inches

| | | |
|---|---|---|
| 5 | 0.0388 | 0.0517 |
| 10 | 0.0776 | 0.1035 |
| 20 | 0.1552 | 0.2070 |
| 40 | 0.3105 | 0.4140 |
| 80 | 0.6210 | 0.8279 |
| 100 | 0.7762 | 1.0349 |

Flow Rate = 10 MMSCFD, D = 6 inches

| | | |
|---|---|---|
| 5 | 0.0582 | 0.0776 |
| 10 | 0.1164 | 0.1552 |
| 20 | 0.2329 | 0.3105 |
| 40 | 0.4657 | 0.6210 |
| 80 | 0.9314 | 1.2419 |
| 100 | 1.1643 | 1.5524 |

Flow Rate = 5 MMSCFD, D = 4 inches

| Length (Miles) | Cost $10,000/in·mile ($/MSCF) | Cost $15,000/in·mile ($/MSCF) |
|---|---|---|
| 5 | 0.0259 | 0.0388 |
| 10 | 0.0517 | 0.0776 |
| 20 | 0.1035 | 0.1552 |
| 40 | 0.2070 | 0.3105 |
| 80 | 0.4140 | 0.6210 |
| 100 | 0.5174 | 0.7762 |

Table 27
(1)

LATERAL PIPELINE CALCULATIONS
(SUMMARY TABLE - CONT.)

Flow Rate = 2 MMSCFD, D = 3 inches

| Length (Miles) | Cost $10,000/in·mile ($/MSCF) | Cost $15,000/in·mile ($/MSCF) |
|---|---|---|
| 5 | 0.0970 | 0.1455 |
| 10 | 0.1940 | 0.2911 |
| 20 | 0.3881 | 0.5820 |
| 40 | 0.7762 | 1.1643 |
| 80 | 1.5524 | 2.3286 |
| 100 | 1.9404 | 2.9107 |

Flow Rate = 1 MMSCFD, D = 2 inches

| | | |
|---|---|---|
| 5 | 0.0648 | 0.0970 |
| 10 | 0.1294 | 0.1940 |
| 20 | 0.2587 | 0.3881 |
| 40 | 0.5175 | 0.7762 |
| 80 | 1.0349 | 1.5524 |
| 100 | 1.2936 | 1.9405 |

Table 27
(2)

SUMMARY OF TRACTOR & TRAILER REQUIREMENTS

FOR $CO_2$ TRANSPORT

Working Pressure = 300 psia

Haul Distances

| Daily Requirement - MMCSFD | 5 Tractor/ Trailers | 10 Tractor/ Trailers | 20 Tractor/ Trailers | 40 Tractor/ Trailers | 80 Tractor/ Trailers | 100 Tractor/ Trailers |
|---|---|---|---|---|---|---|
| 0.5 | 0.265/ 1.338 | 0.371/ 1.409 | 0.582/ 1.552 | 1.006/ 1.837 | 1.853/ 2.408 | 2.277/ 2.694 |
| 1.0 | 0.530/ 1.623 | 0.741/ 1.766 | 1.165/ 2.051 | 2.012/ 2.622 | 3.707/ 3.764 | 4.554/ 4.334 |
| 2.0 | 1.059 2.194 | 1.483/ 2.480 | 2.330/ 3.050 | 4.024/ 4.192 | 7.413/ 6.475 | 9.107/ 7.616 |
| 5.0 | 2.648/ 3.907 | 3.707/ 4.620 | 5.825/ 6.047 | 10.061/ 8.910 | 18.533/ 14.608 | 22.769/ 17.462 |
| 10.0 | 5.295/ 6.760 | 7.413/ 8.187 | 11.649/ 11.041 | 20.121/ 16.749 | 37.065/ 28.164 | 45.537/ 33.871 |

Table 28

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Tractor Unit Investment: $50,000

90% of original investment depreciated straight-line over 7 years to zero salvage

7-year evaluation life

15% discount rate

50% effective tax rate

10% investment tax credit

Annual investment cost: $50{,}000(0.9)\underset{0.24036}{(A/P_{15,7})} =$

$10,800/year
(not subject to tax)

Tax deductible costs:

Fixed cost: $6,000/year

Driver labor
(3 shifts, 4 drivers): 4x30,000 = $120,000/year

Operation: ($0.35/mi) (107 hr/wk) (52 wk/year)

$$\left(\frac{D/15}{D/15 + 0.5}\right)\left(\frac{340}{365}\right)\left(30\text{mph}\right) = \$\left[\frac{3628D}{D/15 + 0.5}\right]/\text{year}$$

Total tax-deductible cost: $\$\left[126{,}000 + \dfrac{3628D}{D/15 + 0.5}\right]/\text{year}$

Depreciation tax savings: 50,000 (0.9)(1/7)(0.5) = $3,200/year

After-tax tractor unit cost: $-3{,}200 + \left[126{,}000 + \dfrac{3628D}{D/15 + 0.5}\right] + 10{,}800/0.5$

$$= \$\left[144{,}400 + \frac{3628D}{D/15 + 0.5}\right]/\text{year}$$

Table 29

Development of Annual Tractor Unit Cost

Trailer Unit Investment: \$45,000

90% of original investment depreciated straight-line over 7-years to zero salvage

7-year evaluation life

15% discount rate

50% effective tax rate

10% investment tax credit

Annual investment cost: $45{,}000(0.9)\underset{0.24036}{(A/P_{15,7})} = \$9{,}700/\text{year}$

Tax deductible costs:

Fixed costs: \$1,000/year

Operation: (\$0.10/mi)(0.95)(24 hr/day)(340 days/year)

$$\times \left(\frac{D/15}{1 + D/15 + 2.459/Q}\right)\left(30 \text{ mph}\right)$$

$$= \$ \left[\frac{1550D}{1 + D/15 + 2.459/Q}\right] /\text{year}$$

Total tax-deductible cost: $\$ \left[1{,}000 + \frac{1550D}{1 + D/15 + 2.459/Q}\right]/\text{year}$

Depreciation tax savings: $45{,}000\ (0.9)(1/7)(0.5) = \$2{,}900/\text{year}$

After-tax trailer unit cost:

$$-2{,}900 + \left[1{,}000 + \frac{1550D}{1 + D/15 + 2.459/Q}\right] + 9{,}700/0.5$$

$$= \$ \left[17{,}600 + \frac{1550D}{1 + D/15 + 2.459/Q}\right] /\text{year}$$

Table 30

Development of Annual Trailer Unit Cost

Table 31 Summary of Transport costs for Truck-Delivered $CO_2$ ($/MCF)
300 psia, $0^{o}F$, saturated vapor

| Daily Requirement - MMSCFD | Haul Distance (mi) 5 | 10 | 20 | 40 | 80 | 100 |
|---|---|---|---|---|---|---|
| 0.5 | 0.407 | 0.548 | 0.830 | 1.394 | 2.522 | 3.086 |
| 1.0 | 0.353 | 0.494 | 0.776 | 1.340 | 2.468 | 3.032 |
| 2.0 | 0.325 | 0.466 | 0.748 | 1.312 | 2.440 | 3.004 |
| 5.0 | 0.309 | 0.450 | 0.732 | 1.296 | 2.424 | 2.988 |
| 10.0 | 0.304 | 0.445 | 0.727 | 1.291 | 2.419 | 2.982 |

# APPENDIX A

Tractor and Trailer Requirements and Costs for $CO_2$ Distribution
By Truck From A Trunk Pipeline

(Based on $45,000 Trailer Used at $0^oF$ and 300 psi, s.g.= 0.0688)

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 0.50 | 5.00 | 0.265 | $166,168.00 | $43,993.10 | 1.338 | $18,839.73 | $25,207.44 | $0.407/Mcf |
| 0.50 | 10.00 | 0.371 | $175,497.13 | $65,048.27 | 1.409 | $19,953.95 | $28,121.85 | $0.548/Mcf |
| 0.50 | 15.00 | 0.477 | $180,680.00 | $86,103.47 | 1.481 | $20,960.80 | $31,036.25 | $0.689/Mcf |
| 0.50 | 20.00 | 0.582 | $183,978.16 | $107,158.64 | 1.552 | $21,875.07 | $33,950.66 | $0.830/Mcf |
| 0.50 | 25.00 | 0.688 | $186,261.53 | $128,213.84 | 1.623 | $22,708.99 | $36,865.06 | $0.971/Mcf |
| 0.50 | 30.00 | 0.794 | $187,936.00 | $149,269.03 | 1.695 | $23,472.70 | $39,779.46 | $1.112/Mcf |
| 0.50 | 35.00 | 0.900 | $189,216.47 | $170,324.25 | 1.766 | $24,174.70 | $42,693.87 | $1.253/Mcf |
| 0.50 | 40.00 | 1.006 | $190,227.34 | $191,379.41 | 1.837 | $24,822.18 | $45,608.27 | $1.394/Mcf |
| 0.50 | 45.00 | 1.112 | $191,045.69 | $212,434.59 | 1.909 | $25,421.26 | $48,522.66 | $1.535/Mcf |
| 0.50 | 50.00 | 1.218 | $191,721.72 | $233,489.78 | 1.980 | $25,977.17 | $51,437.07 | $1.676/Mcf |
| 0.50 | 55.00 | 1.324 | $192,289.59 | $254,545.00 | 2.051 | $26,494.41 | $54,351.46 | $1.817/Mcf |
| 0.50 | 60.00 | 1.430 | $192,773.31 | $275,600.13 | 2.123 | $26,976.89 | $57,265.87 | $1.958/Mcf |
| 0.50 | 65.00 | 1.536 | $193,190.34 | $296,655.38 | 2.194 | $27,427.99 | $60,180.27 | $2.099/Mcf |
| 0.50 | 70.00 | 1.641 | $193,553.53 | $317,710.56 | 2.265 | $27,850.68 | $63,094.66 | $2.240/Mcf |
| 0.50 | 75.00 | 1.747 | $193,872.72 | $338,765.75 | 2.337 | $28,247.55 | $66,009.06 | $2.381/Mcf |
| 0.50 | 80.00 | 1.853 | $194,155.41 | $359,820.94 | 2.408 | $28,620.92 | $68,923.47 | $2.522/Mcf |
| 0.50 | 85.00 | 1.959 | $194,407.56 | $380,876.13 | 2.479 | $28,972.79 | $71,837.86 | $2.663/Mcf |
| 0.50 | 90.00 | 2.065 | $194,633.84 | $401,931.31 | 2.551 | $29,304.98 | $74,752.27 | $2.804/Mcf |
| 0.50 | 95.00 | 2.171 | $194,838.03 | $422,986.44 | 2.622 | $29,619.10 | $77,666.66 | $2.945/Mcf |
| 0.50 | 100.00 | 2.277 | $195,023.25 | $444,041.75 | 2.694 | $29,916.57 | $80,581.05 | $3.086/Mcf |

| $CO_2$ Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | $CO_2$ Cost |
|---|---|---|---|---|---|---|---|---|
| 1.00 | 5.00 | 0.530 | $166,168.00 | $87,986.20 | 1.623 | $19,643.59 | $31,889.16 | $0.353/Mcf |
| 1.00 | 10.00 | 0.741 | $175,497.13 | $130,096.55 | 1.766 | $21,356.97 | $37,718.00 | $0.494/Mcf |
| 1.00 | 15.00 | 0.953 | $180,680.00 | $172,206.94 | 1.909 | $22,814.17 | $43,546.83 | $0.635/Mcf |
| 1.00 | 20.00 | 1.165 | $183,978.16 | $214,317.28 | 2.051 | $24,068.66 | $49,375.65 | $0.776/Mcf |
| 1.00 | 25.00 | 1.377 | $186,261.53 | $256,427.69 | 2.194 | $25,159.99 | $55,204.48 | $0.917/Mcf |
| 1.00 | 30.00 | 1.589 | $187,936.00 | $298,538.06 | 2.337 | $26,118.04 | $61,033.29 | $1.058/Mcf |
| 1.00 | 35.00 | 1.800 | $189,216.47 | $340,648.50 | 2.480 | $26,965.83 | $66,862.11 | $1.199/Mcf |
| 1.00 | 40.00 | 2.012 | $190,227.34 | $382,758.81 | 2.622 | $27,721.35 | $72,690.91 | $1.340/Mcf |
| 1.00 | 45.00 | 2.224 | $191,045.69 | $424,869.19 | 2.765 | $28,398.88 | $78,519.72 | $1.481/Mcf |
| 1.00 | 50.00 | 2.436 | $191,721.72 | $466,979.56 | 2.908 | $29,009.92 | $84,348.52 | $1.622/Mcf |
| 1.00 | 55.00 | 2.648 | $192,289.59 | $509,090.00 | 3.050 | $29,563.79 | $90,177.33 | $1.763/Mcf |
| 1.00 | 60.00 | 2.859 | $192,773.31 | $551,200.25 | 3.193 | $30,068.16 | $96,006.13 | $1.904/Mcf |
| 1.00 | 65.00 | 3.071 | $193,190.34 | $593,310.75 | 3.336 | $30,529.38 | $101,834.94 | $2.045/Mcf |
| 1.00 | 70.00 | 3.283 | $193,553.53 | $635,421.13 | 3.478 | $30,952.75 | $107,663.73 | $2.186/Mcf |
| 1.00 | 75.00 | 3.495 | $193,872.72 | $677,531.50 | 3.621 | $31,342.76 | $113,492.53 | $2.327/Mcf |
| 1.00 | 80.00 | 3.707 | $194,155.41 | $719,641.88 | 3.764 | $31,703.20 | $119,321.34 | $2.468/Mcf |
| 1.00 | 85.00 | 3.918 | $194,407.56 | $761,752.25 | 3.906 | $32,037.30 | $125,150.13 | $2.609/Mcf |
| 1.00 | 90.00 | 4.130 | $194,633.84 | $803,862.63 | 4.049 | $32,347.86 | $130,978.94 | $2.750/Mcf |
| 1.00 | 95.00 | 4.342 | $194,838.03 | $845,972.88 | 4.192 | $32,637.27 | $136,807.69 | $2.891/Mcf |
| 1.00 | 100.00 | 4.554 | $195,023.25 | $888,083.50 | 4.334 | $32,907.63 | $142,636.50 | $3.032/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 1.50 | 5.00 | 0.794 | $166,168.00 | $131,979.28 | 1.909 | $20,207.09 | $38,570.90 | $0.334/Mcf |
| 1.50 | 10.00 | 1.112 | $175,497.13 | $195,144.84 | 2.123 | $22,288.45 | $47,314.18 | $0.475/Mcf |
| 1.50 | 15.00 | 1.430 | $180,680.00 | $258,310.41 | 2.337 | $23,988.53 | $56,057.44 | $0.616/Mcf |
| 1.50 | 20.00 | 1.747 | $183,978.16 | $321,475.94 | 2.551 | $25,403.32 | $64,800.70 | $0.757/Mcf |
| 1.50 | 25.00 | 2.065 | $186,261.53 | $384,641.50 | 2.765 | $26,599.07 | $73,543.94 | $0.898/Mcf |
| 1.50 | 30.00 | 2.383 | $187,936.00 | $447,807.06 | 2.979 | $27,622.99 | $82,287.16 | $1.039/Mcf |
| 1.50 | 35.00 | 2.700 | $189,216.47 | $510,972.75 | 3.193 | $28,509.64 | $91,030.41 | $1.180/Mcf |
| 1.50 | 40.00 | 3.018 | $190,227.34 | $574,138.25 | 3.407 | $29,284.89 | $99,773.61 | $1.321/Mcf |
| 1.50 | 45.00 | 3.336 | $191,045.69 | $637,303.75 | 3.621 | $29,968.48 | $108,516.83 | $1.462/Mcf |
| 1.50 | 50.00 | 3.654 | $191,721.72 | $700,469.38 | 3.835 | $30,575.78 | $117,260.03 | $1.603/Mcf |
| 1.50 | 55.00 | 3.971 | $192,289.59 | $763,635.00 | 4.049 | $31,118.87 | $126,003.23 | $1.744/Mcf |
| 1.50 | 60.00 | 4.289 | $192,773.31 | $826,800.50 | 4.263 | $31,607.43 | $134,746.41 | $1.885/Mcf |
| 1.50 | 65.00 | 4.607 | $193,190.34 | $889,966.25 | 4.477 | $32,049.28 | $143,489.63 | $2.026/Mcf |
| 1.50 | 70.00 | 4.924 | $193,553.53 | $953,131.75 | 4.691 | $32,450.81 | $152,232.81 | $2.167/Mcf |
| 1.50 | 75.00 | 5.242 | $193,872.72 | $1,016,297.25 | 4.905 | $32,817.30 | $160,976.03 | $2.308/Mcf |
| 1.50 | 80.00 | 5.560 | $194,155.41 | $1,079,462.75 | 5.119 | $33,153.14 | $169,719.25 | $2.449/Mcf |
| 1.50 | 85.00 | 5.877 | $194,407.56 | $1,142,628.50 | 5.333 | $33,462.03 | $178,462.44 | $2.590/Mcf |
| 1.50 | 90.00 | 6.195 | $194,633.84 | $1,205,794.00 | 5.547 | $33,747.08 | $187,205.63 | $2.731/Mcf |
| 1.50 | 95.00 | 6.513 | $194,838.03 | $1,268,959.50 | 5.761 | $34,010.95 | $195,948.81 | $2.872/Mcf |
| 1.50 | 100.00 | 6.831 | $195,023.25 | $1,332,125.25 | 5.975 | $34,255.92 | $204,692.00 | $3.013/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 2.00 | 5.00 | 1.059 | $166,168.00 | $175,972.41 | 2.194 | $20,624.00 | $45,252.65 | $0.325/Mcf |
| 2.00 | 10.00 | 1.483 | $175,497.13 | $260,193.09 | 2.480 | $22,951.90 | $56,910.38 | $0.466/Mcf |
| 2.00 | 15.00 | 1.906 | $180,680.00 | $344,413.88 | 2.765 | $24,799.25 | $68,568.06 | $0.607/Mcf |
| 2.00 | 20.00 | 2.330 | $183,978.16 | $428,634.56 | 3.050 | $26,300.94 | $80,225.75 | $0.748/Mcf |
| 2.00 | 25.00 | 2.753 | $186,261.53 | $512,855.38 | 3.336 | $27,545.67 | $91,883.41 | $0.889/Mcf |
| 2.00 | 30.00 | 3.177 | $187,936.00 | $597,076.13 | 3.621 | $28,594.21 | $103,541.06 | $1.030/Mcf |
| 2.00 | 35.00 | 3.601 | $189,216.47 | $681,297.00 | 3.906 | $29,489.54 | $115,198.69 | $1.171/Mcf |
| 2.00 | 40.00 | 4.024 | $190,227.34 | $765,517.63 | 4.192 | $30,262.97 | $126,856.31 | $1.312/Mcf |
| 2.00 | 45.00 | 4.448 | $191,045.69 | $849,738.38 | 4.477 | $30,937.79 | $138,513.91 | $1.453/Mcf |
| 2.00 | 50.00 | 4.871 | $191,721.72 | $933,959.13 | 4.763 | $31,531.75 | $150,171.53 | $1.594/Mcf |
| 2.00 | 55.00 | 5.295 | $192,289.59 | $1,018,180.00 | 5.048 | $32,058.55 | $161,829.16 | $1.735/Mcf |
| 2.00 | 60.00 | 5.719 | $192,773.31 | $1,102,400.50 | 5.333 | $32,528.96 | $173,486.72 | $1.876/Mcf |
| 2.00 | 65.00 | 6.142 | $193,190.34 | $1,186,621.50 | 5.619 | $32,951.60 | $185,144.38 | $2.017/Mcf |
| 2.00 | 70.00 | 6.566 | $193,553.53 | $1,270,842.25 | 5.904 | $33,333.38 | $196,801.94 | $2.158/Mcf |
| 2.00 | 75.00 | 6.989 | $193,872.72 | $1,355,063.00 | 6.189 | $33,679.95 | $208,459.53 | $2.299/Mcf |
| 2.00 | 80.00 | 7.413 | $194,155.41 | $1,439,283.75 | 6.475 | $33,995.97 | $220,117.16 | $2.440/Mcf |
| 2.00 | 85.00 | 7.837 | $194,407.56 | $1,523,504.50 | 6.760 | $34,285.31 | $231,774.75 | $2.581/Mcf |
| 2.00 | 90.00 | 8.260 | $194,633.84 | $1,607,725.25 | 7.046 | $34,551.21 | $243,432.34 | $2.722/Mcf |
| 2.00 | 95.00 | 8.684 | $194,838.03 | $1,691,945.75 | 7.331 | $34,796.41 | $255,089.91 | $2.863/Mcf |
| 2.00 | 100.00 | 9.107 | $195,023.25 | $1,776,167.00 | 7.616 | $35,023.23 | $266,747.50 | $3.004/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 2.50 | 5.00 | 1.324 | $166,168.00 | $219,965.50 | 2.480 | $20,944.94 | $51,934.41 | $0.320/Mcf |
| 2.50 | 10.00 | 1.853 | $175,497.13 | $325,241.38 | 2.836 | $23,448.47 | $66,506.59 | $0.461/Mcf |
| 2.50 | 15.00 | 2.383 | $180,680.00 | $430,517.31 | 3.193 | $25,392.60 | $81,078.72 | $0.602/Mcf |
| 2.50 | 20.00 | 2.912 | $183,978.16 | $535,793.25 | 3.550 | $26,945.98 | $95,650.83 | $0.743/Mcf |
| 2.50 | 25.00 | 3.442 | $186,261.53 | $641,069.25 | 3.906 | $28,215.66 | $110,222.91 | $0.884/Mcf |
| 2.50 | 30.00 | 3.971 | $187,936.00 | $746,345.13 | 4.263 | $29,272.86 | $124,794.95 | $1.025/Mcf |
| 2.50 | 35.00 | 4.501 | $189,216.47 | $851,621.13 | 4.620 | $30,166.79 | $139,367.00 | $1.166/Mcf |
| 2.50 | 40.00 | 5.030 | $190,227.34 | $956,897.00 | 4.977 | $30,932.57 | $153,939.03 | $1.307/Mcf |
| 2.50 | 45.00 | 5.560 | $191,045.69 | $1,062,173.00 | 5.333 | $31,595.91 | $168,511.06 | $1.448/Mcf |
| 2.50 | 50.00 | 6.089 | $191,721.72 | $1,167,449.00 | 5.690 | $32,176.07 | $183,083.06 | $1.589/Mcf |
| 2.50 | 55.00 | 6.619 | $192,289.59 | $1,272,725.00 | 6.047 | $32,687.79 | $197,655.09 | $1.730/Mcf |
| 2.50 | 60.00 | 7.148 | $192,773.31 | $1,378,000.75 | 6.403 | $33,142.48 | $212,227.06 | $1.871/Mcf |
| 2.50 | 65.00 | 7.678 | $193,190.34 | $1,483,277.00 | 6.760 | $33,549.20 | $226,799.13 | $2.012/Mcf |
| 2.50 | 70.00 | 8.207 | $193,553.53 | $1,588,552.50 | 7.117 | $33,915.13 | $241,371.06 | $2.153/Mcf |
| 2.50 | 75.00 | 8.737 | $193,872.72 | $1,693,828.75 | 7.474 | $34,246.14 | $255,943.06 | $2.294/Mcf |
| 2.50 | 80.00 | 9.266 | $194,155.41 | $1,799,104.75 | 7.830 | $34,546.98 | $270,515.00 | $2.435/Mcf |
| 2.50 | 85.00 | 9.796 | $194,407.56 | $1,904,380.75 | 8.187 | $34,821.62 | $285,087.00 | $2.576/Mcf |
| 2.50 | 90.00 | 10.325 | $194,633.84 | $2,009,656.75 | 8.544 | $35,073.32 | $299,659.00 | $2.717/Mcf |
| 2.50 | 95.00 | 10.855 | $194,838.03 | $2,114,932.50 | 8.901 | $35,304.84 | $314,231.06 | $2.858/Mcf |
| 2.50 | 100.00 | 11.384 | $195,023.25 | $2,220,208.50 | 9.257 | $35,518.52 | $328,803.00 | $2.999/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 3.00 | 5.00 | 1.589 | $166,168.00 | $263,958.56 | 2.765 | $21,199.63 | $58,616.17 | $0.316/Mcf |
| 3.00 | 10.00 | 2.224 | $175,497.13 | $390,289.69 | 3.193 | $23,834.08 | $76,102.80 | $0.457/Mcf |
| 3.00 | 15.00 | 2.859 | $180,680.00 | $516,620.81 | 3.621 | $25,845.66 | $93,589.38 | $0.598/Mcf |
| 3.00 | 20.00 | 3.495 | $183,978.16 | $642,951.88 | 4.049 | $27,431.91 | $111,075.91 | $0.739/Mcf |
| 3.00 | 25.00 | 4.130 | $186,261.53 | $769,283.00 | 4.477 | $28,714.83 | $128,562.39 | $0.880/Mcf |
| 3.00 | 30.00 | 4.766 | $187,936.00 | $895,614.13 | 4.905 | $29,773.84 | $146,048.84 | $1.021/Mcf |
| 3.00 | 35.00 | 5.401 | $189,216.47 | $1,021,945.50 | 5.333 | $30,662.85 | $163,535.34 | $1.162/Mcf |
| 3.00 | 40.00 | 6.036 | $190,227.34 | $1,148,276.50 | 5.761 | $31,419.75 | $181,021.78 | $1.303/Mcf |
| 3.00 | 45.00 | 6.672 | $191,045.69 | $1,274,607.50 | 6.189 | $32,071.96 | $198,508.19 | $1.444/Mcf |
| 3.00 | 50.00 | 7.307 | $191,721.72 | $1,400,938.75 | 6.618 | $32,639.79 | $215,994.63 | $1.585/Mcf |
| 3.00 | 55.00 | 7.943 | $192,289.59 | $1,527,270.00 | 7.046 | $33,138.61 | $233,481.03 | $1.726/Mcf |
| 3.00 | 60.00 | 8.578 | $192,773.31 | $1,653,601.00 | 7.474 | $33,580.30 | $250,967.44 | $1.867/Mcf |
| 3.00 | 65.00 | 9.213 | $193,190.34 | $1,779,932.50 | 7.902 | $33,974.13 | $268,453.81 | $2.008/Mcf |
| 3.00 | 70.00 | 9.849 | $193,553.53 | $1,906,263.50 | 8.330 | $34,327.48 | $285,940.19 | $2.149/Mcf |
| 3.00 | 75.00 | 10.484 | $193,872.72 | $2,032,594.50 | 8.758 | $34,646.29 | $303,426.63 | $2.290/Mcf |
| 3.00 | 80.00 | 11.120 | $194,155.41 | $2,158,925.50 | 9.186 | $34,935.38 | $320,912.94 | $2.431/Mcf |
| 3.00 | 85.00 | 11.755 | $194,407.56 | $2,285,257.00 | 9.614 | $35,198.73 | $338,399.31 | $2.572/Mcf |
| 3.00 | 90.00 | 12.390 | $194,633.84 | $2,411,588.00 | 10.042 | $35,439.63 | $355,885.75 | $2.713/Mcf |
| 3.00 | 95.00 | 13.026 | $194,838.03 | $2,537,919.00 | 10.470 | $35,660.84 | $373,372.19 | $2.854/Mcf |
| 3.00 | 100.00 | 13.661 | $195,023.25 | $2,664,250.50 | 10.898 | $35,864.66 | $390,858.50 | $2.995/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 3.50 | 5.00 | 1.853 | $166,168.00 | $307,951.69 | 3.050 | $21,406.66 | $65,297.93 | $0.314/Mcf |
| 3.50 | 10.00 | 2.595 | $175,497.13 | $455,337.88 | 3.550 | $24,142.19 | $85,699.03 | $0.455/Mcf |
| 3.50 | 15.00 | 3.336 | $180,680.00 | $602,724.25 | 4.049 | $26,202.92 | $106,100.03 | $0.596/Mcf |
| 3.50 | 20.00 | 4.077 | $183,978.16 | $750,110.38 | 4.549 | $27,811.13 | $126,500.98 | $0.737/Mcf |
| 3.50 | 25.00 | 4.818 | $186,261.53 | $897,497.00 | 5.048 | $29,101.12 | $146,901.91 | $0.878/Mcf |
| 3.50 | 30.00 | 5.560 | $187,936.00 | $1,044,883.25 | 5.547 | $30,158.84 | $167,302.78 | $1.019/Mcf |
| 3.50 | 35.00 | 6.301 | $189,216.47 | $1,192,269.50 | 6.047 | $31,041.84 | $187,703.66 | $1.160/Mcf |
| 3.50 | 40.00 | 7.042 | $190,227.34 | $1,339,655.75 | 6.546 | $31,790.12 | $208,104.50 | $1.301/Mcf |
| 3.50 | 45.00 | 7.784 | $191,045.69 | $1,487,042.25 | 7.046 | $32,432.31 | $228,505.31 | $1.442/Mcf |
| 3.50 | 50.00 | 8.525 | $191,721.72 | $1,634,428.50 | 7.545 | $32,989.48 | $248,906.13 | $1.583/Mcf |
| 3.50 | 55.00 | 9.266 | $192,289.59 | $1,781,814.75 | 8.044 | $33,477.48 | $269,306.94 | $1.724/Mcf |
| 3.50 | 60.00 | 10.008 | $192,773.31 | $1,929,201.25 | 8.544 | $33,908.43 | $289,707.75 | $1.865/Mcf |
| 3.50 | 65.00 | 10.749 | $193,190.34 | $2,076,588.00 | 9.043 | $34,291.78 | $310,108.56 | $2.006/Mcf |
| 3.50 | 70.00 | 11.490 | $193,553.53 | $2,223,973.50 | 9.543 | $34,635.00 | $330,509.25 | $2.147/Mcf |
| 3.50 | 75.00 | 12.232 | $193,872.72 | $2,371,360.50 | 10.042 | $34,944.09 | $350,910.06 | $2.288/Mcf |
| 3.50 | 80.00 | 12.973 | $194,155.41 | $2,518,746.50 | 10.541 | $35,223.88 | $371,310.88 | $2.429/Mcf |
| 3.50 | 85.00 | 13.714 | $194,407.56 | $2,666,133.00 | 11.041 | $35,478.38 | $391,711.69 | $2.570/Mcf |
| 3.50 | 90.00 | 14.455 | $194,633.84 | $2,813,519.50 | 11.540 | $35,710.84 | $412,112.50 | $2.711/Mcf |
| 3.50 | 95.00 | 15.197 | $194,838.03 | $2,960,905.50 | 12.040 | $35,924.01 | $432,513.25 | $2.852/Mcf |
| 3.50 | 100.00 | 15.938 | $195,023.25 | $3,108,292.00 | 12.539 | $36,120.20 | $452,914.00 | $2.993/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 4.00 | 5.00 | 2.118 | $166,168.00 | $351,944.81 | 3.336 | $21,578.27 | $71,979.70 | $0.312/Mcf |
| 4.00 | 10.00 | 2.965 | $175,497.13 | $520,386.19 | 3.907 | $24,394.02 | $95,295.25 | $0.453/Mcf |
| 4.00 | 15.00 | 3.812 | $180,680.00 | $688,827.75 | 4.477 | $26,491.86 | $118,610.69 | $0.594/Mcf |
| 4.00 | 20.00 | 4.660 | $183,978.16 | $857,269.13 | 5.048 | $28,115.30 | $141,926.06 | $0.735/Mcf |
| 4.00 | 25.00 | 5.507 | $186,261.53 | $1,025,710.75 | 5.619 | $29,408.92 | $165,241.38 | $0.876/Mcf |
| 4.00 | 30.00 | 6.354 | $187,936.00 | $1,194,152.25 | 6.190 | $30,463.96 | $188,556.66 | $1.017/Mcf |
| 4.00 | 35.00 | 7.201 | $189,216.47 | $1,362,594.00 | 6.760 | $31,340.84 | $211,871.97 | $1.158/Mcf |
| 4.00 | 40.00 | 8.048 | $190,227.34 | $1,531,035.25 | 7.331 | $32,081.19 | $235,187.22 | $1.299/Mcf |
| 4.00 | 45.00 | 8.896 | $191,045.69 | $1,699,476.75 | 7.902 | $32,714.58 | $258,502.44 | $1.440/Mcf |
| 4.00 | 50.00 | 9.743 | $191,721.72 | $1,867,918.25 | 8.473 | $33,262.63 | $281,817.63 | $1.581/Mcf |
| 4.00 | 55.00 | 10.590 | $192,289.59 | $2,036,360.00 | 9.043 | $33,741.50 | $305,132.81 | $1.722/Mcf |
| 4.00 | 60.00 | 11.437 | $192,773.31 | $2,204,801.00 | 9.614 | $34,163.52 | $328,448.06 | $1.863/Mcf |
| 4.00 | 65.00 | 12.284 | $193,190.34 | $2,373,243.00 | 10.185 | $34,538.23 | $351,763.25 | $2.004/Mcf |
| 4.00 | 70.00 | 13.132 | $193,553.53 | $2,541,684.50 | 10.756 | $34,873.17 | $375,078.44 | $2.145/Mcf |
| 4.00 | 75.00 | 13.979 | $193,872.72 | $2,710,126.00 | 11.326 | $35,174.36 | $398,393.63 | $2.286/Mcf |
| 4.00 | 80.00 | 14.826 | $194,155.41 | $2,878,567.50 | 11.897 | $35,446.65 | $421,708.88 | $2.427/Mcf |
| 4.00 | 85.00 | 15.673 | $194,407.56 | $3,047,009.00 | 12.468 | $35,694.01 | $445,024.06 | $2.568/Mcf |
| 4.00 | 90.00 | 16.521 | $194,633.84 | $3,215,450.50 | 13.039 | $35,919.70 | $468,339.13 | $2.709/Mcf |
| 4.00 | 95.00 | 17.368 | $194,838.03 | $3,383,891.50 | 13.609 | $36,126.48 | $491,654.38 | $2.850/Mcf |
| 4.00 | 100.00 | 18.215 | $195,023.25 | $3,552,334.00 | 14.180 | $36,316.60 | $514,969.50 | $2.991/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 4.50 | 5.00 | 2.383 | $166,168.00 | $395,937.88 | 3.621 | $21,722.82 | $78,661.47 | $0.310/Mcf |
| 4.50 | 10.00 | 3.336 | $175,497.13 | $585,434.50 | 4.263 | $24,603.71 | $104,891.48 | $0.451/Mcf |
| 4.50 | 15.00 | 4.289 | $180,680.00 | $774,931.25 | 4.905 | $26,730.38 | $131,121.34 | $0.592/Mcf |
| 4.50 | 20.00 | 5.242 | $183,978.16 | $964,427.75 | 5.547 | $28,364.72 | $157,351.16 | $0.733/Mcf |
| 4.50 | 25.00 | 6.195 | $186,261.53 | $1,153,924.75 | 6.190 | $29,659.96 | $183,580.94 | $0.874/Mcf |
| 4.50 | 30.00 | 7.148 | $187,936.00 | $1,343,421.50 | 6.832 | $30,711.72 | $209,810.59 | $1.015/Mcf |
| 4.50 | 35.00 | 8.101 | $189,216.47 | $1,532,918.00 | 7.474 | $31,582.76 | $236,040.31 | $1.156/Mcf |
| 4.50 | 40.00 | 9.055 | $190,227.34 | $1,722,414.50 | 8.116 | $32,315.97 | $262,269.94 | $1.297/Mcf |
| 4.50 | 45.00 | 10.008 | $191,045.69 | $1,911,911.50 | 8.758 | $32,941.66 | $288,499.50 | $1.438/Mcf |
| 4.50 | 50.00 | 10.961 | $191,721.72 | $2,101,408.00 | 9.400 | $33,481.87 | $314,729.19 | $1.579/Mcf |
| 4.50 | 55.00 | 11.914 | $192,289.59 | $2,290,905.00 | 10.042 | $33,953.00 | $340,958.81 | $1.720/Mcf |
| 4.50 | 60.00 | 12.867 | $192,773.31 | $2,480,401.50 | 10.684 | $34,367.50 | $367,188.44 | $1.861/Mcf |
| 4.50 | 65.00 | 13.820 | $193,190.34 | $2,669,898.50 | 11.326 | $34,735.00 | $393,418.00 | $2.002/Mcf |
| 4.50 | 70.00 | 14.773 | $193,553.53 | $2,859,395.00 | 11.968 | $35,063.07 | $419,647.56 | $2.143/Mcf |
| 4.50 | 75.00 | 15.726 | $193,872.72 | $3,048,892.00 | 12.610 | $35,357.73 | $445,877.19 | $2.284/Mcf |
| 4.50 | 80.00 | 16.679 | $194,155.41 | $3,238,388.50 | 13.253 | $35,623.84 | $472,106.75 | $2.425/Mcf |
| 4.50 | 85.00 | 17.632 | $194,407.56 | $3,427,885.00 | 13.895 | $35,865.35 | $498,336.38 | $2.566/Mcf |
| 4.50 | 90.00 | 18.586 | $194,633.84 | $3,617,382.50 | 14.537 | $36,085.52 | $524,565.88 | $2.707/Mcf |
| 4.50 | 95.00 | 19.539 | $194,838.03 | $3,806,878.50 | 15.179 | $36,287.07 | $550,795.50 | $2.848/Mcf |
| 4.50 | 100.00 | 20.492 | $195,023.25 | $3,996,375.50 | 15.821 | $36,472.27 | $577,025.13 | $2.989/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 5.00 | 5.00 | 2.648 | $166,168.00 | $439,931.00 | 3.907 | $21,846.27 | $85,343.25 | $0.309/Mcf |
| 5.00 | 10.00 | 3.707 | $175,497.13 | $650,482.75 | 4.620 | $24,781.02 | $114,487.70 | $0.450/Mcf |
| 5.00 | 15.00 | 4.766 | $180,680.00 | $861,034.63 | 5.333 | $26,930.60 | $143,632.00 | $0.591/Mcf |
| 5.00 | 20.00 | 5.825 | $183,978.16 | $1,071,586.50 | 6.047 | $28,572.93 | $172,776.25 | $0.732/Mcf |
| 5.00 | 25.00 | 6.884 | $186,261.53 | $1,282,138.50 | 6.760 | $29,868.61 | $201,920.41 | $0.873/Mcf |
| 5.00 | 30.00 | 7.943 | $187,936.00 | $1,492,690.25 | 7.474 | $30,916.91 | $231,064.53 | $1.014/Mcf |
| 5.00 | 35.00 | 9.002 | $189,216.47 | $1,703,242.25 | 8.187 | $31,782.51 | $260,208.63 | $1.155/Mcf |
| 5.00 | 40.00 | 10.061 | $190,227.34 | $1,913,794.00 | 8.901 | $32,509.34 | $289,352.69 | $1.296/Mcf |
| 5.00 | 45.00 | 11.120 | $191,045.69 | $2,124,346.00 | 9.614 | $33,128.30 | $318,496.69 | $1.437/Mcf |
| 5.00 | 50.00 | 12.179 | $191,721.72 | $2,334,898.00 | 10.327 | $33,661.73 | $347,640.75 | $1.578/Mcf |
| 5.00 | 55.00 | 13.238 | $192,289.59 | $2,545,450.00 | 11.041 | $34,126.23 | $376,784.75 | $1.719/Mcf |
| 5.00 | 60.00 | 14.297 | $192,773.31 | $2,756,001.50 | 11.754 | $34,534.34 | $405,928.75 | $1.860/Mcf |
| 5.00 | 65.00 | 15.356 | $193,190.34 | $2,966,554.00 | 12.468 | $34,895.74 | $435,072.75 | $2.001/Mcf |
| 5.00 | 70.00 | 16.415 | $193,553.53 | $3,177,105.00 | 13.181 | $35,218.02 | $464,216.81 | $2.142/Mcf |
| 5.00 | 75.00 | 17.474 | $193,872.72 | $3,387,657.50 | 13.895 | $35,507.20 | $493,360.75 | $2.283/Mcf |
| 5.00 | 80.00 | 18.533 | $194,155.41 | $3,598,209.50 | 14.608 | $35,768.14 | $522,504.69 | $2.424/Mcf |
| 5.00 | 85.00 | 19.592 | $194,407.56 | $3,808,761.50 | 15.322 | $36,004.78 | $551,648.75 | $2.565/Mcf |
| 5.00 | 90.00 | 20.651 | $194,633.84 | $4,019,313.50 | 16.035 | $36,220.36 | $580,792.75 | $2.706/Mcf |
| 5.00 | 95.00 | 21.710 | $194,838.03 | $4,229,865.00 | 16.748 | $36,417.57 | $609,936.63 | $2.847/Mcf |
| 5.00 | 100.00 | 22.769 | $195,023.25 | $4,440,417.00 | 17.462 | $36,598.67 | $639,080.63 | $2.988/Mcf |

| $CO_2$ Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | $CO_2$ Cost |
|---|---|---|---|---|---|---|---|---|
| 5.50 | 5.00 | 2.912 | $166,168.00 | $483,924.06 | 4.192 | $21,952.89 | $92,025.00 | $0.308/Mcf |
| 5.50 | 10.00 | 4.077 | $175,497.13 | $715,531.00 | 4.977 | $24,932.91 | $124,083.94 | $0.449/Mcf |
| 5.50 | 15.00 | 5.242 | $180,680.00 | $947,138.25 | 5.761 | $27,101.08 | $156,142.69 | $0.590/Mcf |
| 5.50 | 20.00 | 6.407 | $183,978.16 | $1,178,745.00 | 6.546 | $28,749.38 | $188,201.34 | $0.731/Mcf |
| 5.50 | 25.00 | 7.572 | $186,261.53 | $1,410,352.50 | 7.331 | $30,044.77 | $220,259.91 | $0.872/Mcf |
| 5.50 | 30.00 | 8.737 | $187,936.00 | $1,641,959.25 | 8.116 | $31,089.63 | $252,318.44 | $1.013/Mcf |
| 5.50 | 35.00 | 9.902 | $189,216.47 | $1,873,566.50 | 8.901 | $31,950.24 | $284,376.94 | $1.154/Mcf |
| 5.50 | 40.00 | 11.067 | $190,227.34 | $2,105,173.50 | 9.685 | $32,671.38 | $316,435.38 | $1.295/Mcf |
| 5.50 | 45.00 | 12.232 | $191,045.69 | $2,336,780.50 | 10.470 | $33,284.41 | $348,493.81 | $1.436/Mcf |
| 5.50 | 50.00 | 13.396 | $191,721.72 | $2,568,387.50 | 11.255 | $33,811.95 | $380,552.31 | $1.577/Mcf |
| 5.50 | 55.00 | 14.561 | $192,289.59 | $2,799,995.00 | 12.040 | $34,270.71 | $412,610.69 | $1.718/Mcf |
| 5.50 | 60.00 | 15.726 | $192,773.31 | $3,031,601.50 | 12.825 | $34,673.34 | $444,669.13 | $1.859/Mcf |
| 5.50 | 65.00 | 16.891 | $193,190.34 | $3,263,209.50 | 13.609 | $35,029.52 | $476,727.50 | $2.000/Mcf |
| 5.50 | 70.00 | 18.056 | $193,553.53 | $3,494,816.50 | 14.394 | $35,346.86 | $508,785.88 | $2.141/Mcf |
| 5.50 | 75.00 | 19.221 | $193,872.72 | $3,726,423.50 | 15.179 | $35,631.38 | $540,844.13 | $2.282/Mcf |
| 5.50 | 80.00 | 20.386 | $194,155.41 | $3,958,030.00 | 15.964 | $35,887.94 | $572,902.63 | $2.423/Mcf |
| 5.50 | 85.00 | 21.551 | $194,407.56 | $4,189,637.50 | 16.748 | $36,120.45 | $604,961.00 | $2.564/Mcf |
| 5.50 | 90.00 | 22.716 | $194,633.84 | $4,421,245.00 | 17.533 | $36,332.15 | $637,019.38 | $2.705/Mcf |
| 5.50 | 95.00 | 23.881 | $194,838.03 | $4,652,852.00 | 18.318 | $36,525.70 | $669,077.63 | $2.846/Mcf |
| 5.50 | 100.00 | 25.046 | $195,023.25 | $4,884,459.00 | 19.103 | $36,703.35 | $701,136.00 | $2.987/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 6.00 | 5.00 | 3.177 | $166,168.00 | $527,917.13 | 4.477 | $22,045.93 | $98,706.77 | $0.307/Mcf |
| 6.00 | 10.00 | 4.448 | $175,497.13 | $780,579.38 | 5.333 | $25,064.48 | $133,680.16 | $0.448/Mcf |
| 6.00 | 15.00 | 5.719 | $180,680.00 | $1,033,241.63 | 6.190 | $27,247.97 | $168,653.34 | $0.589/Mcf |
| 6.00 | 20.00 | 6.989 | $183,978.16 | $1,285,903.75 | 7.046 | $28,900.81 | $203,626.44 | $0.730/Mcf |
| 6.00 | 25.00 | 8.260 | $186,261.53 | $1,538,566.00 | 7.902 | $30,195.48 | $238,599.41 | $0.871/Mcf |
| 6.00 | 30.00 | 9.531 | $187,936.00 | $1,791,228.25 | 8.758 | $31,237.03 | $273,572.38 | $1.012/Mcf |
| 6.00 | 35.00 | 10.802 | $189,216.47 | $2,043,891.00 | 9.614 | $32,093.08 | $308,545.25 | $1.153/Mcf |
| 6.00 | 40.00 | 12.073 | $190,227.34 | $2,296,553.00 | 10.470 | $32,809.13 | $343,518.13 | $1.294/Mcf |
| 6.00 | 45.00 | 13.343 | $191,045.69 | $2,549,215.00 | 11.326 | $33,416.92 | $378,491.00 | $1.435/Mcf |
| 6.00 | 50.00 | 14.614 | $191,721.72 | $2,801,877.50 | 12.182 | $33,939.30 | $413,463.88 | $1.576/Mcf |
| 6.00 | 55.00 | 15.885 | $192,289.59 | $3,054,540.00 | 13.039 | $34,393.06 | $448,436.63 | $1.717/Mcf |
| 6.00 | 60.00 | 17.156 | $192,773.31 | $3,307,202.00 | 13.895 | $34,790.91 | $483,409.38 | $1.858/Mcf |
| 6.00 | 65.00 | 18.427 | $193,190.34 | $3,559,865.00 | 14.751 | $35,142.59 | $518,382.25 | $1.999/Mcf |
| 6.00 | 70.00 | 19.698 | $193,553.53 | $3,812,527.00 | 15.607 | $35,455.67 | $553,355.00 | $2.140/Mcf |
| 6.00 | 75.00 | 20.968 | $193,872.72 | $4,065,189.00 | 16.463 | $35,736.20 | $588,327.75 | $2.281/Mcf |
| 6.00 | 80.00 | 22.239 | $194,155.41 | $4,317,851.00 | 17.319 | $35,988.98 | $623,300.63 | $2.422/Mcf |
| 6.00 | 85.00 | 23.510 | $194,407.56 | $4,570,514.00 | 18.175 | $36,217.96 | $658,273.38 | $2.563/Mcf |
| 6.00 | 90.00 | 24.781 | $194,633.84 | $4,823,176.00 | 19.031 | $36,426.33 | $693,246.00 | $2.704/Mcf |
| 6.00 | 95.00 | 26.052 | $194,838.03 | $5,075,838.00 | 19.888 | $36,616.77 | $728,218.88 | $2.845/Mcf |
| 6.00 | 100.00 | 27.322 | $195,023.25 | $5,328,501.00 | 20.744 | $36,791.48 | $763,191.63 | $2.986/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 6.50 | 5.00 | 3.442 | $166,168.00 | $571,910.38 | 4.763 | $22,127.82 | $105,388.55 | $0.306/Mcf |
| 6.50 | 10.00 | 4.818 | $175,497.13 | $845,627.63 | 5.690 | $25,179.56 | $143,276.41 | $0.447/Mcf |
| 6.50 | 15.00 | 6.195 | $180,680.00 | $1,119,345.25 | 6.618 | $27,375.86 | $181,164.03 | $0.588/Mcf |
| 6.50 | 20.00 | 7.572 | $183,978.16 | $1,393,062.50 | 7.545 | $29,032.19 | $219,051.53 | $0.729/Mcf |
| 6.50 | 25.00 | 8.949 | $186,261.53 | $1,666,780.00 | 8.473 | $30,325.89 | $256,938.94 | $0.870/Mcf |
| 6.50 | 30.00 | 10.325 | $187,936.00 | $1,940,497.75 | 9.400 | $31,364.29 | $294,826.31 | $1.011/Mcf |
| 6.50 | 35.00 | 11.702 | $189,216.47 | $2,214,215.00 | 10.328 | $32,216.18 | $332,713.63 | $1.152/Mcf |
| 6.50 | 40.00 | 13.079 | $190,227.34 | $2,487,932.00 | 11.255 | $32,927.66 | $370,600.94 | $1.293/Mcf |
| 6.50 | 45.00 | 14.455 | $191,045.69 | $2,761,649.50 | 12.182 | $33,530.81 | $408,488.19 | $1.434/Mcf |
| 6.50 | 50.00 | 15.832 | $191,721.72 | $3,035,367.50 | 13.110 | $34,048.62 | $446,375.38 | $1.575/Mcf |
| 6.50 | 55.00 | 17.209 | $192,289.59 | $3,309,085.00 | 14.037 | $34,498.01 | $484,262.69 | $1.716/Mcf |
| 6.50 | 60.00 | 18.586 | $192,773.31 | $3,582,802.00 | 14.965 | $34,891.68 | $522,149.75 | $1.857/Mcf |
| 6.50 | 65.00 | 19.962 | $193,190.34 | $3,856,520.50 | 15.892 | $35,239.41 | $560,037.00 | $1.998/Mcf |
| 6.50 | 70.00 | 21.339 | $193,553.53 | $4,130,237.00 | 16.820 | $35,548.80 | $597,924.13 | $2.139/Mcf |
| 6.50 | 75.00 | 22.716 | $193,872.72 | $4,403,954.00 | 17.747 | $35,825.84 | $635,811.25 | $2.280/Mcf |
| 6.50 | 80.00 | 24.092 | $194,155.41 | $4,677,672.00 | 18.675 | $36,075.36 | $673,698.50 | $2.421/Mcf |
| 6.50 | 85.00 | 25.469 | $194,407.56 | $4,951,390.00 | 19.602 | $36,301.27 | $711,585.63 | $2.562/Mcf |
| 6.50 | 90.00 | 26.846 | $194,633.84 | $5,225,108.00 | 20.530 | $36,506.77 | $749,472.75 | $2.703/Mcf |
| 6.50 | 95.00 | 28.223 | $194,838.03 | $5,498,825.00 | 21.457 | $36,694.51 | $787,359.88 | $2.844/Mcf |
| 6.50 | 100.00 | 29.599 | $195,023.25 | $5,772,542.00 | 22.385 | $36,866.69 | $825,247.13 | $2.985/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 7.00 | 5.00 | 3.707 | $166,168.00 | $615,903.38 | 5.048 | $22,200.45 | $112,070.33 | $0.306/Mcf |
| 7.00 | 10.00 | 5.189 | $175,497.13 | $910,675.75 | 6.047 | $25,281.05 | $152,872.63 | $0.447/Mcf |
| 7.00 | 15.00 | 6.672 | $180,680.00 | $1,205,448.50 | 7.046 | $27,488.21 | $193,674.69 | $0.588/Mcf |
| 7.00 | 20.00 | 8.154 | $183,978.16 | $1,500,220.75 | 8.045 | $29,147.26 | $234,476.59 | $0.729/Mcf |
| 7.00 | 25.00 | 9.637 | $186,261.53 | $1,794,994.00 | 9.043 | $30,439.83 | $275,278.44 | $0.870/Mcf |
| 7.00 | 30.00 | 11.120 | $187,936.00 | $2,089,766.50 | 10.042 | $31,475.27 | $316,080.19 | $1.011/Mcf |
| 7.00 | 35.00 | 12.602 | $189,216.47 | $2,384,539.00 | 11.041 | $32,323.37 | $356,881.94 | $1.152/Mcf |
| 7.00 | 40.00 | 14.085 | $190,227.34 | $2,679,311.50 | 12.040 | $33,030.74 | $397,683.56 | $1.293/Mcf |
| 7.00 | 45.00 | 15.567 | $191,045.69 | $2,974,084.50 | 13.039 | $33,629.74 | $438,485.25 | $1.434/Mcf |
| 7.00 | 50.00 | 17.050 | $191,721.72 | $3,268,857.00 | 14.037 | $34,143.50 | $479,286.88 | $1.575/Mcf |
| 7.00 | 55.00 | 18.533 | $192,289.59 | $3,563,629.50 | 15.036 | $34,589.00 | $520,088.50 | $1.716/Mcf |
| 7.00 | 60.00 | 20.015 | $192,773.31 | $3,858,402.50 | 16.035 | $34,979.00 | $560,890.13 | $1.857/Mcf |
| 7.00 | 65.00 | 21.498 | $193,190.34 | $4,153,176.00 | 17.034 | $35,323.27 | $601,691.75 | $1.998/Mcf |
| 7.00 | 70.00 | 22.980 | $193,553.53 | $4,447,947.00 | 18.033 | $35,629.39 | $642,493.38 | $2.139/Mcf |
| 7.00 | 75.00 | 24.463 | $193,872.72 | $4,742,721.00 | 19.031 | $35,903.38 | $683,294.75 | $2.280/Mcf |
| 7.00 | 80.00 | 25.946 | $194,155.41 | $5,037,493.00 | 20.030 | $36,150.05 | $724,096.38 | $2.421/Mcf |
| 7.00 | 85.00 | 27.428 | $194,407.56 | $5,332,266.00 | 21.029 | $36,373.28 | $764,898.00 | $2.562/Mcf |
| 7.00 | 90.00 | 28.911 | $194,633.84 | $5,627,039.00 | 22.028 | $36,576.27 | $805,699.63 | $2.703/Mcf |
| 7.00 | 95.00 | 30.394 | $194,838.03 | $5,921,811.00 | 23.027 | $36,761.65 | $846,501.00 | $2.844/Mcf |
| 7.00 | 100.00 | 31.876 | $195,023.25 | $6,216,584.00 | 24.026 | $36,931.62 | $887,302.63 | $2.985/Mcf |

| $CO_2$ Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | $CO_2$ Cost |
|---|---|---|---|---|---|---|---|---|
| 7.50 | 5.00 | 3.971 | $166,168.00 | $659,896.50 | 5.334 | $22,265.30 | $118,752.09 | $0.305/Mcf |
| 7.50 | 10.00 | 5.560 | $175,497.13 | $975,724.25 | 6.404 | $25,371.24 | $162,468.88 | $0.446/Mcf |
| 7.50 | 15.00 | 7.148 | $180,680.00 | $1,291,552.00 | 7.474 | $27,587.68 | $206,185.34 | $0.587/Mcf |
| 7.50 | 20.00 | 8.737 | $183,978.16 | $1,607,379.75 | 8.544 | $29,248.88 | $249,901.69 | $0.728/Mcf |
| 7.50 | 25.00 | 10.325 | $186,261.53 | $1,923,207.50 | 9.614 | $30,540.25 | $293,617.94 | $0.869/Mcf |
| 7.50 | 30.00 | 11.914 | $187,936.00 | $2,239,035.50 | 10.684 | $31,572.91 | $337,334.13 | $1.010/Mcf |
| 7.50 | 35.00 | 13.502 | $189,216.47 | $2,554,863.50 | 11.754 | $32,417.55 | $381,050.25 | $1.151/Mcf |
| 7.50 | 40.00 | 15.091 | $190,227.34 | $2,870,691.00 | 12.825 | $33,121.21 | $424,766.31 | $1.292/Mcf |
| 7.50 | 45.00 | 16.679 | $191,045.69 | $3,186,518.50 | 13.895 | $33,716.48 | $468,482.44 | $1.433/Mcf |
| 7.50 | 50.00 | 18.268 | $191,721.72 | $3,502,347.00 | 14.965 | $34,226.62 | $512,198.44 | $1.574/Mcf |
| 7.50 | 55.00 | 19.856 | $192,289.59 | $3,818,175.00 | 16.035 | $34,668.66 | $555,914.50 | $1.715/Mcf |
| 7.50 | 60.00 | 21.445 | $192,773.31 | $4,134,003.00 | 17.105 | $35,055.39 | $599,630.38 | $1.856/Mcf |
| 7.50 | 65.00 | 23.033 | $193,190.34 | $4,449,831.00 | 18.175 | $35,396.58 | $643,346.38 | $1.997/Mcf |
| 7.50 | 70.00 | 24.622 | $193,553.53 | $4,765,658.00 | 19.246 | $35,699.82 | $687,062.25 | $2.138/Mcf |
| 7.50 | 75.00 | 26.210 | $193,872.72 | $5,081,486.00 | 20.316 | $35,971.12 | $730,778.38 | $2.279/Mcf |
| 7.50 | 80.00 | 27.799 | $194,155.41 | $5,397,314.00 | 21.386 | $36,215.27 | $774,494.38 | $2.420/Mcf |
| 7.50 | 85.00 | 29.387 | $194,407.56 | $5,713,142.00 | 22.456 | $36,436.14 | $818,210.38 | $2.561/Mcf |
| 7.50 | 90.00 | 30.976 | $194,633.84 | $6,028,970.00 | 23.526 | $36,636.91 | $861,926.13 | $2.702/Mcf |
| 7.50 | 95.00 | 32.564 | $194,838.03 | $6,344,798.00 | 24.596 | $36,820.23 | $905,642.25 | $2.843/Mcf |
| 7.50 | 100.00 | 34.153 | $195,023.25 | $6,660,626.00 | 25.666 | $36,988.25 | $949,358.25 | $2.984/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 8.00 | 5.00 | 4.236 | $166,168.00 | $703,889.63 | 5.619 | $22,323.57 | $125,433.89 | $0.305/Mcf |
| 8.00 | 10.00 | 5.930 | $175,497.13 | $1,040,772.38 | 6.760 | $25,451.91 | $172,065.09 | $0.446/Mcf |
| 8.00 | 15.00 | 7.625 | $180,680.00 | $1,377,655.50 | 7.902 | $27,676.38 | $218,696.03 | $0.587/Mcf |
| 8.00 | 20.00 | 9.319 | $183,978.16 | $1,714,538.25 | 9.043 | $29,339.27 | $265,326.81 | $0.728/Mcf |
| 8.00 | 25.00 | 11.014 | $186,261.53 | $2,051,421.50 | 10.185 | $30,629.41 | $311,957.50 | $0.869/Mcf |
| 8.00 | 30.00 | 12.708 | $187,936.00 | $2,388,304.50 | 11.326 | $31,659.49 | $358,588.06 | $1.010/Mcf |
| 8.00 | 35.00 | 14.402 | $189,216.47 | $2,725,188.00 | 12.468 | $32,500.95 | $405,218.63 | $1.151/Mcf |
| 8.00 | 40.00 | 16.097 | $190,227.34 | $3,062,070.50 | 13.609 | $33,201.24 | $451,849.06 | $1.292/Mcf |
| 8.00 | 45.00 | 17.791 | $191,045.69 | $3,398,953.50 | 14.751 | $33,793.16 | $498,479.56 | $1.433/Mcf |
| 8.00 | 50.00 | 19.486 | $191,721.72 | $3,735,836.50 | 15.892 | $34,300.03 | $545,109.88 | $1.574/Mcf |
| 8.00 | 55.00 | 21.180 | $192,289.59 | $4,072,720.00 | 17.034 | $34,738.98 | $591,740.38 | $1.715/Mcf |
| 8.00 | 60.00 | 22.875 | $192,773.31 | $4,409,602.00 | 18.175 | $35,122.78 | $638,370.63 | $1.856/Mcf |
| 8.00 | 65.00 | 24.569 | $193,190.34 | $4,746,486.00 | 19.317 | $35,461.23 | $685,001.13 | $1.997/Mcf |
| 8.00 | 70.00 | 26.263 | $193,553.53 | $5,083,369.00 | 20.458 | $35,761.91 | $731,631.50 | $2.138/Mcf |
| 8.00 | 75.00 | 27.958 | $193,872.72 | $5,420,252.00 | 21.600 | $36,030.80 | $778,262.00 | $2.279/Mcf |
| 8.00 | 80.00 | 29.652 | $194,155.41 | $5,757,135.00 | 22.741 | $36,272.70 | $824,892.13 | $2.420/Mcf |
| 8.00 | 85.00 | 31.347 | $194,407.56 | $6,094,018.00 | 23.883 | $36,491.48 | $871,522.75 | $2.561/Mcf |
| 8.00 | 90.00 | 33.041 | $194,633.84 | $6,430,901.00 | 25.024 | $36,690.30 | $918,152.88 | $2.702/Mcf |
| 8.00 | 95.00 | 34.735 | $194,838.03 | $6,767,783.00 | 26.166 | $36,871.77 | $964,783.38 | $2.843/Mcf |
| 8.00 | 100.00 | 36.430 | $195,023.25 | $7,104,668.00 | 27.307 | $37,038.07 | $1,011,413.63 | $2.984/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 8.50 | 5.00 | 4.501 | $166,168.00 | $747,882.75 | 5.904 | $22,376.20 | $132,115.66 | $0.304/Mcf |
| 8.50 | 10.00 | 6.301 | $175,497.13 | $1,105,820.75 | 7.117 | $25,524.49 | $181,661.31 | $0.445/Mcf |
| 8.50 | 15.00 | 8.101 | $180,680.00 | $1,463,758.75 | 8.330 | $27,755.97 | $231,206.72 | $0.586/Mcf |
| 8.50 | 20.00 | 9.902 | $183,978.16 | $1,821,697.00 | 9.543 | $29,420.21 | $280,751.88 | $0.727/Mcf |
| 8.50 | 25.00 | 11.702 | $186,261.53 | $2,179,635.50 | 10.756 | $30,709.11 | $330,297.00 | $0.868/Mcf |
| 8.50 | 30.00 | 13.502 | $187,936.00 | $2,537,574.00 | 11.969 | $31,736.77 | $379,842.00 | $1.009/Mcf |
| 8.50 | 35.00 | 15.303 | $189,216.47 | $2,895,512.00 | 13.181 | $32,575.32 | $429,386.94 | $1.150/Mcf |
| 8.50 | 40.00 | 17.103 | $190,227.34 | $3,253,449.50 | 14.394 | $33,272.55 | $478,931.75 | $1.291/Mcf |
| 8.50 | 45.00 | 18.903 | $191,045.69 | $3,611,388.00 | 15.607 | $33,861.41 | $528,476.63 | $1.432/Mcf |
| 8.50 | 50.00 | 20.704 | $191,721.72 | $3,969,326.00 | 16.820 | $34,365.36 | $578,021.63 | $1.573/Mcf |
| 8.50 | 55.00 | 22.504 | $192,289.59 | $4,327,265.00 | 18.033 | $34,801.51 | $627,566.25 | $1.714/Mcf |
| 8.50 | 60.00 | 24.304 | $192,773.31 | $4,685,203.00 | 19.246 | $35,182.69 | $677,111.13 | $1.855/Mcf |
| 8.50 | 65.00 | 26.105 | $193,190.34 | $5,043,141.00 | 20.458 | $35,518.67 | $726,656.00 | $1.996/Mcf |
| 8.50 | 70.00 | 27.905 | $193,553.53 | $5,401,080.00 | 21.671 | $35,817.05 | $776,200.75 | $2.137/Mcf |
| 8.50 | 75.00 | 29.705 | $193,872.72 | $5,759,018.00 | 22.884 | $36,083.79 | $825,745.50 | $2.278/Mcf |
| 8.50 | 80.00 | 31.505 | $194,155.41 | $6,116,956.00 | 24.097 | $36,323.69 | $875,290.38 | $2.419/Mcf |
| 8.50 | 85.00 | 33.306 | $194,407.56 | $6,474,895.00 | 25.310 | $36,540.59 | $924,835.13 | $2.560/Mcf |
| 8.50 | 90.00 | 35.106 | $194,633.84 | $6,832,832.00 | 26.523 | $36,737.66 | $974,379.75 | $2.701/Mcf |
| 8.50 | 95.00 | 36.906 | $194,838.03 | $7,190,770.00 | 27.735 | $36,917.48 | $1,023,924.38 | $2.842/Mcf |
| 8.50 | 100.00 | 38.707 | $195,023.25 | $7,548,710.00 | 28.948 | $37,082.25 | $1,073,469.25 | $2.983/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 9.00 | 5.00 | 4.766 | $166,168.00 | $791,875.75 | 6.190 | $22,423.98 | $138,797.44 | $0.304/Mcf |
| 9.00 | 10.00 | 6.672 | $175,497.13 | $1,170,869.00 | 7.474 | $25,590.15 | $191,257.59 | $0.445/Mcf |
| 9.00 | 15.00 | 8.578 | $180,680.00 | $1,549,862.50 | 8.758 | $27,827.77 | $243,717.38 | $0.586/Mcf |
| 9.00 | 20.00 | 10.484 | $183,978.16 | $1,928,855.50 | 10.042 | $29,493.09 | $296,177.00 | $0.727/Mcf |
| 9.00 | 25.00 | 12.390 | $186,261.53 | $2,307,849.50 | 11.326 | $30,780.77 | $348,636.50 | $0.868/Mcf |
| 9.00 | 30.00 | 14.297 | $187,936.00 | $2,686,843.00 | 12.611 | $31,806.18 | $401,095.94 | $1.009/Mcf |
| 9.00 | 35.00 | 16.203 | $189,216.47 | $3,065,836.00 | 13.895 | $32,642.05 | $453,555.31 | $1.150/Mcf |
| 9.00 | 40.00 | 18.109 | $190,227.34 | $3,444,829.00 | 15.179 | $33,336.48 | $506,014.56 | $1.291/Mcf |
| 9.00 | 45.00 | 20.015 | $191,045.69 | $3,823,823.00 | 16.463 | $33,922.58 | $558,473.88 | $1.432/Mcf |
| 9.00 | 50.00 | 21.921 | $191,721.72 | $4,202,816.00 | 17.747 | $34,423.85 | $610,933.13 | $1.573/Mcf |
| 9.00 | 55.00 | 23.828 | $192,289.59 | $4,581,810.00 | 19.032 | $34,857.47 | $663,392.13 | $1.714/Mcf |
| 9.00 | 60.00 | 25.734 | $192,773.31 | $4,960,803.00 | 20.316 | $35,236.27 | $715,851.38 | $1.855/Mcf |
| 9.00 | 65.00 | 27.640 | $193,190.34 | $5,339,797.00 | 21.600 | $35,570.03 | $768,310.63 | $1.996/Mcf |
| 9.00 | 70.00 | 29.546 | $193,553.53 | $5,718,790.00 | 22.884 | $35,866.34 | $820,769.88 | $2.137/Mcf |
| 9.00 | 75.00 | 31.453 | $193,872.72 | $6,097,784.00 | 24.168 | $36,131.14 | $873,228.88 | $2.278/Mcf |
| 9.00 | 80.00 | 33.359 | $194,155.41 | $6,476,777.00 | 25.453 | $36,369.23 | $925,688.13 | $2.419/Mcf |
| 9.00 | 85.00 | 35.265 | $194,407.56 | $6,855,770.00 | 26.737 | $36,584.45 | $978,147.25 | $2.560/Mcf |
| 9.00 | 90.00 | 37.171 | $194,633.84 | $7,234,765.00 | 28.021 | $36,779.95 | $1,030,606.63 | $2.701/Mcf |
| 9.00 | 95.00 | 39.077 | $194,838.03 | $7,613,757.00 | 29.305 | $36,958.30 | $1,083,065.50 | $2.842/Mcf |
| 9.00 | 100.00 | 40.984 | $195,023.25 | $7,992,751.00 | 30.589 | $37,121.68 | $1,135,524.50 | $2.983/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 9.50 | 5.00 | 5.030 | $166,168.00 | $835,869.00 | 6.475 | $22,467.55 | $145,479.22 | $0.304/Mcf |
| 9.50 | 10.00 | 7.042 | $175,497.13 | $1,235,917.25 | 7.831 | $25,649.82 | $200,853.81 | $0.445/Mcf |
| 9.50 | 15.00 | 9.054 | $180,680.00 | $1,635,966.00 | 9.186 | $27,892.88 | $256,228.03 | $0.586/Mcf |
| 9.50 | 20.00 | 11.067 | $183,978.16 | $2,036,014.50 | 10.542 | $29,559.07 | $311,602.06 | $0.727/Mcf |
| 9.50 | 25.00 | 13.079 | $186,261.53 | $2,436,063.00 | 11.897 | $30,845.56 | $366,976.00 | $0.868/Mcf |
| 9.50 | 30.00 | 15.091 | $187,936.00 | $2,836,111.50 | 13.253 | $31,868.87 | $422,349.88 | $1.009/Mcf |
| 9.50 | 35.00 | 17.103 | $189,216.47 | $3,236,160.50 | 14.608 | $32,702.27 | $477,723.63 | $1.150/Mcf |
| 9.50 | 40.00 | 19.115 | $190,227.34 | $3,636,209.00 | 15.964 | $33,394.13 | $533,097.25 | $1.291/Mcf |
| 9.50 | 45.00 | 21.127 | $191,045.69 | $4,036,257.50 | 17.319 | $33,977.69 | $588,470.88 | $1.432/Mcf |
| 9.50 | 50.00 | 23.139 | $191,721.72 | $4,436,306.00 | 18.675 | $34,476.53 | $643,844.50 | $1.573/Mcf |
| 9.50 | 55.00 | 25.151 | $192,289.59 | $4,836,354.00 | 20.030 | $34,907.85 | $699,218.13 | $1.714/Mcf |
| 9.50 | 60.00 | 27.164 | $192,773.31 | $5,236,403.00 | 21.386 | $35,284.50 | $754,591.75 | $1.855/Mcf |
| 9.50 | 65.00 | 29.176 | $193,190.34 | $5,636,452.00 | 22.741 | $35,616.24 | $809,965.38 | $1.996/Mcf |
| 9.50 | 70.00 | 31.188 | $193,553.53 | $6,036,500.00 | 24.097 | $35,910.66 | $865,339.00 | $2.137/Mcf |
| 9.50 | 75.00 | 33.200 | $193,872.72 | $6,436,550.00 | 25.453 | $36,173.72 | $920,712.50 | $2.278/Mcf |
| 9.50 | 80.00 | 35.212 | $194,155.41 | $6,836,598.00 | 26.808 | $36,410.18 | $976,086.25 | $2.419/Mcf |
| 9.50 | 85.00 | 37.224 | $194,407.56 | $7,236,647.00 | 28.164 | $36,623.88 | $1,031,459.75 | $2.560/Mcf |
| 9.50 | 90.00 | 39.236 | $194,633.84 | $7,636,696.00 | 29.519 | $36,817.94 | $1,086,833.00 | $2.701/Mcf |
| 9.50 | 95.00 | 41.248 | $194,838.03 | $8,036,744.00 | 30.875 | $36,994.97 | $1,142,206.75 | $2.842/Mcf |
| 9.50 | 100.00 | 43.260 | $195,023.25 | $8,436,792.00 | 32.230 | $37,157.10 | $1,197,580.25 | $2.983/Mcf |

| CO2 Demand (MMcfd) | Haul Distance (miles) | Tractors Required | Tractor Annual Unit Cost | Tractor Cost Per Year | Trailers Required | Trailer Annual Unit Cost | Trailer Cost Per Year | CO2 Cost |
|---|---|---|---|---|---|---|---|---|
| 10.00 | 5.00 | 5.295 | $166,168.00 | $879,862.00 | 6.760 | $22,507.44 | $152,160.97 | $0.304/Mcf |
| 10.00 | 10.00 | 7.413 | $175,497.13 | $1,300,965.50 | 8.187 | $25,704.29 | $210,450.00 | $0.445/Mcf |
| 10.00 | 15.00 | 9.531 | $180,680.00 | $1,722,069.25 | 9.614 | $27,952.20 | $268,738.69 | $0.586/Mcf |
| 10.00 | 20.00 | 11.649 | $183,978.16 | $2,143,173.00 | 11.041 | $29,619.07 | $327,027.19 | $0.727/Mcf |
| 10.00 | 25.00 | 13.767 | $186,261.53 | $2,564,277.00 | 12.468 | $30,904.42 | $385,315.56 | $0.868/Mcf |
| 10.00 | 30.00 | 15.885 | $187,936.00 | $2,985,380.50 | 13.895 | $31,925.77 | $443,603.81 | $1.009/Mcf |
| 10.00 | 35.00 | 18.003 | $189,216.47 | $3,406,484.50 | 15.322 | $32,756.88 | $501,891.94 | $1.150/Mcf |
| 10.00 | 40.00 | 20.121 | $190,227.34 | $3,827,588.00 | 16.749 | $33,446.38 | $560,180.00 | $1.291/Mcf |
| 10.00 | 45.00 | 22.239 | $191,045.69 | $4,248,692.00 | 18.175 | $34,027.61 | $618,468.13 | $1.432/Mcf |
| 10.00 | 50.00 | 24.357 | $191,721.72 | $4,669,796.00 | 19.602 | $34,524.23 | $676,756.25 | $1.573/Mcf |
| 10.00 | 55.00 | 26.475 | $192,289.59 | $5,090,900.00 | 21.029 | $34,953.45 | $735,044.25 | $1.714/Mcf |
| 10.00 | 60.00 | 28.593 | $192,773.31 | $5,512,003.00 | 22.456 | $35,328.13 | $793,332.13 | $1.855/Mcf |
| 10.00 | 65.00 | 30.711 | $193,190.34 | $5,933,108.00 | 23.883 | $35,658.04 | $851,620.13 | $1.996/Mcf |
| 10.00 | 70.00 | 32.829 | $193,553.53 | $6,354,210.00 | 25.310 | $35,950.74 | $909,908.00 | $2.137/Mcf |
| 10.00 | 75.00 | 34.947 | $193,872.72 | $6,775,315.00 | 26.737 | $36,212.21 | $968,196.13 | $2.278/Mcf |
| 10.00 | 80.00 | 37.065 | $194,155.41 | $7,196,419.00 | 28.164 | $36,447.18 | $1,026,484.00 | $2.419/Mcf |
| 10.00 | 85.00 | 39.183 | $194,407.56 | $7,617,523.00 | 29.590 | $36,659.49 | $1,084,772.00 | $2.559/Mcf |
| 10.00 | 90.00 | 41.301 | $194,633.84 | $8,038,627.00 | 31.017 | $36,852.27 | $1,143,060.00 | $2.700/Mcf |
| 10.00 | 95.00 | 43.419 | $194,838.03 | $8,459,730.00 | 32.444 | $37,028.09 | $1,201,347.75 | $2.841/Mcf |
| 10.00 | 100.00 | 45.537 | $195,023.25 | $8,880,834.00 | 33.871 | $37,189.09 | $1,259,635.75 | $2.982/Mcf |

APPENDIX B

In order to determine the upstream pressure necessary to provide the necessary cooling through expansion, the following procedures should be followed:

> The outlet conditions from the turboexpander should be determined from a plot of critical pressure vs. critical temperature for $CO_2$ - $CH_4$ composition of the feed gas.
>
> This plot shown in Figure ( ) gives the critical pressure vs. critical temperature for $CO_2$ - $CH_4$ mixtures. For the $CO_2$-$CH_4$ mixtures of interest (% $CO_2$ greater than 30 but less than 80) the critical point can be determined from this figure. In order to obtain the most $CO_2$ in the liquid stream out of the liquid-vapor separator, the outlet pressure out of the turboexpander should be higher than the critical pressure for the feed, and the outlet temperature should be 10, 20, or 30 degrees lower than the critical temperature of the mixture. The limit on how much the outlet temperature should be lower than the critical temperature depends the amount of $CH_4$ desired in the $CO_2$ stream and also the temperature at which solid $CO_2$ may appear (which can cause severe operational problems).

- Once the outlet pressure and temperature ($P_2$ and $T_2$) are set and since the inlet temperature ($T_1$) is known, Figures B-1 through B-4 can be used to calculate the upstream pressure as follows:

STEP 1. Calculate ($T_1$ - $T_2$) and from figure B-1. Knowing $T_1$ determine the factor A

STEP 2. From Figure B-2, using A determined in step 1, and a proper efficiency, determine the factor B. The efficiency is varied from 60 to 90%. The efficiency depends on a few factors, among them the amount of liquid produced.

STEP 3. From Figure B-3, using B obtained in Step 2 and the proper 1/m, determine the factor C. As explained in the main text:

$$1/m = k/(k-1)$$

when, $k = C_p/C_v$

and, $1/m = C_p/(C_p - C_v)$

1/m values depend on methane-carbon dioxide mixture composition, temperature, and pressure. The 1/m values can be obtained from Tables

STEP 4. From Figure B-4, using the factor C obtained in Step 3 and knowing the outlet pressure ($P_2$), determine the upstream pressure $P_1$.

Example:

Data: Desired outlet conditions $T_2 = -35^oF$, $P_2 = 1000$ psi
Inlet Temperature $T_1 = 100^oF$

Calculate: Upstream pressure $P_1 = ?$

The solution is sketched through Figures B-1 to B-4.

| Figure No. | Known Factors | Obtained Factor |
|---|---|---|
| B-1 | $T_1 - T_2 = 135$, $T_1 = 100$ | A = 0.24 |
| B-2 | A = 0.24, E = 0.7 | B = 0.34 |
| B-3 | B = 0.34, 1/m = 2.5 | C = 2.8 |
| B-4 | C = 2.8, $P_2 = 1000$ | $P_1 = 2750$ |

Upstream Pressure = 2750 Psi

Procedure for Calculating Work Out of the Turboexpander

The procedure for calculating the amount of work expected out of a turboexpander is described through the use of an example:

- Basis: 1 lb mole of feed 60 mole% methane
  40 mole % carbon dioxide

  Outlet temperature ($T_2$) = $-35^oF$
  Outlet Pressure ($P_2$) = 500 psi

  Data: Inlet temperature ($T_1$) = $100^oF$
  Inlet Pressure ($P_1$) = 1400 psi

STEP 1. Knowing $T_2 = -35^oF$, and $P_2 = 500$ psi, the liquid mole fraction L is obtained from Figure B-5.

L = 0.406

STEP 2. Knowing $T_1 = 100^oF$, and $P_1 = 1400$ psi, the inlet enthalpy ($H_{in}$) is obtained from Figure B-6

$H_{in} = 92$ Btu/lb

STEP 3. Knowing $T_2 = -35^oF$, $P_2 = 500$ psi, and the liquid fraction of the outlet stream L = 0.406, the enthalpy of the outlet stream ($H_{out}$) is determined from Figure B-7.

$$H_{out} = 12 \text{ Btu/lb}$$

STEP 4. Knowing $H_{in} = 92$ Btu/lb (Step 2) and $H_{out} = 12$ Btu/lb (step 3), the change in enthalpy or theoretical work is obtained from Figure B-8.

Theoretical Work = 80 Btu/lb

STEP 5. Knowing theoretical work = 80 Btu/lb (Step 4), and an efficiency of 70%, the actual work is obtained from Figure B-9

Actual Work = 0.022 Horsepower-Hour/lb

B-1 A Plot of Difference in Inlet ($T_1$) and Outlet ($T_2$) Temperatures vs. An Intermediate Variable (A) for Various Values of the Inlet Temperature ($T_1$)

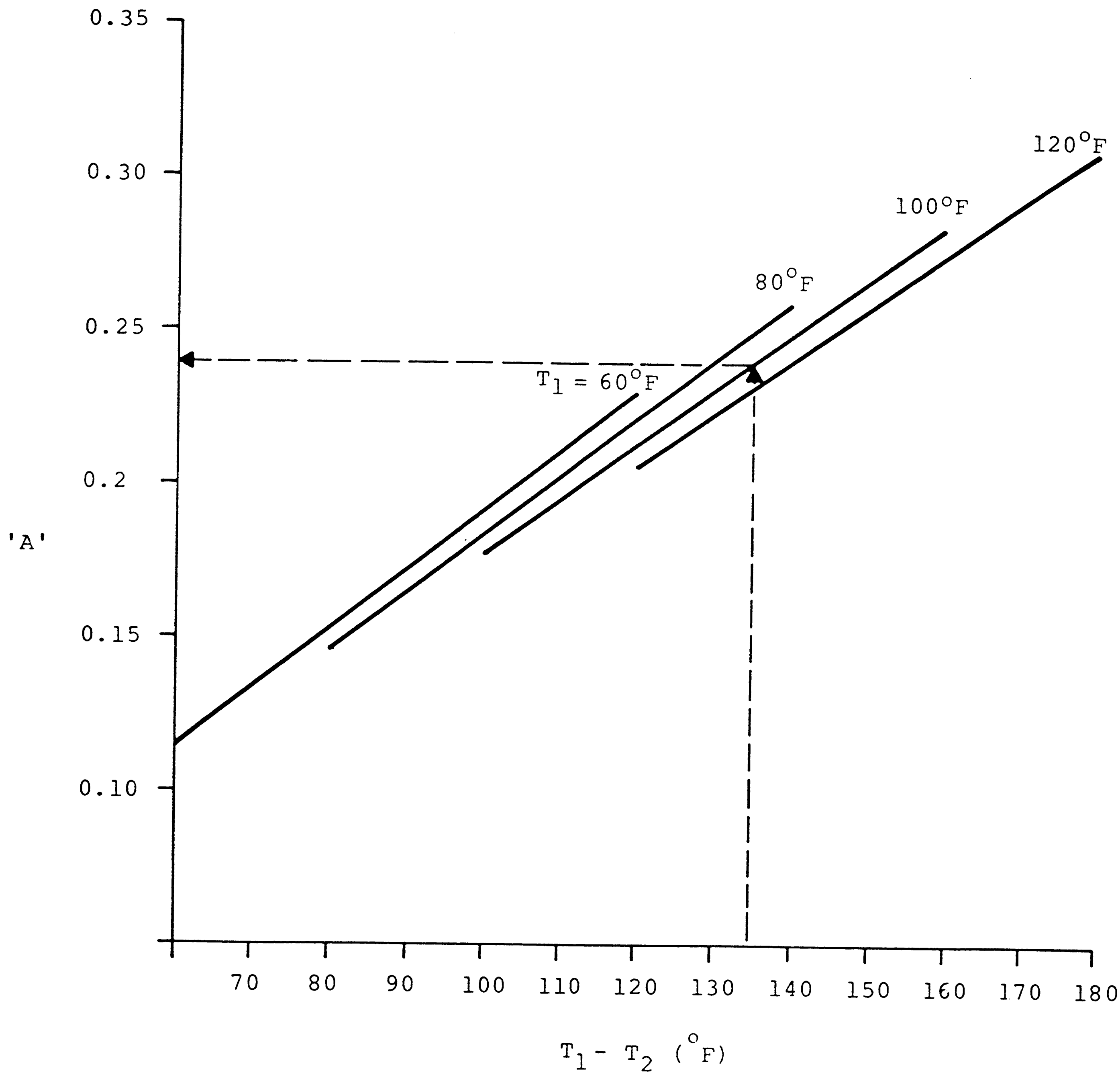


Figure B-2 A Plot of B (An Intermediate Variable) vs. A (Determined from Figure ) for Various Efficiencies of 60-90%

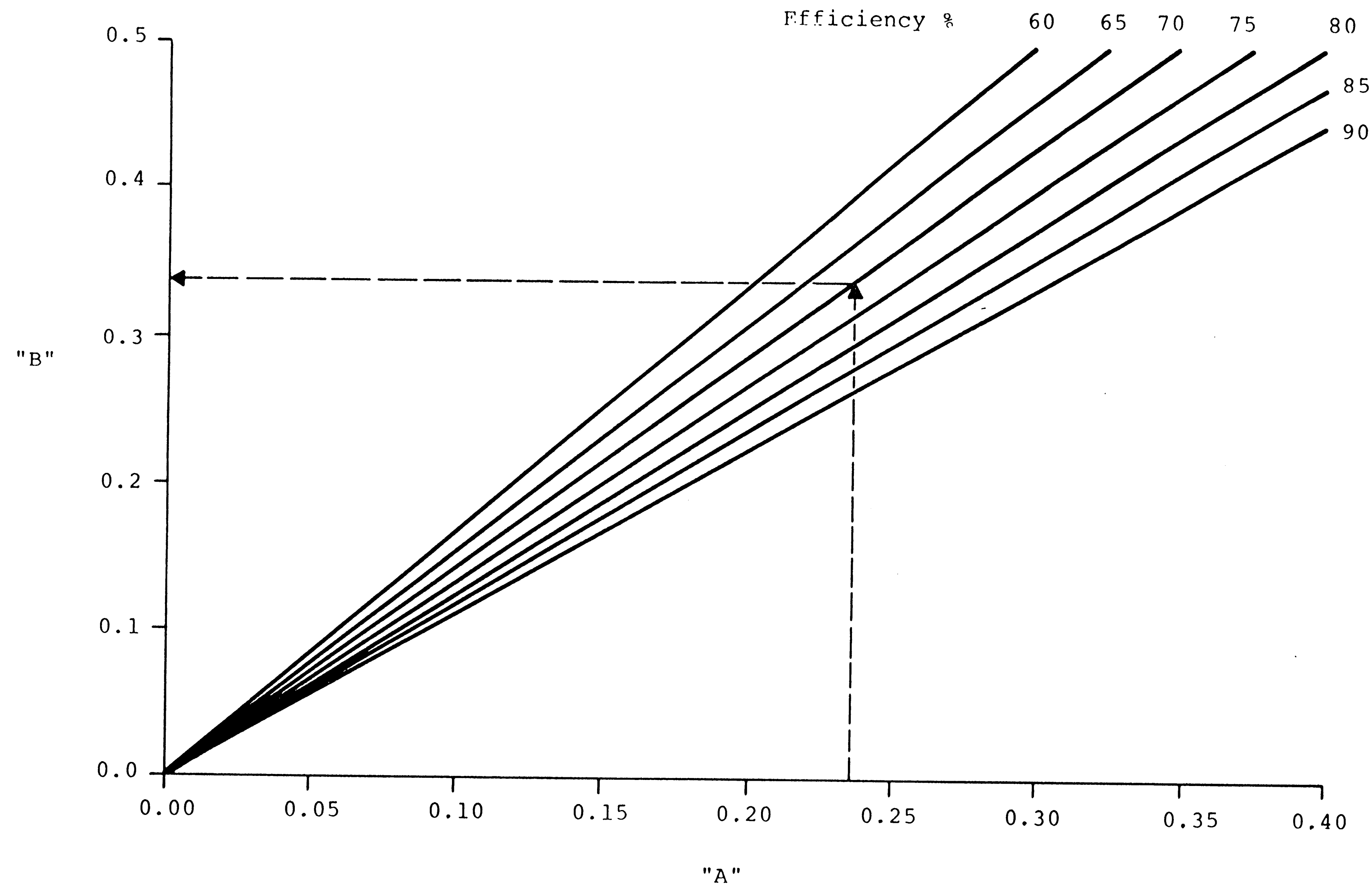


Figure B-3 A plot of C (An Intermediate Variable) vs. B (Determined from Figure ) for Various 1/m Values of 1.0 - 8.0

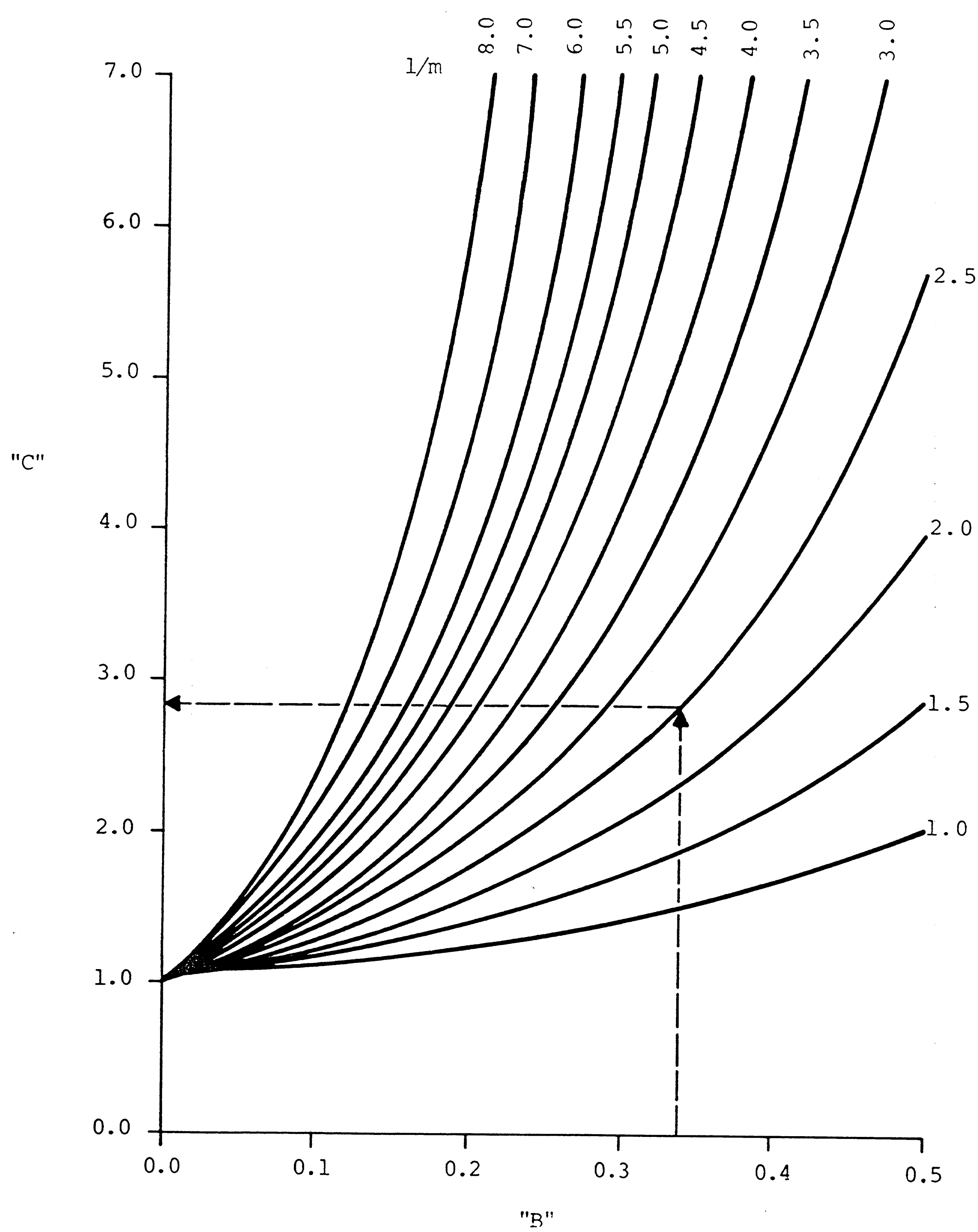


Figure B-4 A Plot of C (Determined from Figure ) vs. Inlet Pressure ($P_1$) for Various Outlet Pressures ($P_2$) 500-1000 psia

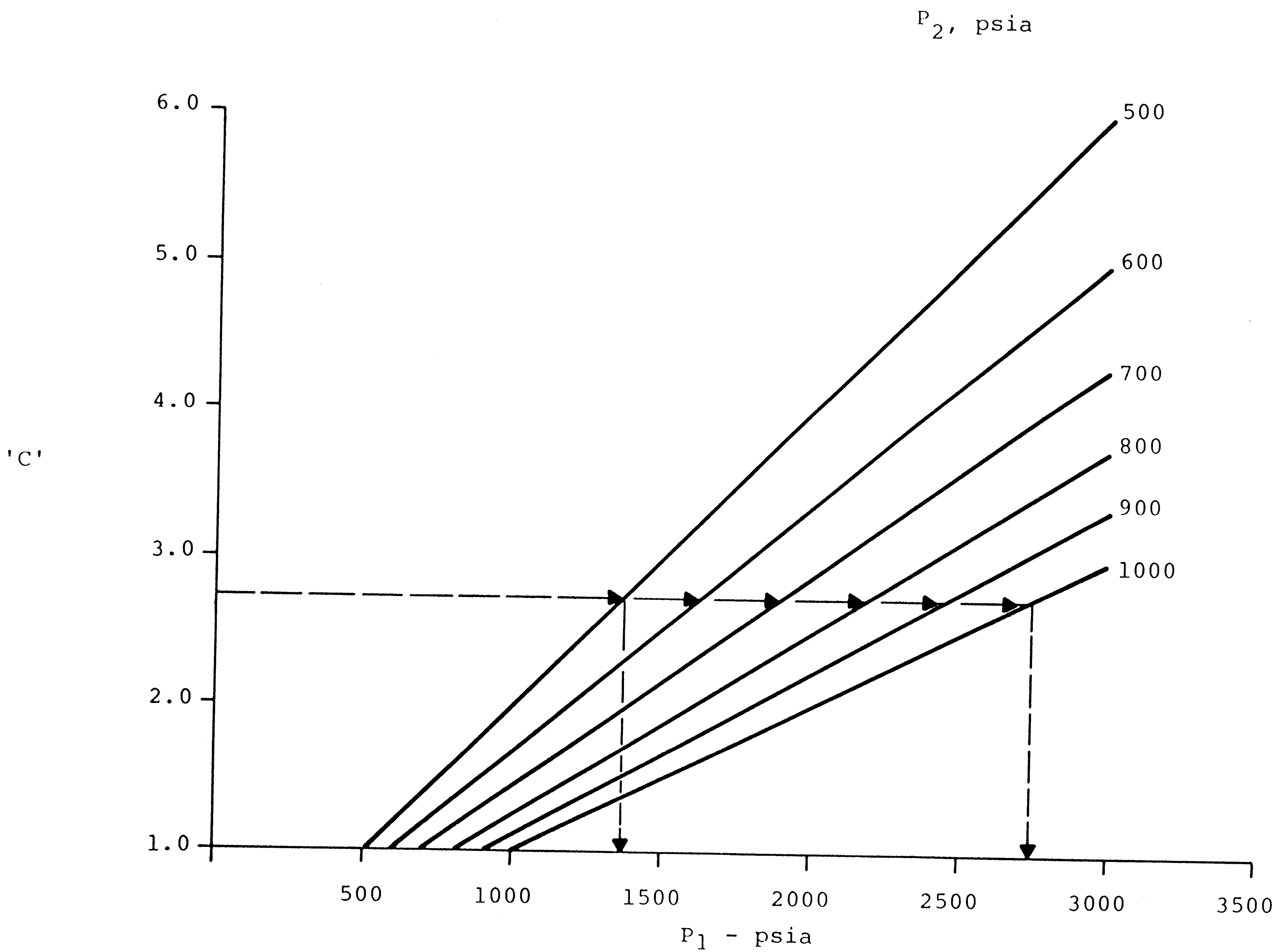


Figure B-5 Liquid Mole Fraction vs. Turboexpander Outlet Temperature ($T_2$) at Pressures of 500-1000 psia for a Misture of 60% methane, 40% $CO_2$

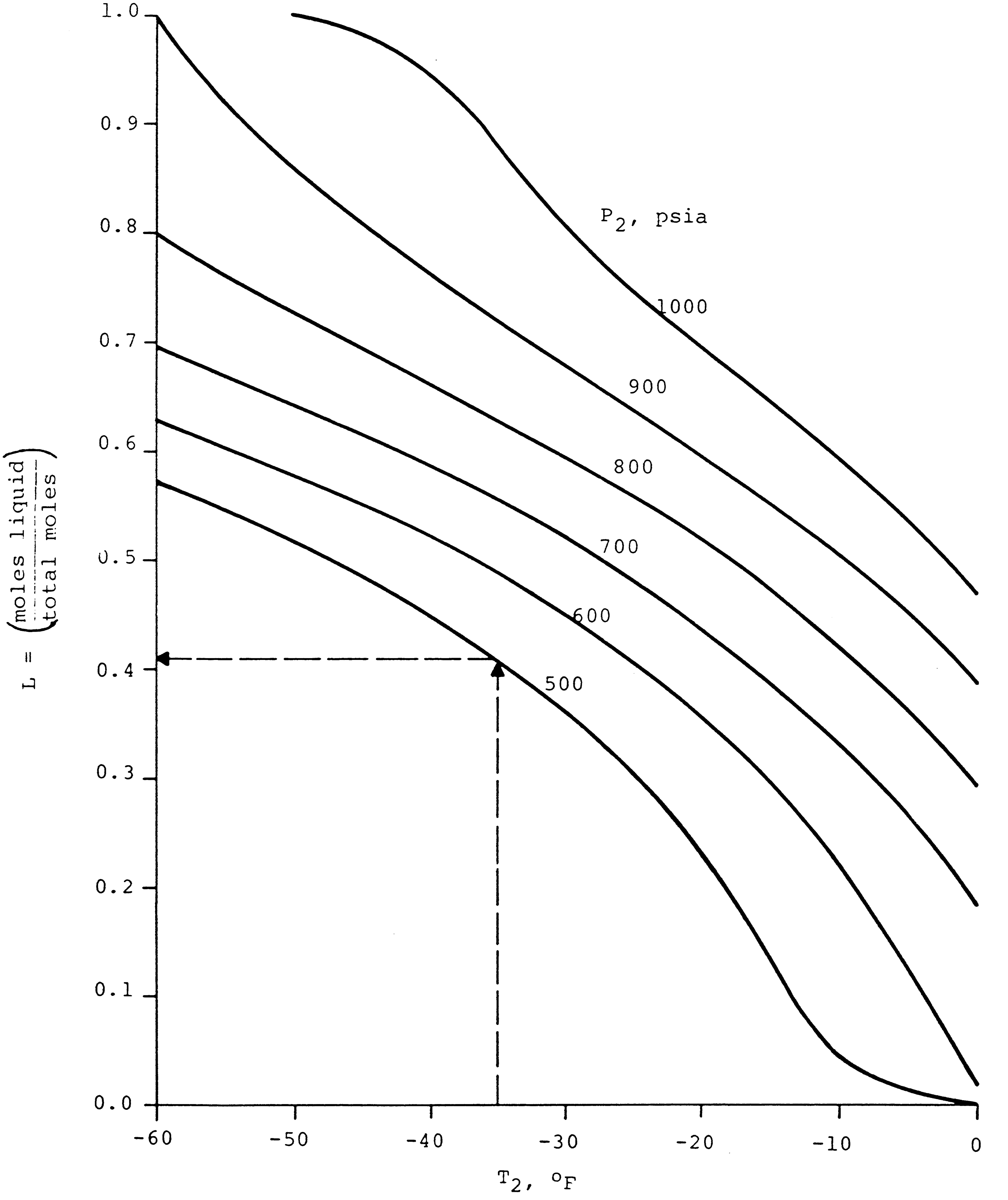


Figure B-6 Enthalpy of the Inlet Stream to the Turboexpander vs. Inlet Pressure at Inlet Temperatures of 60-120F

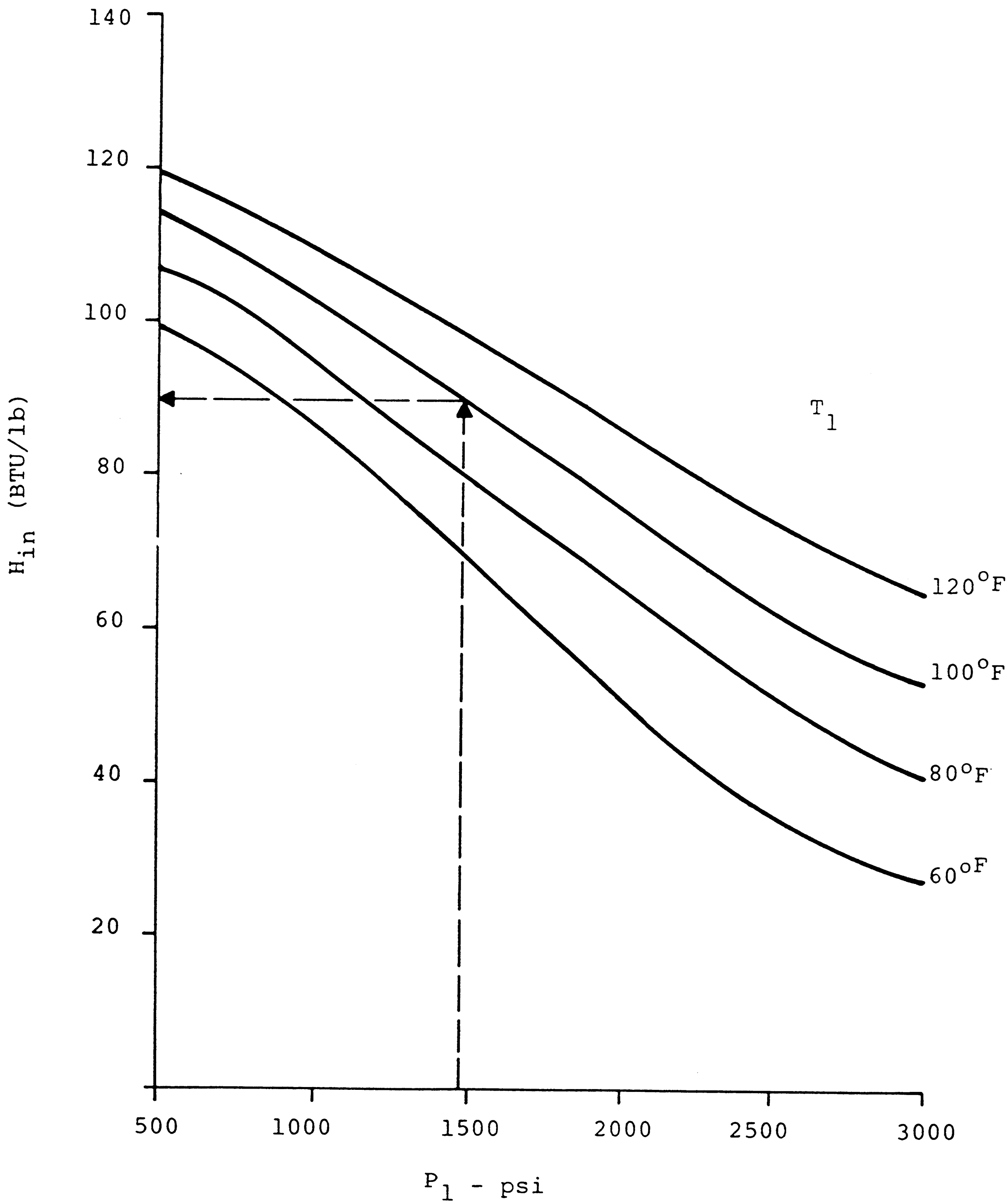
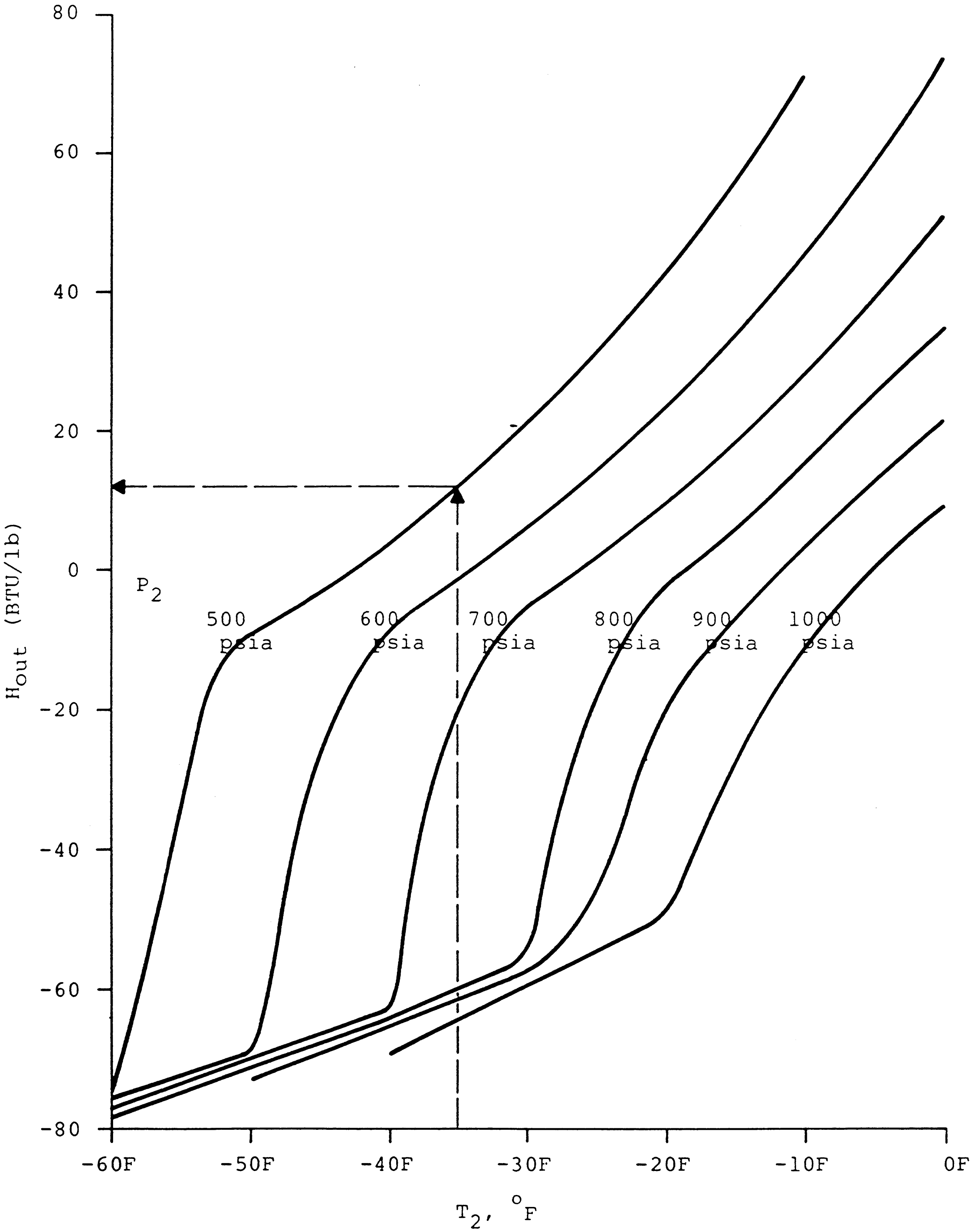
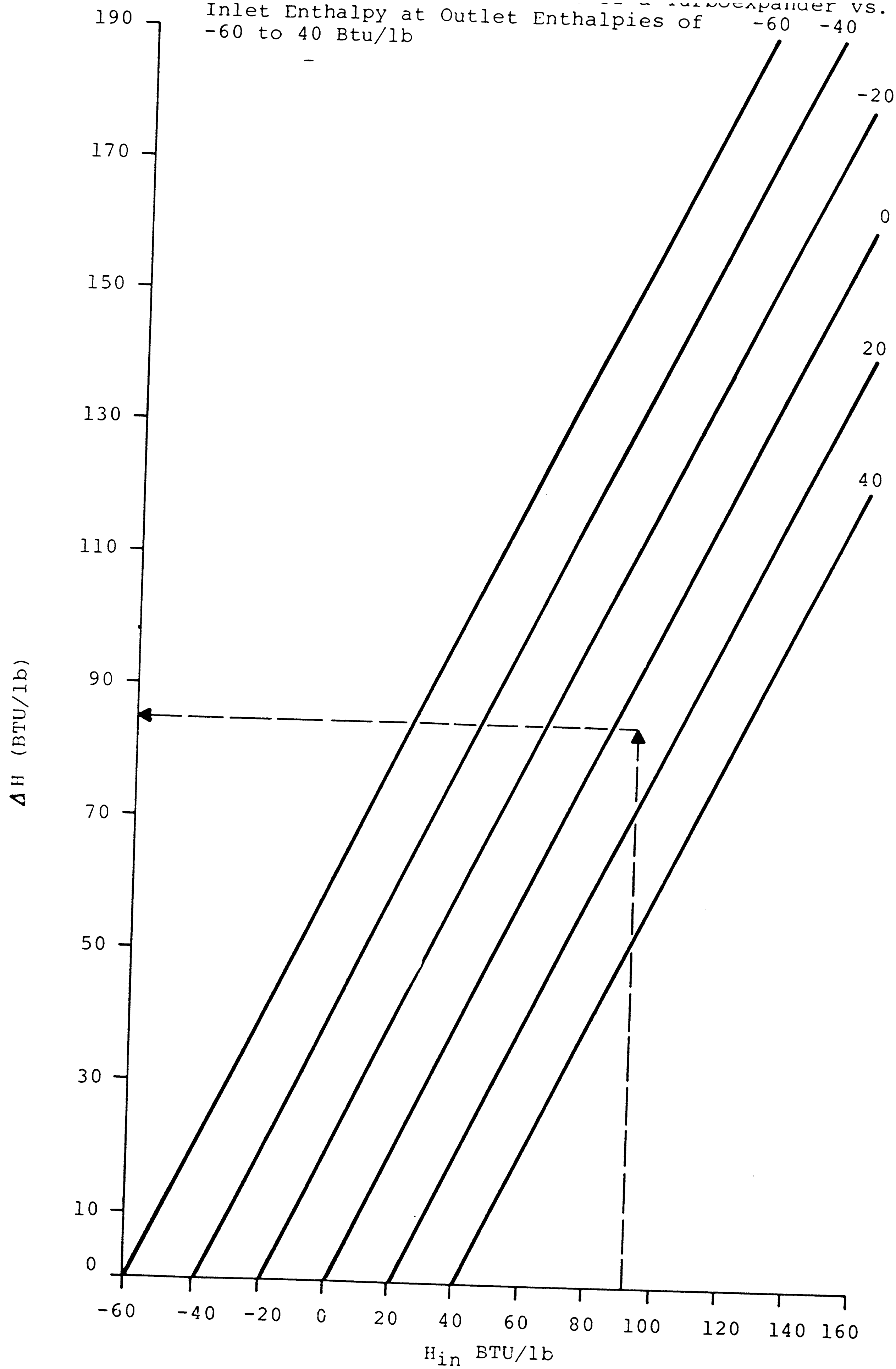


Figure B-7 Enthalpy of the Outlet Stream from the Turboexpander vs. Outlet Temperature at Outlet Pressures of 500-1000 psia



Inlet Enthalpy at Outlet Enthalpies of
-60 to 40 Btu/lb



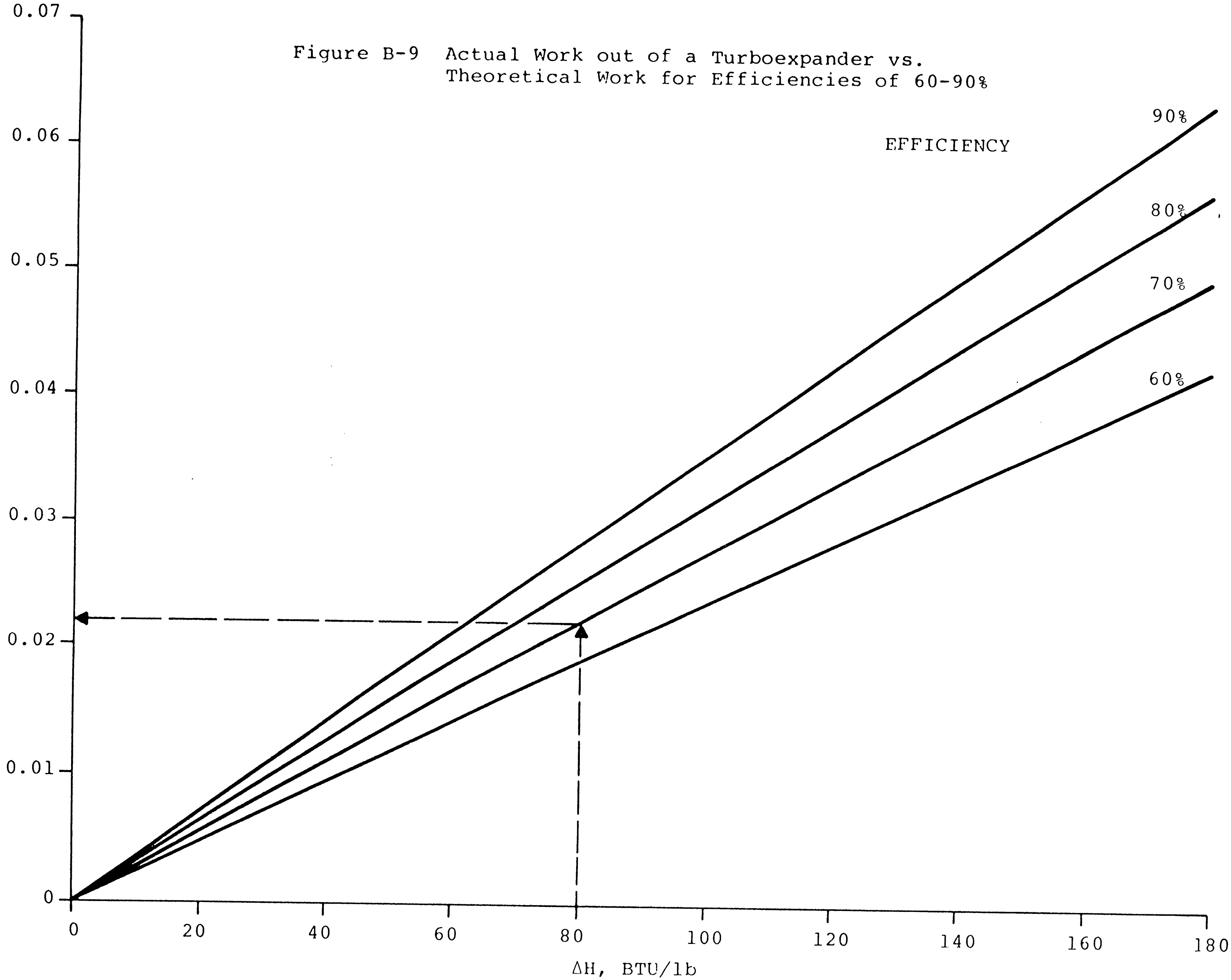



Figure B-9 Actual Work out of a Turboexpander vs. Theoretical Work for Efficiencies of 60-90%



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