

**A STUDY OF CO₂ RECOVERY AND
TERTIARY OIL PRODUCTION ENHANCEMENT
IN THE LOS ANGELES BASIN**

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**Lawrence-Allison & Associates Corporation
Santa Ana, California**



**National Petroleum Technology Office
U.S. DEPARTMENT OF ENERGY
Tulsa, Oklahoma**

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TERTIARY OIL PRODUCTION ENHANCEMENT
IN THE LOS ANGELES BASIN**

Final Report

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INTRODUCTION

Since the onset of the energy crisis, United States industry and government have placed special emphasis on developing ways to decrease dependence on foreign oil. Increasing the fraction of the oil which can be recovered from domestic oil fields is certainly one means toward that end. A promising Tertiary Recovery technique is CO₂ flooding to recover residual oil left by primary recovery and conventional secondary recovery methods such as gas injection and water flood.

As part of a continuing program to develop enhanced oil recovery, the United States Department of Energy commissioned Lawrence-Allison Western Division, (formerly Ameron) to conduct this study.

The tasks included in this study are as follows:

- Evaluate carbon dioxide sources in the Los Angeles Basin.
- Determine the requirements for upgrading and transmitting the gas.
- Write the necessary material specifications.
- Study the Los Angeles Basin oil fields to determine where carbon dioxide may be effectively utilized as an enhanced recovery agent, estimate recovery performance, and evaluate potential economic benefits.

SOURCES OF CARBON DIOXIDE (CO₂) IN THE LOS ANGELES BASIN

The Los Angeles Basin is known primarily as a land-locked air basin, but it also includes approximately 45 operating oil fields. Figure 1 shows the location of the oil fields. The grid system is composed of township lines, nominally 6 miles apart.

In our quest for CO₂, we conducted a search of published data regarding potential sources which included oil refineries, high CO₂ wells, ammonia plants, chemical plants, and electrical power plants. The literature search primarily included the following:

- Publications and data bank of the South Coast Air Quality Management District, (SCAQMD)
- Oil and Gas Journal
- Bulletins of the United States Bureau of Mines

We contacted personnel of the 15 oil refineries in the Los Angeles Basin regarding hydrogen plants, fluid catalytic crackers, and any possible rich CO₂ streams. We also talked to the superintendent of the only ammonia plant in the Los Angeles Basin, the Quality Control Officer of the Southern California Gas Company, plant managers for three of the principal CO₂ distributing companies, and the manager of the gas lab for a major oil company.

Results and Conclusions

The following two groups of CO₂ sources in the Los Angeles Basin appear to be the most feasible and the most economically promising:

- Six oil refineries and one ammonia plant (all near Los Angeles Basin oil fields) have hydrogen plants with by-product streams of concentrated CO₂. The total available (uncommitted) CO₂ from these streams is over 3,000 tons per day. Any or all of these streams could be processed by conventional methods.
- There are seven major electric power plants, all located near oil fields in the Los Angeles Basin. The six larger plants discharge a combined total of 70,000 tons per day of CO₂ from 27 large boilers. Average CO₂ concentration in the flue gas is approximately 14 percent on a dry basis. This very large volume is a good potential source of CO₂ if an economically feasible CO₂ recovery method can be worked out. These sources are listed in table 1 and their proximity to many of the important oil fields is shown in figure 1.

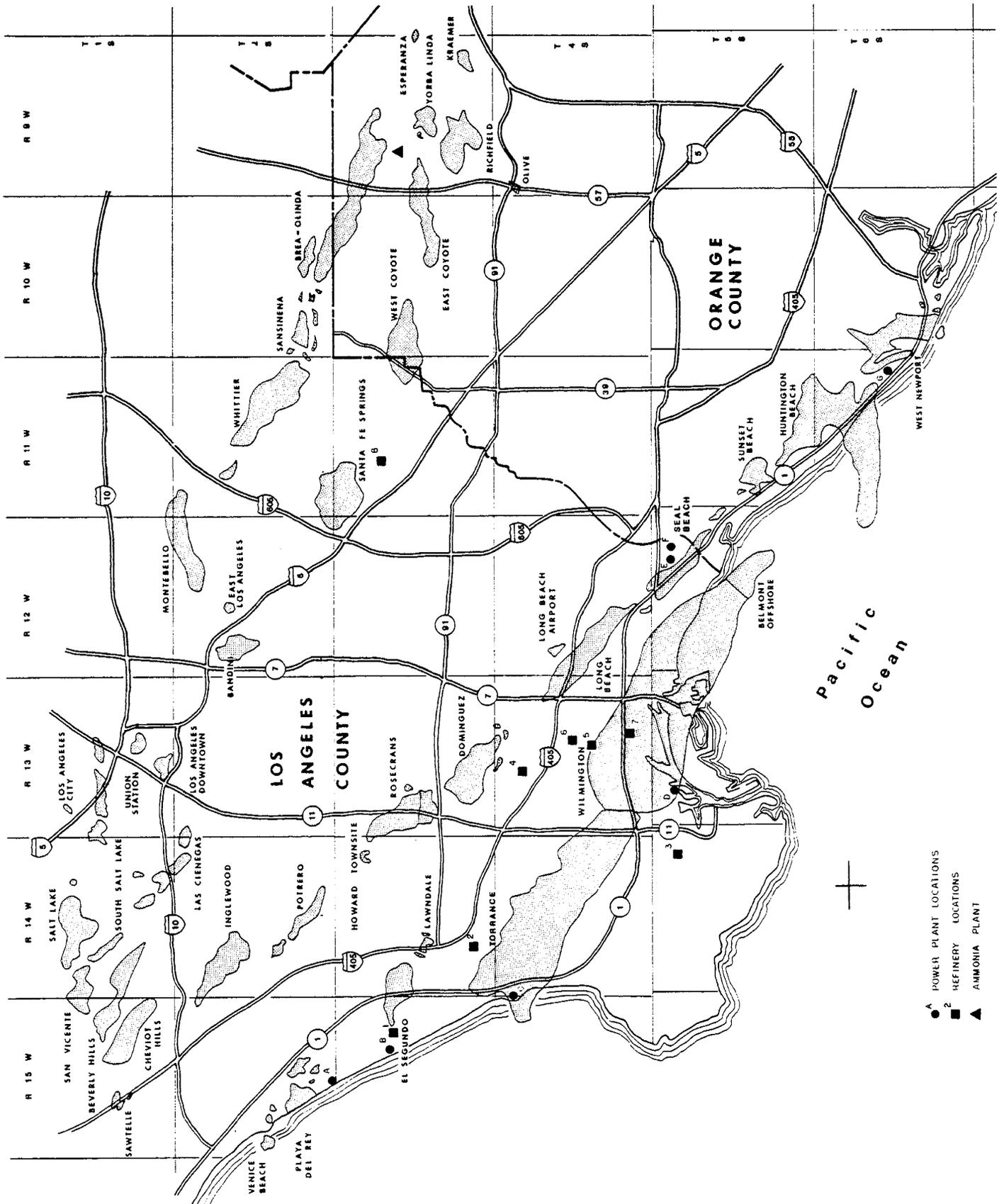


Figure 1. Map of Los Angeles Basin oil fields and CO₂ sources.

TABLE 1
POTENTIAL CO₂ SOURCES

	CO ₂ Tons/day	
Major oil refineries	Committed	Available
1. Chevron	—	1,950
2. Mobil	300	400
3. Union	300	—
4. Shell (No H ₂ plant)	—	—
5. Shell (No H ₂ plant)	—	—
6. Arco	225	—
7. Texaco	—	850
8. Gulf	—	160
<u>Ammonia plant</u>		
Collier (Union Oil)	300	—
	1,125	3,360

	No. of Large boilers	Megawatts	Potential CO ₂ Tons/day
A. LADWP, Scattergood	3	667	4,400
B. Edison, El Segundo	4	1,020	9,400
C. Edison, Redondo	4	1,310	12,900
D. LADWP, Harbor	None	—	—
E. Edison, Alamos	6	1,950	18,100
F. LADWP, Haynes	6	1,599	17,600
G. Edison, Huntington Beach	4	835	8,700
		7,381	71,100

The following potential sources of CO₂ were investigated and deemed insignificant:

- Dozens of combustion stacks in every refinery emit gas containing the usual 14 percent of CO₂. However, these are small individual volumes compared to the stacks in an electric power plant.
- Eight of the oil refineries have fluid catalytic crackers (FCC). Off-gas from an FCC is burned (sometimes in a CO boiler), reportedly resulting in a higher-than-normal CO₂ concentration (as high as 17 percent). After investigation, it was determined that this concentration difference, if any, is insignificant and the FCC-CO boiler was omitted as a potential source of CO₂.
- Lawrence-Allison & Associates found no record of high CO₂ wells in the Los Angeles Basin.
- Fermentation plants (breweries), first thought to be sources of CO₂, proved to be consumers.

REFINERY STREAMS

The available oil refinery streams of CO₂, as shown in table 1, are further described as follows:

<u>Source</u>	<u>Tons/Day</u>	<u>% Purity</u>	<u>Diluent</u>
Mobil	400	99+	
Gulf	160	99+	
Chevron	780	99+	
Chevron	460	93+	Methane & Hydrogen
Texaco	850	81	Nitrogen
Chevron	<u>710</u>	75+	Nitrogen
	3,360		

All the streams are saturated with water and must be dried to prevent corrosion of pipelines and other equipment by the CO₂. If the CO₂ is to be liquefied for tank-truck shipment, it must be very dry to prevent ice formation at the 300-psi, 0°F conditions at which it is transported.

TABLE 2

IMPORTANT LOS ANGELES BASIN OIL FIELDS

(Data from California D.O.G.)

<u>Field</u>	<u>1977 production 10⁶bbl</u>	<u>Reserves* 10⁶bbl</u>
Belmont Offshore	2.07	40
Beverly Hills	4.00	41
Brea-Olinda	3.11	35
East Coyote	0.75	10
West Coyote	1.75	14
Dominguez	0.96	13
Huntington Beach	12.80	128
Inglewood	3.64	26
Las Cienegas	1.42	14
Long Beach	2.58	57
West Newport	1.24	22
Richfield	1.80	25
Santa Fe Springs	0.78	20
Seal Beach	1.04	13
Torrance	2.41	27
Wilmington	52.80	595
Yorba Linda	3.16	38
Total of above	96.31	1,118
Total Los Angeles Basin	101.84	1,179

*Based upon conventional primary and secondary recovery methods. CO₂ flooding could increase these figures.

After drying, the first four streams probably need no further upgrading, only compression and cooling to make them suitable for transporting and injection for enhanced oil recovery.

The last two streams may need further processing to remove most of the diluting nitrogen to produce CO₂ suitable for injection. Recent studies show that the light gases CH₄ and N₂ decrease the solubility of CO₂ in reservoir oil. All of these refinery streams of CO₂ are free of sulfur compounds such as SO₂ that can foul many of the absorbing solvents used in CO₂ purification processes.

POWER PLANT FLUE GAS

Outlined in table 1 is the massive quantity of CO₂ discharged from the stacks of Los Angeles Basin electric power plants. Unfortunately, its concentration is only approximately 14 percent (dry basis). A typical flue gas analysis is as follows:

	<u>Percentage</u>
Nitrogen	74.62
Oxygen	3.30
Carbon Dioxide	12.27
Water	9.80
Sulfur Dioxide	0.01
	<hr/>
	100.00

This source of CO₂, unlike the previously described refinery streams, contains sulfur dioxide, which can foul the solvent used for CO₂ recovery.

SUPPLY FORECAST

The refineries have no immediate plans for modifications which would increase their CO₂ output. CO₂ output from these sources is expected to remain level for the next several years.

Southern California Edison has again asked the State for permission to build a 1,290-megawatt power plant. If the new plant is built at Huntington Beach, one of four possible locations, it would be the least-expensive power plant site for a CO₂ recovery plant. The CO₂ plant would be original equipment (not retro-fit), and the Air Resources Board would require SO₂ scrubbers on the power-plant stacks, meaning that the CO₂ plant would not be required to bear the cost of sulfur removal from the flue gas. This appears to be the only new power plant that may be constructed in the Los Angeles Basin in the next few years and it isn't at all certain.

Overall, no substantial modifications are certain in either refineries or power plants. Both activities are expected to continue at near their present level. We therefore, forecast that the potential CO₂ supply will remain substantially constant during the next several years.

PROCESSING OF CO₂ SOURCES

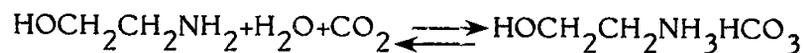
The off-gases discharged by the CO₂ sources must be processed to permit both CO₂ transmission and injection. Processing would consist of compression, drying, and in most cases, purification. The exact order of these steps would depend upon the type of CO₂ source treated and the selected means of purification.

Carbon dioxide compression, which in most cases represents the largest single component of the injected product cost, is absolutely necessary as all sources are at atmospheric pressure. For purposes of estimating compression costs, which are included in the cost summaries in the Transmission Systems section, a wellhead CO₂ injection pressure of 1,500 psi was assumed.

The formation of corrosive carbonic acid in systems containing CO₂ and liquid water dictates drying of all sources containing water vapor. Raw-gas drying is also a necessary prelude to certain purification processes and to the transmission of liquid CO₂ below the freezing point of water.

Depending upon the quality of the source, carbon dioxide purification may be necessary to insure good miscibility of the CO₂ injected into an oil reservoir. Also, as discussed in Transmission Systems, high purity is essential if the product CO₂ is to be hauled by truck or rail to its destination.

The purity of the source is the primary determinant in selecting a purification process. Chemical absorption of CO₂ from the diluent gases is the most viable process for low-pressure, low-purity power plant stack gases. Among the many reactants used by chemical absorption systems, monoethanolamine (MEA) seems to be the most promising because of its high CO₂ capacity, low cost, high stability, and relative ease of regeneration. Purification Alternate No. 1 shows a process flow sketch of CO₂ recovery from power-plant combustion gases via absorption with MEA. The flue gases are first drawn into a fan that boosts the gas pressure by approximately 1.5 psi, thereby imparting the driving force needed to move the gas through the purification facility and out the power plant stack. Next, the incoming gases are cooled to approximately 200°F by heat exchange with the CO₂-depleted diluent gases leaving the purifier. The flue gas then enters the bottom of the vertical absorption tower. The vaporization of water from the MEA absorption solution further cools the gas to 77°F as it rises to the top of the column. Lean absorber liquid containing 15 to 20 percent by weight MEA enters the column at the top tray and flows downward, countercurrent to the rising gas. The overall chemical reaction by which CO₂ combines with water and the MEA absorbent to form a complex is as follows:



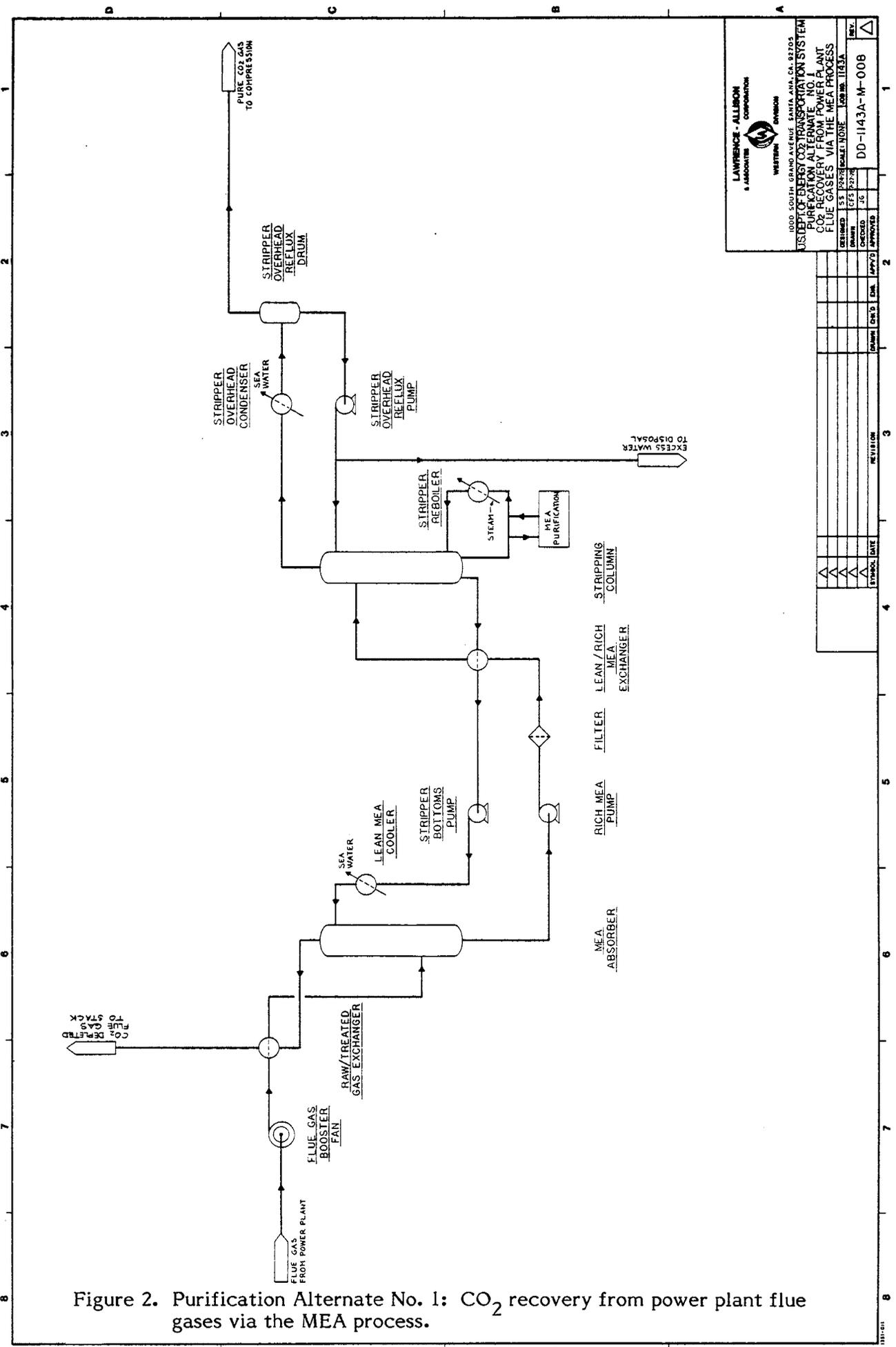


Figure 2. Purification Alternate No. 1: CO₂ recovery from power plant flue gases via the MEA process.

LAWRENCE - ALLISON CORPORATION
 WESTBROOK DIVISION
 1000 SOUTH GRAND AVENUE, SANTA ANA, CALIF. 92705
 U.S. DEPT. OF ENERGY CO₂ TRANSPORTATION SYSTEM
 FLUE GAS RECOVERY FROM POWER PLANT
 CO₂ RECOVERY VIA THE MEA PROCESS
 SCALE: NONE 1:500 000 11/83A
 DRAWN: J.S. 12/27/78 CHECKED: J.C. APPROVED: J.C.
 DD-1143A-M-008 REV. 1

SYMBOL	DATE	REVISION	BY	CHK'D	APP'D
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Lean gas containing a residual CO₂ concentration of 2 to 3 percent (dry basis) leaves the absorber. It is then warmed to approximately 170°F by exchange with the absorber feed gas to provide the buoyancy needed for dispersal after discharge by the power plant stack.

The absorbed CO₂ leaves the column as part of the rich MEA solution pumped from the bottom. This liquid must first be filtered to remove suspended solids such as scale and fly ash, which cause foaming of the MEA. The rich solution is then warmed from 137°F to approximately 195°F by exchange with hot, regenerated MEA before entering the stripping column at the fourth tray. The steam reboiler at the bottom of the stripper adds heat to the MEA solution, causing the above absorption reaction to reverse. The liberated water vapor and CO₂ leave the top of the stripper, while regenerated MEA is concentrated in the bottom. The overhead is cooled with sea water to cause condensation of the water vapor. The pure carbon dioxide remaining in the gas phase can then be separated from the condensate and sent on to compression, drying, and transmission. Part of the condensate is returned to the top of the column to act as reflux while the remaining excess is sent to disposal. Regenerated MEA solution leaves the stripper bottom at 235°F and is cooled to approximately 180°F by exchange with the stripper feed. The lean absorbent is then pumped through a sea water-cooled exchanger and finally returned to the top of the absorber at approximately 75°F.

The 0.25 percent sulfur fuel oil currently burned by coastal power plants produces a flue gas containing 150 ppm SO₂. Sulfur dioxide is a stronger acid gas than CO₂ and reacts rapidly with MEA to form compounds which are thermally stable. These compounds, as well as MEA degradation products produced in the stripper reboiler and MEA oxidation compounds, degrade the CO₂ absorption capacity of the MEA solution. Restoration of MEA reactivity requires either reclamation or purging of the unwanted sulfur compounds from the absorbent solution. One possible reclamation technique is the addition of a strong base such as sodium carbonate or sodium hydroxide to an MEA stream drawn from the stripper reboiler. The MEA, which is a weak base, should be displaced from the SO₂ by the strong-base additive. Subsequent cooling of the solution should cause precipitation of the newly formed sulfates and sulfites, which can then be removed by filtration. The reclaimed MEA remaining in the liquid phase can then be returned to the stripper. Although this process appears feasible, thorough laboratory and pilot-scale testing must first be performed to demonstrate commercial viability. If a satisfactory MEA reclamation process cannot be found, continuous purging of degraded MEA compounds from the recirculating solution or installation of a flue-gas SO₂ scrubber upstream of the MEA unit may be necessary.

The installation of flue-gas SO₂ scrubbing facilities for the sole purpose of permitting CO₂ recovery from power plant gases would be prohibitively expensive. However, future reduction of power-plant SO₂ emissions for air quality reasons, either by further curbing of the fuel-oil sulfur content or by flue-gas scrubbing, is a definite possibility. Standard Oil Company of Ohio (SOHIO) has been attempting to secure approval for an Alaskan

crude-oil terminal at Wilmington. As a condition of approval, the California Air Resources Board requires that SOHIO eliminate 1.2 lb of a neighbor's pollutant for every 1 lb of pollutant that the SOHIO facility would emit. SOHIO and the Air Resources Board have been negotiating with Southern California Edison to install an SO₂ scrubber to treat approximately two-thirds of the flue gases from the 480-megawatt Alamitos Unit No. 6. The total cost of this proposal, including installation of an NO_x-abatement system and the present worth of future scrubber operation costs, is estimated to be \$80,000,000. The actual installation of these facilities would make the Alamitos Station the most desirable location for CO₂ recovery from power-plant flue gases. However, as of this date, the negotiations are in a preliminary stage and no firm commitments have been made.

Regardless of decisions made concerning SO₂ scrubbers, the economics of CO₂ recovery from power-plant flue gases remain unattractive. In the Appendix, under Purification Alternate No. 1, are given capital and operating costs for CO₂ recovery from a 225-megawatt power plant using MEA absorption, followed by compression, drying, and transmission of the purified product. The processing facility (i.e. CO₂ purification, compression, and drying) capital cost, excluding MEA reclamation or SO₂ scrubbing units, is estimated to be \$30,900,000. The combined operating, maintenance, and capital-related costs for such a facility yield a CO₂ product cost of \$1.47 per Mscf. Pipeline transmission adds another \$.06 per Mscf to the product cost, bringing the total field delivered cost to \$1.53 per Mscf.

Regeneration of the MEA absorbent via steam heating in the stripper reboiler represents the largest single component of the product cost. It is estimated that an MEA plant extracting CO₂ from the flue gas from a boiler will use 31 percent of the steam output from that boiler to operate the stripper. Energy trade-off calculations are as follows:

- Burning 1 bbl of fuel oil produces approximately 1,100 lb (9,500 scf) of CO₂ in its flue gas.
- Thirty-one percent of the heating value (0.31 bbl) will be used to provide steam for the stripper.
- Current estimated requirements for a CO₂ drive are 12,000 to 25,000 scf of CO₂ per barrel of crude oil recovered.
- Combining these figures, every barrel of reservoir oil recovered by CO₂ drive may require the burning of 0.39 to 0.8 bbl of fuel oil to provide steam for the MEA stripper.

A further impediment to stack-gas CO₂ recovery is the scarcity of land at coastal power plants. Personnel in the Steam Generation Department of Southern California Edison indicate that what little vacant land is available at the Los Angeles Basin plants can most profitably be used for expansion of generating capacity or fuel storage rather than CO₂ recovery.

The second major CO₂ source, refinery effluent gases, can be upgraded via the three processes listed below:

- Modification of units that produce the CO₂ off-gas.
- Purification via physical absorption of CO₂ in an organic solvent.
- Low-temperature separation, which yields a liquid CO₂ product.

The first process, which may not be possible for all refinery sources, can best be illustrated by examining the CO₂ production at the Chevron Refinery. A hydrogen unit converts steam and methane into hydrogen and carbon dioxide, which is a waste by-product. The hydrogen stream is purified by physical absorption of the CO₂ using a Purisol solvent at high pressures. Solvent regeneration is accomplished first by pressure reduction, then by stripping with nitrogen, thereby introducing a diluent into the desorbed CO₂, which is currently vented to the atmosphere. Alternatively, the purity of the CO₂ could be maintained by slightly warming the solvent after flashing to drive the remaining gases out of solution. The regenerated solvent would then be cooled prior to reuse in the hydrogen purifier. Although this technique would increase the hydrogen unit operation costs, it would completely eliminate the need for downstream CO₂ purification facilities. Where technically possible, this approach promises to be the least-expensive method of obtaining pure CO₂ from that refinery source.

The two remaining purification processes each require upstream compression and drying of the refinery off-gas. Two centrifugal compressors acting in series would be needed to raise the Chevron CO₂ off-gases to the desired purification pressure. Most of the water vapor contained in the first compressor feed would be condensed and separated from the gas phase in the intercoolers and aftercoolers associated with each compressor. Final dehydration can be achieved by running the gas through a short, countercurrent triethylene glycol (TEG) drying column. The dried gas leaving the top of the column is then ready for purification. The six process flow sketches (figures 5 through 10), associated with the transmission alternates in the next section each show the compression and drying equipment, and how they are situated relative to purification and transmission facilities.

Purification via physical absorption is very effective for bulk removal of CO₂ from high-pressure gas streams. Because the CO₂ dissolves in the absorbent but does not react with it, the treated gas must have a high CO₂ partial pressure to force the solute gas into solution. For this reason physical solvents are not recommended if complete CO₂ removal from a gas stream is desired. Chemical absorbents, on the other hand, actually react with the solute gas and are effective even at low partial pressures. However, as noted above, regeneration of chemical absorbents requires expensive high-temperature heat to break chemical bonds, whereas physical solvents are regenerated by pressure reduction or the application of low-temperature heat. This translates into a considerable cost savings for physical absorption processes.

The several commercially available physical absorption processes differ essentially only in the selection of a solvent. Propylene carbonate appears to be the best solvent for refinery off-gas treating because it has a high CO₂ capacity, high chemical stability, low vapor pressure, and low corrosivity toward ordinary steel. Also, it will not dissolve diluent gases such as nitrogen, thereby assuring that a high-purity CO₂ product will be desorbed during solvent regeneration.

Purification Alternate No. 2 shows a process flow sketch of a CO₂ recovery facility using propylene carbonate absorption. Not shown are such necessary support systems as a refrigeration unit and a cooling tower, which are included in the plant cost estimate. Dried refinery gas enters the plant at 600 psi and approximately ambient temperature. It is cooled to approximately 40°F before entering the absorption column by exchange with the cold diluent gases being routed through power recovery turbines. The feed gas then enters the bottom of the absorber and rises countercurrent to the falling propylene carbonate solvent introduced at the top of the column. While in the column, the solvent will rise in temperature from approximately 10°F to 40°F due to the absorption of the latent heat of vaporization released by CO₂ as it is dissolved into the liquid phase. The feed-gas diluent (primarily nitrogen), plus a small amount of unabsorbed CO₂, leave the top of the absorber and pass alternately through power-recovery turbines and exchangers that cool the incoming feed gas. This addition of heat to the waste gases also acts to enhance the total power recoverable from the gas before it is vented to the atmosphere. The recovered CO₂ leaves the bottom of the column dissolved in the rich propylene carbonate. The rich solvent can be regenerated in one of two ways:

- Usually the CO₂ is desorbed by flashing the solvent to atmospheric pressure, either through throttling valves or power-recovery turbines. The use of turbines produces solvent cooling through autorefrigeration, as well as useable power to pump the lean solvent back to the absorber. The autorefrigeration effect helps to reduce the requirement for external refrigeration of the regenerated solvent. The CO₂ product leaves the system at atmospheric pressure and must then be recompressed to permit transmission and injection.
- An alternate regeneration technique that minimizes the overall product cost has been selected. As shown on the Purification Alternate No. 2 diagram, rich solvent from the absorber is first warmed by exchange with the regenerated solvent and desorbed CO₂ gas streams. The solvent is then warmed further to approximately 120°F using waste heat from the upstream raw-gas compression section. Passing through a throttling valve, the propylene carbonate is flashed to 300 psi. Pure CO₂ gas leaves solution and passes overhead as the regenerated solvent is collected in the bottom of a flash drum. At this pressure, the gaseous CO₂ requires no further compression to permit condensation at -10°F and subsequent hauling in pressurized truck trailers. From the flash drum, the separate gas and liquid streams can be cooled to 50°F by exchange with the rich solvent. The CO₂ product stream then leaves the purification unit, while the lean solvent is cooled to 10°F in a refrigerated chiller before returning to the absorber. This chilling, in effect, removes in advance the latent heat of condensation that the solvent must absorb as a consequence of dissolving gaseous CO₂.

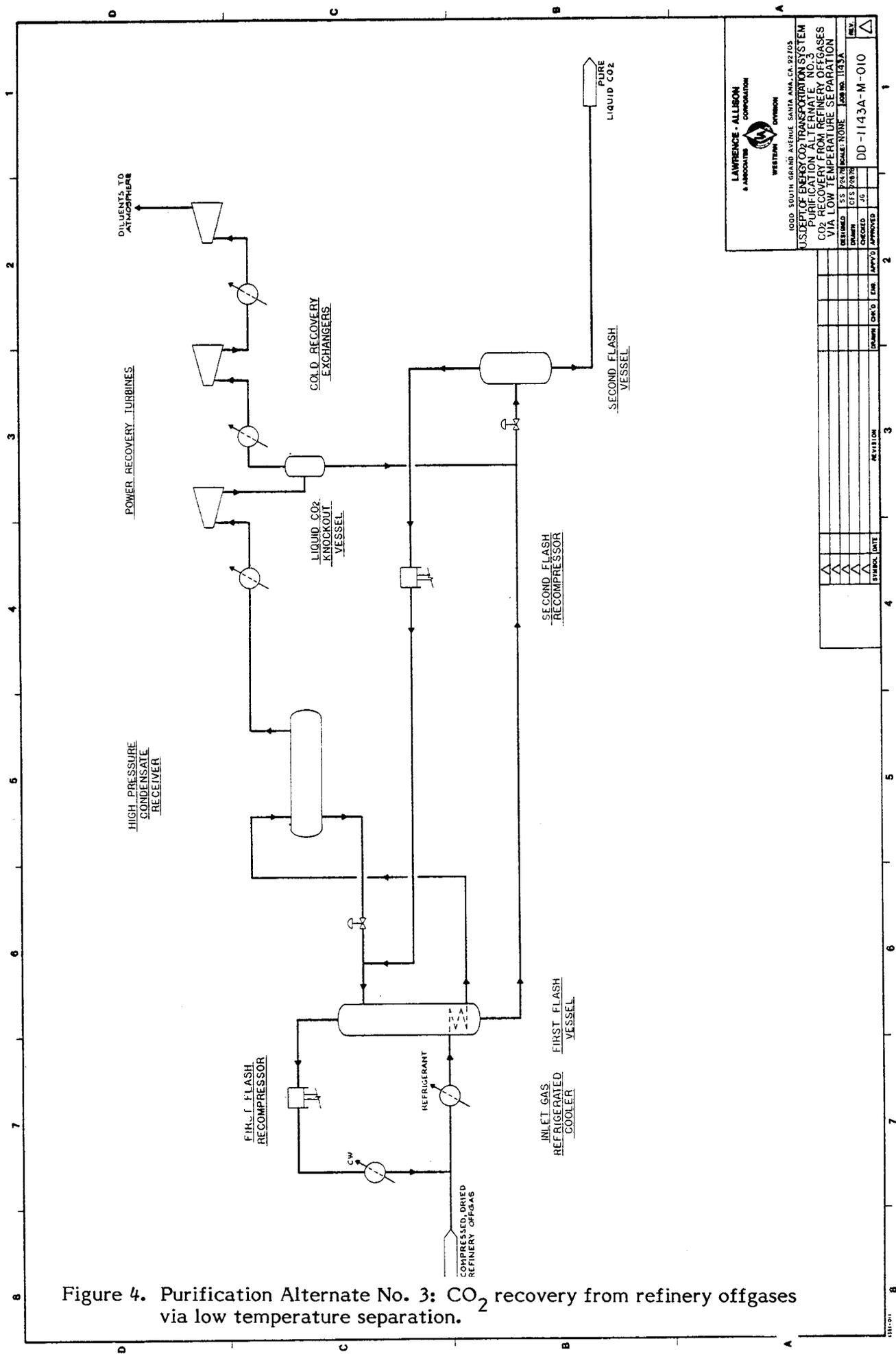
The costs associated with CO₂ purification via absorption in propylene carbonate are computed in the Appendix as part of Transmission Alternates Nos. 1, 3, 5, and 6. These alternates present the entire cost of compressing, drying, purifying, transmitting, and injecting the uncommitted CO₂ output from the Chevron refinery. This gas totals 1,950 tons per day of CO₂ (33.6 MMscfd), with a dry basis composition of 87.5 percent CO₂ and 12.5 percent N₂. It is also saturated with water at atmospheric pressure and 100°F. The capital cost of the purification unit alone, including a refrigeration unit, but not the land value of the plant site, is estimated to be \$4,368,000. The inclusion of this unit adds about \$.31 to \$.32 per Mscf to the product CO₂ cost, assuming 98 percent CO₂ recovery in the purifier. The total processing facility capital costs have been estimated to be \$15,000,000 for production of pure CO₂ gas, and \$16,200,000 to \$17,300,000 for liquid CO₂ production. Corresponding all-inclusive product costs, excluding transmission, are \$.96 to \$.98 per Mscf and \$1.02 to \$1.10 per Mscf.

The third and final refinery-gas purification technique—low-temperature separation—requires both a high CO₂ partial pressure and a high system pressure. It is absolutely essential that the feed gas be anhydrous, since the process operates well below the freezing point of water.

Purification Alternate No. 3 shows a process flow sketch of CO₂ purification via low-temperature separation. Once again, the refrigeration unit and cooling tower that support the process are not shown, but are included in the plant cost estimate. After preliminary compression and drying, the refinery off-gas enters the unit at about 750 psi and ambient temperature.

The feed stream undergoes cooling and partial condensation at -4°F, first in the inlet gas refrigerated cooler then in a set of coils in the first flash vessel. This partial condensation concentrates the less-volatile component, CO₂ in the liquid phase, while the more volatile nitrogen diluent is enriched in the vapor phase. The two phases are separated in the high-pressure condensate receiver. The overhead vapor phase contains the entire nitrogen purge from the system, while the liquid contains the CO₂ product plus some dissolved nitrogen diluent. Since the two phases are in equilibrium, their compositions and relative amounts will be determined by the feed composition and by the receiver temperature and pressure. Assuming the feed from the Chevron Refinery enters the receiver at about 750 psi and -4°F, the overhead vapor would be about 50 percent CO₂ and 50 percent N₂. This estimate was based on extrapolations of experimentally obtained equilibrium data for CO₂ - N₂ systems. More complete data would be required to permit the actual detailed design of a low-temperature separation facility.

The liquid phase from the receiver, while more enriched in CO₂ than the feed gas, still requires further upgrading to reach the desired purity. From the receiver, this liquid passes through a throttling valve to the first flash vessel operating at 300 psi. This reduction in pressure causes the volatile nitrogen and some CO₂ to flash out of solution effecting a further purification of the liquid phase. Heat supplied by the incoming feed prevents the liquid temperature from falling below -20°F as a result of latent heat of



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US DEPT OF ENERGY CO₂ TRANSPORTATION SYSTEM
PURIFICATION ALTERNATE NO. 3
CO₂ RECOVERY FROM REFINERY OFFGASES
VIA LOW TEMPERATURE SEPARATION

DESIGNED: S.S. PETERSON SCALE: NONE JOB NO. 1143A
DRAWN: J.C. GIBSON
CHECKED: J.C.
APPROVED: J.C.

DD-1143A-M-010

REVISION	DATE	BY	CHK'D	APP'D
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3				
4				

Figure 4. Purification Alternate No. 3: CO₂ recovery from refinery offgases via low temperature separation.

vaporization removal by the desorbed gases. This heat exchange prevents metallurgical problems encountered below -20°F , as well as increasing the desorption of gases in the first flash vessel and minimizing the refrigeration duty required to partially condense the incoming feed. The overhead vapor from the first flash is recompressed and combined with the warm feed gas to permit maximum CO_2 recovery. Liquid from the first flash undergoes final purification in the second flash vessel operating at about 275 psi and -20°F . The product liquid leaving this vessel is 99 percent pure CO_2 , suitable for rail or truck hauling. Gas from the second flash is recompressed and combined with the first flash-vessel feed.

The vapor purge from the condensate receiver also undergoes further processing before leaving the unit. Cold recovery exchangers warm the waste gas, thereby increasing the work recovered by then passing the gas through power recovery turbines. Expansion through the turbines recools the gas and may cause some of the CO_2 to condense. This liquid can be separated in a knockout vessel downstream of the first turbine and then routed to the second flash vessel, thereby increasing the recovery of liquid CO_2 product. The exchangers, meanwhile, can be used to precool the unit feed gas upstream of the refrigerated chiller.

It is estimated that 85 percent of the CO_2 in the feed gas will be recovered as a liquid product. The capital cost of a facility designed to produce and handle 99 percent pure liquid CO_2 from the Chevron refinery effluent gases would be about \$13,700,000. A comparable facility using purification via propylene carbonate absorption would cost about \$17,300,000. These costs include off-gas compression and drying, the purification facilities themselves, as well as a refrigeration unit and cooling tower but not to the land value of the plant site. Combining operating, maintenance, and capital-related costs, all inclusive CO_2 product costs of \$1.04 per Mscf and \$1.10 per Mscf were calculated for the respective processes. Low-temperature separation is more economical, at least for liquid CO_2 production, because it requires much less refrigerated cooling. The latent heat of vaporization for CO_2 is removed by the refrigerant only once in the inlet gas refrigerated cooler. The propylene carbonate absorption system, on the other hand, requires about twice this refrigerated cooling duty. The latent heat is first removed from the lean solvent to permit gaseous CO_2 to enter the liquid phase in the absorber. This heat must then be removed a second time to condense the pure CO_2 gas stream leaving the purification unit (see process sketch and Appendix entry for Transmission Alternate No. 1). The capital-cost differential between these two processes is entirely due to the difference in required refrigeration capacities. This savings is partially eroded by the higher raw-gas compression costs and lower product recovery associated with low-temperature separation. The \$.06 per Mscf differential in the product CO_2 cost will be further reduced if a gaseous, rather than a liquid, product is desired.

Itemized capital costs and a computation of the product CO_2 cost for Purification Alternate No. 3 appear in the Appendix.

TRANSMISSION SYSTEMS

This section addresses the problems involved in transmitting carbon dioxide from the identified sources to various oil fields in the Los Angeles basin. Competing transmission systems shall first be described in general terms, then those systems deemed viable will be evaluated with respect to a specific transportation problem involving an actual CO₂ source and a promising nearby oil field. This will permit direct cost comparisons among the transmission alternates. The carbon dioxide source selected for this exercise is the Standard Oil Company of California (Chevron) refinery in El Segundo. Chevron's uncommitted CO₂ output from three streams totals 1,950 tons per day (33.6 MMscfd) and is, by far, the largest of any refinery identified by this report. The hypothetical test oil field will be the Torrance Field, located approximately six or seven miles southeast of the Chevron Refinery.

In a general sense, there are only two means of transmission appropriate for the situations encountered in the Los Angeles basin; namely, hauling of liquefied CO₂ by truck or rail, and pipeline transmission. Each option has a unique set of advantages and disadvantages, as listed below.

- Truck or Rail Hauling

- Advantages

1. Hauling, especially by truck, can easily accommodate changes in routing or destinations.
2. No capital investment in the transmission system itself is required, since trucks and rail cars can be leased. This makes hauling cheaper for short-duration projects.

- Disadvantages

1. Over a long term, hauling is more expensive than pipeline transmission.
2. Logistics problems make hauling impractical for continuous delivery of large volumes of product.
3. The CO₂ product must be transported as a liquid. Therefore, it must be sufficiently pure to condense within the pressure and temperature limits of the hauling vehicles. Thus, the carbon dioxide source may require purification facilities that might not otherwise be necessary.
4. Hauling is relatively labor intensive, and so is vulnerable to interruptions resulting from labor disputes.

- Pipeline Transmission

Advantages

1. Over a long term, pipeline transmission is much cheaper than hauling.
2. Pipelines are well suited to continuous delivery of large volumes of product.
3. The CO₂ can be moved either as a gas or as a liquid.
4. The CO₂ need not be purified before transmission.
5. The operation of a pipeline is relatively immune to disruptions caused by labor problems, foul weather, etc.

Disadvantages

1. A large capital investment is required to construct a new pipeline.
2. A given pipeline serves only a single fixed source and a single destination. A change in source or destination will require at least some additional new pipeline investment.

Truck Hauling—General Discussion

According to several local trucking companies, liquid carbon dioxide is typically hauled in specially designed, insulated, carbon-steel trailers made by the Lubbock Manufacturing Company of Lubbock, Texas. The trailers, which cost approximately \$43,000 each, have a working pressure of approximately 275 psi and a minimum permissible operating temperature of -20°F. Product CO₂ must, therefore, be sufficiently pure to liquefy within these pressure and temperature limits. For CO₂-N₂ systems, 99 percent purity will be required. Such a product could not be hauled in double-walled cryogenic trailers designed to carry other liquefied gases at very low temperatures. The maximum permitted internal pressure of cryogenic trailers is only 35 psig, which is less than the 75.1-psia vapor pressure of pure CO₂ at its freezing point of -69.9°F. Thus, truck hauling of CO₂ will require upgrading of the source to 99 percent purity. This purity requirement may exceed that mandated by the enhanced oil-recovery performance of the CO₂ injected in the field.

The product capacity of a Lubbock CO₂ trailer is 20 tons. Therefore, hauling the 80 tons per hour (1,378 Mscfh) of liquid CO₂ that the Chevron Refinery is capable of producing would require the loading of four truckloads per hour. A local trucking company has estimated that delivery of such a volume to the Torrance oil field on a continuous basis

would require a fleet of 10 to 12 trucks. The per-truck rental cost, including the driver, maintenance, fuel, and insurance, would be approximately \$26 to \$30 per hour. This translates into a hauling cost of approximately \$4.50 per ton, or \$0.276 per MSCF CO₂. As currently designed, the trailer fill and drain lines permit complete loading or unloading in approximately one hour. Enlargement of these lines would be essential to permit faster turnaround times and, thereby, achieve greater truck utilization.

Rail Hauling—General Discussion

Numerous problems and high costs make hauling large volumes of liquefied CO₂ by rail an unattractive transportation alternative.

Delivery times between points within the Los Angeles basin are long and may vary greatly, depending upon the exact locations of the loading and unloading points. For example, a 7-day turnaround was required for the 10-mile round trip between Watson and Wilmington during a recent small-scale test of CO₂ injection in the Wilmington oil field. The railroad charges alone for this project were approximately \$.98 per 100 lb of CO₂ delivered, or \$1.20 per Mscf. These high costs and long delivery times were caused by extensive double handling and transferring of tank cars from one railroad to another, as different jurisdictions were crossed by the CO₂ cargo enroute to its destination. Rail delivery between Chevron's El Segundo Refinery and the Torrance oil field, on the other hand, would require only one day according to personnel with the Santa Fe Railroad. Assuming each tankcar to contain at least 100,000 lb of CO₂ product, a freight-of-all-kinds rate of \$.26 per 100 lb (\$.319/Mscf CO₂) would apply.² Backhauling of empty tankcars from Torrance to El Segundo is included in this freight charge; therefore, no additional costs would be incurred.

To qualify for this relatively low freight rate, large tankcars with a minimum 50-ton net product weight must be used. The largest permitted tankcar of the type used to haul liquefied gases such as CO₂, chlorine, etc., is the Department of Transportation Class 105A500W, which has an 83.25-ton maximum product weight. The car has a 375-psi working pressure and is insulated with 5-in.-thick polyurethane foam encased in an outer steel jacket. Lease costs are approximately \$700 per month, including insurance.

Chevron's potential daily liquefied CO₂ output would fill 24 of these DOT Class 105A500W tankcars. Based upon the optimistic assumption of a one-day round trip between El Segundo and Torrance, a fleet of at least 50 such tankcars is deemed necessary to permit simultaneous CO₂ loading at the refinery and unloading in the oil field. Because of the extensive record keeping, maintenance, and safety responsibilities associated with tankcar ownership, most users choose to lease cars from companies such as General American Transportation Corporation (GATX), which specializes in railcar ownership. The chronic

nationwide railcar shortage, however, poses serious problems to the acquisition of large numbers of specialized tankcars. Sales representatives of GATX, for example, report that 99 percent of their company's cars are currently committed. Construction of new tankcars by GATX requires a minimum five-year commitment from the lessee. Another option would be sub-leasing tankcars for short durations from lessee companies experiencing a temporary surplus of cars. The availability of adequate numbers of cars via this option is, however, uncertain.

Pipeline Transmission—General Discussion

Pipelines are the most economical means of continuously transmitting large volumes of either gaseous or liquefied carbon dioxide over an extended length of time. This conclusion is based upon cost data summarized in table 3 and discussed more fully under Specific Transmission Alternates.

In cases where existing unused but serviceable oil and gas lines can be converted to CO₂ service, pipeline transmission would be competitive for even short-duration projects. Use of such lines eliminates the capital investment that must be amortized over several years to economically justify new pipeline construction. Unused lines can be found with the aid of the Oil Handling Facility Maps of the Greater Los Angeles and Long Beach-Wilmington Areas, obtainable from the Western Oil & Gas Association. These two maps show the routing, product carried, size, and ownership of in-service oil and gas lines. Pipelines available for CO₂ service can be located through inquiries with specific companies shown to have lines in a given area. For example, contacts with the Pipeline Department of the Standard Oil Company of California revealed that there are currently two unused lines in good condition running from the Chevron Refinery to San Pedro via the Torrance Field. One of the lines is 8 in. in diameter over its entire length, while the other varies from 8 to 12 in. The maximum allowable operating pressures of 700 psi and 550 psi, respectively, are too low to permit transmission of liquefied CO₂ at ambient temperatures. Use of both lines, however, would permit gaseous transmission of most, if not all, of the carbon dioxide now being discharged by the Chevron Refinery. The Southern California Gas Company (SCGC) was also contacted. However, SCGC indicated that they currently have no unused transmission or distribution lines suitable for CO₂ service.

A new pipeline must be laid when existing lines are unavailable, undersized, or not suitable for pressurized CO₂ transmission. Before construction may begin, a route or right-of-way must be obtained for the new line. City streets and railroad rights-of-way are the two most viable routes found by this report. Because railroad rights-of-way need little, if any, repaving, they permit lower pipeline construction costs and so are the preferred route for new lines. According to personnel at the Southern Pacific Land Company (a subsidiary of the Southern Pacific Railroad), railroads typically rent longitudinal sections of track rights-of-way, in minimum widths of ten feet, to pipeline users. The minimum width must

be paid for regardless of the number of lines installed or the amount of space actually used. Rental costs for both surface and subsurface use of a right-of-way are 10 percent of the land's value per year. Subsurface rental, which applies to a buried pipeline, is half of that amount. Given that land values in the El Segundo and Torrance areas range from \$4 to \$6 per square foot, the rental for subsurface use of an 8.9-mile-long, 10-ft.-wide strip of land between the Chevron Refinery and the Torrance oil field would be \$117,500 per year. Based upon the Chevron CO₂ output, this translates into a cost of \$.011 per Mscf of CO₂ transmitted. Since a strip of land 10 ft. wide can actually accommodate at least three moderately sized pipelines, the right-of-way rental could be reduced by one-half or two-thirds by sharing the land with another pipeline user and splitting the costs. Unfortunately, use of electric utility rights-of-way is not possible. According to the Rights-of-Way and Land Department of the Southern California Edison Company, longitudinal encroachment of pipelines into overhead power transmission rights-of-way is not permitted. Such an encroachment would compromise Edison's full use of a right-of-way and interfere with possible future undergrounding of the transmission facilities. Perpendicular encroachment to cross a right-of-way, however, is acceptable.

Questions regarding the safety of high-pressure CO₂ transmission lines were discussed with personnel familiar with the design and operation of the SACROC CO₂ pipeline system. They indicated that carbon dioxide may be handled with ordinary, unlined pipeline steel, and that it is no more hazardous to transmit than natural gas. If a pipeline were to break or rupture, the escaping CO₂ would undergo an adiabatic expansion and cool via autorefrigeration to form dry ice. Because most of this cooling will occur at least 1 ft. the rupture, the pipeline will not be further damaged by exposure to very low temperatures. The greatest hazard posed by such an accident would be the settling of the cold, heavier-than-air CO₂ into nearby populated low spots causing suffocation. However, it is unlikely that enough CO₂ to cause such a calamity would be vented by the rupture of a line only 9 miles long. The only major restrictions that must be observed to safely transmit CO₂ are that it must be dry, and that threaded pipe cannot be used because CO₂ washes out thread lubricants. Thus, it is safe to convert existing oil and gas lines to carbon dioxide service as long as these requirements and the maximum allowable working pressures of the lines are observed.

Estimated installed costs for various-sized pipelines over possible transmission routes are summarized below.

	Line size (")	Total installed cost (\$)	Cost per foot (\$)
Chevron Refinery (El Segundo) to Torrance oil field, 8.9 miles (diameter sized for gas flow)	12	2,063,000	43.90
(diameter sized for liquid flow)	8	1,845,000	39.26
Southern California Edison, Redondo Beach Generating Station to Dominguez Hills Field, 14.0 miles	10 18	2,916,000 3,992,000	39.50 - 43.50 54.00 - 59.50
Edison El Segundo to LADWP Scattergood Station, 2.2 miles	16	629,000	54.25 - 59.75
Los Angeles Dept. of Water & Power (LADWP) Scattergood to Inglewood, 10.0 miles	10 18	2,110,000 2,886,000	39.80 - 43.80 54.50 - 60.00
Inglewood Field to Las Cienegas and Beverly Hills Fields, 12.5 miles	10	2,588,000	39.20 - 43.10
LADWP Haynes to Long Beach, 7.5 miles	12 24	1,744,000 2,865,000	44.00 - 48.50 72.35 - 79.60
Edison Alamitos to Wilmington, 7.5 miles	12 24	1,744,000 2,865,000	44.00 - 48.50 72.35 - 79.60
Edison Alamitos to West Coyote Field, 17.6 miles	12 24	4,088,000 6,968,000	44.00 75.00
West Coyote to Santa Fe Springs, Brea Olinda and East Coyote (3 runs of 9 miles each = 27 miles)	10	5,704,000	40.00
East Coyote to Yorba Linda, 4.5 miles	10	952,000	40.00

Specific Transmission Alternates

Each of these alternates represent a technically feasible solution to the problem of carbon dioxide transmission between the Chevron Refinery and the Torrance oil field. An estimate of the total capital cost, itemized by each processing step, as well as an estimated CO₂ product cost for each alternate, appears in table 3. This CO₂ cost is all-inclusive, representing a total of labor, maintenance, utility, tax, depreciation, and return on investment charges. A detailed computation of this product cost for each alternate, showing operating costs and capital cost estimates for each piece of equipment, is given in the Appendix. Finally, simplified process flow sketches for each of the six alternates are presented in figures 5 through 10.

Alternate No. 1—Transport Liquid CO₂ by Truck

As figure 5 indicates, raw CO₂ off-gas from the Chevron Refinery is first compressed to approximately 600 psi and then dehydrated in a triethylene glycol drying unit. These first two processing steps are common to all of the transmission alternates. The water-free gas is then purified. For the sake of consistency, the physical absorption process using a propylene carbonate solvent discussed in the Processing Section, has been selected for each of the transmission alternates requiring CO₂ purification. This will facilitate cost comparisons among the transmission systems only. The upgraded CO₂ gas leaving the purification unit at 300 psi and 50°F is then condensed by cooling to approximately -10°F. For purposes of calculating the refrigeration compressor horsepower, and thus its capital and operating costs, the refrigerant was assumed to be propane. It is anticipated that a single refrigeration unit will handle the cooling duties of both the purification unit and the CO₂ product condenser. The capital costs assigned to the purification and condensation process steps include a proportionate share of the refrigeration unit cost, based upon the respective shares of the total cooling duty. Condensed CO₂ product is sent to pressurized and refrigerated storage vessels to await shipment. The truck trailers used to haul the product are insulated, but not refrigerated, so a small amount of CO₂ will vaporize and be lost by venting during transit. Upon arrival in the oil field, the trucks will be unloaded into similar insulated storage vessels, which provide a surge capacity that permits continuous CO₂ injection at a constant rate. It is estimated that a single such vessel designed to contain 2 hrs. liquid carbon dioxide production will measure 12 ft. in diameter by 40 ft. tangent-to-tangent. A high head pump taking suction on the storage vessel discharges the CO₂ at the desired 1,500-psi injection pressure. The liquid is then warmed to approximately 45°F to avoid the formation of ice in the injection well that would occur if cold CO₂ were used. The CO₂ product may then be routed to a suitable well for injection.

As indicated in table 3, the overall CO₂ product cost associated with this transmission alternate is estimated to be \$1.39 per Mscf. The reason for including the costs of compression, purification, condensation, storage, etc., in the product cost is that these processing steps are made necessary by the selected means of transmission—in this case, hauling by truck as a low-temperature liquid.

Alternate No. 2—Transport Unpurified CO₂ Gas via Existing Pipelines

This alternate assumes that either, (1) the refinery units that vent CO₂ off-gas can be modified to produce a pure gas stream rather than one containing a nitrogen diluent, or (2) the presence of the nitrogen diluent will not significantly degrade the oil recovery performance of the injected CO₂. Since this option represents the minimum possible investment in processing and transmission systems, it also provides the CO₂ product to the oil field at the minimum cost.

Like the other transmission alternates, this system includes raw gas compression and drying facilities. The dehydrated gas will enter the two existing pipelines at their respective maximum operating pressures of 550 psig and 700 psig. These lines are actually somewhat undersized for this service and may cause a pressure drop of as much as 20 psi per mile. In the oil field, a recompressor will raise the gas pressure to approximately 1,500 psi to permit injection. The all-inclusive product cost of CO₂ delivered to the wellhead is estimated to be \$.68 per Mscf.

Alternate No. 3—Transport Purified CO₂ Gas via Existing Pipelines

This alternate assumes that pure CO₂ will be required for oilfield injection and that modification of the refinery units to produce pure CO₂ will not be possible. It is identical to Alternate No. 2, except that a purification unit and booster compressor have been added upstream of the pipelines. The booster compressor is needed to raise the CO₂ pressure from 300 psi as it leaves the purification unit to the pipeline transmission pressures of 550 psig and 700 psig. The all-inclusive CO₂ product cost will be \$.99 per Mscf.

Alternate No. 4—Transport Unpurified CO₂ Gas via a New Pipeline

This alternate makes the same assumptions as Alternative No. 2, differing only in the selected means of transmission. The use of a new, high-pressure pipeline allows transmission of the gas at approximately 1,600 psi and thus eliminates the need for an oilfield recompressor. A booster compressor in the refinery, just upstream of the pipeline, essentially replaces the recompressor. The all inclusive CO₂ product cost will be approximately \$.71 per Mscf.

Alternate No. 5—Transport Purified Gas via a New Pipeline

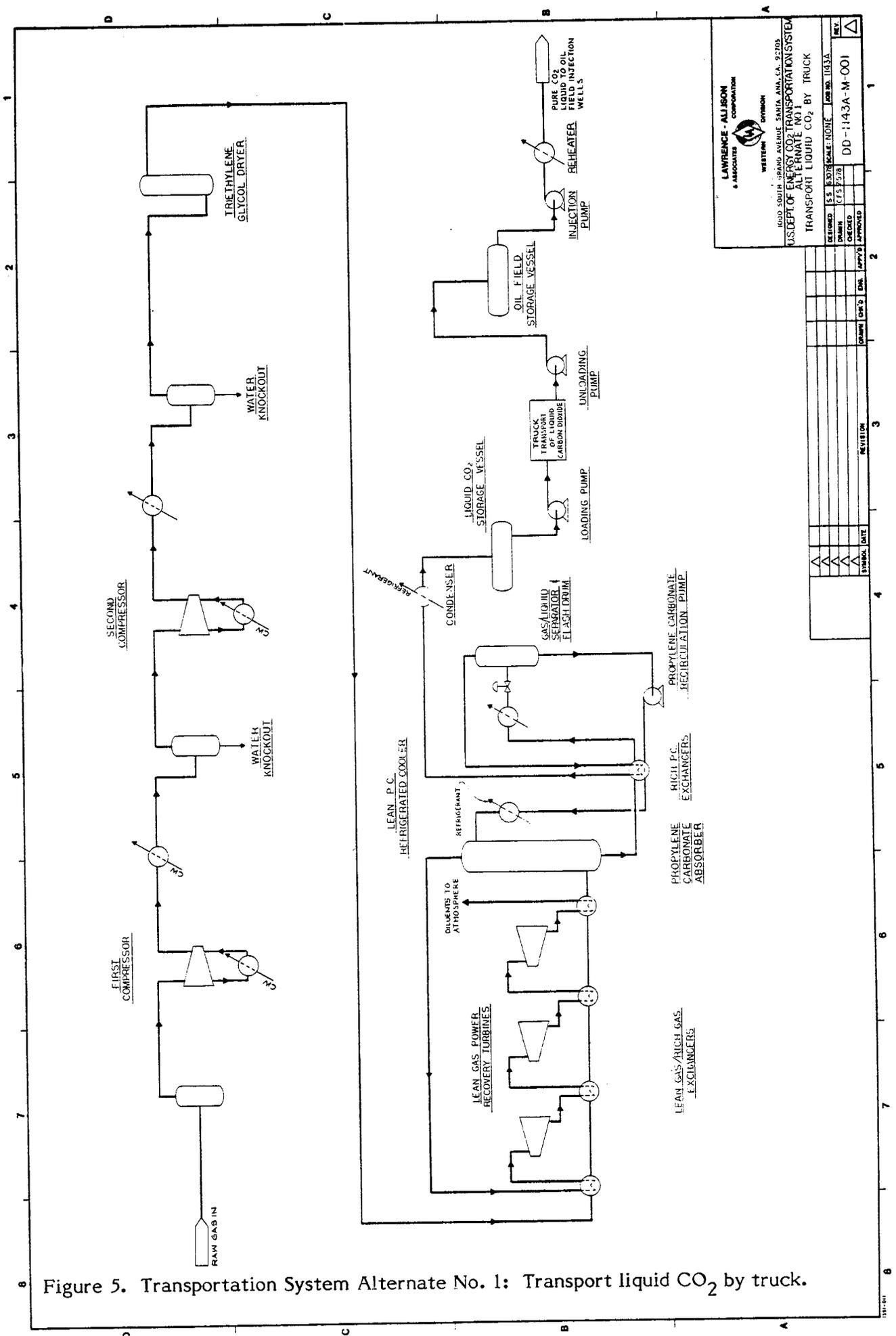
Assumptions stated for Alternate No. 3 are also made here. The use of a new pipeline permits elimination of the oilfield recompressor contained in Alternate No. 3, while the booster compressor must now raise the pure CO₂ pressure from 300 psi to 1,600 psi before transmission. The estimated CO₂ cost delivered to the injection well is \$1.03 per Mscf.

Alternate No. 6—Transport Pure Liquified CO₂ via a New Pipeline

This alternate makes the same assumptions as Alternate Nos. 3 and 5. It differs from Alternate No. 5 only in that a pure liquid, rather than a pure gas, is transmitted by a new pipeline. The processing steps are, therefore, identical up to and including the purification unit. In this alternate, however, the pure CO₂ leaving the purification unit at 300 psi and 50°F is condensed at -10°F by refrigerated cooling. The liquid is collected in a receiver from which it is then drawn by the pipeline feed pump, which discharges the product at approximately 1,600 psi. Just before entering the pipeline it is warmed to about 45°F by heat exchange with the CO₂ gas from the purifier. This cold recovery step removes sensible heat and causes a partial condensation of the pure CO₂ gas before it enters the refrigerated condenser, thus reducing the refrigeration unit cooling duty. Upon arrival in the oil field, the product CO₂ can be routed to the injection wells without further processing. The estimated CO₂ product cost for this alternate is \$1.09 per Mscf.

Analysis of Transmission Alternates

Considerable insight may be obtained by analyzing the cost differentials between various transmission alternates. For example, comparison of Alternate No. 4 with No. 2 and Alternate No. 5 with No. 3 suggests that CO₂ transmission via a new pipeline, rather than an existing one, will add \$.03 to \$.04 per Mscf to the cost of the product CO₂ delivered to the wellhead, assuming all other processing steps are the same. This cost differential is based on 10-year straight-line depreciation of a new pipeline, and capital-related costs as explained in the Appendix. Likewise, comparison of Alternate No. 3 with No. 2 and Alternate No. 5 with No. 4 indicates the CO₂ purification via physical absorption with a propylene carbonate solvent adds \$.31 to \$.32 per Mscf to the product cost. The largest single component of the product cost is the \$.68 per Mscf (Alternate No. 2) required to compress the CO₂ source from atmospheric pressure to the 1,500-psi injection pressure and to overcome pressure losses throughout the system. The cost differential between Alternates No. 6 and No. 5 indicates that, at least when propylene carbonate purification is used, the cost of producing and transmitting a pure liquified CO₂ product is \$.06 per Mscf greater than the cost of a pure gas. The difference arises from the higher refrigeration duty required to make a liquified product. Finally, the \$.30 per Mscf cost differential between Alternates No. 1 and 6 is caused entirely by the higher costs associated with truck hauling relative to pipeline transmission of the same product; i.e. liquified CO₂.



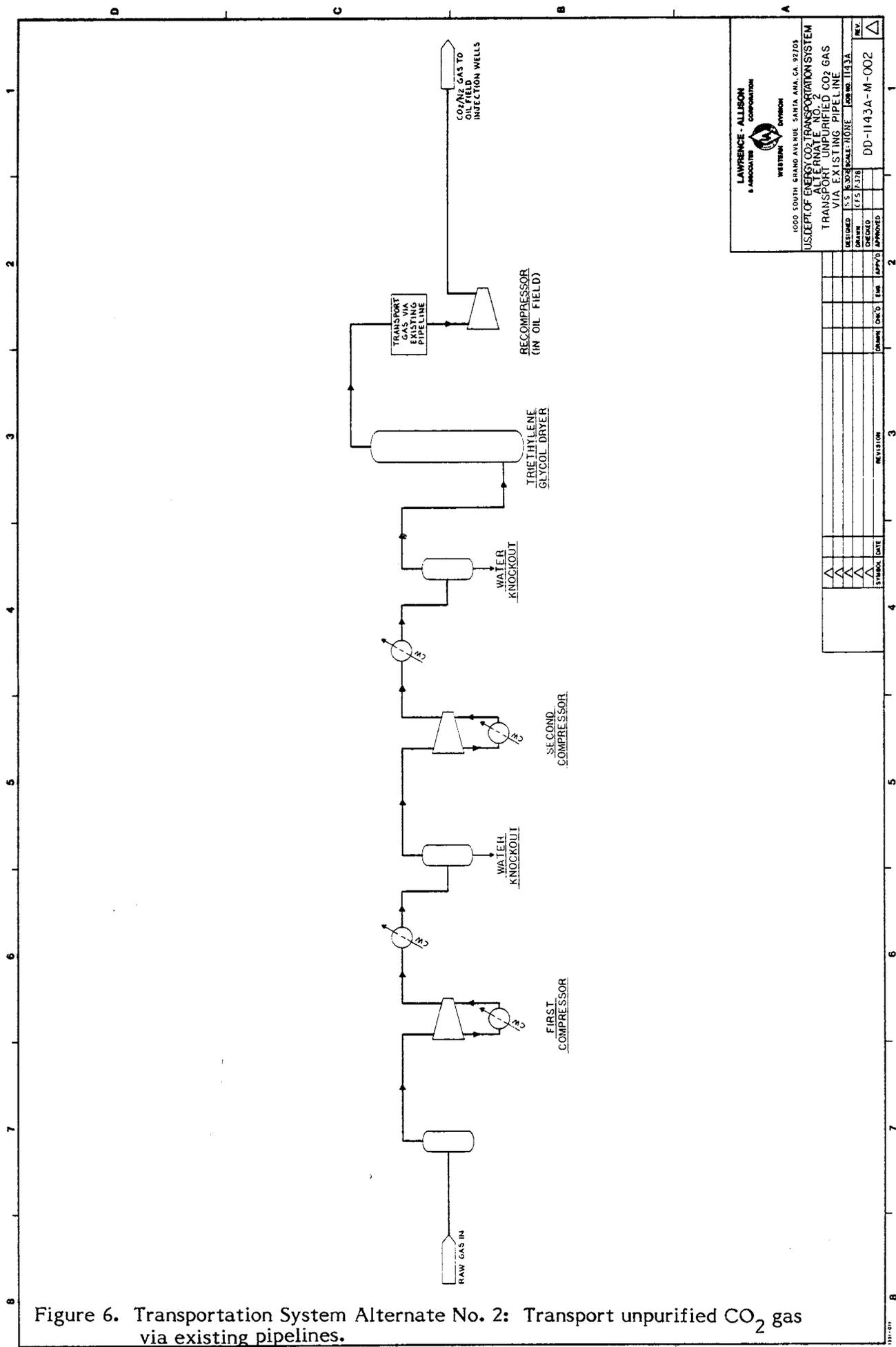
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 & ASSOCIATES
 WESTBROOK, CONNECTICUT

1000 SOUTH GRAND AVENUE SANTA ANA, CA. 92705
 US DEPT OF ENERGY CO₂ TRANSPORTATION SYSTEM
 ALTERNATE NO. 1
 TRANSPORT LIQUID CO₂ BY TRUCK

REVISED: 5.5.83 BY: S.M.E./M.C.H./N.H.E. JOB NO. 1143A
 DRAWN: C.F.S./S.B. DATE: 11/83
 CHECKED: J.M. DATE: 1/84
 APPROVED: DD-1143A-M-001

NO.	DATE	BY	APP'D	REVISION
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3				
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Figure 5. Transportation System Alternate No. 1: Transport liquid CO₂ by truck.



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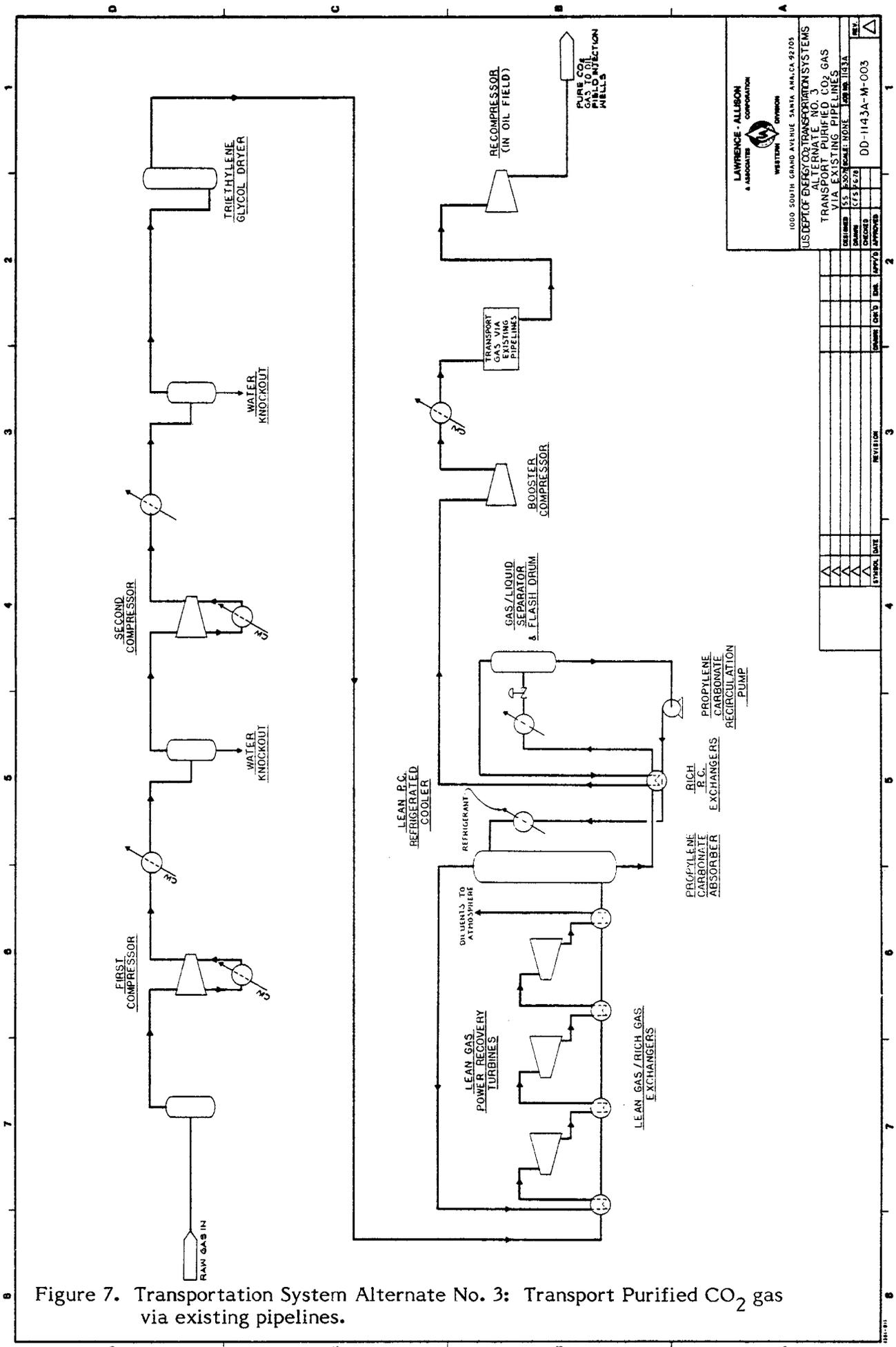
U.S. DEPT. OF ENERGY CO₂ TRANSPORTATION SYSTEM
 ALTERNATE NO. 2
 TRANSPORT UNPURIFIED CO₂ GAS
 VIA EXISTING PIPELINE

DESIGNED	U.S. E.O. 12958	SCALE: HOSE	JOB NO. 1143A
DRAWN	CF5	1/78	
CHECKED			
APPROVED			

DD-1143A-M-002

REV.	DATE	BY	CHK'D	APP'D
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Figure 6. Transportation System Alternate No. 2: Transport unpurified CO₂ gas via existing pipelines.



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 WESTERN DIVISION
 1000 SOUTH GRAND AVENUE SANTA ANA, CALIF. 92705
 US DEPT OF ENERGY CO2 TRANSPORTATION SYSTEMS
 ALTERNATE NO. 3
 TRANSPORT PURIFIED CO2 GAS
 VIA EXISTING PIPELINES

REV.	DATE	BY	CHK'D	APP'D	REVISION
1	11/13/83	LS	LS	LS	
2	1/15/84	LS	LS	LS	
3	1/15/84	LS	LS	LS	
4	1/15/84	LS	LS	LS	

Figure 7. Transportation System Alternate No. 3: Transport Purified CO₂ gas via existing pipelines.

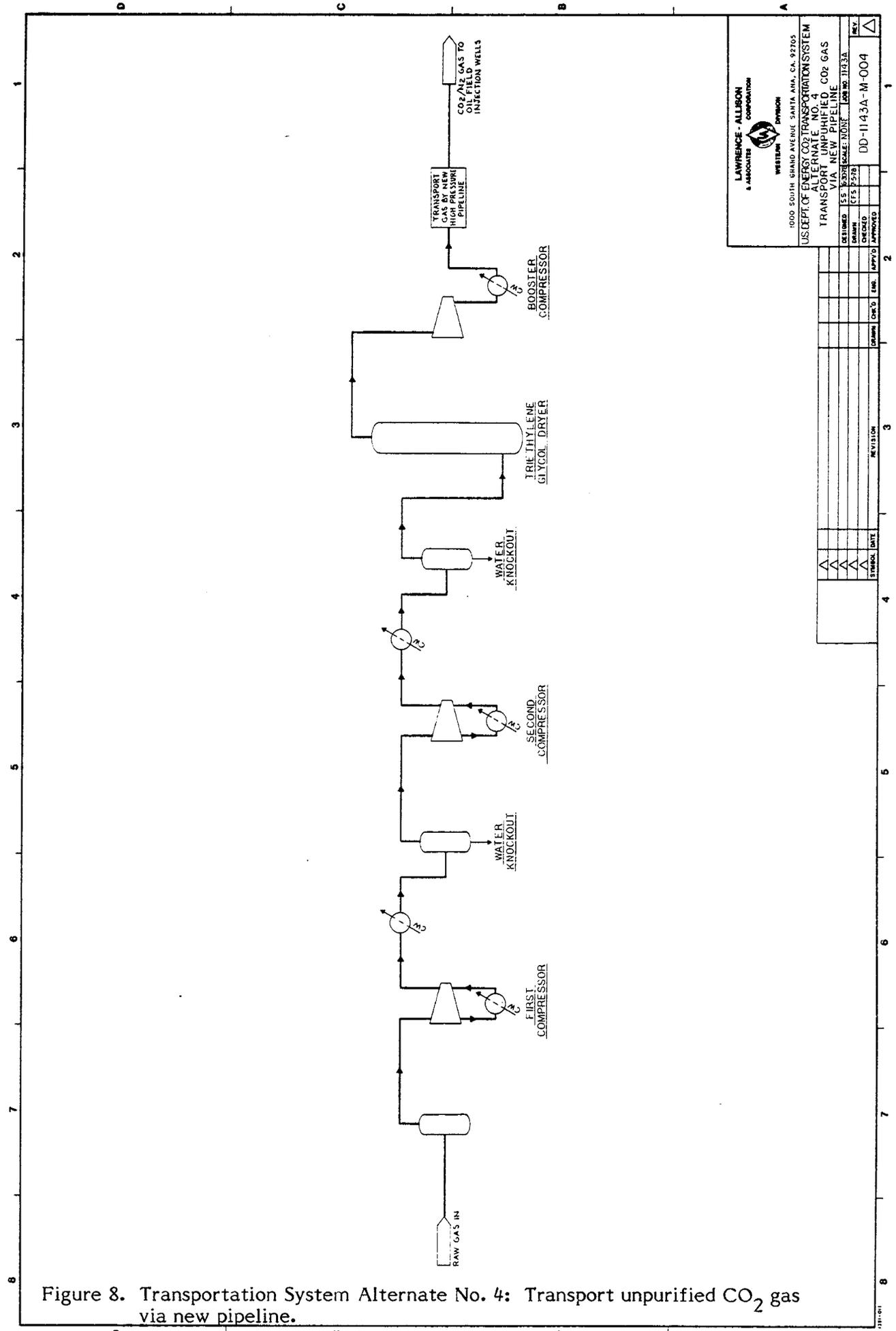


Figure 8. Transportation System Alternate No. 4: Transport unpurified CO₂ gas via new pipeline.

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 WESTBURN DIVISION
 1000 SOUTH GRAND AVENUE, SANTA ANA, CA. 92705

US DEPT. OF ENERGY CO₂ TRANSPORTATION SYSTEM ALTERNATE NO. 4
 TRANSPORT UNPURIFIED CO₂ GAS VIA NEW PIPELINE
 PROJECT NO. 1193A
 DRAWING NO. 2-278

DESIGNED BY: []
 CHECKED BY: []
 APPROVED BY: []

DD-1143A-M-004

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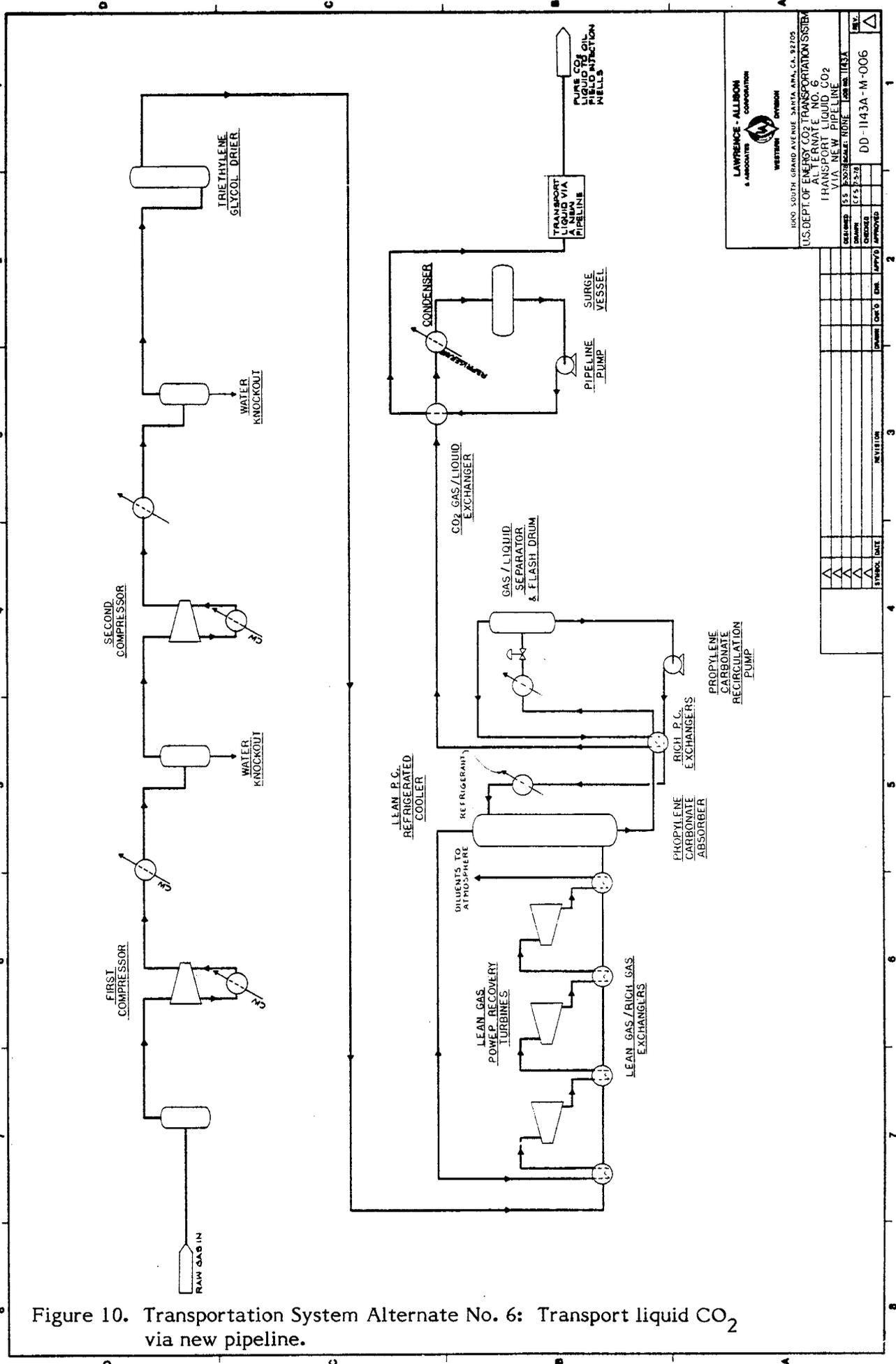
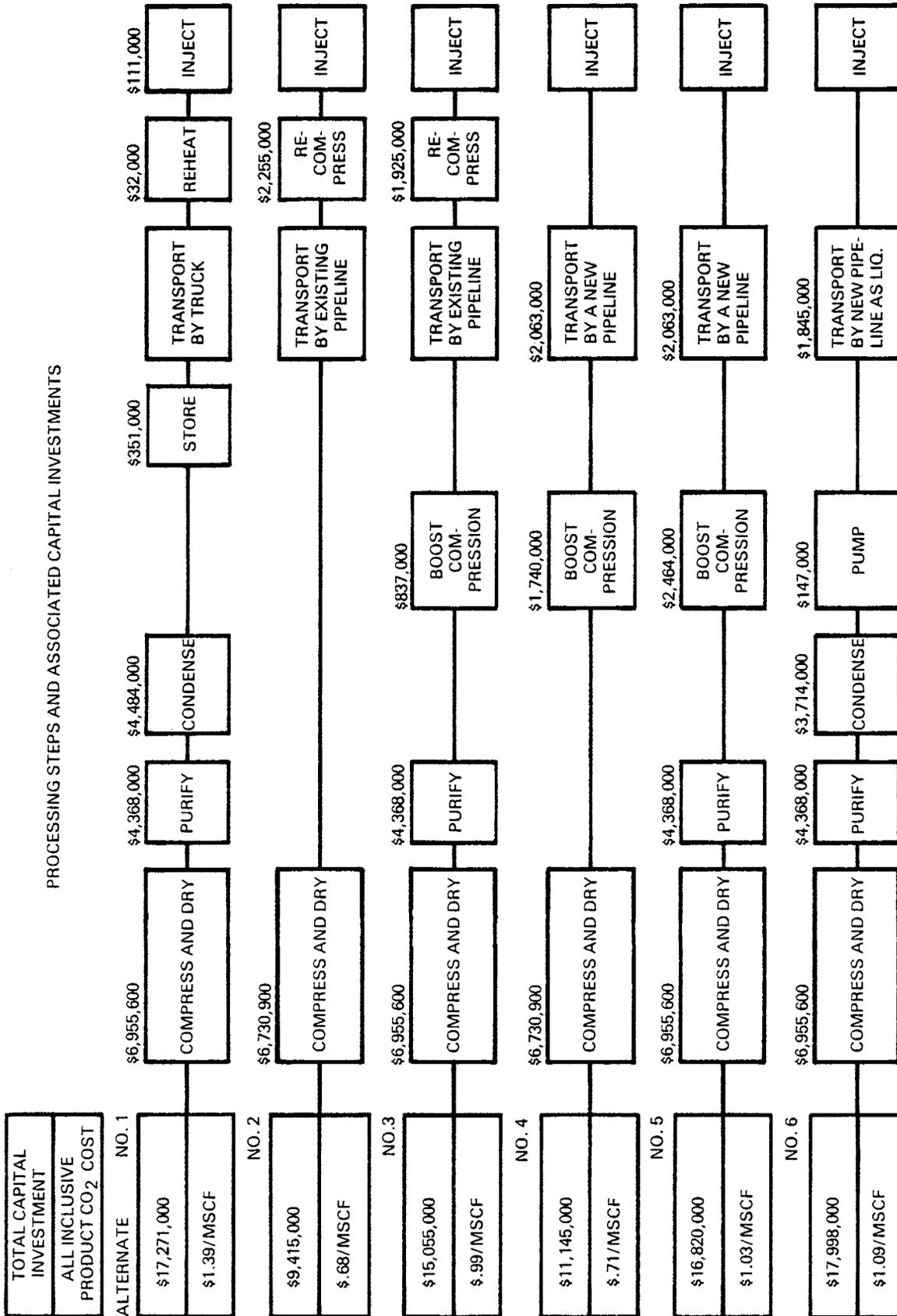


Figure 10. Transportation System Alternate No. 6: Transport liquid CO₂ via new pipeline.

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 1000 SOUTH GRAND AVENUE, SANTA ANA, CA 92705
 U.S. DEPT. OF ENERGY CO₂ TRANSPORTATION SYSTEM
 TRANSPORT LIQUID CO₂ VIA NEW PIPELINE
 DRAWN: J.S. CHECKED: C.F.S. DATE: 11/83
 DD-1143A-M-006

SYMBOL	DATE	REVISION	DESIGNER	ENGR.	APPROVED
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PROCESSING STEPS AND ASSOCIATED CAPITAL INVESTMENTS



TRANSPORTATION SYSTEMS COST SUMMARY

Table 3

MATERIALS

Material specifications for equipment handling CO₂ are included in the Appendix. The specifications can be summarized as follows:

Metals in dry service (CO₂ with a water content less than the saturation value at the operating conditions).

Under dry conditions CO₂ is noncorrosive to iron-base alloys, and the design of handling and distribution systems can be based upon the use of plain carbon steels according to the following codes as they pertain.

- ASME, Boiler and Pressure Vessel Code, Section 8
- ANSI B31.3, Petroleum Refinery Piping
- ANSI B31.8, Gas Transmission and Distribution Piping Systems

Metals in wet service (CO₂ with a water content greater than the saturation value at the operating conditions, i.e. free water).

CO₂ is wet in the dehydration facilities and the lines feeding this equipment.

When free water is present, CO₂ dissolves and forms carbonic acid (H₂CO₃), causing a reduction in pH of the water, making it quite corrosive to plain carbon steel.

All vessels exposed to wet CO₂ shall be designed and constructed according to ASME Section 8 of materials as follows:

- Annealed 304L stainless steel is the most economical of the stainless steels, but 316L and Carpenter 20 Cb-3 may also be considered.
- Carbon steel internally coated with a minimum of 12 mils of a baked epoxy phenolic may be considered for large vessels, but close attention must be given to the preparation of the metal and the application of the coatings to be sure the final coating is free of pinholes. Internals should be made of stainless steel.

Piping, valves, and fittings should be according to Material Specification 1143A-19-C1A and C3A attached.

- Austenitic stainless steel stabilized with columbium or extra-low-carbon stainless steel shall be used to avoid intergranular corrosion.
- Dissimilar metals shall not be used unless they can be selected and arranged to minimize electrolytic corrosion.
- Weld neck flanges shall be used. Slip-on flanges shall not be used.
- Seal-welded threaded connections are not acceptable.

Protection of Well Pipe

Although the gas injected into the well will be dry, the well bore in general and the producing zone in particular will probably contain excessive quantities of free water which, when exposed to CO₂, will produce a corrosive atmosphere. In the absence of a packer, the casing-tubing annulus is exposed to both the wet reservoir gases and the CO₂. As the gas diffuses up the hole, the water condenses on the surface of the casing at areas such as the cool shallow water zone which are cooled below the dewpoint. CO₂ dissolves in the water droplets and corrosion results. The higher operating temperature of the tubing should prevent corrosion to the outside of the tubing during CO₂ injection; however, if the well is used alternately for injecting CO₂ and water, a corrosive environment could result inside of the tubing during periods of changeover and to the outside of the tubing during the cooler operation when injecting water.

- In the absence of a packer, the use of a corrosion inhibitor should be considered. As effective treatment will vary with field and well, the major local inhibitor suppliers should be contacted for an effective inhibitor and treatment.
- If corrosion is expected inside and/or outside of the tubing, consideration should be given to the application of a baked epoxy phenolic such as Kordell 600 or equal. This material is applied 1 mil at a time and baked before applying the next coat. Six mils should provide a spark-free lining capable of providing the required corrosion resistance; however, this type of coating is subject to pinholes, so careful inspection is required.
- If the alternate injection of water is anticipated, the tubing head should include a secondary seal, and valves and fittings of the xmas tree should be internally coated with a baked epoxy phenolic.

Nonmetallic Seals²⁴

In both wet and dry service, CO₂ under pressure presents a problem due to the solubility and diffusivity of the gas in elastomers. A good nonmetallic seal should have the following properties:

- Resistance to inflation with prolonged exposure to the high-pressure gas.
- Ability to withstand rapid decompression without blistering or tearing.
- Ability to release rapidly any absorbed gas and restore quickly to original dimensions.

Teflon, nylon, EPR, polyimide, and semirigid polyurethane meet these criteria for both wet and dry CO₂ service. Although some varieties of Buna-N, Hypalon, Viton, and polyurethane elastomers will swell excessively and blister, they may be used as captive seals; i.e., enclosed seals that are not required to flex during service.

Use teflon tape for threads. Ordinary thread dope is washed away by CO₂.

RESERVOIRS

There are 45 operating oil fields recognized by the Conservation Committee of California Oil Producers and the Classification Committee of the American Association of Petroleum Geologists (AAPG) in the Los Angeles Basin. The Conservation Committee and the AAPG have further classified the 45 oil fields into 244 producing pools, and the monthly oil production figures are reported by pool.

The Los Angeles Basin fields range from the largest (Wilmington) with 2,200 active wells producing 130,000 b/d from 69 pools, to the one-well, 5 b/d Alondra Field. Seventeen of the fields are rated as important based upon an arbitrary yearly production of more than one million bbl. of oil and/or remaining reserves of over 10 million bbl. All 17 of them are being produced by secondary recovery methods; 15 by water flood, one by in situ combustion, and one by steaming. Overall water cut for the group is now more than 89 percent.

Figure 1 shows all the Los Angeles Basin oil fields and indicates the proximity of the sources of CO₂ to many of the important oil fields. Table 2 lists the major oil fields, their 1977 production, and their remaining reserves.

There are 20 oil companies with a significant interest in one or more of the 17 major Los Angeles Basin oil fields. We designed a reservoir questionnaire (a copy of which is included in the Appendix) and sent a total of 41 questionnaires to cognizant personnel at all 20 oil companies. It is difficult in a survey of this type to get a massive response, since many operators are aware of the mandatory questionnaires and the information they must supply to DOE and other government agencies. However, we have used the data from the 18 questionnaires which were returned to us, supplemented by Society of Petroleum Engineers publications, California Division of Oil and Gas records and publications, reports from the Conservation Committee of California Oil Producers, well logs and conversations with several oil-company reservoir engineers, as input to our reservoir assessment.

CO₂ REQUIREMENTS—RESERVOIR EXPERIENCE IN OTHER AREAS

Field experience with the use of CO₂ to recover additional oil is now becoming available.

The Little Creek Field pilot test in Mississippi (Shell Oil Co., operator) has been reported extensively in the "Improved Oil Recovery Field Reports."⁴³ It is a pilot test in a completely watered out reservoir, a genuine tertiary operation. The residual oil is immobile. Over 100,000 bbls. of crude were produced at a ratio of injected CO₂ to produce crude oil of 27 Mscf/bbl. Even allowing for some escape of CO₂ from the pilot, which in fact was probably offset by production of some crude from beyond the confines of the pilot area, the minimum ratio that could be assigned would be 20 Mscf/bbl.

The only other completed tertiary pilot operation is that operated by Chevron in the watered out area of the Canyon Reef reservoir in the Sacroc Unit (Kelly-Snyder Field) The reported ratio is 35.9 Mscf/bbl.

Several secondary demonstration full-scale CO₂ injection operations have been undertaken. Of greatest significance are those at Crossett (North Cross Unit, Texas, operated by Shell) and the Sacroc Center Line Flood Area Project, again operated by Chevron. Both of these fields had mobile oil when CO₂ injection was initiated. At Crossett, the CO₂ project is the secondary recovery scheme. At Sacroc, water flooding was interrupted prior to the field becoming watered out, and secondary operations were continued by a combination of CO₂ and water injection.

North Cross is a very low permeability reservoir. The carbon dioxide/produced oil ratio at this time is 21 Mscf/bbl. The operator anticipated a ratio of 7 Mscf at the completion of the flood. In order to achieve this, the operator will have to produce an additional (over primary) 8.7 million barrels of oil upon the injection of the scheduled additional quantity of 36 billion cubic feet. Achieving a ratio of 4.1 Mscf in the future when the ratio has been 21 in the first half of the project is questionable.

The Sacroc secondary operation is the most reknown of all the CO₂ operations undertaken to date. It was not conducted as an experiment, but as a real attempt to increase recovery rate. Unfortunately, it is very difficult to interpret.

To understand the nature of the results achieved at Sacroc it is necessary to review the history of the field. Water injection was started in 1954 at approximately 90,000 bbls./day. Starting late in 1965, the water injection rate was gradually increased, reaching a value of 150,000 b/d by mid 1970 at which time the oil rate has increased to 120,000 b/d from 40,000 b/d in 1965. Between 1970 and 1972, the water injection rate was dramatically increased to 380,000 b/d with oil production climbing to over 180,000 b/d by 1972 when CO₂ injection was initiated into the Stage I area. The oil rate peaked at only 215,000 b/d in late 1973 even though CO₂ injection was continuing and the water rate was further increased to some 600,000 b/d.

After CO₂ injection was initiated, a number of infill wells were completed in the unit (complementary injectors) and numerous recompletions were undertaken in order to isolate zones of high fluid conductivity.

A. E. Kane, in his paper (SPE 7091, Tulsa, Oklahoma, 1978)⁴⁵ had to estimate what the recovery would have been under continued high pressure water injection, and what it will be under CO₂ injection in order to estimate the additional oil production due to CO₂ injection. It is not clearly stated how these estimates were made; the reader is left to assume that it is by simulation and analogy. Regardless of the basis of the estimates, the estimates must be considered to be tenuous, at best, and lead to an estimated ratio of 6.1 Mscf/bbl. of additional oil. An independent estimate, doing the best that can be done with the decline curves, leads to calculated ratio of 12 Mscf/bbl.

Again, it must be emphasized that the water flood had not been completed when the CO₂ injection was initiated. The oil was still mobile. In fact, the recovery at the start of 1972 amounted to only 18 percent of the original oil in place whereas at the end of 1977 it had climbed to 44 percent and still must get to 56 percent to reach the estimated ultimate value.

We estimate that it will require no less than 12 Mscf to recover an additional barrel of oil and may require as much as 20 Mscf/bbl. The upper limit is less than observed in the only reported tertiary projects, and the lower limit is the same order of magnitude that is estimated from the giant secondary operation at Sacroc.

The fact that these ratios are higher than those earlier estimated is simply due to the fact that the early proponents, who claimed theoretical displacement volumetrics would hold, ignored reservoir heterogeneity and the heterogeneous distribution of residual oil, as well as the instability of the process due to unfavorable viscous and gravity forces. Subsequent estimates of 5 to 10 Mscf did take into account some of these factors, but apparently missed on estimating their severity, although the highest ratios probably may be encountered in the best operations. In actual fact, further understanding of the process may lead to improved mobility control and recompletions that will permit creaming of the applicable reservoirs so that reasonably low ratios will be obtained at the expense of higher recovery.

GRADING RESERVOIRS FOR AMENABILITY TO CARBON DIOXIDE FLOODING

The recovery of crude oil by CO₂ flooding is a fluid-fluid displacement process. Just as in conventional oil production operations, the efficiency of the process will be controlled by conventional factors such as mobility ratio, oil and water saturation in the reservoir, and reservoir heterogeneity.

Carbon dioxide may be miscible with the crude oil, or by interaction with the crude may develop solutions which are miscible with the reservoir crude. It has even been assumed that the recovery process is aided by volatilization of the crude oil into a dense gaseous phase of carbon dioxide. No matter which of these phenomena occur, the aforementioned mobility ratio, oil, and water saturation, and reservoir homogeneity are no less important in controlling the displacement efficiency.

If miscibility is achieved, the CO₂ reduces the viscosity and increases the volume of the crude. This, in turn, increases the oil saturation in the sand which is a plus for displacement efficiency. Miscibility also results in the interfacial tension vanishing between the displaced and displacing phases, and this is a plus since it removes the special resistance to crude-oil flow out of a capillary bottleneck. (However, it must be born in mind that the velocity of fluid flow through this bottleneck will still be slower than that through a wider pore.) Miscibility is known to be aided by high pressures and low temperatures.

Without having at our disposal any hard, factual data on miscibility in systems of California crude oils and carbon dioxide, we made projections based on reported information concerning other crude oils. In the absence of any extensive field studies on carbon dioxide for the recovery of crude oil, specifically in connection with volumetric conformance, we estimated relative sweep efficiency in California reservoirs based on our knowledge of California reservoirs' response to water flooding. We did seek out the assistance of various operators of oil fields in the Los Angeles Basin, and used their input whenever it was available.

The April 1976 Lewin Report²² includes a "Screening Guide - Criteria for the Application of Selected Enhanced Oil Recovery Methods," which is a compendium of expert opinion agreeing generally with the body of published literature. The screening guide criteria for CO₂ injection (for California Codes) advises:

- Viscosity in reservoir should be less than 12 cp
- Gravity higher than 26°API
- Residual oil saturation, 25 percent PV or more
- Depth greater than 3,000 ft.
- Original reservoir pressure greater than 1,500 psi

Following are the major criteria we used in classifying Los Angeles Basin reservoirs in terms of their amenability to carbon dioxide flooding.

1. The oil saturation in the reservoir. The higher it is at the time of initiating a carbon dioxide flood, the better will be the anticipated performance. At smaller degrees of depletion it is fair to anticipate that it will be less likely to encounter strata of high water or gas saturation, which may act as thief zones for the injected carbon dioxide. Further, the higher the oil saturation, the more likely it will be that a bank of mobile, swollen crude oil can be developed ahead of the displacing front. Such an increase in saturation and mobility will of course lower the mobility of other phases.
2. The depth of the reservoir. We estimated that a pressure of no less than 1,500 psi, and probably 2,000-2,500 psi, would be required to develop miscibility of carbon dioxide with the naphthenic, currently low gas-bearing crudes in the Los Angeles Basin. Thus, a depth of at least 2,000 ft., and probably 3,000 ft. would be required for a good prospect. Only minor concern was paid to whether the reservoir had been maintained at such pressure, since it is believed that carbon dioxide would have to displace the gas or water that is now responsible for the pressure, to secure good recovery.
3. Potential for Zone Control. Zone control, or profile control of injected fluids is a congenital problem in California reservoirs. Long intervals, high energy depositional environments, sloughing shales, significant interstitial clays, and faulting make profile control a difficult operation. Therefore, the lithology of the subject reservoirs was scanned to see if distinct, somewhat uniform intervals could be delineated for carbon dioxide injection. It should also be noted that thick intervals which are very deep (and are therefore drilled up on wide spacing) are not desirable because of the potential for gravity segregation of the injected carbon dioxide.

The results of our scanning of Los Angeles Basin reservoirs led to the development of three categories: 1) Prime class, 2) Possible, but information on the reservoir limited, or behavior of the crude with carbon dioxide difficult to project because of gravity, and 3) Rejected because of low oil saturation, shallow depth, apparent fracture system or heterogeneity of sand development or very viscous crude. It is obvious that we do not believe this classification, at arm's length, is cast in concrete.

Our rating of selected reservoirs follows:

<u>Field</u>	<u>Zone</u>	<u>Remarks</u>
<u>1. Prime</u>		
Torrance	Main	(Upper Terminal equivalent)
Wilmington	Upper Terminal	
Belmont	Surfside	
Dominguez	various	
West Coyote	Main and 99	
Seal Beach	Wasem and McGrath	
Huntington Beach	Main	
Richfield	Kraemer	
<u>2. Possible</u>		
Beverly Hills	various	city location
Santa Fe Springs	various	oil saturation unknown
Brea Olinda	various	very low pressure
Inglewood	Vickers	need more information
East Coyote	various	need more information
<u>3. Rejected</u>		
Huntington Beach	Jones	too viscous
Inglewood	Rindge and Rubel	water-flood problems
Las Cienegas	various	water-flood problems
Long Beach	various	water-flood problems
West Newport	Main area	too viscous, successful in situ combustion
Yorba Linda	Shallow	too viscous, successful in steaming

CONCLUSIONS AND RECOMMENDATIONS

SOURCES

There are two major sources of CO₂ in the Los Angeles Basin. Six oil refineries and one ammonia plant (all near Los Angeles Basin oil fields), have hydrogen plants with by-product streams of concentrated CO₂. The total available (uncommitted) CO₂ from these streams is about 3,000 tons per day.

Six major electric power plants, all near L.A. Basin oil fields, discharge a combined total of 70,000 tons per day of CO₂ from 27 large boilers. Average CO₂ concentration in the flue gas is about 14 percent on a dry basis.

PROCESSING

CO₂ purification methods have been selected. We recommend three alternates for purifying the hydrogen plant off-gases, i.e. separating the CO₂ from the nitrogen.

- Modify the existing hydrogen-CO₂ separation system, so that nitrogen is not used for stripping and therefore does not need to be removed.
- Use propylene carbonate absorption.
- Use low-temperature separation.

For CO₂ extraction from flue gas, we recommend monoethanolamine (MEA) absorption.

All the CO₂ source streams are wet and must be dehydrated. We would use triethylene glycol for drying the gas.

TRANSPORTATION

Alternate transportation systems have been compared and evaluated herein. A pipeline is the most practical and the most economical method for moving CO₂ from source to oil field. If an existing pipeline is available, (and there appear to be unused lines along some of the projected routes), considerable capital expenditure can be saved.

MATERIALS

There are no real difficulties in selecting materials for handling CO₂. Plain carbon steel can be used where the CO₂ is dry. In wet service 304 stainless steel can be used, or plain carbon steel can be protected with a properly selected coating or a suitable corrosion inhibitor. These techniques are well known in industry, and are covered in more detail by the material specifications section in the Appendix. The specifications also list a number of elastomers which are suitable for sealing against CO₂ under high pressure.

RESERVOIRS

Several reservoirs have been identified and are listed as prime candidates for CO₂ injection, using the major criteria of high oil saturation in the reservoir, suitable depth of the reservoir, and a good potential for zone control.

A typical California reservoir might have started with 1,500-1,600 bbl/acre-ft. of original oil in place. CO₂ injection will recover an estimated 6 to 7 percent additional oil over and above what might be expected from water-flood. That means that CO₂ injection could be responsible for recovering an additional 100 bbl/acre-ft.

We estimate that from 12 to 20 Mscf of CO₂ will be required per barrel of recovered oil. Under optimum conditions, it is possible to be at the lower end of the range.

CO₂ from the Chevron refinery, dried, purified (separated from the nitrogen), moved to the nearby Torrance oil field by the most economical transportation system (existing pipelines), and compressed to 1,500 psi will cost \$0.99/Mscf. CO₂ from a power plant flue gas (also purified, dried, compressed to 1,500 psi, and delivered to a nearby oil field) is estimated to cost \$1.53/Mscf.

The costs of CO₂ from the two sources are compared below:

COSTS OF CO₂

<u>SOURCE</u>	<u>PER MSCF</u>	<u>PER BBL OF OIL RECOVERED</u>
Refinery	\$0.99	\$12 to 20
Power plant flue gas	\$1.53	\$18 to 31

It is obvious that the engineers contemplating a CO₂ flood must do everything possible to assure that the costs will be at the low end of the range. Even after they decide that a given reservoir is a prime candidate for CO₂ flooding, there will still be at least 3 years of

work, miscibility and solubility studies, reservoir simulation, design work, material and equipment selection, and plans for corrosion mitigation, etc. before the first CO₂ is injected.

To all that work which is certainly necessary, we would also add:

- Plans should be made for handling the mixture of CO₂ and hydrocarbon gas which is produced with the oil. As this mixture becomes richer in CO₂ the economics may be good to separate the CO₂ and reinject it, or depending on the composition, reinject it, as is. The cost of CO₂ is a large fraction of the cost of the project, and this may be a cheaper "source" of more CO₂.
- Consider injecting the CO₂ - N₂ mixtures available from the Chevron and Texaco refineries. Leaving out the "purification" step reduces the cost of the delivered CO₂ by almost one third.
- Consider Huff-Puff. One Los Angeles Basin operator has injected CO₂ into a 13°API tar zone at about 600 psi, and produced oil back by Huff-Puff, rather than by CO₂ drive. They have obtained "very favorable results", but as yet the economics are unknown.
- Consider combinations of the above.

APPENDICES

MATERIAL SPECIFICATION FOR PIPING AND ASSOCIATED
EQUIPMENT EXPOSED TO CO₂

1.0 CARBON DIOXIDE-DRY

CO₂ with a water content less than the saturation value at operating conditions.

1.1 Metals

Under dry conditions CO₂ is noncorrosive to iron-base alloys and the design of handling and distribution systems can be based upon the use of plain carbon steels according to the following codes as they pertain.

- 1.1.1 ASME, Boiler and Pressure Vessel Code, Section 8.
- 1.1.2 ANSI B31.3, Petroleum Refinery Piping.
- 1.1.3 ANSI B31.8, Gas Transmission and Distribution Piping Systems.

1.2 Nonmetallic Seals²⁴

Although dry CO₂ does not present a corrosion problem, CO₂ under pressure presents a problem due to the solubility and diffusivity of the gas in elastomers. A good nonmetallic seal should have the following properties:

- 1.2.1 Resistance to inflation with prolonged exposure to the high-pressure gas.
- 1.2.2 Ability to withstand rapid decompression without blistering or tearing.
- 1.2.3 Ability to release rapidly any absorbed gas and restore quickly to original dimensions.

Teflon, nylon, EPR, polyimide, and semirigid polyurethane meet these criteria. Although some varieties of Buna-N, Hypalon, Viton, and polyurethane elastomers will swell excessively and blister, they may be used as captive seals; i.e., enclosed seals that are not required to flex during service.

1.0 CARBON DIOXIDE-DRY (Continued)

1.3 Inhibitors and/or Coatings

Although the gas injected into the well will be dry, the well bore in general and the producing zone in particular will probably contain excessive quantities of free water which, when exposed to CO₂, will produce a corrosive atmosphere. In the absence of a packer, the casing-tubing annulus is exposed to both the wet reservoir gases and the CO₂. As the gas diffuses up the hole, the water condenses on the surface of the casing at areas such as the cool shallow water zone which are cooled below the dewpoint. CO₂ dissolves in the water droplets and corrosion results. The higher operating temperature of the tubing should prevent corrosion to the outside of the tubing during CO₂ injection; however, if the well is used alternately for injecting CO₂ and water a corrosive environment could result inside of the tubing during periods of changeover and to the outside of the tubing during the cooler operation when injecting water.

1.3.1 In the absence of a packer, the use of a corrosion inhibitor should be considered. As effective treatment will vary with field and well, the major local inhibitor suppliers should be contacted for an effective inhibitor and treatment.

1.3.2 If corrosion is expected inside and/or outside of the tubing, consideration should be given to the application of a baked epoxy phenolic such as Kordell 600 or equal. This material is applied one mil at a time and baked before applying the next coat. Six mils should provide a spark-free lining capable of providing the required corrosion resistance; however, this type of coating is subject to pinholes, so careful inspection is required.

1.3.3 If the alternate injection of water is anticipated, the tubing head should include a secondary seal, and valves and fittings of the xmas tree should be internally coated with a baked epoxy phenolic.

2.0 CARBON DIOXIDE-WET

CO₂ with a water content in excess of the saturation value at operating conditions.

With the exception of well conditions referred to in Subsection 1.3 above, the only exposure of wet CO₂ is in the dehydration facilities and gathering lines feeding this equipment.

2.0 CARBON DIOXIDE-WET (Continued)

2.1 Metals

When free water is present, CO₂ dissolves and forms carbonic acid (H₂CO₃) causing a reduction in pH of the water, making it quite corrosive to steel.

2.1.1 All vessels exposed to wet CO₂ shall be designed and constructed according to ASME Section 8 of materials as follows:

- Annealed 304L stainless steel is the most economical of the stainless steels, but 316L and Carpenter 20 Cb-3 may also be considered.
- Carbon steel internally coated with a minimum of 12 mils of a baked epoxy phenolic may be considered for large vessels, but close attention must be given to the preparation of the metal and the application of the coatings to be sure the final coating is free of pinholes. Internals should be made of stainless steel.

2.1.2 Piping, valves, and fittings should be according to Material Specification 1143A-19-C1A and C3A attached.

2.2 Nonmetallic Seals

Teflon, nylon, EPR, polyimide, and semi-rigid polyurethane should give satisfactory service in wet service as well as dry.

MATERIAL SPECIFICATION

PIPING

1143A-19-C1A

1.0 SERVICE

- 1.1 Wet CO₂ service systems within the pressure-temperature limits of 150-lb ANSI flange ratings, designated CIA on pressure temperature chart.

2.0 GENERAL

- 2.1 All piping, fittings, and valves shall comply with the requirements of ANSI B32.3, code for petroleum refinery piping, latest edition and addendum, as supplemented or replaced by the requirements of this specification.
- 2.2 Herein, "or equal" shall mean operator-approved equal.

3.0 MATERIALS

3.1 General

- 3.1.1 No carbon steel shall be used for piping, fittings, and fabrications directly exposed to wet CO₂.
- 3.1.2 Free machining grade stainless steel shall not be used.
- 3.1.3 Austenitic stainless steel stabilized with columbium or extra-low-carbon stainless steel shall be used to avoid intergranular corrosion.
- 3.1.4 Dissimilar metals shall not be used unless they can be selected and arranged to minimize electrolytic corrosion.

3.2 Bolting

- 3.2.1 150-lb flange bolting shall be ASTM A193, B7.

3.0 MATERIALS (Continued)

3.2.2 Nuts shall be ASTM A194 EH heavy hex series

3.2.3 All bolting shall be aluminized by Bethalume process or equal. Zinc and cadmium plating shall not be used.

3.3 Pipe Flanges and Fittings

3.3.1 ASTM A312, Grade TP304L stainless steel pipe and ASTM A403, Grade WP304L stainless steel fittings will be acceptable. Weld neck flanges, ASTM A182, Grade F304L will be used.

3.3.2 If valves or fittings are not readily available in 304L stainless steel, any equal stainless steel with 12 percent chromium or higher may be used.

3.3.3 Corrosion allowance shall be 1/16 inch.

3.3.4 Seal or back-welded threaded connections are not acceptable substitutes for socket weld construction.

3.3.5 Slip-on flanges shall not be used.

3.3.6 All asbestos gaskets shall be soaked with sodium silicate (water glass) solution when making up to reduce the possibility of leaks.

3.4 Valves

3.4.1 Valves shall be manufacturers standard pattern with sealant recommended by the manufacturer for exposure to wet CO₂.

3.4.2 Diaphragm-type actuators are acceptable.

3.4.3 Seals will be made from teflon, nylon, EPR, polyimide and semi-rigid polyurethane.

3.4.4 Seals made from Viton, Hypalon and Buna-N rubbers will not be used unless they are captive (enclosed) seals. Captive seals will require replacement each time they are decompressed.

3.5 Pressure Gauges

3.5.1 Pressure gauges shall have a type 304L or equal stainless steel Bourdon tube, socket, and tip. Monel shall not be used. Safety cases shall be provided.

4.0 ASSEMBLY METHOD

1½ in. and smaller	Threaded or socket weld
2 in. and larger	Welded, flanged 150-lb RF ANSI

5.0 PIPE WALL THICKNESS

2 in. and smaller	Schedule 80S
3 in. to 10 in. inclusive	Schedule 40S
12 in. to 14 in. inclusive	3/8 in.
16 in. and larger	Calculated wall thickness per ANSI B31.3

6.0 FITTINGS

6.1 Unions and Flanges

1½ in. and smaller	3,000-lb ground joint union with metal-to-metal seat
2 in. and larger	Raised face weld-neck flange with bore to match pipe I.D.

6.2 Closures

1½ in. and smaller	Bull plug
2 in. and larger	Raised-face blind flange

6.3 Reducers

1½ in. and smaller	Swaged nipple
2 in. and larger	Reducer

6.4 Turns

All Sizes	Long-radius elbow. Short-radius elbow permitted only where space is limited.
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6.5	Branch Lines	
	Full Size	TEE
	1½ in. and smaller	Threadolet, full or half coupling, or Schedule 80S TEE
	2 in. and larger	Reducing TEE or standard weight-reducing weldolet
7.0	GASKETS AND THREAD MAKEUP	
	All flanges	1/16-in. compressed asbestos JM 60 or equal
	Threads	Teflon tape
8.0	VALVES	
	8.1 Gate	
	1½ in. and smaller	800-lb FSS OS&Y bolted bonnet, screwed Vogt 12401 or equal. Socket weld Vogt SW-12401 or equal
	2 in. and larger	150-lb ACS flanged. OS&Y bolted bonnet, Crane 21776-API and 21776-AL-API, or equal
	8.2 Plug	
	½ in. through 1½ in.	300-lb ACS screwed, wrench-operated, Rockwell Fig. 1714MA-4 with No. 421 sealant or equal
	2 in. through 4 in.	150-lb ACS flanged, wrench-operated Rockwell Fig. 1625 MA-4 with No. 421 sealant or equal
	8.3 Globe	
	1½ in. and smaller	800-lb FSS OS&Y bolted bonnet, screwed Vogt 12501 or equal. Socket welded Vogt SW12501 or equal.

	2 in. through 6 in.	150-lb ACS flanged, OS&Y, bolted bonnet, Crane 21376-HB and 21376-HC or equal
8.4	Needle	
	1/8 in. through 1 in.	3,000-lb stainless steel, Crane 2224 or equal
8.5	Check	
	1½ in. and smaller	300-lb ACS, bolted cap, swing check screwed - Ladish CF3 No. 5370 or equal
	2 in. through 12 in.	150-lb ACS, flanged, swing check, bolted cap. Ladish CF3 No. 5272 and No. 5373, or equal

MATERIAL SPECIFICATION

PIPING

1143A-19-C3A

1.0 SERVICE

Wet CO₂ service systems within the pressure-temperature limits of 300-lb ANSI flange ratings, designated C3A on pressure temperature chart.

2.0 GENERAL

2.1 All piping, fittings, and valves shall comply with the requirements of ANSI B31.3, code for petroleum refinery piping, latest edition and addendum, as supplemented or replaced by the requirements of this specification.

2.2 Herein, "or equal" shall mean operator approved equal.

3.0 MATERIALS

3.1 General

3.1.1 No carbon steel shall be used for piping, fittings and fabrications directly exposed to wet CO₂.

3.1.2 Free machining grade stainless steel shall not be used.

3.1.3 Austenitic stainless steel stabilized with columbium or "extra-low-carbon" stainless steel shall be used to avoid intergranular corrosion.

3.1.4 Dissimilar metals shall not be used unless they can be selected and arranged to minimize electrolytic corrosion.

3.2 Bolting

3.2.1 All flange bolts, valve bonnet bolts, and pressure equipment closure bolts in piping beyond the pressure and temperature limits of C1A (see chart) shall conform to the following modification of ASTM A193 B7. Stock shall be drawn at

3.0 MATERIALS (Continued)

1,275^oF to 1,325^oF to produce a hardness of Rockwell C-21 or less, a minimum yield strength of 80,000 psi, and a minimum tensile strength of 100,000 psi. Written certification based on a minimum 10 percent random testing shall be required. Bolts shall be stenciled with "B-7M" designation.

3.2.2 Nuts shall be ASTM A194 2H heavy hex series.

3.2.3 All bolting shall be aluminized by Bethalume process or equal. Zinc and cadmium plating shall not be used.

3.3 Pipe Flanges and Fittings

3.3.1 ASTM A312, Grade TP304L stainless-steel pipe and ASTM A403, Grade WP304L stainless-steel fittings will be acceptable. Weld neck flanges, ASTM A182 Grade F 304L will be used.

3.3.2 If valves or fittings are not readily available in 304L stainless steel, any equal stainless steel with 12 percent chromium or higher may be used.

3.3.3 Corrosion allowance shall be 1/16 in.

3.3.4 Seal or back-welded threaded connections are not acceptable substitutes for socketweld construction.

3.3.5 Slip-on flanges shall not be used.

3.3.6 All asbestos gaskets shall be soaked in sodium silicate (water glass) solution when making up to reduce the possibility of leaks.

3.4 Valves

3.4.1 Valves shall be manufacturers standard pattern with sealant recommended by manufacturer for exposure to wet CO₂.

3.4.2 Diaphragm-type actuators are acceptable.

3.4.3 Seals will be made from teflon, nylon, EPR, polyimide and semi-rigid polyurethane.

3.4.4 Seals made from Viton, Hypalon, and Buna-N rubbers will not be used unless they are captive (enclosed) seals. Captive seals will require replacement each time they are decompressed.

3.0 MATERIALS (Continued)

3.5 Pressure Gauges

Pressure gauges shall have a type 304L or equal stainless steel Bordon tube, socket, and tip. Monel shall not be used. Safety cases shall be provided.

4.0 ASSEMBLY METHOD

1½ in. and smaller

Threaded or socketweld

2 in. and larger

Welded, flanged 300-lb RF ANSI

5.0 PIPE WALL THICKNESS

2 in. and smaller

Schedule 80S

3 to 10 in. inclusive

Schedule 40S

12 to 14 in. inclusive

3/8 in.

16 in. and larger

Calculated wall thickness per ANSI B31.3

6.0 FITTINGS

6.1 Unions and Flanges

1½ in. and smaller

3,000-lb ground joint union with metal-to-metal seat

2 in. and larger

Raised face welding neck flange with bore to match pipe I.D.

6.2 Closures

1½ in. and smaller

Bull plug

2 in. and larger

Raised face blind flange

6.3 Reducers

1½ in. and smaller

Swaged nipple

2 in. and larger

Reducer

6.0 FITTINGS (Continued)

6.4 Turns

All sizes

Long-radius elbow. Short-radius elbow permitted only where space is limited

6.5 Branch Lines

Full size

TEE

1½ in. and smaller

Threadolet, full or half coupling, or Schedule 80S TEE

2 in. and larger

Reducing TEE or standard weight-reducing weldolet

7.0 GASKETS AND THREAD MAKEUP

Flanges 350°F and lower

1/16-in. compressed asbestos, JM60 or equal

Flanges above 350°F

Spiral-wound metal with asbestos filling. Flexitalic Style CG or equal

Threads

Teflon tape

8.0 VALVES

8.1 Gate

1½ in. and smaller

800-lb FSS OS&Y bolted bonnet, screwed Vogt 12401. Socket weld Vogt SW-12401

2 in. and larger

300-lb ACS flanged. OS&Y bolted bonnet, Crane 23776-API and 23176AM-API

8.2 Plug

½ through 1½ in.

300-lb ACS screwed, wrench operated Rockwell Fig. 1714 MA-4 with No. 421 sealant or equal

8.0 VALVES (Continued)

	2 through 4 in.	300-lb ACS flanged, wrench-operated Rockwell Fig. 1725 MA-4 with No. 421 sealant or equal
8.3	Globe	
	1½ in. and smaller	800-lb FSS OS&Y bolted bonnet, screwed Vogt 12501. Socket weld Vogt SW-12501
	2 through 10 in.	300-lb ACS OS&Y bolted bonnet flanged Ladish CF-3 No. 7372 or equal
8.4	Needle	
	1/8 through 1 in.	6,000-lb stainless steel Ladish CF-3 No. 2000 or equal
8.5	Check	
	1½ in. and smaller	600-lb ACS, bolted cap, swing check, screwed. Ladish CF-3 No. 5670 or equal
	2 through 12 in.	300-lb ACS, bolted cap, swing check, flanged, Ladish CF-3 No. 5372 and No. 5375 or equal

LAWRENCE-ALLISON & ASSOCIATES CORPORATION, WESTERN DIVISION
 QUESTIONNAIRE TO ELICIT DATA ON LOS ANGELES BASIN OIL FIELDS
 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

	BEVERLY HILLS		
1. Field			
2. Reservoir	A. <u>Rancho</u>	B. <u>Main Zone</u>	C. _____
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Primary</u>	B. <u>Primary</u>	C. _____
4. Consolidation (Good, Fair, Unconsolidated)	A. <u>Fair</u>	B. <u>Fair</u>	C. _____
5. Is there significant carbonate cement? (Yes/No)	A. <u>No</u>	B. <u>No</u>	C. _____
6. Subsurface Depth, ft	A. <u>9000' ss</u>	B. <u>7200'ss</u>	C. _____
7. Gross Section, ft	A. <u>400'</u>	B. <u>300-400'</u>	C. _____
8. Net, ft	A. <u>200</u>	B. <u>150</u>	C. _____
9. Is the net uniformly or erratically distributed?	A. <u>Uniformly</u>	B. <u>Erratic</u>	C. _____
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>10 md [±]</u>	B. <u>70 md [±]</u>	C. _____
11. Permeability, vertical, mds	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
12. Dykstra Coefficient of Permeability Distribution	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
13. Porosity, at atmospheric %	A. <u>16</u>	B. <u>21</u>	C. _____
14. Porosity, restored state, %	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
15. Current Reservoir Pressure, psi	A. <u>Approx. 1100 psi</u>	B. <u>Approx. 800 psi</u>	C. _____
16. Reservoir Temperature °F	A. <u>210°F</u>	B. <u>N.A.</u>	C. _____
17. Acres	A. <u>200</u>	B. <u>150</u>	C. _____
18. Acres/Producing Well	A. <u>—</u>	B. <u>—</u>	C. _____
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>—</u>	B. <u>—</u>	C. _____

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 QUESTIONNAIRE TO ELICIT DATA ON LOS ANGELES BASIN OIL FIELDS
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	SEAL BEACH		
1. Field	-----		
2. Reservoir	A. <u>Wasem</u>	B. <u>South McGrath</u>	C. <u>North McGrath</u>
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Waterflood & Primary</u>	B. <u>Waterflood & Primary</u>	C. <u>Waterflood</u>
4. Consolidation (Good, Fair, Unconsolidated)	A. <u>Good</u>	B. <u>Good</u>	C. <u>Good</u>
5. Is there significant carbonate cement? (Yes/No)	A. <u>Yes</u>	B. <u>Yes</u>	C. <u>Yes</u>
6. Subsurface Depth, ft	A. <u>5075'</u>	B. <u>5800'</u>	C. <u>6600'</u>
7. Gross Section, ft	A. <u>1200'</u>	B. <u>1870'</u>	C. <u>2019'</u>
8. Net, ft	A. <u>410'</u>	B. <u>649'</u>	C. <u>1207'</u>
9. Is the net uniformly or erratically distributed?	A. <u>Erratically</u>	B. <u>Erratically</u>	C. <u>Erratically</u>
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>90</u>	B. <u>10.2</u>	C. <u>70.7</u>
11. Permeability, vertical, mds	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
12. Dykstra Coefficient of Permeability Distribution	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
13. Porosity, at atmospheric %	A. <u>21.8</u>	B. <u>20.0</u>	C. <u>21.4</u>
14. Porosity, restored state, %	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
15. Current Reservoir Pressure, psi	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
16. Reservoir Temperature °F	A. <u>160°F</u>	B. <u>230°F</u>	C. <u>230°F</u>
17. Acres	A. <u>112</u>	B. <u>90</u>	C. <u>46</u>
18. Acres/Producing Well	A. <u>3.6</u>	B. <u>3.8</u>	C. <u>4.2</u>
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>5 Spot Pilot Lower Wasem</u>	B. <u>5 Spot Pilot</u>	C. <u>Combination</u>

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 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

1. Field	SANTA FE SPRINGS		TORRANCE
	A. Meyer Zone	B. Clark-Hathaway Zone	C. Del Amo Zone
2. Reservoir	A. Meyer Zone	B. Clark-Hathaway Zone	C. Del Amo Zone
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. WF-Pilot	B. WF-Pilot	C. Primary
4. Consolidation (Good, Fair, Unconsolidated)	A. Consolid.	B. Consolid.	C. Good to Poor
5. Is there significant carbonate cement? (Yes/No)	A. No	B. No	C. No Data
6. Subsurface Depth, ft	A. 4220	B. 7210	C. 4200-4500
7. Gross Section, ft	A. 750	B. 1000	C. 350 (Avg.)
8. Net, ft	A. 105 Avg. (Floodable)	B. 266 Avg. (Floodable)	C. 100 Avg.
9. Is the net uniformly or erratically distributed?	A. Unif. 20' to 350'	B. Unif. 20' to 500'	C. Uniformly
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A. 720	B. 130	C. 43 (Median)
11. Permeability, vertical, mds	A. N.A.	B. N.A.	C. No Data
12. Dykstra Coefficient of Permeability Distribution	A. 0.5	B. 0.6	C. 0.73
13. Porosity, at atmospheric %	A. 32.3	B. 21.2	C. 27.5 (Avg.)
14. Porosity, restored state, %	A. N.A.	B. N.A.	C. No Data
15. Current Reservoir Pressure, psi	A. 200 Est.	B. 400 Est.	C. 1675
16. Reservoir Temperature °F	A. 150	B. 210	C. 180 375 (Potential Unit Area-Not Now Unitized)
17. Acres	A. 916 (Floodable) 16 (Current Prod. Ratio)	B. 351 (Floodable) 10 (Current Prod. Ratio)	C. 375 (Potential Unit Area-Not Now Unitized)
18. Acres/Producing Well	A. 16 (Current Prod. Ratio)	B. 10 (Current Prod. Ratio)	C. 23.4
19. Water Flood Configuration (5-spot, 7-spot, line drive)	A. 5 Spot & Periphery	B. One 5 Spot	C. —

20. Current Oil in Place, Bbls/Net Acre Foot
A. N.A. B. N.A. C. 1139
21. Current Oil Saturation pore volume %
A. N.A. B. N.A. C. 56
22. How was above estimated? Material Balance, Recent Coring or Logging)
A. Material Balance B. Material Balance C. Material Balance
23. Current Formation Volume Factor
A. 1.00 B. 1.06 C. 1.14
24. Current Solution Gas, SCF/Bbl
A. 50 B. 150 C. 136
25. Current Producing GOR, SCF/Bbl
A. N.A. B. N.A. C. 136
26. Current Produced Gas Density
A. N.A. B. N.A. C. No Data
27. Current Reservoir Oil Viscosity, cp
A. 2.7 (Beal's Corr.) B. 1.05 (Beal's Corr.) C. 4.6
28. Stock Tank Oil Viscosity, cp
A. N.A. B. N.A. C. 14
29. Stock Tank Oil Gravity, °API
A. 33 B. 31 C. 24° (Avg.)
30. Is there a natural water drive, or significant bottom water?
A. No B. No C. Yes
31. If on water flood, has it been easy to maintain good injection profiles?
Comment:
A. Fair Profiles B. No* C. —
 *B. Injecting over long interval and slotted liner completion open top to bottom.
32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment?
A. N.A. B. N.A. C. 56
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume %
A. N.A. B. N.A. C. 56
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide?
A. No B. No C. No

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		TORRANCE		
1.	Field			
2.	Reservoir	A. <u> Main </u>	B. <u> Main </u>	C. <u> Main </u>
3.	Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u> Waterflood </u>	B. <u> Waterflood </u>	C. <u> Waterflood </u>
4.	Consolidation (Good, Fair, Unconsolidated)	A. <u> Unconsol. </u>	B. <u> Poorly Consol. </u>	C. <u> Unconsol. </u>
5.	Is there significant carbonate cement? (Yes/No)	A. <u> No </u>	B. <u> No </u>	C. <u> No </u>
6.	Subsurface Depth, ft	A. <u> 3750 </u>	B. <u> 3200-3500 </u>	C. <u> 3600 </u>
7.	Gross Section, ft	A. <u> 300 </u>	B. <u> 300 (Avg.) </u>	C. <u> 200 </u>
8.	Net, ft	A. <u> 110 </u>	B. <u> 100 (Avg.) </u>	C. <u> 76 </u>
9.	Is the net uniformly or erratically distributed?	A. <u> Uniformly </u>	B. <u> Uniformly </u>	C. <u> Uniformly </u>
10.	Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u> 254 md geom. avg. 422 md arith. avg. </u>	B. <u> 302 (Avg.) </u>	C. <u> 300 </u>
11.	Permeability, vertical, mds	A. <u> — </u>	B. <u> No Data </u>	C. <u> 210 </u>
12.	Dykstra Coefficient of Permeability Distribution	A. <u> — </u>	B. <u> 0.64 </u>	C. <u> N.A. </u>
13.	Porosity, at atmospheric %	A. <u> .35 </u>	B. <u> 33.9 (Avg.) </u>	C. <u> 34 </u>
14.	Porosity, restored state, %	A. <u> — </u>	B. <u> No Data </u>	C. <u> N.A. </u>
15.	Current Reservoir Pressure, psi	A. <u> 1600 (Est.) </u>	B. <u> 1400 </u>	C. <u> 1050 </u>
16.	Reservoir Temperature °F	A. <u> 160 </u>	B. <u> 170 </u>	C. <u> 130 </u>
17.	Acres	A. <u> 839 </u>	B. <u> 1099 </u>	C. <u> 225 </u>
18.	Acres/Producing Well	A. <u> 16 </u>	B. <u> 21.5 </u>	C. <u> 10 </u>
19.	Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u> Peripheral </u>	B. <u> 5 Spot </u>	C. <u> Line Drive </u>

20. Current Oil in Place, Bbls/Net Acre Foot A. 1151.5 B. 1080 C. N.A.
21. Current Oil Saturation pore volume % A. 45 B. 51 C. N.A.
22. How was above estimated? Material Balance, Recent Coring or Logging) A. Material Balance B. Material Balance C. —
23. Current Formation Volume Factor A. 1.056 B. 1.14 C. 1.03
24. Current Solution Gas, SCF/Bbl A. 110 B. 55 C. N.A.
25. Current Producing GOR, SCF/Bbl A. 110 B. 55 C. 160
26. Current Produced Gas Density A. .75 B. No Data C. N.A.
27. Current Reservoir Oil Viscosity, cp A. 15 B. 5.7 C. 55
28. Stock Tank Oil Viscosity, cp A. 110 (Est.) @ 70°F B. 14 C. 230
29. Stock Tank Oil Gravity, °API A. 19.5 B. 24(Avg.) C. 18
30. Is there a natural water drive, or significant bottom water? A. No B. Yes C. No
31. If on water flood, has it been easy to maintain good injection profiles?
Comment: A. — B. No C. No
- A. Profiles on a whole have not been too bad. Not many efforts have been made to restrict injection other than reducing injection into a well or plugging back from the bottom with sand
- B. Upper Zone tends to become a thief: Perfs tend to plug (despite filtering and treating injection water).
- C. Injectors are converted producers.
32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment? A. — B. 51 C. N.A.
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume % A. — B. 49 C. N.A.

34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide?

A. No B. No C. No

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		WILMINGTON-BLOCK I NON-UNIT		
1.	Field			
2.	Reservoir	A. <u>Ranger</u>	B. <u>Upper Terminal</u>	C. _____
3.	Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Waterflood</u>	B. <u>Waterflood</u>	C. _____
4.	Consolidation (Good, Fair, Unconsolidated)	A. <u>Unconsol.</u>	B. <u>Unconsol.</u>	C. _____
5.	Is there significant carbonate cement? (Yes/No)	A. <u>No</u>	B. <u>No</u>	C. _____
6.	Subsurface Depth, ft	A. <u>2700</u>	B. <u>3100</u>	C. _____
7.	Gross Section, ft	A. <u>400</u>	B. <u>320</u>	C. _____
8.	Net, ft	A. <u>120</u>	B. <u>200</u>	C. _____
9.	Is the net uniformly or erratically distributed?	A. <u>Uniformly</u>	B. <u>Uniformly</u>	C. _____
10.	Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>700</u>	B. <u>350</u>	C. _____
11.	Permeability, vertical, mds	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
12.	Dykstra Coefficient of Permeability Distribution	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
13.	Porosity, at atmospheric %	A. <u>32</u>	B. <u>30</u>	C. _____
14.	Porosity, restored state, %	A. <u>N.A.</u>	B. <u>N.A.</u>	C. _____
15.	Current Reservoir Pressure, psi	A. <u>500</u>	B. <u>665</u>	C. _____
16.	Reservoir Temperature °F	A. <u>125</u>	B. <u>130</u>	C. _____
17.	Acres	A. <u>77</u>	B. <u>77</u>	C. _____
18.	Acres/Producing Well	A. <u>5</u>	B. <u>5</u>	C. _____
19.	Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>Line Drive</u>	B. <u>Line Drive</u>	C. _____

20. Current Oil in Place, Bbls/Net Acre Foot A. N.A. B. N.A. C.
21. Current Oil Saturation pore volume % A. N.A. B. N.A. C.
22. How was above estimated? Material Balance, Recent Coring or Logging) A. Coring B. Coring C.
23. Current Formation Volume Factor A. 1.020 B. 1.032 C.
24. Current Solution Gas, SCF/Bbl A. N.A. B. N.A. C.
25. Current Producing GOR, SCF/Bbl A. 45 B. 60 C.
26. Current Produced Gas Density A. N.A. B. 0.7219 C.
27. Current Reservoir Oil Viscosity, cp A. N.A. B. 53 C.
28. Stock Tank Oil Viscosity, cp A. 280 B. 230 C.
29. Stock Tank Oil Gravity, °API A. 16.2 B. 18.0 C.
30. Is there a natural water drive, or significant bottom water? A. No B. No C.
31. If on water flood, has it been easy to maintain good injection profiles? Comment: A. Yes B. Yes C.
32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment? A. N.A. B. N.A. C.
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume % A. N.A. B. N.A. C.
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide? A. No B. No C.

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		WILMINGTON FIELD BLOCKS II THRU V		
1. Field				
2. Reservoir	A.	<u>Tar</u>	B. <u>Ranger</u>	C. <u>Upper Terminal</u>
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A.	<u>Waterflood</u>	B. <u>Waterflood</u>	C. <u>Waterflood</u>
4. Consolidation (Good, Fair, Unconsolidated)	A.	<u>Unconsol.</u>	B. <u>Unconsol.</u>	C. <u>Unconsol.</u>
5. Is there significant carbonate cement? (Yes/No)	A.	<u>No</u>	B. <u>No</u>	C. <u>No</u>
6. Subsurface Depth, ft	A.	<u>2000</u>	B. <u>2350</u>	C. <u>2850</u>
7. Gross Section, ft	A.	<u>350</u>	B. <u>500</u>	C. <u>570</u>
8. Net, ft	A.	<u>140</u>	B. <u>150</u>	C. <u>340</u>
9. Is the net uniformly or erratically distributed?	A.	<u>Erratic</u>	B. <u>Erratic</u>	C. <u>Erratic</u>
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A.	<u>2500</u>	B. <u>1650</u>	C. <u>740</u>
11. Permeability, vertical, mds	A.	<u>—</u>	B. <u>—</u>	C. <u>—</u>
12. Dykstra Coefficient of Permeability Distribution	A.	<u>.65 approx.</u>	B. <u>.65 approx</u>	C. <u>.65 approx.</u>
13. Porosity, at atmospheric %	A.	<u>36.0</u>	B. <u>36.9</u>	C. <u>34.2</u>
14. Porosity, restored state, %	A.	<u>28.0</u>	B. <u>27.0</u>	C. <u>27.0</u>
15. Current Reservoir Pressure, psi	A.	<u>710</u>	B. <u>830</u>	C. <u>1010</u>
16. Reservoir Temperature °F	A.	<u>118</u>	B. <u>130</u>	C. <u>146</u>
17. Acres	A.	<u>2160</u>	B. <u>3930</u>	C. <u>1690</u>
18. Acres/Producing Well	A.	<u>7 approx.</u>	B. <u>5-10</u>	C. <u>8</u>
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A.	<u>Modified Line-Peripheral</u>	B. <u>Modified 5-Spot</u>	C. <u>Peripheral</u>

20. Current Oil in Place, Bbls/Net Acre Foot
 A. 1120 B. 800 C. 600
21. Current Oil Saturation pore volume %
 A. 53 B. 39 C. 30
22. How was above estimated? Material Balance, Recent Coring or Logging)
 A. Material Balance & Recent Logging B. Material Balance & Recent Logging C. Material Balance & Recent Logging
23. Current Formation Volume Factor
 A. 1.03 B. 1.035 C. 1.03
24. Current Solution Gas, SCF/Bbl
 A. 38 B. 50 C. 20
25. Current Producing GOR, SCF/Bbl
 A. 38 B. 50 C. 20
26. Current Produced Gas Density
 A. .89 to .92 B. .89 to .92 C. .89 to .92
27. Current Reservoir Oil Viscosity, cp
 A. 700 B. 50 C. 26
28. Stock Tank Oil Viscosity, cp
 A. 1500 B. 350 C. 130
29. Stock Tank Oil Gravity, °API
 A. 12-14 B. 12-21 C. 17-23
30. Is there a natural water drive, or significant bottom water?
 A. Yes B. No C. Yes
31. If on water flood, has it been easy to maintain good injection profiles?
 Comment: A. No B. No C. No

Because of our unconsolidated sands, most of injectors are gravel packed for sand control, making it very difficult for selective control. Our artificially controlled lower tier oil price has not allowed us to do redrill or major remedial work.

32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment?
 A. — B. — C. —
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume %
 A. — B. — C. —
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide?
 A. — B. — C. —

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 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

	WILMINGTON FIELD		
1. Field			
2. Reservoir	A. <u>Lower Terminal</u>	B. <u>Union Pacific</u>	C. <u>Ford</u>
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Waterflood</u>	B. <u>Primary</u>	C. <u>Primary</u>
4. Consolidation (Good, Fair, Unconsolidated)	A. <u>Unconsol.</u>	B. <u>Fair</u>	C. <u>Fair</u>
5. Is there significant carbonate cement? (Yes/No)	A. <u>No</u>	B. <u>No</u>	C. <u>No</u>
6. Subsurface Depth, ft	A. <u>3420</u>	B. <u>4130</u>	C. <u>4900</u>
7. Gross Section, ft	A. <u>750</u>	B. <u>730</u>	C. <u>1130</u>
8. Net, ft	A. <u>450</u>	B. <u>160</u>	C. <u>340</u>
9. Is the net uniformly or erratically distributed?	A. <u>Erratic</u>	B. <u>Erratic</u>	C. <u>Erratic</u>
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>450</u>	B. <u>350</u>	C. <u>275</u>
11. Permeability, vertical, mds	A. <u>—</u>	B. <u>—</u>	C. <u>—</u>
12. Dykstra Coefficient of Permeability Distribution	A. <u>.65 approx.</u>	B. <u>.65 approx.</u>	C. <u>.65 approx.</u>
13. Porosity, at atmospheric %	A. <u>32.1</u>	B. <u>31.0</u>	C. <u>29.0</u>
14. Porosity, restored state, %	A. <u>26.0</u>	B. <u>25.0</u>	C. <u>25.0</u>
15. Current Reservoir Pressure, psi	A. <u>1210</u>	B. <u>600</u>	C. <u>600</u>
16. Reservoir Temperature °F	A. <u>164</u>	B. <u>188</u>	C. <u>218</u>
17. Acres	A. <u>1000</u>	B. <u>575</u>	C. <u>1000</u>
18. Acres/Producing Well	A. <u>10</u>	B. <u>10</u>	C. <u>10</u>
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>Peripheral</u>	B. <u>—</u>	C. <u>—</u>

20. Current Oil in Place, Bbls/Net Acre Foot A. 550 B. 650 C. 650
21. Current Oil Saturation pore volume % A. 28 B. 35 C. —
22. How was above estimated? Material Balance, Recent Coring or Logging) A. — B. — C. —
23. Current Formation Volume Factor A. 1.03 B. 1.05 C. 1.05
24. Current Solution Gas, SCF/Bbl A. 30 B. 60 C. 60
25. Current Producing GOR, SCF/Bbl A. 30 B. 60 C. 60
26. Current Produced Gas Density A. — B. — C. —
27. Current Reservoir Oil Viscosity, cp A. 4.5 B. 2.0 C. 2.0
28. Stock Tank Oil Viscosity, cp A. 25 B. 14 C. 14
29. Stock Tank Oil Gravity, °API A. 26-28 B. 29 C. 29
30. Is there a natural water drive, or significant bottom water? A. Yes B. Yes C. Yes
31. If on water flood, has it been easy to maintain good injection profiles? Comment: A. No B. — C. —

32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment? A. — B. — C. —
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume % A. — B. — C. —
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide? A. — B. — C. —

LAWRENCE-ALLISON & ASSOCIATES CORPORATION, WESTERN DIVISION
 QUESTIONNAIRE TO ELICIT DATA ON LOS ANGELES BASIN OIL FIELDS
 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

		EAST WILMINGTON - LONG BEACH UNIT		
1. Field				
2. Reservoir	A.	Ranger	B.	Upper Terminal
			C.	Terminal East
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A.	Waterflood	B.	Waterflood
			C.	Waterflood
4. Consolidation (Good, Fair, Unconsolidated)	A.	Unconsol.	B.	Unconsol.
			C.	Unconsol.- Fair
5. Is there significant carbonate cement? (Yes/No)	A.	Not as cement.	B.	Not as cement.
			C.	Not as cement.
6. Subsurface Depth, ft	A.	2500	B.	3200
			C.	3600
7. Gross Section, ft	A.	850	B.	650
			C.	900
8. Net, ft	A.	300	B.	300-400
			C.	325-600
9. Is the net uniformly or erratically distributed?	A.	Erratically	B.	Erratically
			C.	Erratically
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A.	500-4000 (Unrestored)	B.	500-4000 (Unrestored)
			C.	100-2000 (Unrestored)
11. Permeability, vertical, mds	A.	N.A.	B.	N.A.
			C.	N.A.
12. Dykstra Coefficient of Permeability Distribution	A.	N.A.	B.	N.A.
			C.	N.A.
13. Porosity, at atmospheric %	A.	35-40	B.	30-40
			C.	30-40
14. Porosity, restored state, %	A.	24-27	B.	N.A.
			C.	N.A.
15. Current Reservoir Pressure, psi	A.	1100	B.	1000
			C.	1400
16. Reservoir Temperature °F	A.	125	B.	150
			C.	160
17. Acres	A.	6200	B.	1040
			C.	530
18. Acres/Producing Well	A.	8-18	B.	37
			C.	33
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A.	Line Drive	B.	Peripheral
			C.	Peripheral

20. Current Oil in Place, Bbls/Net Acre Foot
 A. 1200 B. 1900* C. 1600*
21. Current Oil Saturation pore volume %
 A. 56 B. 67* C. 75*
22. How was above estimated? Material Balance, Recent Coring or Logging)
 A. Material Balance B. Material Balance C. Material Balance
23. Current Formation Volume Factor
 A. 1.06 B. 1.06 C. 1.09
24. Current Solution Gas, SCF/Bbl
 A. 110 B. 85 C. 160
25. Current Producing GOR, SCF/Bbl
 A. 180 B. 130 C. 160
26. Current Produced Gas Density
 A. N.A. B. N.A. C. N.A.
27. Current Reservoir Oil Viscosity, cp
 A. 20-70 B. 30 C. 12
28. Stock Tank Oil Viscosity, cp
 A. N.A. B. N.A. C. N.A.
29. Stock Tank Oil Gravity, °API
 A. 14-26 B. 17 C. 20
30. Is there a natural water drive, or significant bottom water?
 A. No B. No C. No
31. If on water flood, has it been easy to maintain good injection profiles?
 Comment: A. No B. No C. No
Permeability profiles not uniform, usually become worse with time.
32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment?
 A. 0.42 B. 0.60* C. 0.57*
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume %
 A. N.A. B. N.A. C. N.A.
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide?
 A. No B. No C. No

*Based on Unrestored Porosity

LAWRENCE-ALLISON & ASSOCIATES CORPORATION, WESTERN DIVISION
 QUESTIONNAIRE TO ELICIT DATA ON LOS ANGELES BASIN OIL FIELDS
 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

		EAST WILMINGTON - LONG BEACH UNIT (Continued)		
1.	Field			
2.	Reservoir	A. <u>UP-Ford</u>	B. _____	C. _____
3.	Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Partial Waterflood</u>	B. _____	C. _____
4.	Consolidation (Good, Fair, Unconsolidated)	A. <u>Fair-Good</u>	B. _____	C. _____
5.	Is there significant carbonate cement? (Yes/No)	A. <u>No</u>	B. _____	C. _____
6.	Subsurface Depth, ft	A. <u>4200-6500</u>	B. _____	C. _____
7.	Gross Section, ft	A. <u>2000</u>	B. _____	C. _____
8.	Net, ft	A. <u>450-750</u>	B. _____	C. _____
9.	Is the net uniformly or erratically distributed?	A. <u>Erratically</u>	B. _____	C. _____
10.	Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>100</u>	B. _____	C. _____
11.	Permeability, vertical, mds	A. <u>N.A.</u>	B. _____	C. _____
12.	Dykstra Coefficient of Permeability Distribution	A. <u>N.A.</u>	B. _____	C. _____
13.	Porosity, at atmospheric %	A. <u>20.5</u>	B. _____	C. _____
14.	Porosity, restored state, %	A. <u>N.A.</u>	B. _____	C. _____
15.	Current Reservoir Pressure, psi	A. <u>1700</u>	B. _____	C. _____
16.	Reservoir Temperature °F	A. <u>200</u>	B. _____	C. _____
17.	Acres	A. <u>2680</u>	B. _____	C. _____
18.	Acres/Producing Well	A. <u>40</u>	B. _____	C. _____
19.	Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>Peripheral</u>	B. _____	C. _____

20. Current Oil in Place, Bbls/Net Acre Foot A. 750 B. _____ C. _____
21. Current Oil Saturation pore volume % A. 58 B. _____ C. _____
22. How was above estimated? Material Balance, Recent Coring or Logging) A. Material Balance B. _____ C. _____
23. Current Formation Volume Factor A. 1.20 B. _____ C. _____
24. Current Solution Gas, SCF/Bbl A. 300 B. _____ C. _____
25. Current Producing GOR, SCF/Bbl A. 700 B. _____ C. _____
26. Current Produced Gas Density A. N.A. B. _____ C. _____
27. Current Reservoir Oil Viscosity, cp A. 1.5 B. _____ C. _____
28. Stock Tank Oil Viscosity, cp A. N.A. B. _____ C. _____
29. Stock Tank Oil Gravity, °API A. 28 B. _____ C. _____
30. Is there a natural water drive, or significant bottom water? A. No B. _____ C. _____
31. If on water flood, has it been easy to maintain good injection profiles? Comment: A. No B. _____ C. _____
Permeability profiles not uniform, usually become worse with time.
32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment? A. 0.53 B. _____ C. _____
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume % A. N.A. B. _____ C. _____
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide? A. No B. _____ C. _____

LAWRENCE-ALLISON & ASSOCIATES CORPORATION, WESTERN DIVISION
 QUESTIONNAIRE TO ELICIT DATA ON LOS ANGELES BASIN OIL FIELDS
 FOR A PRELIMINARY ASSESSMENT OF AMENABILITY TO CARBON DIOXIDE FLOODING

	YORBA LINDA		
1. Field			
2. Reservoir	A. <u>Shallow</u>	B. <u>Main Page</u>	C. <u>Repetto</u>
3. Current Exploitation (Primary, Water Flood or other, specify the latter.)	A. <u>Steam Stimulation</u>	B. <u>Steam Stimulation</u>	C. <u>Primary</u>
4. Consolidation (Good, Fair, Unconsolidated)	A. <u>Unconsol.</u>	B. <u>Unconsol.</u>	C. <u>Unconsol.</u>
5. Is there significant carbonate cement? (Yes/No)	A. <u>No</u>	B. <u>No</u>	C. <u>No</u>
6. Subsurface Depth, ft	A. <u>300-1000</u>	B. <u>1450</u>	C. <u>2200</u>
7. Gross Section, ft	A. <u>—</u>	B. <u>—</u>	C. <u>—</u>
8. Net, ft	A. <u>400</u>	B. <u>70</u>	C. <u>70</u>
9. Is the net uniformly or erratically distributed?	A. <u>Erratically</u>	B. <u>Erratically</u>	C. <u>Erratically</u>
10. Permeability, horizontal to air (or if to water or oil, specify), mds	A. <u>1000</u>	B. <u>1000</u>	C. <u>1000</u>
11. Permeability, vertical, mds	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
12. Dykstra Coefficient of Permeability Distribution	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
13. Porosity, at atmospheric %	A. <u>30</u>	B. <u>30</u>	C. <u>30</u>
14. Porosity, restored state, %	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
15. Current Reservoir Pressure, psi	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>
16. Reservoir Temperature °F	A. <u>90</u>	B. <u>100</u>	C. <u>N.A.</u>
17. Acres	A. <u>100</u>	B. <u>N.A.</u>	C. <u>300</u>
18. Acres/Producing Well	A. <u>1.25</u>	B. <u>N.A.</u>	C. <u>10</u>
19. Water Flood Configu- ration (5-spot, 7-spot, line drive)	A. <u>N.A.</u>	B. <u>N.A.</u>	C. <u>N.A.</u>

20. Current Oil in Place, Bbls/Net Acre Foot
 A. 1000 B. 1000 C. 1000
21. Current Oil Saturation pore volume %
 A. 60 B. 50 C. 50
22. How was above estimated? Material Balance, Recent Coring or Logging)
 A. Material Balance B. Material Balance C. Material Balance
23. Current Formation Volume Factor
 A. N.A. B. N.A. C. N.A.
24. Current Solution Gas, SCF/Bbl
 A. N.A. B. N.A. C. N.A.
25. Current Producing GOR, SCF/Bbl
 A. N.A. B. N.A. C. 200
26. Current Produced Gas Density
 A. N.A. B. N.A. C. N.A.
27. Current Reservoir Oil Viscosity, cp
 A. N.A. B. N.A. C. N.A.
28. Stock Tank Oil Viscosity, cp
 A. 10,000 B. N.A. C. N.A.
29. Stock Tank Oil Gravity, °API
 A. 13 B. N.A. C. 16
30. Is there a natural water drive, or significant bottom water?
 A. Yes B. Yes C. Yes
31. If on water flood, has it been easy to maintain good injection profiles?
 Comment:
 A. N.A. B. N.A. C. N.A.

32. At current oil prices and under current operations, what would the residual oil saturation be at abandonment?
 A. N.A. B. N.A. C. N.A.
33. Under world oil prices but with current technology, what would be the residual oil saturation at abandonment? Pore volume %
 A. N.A. B. N.A. C. N.A.
34. Have any miscibility, solubility, or other studies of this crude been made with carbon dioxide?
 A. No B. No C. No

PROCESSING AND TRANSMISSION SYSTEMS COST DATA

PURIFICATION ALTERNATE NO. 1

Carbon Dioxide Recovery from Power Plant Flue Gases via Absorption with MEA
 Product: 73.2 MMscfd Pure CO₂

CAPITAL COSTS

<u>CO₂ purification</u>	<u>Installed cost</u> (<u>\$</u>)	<u>Cost to</u> <u>Customer</u> (<u>\$</u>)
Flue gas fan	670,000	(see Note 4)
Raw/treated gas exchanger	160,300	
Absorber column	146,300	
Rich MEA pump	400,000	
Rich MEA filter	100,000	
Lean/rich MEA exchanger	1,405,700	
Stripper column	150,400	
Stripper bottoms pump	400,000	
Lean MEA cooler	2,683,200	
Stripper reboiler	3,015,300	
Stripper overhead condenser	1,057,500	
Stripper overhead reflux drum	100,000	
Stripper overhead reflux pump	266,000	
Seawater cooling circulation pump	1,662,500	
MEA reclamation unit	<u>(unknown)</u>	
Subtotal	<u>12,217,200</u>	<u>13,438,920</u>
 <u>Compression and drying of pure CO₂</u>		
First compressor	6,250,000	
First compressor coolers	170,430	
Miscellaneous knockout drums	50,000	
Second compressor	8,800,000	
Second compressor coolers	183,540	
Miscellaneous knockout drums	50,000	
Dehydration	<u>400,900</u>	
Subtotal	<u>15,904,970</u>	<u>17,495,357</u>
TOTAL CAPITAL INVESTMENT	<u>28,122,170</u>	<u>30,934,277</u>

UTILITY COSTS

	<u>Total Utility Use</u>	<u>Use per MSCF CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>CO₂ purification</u>			
Inlet fan	4,990 kW	1.635 kWh	0.051
Miscellaneous pumps	1,193 kW	.391 kWh	0.012
Stripper reboiler (steam)	709.9 mlb/h	232.6 lb	0.637
Seawater cooling	1,864 kW	.61 kWh	0.019
MEA reclamation	—	—	<u>(unknown)</u>
Subtotal			<u>0.719</u>
<u>Compression & drying of pure CO₂</u>			
First compressor	9,320 kW	3.05 kWh	0.095
Second compressor	13,420 kW	4.40 kWh	<u>0.137</u>
Subtotal			<u>0.232</u>
TOTAL UTILITY COSTS			<u>0.951</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Purification, Compression, and Drying

Capital Investment = \$30,934,277

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.314
Steam	lb	0.00274/lb	<u>0.637</u>
Subtotal			<u>0.951</u>
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.003
Operating supplies—1/3 operating labor			0.001
Supervision & overhead—100 percent operating labor			0.003
Maintenance labor & supplies—3½ percent CI/year			0.040
Capital-related costs (see notes)			<u>0.472</u>
Subtotal			<u>0.519</u>
TOTAL CO ₂ PROCESSING COSTS			<u>1.470</u>
 Transmission costs via 18-in. diameter, 14-mile-long pipeline			
<u>Capital Investment</u> = \$3,992,000			
Right-of-way rental (\$80,000/yr)			0.003
Capital-related costs (see notes)			<u>0.061</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.064</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI			<u>1.534</u>

NOTES: (1) Yearly capital-related costs consist of the following:

Depreciation (assuming straight-line depreciation over a 10-year life)	10%
Interest charges	5%
Taxes and insurance	3%

Return on investment	<u>20%</u>
Total, as a percentage of initial capital investment	<u>38%</u>

- (2) Plant assumed to operate 340 days/year
- (3) The total capital investment does include the cost of coolingwater pumps and a cooling tower, if required by the design.
- (4) The "Cost to Customer" figures include a contractor profit and overhead allowance equal to 10 percent of the total "Installed Cost" numbers. The installed costs do not include the costs of land, access roads, environmental reports, permits, royalties, initial charge of chemicals, lubricants, and other similar supplies, specialties or services, nor office buildings, warehouses or shops.

PURIFICATION ALTERNATE NO. 3

Carbon Dioxide Recovery from Refinery Off-Gases via Low-Temperature Separation
 Product: 26.6 MMscfd Pure CO₂

CAPITAL COSTS

<u>Compress and dehydrate</u>	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
First knockout drum	16,000	
First compressor	3,150,000	
First compressor coolers	109,250	
Second knockout drum	12,400	
Second compressor	3,150,000	
Second compressor coolers	91,770	
Third knockout drum	7,900	
Dehydration	<u>286,000</u>	
Subtotal	<u>6,823,320</u>	<u>7,505,652</u>
 <u>Purify</u>		
Inlet gas refrigerated cooler	92,650	
Condensate receiver	40,200	
First flash vessel	40,000	
First & second flash recompressors	500,000	
First flash recompressor aftercooler	3,930	
Second flash vessel	6,800	
Power recovery turbines	700,000	
Cold recovery exchangers	14,420	
Liquid CO ₂ knockout vessel	6,800	
Refrigeration unit	<u>3,192,000</u>	
Subtotal	<u>4,596,800</u>	<u>5,056,480</u>
 <u>Store liquid CO₂</u>		
Liquid CO ₂ storage vessel	150,300	
Loading pump	9,300	
Oil field CO ₂ storage	150,300	
Unloading pump	<u>9,300</u>	
Subtotal	<u>319,200</u>	<u>351,120</u>
 <u>Reheat liquid CO₂</u>		
Oil field reheater	29,100	32,000

	<u>Installed cost</u> (<u>\$</u>)	<u>Cost to</u> <u>Customer</u> (<u>\$</u>)
<u>Inject CO₂</u>		
Oil field injection pump	101,100	111,200
<u>Utilities & offsites</u>		
Cooling tower	465,500	
Coolingwater pump	<u>150,500</u>	
Subtotal	<u>616,000</u>	<u>677,600</u>
TOTAL CAPITAL INVESTMENT	<u>12,485,500</u>	<u>13,734,052</u>

UTILITY COSTS

	<u>Total</u> <u>Utility</u> <u>Use</u>	<u>Use per</u> <u>Mscf</u> <u>CO₂</u>	<u>Cost per</u> <u>Mscf</u> <u>CO₂</u> (<u>\$</u>)
<u>Compress and dehydrate</u>			
Total power first and second compressors	9,694 kW	8.73 kWh	0.273
<u>Purify</u>			
Recompressors	745 kW		
Less power recovered via turbines	-745 kW		
Electric power refrigeration unit	3,400 kW	3.06 kWh	0.096
<u>Store & handle</u>			
Loading & unloading pumps	21 kW	.02 kWh	0.001
<u>Reheat</u>			
Steam to reheater	5,660 lb/h	5.1 lb	0.014
<u>Inject</u>			
Injection pump power	100 kW	.087 kWh	0.003
<u>Utilities & offsites</u>			
Coolingwater pump	200 kW	.18 kWh	<u>0.006</u>
TOTAL UTILITY COSTS			<u>0.393</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, Low-Temperature Purification

Capital Investment = \$13,734,052

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)²</u>
Electric power	kWh	0.03125/kWh	0.379
Steam	lb	0.00274/lb	<u>0.014</u>
Subtotal			<u>0.393</u>
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.009
Operating supplies—1/3 operating labor			0.003
Supervision & overhead—100 percent operating labor			0.009
Maintenance labor & supplies—3½ percent CI/yr			0.053
Capital-related costs (see notes Purification Alternate No. 1)			<u>0.577</u>
Subtotal			<u>0.651</u>
TOTAL CO ₂ PROCESSING COSTS			<u>1.044</u>
 Transmission costs: Truck Hauling			
<u>Capital Investment</u> = None			
			 Cost per Mscf CO ₂ (\$) ²
<u>Direct costs</u>			
Truck hauling (all inclusive)			0.276
Operating labor			<u>0.014</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.290</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED			<u><u>1.334</u></u>

TRANSMISSION ALTERNATE NO. 1

Transport Liquid CO₂ by Truck

Product: 33.1 MMscfd Pure CO₂

CAPITAL COSTS

<u>Compress and dehydrate</u>	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
First knockout drum	16,000	(see Note 4)
First compressor	2,900,000	
First compressor coolers	109,250	
Second knockout drum	12,400	
Second compressor	2,900,000	
Second compressor coolers	91,770	
Third knockout drum	7,900	
Dehydration	<u>286,000</u>	
Subtotal	<u>6,323,320</u>	<u>6,955,652</u>
 <u>Purify</u>		
Lean/rich gas exchangers	7,900	
Absorber column	26,400	
Rich propylene carbonate exchangers	285,800	
Gas/liquid separator & flash drum	15,000	
Propylene carbonate recirculation pump	491,000	
Lean propylene carbonate refrigerated cooler	95,220	
Lean gas power recovery turbines	393,400	
Refrigeration unit	<u>2,656,500</u>	
Subtotal	<u>3,971,220</u>	<u>4,368,342</u>
 <u>Condense pure CO₂</u>		
CO ₂ condenser	96,600	
Refrigeration unit	<u>3,979,500</u>	
Subtotal	<u>4,076,100</u>	<u>4,483,710</u>
 <u>Store & handle liquid CO₂</u>		
Liquid CO ₂ storage vessel	150,300	
Loading pump	9,300	
Oil field CO ₂ storage	150,300	
Unloading pump	<u>9,300</u>	
Subtotal	<u>319,200</u>	<u>351,120</u>

	<u>Installed cost</u> <u>(\$)</u>	<u>customer</u> <u>(\$)</u>
<u>Reheat liquid CO₂</u>		
Oil field reheater	29,100	32,000
<u>Inject liquid CO₂</u>		
Oil field injection pump	101,100	111,200
<u>Utilities & offsites</u>		
Cooling tower	665,000	
Coolingwater pump	<u>215,460</u>	
Subtotal	<u>880,460</u>	<u>968,500</u>
TOTAL CAPITAL INVESTMENT	<u>15,700,500</u>	<u>17,270,550</u>

UTILITY COSTS

	<u>Total</u> <u>Utility</u> <u>Use</u>	<u>Use per</u> <u>Mscf</u> <u>CO₂</u>	<u>Cost per</u> <u>Mscf</u> <u>CO₂</u> <u>(\$)</u>
<u>Compress and dehydrate</u>			
Electric power for compressors	8,948 kW	6.85 kWh	0.214
<u>Purify</u>			
Pump power	550 kW		
Less power recovered via turbines	<u>-420 kW</u>		
Net Power Use	130 kW	0.0995 kWh	0.003
Refrigeration unit power	2,830 kW	2.17 kWh	0.068
<u>Condense pure CO₂</u>			
Refrigeration unit power	4,240 kW	3.25 kWh	0.101
<u>Store & handle liquid CO₂</u>			
Loading & unloading pump electric power	26 kW	0.02 kWh	0.001
<u>Reheat liquid CO₂</u>			
Steam to reheater	6,660 lb/h	5.1 lb	0.014

	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>Inject liquid CO₂</u>			
Electric power	113 kW	.087 kWh	0.003
<u>Utilities & offsites</u>			
Coolingwater pump	242 kW	.18 kWh	<u>0.006</u>
TOTAL UTILITY COSTS			<u>0.410</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, Purification, Condensation, Liquid Handling, and Injection

Capital Investment = \$17,270,550

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.396
Steam	lb	0.00274/lb	<u>0.014</u>
Subtotal			<u>0.410</u>
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.057
Capital-related costs (see notes)			<u>0.616</u>
Subtotal			0.689
TOTAL CO ₂ PROCESSING COSTS			<u>1.099</u>

Transmission costs: Truck Hauling

Capital Investment = None

<u>Direct costs</u>	<u>Cost per Mscf CO₂ (\$)</u>
Truck hauling (all inclusive)	0.276
Operating labor	<u>0.014</u>
TOTAL CO ₂ TRANSMISSION COSTS	<u>0.290</u>
TOTAL COST OF LIQUID CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI	<u><u>1.389</u></u>

NOTES: (1) Yearly capital-related costs consist of the following:

Depreciation (assuming straight-line depreciation over a 10-year life)	10%
Interest charges	5%
Taxes and insurance	3%
Return on investment	<u>20%</u>
Total, as a percentage of initial capital investment	<u>38%</u>

- (2) Plant assumed to operate 340 days/year
- (3) The total capital investment does include the cost of a cooling tower and coolingwater pumps, if required by the specific design.
- (4) The "Cost to Customer" figures include a contractor profit and overhead allowance equal to 10 percent of the total "Installed Cost" numbers. The installed costs do not include the costs of land, access roads, environmental reports, permits, royalties, initial charge of chemicals, lubricants, and other similar supplies, specialties or services, nor office buildings, warehouses or shops.

TRANSMISSION ALTERNATE NO. 2

Transport Unpurified CO₂ Gas via Existing Pipelines

Product: 37.8 MMscfd 87.5% Pure CO₂

CAPITAL COSTS

<u>Compress and dehydrate</u>	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
First knockout drum	16,000	
First compressor	2,800,000	
First compressor coolers	106,900	
Second knockout drum	12,400	
Second compressor	2,800,000	
Second compressor coolers	89,800	
Third knockout drum	7,900	
Dehydrate	<u>286,000</u>	
Subtotal	<u>6,119,000</u>	<u>6,730,900</u>
 <u>Recompress after transmission</u>		
Recompressor	2,050,000	2,255,000
 <u>Utilities & offsites</u>		
Cooling tower	300,000	
Coolingwater pump	<u>90,440</u>	
Subtotal	<u>390,440</u>	<u>429,484</u>
TOTAL CAPITAL INVESTMENT	<u>8,559,440</u>	<u>9,415,384</u>

UTILITY COSTS

<u>Compress and dehydrate</u>	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	8,800 kW	6.64 kWh	0.208

	<u>Total Utility</u>	<u>Use per Mscf</u>	<u>Cost per Mscf CO₂</u>
<u>Recompress</u>			
Recompressor power	3,057 kW	2.34 kWh	0.073
<u>Utilities & offsites</u>			
Power coolingwater pump	100 kW	.08 kWh	<u>0.002</u>
TOTAL UTILITY COSTS			<u>0.283</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, and Recompression

Capital Investment = \$9,415,384

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.283
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.029
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.336</u>
Subtotal			<u>0.381</u>
TOTAL CO ₂ PROCESSING COSTS			<u>0.664</u>

Transmission costs: Existing Pipelines

Capital Investment: None

<u>Direct costs</u>	<u>Cost per Mscf CO₂ (\$)</u>
Pipeline rental	0.007
Right-of-way rental	<u>0.007</u>
TOTAL CO ₂ TRANSMISSION COSTS	<u>0.014</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI	<u>0.678</u>

TRANSMISSION ALTERNATE NO. 3

Transport Purified CO₂ Gas via Existing Pipelines

Product 33.1 MMscfd Pure CO₂

CAPITAL COSTS

<u>Compress and dehydrate</u>	<u>Installed cost</u> <u>(\$)</u>	<u>Cost to</u> <u>Customer</u> <u>(\$)</u>
First knockout drum	16,000	
First compressor	2,900,000	
First compressor coolers	109,250	
Second knockout drum	12,400	
Second compressor	2,900,000	
Second compressor coolers	91,770	
Third knockout drum	7,900	
Dehydrate	<u>286,000</u>	
Subtotal	<u>6,323,320</u>	<u>6,955,652</u>
 <u>Purify</u>		
Lean/rich gas exchangers	7,900	
Absorber column	26,400	
Rich propylene carbonate exchangers	285,800	
Gas/liquid separator & flash drum	15,000	
Propylene carbonate recirculation pump	491,000	
Lean propylene carbonate refrigerated cooler	95,220	
Lean gas power recovery turbines	393,400	
Refrigeration unit	<u>2,656,500</u>	
Subtotal	<u>3,971,220</u>	<u>4,368,342</u>
 <u>Boost compression</u>		
Booster compressor	750,000	
Aftercooler	<u>11,275</u>	
Subtotal	<u>761,275</u>	<u>837,400</u>
 <u>Recompress CO₂ after transmission</u>		
Recompressor	1,750,000	1,925,000

	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
<u>Utilities & offsites</u>		
Cooling tower	665,000	
Coolingwater pump	<u>215,460</u>	
Subtotal	<u>880,460</u>	<u>968,500</u>
TOTAL CAPITAL INVESTMENT	<u>13,686,275</u>	<u>15,054,902</u>

UTILITY COSTS

	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>Compress and dehydrate</u>			
Total power first and second compressors	8,948 kW	6.85 kWh	0.214
<u>Purify</u>			
Pump power	550 kW		
Less power recovered via turbines	<u>-420 kW</u>		
Net Power Use	130 kW	0.099 kWh	0.003
Refrigeration unit power	2,830 kW	2.170 kWh	0.068
<u>Boost compression</u>			
Booster compressor power	1,120 kW	.86 kWh	0.027
<u>Recompress</u>			
Recompressor power	2,610 kW	2.0 kWh	0.062
<u>Utilities & offsites</u>			
Power coolingwater pump	193 kW	.15 kWh	<u>0.005</u>
TOTAL UTILITY COSTS			<u>0.379</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, Purification, Boosting, and Recompression

Capital Investment = \$15,054,902

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.379
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.048
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.537</u>
Subtotal Indirect Costs			<u>0.601</u>
TOTAL CO ₂ PROCESSING COSTS			<u>0.980</u>
 Transmission costs: Existing Pipelines			
<u>Capital Investment</u> = None			
<u>Direct costs</u>			<u>Cost per Mscf CO₂ (\$)</u>
Pipeline rental			0.007
Right-of-way rental			<u>0.007</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.014</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI			<u><u>0.994</u></u>

TRANSMISSION ALTERNATE NO. 4

Transport Unpurified CO₂ Gas via New Pipeline

Product: 37.8 MMscfd 87.5% Pure CO₂

CAPITAL COSTS

	<u>Installed cost</u> <u>(\$)</u>	<u>Cost to</u> <u>Customer</u> <u>(\$)</u>
<u>Compress and dehydrate</u>		
First knockout drum	16,000	
First compressor	2,800,000	
First compressor coolers	98,300	
Second knockout drum	12,400	
Second compressor	2,800,000	
Second compressor coolers	98,300	
Third knockout drum	7,900	
Dehydrate	<u>286,000</u>	
Subtotal	<u>6,118,900</u>	<u>6,730,790</u>
<u>Boost compression</u>		
Booster compressor	1,550,000	
Aftercooler	<u>31,900</u>	
Subtotal	<u>1,581,900</u>	<u>1,740,090</u>
<u>Utilities & offsites</u>		
Cooling tower	347,800	
Coolingwater pump	<u>107,730</u>	
Subtotal	<u>455,530</u>	<u>501,083</u>
TOTAL PROCESSING CAPITAL INVESTMENT	<u>8,256,330</u>	<u>9,081,963</u>
<u>Transmission</u>		
Install new 12" pipeline		2,063,150
TOTAL CAPITAL INVESTMENT		<u>11,145,113</u>

UTILITY COSTS

	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>Compress and dehydrate</u>			
Total power first and second compressors	8,800 kW	6.64 kWh	0.208
<u>Boost compression</u>			
Booster compressor electric power	2,310 kW	1.77 kWh	0.055
<u>Utilities & offsites</u>			
Power coolingwater pump	120 kW	.09 kWh	<u>0.003</u>
TOTAL UTILITY COSTS			<u>0.266</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, and Compression Boosting

Capital Investment = \$9,081,963

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.266
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.028
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.324</u>
Subtotal			<u>0.368</u>
TOTAL CO ₂ PROCESSING COSTS			<u>0.634</u>
Transmission costs: Via new 12-in. diameter, 8.9-mile-long pipeline			
<u>Capital Investment</u> = \$2,063,150			
Right-of-way rental			0.004
Capital-related costs (see notes Transmission Alternative No. 1)			<u>0.074</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.078</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI			<u><u>0.712</u></u>

TRANSMISSION ALTERNATE NO. 5

Transport Purified CO₂ Gas via New Pipeline

Product: 33.1 MMscfd Pure CO₂

CAPITAL COSTS

	Installed cost (\$)	Cost to customer (\$)
<u>Compress and dehydrate</u>		
First knockout drum	16,000	
First compressor	2,900,000	
First compressor coolers	109,250	
Second knockout drum	12,400	
Second compressor	2,900,000	
Second compressor coolers	91,770	
Third knockout drum	7,900	
Dehydrate	<u>286,000</u>	
Subtotal	<u>6,323,320</u>	<u>6,955,652</u>
<u>Purify</u>		
Lean/rich gas exchangers	7,900	
Absorber column	26,400	
Rich propylene carbonate exchangers	285,800	
Gas/liquid separator & flash drum	15,000	
Propylene carbonate recirculation pump	491,000	
Lean propylene carbonate refrigerated cooler	95,220	
Lean gas power recovery turbines	393,400	
Refrigeration unit	<u>2,656,500</u>	
Subtotal	<u>3,971,220</u>	<u>4,368,342</u>
<u>Boost compression</u>		
Booster compressor	2,200,000	
Aftercooler	<u>40,200</u>	
Subtotal	<u>2,240,200</u>	<u>2,464,220</u>
<u>Utilities & offsites</u>		
Cooling tower	665,000	
Coolingwater pump	<u>215,460</u>	
Subtotal	<u>880,460</u>	<u>968,500</u>
TOTAL PROCESSING CAPITAL INVESTMENT	<u>13,415,200</u>	<u>14,756,720</u>

<u>Transmission</u>	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
Install new 12" pipeline		2,063,150
TOTAL CAPITAL INVESTMENT		<u>16,819,879</u>

UTILITY COSTS

	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>Compress and dehydrate</u>			
Total power first and second compressors	8,948 kW	6.85 kWh	0.214
<u>Purify</u>			
Pump power	550 kW		
Less power recovered via turbines	<u>-420 kW</u>		
Net Power Use	130 kW	0.099 kWh	0.003
Refrigeration unit power	2,830 kW	2.17 kWh	0.068
<u>Boost compression</u>			
Booster compressor power	3,280 kW	2.51 kWh	0.078
<u>Utilities & offsites</u>			
Power coolingwater pump	209 kW	.16 kWh	<u>0.005</u>
TOTAL UTILITY COSTS			<u>0.368</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, Purification, and Compression Boosting

Capital Investment = \$14,756,720

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.368
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.046
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.526</u>
Subtotal Indirect Costs			<u>0.588</u>
TOTAL CO ₂ PROCESSING COSTS			<u>0.956</u>
 Transmission costs: Via new 12-in. diameter, 8.9-mile-long pipeline			
<u>Capital Investment</u> = \$2,063,150			
Right-of-way rental			0.004
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.074</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.078</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL AT 1,500 PSI			<u>1.034</u>

TRANSMISSION ALTERNATE NO. 6

Transport Liquid CO₂ via New Pipeline

Product: 33.1 MMscfd Pure CO₂

CAPITAL COSTS

	Installed cost (\$)	Cost to Customer (\$)
<u>Compress and dehydrate</u>		
First knockout drum	16,000	
First compressor	2,900,000	
First compressor coolers	109,250	
Second knockout drum	12,400	
Second compressor	2,900,000	
Second compressor coolers	91,770	
Third knockout drum	7,900	
Dehydrate	<u>286,000</u>	
Subtotal	<u>6,323,320</u>	<u>6,955,652</u>
<u>Purify</u>		
Lean/rich gas exchangers	7,900	
Absorber column	26,400	
Rich propylene carbonate exchangers	285,800	
Gas/liquid separator & flash drum	15,000	
Propylene carbonate recirculation pump	491,000	
Lean propylene carbonate refrigerated cooler	95,220	
Lean gas power recovery turbine	393,400	
Refrigeration unit	<u>2,656,500</u>	
Subtotal	<u>3,971,220</u>	<u>4,368,342</u>
<u>Condense</u>		
CO ₂ Gas/liquid exchanger	21,800	
CO ₂ condenser	74,800	
Refrigeration unit	<u>3,279,500</u>	
Subtotal	<u>3,376,100</u>	<u>3,713,710</u>
<u>Pump</u>		
Pipeline pump	118,370	
Surge vessel	<u>15,000</u>	
Subtotal	<u>133,370</u>	<u>146,700</u>

	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
<u>Utilities & offsites</u>		
Cooling tower	665,000	
Coolingwater pump	<u>215,460</u>	
Subtotal	<u>880,460</u>	<u>968,500</u>
TOTAL PROCESSING CAPITAL INVESTMENT	<u>14,684,470</u>	<u>16,152,917</u>

	<u>Installed cost (\$)</u>	<u>Cost to Customer (\$)</u>
<u>Transmission</u>		
Install new 8" pipeline		1,845,260
TOTAL CAPITAL INVESTMENT		<u>17,998,177</u>

UTILITY COSTS

	<u>Total Utility Use</u>	<u>Use per Mscf CO₂</u>	<u>Cost per Mscf CO₂ (\$)</u>
<u>Compress and dehydrate</u>			
Total power first and second compressors	8,948 kW	6.85 kWh	0.214
<u>Purify</u>			
Pump power	550 kW		
Less power recovered via turbines	<u>-420 kW</u>		
Net Power Use	130 kW	0.099 kWh	0.003
Refrigeration unit power	2,830 kW	2.17 kWh	0.068
<u>Condense</u>			
Refrigeration unit power	3,490 kW	2.67 kWh	0.084
<u>Pump</u>			
Pipeline pump power	133 kW	.10 kWh	0.003
<u>Utilities & offsites</u>			
Coolingwater pump	242 kW	.18 kWh	<u>0.006</u>
TOTAL UTILITY COSTS			<u>0.378</u>

TOTAL COST OF CO₂ PRODUCT DELIVERED TO OIL FIELD

Processing Costs: Compression, Drying, Purification, Condensing, and Pumping

Capital Investment = \$16,152,917

<u>Direct costs</u>	<u>Utility Unit</u>	<u>Unit Cost (\$)</u>	<u>Cost per Mscf CO₂ (\$)</u>
Electric power	kWh	0.03125/kWh	0.378
 <u>Indirect costs</u>			
Operating labor—1 man/shift @ \$9/hr			0.007
Operating supplies—1/3 operating labor			0.002
Supervision & overhead—100 percent operating labor			0.007
Maintenance labor & supplies—3½ percent CI/yr			0.053
Capital-related costs (see notes Transmission Alternate No. 1)			<u>0.576</u>
Subtotal			<u>0.645</u>
TOTAL CO ₂ PROCESSING COSTS			<u>1.023</u>
Transmission costs: Via new 8-in.-diameter, 8.9-mile-long pipeline			
<u>Capital Investment</u> = \$1,845,260			
Right-of-way rental			0.004
Capital-related costs (see note Transmission Alternate No. 1)			<u>0.066</u>
TOTAL CO ₂ TRANSMISSION COSTS			<u>0.070</u>
TOTAL COST OF CO ₂ PRODUCT DELIVERED TO INJECTION WELL			<u>1.093</u>

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