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FOSSIL ENERGY

SOURCES AND DELIVERY OF CARBON DIOXIDE FOR ENHANCED OIL RECOVERY

Final Report for October 1977—December 1978

By
Mike Hare
Herbert Perlich
Raymond Robinson
Meghji Shah
Frederick Zimmerman

December 1978
Date Published

Work Performed Under Contract No. EX-76-C-01-2515

Pullman Kellogg
Division of Pullman Incorporated
Houston, Texas

U. S. DEPARTMENT OF ENERGY

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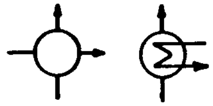
LIST OF ABBREVIATIONS

B - Billion (10^9)
BBL - Barrel (42 U.S. Gal.)
BPD - Barrels per day
BTU - British Thermal Unit
CAOF - Calculated absolute open flow
Cap. - Capacity
CO_2 - Carbon dioxide
Cp - Heat capacity at constant pressure
DCFRR (DCF-ROR) - Discounted cash flow rate of return
DOE - U. S. Department of Energy
E_R, RE - Recovery Efficiency
Engr. - Engineering
EOR - Enhanced oil recovery
Est. - Estimate, estimated
°F - Degrees Fahrenheit
FCCU - Fluidized catalytic cracking unit
FWHP - Flowing well head pressure
G - Gas in place
Gal. - U. S. Gallon
GPM - Gallons per minute
Gr - Grains
GURC - Gulf Universities Research Consortium
KWH - Kilowatt - hour
LB - Pound
LB MOL - Pound mole
LHV - Lower heating valve
M - Thousand (10^3)
MM - Million (10^6)
Md - Millidarcies
MEA - Monoethanol amine
Mega - 10^6
MW - Mega watts, molecular weight
NCI - National Cryo- Chemics, Inc.
N.G. - Natural gas
OGIP - Original gas in place
OOIP - Original oil in place
Pa - Abandonment pressure
Pi - Initial pressure
Psc - Pressure at standard conditions 14.7 psia
Prod. - Production
PSI - Pounds per square inch; A - absolute; G - gauge (base 14.7 psia)
Sg - Gas saturation
SACROC - Scurry area Canyon Reef Operations Committee
SCF - Standard cubic feet
SIBHP - Shut-in bottom hole pressure
SIP - Shut in pressure
SNG - Synthetic (or substitute) natural gas
STPD - Short tons per day
T - Trillion (10^{12})
Ti - Initial Temperature
TEG - Triethylene glycol
TPY - Tons per day
Z_a - Gas compressibility factor (abandonment)

Z_i - Gas compressibility factor (initial)
% - Percent
∴ - Therefore
Ø - Porosity

FLOW SHEET SYMBOLS

EXCHANGERS

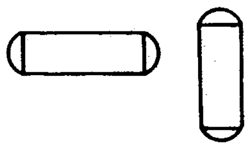

PROCESS HEAT EXCHANGER


COOLING WATER EXCHANGER


STEAM REBOILER
(KETTLE TYPE)


AIR COOLED EXCHANGER

VESSELS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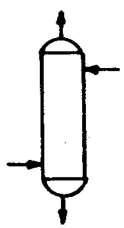
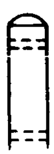
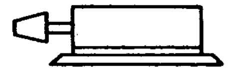

TOWER


PACKED TOWER


TRAYED TOWER
(SIEVE TRAYS)

ROTATING EQUIPMENT

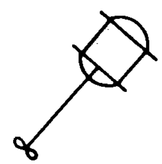

CENTRIFUGAL COMPRESSOR WITH DRIVER
(MAY HAVE MORE THAN ONE COMPRESSOR
PER DRIVER)


TURBINE, STEAM OR HYDRAULIC


RECIPROCATING COMPRESSOR


BLOWER

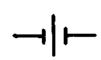

CENTRIFUGAL PUMP


MIXER

OTHER


FILTER


VALVE


RESTRICTION ORFICE


RAW MATERIAL FEED

ABSTRACT

This report presents results from a comprehensive study by Pullman Kellogg, with assistance from Gulf Universities Research Consortium (GURC) and National Cryo-Chemics Incorporated (NCI), of the carbon dioxide supply situation for miscible flooding operations to enhance oil recovery.

A survey of carbon dioxide sources within the geographic areas of potential EOR are shown on four regional maps with the tabular data for each region to describe the sources in terms of quantity and quality

Evaluation of all the costs, such as purchase, production, processing, and transportation, associated with delivering the carbon dioxide from its source to its destination are presented.

Specific cases to illustrate the use of the maps and cost charts generated in this study have been examined.

SUMMARY

PART 1

1. SUMMARY

This report concludes the study by Pullman Kellogg on sources and delivery of carbon dioxide for enhanced oil recovery (EOR). Included in this study are all significant sources of aboveground and naturally occurring carbon dioxide. These are identified on four regional maps. Quantity and quality of these sources have been tabulated to the extent of availability of the data and the costs associated with delivering the carbon dioxide from its source to its destination for EOR have been developed and are presented, as far as possible, graphically. These costs include purchase, production, processing of impure sources and transportation costs. The oil fields suitable for EOR by CO_2 miscible flooding in West Texas, South Mississippi and South Louisiana have been used as specific cases to show examples of the use of the maps and cost charts developed for evaluating the cost of CO_2 for EOR.

A discussion and map of CO_2 sources in the Los Angeles Basin area is included in this report. This information is presented from the study by Lawrence-Allison and Associates[28] Corporation under DOE contract No-EF-77-C-03-1582.

The results of the survey of CO_2 sources are shown on four regional maps for convenience in establishing practical

CO_2 supplier-user pairs. The sources considered are natural deposits, industrial sources such as process and cement plants, natural gas treating plants, SNG plants, refineries and power plants. The survey indicates that although the potential peak CO_2 demand of 7-11 BSCFD could be met from the total CO_2 supply of 28.4 BSCFD, shortages of CO_2 for the West Texas oil fields could develop if actual demand for CO_2 reaches levels projected in this report.

GURC has made a geologic appraisal and economic analysis of naturally occurring CO_2 sources of the McElmo Dome and Sheep Mountain areas in Colorado, Northeast New Mexico area and Jackson Dome area in Mississippi. The effect of the well flow rate and the well reserve size on the DCFRR at CO_2 price of $0.25 and $0.50 per MSCF is presented graphically.

The present cost of purchasing CO_2 varies from a low of $1-2/Ton for power plant stack source to a high of $15-20/Ton F.O.B. plant for a liquid CO_2 plant source. High purity CO_2 sources such as ammonia plant vents command higher prices for CO_2 than low purity sources like flue gas stacks from power plants.

The costs of purification of contaminated naturally occurring CO_2 sources are developed using physical solvent process, called selexol system.* These costs are presented

* Allied Chemical process.

graphically to show the effect of source pressure, source quality and the capacity of the processing plant on DCFRR. The purification costs for recovering CO_2 from power plant stacks are presented and compared using chemically reactive system (MEA) and physical solvent system (selexol). The recovery costs for 125 MMSCFD of CO_2 capacity vary from $0.60/MSCF to $1.03/MSCF for a MEA system and from $1.97/MSCF to $2.96/MSCF for a selexol unit using a 10% - 25% DCFRR. The cost of compression for CO_2 source that is of high purity but needs compression from near atmospheric pressure to 2000 psig is presented graphically. The procedure is given to evaluate the cost of compression for the high purity source with higher than atmospheric initial pressure.

For evaluating transportation costs, three pipeline systems are examined as possible candidates for transporting CO_2 for enhanced oil recovery. The three pipeline systems studied are:

1. The supercritical pipeline system
2. The subcritical pipeline system
3. The liquid pipeline system

Each pipeline system has been designed to transport CO_2 from the source area to the oil recovery site. From these designed pipeline systems, the most feasible pipeline system for CO_2 transportation is determined using an economic analysis which compares fixed capital costs, as well as direct and indirect operating costs. The conclusion from this analysis is, as may be expected, the use of the supercritical pipeline system for CO_2 transportation. Figures and graphs are presented to determine CO_2 pipeline transportation costs using the supercritical pipeline system. This data is presented in such a manner that virtually all CO_2 transportation costs, for pipeline lengths up to 500 miles, can be estimated, regardless of terrain and elevation effects.

Specific cases presented in this report conclude that the investment costs for producing natural CO_2 can not be estimated with high degree of confidence because of the uncertain factors such as, well reserve size, flowing well head pressure, well deliverability, and actual drilling costs. Total cost of natural CO_2 contaminated with H_2S is high because of the high purification cost. Power plant stack gas source in the vicinity of the oil field could be used as a CO_2 source but a single power plant source probably can not supply sufficient CO_2 for a complete EOR project. CO_2 from ammonia plant vents is of high purity but compression costs and purchase costs can be substantial. Alternate sources of CO_2 such as power plant stacks should be considered for economic analysis.

INTRODUCTION

PART 2

2.1 GENERAL DESCRIPTION

Miscible flooding has been investigated as a possible alternate for recovering oil from depleted, water-out reservoirs. Data from the laboratory tests and field tests appear to confirm the applicability of carbon dioxide (CO_2) as a material capable of recovering additional quantities of oil from such reservoirs. Preliminary extimates indicate that 5 - 10 billion barrels of oil could be produced from enhanced oil recovery by CO_2 flooding. This might require upwards of 40 - 50 trillion standard cubic feed (TSCF) of CO_2.

Current research and field tests should resolve some of the questions concerning the process of miscible flooding by CO_2 so that economic evaluations can be made more reliable. Future plans for EOR by CO_2 flooding will depend on the availability and the cost of CO_2. This report by Pullman Kellogg is a comprehensive evaluation of the supply and cost of CO_2 for EOR.

2.2 FIRST PHASE SUMMARY

Pullman Kellogg, with assistance from Gulf Universities Research Consortum (GURC), completed the first study [19] on the supply of carbon dioxide for enhanced oil recovery (EOR), in September, 1977. This study was conducted for the United States Department of Energy (DOE) under the contract EX-76-C-01-2515. The study dealt with the identification of candidate oil reservoirs suitable for EOR by use of CO_2 sources, costs associated with the purification of impure CO_2 sources, and with the cost of transportation of CO_2 to candidate oil reservoir for EOR. In addition, the study included an investigation of sources for physical and equilibrium properties of carbon dioxide.

Time and funding had limited the extent of the study to a broadscope, orientation type report. The candidate oil fields for EOR were selected by GURC on the basis of criteria pertaining to rock and fluid properties and reservoir conditions. These fields, located in thirteen states, were mapped and tabulated and were included in the first phase report. Aboveground CO_2 sources, consisting of power plant stack gas, cement plant stack gas and chemical plant vents, were tabulated to indicate the approximate total quantity of CO_2 available in each of the thirteen states considered. Exact location of each individual source, its quantity and quality were not identified. An exhaustive survey for naturally occurring sources of CO_2 was not done and the sources were depicted on the map only to approximate geographic location with some qualitative assessments.

The purification of power plant flue gas, to recover CO_2, was discussed using a regenerative solution of MEA. A physical solvent process (Selexol) was used to purify naturally occurring sources. Two hypothetical cases studied were sources containing 50% CO_2 and 98% CO_2 with and without 2% H_2S impurity, and the balance being Methane (CH_4). Costs of development of natural sources were not determined. Transportation costs were determined for a few hypothetical cases. Economic analysis, using simplifying assumptions, was done to determine the approximate total cost of CO_2 for EOR by miscible flooding operations. This analysis was done for the various aboveground and underground hypothetical sources of CO_2.

Although limited in scope, several conclusions were drawn from first phase study. Review of the carbon dioxide supply situation in the thirteen-state region indicated that sufficient CO_2 is available from aboveground sources to satisfy the projected future demand for EOR. Power plant and cement plant stack gas sources are the most widespread and abundant aboveground sources. However, the quality of these sources is low, typically less than 20% CO_2, resulting in higher purification costs. The highest quality aboveground sources of CO_2 are process vents from chemical plants, such as ammonia plants.

The study also concluded that natural sources of CO_2 have great potential for EOR application. The Four Corners area (Utah, Colorado, Arizona, New Mexico), Southeast Colorado, Northeast New Mexico and Central Mississippi areas seemed to have potential naturally occurring carbon dioxide. However, the study concluded that natural sources alone might not supply all the CO_2 required for EOR efforts.

2.3 SCOPE OF THIS REPORT

The first study provided a general overview of the CO_2 supply situation for EOR by miscible flooding operation. However, a more comprehensive picture is needed with a fully developed economic study.

This report is the result of the second phase study by Pullman Kellogg, assisted by GURC and National Cryo-Chemics, Inc. (NCI), conducted under contract EX-76-C-01-2515. It represents an extensive effort in developing carbon dioxide supply situation for enhanced oil recovery by miscible flooding. The objectives of the study are threefold and include:

1. Mapping of all significant sources of carbon dioxide on regional maps in a quantitative manner so that through the use of the maps and legends a description of the sources in terms of quantity and quality is available.
2. Evaluation of all the costs associated with delivering the carbon dioxide from its source to its destination. These include purchase, production, processing, and transportation costs.

   As far as possible, the cost information will be presented graphically so that by working with an initial description of a gas source all costs associated with delivery of carbon dioxide from that source can be developed.
3. Examination of several specific cases to show examples of the use of the maps and cost charts generated in the above described objectives.

Emphasis has been placed on preparation of detailed regional maps of the carbon dioxide sources consisting of power plant stack gases, cement plant stack gases, process vents, and naturally occurring CO_2. By-product CO_2 from fermentation processes and phosphate manufacturing is not considered since these are determined to be sources with high purification costs and low availability. The regions of CO_2 sources selected for the study are believed to be areas of greatest industry interest in EOR by CO_2 miscible flooding.

A separate map showing the carbon dioxide sources in the Los Angeles Basin is included in this report. This map was developed by Lawrence-Allison & Associates [28] Corporation under DOE contract NO-EF-77-C-03-1582 for the study of CO_2 Recovery & Tertiary Oil Production Enhancement in the Los Angeles Basin.

This study has focused on determining availability and cost of production of natural sources of CO_2. Industry interest in natural sources is also very high as witnessed by the ongoing exploration efforts of major oil companies and others to locate, define and characterize natural CO_2 reservoirs in terms of production rates and reserves. Several states, including New Mexico, Mississippi, Utah and Colorado, presently are believed to have sufficient reserves to justify commercial interest.

A comprehensive picture of all the costs, associated with supplying CO_2 from its source to a candidate oil reservoir, is presented. These costs consisting of the purchase, production, processing and transportation are shown so that an oil reservoir operator can determine the various costs involved for a particular CO_2 source. Therefore, the total cost of CO_2 for EOR by miscible flooding can be evaluated. An operator can compare total cost of the various sources available to him and can determine the most economical source. The intent of this report is for general guidance and for use as a screening tool to locate available sources of CO_2 and for initial economic evaluation. An operator considering EOR by CO_2 miscible flooding still will have to do the final evaluation. All investment costs and economic evaluations reflect 4th quarter, 1978 price levels with a ± 25% accuracy.

Impure sources of CO_2 will require processing to upgrade the quality of CO_2. The capital investment costs for the processing units and the results of economic analysis are presented graphically for CO_2 capacities from 125 MMSCFD to 500 MMSCFD. Sensitivity of the purification cost to the quality of the source and to the pressure at which the source is available is also depicted graphically.

Investment costs and economic analysis for the transportation of CO_2 by supercritical pipeline, subcritical pipeline and liquid CO_2 pipeline are presented. Transportation costs for pipeline capacities from 50 MMSCFD to 500

MMSCFD with pipeline lengths from 50 miles to 500 miles are evaluated.

Three specific (typical) cases are presented to illustrate the use of maps and cost charts for evaluating the total cost of carbon dioxide. The sources considered for these illustrations include natural CO_2 sources, a power plant stack gas source and ammonia plant vent sources.

CO_2 SOURCES AND AVAILABILITY

PART 3

3.1 INTRODUCTION

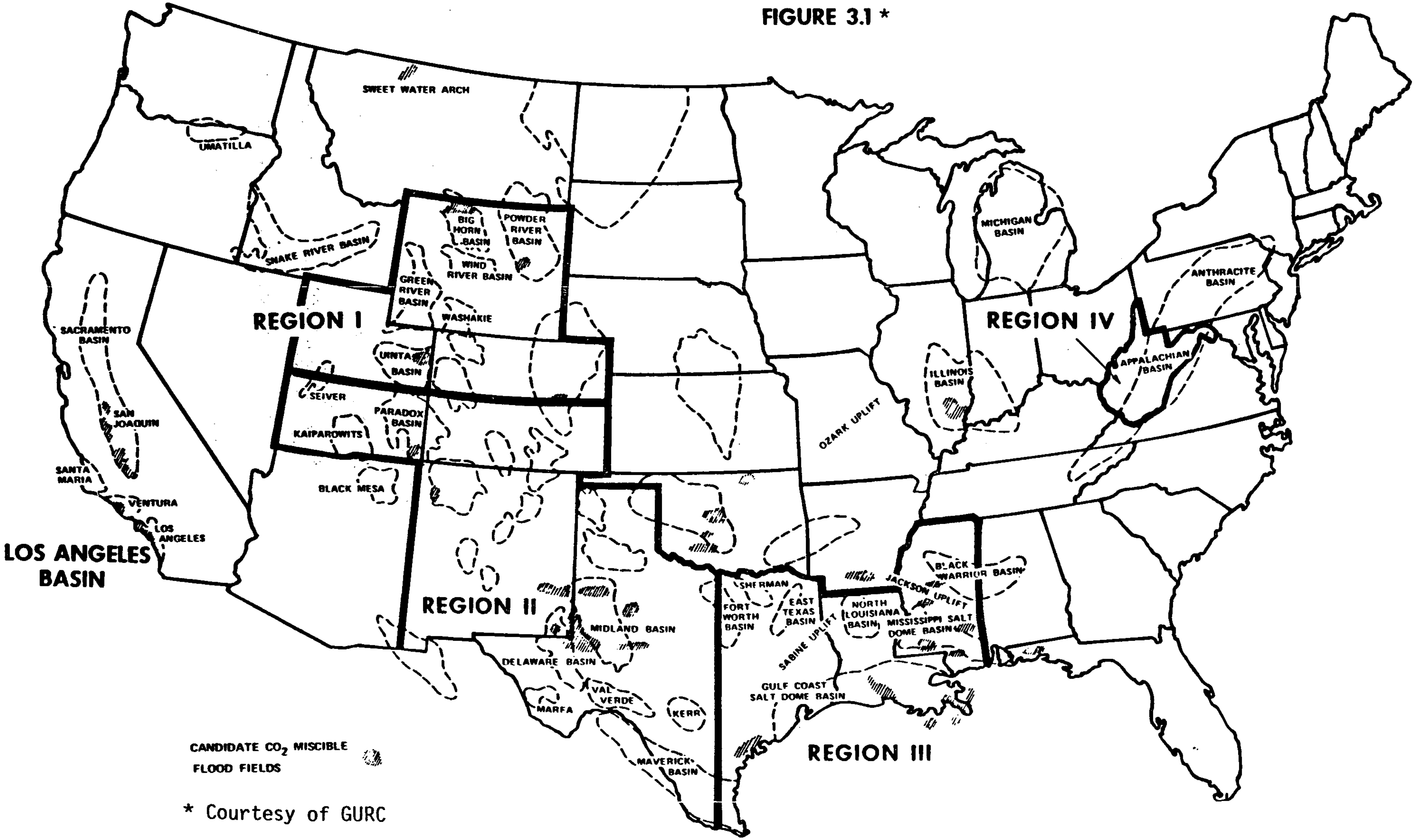
One of the primary purposes of this report is to present a comprehensive survey of all significant CO_2 sources, both naturally occurring and industrial, within the areas determined to have CO_2 enhanced oil recovery potential. Those states having candidate oil fields meeting criteria established by GURC in the first phase of this report (Figure 3.1) define the regions within which the CO_2 survey is generally confined. For convenience in establishing practical CO_2 supplier-user pairs, the areas of interest have been divided into four regions.

Region I: Wyoming, North Utah and North Colorado
Region II: South Utah, South Colorado, New Mexico, West Texas
Region III: East Texas, Louisiana, Mississippi
Region IV: West Virginia

Some candidate oil producing areas presented in the first phase study [19] are not considered because of the low likelihood of wide-spread field development. In addition to the four region survey presented here, data for the Los Angeles basin area of California as provided in the Lawrence-Allison report[28] will be included.

CANDIDATE CO_2 MISCIBLE FLOOD OIL FIELDS AND REGION DEFINITIONS

FIGURE 3.1 *



* Courtesy of GURC

3.1.1 Demand for CO_2 as an Oil Recovery Agent

At present there are approximately twelve on-going CO_2 enhanced oil recovery projects in Colorado, Louisiana, Texas and West Virginia listed in Table 3.1. All of these are test projects with the exception of the commercial Kelly-Snyder/Sacroc Unit in West Texas and have an aggregate enhanced production potential of approximately 40000 BPD.[31] These projects alone demand 240 MMSCFD CO_2 with 200 MMSCFD dedicated to the West Texas projects. These amounts are only a fraction of the possible CO_2 demand if substantial enhanced recovery projects are in the future.

An attempt to forecast the future demand for CO_2 will be somewhat conjectural since the demand on CO_2 is, of course, strongly related to the amount of oil that could be recovered by miscible flooding. However, it still will be useful to establish a range for potential demand. Studies in the year 1976 by Lewin and Associates,[32] the National Petroleium Council[35] and the Office of Technology Assessment[36] have projected that potential CO_2 recoverable oil could be in the billions of barrels. For example, at an oil price of $13.75/BBL all reports estimate about 5 billion barrels of recovery. However, care should be taken in using these projections because of the high sensitivity to oil price, CO_2 costs, recovery performance, and risk weighted DCFRR requirements. At the $13.75/BBL price level expected oil recovery ranges from 2 billion BBL for 20% DCFRR to 4.8 billion BBL for the base case 10% DCFRR on after tax income. Doubling or halving the base CO_2 costs alters the recovery range from 2.5 to 6.5 billion BBL and assumptions about performance can vary the range from 2 to 7 billion BBL.[33]

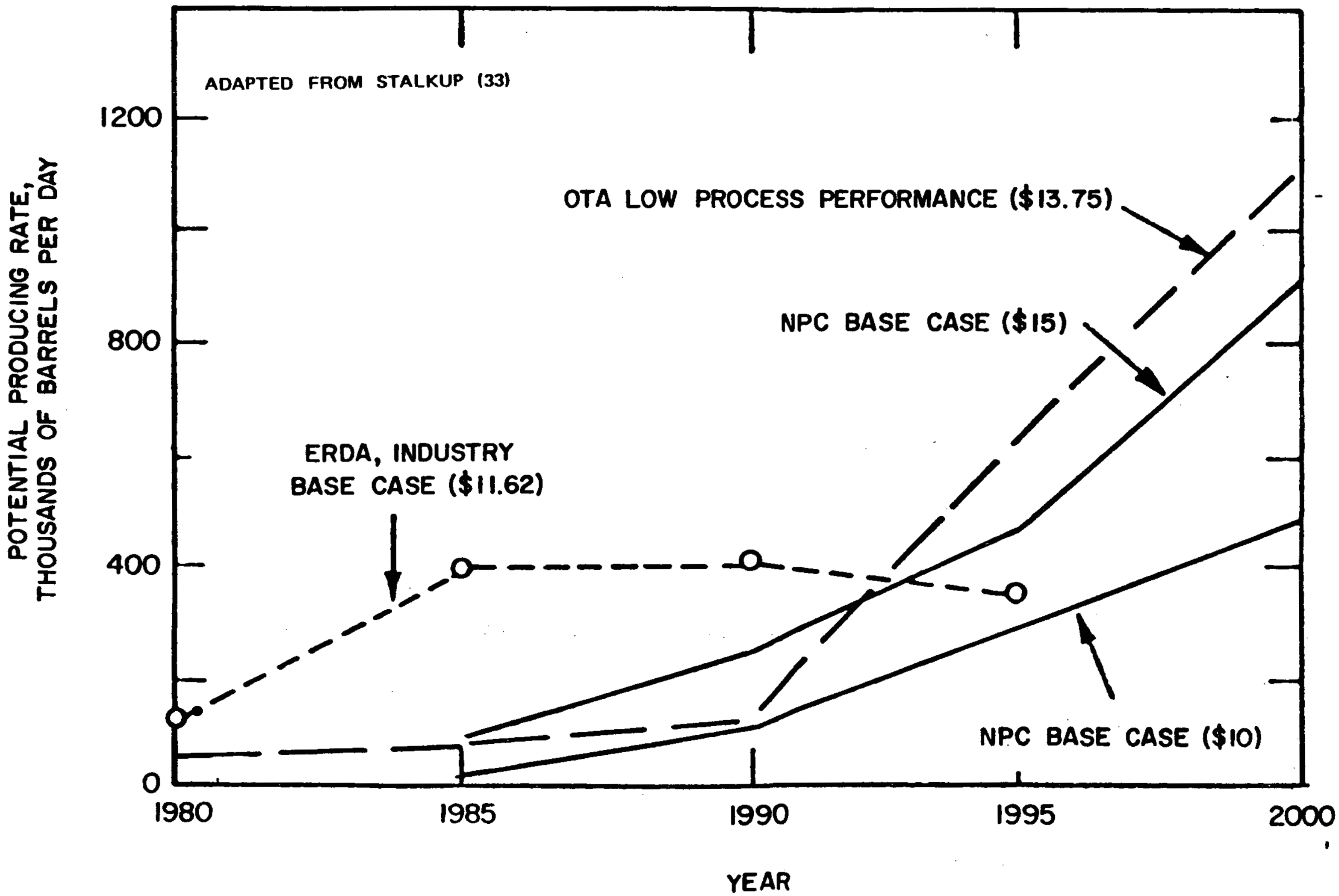
TABLE 3.1

CURRENT CARBON DIOXIDE ENHANCED OIL RECOVERY PROJECTS

State	County	Field	Operator	Enhanced Production, BPD	CO_2 Requirement MMSCFD
Colorado	Jackson	McCallum	Conoco	-	10
Louisiana	Iberia	Weeks Island	Shell/DOE	500 (projected)	2.3
Texas	Crane & Upton	Crossett	Shell	1450	15-20
	Ector	N. Cowden	Amoco	45 (gross prod.)	-
	Hockley	Levelland	Amoco	66 (gross prod.)	-
	Loving, Reeves, Ward	Two Freds	Houston N.G.	556	10
	Scurry	Kelly-Snyder/Sacroc	Chevron	37000	150-200
		Wasson/Denver	Shell	-	3 MMSCFD for 100 days
West Virginia	-	Granny's Creek	Columbia Gas	35	0.45 until June 77
		Griffithville	Guyan Oil/DOE	-	-
		Rock Creek	Pennzoil/DOE	2	0.25-0.35
Utah		Aneth	Phillips		

FIGURE 3.2

INCREMENTAL OIL PRODUCTION FROM CARBON DIOXIDE MISCIBLE FLOODING



In any case, the potential of production from enhanced recovery projects probably will not be significant until the late 1990's as shown in Figure 3.2. Between now and the 1990's the NPC and OTA studies estimate potential incremental production rates from 100,000 BPD to 200,000 BPD for $13.75 oil, with peak production in the late 1990's possibly reaching 700,000 to 1,100,000 BPD.

Cumulative requirements upwards of 40 to 50 TSCF of CO_2 may be necessary to recover the amount of oil projected in Figure 3.2.[17] To achieve this supply could require the development of 20 to 30 TSCF of reserves within the next ten years to be delivered at rates of 0.5 to 1.0 BSCFD beginning the early to mid 1980's.[33] This supply rate would probably have to increase to rates of 3 to 4 BSCFD during the last fifteen years of this century and be at a cost that would permit economically attractive floods. Depending on the success of first phase commercial ventures, an additional 20 TSCF of CO_2 development might be required in order to fully maximize the recovery potential and CO_2 miscible flooding in the years beyond 2000.

At least 80% and possibly up to 90% of the oil recovered by the CO_2 flooding process will come from the West Texas oil fields, hence the largest demands for CO_2 are going to be located in this area. The importance of the West Texas oil fields in enhanced recovery efforts is rather evident from the level of activity here by the major oil companies to test the flooding process and from announcements of intent to construct large CO_2 supply lines from Colorado and New Mexico. Estimates by GURC put the potential range of CO_2 enhanced recovery in West Texas at 2 to 8 billion barrels with a median value of 5 billion barrels. The net new CO_2

requirement for this area is anticipated to range from 4 to 12 MSCF/BBL oil resulting in an expected gross CO_2 requirement of 40 to 50 TSCF.

There are lesser, but nevertheless significant, potential demands for CO_2 in South Louisiana and Mississippi. GURC estimates that the potential amounts of CO_2 recoverable oil for South Louisiana range from 50 to 300 million BBL with an average net new CO_2 requirement of 6 MSCF/BBL. Total South Louisiana CO_2 demand is then expected to range from 0.3 to 1.8 BSCF. Estimated potential CO_2 recoverable oil from Mississippi is 100 million BBL with a net new CO_2 requirement in the range of 10 MSCF/BBL. Therefore, Mississippi CO_2 demand could be on the order of 1 TSCF.

3.1.2 Aggregate Carbon Dioxide Supply Situation

Because of the relatively low number of candidate CO_2 flood fields it seems more appropriate to assess the aggregate carbon dioxide supply situation for a specific field or region rather than the nation as a whole. In fact, drawing conclusions about the national supply of CO_2 versus potential demand could be totally misleading. For example, in the preceding section it was stated that potential incremental production could reach 700,000 barrels per day, requiring around 7 BSCFD CO_2 (at 10 MSCF/BBL). With a total CO_2 supply of 28.4 BSCFD available (Table 3.2) it appears that peak CO_2 demand can be met. However, as pointed out, it is expected that 90% of the incremented oil could come from West Texas; hence, West Texas CO_2 demand could reach 6.3 BSCFD. From Table 3.2 it appears that only 6.0 BSCFD are available at this time from all sources in Region II at any cost; hence, there could be an aggregate shortage of CO_2 at projected peak recovery.

TABLE 3.2

AGGREGATE SOURCES OF CARBON DIOXIDE SURVEYED IN THIS REPORT - MMSCFD

	TOTAL	POWER PLANT	CEMENT PLANT	PROCESS PLANT	NATURAL DEPOSITS
REGION I					
Wyoming	1748.0	1403.3	27.2	317.5*	3-6 MMSCFD/WELL
N. Utah	458.1	378.0	29.1	11.0	40
N. Colorado	999.6	952.2	25.7	3.3	18.4
	3205.7	2733.5	82.0	331.8	58.4
REGION II					
S. Utah	105.0	55.0	0	0	50
S. Colorado	1075.0	393.6	31.4	0	600
New Mexico	2988.9	1510.1	38.0	440.8 *	650
W. Texas	2234.8	1421.9	253.0	237.9	322
	6003.7	3380.6	322.4	678.7	1622
REGION III					
E. Texas	10541.6	9831.2	316.6	393.8	0
Louisiana	1966.5	1520.6	0	445.9	not significant
Mississippi	970.1	542.5	191.0	111.6	125
	13478.2	11894.3	507.6	951.3	125
REGION IV					
West Virginia	4380.2	3556.8	0	793.6	29.8
CALIFORNIA	1284.2	1226.2	n/a	58.0	0
Grand Totals	28351.8	22791.2	912.0	2813.4	1835.2

*Includes Proposed SNG Plant

TABLE 3.3

SUMMARY OF NATURAL CO_2 RESERVES SURVEYED

	BSCF CO_2	Reference
Region I		
Wyoming	3900	Speculative
N. Utah	1600	NCI (10), Campbell (12)
N. Colorado	100	Oil Daily (30)
TOTAL	5600	
Region II		
S. Utah	1000	NCI (10)
S. Colorado	3900-5900	GURC (17)
New Mexico	7000-10600	GURC (17)
W. Texas	600+	West (26)
TOTAL	12500-18100	
Region III		
E. Texas	None	NCI (10)
Louisiana	-	Finney (9)
Mississippi	2000-4200	GURC (17), NCI (10)
TOTAL	2000-4200	
Region IV		
West Virginia	330	NCI (10)
Los Angeles	None	Lawrence-Allison (28)

It is for reasons such as this that geographical divisions have been set up for study in this report. The regional definitions are only for convenience and the CO_2 source maps have been drawn with overlap to allow flexibility in source-field matching. A summary of the potential supply situation for each region follows.

Supply Situation - Region I

The potential for CO_2 flooding in Region I is very uncertain at this time since to date there have been no field pilot tests. Chevron is looking near Rangely, Colorado, which has the most promising prospect, but no data is available at this time. GURC recommends against making extrapolations on potential CO_2 demand at this time for Region I. Current aggregate supply of CO_2 for this area is around 5.6 TSCF available at 3 BSCFD.

Supply Situation - Region II

GURC has estimated that 2 to 8 billion barrels of oil might be recovered from West Texas by CO_2 flooding, with a median estimate of 5 billion barrels. This figure compares favorably with 4.7 billion estimated by Gill in his table of major carbon dioxide recovery projects (34), within the range of Doscher & Wise's figure of 4 billion barrels (37), and with the NPC estimate of 5.5 billion BBLS. Gill reports that Chevron has determined an average CO_2 requirement of 6.7 MSCF/BBL for their Sacroc/Kelly-Snyder project. However, Chevron engineers say this is at the low range of possible requirements and that the CO_2 requirement is expected to increase to 15 MSCF CO_2/BBL recovered. Assuming an average requirement of 10 MSCF/BBL, upwards of 50 TSCF could ultimately be required with a peak delivery rate of 6.3 BSCFD. From Table 3.2 , peak deliverable CO_2 is estimated to be 6.0 BSCFD, including contribution from two proposed SNG plants. The supply of natural CO_2 reserves within Region II is estimated at 12-18 TSCF, far short of the 50 TSCF - that might be demanded.

Supply Situation - Region III

GURC estimates for Louisiana that incremented oil production could be 50 - 300 million BBL's with a net new CO_2 requirement of 6 MSCF CO_2/BBL. NPC[35] estimates the technically recoverable oil for Louisiana at 300 million barrels. This places the Louisiana contribution to Region III potential demand between 0.3 to 1.8 TSCF. Potential incremental oil production for Mississippi is estimated to be 100 million BBL at a net CO_2 requirement of 10 MSCF CO_2/BBL, placing Mississippi demand at 1 TSCF. Little oil production is expected from South or East Texas by CO_2 injection alone.

Total Region III demand could be on the order of 2.8 TSCF CO_2. From Table 3.2 , based on current information, it is estimated that 3 TSCF natural CO_2 could be available, all from the Mississippi Jackson Dome area. After satisfying local Mississippi demand there should be around 2 TSCF CO_2 available for transport (via pipeline) from Mississippi to South Louisiana. The potential South Louisiana demand of 1.8 TSCF can be met also from high purity ammonia and hydrogen plant CO_2 vents. At current operating levels there is approximately 390 MMSCFD of 98+% CO_2 being vented which is available for collection and transportation via pipeline to the South Louisiana oil fields. One such pipeline, 151 MMSCFD, 120 miles is presented in Section 8.4 of this report and a first pass economic evaluation has been made.

Based on current CO_2 supplies it could take as many as 50 years to recover the potential 300 million barrels of incremental oil once enhanced recovery efforts are initiated.

Supply Situation - Region IV

The development of CO_2 enhanced oil recovery projects in West Virginia is in the preliminary stages at this time. There have been only three test projects to date: Columbia Gas Transmission at Granny's Creek, Clay County; Guyan Oil at Griffithville, Lincoln County; and Pennzoil at Rock Creek, Roane County. Results from these tests have not been encouraging.

Columbia has recovered 8500 BBL of incremental oil (35 BPD) as of April, 1978, consuming 170 MMSCF CO_2 in one year. Whereas the project was technically successful, the project manager was quoted as saying "It wouldn't have paid out at \$100/BBL"[38]. Despite this Columbia plans to develop the natural deposits in the Indian Creek Field in Kanawah County for both CO_2 and natural gas content. Potential CO_2 reserves are 130 Billion SCF delivered at a rate of 20 MMSCFD CO_2 over the next 14 years to the Granny Creek project.

Guyan Oil estimates incremental oil recovery by CO_2 to be 8100 BBL/acre for their 90 acre pilot project in the Berea sand fields of Lincoln County[39]. Since the test is in the infant stage no recovery efficiencies are available yet. Assuming an inefficient CO_2 requirement of 25,000 SCF CO_2/BBL for the test will result in a net CO_2 requirement of 18 BSCF. If the entire 10,000 acre field is productive at 8100 BBL/acre and requires no more than 15,000 SCF CO_2/BBL, then up to 1.2 TSCF CO_2 could ultimately be required. The 81 million barrels of oil so produced would represent 23% of the OOIP and as such this may be a somewhat optimistic projected recovery. Discovered in 1907, cummulative production for the Berea field has been only 14 million barrels.[47]

The Pennzoil Rock Creek field is potentially the largest test project planned by DOE/Industry in West Virginia. There are 11,200 productive acres involved which could yield as much as 40 million BBLS of incremental oil if the CO_2 test is successful,[39] requiring 600 BSCF CO_2 if a CO_2 requirement of 15,000 SCF CO_2/BBL is assumed. However, as of March, 1978, production only averaged 2 BPD, so until test rates are accelerated it will be difficult to make accurate projection on recoverability.

All of these recovery projections in West Virginia are highly conjectural since the test projects are in early stages of development. Nominally, it appears that near 2 TSCF CO_2 could be required if all projects test out as successfully and economically as the operators hope and the candidate fields are as large as expected. However, the consensus of industry is that the West Virginia area has little to offer in the way of CO_2 enhanced incremental production which will significantly affect the natural energy supply.

Supply Situation - Los Angeles Basin

Data on the supply and demand situation for CO_2 in the Los Angeles Basin area has been taken from the Lawrence-Allison report.[28] This report identified refinery and power plants as the two major sources of CO_2 and found no evidence of naturally occurring CO_2.

Total CO_2 supply is estimated to be 1.28 BSCFD with approximately 1.23 BSCFD available from power plants and 0.5 BSCFD available from refinery and ammonia plants. Lawrence-Allison forecasts the potential CO_2 supply to remain substantially constant during the next several years. The largest refinery source of CO_2 in the area is the three-unit Chevron refinery with a combined CO_2 output of 33.6 MMSCFD. It is located 6-7 miles from the more promising Torrence Field.

Little was concluded by Lawrence-Allison on the expected incremental recoverable oil by CO_2 injection. Estimates include projections that CO_2 injection could be responsible for recovering 100 BBL/acre-foot with a net CO_2 requirement of 12-25 MSCF CO_2/BBL oil. It is conceivable on evaluating a Torrence Field-Chevron project that 1,340 BPD incremental oil could be recovered using Chevron as the CO_2 source and a CO_2 requirement of 25 MSCF CO_2/BBL. This would increase production 20% over the 1977 levels of 6,600 BPD using water flood secondary recovery efforts alone. This is somewhat higher than Chevron's and Lawrence-Allison's expectations of a 6-7% increase in recovery of OOIP due to CO_2 injection after water flood. So it seems likely that CO_2 demand can be met easily in the Los Angeles area primarily because of the projected low demand on CO_2 for enhanced recovery efforts.

3.1.3 Sources of Survey Data

The sources of CO_2 considered for this study include both natural and industrial by-product CO_2. Information on naturally occurring CO_2 within the four-region area is supplied from the U.S. Bureau of Mines gas analysis data base, GURC REPORT [17] No. 165 on naturally occurring CO_2, various state boards, published CO_2 survey reports, limited industrial contact, and from National Cryo-Chemics, Inc. through a subcontract developed in connection with this report.[10] Selected portions of the Cryo-Chemics report are included in Appendix A2. Data for the Los Angeles Basis is derived from the Lawrence-Allison report. [28]

The role of Pullman-Kellogg in the determination of the quantities of naturally occurring carbon dioxide is limited primarily to the consolidation information contained in these various reports and studies and to the resolution of conflicting data, to the best of its ability. Pullman-Kellogg has neither attempted to reevaluate geological data nor to establish criteria to assist in the prediction of recoverable CO_2 reserves.

Pullman-Kellogg is in a position, however, to give a reasonably accurate estimate of the amount of CO_2 available as industrial by-product. The industrial sources considered and tabulated in this survey include:

1. Power plant stack gases
2. Flue gases from cement plants
3. Vent gases from ammonia and chemical plants
4. Vent gases from refinery and hydrogen plants
5. Vent gases from natural gas processing plants
6. Vent gases from proposed SNG plants.

For each of these sources as an exhaustive search as possible has been undertaken given the time frame within which this report had to be prepared. Furthermore, as many point sources as possible have been given individual attention to determine the actual amount of CO_2 that could be expected to be available from each source. Prorating factors applied to classes of sources to determine CO_2 availability have been avoided whenever possible.

Power Plant Stack Gases

The list of current and projected power plants is taken from the 1977 edition of the National Coal Association's publication, Steam Electric Plant Factors. This publication contains data on a plant-by-plant basis regarding generation capacity, actual generation for 1976, type, quality and quantity of fuel consumed, and origin of coal for coal fired plants. Using this data and information contained within the Keystone Coal Industry Manual[16], it is possible to make a reasonable estimate of the composition of the fuel being burned in each plant. For example, coal based power plants have been matched to type of coal used since this determines the quantity of carbon dioxide produced as well as the level of sulfur oxides in the stack gas. Some power plants, e.g. the coal fired plants of Texas, have even been individually contacted in order to get actual data on fuel composition and consumption; however, it is not practical to contact all plants considering the number involved.

From actual fuel consumption an accurate estimate of the quantity of CO_2 available in the stack gases can be determined. For each fuel the assumption of burning at 30% excess air is made for purposes of establishing flue gas CO_2 concentration; however, excess air does not contribute significantly to the ultimate amount of CO_2 available from a given fuel. The quantities of power plant CO_2 presented in the tabular results in the appendix reflect the calculated actual CO_2 production for the year 1976.

Flue Gases from Cement Plants

The list of cement plants in operation throughout the four regions surveyed has been taken from the Portland Cement Association publication U. S. Portland Cement Industry: Plant Information Summary, December 31, 1977,[3] and from the directory of cement plants in the Keystone Coal Industry Manual[16] which contains information on plant capacity and fuel consumption. Each cement plant operator was sent a questionaire soliciting information on plant capacity, actual production rate, fuel quality (composition) and consumption, the amount of calcined calcium carbonate, flue gas analysis and vent rate. Response was moderate (about 30%) and the data obtained has been used to estimate the amounts of CO_2 available from those operators who did not respond, by closest matching based on capacity and fuel type and then prorating based on capacity.

Vent Gases from Ammonia and Chemical Plants

The list of current ammonia and other chemical plants producing carbon dioxide as a by-product is obtained from the 1974 edition of the Stanford Research Institute - Chemical Economics Handbook (CEH),[5] the TVA Fertiliza Information System,[4] the Corneli paper on Production Economics for Hydrogen, Ammonia and Methanol During the 1980-2000 Period,[18] and from in-house information drawing on over 40 years experience in the ammonia/hydrogen business. The Chemical Economics Handbook reports carbon dioxide generation capacity by company, location, and source stating the total CO_2 generated, percent vented and percent recovered. The quantities tabulated in Appendix A1 are the vented amounts. The chief industries covered in the CEH are ammonia, hydrogen, and ethylene oxide plants.

For those plants which explicit CO_2 data is unavailable certain assumptions had to be made concerning available CO_2. First, if the amount recaptured or vented is unknown, then it is assumed that all the potential CO_2 is available for use. Second, for the ammonia plants with unknown vent rates it is assumed that 22,000 SCF CO_2/TON NH_3 capacity is available. This figure is derived from the average of a number of actual gas-feed ammonia plants designed by Pullman-Kellogg; there is little variation, the range being from 21,000 to 23,000 SCF CO_2/TON NH_3. Carbon dioxide purity from ammonia vents is typically greater than 98% and is usually available at near atmospheric to a few psig positive pressure.

Vent Gases from Refinery and Hydrogen Plants

The contribution of refinery by-product CO_2 to the aggregate CO_2 supply picture is small when compared to other classes of sources. Typically the quantities available are less than 10 MMSCFD at a given refinery, but range upwards to 34 MMSCFD. Exact CO_2 quantities from a refinery are difficult to determine because each refinery is different in feedstock, processing steps, etc. The only certain way to establish the qualities of CO_2 available would be to contact the chief engineer of each refinery. Since this is not practical given the time frame of this report, estimates of potential CO_2 available will be made based on extensive in-house experience in refinery design.

The two largest sources of CO_2 from a refinery are power generation flue gases and catalytic cracking catalyst regeneration by-product. From Nelson[40] the average heat requirement for a refinery can be estimated as 691,400 BTU/BBL crude oil refined. Assuming that one-half of this heat can be recovered and that an average of 3.84 MMSCFD CO_2 is available in the flue gas from burning 100MM BTU/hr fuel oil then we have calculated that 550 SCF CO_2 will be generated per barrel of crude refined due to power generation. For catalyst regeneration it is estimated that 205 SCF CO_2 will be generated per barrel of crude feed to the catalytic cracking units.[8] The concentration of CO_2 in regeneration vent gas is approximately 8%.

Collecting and cleaning the CO_2 from refinery stacks will be difficult because plot space for additional processing equipment is generally lacking and because of pipe routing problems. Compounding this problem is the difficulty of collecting all the different power generating flue gases from within the refinery for routing to a central processing point. In general, however, the gases available from the catalyst regeneration step are available from a single point and are really the only refinery source that can be given any practical consideration. The quantities of CO_2 reported from refineries listed in the tables in the appendix are the potential CO_2 available from the regeneration step only and are calculated based on the reported feeds to the catalytic cracking units,[41] noted in the comments column. If the CO_2 from the flue gases is to be considered, then one should use the 550 SCF CO_2/BBL crude factor applied to the total capacity of the refinery. Use of these factors to estimate the amount of CO_2 available rather than determining actual vent rates should not significantly distort the aggregate CO_2 supply. These figures will serve primarily as a guide to potentially significant CO_2 contributions from refineries.

The bulk of the hydrogen produced in the U. S. is consumed in ammonia manufacture and in petroleum refineries, principally for hydrodesulfurization and hydrocracking of heavy hydrocarbons. Hydrogen is produced primarily by the steam reforming of natural gas or some other light hydrocarbon feedstock. From Pullman-Kellogg's experience in having built over 12 hydrogen plants, it is known that the amount of carbon dioxide generated differs little from 33 SCF CO_2 for every 100 SCF hydrogen produced. This factor is used to estimate the quantities of CO_2 available based on the published plant capacities[4,18] when the vented quantities are not explicitly known.

Vent Gases from Natural Gas Processing Plants

Information on the amount of CO_2 available from natural gas treating plants comes primarily from contact with gas plant operators. At the beginning of this project a data base of naturally-occurring carbon dioxide was developed by GURC, based in part on data originally published by the U. S. Bureau of Mines Helium Division, which included over 11,000 analyses of gas samples from oil and gas wells.[42] GURC edited the original information for errors, deleted all samples containing less than 5% carbon dioxide and expanded the data base with additional data from literature and with information from personal communication with state geologists. The resulting data base presently contains over six hundred records representing occurrences of natural carbon dioxide greater than 5 mole percent in 27 states.

The data base developed by GURC has been used as a tool to facilitate both the study of natural sources of carbon dioxide and to study the availability of carbon dioxide from acid gas processing plants. A list was prepared of gas treatment plants[41] located in counties having samples in the GURC data base. Over seventy plants, each of which processes at least 50 MMSCFD of natural gas, were surveyed by mail and phone contacts. Additionally, the list was supplemented with information obtained from acid gas removal process licensors and from engineering firms engaged in the design and construction of gas treating plants.

Results of this survey indicate that, in general, acid gas treatment plants are not a significant source of carbon dioxide for enhanced oil recovery. There are isolated exceptions, however. In general, the amount of carbon dioxide available is small, typically <10 MMSCFD, and widely distributed. In some cases the gas is not available since operators are already injecting the carbon dioxide - rich gas into nearby oil reserves. Those plants reported to vent carbon dioxide as a by-product are listed in the appendix as "N.G. Processing" facilities.

Vent Gases from Proposed SNG Plants

The CO_2 vent gases from a coal gasification (SNG) plant represent the largest single potential industrial source of carbon dioxide. The attractiveness of SNG by-product CO_2 is attributed to its anticipated purity, typically greater than 85% CO_2, with the balance composed primarily of nitrogen and oxygen.[8] There could be a few ppm sulfur depending on the coal being gasified.

The list of proposed coal gasification projects is constantly changing, with fewer projects surviving the list each time it is up-dated.

The source from which the list of currently active projects has been taken is the July 12, 1978, Chemical Week.[49] Only three plants in the areas covered by this study have advanced far enough to warrent serious consideration as potential sources of CO_2. These are:

1.	El Paso Natural Gas	Four Corners area, N. M.	72 MMSCFD SNG
2.	Texas Eastern Trans./ Pacific Lighting	Four Corners area, N. M.	275 MMSCFD SNG
3.	Panhandle Eastern Pipeline	Converse Co., Wyoming	275 MMSCFD SNG

From studies done by Pullman-Kellogg in connection with one of the proposed SNG projects it has been estimated that 1.14 SCF CO_2 will be generated per SCF of SNG gas produced. This factor, then, has been applied to the three projects listed in the table to estimate the amount of CO_2 that might become available.

The significance of SNG by-product CO_2 lies in its potential as a single source of CO_2 for an enhanced recovery project. As an example, the Texas Eastern project could be a sole source of CO_2 for a 300 MMSCFD pipeline to supply the oil fields of West Texas. It is not known, however, at this time whether further purification of the carbon dioxide would be required in order for it to be functional as an enhanced recovery agent. Depending on the sulfur content of the coal gasified there could also be enough H_2S present in the by-product that would exceed the interstate pipeline standard of 1/4 grain/100 SCF. Purification of this quantity of CO_2, if the project goes to fruition, could significantly affect its competetive position with natural sources occurring in the area.

3.2 PRESENTATION OF CARBON DIOXIDE SOURCE MAPS

This section contains the maps of all the carbon dioxide sources surveyed. Each region, as previously defined, is on an individual map. The purpose of the maps is to enable one to spot quickly potential CO_2 sources for a given candidate CO_2 miscible flood oil field and has been so used in the preparation of the specific cases described in Section 8 of this report. The regions have been somewhat arbitrarily defined with consideration given to current commercial interest in CO_2 EOR projects, and the maps have been drawn with considerable overlap so as not to impair the discovery of suitable CO_2 supplies for a given project.

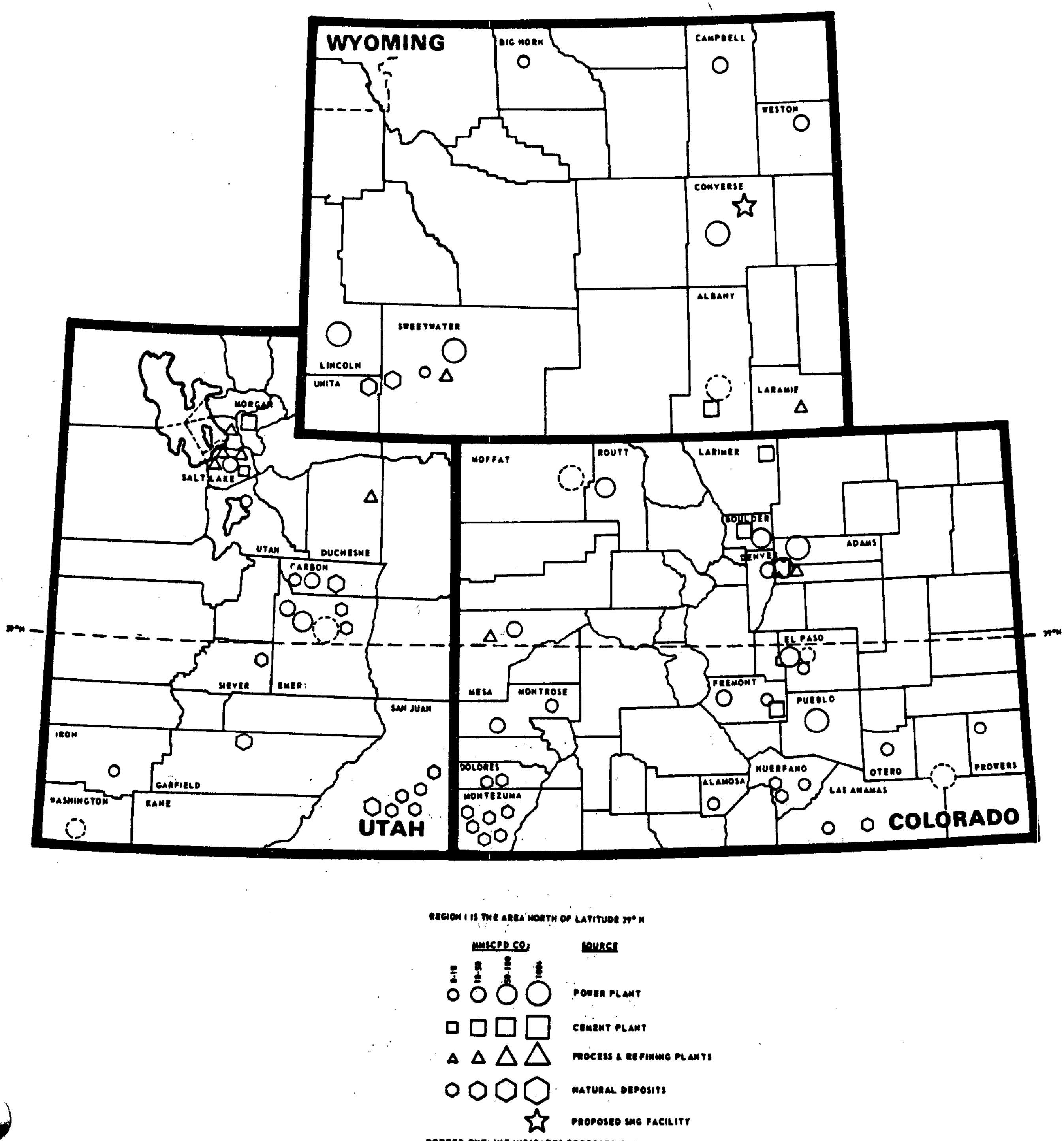
The maps are drawn with size related symbols indicating approximate quantities of CO_2 expected from a given source. Each county within which a source is located is identified. Details on each source noted are found in the appendix in the Key to Carbon Dioxide Source Maps. The sources identified are those which are known to have uncommitted CO_2 available, except where noted. If the availability of CO_2 from a given source is unconfirmed, then the total amount expected to be available is listed. It is expected, of course, that any one seriously considering one of the sources listed as a potential CO_2 supply will contact the operator to confirm the figures given and to be advised of special problems that might be associated with the source.

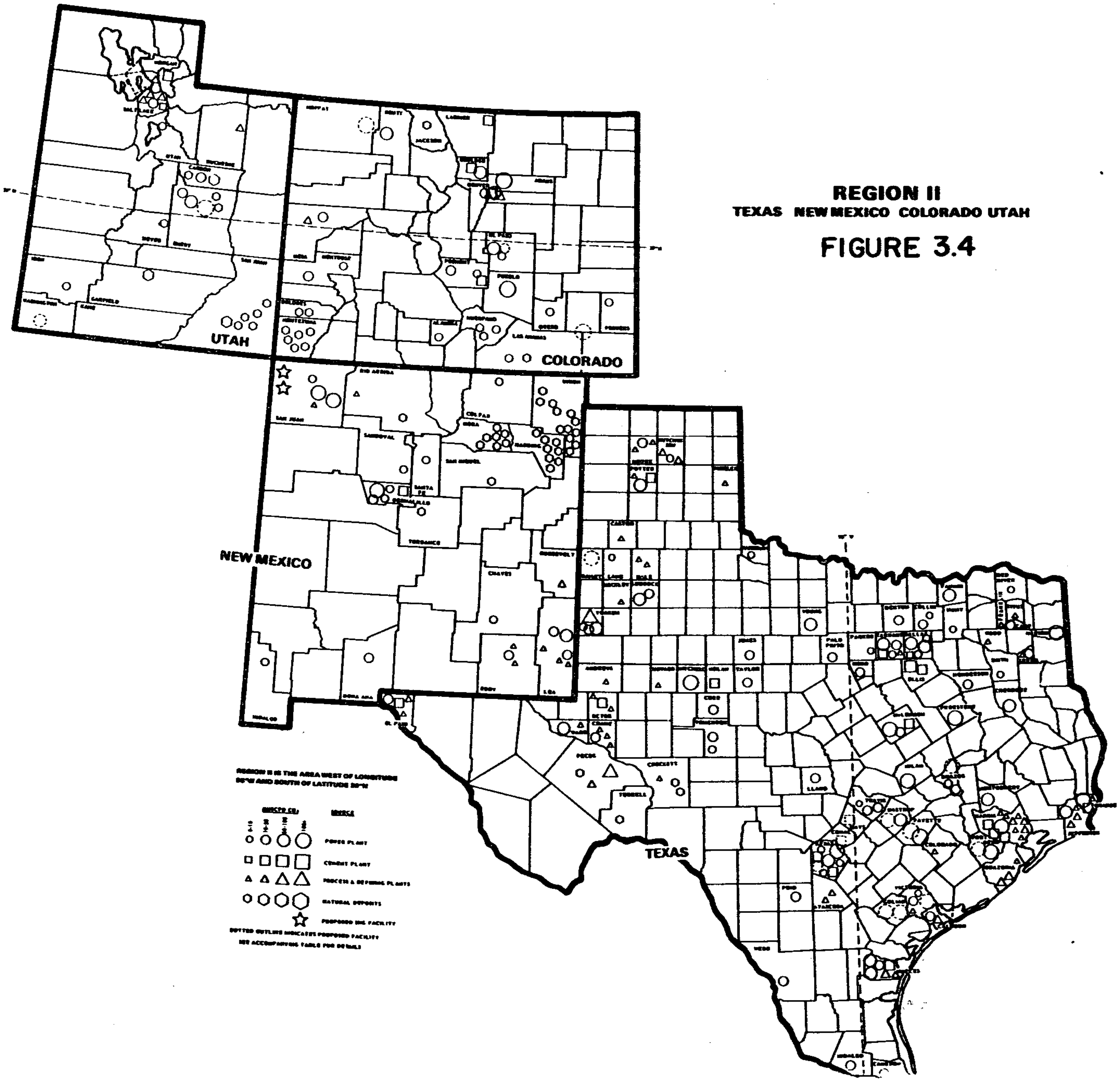
The map and information related to the Los Angeles Basin has been provided by the Lawrence-Allison Corp., who issued a separate report on CO_2 enhanced oil recovery for the Los Angeles Basin.[28] Information for this area is reproduced without modification.

REGION I

WYOMING, COLORADO, UTAH

FIGURE 3.3





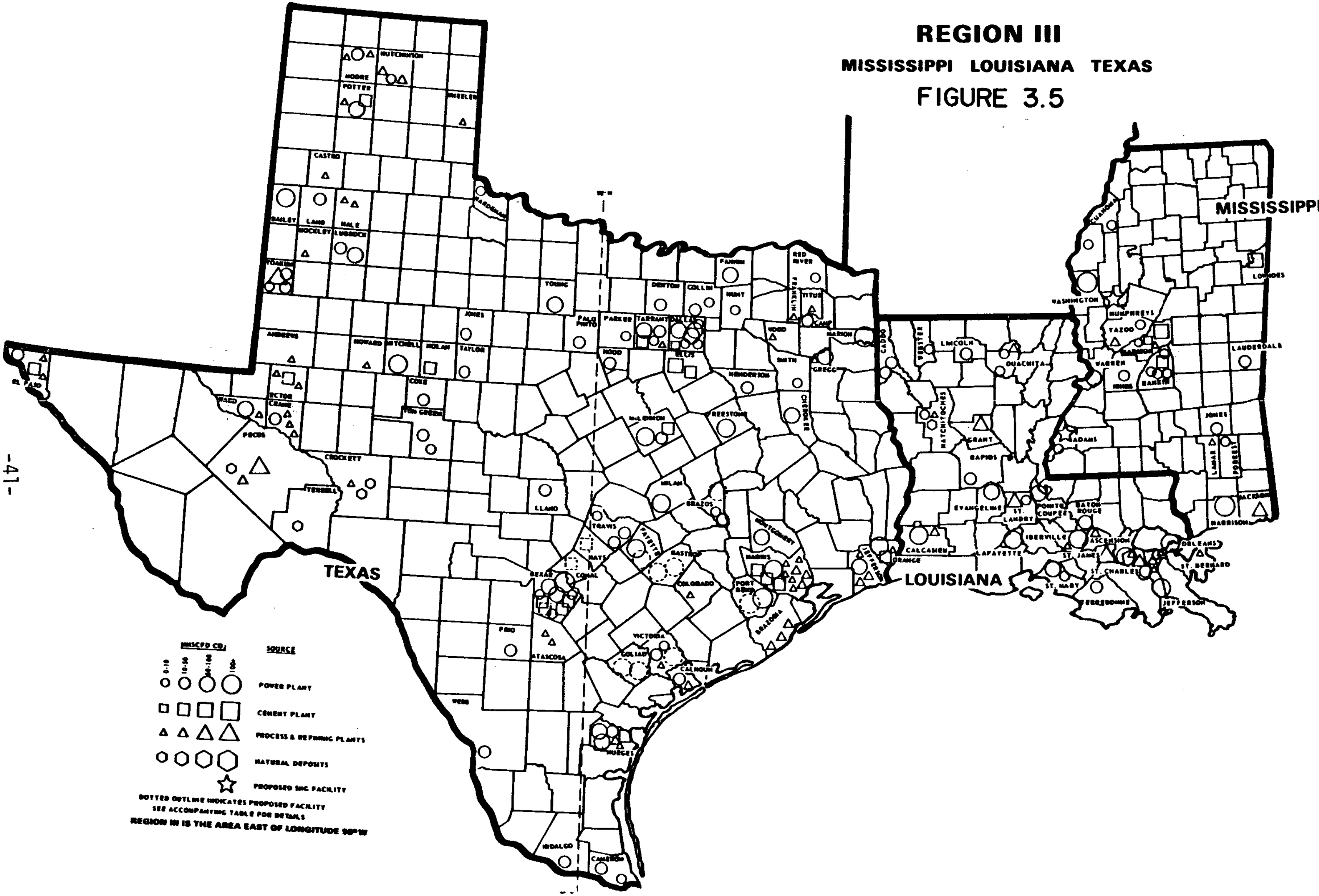
REGION II
TEXAS NEW MEXICO COLORADO UTAH

FIGURE 3.4

REGION III

MISSISSIPPI LOUISIANA TEXAS

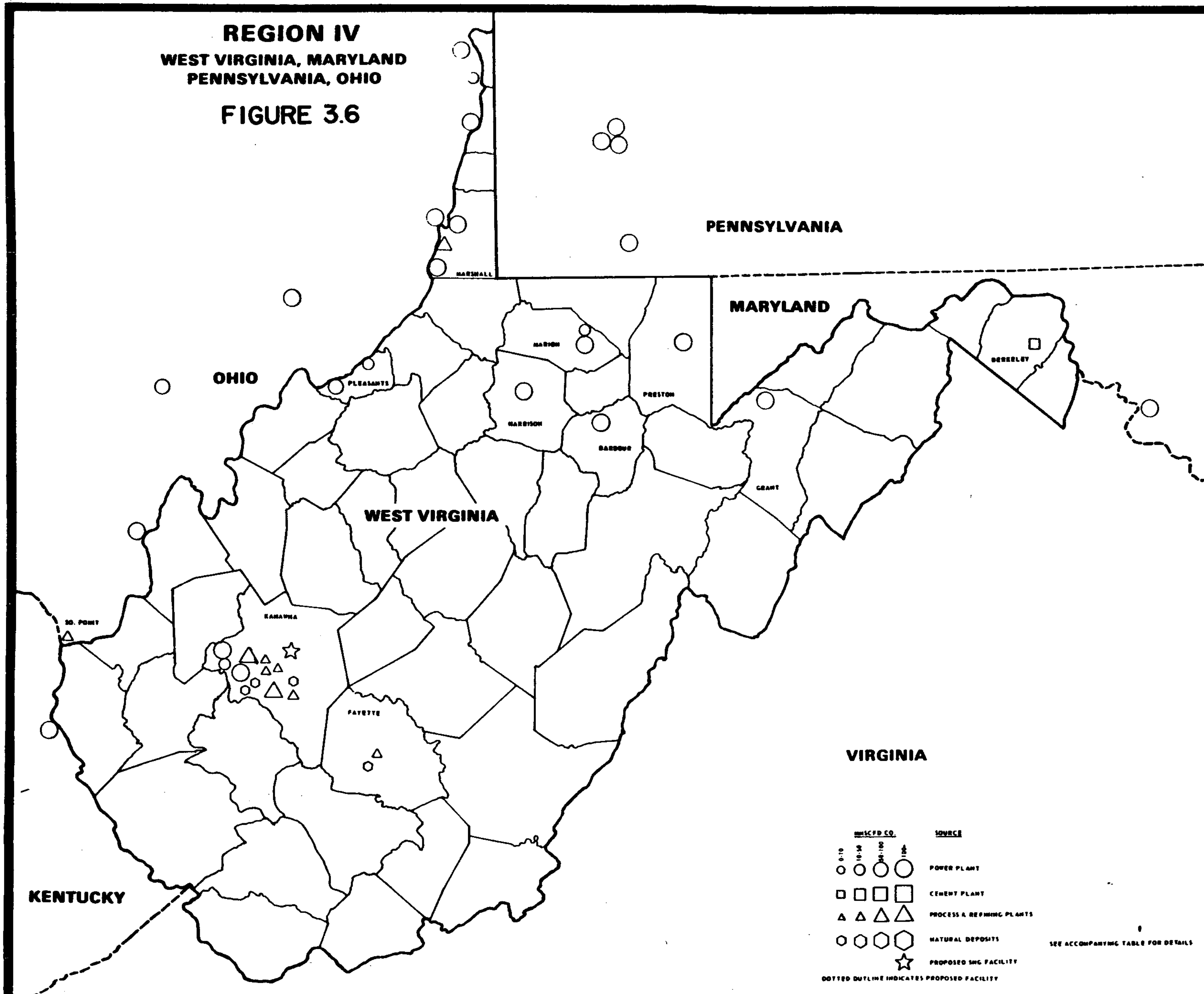
FIGURE 3.5



REGION IV

WEST VIRGINIA, MARYLAND PENNSYLVANIA, OHIO

FIGURE 3.6



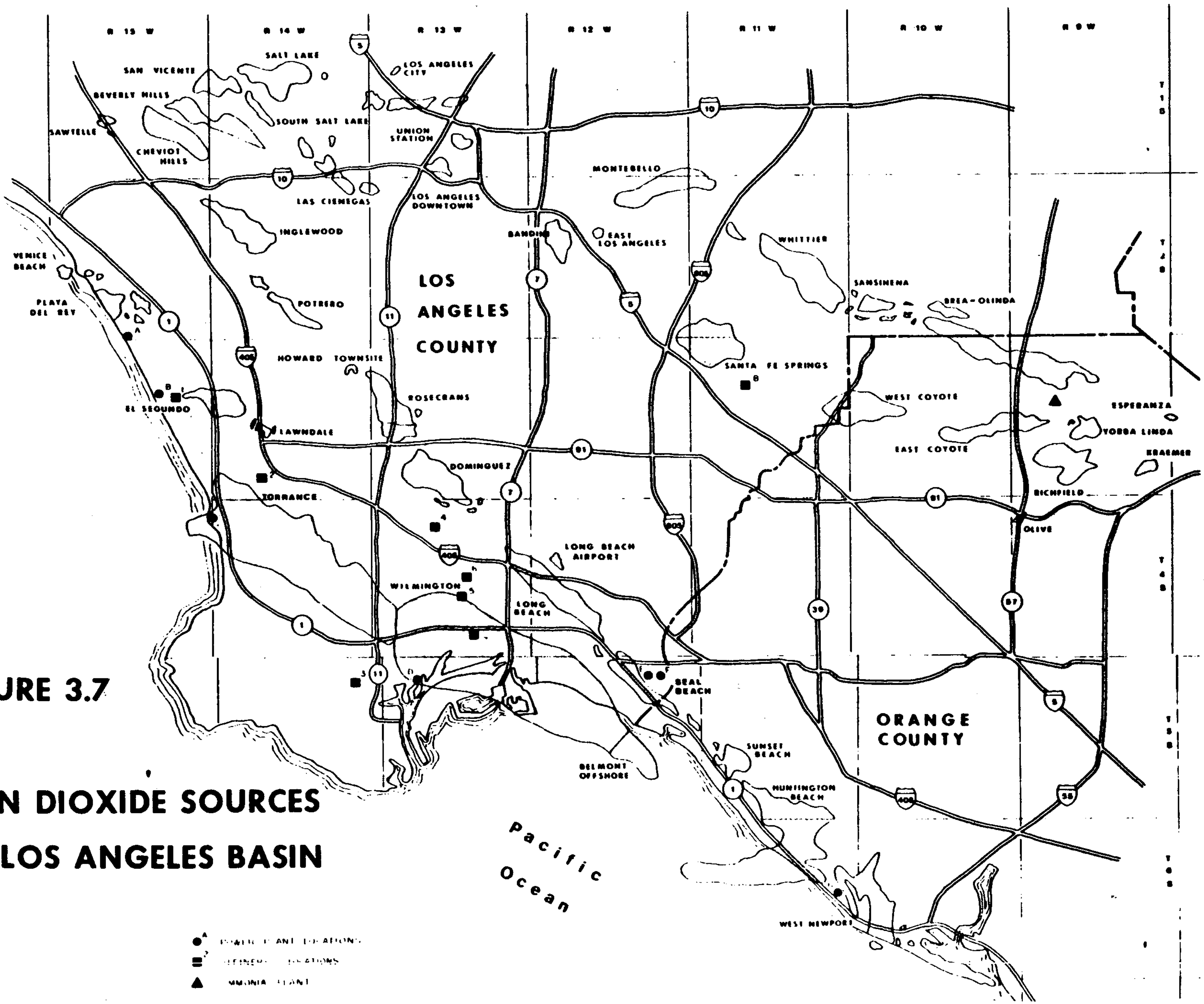


FIGURE 3.7

CARBON DIOXIDE SOURCES IN THE LOS ANGELES BASIN

3.3 CONCLUSION

The primary purpose of this section is to locate all potential CO_2 sources within those areas that could have oil fields amenable to carbon dioxide miscible flooding. Those sources considered are natural deposits, industrial sources such as ammonia, hydrogen, ethylene oxide, and natural gas treating plants and refineries, as well as power plants, cement plants, and proposed SNG plants. These sources have been mapped and tabulated. Also in conjunction with this an estimate of potential CO_2 demand for enhanced oil recovery is indicated.

Comparison of supply and demand indicates that the potential peak CO_2 demand of 7 - 11 BSCFD could be met from the total CO_2 supply of 28.4 BSCFD; however, the source of CO_2 is not always convenient to the candidate oil fields. A potential CO_2 shortage in the West Texas fields does exist as the potential demand of 6.3 BSCFD could have to be supplied from an estimated maximum availability of only 6.0 BSCFD. The single largest sources considered are proposed SNG plants and natural carbon dioxide deposits in Colorado and New Mexico. Natural CO_2 reserves are expected to range from 20 - 28 TSCF with 12 - 18 TSCF to be available for the West Texas oil fields.

It is recommended that if serious consideration be given to any particular source of CO_2 that personal contact be made with the proprietor of that source. Individual sources could have circumstances rendering the amount carbon dioxide listed unavailable or unfit for use. It is beyond the scope of this report to double-check the availability of each source listed. These figures are best suited for project planning estimates and care has been taken to ensure the best data possible.

COST OF PURCHASING

PART 4

4.1 COST OF PURCHASING

Costs shown in this section represent survey of purchase price of carbon dioxide prevailing in today's market place. This is the price that an oil reservoir operator may pay to the producer of carbon dioxide. It should be recognized that purchase costs for CO_2 could be escalated rapidly by an increase in demand of CO_2 for EOR efforts. Many chemical companies and the ammonia producers were contacted to establish price of CO_2. The survey was then extended to the natural gas processing plants and the utility companies. Wellhead prices for natural carbon dioxide are based on reports from Colorado but it is uncertain whether any such sales have taken place.

Factors influencing the costs are gas purity and location to some degree. High purity CO_2 sources, such as, ammonia plant vents and SNG plants command higher prices for CO_2 than low purity sources like flue gas stacks from power plants and cement plants. Low purity sources will require significant clean-up before CO_2 from such sources can be used for EOR by miscible flooding. Wellhead prices for natural carbon dioxide will depend on purity, pressure of the source and on location of the source. The present cost of purchasing carbon dioxide from the sources studied varies from a low of $1-2/Ton for sources such as power plant stacks to a high of $15-20/Ton F.O.B. plant for a liquid CO_2 plant. These costs are summarized in the Table 4.1.

TABLE 4.1

CURRENT ESTIMATE OF

COST OF PURCHASING CO_2

	COST	
ABOVE GROUND SOURCES	$/TON	CENTS/1000 SCF
PROCESS VENTS (PURE, LOW PRESSURE)*	4-8	25-50
COMBUSTION VENTS (LOW PURITY, LOW PRESSURE)	1-2	6-12
AS LIQUID F. O. B. PLANT	15-20	85-115
NATURAL SOURCES		
WELLHEAD PRICE**	4-6 (10)	25-35 (10)

*AMMONIA, SNG, HYDROGEN PLANTS - NATURAL GAS, LPG, NAPHTHA FEEDSTOCKS, PROCESSES BASED ON PARTIAL OXIDATION OF FEEDSTOCKS OR GASIFICATION OF COAL SHOULD COMMAND A LOWER PRICE DUE TO LOWER PURITY.

**COLORADO AREA

COST OF NATURAL CO_2 PRODUCTION

PART 5

5.1 INTRODUCTION

The first phase study concluded that natural sources of CO_2 have perhaps the greatest potential as a future source of low cost CO_2 for EOR projects but the costs involved in the development and production of natural CO_2 were not determined. Thus, in the second phase study one of the objectives is to estimate the cost of producing natural CO_2.

For naturally occurring CO_2 it was agreed that GURC would assist the Pullman Kellogg Co. in locating and characterizing the major source areas and estimate costs to drill and produce the CO_2 into a trunk-line. Therefore, this section of the second phase study report is based on GURC's report.[17] Pullman Kellogg provided investment and operating cost estimates for compression and treatment.

GURC's overall responsibility include the following three tasks:

1. Evaluation of reservoir rock properties and gas composition of potential CO_2 sources
2. Estimate of costs for drilling and producing naturally occurring CO_2 reservoirs
3. Field utilization of CO_2 in EOR projects (Information used in Section 8)

5.1.1 Data Base Support

A Carbon Dioxide Data Base was constructed, early in the project, to facilitate data acquisition, storage, and analysis. The Bureau of Mines Helium Division has collected over 11,500 analyses of gas samples from oil and gas wells since 1917 and reported them in its publication, "Analysis of Natural Gases, 1917-74". GURC obtained a magnetic tape of this data and the 1975 and 1976 updates. All sample data with carbon dioxide content greater than 5 mole percent were stripped from this tape. The data were edited for errors, and numerous records were amended and supplemented with additional data to compile the final version of the CO_2 Data Base.

5.1.2 Source Delineation

Initial delineation of areas containing potential reserves of CO_2 was conducted using the gas quality and reservoir characteristic data in the CO_2 Data Base.

The regions considered include West Virginia, the Jackson Dome area in Mississippi, the Delaware-Val Verde Basins in West Texas, and 9 sub-areas of the Rocky Mountains, as follows:

1. Southwest Colorado, McElmo Dome area
2. South Central Colorado, Sheep Mountain area
3. North Central Colorado, North Park Basin
4. Northeast New Mexico, Sierra Grande Arch area
5. Northwest New Mexico, San Juan Basin
6. Central Utah, Farnham Dome area
7. Central Utah, Gordon Creek area

8. Southeast Utah, Paradox Basin

9. Wyoming, Sweetwater and Uinta Counties area

Gas compositions in the above areas are varied.

All the Rocky Mountain sub-areas, however, contain CO_2 greater than 85 mole percent, with a few mole percent methane and/or nitrogen as contaminants.

Gas sampled from the Smackover and Norphlet trends in the Jackson Dome area contain CO_2 concentrations from 70-99 percent, associated with considerable percentages of H_2S in some localities. Commercial hydrocarbon production occurs in the downdip Smackover and Norphlet with lower concentrations of CO_2 reported. Gas containing CO_2 in West Virginia ranges from 10 to 83 percent CO_2 and is associated with a methane fraction from 40 to 80 percent. Small percentages of nitrogen are usually present.

5.1.3 Selection Criteria for Detailed Study

Three criteria were established to determine which of these source areas warranted detailed reservoir and economic evaluation:

1. Reserves apparently sufficient for pipeline consideration
2. Proximity to miscible flood candidate reservoirs
3. Evidence of strong industry interest

On this basis four areas were chosen for further geologic, reservoir, and economic analysis:

1. McElmo Dome area, southwest Colorado

2. Sheep Mountain area, south central Colorado

3. Northeast New Mexico

4. Jackson Dome area, Mississippi

These areas, each representing a unique geologic environment and posing varying drilling and production conditions, are reasonably representative of the range of expected naturally-occurring CO_2 sources. These areas are discussed in Sections 5.4 through 5.7.

5.2 RESERVOIR CHARACTERIZATION METHODOLOGY

Reservoir rock properties such as permeability, porosity, effective net pay, bottom hole pressure, and temperature are critical parameters for estimating original gas in place (OGIP) and sustained gas deliverability. Detailed site specific reservoir description and well test data for naturally-occurring carbon dioxide are not readily available in the public domain. Thus, visits were made to state geological and regulatory offices to obtain well data from records available for public inspection.

In cases where local reservoir data were unavailable, reservoir properties were estimated from regional data. Bottom hole temperatures were computed from the AAPG Geothermal Gradient Map of North America[44]. Initial reservoir pressures were estimated from pressure versus depth plots using regional data available in the CO_2 Data Base. To assess the influence on OGIP of variations in porosity, net pay and gas saturation, several sensitivity studies were conducted.

Reservoir Analysis Methodology

For each area studied, the OGIP (pure CO_2) for one acre-foot of pore space was calculated, as a function of initial BHP, temperature and Z factor using the relation:

$$G = 43{,}560\ T_{sc}P_i/P_{sc}Z_iT_i$$

OGIP per acre-foot of pore space was plotted versus depth, as illustrated in Figure 5.1. As shown, the OGIP/acre-foot exhibits a rapid increase with depth down to approximately 6,000 feet. Below 6,000 feet OGIP/acre-foot increases minimally.

Each of the geologic areas of interest was characterized by the depths at which CO_2 occurs and the anticipated range of net pay thickness.

Effective net pay was estimated after analysis of the perforated interval, core analyses, and electric log interpretation. In each area, net pay distribution of wells drilled to date approximates log normal. The 10th, 50th, and 90th percentile were assumed representative of distribution of pay in the "worst likely", "modal", and "best likely" wells, respectively, when each reservoir or producing area is fully developed.

Porosities were estimated from core data and compensated neutron, sonic and density logs.

In a given reservoir the OGIP per 640 acre-spaced well is a function of the gas saturation porosity product (SgØ) and net pay. As shown by Figure 5.3, in a typical 8,000 foot deep reservoir, the OGIP per 640 acre-spaced well could vary from about 25 to 230 BSCF, for net thicknesses of 60 to 150 feet and reasonable values of SgØ for commercial wells.

It is evident that, in each area, there will be variation in reservoir rock quality and, thus, in OGIP and deliverability. We have elected to represent the anticipated variation of each area as the performance of a set of wells of varying quality. Thus, in each area, OGIP has been estimated for a "best likely well", a "modal well", and a "worst likely well", considering observed variations in net pay, porosity-gas saturation product, and bottom hole pressure.

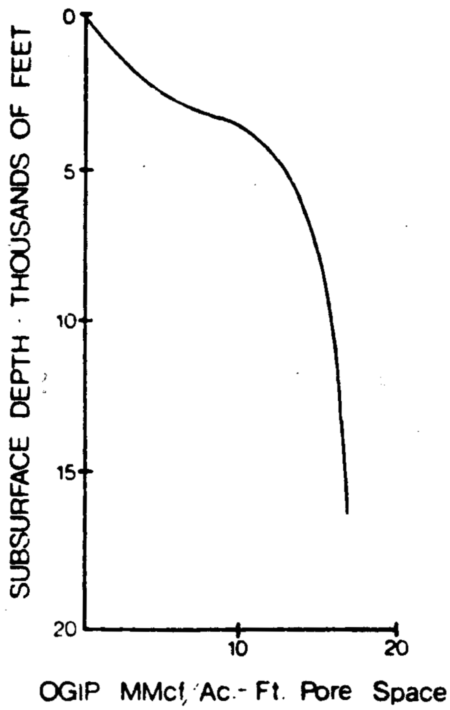
Field test data, including flow rates, flowing tubing pressures and drawdowns were then analyzed. For each area a range of average sustained deliverabilities and flowing tubing pressures was estimated. Production was assumed to be by depletion drive for all areas investigated.

Estimated recovery efficiencies for each area were calculated from abandonment pressures determined after discussions with industry personnel. In a depletion drive reservoir, recovery efficiencies are a function of initial BHP and abandonment pressures as indicated by the following equation:

$$E_R = 1 - \frac{P_a}{z_a} \bigg/ \frac{P_i}{z_i}$$

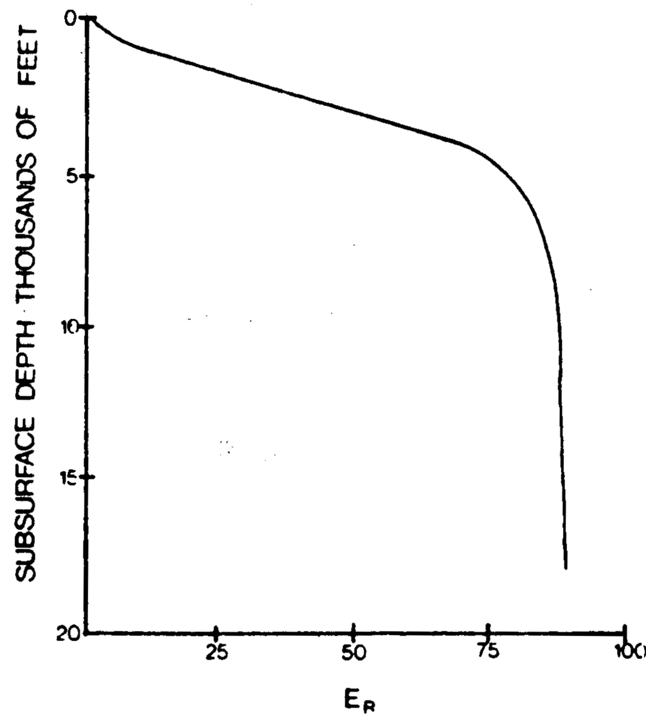
As shown in Figure 5.2, recovery efficiency, calculated for a wellhead abandonment pressure (P_a) of 400 psig, increases rapidly with depth to about 88 percent at 4,000 feet, then stabilizes. Wellhead abandonment pressures for each area were converted to bottom hole abandonment pressures to account for regional variations in depth. Actual abandonment pressures in any area will be a function of compression intake pressure and drawdown.

For each area, a set of well "models" spanning the probable variation in rock quality, OGIP, and sustained productivity was created. Flowing bottom hole pressure decline curves were constructed as a function of cumulative gas produced, allowing easy correlation with varying flow rates and recovery efficiencies. Individual well economics on these representative well models were then calculated.



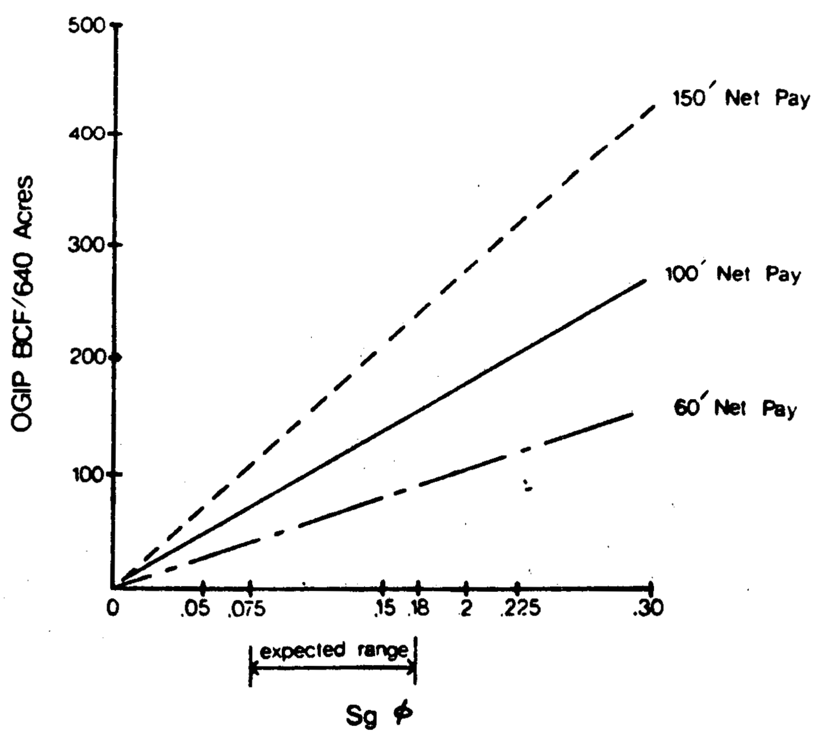
OGIP vs. DEPTH

FIGURE 5.1



Recovery Efficiency vs. Depth
for a Wellhead Abandonment
Pressure of 400 psig

FIGURE 5.2



OGIP vs. Gas Saturation X Porosity
for Varying Thicknesses for
a Reservoir at a Depth of 8,000 feet

FIGURE 5.3

5.3 ECONOMIC ANALYSIS METHODOLOGY

A simple economic "model" was constructed using estimated investment and operating costs, and reservoir sensitivity studies as input. The objective was to develop a range of per-mcf costs for CO_2, dehydrated and delivered to an interstate pipeline at specified pressures for each of the analyzed areas.

In an effort to estimate the probable costs of drilling CO_2 wells, cost data were collected on typical natural gas wells through communication with the AAPG, AGA, API, and numerous industry personnel. Communication with industry personnel indicated natural gas well cost data were insufficient for accurate estimation of CO_2 well costs. Reported drilling costs for CO_2 wells in the areas of interest were found to be substantially higher than originally anticipated. These higher drilling and completion costs are explained in part by the following:

1. Road and site costs, rugged topography
2. Remote locations; distances from traditional oil field services and supply centers
3. Problems with hole collapse in shale formations, requiring high cost drilling fluids (oil base muds)
4. Drilling with air to limit formation damage in low permeability zones

5. Encountering thick sections of hard igneous rock
6. Large casing and tubing sizes
7. Highly corrosive nature of CO_2 requiring special coatings and stainless steel wellheads
8. Presence of archeological sites and required studies
9. High cost directional drilling to deviate holes for topographic and environmental reasons.
10. Requirement for buried gathering lines.

Drilling costs, thus, are highly site specific and regionally variant due to topographic and geologic variations. Completed well costs herein were based on those reported by operators in each area. A dry hole "cost" was estimated in each area, based on historical dry hole ratios and estimated intangible well costs, for both exploration and development drilling phases.

Lease bonus fees and estimated probably operating costs for CO_2 wells were obtained from industry personnel. Investment and operating costs for surface processing equipment were supplied by Pullman Kellogg. For all areas other than New Mexico, surface equipment design was based on 50 MMSCFD capacity "modules" composed of a dehydration, compression, hydrate inhibiting, and gathering system. For the New Mexico area economic analysis compression system investment and operating costs were omitted since compression was assumed to be undertaken by the pipeline.

Each well was assigned a pro rata expenditure for surface processing equipment based on its production contribution to the 50 MMSCFD module. The compression system, based on pipeline delivery pressure of about 2000 psig, required one stage of compression for flowing wellhead pressures between 565-2000 psig, two stages for flowing pressures of 185-565 and three stages for flowing pressures of 50-185. Actual surface processing costs may vary considerably if two-phase flow is encountered at the surface.

Discounted cash flow economic evaluations considering lease bonus, exploration cost, drilling and development expenses, gathering, treating, and compression costs, and probably time delays (as discussed below) were calculated using a commercial computer package.[45]

Earning power was calculated on an after Federal income tax basis, with a tax rate of 48 percent and no depletion allowance for all cases. A 12.5 percent royalty was assumed.

Economics were calculated in constant value 1978 dollars and assume that inflation in investment and operating expenses will not exceed inflation in gas price.

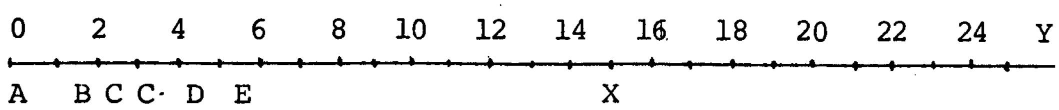
In the McElmo Dome and Sheep Mountain areas a time span of five years was selected as probably representative of the length of time between initial lease expenditures and first production and delivery of the CO_2 to the consuming EOR operator. The actual time frame may be expected to have negative impact on present value economics. The time period from initial lease bonus expenditure until CO_2 enters the pipeline in year 6 is shown in Figure 6.4. It provides for exploration and proving of reserves in years 2-4, building of the pipeline in years 3 through 4, and development drilling and installation of surface processing equipment in year 5. Production of CO_2 was assumed to commence in year 6 and was sustained until the calculated reserves were produced.

For the Jackson Dome and Northeast New Mexico areas an additional year was added for the development phase. This allowed major capital expenditures for development drilling and surface processing equipment to be budgeted over years 5 and 6. Initial CO_2 production begins in year 7.

At variable times, dependent on each well's pressure decline behavior, additional compression stages may be introduced, and are represented by X. Economics were considerably affected by the time table for compression stages in all areas.

All economics were calculated based on well spacing of 640 acres per well. Ultimate well spacing may be greater or smaller depending on actual well and reservoir performance and cost experience.

Results presented in the following sections must be considered speculative and depend on sustained production history for confirmation.



EVENT	YEARS	COMMENTS
A	1	Lease bonus expenditures and geologic-geophysical evaluation begins.
B	2-4	Exploration commences and reservoirs defined
C	3	Pipeline construction - initiated
C	4	Pipeline building continues
D	5	Development drilling; surface processing equipment and gathering system installed
E	6	First CO_2 produced into pipeline and injected by EOR operator
	6-Y	Variable years of production until reserve is produced
X	Variable	Years in which additional compression stages may be required

TYPICAL TIME FRAME FOR USE IN PRESENT VALUE ECONOMIC

EVALUATION OF CO_2 EXPLORATION PRODUCTION PROGRAM

FIGURE 5.4

5.4 SOUTHWESTERN COLORADO - McElmo Dome Area

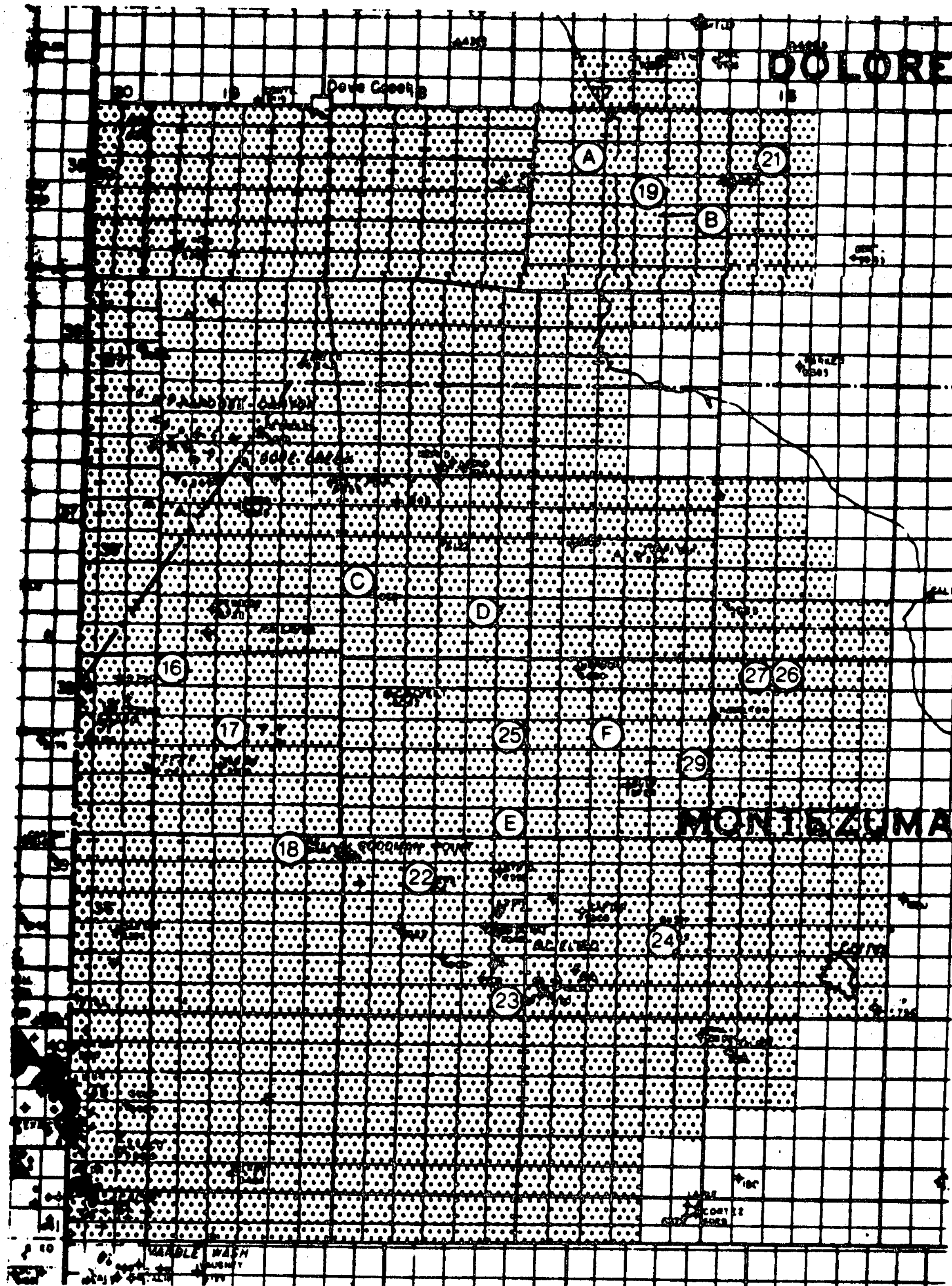
5.4.1 Geologic Introduction

Exploration for CO_2 reservoirs in southwestern Colorado has focused on the McElmo Dome Area in Montezuma County and the Doe Canyon-Dove Creek Area of Dolores County. The two CO_2 productive zones in this portion of the Paradox Basin are the Leadville (Mississippian) and the Ouray (Devonian), found at depths of 6500-9000 feet subsurface. Both formations are composed of limestones and dolomites, with the Leadville indicating greater productivity.

Analysis of produced gas from both formations indicates a composition of 96-99 percent CO_2 with 1-4 percent nitrogen.

Exploratory wells, identified on Figure 5.5[10] appear to have delineated a productive area in the southern portion of the structure. Wells identified as 18, 22, 25, C, D, E, and F reportedly were completed as potential producers. Data is unavailable regarding CO_2 productivity in well numbers 16 and 27. The remaining six wells reportedly were plugged as dry holes. Productive area at the southern end of the structure is estimated to be roughly 100 sections, or 64,000 acres.

Although at least one productive well, identified as well A on Figure 5.5 has been completed in the northern Dove Creek portion of the structure, control is insufficient for accurate reservoir delineation in the northern area. Of the remaining wells, well 19 is apparently dry, while well B reportedly encountered a faulted section in the Mississippian and is being sidetracked for recompletion. No data is available on well 21.



SOUTHWESTERN COLORADO – CO_2 SOURCES

FIGURE 5.5

5.4.2 Reservoir Analysis

Data on exploratory CO_2 wells in the McElmo area were acquired from public records in the office of the Colorado Oil and Gas Commission, Denver, Colorado. Data obtained includes well completion reports, back pressure tests, core analyses, and electric logs. Communication with various industry personnel provided insight into difficulties encountered in determining productivity during evaluation of the area.

Net pay thicknesses were determined by interpretation of core analysis reports, compensated neutron logs, formation density logs, and perforation records in completion reports.

Porosity and permeability values were estimated from logs and core analysis reports. Gas saturation was estimated to be 80 percent.

Reservoir temperatures and initial bottom hole pressures were derived from regional data available in the CO_2 Data Base.

It is anticipated that with one stage of compression bottom hole abandonment pressures of around 1200 psig can be achieved. This would result in recovery efficiencies on the order of 65-70 percent OGIP.

Assuming the reservoir has an areal extent of 100 productive sections, and from 75 to 100 feet net pay thickness with approximately 7 percent porosity, OGIP is estimated to be from 3.8 to 4.9 TSCF. Ultimate recovery (65-70% RE) might range from 2.6 to 3.4 TSCF.

5.4.3 Well Analysis

Calculated absolute open flows (CAOF) for individual wells exhibit wide variation, reportedly ranging from 17 to 115 MMSCFD. Sustained deliverability is expected to be significantly less than CAOF. Estimate of probably sustained deliverability was complicated by complex reservoir geology and evidence of substantial rock heterogeneity. Core analyses indicate two possible controls on reservoir porosity and permeability. Both fracture and matrix porosity and permeability have been reported. Fracture controlled permeability, frequently found in carbonate rocks, often results in a rapid initial decline of wellbore pressures and flow rates. Following the initial decline, pressures and flow rates tend to stabilize at low levels for long periods. Matrix controlled permeability, on the other hand, allows a more "typical" wellbore pressure decline and, for the same initial CAOF, would tend to sustain higher flowing rates and pressures than fracture controlled permeability.

Sustained well deliverabilities are anticipated in the 3-12 MMSCFD range, and flowing wellhead pressures from 600-850 psig. If flowing wellhead pressures are in this range, then only one stage of compression should be required for delivery into a pipeline at around 2000 psig.

5.4.4 Economic Analysis

Economic sensitivity analyses were conducted with a commercial computer package. Sensitivity of earning power to sustained flow rates and reserves per well were evaluated considering three rate cases and three reserve cases. Thus, the following "models" wells were created; a "worst likely well" producing 3MMSCFD, a "model well" producing 6 MMSCFD, and a "best likely well" producing

12 MMSCFD. Economics of each of the model wells was examined for reserves of 15 BSCF, 30 BSCF, and 60 BSCF.

Capital expenditures for each well "model" are summarized in Table 5.1. The completed well cost, pro rata dry hole cost, lease bonus fees, and gathering system cost were considered independent of the production rate and were held constant for all three well "models". Cost estimates for surface processing equipment were based on 50 MMSCFD modules, and investment for compression and dehydration equipment for each well was prorated based on each well's share of the entire module. As noted previously, one stage of compression was estimated to be adequate to achieve recovery efficiencies in the range of 65-70 percent.

Each producing well was burdened with a pro rata share of the total intangible cost of dry holes and lease bonus. Based on observed experience in the area, a dry hole ratio of 1 in 3 during exploration drilling was utilized. The ratio was assumed to be 1 in 4 during development drilling.

Total investment (1978 dollars) for a 3 MMSCFD well was estimated at $1.4 million, for a 6 MMSCFD well at $1.6 million, while a 12 MMSCFD well's estimated cost was $2.1 million. The substantial increase in investment as production capacity increases is proportional to increased compression horsepower required. Operating costs utilized are indicated in Table 5.2.

TABLE 5.1

INVESTMENT - McELMO DOME, COLORADO

Flow Rate	Compression B.H.P. Required	Compression* Investment In $	Dehydration Cost In $	Well Cost In $	Dry** Hole Cost (Pro Rata)	Gather-*** Ing System Cost In $	Total[I]
3 MMSCFD	387	209,290	24,000	750,000	112,500	250,000	1,345,790
6 MMSCFD	774	418,580	48,000	750,000	112,500	250,000	1,579,080
12 MMSCFD	1,548	837,150	96,000	750,000	112,500	250,000	2,045,650

*Based on $540.8/BHP.

**Pro rata dry hole cost assigned on intangible investment for 25 dry wells per 75 producing wells.

***Gathering system based on 1 mile of 4" S.S. quality pipe. Also includes $65,000/well for hydrate control.

[I] Total excludes $25,600/640 acres lease fee.

TABLE 5.2

OPERATING COSTS - McELMO DOME, COLORADO

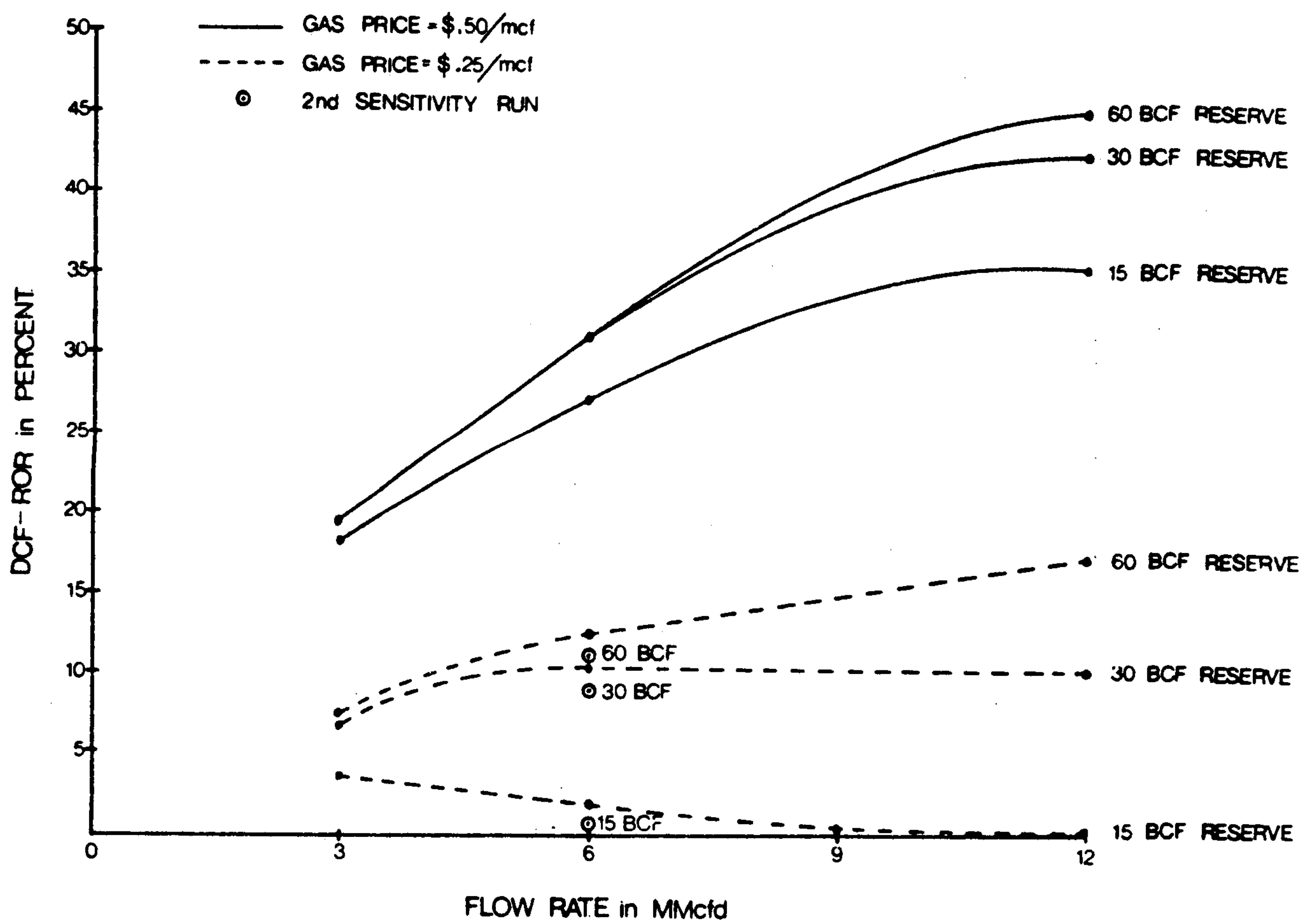
$/well/month = 300

¢/MSCF for compression from 565 to 2000 psig = 7.7

¢/MSCF for dehydration = .53

¢/MSCF for hydrate control = .4

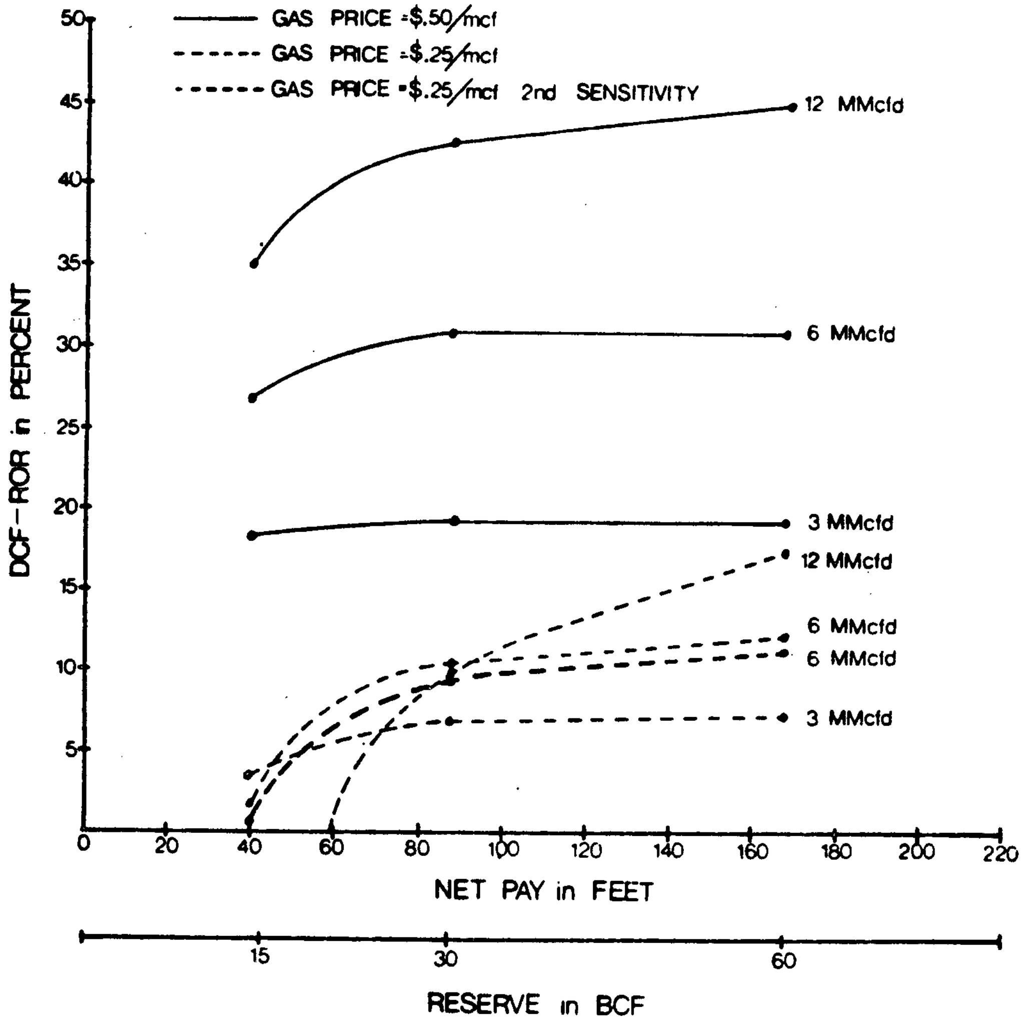
TOTAL COST $/UNIT = .0863/MSCF



AFIT DCF-ROR vs. FLOW RATE VARYING PER WELL RESERVE AND GAS PRICE

McELMO DOME, COLORADO

FIGURE 5.6



AFIT DCF-ROR vs. PER WELL RESERVE FOR VARYING FLOW RATES AND GAS PRICE

McELMO DOME, COLORADO

FIGURE 5.7

The investment schedule assumes expenditures will occur over a 5-year period prior to actual CO_2 production. Discounted cash flow rates of return (DCF-ROR) calculated assuming gas prices of $.25/MSCF and$.50/MSCF are presented in Figures 5.6 and 5.7.

In Figure 5.6 AFIT DCF-ROR is plotted versus flow rate for various gas reserves and prices. At $.25/MSCF, earning power for a well with less than 30 BSCF will be lower at a production rate of 12 MMSCFD than at a rate of 6 MMSCFD. This phenomenon is attributed to the substantially larger front end investment required for compression and dehydrating equipment for the higher production rate. The extra investment will pay out only with higher reserves or wellhead prices. Similarly, if a well has a reserves of 15 BSCF or less, a production rate of 3 MMSCFD would yield a higher earning power than one of 6 MMSCFD.

Figure 5.7 presents the same data as Figure 5.6, but plots DCF-ROR versus reserve for various flow rates and prices. For a well with 60 BSCF as a reserve, DCF-ROR increases with production rate increases, up to the maximum rate evaluated of 12 MMSCFD. However, as the per well reserve drops to 30 BSCF the DCF-ROR is relatively insensitive to rate about 6 MMSCFD.

As per well reserves fall to 15 BSCF the "optimum" rate appears to be less than 3 MMSCFD, with additional compression to produce at higher rates apparently being unwarranted.

At a gas price of $.50/MSCF, the additional revenue gained tends to overcome the front-end investment, and higher production rates show a positive correlation with increased DCF-ROR at all reserve per well figures.

These calculations indicate that because of high front-end expenditures required for compression and surface processing equipment, sustained rate and ultimate gas reserve become critical parameters in determining production economics.

The McElmo Dome area development is occurring in an area described as rich in Indian artifacts. Expensive archeological studies reportedly have been performed. Gathering system flow lines may have to be buried throughout the area. Some wells reportedly have been deviated for archeological or environmental reasons. These difficulties are time and capital consumptive for the operator.

A second set of economic analyses was conducted to study earning power sensitivity to variation in capital investment and additional time delay. To account for additional delays incurred in development of the area due to the above factors, the time required for development was increased by one year. To assess the influence of increased overhead an additional operating cost estimated at 20 percent of the well cost was budgeted over the final 2 years of development. Capital expenditures for development were allocated over years 5 and 6.

This sensitivity analysis examined a flow rate of 6 MMSCFD for the same three per well reserve cases analyzed previously, 15 BSCF, 30 BSCF and 60 BSCF. Results presented in Figure 5.6 indicate the additional expenditures and time delays will have only a slight negative effect on earning power.

All McElmo Dome area analyses assumed 640 acre drainage. If the actual drainage area is substantially less, the reserves will be proportionately smaller. These economic calculations indicate little inducement for infill drilling if drainage is less than 640 acres/well. Figure 5.6 indicates that, depending on wellhead price, the earning power will be particularly affected as per well reserves drop below about 30 BSCF, due to the large capital expenditures required for wells and surface processing equipment.

Operators have reported initial spacing is expected to be one well per 640 acres(46). However, if after sustained production, drainage is indicated to be considerably less than 640 acres, and if economics will support it, infill drilling may be required.

Recent literature suggests industry is considering a pipeline with a capacity of 300 MMSCFD from the McElmo area to West Texas.

For the "modal" reserve case of 30 BSCF per well, a well draining 640 acres, with an average sustained deliverability of 6 MMSCFD, would produce its reserve in 13 years. If the average sustained deliverability per well was 6 MMSCFD and 100 sections were productive, then a pipeline demand of 300 MMSCFD could probably be met for 25 years.

5.5 SOUTHCENTRAL COLORADO - SHEEP MOUNTAIN AREA

5.5.1 Geologic Introduction

The Sheep Mountain area, located in Huerfano Co., Colorado, is CO_2 productive from two stratigraphic units. The Dakota (Cretaceous), and a smaller reservoir, the Entrada (Jurassic), produce gas composed of 97 to 99.6 percent CO_2 with minor amounts of N_2 and methane present.

The Dakota and Entrada are porous sand bodies forming narrow elongate reservoirs. Core analyses and log data indicate rapid lateral shaling out of both reservoirs. Additionally, in Huerfano County the sands appear to be located on two large structural highs, thus forming combined stratigraphic and structural traps.

Drilling in the area has been considerably more difficult than anticipated. Surface topography is generally rough, creating siting and operational difficulties. Many wells have had to be deviated. Sections of up to 1000 feet of igneous rock have been encountered. Hole collapse problems in the Pierre Shale have necessitated the use of an oil base mud for portions of the drilling.

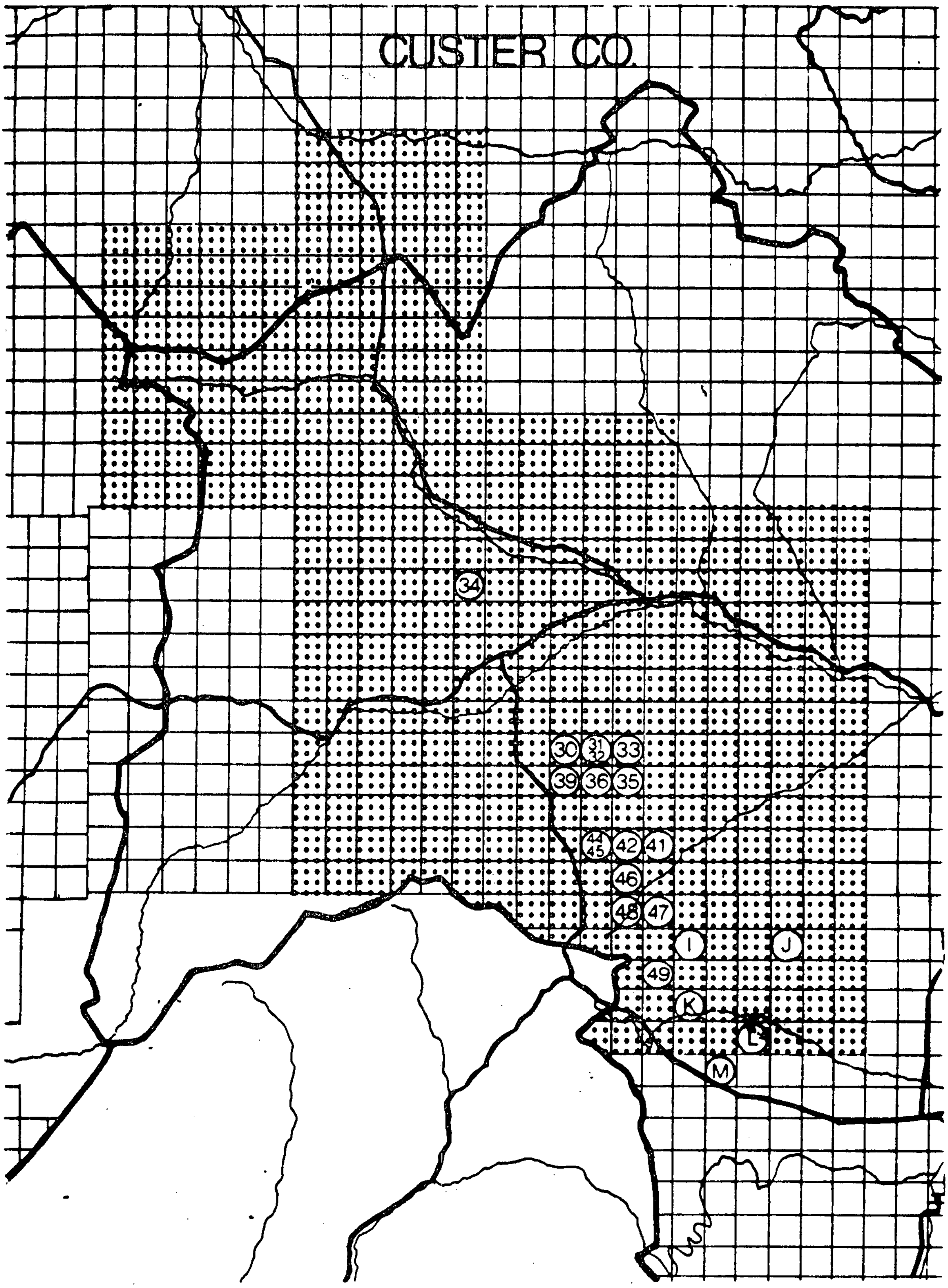
Dakota and Entrada beds exhibit dips of up to 20 degrees in the area. Evidence of substantial vertical fault displacement has been observed in several wells.

The top of the Dakota is encountered from 3,000 to 8,000 feet subsurface. Gross thickness of the Dakota formation in the Sheep Mountain area ranges from 150-350 feet.

The top of the Entrada is usually found about 300 feet stratigraphically below the Dakota. The Entrada formation exhibits a similar range of gross thickness. The Entrada apparently is a smaller reservoir areally than the Dakota. The Entrada was not analyzed in as great a detail as the Dakota, since initial indications are that Entrada production will not be comingled with Dakota production.

Figure 5.8[10] identifies well numbers 30, 31, 34, 36, 42, 44, 46, 47, 49 and L as potential Dakota producers. Well numbers 33, 35, 39, J and M have been reported as dry holes. Data is lacking for wells 32, 41, 45, 48, I and K.

The productive Dakota area on the southern half of the structure appears to have been effectively delineated by the wells shown on Figure 5.8. A possibly productive area in the northern portion has not yet been fully evaluated. One well, number 34 on Figure 5.8, exhibits good porosity and permeability in the Dakota and appears promising as a CO_2 producer. Available geologic control suggests a reservoir with a probable minimum of 25 productive sections, or 16,000 acres and a possible upper limit estimated at around 30 sections or 19,000 acres.



SOUTHCENTRAL COLORADO – CO_2 SOURCES

FIGURE 5.8

5.5.2 Reservoir Analysis

Reservoir analysis was based on data obtained from public files of the Colorado Oil and Gas Commission offices in Denver, Colorado. Well completion reports, core analyses, and daily drilling summaries provided considerable data and insight into the results obtained in the exploration program. Additional data was obtained from industry personnel.

Variation in net productive pay thickness of the Dakota was determined by analysis of core data, perforation records, and electric log interpretation.

Porosity and permeability data from available core reports were statistically analyzed to determine distribution and appropriate means. Individual mean porosity values for analyzed wells ranged from 13.3 to 19.6 percent. Permeability means ranged from .35 md to 497 md.

Bottom hole pressures and temperatures were obtained from completion reports and daily drilling reports. Gas saturation was estimated to be 80 percent.

If productive area of the Dakota reservoir ranges from 25 to 30 sections, and has 75 to 125 feet of net pay with an average porosity of 17 percent, OGIP in the Dakota is estimated to be 1.9 TSCF to 3.8 TSCF. Ultimate recovery (65% RE) might range from 1.3 TSCF to 2.5 TSCF.

5.5.3 Well Analysis

Wells reportedly tested at flow rates from 1 MMSCFD to 27 MMSCFD. Flowing tubing pressures varied from 80 psig to 880 psig. Wells yielding the lower flow rates and tubing pressures reportedly have had wellbore damage, so test results may not reflect their true potential.

Per well average sustained deliverability is estimated to range from 3 to 12 MMSCFD. The wells are expected to be produced at sufficient flowing pressures to enable delivery to a pipeline (1500-2000 psig), with only one stage of compression.

Recovery efficiencies were estimated from shut in and flowing bottom hole pressure decline curves as a function of cumulative gas produced. Average pressure drawdown between shut in pressure and flowing pressure was estimated from flow test data. Abandonment was presumed to occur when wellhead flowing pressures required more than one stage of compression for pipeline delivery. Calculations indicate recovery efficiencies on the order of 65 percent of OGIP probably can be achieved with one stage of compression.

5.5.4 Economic Analysis

Sensitivity of earning power to three sustained flow rate cases and various per well reserves was conducted with a commercial computer package. Average sustained production rates of 3 MMSCFD, 6 MMSCFD and 12 MMSCFD were assumed representative of field wells when development is completed. Economic sensitivities for these "model" wells were analyzed for per well reserves of 36 BSCF, 70 BSCF, and 137 BSCF. Drilling was assumed to occur on 640 acre spacing.

Capital expenditures for each completed well model are presented in Table 5.3. The completed well cost, pro rata dry hole cost, lease bonus fees, and gathering system cost were considered fixed fees independent of production rate.

Producing wells were assigned, on a pro rata basis, the total intangible cost of dry holes. Exploratory drilling in Sheep Mountain thus far has yielded a dry hole ratio of 1 in 3. A similar ratio was predicted for development drilling.

Surface processing equipment costs were developed on the basis of 50 MMSCFD modules, with compression and dehydration equipment investment for each well proportional to the well's contribution to the module. It was assumed that with one stage of compression recovery efficiencies on the order of 65 percent could be achieved.

Total investment for a 3 MMSCFD well "model" was placed at $1.4 million, for a 6 MMSCFD well "model" at $1.7 million, and for a 12 MMSCFD well "model" at $2.1 million. The variation in investment as rate increases is related to the increased capital expenditures required for compression and dehydration equipment capacity.

TABLE 5.3

INVESTMENT - SHEEP MOUNTAIN, COLORADO

Flow Rate	Compression B.H.P. Required	Compression Investment In $ *	Dehydration Cost In $	Well Cost In $	Dry Hole Cost (Pro-Rata In $ **	Gathering System Cost In $ ***	Total I
3 MMSCFD	387	209,290	24,000	650,000	128,700	250,000	1,261,990
6 MMSCFD	774	418,580	48,000	650,000	128,700	250,000	1,495,280
12 MMSCFD	1,548	837,150	96,000	650,000	128,700	250,000	1,961,850

*Based on $540.8/BHP.

**Prorata dry hole cost assigned on intangible investment for 15 dry wells per 30 producing wells.

***Gathering system based on 1 mile of 4" S.S. quality pipe. Also includes $65,000/well for hydrate control.

ITotal excludes 75,000/640 acres lease bonus.

TABLE 5.4

OPERATING COSTS - SHEEP MOUNTAIN, COLORADO

$/well/month = 300

¢/MSCF for compression = 7.7

¢/MSCF for dehydration = .53

¢/MSCF for hydrate control = .4

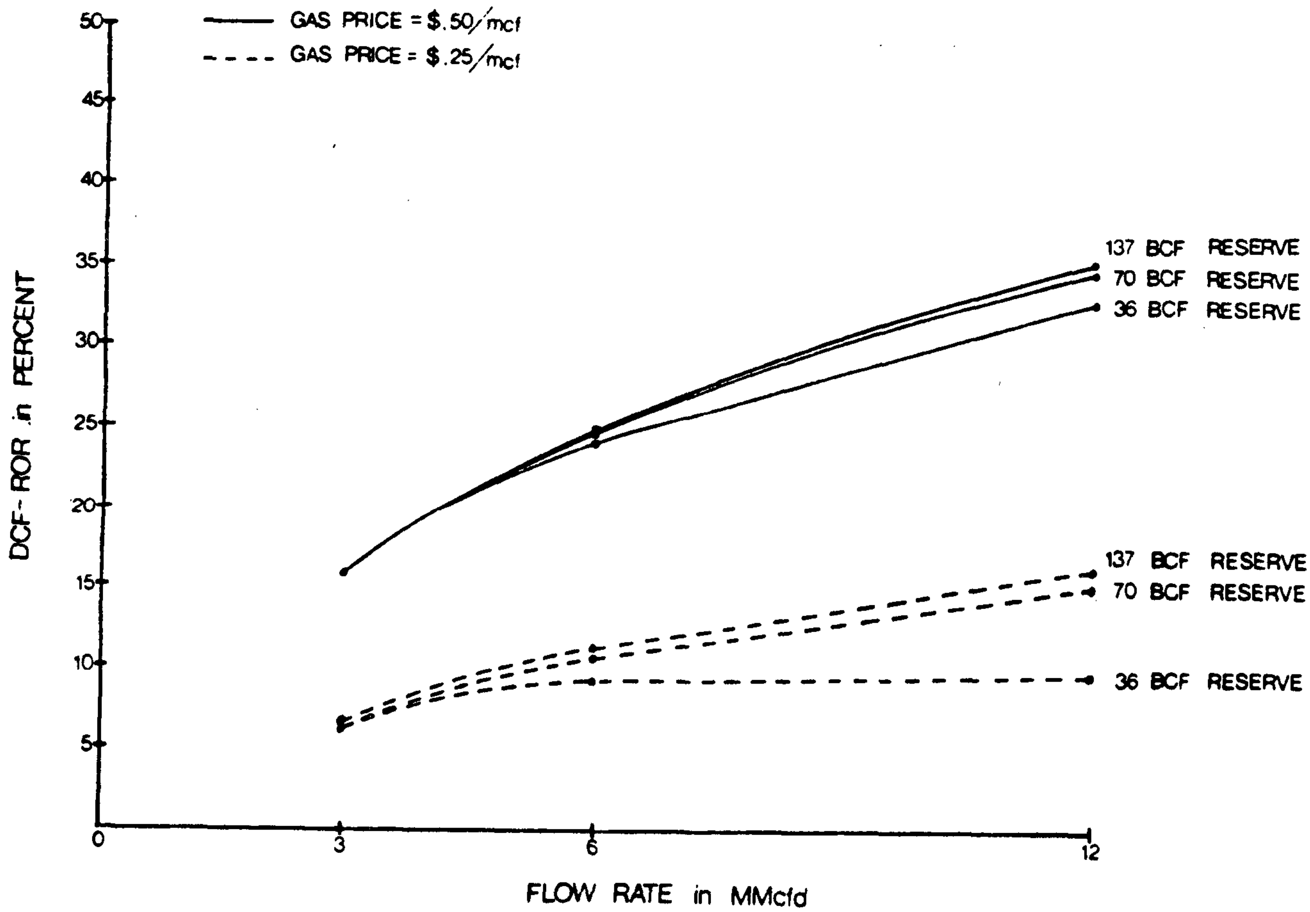
TOTAL COST $/UNIT = .0863/MSCF

Operating costs for the well and surface processing are summarized in Table 5.4.

Capital expenditures for Sheep Mountain were scheduled over five years, as indicated in Figure 5.4 with initial CO_2 production occurring in year 6.

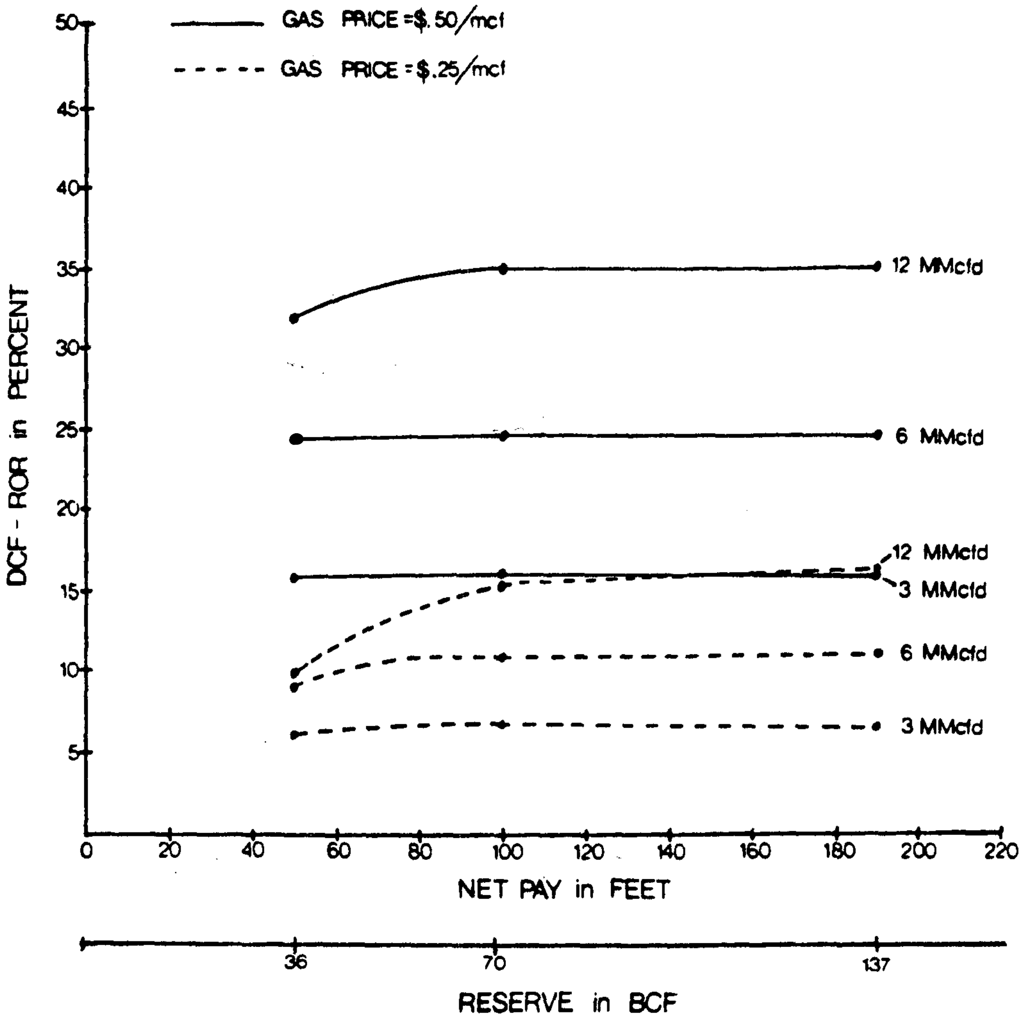
Results of discounted cash flow rate of return (AFIT DCF-ROR) calculations using the above input and gas prices of $.25/MSCF and $.50/MSCF, are presented in Figures 5.9 and 5.10.

Figure 5.9 shows AFIT DCF-ROR versus flow rate for various gas reserve and prices. When gas is priced at $.25/MSCF, for the lowest flow rate examined, 3 MMSCFD, DCF-ROR increases only slightly as per well reserve increases from 36 BSCF to 70 BSCF. DCF-ROR then increases only slightly as reserve increases from 70 BSCF to 137 BSCF. For a rate of 6 MMSCFD, DCF-ROR shows a mild upturn as reserve increases from 36 BSCF to 70 BSCF. The 12 MMSCFD rate case yields an increase in DCF-ROR each time reserve is increased. Figure 5.9 indicates for areas with thicker pay, i. e., if per well reserve on 640 acre spacing is in excess of 70 BSCF, then infill drilling, possibly on 320 acres per well, could yield greater field production capacity with little loss of earning power per well for all flow rates examined.



AFIT DCF-ROR vs. FLOW RATE FOR VARYING PER WELL RESERVE AND GAS PRICE
SHEEP MT., COLORADO

FIGURE 5.9



DCF-ROR vs. WELL RESERVE FOR VARYING FLOW RATES AND GAS PRICE
SHEEP MT, COLORADO

FIGURE 5.10

Figure 5.10 is derived from the same data as Figure 5.9, but shows AFIT DCF-ROR versus per well reserve for various flow rates and prices.

Figure 5.9, for a gas price of $.25/MSCF, shows a flow rate of 3 MMSCFD yields an insignificant change in earning power as per well reserves increase from 36 BSCF to 70 BSCF and then from 70 BSCF to 137 BSCF. Earning power shows a modest gain when reserve increases from 36 BSCF to 70 BSCF for the median rate of 6 MMSCFD. Little enhancement of earning power is seen with a reserve greater than 70 BSCF for a rate of 6 MMSCFD. A rate of 12 MMSCFD shows a significant increase in earning power when reserves are increased from 36 BSCF to 70 BSCF, but very little gain when per well reserve increases from 70 BSCF to 137 BSCF.

When gas price is increased to $.50/MSCF, the front-end investment is recovered rapidly causing flow rate to become the predominant influence on DCF-ROR. As indicated in Figures 5.9 and 5.10 little change in DCF-ROR is seen in any of the three rate cases as per well reserve increases. For the higher price case examined, $.50/MSCF, Figure 5.10 clearly shows gain in DCF-ROR as a function of an increase in flow rate rather than an increase in reserve per well.

Figures 5.9 and 5.10 appear to indicate that for areas exhibiting thicker net pay, a drilling program on a well spacing of 320 acres per well would cause little reduction in earning power, while significantly increasing overall field production capacity. Additionally, although test data indicate permeability is well above average in most wells, if drainage areas were found to be significantly less than 640 acres per well, a 65 percent recovery efficiency could be overly optimistic. A lower RE per 640 acres could necessitate infill drilling.

5.6 NORTHEAST NEW MEXICO AREA

5.6.1 Geologic Introduction

Several areas in Northeastern and North Central New Mexico, including locations near Bueyeros, Wagon Mound and Turkey Mound, have been recent exploration targets for commercial CO_2 reserves.

The most rewarding drilling activity to date has occurred on a large structural nose on the eastern flank of the Sierra Grande Uplift. Covering portions of Union, Harding, and Quay Counties, the structure encompasses the Bueyeros CO_2 field, which has produced CO_2 since the early 1950's for the manufacture of dry ice.

Recent exploratory drilling in the area indicates high grade CO_2 of 99.1 to 100 percent purity, with a maximum of .3 percent N_2 and .2 percent hydrocarbons as contaminants.

Several stratigraphic units in the area have exhibited shows of CO_2. The Glorietta (Permian), Tubb (Permian), and a granite wash member appear to be CO_2 productive.

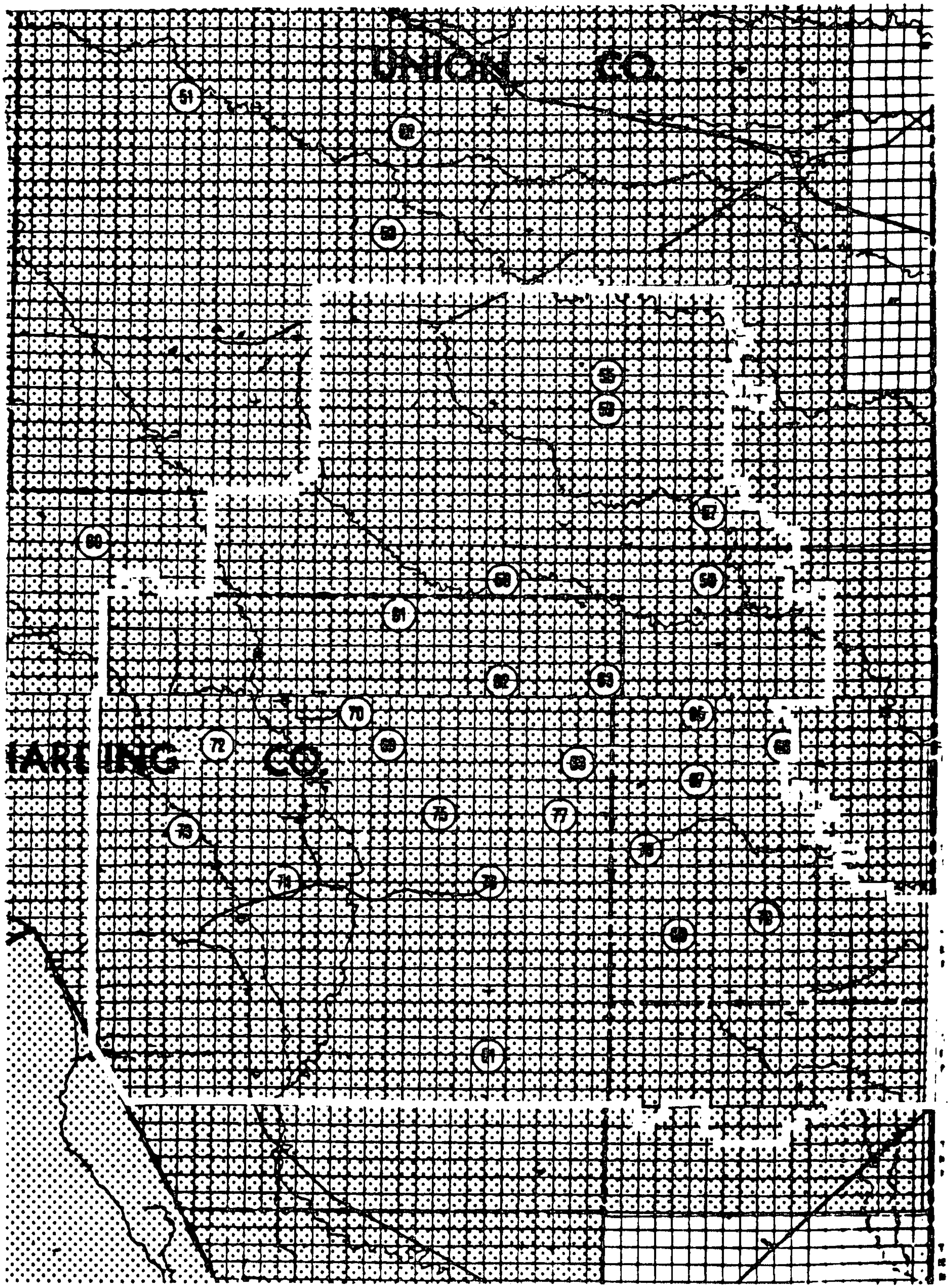
The top of the Tubb, the primary producer, may be clearly identified directly below the Cimarron Anhydrite on electric logs and is found at depths from 1900 feet to 2700 feet subsurface.

Tubb deposition apparently did not occur to the Northwest of the structure, resulting in a depositional pinch-out on the Sierra Grande Uplift. Tubb thickness increases rapidly downdip, to a maximum of 500 feet in the south-eastern portion of the structure.

The Tubb pinch-out limits productivity in the northwestern portion of the structure. An apparent gas water contact limits productivity downdip in the southeastern and southwestern portions of the structure.

Preliminary application has been made to the State of New Mexico by a major operator in the area for a unit containing nearly 1.2 million acres, to be called the Bravo Dome Carbon Dioxide Unit. The proposed unit boundaries are indicated on Figure 5.11[10].

Twenty-four exploratory wells, as identified in Figure 5.11[10] apparently have delineated a large CO_2 productive area. Wells 55, 58, 59, 61-65, 67-69, and 71-80 were reportedly productive. Numbers 53, 57, 60, and 66 were reported as dry holes. Well number 81 was reported as a mechanical failure. No data are available on wells 56 and 70. Although well control for such a large area is poor, averaging 1 well per 40,000 acres, there appears to be a potentially productive area of approximately 880,000 acres or roughly 38 townships. Productive area is defined as that acreage containing a minimum of 20 net feet of pay.



NORTHEAST NEW MEXICO – CO_2 SOURCES

FIGURE 5.11

5.6.2 Reservoir Analyses

Data utilized in the analyses herein were obtained from public files of the New Mexico Oil Conservation Commission in Santa Fe. Data were obtained from completion reports, scout cards, production records and electric logs. The absence of meaningful CO_2 production records for the Bueyeros field and poor well control for such a large area compounded difficulties in interpreting complex reservoir characteristics.

Distribution of net pay thickness was estimated from perforation records in completion reports and available logs of the section. Other investigators have reported considerable difficulty in correlation of specific porosity zones or net pay in geologic cross sections.

Initial bottom hole pressures and porosity values were determined from working maps of the area obtained in State offices.

Reservoir temperatures were estimated from regional data available in the CO_2 Data Base. Gas saturation was estimated to be 80 percent.

Porosity values averaged 20 percent over the area. Porosity values are lower in the updip section, and generally increase in the thicker downdip section. Permeability, while following the same general trend of porosity, is generally low, contributing to low deliverability and flowing tubing pressures.

High purity CO_2 gas, containing 98-99.9% CO_2, has been produced from Bueyeros field since the early 1950's for the manufacture of dry ice. Flow rates from several of these early wells have reportedly exceeded 1 MMSCFD, at flowing tubing pressures of 300-330 psig.

Although production records are inadequate, some sources familiar with the area have observed that little or no reservoir pressure decline has occurred in the wells produced in the Bueyeros area in over 20 years.

Bottom hole pressures in the Bravo Dome area do not exhibit normal correlation with subsea depth. Bottom hole pressures observed in recent exploratory wells ranged from 314 psig, at 2821 feet subsea, in well 72 on Figure 5.11, to 990 psig, at 2606 feet subsea, in well 61 on Figure 5.11. Corresponding subsea gradients range from .137 psi/ft. to .396 psi/ft. Thus, both bottom hole pressures and subsea gradients are generally higher in the southwest portion of the area and decrease in a northeasterly direction.

Several theories have been advanced for the indicated bottom hole pressure variations, including:

1. Possibility of a CO_2 source on the southwestern side of reservoir actively charging the reservoir
2. Possible leaks into Palo Duro or Dalhart basin on the northeastern side of the structure

The variation in observed bottom hole pressure complicated analysis of probable well and reservoir behavior.

Calculated recovery efficiencies in the area should be considered uncertain at this time. Based on bottom hole abandonment pressures of 225 psig, recovery efficiencies on the order of 65% can be calculated. Such bottom hole abandonment pressures would require two stages of compression. If economic justification could be found for three stages of compression, then recovery efficiencies might reach 85%.

Assuming the productive area covers 880,000 acres, or roughly 38 townships, with an average of 40 to 60 feet net pay, and approximately 20 percent porosity, OGIP is estimated to range from 10.9 TSCF to 16.3 TSCF. The ultimate recovery (65% RE) would then range from 7.0 TSCF to 10.6 TSCF.

5.6.3 Well Analysis

Exploratory wells identified as potential producing wells in Figure 5.11, tested flow rates from .25 MMSCFD to 1.9 MMSCFD. Flowing tubing pressures were generally low, ranging from 30 psig to 250 psig. Most wells exhibiting initial flow rates less that .5 MMSCFD were reportedly acidized to improve performance.

Anticipated sustained per well deliverabilities range from .3 MMSCFD to 1.2 MMSCFD, with flowing tubing pressures of at least 140 psig. With sustained flowing tubing pressures in excess of 140 psig, two stages of compression should be sufficient for delivery to the pipeline.

5.6.4 Economic Analysis

Economic sensitivity analyses were conducted with a commercial computer package. Economics were calculated for three sets of model wells and assumed producing rates of .3 MMSCFD, .6 MMSCFD, and 1.2 MMSCFD. Economics for each set of model wells were examined for reserves of 3.0 BSCF, 6.5 BSCF, and 14.3 BSCF. The range of per well reserve estimates is based on a recovery efficiency of 65% from a 640 acre drainage area.

Estimated capital expenditures, representing lease bonus costs, completed well cost, pro rata dry hole cost, gathering system and dehydrator investment, are presented in Table 5.5.

A dry hole ratio of 1 in 4 during exploratory drilling was based on historical data derived from the exploratory program while a dry hole ratio of 1 in 5 was assumed for development drilling. Each producing well was burdened with a pro rata share of the total intangible cost of dry holes and lease bonus fees.

As it is expected the pipeline will be responsible for compression of the gas, economic analyses were based only on the production, drying, and gathering of CO_2 and did not include compression expenses. (It is uncertain at this time whether well performance will warrant the installation of three stages of compression. Optimization of compression sizing and scheduling this low pressure gas resource is beyond the scope of this report.)

Table 5.6 summarizes well operating costs, gas processing costs, and

TABLE 5.5

INVESTMENT - NORTHEAST NEW MEXICO

Flow Rate	Dehydration Cost In $	Well Cost In $	Dry Hole* Cost In $ (Prorata)	Gathering** System Cost In $	Total[I]
0.3 MMSCFD	2,400	150,000	18,000	190,000	360,400
0.6 MMSCFD	4,800	150,000	18,000	190,000	362,800
1.2 MMSCFD	9,600	150,000	18,000	190,000	367,600

*Prorata dry hole cost assigned on intangible investment for 25 dry wells per 100 producing.

**Gathering system based on 1 mile of 2" S.S. quality pipe. Also includes $32,500/well for hydrate control.

[I]Total excludes $22,000/640 acres lease fee.

TABLE 5.6

OPERATING COSTS - NORTHEAST NEW MEXICO

\$/well/month = 300

¢/MSCF for dehydration = 0.53

¢/MSCF for hydrate control = 0.4

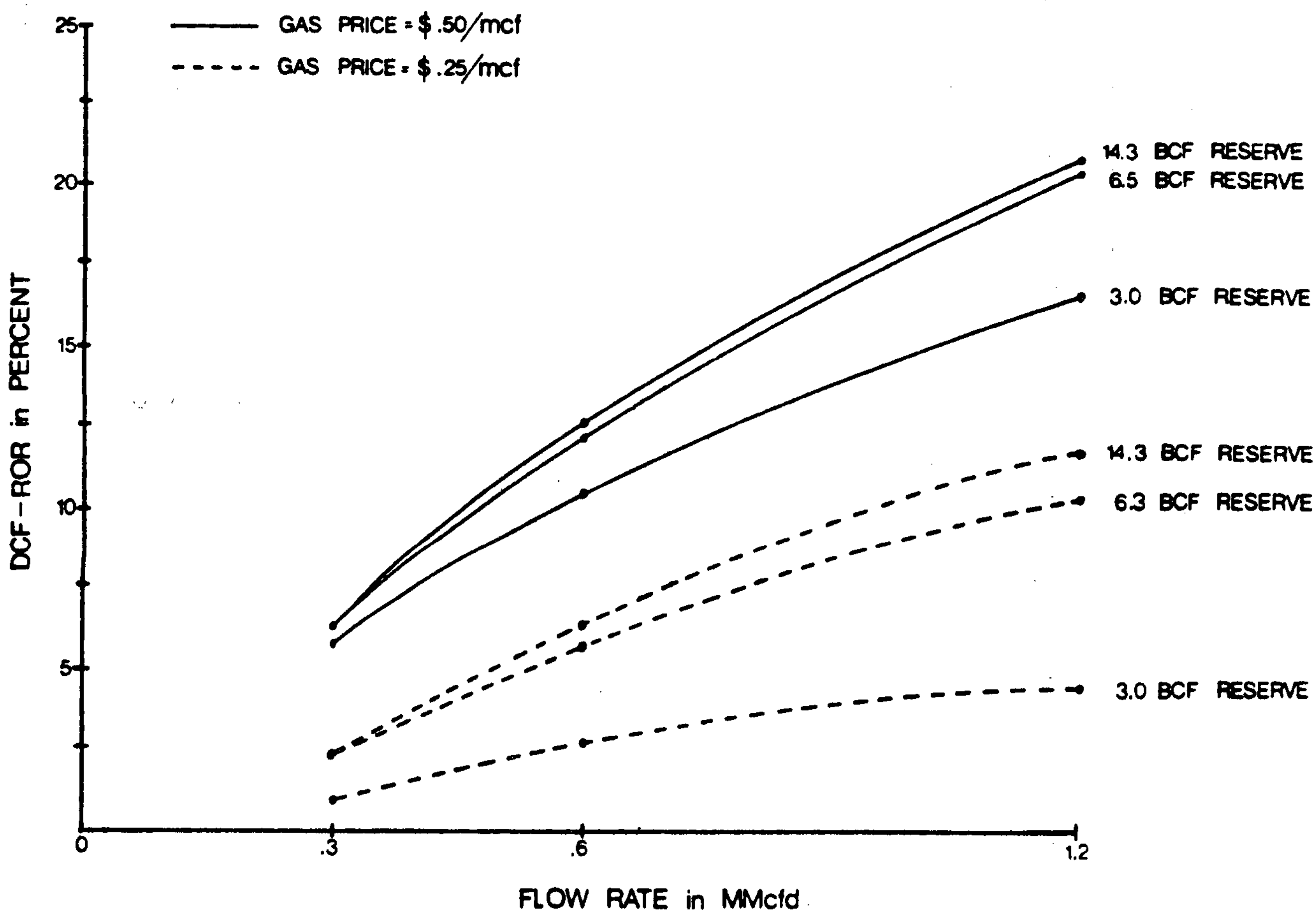
TOTAL COST \$/UNIT = .0093/MSCF

hydrate inhibiting costs.

Investment was assumed to occur over a six year period prior to actual CO_2 production. Development capital expenditures were scheduled over the two years prior to CO_2 production. Figures 5.12 and 5.13 present results of economic analyses for gas prices of \$.25/MSCF and \$.50/MSCF.

Shown on Figure 5.12 is AFIT DCF-ROR versus flow rate for various per-well gas reserve and prices. For a flow rate of .3 MMSCFD, with a gas price of \$.25/MSCF and the lowest per well reserve analyzed, 3.0 BSCF, the earning power is one percent. Flow rates of .6 or 1.2 MMSCFD will both have earning powers of less than five percent for a reserve of 3.0 BSCF. As reserve increases to 6.5 BSCF per well, earning power increases rapidly for all three flow rates. When reserve per well rises to 14.3 BSCF, earning power, for a flow rate of .3 MMSCFD, remains constant and shows no increase over the 6.5 BSCF reserve case. However, earning powers for the higher production rates of .6 MMSCFD and 1.2 MMSCFD continue to show moderate increases as per well reserves increase from 6.5 BSCF to 14.3 BSCF.

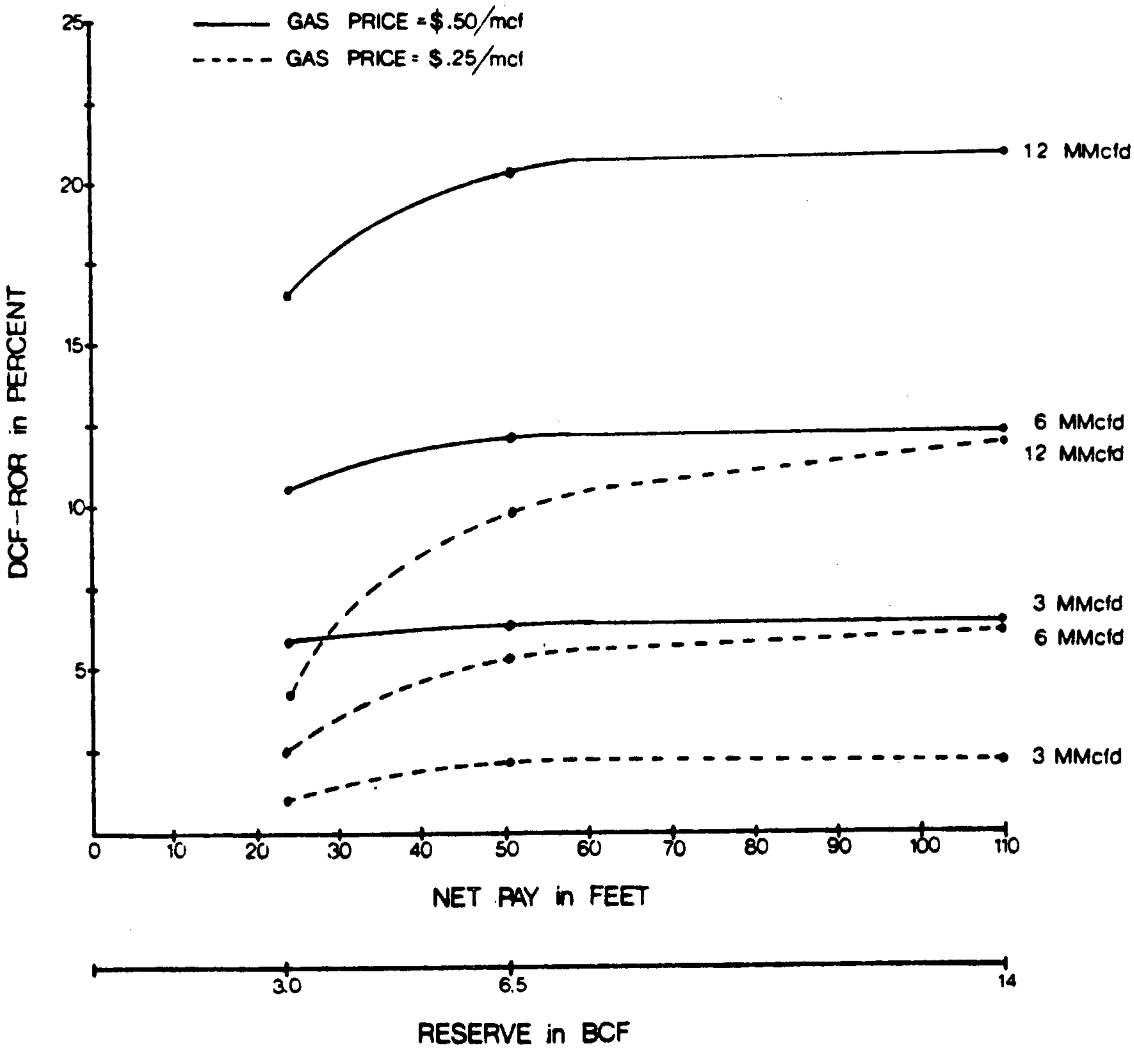
Plotted on Figure 5.13 is AFIT DCF-ROR versus per well reserve for various flow rates and prices. Figure 5.13 indicates earning power is more sensitive to flow rate than reserve when reserve is greater than 6.5 BSCF. For a gas price of \$.25/MSCF as the flow rate increases from .3 MMSCFD to 1.2 MMSCFD, the earning power increases from 1 to almost 5 percent for a reserve of 3.0 BSCF per well. For reserve cases of 6.5 BSCF and 14.3 BSCF earning power shows substantial increase as flow rate increases from .3 to .6 MMSCFD.



AFIT DCF-ROR vs. FLOW RATE FOR VARYING PER WELL RESERVE AND GAS PRICE

N.E. NEW MEXICO

FIGURE 5.12



AFIT DCF-ROR vs. PER WELL RESERVE FOR VARYING FLOW RATES AND GAS PRICE

N.E. NEW MEXICO

FIGURE 5.13

As flow rate increases from .6 MMSCFD to 1.2 MMSCFD, earning power shows comparable increases.

For gas prices of \$.50/MSCF the earning power increases for each case examined. For flow rates of .3 MMSCFD and .6 MMSCFD with gas priced at \$.50/MSCF, Figures 5.12 and 5.13 show insignificant gain in earning power as per well reserve increases from 3.0 BSCF to 14 BSCF. For a higher flow rate of 1.2 MMSCFD, earning power increases up to a reserve of 6.5 BSCF, then stabilizes with increasing reserve.

Figures 5.12 and 5.13 should be utilized cautiously. Considering the variability of initial bottom hole pressures, and the uncertainty in reservoir characteristics and reservoir continuity, economic interpretation prior to sustained production performance must be considered rather speculative. Cost of CO_2 delivered to the user must reflect a wellhead price that will earn an acceptable profit to the lease operator and must also reflect compression and transportation costs.

The economics reflected by Figures 5.12 and 5.13 assume 640 acre spacing and drainage areas. Depending on well performance and pricing, and CO_2 demand, closer spacing might be warranted. At this stage of development, however, further calculations are not justified.

Peak production rates discussed thus far from this accumulation have been in the 600 MMSCFD range. If the average per well sustained delivery were

.6 MMSCFD, this would require a minimum of 1000 wells and would result in the production of 5 TSCF of CO_2 over a period of 25 years. When this total production is compared to the possible 7.0 to 10.6 TSCF original reserve, it seems apparent there is adequate gas in place to sustain this demand.

Well control in northeastern New Mexico is sparse. Effective drainage radii are not known. Pressure data obtained from the exploratory drilling program does not appear to conform to geologic and engineering norms. Interpretation of future productivity, reservoir behavior, and economics, thus, is highly speculative. Improved reservoir and economic evaluations can only be made with increased well density or control and sustained production data.

5.7 CENTRAL MISSISSIPPI - JACKSON DOME AREA

5.7.1 Geologic Introduction

Jurassic sediments in portions of Rankin, Madison, and Scott Counties, Mississippi, have tested sour gas over the past 20 years of exploration. A corridor of CO_2 and H_2S rich gas is located on the east-northeast flank of the Jackson Uplift and extends in an arc from north central Madison County through central Rankin County.

Two stratigraphic units, the Smackover (Jurassic) and the Norphlet (Jurassic) have indicated strong CO_2 potentials during exploratory drilling and testing.

The sour gas productive trend also encompases several hydrocarbon productive fields including Loring, Pelahatchie, Piney Woods and Thomasville.

Wells in the area have tested carbon dioxide concentrations ranging from 65 percent to 99.6 percent from the Jurassic section. The gas is associated with up to 10 percent H_2S and varied quantities of methane.

The carbon dioxide resource area of prime interest in this investigation is limited to areas penetrated by wells 92, 93, 94, 95, 96, 98, G, and H on Figure 5.14. These wells appear to be located on four separate closures, some of which may have common accumulations. Tests in these wells indicate gas with a minimum of 98 percent CO_2 and a maximum of 1.5 percent H_2S. Several tests have indicated H_2S concentrations less than 1 grain per 100 cubic feet.

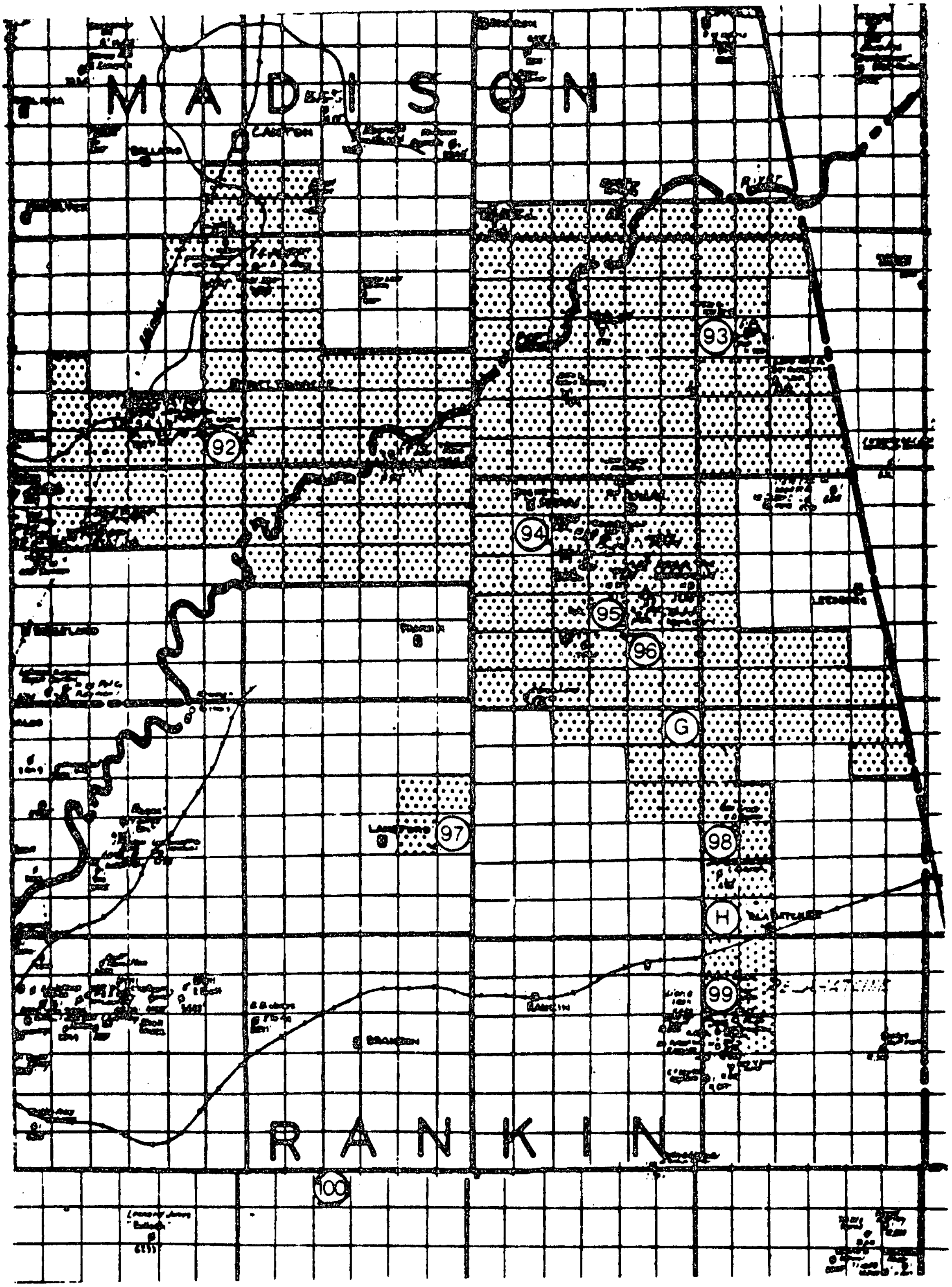
Available data indicate CO_2 concentrations in Jurassic gas tested by wells 97 and 99 is considerably less than 98 percent while an additional CO_2 resource of considerable magnitude may exist with higher H_2S concentrations, it will not be examined in this report.

Estimate of gross thickness in the Smackover and Norphlet is complicated by differences of opinion among various investigators regarding Jurassic stratigraphic interpretation. Differentiation of the basal Smackover sands and the upper Norphlet sand is reportedly difficult.

The top of the Smackover in the Jackson Dome area is found between 14,000 to 16,000 feet subsurface. The Norphlet is located directly below the lower Smackover. The Smackover in this area is primarily carbonate, composed of brown to grey limestones and dolomites with interbedded dolomitic sands. A porous dolomitic basal sand member is usually present.

Gross thickness of the Smackover in the study area is estimated to be from 1,000 to 2,000 feet. Smackover porosity and permeability have been described as highly varied in the carbonate section, ranging from porous oolitic, to intergranular, to vuggy and fractured[48].

The Norphlet is described as a sequence of primarily fine grained sands. Gross thickness for the Norphlet is a minimum of 300 feet.



CENTRAL MISSISSIPPI — CO_2 SOURCE

FIGURE 5.14

5.7.2 Reservoir Analysis

Data for reservoir analysis was obtained from public records on file in the Mississippi Oil and Gas Board Office in Jackson, Mississippi, and through a GURC consultant. All available well completion reports, electric logs and scout cards were utilized to interpret reservoir characteristics. Core analyses, although of limited availability, were examined whenever possible.

Net pay thickness was estimated by analysis of perforated intervals reported in well completion reports and scout cards, and by correlation with subsurface logs.

Porosities and permeabilities were estimated from available core analysis data and from literature on the Jurassic section in the area. Gas saturation was estimated to be 80 percent.

Reservoir temperatures and initial bottom hole flowing and shut in pressures were determined from well completion reports and scout cards. Pressure gradients exceed geo-pressured in some portions of the reservoir. Pressure gradients ranged from .485 psi/ft. at 12,130 feet subsurface to .713 psi/ft. at 16,801 feet subsurface, yielding the maximum subsurface pressure gradient of .713 psi/ft.

Seven of the wells identified on Figure 5.14, wells 92, 93, 94, 95, 96, 98, and H, reportedly tested gas with CO_2 in excess of 98 percent. These wells provide a bare minimum of well control. All seven of the indicated wells, drilled between 1951 and 1978, were drilled on structural highs, probably related to deep-seated salt features. Of these seven wells, all but number 96 were drilled for hydrocarbon exploration.

Wells 94, 95, and 96 appear to be located on a single, fairly large elongate structured feature. The objective of well number 96, drilled and completed as a potential CO_2 producer by a major operator, was reportedly to test the Jurassic CO_2 potential on the southern flank of the feature. Well number G is being drilled as a confirmation CO_2 well to number 96 and, when completed, will provide additional data as to the productivity of this structure to the south.

Well 92 appears to be located on an uplift of limited areal extent. The operator of well 92 has applied to the Oil and Gas Board of the State of Mississippi for creation of a 1,760 acre gas unit. Well spacing of 880 acres per well was requested in the application.

Similarly, well 93 is located on a small structure in which the extent of CO_2 productivity is unknown.

Well 98 was apparently completed on the crest of another separate structure. Well H is reportedly on the southern edge of this structure and was CO_2 productive in several intervals of the Jurassic section.

Available data indicate CO_2 concentrations in Jurassic gas tested by wells 97 and 99 is considerably less than 98 percent.

Estimate of productive area is hampered by insufficient well control on the structural features of interest. Most control wells were drilled as hydrocarbon exploratory wells on seismic and gravity data. Lateral extent of productivity on the four structures discussed above is unknown at this time.

Preliminary estimates, based on limited well control, suggest the four features contain a minimum combined productive area of 27 sections, or 17,280 acres. Better estimates of productive area will require additional subsurface data.

Assuming combined productive area of these four structural features is a minimum of 27 sections, and total net pay of the Jurassic section is 120 feet to 250 feet, with 13 percent average porosity, minimum original gas in place might range from 3.4 TSCF to 7.0 TSCF. Minimum ultimate recovery (60% RE) would then range from 2.0 TSCF to 4.2 TSCF.

5.7.3 Well Analysis

Wells designated as control wells on Figure 5.13 in Section 5.7.2 reportedly tested varying ranges of flow rates and tubing pressures from the Smackover and Norphlet as exhibited below.

	Flow Rate	Flowing Tubing Pressure
Smackover	2 MMSCFD	3,375 psig
	12 MMSCFD	3,450 psig
Norphlet	10.7 MMSCFD	3,950 psig
	20.5 MMSCFD	3,565 psig

Several wells exhibited productivity in both the Norphlet and Smackover.

Sustained per well deliverabilities from both the Smackover and Norphlet, individually are expected to range from 4 MMSCFD to 16 MMSCFD. Initial flowing tubing pressures are expected to range from 2,000 psig to 4,500 psig. It seems unlikely that dual completions will be made in these two formations.

As flowing tubing pressures decline below 2,000 psig, a single stage of compression probably will be required to meet pipeline delivery pressures and to achieve recovery efficiencies on the order of 60 percent.

Abandonment pressures were calculated from shut-in and flowing bottom hole pressure decline curves as a function of cumulative gas produced. Abandonment was assumed to occur when surface flowing pressures declined below 560 psig and more than one stage of compression was required to meet pipeline delivery pressures. Estimates indicate bottom hole abandonment pressures of around 2,300 psig should yield recovery efficiencies on the order of 60 percent of gas in place.

5.7.4 Economic Analysis

Economics were calculated for three sets of model wells, which assumed average sustained producing rates of 4 MMSCFD, 8 MMSCFD, and 16 MMSCFD. Economics for each set of the model wells were evaluated for per well reserves of 17 BSCF, 51 BSCF, and 152 BSCF. This range of per well reserve estimates is based on recovery efficiencies of 60 percent from a 640 acre drainage area and net pays of 25, 75, and 225 feet, respectively.

Capital expenditures for each well model are summarized in Table 5.7. The completed well cost, pro rata dry hole cost, lease bonus fees, and gathering system cost were considered fixed costs, independent of the production rate.

Based on experience to date in the area, a dry hole ratio of one in five was assumed for both exploration and development drilling.

Cost estimates for surface processing equipment were based on 50 MMSCFD modules. Investment for compression and dehydration facilities was prorated on the basis of each well's contribution to the entire module. Flowing tubing pressures will, initially, enable production without compression. At some point as flowing tubing pressures decline, depending on the production rate and reserve, one stage of compression will be required to meet a pipeline delivery pressure of 2,000 psig.

TABLE 5.7

INVESTMENT - JACKSON DOME AREA, MISSISSIPPI

FLOW RATE	COMPRESSION B.H.P. REQUIRED	COMPRESSION INVESTMENT IN $ *	DEHYDRATION COST IN $	WELL COST IN $	DRY HOLE COST (PRO-RATA) **	GATHER-ING SYSTEM COST IN $ ***	TOTAL $ I
4 MMSCFD	516	279,050	32,000	2,500,000	360,000	250,000	3,171,050
8 MMSCFD	1,032	558,090	64,000	2,500,000	360,000	250,000	3,482,090
16 MMSCFD	2,064	1,116,170	128,000	2,500,000	360,000	250,000	4,104,170

* Based on $540.8/BHP.

** Pro-rata dry hole cost assigned on intangible investment for 20 dry wells per 80 producing wells.

*** Gathering system based on 1 mile of 4" S.S. quality pipe. Also includes $65,000/well for hydrate control.

I Total excludes $64,000/640 acres lease fee.

Well operating costs and surface processing operating costs, including hydrate inhibition, drying and compressing the gas, are listed in Table 5.8. It is assumed that costs of purifying the CO_2, i.e., removal of H_2S, will be borne by the pipeline.

The investment schedule assumed capital expenditures will occur over a 6-year period prior to actual CO_2 production. Investment for development drilling and surface processing equipment was scheduled over the final two years (years 5 and 6) prior to CO_2 production.

Results of discounted cash flow rate of return (DCF-ROR) calculations using the above reservoir input and gas prices of $.25/MSCF and $.50/MSCF are presented in Figures 5.15 and 5.17.

Figure 5.16 shows AFIT DCF-ROR versus flow rate for various per well reserves and gas prices. When gas is priced at $.25/MSCF, flow rates of 4 MMSCFD, 8 MMSCFD, or 16 MMSCFD, yield a zero rate of return for a reserve of 17 BSCF. As per well reserve increases to 51 BSCF yields substantial improvement in the rate of return, generating a DCF-ROR of 6.6 percent for a flow rate of 4 MMSCFD and 20 percent for a flow rate of 16 MMSCFD.

Figure 5.16 plots AFIT DCF-ROR versus per well gas reserve for varying flow rates and gas prices. As previously indicated, for a gas price of $.25/MSCF, the rate of return for evaluated flow rates is zero when per well gas reserve is 17 BSCF. An increase in reserve per well to 57 BSCF generates small rates of return at all three flow rates.

TABLE 5.8

OPERATING COSTS - JACKSON DOME AREA, MISSISSIPPI

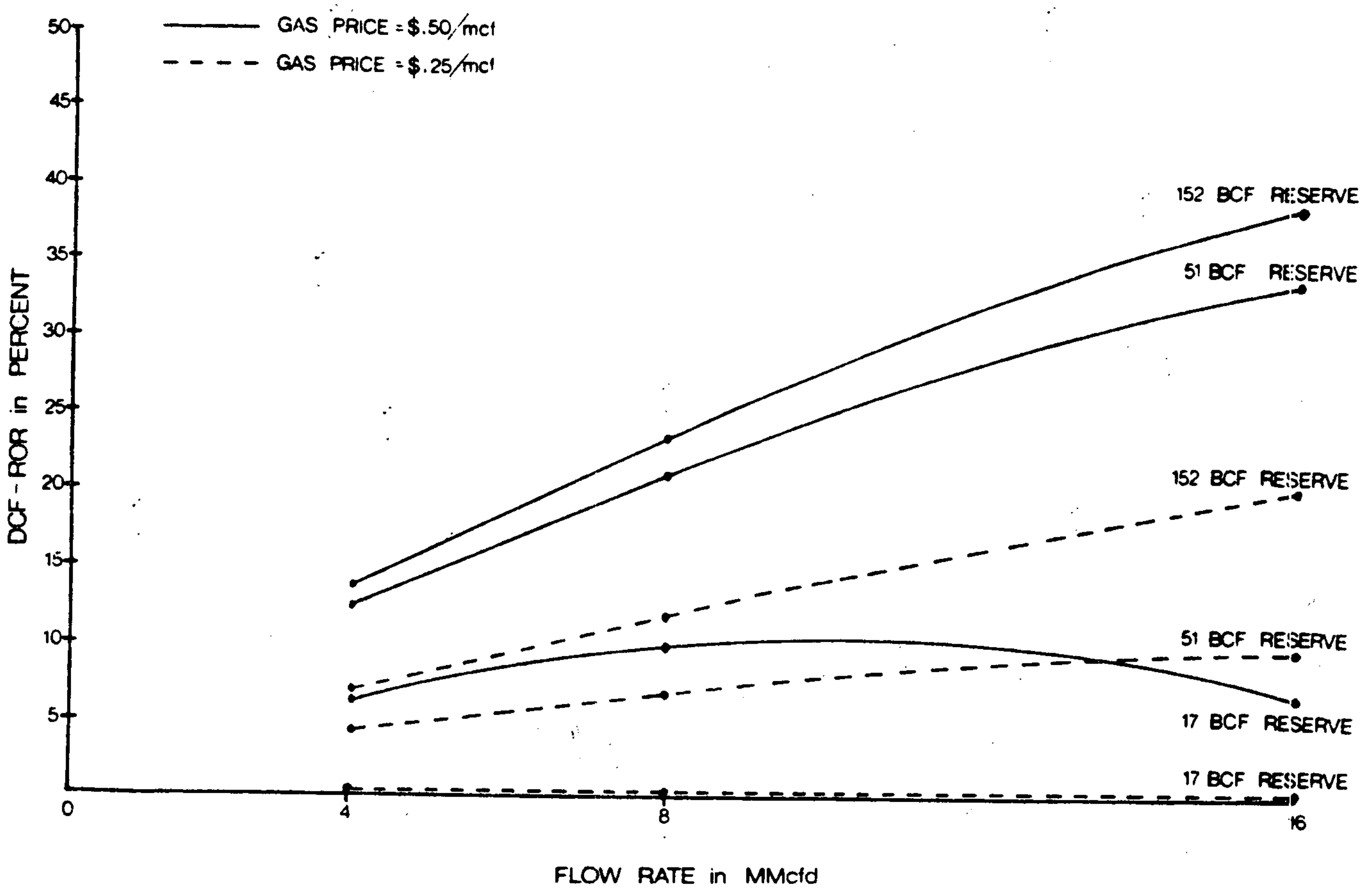
$/well/month = 300

¢/MSCF for compression from 565 psig - 2000 psig when required - 7.7

¢/MSCF for dehydration - .53

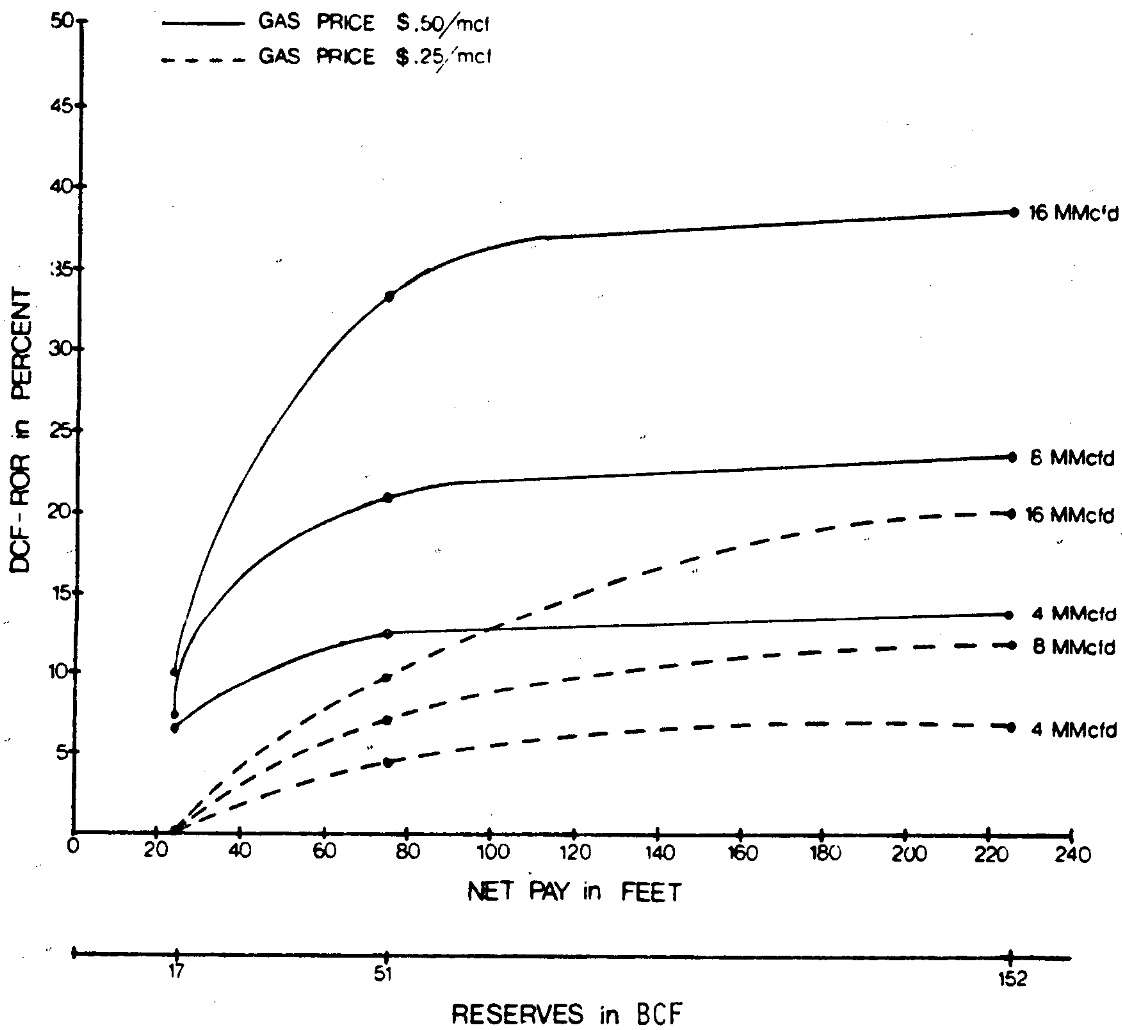
¢/MSCF for hydrate control - .4

TOTAL COST$/UNIT - .0863/MSCF



DCF-ROR vs. FLOW RATE FOR VARYING PER WELL RESERVES AND GAS PRICE
JACKSON DOME AREA, MISSISSIPPI

FIGURE 5.15



AFIT DCF-ROR vs. PER WELL RESERVE FOR VARYING FLOW RATES AND GAS PRICE
JACKSON DOME AREA, MISSISSIPPI

FIGURE 5.16

When a per well reserve of 152 BSCF was evaluated for a gas price of $.25/MSCF, flow rates of 8 MMSCFD and 16 MMSCFD yielded rates of return of 12 percent and 20 percent respectively. A flow rate of 4 MMSCFD yielded only 6.5 percent, even with the large reserve.

Referring again to Figure 5.15, when gas price is increased to $.50/MSCF, the lowest reserve case analyzed, 17 BSCF, will yield an increasing DCF-ROR as rate increases from 4 MMSCFD to 8 MMSCFD. However, as flow rate increases beyond 8 MMSCFD, DCF-ROR begins to decline, indicating an "optimum" producing rate around 8 MMSCFD. This "optimum" is due to the increased surface processing equipment investment required with increased production rates. Thus at $.50/MSCF a reserve of 17 BSCF is not sufficient to pay out the large incremental capital expenditures required for increased surface processing capacity over a short productive period. When per well reserve is increased to 51 BSCF, a positive correlation between increased flow rate and increases in DCF-ROR is seen in all cases. This trend continues as reserve increases to 152 BSCF per well.

When gas if priced at $.50/MSCF, Figure 5.16 depicts, for a per well reserve of 51 BSCF, a sharp increase in DCF-ROR for all flow rates examined. When per well reserve increases from 51 BSCF to 152 BSCF, the curve flattens with increased reserve, and increases in flow rate become the predominant factor in generating increased DCF-ROR.

To date, exploration for CO_2 in Mississippi has been limited. Data utilized herein was generated by wells intended for hydrocarbon exploration rather than for CO_2 exploration. Increased well density and resulting subsurface data following further CO_2 exploration will be necessary to make reasonable forecasts and to evaluate production methodologies and economics.

These economics could be affected adversely if removal of the small amounts of H_2S requires significant reduction in wellhead flowing pressures, necessitating multi-stage compression for pipeline delivery. Such an investigation is beyond the scope of our study.

COST OF GAS PROCESSING

PART 6

6.1 BASIS OF WORK

The maps of CO_2 sources presented in Part 3 show some sources at a sufficiently high CO_2 purity such that the only processing required for use by EOR techniques is compressing and drying of the gas to pipeline quality. Sources of these high purity sources are the stack gases from ammonia (NH_3) and hydrogen (H_2) plants and some naturally occurring deposits. If the gas is naturally occurring and only compression and dehydration is necessary to obtain pipeline quality gas, then these associated costs are included in the production costs of these gases (see Part 4). If the high purity is a by-product from either a NH_3 or a H_2 plant, then the compression and dehydration are considered additional costs and are presented in Section 6.5.

However, most sources shown in Part 3 require additional processing to remove their impurities and concentrate the CO_2 for EOR use. The above-ground sources of impure CO_2 include the flue gases of power plants and cement plants plus some off gas from natural gas processing plants. The recovery of CO_2 from the flue gas of a power plant is discussed in Section 6.4.

Naturally occurring deposits of CO_2 can also contain sufficiently high concentrations of impurities to require purification. Examples of naturally occurring deposits that require processing to pipeline quality are some of the deposits found in Mississippi, West Virginia and Wyoming. The discussion of the purification of these sources is in Section 6.3.

Some of the impurities found in the sources include nitrogen (N_2), hydrogen sulfide (H_2S), methane (CH_4) and heavier hydrocarbons, oxygen (O_2), sulfur dioxide (SO_2), sulfur trioxide (SO_3), and water vapor (H_2O). The impure CO_2 sources have large variations in gas composition and other characteristics. The concentrations of the impurities in the various sources range from zero to 90+%. As well, the sources are available for purifying at pressures ranging from atmospheric to 1000+ psig.

The sources of CO_2 are available at a wide range of compositions and pressures, and a variety of processes can be considered to obtain the most economic method. The commercially available processes to purify CO_2 are:

1. Chemically reactive absorbent processes such as amine solutions and hot carbonates
2. Physical absorbent systems such as Selexol, Rectisol and Sulfinol
3. Dry bed absorption with molecular sieves
4. Cryogenic separation techniques

The process chosen for each source must be capable of handling a large volume of gas (50-500 MMSCFD) in an economic manner to produce gas with a content of at least 98% CO_2 and less than 0.25 grains of H_2S (if present) per 100 SCF of product gas. Due to the large volume of gas and the high purity required, dry bed absorption and cryogenic separation may be eliminated from consideration due to economic reasons.

Thus, chemically reactive processes and physical solvents are the only viable processes to be considered for the CO_2 purification.

Chemical reactive absorbent systems utilize basic compounds which, in aqueous solutions, react reversibly with acidic constituents in a sour gas stream such as CO_2 or H_2S. In the absorber these reactions form soluble carbonate and sulfide salts. These salts decompose thermally in the stripper to release the acid gases. Compounds normally employed as absorbents in these systems include ethanolamines and promoted carbonates (primarily potassium carbonate). Of the chemical reactive processes, alkanolamine-monoethanolamine (MEA) is particularly well suited for acid gas removal when the gas is available at low pressure.

MEA is preferable to the other alkanolamines (diethanolamine, triethanolamine, etc.) which are used in similar processes because of its higher CO_2 absorption capacity and ease of regeneration.[20] These combine to result in reduced equipment sizes and lower heat requirements for solution regeneration in MEA systems. MEA is considered more corrosive than other amines; however, this difficulty can be minimized by limiting the concentration of the MEA in the aqueous solution to about 20% by weight or by using suitable corrosion inhibitors.

Physical absorbents are primarily neutral organic compounds containing minimal quantities of water and other ingredients. Absorption of the acid gases depends on their solubilities in these compounds. Therefore, physical absorption systems are particularly well suited to feed gas streams where acid gases are present at high partial pressures and moderate CO_2 levels (about 1%) can be tolerated in the treated gas. Since there is no chemical reaction between the acid gas and the solvent in physical absorption systems, the solvent can be regenerated by simply flashing the absorbed acid gas from the solvent at reduced pressure. Thus, input of reboiler heat is not required for solvent regeneration.

Because heat is not required to regenerate the solvent, a physical absorbent system operates at lower temperatures as compared to a chemical system. Since the solvent is noncorrosive, the system may be constructed of carbon steel. Therefore, physical absorbent systems can offer capital savings as well as energy savings for the treatment of gas streams. This report uses Allied Chemical Company's Selexol process as the physical absorbent systems.

Two primary types of CO_2 source indicated in Part 3 are the flue gases from utility power plants and from cement plants. The two processes proposed to purify these sources are the MEA system (chemically reactive absorbent), and the Selexol system (physical absorbent).

Other important sources of CO_2 are naturally occurring gases. The proposed purification processes for these sources (if necessary) are the Selexol system and the MEA system with 35% by weight MEA solution.

The transportation of the product CO_2 via supercritical pipeline (as discussed in Part 7) requires a 2000 psig inlet pressure to the pipeline. Since the purification processes produce gas at a range of pressures from below atmospheric to several hundred psig, compression of the product gas is necessary. Therefore, the cost of purification includes the processing of the gas to 98+% CO_2 and compression to 2000 psig. If the source is naturally occurring and of high purity, the compression and dehydration cost is included in the cost of production. If the source is an aboveground one of high purity, then the compression and dehydration cost is presented separately in Section 6.5.

6.2 CO_2 RECOVERY FROM FLUE GAS

6.2.1 Introduction

The maps and tables in Part 3 reveal that the majority of the available aboveground CO_2 sources are the flue gases of power plants and cement plants. These sources also represent the largest quantity of total available CO_2. The greatest CO_2 concentrations in the stack gases are those of the coal-fired plants; however, these flue gases also contain the impurities (sulfur oxides) that present additional problems.

The future will undoubtedly see more power plants constructed or modified to burn coal because of its relative abundance compared to other potential fossil fuel sources. The importance of this source of CO_2 is expected to increase relative to power plants based on alternate fuels. This section concerns the costs associated with this source. It is also expected that recovering CO_2 from coal-fired power plants will represent worst case economics. Thus, an upper limit on costs for other sources such as cement plant stack gases or natural gas-fired power plants is established.

For a power plant, the concentration of CO_2 in the flue gas is limited by the dilution effect of the excess combustion air. The flue gas contains in addition to CO_2, quantities of N_2 and O_2 plus the water vapor produced during combustion. In the case of a coal-fired power plant, quantities of fly ash and oxides of sulfur of which are primarily sulfur dioxide (SO_2) with some sulfur trioxide (SO_3) are present. The flue gases are normally available at a few inches of water positive pressure.

The flue gas composition below is the basis of design and is from a job engineered by Pullman Kellogg [21] utilizing the Kellogg/Weir magnesium promoted lime slurry process for bulk sulfur oxides removal of a coal fired power plant. The composition represents a typical flue gas which could be available from a coal fired power generating facility after bulk sulfur oxides removal.

Component	Mole Percent
Carbon Dioxide	16.5
Nitrogen	64.6
Oxygen	5.6
Water	13.3
Total	100.0
Fly Ash, Grains/SCF	0.03
Sulfur Dioxide, ppm (v)	212.00
Temperature, ^oF	125.00
Pressure, psia	14.46
Ambient Pressure, psia	14.31

To recover 125 MMSCFD of CO_2, treatment of approximately 26% of the flue gas exiting the sulfur scrubbers from the reference 917 megawatt power station is necessary. This amount is slightly more than the output of one of the four parallel units operating (five installed units, including a spare).

6.2.2 Purification Processes

Numerous purification processes are available to recover CO_2 from flue gases.[15] As discussed in Section 6.1, these include chemically reactive systems, physical absorbent systems, dry bed absorption systems and cryogenic systems. The low concentration of CO_2, the high volume of gases to be processed and the low available pressure limit the number of effective processes to two types. Those are the chemically reactive system (MEA) and the physical absorbent system (Selexol). Both systems are affected by the presence of sulfur oxides in the feed gas.

MEA reacts with sulfur oxides and will scrub them along with carbon dioxide from the flue gases. Recovery of MEA after its reaction with sulfur oxides is not as straightforward as recovery after reaction with CO_2. In the case of the CO_2 reaction product, the MEA is regenerated and the CO_2 product is obtained by simply heating the solution and stripping out the product with steam. Chemical treatment is necessary to recover the MEA which reacts with the sulfur oxides. Sodium carbonate (Na_2CO_3) can be used for this (see later discussion in Section 6.2.3).

Equipment and chemical costs to recover MEA could become quite significant if the content of sulfur oxides in the flue gas is high. For purposes of this study, the bulk of any sulfur oxides present in the flue gas are assumed to be removed by a pretreatment step. This is a logical assumption with the continuing trend to clean up power plant flue gases as a means of protecting the environment.

A wetscrubbing system for removal of sulfur oxides from flue gases removes about 90% of the amount originally present. The treated flue gas typically contains about 200 ppm residual sulfur oxides which are essentially all SO_2.

The sulfur oxides also affect the Selexol system because they are absorbed simultaneously with the CO_2. The Selexol is hydrophilic and is designed to contain 5% by weight water. The absorbed sulfur oxides combine with the water to form acids. Thus, the normally noncorrosive Selexol system becomes a corrosive one in the presence of sulfur oxides. In addition, sulfuric acid (H_2SO_4, formed by combination of SO_3 and water) irreversibly degrades the Selexol solution.

To develop the economics for supplying CO_2 recovered from power plant flue gas, this report assumes the purification facilities are additions to an existing power plant. No attempt is made to integrate the utility systems with the power plant (e.g., by use of co-generated steam).

The following are descriptions of both the MEA process and the Selexol process for recovery of CO_2 from a coal-fired power plant flue gas.

6.2.3 Process Descriptions

MEA

Reference is to Figure 6.1 which illustrates the process described.

Carbon dioxide is recovered from the flue gas by use of an aqueous solution of 20% by weight MEA. Flue gas at the inlet of the treatment facilities, after pervious bulk sulfur oxides removal, is at $125^{\circ}F$ and 14.46 psia. To overcome pressure loss created by the MEA absorption tower, the flue gas is first compressed to about 16.0 psia before it enters the tower. The absorption tower includes water wash trays at the top to reduce MEA losses. As the flue gas passes through the absorption tower, the CO_2 is removed by reacting with the MEA solution according to the following reversible reaction:

$$HO(CH_2)\ NH_2 + H_2) + CO_2 = HO(CH_2)NH_3HCO_3$$

The forward reaction, taking place in the absorber, is favored by moderate operating temperatures in the range of $100-175^{\circ}F$.

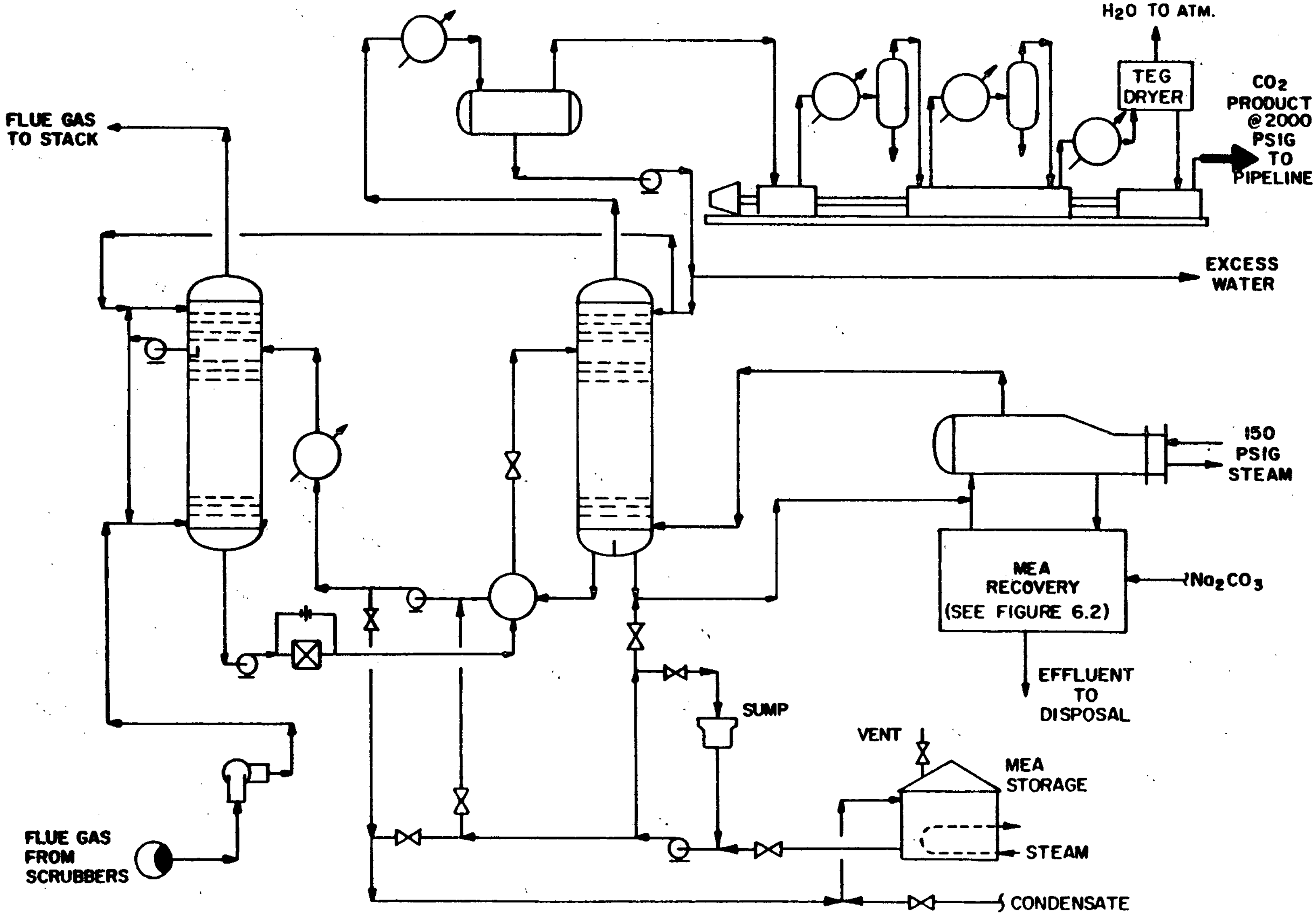
The MEA solution loaded with CO_2 flows from the bottom of the absorber via a transfer pump to the top of the regenerator tower. A mechanical filter operating on about 25% of the spent solution removes solids such as pipe scale, fly ash, and degradation products from the solution.

Before entering the regenerator, the CO_2 loaded solution exchanges heat with the regenerated MEA solution exiting the bottom of the regenerator tower.

The bottom of the regenerator tower operates at 10.5 psig. A steam heated reboiler supplies the necessary heat input to the regenerator. Elevated temperature combined with steam stripping in the regenerator favors the reverse of the reaction that takes place in the absorber, thus regenerating the MEA solution.

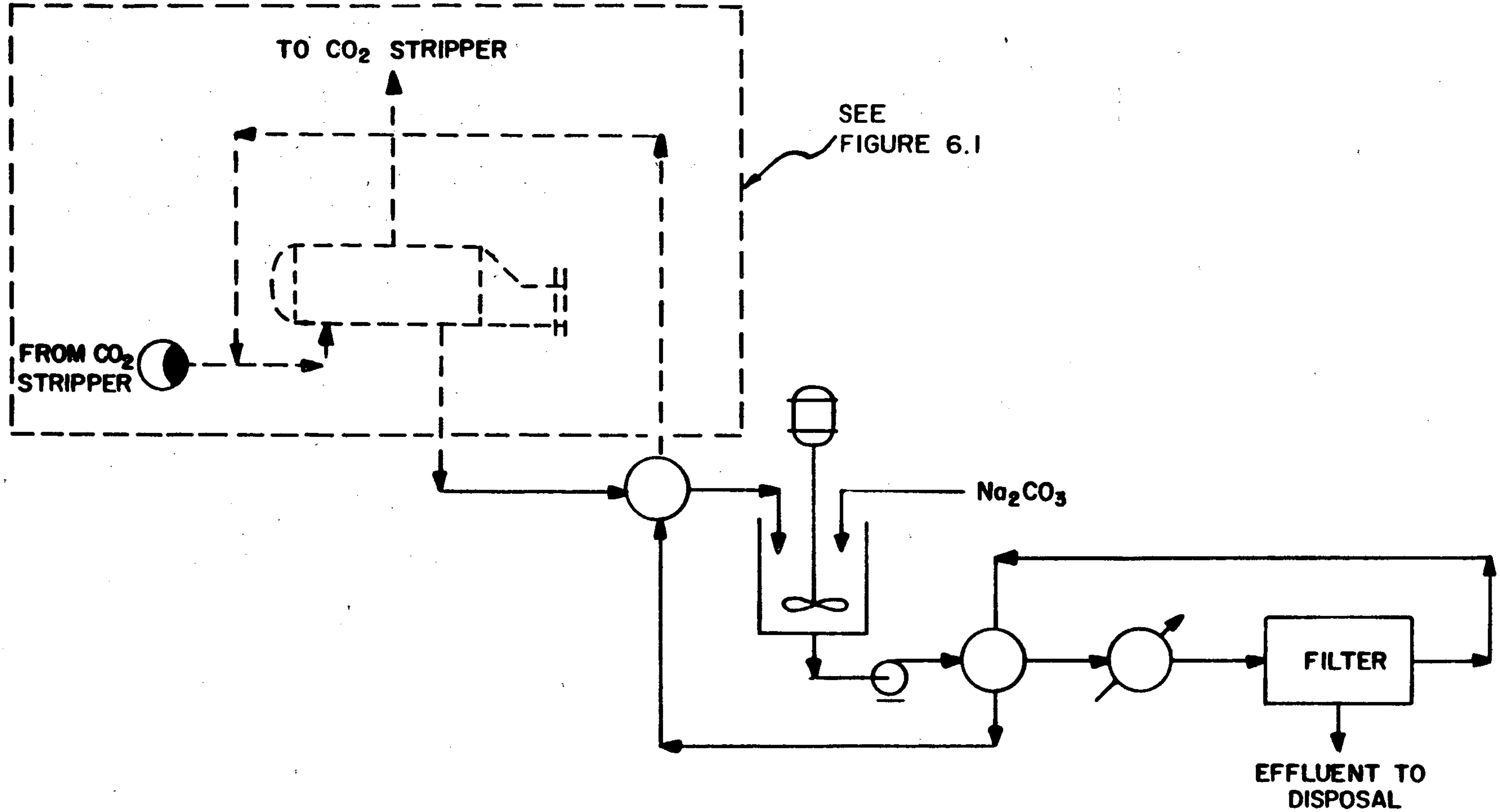
A kettle reboiler with total solution vaporization provides a high degree of concentration for the MEA solution, which is sent to MEA recovery (see later discussion).

Carbon dioxide is stripped from the MEA solution and passes overhead of the tower through a condenser where the bulk of the water is condensed. Most of the water returns to the regenerator as reflux. Some flows to the absorber wash trays which include a pumparound system. Excess water resulting from cooling the water saturated flue gas in the absorber is pumped to disposal.



FLOW DIAGRAM FOR MEA PROCESS TO RECOVER CO_2 FROM FLUE GAS

FIGURE 6.1



CONTINUOUS MEA RECOVERY WITH Na_2CO_3

FIGURE 6.2

MEA solution from the bottom of the regenerator tower cools by heat exchange, first with the spent MEA from the absorber tower, and finally with cooling water before the solution returns to the top of the absorber.

The system includes a sump used for chemical mixing and for collecting drips and spills which occur during operation. Facilities also provide for MEA storage during initial charging of the system and during shutdowns. Facilities are necessary to recover MEA which has reacted with the sulfur oxides present in the flue gas. These recovery facilities are discussed below.

MEA Recovery (Proposed)

Solution losses occur in amine systems due to formation of degradation products which are stable at the conditions employed for regeneration of the MEA solution. They also occur due to vaporization of MEA in the absorber and regenerator (wash trays are provided to minimize vaporization losses), and by leaks and spills from the system.

Formation of degradation products may occur for a number of reasons such as overheating the solution and chemical reaction of the amine with a component in the gas being scrubbed. In the present case of a flue gas from a coal fired power plant, MEA reacts with the residual sulfur oxides (and may also react with the residual oxygen[22] to form stable degradation products.

The usual method of controlling degradation products in an MEA system is the installation of a reclaimer. Degradation products concentrate in the liquid with boiling and are removed periodically from the reclaimer for disposal. Sodium carbonate or sodium hydroxide is sometimes added to the reclaimer to enhance MEA recovery and/or prevent corrosion[22].

For the MEA system processing flue gas, the larger quantity of potential degradation products formed by reaction of MEA with sulfur oxides requires a different approach to recover the MEA solution. Addition to the solution of a compound that is chemically a stronger base than MEA liberates the MEA from the thermally stable reaction product formed by MEA and sulfur oxides. Either sodium hydroxide (NaOH) or sodium carbonate (Na_2CO_3) serves this purpose. The quantity of sulfur oxides remaining in a flue gas after bulk removal ties up significant quantities of the MEA, and a continuous reclaiming process is necessary.

In the reclamation process, Na_2CO_3 is preferred because NaOH displaces MEA from its reaction product with CO_2 equally as well as from the undesirable degradation products. This leads to high and unnecessary chemical consumption.

Figure 6.2 shows the MEA recovery process utilizing Na_2CO_3. A slipstream of MEA solution flows from the regenerator reboiler and cools by heat exchange against the reclaimed solution returning to the reboiler. Sodium carbonate is added to the solution in a mix tank which also serves as a hold tank for the mixture. Thus, residence time is provided to allow the dissociation reactions to liberate the MEA. The solution from the mix tank cools, causing precipitation of sodium sulfate and sodium sulfite which are removed in a filter along with other solids and degradation products. Purified MEA solution returns to the system.

The proposed MEA recovery system is not well enough defined for an estimate of its cost to be included in this study. Additional work, including laboratory study, is believed necessary to prepare a final design for such a system.

Compression and Drying

The product CO_2 is delivered to the compression step at about 5 psig and 110°F saturated with water vapor. The gas is compressed in a three case centrifugal compressor powered by a steam turbine to a final discharge pressure of 2,000 psig for delivery to the gas pipeline. As the gas is compressed and cooled between stages, water condenses and is removed in knockout vessels. Between the second and third case the gas is dried to final pipeline quality by a triethylene glycol (TEG) dryer.

Utility Systems

Utility costs used in Part 6.2.4 are based on using steam driven turbines for the CO_2 product and recycle compressors and the flue gas blowers, and electric drivers for all remaining rotating equipment. Small additional steam requirements are also included for regeneration of the glycol solution in the triethylene glycol dryer.

Two steam pressure levels were selected. Steam imported from offsite is assumed available at 900 psig and 900°F. A second level at 175 psig supplies heat to the amine reboiler through a letdown station at 150 psig.

The turbine drivers for the CO_2 product and recycle compressors as well as the flue gas blowers are driven by 900 psig steam and exhaust to the 175 psig header.

The MEA system requires reboiler heat for solution regeneration. The exhaust steam from the turbines provides most of the amine reboiler heat; however, some additional 900 psig letdown to the 175 psig level is required. A desuperheat station is provided at the inlet of the amine reboiler. The 150 psig steam condensate returns to battery limits from the amine reboiler.

Cooling water is assumed available at battery limits at 90°F.

Environmental Considerations

Effluent streams are produced during normal operation of the recovery and gas transmission facilities. The following is a brief discussion of the expected effluent streams, and possible methods of handling them. Cost for treatment facilities are not included in plant investment.

Excess water is produced in the recovery plant by condensation of water vapor from the flue gas as it passes to the absorber. This effluent stream is relatively clean. It contains dissolved CO_2 plus traces of other gaseous components absorbed from the raw flue gas. This water can be reused in a process plant, usually after steam stripping, for boiler feedwater makeup or as makeup water to a cooling tower.

If present, the MEA stripper overhead which is rich in MEA can be recycled to the process.

Because the process for MEA recovery using Na_2CO_3 is not fully defined, it is impossible to define in detail the effluent expected from the process. The major component should be sodium sulfite with perhaps some sodium sulfate. If a crystallization process is used, the product could be reasonably pure. The degree of purification could be improved (as necessary) by further processing such as recrystallization or washing.

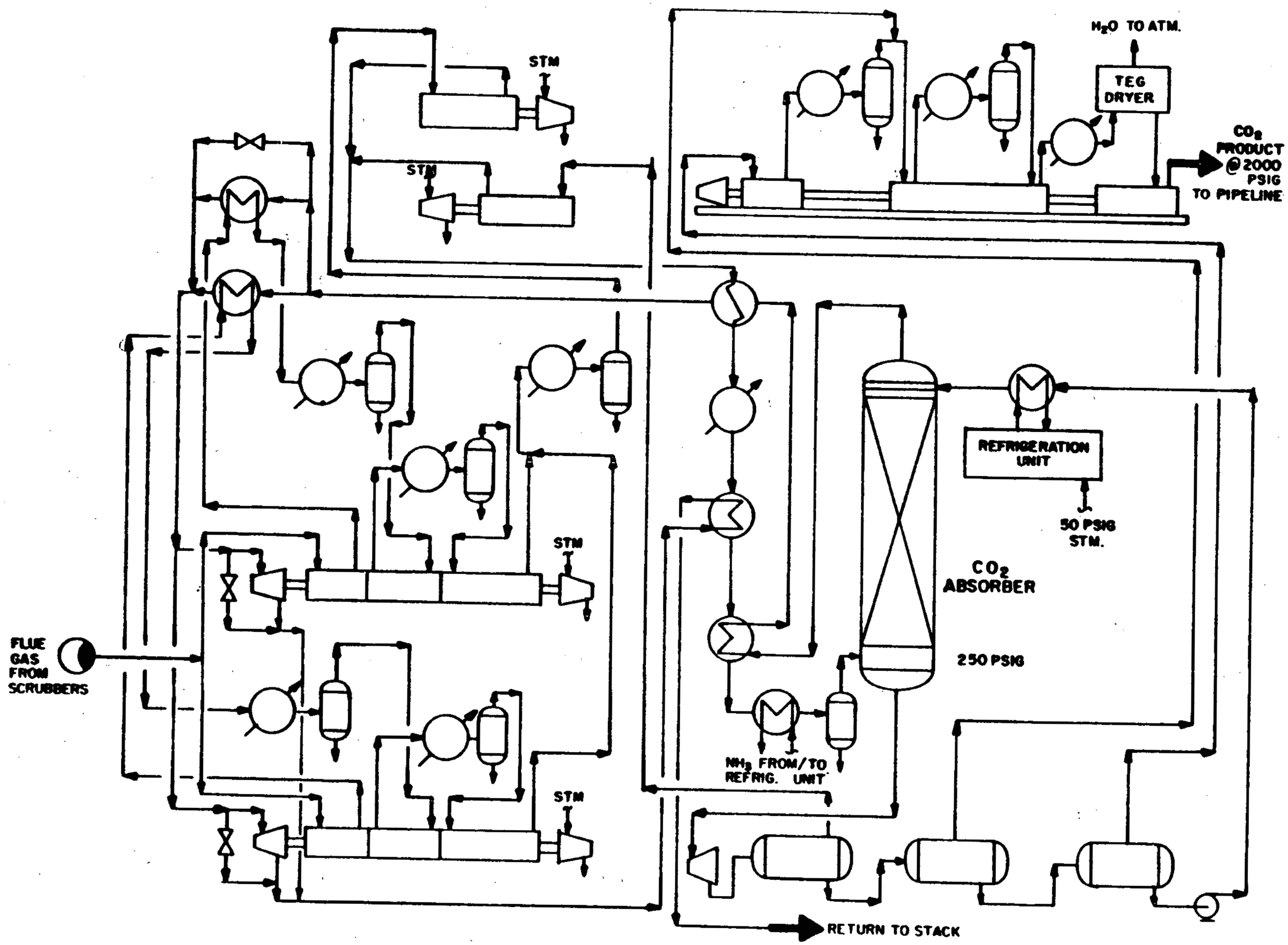
Water will be produced from the product gas as it is compressed and cooled between stages. This water can be combined with the excess water from the process system for disposal.

Regeneration of the TEG solution in the CO_2 product gas dryer will produce a gaseous effluent consisting mostly of water vapor plus traces of TEG. This effluent stream can be vented to the atmosphere.

Selexol

Reference is to Figure 6.3 which illustrates the process described below.

Recovery of CO_2 from the flue gases of power plants may also be accomplished by the use of a physical absorption system. Prior to entering the absorber, the flue gas must be compressed. Physical absorption systems utilize partial pressure as a driving force and a gas with 16.5% CO_2 at atmospheric pressure does not provide sufficient partial pressure for the recovery to be practical or economical. A minimum absorber operating pressure of 250 psig has been assumed. Therefore, the flue gas is compressed to an intermediate pressure in two parallel, three section, axial compressors driven by turbo expanders supplemented by steam turbines. After each section the gas is cooled by heat exchange with cooling water and knockout drums remove the condensed water. The parallel stream flows then combine and are compressed to the final necessary pressure in a one section centrifugal compressor.



FLOW DIAGRAM FOR SELEXOL PROCESS TO RECOVER CO_2 FROM FLUE GAS

FIGURE 6.3

The feed gas then combines with the recycle gas from the recycle compressor and this combination is cooled prior to entering the absorber. The combined absorber feed cools by heat exchange with the absorber overhead, cooling water, the returning flue gas from the exhaust of the turbo-expanders and with ammonia refrigerant. As the absorber feed cools, water condenses and is removed in a knockout drum prior to entering the absorber.

The absorption tower is a conventional, counter-current, packed bed absorber. In the absorber, CO_2, H_2O, N_2 and O_2 dissolve in the Selexol solution and exit the absorber to flash through a hydraulic turbine. The turbine connects to the lean solution circulation pumps and serves to recover pumping energy. This minimizes the net process energy requirements.

The flashed vapor and liquid enter the first of several staged flash drums. The vapor from the first flash drum returns to the absorber as recycle vapor. The recycle vapor contains much of the absorbed N_2 and O_2 and ensures that the CO_2 product purity is met.

Regeneration of the Selexol solution and the production of the CO_2 product occurs in the staged adiabatic flashing of the liquid from the recycle flash drum. The staging of the product flashes reduces the load on the product CO_2 compressor and consequently the energy consumption. The last flash drum pressure is sufficiently low enough to regenerate the Selexol to achieve the specified absorber overhead CO_2 level.

The initial adiabatic flash is the only flash in the process system that utilizes a hydraulic turbine for energy recovery. Feedback from pump manufacturing representatives indicates that any other utilization of the hydraulic turbines will be economically unjustifiable.

Centrifugal pumps recycle the regenerated lean solution back to the top of the absorber to complete the cycle. The pumps are driven by the hydraulic turbine with supplementary power provided by electric motors. An ammonia absorption unit provides refrigerant for cooling of the lean solution and absorber feed gas. The refrigerant maintains the overall process heat balance and provides cold absorber feeds that enhance the absorption operation.

The combined feed gas to the absorber and the compressed gas from the first section of the two parallel feed gas compressors heat the cold absorber overhead gas. The heated overhead gas then expands through two parallel turbo-expanders which drive the two parallel feed gas compressors. The turbo-expanders recover approximately 25% of the required feed gas compression horsepower. The cold exhaust from the turbo-expander then heat exchanges with the combined feed gas to the absorber and is returned to the stack at $100^{\circ}F$.

Compression and Drying

The product CO_2 enters the compressor from below atmospheric pressure to 20 psig via the staged product flashes. Compression to the required 2000 psig pipeline inlet pressure occurs in a multi-stage, centrifugal compressor with a steam turbine driver. During compression and interstage cooling by air coolers, water produced with the CO_2 condenses and separates from the system through the knockout vessels. Prior to the last stage of compression final dehydration of the CO_2 product to pipeline quality occurs in the TEG dryer.

Utility Systems

Utility costs presented later in this part are based on use of steam driven turbines for the CO_2 product, feed gas and recycle compressors, and electric drives for all remaining rotating equipment. Small additional steam requirements are also included for regeneration of the glycol solution in the triethylene glycol dryer.

Two steam pressure levels were selected. Steam imported from offsite is assumed available at 900 psig - 900^{o}F. A second level at 75 psig supplies heat to the ammonia absorption unit through a letdown station to 50 psig.

The turbine drivers for the CO_2 product, feed gas and recycle compressors are driven by the 900 psig steam that condenses to exhaust at a pressure of four inches of mercury.

The ammonia absorption unit requires reboiler heat for its operation. The 75 psig steam provides the reboiler heat through a letdown station to

50 psig. A desuperheat station is provided at the inlet of the unit's reboiler. The 50 psig steam condensate returns to battery limits from the unit.

Cooling water is assumed available at battery limits at 90°F.

Environmental Considerations

Effluent streams are produced during normal operation of the recovery and gas transmission facilities. The following is a brief discussion of the expected effluent streams, and possible methods of handling them. Costs for treatment facilities are not included in plant investment.

Excess water is produced in the recovery plant by condensation of water vapor from the flue gas as it passes to the absorber. This effluent stream is relatively clean. It contains dissolved CO_2 plus traces of other gaseous components absorbed from the raw flue gas. This water can be reused in the power plant, usually after steam stripping, for boiler feedwater makeup or as makeup water to a cooling tower.

Water will be produced from the product gas as it is compressed and cooled between stages. This water can be combined with the excess water from the process system for disposal.

Regeneration of the TEG solution in the CO_2 product gas dryer will produce a gaseous effluent consisting mostly of water vapor plus traces of TEG. This effluent stream can be vented to the atmosphere.

6.2.4 Cost of CO_2 Recovery

As discussed in the previous sections, two processes have potential for CO_2 recovery from power plant flue gas. The investment cost, annual operating cost and resulting price of the recovered gas at various DCFRR's are presented in this section.

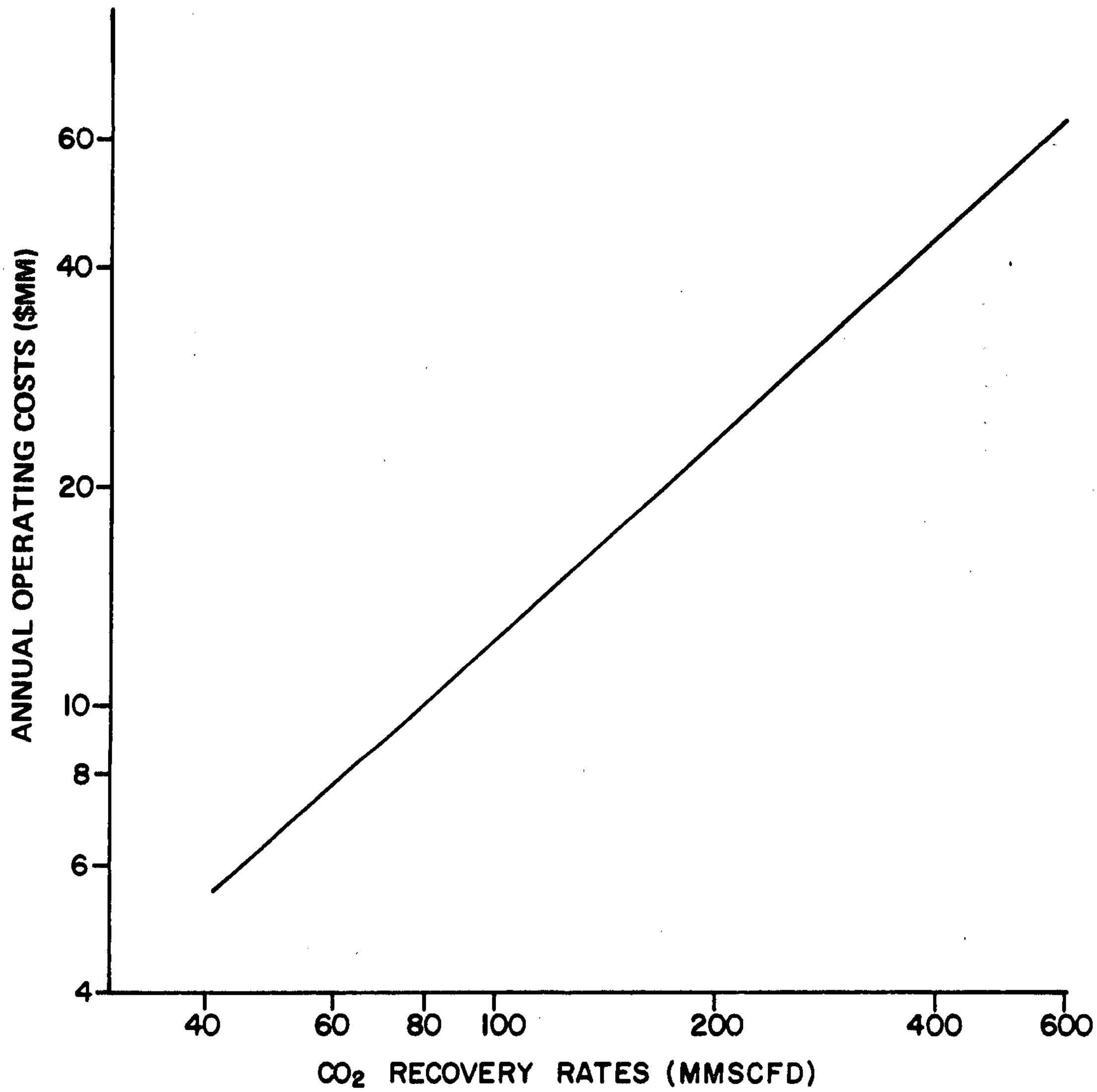
The basis of the total investment cost is outlined in Appendix B1 the basis for determining the annual operating expenses is outlined in Appendix B2. The basis for the calculations of the resulting price of the recovered gas is:

1. 20 year project life (after start-up)
2. 20 year straight line depreciation
3. 100% equity capital (no debt)
4. The salvage value is zero
5. The working capital is $16,000/MMSCFD CO_2

The annual operating costs for the MEA process at various CO_2 recovery rates are shown in Figure 6.4. Figure 6.5 illustrates the corresponding investment costs associated with the various recovery rates and Figure 6.6 shows the resulting CO_2 recovery costs for these rates at the different DCFRR's. The costs do not include the costs for the proposed MEA recovery system, any cooling water circulation pumps, or the disposal of condensed process water.

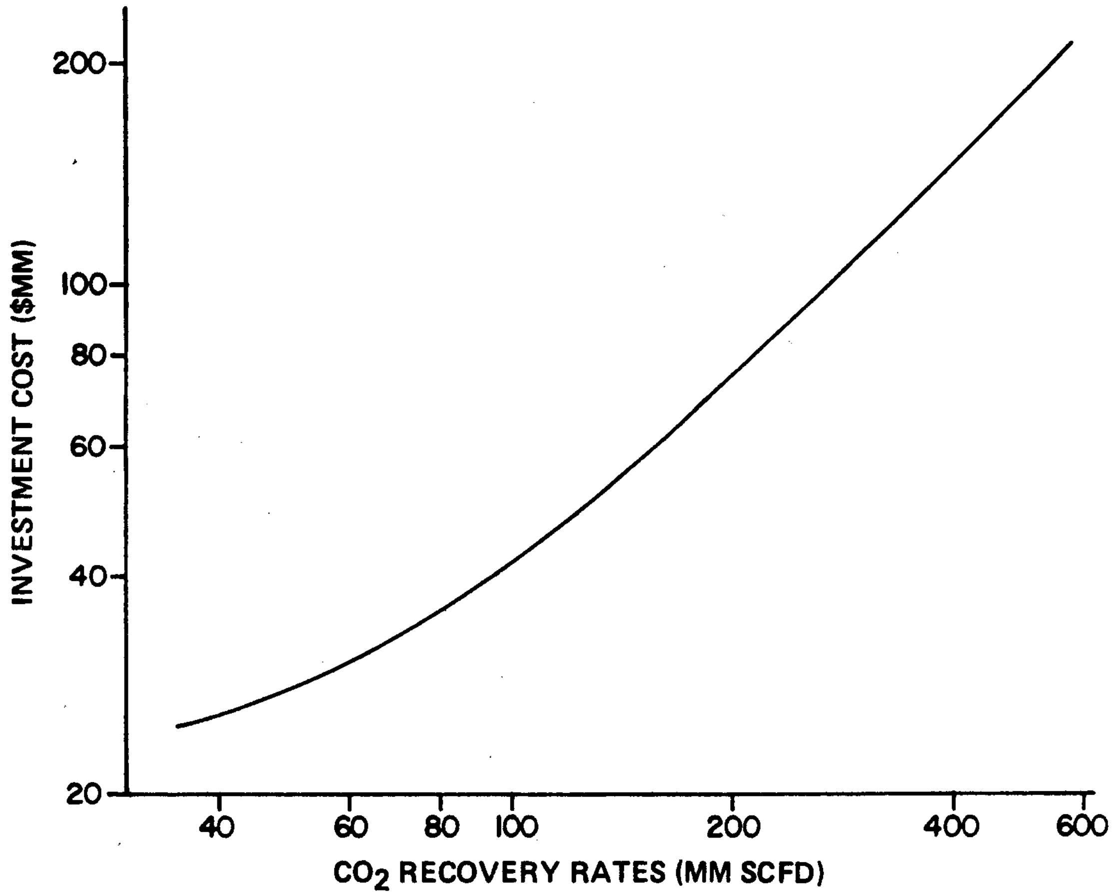
As Figure 6.6 shows, the CO_2 recovery costs for a recovery rate of 125 MMSCFD of CO_2 range from $0.60/MSCF - $1.03/MSCF for 10% - 25% DCFRR.

For comparison of the Selexol process to the MEA process, a 125 MMSCFD recovery unit was designed and cost estimated. A recovery unit of this capacity using Selexol has an investment cost of approximately $114,000,000 and an annual operating cost of approximately $60,700,000. The costs do not include any cooling water circulation pumps, or the disposal of condensed process water. The CO_2 recovery costs for the Selexol process recovering 125 MMSCFD CO_2 range from $1.97 - $2.96/MSCF for 10% - 25% DCFRR.



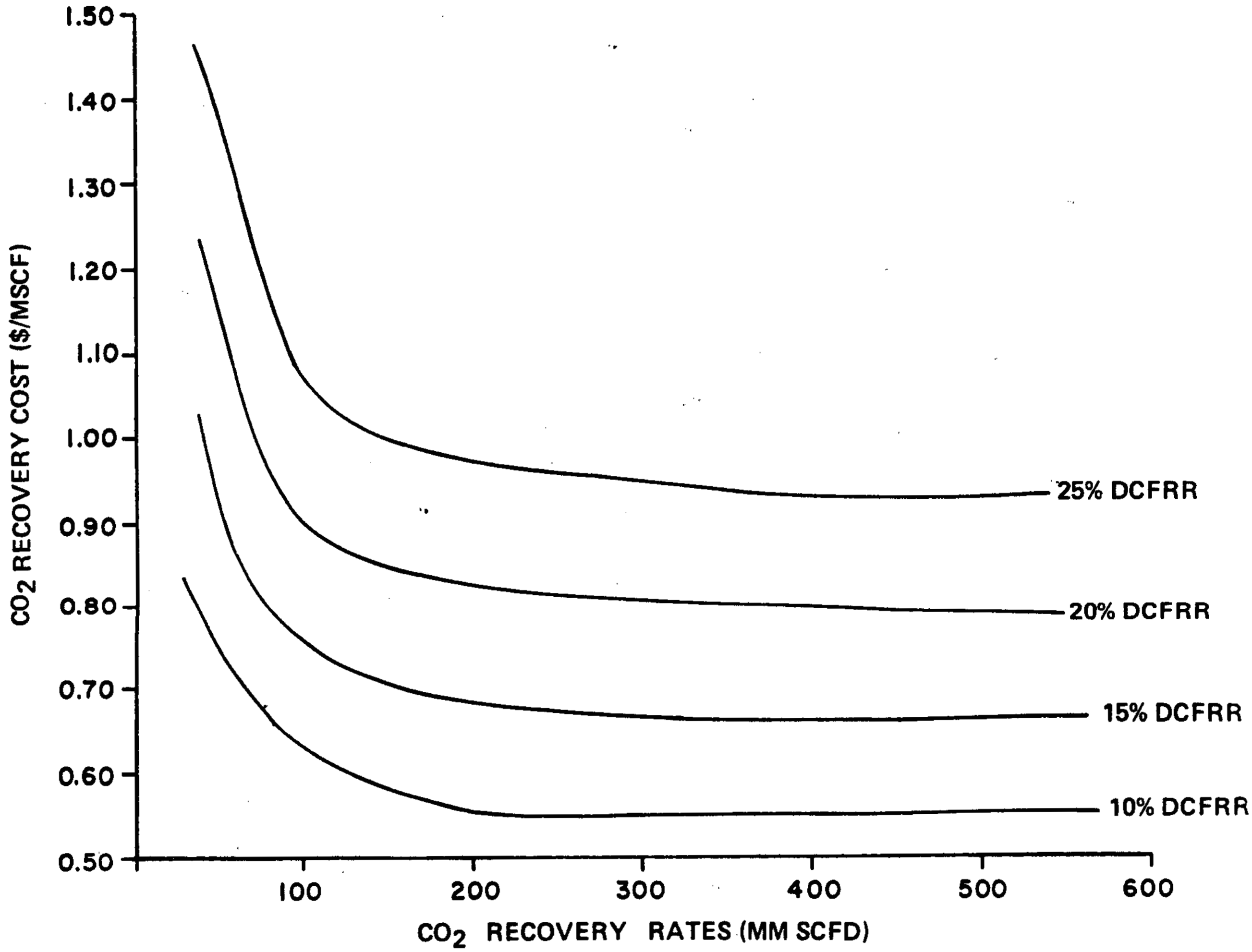
ANNUAL OPERATING COSTS FOR CO_2 RECOVERY FROM POWER PLANT FLUE GASES USING THE MEA PROCESS VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.4



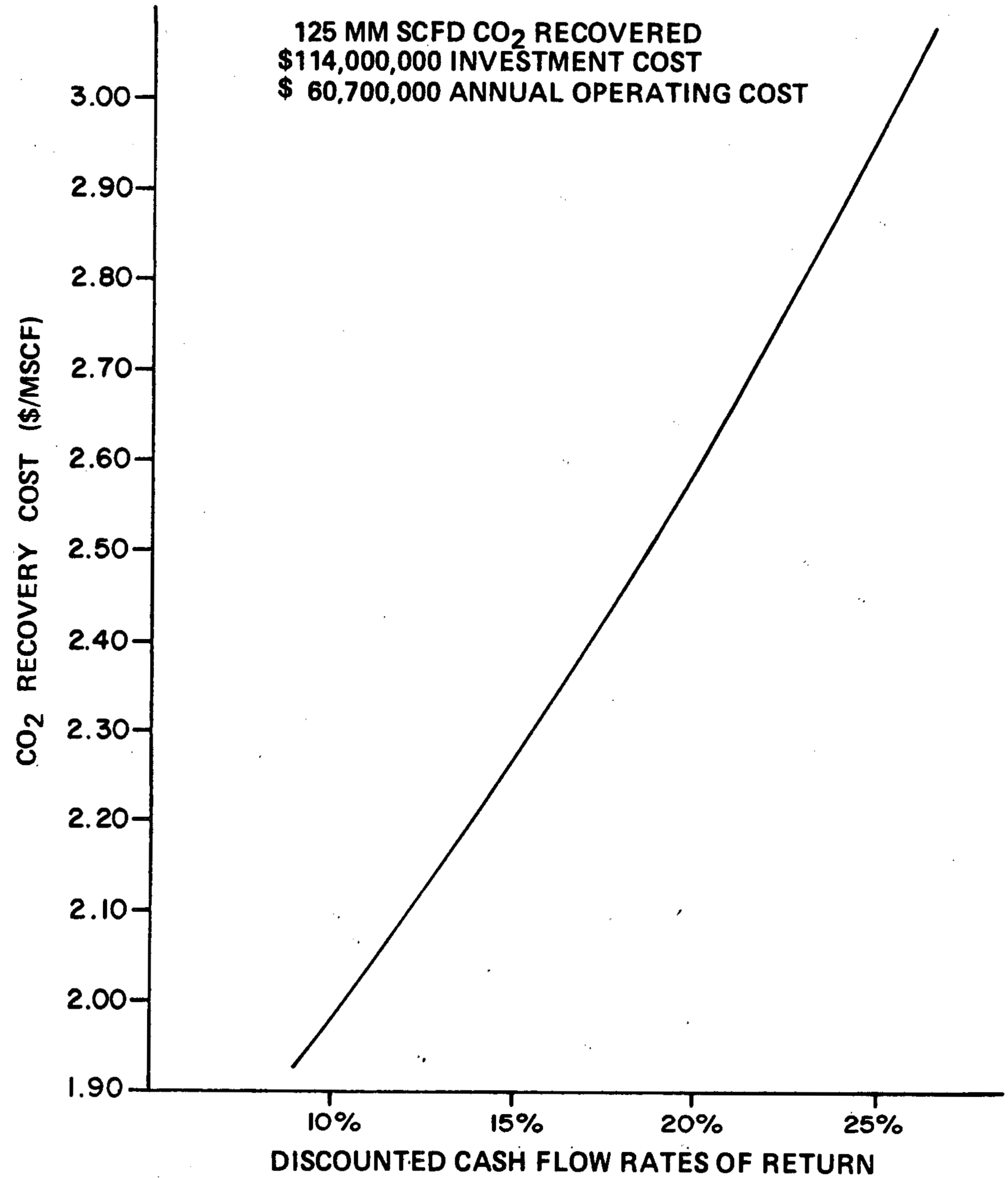
INVESTMENT COSTS FOR CO_2 RECOVERY FROM POWER PLANT FLUE GASES USING THE MEA PROCESS VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.5



COST OF CO_2 RECOVERY FROM POWER PLANT FLUE GASES USING THE MEA PROCESS FOR VARIOUS CO_2 RECOVERY RATES AND DISCOUNTED CASH FLOW RATES OF RETURN

FIGURE 6.6



COST OF CO_2 RECOVERY FROM POWER PLANT FLUE GASES USING THE SELEXOL PROCESS VERSUS VARIOUS DISCOUNTED CASH FLOW RATES OF RETURN

FIGURE 6.7

6.2.5 Conclusions

As previously stated in Section 6.2.4, the recovery costs for 125 MMSCFD of CO_2 are from \$0.60/MSCF - \$1.03/MSCF for a MEA unit and from \$1.97/MSCF - \$2.96/MSCF for a Selexol unit using a 10% - 25% DCFRR range. Though the MEA costs do not include the costs for a MEA recovery unit, the total cost for recovery via the MEA process will be approximately a third that necessary for recovery via Selexol.

One reason for the extra cost of the Selexol unit is the necessity to raise the partial pressure of the CO_2 to enable absorption to occur. The requirement for raising the partial pressure is discussed in Section 6.3.1. The additional capital and utility costs associated with the compression of the flue gas make the Selexol unit economically undesirable. Therefore, referring to Figure 6.6 in Section 6.2.4 will provide the cheapest costs of CO_2 recovery from power plant flue gases.

6.3 PURIFICATION OF NATURAL SOURCES OF CO_2

6.3.1 Introduction

Naturally occurring sources of CO_2 are available at a variety of pressures and CO_2 concentrations. Because of the wide variations in source characteristics found (see Part 3), this discussion presents the associated costs for purification of these sources as a function of CO_2 concentration and available pressure. Therefore, the associated costs of purification are presented graphically in part 6.3.4 of this report.

6.3.2 Purification Processes

Limitations exist on the quality of the source purified. Low concentrations of CO_2 (thus low CO_2 partial pressure) produce an additional load on the purification process when compared with sources containing higher concentrations. As discussed previously, two processes may be considered for purification purposes; a chemically reactive system, MEA, and a physical absorbent, Selexol. Literature(23)(24)(25) and operating experience show that for CO_2 partial pressures above 200 psig, Selexol is the most economical process of the two. For lower partial pressures, the MEA system should be considered.

Therefore, the results in Part 6.3.4 discuss Selexol system utilizing various operating pressures for feed gas at 75°F with 25%, 50%, 75% and 90%

CO_2 concentrations. Also, the Selexol system is compared with a MEA system for a feed gas of 10% CO_2 operating at 1000 psig. Any natural sources with concentration significantly higher than 90% CO_2 (and without H_2S) may be utilized directly after dehydration.

Gases containing toxic impurities such as H_2S require an additional purification step than those gases without H_2S. The H_2S in the feed gas must be removed from both the CO_2 and CH_4 product streams in this case. The quantity in the feed gas is too small to use a direct sulfur removal process such as a Claus process. In addition, the partial pressure of the CO_2 is above the current design experience for a process such as Stretford.

Therefore, the process selected must have two distinct functions. First it must upgrade the H_2S content to a level permitting recovery, and second it must recover the CO_2 and CH_4 economically.

One of the properties of a physical solvent such as Selexol is its natural selectivity between the various acid gas constituents. For Selexol, absorption of H_2S is preferred over CO_2 by a ratio of about seven to one.

After review of alternate process schemes, the nature of a physical solvent process such as Selexol seemed idealy suited to this case. In addition, such a process offers low utilities and investment costs.

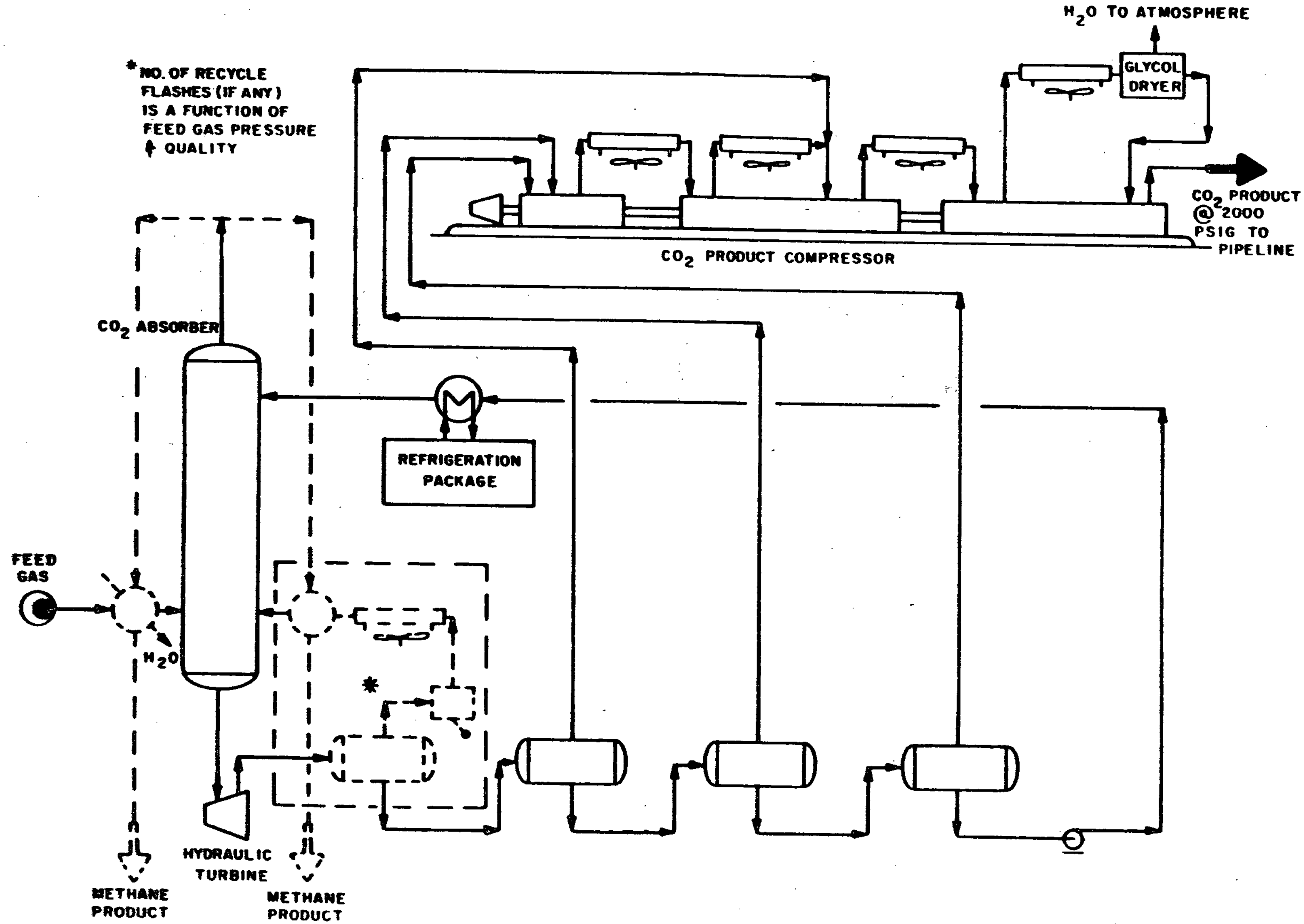
The H_2S is available from the Selexol process as approximately a 75% gaseous mixture with CO_2. It is possible to separate the H_2S from the CO_2 in this gas, and recover the sulfur using a Stretford process. The reason this is possible is that the gas stream is at low pressure (18 psia). Although the CO_2 still interferes with the process, the partial pressure of CO_2 is within demonstrated design limits for the Stretford process.

6.3.3 Process Descriptions

Selexol

Reference is to Figure 6.8, A Generalized Selexol Process Flowsheet.

Naturally occurring CO_2 and CH_4 saturated with water vapor at 75°F enters the CO_2 absorption tower. The tower is a conventional, countercurrent, packed bed absorber. If recycle is not necessary, the feed gas cools by heat exchange against the absorber overhead gas. This serves to recover some of the refrigeration required in the process and tends to lower the absorber operating temperature thus enhancing the CO_2 absorption. The overhead gas is the CH_4 product containing approximately 2% CO_2 and is essentially dry. About 98% of the total CH_4 is recovered in the overhead product stream.



A GENERALIZED SELEXOL PROCESS FLOWSHEET

FIGURE 6.8

The absorbed CO_2 and CH_4 exits the absorber tower dissolved in the rich Selexol solution and flashes through a hydraulic turbine. The turbine connects to the lean solution circulation pumps and serves to recover pumping energy. This minimizes the net process energy requirements.

The flashed vapor and liquid enter the first of several staged flash drums. If the feed CO_2 content is less than 85%, recycle vapor is necessary to reduce the CH_4 losses and secure CO_2 product purity. The lowest pressure of these adiabatic recycle flashes is set to assure the product purity. The number of recycle flashes is a function of the absorber operating pressure and the final recycle pressure required. The staging reduces the load (therefore, the energy consumption) on the recycle compressor. A gas turbine drives the multistage, intercooled recycle compressor. Electric motor driven air coolers provide interstage cooling and cool the recycle gas exiting the machine. The recycle gas cools further by heat exchange with the absorber tower overhead product. For an equipment list for the various feed gas qualities and pressures studied, refer to Appendix B4.

Regeneration of the Selexol solution and the production of the CO_2 product occurs in the staged adiabatic flashing of the liquid from the last recycle drum (or in the cases with no recycle, after the hydraulic turbine). The staging of the product flashes reduces the load on the product CO_2 compressor and consequently the energy consumption. The number of product flashes is dependent on the necessary pressure to sufficiently regenerate the Selexol solution enough to achieve the specified absorber overhead CO_2 level and the last recycle flash pressure (if there is recycle) or the absorber operating pressure (if there is no recycle).

The initial adiabatic flash is the only flash in the process system that utilizes a hydraulic turbine for energy recovery. Comments from pump company representatives indicate that any other utilization of the hydraulic turbines is economically unjustifiable.

Centrifugal pumps recycle the regenerated lean solution back to the top of the absorber to complete the cycle. The pumps are driven by the hydraulic turbine with supplementary power provided by electric motors. An ammonia absorption unit utilizing the exhaust heat from the gas turbines driving the recycle compressor and the CO_2 product compressor provides refrigerant for cooling of the lean solution. The refrigerant maintains the overall process heat balance and provides a cold lean solution that enhances the absorption operation.

Compression and Drying

The product CO_2 enters the compressor from approximately atmospheric pressure (or slightly below) to pressures of 350 psig via the staged product flashes. Compression to the required 2000 psig pipeline inlet pressure occurs in a multi-stage, centrifugal compressor with a gas turbine driver. During compression and interstage cooling by air coolers, water produced with the CO_2 condenses and separates from the system through the knockout vessels. Prior to the last stage of compression, final dehydration of the CO_2 product to pipeline quality occurs in the TEG dryer.

MEA

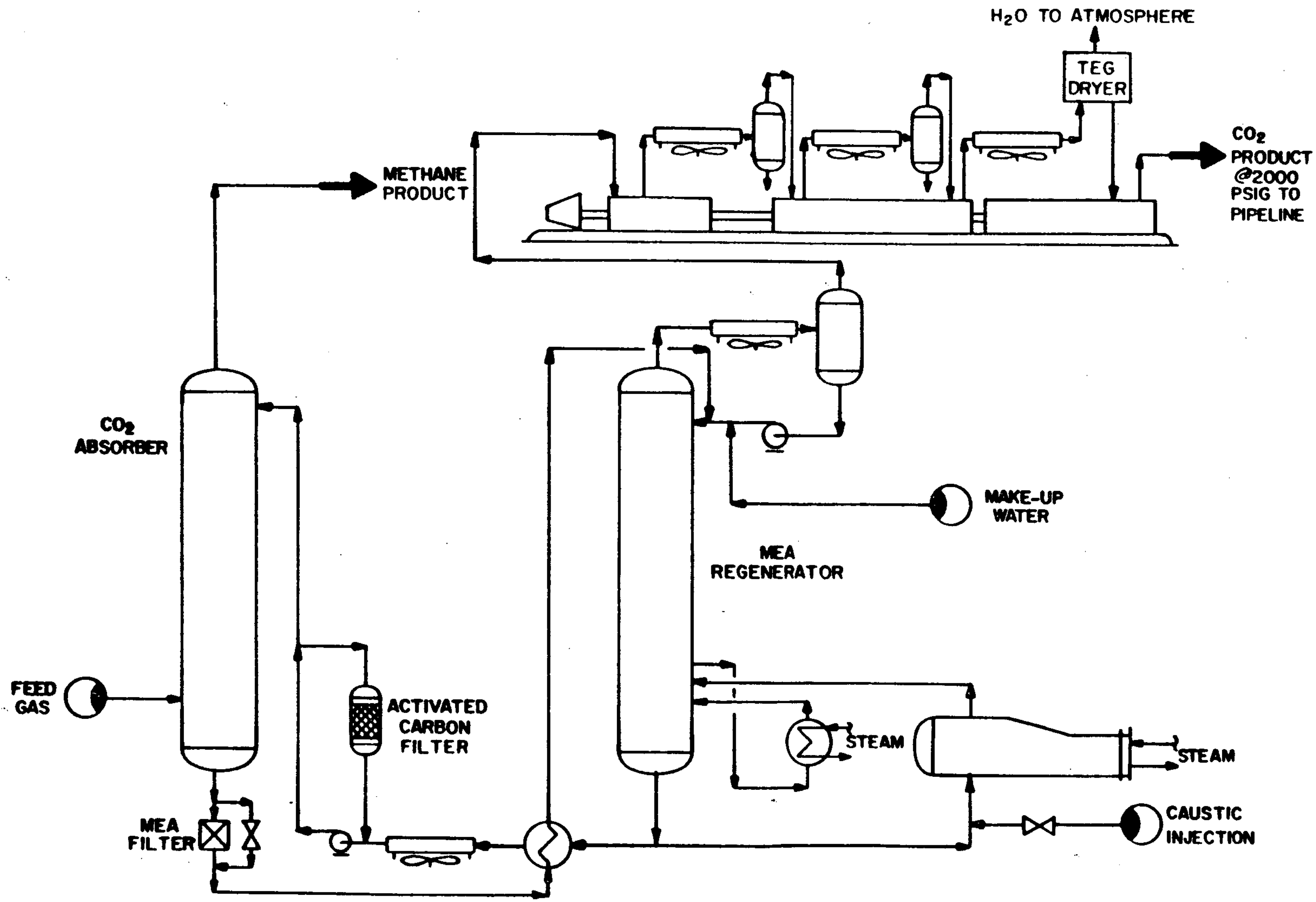
This process is for the purification of a natural source with a composition of 10% CO_2 and 90% CH_4, saturated with water vapor and available to the treatment facilities inlet at 1000 psig and 75^{o}F. The following description refers to Figure 6.9 .

The CO_2 recovery is with the use of 35 weight % MEA solution in water. The CO_2 removal from the feed gas occurs in a counter-current, trayed MEA absorption tower. As the feed gas passes through the absorption tower, the CO_2 is removed by reacting with the MEA solution according to the following reversible reaction:

$$HO(CH_2)NH_2 + H_2O + CO_2 = HO(CH_2)NH_3HCO_3$$

The forward reaction, taking place in the absorber, is favored by moderate operating temperatures in the range of 100-175^{o}F.

The MEA solution loaded with CO_2 flows from the bottom of the absorber via a transfer pump to the top of the regenerator tower. A



FLOW DIAGRAM FOR MEA PROCESS TO RECOVER CO_2 FROM NATURAL SOURCES

FIGURE 6.9

mechanical filter operating on about 25% of the spent solution removes solids such as pipe scale, fly ash, and degradation products from the solution.

Before entering the regenerator the CO_2 loaded solution heat exchanges against the regenerated MEA solution from the bottom flow of the regenerator tower.

The bottom of the regerator tower operates at 10.5 psig. A steam heated reboiler supplies the necessary heat input to the regenerator. Elevated temperature combined with steam stripping in the regenerator favors the reverse of the reaction that takes place in the absorber, thus regenerating the MEA solution.

In the regenerator, CO_2 is stripped from the MEA solution and passes overhead through an air cooler where the bulk of the water condenses. The condensate returns to the regenerator as reflux. To maintain the water balance of the system and provide sufficient stripping steam, make-up water is added to the condensate reflux.

MEA solution from the bottom of the regenerator cools by heat exchange against the spent MEA from the absorber and by an air cooler before the solution flows to the top of the absorber.

The MEA solution is inhibited with Union Carbide's "Amine Guard" to protect against corrosion. A carbon filter, operating on 1.4% of the lean MEA flow to the top of the absorber is provided.

The system includes a sump used for chemical mixing and for collecting drips and spills which occur during operation. Facilities also provided for MEA storage during initial charging of the system and during shut-downs.

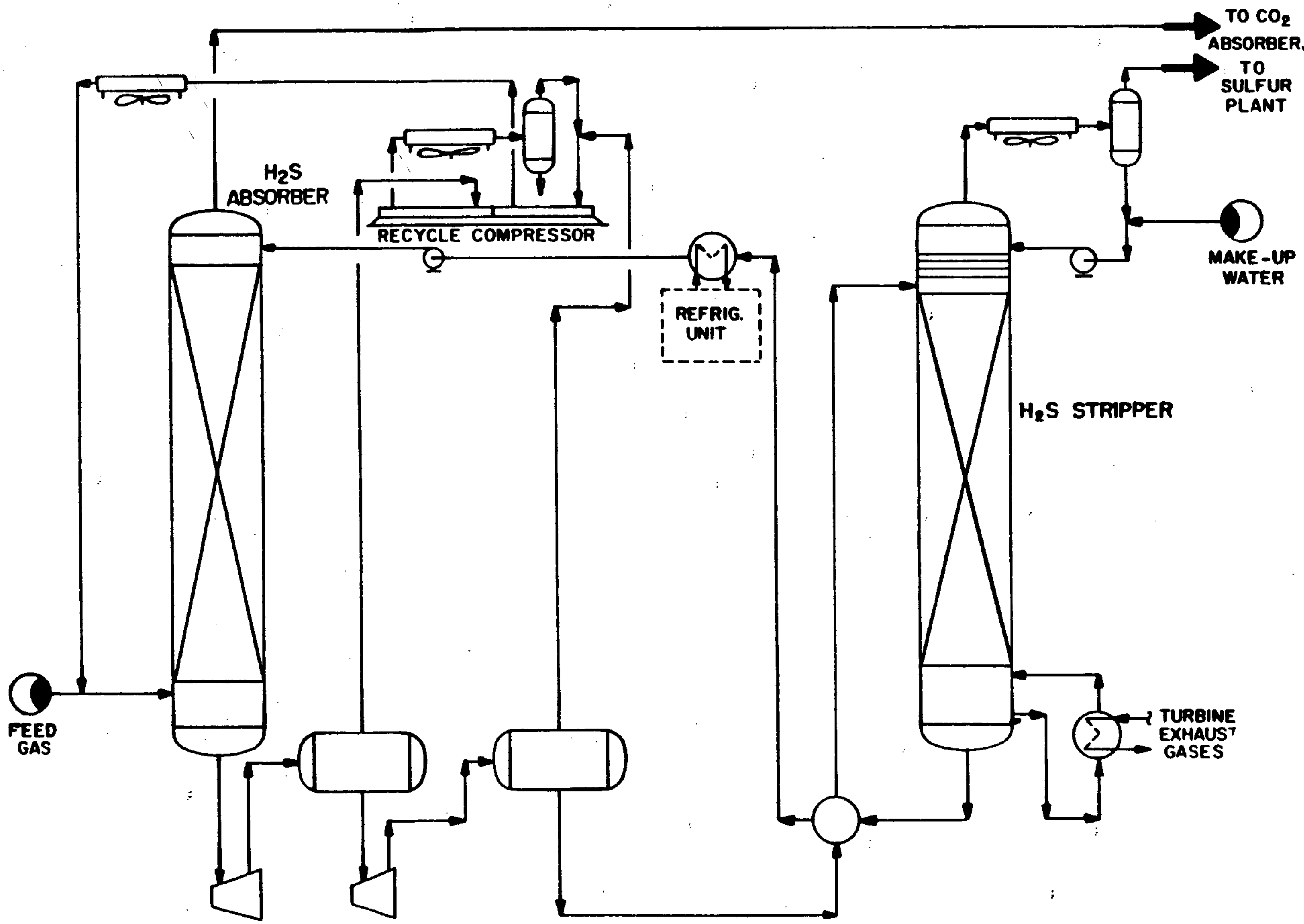
Compression and Drying

Carbon dioxide is delivered to the compression step at about 5 psig and 110^{o}F saturated with water vapor. It is compressed in a three case centrifugal compressor powered by a steam turbine to a final discharge pressure of 2,000 psig for delivery to the gas pipeline. As the gas compresses and cools between stages, water condenses and is removed in knockout vessels. Between the second and third case the gas is dried to final pipeline quality by a TEG dryer.

Hydrogen Sulfide Removal

The first stage of purification for sources containing H_2S is the selective separation and removal of the H_2S. The majority of the H_2S with a portion of the CO_2 separates in the first of two Selexol units and the final removal of the H_2S from this associated CO_2 is in the Stretford unit where the H_2S is converted to sulfur. The bulk of the CO_2 is then removed in the second Sel[illegible]

The feed gas containing CO_2, CH_4, H_2S and saturated with water vapor at 75^{o}F enters the conventional, counter-current, packed absorption tower



PROCESS FLOW DIAGRAM FOR H_2S SEPARATION UNIT

FIGURE 6.10

in the Selexol unit. A Selexol solution enters the top of the tower and absorbs essentially all of the H_2S present. The overhead gas from this tower, containing mostly CO_2 and CH_4, less than 1/4 grain of H_2S per 100 SCF of gas and very little water vapor, flows to the bulk CO_2 removal section.

The CO_2 and, to a lesser extent, CH_4 are also soluble in Selexol and tend to be absorbed with the H_2S. However, the H_2S is by far the more soluble component in the feed gas, about seven times more soluble than CO_2. The absorber is designed so the H_2S acts as the controlling component in the absorption. The Selexol solvent circulation rate is limited to the absolute minimum required to remove the H_2S down to the desired level. Restricting the circulation rate in this manner will serve to minimize the quantities of other gases absorbed.

The H_2S content in the feed gas is only 2% and after absorption, the H_2S comprises less than the needed 25% of the dissolved gases in the loaded solution. The 25% or greater H_2S content is necessary for the Stretford units performance. To increase the H_2S level, the loaded solution flashes across a hydraulic turbine to a recycle flash drum. The hydraulic turbine is used for energy recovery and supplies power to the lean solution recirculation pump.

Because of their lower solubilities the CO_2 and CH_4 will preferentially desorb from the solution in the recycle flash drum. These flashed vapors are recompressed, air-cooled, and then recycled back to the H_2S absorber. This recycle drum serves to further enrich the final H_2S stream. The pressure level in the recycle drum is set so that the dissolved gases remaining in the flashed liquid stream will contain over 25% H_2S. At this point the H_2S can be stripped from the solution and the sulfur recovered in a sulfur plant.

With the low leakages specified in the absorber overhead, a solvent very lean in H_2S is required at the absorber top. The high solubility of H_2S in the solvent prevents obtaining this degree of leanness by simple flashing; therefore, steam stripping is required to completely regenerate the solvent.

The liquid stream from the recycle drum is heated by cross-exchange with the stripper bottoms and then flashed to the top of the stripper. The H_2S and CO_2 remaining in solution are stripped out by steam as the liquid descends through the tower. The stripping steam results from the boiling of the water contained in the Selexol solution. The waste heat recovered from the gas turbine drivers on the recycle and product compressors furnishes the required reboiler duty.

The lean solution exiting the stripper bottom cools by heat exchange with the incoming stripper liquid and further by exchange with refrigerant from the ammonia-water absorption package in the CO_2 recovery unit. This final exchange maintains the overall heat balance. The lean solution then is pumped back to the top of the absorber.

The stripper overhead gas is cooled by heat exchange with air in an overhead condenser. The condensed steam is separated from the gas in a knockout drum and, together with makeup water, returned to the stripper top. It is necessary to maintain a 95 wt. percent Selexol - 5 wt. percent water solvent solution in the process to enable steam generation in the stripper bottom. At water contents lower than 5 percent, temperature limitations would prohibit the use of the gas turbines waste heat in the reboiler or provide insufficient stripping. Higher water contents would add significantly to the energy requirements and equipment sizes with no corresponding increase in absorption capacity or selectivity.

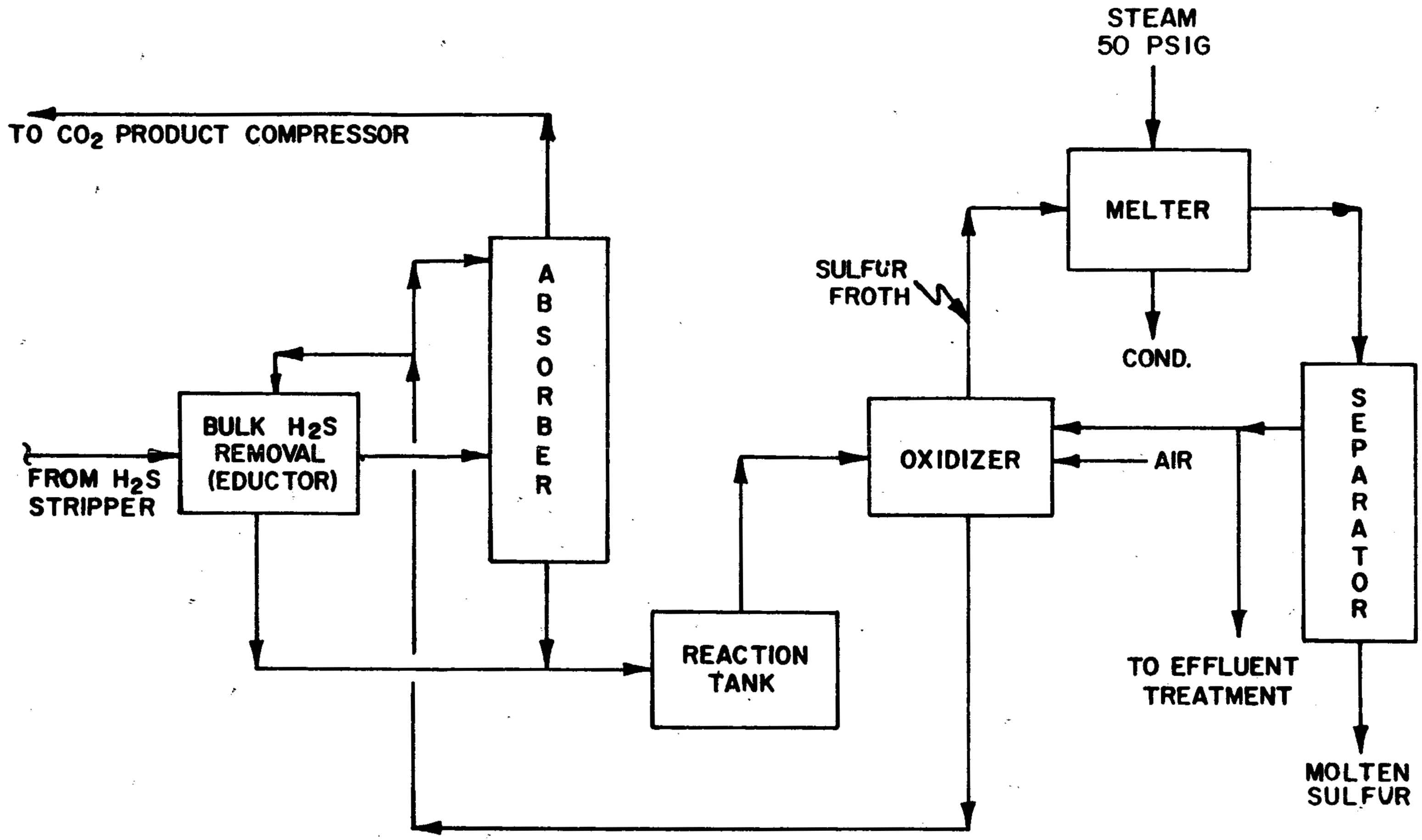
The stripper overhead gas exiting the knockout drum is fed to the Stretford sulfur plant for processing. The CO_2 stream from the sulfur plant is essentially pure CO_2 (only 1/4 gr H_2S/100 SCF gas). This effluent gas is combined with the product streams from the bulk CO_2 removal section and sent to the CO_2 product compressor for compression and drying.

Sulfur Recovery (Stretford Process)

Figure 6.11 presents a block flow diagram for the Stretford process described below.

In the Stretford process H_2S is converted to elemental sulfur (S) according to the following overall equation:

$$2H_2S + O_2 = 2H_2O + 2S$$



BLOCK FLOW DIAGRAM OF THE STRETFORD PROCESS

FIGURE 6.11

The reaction is initiated in an absorber. Here an alkaline solution of sodium carbonate (NA_2CO_3) containing a vanadium salt (V^{+5}) along with anthraquinone - disulfonic acid (ADA), absorbs the H_2S gas and converts it to hydrosulfide ion:

$$Na_2CO_3 + H_2S = NaHS + NaHCO_3$$

The solution then flows to a reaction vessel where elemental sulfur is formed,

$$HS- + 2V^{+5} = 2V^{+4} + S + H^+$$

In the third state of the process, an oxidation step employing air is used to regenerate the solution, and the S is removed as a floating froth:

$$2V^{+4} + ADA\ (oxidized) = 2V^{+5} + ADA\ (reduced)$$

The regenerated solution is then returned to the absorber tower completing the absorption step.

It is usually more economical when treating large gas quantities, as in the present case, to remove the bulk of the sulfur in a pretreatment step using liquid eductors. Here the inlet gas is brought into intimate contact with about 80% of the solution flow. Liquid from the eductors flows to the reaction vessel where the reaction to elemental sulfur takes place. The gas is further contacted in an absorber to remove the H_2S to a level of less than 1/4 gr H_2S/100 SCF.

The sulfur froth typically contains about 7-10% S. A variety of methods are available to recover the sulfur from the froth. These include filtrating, centrifuging and more recently, the use of a sulfur melter.

In the sulfur melter method, the sulfur froth is heated and the molten sulfur separates by gravity from the Stretford solution which is returned to the process. It is possible to recover 99.5% pure S by this technique. The utilities and costs for the Stretford process in this study assume that a sulfur melter technique is to be used. Assuming a 2% H_2S content in the entering feed gas, approximately 0.8 short tons of S per MM SCF of feed gas are recovered.

Utility Systems

Utility costs used in Section 6.3.4 are based on using gas turbine drivers for the CO_2 product and recycle compressors in the cases utilizing the Selexol process and a steam turbine driver for the CO_2 product compressor in the alternate MEA process case. Electric motors drive all remaining rotating equipment.

In the Selexol process cases, the waste heat from the gas turbine exhaust is recovered for use by the NH_3-H_2O absorption refrigeration system and in the regeneration of the glycol solution in the TEG drying system.

For the alternate MEA process, steam generating facilities are included in the investment cost. These facilities provide steam at 900 psig and 900°F to the CO_2 product compressor turbine. A second level of 50 psig steam supplies heat to the MEA regenerator reboiler and is provided by extraction from the turbine. The 50 psig steam also supplies heat to the glycol reboiler in the TEG drying system. The air coolers are designed to use 90°F ambient air.

Environmental Considerations

Effluent streams will be produced during normal operation of the recovery and gas transmission facilities. The following is a brief discussion of the expected effluents, and possible methods of handling these. Costs for any treatment facilities which may be required are not included in plant investment.

Water is produced from the product gas as it is compressed and cooled between stages. Also water may condense from the inlet raw gas feed. If a H_2S removal unit is not necessary, then the excess water stream is effluent and must be disposed of. The effluent water stream contains traces of hydrocarbons which must be removed before disposal.

If a H_2S removal system is necessary, then the water balance of the system is such that a slight excess of water above that obtained from these sources will be required as makeup to the H_2S stripper. Therefore, the water streams may recycle back to the process to eliminate the effluents and to reduce the makeup water requirements.

Regeneration of the glycol solution in the TEG dryer will produce a gaseous effluent consisting mostly of water vapor plus traces of TEG. This effluent stream can be vented to the atmosphere.

In the Stretford units, the formation of byproduct salts such as sodium thiosulfate and sodium sulfate require a purge stream of Stretford solution from the sulfur recovery plant. Usually this can be avoided during the first six months to one year of operation, but then must be done on a routine basis. In this manner these salts are prevented from building up in the process and plugging equipment such as the absorber.

Treatment of this purge stream will almost certainly be required from an environmental and economic standpoint, and a number of different processes are available to do this, each in various stages of development. Among these are processes offered by Peabody Engineering (Holmes Stretford), and Sun Oil (J. F. Pritchard Co. Licensee).

Methane gas is a byproduct of these cases. This will be highly marketable and is assumed to be transmitted in the usual manner for natural gas; therefore, not posing a problem.

6.3.4 Purification Costs

As the previous sections discussed, natural sources are available at a wide range of pressures and qualities. Because of the different feed conditions, several processes should be considered. Initially, this discussion will exam natural sources containing only CO_2, CH_4 and H_2O with concentrations of 50% and 90% CO_2 for various CO_2 production rates with emphasis on feed gas pressures of 250, 500 and 1,000 psig. The Selexol process was chosen for the purification of these sources. As well, a case of 10% CO_2 at 1,000 psig using Selexol for various CO_2 production rates and a case with MEA for these inlet conditions at a CO_2 production rate of 125 MMSCFD will be examined.

Also the costs for purification of natural sources containing 25% CO_2 (75% CH_4) and 75% CO_2 (25% CH_4) by Selexol process are presented for a CO_2 production rate of 125 MMSCFD and various feed gas pressures.

Natural sources containing 2% H_2S are also examined for cases of 50% CO_2 (48% CH_4) and 90% CO_2 (8% CH_4) at inlet pressures of 250 and 1,000 psig and at various CO_2 production rates.

For all cases discussed, the investment cost, annual operating cost and resulting price of the purified CO_2 at various DCFRR's are presented in this section.

The basis of the total investment cost is outlined in Appendix B1 and the basis for determining the annual operating expenses is outlined in Appendix B2. The basis for the calculations of the resulting price of the gas is:

1. 20 year project life (after start-up)
2. 20 year straight line depreciation
3. 100% equity capital (no debt)
4. The salvage value is zero
5. The working capital is \$16,000/MMSCFD CO_2

All cases of natural sources discussed in this section have 98% purity CH_4 as an additional product. To provide only the cost of the CO_2 purification, credit was not taken for the CH_4 produced. However, the gas consumed in the gas turbines was considered an expense. The method provides the true cost of the CO_2 purification and may be applicable to similar natural sources containing non-marketable gases such as N_2 rather than CH_4. If credit is given for the CH_4 product, then the price of the CO_2 purification is lower than presented (particularly in the 10% and 50% CO_2 cases).

The investment costs and annual operating costs for the purification of natural sources with 50% and 90% CO_2 (the remainder CH_4 and H_2O) and available at 250 psig are presented for various CO_2 production rates in Figures 6.12 and 6.13 respectively. Similarly, the costs for the same concentrations available at 500 psig are presented for various CO_2 production rates in Figures 6.14 and 6.15. The investment costs and annual operating costs for feeds with 10%, 50% and 90% CO_2 and available at 1,000 psig are shown in Figures 6.16 and 6.17.

The resulting purification costs for the sources with 50% and 90% CO_2 are illustrated for feed pressures of 250, 500 and 1,000 psig in Figures 6.18, 6.19 and 6.20 respectively. The associated purification costs for the sources of 10% CO_2 at 1,000 psig are presented in Figure 6.21.

The Figures 6.18 - 6.21 show that depending on the amount of CO_2 in the feed, the amount of CO_2 produced, the feed gas pressure and DCFRR used, the price for the purification ranges from $0.22 - 1.85/MSCF.

Since the partial pressure of the CO_2 for the 10% CO_2 content case is low (see comments made in Section 6.3.2), an alternate MEA case for 10% CO_2 at 1,000 psig for the production of 125 MMSCFD of CO_2 was designed and estimated. The investment costs for this unit are approximately $75,319,000 and the annual operating expenses are approximately $26,744,000. The investment costs and annual operating costs include the necessary steam generating facilities and boiler feedwater treatment. The table below, Table 6.1, lists the comparative purification costs for the MEA unit and a similar Selexol unit.

TABLE 6.1

COMPARISON OF PURIFICATION COSTS FOR A SELEXOL UNIT AND A MEA UNIT PURIFYING FEED GAS AT 10% CO_2 AND 1,000 PSIG TO PRODUCE 125 MMSCFD CO_2

	Purification Costs ($/MSCF)	
DCFRR	SELEXOL	MEA
10%	1.01	1.00
15%	1.25	1.19

TABLE 6.1 (contd.)

DCFRR	SELEXOL	MEA
20%	1.51	1.41
25%	1.81	1.65

To illustrate the effects of the feed gas quality and pressure on the investment costs and annual operating costs of the Selexol process units for a specified CO_2 production rate of 125 MMSCFD, these costs are presented in Figures 6.22 and 6.23. The resulting purification costs ($/MSCF) for the various feed gas qualities and pressures are shown in Figure 6.24 for 10% DCFRR, in Figure 6.25 for 15% DCFRR, in Figure 6.26 for 20% DCFRR and in Figure 6.27 for 25% DCFRR.

As discussed previously, natural sources containing H_2S require additional processing to first remove the H_2S from the feed gas and then reduce it to elemental sulfur. Hence, the investment and annual operating costs for sources with H_2S contamination are higher than those sources without H_2S. The investment and annual operating costs for sources containing 2% H_2S and 50% or 90% CO_2 (the remainder is CH_4) at 250 psig are given in Figures 6.28 and 6.29 respectively. The costs do not include any treatment of the Stretford unit's purge stream or any credit for either the elemental sulfur or the CH_4 that is also produced during CO_2 purification. The resulting costs of purification of the 250 psig feed gas containing 50% CO_2 48% CH_4 and 2% H_2S are shown for various DCFRR's and CO_2 production rates in Figure 6.30. Figure 6.31 shows the purification costs associated with 90% CO_2, 250 psig source.

As presented for sources available at 250 psig, the investment and annual operating costs for sources at 1,000 psig are shown in Figures 6.32 and 6.33. The resulting purification costs for the 50% CO_2 feed at various DCFRR's are presented in Figure 6.34. Likewise, the purification costs of the 90% CO_2 sources are illustrated in Figure 6.35.

Because of the high volume of gas processed, only one Selexol case was examined. The "best case" (i.e. the economics on this case represent the best possible for the various feed source pressures studied) of a feed source pressure available at 1,000 psig and a CO_2 production rate of 125 MMSCFD were chosen for study of sources containing 10% CO_2, 88% CH_4 and 2% H_2S. The investment cost for this case is approximately $232,091,000 and the annual operating cost is about $93,317,000. The resulting purification costs are $3.31/MSCF for 10% DCFRR, $3.89/MSCF for 15% DCFRR, $4.55/MSCF for 20% DCFRR and $5.29/MSCF for 25% DCFRR.

6.3.5 Conclusions

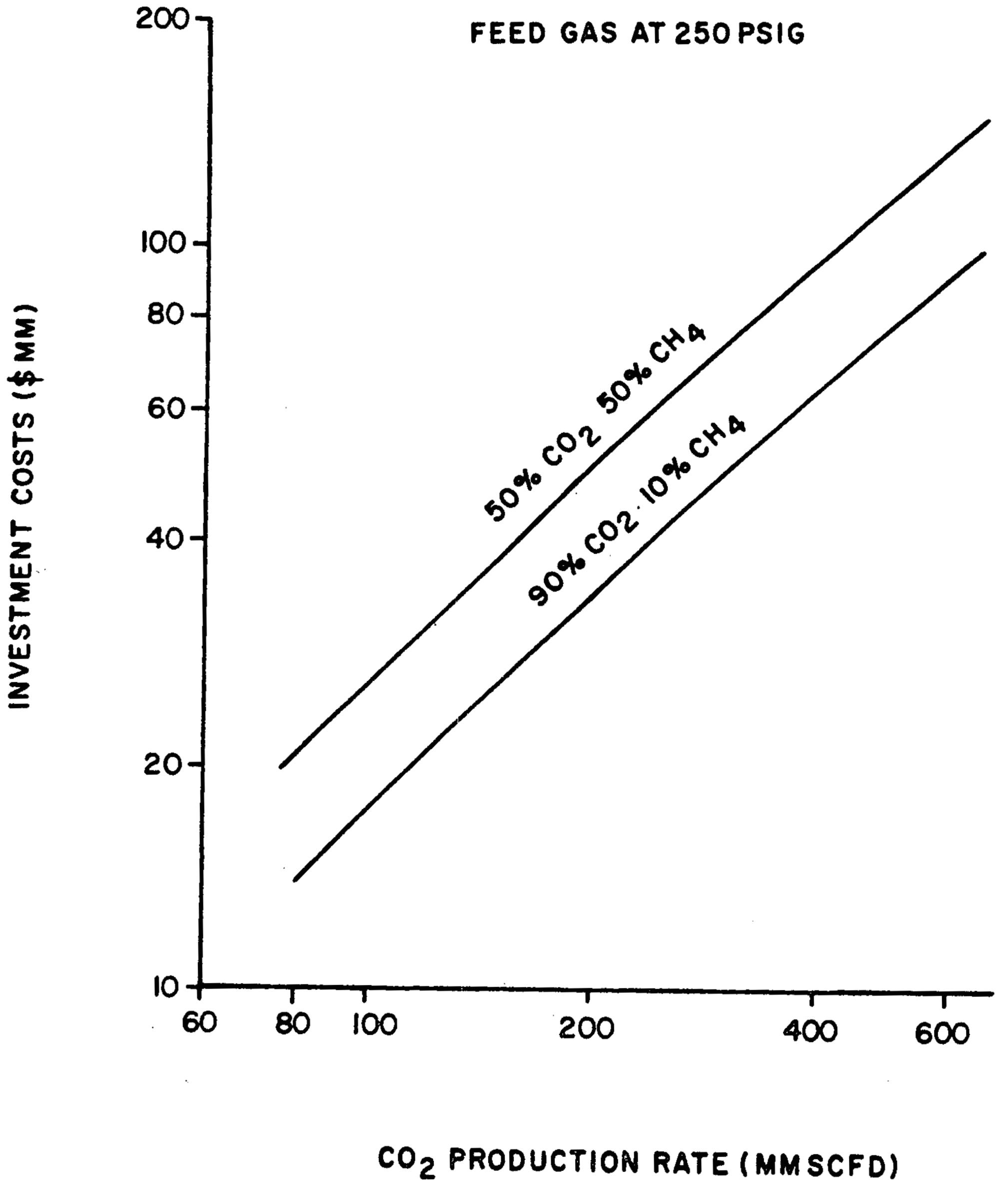
In the case of natural sources, investment cost, annual operating cost and resulting price of the purified CO_2 are dependent upon the sources' composition and available pressure, the CO_2 production rate and the DCFRR used for price determination. The figures in Section 6.3.4 show that depending upon the above mentioned variables, the purification costs may range from \$0.34/MSCF to \$5.30/MSCF. With this wide range and the number of independent variables conclusions are difficult.

For natural sources without H_2S, but with lower concentrations of CO_2 (lower than about 85%), Figures 6.24 - 6.27 show that processing at about 500 psig is the most economical pressure. The lower quality sources (less than about 85% CO_2) require a recycle compressor (see discussion in Section 6.3.3) and at absorber pressures higher than about 500 psig, the recycled gas raises the investment costs.

Purification costs of natural sources with very low CO_2 concentrations (10% CO_2) are compared previously in Table 6.1. Both the MEA process and the Selexol process are studied. The investment costs for a comparable MEA unit are significantly lower than those for a Selexol one, but the MEA unit's annual operating costs are much higher. The resulting gas purification costs for the MEA unit are only slightly less than those for the Selexol unit.

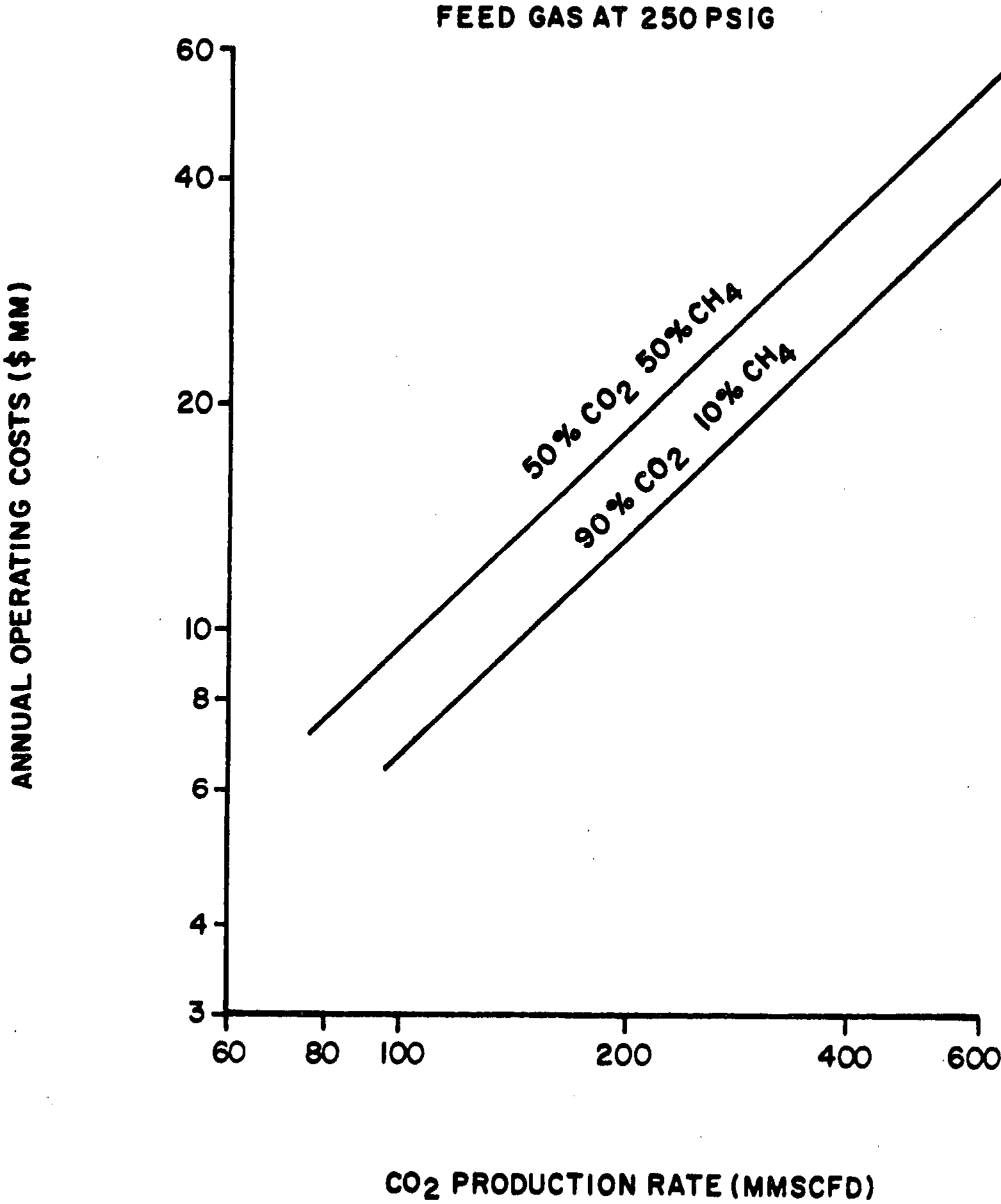
However, if the source quality is high enough, then a recycle compressor is not necessary and both the investment and purification costs decrease as the absorber pressure rises. The 90% CO_2, 10% CH_4 curves in Figures 6.24 - 6.27 illustrate this.

Natural sources containing H_2S require the extra expense for the removal and disposal of the H_2S from both the product CO_2 and CH_4 streams. The extra expense due to the presence of H_2S translates to CO_2 purification costs of 2 - 3 times those of similar sources without H_2S. If credit were given for the by-product sulfur produced during the disposal of the H_2S, the CO_2 purification costs would decrease slightly, but would still remain more than those for sources without H_2S.



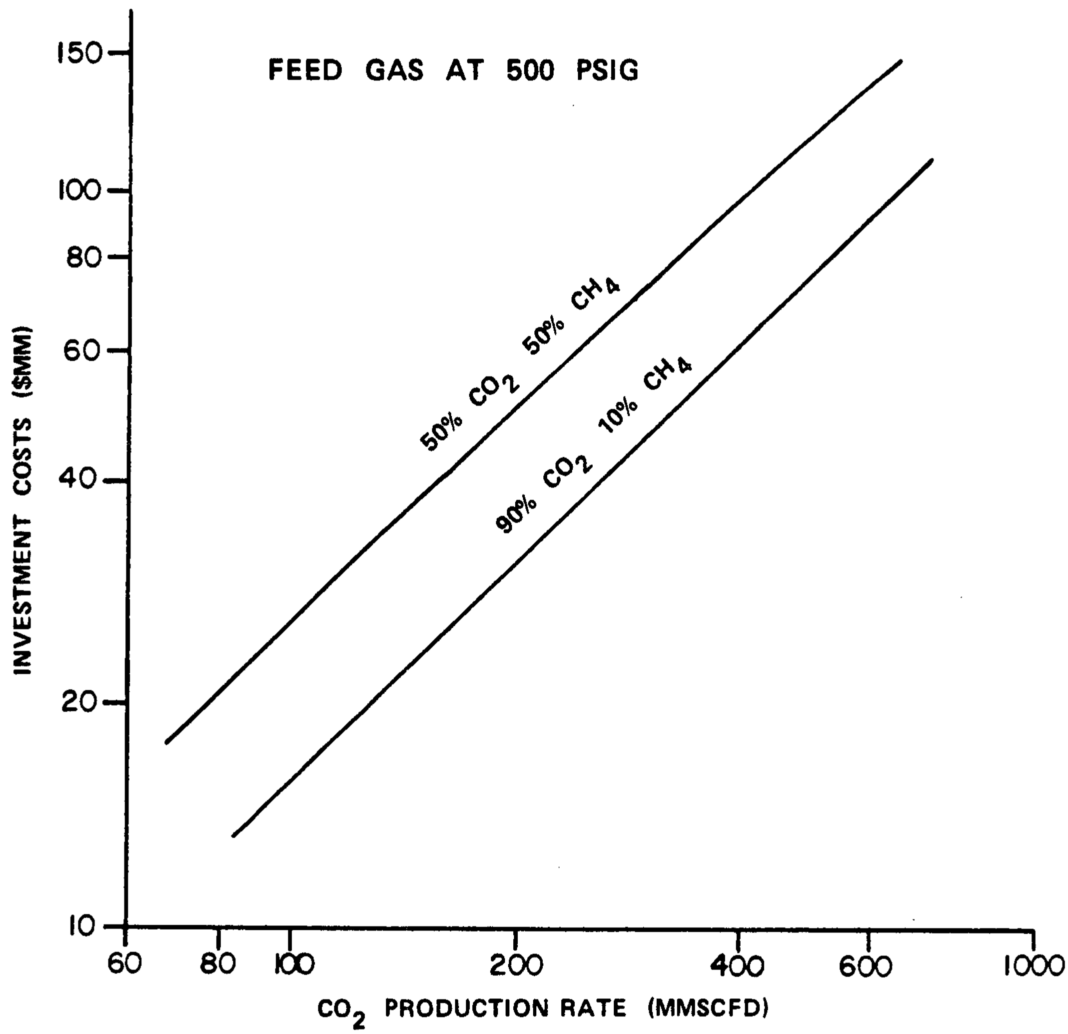
INVESTMENT COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 250 PSIG FOR FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.12



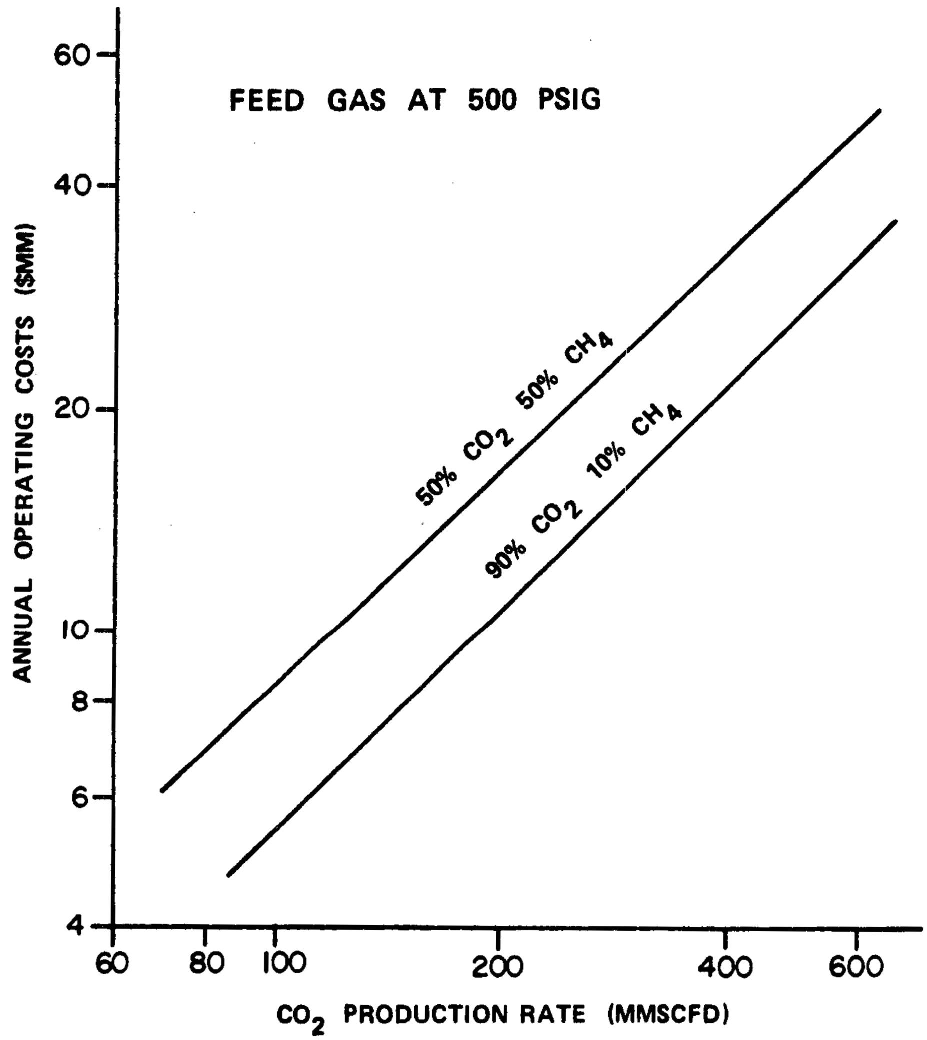
ANNUAL OPERATING COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 250 PSIG FOR FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.13



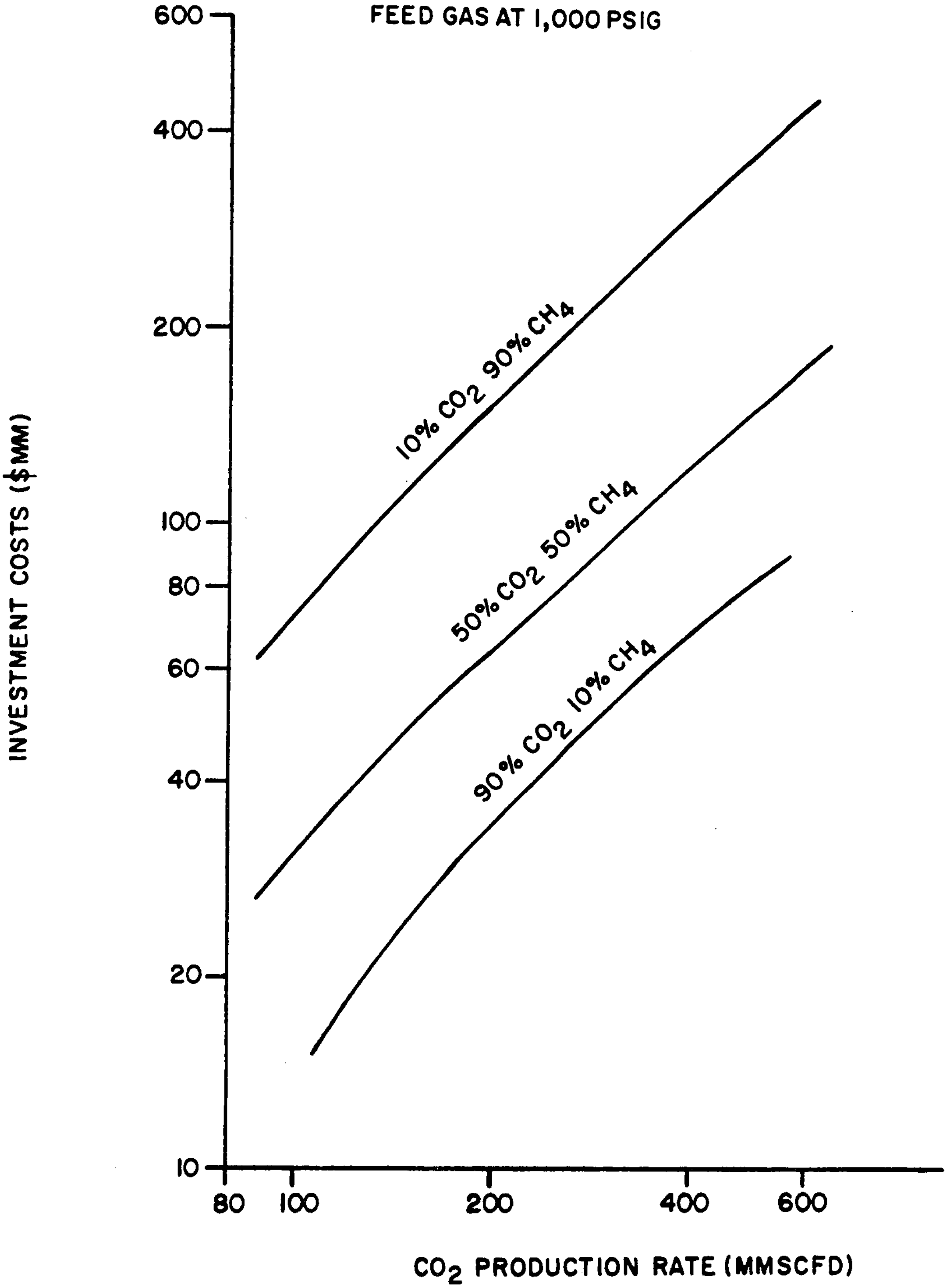
INVESTMENT COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 500 PSIG FOR FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.14



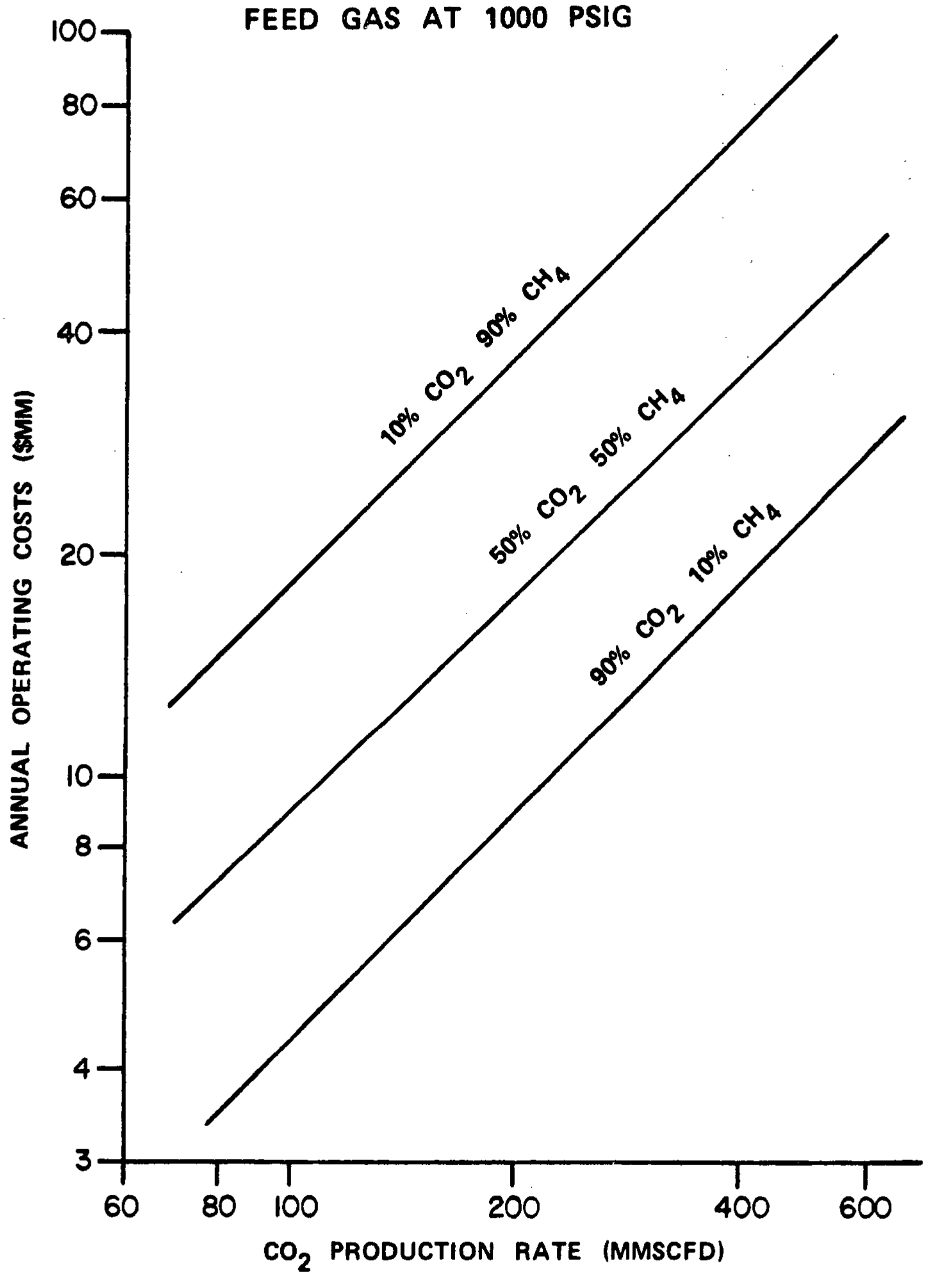
ANNUAL OPERATING COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 500 PSIG FOR FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.15



INVESTMENT COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 1,000 PSIG FOR FEED GAS OF 10%, 50%, AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

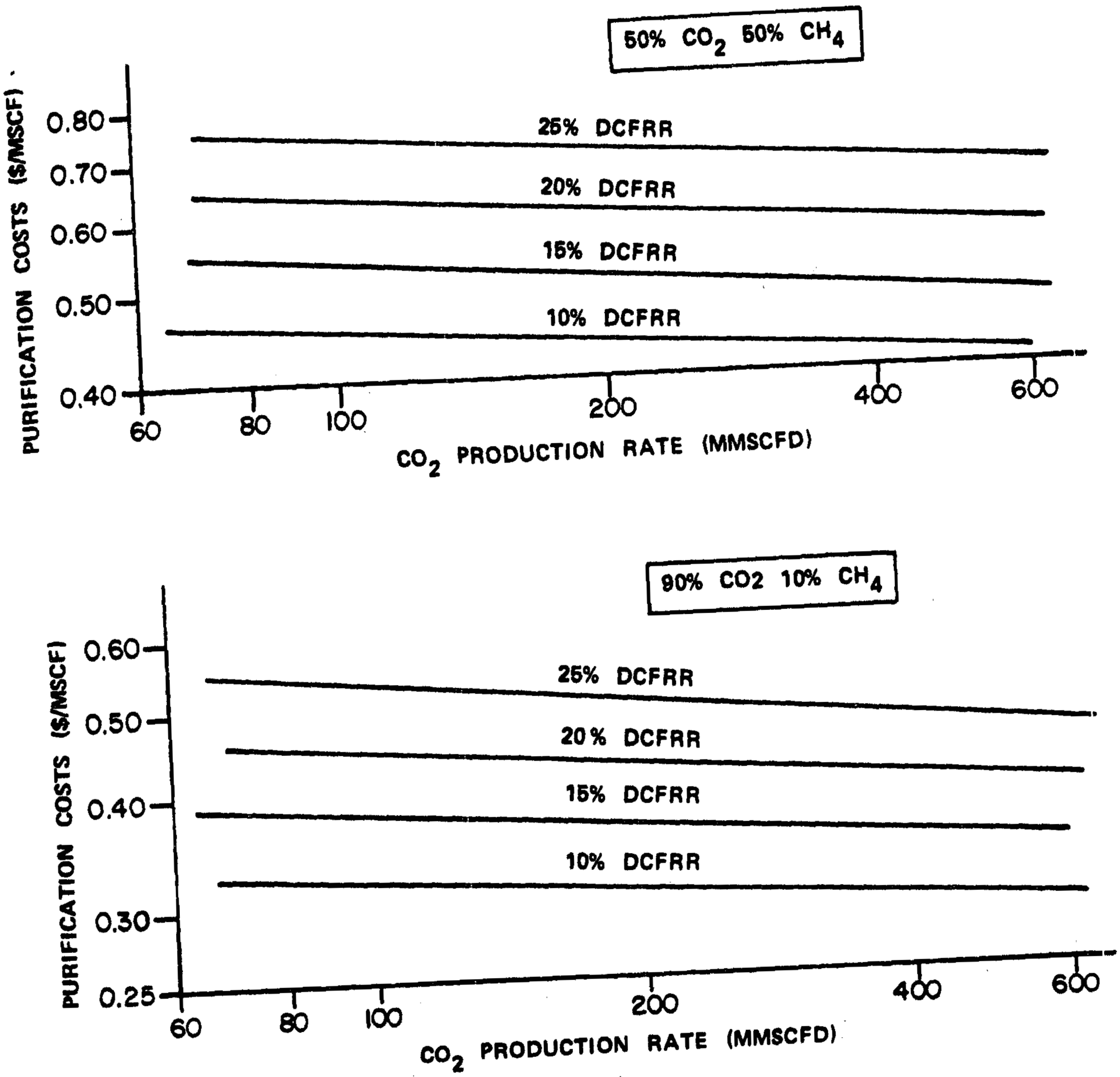
FIGURE 6.16



ANNUAL OPERATING COSTS FOR PURIFICATION FROM NATURAL SOURCES AVAILABLE AT 1,000 PSIG FOR FEED GASES OF 10%, 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.17

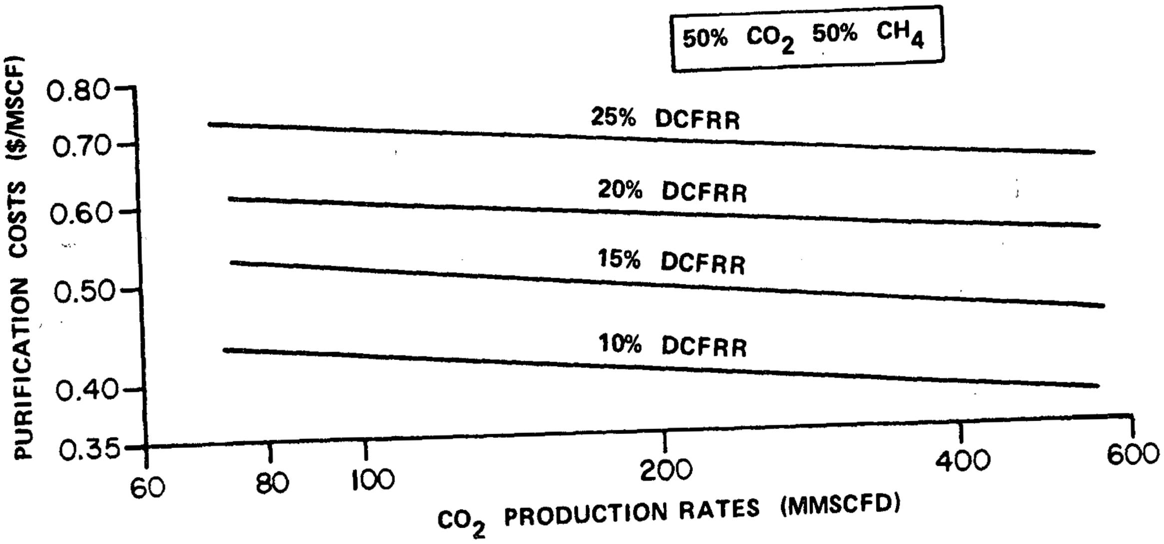
FEED GAS AT 250 PSIG

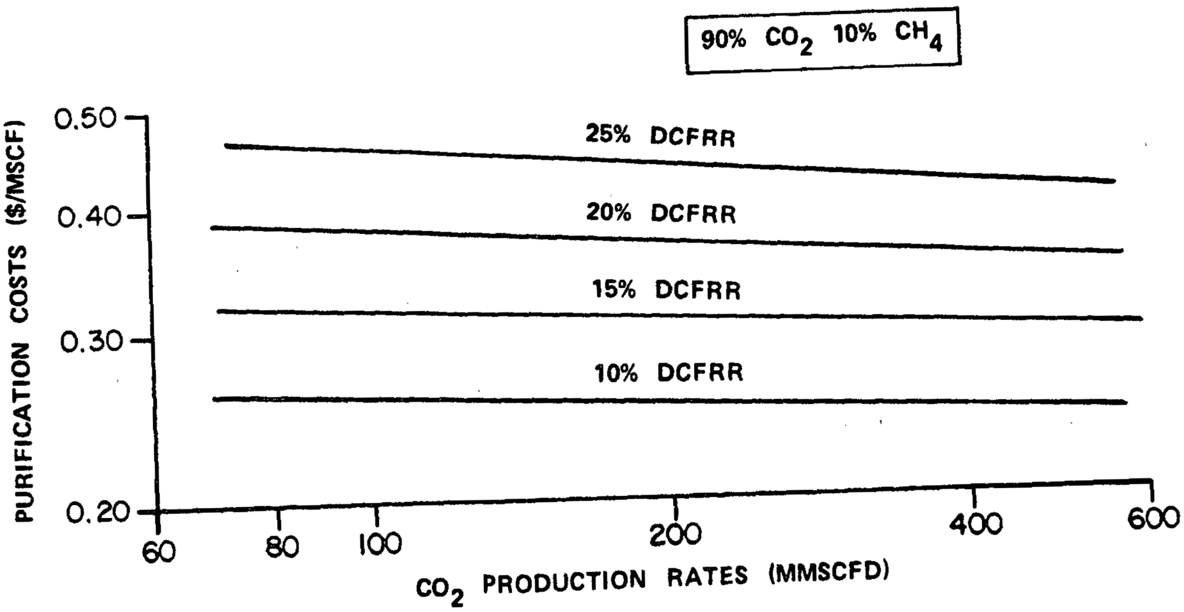


PURIFICATION COSTS FOR NATURAL SOURCES AVAILABLE AT 250 PSIG WITH FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.18

FEED GAS AT 500 PSIG

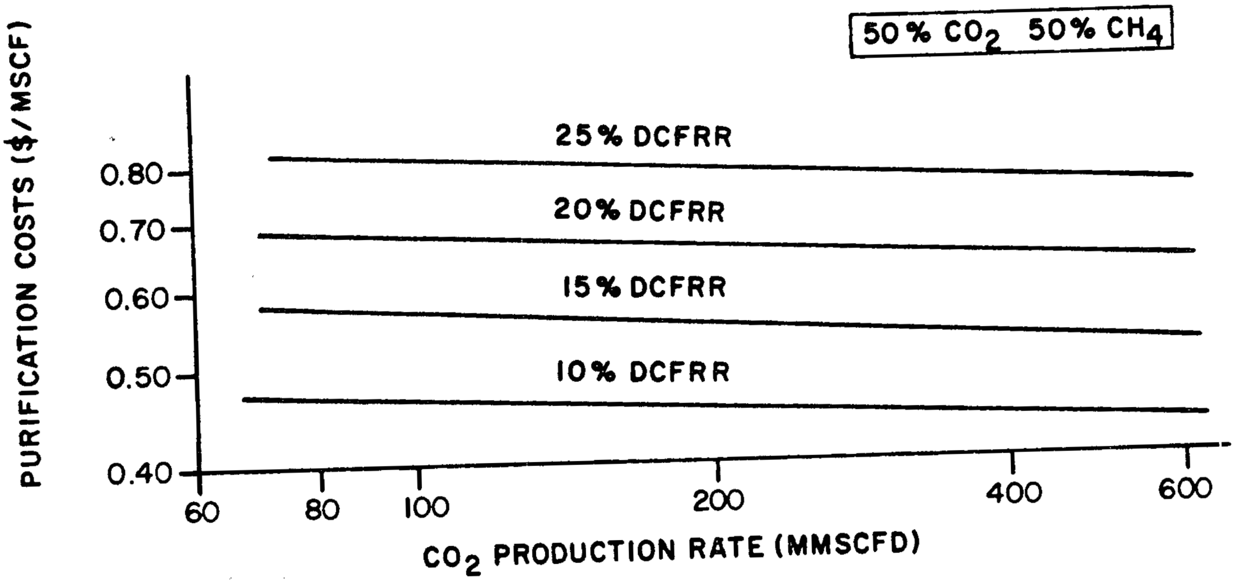


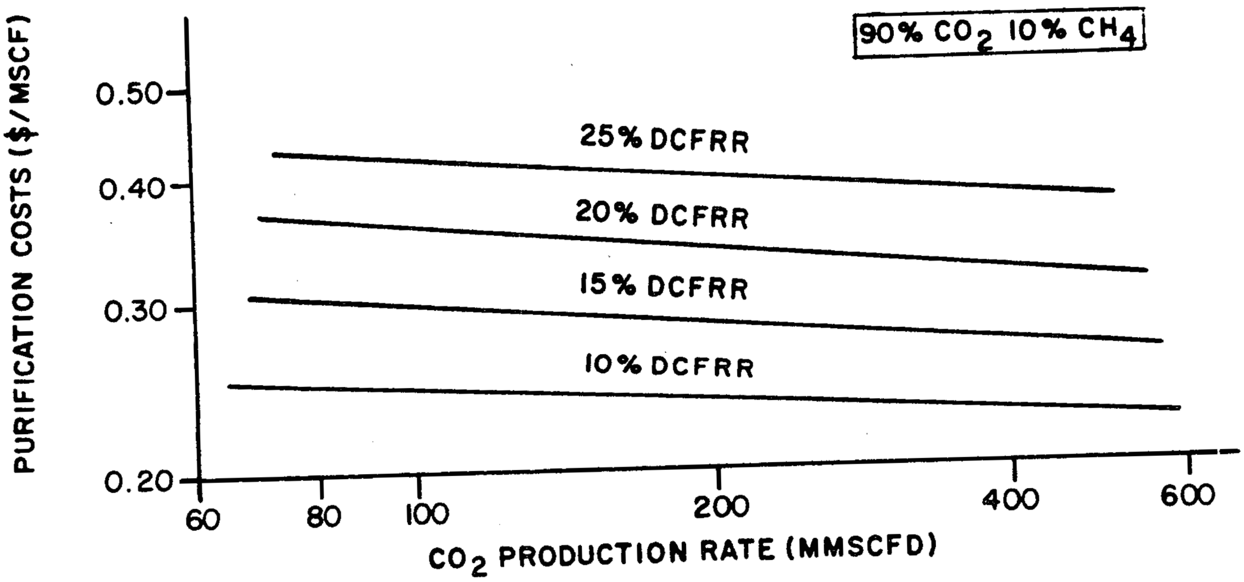


PURIFICATION COSTS FOR NATURAL SOURCES AVAILABLE AT 500 PSIG WITH FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.19

FEED GAS AT 1,000 PSIG

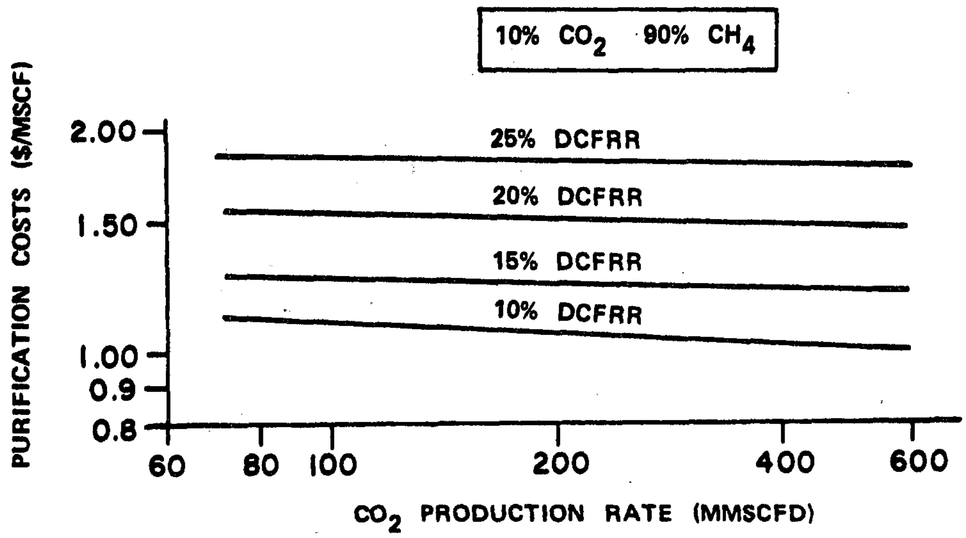




PURIFICATION COSTS FOR NATURAL SOURCES AVAILABLE AT 1,000 PSIG WITH FEED GASES OF 50% AND 90% CO_2 VERSUS VARIOUS CO_2 PRODUCTION RATES

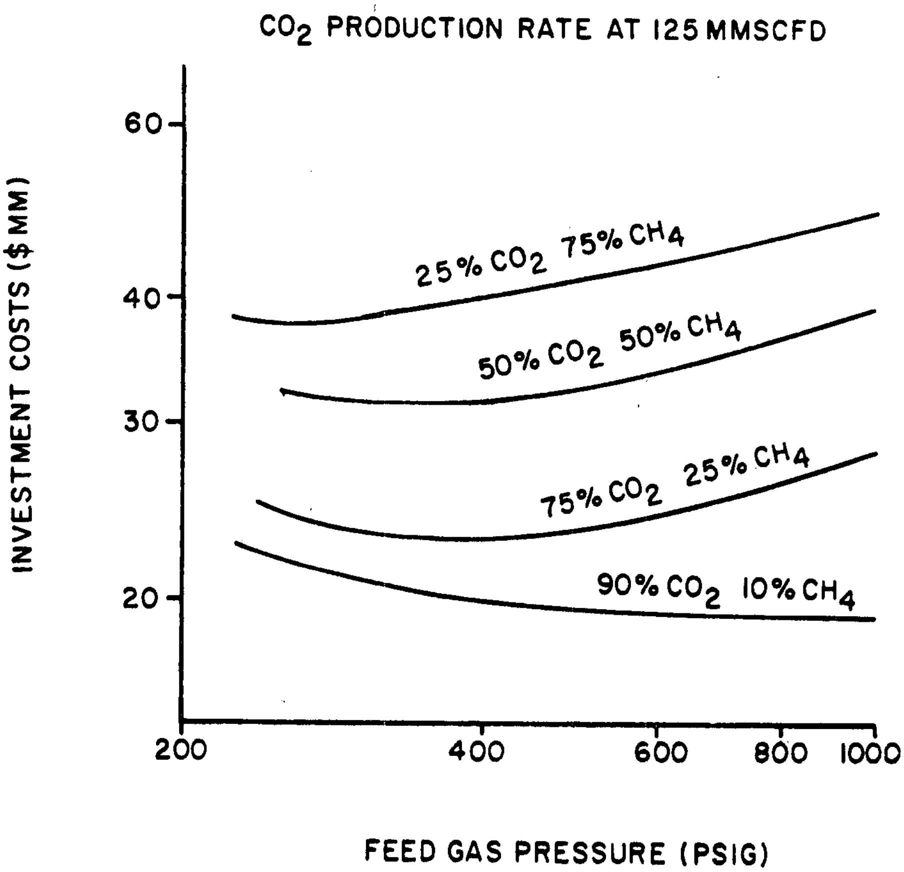
FIGURE 6.20

FEED GAS AT 1000 PSIG



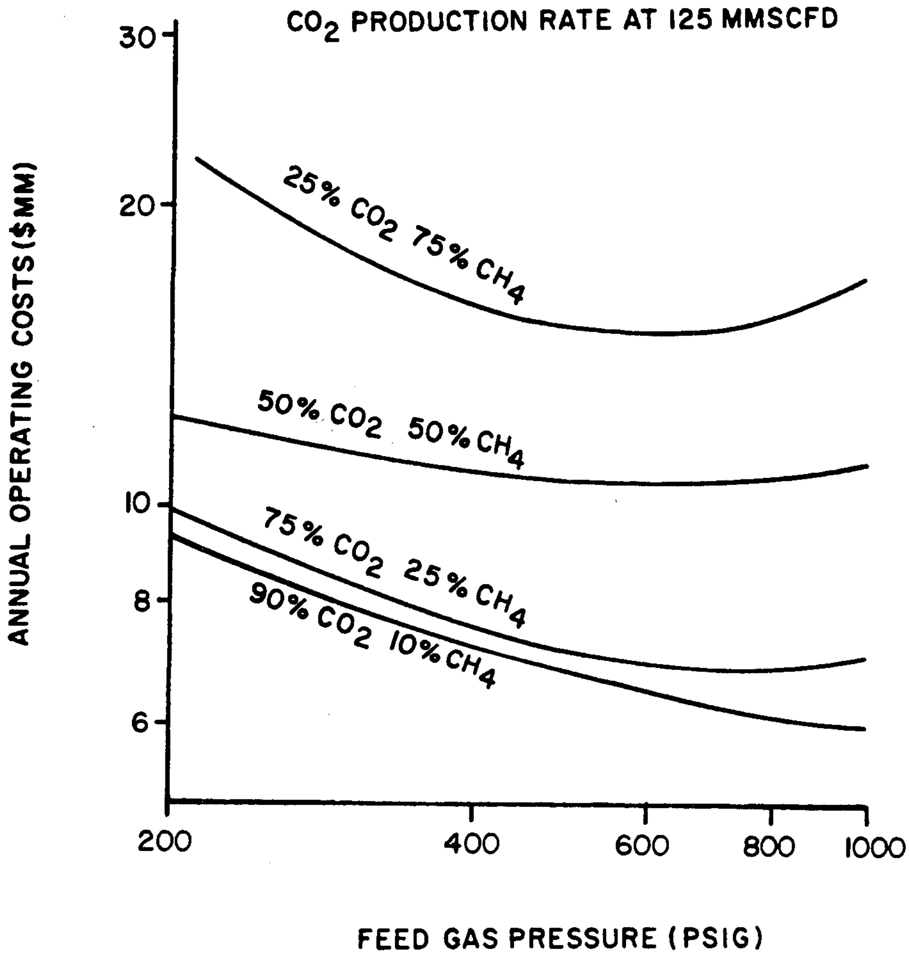
PURIFICATION COSTS FOR NATURAL SOURCES AVAILABLE AT 1,000 PSIG WITH FEED GAS OF 10% CO_2, 90% CH_4 VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.21



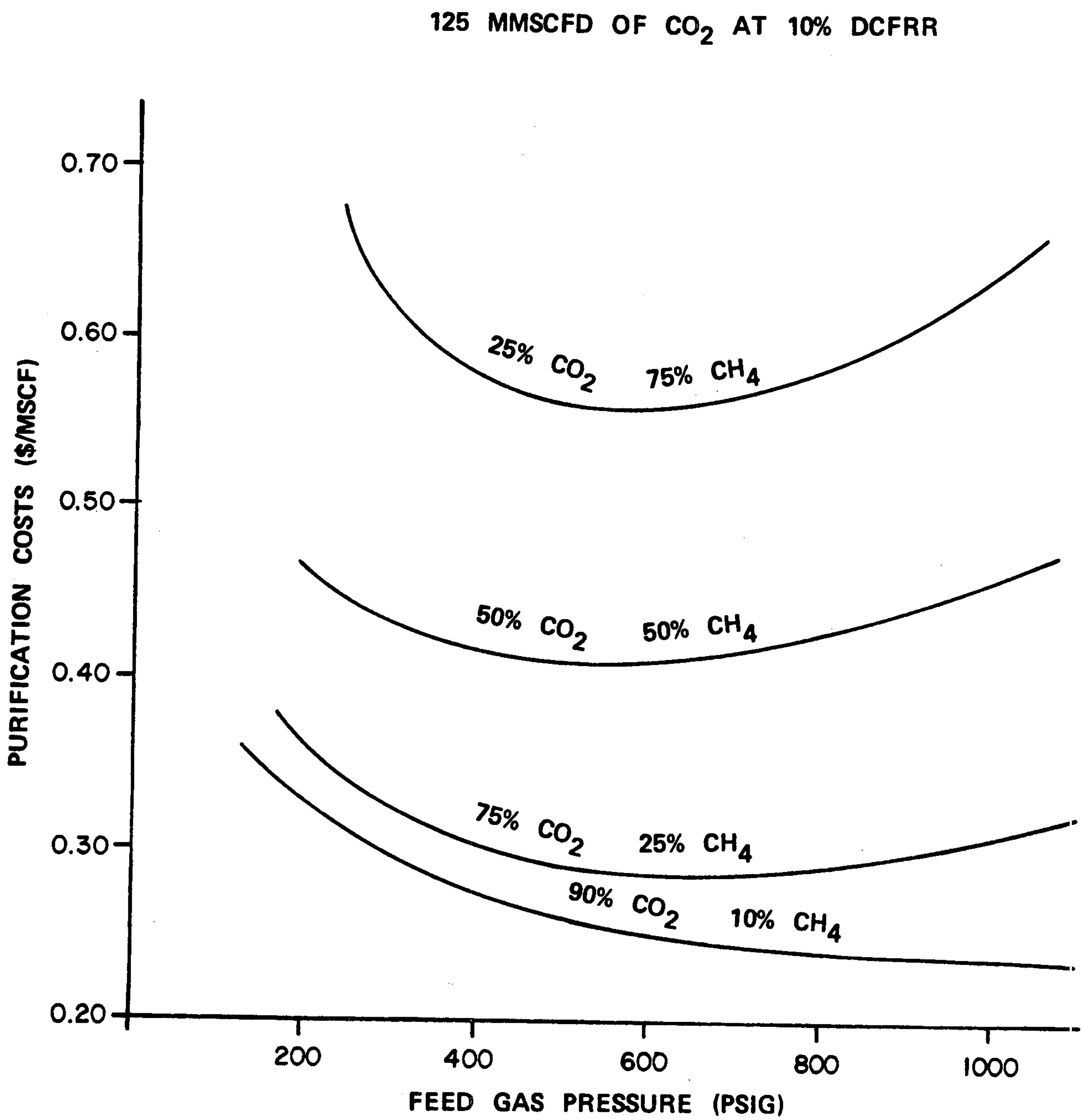
INVESTMENT COSTS FOR PURIFICATION FROM NATURAL SOURCES
AT A CO_2 PRODUCTION RATE OF 125 MMSCFD
FOR VARIOUS FEED GAS QUALITIES
VERSUS VARIOUS FEED GAS PRESSURES

FIGURE 6.22



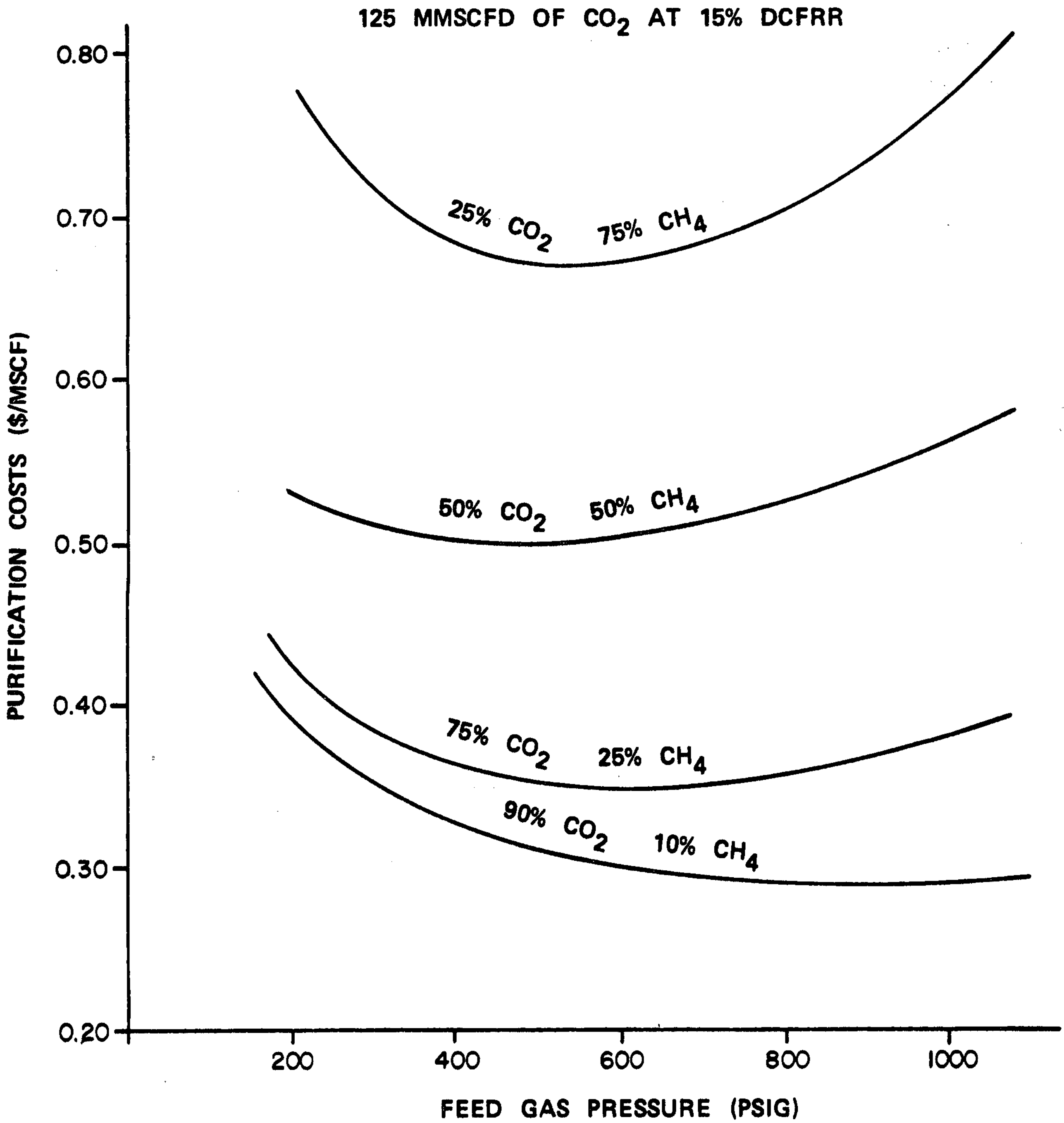
ANNUAL OPERATING COSTS FOR CO_2 PURIFICATION FROM NATURAL SOURCES AT A CO_2 PRODUCTION RATE OF 125 MMSCFD VERSUS VARIOUS FEED GAS PRESSURES

FIGURE 6.23



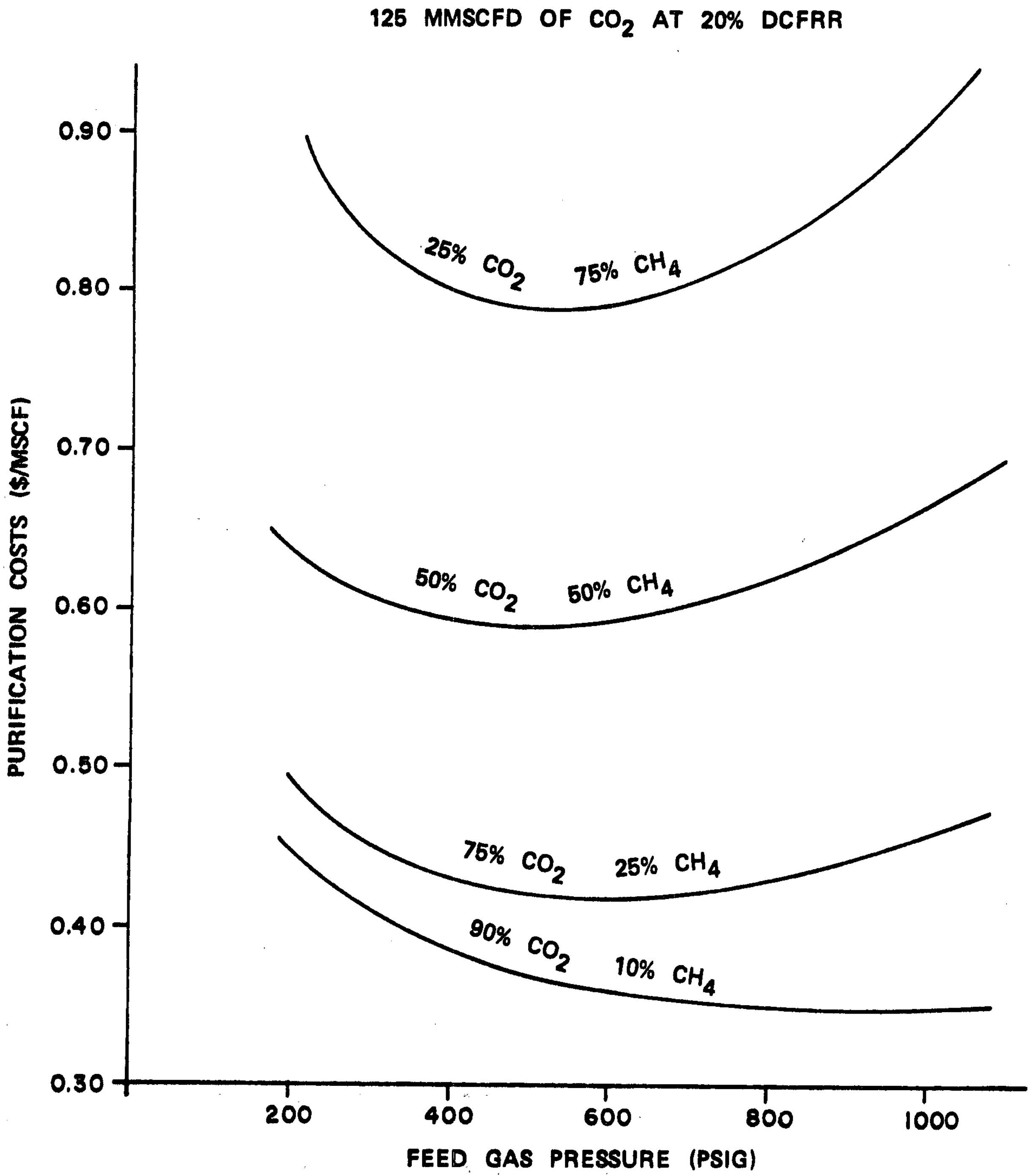
PURIFICATION COSTS AT 10% DCFRR OF VARIOUS QUALITY FEED GASES AT A CO_2 PRODUCTION RATE OF 125MMSCFD VERSUS FEED GAS PRESSURE

FIGURE 6.24



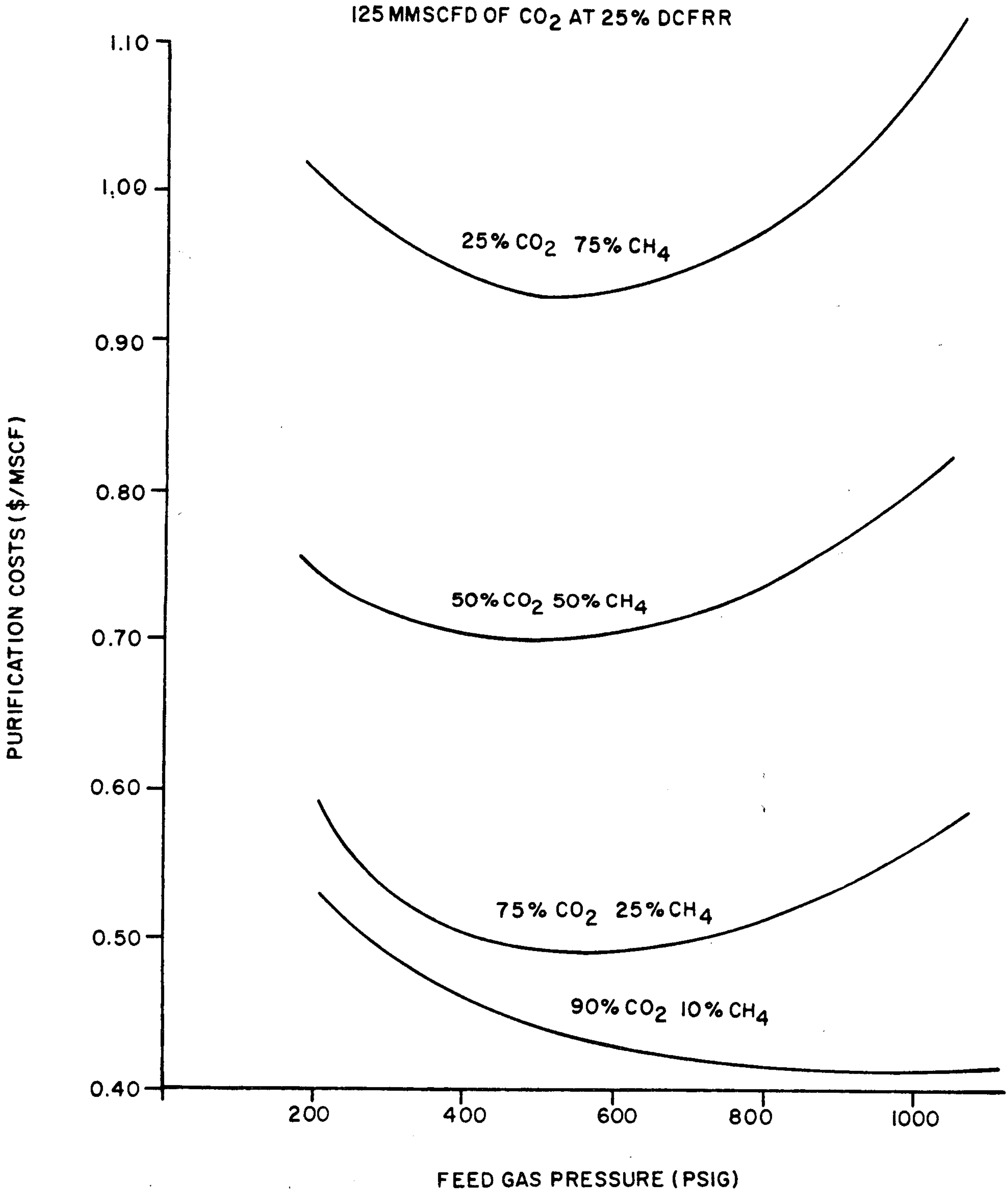
PURIFICATION COSTS AT 15% DCFRR OF VARIOUS QUALITY FEED GASES AT A CO_2 PRODUCTION RATE OF 125MMSCFD VERSUS FEED GAS PRESSURE

FIGURE 6.25



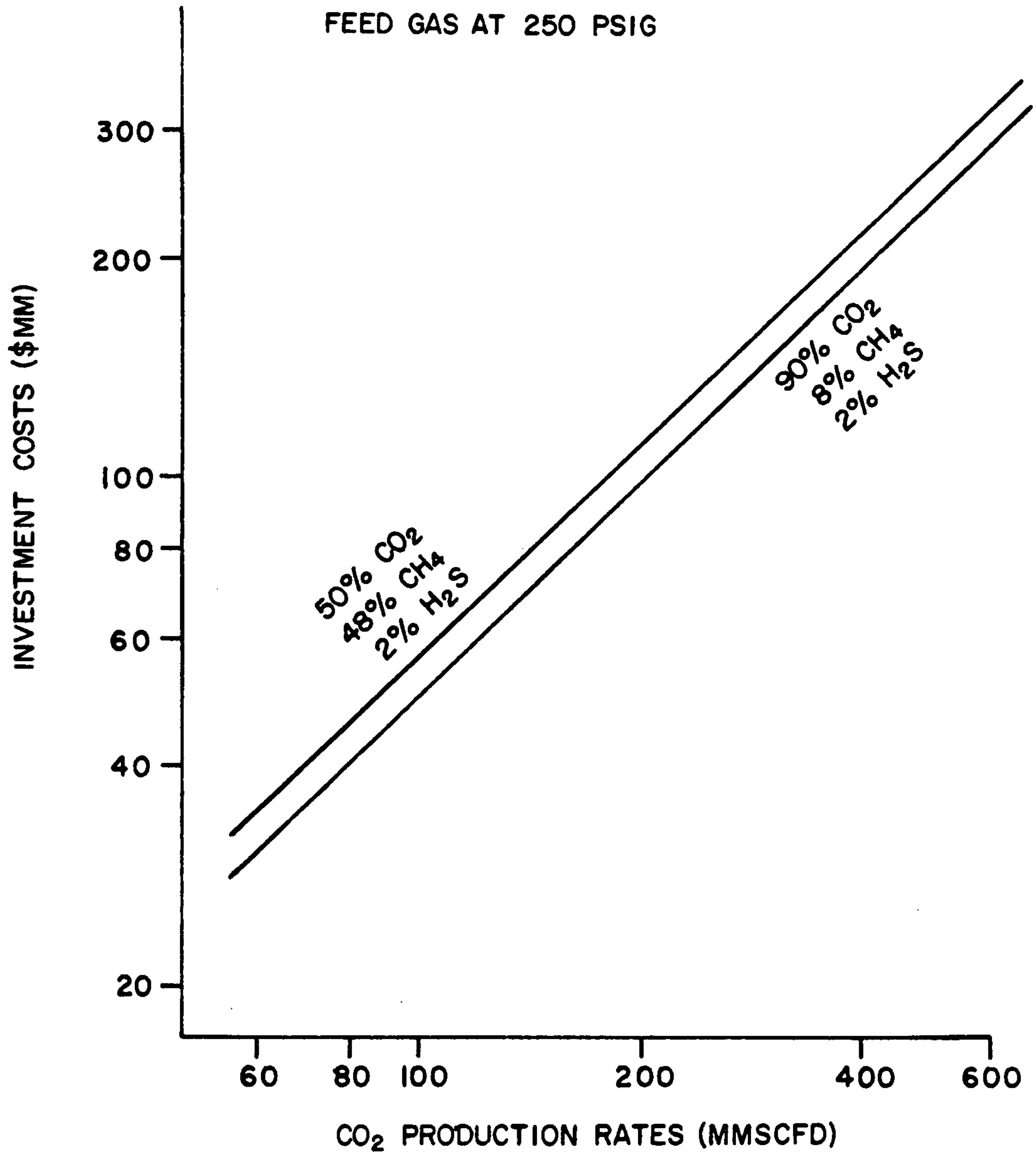
PURIFICATION COSTS AT 20% DCFRR OF VARIOUS QUALITY FEED GASES AT A CO_2 PRODUCTION RATE OF 125 MMSCFD VERSUS FEED GAS PRESSURE

FIGURE 6.26



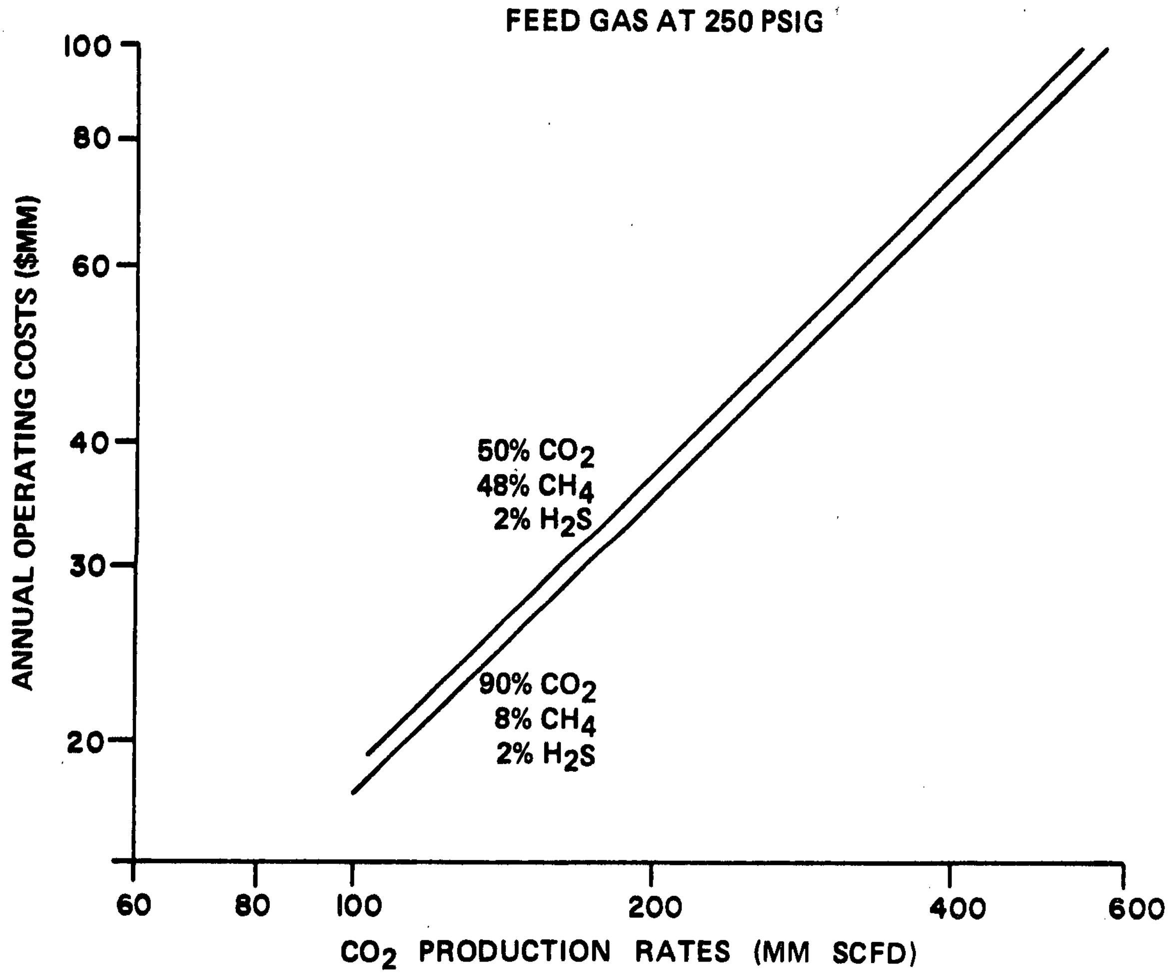
PURIFICATION COSTS AT 25% DCFRR
OF VARIOUS QUALITY FEED GASES AT A CO_2
PRODUCTION RATE
OF 125 MMSCFD VERSUS FEED GAS PRESSURE

FIGURE 6.27



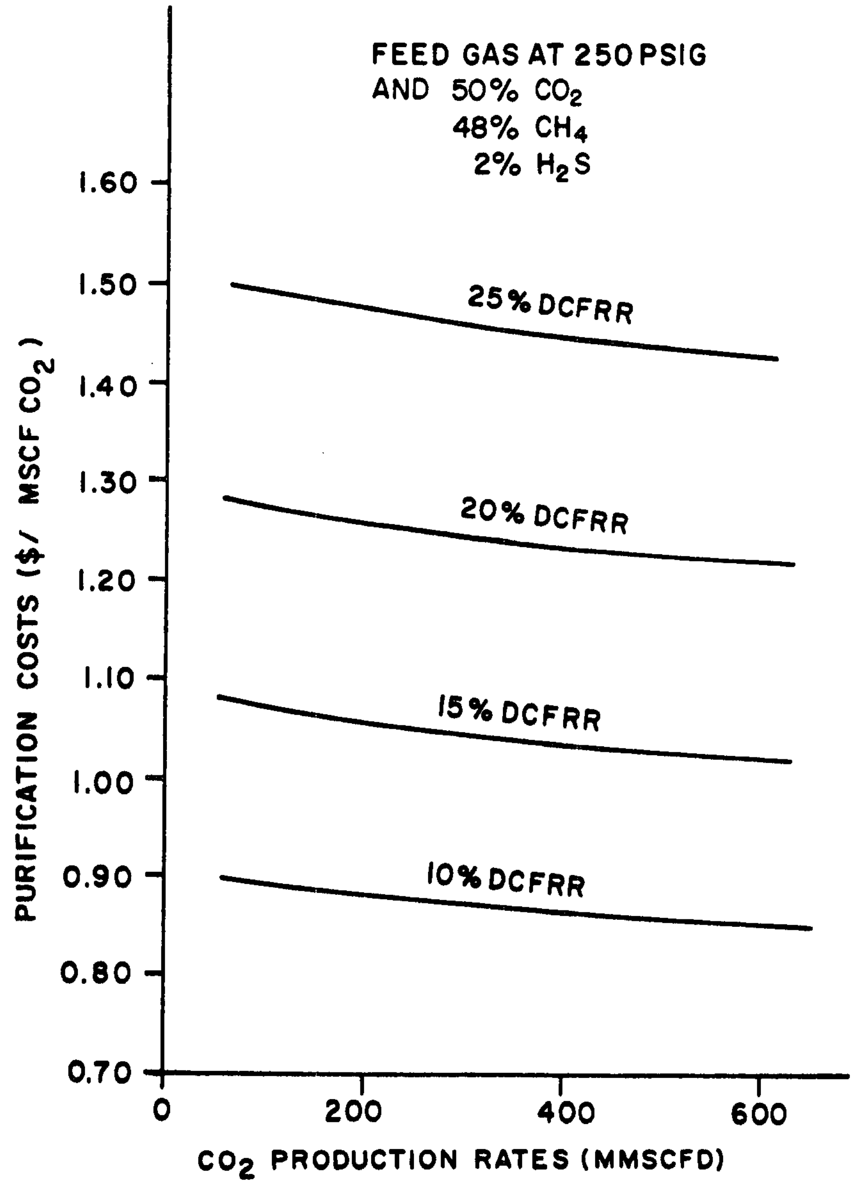
INVESTMENT COSTS FOR CO_2 PURIFICATION FROM NATURAL SOURCES CONTAINING H_2S AND AVAILABLE AT 250 PSIG VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.28



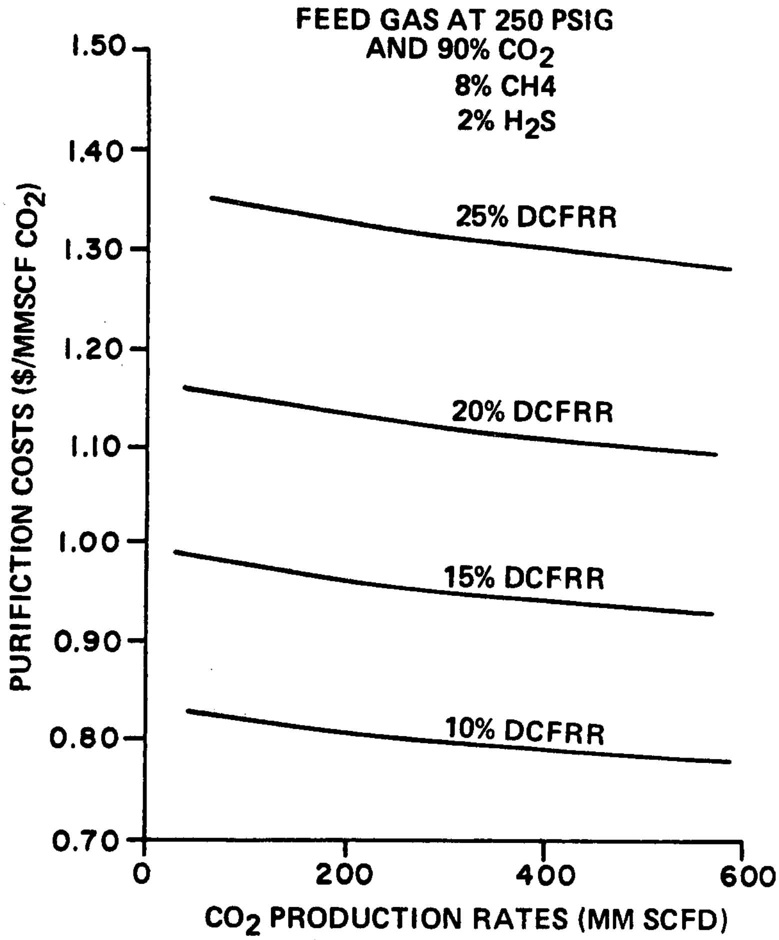
ANNUAL OPERATING COSTS FOR CO_2 PURIFICATION FROM NATURAL SOURCES CONTAINING H_2S AND AVAILABLE AT 250 PSIG VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.29



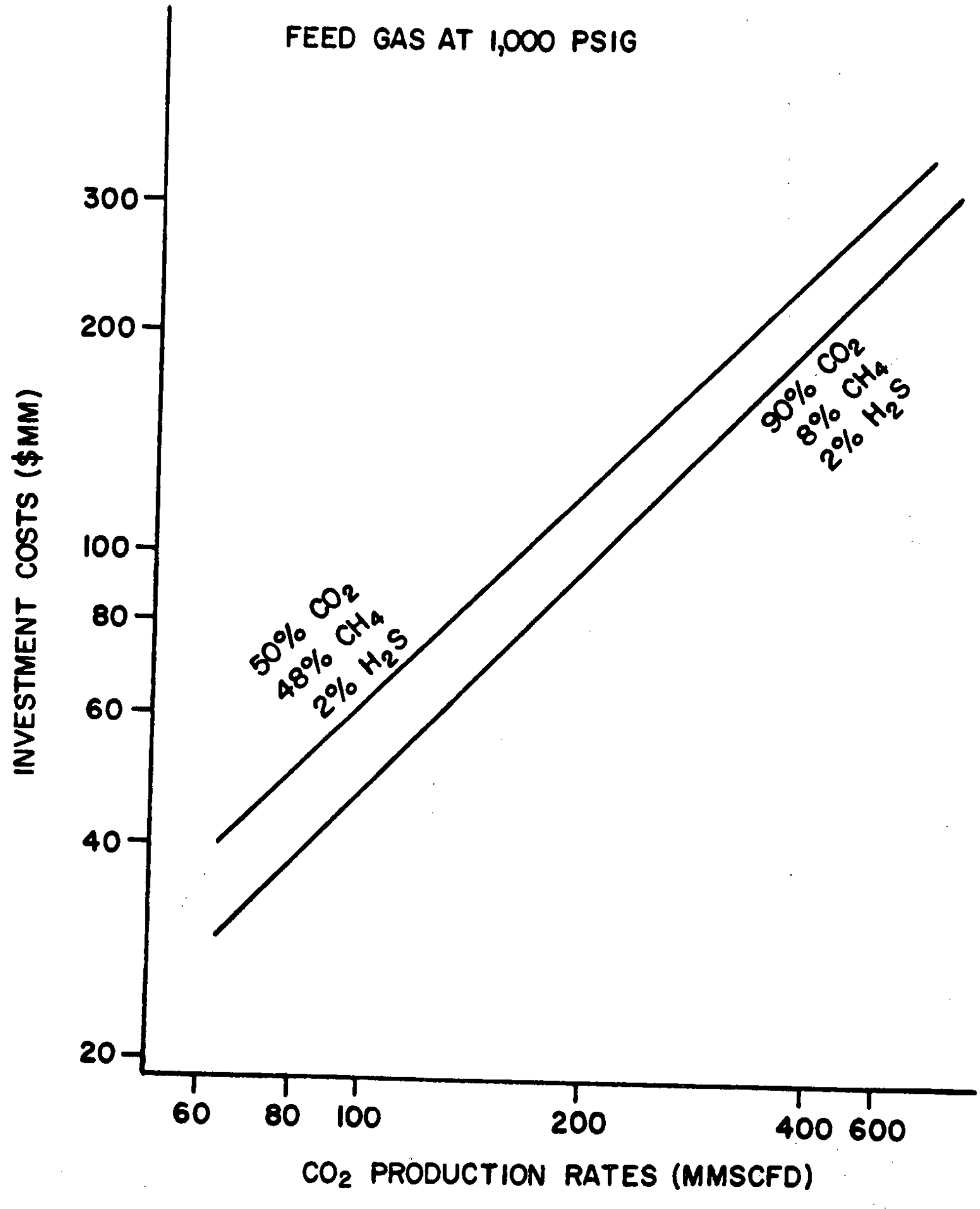
PURIFICATION COSTS OF FEED GAS CONTAINING 50% CO_2 AND 2% H_2S FROM NATURAL SOURCES AND AVAILABLE AT 250 PSIG FOR VARIOUS DISCOUNTED CASH FLOW RATES OF RETURN VERSUS CO_2 PRODUCTION RATES

FIGURE 6.30



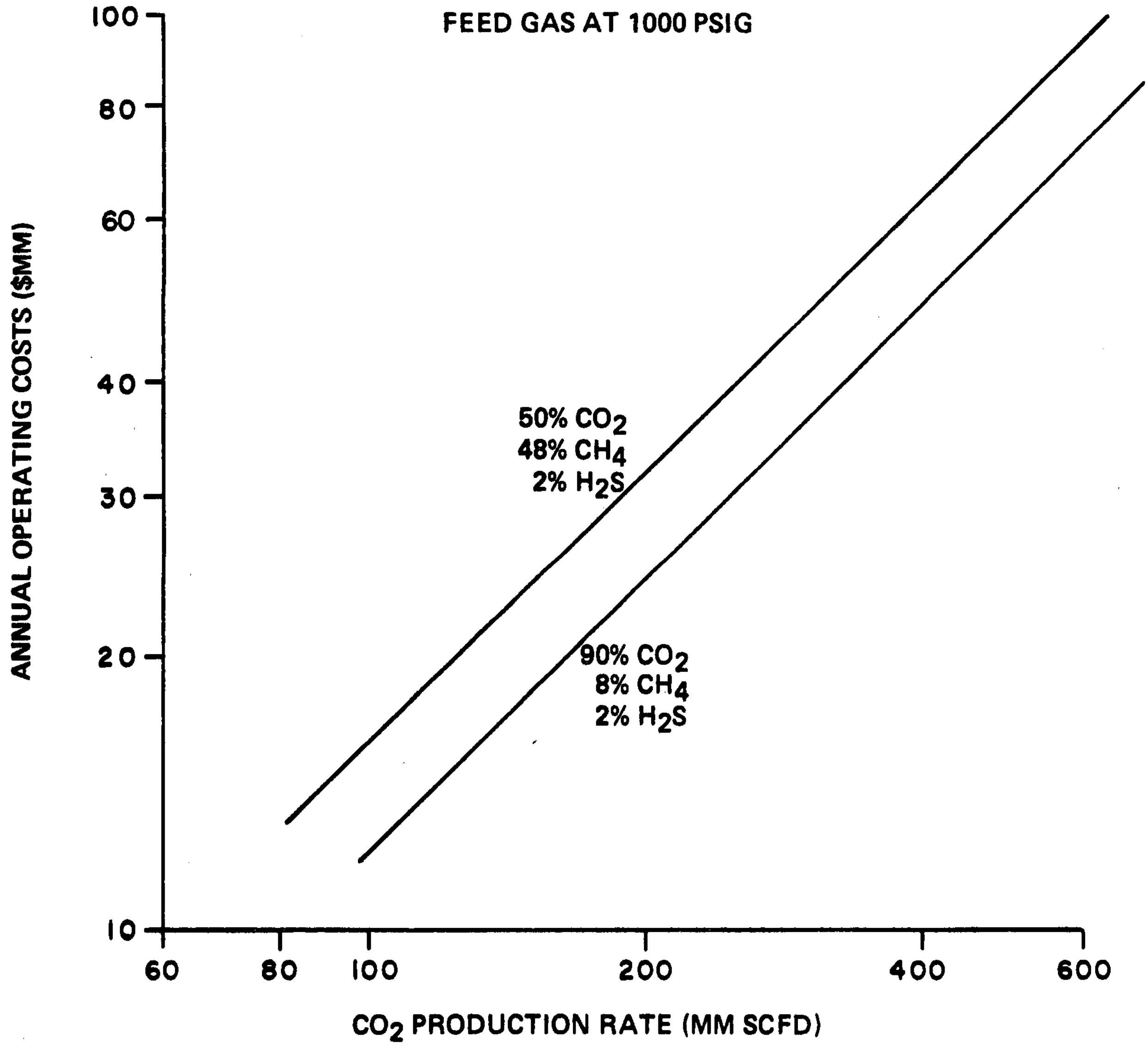
PURIFICATION COSTS OF FEED GAS CONTAINING 90% CO_2 AND 2% H_2S FROM NATURAL SOURCES AND AVAILABLE AT 250 PSIG FOR VARIOUS DISCOUNTED CASH FLOW RATES OF RETURN VERSUS CO_2 PRODUCTION RATES

FIGURE 6.31



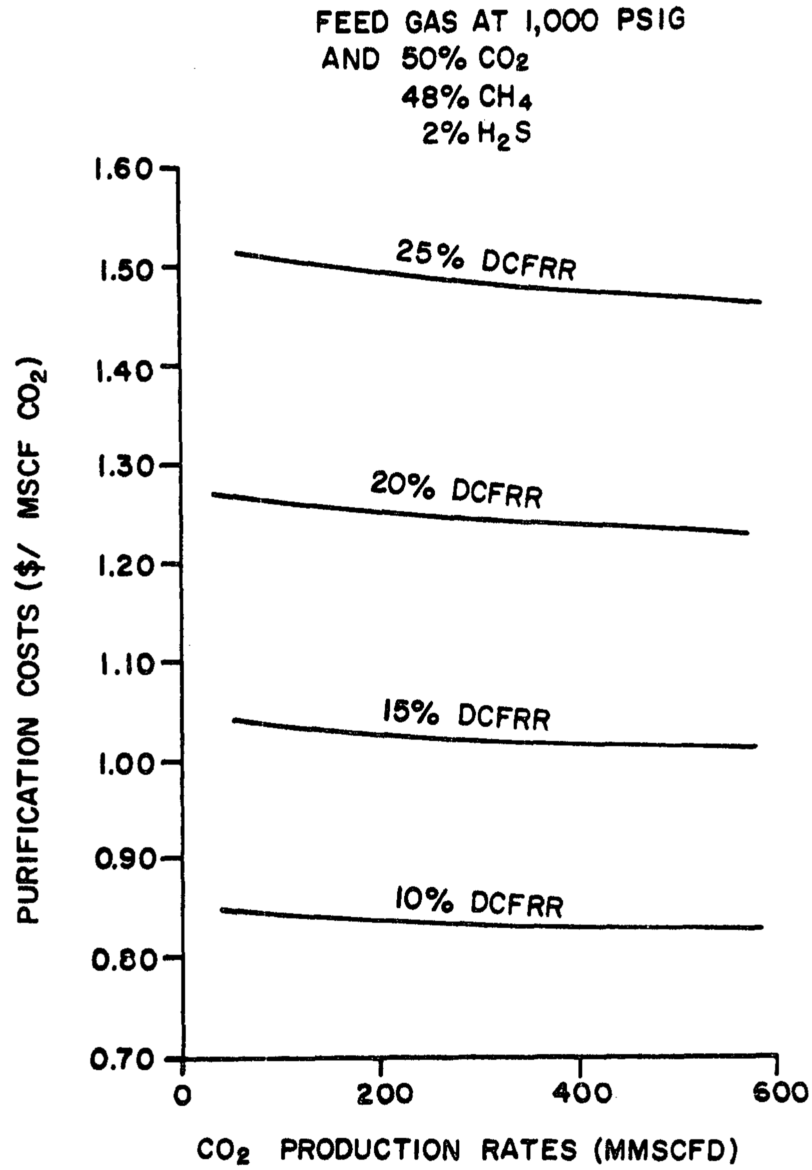
INVESTMENT COSTS FOR CO_2 PURIFICATION FROM NATURAL SOURCES CONTAINING H_2S AND AVAILABLE AT 1,000 PSIG VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.32



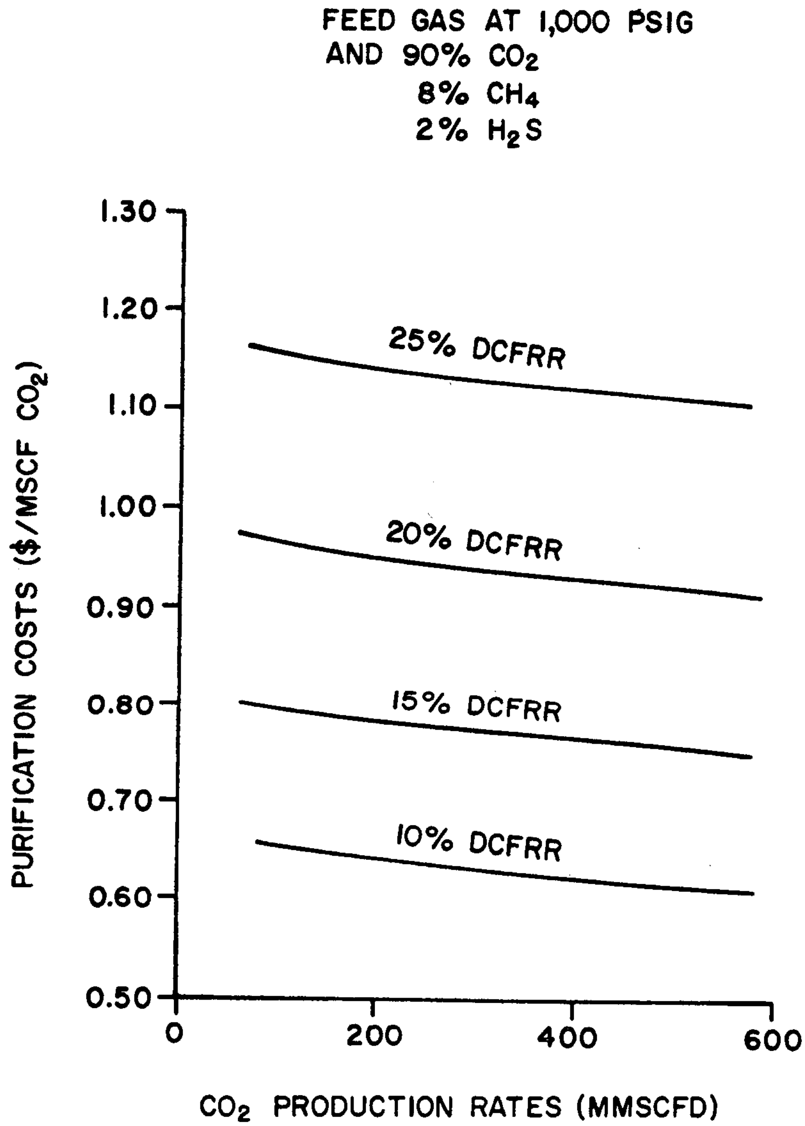
ANNUAL OPERATING COSTS FOR CO_2 PURIFICATION FROM NATURAL SOURCES CONTAINING H_2S AND AVAILABLE AT 1000 PSIG VERSUS VARIOUS CO_2 PRODUCTION RATES

FIGURE 6.33



PURIFICATION COSTS OF FEED GAS CONTAINING 50% CO_2 AND 2% H_2S FROM NATURAL SOURCES AND AVAILABLE AT 1,000 PSIG FOR VARIOUS DISCOUNTED CASH FLOW RATES OF RETURN VERSUS CO_2 PRODUCTION RATES

FIGURE 6.34



PURIFICATION COSTS OF FEED GAS CONTAINING 90% CO_2 AND 2% H_2S FROM NATURAL SOURCES AND AVAILABLE AT 1,000 PSIG FOR VARIOUS DISCOUNTED CASH FLOW RATES OF RETURN VERSUS CO_2 PRODUCTION RATES

FIGURE 6.35

6.4 PURIFICATION OF OTHER CO_2 SOURCES

The preceding discussions have not considered the purification costs for all potential CO_2 sources. The section on recovery from flue gas (6.3) examined those costs of a coal-fired power plant excluding gas-fired power plant and cement plant flue gases. Additionally, the recovery costs of CO_2 from flue gases exiting refinery fluid catalytic cracking units were not directly established. All of these sources have characteristics similar to the coal-fired power plants in respect to very low CO_2 quality and low (atmospheric) pressure. Therefore, their cost of purification may be approximated by the cost of the similar coal-fired power plant flue gas.

The concept of applying the purification cost of one of the sources examined in this report to another source that has similar gas composition and pressure may be utilized to expand this report to encompass all potential CO_2 sources. A basic similarity of source quality will provide a basis for which an estimate of the purification cost of another source may be made.

6.5 PROCESSING OF HIGH PURITY SOURCES

6.5.1 High Purity Sources Available At Atmospheric Pressure

The maps of CO_2 sources in Part 3 reveal that several aboveground CO_2 sources contain high purity gas at a low pressure. Ammonia and hydrogen plants are good examples of sources with the gas available at 98% CO_2 purity and at atmospheric pressure. For the processing of these gases, purification of the CO_2 is not necessary, but compression and dehydration to pipeline pressure is needed.

The investment cost and annual operating cost for the required compression/dehydration with inlet conditions at atmospheric pressure and the resulting gas price ($/MSCF) for various DCFRR's are shown in Figures 6.36 and 6.37 respectively. The compressors are two body, four section, centrifugal machines. A steam turbine driver is assumed since the machine will be located within the confines of a chemical plant and 900 psig - 900^{o}F steam can be assumed to be available. As the gas is compressed and cooled between stages, water condenses and is removed in knockout vessels. Between the third and fourth section of the compressor the gas is dried to final pipeline quality by a TEG drying unit.

6.5.2 High Purity Sources Available At Pressures Greater Than Atmospheric

As indicated in Part 5 of this report, the Northeast New Mexico area contains high purity CO_2 sources at pressures in the range of 300 to 900 psig. As a result, compression costs incurred in compressing CO_2 from high pressure sources for pipeline transmission, are not as large as compression costs incurred in compressing CO_2 from atmospheric sources. Therefore, Equation 6.1 is shown below to aid in determining compression costs for various inlet pressure conditions.

$$BHP = 2.6 \times 10^{-4}\, Q \left(\left(\frac{2014.7}{P_1}\right)^{.18} - 1 \right) \qquad (6.1)$$

where

BHP = Required Compressor Horsepower (HP)

Q = Compressor Inlet Flow Rate (SCFD)

P_1 = Compressor Inlet Pressure (PSIA)

As a result, using Figures 6.36 and 6.37 in conjunction with Equation 6.1, it is possible to determine base compressor investment costs and base compressor operating costs for high purity CO_2 sources. The procedure outlined below demonstrates the use of Equation 6.1 in determining these investment and operating costs:

1. Using Equation 6.1, calculate BHP based on actual compressor inlet conditions (e.g. P_1 = 700 psia and 200 MMSCFD CO_2).

2. Then, recalculate a new flow rate (Q) using Equation 6.1 based on the calculated BHP from Step 1 above and P_1 = 14.7 psia (e.g. if BHP = 11,000 @ 700 psia and 200 MMSCFD CO_2, then @ BHP = 11,000 and 14.7 psia, Q = 31.5 MMSCFD CO_2).

3. Use Figure 6.36 to determine the base compressor investment cost at the flow rate determined in Step 2 above (e.g. if Q = 31.5 MMSCFD CO_2, then the compression investment cost = $7,000,000).

4. Use Figure 6.37 to determine the base compressor operating costs at the flow rate determined in Step 2 above (e.g. if Q = 31.5 MMSCFD CO_2, then the compression operating cost = $3,100,000/YR).

If the high purity CO_2 source, for example, is a processing plant, then the compression cost can be determined as part of the processing plant if the base compressor is located within the plant. Consequently, the rate-of-return on investment is not limited. Figure 6.38 shows these compression costs for the high purity CO_2 sources based on DCFRR's of 10%, 15%, 20%, and 25%. However, if the high purity CO_2 source, for example, is a natural source and does not require processing, the base compressor could be considered part of the pipeline, and in as much, regulated by the Federal Energy Regulatory Commission. Consequently, an 8% or 10% rate-of-return on investment using the utility financing method, limits and governs these compression costs. Figure 6.39 shows these compression costs for high purity CO_2 sources based on an 8% and 10% rate-of-return using the Utility Financing Method.

All figures in Section 6.3.4 were drawn for the case of sources composed of CO_2, CH_4, H_2S (if present) and H_2O without credit for the CH_4 that is co-produced. Therefore, these figures will provide a good estimate of the processing costs for sources containing non-saleable gases (such as N_2) or other light hydrocarbons.

High quality sources (those not needing purification) are discussed in Section 6.5. Other sources that are of similar composition to those previously examined in Sections 6.2 and 6.3 are discussed in Section 6.4.

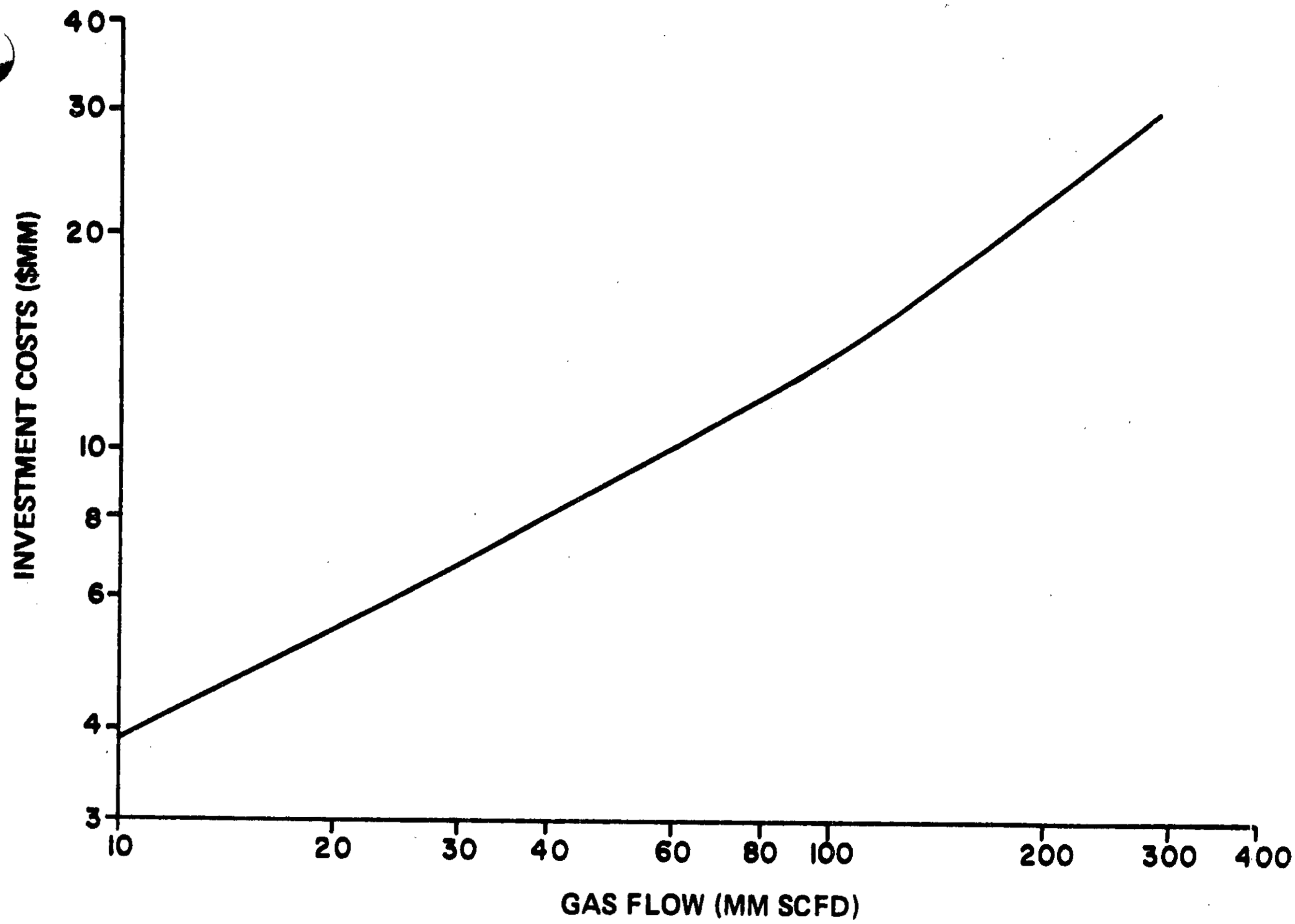
6.6 APPLICATION OF INFORMATION

The following discussion is an effort to explain how to use the information contained in the preceeding sections and extend it to other CO_2 sources not specifically mentioned. Part 6 includes the processing costs associated with gas sources composed of 10% to 98+% CO_2 and available from pressures of atmospheric to over 1,000 psig. Thus a wide range of situations are presented for use in determining processing costs.

If the source is poor quality (10% - 25% CO_2) and available at low pressures, the costs in Figures 6.4, 6.5 and 6.6 will provide a good estimate for this type of source's processing costs.

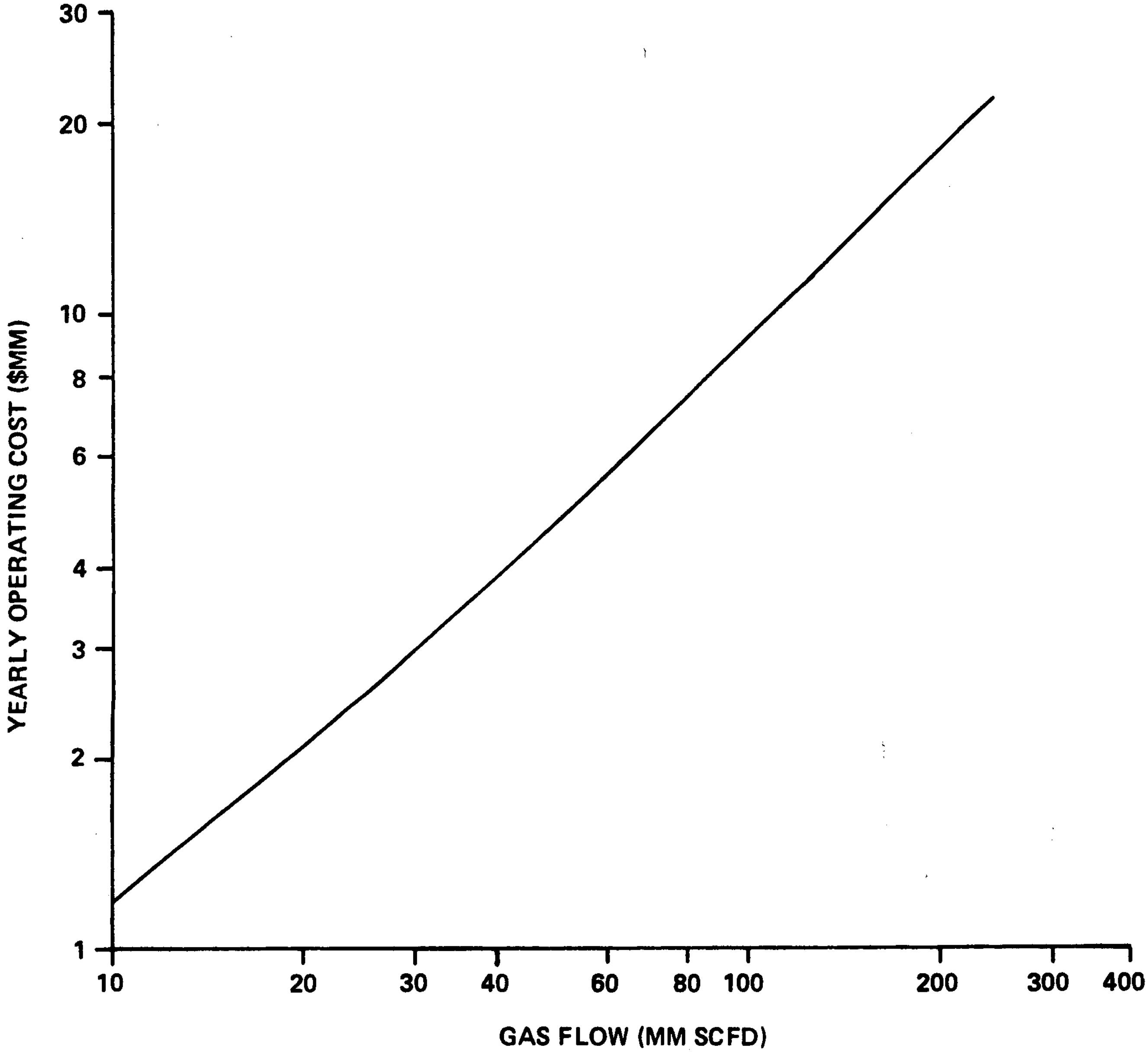
However, if the poor quality source is available at much higher pressures, then the cost for the processing 125 MMSCFD of CO_2 may be established from Table 6.1. For CO_2 production rates other than 125 MMSCFD CO_2, the relationship of the purification costs as a function of CO_2 rates may be established from Figures 6.6 and 6.21. From this established relationship, the processing costs of different CO_2 production rates may be established.

The processing costs of low (25% - 75% CO_2) and medium (75% - 92% CO_2) quality sources available at various pressures and production rates are contained in Figures 6.12 - 6.27 in Section 6.3.4. Careful extrapolation may be used to extend the ranges of these curves slightly.



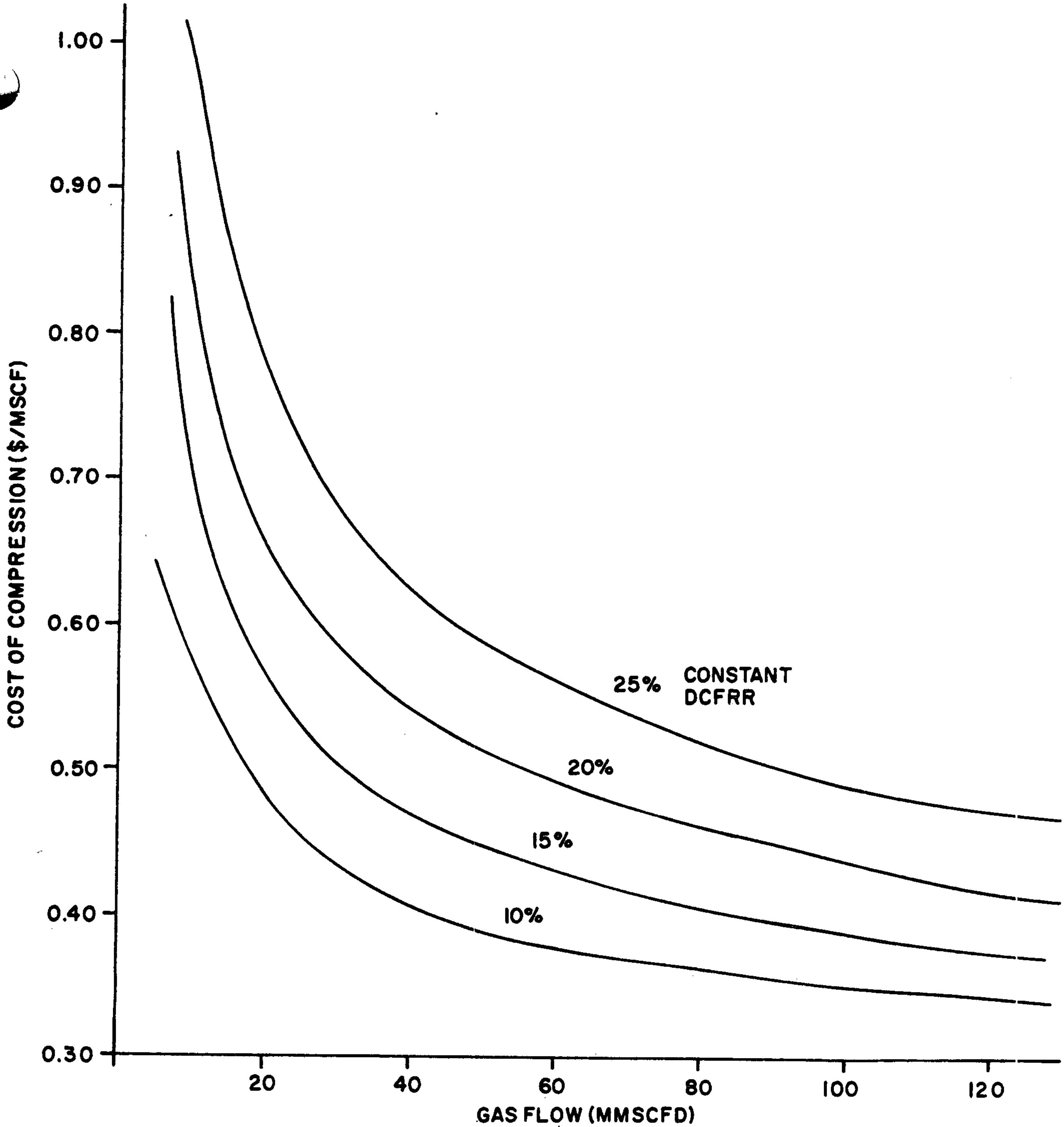
INVESTMENT COST OF BASE PIPELINE COMPRESSOR FOR HIGH PURITY CO_2 SOURCES

FIGURE 6.36



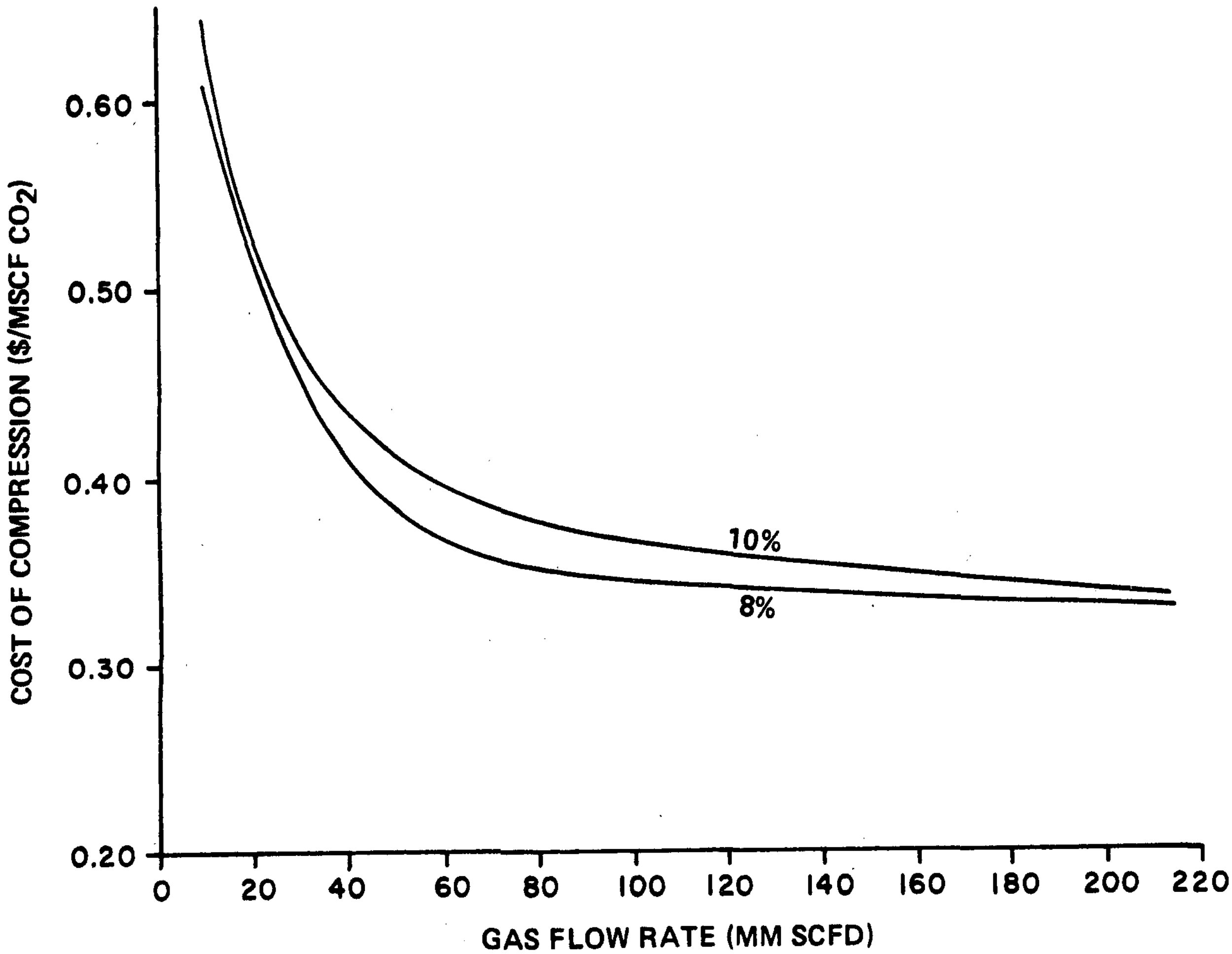
<u>ANNUAL OPERATING COST OF BASE PIPELINE COMPRESSOR FOR HIGH PURITY CO_2 SOURCES</u>

FIGURE 6.37



COMPRESSION COSTS FOR HIGH PURITY, LOW PRESSURE SOURCES AT VARIOUS DCFRR's

FIGURE 6.38



COMPRESSION COSTS FOR HIGH PURITY, LOW PRESSURE SOURCES AT 8% AND 10% RATE OF RETURN USING UTILITY FINANCING

FIGURE 6.39

COST OF TRANSPORTATION

PART 7

7.1 INTRODUCTION

This portion of the report is concerned with evaluating the most economical method of transporting carbon dioxide (CO_2) for enhanced oil recovery. Transport of CO_2 by tank truck or by railcar is not considered in this report. A few reasons for not considering the tank truck and railcar systems are listed below:

1. Previous studies concerned with CO_2 transmission indicate that tank truck and railcar transport systems are more expensive than pipeline systems.[28]
2. Continuous transport of CO_2 using tank truck and railcar is labor dependent. A possible strike could curtail the transport of CO_2 to the oil recovery site.
3. A shortage of tank trucks and railcars could interrupt a continual flow of CO_2 to the oil recovery site if such tankers were not available.

Therefore, this report will concentrate only on CO_2 transmission utilizing pipeline systems. Three pipeline systems are chosen as candidates for the most economical method of CO_2 transportation. These are:

1. A supercritical pipeline system
2. A subcritical pipeline system
3. A liquid pipeline system

Each method of transportation is defined in more detail later in this report.

The pipeline systems mentioned above will transport CO_2 from the natural CO_2 well site or the CO_2 production plant site, to the candidate oil reserve site for enhanced oil recovery. At the oil recovery site, the CO_2 is injected into the oil reservoir to create whatever pressure criterion selected for complete miscible flooding within the oil reservoir. Carbon dioxide pressure at the discharge of the injection compressor is assumed to be 2000 psig for this study.

In this report, Section 7.2 describes the three candidate pipeline systems. Each pipeline system is a functional and workable unit, in that each system transports CO_2 from the source area to the oil recovery site. However, it is our opinion that two of these pipeline systems are not economically practical. Therefore, these three systems will be analyzed in Section 7.3 and the analyzed results will demonstrate which system is best for the transportation of CO_2. Additionally, Section 7.3 contains material which can determine CO_2 transportation costs for various pipeline capacities and transmission lengths utilizing the most economical method of transportation.

7.2 MODES OF CO_2 TRANSPORTATION

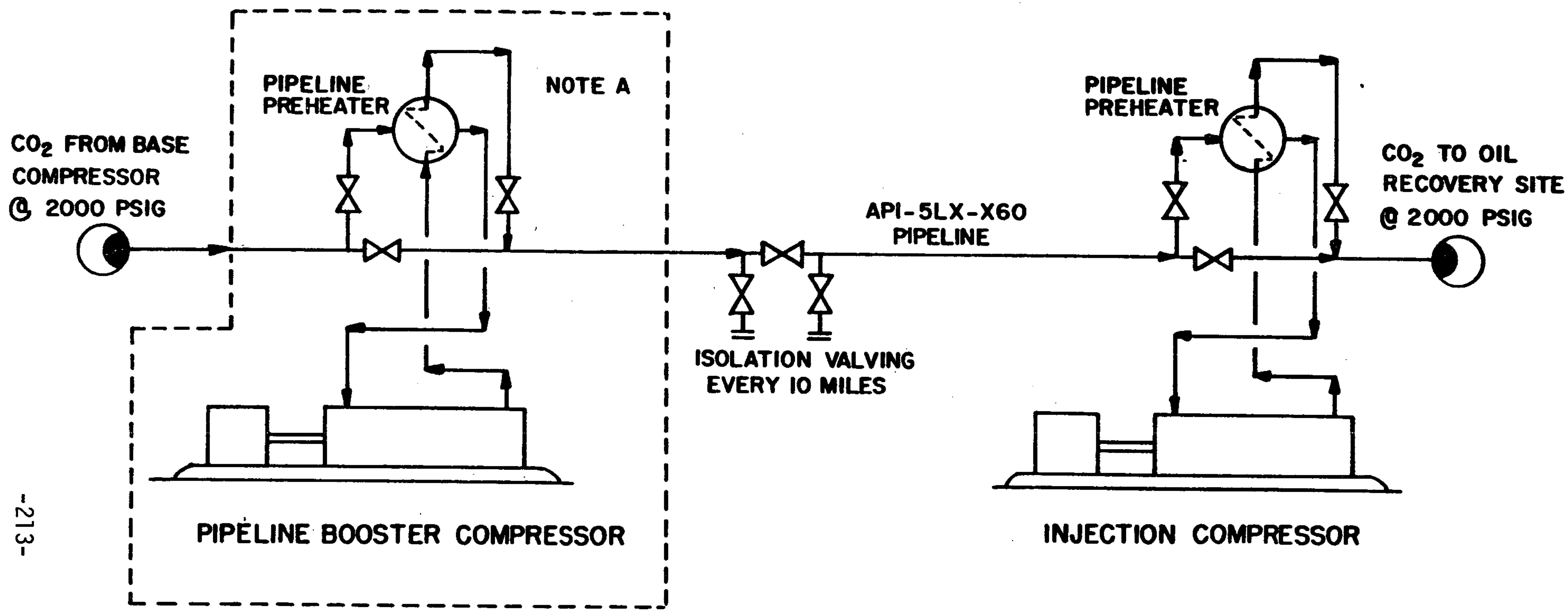
Three pipeline systems are considered for the transportation of carbon dioxide (CO_2) for enhanced oil recovery. This section will describe each pipeline system and explain the various components that make up that particular pipeline.

7.2.1 The Supercritical Pipeline

The first system discussed is the supercritical pipeline. Supercritical is defined as being above or greater than critical pressure. The critical pressure of CO_2 is 1071 psia, therefore, in a supercritical CO_2 pipeline system, the operating pressure is greater than 1071 psia.

In this report, the supercritical pipeline system consists of the intermediate pipeline booster compressors, the pipeline compressor preheaters, the injection compressor, and the pipeline. Figure 7.1 shows a general schematic of the supercritical pipeline.

A base compressor, located either at the natural CO_2 well site or at the CO_2 production plant site, injects the CO_2 into the pipeline at 2000 psig. Intermediate pipeline booster compressors are required to maintain pressure in the range of 1400 to 2000 psig throughout the pipeline. At each intermediate pipeline booster compressor, a pipeline



NOTE A: THE NUMBER OF REQUIRED PIPELINE BOOSTER COMPRESSORS VARIES WITH PIPELINE LENGTH AND CO_2 FLOWRATE.

SUPERCRITICAL PIPELINE ARRANGEMENT

FIGURE 7.1

preheater is provided to heat the CO_2 from the equilibrated pipeline temperature to 130^{O}F. This is done to satisfy compressor manufacturer's guarantees.* The injection compressor, for all practical purposes, is an intermediate pipeline booster compressor and is located at the oil recovery site. Its function is to inject CO_2 into the well at 2000 psig.

In Section 7.3 the supercritical pipeline economics are discussed. However, as a prelude one should be made aware that not all possible supercritical pipeline systems have been economically evaluated. A preliminary screening procedure eliminates many systems which are economically unrealistic. The results from this preliminary screening procedure are the optimum supercritical pipeline systems shown in Table 7.1 on the following page. These optimum supercritical pipeline systems are the ones analyzed in Section 7.3.

* The compressibility factor for CO_2 above 1400 psig and 60^{O}F is less than 0.3. The compressor manufacturers desire a minimum compressibility factor of 0.5 at 1400 psig. Therefore, a compressor inlet temperature of 130^{O}F solves this minimum compressibility factor requirement.

TABLE 7.1

OPTIMUM SUPERCRITICAL PIPELINE SYSTEMS

STUDIED FOR CO_2 TRANSMISSION

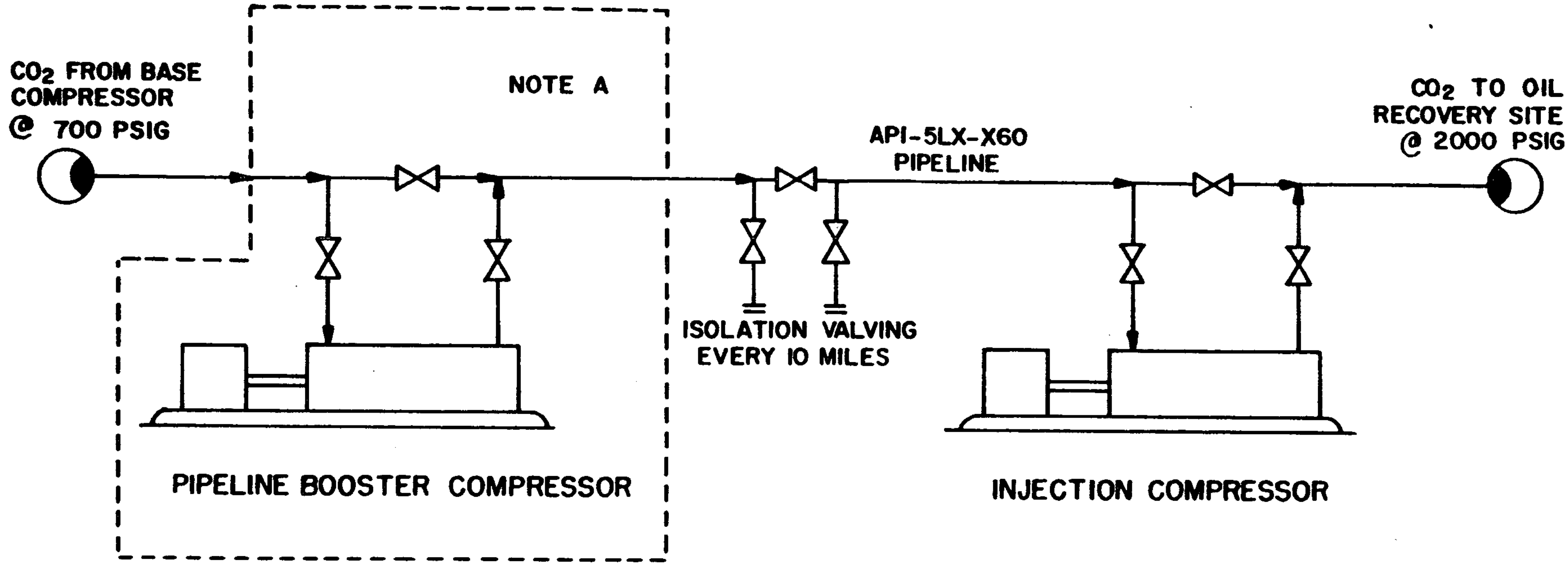
CO_2 Flowrate	Pipeline Diameter	Pipeline Length
50 MMSCFD	8"	50 mile to 500 mile
125 MMSCFD	10"	50 mile to 500 mile
250 MMSCFD	14"	50 mile to 500 mile
500 MMSCFD	18"	50 mile to 500 mile

7.2.2 The subcritical Pipeline

The second candidate for CO_2 transmission is the subcritical pipeline. Subcritical is defined as lower or less than critical pressure. Therefore, any operating pressure lower than critical pressure of CO_2 (1071 psia) in a pipeline is subcritical.

As shown in Figure 7.2, the subcritical pipeline is very similar to the supercritical pipeline. The subcritical pipeline consists of the intermediate pipeline booster compressors, the injection compressor, and the pipeline. However, the subcritical pipeline booster cmpressors do not require pipeline preheaters as required in the supercritical pipeline.

A base compressor, located either at the nautral CO_2 well site or at the CO_2 production site, injects the CO_2 into the pipeline at



NOTE A: THE NUMBER OF REQUIRED PIPELINE BOOSTER COMPRESSORS VARIES WITH PIPELINE LENGTH AND CO_2 FLOWRATE.

SUBCRITICAL PIPELINE ARRANGEMENT

FIGURE 7.2

700 psig. The intermediate pipeline booster compressors maintain a 400 to 700 psig pressure range throughout the pipeline. Once the CO_2 reaches the oil recovery site, it is injected into the well at 2000 psig by the injection compressor.

In Section 7.3 the subcritical pipeline economics is discussed. However, a preliminary screening procedure for the subcritical pipeline, identical to that used for the supercritical pipeline, eliminates pipeline systems which are economically unrealistic. As a result of this preliminary screening procedure, only the optimum subcritical pipeline systems are analyzed in Section 7.3. In addition, it is evident from the preliminary screening procedures, that the subcritical pipeline at longer transmission lengths, compared to the supercritical pipeline, are not as economical. Therefore, only the shorter CO_2 pipelines in the subcritical case are analyzed. The systems analyzed in Section 7.3 are shown in Table 7.2 below.

TABLE 7.2

OPTIMUM SUBCRITICAL PIPELINE SYSTEMS

STUDIED FOR CO_2 TRANSMISSION

CO_2 Flowrate	Pipeline Diameter	Pipeline Length
50 MMSCFD	12"	50 mile, 100 mile, 300 mile
125 MMSCFD	18"	50 mile, 300 mile

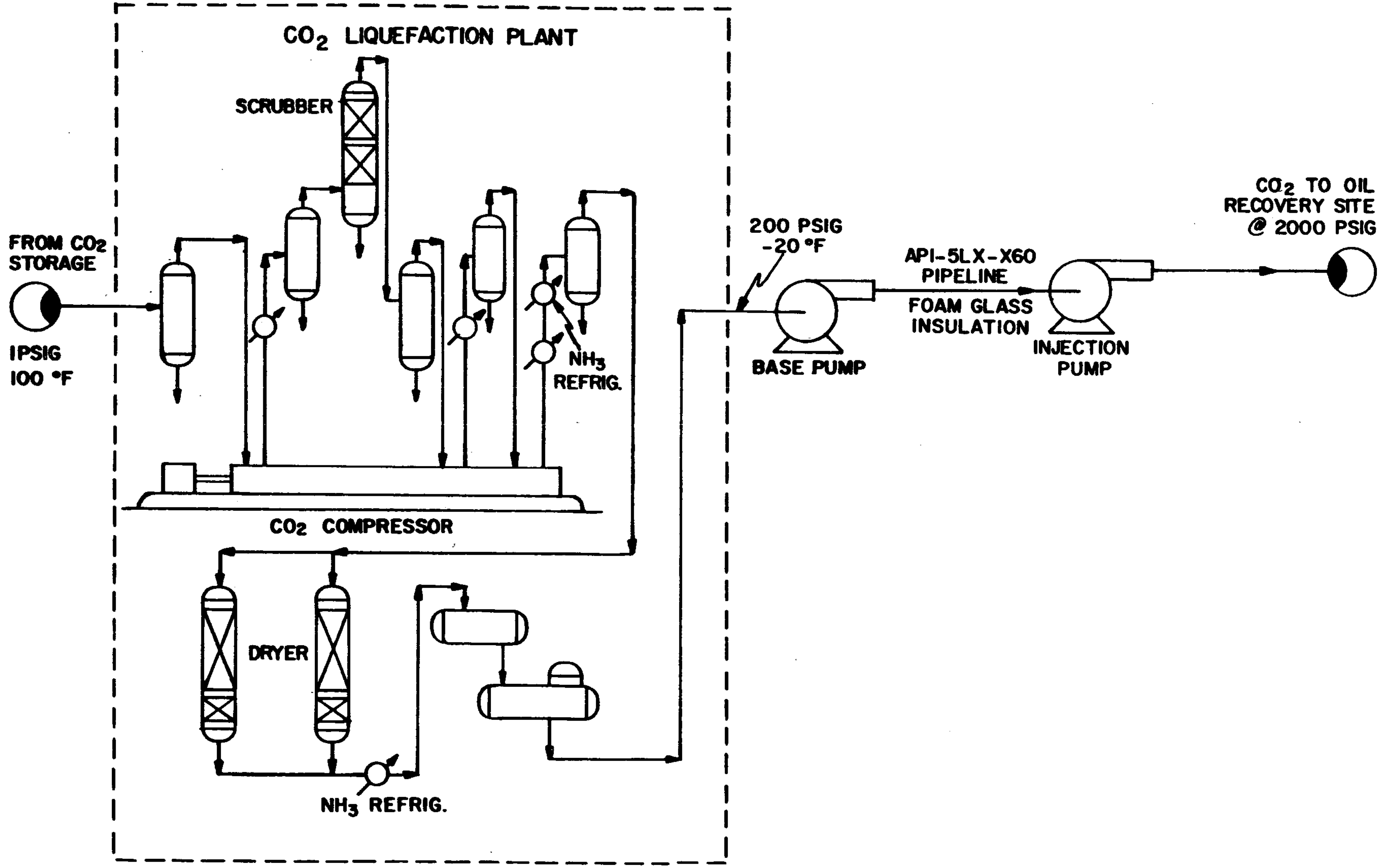
For cost estimating purposes, all the intermediate pipeline booster compressors and injection compressors in the supercritical and subcritical pipelines have electric motor drivers. Also, sectionalizing block valves (isolation valves) have been installed every ten miles in the pipeline. An average value of ten miles was derived from various spacing requirements conforming to geographic classification guidelines for sectionalizing block valves in pipelines.

7.2.3 The Liquid Pipeline

The third candidate system for CO_2 transmission is the liquid pipeline. The two previously mentioned pipeline systems, the supercritical and the subcritical systems, are essentially vapor flow pipelines. Whereas, the liquid pipeline system, as the name implies, is a liquid flow pipeline.

Basically, there are four major components required in making the liquid pipeline a functional system. These components are the CO_2 liquefaction plant, the base pump, the injection pump, and the insulated pipeline. These four major liquid pipeline components are shown in a general schematic found in Figure 7.3.

The CO_2 liquefaction plant is the key ingredient for the liquid transport of CO_2 in this pipeline. Initially, CO_2 enters the liquefaction plant from the natural CO_2 site or production plant site in the



LIQUID PIPELINE ARRANGEMENT AT 100 MILES FOR 8", 10", AND 12"

FIGURE 7.3

vapor state. Through a series of compression and cooling steps, the CO_2 changes from the vapor state to the liquid state. Once in the liquid state, the CO_2 is charged into the liquid pipeline at a pressure of 200 psig and -20F.

A base pump transports the liquified CO_2 from the liquefaction plant through the pipeline to an injection pump. The injection pump, located at the oil recovery site, injects the liquid CO_2 into the well at 2000 psig.

Table 7.3 below summarizes the liquid pipeline systems analyzed in Section 7.3.

TABLE 7.3

LIQUID PIPELINE SYSTEMS

STUDIED FOR CO_2 TRANSMISSION

CO_2 Flow Rate	Pipeline Diameter	Pipeline Length
50 MMSCFD	8"	100 miles
50 MMSCFD	10"	100 miles
50 MMSCFD	12"	100 miles

From these liquid pipeline systems listed above, an optimum system is determined. The optimum liquid pipeline system at 50 MMSCFD and 100 miles is then compared to the supercritical and the subcritical pipeline systems for CO_2 transmission. The results follow in Section 7.3.

7.3 ECONOMIC ANALYSIS OF CO_2 TRANSPORTATION

This section is concerned with the economics of CO_2 transmission in a system of pipelines, specifically the supercritical, the subcritical, and the liquid pipeline systems. The preliminary pipeline screening procedure used in the selection of the optimum pipeline systems found in this report, is described in this section. In addition, an economic evaluation using discounted cash flow methods and utility financing methods for these selected optimum pipeline systems, is contained within this section as well. Based on the economic analyses mentioned above, the three pipeline systems, supercritical, subcritical, and liquid, are directly compared. The direct comparison between these three systems will, in turn, demonstrate which pipeline system is best for transporting CO_2. Finally, this section contains material that can serve as a basis for determining CO_2 transportation costs consistent with flow rate, transmission length, pipeline differential elevation, and geographic terrain.

7.3.1 Preliminary Pipeline Screening Procedure

As previously stated in Section 7.2, a preliminary screening procedure eliminated many unrealistic pipeline systems with the intent of analyzing only the optimum pipeline system. A descriptive summary of the screening procedure follows below. For brevity, only the supercritical pipeline system is discussed. The other pipeline systems, subcritical and liquid, were analyzed using the same method.

Initially, a basis was established for qualifying three parameters in the CO_2 pipeline system. The first parameter requiring a basis was the pipeline transmission length. These lengths were set at:

1. 50 miles
2. 100 miles
3. 300 miles
4. 500 miles

The second parameter requiring a basis was the pipeline capacity (CO_2 flow rate). These capacities were set at:

1. 50 MMSCFD CO_2
2. 125 MMSCFD CO_2
3. 250 MMSCFD CO_2
4. 500 MMSCFD CO_2

The third parameter requiring a basis was the pipeline operating pressure. For the supercritical pipeline the minimum operating pressure was set at 1400 psig[26] and the maximum operating pressure was set at 2000 psig. Once the parameters outlined above were established, the pipeline materials of construction were specified. Based on conservative engineering practice coupled with essentially a water-free CO_2 flow composition, API-5LX-X60[26] was chosen as the standard specification for the CO_2 pipeline. Minimum pipeline wall thicknesses were then calculated for various pipeline diameters based on the operating pressures described above and API-5LX-X60. All pipeline design calculations were based on the code ANSI-B31.8, Gas Transmission and Distribution Systems .

Once the qualifying parameters of the pipeline were established (i.e. flow rate, transmission length, pressure limits, pipe diameter, and wall thickness) it was then possible to design several functional CO_2 pipeline systems. These "functional pipelines" were designed using the "New Panhandle" equation.[27] In this report, a "functional pipeline" is defined as a pipeline system which transports the material (CO_2) from point A to point B using any number of compressors required for this transport in a specific pipe size. An example of a "functional pipeline" system follows:

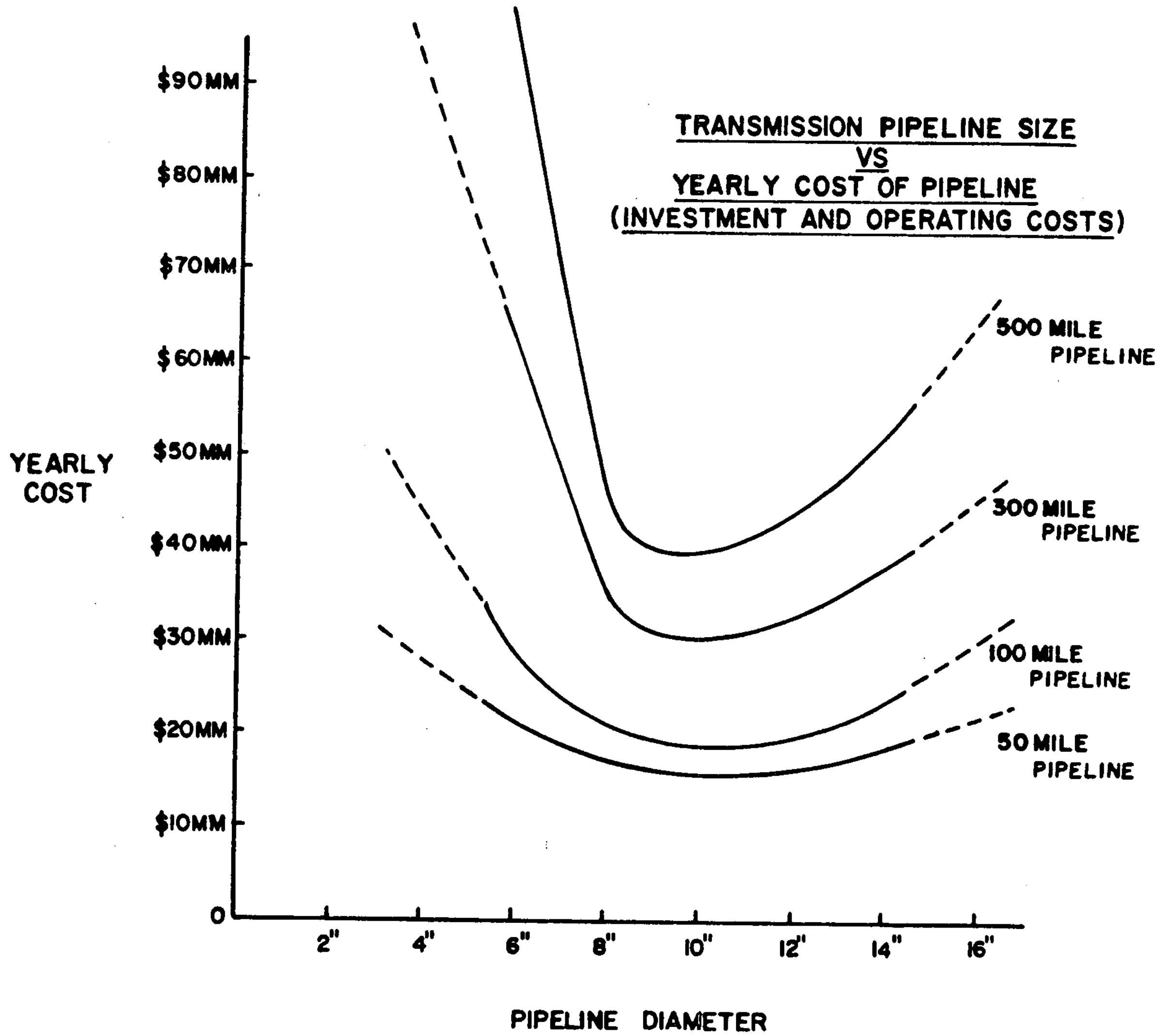
> For a supercritical pipeline system (1400 to 2000 psig), 87 pipeline booster compressors are required to transport 125 MMSCFD CO_2 at a transmission length of 500 miles in a 6-inch diameter pipeline. (See Figure 7.4 for relative costs between "functional" and optimum pipeline systems.)

After designing the "functional pipeline" systems from the parameters established earlier, the optimum pipeline system for each transmission length was then extracted using a preliminary economic analysis. The basis for this preliminary economic analysis involved using both direct and indirect operating costs, as well as the initial investment (the installed purchase price) of the pipeline system. Curves for the optimum pipeline systems extracted using this method of analysis are shown in Figure 7.4 below at a flow rate of 125 MMSCFD CO_2. Each optimum pipeline system extracted in this manner was then reanalyzed, with greater emphasis placed on accurate fixed cost estimates. In addition, each optimum pipeline system was reevaluated using utility financing and discounted cash flow rate-of-return on investment techniques. The analyzed and evaluated results for the optimized pipeline systems immediately follow.

7.3.2 The Best CO_2 Pipeline Transportation System

In the preceding section (7.3.1), the general screening procedure employed for selecting optimum pipeline systems for economic evaluation was discussed. This section, in turn, will take these selected pipeline systems and evaluate their investment costs, analyze their operating costs, and compare their overall combined costs (investment and operating costs). The CO_2 transportation costs, based on a fixed rate-of-return using discounted cash flow and utility financing methods, will in turn, indicate which mode of CO_2 transportation is the most economical.

Tables 7.4, 7.5, and 7.6 show the total investment costs for the supercritical, the subcritical, and the liquid pipeline systems, respect-



PRELIMINARY SCREENING RESULTS AT 125MMSCFD CO_2

FIGURE 7.4

TABLE 7.4

SUPERCRITICAL PIPELINE — Basis: Flat Terrain

TOTAL INVESTMENT COSTS

Pipeline Capacity ($MMSCFD\ CO_2$)	Pipeline Length (Miles)	Total Investment Cost ($MM)	Total Investment Cost/Unit CO_2 ($MM/MMSCFD CO_2 Capacity)
50	50	7	0.14
	100	13	0.26
	300	36	0.72
	500	59	1.18
125	50	10	0.08
	100	19	0.15
	300	52	0.42
	500	85	0.68
250	50	14	0.06
	100	27	0.11
	300	73	0.29
	500	123	0.49
500	50	20	0.04
	100	39	0.08
	300	114	0.23
	500	189	0.38

TABLE 7.5

SUBCRITICAL PIPELINE BASIS: Flat Terrain

TOTAL INVESTMENT COSTS

Pipeline Capacity (MMSCFD CO_2)	Pipeline Length (Miles)	Total Investment Cost ($MM)	Total Investment ($MM/MMSCFD CO_2 Capacity)
50	50	10	0.20
	100	19	0.38
	300	62	1.24
125	50	15	0.12
	300	75	0.60

TABLE 7.6

LIQUID PIPELINE BASIS: Flat Terrain

TOTAL INVESTMENT COSTS

Pipeline Capacity (MMSCFD CO_2)	Pipeline Length (Miles)	Total Investment Cost ($MM)	Total Investment ($MM/MMSCFD CO_2 Capacity)
50	100 (8" dia.)	37	0.74
	100 (10" dia.)	40	0.80
	100 (12" dia.)	43	0.86

ively. The total investment costs are based on 1978 fourth quarter dollars. (For further information concerning the total investment costs used in this report for the supercritical, the subcritical, and the liquid systems, see Appendix B1.)

Table 7.7 compares the total investment costs for three candidate pipeline systems. For this comparison, the CO_2 pipeline flow rate is 50 MMSCFD and the pipeline transmission length is 100 miles. Note the various equipment costs for each pipeline system. These costs are substantially less than the pipeline costs. Also, note which pipeline system requires less total capital for investment. Based on total investment costs, the supercritical pipeline system appears, thus far, to be more economical than the liquid and subcritical pipelines. However, utility and operating costs need to be analyzed and compared as well. These cost comparisons follow.

Tables 7.8, 7.9, and 7.10 show the total direct and indirect operating costs for the supercritical, the subcritical, and the liquid pipeline systems, respectively. From these tables, for the CO_2 pipeline compared earlier (flow rate of 50 MMSCFD at a transmission length of 100 miles), it is evident that the operating costs for the liquid pipeline are far greater than the operating costs for either the supercritical or the subcritical pipeline systems. (For information concerning utility and operating cost basis used in this report, see Appendix B2.) Even though the operating cost for pumping liquid CO_2 is relatively low, the

TABLE 7.7

COMPARISON OF

TOTAL INVESTMENT COSTS

FOR 50 MMSCFD CO_2 @ 100 MILES

BASIS: Flat Terrain

Equipment List	Supercritical Total Investment Cost	Subcritical Total Investment Cost	Liquid Total Investment Cost
COMPRESSORS	$2,000,000*	$5,000,000	-
PUMPS	-	-	$2,000,000
PIPE	$11,000,000	$14,000,000	$11,000,000
PIPE INSULATION	-	-	$5,000,000
LIQUEFACTION PLANT	-	-	$19,000,000
TOTAL	$13,000,000	$19,000,000	$37,000,000
TOTAL INVESTMENT \$MM Per MMSCFD CO_2 CAPACITY	$0.26	$0.38	$0.74

*WITH COMPRESSOR PREHEATERS INCLUDED

TABLE 7.8

SUPERCRITICAL PIPELINE

BASIS: Flat Terrain

OPERATING COSTS

Pipeline Capacity ($MMSCFD\ CO_2$)	Pipeline Length (Miles)	Total Operating Cost ($MM/YR)	Operating Cost ($/MSCF CO_2)
50	50	0.7	0.04
	100	1.1	0.06
	300	3.0	0.18
	500	5.0	0.29
125	50	1.1	0.03
	100	2.1	0.05
	300	6.0	0.14
	500	9.8	0.23
250	50	1.3	0.02
	100	2.7	0.03
	300	6.8	0.08
	500	11.9	0.14
500	50	2.7	0.02
	100	5.4	0.03
	300	17.5	0.10
	500	29.2	0.17

TABLE 7.9

SUBCRITICAL PIPELINE BASIS: Flat Terrain

OPERATING COSTS

Pipeline Capacity ($MMSCFD\ CO_2$)	Pipeline Length (Miles)	Total Operating Cost ($MM/YR)	Operating Cost ($/MSCF CO_2)
50	50	2.0	0.12
	100	3.1	0.18
	300	5.3	0.31
125	50	3.8	0.09
	300	9.1	0.21

TABLE 7.10

LIQUID PIPELINE BASIS: Flat Terrain

OPERATING COSTS

Pipeline Capacity ($MMSCFD\ CO_2$)	Pipeline Length (Miles)	Total Operating Cost ($MM/YR)	Operating Cost ($/MSCF CO_2)
50	100 (8" dia.)	8.1	0.48
	100 (10" dia.)	8.2	0.48
	100 (12" dia.)	8.4	0.49

major operating costs generated from liquefying CO_2 far exceeds the pumping costs and discourages the use of a liquid CO_2 pipeline. The operating costs for the supercritical and the subcritical pipeline systems, however, are relatively close. Consequently, no decision can be made to definitely support or negate either system without further examination.

In order to select the best pipeline system, CO_2 transportation costs, based on various rates-of-return using discounted cash flow and utility financing methods, require analyzing. Table 7.11 shows the results of both the discounted cash flow and the utility financing methods for various rates-of-return on investment. For both the discounted cash flow and the utility financing methods, the basis for economic evaluation are listed below.

1. 20 year project life
2. 20 year straight line depreciation on total capital requirement
3. 48% federal income tax
4. 100% equity capital (no debt)

(For further information concerning the discounted cash flow method and the utility financing method, see Appendix B3.) Rates-of-return on investment were set at 10%, 15%, 20%, and 25% for the discounted cash

TABLE 7.11

CO_2 TRANSPORTATION COSTS BASED ON "DCFRR" AND "UTILITY FINANCE" METHODS

BASIS: Flat Terrain

	Pipeline Capacity (MMSCFD CO_2)	Pipeline Length (MILES)	Gas Cost Based On Discounted Cash Flow ($/MSCF CO_2)				Gas Cost Based On Utility Financing ($/MSCF CO_2)	
			@10%	@15%	@20%	@25%	@8%	@10%
Supercritical	50	50	0.12	0.17	0.22	0.27	0.13	0.14
		100	0.21	0.29	0.38	0.48	0.22	0.24
		300	0.61	0.83	1.09	1.38	0.61	0.69
		500	1.00	1.36	1.78	2.25	1.00	1.14
	125	50	0.07	0.09	0.12	0.15	0.07	0.08
		100	0.14	0.19	0.24	0.30	0.14	0.16
		300	0.39	0.52	0.67	0.83	0.39	0.44
		500	0.63	0.84	1.08	1.35	0.64	0.71
	250	50	0.05	0.07	0.09	0.11	0.05	0.06
		100	0.10	0.13	0.17	0.21	0.10	0.11
		300	0.25	0.35	0.45	0.57	0.26	0.29
		500	0.43	0.59	0.76	0.96	0.44	0.49
	500	50	0.04	0.05	0.07	0.08	0.04	0.05
		100	0.08	0.10	0.13	0.16	0.08	0.09
		300	0.24	0.31	0.39	0.48	0.24	0.27
		500	0.40	0.51	0.65	0.80	0.40	0.44
Subcritical	50	50	0.23	0.29	0.36	0.44	0.23	0.25
		100	0.41	0.52	0.66	0.81	0.41	0.45
		300	1.05	1.43	1.86	2.35	1.05	1.19
	125	50	0.16	0.20	0.24	0.29	0.16	0.17
		300	0.57	0.76	0.97	1.21	0.57	0.64
Liquid	50	100	0.92	1.15	1.42	1.72	0.93	1.01
		100	0.95	1.20	1.48	1.79	0.96	1.05
		100	1.01	1.27	1.58	1.92	1.01	1.11

flow method, from which the CO_2 transportation cost was directly determined. Similarly, using the utility financing method, a CO_2 transportation cost was determined for rates-of-return of 8% and 10%. Table 7.12 compares the CO_2 transportation costs for the three candidate pipeline systems at 50 MMSCFD CO_2 and 100 miles. It is evident, from this table, that the liquid pipeline system is, by far, the worst system economically. The subcritical pipeline system is more economical than the liquid pipeline system, but the supercritical pipeline system is, in turn, more economical than the subcritical pipeline system. For the case shown in Table 7.12, the CO_2 transportation costs for both the supercritical and the subcritical pipeline systems are relatively close. But, at longer pipeline transmission lengths (greater than 300 miles) the transportation cost differential between these two systems becomes significantly larger.

Based on the data presented above, the obvious conclusion is that the supercritical pipeline system is the most economical mode of CO_2 transportation. However, that is not to say that a subcritical pipeline system should not be used for transporting CO_2. There may be specific cases where the subcritical pipeline system is more economical. However, for such specific cases, detailed engineering is required and does not fall within the scope of this report. Therefore, the remaining sections of this report will only consider the use of the supercritical pipeline system for CO_2 transmission.

TABLE 7.12

COMPARISON OF CO_2 TRANSPORTATION COSTS   BASIS: Flat Terrain

FOR 50 MMSCFD OF CO_2 @ 100 MILES

Method of Analysis And Rate-Of-Return	Supercritical Transport Cost ($/MSCF CO_2)	Subcritical Transport Cost ($/MSCF CO_2)	Liquid Transport Cost ($/MSCF CO_2)
DISCOUNTED CASH FLOW @ 10%	0.21	0.41	0.86
DISCOUNTED CASH FLOW @ 15%	0.29	0.52	1.07
DISCOUNTED CASH FLOW @ 20%	0.38	0.66	1.30
DISCOUNTED CASH FLOW @ 25%	0.43	0.81	1.57
UTILITY FINANCING (FERC) @ 8%	0.22	0.41	0.86
UTILITY FINANCING (FERC) @ 10%	0.24	0.45	0.94

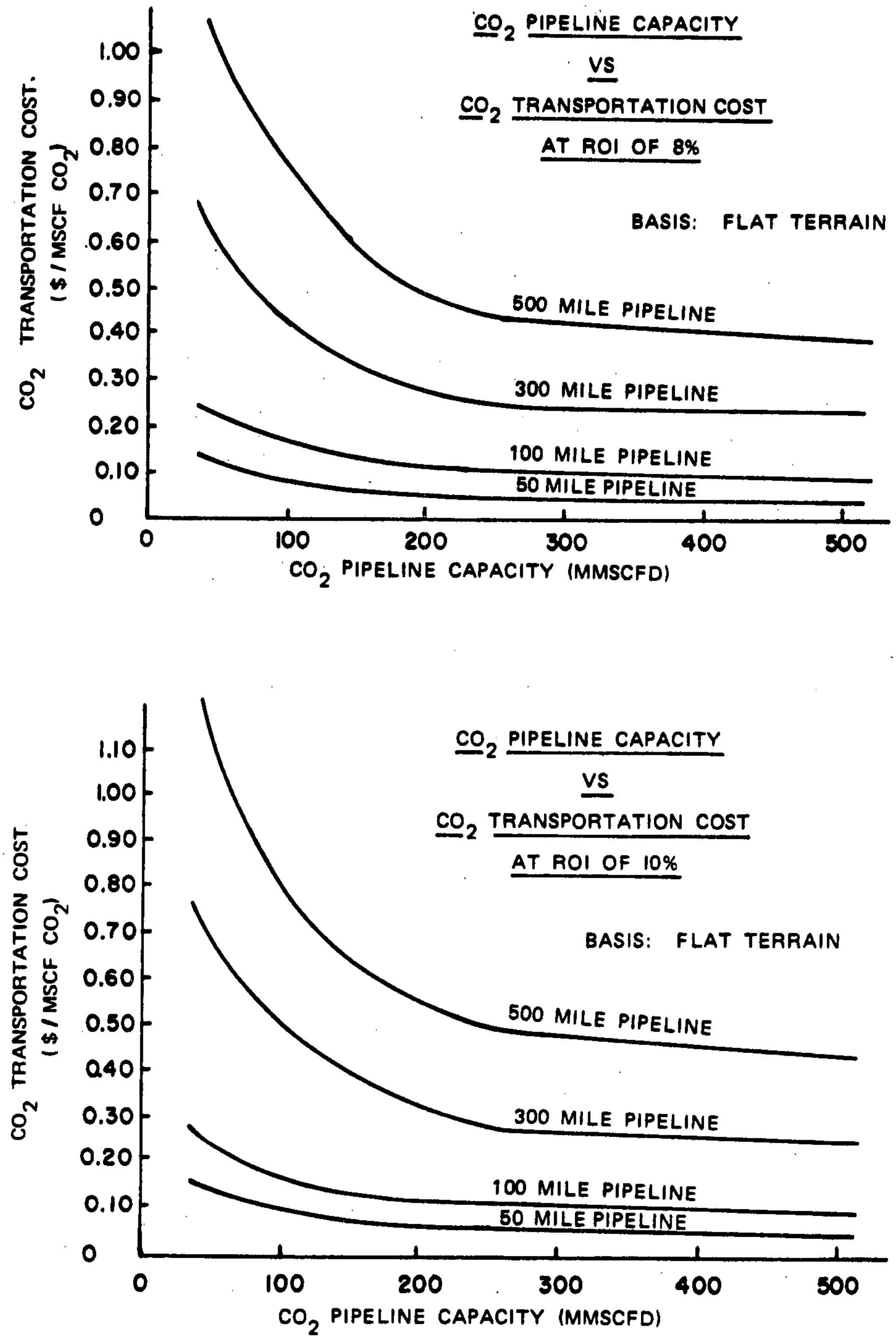
7.3.3 CO_2 Transportation Costs

In Section 7.3.2, the supercritical pipeline system was shown as the most economical method for transporting CO_2. Figures 7.5a and 7.5b show the CO_2 transportation costs associated with transporting CO_2 at various flow rates and at various transmission lengths. These curves are based on a flat terrain pipeline system at an 8% and 10% rate-of-return using the utility financing method and at a 10%, 15%, 20% and 25% rate-of-return using the discounted cash flow method.

Within this section are graphs which can estimate CO_2 transportation costs for any supercritical pipeline system, regardless of terrain and geographic affect. The general procedure involved in estimating CO_2 transportation costs and a general explanation concerning the application of estimating terrain adjustment factors follows below. In Part 8 of this report, these estimating terrain adjustment factors are used directly to determine CO_2 transportation costs for specific cases.

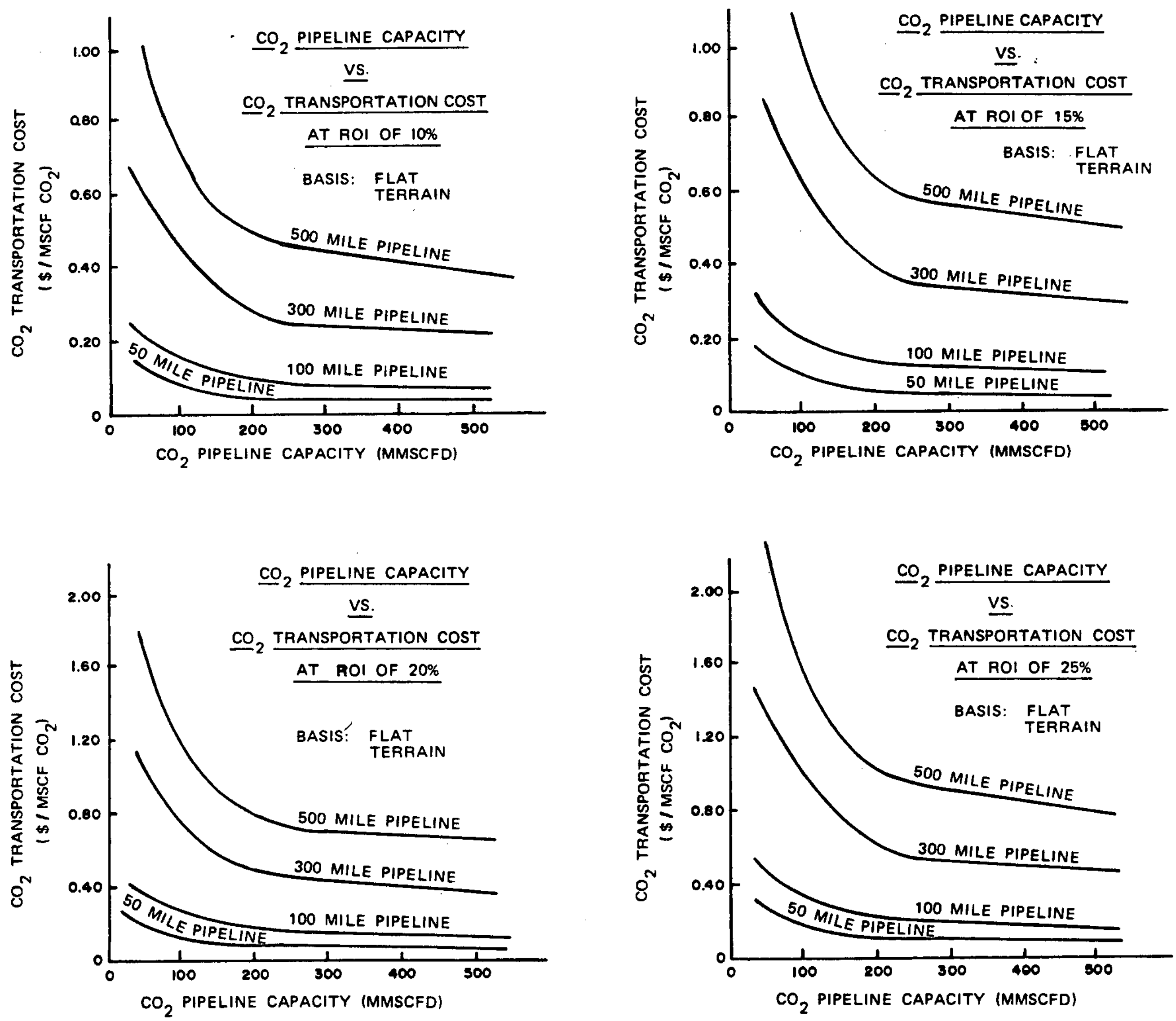
Before any material in this section can be utilized, certain pipeline variables and parameters require defining. These variables and parameters are listed below.

1. CO_2 pipeline capacity (MMSCFD CO_2)
2. Total pipeline transmission length (miles)
3. Elevation at CO_2 source site (feet)



RATE-OF-RETURN CURVES FOR SUPERCRITICAL PIPELINE BASED ON UTILITY FINANCING METHOD

FIGURE 7.5a



RATE-OF-RETURN CURVES FOR SUPERCRITICAL PIPELINE BASED ON DISCOUNTED CASH FLOW

FIGURE 7.5b

4. Elevation at CO_2 user site (feet)

5. Total pipeline transmission length in rolling hill terrain (miles)

6. Total pipeline transmission length in rugged terrain (miles)

7. Total pipeline transmission length crossing rivers (miles)

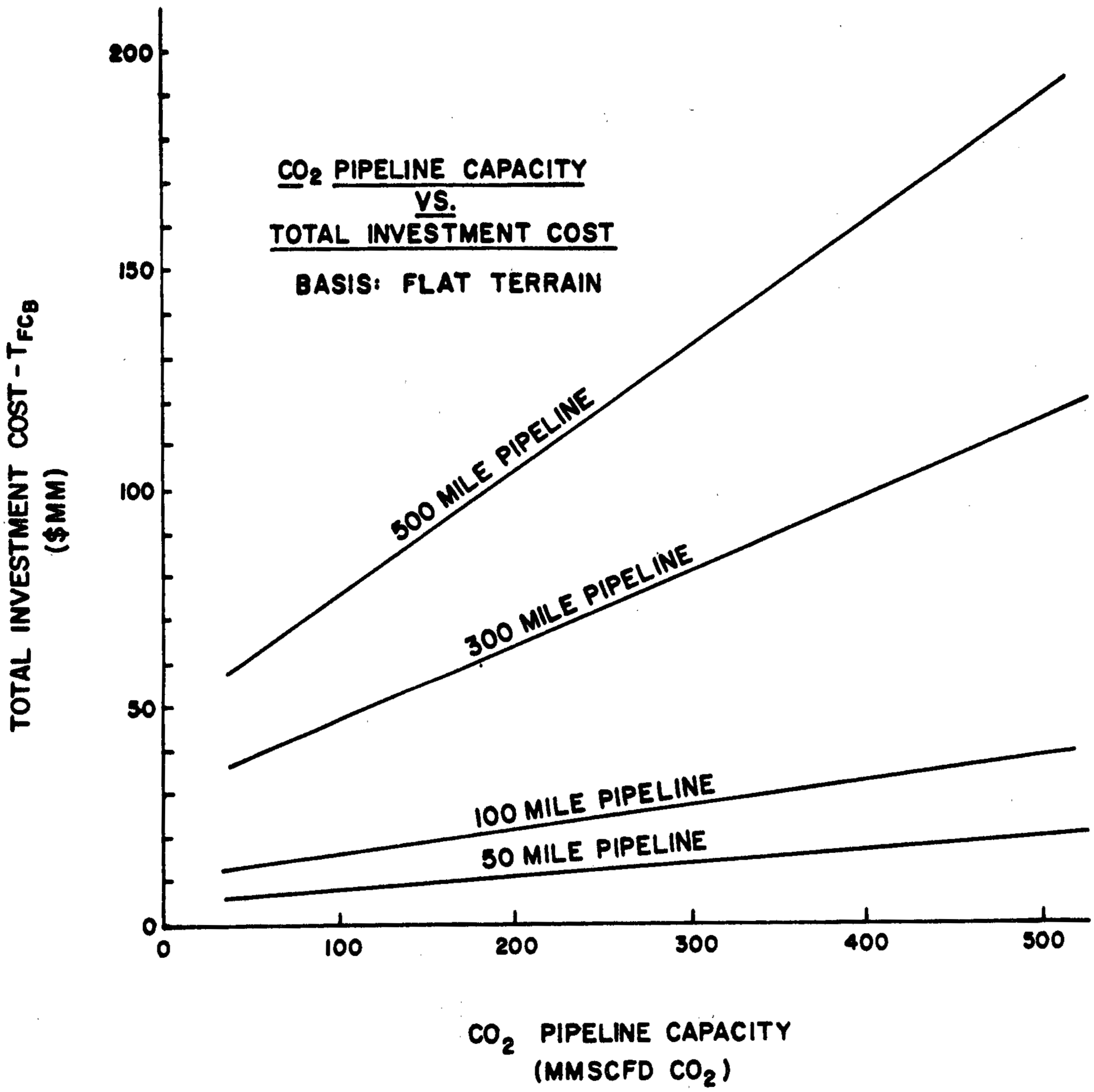
The initial step in estimating CO_2 transportation costs is to determine a total fixed capital investment cost (base cost) for the pipeline system. Figures 7.6 and 7.7 show these investment costs. Figure 7.6 shows the total pipeline investment cost as a function of CO_2 flow rate and transmission length. Figure 7.7 shows an average per mile investment cost as a function of CO_2 flow rate. Since the investment costs obtained from Figure 7.7 are average costs per mile, these costs are not as accurate as those costs shown in Figure 7.6 where transmission length is a parameter. Nonetheless, Figure 7.7 does provide a good investment cost estimate and can be used if desired in determining the pipeline's base cost.

Estimated terrain adjustment factors are added directly to the base costs which, in turn, reestimates the total fixed capital investment costs used in determining the CO_2 transportation costs. Equation 7.1 below shows this relationship.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \qquad (7.1)$$

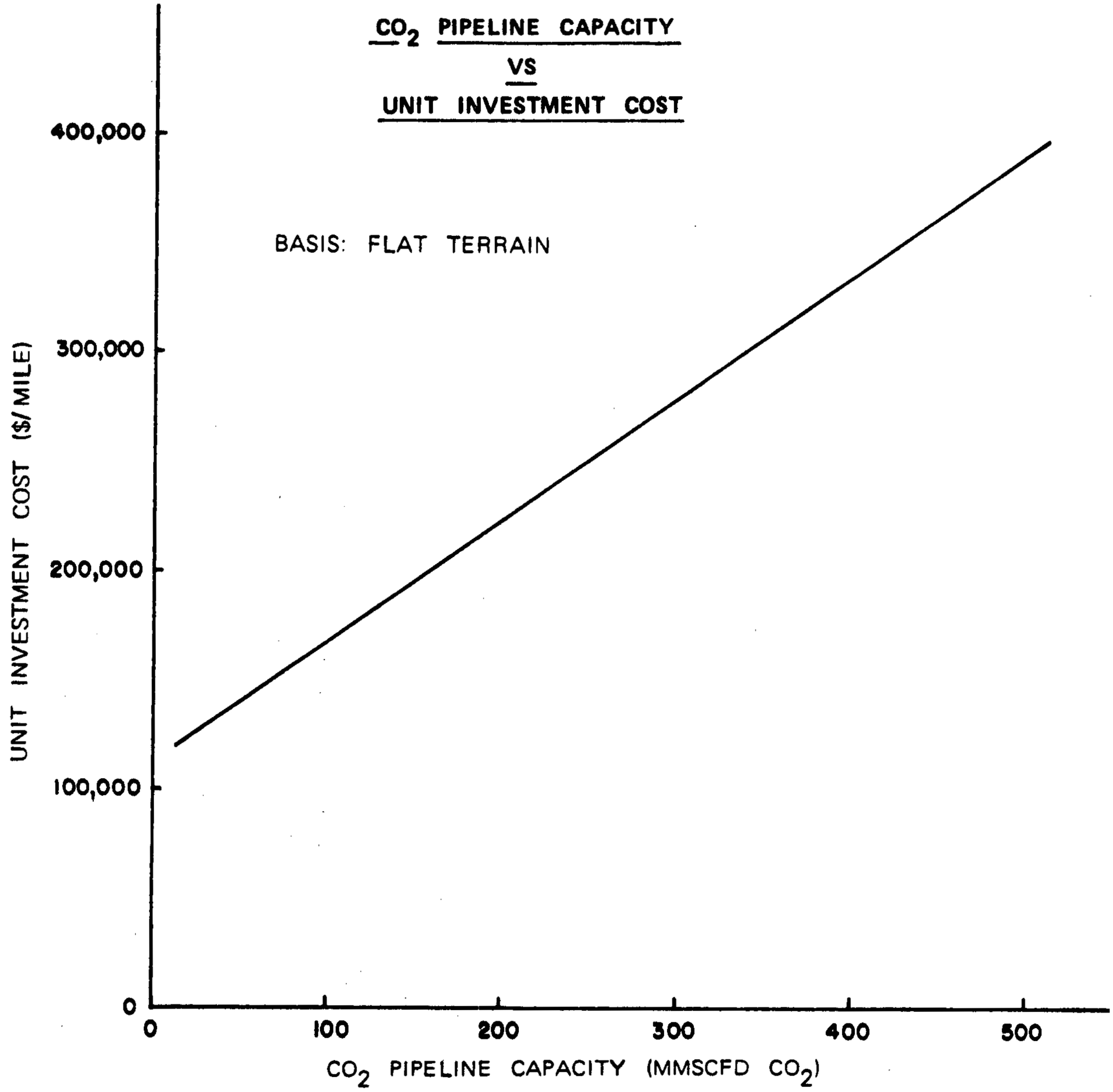
Where

T_{FC_A} = Total fixed capital investment cost with terrain adjustment



TOTAL PIPELINE INVESTMENT COST - T_{FC_B}
FOR A SUPERCRITICAL CO_2
PIPELINE

FIGURE 7.6



PER MILE PIPELINE INVESTMENT COSTS WITHOUT TERRAIN ADJUSTMENT FOR A SUPERCRITICAL PIPELINE

FIGURE 7.7

T_{FC_B} = Total fixed capital investment cost without terrain adjustment (See Figures 7.6 & 7.7)

$M_{R.T.}$ = Total miles of pipeline in rugged terrain

$\$\Delta R.T.$ = Rugged terrain adjustment factor (See Figure 7.10)

$M_{R.H.}$ = Total miles of pipeline in rolling hill terrain

$\$\Delta R.H.$ = Rolling hill terrain adjustment factor (See Figure 7.9)

$M_{R.C.}$ = Total miles of pipeline crossing rivers

$\$\Delta R.C.$ = River crossing adjustment factor (See Figure 7.11)

$\$\Delta H$ = Pipeline differential elevation cost factor (See Figures 7.8a & 7.8b)

The remainder of this section is devoted to determining CO_2 transportation costs using terrain adjustment factors and applying these terrain adjustment factors to Equation 7.1.

Figures 7.8a and 7.8b determine the pipeline differential elevation cost factor ($\$\Delta H$). Specifically, these two figures determine the change in compressor station costs in relation to the base cost. The steps outlined below determine this pipeline differential elevation cost factor ($\$\Delta H$).

Use of Figure 7.8a -

1. Based on a flow rate closest to the one in question, choose either the 50 MM, the 125 MM, the 250 MM or the 500 MMSCFD flow rate curve.
2. Determine the differential elevation of the pipeline.

   $\Delta H = H_{in} - H_{out}$

where

H_{in} = pipeline elevation @ CO_2 source site (feet)

H_{out} = pipeline elevation @ CO_2 user site (feet)

3. Locate the differential elevation of the pipeline on the "ΔH (Elevation Difference)" axis and select the appropriate pipeline transmission length. From these parameters, the number of required pipeline compressors is found on the vertical axis (always go to the next higher number of required compressors if a fraction).

   Example: Flow rate = 125 MMSCFD CO_2

   ΔH = 3000 feet

   Pipeline Length = 300 miles

   $\therefore$ Number of Required Compressors = 3

Use of Figure 7.8b -

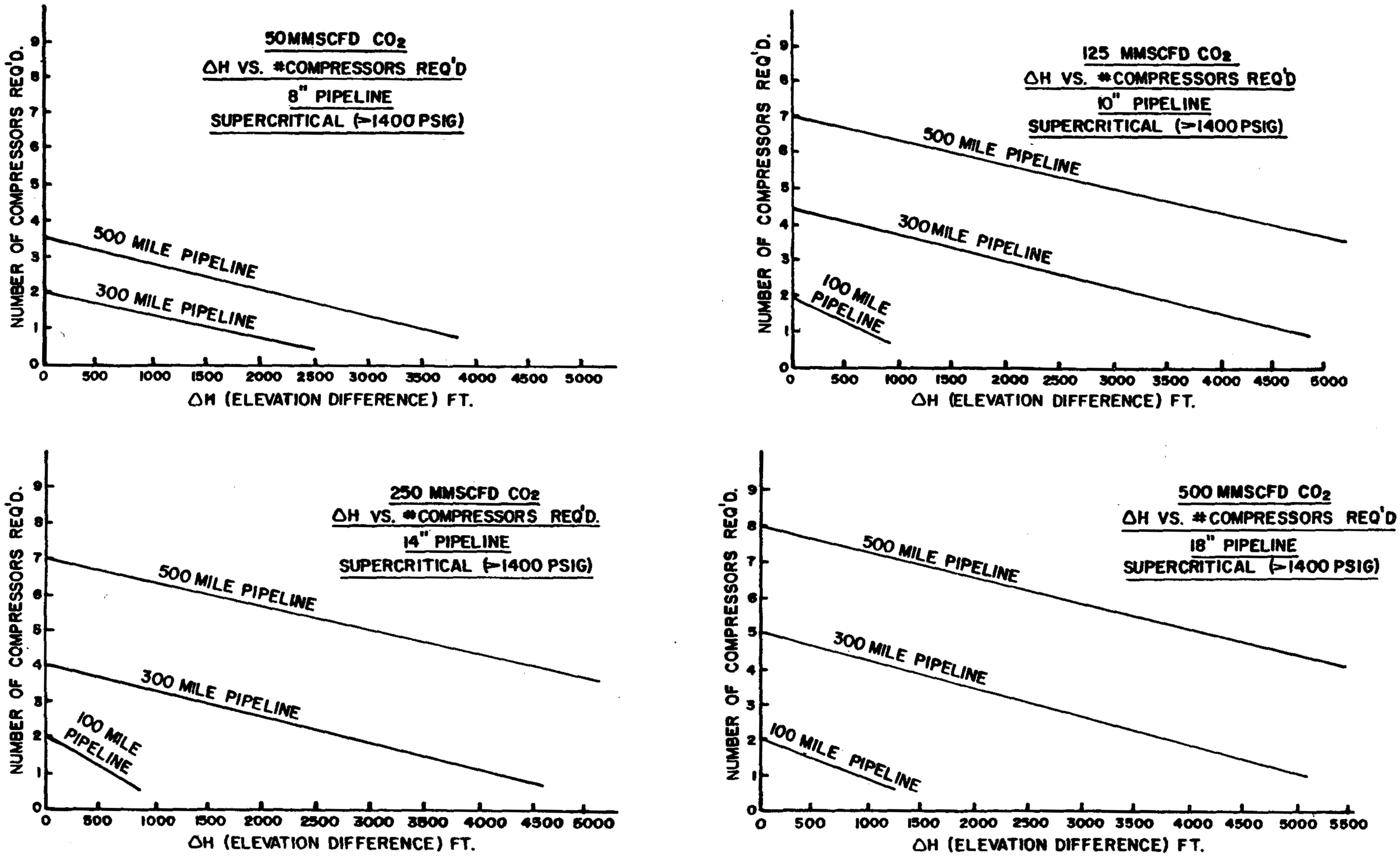
1. Locate the number of required pipeline compressors on the horizontal axis and select the appropriate pipeline transmission length. From these parameters, the pipeline differential elevation cost factor (\$$\Delta$H) is found on the vertical axis.

   Example: Flow Rate = 125 MMSCFD CO_2

   Number of Required Compressors = 3

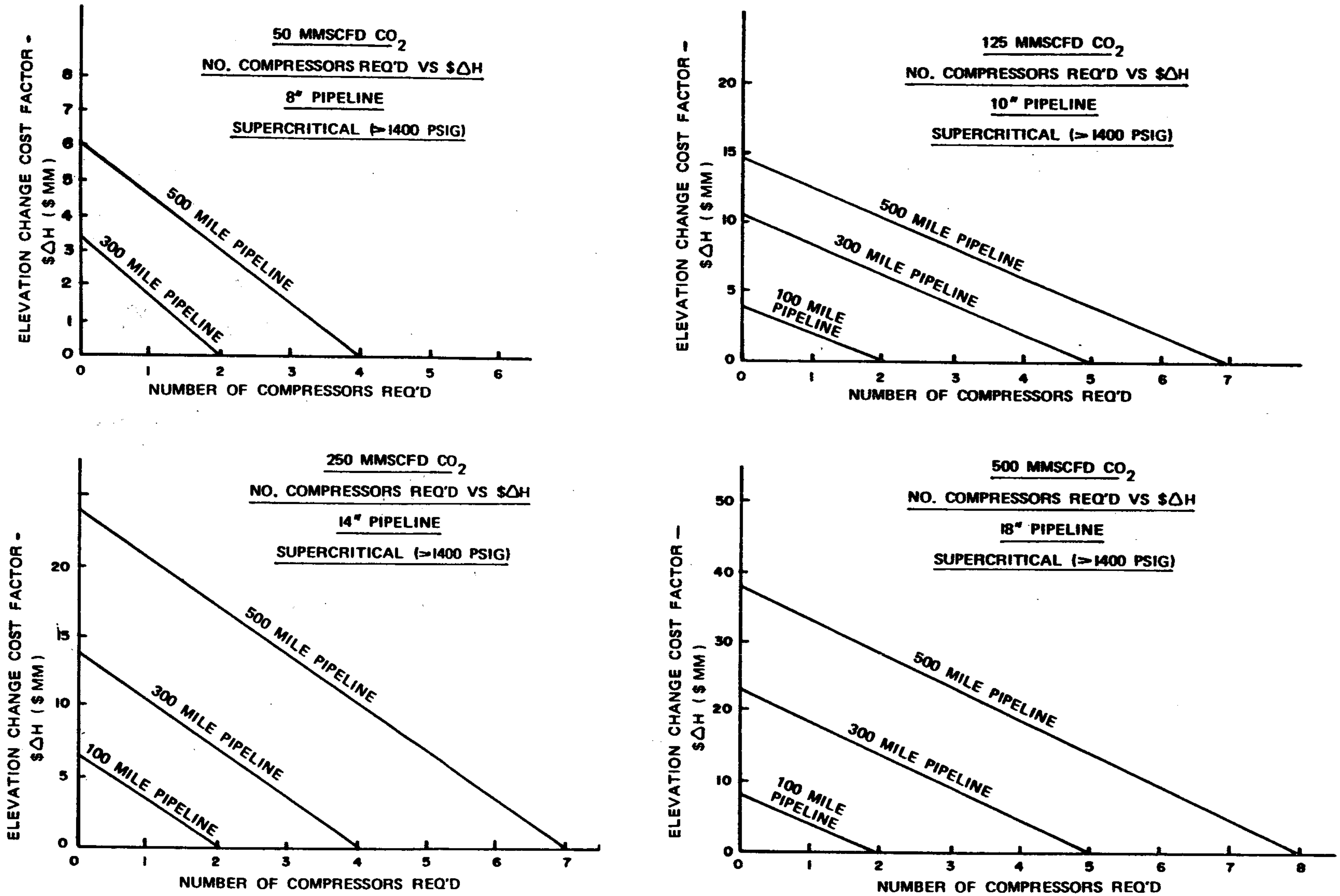
   Pipeline Length = 300 miles

   $\therefore$ \$$\Delta$H = \$4,000,000



PIPELINE DIFFERENTIAL ELEVATION AFFECT

FIGURE 7.8a



PIPELINE DIFFERENTIAL ELEVATION COST FACTOR

FIGURE 7.8b

Figures 7.9, 7.10 and 7.11 determine the rolling hill terrain cost factor ($ΔR.H.), the rugged terrain cost factor ($ΔR.T.), and the river crossing terrain cost factor ($ΔR.C.), respectively. The method used in determining the three terrain cost factors is similar for each cost factor. Therefore, the instructions described below, outline the procedure used in determining each terrain cost factor. Use of Figures 7.9, 7.10, and 7.11 -

1. Locate the known CO_2 pipeline flow rate (capacity) on the horizontal axis.
2. Locate the intersection of the CO_2 pipeline capacity on the average pipeline transmission length line.
3. The terrain cost factor is located on the vertical axis.

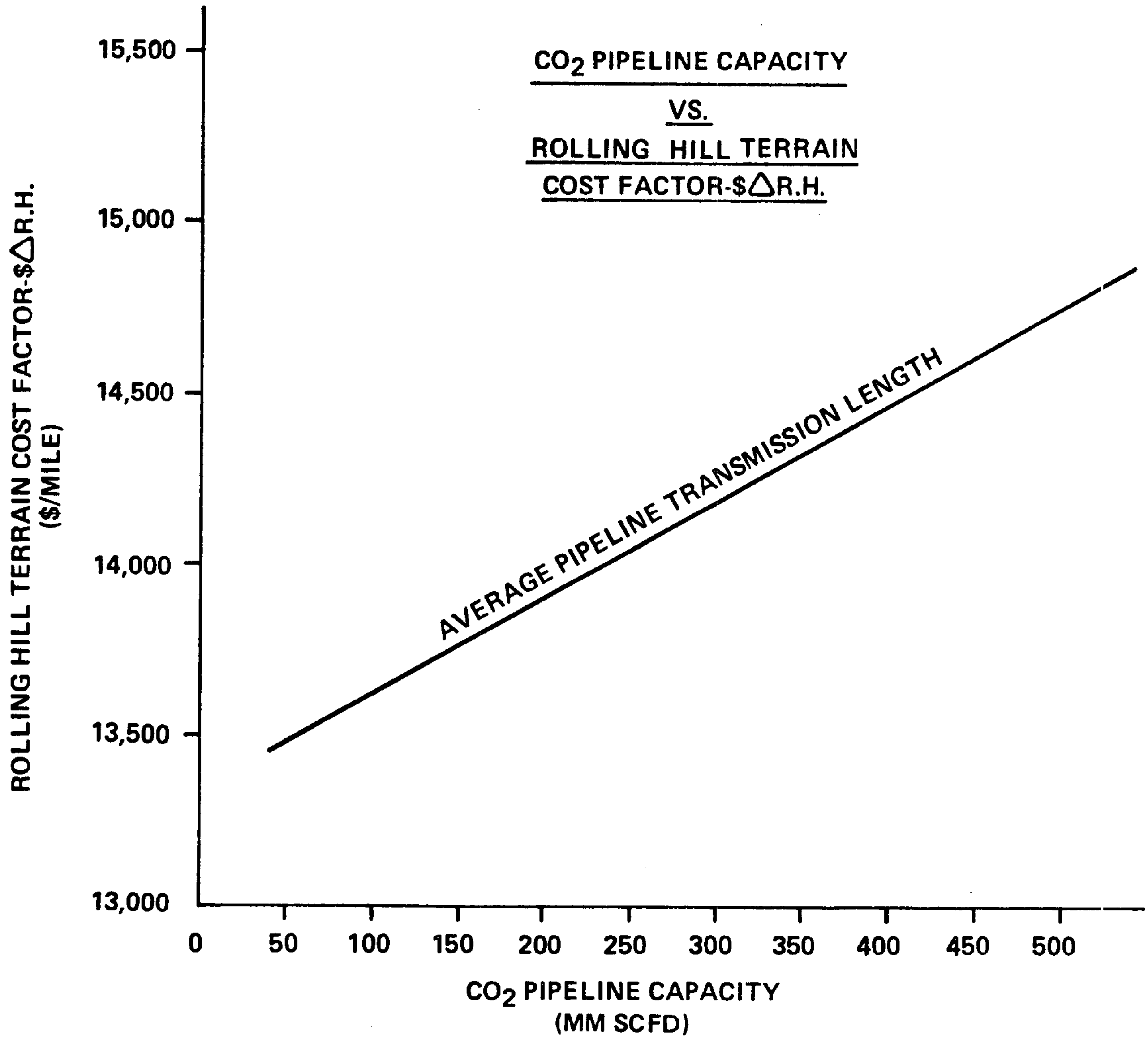
   Example: 20 miles of rolling hill terrain

   Flow rate = 125 MMSCFD CO_2

   Pipeline length = 300 miles

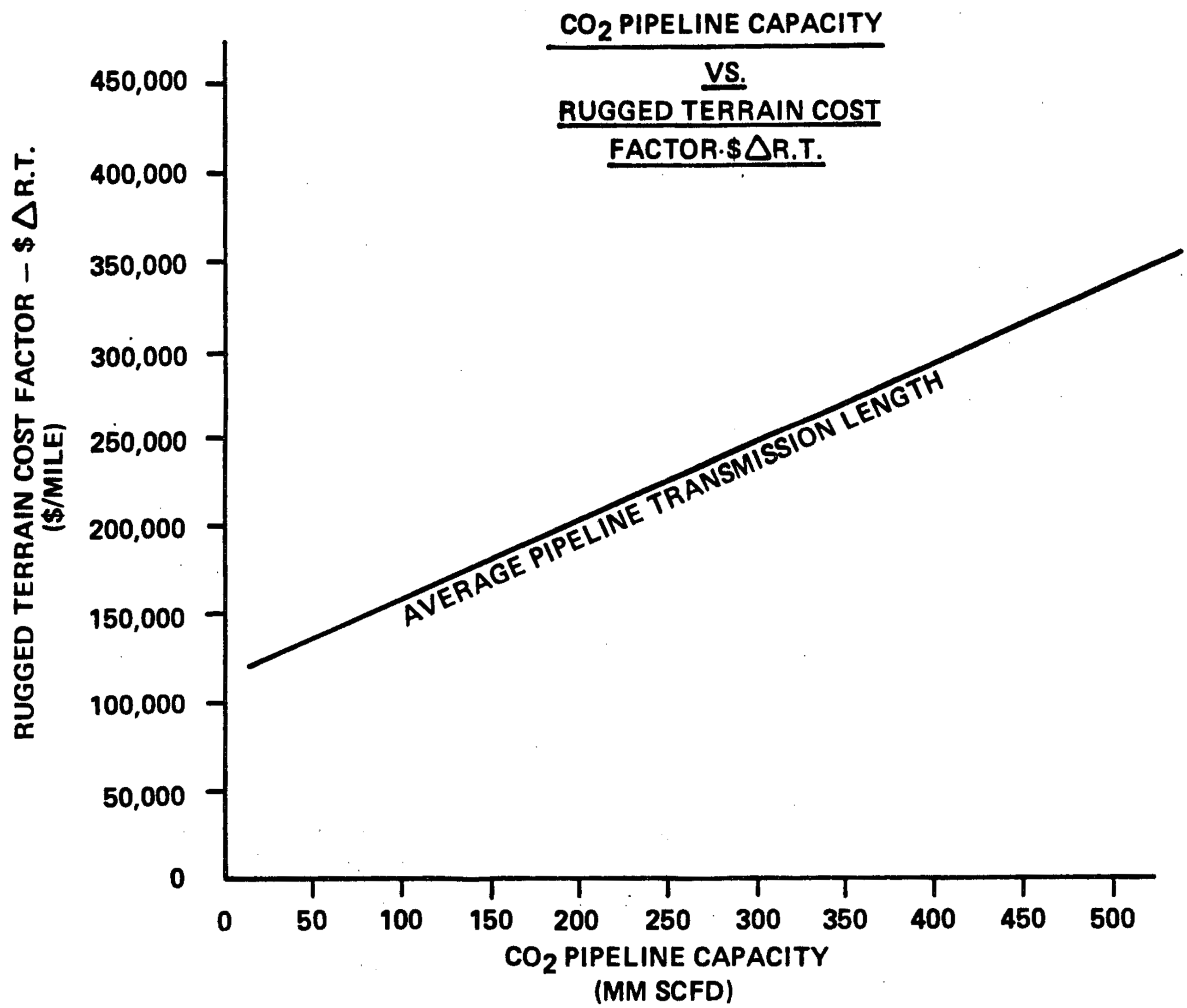
   $ΔR.H. = $13,700 miles

Once the terrain cost factors are determined and the total fixed capital investment is adjusted (See Equation 7.1), representative CO_2 transportation costs for a supercritical pipeline can be determined from Figures 7.12a and 7.12b. The following example determines the CO_2 transportation cost via supercritical pipeline and demonstrates the use of Equation 7.1 and Figures 7.6, 7.12a and 7.12b.



ROLLING HILL TERRAIN COST FACTOR-
$ΔR.H.

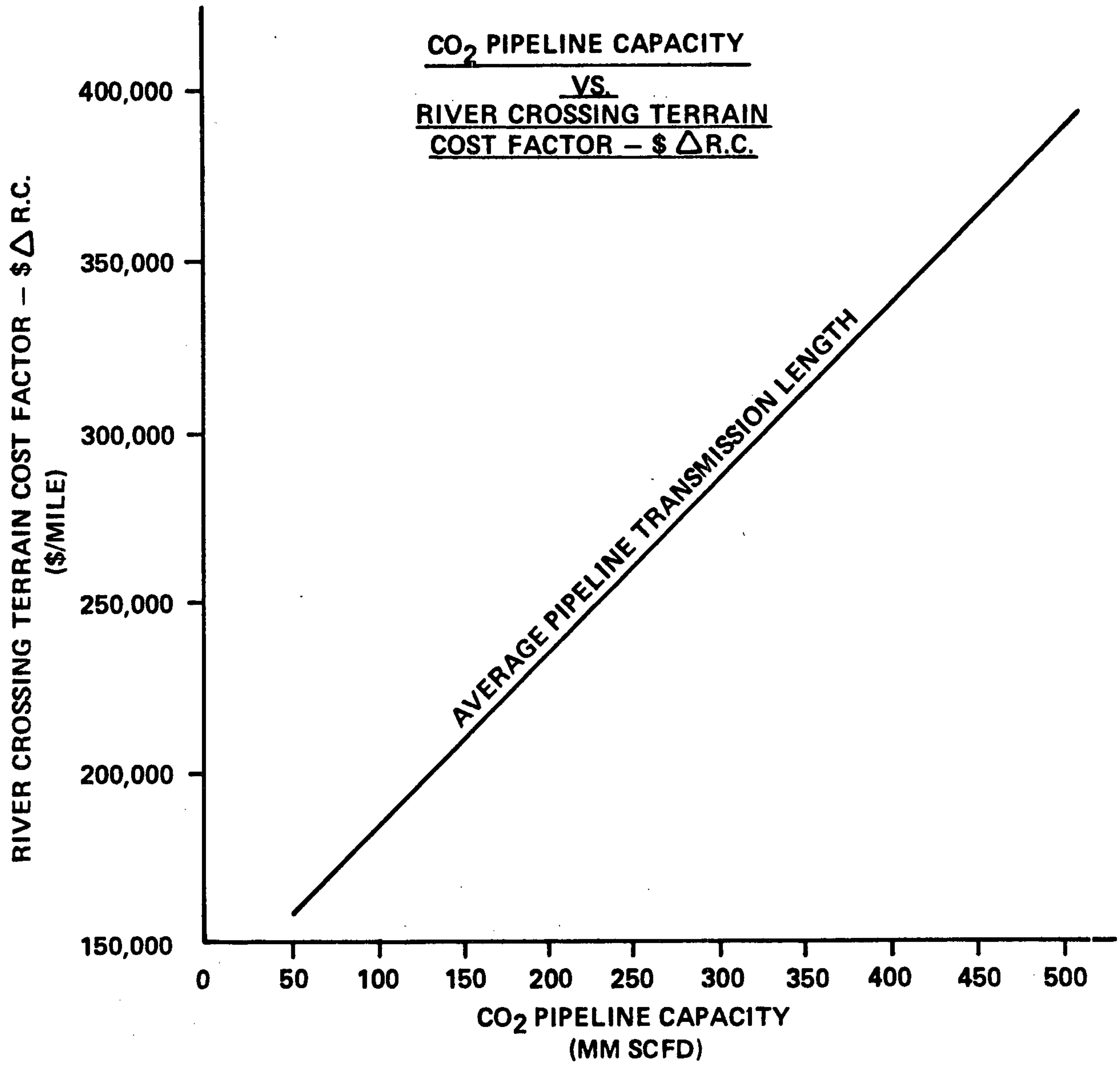
FIGURE 7.9



RUGGED TERRAIN COST FACTOR–

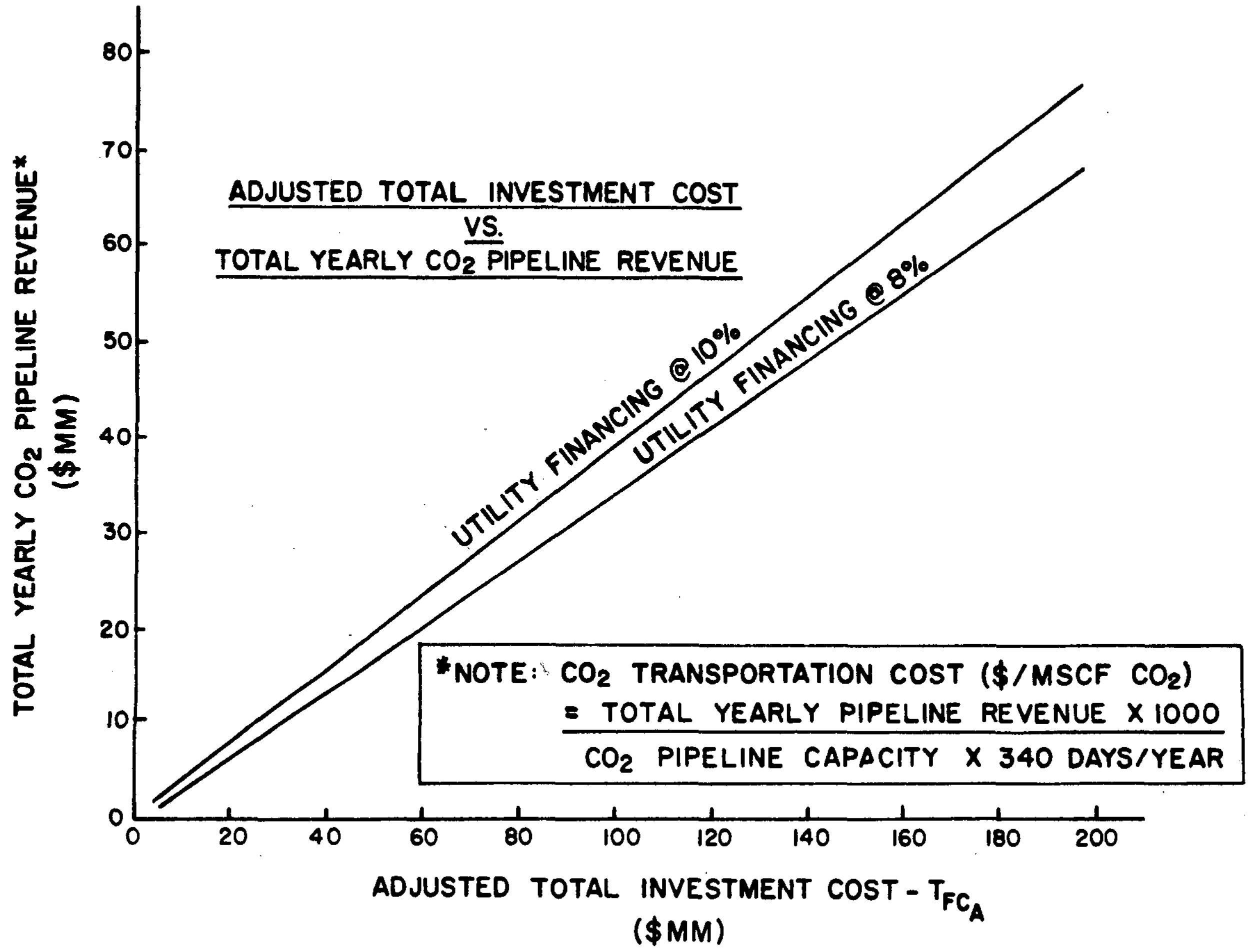
$\$\Delta$R.T.

FIGURE 7.10



RIVER CROSSING TERRAIN COST FACTOR–
$ΔR.C.

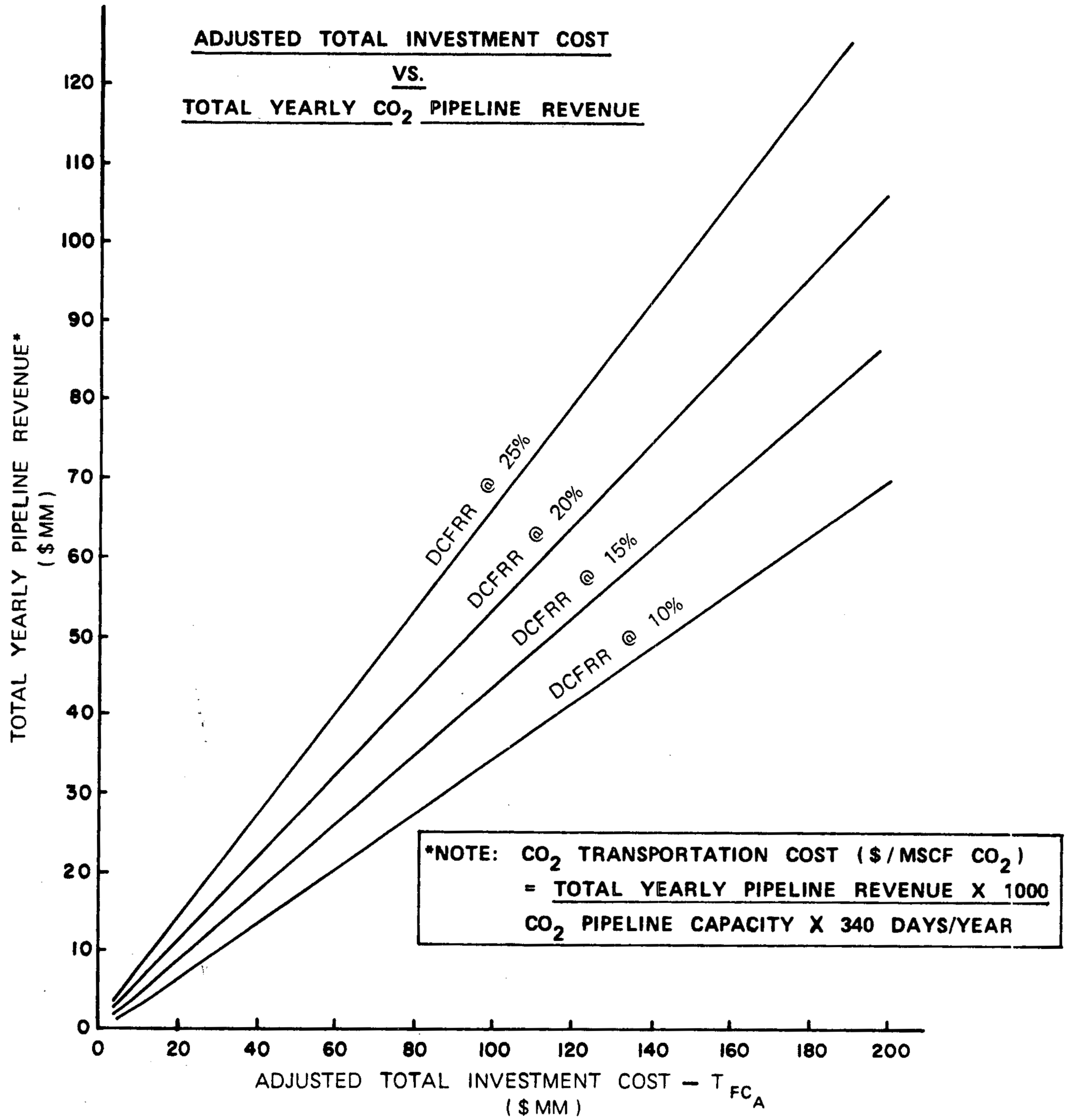
FIGURE 7.11



$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H. \qquad \text{EQ. 7.1}$$

CO_2 TRANSPORTATION COST FOR A SUPERCRITICAL PIPELINE BASED ON UTILITY FINANCING OF 8% AND 10%

FIGURE 7.12a



$$T_{FC_A} = T_{FC_B} + M_{R.T.}\ (\$\Delta R.T.) + M_{R.H.}\ (\$\Delta R.H.) + M_{R.C.}\ (\$\Delta R.C.) - \$\Delta H \qquad \text{EQ. 7.1}$$

CO_2 TRANSPORTATION COST FOR A SUPERCRITICAL PIPELINE BASED ON DCFRR OF 10%, 15%, 20% AND 25%

FIGURE 7.12 b

Example
Problem:

Determine the CO_2 transportation cost at a rate-of-return of 10% using both discounted cash flow and utility financing methods for a 275 mile, 200 MMSCFD CO_2 supercritical pipeline. The following pipeline parameters and terrain factors are given below for Equation 7.1:

- $\$\Delta H = \$2,000,000$
- $M_{R.T.} = 0$ miles
- $M_{R.H.} = 75$ miles @ $\$\Delta R.H. = \$13,850/\text{mile}$
- $M_{R.C.} = 15$ miles @ $\$\Delta R.C. = \$226,000/\text{mile}$

Equation 7.1 is:

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H$$

Example
Solution:

$$T_{FC_A} = T_{FC_B} + 75 \text{ miles } (\$13,850/\text{mile}) + 15 \text{ miles } (\$226,000/\text{mile}) - \$2,000,000$$

$$T_{FC_A} = T_{FC_B} + \$2,429,000$$

From Figure 7.6 -

$$T_{FC_B} = \$55,000,000$$

$$T_{FC_A} = \$55,000,000 + \$2,429,000$$

$$= \$57,429,000$$

Using T_{FC_A} = \$57,429,000 and Figure 7.12a

- Total yearly CO_2 pipeline revenue = \$22,000,000 @ 10%

- CO_2 transportation cost = $\dfrac{\$22,000,000 \times 1000}{200 \text{ MMSCFD} \times 340 \text{ Day/YR}}$

  = \$0.32/1000 SCF CO_2

AND

Using T_{FC_A} = \$57,429,000 and Figure 7.12b

- Total yearly CO_2 pipeline revenue = \$19,000,000 @ 8%

- CO_2 transportation cost = $\dfrac{\$19,000,000 \times 1000}{200 \text{ MMSCFD} \times 340 \text{ Day/YR}}$

  = \$0.28/1000 SCF CO_2

Example
<u>Results</u>:

CO_2 transportation cost @ 8%
using utility financing = <u>\$0.32/1000 SCF CO_2</u>

CO_2 transportation cost @ 10%
using discounted cash flow = <u>\$0.28/1000 SCF CO_2</u>

7.4 CONCLUSION

In Part 7.0 of this report, we discussed pipeline systems used in transporting CO_2 for enhanced oil recovery. The three pipeline systems studied are:

1. The supercritical pipeline system (operating pressure greater than 1400 psig)
2. The subcritical pipeline system (operating pressure less than 1000 psig)
3. The liquid pipeline system

Each system was studied and evaluated on fixed capital investments, on direct and indirect operating costs, and on rates-of-return using discounted cash flow and utility financing methods. Based on these evaluations, we conclude that the supercritical pipeline system is the most economical pipeline system available for transporting CO_2 for enhanced oil recovery.

CO_2 transportation costs incurred via the supercritical pipeline system vary considerably depending on flow rate, transmission length and return on investment. To illustrate, at a flow rate of 125 MMSCFD CO_2 and a rate-of-return on investment of 15%, the CO_2 transportation costs are \$.09/MSCF CO_2 and \$.84/MSCF CO_2 for a 50 mile and 500 mile supercritical pipeline system, respectively. In addition, at a flow rate of 250 MMSCFD CO_2 and a rate-of-return on investment of 15%, the CO_2 transportation costs are \$.07/MSCF CO_2 and \$.59/MSCF CO_2 for a 50 mile and 500 mile supercritical pipeline system, respectively. Therefore, the results dictate the difficulty in predicting generalized CO_2 transportation costs. However, the preceding results demonstrate that the greater the CO_2 pipeline capacity and the shorter the pipeline transmission length, then the less expense involved in transporting CO_2.

Therefore, we recommend using a <u>supercritical pipeline system</u> and, when possible, transporting large CO_2 quantities at short distances.

SPECIFIC CASES

PART 8

8.1 INTRODUCTION

The objective of this section is to demonstrate the use of the information developed in the preceding sections in evaluating the total cost of CO_2 for EOR. This is illustrated by formulating some specific (typical) cases of CO_2 sources and the candidate oil reservoirs for potential EOR. The information used in such evaluations include the survey of CO_2 sources and, depending on the type of selected source, the associated costs of CO_2 supply to the candidate oil reservoir. The costs may consist of purchase costs, production and processing costs, and transportation costs. The CO_2 injection cost in the oil field is not included and is not within the scope of this report.

The total cost developed for the cases in this section is for general guidance. There are some undefined factors so accurate analysis is not possible in the scope of this report. For example, the cost of purchase of CO_2 from a process plant source or power plant flue gas will depend on the demand of CO_2. The accurate cost for transporting CO_2 will require an alternate pipeline route study consistent with topography. Pipeline routes selected for the specific cases follow, as much as possible, existing product pipeline routes. In addition, pipeline lengths have been approximated based on available geographical data. These pipeline lengths, in turn, are used in conjunction with the various terrain factors included in this report. The cost of transportation will also depend on the economic analysis procedure (Utility Financing or DCFRR) used by a pipeline company.

The cases selected in this section are based on the information developed by GURC. GURC evaluated candidate oil reservoirs in terms of estimated EOR by CO_2 miscible flooding and the estimated net new CO_2 requirement per barrel of recovered oil. It is appropriate to mention at this point that these are planning estimates of EOR and net CO_2 requirement and probably will change when current research and field tests resolve some of the questions concerning the process.

Three cases, as described and evaluated in Sections 8.2, 8.3, and 8.4, are selected on the basis of industry interest. These cases are:

1. CO_2 supply to West Texas oil reservoir
2. CO_2 supply to South Mississippi oil reservoir
3. CO_2 supply to South Louisiana oil reservoir

8.2 CO_2 SUPPLY TO WEST TEXAS OIL RESERVOIR

Oil fields in Texas have 30% of the U. S. oil reserves. West Texas fields hold the majority of this oil. These fields have the greatest potential for EOR by CO_2 miscible flooding. There is substantial industry interest for EOR in this area. GURC estimates that there are 2 to 8 billion BBL of oil that is potentially recoverable by CO_2 flooding in West Texas. This is about 5 - 20% of the original oil in place in West Texas. Net new CO_2 requirement* for flooding of these fields is estimated to be 4 - 12 MSCF/BBL of recovered oil. This amounts to a potential new CO_2 requirement of 8 - 96 TSCF of potential net new CO_2 required.

In order to determine the total cost of CO_2 to these fields, three different CO_2 sources are evaluated separately in Sections 8.2.1, 8.2.2 and 8.2.3. The delivery point of CO_2 from the interstate pipeline to the candidate reservoirs in the area, as shown in Figure 8.1, is assumed to be at the intersection of Yoakum, Terry and Gaines counties in West Texas. An average net CO_2 requirement of 8 MSCF/BBL of

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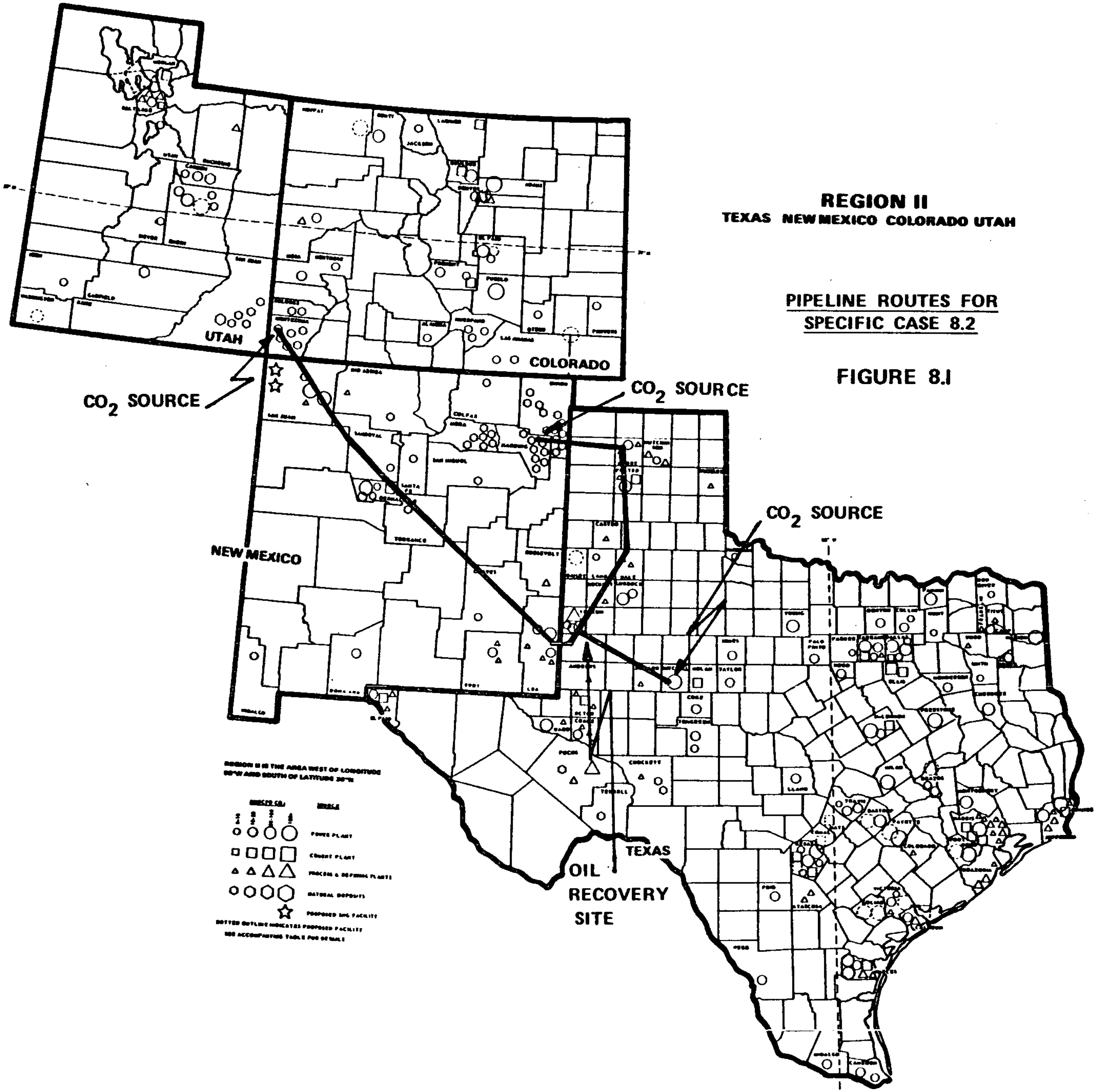
* Net new CO_2 = CO_2 actually injected - CO_2 recovered with production and recycled

= CO_2 remaining in reservoir - CO_2 supplied to field recovered oil will be used for this evaluation. The following three CO_2 sources are considered:

1. CO_2 from Southwest Colorado
2. CO_2 from Northeast New Mexico
3. CO_2 from power plant flue gas, Mitchell County, West Texas



REGION II
TEXAS NEW MEXICO COLORADO UTAH

PIPELINE ROUTES FOR
SPECIFIC CASE 8.2

FIGURE 8.1

8.2.1. CO_2 From Southwest Colorado Source

Source

Natural CO_2 reservoirs in Southwest Colorado are located in McElmo Dome area in Montezuma County and the Doe Canyon-Dove Creek area in Dolores County. For the purpose of this specific case, the delivery point of the CO_2 to an interstate pipeline is assumed to be in McElmo Dome area in Montezuma County at 37N - 18W. CO_2 available from this source is of high purity (96+% CO_2) and will not need any purification with the exception of dehydration of the gas.

Basis

Net CO_2 requirement	: 8,000 SCF/BBL
Total net CO_2 requirement	: 300 MMSCFD
Total pipeline length	: 500 miles
Pipeline length in rugged terrain	: 328 miles
Pipeline length in rolling hill terrain	: 165 miles
Number of river crossings	: 7
Total length of river crossings	: 7 miles, approximate
Elevation difference between CO_2 source and an oil field	: 3,500 feet

Capital Investment - Costs

These investment costs represent capital investment by a producer of natural CO_2 in McElmo Dome area and by a pipeline company to transport CO_2 from McElmo Dome area to a candidate oil reservoir in West Texas.

The investment costs associated with the production of CO_2 are subject at this time to many uncertainties which significantly affect the value or cost of the CO_2 produced. Among these uncertainties are the cost of actual drilling, expected deliverability, and expected life of the fields. GURC has provided a range of costs in Section 5 that could be anticipated for this area. For application to the specific cases chosen we will assume a median CO_2 cost of $0.38/MSCF.

The pipeline investment cost incurred in transporting natural CO_2 from the McElmo Dome area to West Texas (see Figure 8.1 for pipeline route) is determined from the data found in Section 7.3.3. This data is based on a supercritical pipeline system, which is defined and shown in Part 7 as being the most economical pipeline system.

As shown in Section 7.3.3, Equation 7.1 (shown below for reference) is used to determine pipeline investment costs with adjustments for geographic terrain and elevation affects.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \qquad (7.1)$$

The data presented and defined below comes from the material found in Section 7.3.3:

1. Unadjusted pipeline investment cost -

   T_{FC_B} (Figure 7.6) = \$133,000,000

2. Length of pipeline in rugged terrain -

   $M_{R.T.}$ = 328 miles

3. Rugged terrain adjustment factor -

   $\$\Delta R.T.$ (Figure 7.10) = \$250,000/mile

4. Length of pipeline in rolling hill terrain -

   $M_{R.H.}$ = 165 miles

5. Rolling hill terrain adjustment factor -

   $\$\Delta R.H.$ (Figure 7.9) = \$14,200/mile

6. Length of pipeline crossing rivers -

   $M_{R.C.}$ = 7 miles

7. River crossing terrain adjustment factor -

   $\$\Delta R.C.$ (Figure 7.11) = \$285,000/mile

8. Pipeline elevation change adjustment factor -

   $\$\Delta H$ (Figures 7.8a and 7.8b) = \$7,500,000

9. Adjusted pipeline investment cost -

$$T_{FC_A} = \underline{\$212,000,000}$$

Therefore, the total pipeline investment cost in transporting natural CO_2 from the McElmo Dome area to West Texas is $212,000,000.

Total Cost of CO_2 To a Candidate Oil Reservoir Operator

The total cost of CO_2 to a candidate oil reservoir operator who is using CO_2 from McElmo Dome area, Southwest Colorado, is the sum of prices demanded by a producer of CO_2 and by a pipeline company transporting CO_2 to an oil field.

Assume that a producer of CO_2 has set the price of CO_2 at $0.38/MSCF, and delivers 300 MMSCFD of CO_2 at 2000 psig at pipeline inlet.

From Figure 7.12a -

Based on a rate-of-return of 8% using the Utility Financing Method and at a total pipeline investment cost of $212,000,000, the total yearly pipeline revenue received is $69,000,000. Hence, the transportation cost for a

300 MMSCFD CO_2 pipeline from the McElmo Dome area to West Texas is $0.68/MSCF CO_2.

Thus the total cost of CO_2 to a candidate oil reservoir operator in West Texas, with CO_2 source in McElmo Dome area, Southwest Colorado, is as follows:

Cost		$/MSCF
Production		0.38
Transportation		0.68
	Total	1.06

8.2.2 CO_2 From Northeast New Mexico

As discussed in Section 5.6, the Northeast New Mexico area appears to have substantial reserves of natural CO_2. These reserves cover portions of Union, Harding and Quay Counties. For the purpose of this specific case, the delivery point of CO_2 to an interstate pipeline is assumed to be in Harding County at 20N - 32E. CO_2 available from this source is of high purity (99+% CO_2) and will not need any purification with the exception of dehydration of the gas.

Basis

Net CO_2 requirement	: 8,000 SCF/BBL
Total net CO_2 requirement	: 500 MMSCFD
Total pipeline length	: 320 miles
Pipeline length in rolling hill terrain	: 316 miles
Number of river crossings	: 4
Total length of river crossings	: 4 miles, approximate
Elevation difference between CO_2 source and an oil field	: 1,750 feet

Capital Investment Costs

These investment costs represent capital investment by a producer of natural CO_2 in Northeast New Mexico and by a pipeline company for transporting CO_2 from the production site to a candidate oil reservoir in West Texas. As discussed in Section 5.6.4, compression of CO_2 is not included in the production of CO_2 which is produced at an average pressure of 140 psig. Therefore, for natural CO_2 produced in Northeast New Mexico, it is assumed that the compression would be done by a pipeline company.

For reasons outlined in Section 8.2.1 it is difficult at this time to evaluate the investment costs that are associated with CO_2 production. For New Mexico we will assume a CO_2 cost of $0.50/MSCF, which is the upper value analyzed by GURC and which spans a reasonable range of expected DCFRR values.

The pipeline investment compression cost is based on compressing 500 MMSCFD CO_2 from 140 psig to 2000 psig. As discussed in Section 6.5 of this report, compression costs can be determined using Equation 6.1 (shown below for reference) if compressor inlet pressures are greater than atmospheric, as in this case.

$$BHP = 2.6 \times 10^{-4}\ Q\ \{(\frac{2014.7}{P_1})^{0.18} - 1\} \qquad (6.1)$$

The procedure outlined in Section 6.5.2, essentially uses Equation 6.1 to determine a new compression flow rate, which, in turn, can be used to determine a pipeline investment compression cost. The basic steps and results shown below explain this procedure:

1. Calculate the actual compressor horsepower (BHP) required based on P_1 = 140 psig (155 psia) and Q = 500 MMSCFD CO_2 using Equation 6.1. BHP = 76,300
2. Using the BHP determined above, calculate a new CO_2 flow rate at P_1 = 14.7 psia. (This step factors

the flow rate so that Figure 6.36 can be used to determine pipeline investment compression costs which are based on 14.7 psia.)

$Q = 206$ MMSCFD CO_2

3. Determine the pipeline investment compression cost using Figure 6.36 and $Q_1 = 206$ MMSCFD CO_2.

   Pipeline compression cost = $23,000,000

Hence, in going from a source pressure of 140 psig to pipeline pressure of 2000 psig, an initial pipeline investment compression cost of $23,000,000 is incurred at 500 MMSCFD CO_2.

The pipeline investment cost incurred in transporting natural CO_2 from the Northeast New Mexico area to West Texas (see Figure 8.1 for pipeline route) is determined from the data found in Section 7.3.3. This data is based on a supercritical pipeline system, which is defined and shown in Part 7 as being the most economical pipeline system.

As shown in Section 7.3.3, Equation 7.1 (shown below for reference) is used to determine pipeline investment costs with adjustments for geographic terrain and elevation affects.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \qquad (7.1)$$

The data presented and defined below comes from the material found in Section 7.3.3:

1. Unadjusted pipeline investment cost -

   T_{FC_B} (Figure 7.6) = \$118,000,000

2. Length of pipeline in rolling hill terrain -

   $M_{R.H.}$ = 316 miles

3. Rolling hill terrain adjustment factor -

   $\$\Delta R.H.$ (Figure 7.9) = \$14,750/mile

4. Length of pipeline crossing rivers -

   $\$M_{R.C.}$ = 4 miles

5. River crossing terrain adjustment factor -

   $\$\Delta R.C.$ (Figure 7.11) = \$390,000/mile

6. Pipeline elevation change adjustment factor -

   $\$\Delta H$ (Figures 7.8a and 7.8b) = \$5,000,000

7. Adjusted pipeline investment cost -

   T_{FC_A} = $\underline{\$119,000,000}$

Therefore, the estimated total pipeline investment cost in transporting 500 MMSCFD of natural CO_2 from Northeast New Mexico to West Texas is $\underline{\$119,000,000}$.

Total Cost of CO_2 To A Candidate Oil Reservoir Operator

The total cost of CO_2 to an operator who is using CO_2 from the Northeast New Mexico area, is the sum of the prices demanded by a producer of CO_2 and by a pipeline company transporting CO_2 to an oil field in West Texas.

Assume that a producer of CO_2 has set the price of CO_2 at $0.50/MSCF and delivers 500 MMSCFD of CO_2 at 140 psig to a pipeline company.

From Figure 6.39 -

Based on a rate-of-return of 8% using the Utility Financing Method and a factored flow rate of 206 MMSCFD CO_2, the resulting total pipeline compression cost is $0.33/MSCF CO_2. Hence, in going from a source pressure of 140 psig to a supercritical pipeline pressure of 2000 psig, an initial compression cost of $0.33/MSCF CO_2 is incurred at 500 MMSCFD CO_2.

From Figure 7.12a -

Based on a rate-of-return of 8% using the Utility Financing Method and at a total pipeline investment cost of $119,210,000, the total yearly pipeline revenue received is $40,500,000.

Hence, the transportation cost from a 500 MMSCFD CO_2 pipeline from Northeast New Mexico to West Texas is <u>$0.24/MSCF CO_2</u>.

Thus, the total cost to a candidate oil reservoir operator in West Texas, receiving CO_2 from Northeast New Mexico, is as follows:

Cost		$/MSCF
Production		0.50
Compression		0.33
Transportation		0.24
	Total	1.07

8.2.3 <u>CO_2 From Power Plant Flue Gas, Mitchell County, West Texas</u>

<u>Source</u>

The power plant in Mitchell County, West Texas, is owned by Texas Electric Service and has a generation capacity of 827 MW. The plant operated at about 54% capacity in the year 1976. Available CO_2 in the flue gas from this plant at the operating capacity of 54% amounts to about 112 MMSCFD. Concentration of CO_2 in the flue gas varies from 8 to 12%.

Balance of the flue gas consists of nitrogen, oxygen and water vapor. Therefore, this source will require purification to produce CO_2 with 98% purity.

Basis

Net CO_2 requirement	: 8,000 SCF/BBL
Total net CO_2 requirement	: 112 MMSCFD
Total pipeline length	: 105 miles
Pipeline length in rolling hill terrain	: 75 miles
Pipeline length in flat terrain	: 30 miles
Number of river crossings	: None
Elevation difference between the source and an oil field	: Not appreciable

Capital Investment Costs

These investment costs represent capital investments by a company that builds the purification plant for recovery of CO_2 from flue gas and by a pipeline company for transporting CO_2 to a candidate oil reservoir. As discussed in Section 6.2, purification of flue gas is accomplished by using a regenerative solution of MEA. Investment cost for purification includes compression of product CO_2 to 2000 psig.

For 112 MMSCFD of CO_2 for recovery,

Plant investment cost (from Figure 6.5, Section 6.2.4) = <u>45 MM dollars</u>.

The pipeline investment cost incurred in transporting CO_2 from the power plant in Mitchell County, West Texas (see Figure 8.1 for pipeline route) is determined from the data found in Section 7.3.3. This data is based on a supercritical pipeline system, which is defined and shown in Part 7 as being the most economical pipeline system.

As shown in Section 7.3.3, Equation 7.1 (shown below for reference) is used to determine pipeline investment costs with adjustments for geographic terrain and elevation effects.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \qquad (7.1)$$

The data presented and defined below comes from the material found in Section 7.3.3:

1. Unadjusted pipeline investment cost -
   T_{FC_B} (Figure 7.6) = \$15,000,000
2. Length of pipeline in rolling hill terrain -
   $M_{R.H.}$ = 75 miles

3. Rolling hill terrain adjustment factor -

$\$\Delta R.H.$ (Figure 7.9) = \$13,700/mile

4. Adjusted pipeline investment cost -

T_{FC_A} = $\underline{\$16,000,000}$

Therefore, the total pipeline investment cost in transporting natural CO_2 from the power plant is $\underline{\$16,000,000}$.

Total Cost of CO_2 To A Candidate Oil Reservoir Operator

The total cost of CO_2 to an oil field operator who is using CO_2 from a flue gas purification plant in Mitchell County, West Texas, is the sum of the prices demanded by the power plant for sale of flue gas, by a processor of flue gas to recover CO_2 and by a pipeline company for transporting CO_2 to an oil field in West Texas.

As discussed in Section 4.1, purchase cost of a CO_2 source will depend on the demand of CO_2 for EOR efforts. Table 4.1 shows that the purchase cost of flue gas can vary from \$0.06 - \$0.12 per MSCF of CO_2.

For this case, assume flue gas cost = $\underline{\$0.09/\text{MSCF}}$.

Assume that a processor of flue gas sets the price of CO_2 based on 20% DCFRR and delivers 112 MMSCFD of CO_2 at 2000 psig.

From Figure 6.6 -

At 20% DCFRR, cost of purification = $0.88/MSCF of CO_2

From Figure 7.12a -

Based on a rate-of-return of 8% using the Utility Financing Method and at a total pipeline investment cost of $16,000,000, the total yearly pipeline revenue received is $4,000,000. Hence, the transportation cost for the 112 MMSCFD CO_2 pipeline from the power plant is $0.11/MSCF CO_2.

Thus, the total cost to a candidate oil reservoir operator in West Texas, receiving CO_2 recovered from the flue gas of a power plant located in Mitchell County, West Texas, is as follows:

Cost	$/MSCF
Flue gas purchase cost	0.09
Purification cost	0.88
Transportation cost	0.11
Total	1.08

8.3 CO_2 SUPPLY TO SOUTH MISSISSIPPI OIL RESERVOIR

GURC estimates that in Southern Mississippi there could be 100 million barrels of oil potentially recoverable by CO_2 flooding. Net new CO_2 requirement for flooding of these fields is estimated to be about 10 MSCF/BBL of recovered oil. This amounts to a potential net new CO_2 requirement of up to 1 TSCF. The delivery point of the CO_2 from an intrastate pipeline to the candidate oil reservoir as shown in Figure 8.2, is assumed to be in Lincoln County, located at 6N - 8E. The total cost of CO_2 to these fields is determined using the natural CO_2 from the Jackson Dome area of Rankin County in Mississippi. This evaluation is discussed in Section 8.3.1.

8.3.1 CO_2 From Mississippi

Source

The major high purity natural CO_2 deposits in Mississippi are located in Rankin and Madison Counties in the central part of the state. The higher purity gas lies in the northern portion of these deposits. The extreme southern portion contains high concentrations of hydrogen sulfide and hydrocarbons. Hydrogen sulfide concentration varies from a few ppm to about 10%.

For this specific case, it is assumed that a source with CO_2 concentration of about 90% and with hydrogen sulfide concentration of 2% will be used. Thus, CO_2 available from this source will need purification to produce CO_2 with 98% purity. Delivery point of the CO_2 to an intrastate pipeline is assumed to be in Rankin County, Mississippi, located at 7N - 4E.

Basis

Net CO_2 requirement	: 10,000 SCF/BBL
Total net CO_2 requirement	: 117 MMSCFD
Total pipeline length	: 80 miles
Pipeline length in rolling hill terrain	: 60 miles
Pipeline length in flat terrain	: 19 miles
Number of river crossings	: one
Length of river crossings	: 1 mile
Elevation difference between the source and the oil field	: not appreciable

Capital Investment Costs

These investment costs represent capital investments by a producer of natural CO_2 in Mississippi, by a company that builds the purification plant for cleaning up the CO_2, and by a pipeline company for transporting CO_2 from a purification plant site to a candidate oil reservoir in

Southern Mississippi. At this time the variables which most affect production costs, namely drilling costs, expected deliverability, and expected life of the fields have not been well enough defined to give a most likely cost of CO_2 produced. GURC has analyzed in Section 5.7.4 the conditions associated with CO_2 costing \$0.25 and \$0.50/MSCF for delivery at 2000 psig. It is felt that CO_2 will have to be priced at the higher of these two values in order for the producer to achieve a reasonable DCFRR. So for this analysis we will use <u>\$0.50/MSCF</u> as the cost associated with CO_2 production.

As discussed in Section 6.3 purification of H_2S laden natural CO_2 is best accomplished by selective absorption with Selexol. Investment cost for the purification of a 90% CO_2, 2% H_2S gas available at a minimum inlet pressure of 1000 psig is

For 117 MMSCFD of CO_2 for recovery,
Plant investment cost (Figure 6.32) = <u>50 MM dollars</u>

The pipeline investment cost incurred in transporting natural CO_2 from Mississippi (see Figure 8.2 for pipeline route) is determined from the data found in Section 7.3.3. This data is based on a supercritical pipeline system, which is defined and shown in Part 7 as being the most economical pipeline system.

As shown in Section 7.3.3, Equation 7.1 (shown below for reference) is used to determine pipeline investment costs with adjustments for geographic terrain and elevation affects.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \quad (7.1)$$

The data presented and defined below comes from the material found in Section 7.3.3:

1. Unadjusted pipeline investment cost -

   T_{FC_B} (Figure 7.6) = \$12,000,000
2. Length of pipeline in rolling hill terrain -

   $M_{R.H.}$ = 60 miles
3. Rolling hill terrain adjustment factor -

   $\$\Delta R.H.$ (Figure 7.9) = \$13,700/mile
4. Length of pipeline crossing rivers -

   $\$M_{R.C.}$ = 1 mile
5. River crossing terrain adjustment factor -

   $\$\Delta R.C.$ (Figure 7.11) = \$195,000/mile
6. Adjusted pipeline investment cost -

   T_{FC_A} = \$13,000,000

Therefore, the total pipeline investment cost in transporting natural CO_2 is <u>\$13,000,000</u>.

<u>Total Cost of CO_2 To A Candidate Oil Reservoir Operator</u>

The total cost of CO_2 to a candidate oil reservoir operator who is using CO_2 from Jackson Dome, Mississippi is the sum of prices demanded by a producer of CO_2, by a processor of the contaminated gas to recover the CO_2, and by a pipeline company transporting CO_2 to an oil field in South Mississippi.

Assume that the cost of CO_2 produced and delivered at 2000 psig to the pipeline inlet is $0.50/MSCF.

Assume that a processor of natural gas sets the cost of purifying the CO_2 based on 20% DCFRR and processes 117 MMSCFD of CO_2 at 1000 psig.

From Figure 6.35 -

At 20% DCFRR, cost of purification = $0.97/MSCF of CO_2

From Figure 7.12a -

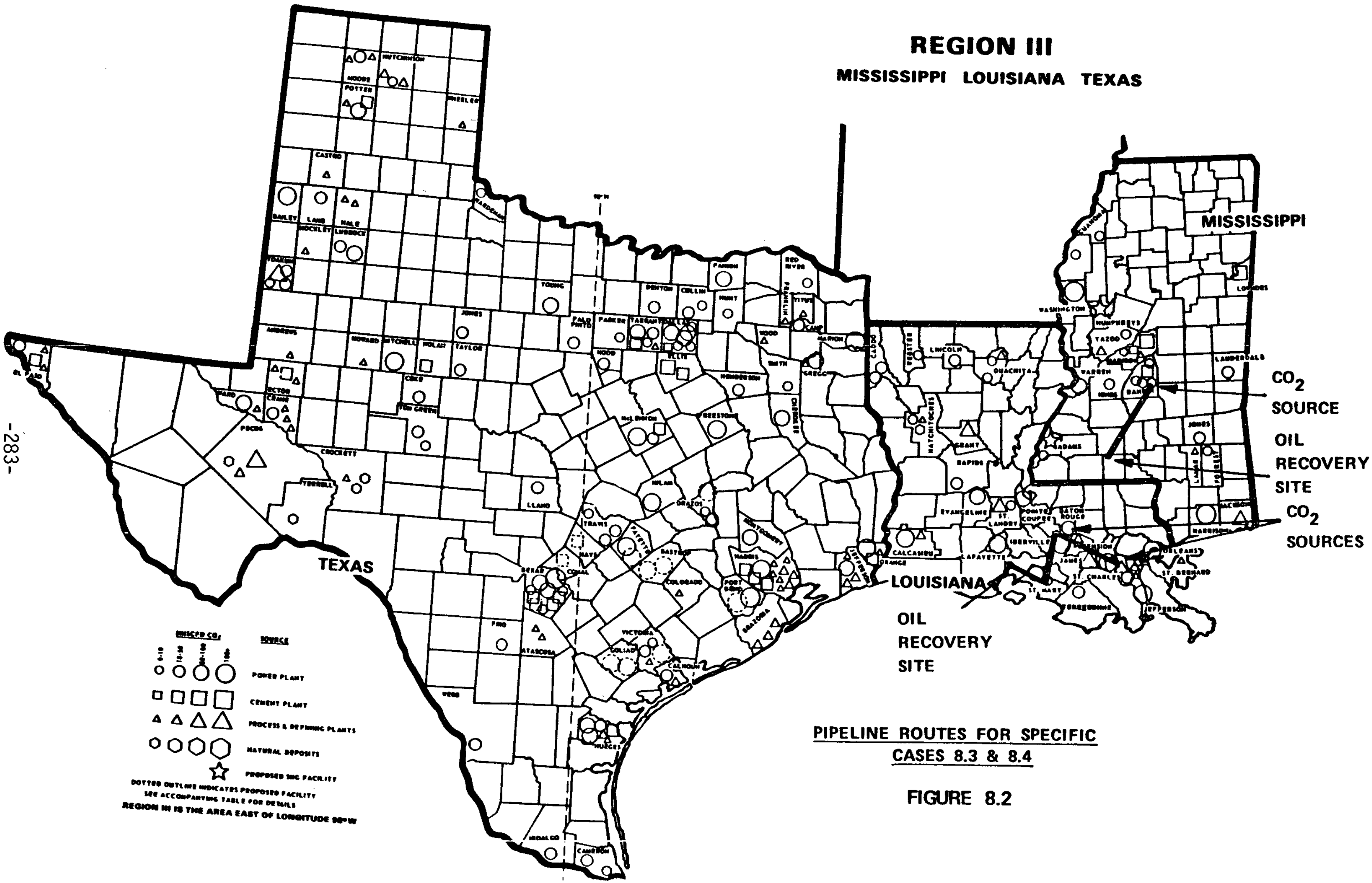
Based on a rate-of-return of 8% using the Utility Financing Method and at a total pipeline investment cost of $13,000,000, the total yearly pipeline revenue received is $4,000,000. Hence, the transportation cost for a 117 MMSCFD CO_2 pipeline from the Jackson Dome area to South Mississippi is $0.10/MSCF CO_2.

Thus, the total cost of CO_2 to a candidate oil reservoir operator in South Mississippi, receiving CO_2 from Rankin County, Mississippi, is as follows:

Cost		$/MSCF
Production		0.50
Purification		0.97
Transportation		0.10
	Total	1.57

REGION III

MISSISSIPPI LOUISIANA TEXAS



PIPELINE ROUTES FOR SPECIFIC CASES 8.3 & 8.4

FIGURE 8.2

8.4 CO_2 SUPPLY TO SOUTH LOUISIANA OIL RESERVOIR

GURC estimates that in Southern Louisiana there could be 50 to 300 million barrels of oil that is potentially recoverable by CO_2 flooding. Net new CO_2 requirement for flooding of these fields is estimated to be about 6 MSCF/BBL of recovered oil. This amounts to a potential net new CO_2 requirement of 300 - 1800 BSCF.

Delivery point of the CO_2 from an intrastate pipeline to the candidate oil reservoirs for this specific case, as shown in Figure 8.2, is assumed to be in Iberia Parish at 14S-7E. The most viable source of CO_2 for these fields in Southern Louisiana is high purity CO_2 available as a by-product from ammonia plants, located about 50 miles from these fields. These ammonia plants in Louisiana are located on a 70 mile stretch along the Mississippi River in Donaldsonville-Geismar and Luling-Taft-Avondale. Total CO_2 available from these ammonia plants amounts to about 194 MMSCFD. For this specific case, CO_2 from ammonia plants located in Luling and Donaldsonville are considered since plants located in these cities produce about 151 MMSCFD of the total 194 MMSCFD. The description of these sources is presented in Sections 3.1.3 and 3.2.3 and in Appendix tables of CO_2 sources for Region III.

8.4.1 CO_2 From Ammonia Plants Vents In Luling And Donaldsonville

Source

Three ammonia plants, located in Luling (St. Charles County), are owned by the Monsanto Company. These plants produce 53.5 MMSCFD by-product CO_2. Plants located in Donaldsonville (St. James County) are owned by C.F. Industries (four plants) producing 72.3 MMSCFD of CO_2, and by the First Mississippi Corporation, producing 24.7 MMSCFD. CO_2 from these plants is of 98% purity and is saturated with water at 3.5 psig. Therefore, these sources will not need purification with the exception of compression and dehydration.

Basis

Net CO_2 requirement	: 6,000 SCF/BBL
Total net CO_2 requirement	: 151 MMSCFD
Total length of pipeline	: 120 miles
River crossings	: 40 miles (to compensate for marsh land)
Length of pipe in flat terrain	: 80 miles
Elevation difference between the CO_2 sources and an oil field	: not appreciable

Capital Investment Costs

These investment costs represent capital investment by a processor who installs compression and dehydration equipment and by a pipeline company to transport CO_2 to an oil field in southern Louisiana.

Investment cost for the compression stations will be for two stations. The capacity of the compression station in Luling is 60 MMSCFD and for Donaldsonville is 110 MMSCFD.

From Figure 6.36 -

In going from a source pressure of 0 psig to pipeline pressure of 2000 psig, an initial pipeline investment compression cost of $10,500,000 is incurred at 60 MMSCFD CO_2 for the Luling station.

Additionally, the pipeline investment compression cost for the Donaldsonville station is based on compressing 110 MMSCFD CO_2 from 0 psig to 2000 psig.

From Figure 6.36 -

Hence, in going from a source pressure of 0 psig to pipeline pressure of 2000 psig, an initial pipeline investment compression cost of $15,000,000 is incurred at 110 MMSCFD CO_2.

The pipeline investment cost incurred in transporting CO_2 from Luling and Donaldsonville (see Figure 8.2 for pipeline route) is determined from the data found in Section 7.3.3. This data is based on a supercritical pipeline system, which is defined and shown in Part 7 as being the most economical pipeline system.

As shown in Section 7.3.3, Equation 7.1 (shown below for reference) is used to determine pipeline investment cost with adjustments for geographic terrain and elevation affects.

$$T_{FC_A} = T_{FC_B} + M_{R.T.}(\$\Delta R.T.) + M_{R.H.}(\$\Delta R.H.) + M_{R.C.}(\$\Delta R.C.) - \$\Delta H \quad (7.1)$$

The data presented and defined below comes from the material found in Section 7.3.3:

1. Unadjusted pipeline investment cost -

   T_{FC_B} (Figure 7.6) = $24,000,000

2. Length of pipeline crossing rivers -

   $M_{R.C.}$ = 40 miles

3. River crossing terrain adjustment factor -

$\$\Delta R.C.$ (Figure 7.11) = $212,000/mile

4. Adjusted pipeline investment cost -

T_{FC_A} = <u>$32,000,000</u>

Therefore, the total pipeline investment cost in transporting the vented ammonia plant CO_2 is <u>$32,000,000</u>.

<u>Total Cost of CO_2 to a Candidate Oil Reservoir Operator</u>

The total cost of CO_2 to an oil field operator who is using CO_2 from ammonia plants in Luling and Donaldsonville, is the sum of the prices demanded by producers of ammonia for the sale of CO_2 by-product, by a processor for compression of CO_2 and by a pipeline company for transporting CO_2.

As discussed in Section 4.1, purchase cost of high purity CO_2 from ammonia plants will depend on the demand of CO_2 and on ammonia producer's inhouse future requirement of CO_2 for production of urea and for other uses. Table 4.1 shows the purchase cost of high purity process vent sources between 25 - 50 cents per MSCF.

For this case assume the purchase cost of CO_2 = <u>$0.35/MSCF</u>.

Assume that a processor (it may be that the ammonia plant owner would install compression equipment) who compresses CO_2 sets the price of CO_2 based on 20% DCFRR and delivers a total of 151 MMSCFD of CO_2 at 2000 psig.

From Figure 6.38 -

Based on a rate-of-return of 20% using the discounted cash flow method and a flow rate of 60 MMSCFD CO_2 and 110 MMSCFD CO_2 for the Luling and Donaldsonville sites, respectively, the resulting total pipeline cmpression cost is as follows:

1. \$0.50/MSCF CO_2 (Luling)
2. \$0.43/MSCF CO_2 (Donaldsonville)

Hence, in going from the source sites at 0 psig to a supercritical pipeline pressure of 2000 psig, an initial compression cost of \$0.45/MSCF CO_2 is incurred.

$$\frac{0.50 \times 60 + 0.43 \times 110}{170} = \$0.45/\text{MSCF}$$

From Figure 7.12a

Based on a rate-of-return of 8% using the Utility Financing Method and at a total pipeline investment cost of \$32,000,000;

the total yearly pipeline revenue received is \$10,000,000. Hence, the transportation cost for a 151 MMSCFD CO_2 pipeline from the Luling and Donaldsonville area is <u>\$0.20/MSCF CO_2</u>.

Thus, the total cost to a candidate oil reservoir operator in Southern Louisiana, receiving CO_2 from ammonia plants in Luling and Donaldsonville, is as follows:

Cost		\$/MSCF
Purchase cost of CO_2 from ammonia plant		0.35
Compression cost		0.45
Transportation cost		0.20
	Total	1.00

8.5 CONCLUSIONS

Investment costs for producing natural CO_2 in Southwest Colorado, Northeast New Mexico and in Mississippi can not be estimated with high degree of confidence because of the uncertain factors such as, well reserve size, flowing well head pressure, well deliverability, and actual drilling costs. Compression investment cost will depend on the expected rate of decline in flowing well head pressure. The scheduled number of wells to be drilled to produce a fixed amount of CO_2 will depend on well deliverability and well reserve size.

For oil fields in West Texas, natural CO_2 sources in Southwest Colorado and Northeast New Mexico appear to be the economical choice because of low CO_2 cost and sufficient supply. Power plant stack gas sources in the vicinity of the oil fields could be used as supplementary sources to the natural sources, but a single power plant source probably can not supply sufficient CO_2 for a complete EOR project.

Total cost of natural CO_2 contaminated with H_2S is high because of the high purification cost. If oil fields in Southern Mississippi use high purity CO_2 from Madison County deposits, purification costs can be reduced substantially.

Although CO_2 from ammonia plant vents is of high purity, compression costs and purchase costs can be substantial. Specific cases studied in this section indicate that the cost difference between CO_2 from power plants and ammonia plants is marginal due primarily to the high purchase cost for CO_2 from ammonia plants. Were this purchase cost to drop somewhat then ammonia plant sources would enjoy a substantial economic advantage over power plant sources.

CONCLUSIONS AND RECOMMENDATIONS

PART 9

9. CONCLUSIONS AND RECOMMENDATIONS

This study on carbon dioxide for enhanced oil recovery originally had a two-fold purpose, (1) to conduct a survey of all known carbon dioxide sources within regions believed to have oil reservoirs amenable to CO_2 miscible flood enhanced oil recovery and (2) to describe a generalized procedure applicable to all sources for determing the costs associated with delivering CO_2 to a reservoir. To a large extent these objectives have been met.

We believe the data collected on carbon dioxide availability to be the most comprehensive survey taken to date. We have included virtually all known significant sources of CO_2. Among these are natural geological deposits, power plant and cement plant stacks, process plant and gas treating plant vents, refineries, proposed SNG plants and other industrial sources. Each source is characterized as to type, expected capacity and CO_2 production rate, and quality of source gas. Information as to proprietor or operator and location of each source as well as any known unusual feature is noted. This information is mapped and tabulated to facilitate the location of potential CO_2 supplies for a given prospective CO_2 EOR project.

Although not in the original scope of this report we feel it necessary to include an analysis of potential demand for CO_2 as an enhanced oil recovery agent. We believe this is necessary in order for one to get a good perspective for the relative amount of CO_2 available and also to help pinpoint areas that might develop potential shortages of CO_2. We are not in a position to give an independent estimate of potential demand, however, we have consolidated estimates made by four other investigators and we feel that the range of CO_2 demand reported here spans the more likely values. One of the more significant conclusions drawn from these projections is the potential for a shortage of CO_2 that might develop in the West Texas oilfields. At best it seems that there is an even match between CO_2 supplies and potential demand for the area.

The original concept developed for determining a CO_2 delivery cost was to describe graphically and in general the various cost associated with processing and transporting CO_2 so that a person with a known carbon dioxide source and destination could determine a delivered cost of the carbon dioxide to the oil reservoir. These costs were to include the cost of CO_2 production, purchase, purification, compression and transportation. Conversely, a person with a known candidate oil reservoir and ceiling price for the delivered CO_2 would be able to establish the geographic area wherein the carbon dioxide could be found. To a large extent the first concept has been successfully developed.

We became aware early in the study, though, that due to the limited number of economically potential CO_2 EOR projects industry and the government would be better served by developing cost data applicable to these projects that have a reasonable chance to reach fruition rather than developing data for hypothetical cases. For this reason the development of the converse objective stated above would not be of real value.

Early in the study GURC characterized the naturally occurring sources in Colorado, New Mexico and Mississippi. This information was necessary in order to allow development of purification and transportation systems tailored to these characteristics. Among these variables considered in system development were quality of source including known contaminants, available pressure, flow rates, and certain environmental considerations. Conditions applicable to industrial sources were also taken into consideration in the development of the system cost data.

We have investigated various carbon dioxide removal systems for each characteristic source. The two systems that merit the most consideration are a chemically reactive MEA system and physical solvent Selexol system. We have determined that the Selexol system is more economical for most natural CO_2 sources whereas an MEA system is more economical for low concentration industrial sources such as power plant stacks. For those natural sources containing H_2S a two stage Selexol system with a Stretford sulfur removal plant has been considered.

The transportation systems evaluated are supercritical, subcritical and liquid pipelines. In all cases we find the supercritical pipeline operating within the range of 1400-2000 psig to be the most economical. Graphs have been developed to determine the costs for transporting various quantities of CO_2 over the range of 50-500 MMCFD and for distance spanning 50-500 miles inclusive of terrain and elevation effects. Final delivery pressure for the CO_2 is set at 2000 psig.

Three specific cases are developed to demonstrate the application of data contained within this report. These are (1) CO_2 supply to a West Texas oil reservoir using CO_2 from the McElmo Dome area of Southern Colorado and the Bravo Dome area of Northeast New Mexico, supplemented with CO_2 from a power plant in Mitchell County, Texas, (2) CO_2 supply to a South Mississippi oil reservoir using CO_2 from the Jackson Dome area in Rankin County, Mississippi and (3) CO_2 supply to a South Louisiana oil reservoir using CO_2 from ammonia plants in Luling and Donaldsonville. We have concluded that the investment costs for producing natural CO_2 can not be determined with a high enough degree of confidence because of uncertain factors such as well reserve, deliverability, flowing well head pressure, and drilling costs. Additionally, the total cost of purified natural CO_2 laden with H_2S is high because of high purification costs.

Surprisingly CO_2 from certain low purity, low pressure power plant stacks could be competitive with higher purity, higher pressure sources because of the general proximity of some power plants to the candidate oil fields. In other words the expense of purification and compression of power plant stack gas is almost equivalent to the transportation and purchase costs borne by distant high purity sources.

It is difficult to draw general conclusions about the merits of one source of CO_2 supply over another. Each CO_2 EOR candidate oil reservoir operator is faced with a unique combination of conditions which might direct him to consider using a variety of CO_2 sources. We cannot state which CO_2 source is best in a given situation. The final cost for CO_2 to an operator is the sum of several terms, any of which can have a significant influence on the final total. We hope and feel however that sufficient information is contained within this report to enable someone to make a reasonable determination of the final delivered cost of CO_2 from any given source.

We highly recommend that if serious consideration be given to any particular CO_2 source listed in this report that the owner of that source be personally contacted before any preliminary work begins. Market conditions are constantly changing which could affect the availability of the source under consideration. Additionally some sources might have characteristics which could make the CO_2 difficult to purify or collect. Some owners might not be willing to enter into such a sale arrangement. The incentives for doing a little preliminary investigation are numerous.

For sources of CO_2 under consideration other than high purity ones we recommend a complete analysis for contaminants. There are some components which can have deleterious effects on the operation of a purification system and some others might become sufficiently enough concentrated to present potential health hazards. We do not intend to alarm anyone but want rather to point out that there could be potential problems, however remote.

We feel that this report will be of significant value to those who are involved in CO_2 enhanced oil recovery efforts, both industry and government. We have taken every reasonable care to ensure accurate, realistic, and current data and evaluations. We feel the range of conditions investigated span most of those conditions expected to be encountered in CO_2 EOR projects. This report is quite comprehensive for determining CO_2 delivery costs. We have not attempted to evaluate the economics of recovering oil by CO_2 miscible flooding. This involves factors such as reservoir properties, CO_2 injection requirements, oil pricing, and tax structures. In addition extensive field testing for each reservoir is usually undertaken to determine the ultimate feasibility of a full scale CO_2 EOR project. We are not in a position at this time to make such evaluations, which will ultimately determine the future of CO_2 enhanced oil recovery.

APPENDICES

APPENDIX A

SOURCES OF CO_2

APPENDIX A1. TABLES OF SOURCES

KEY TO CARBON DIOXIDE SOURCE MAP

Region I

WYOMING

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Albany	Laramie	Tri-State Gen&Trans.	Power Plant-Coal	3 x 500 MW	15.3	325	on 1980,83	2
		Monolith	Cement	200000 TPY	19	27.2	Average	1
Big Horn	Sheridan	Montana-Dakota Util.	Power Plant-Coal	8 MW	15.2	1.6	20% cap.	2
Campbell	Wyodak	Black Hills P&L	Power Plant-Coal	27 MW	15.2	12.4	63% cap.	2
Converse	Glen Rock	Pacific P&L	Power Plant-Coal/Oil	750 MW	15.2	291.1	60% cap.	2
	Converse Co.	Panhandle Eastern	SNG-Coal	275 MMSCFD	85.5	314	Proposed	11
Laramie	Cheyenne	Wycon Chem. Co.	Ammonia Plant	457 STPD	99	6.5	Average	4
Lincoln	Kemmerer	Utah P&L	Power Plant-Coal	707 MW	15.4	239.6	68% cap.	2
Sweet- water	Sweetwater	Pacific P&L	Power Plant-Coal/Oil	1525 MW	15.3	383.5	47% cap.	2
				508 MW	15.3	127.4	on 1979	2
	Green River	Pacific P&L	Power Plant-Oil/Gas	15 MW	-	2.5	45% cap.	2
	Rock Springs	Colo. Interstate Gas	N.G. Processing	60 MMSCFD	40	6.0	Estimate	7
Uinta		Church Buttes	Natural Deposit	3.9 TSCF @ 93% recovery	83.38	15x10.5 ea.	2.09% H_2S 15 psi FWHP Speculative	43
Weston	Osage	Black Hills P&L	Power Plant-Coal	34 MW	15.2	20.2	83% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region I

UTAH

County	City /Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Carbon	Gordon Cr. Area	Pacific-Western	Natural Deposits	1.27 TSCF	98.8	8.5-8.8/well	3500 psig	10,12
		Skelly	Natural Deposits	130 BSCF	91.6	6.5/well	3500 psig	10,12
	Price	Utah P&L	Power Plant-Coal	415 MW	-	45.8		2
	Farnum Dome Area	Equity Oil	Natural Deposit	200 BSCF	98.6	2.8-12/well	1000-1500 psig	10,12
Duchesne	Roosevelt	Plateau, Inc.	Refinery-FCCU	8500 BPD	-	1.07	5200 BPD to FCCU	6,41
Emery	S.E. Mounds	Carter Oil	Natural Deposit	0.76 BSCF	91.2	1.6/well	925 psig	12
	Cedar Mtn.	El Paso N.G.	Natural Deposit	1.0 BSCF	94.7	-	610 psig	12
	Huntington	Utah P&L	Power Plant-Coal	446 MW	15.1	49.3	25% cap.	2
	Castle Gate	Utah P&L	Power Plant-Coal	188 MW	15.1	50.1	57% cap.	2
	Emery	Utah P&L	Power Plant-Coal	4 x 415 MW	-	4 x 45.8	on 1978,80 83,85	2
Morgan	Woods Cross	Phillips Pet.	Refinery-FCCU	24210 BPD	-	1.6	8000 BPD to FCCU	6,41
	Devils Slide	Ideal Industries	Cement Plant	360000 TPY	-	15.0	Estimated	3,1
Salt Lake	Salt Lake City	Utah P&L	Power Plant-Coal	251 MW	14.6	40.0	38% cap.	2
		Portland Cement of UT	Cement Plant	350000 TPY	-	14.1	Calculated	1
		Amoco Oil	Refinery-FCCU	40400 BPD	-	3.7	18000 BPD to FCCU	6,41
		Chevron	Refinery-FCCU	47368 BPD	-	3.7	18000 BPD to FCCU	6,41

KEY TO CARBON DIOXIDE SOURCE MAP

Region I

UTAH

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Salt Lake	N. Salt Lake	Husky Oil	Refinery-FCCU	24000 BPD	-	0.9	4400 BPD to FCCU	6,41
Utah	Orem	Utah P&L	Power Plant-Coal/Gas	59 MW	14.6	9.6	34% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region I

COLORADO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Adams	Denver	Public Service	Power Plant-Coal	801 MW	15.3	206.4	116% cap.	2
Boulder	Boulder	Public Service	Power Plant-Coal/Gas	281 MW	15.3	56.4	57% cap.	2
	Lyons	Martin Marietta	Cement Plant	400000 TPY	13.7	14.2	1249 ppm SO_2	1
Denver	Denver	Public Service	Power Plant-Coal/Gas	250 MW	15.3	83.0	68% cap.	2
			Power Plant-Oil/Gas	115 MW	-	30.4	67% cap.	2
		Continental Oil Co.	Refinery-FCCU	33500 STPD	-	3.1	15000 BPD to FCCU	6,41
Jackson	McCallum Area	Continental Oil	Natural Deposits	100 BSCF	-	5.2/well		36
		Gulf Oil	Natural Deposits	-	-	5/well	South Canyon unit	9
Larimer	Boettcher	Ideal Basic Ind.	Cement Plant	235000 TPY	-	11.5	Revamping	3,1
Mesa	Cameo	Public Service	Power Plant-Coal/Gas	75 MW	14.6	25.3	79% cap.	2
	Fruita	Gary Operating Co.	Hydrogen Plant	-	98.0	0.2		5
Moffat	Craig	Colo-UTE Elec. Assn.	Power Plant-Coal	2 x 380 MW	14.6	305.4	on 1979	2
			Power Plant-Coal	2 x 190 MW	14.6	152.7	on 1981,82	2
Routt	Hayden	Colo-UTE Elec. Assn.	Power Plant-Coal	465 MW	15.3	92.6	43% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

UTAH

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Garfield	Escalante Anticline	Phillips	Natural Deposit	1.13 BSCF	92.75	13.1-24.8/well	44-138 psig SIP	12
Iron	Cedar City	Calif-Pacific Util.	Power Plant-Coal	7 MW	-	0.4	5% cap.	2
San Juan	Eastland	Western N.G.	Natural Deposit	14.3 BSCF	92	-	Rept'd Strong Blow	12
	Deadman Canyon	Pan Am Pet.	Natural Deposit	13.1 BSCF	95	2.0/well	2580 psig	12
	White Mesa	Phillips	Natural Deposit	18.0 BSCF	97	1.0/well		12
	Bluff	Shell	Natural Deposit	16.0 BSCF	92	2.2-8.7/well	2670 psig	12
	Tank Mesa	Carter Oil	Natural Deposit	9.5 BSCF	93	0.1/well	1883 psig	12
	Lime Ridge	Utah Southern	Natural Deposit	14.3 BSCF	97	-	unreliable data	12
Siever	Emery Field	Skelly	Natural Deposit	7.5 BSCF	92	-	low flow	12
Washington	St. George	LA Dept. Water & Power	Power Plant-Coal	2 x 250 MW	15.2	2 x 27.3	on 1984,85	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

COLORADO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Alamosa	Alamosa	Public Service	Power Plant-Oil/Gas	18 MW	-	4.3	43% cap.	2
Dolores	Dove Creek/ Doe Canyon	Shell Oil	Natural Deposit	-	99	3-12/well	600-800 psig FWHP	10
El Paso	Colorado Springs	Colo. Spgs. Dept. Pub. Util.	Power Plant-Oil/Gas	62 MW	15.3	2.6	12% cap.	2
			Power Plant-Coal/Gas	282 MW	15.3	80.6	64% cap.	2
			Power Plant-Coal	200 MW	15.3	49.2	on 1979	2
Fremont	Canon City	Central Tel. & Util.	Power Plant-Coal/Gas	38 MW	14.6	16.4	13% cap.	2
	Pueblo	Central Tel. & Util.	Power Plant-Oil/Gas	30 MW	-	5.9	15% cap.	2
	Portland	Ideal Basic Ind.	Cement Plant	885000 TPY	-	31.4	Revamping	3,1
Huerfano	Walsenburg	Walsenburg Util.	Power Plant-Coal	11 MW	14.6	6.1	9% cap.	2
	Sheep Mtn. Range	Arco	Natural Deposits	1.3-2.5 TSCF	97-99.6	250 field total	400-600 psig FWHP	10,17
Las Animas	Trinidad	Trinidad EPL	Power Plant-Gas	11 MW	-	0.8	25% cap.	2
	Grande Area	-	Natural Deposit	1-2 TSCF	98.2	low	150 psig SIBHP	10
	Model Dome	U.S. Dept. Interior	Natural Deposit	-	-	0.6		9
Montezuma	McElmo Dome	Shell/Mobil	Natural Deposit	2.6-3.4 TSCF	96-99	300 field total	600-850 psig FWHP	10,17
Montrose	Montrose	Western Colo. Power	Power Plant-Coal	10 MW	14.9	3.9	53% cap.	2
	Nulca	Colo-UTE Elec. Assn.	Power Plant-Coal	34 MW	14.6	10.2	48% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

COLORADO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Otero	Rocky Ford	Central Tel. & Util.	Power Plant-Oil/Gas	7 MW	-	2.1	55% cap.	2
Prowers	Lamar	Lamar Utilities Board	Power Plant-Gas	34 MW	-	3.7	33% cap.	2
	Haxton Unit	Lotus Oil Co.	Natural Deposit	-	-	0.5/well	insignificant	9
Pueblo	Pueblo	Public Service	Power Plant-Coal	778 MW	15.2	207.7	59% cap.	2
Rio Blanco	White River	Texaco	Natural Deposit	-	-	0.25/well	insignificant	9
		Frontier Refiners	Natural Deposit	-	-	0.4/well	insignificant	9

Note: Public Service of Colorado has 2-500 MW power plants planned for S.E. Colorado to come on stream 1984, 86

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

NEW MEXICO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Bernalillo	Alburquerque	Plains Elec. & Trans.	Power Plant-Coal	330 MW	-	109	on 1982	2
		Public Service NM	Power Plant-Gas	114 MW	10.2	18.0	50% cap.	2
			Power Plant-Gas	35 MW	10.2	0.8	5% cap.	2
			Power Plant-Gas	175 MW	10.2	35.0	71% cap.	2
		Ideal Basic Ind.	Cement Plant	500000 TPY	15	38.0	Revamping	1,3
Chaves	Ruswell	S.W. Pub. Serv.	Power Plant-Gas	24 MW	10.2	6.8	78% cap.	2
Colfax	Raton	Raton Pub. Serv.	Power Plant-Coal	12 MW	14.7	2.0	27% cap.	2
	Maxwell Field	York Denton	Natural Deposit	-	99.8	1.53/well	shut in	13,14
	Jaritos Dome	Calif. Florsheim	Natural Deposit	-	67.2	0.25/well	Balance N_2/O_2	13
	S.E. Colfax	Amoco	Natural Deposit	-	99	0.9-1.0/well	Specula-tive	10
Dona Ana	nr. El Paso	El Paso Elec.	Power Plant-Gas/Oil	385 MW	10.2	34.4	28% cap.	2
Eddy	Carlsbad	S.W. Pub. Serv.	Power Plant-Gas	44 MW	10.2	11.4	74% cap.	2
		Baker Ind.	Ammonia Plant	580(+420) STPD	98	13.2(+8.8)	(by 1980)	4,6,18
		N-Ren	Ammonia Plant	100 STPD	98	2.2		4,6
	Eddy Co.	Amoco	N.G. Processing	40 MMSCFD	31	0.6	Frailey Engr.	7
	Artesia	Navajo Refinery	Refinery-FCCU	31474 BPD	-	1.1	5200 BPD to FCCU	6,41
Harding	Harding Co., Union Co., Quay Co.,	Amoco	Natural Deposits	7.0-10.6 TSCF	99.1-100	650 reported maximum field prod. 0.3-1.2/well	>140 psig FWHP	10,13,17

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

NEW MEXICO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2MMSCFD	Comments	Ref
Hidalgo	Lordsburg	Community Pub. Serv.	Power Plant-Oil/Gas	41 MW	10.2	6.7	42% cap.	2
Lea	N. Lovington	Lea Co. Elec. Coop	Power Plant-Gas	49 MW	10.2	12.9	85% cap.	2
		Tipperary Corp.	Ammonia Plant	300 STPD	98	6.6		18
	Hobbs	N.M. Elec. Serv.	Power Plant-Gas	100 MW	10.2	17.7	71% cap.	2
		Northern Nat. Gas	N.G. Processing	220 MMSCFD	-	2.3	0.4% H_2S	1
		S.W. Pub. Serv.	Power Plant-Gas	265 MW	10.2	55.8	83% cap.	2
	Maljamar	Conoco	N.G. Processing	20 MMSCFD	81	0.3	Frailey Engr.	7
	Lea Co.	Getty	N.G. Processing	2 x 0.5 MMSCFD	75	0.75	12-20% H_2S	1
Mora	Turkey Mt.	Cities Serv.	Natural Deposits	-	99	1.0/well	100-150 psig	10
	Wagon Mound Field	Mobil Oil	Natural Deposits	-	99	1.0/well	100-150 psig	10
		Arkansan Fuel & Iron Co.	Natural Deposits	-	92.2	12-25	possibly incorrectly reported	13,14

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

NEW MEXICO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Rio Arriba	Abiquiu	Lowry et al.	Natural Deposits	-	96.5	-	690 psig	13
	San Juan Basin	Southern Union Ref.	N.G. Processing	80 MMSCFD	-	0.5		1
Roosevelt	Milnesand	Cities Serv.	N.G. Processing	37 MMSCFD	67	1.7	1.5% H_2S	1
Sandoval	Algodones	Plains Electric	Power Plant-Gas	51 MW	10.2	8.2	46% cap.	2
San Juan	Farmington	Arizona Pub. Serv.	Power Plant-Coal	2269 MW	15.1	670	58% cap.	2
		Farmington Elec.	Power Plant-Gas	28 MW	-	2.9	41% cap.	2
	Fruitland	Pub. Serv. N.M.	Power Plant-Coal	2 x 338 MW	14.6	100	29% cap.	2
				2 x 468 MW	14.6	418	on 1979,81	2
	Four Corners	El Paso N.G.	SNG-Coal	72 MMSCFD	85.5	82	Proposed	11
		Texas Eastern/ Pacific Lighting	SNG-Coal	275 MMSCFD	85.5	314	Proposed	11
	San Juan Basin	Southern Union Ref.	N. G. Processing	100 MMSCFD	-	0.9		9
		Citgo	N. G. Processing	-	-	5.8		1
San Miquel		Belco	Natural Deposit	-	-	1.2/well		9
Santa Fe	Santa Fe	Publ. Serv. N.M.	Power Plant-Gas	12 MW	10.2	0.5	10% cap.	2
Torrence	Estancia Valley	Estancia Valley Development	Natural Deposits	-	97.9	2.25/well	Plugged	14
		Witt Ice & Gas Co.	Natural Deposits	-	97.9	2.35/well		14

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

NEW MEXICO

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Union	Des Moines Field	Amoco	Natural Deposits	–	98.4	1.0/well		10,14,13
	Sierra Grande	Sierra Grande Oil Co.	Natural Deposits	-	98.6	1-2/well	Plugged	14

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Andrews	Inez Field	Amoco	N.G. Processing	1.1 MMSCFD	21.68	0.167	1.2% H_2S, 2.5% SO_2	1
Atascosa	Fashing Field	Warren Pet.	N.G. Processing	7.6 MMSCFD	51.36	4.0	3.6% H_2S	1
			N.G. Processing	-	-	3.1	0.4% H_2S	1
Bailey	Muleshoe	S.W. Pub. Serv.	Power Plant-Coal	530 MW	14.9	132	on 1985 @ 50% cap.	2
Bexar	San Antonio	S.A. Portland	Cement Plant	470000 TPY	10-20	26		3
		Texas Indus.	Cement Plant	500000 TPY	10-20	27	Uncertain Startup	3
		Capital Cement	Cement Plant	338400 TPY	10-20	20		3
		Kaiser Cement	Cement Plant	490000 TPY	10-20	27		3
		S.A. Pub. Serv. Board	Power Plant-Gas	894 MW	8-12	86.6	35% cap.	2
				263 MW	8-12	9.9	11% cap.	2
				163 MW	8-12	6.4	14% cap.	2
				892 MW	8-12	76.2	35% cap.	2
				494 MW	8-12	7.4	5% cap.	2
	Elmendorf	S.A. Pub. Serv. Board	Power Plant-Coal	836 MW	14-9	269.7		2
Castro	Dimmitt	Good Pasture Inc.	Ammonia Plant	-	98	0.7		5
Coke	Bronte	West Texas Util.	Power Plant-Gas	81 MW	8-12	13.8	65% cap.	2
Comal	New Braunfels	General Portland	Cement Plant	800000 TPY	10-20	45	on 1980	3

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Crane	-	Arco	N.G. Processing	1.6 MMSCFD	91	1.5		1
			N.G. Processing	-	-	1.3	1.8% H_2S	1
	Girvin	West Tex. Util	Power Plant-Gas	132 MW	8-12	26.6	76% cap.	2
	-	-	N.G. Processing	15 MMSCFD	95	0.12	by Tulsa Pro. Quip	7
Crockett	Ozona	-	N.G. Processing	13 MMSCFD	90	0.2	by Stearns Rogers	7
	-	Sun Oil	Natural Deposit	-	-	4.7/well		
	-	Texaco	Natural Deposit	-	-	2.0/well		
Ector	Odessa	El Paso N.G.	Ammonia Plant	350 STPD	98	6.8		5
		Shell Oil	Refinery-FCCU	35000 BPD	8-15	2.1	10500 BPD/FCUU	6,41
		S.W. Portland	Cement Plant	250000 TPY	10-20	14.0		3
El Paso	El Paso	Chevron	Refinery-FCCU	74737 BPD	8-15	4.5	22000 BPD/FCCU	6,41
		El Paso Elec.	Power Plant-Gas	265 MW	8-12	38.9	54% cap.	2
		S.W. Portland	Cement Plant	285000 TPY	10-20	25.0		1
		Texaco	Refinery-FCCU	17895 BPD	8-15	1.4	7000 BPD/FCCU	6,41
Frio	Pearsall	Medina Elec. Coop	Power Plant-Gas	75 MW	8-12	17.8	71% cap.	2
Hale	Plainview	Farmers Nat'l Chem.	Ammonia Plant	60 MMSCFD	98	1.3		18
Hardeman	Quanah	W. Texas Util.	Power Plant-Gas	44 MW	8-12	5.5	38% cap.	2
Hays	Buda	Centex	Cement Plant	470000 TPY	10-20	25	on 1978	3
Hidalgo	Mission	Central P&L	Power Plant-Gas	188 MW	8-12	17.9	33% cap.	2
Howard	Big Springs	WR Grace	Ammonia Plant	300 STPD	98	5.5		5
Hutchinson	Borger	Camex	Ammonia Plant	1000 STPD	98	18.9		5
		Phillips	Refinery-FCCU	105263 BPD	8-15	11.5	56000 BPD/FCCU	6,41
		S.W. Pub. Serv.	PowerPlant-Gas	34 MW	8-12	4.6	61% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Jones	Stamford	W. Texas Util.	Power Plant-Gas	241 MW	8-12	33.4	50% cap.	2
Lamb	Earth	S.W. Pub. Serv.	Power Plant-Gas	434 MW	8-12	46.4	40% cap.	2
Llano		Lower Colo. River Authority	Power Plant-Gas	408 MW	8-12	39.6	38% cap.	2
Lubbock	Lubbock	Lubbock P&L	Power Plant-Gas	103 MW	8-12	15.2	60% cap.	2
		S.W. Pub. Serv.	Power Plant-Gas	512 MW	8-12	57.9	43% cap.	2
Mitchell	Colo. City	Texas Elec. Serv.	Power Plant-Gas	827 MW	8-12	111.9	54% cap.	2
Moore	Dumas	Diamond Shamrock	Ammonia Plant	450 STPD	98	8.2		5
	Sunray	Diamond Shamrock	Refinery-FCCU	53500 BPD	8-15	4.7	23000 BPD to FCCU	6,41
		S.W. Pub. Serv.	Power Plant-Gas	68 MW	8-12	17.7	84% cap.	2
Nolan	Maryneal	Lone Star Ind.	Cement Plant	545000 TPY	10-20	31	not feasible	3,1
Palo Pinto	Gordon	Brazos Elec.	Power Plant-Gas	404 MW	8-12	42.0	40% cap.	2
Pecos	Toyah Field	Lone Star Gas	Natural Deposit	600+ BSCF	-	150 total	committed	7
	Fort Stockton	Coastal States	Natural Deposit	-	-	4.5/well	committed	5
		Trans West. Pipeline	N. G. Processing	180 MMSCFD	99	49.9	Vetrocoke	7
Potter	Amarillo	S.W. Portland	Cement Plant	225000 TPY	10-20	13		3
		S.W. Pub. Serv.	Power Plant-Gas	483 MW	8-12	78.1	65% cap.	2
		Texaco	Refinery-FCCU	21053 BPD	8-15	1.6	8000 BPD/FCCU	6,41
Taylor	Abiline	W. Texas Util.	Power Plant-Gas	146 MW	8-12	22.8	75% cap.	2
Terrell	-	El Paso N.G.	N.G. Deposit	200 MMSCFD	-	106	H_2S	8
	Terrell	El Paso N.G.	N.G. Deposit	-	-	5.2		5

KEY TO CARBON DIOXIDE SOURCE MAP

Region II

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2MMSCFD	Comments	Ref
Tom Green	San Angelo	W. Texas Util.	Power Plant-Gas	108 MW	8-12	15.7	76% cap.	2
			Power Plant-Gas	52 MW	8-12	2.1	11% cap.	2
Ward	Monahans	Texas Elec. Serv.	Power Plant-Gas	700 MW	8-12	75.9	43% cap.	2
Ward-Winkler	S. Wink OBrian		N.G. Processing	-	19.44	0.1		1
Webb	Laredo	Central P&L	Power Plant-Gas	187 MW	8-12	18.7	34% cap.	2
Wheeler	-	-	N.G. Processing	6 MMSCFD	99	0.18	Tulsa Pro. Quip.	7
Yoakum	Harrington	S.W. Pub. Serv.	Power Plant-Coal & Gas	343 MW	15.2	5.2	23% cap.	2
	Harrington	S.W. Pub. Serv.	Power Plant-Coal	700 MW	15.2	42.7	on 1978,80	2
	Denver City	S.W. Pub. Serv.	Power Plant-Gas	80 MW	15.2	13.0	57% cap.	2
		-	N.G. Processing	168 MMSCFD	83	140	Stearns-Rogers	7
Young	-	Texas Elec. Serv.	Power Plant-Gas	634 MW	8-12	70.3	45% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Bastrop	-	S.A. Pub. Serv. Board	Power Plant-Gas	375 MW	8-12	120	on 1988	2
	Bastrop	Lower Colo. River Authority	Power Plant-Gas	565 MW	8-12	50.7	36% cap.	2
Brazoria	Texas City	Amoco	Ammonia Plant	2 x 1000	98	38		4
			Refinery-FCCU	370000 BPD	-	34.2	167000 BPD to FCCU	8,41
		Marathon Oil	Refinery-FCCU	68000 BPD	-	6.1	30,000 BPD to FCCU	8,41
		Texas City Ref. Co.	Refinery-FCCU	80000 BPD	-	5.5	27000 BPD to FCCU	8,41
	Freeport	Dow Chemical	Ethylene Oxide	-	-	3.5		5
			Ammonia Plant	350 STPD	98	7.7		18
	Sweeny	Phillips	Refinery-FCCU	109474 BPD	-	7.0	34000 BPD to FCCU	8,41
Brazos	Bryan	Texas Power Pool	Power Plant-Coal	3 x 400 MW	15.2	409.1	on 1980,82, 84	2
		S.W. Pub. Serv.	Power Plant-Gas	21 MW	8-12	2.9	40% cap.	2
		Bryan Municipal Elec. Systems	Power Plant-Gas	125 MW	8-12	17.5	44% cap.	2
Calhoun	Seadrift	Union Carbide	Ethylene Oxide	-	-	15.0		5
	PT. Comfort	Central P&L	Power Plant-Gas	261 MW	8-12	29.1	44% cap.	2
Cameron	La Paloma	Central P&L	Power Plant-Gas	219 MW	8-12	26.2	46% cap.	2
	Brownsville	Brownsville Pub. Util. Board	Power Plant-Gas	53 MW	8-12	7.8	41% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Camp	Lone Star	S.W. Elec. Power	Power Plant-Gas	58 MW	8-12	0.1	1% cap.	2
Cass	-	Shell Oil	N.G. Processing	-	12	2.9	1% H_2S	1
Cherokee	-	Texas P&L	Power Plant-Gas	703 MW	8-12	85.6	50% cap.	2
Collin	Frisco	Texas P&L	Power Plant-Gas	156 MW	8-12	9.2	20% cap.	2
	-	Garland Elec. Dept.	Power Plant-Gas	347 MW	8-12	27.5	29% cap.	2
Colorado	-	Shell Oil	N.G. Processing	-	80	2.5		1
Dallas	Dallas	Dallas P&L	Power Plant-Gas	223 MW	8-12	4.1	3% cap.	2
				927 MW	8-12	85.0	34% cap.	2
				990 MW	8-12	62.4	24% cap.	2
				708 MW	8-12	72.7	39% cap.	2
				340 MW	8-12	7.8	7% cap.	2
		General Portland	Cement Plant	425000 TPY	-	0.34		1
	Garland	Garland Elec. Dept.	Power Plant-Gas	96 MW	8-12	4.5	14% cap.	2
Denton	Denton	Denton Municipal Util.	Power Plant-Gas	189 MW	8-12	17.8	31% cap.	2
Ellis	Midlothian	Gifford Hill	Cement Plant	846000 TPY	-	47		3
		Texas Industries	Cement Plant	1128000 TPY	-	63		3
Fannin	Savoy	Texas P&L	Power Plant-Gas	1175 MW	8-12	75.7	23% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Fayette	La Grange	Lower Colo. River Authority	Power Plant-Coal	600 MW	8-12	136.2	50% cap.	2
				600 MW	8-12	136.2	on 1979	2
Fort Bend	Richmond	Houston P&L		1225 MW	8-12	185.6	60% cap.	2
			Power Plant-Coal	3 x 640 MW	8-12	3 x 176	on 1978,80 81	2
Franklin	New Hope	Getty Oil	N.G. Processing	-		2.0	0.6% H_2S	1
Freestone	-	Texas P&L	Power Plant-Coal	1186 MW	15.1	445.4	71% cap.	2
Goliad	Fannin	Central P&L	Power Plant-Coal	2 x 550 MW	15.1	2 x 225		2
Gregg	Cherokee Lake	S.W. Elec. Power	Power Plant-Gas/Oil	186 MW	8-12	89	165% cap.	2
	Longview	Eastman Kodak	Ethylene Oxide	-	-	0.9		5
Hale	-	Amoco	N.G. Processing	10 MMSCFD	93	0.4	Frailey Engrg.	7
	Plainview	Occidental Petroleum	Ammonia Plant	150 STPD	98	3.4		5
Harris	Bacliff	Houston L&P	Power Plant-Gas	2314 MW	8-12	330.1	59% cap.	2
	Baytown	Exxon	Refinery-FCCU	405000 BPD	-	27.7	135000 BPD to FCCU	8,41
	Clear Lake	Celanese	Ethylene Oxide	-	-	6.0		5
	Dear Park	Rohm&Haas	Ammonia Plant	100 STPD	98	1.6		5
		Shell Oil	Refinery-Hydrogen	71 MMSCFD	98	17.7		18
			Refinery-FCCU	210000 BPD	-	14.3	70000BPD/FCCU	6,18,41
	Houston	ARCO	Refinery-FCCU	324500 BPD	-	14.1	69000 BPD to FCCU	8,41

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Harris	Houston	Charter Oil	Refinery-FCCU	70000 BPD	-	7.4	36000 BPD to FCCU	8,41
		Crown Cent. Pet.	Refinery-FCCU	103000 BPD	-	8.8	43000 BPD to FCCU	8,41
		Gulf Coast Portland	Cement Plant	600000 TPY	-	34		3
		General Portland	Cement Plant	452000 TPY	-	20		3
		Houston L&P	Power Plant-Gas	2295 MW	8-12	334.1	59% cap.	2
				210 MW	8-12	7.5	10% cap.	2
				353 MW	8-12	30.6	32% cap.	2
				821 MW	8-12	78.8	36% cap.	2
				548 MW	8-12	48.4	34% cap.	2
		Ideal Basic	Cement Plant	620000 TPY	-	35		3
		Lone Star Ind.	Cement Plant	526000 TPY	-	28.3		1
		Nat. Distillers	Chemicals	-	-	0.6	Benfield	7
		Tenneco	Ammonia Plant	600 STPD	98	12.9		5
	La Port	Houston L&P	Power Plant-Gas	826 MW	8-12	119.3	57% cap.	2
	Pasadena	Phillips	Ammonia Plant	660 STPD	98	11.4		5
	Spring	Cosden Oil & Chem.	Refinery-FCCU	68421 BPD	-	4.9	24000 BPD to FCCU	6,41
	Webster	Houston L&P	Power Plant-Gas	614 MW	8-12	80.2	50% cap.	2
Henderson	Henderson	Texas P&L	Power Plant-Gas	412 MW	8-12	17.1	15% cap.	2
Hood	Decordova	Texas P&L	Power Plant-Gas	775 MW	8-12	37.8	19% cap.	2
Hunt	Greenville	Greenville Municipal L&P	Power Plant-Gas	39 MW	8-12	6.1	59% cap.	2
Jefferson	Port Arthur	Texaco	Refinery-FCCU	427368 BPD		27.6	135000 BPD to FCCU	6,41

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Jefferson	Port Arthur	Gulf Oil	Refinery-Hydrogen	23 MMSCFD	98	7.6		18
			Refinery-FCCU	319000 BPD		24.6	120000BPD/FCCU	6,41
		American Petrofina	Refinery-FCCU	115789 BPD	-	6.6	32000BPD/FCCU	6,41
	Fort Neches	Jefferson Chemical	Ethylene Oxide	-	-	10.3		5
	Beaumont	DuPont	Ammonia Plant	1000 STPD	98	18.1		5
		Gulf States Util.	Power Plant-Gas	152 MW	-	72.8	56% cap.	2
		Mobil Oil	Refinery-Hydrogen	60MMSCFD	98	20.0		18
			Refinery-FCCU	-	-	23.3		6
			Ammonia Plant	850 STPD	98	13.7		18
		PPG	Ethylene Oxide	-	-	1.7		5
		Swift Chemical	Ammonia Plant	900 STPD	98	14.7		5
		Union Oil California	Refinery-FCCU	126316 BPD	-	8.0	39000BPD/FCCU	6,41
Marion	Jefferson	S.W. Elec. Power	Power Plant-Gas	881 MW	8-12	138.5	63% cap.	2
			Power Plant-Coal	528 MW	15.1	214.5		2
			Power Plant-Coal	1056 MW	15.1	429	on 1980,84	2
McLennon	Riesel	Texas P&L	Power Plant-Gas	315 MW	8-12	36.8	43% cap.	2
	Waco	Texas P&L	Power Plant-Gas	1379 MW	8-12	115.0	34% cap.	2
		Universal Atlas	Cement Plant	225600 TPY	-	13		3
Milam	Rockdale	Texas P&L	Power Plant-Coal	545 MW	15.1	147	50% cap.	2
Mont-gomery	Montgomery	Gulf States Util.	Power Plant-Coal	542 MW	8-12	83.5	59% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Nueces	Corpus Christi	Centex	Cement Plant	263200 TPY	-	15		3
		Central P&L	Power Plant-Gas	703 MW	8-12	53.4	29% cap.	2
				574 MW	8-12	40.0	28% cap.	2
				595 MW	8-12	52.8	36% cap.	2
		Champlin	Refinery-FCCU	131500 BPD	-	11.0	54000 BPD to FCCU	6,41
		Coastal States	Refinery-FCCU	194737 BPD	-	3.9	19000 BPD to FCCU	6,41
		S.W. Refining	Refinery-FCCU	122450 BPD	-	2.5	12000 BPD to FCCU	6,41
		Sun Petroleum	Refinery-FCCU	60000 BPD	-	4.1	20000 BPD to FCCU	6,41
Orange	Bridge City	Gulf States Util.	Power Plant-Gas	1543 MW	8-12	227.7	57% cap.	2
	Orange	Allied Chemical	Ethylene Oxide	-	-	1.0		5
		Alpha Portland	Cement Plant	338400	-	20		3
Parker	Wetherford	Brazis Elec. Power	Power Plant-Gas	75 MW	8-12	1.2	5% cap.	2
Red River	Bogota	Texas P&L	Power Plant-Gas	112 MW	8-12	5.0	16% cap.	2
Smith	Tyler	La Gloria	Refinery-FCCU	29700 BPD	-	2.0	10000 BPD to FCCU	6,41
Tarrent	Fort Worth	General Portland	Cement Plant	731000 TPY	-	41		3
		Texas Elec. Serv.	Power Plant-Gas	978 MW	8-12	72.9	30% cap.	2
				116 MW	8-12	2.3	6% cap.	2
				706 MW	8-12	26.7	12% cap.	2
		Winston Refining	Refinery-FCCU	20500 BPD	-	0.7	3400 BPD to FCCU	6,41

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

TEXAS

County	City /Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2MMSCFD	Comments	Ref
Titus	Mount Pleasant	American Petrofina	Refinery-FCCU	27368 BPD	-	2.0	9600 BPD to FCCU	6,41
Travis	Austin	Austin Elec. Dept.	Power Plant-Gas	730 MW	8-12	31.4	38% cap.	2
				120 MW	8-12	10.1	24% cap.	2
				555 MW	8-12	42.6	30% cap.	2
Victoria	Nursey	S. Texas Elec. Coop	Power Plant-Gas	25 MW	8-12	5.4	70% cap.	2
			Power Plant-Coal	400 MW	15	287	on 1980	2
	Victoria	Central P&L	Power Plant-Gas	553 MW	8-12	41.5	30% cap.	2
		DuPont	Ammonia Plant	300 STPD	98	6.5		5
Wood	W. Yantis	Amoco	N.G. Processing	-	-	1.6	116 ppm H_2S	1

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

LOUISIANA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Ascension	Geismar	Allied Chemical	Ammonia Plant	1000 STPD	98	10.5		5
		BASF	Ethylene Oxide	425 STPD	-	9.6	Shell, O_2	5
		Bordon	Ammonia Plant	1000 STPD	98	22.0		18
		Shell Oil	Ethylene Oxide	410 STPD	-	6.0	Shell, O_2	5
Baton Rouge	Baton Rouge	Exxon	Refinery-FCCU	528000 BPD	-	34.6	169000BPD/FCCU	6,41
		Gulf States Util.	Power Plant-Gas	253 MW	8-12	20.6	55% cap.	2
				175 MW	8-12	35.5	77% cap.	2
Caddo	Mornings Port	S.W. Elec. Power	Power Plant-Gas	277 MW	8-12	18.4	29% cap.	2
	Shreveport	S.W. Elec. Power	Power Plant-Gas	170 MW	8-12	12.5	29% cap.	2
Calcasieu	Lake Charles	Calcasieu Chemical Corp.	Ethylene Oxide	315 STPD		8.7	Shell, Oxygen	5
		Cities Service	Refinery-FCCU	280000 BPD		25.6	125000BPD/FCCU	6,41
		Conoco	Refinery-FCCU	85000 BPD		5.5	27000BPD/FCCU	6,41
		Olin	Ammonia Plant	2 x 700 STPD	98	23.3		4
	West Lake	Gulf States Util.	Power Plant-Gas	982 MW	8-12	85.7	42% cap.	2
Caldwell	Vixen	Barnwell Drilling	Natural Deposit	-	-	0.7/well		9
		Shell Oil	Natural Deposit	-	-	0.4/well		9
Evange-line	St. Landry	Central LA. Elec. Co.	Power Plant-Gas	483 MW	8-12	39.6	35% cap.	2
Grant	Polluck	Farmland Ind.	Ammonia Plant	1000 STPD	98	22.4		4,8
				1000 STPD	98	26.5		4,8
				1000 STPD	98	26.6		4,8

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

LOUISIANA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Iberville	Plaquemine	Dow Chemical Co.	Ethylene Oxide	678 STPD	-	8.1		5
	St. Gabriel	Gulf States Util.	Power Plant-Gas	2117 MW	8-12	178.9	31% cap.	2
Jefferson	Avondale	American Cynamid	Ammonia Plant	1000 STPD	98	14.5		4
	Michoud	Air Products	Ammonia Plant	600(+300) STPD	98	7.8 (+6.6)	(by 1980)	4,18
	Westwego	LA Power & Light	Power Plant-Gas	1917 MW	8-12	232.1	57% cap.	2
Lafayette	Lafayette	Central LA Elec. Co.	Power Plant-Gas	445 MW	8-12	42.2	40% cap.	2
		Lafayette Util. System	Power Plant-Gas	143 MW	8-12	17.8	51% cap.	2
				42 MW	8-12	4.7	38% cap.	2
Lincoln	Ruston	Ruston Water & Light	Power Plant-Gas	81 MW	8-12	4.4	19% cap.	2
Natchido-ches	Natchidoches	Natchidoches Municipal Water & Light	Power Plant-Gas	42 MW	8-12	3.7	29% cap.	2
		Placid Oil Co.	N.G. Processing	-	98	3.0		1
			N.G. Processing	20 MMSCFD	98	0.76	Tulsa Pro. Quip.	7
Orleans	Metaire	Good Hope Refineries	Refinery-FCCU	70000 BPD		3.3	16000 BPD/FCCU	6
	New Orleans	Air Products	Hydrogen Plant	12.5 MMSCFD	98	4.1		4
		New Orleans Pub. Serv.	Power Plant-Gas	224 MW	8-12	18.8	21% cap.	2
	Michoud	New Orleans Pub. Serv.	Power Plant-Gas	959 MW	8-12	75.2	24% cap.	2
	New Orleans	New Orleans Pub. Serv.	Power Plant-Gas	96 MW	8-12	0.5	1% cap.	2
		New Orleans Sewer & Water	Power Plant-Gas	47 MW	8-12	4.4	22% cap.	2
	Norco	Shell Oil	Hydrogen Plant	51 MMSCFD	98	16.8		18
			Refinery-FCCU	250000 BPD		20.5	100000BPD/FCCU	6

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

LOUISIANA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2MMSCFD	Comments	Ref
Ouachita	Monroe	Monroe Util. Commission	Power Plant-Gas	166 MW	8-12	16.2	33% cap.	2
	Sterlington	IMC	Ammonia Plant	1000 STPD	98	21.5	7.5-11 MMSCFD for EOR	5
		LA. P&L	Power Plant-Gas	1150 STPD	98	24.7		5
		LA. P&L	Power Plant-Gas	523 MW	8-12	61.6	53% cap.	2
Point Coupe	New Roads	Cajun Elec. Power Co.	Power Plant-Coal	540 MW	8-12	130	50% cap. on 1979	2
				540 MW	8-12	130	50% cap. on 1979	2
				230 MW	8-12	31	53% cap.	2
Rapides	Alexander	Alexander Elec. Power	Power Plant-Gas	175 MW	8-12	15.0	28% cap.	2
St. Bernard	Belle Chase	Gulf Oil	Refinery-FCCU	202000 BPD		16.0	78000 BPD/FCCU	6,41
	Chalmette	Tenneco	Hydrogen Plant	22 MMSCFD	98	6.4	Eickmeyer	7,18
			Refinery-FCCU	90000 BPD		4.5	22000BPD/FCCU	6,41
	Meraux	Murphy Oil Co.	Refinery-FCCU	95400 BPD		2.2	10500 BPD/FCCU	6,41
St. Charles	Luling	Monsanto	Ammonia Plant	700 STPD	98	15.3		4
				600	98	13.1		4
				1150 STPD	98	25.1		4
	Montz	LA. P&L	Power Plant-Gas	1250 MW	8-12	171.4	62% cap.	2
		Occidental	Ammonia Plant	330 STPD	98	6.5		4,18
		Union Carbide	Ethylene Oxide	1370 STPD	-	18.6	Carbide, air	5
	Waterford	LA. P&L	Power Plant-Oil/Gas	892 MW	8-12	118.2	39% cap.	2
St. James	Convent	Texaco	Refinery-FCCU	147368 BPD		14.4	70000 BPD/FCCU	6,41
	Donaldsonville	Agrico	Ammonia Plant	1000 STPD	98	4.5		5,8

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

LOUISIANA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
St. James	Donaldsonville	CF Industries	Ammonia Plant	4 x 1000 STPD	98	72.3		5,8
		First Miss. Corp.	Ammonia Plant	1150 STPD	98	24.7		5,8
St. Landry	Opelousas	Opelousas Elec. & Water	Power Plant-Gas	26 MW	8-12	0.8	38% cap.	2
St. Mary	Baldwin	Central LA. Elec. Co.	Power Plant-Gas	427 MW	8-12	41.0	43% cap.	2
	Morgan City	City of Morgan	Power Plant-Gas	33 MW	8-12	5.1	50% cap.	2
Terra Bonne	Houma	Houma Light & Water	Power Plant-Gas	94 MW	8-12	4.5	39% cap.	2
Webster	Minden	City of Minden	Power Plant-Gas	25 MW	8-12	0.8	10% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

MISSISSIPPI

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Adams	Natchez	Miss. P&L	Power Plant-Oil	66 MW	8-12	8.6	29% cap.	2
	Cranfield	CA. Company	Natural Deposits			0.34/well	w/H_2S	
Bolivar	Wright	Greenwood Util.	Power Plant-Coal/ Oil/Gas	23 MW	8-12	4.6	44% cap.	2
	Cleveland	Miss. P&L	Power Plant-Oil	220 MW	8-12	23.9	27% cap.	2
Coahoma	Clarksdale	Clarksdale Pub. Util.	Power Plant-Gas/Oil	12 MW	8-12	0.5	11% cap.	2
			Power Plant-Gas/Oil	11 MW	8-12	0.5	13% cap.	2
Forrest	Hattisburg	Miss. Power Co.	Power Plant-Oil	77 MW	8-12	10.2	25% cap.	2
Hancock	-	Unknown	N.G. Processing	-	99	1.9	Tulsa Pro. Quip.	7
Harrison	Gulf Port	Miss. Power Co.	Power Plant-Coal/Oil	1173 MW	15	220.1	45% cap.	2
	Hinderson	Greenwood Util.	Power Plant-Coal	32 MW	15	9.0	44% cap.	2
Hinds	Jackson	Miss. Power & Light	Power Plant-Gas	383 MW	8-12	15.3	42% cap.	2
		Shell Oil	Natural Deposit	-	77.16	-	White #1	10
Jackson	Pascagoula	Cyso	Ammonia Plant	-	98	13.9		7
		Chevron Chem.	Ammonia Plant	1150 STPD	98	37.6		4
			Refinery-FCCU	294737 BPD		11.5	56000BPD/FCCU	3,41
		Std. Oil-Kentucky	Hydrogen Plant	109 MMSCFD	98	>27.2	Naptha feed	18
		First Chem. Co.	Hydrogen Plant	-	98	2.0		5

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

MISSISSIPPI

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Jones	Moselle	S. Miss. Power Co.	Power Plant-Gas/Oil	177 MW	8-12	12.3	31% cap.	2
Lamar	Purvis	Amerado-Hess	Refinery-FCCU	30000 BPD	-	3.0	14500BPD/FCCU	8,41
Lauder-dale	Meridian	Miss. Power Co.	Power Plant-Oil/Gas	95 MW	8-12	8.8	26% cap.	2
	Jackson	Miss. P&L	Power Plant-Gas	38 MW	8-12	13.5	42% cap.	2
Lowndes	Artesia	Texas Industries	Cement Plant	415000 TPY	-	85.0		3
Madison	Redwood	Miss. Valley Portland	Cement Plant	280000 TPY	-	59.0		3
	Yazoo City	Miscoa	Chem. Plant	-	-	3.4		9
	Virlelia	Conoco	Natural Deposit	-	-	2.8/well		9
	-	Texas Pacific	Natural Deposit	-	99	12 /well	12 ppm H_2S, Yandell #1	10
Rankin	Brandon	Marquette Cement	Cement Plant	238000 TPY	-	47.0		3
		Aurora Pet.	Natural Deposit	-	-	-	Busick #1	10
		Carter Oil	Natural Deposit	-	99.2	10.0/well	2000 ppm H_2S, Kersh #1	10
		Chevron	Natural Deposit	-	99.2	6.0/well	1500 ppm H_2S, COX #1	9,10
		General Crude	Natural Deposit	-	99.2	12.0	2000 ppm H_2S, JA. Spann #1	10
		Lion Oil	Natural Deposit	-	95.2-98.0	6.0	3000 ppm H_2S, Denkman #2	10
		Shell Oil	Natural Deposit	-	99.1	20.0	1.5 ppm H_2S, Hauberg #1	10
					65.3	-	9.4% H_2S, Mashburn #1	10

KEY TO CARBON DIOXIDE SOURCE MAP

Region III

MISSISSIPPI

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2MMSCFD	Comments	Ref
Warren	Vicksburg	Miss. P&L	Power Plant-Oil/Gas	1327 MW	8-12	80.0	20% cap.	2
Washington	Greensville	Miss. P&L	Power Plant-Oil	781 MW	8-12	133.2	54% cap.	2
Yazoo	Yazoo City	Miss. Chem. Corp.	Ammonia Plant	-	-	11.1		8
		Yazoo City. Pub. Serv.	Power Plant-Oil/Gas	19 MW	8-12	2.0	32% cap.	2

KEY TO CARBON DIOXIDE SOURCE MAP

Region IV

WEST VIRGINIA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Barvour	Maidsville	Monongahela	Power Plant-Coal	1152 MW	14.5	313	61% cap.	2
Fayette	-	Ashland Oil	Natural Deposit	-	83	2.8		15,19
	Alloy	Union Carbide	Ferro Alloy	-	-	38	Coal Fired	16,6
Grant	Mt. Storm	Virginia Elec. & Power	Power Plant-Coal/Oil	1662 MW	14.5	459	56% cap.	2
Hancock	Newell	Quaker State	Hydrogen Plant	1.2 MMSCFD	98	0.3		18
Harrison	Haywood	Monongahela Power	Power Plant-Coal/Oil/ Gas	1368	14.5	665	112% cap.	2
Kanawha	-	Cabot Corp.	Natural Deposit	-	51.8	0.3		15
	-	Cities Service	Natural Deposit	-	59.6	0.4		15
	-	Columbia Nat. Gas	Natural Deposit	-	68	3.4		15
	Belle	DuPont	Ammonia Plant	1200 STPD	98	22.8		5
	Cabin Creek	Appalachia Power	Power Plant-Coal/Gas	170 MW	14.5	27.6	26% cap.	2
	Charleston		N.G. Processing	-	-	20	Stearns Rogers	7
	Glasgow	Appalachia Power	Power Plant-Coal/Oil	439 MW	14.5	117	57% cap.	2
		Columbia Gas	Natural Deposit	-	68	3.4		15
	Indian Creek	Columbia Gas	Natural Deposit	130BSCF	66	20 total		10
	Institute	Union Carbide	Plastics	-	-	19	Coal Fired	16,6
	Marmet	Liquid Air	2 Ammonia Plants	-	98	2 x 240		
	Nitro	FMC Corp.	Industrial	-	-	33	Coal Fired	16,6
	St. Albans	Appalachia Power	Power Plant-Coal/Oil	2932 MW	14.5	823	62% cap.	2
	S. Charleston	FMC Corp.	Ammonia Plant	-	98	48	Coal Fired	16,6
		Union Carbide	Chemicals	-	-	25	Coal Fired	16,6

KEY TO CARBON DIOXIDE SOURCE MAP

Region IV

WEST VIRGINIA

County	City/Area	Company	Type of Facility	Capacity/ Reserve	CO_2 Quality%	CO_2 MMSCFD	Comments	Ref
Marion	Graham	Central Operating Co.	Power Plant-Coal/Oil	1105 MW	14.5	259	65% cap.	2
	Rivesville	Monongahela Power	Power Plant-Coal/Oil/ Gas	109	14.5	41.5	65% cap.	2
Marshall	Captina	Ohio Power	Power Plant-Coal/Oil	712 MW	14.5	219	65% cap.	2
	Moundsville	Allied Chemical	Hydrogen Plant	1.8 MMSCFD	98	0.6		5,6
		Ohio Power	Power Plant-Coal/Oil	1632 MW	14.5	461	59% cap.	2
	Natrium	PPG	Ammonia Plant	-	98	76	Coal Fired	16,15,6
Mason	New Haven	Appalachia Power	Power Plant-Coal	1300 MW	14.5		on 1980	2
Plesants	Plesants	Monongahela Power	Power Plant-Coal	2 x 626 MW			on 1979,80	2
	St. Mary's	Monongahela Power	Power Plant-Coal/Oil	215 MW	14.5	63	56% cap.	2
	Willow Island	Am. Cyanamid	Industrial			10	Coal Fired	16,6
Preston	Albright	Monongahela Power	Power Plant-Coal/Oil	287 MW	14.5	108.7	73% cap.	2
Wetzel	New Martins-ville	Mobay Chemical	Hydrogen Plant	60 MMSCFD	98	20.9	Lummus	5,6

KEY TO POTENTIAL CO_2 SOURCES

IN THE LOS ANGELES BASIN

Major Oil Refineries	CO_2 MMSCFD Committed	CO_2 MMSCFD Available	Quality % CO_2
1. Chevron #1	---	13.5	99+
Chevron #2		7.9	93+
Chevron #3		12.2	75+
2. Mobil	5.2	----	
3. Union	5.2	----	
4. Shell (No H_2 plant)	---	----	
5. Shell (No H_2 plant)	---	----	
6. Arco	3.9	----	
7. Texaco	---	14.7	81
8. Gulf	---	2.7	99+
Ammonia Plant			
Collier (Union Oil)	5.2	----	
	19.5	57.9	

Power Plants	No. of Large Boilers	Megawatts	Potential CO_2 MMSCFD
A. LADWP, Scattergood	3	667	75.9
B. Edison, El Segundo	4	1,020	162.1
C. Edison, Redondo	4	1,310	222.5
D. LADWP, Harbor	None	--	--
E. Edison, Alamitos	6	1,950	312.2
F. LADWP, Haynes	6	1,599	303.5
G. Edison, Huntington Beach	4	835	150.0
		7,381	1,226.2

APPENDIX A2. NATIONAL CRYO-CHEMICS, INC.

A STUDY OF THE

NATURAL CARBON DIOXIDE OCCURRENCES

In a

FOUR-REGION AREA OF THE UNITED STATES

PK JOB #5236-X50-109

JUNE, 1978

NATIONAL CRYO-CHEMICS, INC.

Stamford, Connecticut

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CONTENTS

INTRODUCTION

Pullman Kellogg, under Contract EX-76-C-01-2515, has agreed to study for the Department of Energy the use of Carbon Dioxide as an agent for enhanced oil recovery. As one of the tasks in this study, sources of naturally occurring carbon dioxide will be identified.

National Cryo-Chemics, Inc. possesses extensive background in the production and marketing of carbon dioxide by virtue of the experience of Mr. John M. McNeill in these areas. Thus National Cryo-Chemics, Inc. can assist Pullman Kellogg with identification of natural sources of carbon dioxide.

OBJECTIVES

The overall objective of the work to be performed by National Cryo-Chemics, Inc. is to provide Pullman Kellogg professional assistance with the identification and quantification of natural sources of carbon dioxide.

Geographical Regions in which the natural sources of carbon dioxide are to be identified will include the following four regions:

I. Wyoming, Northern half Utah, Northern half Colorado.

II. Southern half Utah, Southern half Colorado, New Mexico, Western half Texas.

III. Eastern Texas, Louisiana and Mississippi.

IV. Kentucky, Ohio, Pennsylvania, Virginia and West Virginia.

In addition to source location, information to be furnished for each source should include:

1. A qualitative description of the source including the approximate content of: carbon dioxide, methane and other hydrocarbons, sulfur compounds -- including hydrogen sulfide, inert materials such as nitrogen, and water content. The average values of components reported for a source should

be a weighted average in accordance with expected flow rates and analysis from individual wells, whenever this information is available.

2. The flow rate expected from the source and an estimate of possible reserves.

3. The pressure of the gas at the source and any information relating to its expected decline.

4. Depth of the source (average).

5. Number of wells per source, and estimated miles of collection piping.

MAIN OBSERVATIONS AND FINDINGS

- The information presented in this report is based upon data gathered from drilling experience, geological, engineering and production estimates of reservoir size and production rates. Because most of the information is based on paper calculations of exploratory drilling and geologic interpretations, it is subject to question.

- Within many of the concerns contacted there is a disagreement as to how their properties should be produced. This will have a definite effect on the economics of the project(s).

- In some cases, in the areas studied, mention is made of fields which do not meet the selection criteria but which might hold interest as a source for a small localized EOR project.

- The concern of the time lag and the cost of same between inception of the project and initiation of cash flow from produced oil (7 years) was universally voiced.

- No consideration has been publically given to the market value of the carbon dioxide co-produced with natural gas.

- In areas of high purity (98% min.) carbon dioxide, with the exception of Mississippi, the only surface treatment necessary will be the dehydration of the gas produced.

- No attempt was made to estimate the amount of surface collection pipe needed at the various areas covered since the spacing and deliverability of the wells has not yet been determined.

- Some of the structures under pilots in West Texas are not taking the carbon dioxide as rapidly as designed. If this becomes an established fact, the daily requirements will have to be downscaled.

- New Mexico is the only state that has established any minimum spacing for CO_2 production. This is 160 acres. All of the other states seem to be leaving it up to the developers of the resource to establish their own minimums.

- There is concern on the behalf of those companies that could be on the eastern pipeline as to the supply of power for compression purposes.

REGION I: - Wyoming, the northern half of Utah, and the northern half of Colorado. The southern boundary for this region was arbitrarily selected as latitude 39° north.

WYOMING

This state has only one large reserve of natural carbon dioxide which would be suitable for oil recovery purposes. That is the Church Buttes Unit located near the center of the Green River Basin in Unita and Sweetwater counties in the southwestern part of the state. The gas analysis as presented below indicates that this Unit represents an interesting Helium prospect.

The particulars on this area are as follows:

Estimated Reserves	7.2 Trillion SCF
Gas Composition:	
Carbon Dioxide	83 ~~00~~.38%
Nitrogen	5.61%
Hydrocarbons	8.68%
Hydrogen Sulfide	2.09%
Helium	0.24%
Well Flow Rates	10.5 MMCFD
Well Flow Pressure	15 psig

The production was from the Madison structure. Well costs were $3,050,000 in 1975 dollars.

Two additional areas of the state were investigated and both have no carbon dioxide of significant quantity to be considered as source potentials. They are the Hamilton Dome Area in Hot Springs county and The Big Piney Field in Sublette county.

NORTHERN UTAH

The northern part of the state has tested high purity carbon dioxide in several locations. There are two that would meet the selection criteria.

The Gordon Creek Area of Carbon county is the largest in the area. Estimated reserves are thought to be in the 1.3 trillion to 2 trillion cubic feet range. This area is one that, as yet, has not been drilled in order to delineate and define the exact size and location of the reservoir. The gas tested to date has analyzed:

Carbon Dioxide	99.5%
Oxygen	0.1%
Hydrocarbons	0.4%
Hydrogen Sulfide	-0-

The expected flow rates and pressures by producing the Sinbad and Coconino zones would be 16 MMCFD at pressures in excess of 3500 psig. The well depths would have to be greater than 12,000 feet and would cost in the range of $1,250,000 (1978) each.

FARNUM DOME AREA

The Farnum Dome Area located approximately 25 miles due east of Gordon Creek in Carbon county has gas with the following analysis:

Carbon Dioxide	98.6%
Nitrogen	0.8%
Hydrocarbons	0.6%
Hydrogen Sulfide	-0-

The wells which were tested flowed at rates of between 10 and 50 MMCFD. Their depths are in the 3000 to 5000 foot range. No information was available regarding their flow pressures. Their shut-in pressures vary between 1000 and 1500 psig.

The reservoir size, based upon available information, is estimated to be 200 billion cubic feet. The structures tested were the Navajo, Sinbad, Kaibab and Coconino.

NORTHERN COLORADO

There has been much speculation about the carbon dioxide at the McCallum Unit in the North Park Field in Jackson county. It has been estimated that there might be as much as 200 billion cubic feet of gas in the Unit. The operators deny any such size reservoir and state unequivocally that their maximum production would be in the 10 to 12 MMCFD range with reserves totaling no more than 50 billion cubic feet.

When first drilled in the late twenties, the gas analyzed:

Carbon Dioxide	92.0%
Nitrogen	3.0%
Hydrocarbons	5.0%
Hydrogen Sulfide	-0-

REGION II: - The southern half of Utah and Colorado. The state of New Mexico and the western half of Texas. The boundary used to separate Utah and New Mexico was latitude 39° north.

SOUTHERN UTAH

The Paradox Basin Area in San Juan county has tested gas with a carbon dioxide content ranging between 92 and 97%. This information was generated during the drilling of six wells in the 1950's. Because of the lack of interest in carbon dioxide at that time, the test data is not complete. However, it does give an indication of the presence and size -- 1 trillion cubic feet - of the carbon dioxide reservoir. The average gas analysis of the wells drilled was:

Carbon Dioxide	93.6%
Nitrogen, Oxygen	5.3%
Hydrocarbons	2.1%
Hydrogen Sulfide	-0-

The wells were drilled to depths of 6900 to 8100 feet. Flow rates varied from 2 MMCFD to 40 MMCFD. The wells were drilled to the Leadville formation

The Escalante Area, Garfield county, has been mentioned as a possible major source of high purity carbon dioxide. The average analysis of two wells drilled in the area is:

Carbon Dioxide	94.6%
Nitrogen, Oxygen	3.9%
Hydrocarbons	1.3%
Helium	0.2%
Hydrogen Sulfide	-0-

The flow rates ranged from 1 MMCFD to 24 MMCFD.

SOUTHERN COLORADO - SOUTHWEST

The McElmo Dome Area in Montezuma county and the Dove Creek-Doe Canyon Area in Dolores county are the two areas of greatest potential in Southwestern Colorado. Montezuma county has experienced the greatest amount of activity to date. Significant finds of high quality carbon dioxide have been reported by the two majors - Shell and Mobil - who have been drilling there. Shell's drilling efforts have been less successful on a well-for-well basis since they have been doing more delineation drilling. To date, Mobil has drilled only in Montezuma county while Shell has also gone into Dolores county.

The area seems to be very rich in high purity carbon dioxide. It is impossible to accurately estimate the size of the reserve due to the confidentiality placed on this information by the companies operating in the area. We believe it to contain 3 - 5 trillion cubic feet of a 99% carbon dioxide, 1% nitrogen. There was no hydrogen sulfide in the stream.

The well depths have been between 8,000 and 9,000 feet. They have flowed at rates between 6 MMCFD and 37 MMCFD, with flow pressures in the 600 to 800 psig range from the Missippian. There has been no decision as to what the eventual production rate of these wells will be but many feel that it will be in the 5 MMCFD to 10 MMCFD range. Both companies are leaning to 640 acre spacings. The Federal Government has established 25,000 acres as a unit size. The estimated well cost in the area is $750,000 (1977).

SOUTHERN COLORADO - SOUTH CENTRAL

The Sheep Mountain Range Area located in Huerfano county has been the center of the carbon dioxide play in this part of the state. Arco has drilled 21 wells to date and feel that they have delineated their reserves, which are estimated to be 1 to 2 trillion cubic feet. Their drilling has followed 160 acre spacings. Their well cost is estimated at $600,000 (1977).

The gas composition is 98% Carbon dioxide with the balance being made up of Nitrogen and Methane. As in the western part of the state, the gas is free of hydrogen sulfide. Well head pressures are in the 400-600 psig range. They will produce from the Dakota and Entrada sands at 3500 to 6500 feet. The well deliverability is expected to be in the 5 MMCFD to 8 MMCFD range with 640 acre spacings.

SOUTHERN COLORADO - SOUTHEAST

The Grande Area in Las Animas county east of Sheep Mountain is thought to contain significant quantities of natural carbon dioxide. The area was originally drilled in the 1940's and has been semi-active since then. Gas analyses, run during 1974 drilling into the Glorietta sand indicated the following composition:

Carbon Dioxide	98.2%
Nitrogen, Oxygen	1.4%
Hydrocarbons	0.4%
Hydrogen Sulfide	-0-

Flow rates and pressures were not measured but are thought to be low. There is an estimated 1-2 trillion cubic feet of gas in the formation. The deliverability and well head pressures seem to be low. The wells drilled are 1800 feet deep with bottom hole pressures in the 150 psig range.

NEW MEXICO

The Northeast Area is the only part of the state known to have large quantities of naturally occurring carbon dioxide. These deposits are believed to be located in Union, Colfax, Harding, Mora and San Miguel counties.

Three majors have drilled in the area with Amoco being the only one to achieve success. Mobil has given up on the Wagon Mound Area and Cities seems very disenchanted with their two wells at Turkey Mountain.

The only data available regarding the area would be Amoco's findings, some of which are still held confidential. They have drilled 32 wells in Union and Harding counties and have completed 21 of them as producers. The gas analysis (average) was:

Carbon Dioxide	99.0%
Nitrogen	1.0%
Hydrogen Sulfide	-0-

Flow rates were in the 0.9 MMCFD to 1.0 MMCFD range at well head pressures of 100-150 psig. The wells were drilled to depths of 2400 to 2900 feet. The estimated cost of each well is $175,000 completed.

New Mexico is the only state studied that has established a minimum well spacing which is 160 acres. From a production point of view, it most likely will be 320 acres.

The acreage under lease in this area, primarily for carbon dioxide production, is about 2 million acres. Estimates of the size of the reserve have varied since there is so little known of the vast area. The quantities mentioned most frequently are between 3 and 10 trillion cubic feet. We believe it to be in the 6 to 10 trillion range.

All of the firms that purchased leases in the area are of the belief that commercial (pressure-volume) quantities of carbon dioxide can be produced on the East and West side of the Sangre de Cristo Arch. To date, the East appears to be the better position. The production in the area is from the Triassic Santa Rosa sandstones, the Permian Glorietta sandstone, and the Permian ABO (Tubb) Arkosic sandstones.

WEST TEXAS

The only large quantities of natural carbon dioxide in this area are already committed to the SACROC Project. This gas is a by-product of the production of natural gas in the Val Verde Basin.

The Elsinore Field in Pecos county presently is producing 40-50 MMCFD of CO_2 from a natural gas stream. The life of the field is not expected to be more than 8 to 10 years which precludes it from consideration as a source. The carbon dioxide represents 47% of the gas in the reservoir which is estimated at 70 billion cubic feet.

There is some co-produced carbon dioxide in the Panhandle but it is too small a quantity for oil recovery uses. It also would be very expensive to collect.

REGION III: - Eastern Texas, Louisiana and Mississippi.

EAST TEXAS

Based upon the criteria established for this study, there are no candidate wells, fields or areas of natural carbon dioxide in this area. There is some produced with and vented from gas operations in the Kilgore Area. However, the quantity is too small to warrant any further study.

Any natural carbon dioxide required for Enhanced Oil Recovery activities in this area will have to be brought in from Mississippi.

LOUISIANA

Unfortunately, Louisiana does not have any significant natural carbon dioxide. Within the state there are many fields which might be candidates for CO_2 injection. However, as in the case of East Texas, the natural carbon dioxide would have to be piped in from Mississippi. Most of the flood candidates are located along the Gulf Coast so a common carrier pipeline could supply both areas with the gas.

MISSISSIPPI

The major high purity natural carbon dioxide deposits in Mississippi are located in Rankin and Madison counties in the central part of the state. There have been some finds in Hinds county but they are lower in carbon dioxide content.

The reservoir is in the shape of an elongated peanut lying northwest to southeast. The higher purity gas lies in the northern portion of the reservoir. The extreme southern portion contains high concentrations of hydrogen sulfide and hydrocarbons.

The information presented in this study will be based upon the northern sector since the lower quality gas to the south requires an economic justification for the separation and sale or disposition of the gases other than carbon dioxide. The average gas analysis of several wells drilled is:

Carbon Dioxide	99.0%
Hydrocarbons	0.4%
Nitroden, Oxygen	0.6%
Hydrogen Sulfide	Less than 200ppm

The flow rates of the wells tested varied from 12 MMCFD to 21 MMCFD, at pressures ranging from 2200 psig to 3600 psig. The wells' depth ranged from 15,000 to 17,000 feet. Production was from the Norphlet and Smackover formations.

The estimated cost to drill and complete a well for CO_2 production is $2,600,000 (1978). Production from such well would be 12 to 15 MMCFD at flow pressures in the 2500-3500 psig range.

A conservative estimate of the reservoir size is a minimum of 3 trillion cubic feet of the higher quality gas. No attempt was made to estimate the amount of carbon dioxide that could be co-produced with hydrogen sulfide and methane. The feeling in the "CO_2 Colony" is that there will be an ample supply of this high purity gas and that the operators will not have to go to the additional expense of producing and separating the lower quality carbon dioxide.

Shell is about to drill their second CO_2 well in an effort to delineate the southern boundary of the structure under their leases. Their first "Carbon Dioxide Well" drilled in May, 1978, flowed 20.5 MMCFD at 3566 psig. The gas composition was:

Carbon Dioxide	99.1%
Nitrogen	0.5%
Methane	0.4%
Hydrogen Sulfide	1.5 ppm

REGION IV: - Kentucky, Ohio, Pennsylvania, Virginia and West Virginia.

WEST VIRGINIA

The two counties in the state that have natural carbon dioxide are Kanawha and Fayette.

The Indian Creek Field located in Kanawah county has large deposits of a carbon dioxide-natural gas mix. Columbia Gas is developing this property for both the CO_2 and natural gas. They expect to produce 200 billion cubic feet of gas over the next 20 years. The composition of this gas is:

Carbon Dioxide	66%
Nitrogen	4%
Hydrocarbons	30%
Hydrogen Sulfide	Trace

The ease of handling the co-produced natural gas in their distribution system makes this a very attractive project.

They plan to produce 30 MMCFD of gas to net them 20 MMCFD of carbon dioxide for their recovery project at Granny Creek. They hope to maintain this level of production for a minimum of 14 years.

The well costs are estimated at $450,000 (1978) each. These will be drilled to depths of 6600 to 7000 feet. Production will be from the Tuscarora formation at the rate of 2 MMCFD per well. Using the gas analysis this would result in a producible reserve figure of 130 billion cubic feet of carbon dioxide.

The commercialization of the carbon dioxide (83%), nitrogen (2%), and natural gas (15%) in Fayette county is reliant upon the regional needs for carbon dioxide. This most likely will not be developed until the Indian Creek Field's CO_2 production is totally committed.

KENTUCKY, OHIO, PENNSYLVANIA & VIRGINIA

All of these states do not have any significant quantities of naturally occurring carbon dioxide.

<u>OTHER AREAS</u>: - Oklahoma, Arkansas.

Nothing was found of any significance in these two states.

Again, the 50 MMCFD requirement was used in evaluating the areas.

APPENDIX A3. AREAS REPROTED TO HAVE SIGNIFICANT CO_2 DEPOSITS

AREAS REPORTED TO HAVE SIGNIFICANT CO_2 DEPOSITS*

STATE	FIELD/ LOCATION	GAS COMPOSITION MOLE PERCENT	PRODUCTIVE STRATIGRAPHIC UNITS	RANGE OF INITIAL POTENTIALS MMSCFD	EXPECTED** SUSTAINED DELIVERABILITY MMSCFD	EXPECTED RANGE OF SHUT IN WELLHEAD PRESSURES PSIG	EXPECTED** RANGE OF INITIAL FLOWING WELLHEAD PRESSURES PSIG	EXPECTED WELL DEPTHS FEET
COLORADO (SOUTHWEST)	MCELMO DOME AREA, DOVE CREEK; MONTEZUMA AND DOLORES CO'S.	>96% CO_2 BALANCE N_2	LEADVILLE, ELBERT	5-37	3-12	2200-2800	600-850	6000-8000
COLORADO (SOUTH CENTRAL)	SHEEP MOUNTAIN AREA; HUERFANO CO.	96-99% CO_2 BALANCE N_2, CH_4	DAKOTA, ENTRADA	3-44	3-12	700-1100	200-700	3400-7000
MISSISSIPPI (CENTRAL)	JACKSON DOME AREA RANKIN & MADISON COS.	70-99% CO_2 MANY WELLS >98% CO_2 MAY CONTAIN FROM TRACE TO 10% H_2S	SMACKOVER, NORPHLET	5-20	4-16	4000-6500	2000-4000	12000-17000
NEW MEXICO (NORTHEAST)	HARDING, UNION, MORA, COLFAX COS.	92-99.7% CO_2 BALANCE N_2	TRIASSIC-SANTA ROSA PERMIAN-GLORIETA, TUBB	.5-6	.1-2.0	250-700	50-500	1500-2800
UTAH (CENTRAL)	GORDON CREEK AREA; CARBON CO.	91-99% CO_2 BALANCE N_2, CH_4	SINBAD, COCONINO	5-16	NA	3900-4200	NA	11000-12000
UTAH (SOUTHWEST)	FARNHAM DOME AREA; CARBON CO.	98% CO_2	NAVAJO, SINBAD, KAIBAB	2.8-16	NA	1000-1500	NA	2800-5000
WEST VIRGINIA (SOUTH CENTRAL)	FAYETTE & KANAWHA COS.	20-60% CO_2 BALANCE N_2, CH_4	TUSCARORA	.5-5	.1-2	800-2900	NA	6000-9000
WYOMING (SOUTHWEST)	CHURCH BUTTES AREA; VINTA & SWEETWATER COS.	83-86% CO_2 2% H_2S BALANCE N_2, CH_4	MADISON	2-10	2-6	NA	LOW	18000-19000

* DATA PRESENTED HEREIN IS INSUFFICIENT FOR RESERVE ESTIMATION
** HIGHLY SPECULATIVE; REQUIRES SUSTAINED PRODUCTION HISTORY FOR CONFIRMATION

APPENDIX B.

THE ASSOCIATED COSTS FOR CO_2 RECOVERY AND TRANSPORTATION

APPENDIX B1.

BASIS OF ESTIMATES FOR INVESTMENT COSTS

TYPE OF ESTIMATE

Analytical estimate of total investment costs, within Process Unit Battery Limits, (U.S. Gulf Coast Location).

DATE OF ESTIMATE

The total investment costs reflect a 4th Quarter, 1978 price level.

GENERAL BASIS OF ESTIMATES

A number of base case estimates throughout the range of study with balance of estimates capacity prorated.

BASIS OF BASE CASE ESTIMATES

EQUIPMENT (Direct)

Budget Quotations Obtained by Engineering:

Pumps, compressors and drivers
(excluding small motor drivers)
Special Equipment

Inhouse Estimates:

Exchangers
Towers
Drums

BULK MATERIALS (Direct)

Factored by Class

FREIGHT

Factored

IN PLACE SUBCONTRACT

Equipment: As Direct Equipment Above
Bulk Materials: Factored by Class

EXCLUSIONS FROM ESTIMATES

Sales & Use Tax
Import Duties
Ocean Freight, Marine Insurance
Other Costs (License Fees, Royalties)
Interest During Construction
Forward Escalation
Land Costs

APPENDIX B2.

BASIS OF OPERATING EXPENSES

UTILITY COSTS

Electric Power		@ $.035/KW-HR		
Cooling Water		@ $0.020/MGAL		
Fuel Gas		@ $2.00/MMBTU		
Selexol		@ $1.00/LB		
Steam	900 psig & 900^{o}F	@ $2.80/MLBS	(1)
	175 psig & 600^{o}F	@ $2.74/MLBS	(1)
	75 psig & 500^{o}F	@ $2.38/MLBS	(1)
	900 psig & 900^{o}F	@ $3.80/MLBS	(2)

INDIRECT COSTS

Operating Labor - 1 Man Per Shift @ $9.00/HR (3)

Operating Supplies - 30% of Operating Labor

Supervision and Overhead - 100% of Operating Labor

Maintenance and Labor - 3.5% of Total Plant Investment Cost

State and Local Taxes - 3.0% of Total Plant Investment Cost and Insurance

Total Operating Cost = Utility Cost + Indirect Cost

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(1) Steam cost based on coal fired steam boiler, and coal price of 50 ¢/MM BTU LHV.

(2) Steam cost based on gas fired steam boiler, and gas price of $2.00/MCF delivered.

(3) 8160 hours of operation (340 days) per year are assumed.

APPENDIX B3.1

COST OF CARBON DIOXIDE BY THE DISCOUNTED CASH FLOW RATE OF RETURN METHOD

The determination of annual revenues (i.e. costs) by the method of a fixed discounted cash flow rate of return (DCFRR) is discussed below. This method includes all cash flows throughout the project's life and adjusts their time-value to one fixed time. The fixed reference point used in all the DCFRR calculations below is the initial time at which full production is achieved (after start-up). For a complete economic analysis, four DCFRR's of 10%, 15%, 20% and 25% are calculated.

BASIS FOR CALCULATIONS

1. 20 year project life (after start-up)
2. 20 year straight line depreciation
3. 48% federal income tax rate
4. 100% equity capital (no debt)
5. Yearly state and local tax plus insurance is equal to 3% of capital investment
6. Discrete interest
7. The salvage value is zero
8. 340 days of operating time per year

DEFINITION OF TERMS

X = Yearly Revenue ($/YR)

O = Yearly Operating Costs ($/YR)

S = Start-Up Costs ($)

I = Total Plant Investment ($)

W = Working Capital ($)

F = Yearly Gas Flow Rate (MMSCF/YR)

G = Gas Cost ($/MMSCF)

Derivations of the 10%, 15%, 20% and 25% DCFRR equations are shown in Tables B3.1 - B3.4. From these tables, the equation for the cost of CO_2 ($/MMSCF) at the specified DCFRR are the following:

$$\text{at 10\% DCFRR, } G = \frac{O + 0.1837I + 0.1175S + 0.1922W}{F}$$

$$\text{at 15\% DCFRR, } G = \frac{O + 0.3072I + 0.1598S + 0.2884W}{F}$$

$$\text{at 20\% DCFRR, } G = \frac{O + 0.4277I + 0.2054S + 0.3847W}{F}$$

$$\text{and at 25\% DCFRR, } G = \frac{O + 0.5620I + 0.2530S + 0.4810W}{F}$$

APPENDIX B3.1

TABLE B3.1

DISCOUNTED CASH FLOW FOR RATE OF RETURN AT 10%

END OF YEAR	(A) GAS REVENUE	(B) TOTAL OPERAT-ING COST	(C) DEPRECIATION	(D) TAXABLE INCOME = A-B-C	(E) NET INCOME AFTER F.I.T. (.52) x D	(F) INVESTMENT	(G) CASH FLOW = C + E - F	(H) DISCOUNT FACTOR	(I) DISCOUNTED CASH FLOW = (G) x (H)
0	--	S	---	-S	-0.52S	(1+0.1)I+W	-0.52S-(1+0.1)I-W	1.0000	-(1.1)I-0.52S-W
1	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.90909	0.4727(X-N)+0.0218I
2	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.8265	0.430(X-N)+0.0198I
3	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.7513	0.391(X-N)+0.0180I
4	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.6830	0.355(X-N)+0.0164I
5	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.6209	0.323(X-N)+0.0149I
6	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.5645	0.294(X-N)+0.0135I
7	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.5132	0.267(X-N)+0.0123I
8	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4665	0.243(X-N)+0.0112I
9	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4241	0.221(X-N)+0.0102I
10	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3855	0.200(X-N)+0.0092I
11	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3505	0.182(X-N)+0.0084I
12	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3186	0.166(X-N)+0.0076I
13	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2897	0.151(X-N)+0.0070I
14	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2633	0.137(X-N)+0.0063I
15	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2394	0.124(X-N)+0.0051I
16	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2176	0.113(X-N)+0.0052I
17	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1978	0.103(X-N)+0.0047I
18	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1799	0.094(X-N)+0.0042I
19	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1635	0.085(X-N)+0.0039I
20	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	-W	0.52(X-N)+0.024I+W	0.1486	0.077(X-N)+0.0036I+0.1486W
Totals	20X	20N+S	1.0I	20X-20N -I-S	10.4X-10.4N -0.52I-0.52S	(1+DCF)I	10.4(X-N)-(0.52+0.1)I -0.52S		4.4287(X-N) +0.8960I-0.52S -0.851W

APPENDIX B3.1

TABLE B3.2

DISCOUNTED CASH FLOW FOR RATE OF RETURN AT 15%

END OF YEAR	(A) GAS REVENUE	(B) TOTAL OPERATING COST	(C) DEPRECIATION	(D) TAXABLE INCOME = A-B-C	(E) NET INCOME AFTER F.I.T. (.52) x D	(F) INVESTMENT	(G) CASH FLOW = C + E - F	(H) DISCOUNT FACTOR	(I) DISCOUNTED CASH FLOW = (G) x (H)
0	--	S	-----	-S	-0.52	(1+0.15)I+W	-0.52S-(1+0.15)I-W	1.000	-1.15I-0.52S-W
1	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.8696	0.4522(x-N)+0.0209I
2	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.7562	0.3932(X-N)+0.0181I
3	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.6575	0.3419(X-N)+0.0158I
4	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.5718	0.2973(X-N)+0.0137I
5	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4972	0.2585(X-N)+0.0119I
6	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4323	0.2248(X-N)+0.0104I
7	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3759	0.1955(X-N)+0.0090I
8	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3269	0.1700(X-N)+0.0078I
9	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2843	0.1478(X-N)+0.0068I
10	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2472	0.1285(X-N)+0.0059I
11	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2149	0.1117(X-N)+0.0052I
12	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1869	0.09719(X-N)+0.0045I
13	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1625	0.0845(X-N)+0.0039I
14	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1413	0.07348(X-N)+0.0034I
15	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1229	0.06391(X-N)+0.0029I
16	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1069	0.05559(X-N)+0.0026I
17	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.09293	0.04832(X-N)+0.0022I
18	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.08081	0.04202(X-N)+0.0019I
19	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.07027	0.03654(X-N)+0.0017I
20	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	-W	0.52(X-N)+0.024I+W	0.06110	0.03177(X-N)+0.0015I+0.061W
Totals	20X	20N+S	1.0I	20X-20N-I-S	10.4X-10.4N-0.52I - 0.52S	(1+0.15)I	10.4(X-N)-(0.52-0.15)I -0.52S		3.255(X-N)-0.998I-0.52S-0.939W

APPENDIX B3.1

TABLE B3.3

DISCOUNTED CASH FLOW FOR RATE OF RETURN AT 20%

END OF YEAR	(A) GAS REVENUE	(B) TOTAL OPERAT- COST	(C) DEPRECIATION	(D) TAXABLE INCOME =A-B-C	(E) NET INCOME AFTER F.I.T. (.52) x D	(F) INVESTMENT	(G) CASH FLOW = C + E - F	(H) DISCOUNT FACTOR	(I) DISCOUNTED CASH FLOW = (G) x (H)
0	--	S	--	-S	-0.52	(1+.020)I+W	-0.52S-(1+0.20)I-W	1.000	-1.20I-0.52S-W
1	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.8333	0.4333(X-N)+0.020I
2	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.6944	0.3611(X-N)+0.017I
3	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.5784	0.3008(X-N)+0.014I
4	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4823	0.2508(X-N)+0.012I
5	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4019	0.2090(X-N)+0.010I
6	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3349	0.1741(X-N)+0.008I
7	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2791	0.145(X-N)+0.007I
8	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2326	0.121(X-N)+0.006I
9	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1938	0.101(X-N)+0.005I
10	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1615	0.084(X-N)+0.004I
11	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1346	0.070(X-N)+0.003I
12	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.1122	0.058(X-N)+0.003I
13	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.0935	0.049(X-N)+0.002I
14	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.07789	0.041(X-N)+0.002I
15	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.06491	0.034(X-N)+0.002I
16	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.05409	0.028(X-N)+0.001I
17	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.04507	0.023(X-N)+0.001I
18	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.03756	0.020(X-N)+0.001I
19	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.03130	0.016(X-N)+0.001I
20	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	-W	0.52(X-N)+0.024I+W	0.02608	0.014(X-N)+0.001I+.026IW
Totals	20X	20N+S	1.0I	20X-20N-I-S	10.4X-10.4N -0.52I-0.52S	(1+0.20)I	10.4(X-N)-(0.52+0.20)I -0.52S		2.532(X-N)-1.083I-0.52S -0.974W

APPENDIX B3.1

TABLE B3.4

DISCOUNTED CASH FLOW FOR RATE OF RETURN AT 25%

END OF YEAR	(A) GAS REVENUE	(B) TOTAL OPERAT-ING COST	(C) DEPRECIATION	(D) TAXABLE INCOME =A-B-C	(E) NET INCOME AFTER F.I.T. (.52) x D	(F) INVESTMENT	(G) CASH FLOW = C + E - F	(H) DISCOUNT FACTOR	(I) DISCOUNTED CASH FLOW = (G) x (H)
0	--	S	--	-S	-0.52S	(1+0.25)I+W	-0.52S-(1+0.25)I-W	1.000	-1.25I-.52S-W
1	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.8000	0.416(X-N)+0.0192I
2	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.6400	0.333(X-N)+0.0154I
3	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.5120	0.266(X-N)+0.0123I
4	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.4096	0.213(X-N)+0.0098I
5	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.3277	0.170(X-N)+0.0079I
6	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2621	0.136(X-N)+0.0063I
7	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.024I	0.2097	0.109(X-N)+0.0050I
8	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.1678	0.0873(X-N)+0.0040I
9	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.1342	0.0698(X-N)+0.0032I
10	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.1074	0.0558(X-N)+0.0026I
11	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.08590	0.0447(X-N)+0.0021I
12	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.06872	0.0357(X-N)+0.0016I
13	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.0550	0.0286(X-N)+0.0013I
14	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.04398	0.0229(X-N)+0.0011I
15	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.03518	0.0183(X-N)+0.0008I
16	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.02815	0.0146(X-N)+0.0007I
17	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.02252	0.0117(X-N)+0.0005I
18	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.01801	0.0094(X-N)+0.0004I
19	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	--	0.52(X-N)+0.24I	0.01441	0.0075(X-N)+0.0003I
20	X	N	0.05I	X-N-0.05I	.52X-.52N-0.026I	-W	0.52(X-N)+0.24I+W	0.01153	0.0060(X-N)+0.0003I+0.0115W
Totals	20X	20N+S	1.0I	20X-20N-I-S	10.4X-10.4N-0.52I -0.52S	(1+0.25)I	10.4(X-N)-(0.52+0.25)I -.52S		2.055(X-N)+1.1551I-0.52S 0.9885W

APPENDIX B3.2

CARBON DIOXIDE TRANSPORTATION COST FROM THE UTILITY FINANCING METHOD

Outlined below is one method for calculating CO_2 transportation costs (annual gas revenue). This method is more commonly known as the "Utility Financing" method as prescribed by the Federal Energy Regulatory Commission (FERC).

BASIS FOR CALCULATIONS

1. 20 year project life
2. Straight line depreciation on total capital requirement excluding working capital
3. 48 percent federal income tax rate
4. 100 percent equity capital (no debt)
5. New pipeline system
6. 340 days per year operating time
7. 3% of investment/yr state and local taxes and insurance

DEFINITION OF TERMS

C = Total Capital Requirement ($)

W = Working Capital ($)

N = Total Net Annual Operating Cost ($)

G = Carbon Dioxide Flow Rate (SCFD)

P = Percent Return On Rate Base (valuation)

RR = Annual Gas Revenue Requirement ($)

The rate base (valuation of a new pipeline project is defined below in Equation 1.:

$$\text{Rate base (valuation)} = C-W \qquad (1)$$

The annual gas revenue requirement, in turn, is calculated using this rate base (valuation) equation. Equation 2. below shows the relationship between the annual gas revenue and the rate base (valuation):

$$RR = P\,(C-W) + 48/52\,P\,(C-W) + 0.05\,(C) + N \qquad (2)$$

The percent return on rate base (valuation) is set at 8% and 10% according to the return allowed for either crude or refined pipeline systems.

Once the annual gas revenue requirement is determined, for comparison reasons only, it is converted into gas costs per unit of CO_2. Equation 3. below shows this procedure:

$$CO_2 \text{ Transportation Cost/MSCF } CO_2 = \frac{RR}{G} \times \frac{1}{340 \text{ Days}} \times 1000 \qquad (3)$$

A sample calculation using this "Utility Financing" method follows.

UTILITY FINANCING SAMPLE CALCULATION

$C = \$18,782,000$

$W = \$0$

$N = \$2,094,000$

$G = 125 \times 10^6$ SCFD CO_2

$P = 8\%$

$$\text{Rate Base (Valuation)} = C-W \qquad (1)$$

$$= \$18,782,000-\$0$$

$$= \$18,782,000$$

$$RR = P\ (C-W) + 48/52\ P\ (C-W) + 0.05\ (C) + N \qquad (2)$$

$$= .08\ (\$18,782,000) + 48/52\ (.08)\ (\$18,782,000) + 0.05\ (\$18,782,000) + \$2,094,000$$

$$= \$5,922,638/\text{YR}$$

$$CO_2 \text{ Transportation Cost} = \frac{RR}{G} \times \frac{1}{340} \times 1000 \qquad (3)$$

$$= \frac{5,922,638}{125 \times 10^6} \times \frac{1}{340} \times 1000$$

$$= \underline{\$0.139/\text{MSCF } CO_2}$$

APPENDIX B4

EQUIPMENT LISTS OF THE SELEXOL PROCESSES FOR VARIOUS NATURALLY OCCURRING CO_2 SOURCES (WITHOUT H_2S)

Feed Gas															
Pressure, psig			250					500					1000		
CO_2 Content, %	10	25	50	75	90	10	25	50	75	90	10	25	50	75	90
Equipment List															
Recycle Gas/CH_4 Product Exchanger	*	*				*	*				*	*	*		
Feed Gas Exchanger			*	*	*			*	*	*				*	*
Recycle Gas Cooler	*	*	*	*		*	*	*	*		*	●	*	*	
Lean Solution Refrig. Exchanger	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*
Recycle Compressor-Interstage Cooler											*	*	●		
CO_2 Absorber	6	*	*	*	*	6	*	*	*	●	6	*	*	*	*
High Pressure Recycle Flash Drum						*	*				*	*			
Medium Pressure Recycle Flash Drum											*	●			
Low Pressure Recycle Flash Drum	*	*	*	*		*	*	*	*		*	*	*	*	
High Pressure Product Flash Drum	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*
Medium Pressure Product Flash Drum		*	*	*	*				*	*				*	*
Low Pressure Product Flash Drum	*	*	*	*	*	*	*	*	*	*	*	*	●	*	*
Compressor Shelter	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*
Recycle Compressor	*	*	*	*		*	*	*	*		*	*	*	*	
Lean Selexol Pump	6	2	2	2	2	6	4	4	2	2	6	2	2	2	2
Lean Selexol Booster Pump	6				2	6	4	4	2	2	6	2	2	2	2
Absorber Bottoms Hydraulic Turbine	2	2	2	2	2	2	2	2	*	*	2	*	*	*	*
CO_2 Product Compressor (& Interstage Coolers & Knockout)	*	*	*	*	*	*	*	●	*	*	*	*	*	*	*
Sub Station	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*
Control House	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*
Absorption Refrigeration Unit	*	*	*	*	*	*	*	*	*	*	*	*	*	*	*

* Denotes 1 item is included, otherwise the number of items required is denoted.

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