PROCESS DEVELOPMENT STUDY NO. 17

CO_2 COMPRESSION FOR ENHANCED OIL RECOVERY

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PROCESS DEVELOPMENT STUDY NO. 17

CO_2 COMPRESSION FOR ENHANCED OIL RECOVERY

1.0 INTRODUCTION

Tri-State Synfuels Company proposes to construct and operate an indirect coal liquefaction plant in Western Kentucky based on Sasol-type technology. The original Feasibility Study identified the availability of large quantities of carbon dioxide from the Stretford and Synthol units in the liquefaction plant. The possibility of selling this carbon dioxide for enhanced oil recovery (EOR) is now under investigation.

The purpose of this process study is to develop the data required to perform an economic evaluation of this possibility. The scope of work for this study (Appendix 1) included the following:

a. A determination of the preferred method of pressurizing the gas and moving it to the plant "battery limits". This entailed a comparative evaluation of three possibilities:

1. Straight compression to 2200 psia.
2. Compression, liquefaction with ammonia absorption refrigeration, pumping to 2200 psia and revaporization. This case is referred to as the ammonia absorption refrigeration scheme.
3. Compression, liquefaction with propylene refrigeration, pumping to 2200 psia and revaporization. This case is referred to as the propylene refrigeration scheme.

b. A process design for compressing and drying the gas to desired battery limit conditions via the method determined above.

c. A determination of the pressure, temperature, and dew-point requirements for the CO_2 stream.

d. An evaluation of the effect on CO_2 compressor plant operations when CO_2 is either not available or demand for it slackens.

e. An estimate of capital and operating costs for the "battery limits" installation.

This work has been completed and this report presents the pertinent findings.

2.0 SUMMARY

2.1   The capital costs given in this report are the costs associated with pressurizing the CO_2 and delivering it to plant battery limits at the following conditions:

Pressure	2200 psia
Temperature	100 °F
Dew Point at 2200 psia	-20 °F

2.2   Straight compression was selected as the preferred method of pressurizing the CO_2 gas and moving it to the plant battery limits. This selection was based on comparative analysis of compression versus liquefaction, pumping and re-vaporization using either propylene refrigeration or ammonia absorption refrigeration to liquify the CO_2.

2.3   The recommended design for compressing and drying the gas consists of a 5 stage compressor that increases CO_2 pressure from 15 psia to 2205 psia. Intercoolers and condensate separators are provided between stages and an aftercooler is also included. Final gas drying requirements are met by operating an ethylene glycol dryer on gas flowing between the 4th and 5th compressor stages. Six parallel compressor trains are provided to accommodate design flow rates.

2.4   The feed rate may be reduced because CO_2 gas is either not available or demand for it slackens. This change in operations will be accomplished by unloading and shutting down individual compressor trains until the desired flow rate is reached. Lack of demand will require venting of any excess CO_2 gas. Incineration prior to venting will be required because of its hydrocarbon content. The required incineration and waste heat recovery equipment were included in the initial Feasibility Study where all of this gas was vented continuously. This equipment must be retained in this arrangement to accommodate periodic venting operations.

2.5 CO_2 sales for EOR will increase coal consumption in the main steam plant. The steam balance in the Feasibility Study utilized waste heat recovered from CO_2 incineration prior to venting. CO_2 sales here would eliminate incineration and waste heat recovery from this gas. To maintain the present steam balance, an additional 572 MM BTU (HHV)/hr of coal must be burned in the coal burning steam plant boilers.

2.6 Production data for the battery limit CO_2 compression plant are estimated as follows:

Maximum Operation	340 days/yr
CO_2 Sales Gas Production Rate	429.87 MMSCFD
HHV Sales Gas	572 MM BTU/hr @ 31.9 BTU/SCF

2.7 The capital cost of the battery limit CO_2 compression plant on a January 1981 instantaneous cost basis with 10 percent contingency and no sales tax is estimated to be $106.58 MM.

2.8 The direct operating requirements for the battery limit CO_2 compression plant are estimated as follows:

Electricity: 114,548 kw @ $0.055/kwh	$51.41 MM/yr
Cooling Water: 51,900 GPM @ $0.053/1000 gal	1.35
Make-up Water: (112.1 GPM) @ $0.50/1000 gal	(.03)
Coal: 572 MM BTU (HHV)/hr @ $2.00/Million BTU	9.34
Ethylene Glycol: 0.10 gal/MM SCF @ $3.35/gal	.05
Operating Labor: 3 men/shift @ $29,000/man yr	.35
Maintenance Labor: 40% x 3.5% x capital cost	1.49
Maintenance Material: 60% x 3.5% x capital cost	2.24
Total Direct Operating Costs	$66.20 MM/yr
	$ 0.45/MSCF

NOTE: Utility costs above are those currently proposed by Fluor and are given in the Process Design Criteria manual.

3.0 PROCESS DESCRIPTIONS

Descriptions of the alternative designs are presented in the following sections.

3.1 Straight Compression Scheme

The configuration used for the design of this scheme is graphically described by the Process Flow Diagram on Fig. 6-1. Here the combined CO_2 gas from the Sulfur Recovery and Synthol units is pressurized from 15 to 2205 psia in a five stage compressor (28-1-C-1). Intercoolers with moisture separators (28-1-E-1, 28-1-E-2, 28-1-E-3, and 28-1-E-4) are provided interstage to minimize compression work and moisture load on the dryer. Sales gas moisture requirements are met by operating an ethylene glycol dryer (28-1-D-1) on the gas flowing between the 4th and 5th compressor stages. Ethylene glycol drying was selected in this case instead of molecular sieves because the pressure is sufficiently high to justify its use. Operating conditions at this point preclude the possible formation of

hydrates. An aftercooler (28-1-E-5) is used to provide a
moderate sales gas temperature. Recovered water is recycled
to the Stretford unit after flash-degassing (28-1-V-1).

Six parallel equipment trains are provided to accommodate
the design flow rate.

3.2 Ammonia Absorption Refrigeration Scheme

Compression of carbon dioxide in this arrangement is graphi-
cally described by the Process Flow Diagram on Fig. 6-2. Here,
feed gases from the Sulfur Recovery and Synthol units are
combined and compressed (28-1-C-1) in three stages to 315 psia.
Intercoolers (28-1-E-1 and 28-1-E-2) including water knock-out
facilities are provided between stages to reduce work requirements.
A molecular-sieve dryer (28-1-D-1) is also included between the
second and third compressor stages to produce a dry gas that
meets product moisture requirements.

Dry gas at 315 psia is then cooled to 20°F in a feed-effluent
exchanger (28-1-E-3) which utilizes the transferred energy to
superheat pressurized CO_2 to about 127°F. The CO_2 gas at 20°F
is further cooled and condensed to a liquid in exchanger
28-1-E-4 which removes the necessary heat by vaporizing liquid
NH_3. The resultant liquid CO_2 is separated (28-1-V-2) from non-
condensible gases and pumped (28-1-P-1) to the desired super-
critical pressure. Refrigeration available in the supercriti-
cal CO_2 is recovered by inter-change (28-1-E-3) as mentioned
above.

Available refrigeration associated with non-condensible gases
leaving separator 28-1-V-2 is recovered by exchange (28-1-E-5)
with liquid NH_3. These gases are then compressed from 300 psia
to 2205 psia in a two stage reciprocating compressor (28-1-C-2)
and combined with the main gas stream as part of the sales gas.
An intercooler (28-1-E-6) and aftercooler (28-1-E-7) are provided
here.

Six parallel equipment trains as shown on Fig. 6-2 are provided
to accommodate the design flow rate.

The NH_3 absorption refrigeration system required to sustain
these compressor trains is graphically depicted by the Process
Flow Diagram on Fig. 6-3. In this arrangement, recycle strong
ammonia liquors are fed into a fractionating column (28-1-V-4)
which separates them into a liquid overhead that is essentially
pure NH_3 (99.9 mol percent) and aqueous bottoms that contain
19 mol percent NH_3. Part of the liquid NH_3 is returned to
the fractionating column as reflux. The remainder is sub-
cooled (28-1-E-5 and 28-1-E-10) prior to Joule-Thompson
expansion in order to reduce the amount of gas flashed in this

auto-refrigeration step. Low temperature flash gas is separated (28-1-V-6) from the remaining liquid which is then vaporized in the CO_2 condenser (28-1-E-4). A small liquid purge stream is withdrawn (28-1-P-5) to prevent trace H_2O impurity buildup in the boiling liquid NH_3. Vaporized NH_3 from the condenser is combined with cold flash gas from the separator and the liquid purge stream. Refrigeration available in this mixed stream is recovered by interchange (28-1-E-10).

Bottoms from fractionating tower 28-1-V-4 are stripped in a separate tower (28-1-V-8) which operates at about 135 psia to produce a 10 mol percent NH_3 bottoms product and a nominal 72 mol percent NH_3 overhead product. These bottoms are cooled by interchange (28-1-E-13) from 300°F to 105°F and used to reconstitute a 25 mol percent NH_3 strong liquor by reabsorption of the NH_3 vapors from 28-1-E-10. The resultant heat of absorption is removed from the system by a water cooled exchanger (28-1-E-12) that also serves as the absorber. The reconstituted strong liquor is then pumped (28-1-P-2) back to the ammonia stripper as feed after preheating in exchanger (28-1-E-13).

Overheads from the bottoms stripper (28-1-V-8) are totally condensed and pumped back (28-1-P-4) to fractionating tower 28-1-V-4 as a second strong liquor feed. Feed heater 28-1-E-11 partially vaporizes this liquor by transferring heat from the steam condensate leaving reboilers 28-1-E-8 and 28-1-E-15.

Two trains as shown on Fig. 6-3 are required for the total system.

3.3   Propylene Refrigeration Scheme

Compression of carbon dioxide in this arrangement is graphically described by the Process Flow Diagram on Fig. 6-4. Here, feed gases from the Stretford and Synthol units are combined and compressed (28-1-C-1) in three stages to 180 psia. Intercoolers (28-1-E-1 and 28-1-E-2) with water knock-outs are provided between stages to reduce work requirements. A molecular-sieve dryer (28-1-D-1) is also included between the second and third compressor stages to produce a dry gas that meets product moisture requirements.

Dry gas at 180 psia is then cooled to -20°F in a feed-effluent exchanger (28-1-E-3) which utilizes the transferred energy to superheat pressurized CO_2 to about 106°F. The CO_2 gas at -20°F is further cooled and condensed to a liquid in exchanger 28-1-E-4 which removes the necessary heat by vaporizing propylene. The resultant liquid CO_2 is separated (28-1-V-2) from non-condensible gases and pumped (28-1-P-1) to the desired supercritical pressure. Refrigeration available in the supercritical CO_2 is recovered in exchanger 28-1-E-3 as mentioned above.

Available refrigeraton in the non-condensible gases leaving
separator 28-1-V-2 is recovered by exchange (28-1-E-5) with
liquid propylene. These gases are then compressed from 166
psia to 2205 psia in a reciprocating compressor (28-1-C-2)
and combined with the main gas stream as part of the sales gas.
An intercooler (28-1-E-6) and aftercooler (28-1-E-7) are also
provided.

Six parallel propylene refrigeration trains are provided to
accommodate the design flow rate.

The propylene refrigeration system required to sustain these
compressor trains is graphically depicted by the Process Flow
Diagram on Fig. 6-5. In this arrangement, recycled propylene
gas is liquified by two stage compression (28-1-C-3) and
condensation (28-1-E-9). An interstage cooler (28-1-E-8) is
included to minimize compressor work. Liquid propylene is then
subcooled (28-1-E-5 and 28-1-E-10) prior to Joule-Thompson
expansion in order to reduce the amount of gas flashed in this
auto-refrigeration step. Low temperature flash gas is separated
(28-1-V-4) from the remaining liquid which is then vaporized
in the CO_2 condenser (28-1-E-4). Revaporized propylene from the
CO_2 condenser is combined with flash gas from the separation
and recycled to the compressor. Refrigeration available in
this mixed stream is recovered by interchange in 28-1-E-10
prior to recompression.

4.0 DISCUSSION AND CONCLUSION

The Capital Cost reported in Section 2.0 of this report does not
represent the total capital required. Tri-State's economic model
should add items such as:

- Pre-Production Costs

- Working Capital

- Construction Loan Interest

- Capital Cost Escalation

- Land

About 12 acres of land would be needed to accommodate this compressor
installation. This area is in addition to the plant site requirements
identified in the Feasibility Study.

Licensor fees and royalties will not be required to construct and
operate this compressor plant.

6

The comparative evaluation of the three processing methods considered in this study utilized Fluor's proprietary, computerized process simulation system to prepare complete heat and material balances for each alternate. This method provided a consistent and reliable basis for defining the energy requirements for each case. The magnitude of the differences identified by this method was adequate to delete the propylene refrigeration scheme from further consideration without additional engineering work.

Table 4-1 summarizes the mechanical work requirements for all cases. By inspection of this table and the appropriate process flow diagrams, it is apparent that the propylene refrigeration scheme is not competitive with straight gas compression for two reasons. First, it requires about 12½ percent more energy. Second, it is more complex mechanically and there are no obvious advantages.

Comparing the NH_3 absorption refrigeration system with straight gas compression is more difficult. The NH_3 absorption refrigeration system has a disadvantage of being more complicated mechanically but has the advantage of using less work. The latter is inherent in the NH_3 absorption refrigeration process which substitutes steam consumption in reboilers for mechanical work in a compressor. To determine if the advantages outweight the disadvantages, it is necessary to consider overall plant operations.

The NH_3 absorption refrigeration system reduces total work requirements by 39,568 HP at the expense of using 807,903 lbs/hr of 120 psig saturated steam. This steam was used to generate electrical power in the feasibility study. It also increases system cooling water requirements nearly threefold as indicated in Table 4-2.

Comparative operating requirements for these two cases are presented in Table 4-3.

The Capital Costs of the two plants on a January 1981 instantaneous cost basis with 10 percent contingency and no sales tax have been estimated as follows:

     Straight Compression Scheme $106.58 MM
     NH_3 Absorption Refrigeration Scheme $125.45 MM

Direct operating costs were estimated for both cases from the above data and the results are presented in Table 4-4. It should be noted that the "bottom line" here is dominated by the steam and power charges. The unit costs used to compute the operating costs are relative to both coal costs and the conversion costs of generating steam and power. Subsequent changes in the cost of one of these cases should produce proportionate changes in the other. Consequently, the selection of the straight gas compression route as the preferred method is expected to remain valid for reasonable changes in these unit costs.

There are no major problems associated with the compression of large quantities of CO_2 to pipeline pressure. The compressor train shown in this report probably represents the maximum desirable size.

TABLE 4-1

SUMMARY OF TOTAL

WORK INPUT (BHP)

User Equipment Number		Propylene Refrigeration	Gas Compression	NH$_3$ Absorption Refrigeration
Compression System				
28-1-C-1:	Stage 1	30,240	35,742	35,742
	Stage 2	28,344	31,326	34,590
	Stage 3	27,348	31,680	32,952
	Stage 4	-	29,550	-
28-1-C-2:	Stage 1	5,148	-	2,790
	Stage 2	4,722	-	2,478
28-1-P-1		4,380	-	4,512
	Sub-total	100,182	153,612	113,064
Refrigeration System				
28-1-C-3:	Stage 1	37,411	-	-
	Stage 2	35,293	-	-
28-1-P-2		2	-	904
28-1-P-3		-	-	3
28-1-P-4		-	-	70
28-1-P-5		-	-	3
	Sub-total	72,706	0	980
	Total	172,888	153,612	114,044

TABLE 4-2

SUMMARY OF TOTAL COOLING

WATER REQUIREMENTS (lbs/hr)

User Exchanger Number	Straight CO_2 Compression	NH_3 Absorption Refrigeration
Compression System		
28-1-E-1	5,150,880	5,150,880
28-1-E-2	3,867,600	4,252,080
28-1-E-3	3,854,160	-
28-1-E-4	4,548,240	-
28-1-E-5	8,565,600	-
28-1-E-6	-	380,880
28-1-E-7	-	686,400
Sub-total	25,986,480	10,470,240
Refrigeration System		
28-1-E-9	-	38,957,000
28-1-E-12	-	9,141,924
28-1-E-14	-	13,696,716
Sub-total	0	61,795,640
Total	25,986,480	72,265,880

TABLE 4-3

SUMMARY OF PROCESS

OPERATING REQUIREMENTS

Expended Item and Measuring Unit	Gas Compression	NH$_3$ Absorption Refrigeration
Electricity - kw	114,548	85,043
Cooling Water - GPM	51,900	144,400
Make-up Water - GPM	(112.1)	(98.3)
120 psig Steam - lbs/hr	0	807,903
Coal - MM BTU (HHV)/hr	972	972
Ethylene Glycol - gal/MMSCF	0.10	0
Operating Labor - men/shift	3	4

TABLE 4-4

SUMMARY OF DIRECT

OPERATING COSTS

Expended Item and Unit Cost	Gas Compression	NH₃ Absorption Refrigeration
Electricity @ 0.055/kwh	$51.41 MM/yr	$38.17 MM/yr
120 psig steam @ $2.45/M lbs	0	16.15
Cooling water @ $0.053/M gal	1.35	3.75
Make-up water @ $0.50/M gal	(.03)	(.02)
Coal @ $2.00/MM BTU (HHV)	9.34	9.34
Ethylene Glycol @ $3.35/gal	.05	0
Operating Labor @ $29,000/man-yr	.35	.47
Maintenance Labor @ 40% x 3.5% x capital cost	1.49	1.76
Maintenance Material @ 60% x 3.5 x capital cost	2.24	2.63

Total Direct Operating Cost: $66.20 MM/yr $72.25 MM/yr

$ 0.45/MSCF $ 0.49/MSCF

5.0 DESIGN BASIS

5.1 Battery Limit Conditions

In all cases, the CO_2 battery limit pressure was set at 2200 psia to allow for some transport pressure loss before dropping below the critical pressure for this gas. Historically, this is consistent with the 2200 psia used by Shell/Texaco/Arco in their Crossett field (West Texas) which has an injection pressure of 1700 psi.

The battery limit temperature was arbitrarily set at 100°F to minimize pipeline expansion problems.

The dew point at battery limit pressure was set at -20°F to eliminate the possibility of water condensation in the pipeline.

Specific sulfur limits were set at 50 ppm for H_2S and no limit on other sulfur compounds. Tri-State should verify the acceptability of these limits. Fluor is not aware of any legal restrictions with pipelining CO_2. Precedents are being established in other EOR operations using CO_2 gas with a higher sulfur content than that considered here. (Arco's Wesson field is an intrastate operation that will use an 89% CO_2 gas containing 5% H_2S. This compares with a total H_2S + COS content of 0.07 percent for this application).

If unforseen future events require this gas to meet natural gas sulfur requirements additional capital expenditure will be required. In this case the COS would be hydrolyzed to CO_2 and H_2S, and the H_2S selectively removed and converted to elemental sulfur. This study does not include this additional processing.

5.2 Liquefaction Conditions

In both refrigeration cases, CO_2 is condensed in a heat exchanger by transferring heat to a boiling refrigerant. The CO_2 condensing pressure is determined by the respective refrigerant evaporation temperature. Good design practice precludes operation at less than atmospheric pressure anywhere in the system in order to eliminate the risk of drawing air into the system. For this study this was accomplished by setting the system pressure for evaporating the refrigerant at nominally 18 psia. The corresponding saturation temperatures for propylene and ammonia are about -46°F and -20°F, respectively. Practical considerations in heat exchanger design translates these temperatures to CO_2 condensation temperatures of about -35°F and -10°F respectively. The corresponding pressures required for CO_2 condensation are nominally 170 psia and 305 psia.

5.3 <u>Gas Drying Methods</u>

Only in the straight gas compression scheme is it possible to achieve a gas drying pressure above 700 psia. This is the nominal lower limit for using an ethylene glycol dryer to achieve the required dew point. The benefits derived from absorption drying accrued specifically to this scheme.

6.0 FLOW DIAGRAMS

The flow diagrams depicting the three processes which were examined in this study are included here as follows:

APPENDIX 1

PROCESS DEVELOPMENT STUDY NO. 17

CO_2 COMPRESSION FOR ENHANCED OIL RECOVERY

SCOPE OF STUDY

TRI-STATE SYNFUELS COMPANY
Indirect Coal Liquefaction Plant
Western Kentucky

FLUOR ENGINEERS AND CONSTRUCTORS, INC.
Contract 835504
April 15, 1981

PROCESS STUDY

CO$_2$ COMPRESSION STUDY

1.0 GENERAL

This study will provide the capital and operation costs for drying and compressing CO$_2$ to be sold for use in enhanced oil fields recovery operations.

2.0 WORK DEFINITION

2.1 Composition of the feed gas is proprietary with the process developer.

2.2 A study will be made to determine the method for pressurizing the carbon dioxide and moving it to the plant battery limits. Two possible routes are: 1) compression, 2) liquefaction and pumping through a vaporizer to the desired pressure. For the liquefaction of CO$_2$, the use of ammonia absorption refrigeration and propylene refrigeration will be investigated.

2.3 The pipeline from battery limits to the CO$_2$ purchaser is not included in this study.

2.4 Determine the pressure, temperature and dew point requirement for the CO$_2$ stream. Condensation must be avoided to prevent excessive corrosion to carbon steel.

TRI-STATE SYNFUELS COMPANY
Indirect Coal Liquefaction Plant
Western Kentucky

FLUOR ENGINEERS AND CONSTRUCTORS, INC.
Contract 835504
April 15, 1981

PROCESS STUDY (Continued)

    2.5 Evaluate the effect of partial feed on plant operations when CO_2 is either not available or demand for CO_2 slackens.

    2.6 Estimate capital cost for the selected process.

    2.7 Estimate operating cost for the system selected.

3.0 **DELIVERABLE TO TRI-STATE**

    A formal report that contains the following:

    3.1 Capital cost estimate.

    3.2 Operating cost estimate.

    3.3 Block Flow diagrams, process description and material balances.

4.0 **SCHEDULE**

    It is estimated that the above work will be completed 10 weeks after the work is started by Fluor.

APPENDIX 2

PROCESS DEVELOPMENT STUDY NO. 17

CO_2 COMPRESSION FOR ENHANCED OIL RECOVERY

ESTIMATE DETAILS

COST ESTIMATE

CLIENT __TRI-STATE SYNFUELS__ DESCRIPTION __"CO$_2$ COMPRESSION__ PROP. NO. _____

__STUDY"__ W.O. NO. _____

LOCATION __WESTERN KENTUCKY__ __GAS COMPRESSION SCHEME__ CONT. NO. __835504__

MADE BY _____

PROJECT __TRI-STATE SYNFUELS__ __JAN 1980 INSTANTANEOUS__ APPROVED _____

A/C NO.	ITEM & DESCRIPTION	MANHOURS x 10^3	ESTIMATED COST U.S.$ x 10^3			
			LABOR	SUB-CONTRACTS	MATERIALS	TOTAL
00-00	Excavation					
10-00	Concrete					
20-00	Structural Steel					
30-00	Buildings					
40-00	Machinery & Equipment					
50-00	Piping					
60-00	Electrical					
70-00	Instruments					
80-00	Painting & Scaffolding					
85-00	Insulation					
	DIRECT FIELD COSTS	818	9 816		56 056	65 872
90-00	International Expense					
91-00	Temporary Construction Facilities					
92-00	Constr. Services, Supplies & Expense					
93-00	Field Staff, Subsistence & Expense					
94-00	Craft Benefits, Payroll Burdens & Insur.					
95-10	Equipment Rental					
95-50	Small Tools					
99-40	Field Staff Overhead Costs					
	INDIRECT FIELD COSTS					19 113
	TOTAL FIELD COSTS					84 985
96-00	Home Office Construction					
	Project Engineering					
	Process Engineering					
	Design					
	Purchasing					
	Business Services					
97-00	Office Expense					
98-00	Office Payroll Burdens					
99-50	Office Overhead Costs					
	TOTAL OFFICE COSTS					12 117
	TOTAL FIELD & OFFICE COSTS					97 102
99-30	Sales Tax					
99-10	Escalation					
99-20	Contingency					9 473
	TOTAL					
99-60	Fee					
	TOTAL PROJECT					106 575

COST ESTIMATE

CLIENT __TRI-STATE SYNFUELS__   DESCRIPTION __"CO2 COMPRESSION__   PROP. NO. _____

__STUDY"__   W.O. NO. _____

LOCATION __WESTERN KENTUCKY__   __NH₃ ABSORPTION SCHEME__   CONT. NO. __835504__

MADE BY _____

PROJECT __TRI-STATE SYNFUELS__   __JAN 1980 INSTANTANEOUS__   APPROVED _____

A/C NO.	ITEM & DESCRIPTION	MANHOURS x 10³	ESTIMATED COST U.S.$ x 10³			
			LABOR	SUB-CONTRACTS	MATERIALS	TOTAL
00-00	Excavation					
10-00	Concrete					
20-00	Structural Steel					
30-00	Buildings					
40-00	Machinery & Equipment					
50-00	Piping					
60-00	Electrical					
70-00	Instruments					
80-00	Painting & Scaffolding					
85-00	Insulation					
	DIRECT FIELD COSTS	1015	12 180		64 401	76 581
90-00	International Expense					
91-00	Temporary Construction Facilities					
92-00	Constr. Services, Supplies & Expense					
93-00	Field Staff, Subsistence & Expense					
94-00	Craft Benefits, Payroll Burdens & Insur.					
95-10	Equipment Rental					
95-50	Small Tools					
99-40	Field Staff Overhead Costs					
	INDIRECT FIELD COSTS					23 659
	TOTAL FIELD COSTS					100 240
96-00	Home Office Construction					
	Project Engineering					
	Process Engineering					
	Design					
	Purchasing					
	Business Services					
97-00	Office Expense					
98-00	Office Payroll Burdens					
99-50	Office Overhead Costs					
	TOTAL OFFICE COSTS					14 054
	TOTAL FIELD & OFFICE COSTS					114 294
99-30	Sales Tax					
99-10	Escalation					
99-20	Contingency					11 152
	TOTAL					
99-60	Fee					
	TOTAL PROJECT					125 446

PRINTED IN U

DATE __9/11/81__ REVISION NO. __1__   REVISION DATE __9/11/81__   PAGE NO. _____   FORM E-154 REV. 1/80

COST ESTIMATE

CLIENT **TRI-STATE SYNFUELS**
LOCATION **WESTERN KENTUCKY**
PROJECT **TRI-STATE SYNFUELS**

DESCRIPTION **"CO₂ COMPRESSION STUDY"**
GAS COMPRESSION SCHEME
JAN 1980 INSTANTANEOUS

PROP. NO. _____
W.O. NO. _____
CONT. NO. **835504**
MADE BY _____
APPROVED _____

A/C NO.	ITEM & DESCRIPTION	MANHOURS x 10³	ESTIMATED COST U.S.$ x 10³			
			LABOR	SUB-CONTRACTS	MATERIALS	TOTAL
00-00	Excavation					
10-00	Concrete					
20-00	Structural Steel					
30-00	Buildings					
40-00	Machinery & Equipment					
50-00	Piping					
60-00	Electrical					
70-00	Instruments					
80-00	Painting & Scaffolding					
85-00	Insulation					
	DIRECT FIELD COSTS	818	9 816		56 056	65 872
90-00	International Expense					
91-00	Temporary Construction Facilities					
92-00	Constr. Services, Supplies & Expense					
93-00	Field Staff, Subsistence & Expense					
94-00	Craft Benefits, Payroll Burdens & Insur.					
95-10	Equipment Rental					
95-50	Small Tools					
99-40	Field Staff Overhead Costs					
	INDIRECT FIELD COSTS					19 113
	TOTAL FIELD COSTS					84 985
96-00	Home Office Construction					
	Project Engineering					
	Process Engineering					
	Design					
	Purchasing					
	Business Services					
97-00	Office Expense					
98-00	Office Payroll Burdens					
99-50	Office Overhead Costs					
	TOTAL OFFICE COSTS					12 117
	TOTAL FIELD & OFFICE COSTS					97 102
99-30	Sales Tax					
99-10	Escalation					
99-20	Contingency					9 473
	TOTAL					
99-60	Fee					
	TOTAL PROJECT					106 575

DATE **9/10/81** REVISION NO _____ REVISION DATE _____ PAGE NO _____

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