SAND78-8195
Unlimited Release

An Assessment of the Use of Chemical Reaction Systems in Electric Utility Applications
Phase II Final Report

Prepared by Sandia Laboratories, Albuquerque, New Mexico 87115 and Livermore, California 94550 for the United States Department of Energy under Contract DE-AC04-76DP00789.

Printed June 1979

Sandia Laboratories
energy report

SF 2900-QQ8(3-74)

Issued by Sandia Laboratories, operated for the United States Department of Energy by Sandia Corporation.

NOTICE

This report was prepared as an account of work sponsored by the United States Government. Neither the United States nor the United States Department of Energy, nor any of their employees, nor any of their contractors, subcontractors, or their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness or usefulness of any information, apparatus, product or process disclosed, or represents that its use would not infringe privately owned rights.

Printed in the United States of America
Available from
National Technical Information Service
U. S. Department of Commerce
5285 Port Royal Road
Springfield, VA 22161
Price: Printed Copy $9.00 ; Microfiche $3.00

AN ASSESSMENT OF THE USE OF CHEMICAL REACTION SYSTEMS

IN ELECTRIC UTILITY APPLICATIONS

SAND 78-8195

Phase II Final Report, June 1979

Prepared By

GILBERT ASSOCIATES, INC.
P.O. Box 1498
Reading, Pennsylvania 19603

Gilbert Project Manager
J.T. Stewart

Principal Investigators
H.T. Chen
J.C. Hsu
W.R. Mounce
F.D. Nankani

Prepared For

SANDIA LABORATORIES
Livermore, California 94550

Under
Contract No. 18-6284

SANDIA PROJECT MANAGER
J.D. Fish

ABSTRACT

This report documents the work performed during the second phase of an assessment of the use of chemical reaction systems (CRS) as a means to store or transport thermal energy. The Phase I study, which was carried out under contract with EPRI, resulted in a catalogue of potential heat sources, end users, and a number of candidate chemical reaction systems. Four promising combinations of source/reaction/user were selected and the final result was a conceptual design and major equipment cost estimate for each system. The Phase I study is documented in EPRI RP 1086-1.

In the current Phase II study, technical and economic assessment of three CRSs were continued under contract with Sandia Laboratories. The objectives were to estimate the life cycle cost of storing or transporting waste heat via these CRSs and to compare it with the cost of heat recovery via conventional technologies. This latter objective was approached by selection and design of a commercially available alternative system for each CRS, as a basis for evaluating attractiveness of CRS.

Estimation of total capital, operating and maintenance costs, and calculation of life cycle cost were then carried out for all systems. In addition, cost sensitivities with respect to system capacity, transmission distance, and plant life were analyzed. Finally, each CRS was contrasted with its conventional alternative and was analyzed for its market potential.

It was found that none of the CRSs studied were economically competitive against a commercially available alternative system nor against fuels (natural gas and oil) at current prices. However, one of the CRSs has the potential for becoming economical in selected applications within the 1980-2000 period, should fuel costs escalate considerably above current levels. In general, CRS appears to have the best potential when coupled with a large, continuous heat source for transport applications within relatively short distances.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

LIST OF TABLES (Cont'd.)

EXECUTIVE SUMMARY

INTRODUCTION

This report documents the work done during the second phase of an assessment of the use of chemical reaction systems (CRSs) in electric utility applications. In the CRS concept, waste heat is used to drive a reversible chemical reaction in the endothermic direction. The reaction products are stored on-site or transported to another location; and when the reverse exothermic reaction is conducted at the time or location of use, thermal energy is recovered for the intended application.

In the Phase I study, various heat sources, potential users of recovered thermal energy, and a broad spectrum of candidate chemical reactions were investigated to determine the most promising source/CRS/user combinations. The screening yielded four systems, and the final result was a conceptual design and major equipment cost estimate for each CRS. It was concluded that one of the four CRSs did not warrant further study due to design and operating complexity and high cost. The three systems recommended for further technical and economic assessment in Phase II were:

CRS	Waste Heat Source	Recovered Energy Use	Mode
Methane/Syngas	Municipal Incin.	Steam	Transport
Benzene/Cyclohexane	Gas Turbine	Steam	Transport
Sulfuric Acid/Water	Fuel Cell	Hot Water	Storage

The Benzene/Cyclohexane system, however, was recommended for coupling with a higher temperature, continuous heat source to render it more efficient and economical.

OBJECTIVE AND SCOPE

The current Phase II study was undertaken to estimate the life cycle cost of storing or transporting waste heat via the three CRSs and to compare it with the cost of heat recovery via commercially available alternative systems. The study included:

1. Redesign of Benzene/Cyclohexane CRS with a more favorable waste heat source.

2. Selection and design of a commercially available alternative system for each CRS, as a basis for evaluating attractiveness of CRS.

3. Estimation of total capital, operating and maintenance costs for each CRS and alternative system. Calculation of life cycle cost for all systems ($\$/10^6$ Btu delivered energy) and analysis of its sensitivity with respect to system capacity, transmission distance, and plant life.

4. Comparison of each CRS with its commercially available alternative.

5. Estimation of market potential for CRSs between 1980 and 2000.

TECHNICAL HIGHLIGHTS (Section 2.0 and Section 4.0)

o Methane/Syngas CRS and Its Alternative System

The M/S CRS design from the Phase I study was used with only minor modifications. This system absorbs waste heat from municipal incinerator flue gas at $1700^{o}F$, transports the energy 25 miles via a pipeline, and produces 600 psia steam ($486^{o}F$) at the user end. Its overall thermal efficiency is 19%.[a]

A commercially available alternative system was designed for the same waste heat source and transmission distance, using Therminol 66 heat transfer fluid. The T-66 system transports sensible heat via an insulated pipeline

---

(a) In this study, the thermal efficiency of a system is calculated in two ways, as shown in Table 1. Efficiencies given in this summary are Definition 2 efficiencies.

and delivers the heat at a maximum temperature of 550°F. The T-66 system has a higher heat recovery efficiency of 53%; however, it can produce only hot water (486°F) or very low pressure steam (50 psia).

Compared to the commercially available system, the CRS has the advantages of being able to produce high pressure steam and of being less sensitive to transmission distance and ambient temperature. The disadvantages are the unproven status of some of its components, its design and operating complexity, its need for a very high waste heat temperature to drive endothermic reaction, and its lower thermal efficiency.

o Benzene/Cyclohexane CRS and Its Alternative System

The B/C CRS was completely redesigned to use a higher temperature and continuous waste heat source - exit gas from a cement kiln at 1300°F. The higher temperature allowed designing the exothermic reactor at the same pressure as the endothermic reactor, thus reducing equipment and power requirement for hydrogen compression. This CRS also transports thermal energy 25 miles to the user end, where 400 psig steam (445°F) is produced. The overall heat recovery efficiency of this CRS is 47%.

A commercially available alternative system was designed for the same waste heat source and transmission distance, using Therminol 60 heat transfer fluid. As with the T-66 system, the T-60 system can deliver sensible heat to the user end at temperatures high enough for hot water production, but cannot produce 400 psia steam. The overall thermal efficiency of this alternative system is 55%.

Compared to the commercially available system, the CRS has the advantage of being able to produce high pressure steam and of being less sensitive to transmission distance and ambient temperature. The disadvantages are the unproven status of some of its components (e.g., catalysts) and its design and operating complexity.

- Sulfuric Acid/Water CRS and Its Alternative System

  The SA/W CRS design from the Phase I study was used without modifications. This CRS is a storage system and the waste heat source is steam at 120 psig from a fuel cell installation. During the 16 hours of the charging period, the waste heat is used to separate 50% acid into water and 87% acid, which are stored in tanks. During the following 8 hours of the discharging period, the water and 87% acid are metered and mixed. The heat of mixing, as well as stored sensible heat, is transferred via heat exchangers to water in a commercial or residential hot water system. The heat recovery efficiency is only 23%.

  A pressurized hot water storage system (HW) was chosen as the alternative design. Using the same heat source and for the same end use application, the HW system stores water at 325^{o}F in a series of insulated tanks. The heat recovery efficiency is 96%.

  Compared to the conventional system, this CRS has only one minor advantage. It can deliver sensible heat at a slightly higher temperature (385^{o}F) than that of the waste heat source (350^{o}F). It can be considered as commercially available; however, its much lower thermal efficiency makes this CRS unattractive.

ECONOMIC ANALYSIS

Base Case Economics (Section 3.1)

For the three CRSs and three alternative systems, the total capital requirement, total operating cost, and life cycle energy cost were developed at the base case conditions (design capacity, 25 mile transmission, and 30-year plant life). The results are summarized in Table 1 and discussed below:

- Among the three CRSs considered, the Benzene/Cyclohexane CRS shows the most economic life cycle cost at $8.40 per million Btu, with the Methane/Syngas CRS being the most expensive at $30.74 per million Btu. The relatively

TABLE 1

ECONOMICS SUMMARY FOR BASE CASES

	CRS			Alternatives		
	M/S	B/C	SA/W	T-66	T-60	HW
Annual Heat Recov., 10^9 Btu	147	771	32	333	770	130
Thermal Efficiency, %[(a)]						
Definition 1	30.4	57.0	23.7	60.3	60.5	95.9
Definition 2	19.4	47.0	23.2	53.0	54.9	95.9
Transmission Distance, Miles	25	25	0	25	25	0
Capital Requirements, 10^6 $						
Plant	6.644	7.366	2.244	12.221	12.014	3.326
Transmission	13.114	15.483	0	18.323	18.323	0
Total	19.758	22.849	2.244	30.544	30.337	3.326
Total Oper. Cost (1st Yr), 10^6 $/Yr	0.8315	1.6171	0.1392	0.5549	0.7462	0.2188
Life Cycle Cost, $/$10^6$ Btu[(b)]	30.74	8.40	18.67	16.59	7.63	7.14

(a) Definition 1 = (Useful Heat x 100)/(Waste Heat + External Energy)
Definition 2 = (Useful Heat - External Energy) x 100/Waste Heat

(b) Based on a 30-year plant life, the levelized fixed charge rate is 0.1463 and the levelized O&M factor is 1.935.

NOTE: All costs are on a mid-1979 basis.

favorable economics of the B/C CRS is attributable to its high system capacity and thermal efficiency, and its continuous mode of operation. The M/S CRS operates only 16 hours a day, 6 days a week, and is much smaller in capacity.

- When each CRS is compared with its alternative, the alternative system is more economical in all cases. The life cycle costs of the Therminol 66 and Hot Water systems are 45 to 60 percent lower than the Methane/Syngas CRS and Sulfuric Acid/Water CRS, respectively. Between the B/C CRS and its alternative Therminol 60 system, the difference is much smaller. The life cycle cost of the B/C CRS is only about 10 percent higher than than of its alternative ($8.40 versus $7.63 per million Btu).

Sensitivity Analysis (Section 3.2)

Sensitivity of various costs was analyzed with respect to changes in system capacity, transmission distance, and plant life for all systems.

The effect of system capacity was analyzed at four relative capacity levels, 0.5, 1(base), 2, and 5. The effect of transmission distance was investigated at 10, 25, and 100 miles for CRSs and at 0, 10 and 25 miles for the alternatives. The effect of plant life was analyzed for 20 and 30 years.

The results of sensitivity analysis are summarized in Table 2 and discussed below:

Effect of System Capacity (Section 3.2.1)

- For each system, the life cycle cost decreases significantly with an increase in system capacity. The life cycle cost of CRSs at five times base capacity is about 40 percent lower than that at the base case capacity.

- In almost all the cases, the life cycle cost of each alternative system remains lower than the respective CRS at all capacities studied. This

TABLE 2

SENSITIVITY ANALYSIS - LIFE CYCLE COST ($\$/10^6$ BTU)

	(a) Capacity (Relative Factor)				(b) Distance (miles)				(c) Life (Yrs.)	
	0.5	1	2	5	0	10	25	100	20	30
CRS										
Methane/Syngas	41.72	30.74	23.51	16.55	-	22.33	30.74	75.69	30.47	30.74
Benz./Cyclohex.	10.82	8.40	6.70	5.18	-	6.39	8.40	20.26	8.11	8.40
S.A./Water	24.45	18.67	14.43	10.89	18.67	-	-	-	18.15	18.67
Alternatives										
T-66	25.06	16.59	12.08	8.87	2.87	8.48	16.59	-	17.01	16.59
T-60	11.56	7.63	6.34	4.90	1.86	3.97	7.63	-	7.74	7.63
Hot Water	7.85	7.14	6.25	5.70	7.14	-	-	-	6.91	7.14

Note: Underlined figures are the base case costs for each sensitivity parameter (relative capacity, transmission distance, and plant life).

All other parameters are the same as base case.

is largely due to the higher thermal efficiency of the alternative systems.

- In all of six systems studied, the capital requirement is more predominant than the operating cost in contributiong to the overall life cycle cost. The degree of predominance is more pronounced in alternative systems than in the CRSs at all capacities studied. (See Table 3 for details.)

Effect of Transmission Distance (Section 3.2.2)

- In the CRSs, the life cycle cost increases significantly with an increase in transmission distance. The increase in life cycle cost is mainly due to the added pipeline cost. The plant capital requirement as well as the operating cost is about constant. (See Table 4 for percentage breakdown by elements.)

- Over the range of distance studied, the life cycle cost is lower for the alternative systems than the CRSs. However, the increase in life cycle cost is more rapid for the alternative system. This is mainly due to the following two factors:

  1) The thermal efficiency of the alternative systems decreases more rapidly with distance because of sensible heat loss through the pipe wall.

  2) Increasing transmission distance requires a proportional increase in initial chemical inventory (i.e. Therminol heat transfer fluids) which is expensive (about $7.50 to $8.50 per gallon).

Effect of Plant Life (Section 3.2.3)

- The cost sensitivity was analyzed for two plant lives, 20 and 30 years. The plant life does not have a significant or consistent effect on life cycle cost. The two alternative transportation systems (T-66 and T-60) show a slight decrease in life cycle cost with plant life, whereas the other systems show a small increase. This is largely due to the

TABLE 3

LIFE CYCLE COST
PERCENTAGE BREAKDOWN BY ELEMENTS

		Relative System Capacity											
CRS		0.5			1.0 (Base)			2.0			5.0		
		Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total
M/S	Plant	19	31	50	22	32	54	23	34	57	30	38	68
	Trans.	46	4	50	43	3	46	41	2	43	31	1	32
	Total	65	35	100	65	35	100	64	36	100	61	39	100
B/C	Plant	14	43	57	17	46	63	20	50	70	24	57	78
	Trans.	40	3	43	35	2	37	29	1	30	21	1	22
	Total	54	46	100	52	48	100	49	51	100	45	58	100
SA/W	Plant	53	47	100	55	45	100	55	45	100	54	46	100
	Trans.	-	-	-	-	-	-	-	-	-	-	-	-
	Total	53	47	100	55	45	100	55	45	100	54	46	100
Alternatives													
T-66	Plant	22	13	35	32	17	49	37	21	58	46	25	71
	Trans.	62	3	65	48	3	51	40	2	42	28	1	29
	Total	84	16	100	80	20	100	77	23	100	74	26	100
T-60	Plant	23	17	40	30	22	52	36	25	61	45	28	73
	Trans.	56	4	60	46	2	48	38	1	39	26	1	27
	Total	79	21	100	76	24	100	74	26	100	71	29	100
HW	Plant	49	51	100	53	47	100	53	47	100	53	47	100
	Trans.	-	-	-	-	-	-	-	-	-	-	-	-
	Total	49	51	100	53	47	100	53	47	100	53	47	100

TABLE 4

LIFE CYCLE COST
PERCENTAGE BREAKDOWN BY ELEMENTS

CRS		Transmission Distance, Miles: 0.5 Cap.	0.5 Op.	0.5 Total	10 Cap.	10 Op.	10 Total	25 Cap.	25 Op.	25 Total	100 Cap.	100 Op.	100 Total
M/S	Plant	-	-	-	29	43	72	22	32	54	9	17	26
	Trans.	-	-	-	23	5	28	43	3	46	68	6	74
	Total	-	-	-	52	48	100	65	35	100	77	23	100
B/C	Plant	-	-	-	21	58	79	17	46	63	8	25	33
	Trans.	-	-	-	19	2	21	35	2	37	60	7	67
	Total	-	-	-	40	60	100	52	48	100	68	32	100
Alternatives													
T-66	Plant	64	36	100	37	25	62	32	17	49	-	-	-
	Trans.	-	-	-	34	4	38	48	3	51	-	-	-
	Total	64	36	100	71	29	100	80	20	100	-	-	-
T-60	Plant	55	45	100	36	29	65	30	22	52	-	-	-
	Trans.	-	-	-	31	4	35	46	2	48	-	-	-
	Total	55	45	100	67	33	100	76	24	100	-	-	-

counterbalancing variations in the fixed charge rate and O&M cost factor. The slight decrease in life cycle cost with increasing plant life in the T-66 and T-60 systems is due to the relatively high capital requirements and low operating costs than their corresponding CRSs.

Market Potential (Section 5.0)

In order to be commercially attractive, a CRS must meet at least two criteria. First, the cost for recovered energy must be lower than the cost of supplying energy by burning fuel. Second, the CRS must recover heat at a lower cost than would be possible with a conventionally available alternative system, or it must recover heat in a more useful form.

The current average cost for natural gas and oil are \$2.00 and \$2.46 per million Btu, respectively. When levelized over a 30 year CRS plant life using 6.0% inflation/0.2% escalation and considering 80% efficiency of fuel use, the cost of fuel would be \$5.50 - \$6.00 per million Btu.

The only CRS to be close to competitive with its alternative system is the Benzene/Cyclohexane system. However, it is not economic when compared to purchased fuel unless the fuel costs (natural gas or oil) escalate significantly over current costs. Assuming 2.4%/year fuel cost escalation in relation to other costs, then the Benzene/Cyclohexane CRS would be economic below about 20 miles.

In summary, the Benzene/Cyclohexane CRS appears to be the only one which has a possibility of economic application in the 1980 - 2000 period. It is the most economical CRS studied. However, the application would have to meet the following criteria:

1. Waste heat source is above 1000^{o}F
2. Waste heat source is essentially continuous
3. Waste heat source is greater than 150×10^{6} Btu/hr
4. Transporting distance is less than about 25 miles
5. User requires steam rather than hot water

6. Temperature inside transmission pipe is kept above 44°F (freezing point of cyclohexane).

Because of all the above qualifications, it is estimated that the market potential for the Benzene/Cyclohexane CRS is rather limited. The need for joint action between owner of waste heat source and energy user on a pioneering venture would be a deterrent to use of CRS. This means that the cost for delivered energy would have to be well below fuel costs before a project would be implemented.

CONCLUSIONS AND RECOMMENDATIONS

Specific Conclusions On CRSs Studied

The Methane/Syngas CRS is not expected to offer economical application for heat recovery and transport within the 1980 - 2000 period. Its efficiency is low and it is restricted to a very high temperature heat source.

The Benzene/Cyclohexane CRS is the most promising of those studied. But, for heat recovery and transport, it is not competitive versus fuels (natural gas or oil) at current prices nor against a commercially available alternative system. However, it can deliver high pressure steam, which the alternative system cannot do. This CRS has the potential for becoming economical in selected applications within the 1980 - 2000 period, should fuel costs escalate considerably above current levels.

The Sulfuric Acid/Water CRS was found to offer no great advantage over a pressurized hot water system for thermal storage. It can deliver heat at a temperature slightly higher than the waste heat source, which the conventional system cannot do. However, its much lower heat recovery efficiency results in higher costs than with the conventional system. It does not appear that this CRS will find use in thermal storage.

General Conclusions On CRS

The concept of CRS for thermal energy storage or transport has been shown to be technically feasible in several individual applications. It can deliver useful heat over a considerable distance. In general, CRS appears to have the best potential when coupled with a large, continuous heat source for transport at 10 to 30 mile distances.

None of the CRSs studied are economically competitive against commercially available alternative system nor against fuels (natural gas and oil) at present prices. Should fuel costs escalate at a relatively high rate within the 1980 - 2000 period, some CRS applications might prove economically competitive.

Limited availability of heat sources at a sufficiently high temperature to drive the endothermic reaction would hinder commercial application. High temperature waste heat would be recovered on-site whenever possible. Only when efficient on-site use cannot be practiced would CRS be needed. Also, working against CRS use is the requirement for a joint project between two locations - source and user. The CRS would have to be more than marginally economical before a project would be undertaken.

Recommendations

- o The Methane/Syngas CRS might be considered for future thermal transport applications at much larger capacities than in present study.

- o Since the Benzene/Cyclohexane CRS appears to have some potential for future applications, further definition of sources of waste heat available to drive endothermic reaction would be useful. Work on defining reactor designs and catalyst development might also be useful.

SECTION 1.0
INTRODUCTION

A Chemical Reaction System (CRS) utilizes a reversible chemical reaction to store or transport thermal energy. In this study, waste heat is used to drive the endothermic reaction. The reaction products are stored on site or transported to another location; and when the reverse exothermic reaction is conducted at the time or location of use, thermal energy is liberated for the intended application.

The Phase I study (EPRI RP 1086-1) included screening of a spectrum of candidate chemical reactions to select those with the most potential for useful application. In addition, various sources of waste heat and potential users of the recovered heat were studied. Four CRSs were selected and matched with a waste heat source and end use. The final result was a conceptual design and major equipment cost estimate for each CRS. The four systems were:

CRS	Waste Heat Source	Recovered Energy Use	Mode
SO_2/SO_3	Electric Furnace	Steam	Storage
Methane/Syngas	Municipal Incin.	Steam	Transport
Benzene/Cyclohexane	Gas Turbine	Steam	Transport
Sulfuric Acid/Water	Fuel Cell	Hot Water	Storage

It was concluded that the SO_2/SO_3 CRS did not warrant further study due to design and operating complexity and high cost. The Benzene/Cyclohezene system was recommended for coupling with a higher temperature, continuous heat source to render it more efficient and economical.

The current Phase II study was undertaken to estimate the life cycle cost of storing or transporting waste heat via the three CRSs and to compare the cost to heat recovery via commercially available alternative systems. The study included:

1. Redesign of Benzene/Cyclohexane CRS with a more favorable waste heat source.

2. Selection and design of a commercially available alternative system for each CRS, as a basis for evaluating attractiveness of CRS.

3. Estimation of total capital, operating and maintenance costs for each CRS and alternative system. Calculation of life cycle cost for all systems ($\$/10^6$ Btu delivered energy) and analysis of its sensitivity with respect to system capacity, transmission distance, and plant life.

4. Comparison of each CRS with its commercially available alternative.

5. Estimation of market potential for CRSs between 1980 and 2000.

SECTION 2.0
CONCEPTUAL DESIGN OF THERMAL ENERGY STORAGE/TRANSPORTATION SYSTEMS

2.1 CHEMICAL REACTION SYSTEMS

Conceptual designs and major element costs are included for the three CRSs under study. The design of the Methane/Syngas was taken from the Phase I report (2-1) with only minor modifications. The Benzene/ Cyclohexane system was completely redesigned to account for a new waste heat source. The original source was 900^{o}F exhaust gas from gas turbines used for meeting peak electric demand. The temperature was too low to allow designing the endothermic reactor at the same pressure as the exothermic reactor. Consequently, there was a large capital and operating cost for compressing hydrogen from the endothermic to the exothermic side pressure. Also the source was available only 3 hours per day, and this had a severe impact on economics. The new source, exit gas from a cement kiln, is continuous and at a higher temperature.

The description of the CRSs is in three main segments:

- o Charging Section - The heat source end.
- o Transmission Section (for transportation applications only) -Pipelines between the charging and discharging section.
- o Discharging Section - The user end.

2.1.1 Methane/Syngas CRS

In this CRS, incinerator waste heat is used to drive the endothermic steam-methane reforming reaction. The source of waste heat is flue gas, at 1700^{o}F, from a 25,000 lb/hr municipal incinerator. The source is assumed to operate 16 hours per day for six days per week (see Table 2.1-1).

The product gas (hydrogen, carbon monoxide, carbon dioxide, water vapor, unconverted methane) is transferred via pipeline to the methanation section, where hydrogen and carbon oxides undergo the exothermic methanation reaction. The heat liberated is used to generate 29,000 lb/hr of 600 psia steam for electric power generation or industrial process use. Since gas storage would be costly, the system was designed for energy transport, with no storage capability. When the heat source is operating, steam is generated at the user end.

Charging Section

As shown in the process flowsheet (Figure 2.1-1), the flue gas from the incinerator passes through the reactor R-1 outside of the tubes and transfers part of its heat to the reactants inside the tubes. The heat absorption rate is 49.8 x 10^6 Btu/hr in R-1. Because the reactor operates at high temperature, the flue gas can be cooled from 1700^{0}F to only 1100^{0}F. The flue gas then passes to waste heat boiler E-1 where more heat, 33.3 x 10^6 Btu/hr, is extracted to generate steam for the reactant mixture. Thus, the total heat absorbed from the flue gas is 83.1 x 10^6 Btu/hr. The flue gas leaves the waste heat boiler at 667^{0}F. It is cooled in an air-fin cooler, E-2, to 400^{0}F and goes to an exhaust blower C-1 needed to overcome the pressure drop in R-1 and E-1.

The methane-rich gas from the methanation section is compressed to reactor pressure, mixed with steam to bring the H_2O/CH_4 ratio to 3, and is passed through the catalyst tubes in the reactor. The catalyst is nickel on alumina. Even though the initial activity of the catalyst will be significantly reduced at the high temperature in the tubes, there will be enough activity to ensure reaching equilibrium at the outlet.

The reactor size is determined by the heat transfer rate. At the outlet conditions (1520^{0}F and 588 psia) the conversions of methane is 63%. The pressure was selected to be the same as in the methanation reactor in

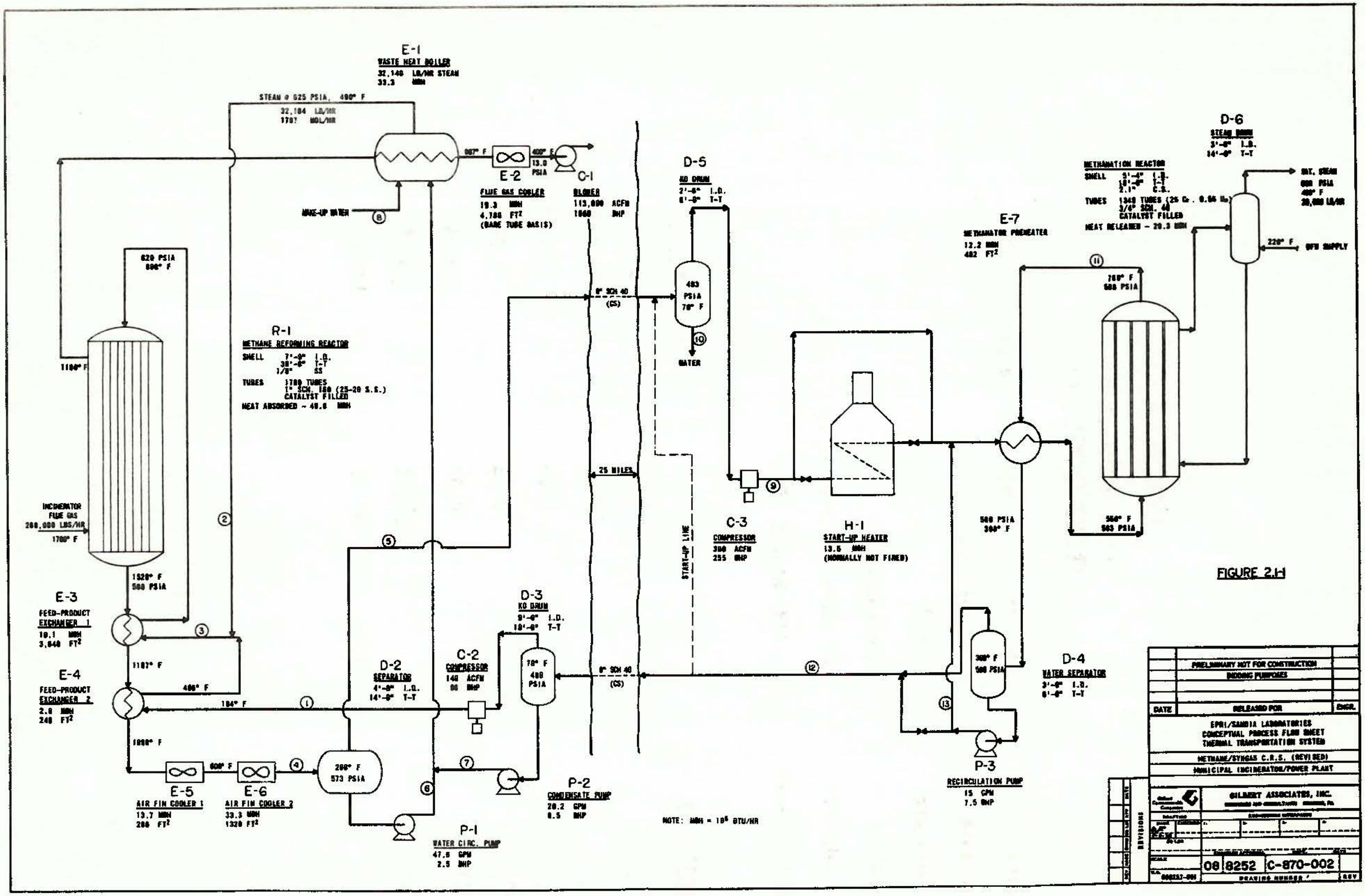
TABLE 2.1-1

HEAT SOURCE DATA

METHANE/SYNGAS CRS

Source:	Municipal Incinerator Flue Gas
Flow Rate:	267,585 lb/hr
Temperature:	1700^{o}F
Pressure:	Atmospheric
Availability:	16 hours per day; 6 days a week
Specific Heat:	0.2785 Btu/lb ^{o}F

Composition		Mole %	Wt. %
CO_2		7.54	11.60
N_2		73.02	71.42
O_2		9.60	10.74
SO_2		0.03	0.05
HCl		0.03	0.04
H_2O		9.78	6.15
	Total	100.00	100.00



E-1
WASTE HEAT BOILER
32,140 LB/HR STEAM
33.3 MBH
STEAM @ 625 PSIA, 490° F
32,184 LB/HR
1787 MOL/HR
E-2
FLUE GAS COOLER
19.3 MBH
4,700 FT²
(BARE TUBE BASIS)
C-1
BLOWER
113,000 ACFM
1060 BHP
13.0 PSIA
MAKE-UP WATER
620 PSIA
1100° F
R-1
METHANE REFORMING REACTOR
SHELL 7'-0" I.D.
38'-0" T-T
1/8" SS
TUBES 1700 TUBES
1" SCH. 160 (25-20 S.S.)
CATALYST FILLED
HEAT ABSORBED ~ 40.0 MBH
INCINERATOR FLUE GAS
280,000 LBS/HR
1700° F
1520° F
500 PSIA
E-3
FEED-PRODUCT EXCHANGER 1
10.1 MBH
3,040 FT²
E-4
FEED-PRODUCT EXCHANGER 2
2.0 MBH
240 FT²
E-5
AIR FIN COOLER 1
13.7 MBH
200 FT²
E-6
AIR FIN COOLER 2
33.3 MBH
1320 FT²
573 PSIA
D-2
SEPARATOR
4'-0" I.D.
14'-0" T-T
P-1
WATER CIRC. PUMP
47.6 GPM
2.5 BHP
C-2
COMPRESSOR
140 ACFM
D-3
KO DRUM
T-T
P-2
CONDENSATE PUMP
20.2 GPM
8" SCH 40
(CS)
25 MILES
START-UP LINE
D-5
KO DRUM
I.D.
T-T
PSIA
WATER
C-3
COMPRESSOR
300 ACFM
255 BHP
H-1
START-UP HEATER
13.5 MBH
(NORMALLY NOT FIRED)
E-7
METHANATOR PREHEATER
12.2 MBH
482 FT²
METHANATION REACTOR
SHELL
TUBES
CATALYST FILLED
HEAT RELEASED – 20.3 MBH
D-6
STEAM DRUM
I.D.
T-T
SAT. STEAM
220° F
BFW SUPPLY
D-4
WATER SEPARATOR
I.D.
T-T
P-3
RECIRCULATION PUMP
15 GPM
7.5 BHP
NOTE: MBH = 10⁶ BTU/HR
PRELIMINARY NOT FOR CONSTRUCTION
BIDDING PURPOSES
DATE
RELEASED FOR
ENGR.
EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
THERMAL TRANSPORTATION SYSTEM
METHANE/SYNGAS C.R.S. (REVISED)
MUNICIPAL INCINERATOR/POWER PLANT
GILBERT ASSOCIATES, INC.
REVISIONS
08 8252 C-870-002
DRAWING NUMBER
REV

FIGURE 2.H

INTENTIONALLY BLANK

order to minimize the power required for gas compression. However, this relatively high pressure limits the conversion. The conversion and compositions used for both reactors were taken from the General Electric report[(2-2)].

The number of tubes is set by pressure drop consideration on the tube side. The tube spacing is set by the allowable pressure drop on the shell side. The reactor length, and, hence, total tube surface is governed by the calculated heat transfer coefficient and ΔT. The relatively low overall heat transfer coefficient 11.2 Btu/hr-ft^2-^{o}F is due to the need to use low velocities on the shell side to keep the pressure drop low (0.7 psi). The coefficient was recalculated from the Phase I report (26.0 Btu/hr-ft^2-^{o}F), and this caused the reactor size to increase. The tubes are 25% Cr -20% Ni stainless steel (HK-40) to provide sufficient allowable stress and resistance to oxidative corrosion at the high tube wall temperature. The shell is 18-8 stainless steel.

The product gas at 1520^{o}F goes to two feed-product exchangers E-3 and E-4. The temperature of the product is still 1090^{o}F exiting the exchangers. While this high temperature heat might be usefully recovered in some actual cases, in the current design the gas is cooled to 200^{o}F with air in E-5 and E-6. At 200^{o}F, almost all of the water vapor in the product gas is condensed. The condensate is converted to steam and recycled to the reactor. The cooled product gas is transferred by pipeline to the methanation section.

Transmission Section

The transmission distance between the charging section (heat source) and the discharging section (end user) is assumed to be 25 miles. Two 8-inch diameter pipes are required. The product gas from the charging section is transmitted through one of the 8-inch pipes to the user end. The other 8-inch pipe is for the return of methanation product gas. Both pipes are Schedule 40 carbon steel and are buried underground. The pipe is wrapped and coated for corrosion protection.

Discharging Section

Product gas from the charging section goes first to knock-out drum D-5, where a small amount of condensate is removed before the compressor C-3. The gas is compressed to reactor pressure, mixed with recycle water, preheated with product gas, and is passed through the catalyst tubes in reactor R-2. A nickel on alumina catalyst is used for methanation. Introduction of recycle water with the feed gas is done to avoid coke formation in the catalyst tubes. Carbon oxides and hydrogen react to form methane, liberating 29.3×10^6 Btu/hr. The conversion of carbon monoxide is essentially complete at the reactor outlet conditions of 760^oF and 588 psia. As with the reforming reactor, it is assumed there would be sufficient catalytic activity to ensure that the products reach equilibrium at the reactor outlet. The reactor size is set by the heat transfer rate.

The product gas is cooled in preheater E-7 against the feed to 398^oF. At this temperature, the desired 6160 lb/hr steam is condensed for recycle to the reactor. The product gas is transferred by pipeline back to the charging section.

The component flow rates and operating conditions for the various process streams are shown in Table 2.1-2.

Heat Recovery Efficiency

Table 2.1-3 shows the overall energy balance around the system. There is 49.8×10^6 Btu/hr of heat removed from the flue gas in the reforming reactor and 33.3×10^6 Btu/hr extracted in the steam boiler, for a total of 83.1×10^6 Btu/hr.

There is an external energy requirement of 13.2×10^6 Btu/hr for pump, blower, and compressor drives. Electric power is converted to equivalent fuel value using 30% efficiency for electric power generation.

TABLE 2.1-2

METHANE/SYNGAS SYSTEM
MATERIAL BALANCE SHEET

	1		2		3		4		5		6	
	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr
CH_4	595.8	9,533	-	-	595.8	9,533	218.2	3,491	218.2	3,491	-	-
CO	-	-	-	-	-	-	216.4	6,059	216.4	6,059	-	-
CO_2	24.2	1,065	-	-	24.2	1,065	184.8	8,131	184.8	8,131	-	-
H_2	94.2	188	-	-	94.2	188	1390.0	2,780	1390.0	2,780	-	-
H_2O	.5	9	1,786.9	32,164	1787.4	32,173	1273.5	22,923	3.5	63	1,270	22,860
	714.7	10,795	1,786.9	32,164	2501.5	42,959	3282.9	43,384	2012.9	20,524	1,270	22,860
Temp., ^{o}F	104		486		486		200		200		200	
Press., psia	630		625		625		573		573		573	

	7		8		9		10		11		12	
	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr	mol/hr	lb/hr
CH_4	-	-	-	-	218.9	3,491	-	-	598.8	9,533	595.8	9,533
CO	-	-	-	-	216.4	6,059	-	-	-	-	-	-
CO_2	-	-	-	-	184.8	8,131	-	-	24.2	1,065	24.2	1,065
H_2	-	-	-	-	1390.0	2,780	-	-	94.2	188	94.2	188
H_2O	539.5	9,723	2.0	36	1.5	27	2.0	36	882.9	15,892	540.6	9,730
	539.5	9,723	2.0	36	2010.9	20,488	2.0	36	1597.1	26,678	1254.8	20,516
Temp., ^{o}F	70		70		109		70		760		398	
Press., psia	480		625		618		483		588		580	

TABLE 2.1-3

METHANE/SYNGAS CRS

HEAT RECOVERY EFFICIENCY

Product	Steam
Pressure/Temp., psia/^{o}F	600/486
Flow rate, lbs/hr	29,000
Waste Heat Absorbed, 10^6 Btu/hr	83.1
External Energy Required[(a)], 10^6 Btu/hr	
Blower (1140 H.P.)	9.7
Compressors, Pumps (408 H.P.)	3.5
	13.2
Useful Heat Recovered, 10^6 Btu/hr	30.4
Efficiency of Heat Recovery, %	
Definition 1[(b)]	30.4
Definition 2[(c)]	19.4

(a) Thermal Energy to electric energy conversion efficiency of 30% is assumed.

(b) Definition 1 = (Useful Heat x 100)/(Waste Heat + Ext. Energy)

(c) Definition 2 = (Useful Heat - Ext. Energy) x 100/(Waste Heat).

At the user end, 29.3 x 10^6 Btu/hr is recovered in the form of steam. The efficiency of the system is expressed in two forms as explained in Table 2.1-3 footnotes b and c. By definition 1, the heat recovery efficiency is about 30.4 percent while by definition 2 the efficiency is only 19.4 percent. The former value represents the recovered heat as a percent of total energy input (waste heat absorbed + external energy required). The latter value represents the recovered heat minus the external energy required as a percent of waste heat absorbed.

Methanation reactors have been operated commercially as part of hydrogen plants. In these cases, the feed gas contains only a small amount of CO and the heat release is relatively small. An adiabatic fixed bed reactor is normally used. A single adiabatic reactor would not be possible for the present case because of the very large heat release. Multiple adiabatic reactors with inter-bed heat transfer is a possibility. However, a single tubular reactor with heat transfer to boiling water was selected because of lower cost and better temperature control.

While there is some surge capacity for feed gas in the pipeline, the methanation section must be lined out shortly after the start-up of the reforming section. Gas would be circulated through the start-up heater H-1, exchanger E-7, reactor R-2, and separator D-4 and back to the suction of the compressor C-3. When the feed gas starts to flow from the reforming section, the recycle line would be closed. As the exothermic reaction proceeds, little or no firing of the heater would be needed.

Operational Life and Maintenance Requirements

Potential corrosion of the methane reforming reactor R-1 by the incinerator flue gas is a factor that might contribute to relatively high maintenance costs.

Periodic replacement of the catalyst would be required, about every two years. Poor temperature control in the methanation reactor R-2, resulting in very high temperature peaks, would necessitate more frequent replacement.

Environmental and Safety Considerations

There do not seem to be any environmental considerations that would prevent use of this chemical reaction system. However, the mixture of gases involved in the system is hazardous and requires proper design and operation to avoid fire or explosion.

Disadvantages, Limitations, Institutional Barriers

Operation of a municipal incinerator is somewhat variable. The temperature of the flue gas depends on the moisture content of the solid waste, which will vary daily depending on the weather. This will change the operation in the reforming reactor R-1.

Incinerator flue gas can contain small amount of contaminants such as chlorides which are corrosive. Tube life in the reforming reactor would be a major concern.

Temperature control in the methanation reactor R-2 is critical. A peak temperature will necessarily occur in the reactor. Changes in feed composition and operation upsets could cause peak temperaures sufficient to damage the catalyst. This occurrence would require an expensive replacement of catalyst.

Electric utilities may be reluctant to engage in chemical processing, which requires a different type of operator than in power plants.

Loss of process gas via accidental or intentional venting would require replacement, which would be done by adding makeup methane to the system.

Equipment Specifications

Descriptions of the major equipment needed to operate the Methane/Syngas system are given in Appendix D-1. Equipment specifications were not developed in complete detail, but only sufficient to allow obtaining budgetary cost estimates.

Equipment Cost

Based on the equipment specifications, the costs of the major equipment were determined. Costs were generally obtained from equipment suppliers, but where necessary, costs were estimated by the Gilbert Cost Engineering Department. Costs for this system were obtained on a third quarter 1978 basis in Phase I, but were updated to a mid-1979 basis and presented in Table 2.1-4. An eight percent inflation/escalation was used.

2.1.2 Benzene/Cyclohexane CRS

In this CRS, waste heat from the cement kiln exit is used to decompose cyclohexane into benzene and hydrogen in a catalytic reactor. The heat source data is presented in Table 2.1-5.

The benzene and hydrogen are transported in separate pipelines over a distance of 25 miles to the user end where they are recombined to regenerate the cyclohexane. The liberated heat is used to produce 400 psia steam, which is suitable for either industrial plant use or for heating boiler feed water in an electric utility plant.

This CRS was designed for thermal energy transportation since hydrogen would be impractical to store due to the huge volume generated.

TABLE 2.1-4

METHANE/SYNGAS CRS
EQUIPMENT COST

Equipment	Qty.	Total Equipment Cost: Due to Capacity, $	Total Equipment Cost: Due to Energy Storage, $	Source
Methane Reforming Reactor, R-1*	1	430,400	-	Vulcan
Steam Boiler, E-1	1	108,000	-	Trane
Flue Gas Air-Fin Cooler, E-2	1	62,600	-	Trane
Exhaust Gas Blower, C-1	1	34,600	-	Buffalo Forge
Preheater 1, E-3	1	98,300	-	Vulcan
Preheater 2, E-4	1	14,000	-	Vulcan
Air-Fin Cooler 1, E-5	1	16,200	-	Trane
Air-Fin Cooler 2, E-6	1	32,400	-	Trane
Vapor-Liquid Separator, D-2	1	7,600	-	Estimated
Water Circulation Pump, D-2	1	1,300	-	Estimated
Knock-Out Drum, D-3	1	3,200	-	RECO
Condensate Pump, D-2	1	3,200	-	Estimated
Reformer Feed Gas Compressor, C-2	1	23,900	-	Curtis
Methanation Reactor, R-2	1	141,000	-	Vulcan
Steam Drum, D-6	1	7,600	-	RECO
Methanation Preheater, E-7	1	28,100	-	Vulcan
Water Separator, D-4	1	2,300	-	RECO
Water Reciculation Pump, P-3	1	3,800	-	Estimated
Start-up Fired Heater, H-1	1	127,400	-	Zurn
Knock-Out Dru, D-5	1	1,600	-	RECO
Methanation Feed Gas Compressor, C-3	1	36,200	-	Curtis
		1,183,700	-	

Total Equipment Cost = $1,183,700 + 0 = $1,183,700

NOTE: Bare equipment costs were obtained in Phase I on a 3rd quarter 1978 basis but updated to a mid-1979 basis, using 8% inflation/escalation.

*This equipment was redesigned in Phase II, hence differs from Phase I in size and cost.

TABLE 2.1-5

HEAT SOURCE DATA
BENZENE/CYCLOHEXANE CRS

Source:	Cement Kiln Exit Gas	
Flow Rate:	679,000 lbs/hr	
Temperature:	$1300^{o}F$	
Pressure:	Atmospheric	
Availability:	Continuous	
Composition	Mole %	Wt. %
CO_2	25.6	35.8
H_2O	7.9	4.6
SO_2	0.2	0.4
O_2	1.6	1.6
N_2	64.7	57.6
Total	100.0	100.0

Charging Section

The process flowsheet is shown in Figure 2.1-2. The endothermic reactors, R-1A and B, are vertical vessels with catalyst filled tubes operated in parallel. The cement kiln exit gas passes through the shell-side of the reactors. The heat absorption rate is 134 x 10^6 Btu/hr, corresponding to the gas being cooled from 1300^{o}F to 600^oF.

The kiln exit gas flow rate corresponds to a cement plant with 1.0 million tons/year clinker production, which is representative of a large plant. It is assumed that there is an existing gas clean-up train at the cement plant which is denoted in the flowsheet by dotted lines. An additional exhaust blower would be needed to overcome the shell-side pressure drop. Power for the blower C-1 represents the major energy input.

The tubes of reactors R-1A and B are filled with catalyst pellets composed of nickel on an alumina base. Cyclohexane is pumped from surge tank TK-1 to feed-product heat exchanger E-1, where it is preheated and completely vaporized. It is split into two streams and enters the reactors, where it undergoes decomposition into benzene and hydrogen. There should be essentially no sulfur contamination in the feed gas, and the catalyst activity would remain high. As a result, the products are assumed to be in chemical equilibrium at the reactor outlet. The reactor size was determined by the heat transfer rate.

The relationship between temperature and chemical equilibrium constant for the benzene-hydrogen-cyclohexane system, used to calculate composition at a given temperature and pressure, is shown in Figure 2.1-3. The reactor outlet conditions of 820°F and 390 psia were selected as optimum for the present case. The pressure was selected to be slightly above the pressure in the exothermic reactor in order to minimize power for hydrogen compression. The 820° is easily

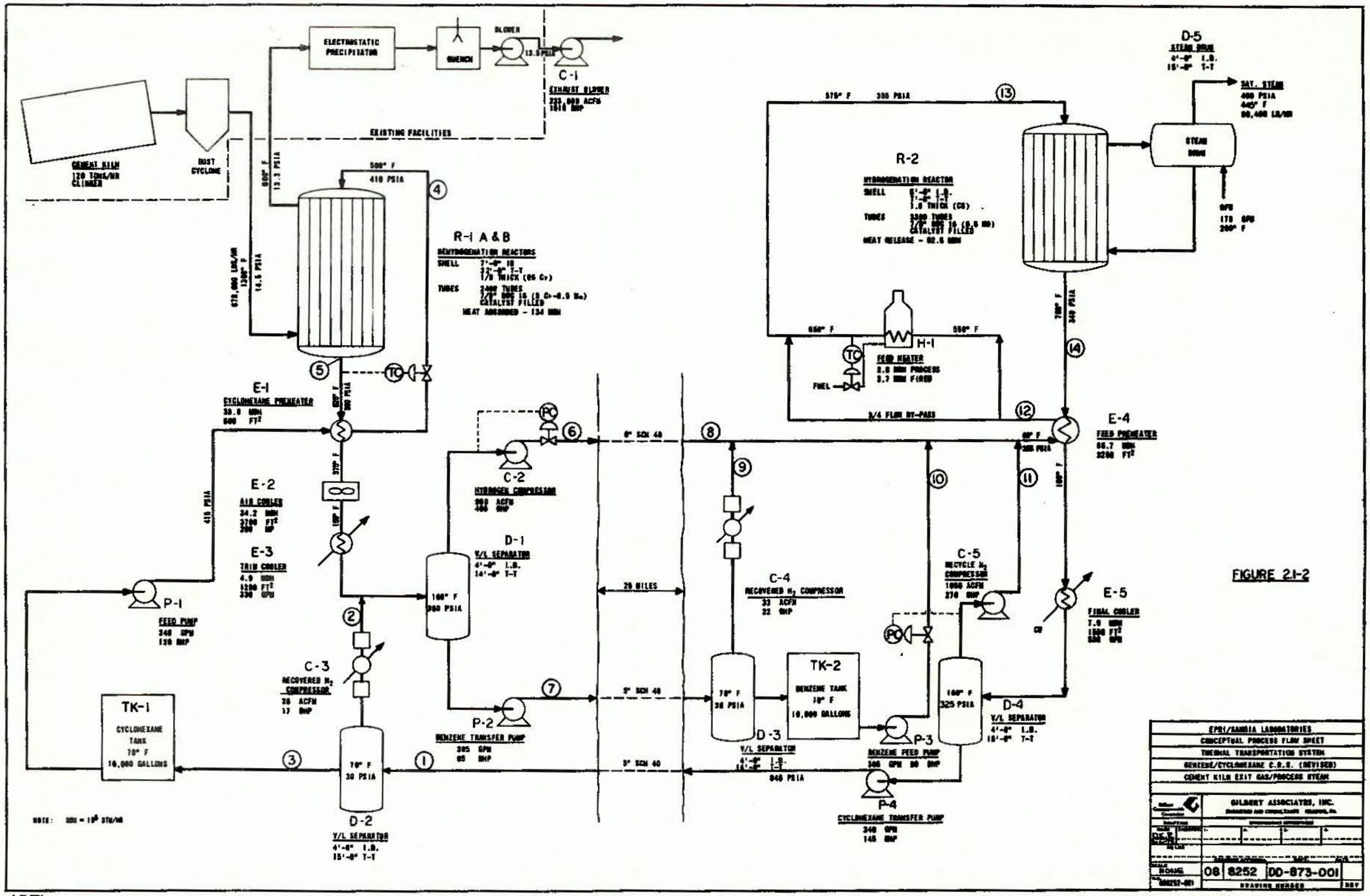

ELECTROSTATIC PRECIPITATOR
QUENCH
BLOWER
C-1
EXHAUST BLOWER
EXISTING FACILITIES
CEMENT KILN
120 TONS/HR CLINKER
DUST CYCLONE
R-1 A & B
R-2
H-1
FEED HEATER
FUEL
3/4 FLOW BY-PASS
D-5
STEAM DRUM
SAT. STEAM
E-1
CYCLOHEXANE PREHEATER
E-2
AIR COOLER
E-3
TRIM COOLER
E-4
FEED PREHEATER
E-5
FINAL COOLER
C-2
HYDROGEN COMPRESSOR
C-3
RECOVERED H_2 COMPRESSOR
C-4
RECOVERED H_2 COMPRESSOR
C-5
RECYCLE H_2 COMPRESSOR
D-1
V/L SEPARATOR
D-2
V/L SEPARATOR
D-3
V/L SEPARATOR
D-4
V/L SEPARATOR
TK-1
CYCLOHEXANE TANK
TK-2
BENZENE TANK
P-1
FEED PUMP
P-2
BENZENE TRANSFER PUMP
P-3
BENZENE FEED PUMP
P-4
CYCLOHEXANE TRANSFER PUMP
20 MILES
EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
THERMAL TRANSPORTATION SYSTEM
BENZENE/CYCLOHEXANE C.R.S. (REVISED)
CEMENT KILN EXIT GAS/PROCESS STEAM
GILBERT ASSOCIATES, INC.
NONE
08 8252 DD-873-001
DRAWING NUMBER

FIGURE 2.1-2

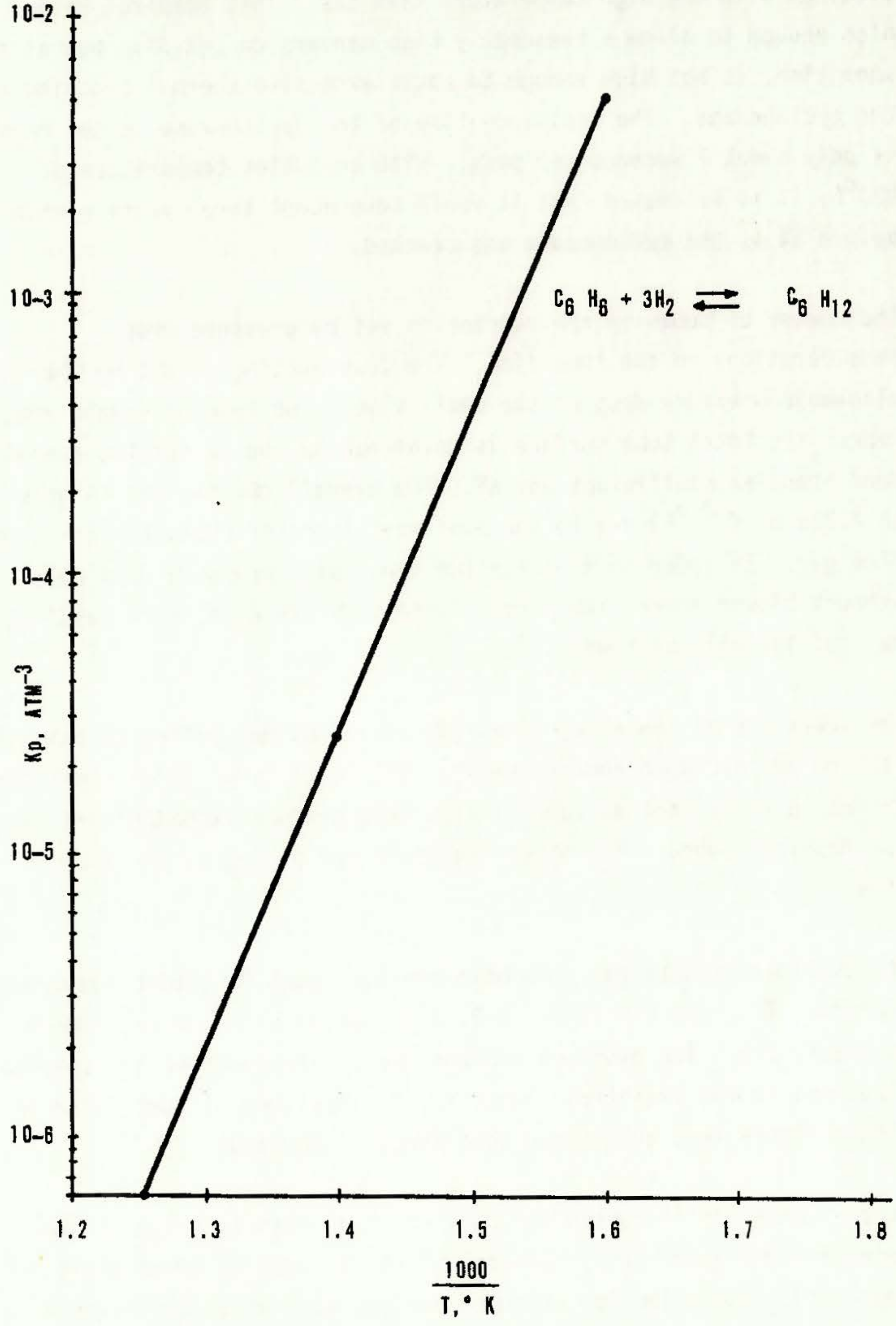


FIGURE 2.1-3
EQUILIBRIUM CONSTANT VS. TEMPERATURE - BENZENE/CYCLOHEXANE SYSTEM

obtained with the high temperature kiln gas. This temperature is high enough to allow a reasonably high conversion (82.3%); but at the same time, is not high enough to cause excessive thermal cracking of the cyclohexane. The residence time of the cyclohexane in the reactor is only about 7 seconds per pass. With an outlet temperature of 820°F, it is estimated that it would take about three years operation before 5% of the cyclohexane was cracked.

The number of tubes in the reactor is set by pressure drop considerations on the tube side. The tube spacing is set by the allowable pressure drop on the shell side. The reactor length and, hence, the total tube surface is goverened by the calculated overall heat transfer coefficient and ΔT. The overall coefficient is only 15.4 Btu/hr-ft^2-°F, due to the poor heat transfer characteristics of flue gas. In order to minimize the shell side pressure drop and exhaust blower power, two parallel reactors are used, each handling half of the kiln exit gas.

The tubes are of low alloy steel (5% Cr - 0.5% Mo) to resist oxidative scaling and hydrogen enbrittlement. The shell is of 9% Cr steel for oxidative resistance at 1300°F. The feed-product exchanger E-1 has 0.5 Mo steel tubes. All other equipment and piping is of carbon steel.

The combined product gas from both reactors goes to feed-product heat exchanger E-1, air-fin cooler E-2, trim cooler E-3, and gas-liquid separator D-1. The hydrogen and benzene are transported in separate pipelines to the exothermic reactor. The hydrogen is compressed to line pressure in a multistage centrifugal compressor, C-2.

Reactor pressure is controlled by a valve on the discharge of the hydrogen compressor, C-2. Control of reactor outlet temperature is important. Operation for extended periods much above 820°F would cause excessive thermal cracking, resulting in contamination of the liquid and hydrogen with by-products. Even short-term operation at temperatures in 850 - 900°F range could cause coke deposition on the catalyst, reducing catalyst activity. Outlet temperature is controlled separately in each reactor by a control valve in the vapor feed line between E-1 and the reactor inlet. If reactor outlet temperature should rise, the control valve will increase the cyclohexane flow to the reactor.

Cyclohexane transported from the exothermic reactor R-2 contains a small amount of hydrogen which was dissolved at high pressure in D-4. The hydrogen comes out of solution as the pressure is reduced. It is recovered with separator D-2 and compressor C-3.

Transmission Section

The transmission distance between the heat source and the user end is assumed to be 25 miles. Three pipelines are required. For the forward line two pipelines are needed, one 8-inch for hydrogen and one 5-inch for benzene. For the return line, one 5-inch pipeline is needed for cyclohexane. All pipes are Sch. 40 and are buried underground.

Discharging Section

Benzene and hydrogen produced in the deyhdrogenation section are combined with recycle hydrogen and with the recovered hydrogen from separator D-3, preheated and sent to reactor R-2.

The reactor contains catalyst filled tubes in which the exothermic benzene hydrogenation reaction occurs. The catalyst is nickel on alumina pellets. Heat is liberated at the rate of 92.6 x 10^6 Btu/hr, and 400 psia saturated steam is generated on the shell side to remove the heat of reaction.

As in the dehydrogenation reactor, the products are assumed to be in chemical equilibrium at the reactor outlet. The outlet conditions (700^{o}F and 340 psia) were selected to provide 82.3% conversion, which is the same as in the endothermic section. The required heat transfer surface area is smaller than for the endothermic reactor, because the heat transfer coefficient on the shell side is much higher for boiling water than for flue gas.

The product gas is cooled in the feed preheater E-4 and the final cooler E-5 and passes to separator D-4. The overhead vapor is mainly hydrogen and is recycled to the reactor. The liquid stream is mainly cyclohexane with some unreacted benzene.

Reactor pressure is controlled by a pressure control valve on the outlet of feed pump P-3. Controlling the feed rate of reactant from P-3 governs the amount of hydrogen reacted and, hence, the pressure in the system. Fired heater H-1 is needed for start-up and for temperature control during normal operation. The point of control would be the reactor feed temperature. The furnace was sized to handle only 1/4 of the feed mixture in order to reduce the size of the furnace tubes and the pressure drop.

Because there is no storage capacity, the hydrogenation section must be started up concurrently with the dehydrogenation section. Cyclohexane is recirculated from separator D-4 through the system, and the reactor inlet temperature is increased to 400^{o}F using the fired heater. When the benzene and hydrogen start flowing, the reactor inlet temperature is raised to 575^{o}F and the feed is changed to benzene from TK-2. A 10,000 gallon tank TK-2 is used for surge capacity and for start-up. There is some surge capacity for hydrogen in the pipeline.

The component flows and operating conditions for the various streams are given in Table 2.1-6.

TABLE 2.1-6

BENZENE/CYCLOHEXANE CRS MATERIAL BALANCE SHEET

Stream No.	1 Mols/Hr	1 Lb/Hr	2 Mols/Hr	2 Lb/Hr	3 Mols/Hr	3 Lb/Hr	4 Mols/Hr	4 Lb/Hr
C_6H_6	242	18,876	-	-	242	18,876	242	18,876
C_6H_{12}	1,370	115,080	-	-	1,370	115,080	1,370	115,080
H_2	9	18	9	18	0	0	0	0
Total	1,612	133,974	9	18	1,612	133,956	1,612	133,956
Temp., ^oF	70		305		70		500	
Press., psia	30		380		30		410	

Stream No.	5 Mols/Hr	5 Lb/Hr	6 Mols/Hr	6 Lb/Hr	7 Mols/Hr	7 Lb/Hr	8 Mols/Hr	8 Lb/Hr
C_6H_6	1,370	106,860	24	1,872	1,346	104,988	24	1,872
C_6H_{12}	242	20,328	4	336	238	19,992	4	336
H_2	3,384	6,768	3,373	6,746	11	22	3,373	6,746
Total	4,996	133,956	3,401	8,954	1,595	125,002	3,401	8,954
Temp., ^oF	820		128		100		70	
Press., psia	390		450		720		365	

TABLE 2.1-6 (Cont'd.)

BENZENE/CYCLOHEXANE CRS
MATERIAL BALANCE SHEET

Stream No.	9		10		11		12	
	Mols/Hr	Lb/Hr	Mols/Hr	Lb/Hr	Mols/Hr	Lb/Hr	Mols/Hr	Lb/Hr
C_6H_6	-	-	1,346	104,988	5	390	1,375	107,250
C_6H_{12}	-	-	238	19,992	27	1,268	269	22,596
H_2	11	22	0	0	3,384	6,768	6,768	13,536
Total	11	22	1,584	124,980	3,416	9,426	8,412	143,382
Temp., ^oF	302		70		119		550	
Press., psia	365		365		365		360	

Stream No.	13		14	
	Mols/Hr	Lb/Hr	Mols/Hr	Lb/Hr
C_6H_6	1,375	107,250	247	19,266
C_6H_{12}	269	22,596	1,397	117,348
H_2	6,768	13,536	3,384	6,768
Total	8,412	143,382	5,028	143,382
Temp., ^oF	575		700	
Press., psia	355		340	

Heat Recovery Efficiency

Table 2.1-7 shows the overall balance around the system. There is 133.7×10^6 Btu/hr extracted from the cement kiln exit gas. The external energy consumed for compressor and pump drives and fired heater amounts to 29.8×10^6 Btu/hr. Electric power is converted to equivalent fuel value using 30% efficiency for electric power generation. At the user end, 92.6×10^6 Btu/hr of useful energy is recovered in the form of steam. The efficiency of heat recovery is expressed in two forms (see footnotes b & c, Table 2.1-7). The recovered heat represents 57% of the total energy input (waste heat plus external energy). The recovered heat minus the external energy is 47% of the waste heat extracted.

Operational Life and Maintenance Requirements

Periodic replacement of the catalyst would be required. This would be about every two years. However, poor temperature control in the reactors, resulting in very high temperature peaks, would necessitate more frequent replacement.

Environmental and Safety Factors

There do not seem to be any environmental considerations that would prevent use of this chemical reaction system.

Both cyclohexane and benzene are flammable liquids and would require the normal safety precautions in design and operation used for these liquids. Hydrogen gas has a wide range of explosive limits but has been safely handled industrially for many years.

TABLE 2.1-7

BENZENE/CYCLOHEXANE CRS
HEAT RECOVERY EFFICIENCY

Product		Steam
Pressure/Temp., psia/^{o}F		400/445
Flow rate, lbs/hr		89,000
Waste Heat Absorbed, 10^6 Btu/hr		133.7
External Energy Required,[a] 10^6 Btu/hr		
Blower, C-1	(1510 BHP)	13.5
Compressor, C-2	(400 BHP)	3.7
Compressors, C-3,C-4	(39 BHP)	0.4
Compressor, C-5	(270 BHP)	2.5
Pumps	(440 BHP)	4.1
Cooler	(200 BHP)	1.9
Fired Heater		3.7
		29.8
Useful Heat Recovered		92.6
Efficiency of Heat Recovery, %		57.0
Definition 1[b]		57.0
Definition 2[c]		47.0

(a) Thermal energy to electric energy conversion of 30% assumed.

(b) Definition 1 = (Useful Heat x 100)/(Waste Heat + Ext. Energy Required).

(c) Definition 2 = (Useful Heat - Ext. Energy Required) x 100/ (Waste Heat).

System Disadvantages, Limitations, Institutional Barriers

Hydrogen gas under pressure is subject to loss from vessels and pipes by diffusion, as well as from imperfect seals. Since this is a closed system, losses of hydrogen would require replenishment from an outside source. A quantity of benzene would be removed from TK-2, and an equivalent amount of purchased cyclohexane would be added to TK-1. Decomposition of the cyclohexane would replenish the hydrogen inventory. Losses of hydrogen due to gross leakage, accidental venting, or through safety release valves would entail an economic penalty in replacement. Likewise, by-product formation would require purging, resulting in costs for chemical make-up.

Even though the cement kiln is classified as a continuous operation, the CRS would be subject to interruptions in the cement plant. It appears that the system would be most suitable when the steam produced represents only a fraction of the total user requirement.

Good coordination between the endothermic and exothermic sections is imperative. The two sections must be started up and shut down concurrently, otherwise hydrogen produced would have to be vented, since there is no hydrogen storage capacity.

The pipeline must be kept above $44^{0}F$ (the freezing point of cyclohexane). For an uninsulated, buried pipeline (present design), this would restrict the geographical area of application. An insulated, above-ground pipeline would be an option to avoid the freezing point problem.

Equipment Specifications

Description of the major equipment needed to operate the benzene/cyclohexane system is given in Appendix D-2. Equipment specifications were not developed in complete detail, but only sufficiently to allow obtaining budgetary cost estimates.

Equipment Cost

Based on the equipment specifications, the costs of the major equipment were determined and are shown in Table 2.1-8. Costs were generally obtained from equipment suppliers. When necessary, costs were estimated by the Gilbert Cost Engineering Department. Costs are on a mid-1979 basis.

The total cost for process equipment is $1.26 million. It should be noted that this cost is bare equipment cost FOB supplier, and it does not include freight, erection, piping, instrumentation, etc. For detailed capital and operating costs of this CRS, refer to Section 3.0.

2.1.3 Sulfuric Acid/Water CRS

In this conceptual design, the Sulfuric Acid/Water CRS is coupled to a Type 1 phosphoric acid fuel cell generator (thermal source) and a commercial/residential district heating system (end user). The fuel cell generator is assumed to consist of five 4.5-MW modules, producing waste heat in the form of saturated steam at 120 psig and 350^oF[2-3]. The generator is assumed to operate 16 hours per day and, during this period, the waste steam is used to concentrate 50 Wt. % sulfuric acid to 87 Wt.% solution.

During the following 8-hour discharging period, the concentrated acid solution and the water separated in the charging process are remixed to form the 50 Wt.% solution. The heat of mixing as well as stored sensible heat are recovered at a rate of 10.96×10^6 Btu/hr, which are used to heat the return water from the district heating system from 100^oF to 280^oF.

The acid/water system is designed for storage application and is assumed to be located in the proximity of the fuel cell generator. Any steam or hot water available from the fuel cell generator during discharging period could also be used directly for district heating purposes.

TABLE 2.1-8

BENZENE/CYCLOHEXANE CRS EQUIPMENT COST

Equipment	Qty.	Due to Capacity $	Due to Energy Storage $	Source
Dehydrogenation Reactor, R-1 A&B	2	465,000	-	Vulcan/GAI
Cyclohexane Storage Tank, TK-1	1	-	17,000	CB&I
Cyclohex. Feed Pump & Driver, P-1	1	5,000	-	Ingersoll-Rand
Benzene Transfer Pump, P-2	1	16,000	-	Ingersoll-Rand
Hydrogen Compressor, C-2	1	146,000	-	Ingersoll-Rand/GAI
Recov. Hydrogen Compressor, C-3	1	11,000	-	Gardner-Denver
Vapor-Liquid Separator, D-1	1	6,000	-	Reco
Vapor-Liquid Separator, D-2	1	4,000	-	Reco
Cyclohexane Preheater, E-1	1	16,000	-	Vulcan/GAI
Trim Cooler, E-3	1	21,000	-	Vulcan/GAI
Air Cooler, E-2	1	73,000	-	GAI
Hydrogenation Reactor, R-2	1	101,000	-	Vulcan/GAI
Benzene Storage Tank, TK-2	1	-	18,000	CB&I
Benzene Feed Pump & Driver, P-3	1	3,000	-	Ingersoll-Rand
Cyclohex. Transfer Pump & Driver, P-4	1	16,000	-	Ingersoll-Rand
Recovered Hydrogen Compressor, C-4	1	14,000	-	Gardner-Denver
Recycle Hydrogen Compressor, C-5	1	106,000	-	Ingersoll-Rand/GAI
Vapor-Liquid Separator, D-3	1	4,000	-	Reco
Vapor-Liquid Separator, D-4	1	6,000	-	Reco
Feed Preheater, E-4	1	42,000	-	Vulcan/GAI
Final Cooler, E-5	1	26,000	-	Vulcan/GAI
Feed Heater, H-1	1	39,000	-	Zurn/GAI
Steam Drum, D-5	1	27,000	-	GAI
Exhaust Blower, C-1	1	81,000	-	Buffalo-Forge/GAI
		1,228,000	35,000	

Total Equipment Cost = $1,228,000 + $35,000 = $1,263,000.

NOTE: All costs are on a mid-1979 basis.

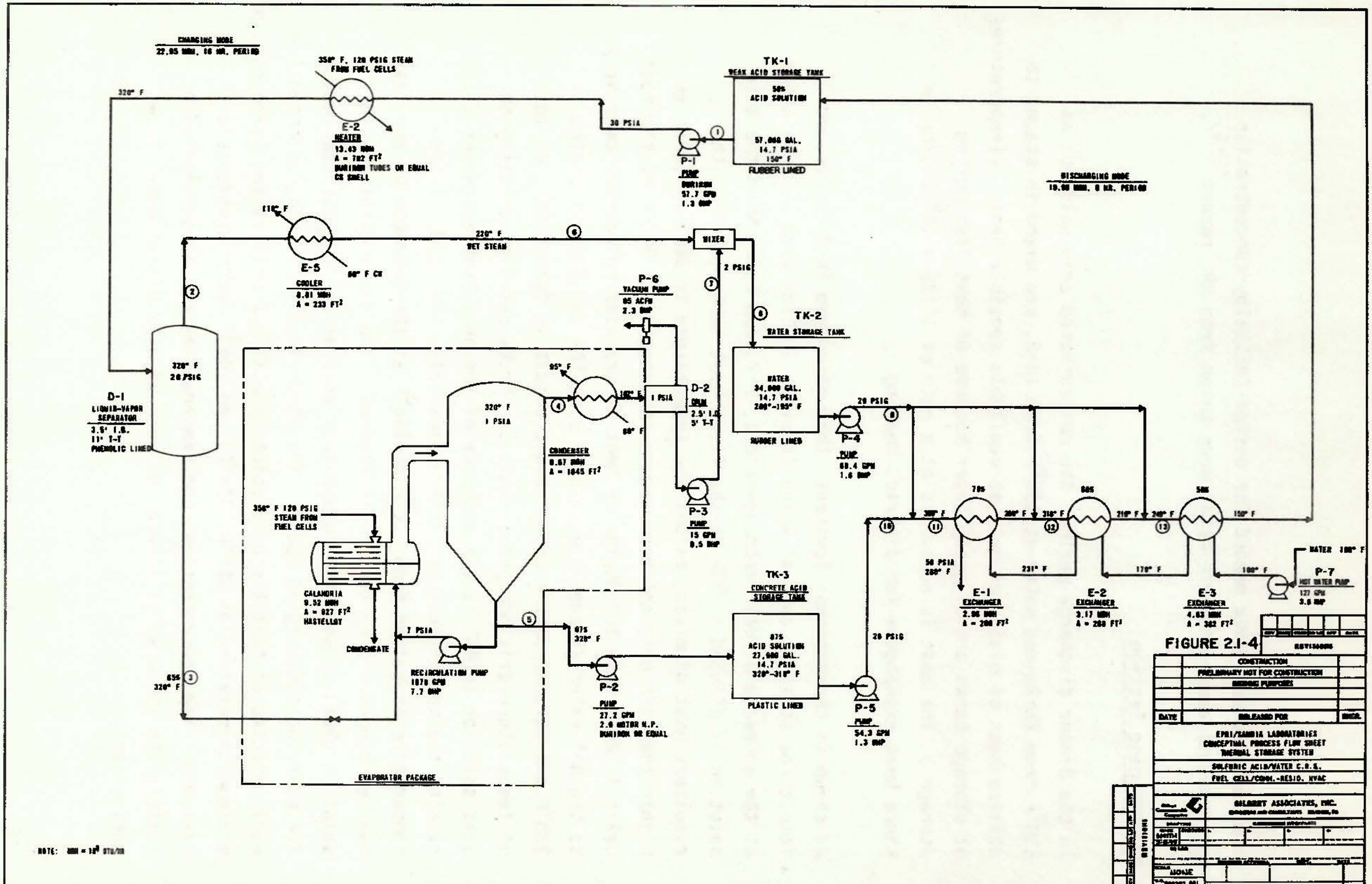
Charging Section

The charging section operates 16 hours a day. As shown in the process flowsheet, Figure 2.1-4, the 50 Wt.% acid solution from tank TK-1 at $150^{o}F$ is pumped to a shell-and-tube heat exchanger E-2 where it is heated to $320^{o}F$ with the saturated steam from the fuel cell generator. The steam condenses in the shell side at $350^{o}F$ and 120 psig, releasing heat at a rate of 13.43×10^{6} Btu/hr.

To retard acid corrosion at high temperatures, the heater tubes are made of Hastelloy and the pump is made of Duriron.

The dilute acid solution which is now partially vaporized enters a separator D-1 with baked phenolic where it is separated into a steam and a 65 Wt% acid solution. The steam goes to a cooler E-5 where it is partially condensed and subsequently mixed with a water stream recovered from the downstream evaporator. Due to the presence of small amounts of acid carried over, the water mixture is stored in a rubber-lined 34,000 gallon tank, TK-2.

The 65 Wt.% solution leaving the vapor-liquid separator, D-1 is sent to a forced-circulation evaporator equipped with submerged, horizontal heating elements. The evaporator system is constructed of Hastelloy and operates at a reduced pressure of 1 psia, the choice of which is dictated by the temperature of cooling water ($80^{o}F$) available at the condensor. The acid is concentrated from 65% to 87% H_2SO_4, using part of the 120 psig steam from the fuel cell. The heat absorption rate here is 9.52×10^{6} Btu/hr. The water vapor is condensed at $102^{o}F$ and the condensate is pumped to a mixer where it joins the wet steam coming from the cooler E-5. The 87% solution is pumped from the evaporator bottom to a 27,000 gallon storage tank, TK-3, lined with plastic.


CHARGING MODE
22.85 MMH, 16 HR. PERIOD
350° F, 120 PSIG STEAM
FROM FUEL CELLS
E-2
HEATER
13.43 MMH
A = 782 FT²
DURIRON TUBES OR EQUAL
CS SHELL
320° F
30 PSIA
TK-1
WEAK ACID STORAGE TANK
58%
ACID SOLUTION
57,000 GAL.
14.7 PSIA
150° F
RUBBER LINED
P-1
PUMP
DURIRON
57.7 GPM
1.3 BHP
DISCHARGING MODE
19.00 MMH, 8 HR. PERIOD
110° F
E-5
COOLER
8.81 MMH
A = 233 FT²
80° F CW
220° F
WET STEAM
MIXER
2 PSIG
P-6
VACUUM PUMP
85 ACFM
2.3 BHP
TK-2
WATER STORAGE TANK
WATER
34,000 GAL.
14.7 PSIA
200°-185° F
RUBBER LINED
D-1
LIQUID-VAPOR
SEPARATOR
3.5' I.D.
11' T-T
PHENOLIC LINED
320° F
20 PSIG
95° F
80° F
1 PSIA
D-2
DRUM
2.5' I.D.
5' T-T
320° F
1 PSIA
CONDENSER
8.67 MMH
A = 1845 FT²
P-3
PUMP
15 GPM
0.5 BHP
20 PSIG
P-4
PUMP
68.4 GPM
1.6 BHP
350° F 120 PSIG
STEAM FROM
FUEL CELLS
CALANDRIA
9.32 MMH
A = 827 FT²
HASTELLOY
CONDENSATE
7 PSIA
RECIRCULATION PUMP
1870 GPM
7.7 BHP
65%
320° F
67%
320° F
P-2
PUMP
27.2 GPM
2.6 MOTOR H.P.
DURIRON OR EQUAL
EVAPORATOR PACKAGE
TK-3
CONCRETE ACID
STORAGE TANK
67%
ACID SOLUTION
27,900 GAL.
14.7 PSIA
320°-310° F
PLASTIC LINED
P-5
PUMP
54.3 GPM
1.3 BHP
20 PSIG
300° F
50 PSIA
280° F
E-1
EXCHANGER
2.86 MMH
A = 200 FT²
231° F
E-2
EXCHANGER
3.17 MMH
A = 280 FT²
170° F
E-3
EXCHANGER
4.63 MMH
A = 382 FT²
150° F
100° F
WATER 100° F
P-7
HOT WATER PUMP
127 GPM
3.6 BHP
REVISIONS
CONSTRUCTION
PRELIMINARY NOT FOR CONSTRUCTION
BIDDING PURPOSES
DATE
RELEASED FOR
EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
THERMAL STORAGE SYSTEM
SULFURIC ACID/WATER C.E.S.
FUEL CELL/COMM.-RESID. HVAC
GILBERT ASSOCIATES, INC.
DRAWING NUMBER
NOTE: MMH = 10⁶ BTU/HR
FIGURE 2.1-4

The thermodynamic data needed for design (enthalpy-concentration diagram and vapor pressure data) were taken from RRC report[2-4].

Discharging Section

In the 8-hour discharge period, the concentrated acid solution at 310°F from TK-3, and water at 195°F from TK-2, are mixed in stages to utilize heat of mixing as well as available sensible heat. (Temperatures at storage tanks are slightly lower because of heat loss during storage.) The heat is recovered at a rate of 10.96 x 10^6 Btu/hr in three heat exchangers for district heating.

As shown in the process flowsheet, the water from TK-2 is divided into three streams, each of which is mixed in-line with acid solution at the entrance of heat exchangers E-1, E-2, and E-3. When the acid solution is diluted to 79% at the first heat exchanger E-1, the resultant heat of mixing raises the temperature to 385°F. (This is higher than the original heat source temperature at 350°F at the fuel cells.) As 2.96 x 10^6 Btu/hr of heat is transferred counter-currently to the hot water stream in exchanger E-1, the solution is cooled to 260°F and further heat transfer would result in "pinching", or loss of temperature-driving-force. This is circumvented by diluting the acid solution to 68% at the entrance of the next heat exchanger E-2, raising the temperature to 316°F. Similarly, the acid solution temperature is raised from 210°F to 246°F at the entrance of the last heat exchanger E-3 by the final dilution. The fully diluted acid solution (50%) is sent to storage tank TK-1 at 150°F, thus completing the charging-discharging cycle. The three levels of acid concentration were selected so that the acid solution will not boil at the corresponding maximum temperature of 385°F, 316°F, and 246°F under atmospheric pressure. The concentrations and temperatures can be controlled by putting the three water streams as well as acid stream under flow controllers.

Table 2.1-9 gives the material balance sheet for the sulfuric acid/water storage system. The corresponding thermal energy balance based on one pound of dilute acid solution (50%) is shown in Table 2.1-10. It can be seen that the net energy storage density is 144.5 Btu/lb while the net energy delivered is 139.5 Btu/lb. The difference is due to heat loss during storage.

Heat Recovery Efficiency

Table 2.1-11 shows the total energy input, useful heat recovered, and heat recovery efficiency for the sulfuric acid/water CRS. The total energy input to the system consists of 367.2 x 10^6 Btu/cycle of waste heat (steam) from fuel cells and 2.4 x 10^6 Btu/cycle of external energy requirements for pumps. Electric power is converted to equivalent thermal valve using 30% efficiency.

Useful energy recovered is 87.68 x 10^6 Btu/cycle. The efficiency of heat recovery is expressed by two definitions as shown in Table 2.1-11. Based on definition 1, the thermal efficiency is 23.7 percent while by definition 2, the efficiency is 23.2 percent. The difference between the two is negligible since the external energy requirements are a small fraction of the total energy input to this system.

Operational Life and Maintenance Requirements

Operation of this chemical reaction system should not be more complicated than that of a sulfuric acid manufacturing plant which has routinely handled concentration of sulfuric acid. Although potential acid corrosion problems exist, especially at high temperatures, maintenance of the storage system should not be excessive if proper corrosion-resistant materials are specified for major equipment such as heaters, pumps or exchangers. High silicon irons such as Duriron or high nickel irons such as Hastelloy have been widely used in

TABLE 2.1-9

SULFURIC ACID/WATER CRS MATERIAL BALANCE SHEET

Stream No.	1	2	3	4	5	6	7
Water, lb/hr	19,629	9,061	10,568	7,636	2,932	9,061	7,636
Acid	19,629	-	19,629	-	19,629	-	-
Total, lb/hr	39,258	9,061	30,197	7,636	22,561	9,061	7,636
Wt. %	50	0	65	0	87	0	0
Temp., ^{o}F	150	320	320	320	320	220	102
Press., psia	14.7	20	20	1	1	17	16.7
Enthalphy, Btu/lb.	-61	1201	5	1205	21	250.6	70

Stream No.	8	9	10	11	12	13
Water, lb/hr	16,697	33,394	5,864	10,434	18,474	39,258
Acid	-	-	39,258	39,258	39,258	39,258
Total, lb/hr	16,697	33,394	45,122	49,692	57,732	78,516
Wt. %	0	0	87	79	68	50
Temp., ^{o}F	200	195	310	385	316	246
Press., psia	14.7	35	35	34	29	22
Enthalphy, Btu/lb.	168	163	16	29.5	-3.12	0.49

TABLE 2.1-10

SULFURIC ACID/WATER CRS
THERMAL ENERGY BALANCE

Basis: one lb. of 50 wt. % acid solution

Charging Period	Btu/lb
Energy In:	
Heater	342.0
Evaporator	252.6
	584.6
Energy Out:	
Cooler	219.3
Condenser	220.8
	440.1
Net Energy Stored	144.5
Discharging Period	
Energy Discharged:	
Heat Exchanger 1, E-1	37.6
Heat Exchanger 2, E-2	40.4
Heat Exchanger 3, E-3	61.5
Net Energy Delivered	139.5
Heat Loss (During Storage)	5.0
	144.5

TABLE 2.1-11

SULFURIC ACID/WATER CRS HEAT RECOVERY EFFICIENCY

Product	District Heating Hot Water	
Pressure/Temp., psia/^{o}F	65/280	
Flowrate[(a)]	63,550 lbs/hr; 1.02 x 10^6 lbs/cycle	
Energy Balance	10^6 Btu/hr	10^6 Btu/Cycle
Waste Heat Absorbed[(b)]	22.95	367.2
External Energy Required[(c)]		
Charging Section[(b)]	0.1231	1.96
Discharging Section[(a)]	0.0552	0.44
Total	0.1783	2.40
Useful Heat Recovered[(a)]	10.96	87.68
Efficiency of Heat Recovery, %		
Definition 1[(d)]		23.7
Definition 2[(e)]		23.2

(a) Available 8 hours/day only.

(b) Available 16 hours/day only.

(c) Thermal energy to electric energy conversion efficiency of 30% is assumed.

(d) Definition 1 = (Useful Heat) x 100/(Waste Heat + Ext. Energy Required).

(e) Definition 2 = (Useful Heat - Ext. Energy Required) x 100/ (Waste Heat).

sulfuric acid service. These materials are practically unaffected by all concentrations of sulfuric acid over a wide range of temperatures. For storage tanks or vessels at relatively low temperatures, carbon steel lined with rubber or plastics should offer excellent service. A relatively high maintenance item might be the reciprocating vacuum pump P-6, which is required to maintain the evaporator system at a reduced pressure.

Environmental and Safety Factors

The acid/water system is designed for closed-loop operation and should pose no environmental problems. The only discharge from the system is at the vacuum pump which removes a very small quantity of water vapor with the air leaked into the system; however, no acid contamination in this stream is expected at the design temperature of $102^{0}F$ at the drum D-2. Normal liquid leakage at pumps or piping flanges is expected, however, and provisions must be made to neutralize or dispose of acid solution.

System Disadvantages, Limitations, Institutional Barriers

The acid/water system offers a moderately improved thermal energy storage density over conventional hot water storage systems. This is achieved at the expense of increased system complexity, and operation under vacuum. The characteristics of the heat source (fuel cells) is such that steam or hot water available from fuel cell cooling circuits might very well be utilized directly for district HVAC applications. If there is mis-match of source/user duty cycle, storage may justify CRS application.

Equipment Specifications

Specifications of major equipment required to operate the sulfuric acid/water system are presented in Appendix D-3.

Equipment Cost

Based on the equipment specifications, the costs of the major equipment were determined. Costs were generally obtained from equipment suppliers, but where necessary, costs were estimated by Gilbert Cost Engineering Department. Costs for this system were obtained on a third quarter 1978 basis in Phase I, but were updated to a mid-1979 basis and presented in Table 2.1-12. An eight percent inflation/ escalation was used.

TABLE 2.1-12

SULFURIC ACID/WATER CRS
EQUIPMENT COST

Equipment	Qty.	Total Equipment Cost: Due to Capacity, $	Total Equipment Cost: Due to Energy Storage, $	Source
Weak Acid Storage, TK-1	1	-	77,000	RECO
Water Storage, TK-2	1	-	35,300	RECO
Conc. Acid Storage, TK-3	1	-	52,400	RECO
Pump, P-1	1	1,600	-	Gould
Pump, P-2	1	5,700	-	Chem Pump
Pump, P-3	1	500	-	Gould
Pump, P-4	1	900	-	Gould
Pump, P-5	1	5,700	-	Chem Pump
Vacuum Pump, P-6	1	1,900	-	Buffalo Forge
Hot Water Pump, P-7	1	10,000	-	Gould
Heat Exchanger, E-1	1	6,400	-	Pfaudler
Heat Exchanger, E-2	1	8,600	-	Pfaudler
Heat Exchanger, E-3	1	11,800	-	Pfaudler
Heater, E-4	1	22,500	-	Pfaudler
Separator, D-1	1	6,400	-	RECO
Cooler, E-5	1	7,500	-	Pfaudler
Evaporator Package (Includes Condenser & Pump)	1	100,000	-	Estimated
Drum, D-2	1	3,200	-	RECO
		192,700	164,700	

Total Equipment Cost - $192,700 + $164,700 = $357,400

NOTE: Bare equipment costs were obtained in Phase I on a third quarter 1978 basis but updated to a mid-1979 basis, using 8% inflation/escalation.

2.2 ALTERNATIVE TECHNOLOGIES

This section describes the design of three alternative thermal energy storage/transportation systems based on currently available technologies. Each of these systems was designed as a viable alternative to one of the CRSs described in Section 2.1, utilizing the same heat source as that of the corresponding CRS. Each alternative design is described following a discussion on the selection of heat transfer medium for specific storage or transportation applications in Section 2.2.1.

2.2.1 Selection of Storage/Transportation Medium

Two of the alternative thermal energy systems involved transportation, while the third is for storage. The selection of the medium is discussed separately below.

A. Selection of Thermal Energy Transportation Medium

Due to the impracticality of pumping solids and the high cost of pumping vapor, only liquid sensible heat systems were considered. Sensible heat systems are the least complex of the various modes of thermal storage/transportation and thus place the least restrictions on the characteristics of the medium. The generally desirable properties of the medium are:

- o High specific heat capacity
- o High temperature capability
- o High density, and
- o Low vapor pressure at elevated temperatures.

Water, as liquid and vapor, has been the most common of all heat transfer fluid due to its superior properties. Above 32°F and below 350°F water is usually the automatic choice as a heat transfer fluid. Unfortunately, above 350°F steam pressures increase rapidly, causing great increases in equipment cost. For example, vapor pressure of steam at $350^{o}F$ is 135 psia; at $650^{o}F$ is 2205 psia and at its critical temperature of $705^{o}F$ the pressure is 3206 psia. Hence, above $350^{o}F$, other heat transfer mediums with low vapor pressures are preferred. Petroleum-derived heat transfer oils are commonly used between $350^{o}F$ to $650^{o}F$.

Table 2.2-1 presents some commercially available heat transfer fluids considered in this study for thermal energy transportation. Choice of a particular medium depends on the process conditions or requirements which are unique to the system. Selection of the medium for the two alternate thermal energy transportation systems is discussed below.

Alternative to Methane/Syngas CRS: The heat source for this system is the flue gas at $1700^{o}F$ from a municipal incinerator, which is available 16 hours a day and 6 days per week. The waste heat is absorbed by a heat transfer fluid and pumped 25 miles to the user end, where the heat is released. The cooled fluid is pumped back to the heat source.

The medium must stay fluid during shut-down periods and also have a low enough pour point to maintain fluidity. For the purpose of this study, it is assumed that the municipal incinerator (Charging Section) and the user of the heat (Discharging Section) are located in areas where the winter temperatures are moderate. As can be seen in Table 2.2-1, Therminol 66 and Dowtherm G are the two liquid phase oils that have the largest operating temperature range of 15^{o} to $650^{o}F$ and 12^{o} to $650^{o}F$, respectively, with a pour point of $-18^{o}F$ for both. Although both fluids are suitable for this system, Therminol 66 (T-66) was chosen over Dowtherm G due to its higher boiling point, $643^{o}F$ versus $575^{o}F$. Table 2.2-2 lists the variation of its properties with temperature.

TABLE 2.2-1

THERMAL ENERGY TRANSPORT MEDIUM

COMMERCIAL NAME	COMPOSITION	TEMPERATURE RANGE, °F	PRINCIPAL USAGE	POUR POINT, °F	BOILING POINT, °F	AUTOIGNITION TEMP, °F
Dowtherm LF	Alkylated diphenyl and diphenyl oxide	-25 to 600	Liquid	<-25	507	>1020
Dowtherm J	Alkylated aromatic	-100 to 575	Vapor, 358 to 575	-100	358	806
Dowtherm G	Di- & Tri-aryl ethers	12 to 650	Liquid	-18	575	>1030
Dowtherm A	Eutectic mixture of diphenyl & diphenyl oxide	60 to 750	Vapor 495 to 750	53.6	495	1150
Mobiltherm 600	Alkylated aromatic	-6 to 600	Liquid	0	-	-
Therminol 44	Modified ester	-50 to 425	Liquid	-85	640	705
Therminol 55	Synthetic Hydrocarbon	0 to 600	Liquid	-40	635	675
Therminol 60	Polyaromatic compounds	-60 to 600	Liquid	-90	550	835
Therminol 66	Modified Terphenyl	15 to 650	Liquid	-18	643	705
Therminol 88	Mixed Terphenyl	300 to 750	Liquid	140	687	>1000
UCAR 17	Ethylene glycol based	-40 to 275	Liquid	9	356	-

TABLE 2.2-2

BASIC PROPERTIES OF THERMINOL 66

TEMPERATURE °F	DENSITY lb/gal	DENSITY lb/ft^3	SPECIFIC HEAT Btu/lb °F	THERMAL CONDUCTIVITY Btu/hr ft. °F	VISCOSITY lb/hr. ft.	VAPOR PRESSURE mm Hg. abs.
0	8.67	64.9	0.320	0.0720	150,000	-
50	8.51	63.6	0.350	0.0711	617	-
100	8.34	62.4	0.380	0.0703	67.8	-
250	7.77	58.0	0.455	0.0678	5.86	-
300	7.59	56.8	0.480	0.0670	3.75	2.0
350	7.39	55.2	0.505	0.0662	2.57	-
400	7.17	53.6	0.530	0.0653	1.88	20.0
450	7.04	52.5	0.555	0.0645	1.40	50.0
500	6.75	50.5	0.580	0.0637	1.08	100.0
550	6.62	49.5	0.605	0.0628	0.87	200.0
600	6.42	48.1	0.630	0.0620	0.82	350.0
650	6.24	46.8	0.655	0.0613	0.65	760.0
700	6.09	45.6	0.680	0.0605	0.49	1,000

Alternative to Benzene/Cyclohexane CRS: The heat source for this system is the exit gases at $1300^{o}F$ from a cement kiln, which is operated continuously 24 hours a day and 7 days a week. The waste heat is absorbed by a heat transfer fluid and pumped 25 miles to the user end, where the heat is released. The cooled fluid is pumped back to the heat source.

For the purpose of this study, it is assumed that the cement plant and the user end are located in the northern region of the country subjected to severe weather. Hence the chosen heat transfer oil should have a very low freezing point and a high boiling point. From Table 2.2-1 three oils appear suitable for this system, namely, Dowtherm LF (-25 to $600^{o}F$), Mobiltherm 600 (-5 to $600^{o}F$) and Therminol 60 (-60 to $600^{o}F$). Of the three, Therminol 60 (T-60) was selected due to its wider range and a higher boiling point. Table 2.2-3 shows the variation of T-60 properties with temperatures.

B. Selection of Thermal Energy Storage Medium

Thermal energy can be stored by phase change, heat of solution, and sensible heat effects. Table 2.2-4 lists some of the representative mediums suitable for thermal energy storage applications.

Alternative to Sulfuric Acid/Water CRS: The heat source for this system is saturated steam at $350^{o}F$ and 120 psig, available from a fuel cell generator. The waste heat is assumed available 16 hours per day at full capacity. The end user is assumed to be in the vicinity of the fuel cell plant, and requiring heat during the 8-hour downtime of the fuel cells. Hence, waste heat is absorbed and stored during the 16-hour operating of the fuel cell and retrieved during the next 8 hours.

TABLE 2.2-3

BASIC PROPERTIES OF THERMINOL 60

TEMPERATURE °F	DENSITY lb/gal	DENSITY lb/ft³	SPECIFIC HEAT Btu/lb °F	THERMAL CONDUCTIVITY Btu/hr ft. °F	VISCOSITY lb/hr ft.	VAPOR PRESSURE mm Hg. abs.
-65	8.72	65.2	0.315	0.0797	9,746	-
-50	8.66	64.8	0.321	0.0793	2,763	-
0	8.50	63.6	0.346	0.0780	160	-
50	8.35	62.5	0.371	0.0768	32	-
100	8.20	61.3	0.395	0.0755	11.7	-
150	8.05	60.2	0.420	0.0743	6.3	-
200	7.90	59.1	0.445	0.0731	4.0	<2.0
250	7.75	57.9	0.470	0.0718	2.8	5
300	7.60	56.9	0.495	0.0705	2.1	12
350	7.43	55.6	0.518	0.0693	1.6	30
400	7.30	54.6	0.543	0.0681	1.3	65
450	7.12	53.3	0.568	0.0668	1.1	130
500	6.95	52.0	0.593	0.0656	0.93	240
550	6.80	50.9	0.618	0.0643	0.81	450
600	6.65	49.7	0.643	0.0630	0.69	760
650	6.50	48.6	0.668	0.0618	0.60	1,200

TABLE 2.2-4

THERMAL ENERGY STORAGE MATERIALS

	USABLE TEMPERATURE RANGE, °F[a]	REQUIRED PRESS., PSIG 300°F/600°F	THERMAL COND. @ 300°F, BTU/HR FT °F	DENSITY @ 300°F LB/FT³	C_p @ 300°F BTU/LB °F	ρC_p @ 300°F BTU/FT³ °F
Water	32 - 600	52/1500	0.40	57	1.0	57
Water-Ethylene Glycol[b]	-35 - 600	19/1170	0.22	62	0.91	56
Organic Liquids						
Dowtherm A	60 - 750	0/30	0.07	59	0.46	27
Dowtherm J	-100 - 575	0/160	0.07	47	0.54	25
Therminol 66	15 - 650	0/0	0.07	57	0.48	27
Molten Salts						
$KNO_3 \cdot NaNO_3 \cdot NaNO_2$	290 - 800	0/0	0.33	120	0.33	40
$NaNO_3 \cdot KNO_3$	428 - N.A.	N.A./0	0.36	117	0.36	42
$NaNO_3$	585 - N.A.	N.A./0	0.35	119	0.44	52
$NaCl \cdot CaCl_2$	928 - N.A.	N.A./0	0.59	112	0.24	27
Solids						
Rock		0/0	~1	100	0.20	20
Cast Iron		0/0	25.8	200	0.11	22

----

[a] Lower limit is freezing point of liquid. Viscoity may be excessive at this temperature.

[b] 60 vol % mixture

The product of heat capacity and density is a measure of the storage capacity of a material (Btu/^{o}F Ft^{3}). Water has the highest volumetric heat capacity. It also has good heat transfer properties (high thermal conductivity and low viscosity). Water is the preferred material for thermal storage at temperatures below 350^{o}F. A mixture of ethylene glycol and water could be used for installation in cold climates to avoid the necessity for steam or electric tracing of pipes. At temperatures higher than about 350^{o}F the vapor pressure of water becomes quite high and relatively thick wall vessels would be required.

Due to the characteristics of the heat source for this system, water was the most logical choice. This system is termed "Pressurized Hot Water System".

2.2.2 Therminol 66 System

This thermal energy transportation system is designed as a viable alternative to the Methane/Syngas chemical reaction system. Waste heat from a municipal incinerator (see Table 2.1-1) is used to heat Therminol 66 which is piped 25 miles to the user end where the heat is recovered via heat exchangers. The cooled oil is pumped back to the heat source end to complete the closed loop. At the user end, the heat is recovered either as steam or hot water depending on user requirements. (This criteria is discussed further under sub-section "Discharging End"). The flowsheet in Figure 2.2-1 shows the overall configuration of the whole system.

Charging Section

The flue gases from the incinerator, on its way to the stack, will be diverted to the Therminol 66 heater (E-1). The flue gas entering the heater at 1700^{o}F is cooled to about 440^{o}F, releasing 93.3 x 10^{6} Btu/hr of energy to the transfer oil. Therminol 66 is heated from 60^{o}F to 650^{o}F. The Therminol 66 heater is a countercurrent heat exchanger,

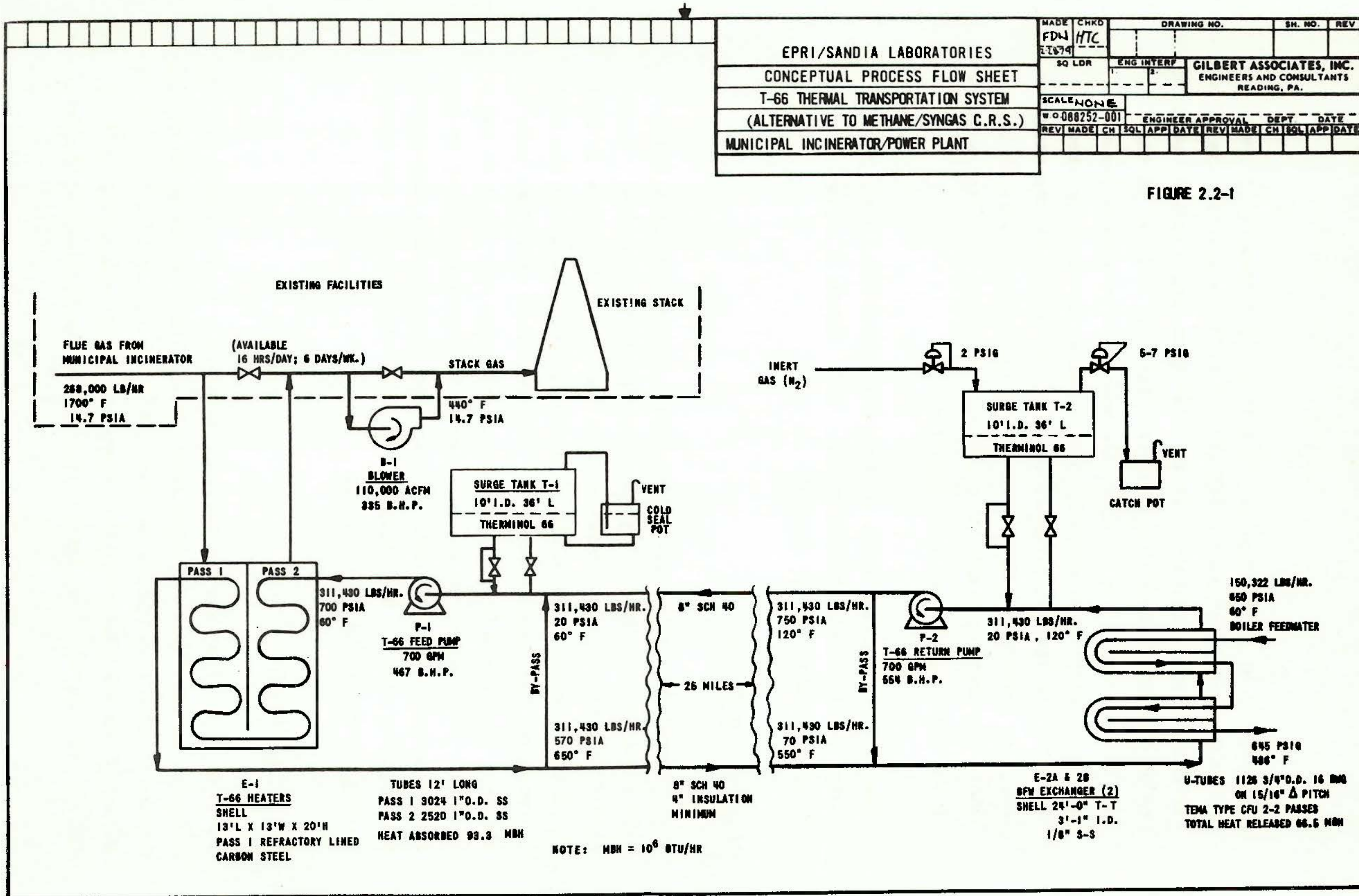

EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
T-66 THERMAL TRANSPORTATION SYSTEM
(ALTERNATIVE TO METHANE/SYNGAS C.R.S.)
MUNICIPAL INCINERATOR/POWER PLANT
MADE CHKD
FDW HTC
DRAWING NO. SH. NO. REV
SQ LDR ENG INTERF
GILBERT ASSOCIATES, INC.
ENGINEERS AND CONSULTANTS
READING, PA.
SCALE NONE
W.O. 088252-001
ENGINEER APPROVAL DEPT DATE
REV MADE CH SQL APP DATE REV MADE CH SQL APP DATE
EXISTING FACILITIES
EXISTING STACK
FLUE GAS FROM MUNICIPAL INCINERATOR
288,000 LB/HR
1700° F
14.7 PSIA
(AVAILABLE 16 HRS/DAY; 6 DAYS/WK.)
STACK GAS
440° F
14.7 PSIA
B-1
BLOWER
110,000 ACFM
335 B.H.P.
SURGE TANK T-1
10'I.D. 36' L
THERMINOL 66
VENT
COLD SEAL POT
INERT GAS (N_2)
2 PSIG
5-7 PSIG
SURGE TANK T-2
10'I.D. 36' L
THERMINOL 66
VENT
CATCH POT
PASS 1
PASS 2
311,430 LBS/HR.
700 PSIA
60° F
P-1
T-66 FEED PUMP
700 GPM
467 B.H.P.
BY-PASS
311,430 LBS/HR.
20 PSIA
60° F
8" SCH 40
25 MILES
311,430 LBS/HR.
750 PSIA
120° F
P-2
T-66 RETURN PUMP
700 GPM
554 B.H.P.
BY-PASS
311,430 LBS/HR.
20 PSIA, 120° F
150,322 LBS/HR.
650 PSIA
60° F
BOILER FEEDWATER
311,430 LBS/HR.
570 PSIA
650° F
311,430 LBS/HR.
70 PSIA
550° F
645 PSIG
496° F
E-1
T-66 HEATERS
SHELL
13'L X 13'W X 20'H
PASS 1 REFRACTORY LINED
CARBON STEEL
TUBES 12' LONG
PASS 1 3024 1"O.D. SS
PASS 2 2520 1"O.D. SS
HEAT ABSORBED 93.3 MBH
8" SCH 40
4" INSULATION
MINIMUM
NOTE: MBH = 10^6 BTU/HR
E-2A & 2B
BFW EXCHANGER (2)
SHELL 24'-0" T-T
3'-1" I.D.
1/8" S-S
U-TUBES 1126 3/4"O.D. 16 BWG
ON 15/16" Δ PITCH
TEMA TYPE CFU 2-2 PASSES
TOTAL HEAT RELEASED 66.5 MBH

FIGURE 2.2-1

with the flue gas flowing on the shellside and Therminol 66 on the tubeside. The shell has two passes while the tubeside has multiple passes. Though both shellside passes contain about equal tube-surface area, over 80 percent of the heat transfer occurs in pass 1 due to higher overall heat transfer coefficient (12.3 versus 6.0 Btu/hr ft^2 ^oF). The lower coefficient in pass 2 is primarily due to the inlet temperature condition of Therminol 66 (60^oF), and the lower log-mean-temperature-difference. Also, due to the high viscosity of Therminol 66 (see Table 2.2-2) at the inlet temperature, the flow is laminar in pass 2 and turbulent in pass 1. The high viscosity also causes high pressure drop through the tubes in pass 2. The Therminol 66 flowrate is controlled to prevent film temperatures above 700^oF where deterioration of the oil can occur. Figure 2.2-2 shows the relationship between temperature and enthalpy for this heater.

Transmission Section

The heated T-66 oil, at 650^oF, is pumped 25 miles to the user end (Discharging Section) via an insulated, aboveground pipeline. The 8-inch pipeline was designed for a Therminol 66 flowrate of 311,400 lbs/hr at an average temperature of 600^oF for the forward line and an average of 60^oF for the return line which need not be insulated. The velocity was held between 4 and 6 ft/sec for both directions. At this velocity, the transmission time is about 7 hours in the forward direction.

Due to the distance of transmission, heavy insulation will be required to minimize heat losses from the forward line. In comparing aboveground and buried pipeline transmission, it was determined that even at a depth of 6 feet below ground, the thermal loss from an uninsulated pipeline would be excessive in comparison with loss from an insulated above-ground line. In order to keep the temperature drop and the thickness of insulation within acceptable limit a 4^oF/mile drop and 4-inch thick insulation (calcium silicate) was specified. Thus, after the 25-mile transmission in an aboveground pipeline, the

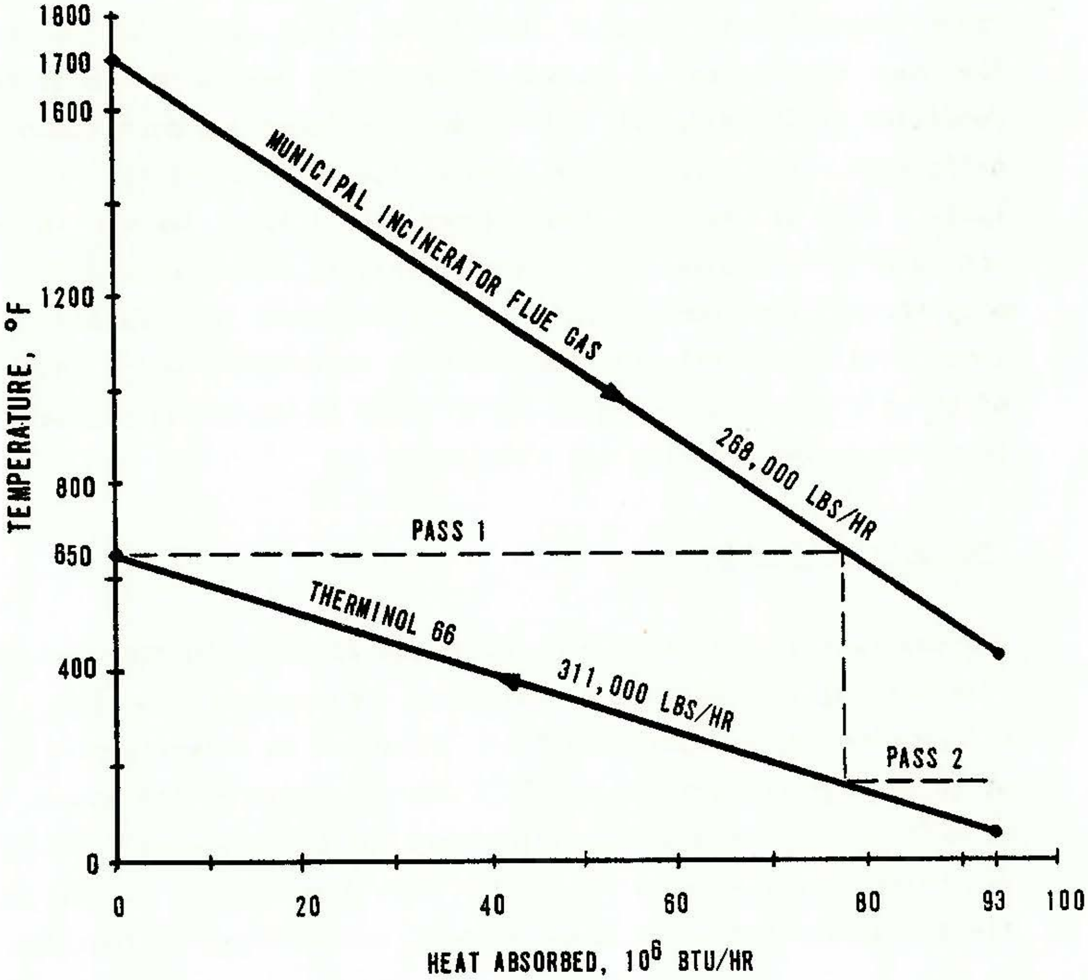


FIGURE 2.2-2
TEMPERATURE-ENTHALPY DIAGRAM
CHARGING SECTION, T-66 SYSTEM

therminol fluid would arrive at the user end at 550^{o}F. (Note: For a detailed discussion on insulation selection criteria, see Appendix A.)

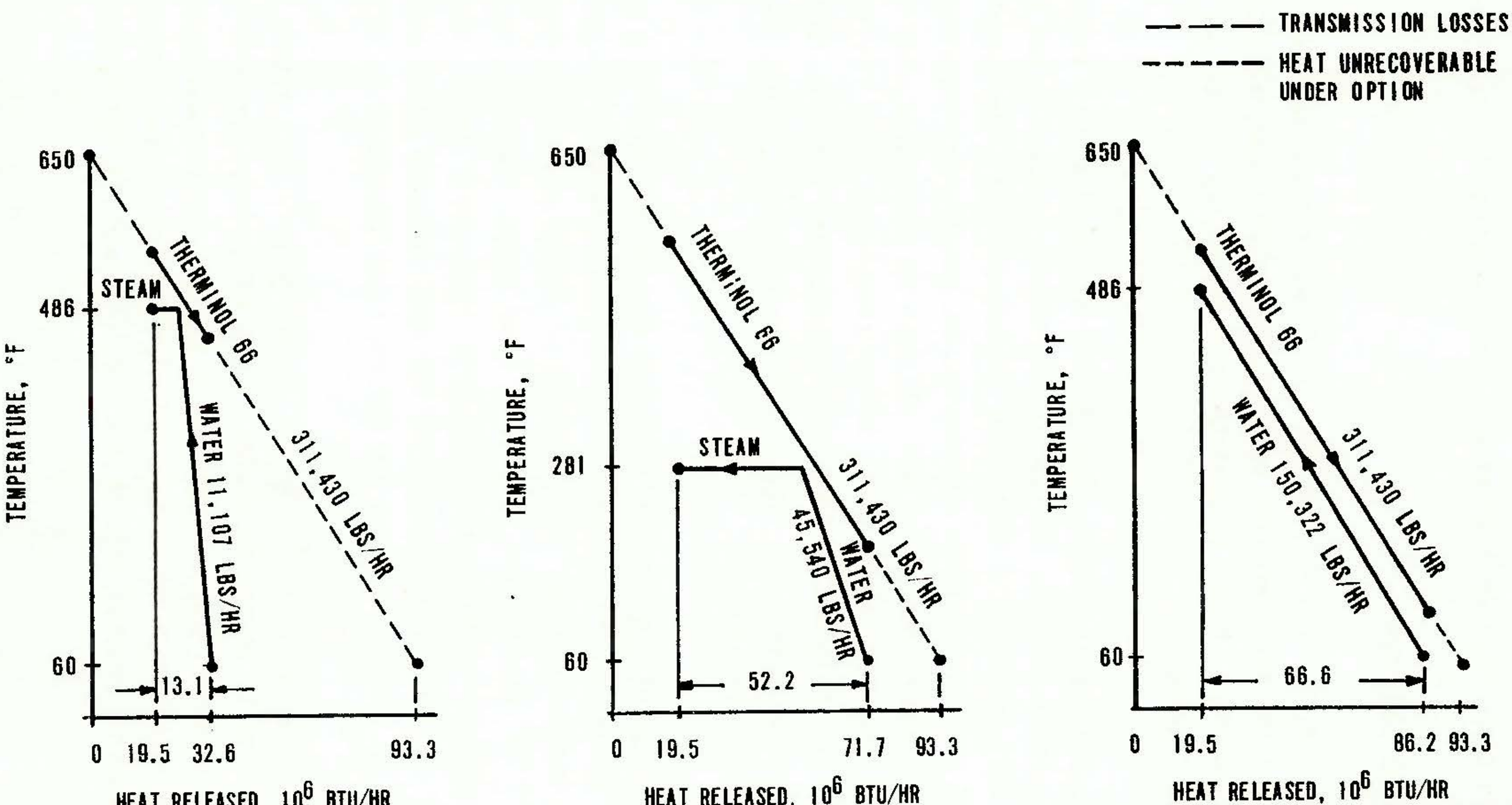
Discharging Section

The user end is assumed to be located 25 miles from the waste heat source. Three heat recovery options, described below, were analyzed; and the option with the highest thermal efficiency was chosen for conceptual design.

Option 1 - 600 psia steam: This option assumes the user requires 600 psia steam (486^{o}F), similar to the steam produced in the Methane/ Syngas CRS. Figure 2.2-3(a) shows graphically that the steam can be produced at a low flowrate of only 11,100 lbs/hr, corresponding to 13.1 x 10^6 Btu/hr heat recovery. As seen in the figure, the limiting factor is the temperature pinch at the cross-over point (in the exchangers). Column 1 of Table 2.2-5 summarizes the heat recovery efficiency for Option 1. This option is not viable, since the efficiency is less than one percent.

Option 2 - 50 psia steam: This option assumes the user requires 50 psia steam (281^{o}F). Figure 2.2-3(b) shows graphically that this product can be produced at a flowrate of 45,540 lbs/hr, corresponding to 52.2 x 10^6 Btu/hr heat recovery. Once again, the limiting factor is the approach at the temperature cross-over point in the exchanger. Column 2 of Table 2.2-5 summarizes the heat recovery efficiency for this option. The efficiency for Option 2 has increased to 42.6 percent.

Option 3 - 600 psia (486^{o}F) Saturated Hot Water: This option assumes that the user is an existing plant using 600 psia steam, and that the plant would accept the 600 psia water into its boiler and supply the required latent heat to produce the high pressure steam. Figure 2.2-3(c) shows graphically that this option will produce 150,320 lbs/hr of hot



(a) OPTION 1: 600 psi STEAM

(b) OPTION 2: 50 psi STEAM

(c) OPTION 3: 600 psi (486° F) WATER

FIGURE 2.2-3

HEAT RECOVERY OPTIONS
DISCHARGING SECTION, T-66 SYSTEM

TABLE 2.2-5

THERMINOL 66 SYSTEM
HEAT RECOVERY EFFICIENCY

	OPTION 1	OPTION 2	OPTION 3[a]
Product	Steam	Steam	Hot Water
Pressure/temperature, psia/^{o}F	600/486	50/281	600/486
Flowrate, lbs/hr (16 hrs/day)	11,100	45,540	150,320
Waste Heat Available[b], 10^6 Btu/hr	93.3	93.3	93.3
Ext. Energy Required[c], 10^6 Btu/hr			
Blower (B-1) (16 hrs/day)	3.18	3.18	3.18
Pump (P-1) (24 hrs/day)	4.24	4.24	4.24
Pump (P-2) (24 hrs/day)	1.98	1.98	5.09
Total	9.40	9.40	12.51
Useful Heat Recovered, 10^6 Btu/hr	13.1	52.2	66.6
Efficiency of Heat Recovery, %			
Definition 1[d]	12.4	49.4	60.3
Definition 2[e]	0.6	42.6	53.0

----

[a] OPTION 3 was selected for conceptual design.

[b] System operates 24 hrs/day, 6 days/wk, but heat source available 16 hrs/day, 6 days/wk.

[c] Thermal energy to electric energy conversion efficiency of 30% is assumed.

[d] Definition 1 = (Useful Heat x 100)/(Waste Heat + Ext. Energy)

[e] Definition 2 = (Useful-Ext. Energy) x 100/(Waste Heat)

water at 600 psia, corresponding to 66.6 x 10^6 Btu/hr heat recovery. Column 3 of Table 2.2-5 summarizes the heat recovery efficiency for Option 3. The efficiency is 53 percent.

Based on the heat recoveries, Option 3 was selected for the preliminary design of the discharging section. The BFW heat exchangers were then designed for a total duty of 66.6 x 10^6 Btu/hr with BFW flowing on the tubeside and Therminol 66 on the shellside. For this duty, two exchangers (E-2A and E-2B) were needed; each has 2 passes on both shellside and tubeside. The BFW is heated from 60^oF to 486^oF as the Therminol 66 cools down from 550^oF to 120^oF.

Operational Life and Maintenance

There would be no items requiring high maintenance in this system. Periodic laboratory tests, as recommended by the manufacturer of Therminol fluids (Monsanto), should be made on Therminol fluid to determine its condition. Periodic maintenance on the circulating pumps (P-1 & P-2) and exchangers (E-1 & E-2) will be needed.

Environmental and Safety Factors

Therminol fluids are virtually non-toxic and non-irritating, posing no special environmental problems. However, organic heat transfer fluids such as Therminol may exhibit a slow oxidation reaction with the air trapped inside the voids of the insulating material when system temperatures reach about 500^oF. Saturated insulation offers a large fuel surface in the face of poor heat dissipation conditions, and this, along with possible catalysis from the insulating material (magnesia, silicate-bonded asbestos or calcium silicate), can cause a temperature build-up in the mass. This temperature build-up can result in ignition of the fluid when the space between the piping and the saturated insulation is exposed to air (i.e., should the insulation be broken for repair, etc.)

System Disadvantages, Limitations, Institutional Barriers

The system design is based on commercially available current technology and hence no major limitations or barriers are anticipated. The limitations percieved at this time is the transfer fluid's upper temperature limit, and the limited transmission distance. Based on the present heat source and the design of pipeline, there is a $4^{O}F$/mile temperature drop. Hence, longer transmission distance (e.g., 100 miles) causes high total temperature drop and lower thermal energy available at the user end.

Equipment Specifications and Cost

Equipment specifications were not developed in complete detail, but sufficiently to obtain manufacturers estimate. Major equipment specifications are given in Appendix D-4. Table 2.2-6 presents manufacturer's bare equipment cost.

2.2.3 Therminol 60 System

This is an alternative system to the Benzene/Cyclohexane CRS. The heat source is the exit gases from a cement kiln previously described in Table 2.1-5. The hot gases heat Therminol 60 which is piped 25 miles to the user end where the heat is recovered via heat exchangers. The cooled oil is pumped back to the heat source to complete the closed loop. At the user end the heat is recovered either as steam or hot water, depending on user requirements. This is discussed further under "Discharging Section".

The flowsheet shown in Figure 2.2-4 gives the overall configuration of the whole system.

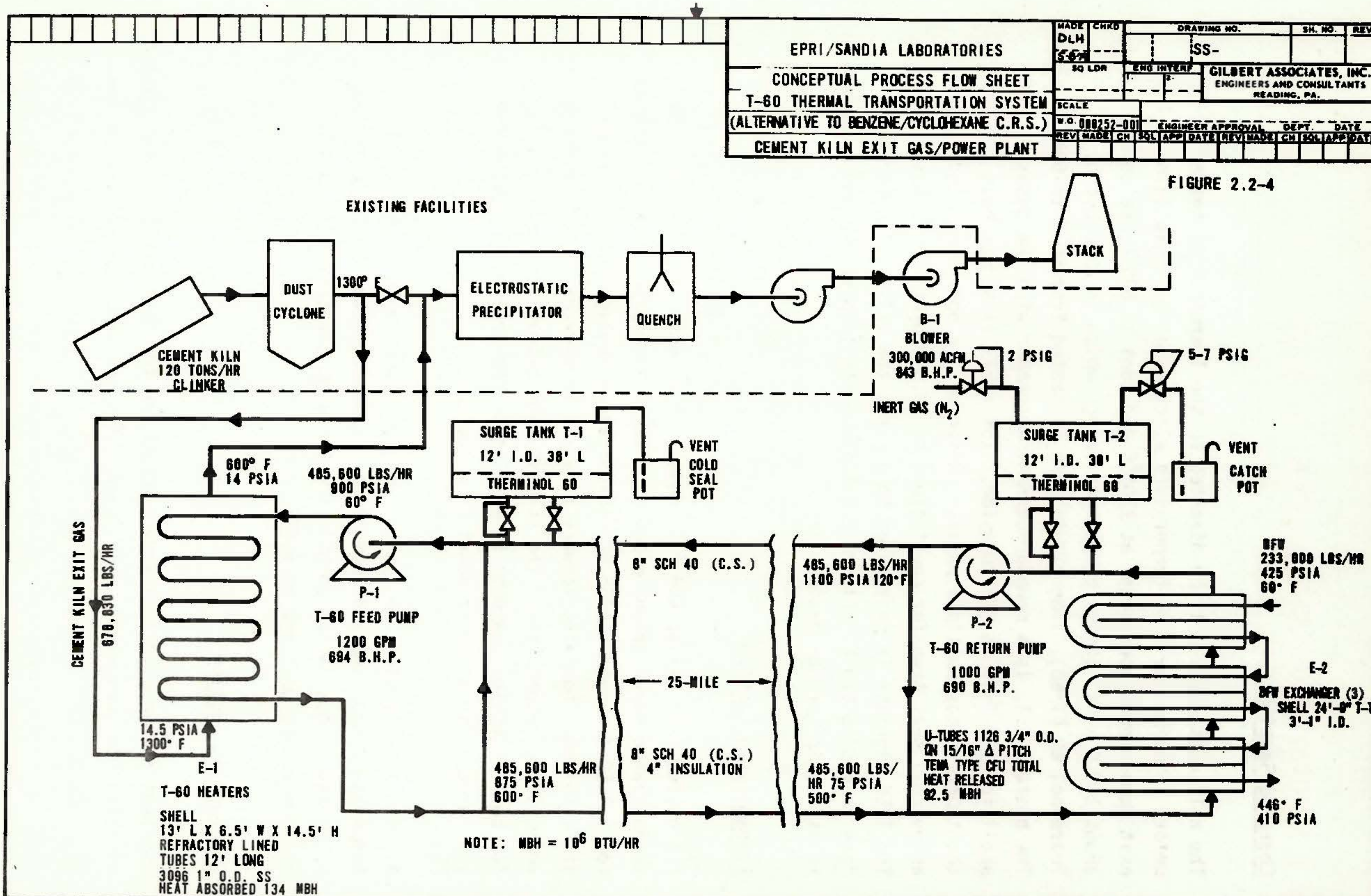
TABLE 2.2-6

THERMINOL 66 SYSTEM
EQUIPMENT COST

Equipment	Qty.	Due to Capacity $	Due to Energy Storage $	Source
Exhaust Blower, B-1	1	48,000	-	Buffalo Forge
Therminol Heater, E-1	1	375,000	-	Zurn
BFW Heat Exchangers, E-2	2	350,000	-	Goulds
Feed Pump, P-1	1	37,000	-	RECO
Return Pump, P-2	1	37,000	-	Goulds
Therminol Surge Tank, T-1	2	-	70,000	YUBA
Therminol Surge Tank, T-2	2	-	70,000	RECO
		847,000	140,000	

Total Equipment Cost = $847,000 + $140,000 = $987,000

NOTE: All costs are on a mid-1979 basis.


EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
T-60 THERMAL TRANSPORTATION SYSTEM
(ALTERNATIVE TO BENZENE/CYCLOHEXANE C.R.S.)
CEMENT KILN EXIT GAS/POWER PLANT
MADE DLH
CHKD
DRAWING NO. SS-
SH. NO.
REV
SQ LDR
ENG INTERF
GILBERT ASSOCIATES, INC.
ENGINEERS AND CONSULTANTS
READING, PA.
SCALE
W.O. 088252-001
ENGINEER APPROVAL DEPT. DATE
REV MADE CH SQL APP DATE REV MADE CH SQL APP DATE
FIGURE 2.2-4
EXISTING FACILITIES
CEMENT KILN 120 TONS/HR CLINKER
DUST CYCLONE
1300° F
ELECTROSTATIC PRECIPITATOR
QUENCH
STACK
B-1 BLOWER 300,000 ACFM 843 B.H.P.
2 PSIG
5-7 PSIG
INERT GAS (N_2)
SURGE TANK T-1 12' I.D. 38' L THERMINOL 60
VENT COLD SEAL POT
SURGE TANK T-2 12' I.D. 38' L THERMINOL 60
VENT CATCH POT
600° F 14 PSIA
485,600 LBS/HR 900 PSIA 60° F
CEMENT KILN EXIT GAS 878,830 LBS/HR
P-1 T-60 FEED PUMP 1200 GPM 694 B.H.P.
14.5 PSIA 1300° F
E-1 T-60 HEATERS
SHELL 13' L X 6.5' W X 14.5' H REFRACTORY LINED
TUBES 12' LONG 3096 1" O.D. SS
HEAT ABSORBED 134 MBH
8" SCH 40 (C.S.)
485,600 LBS/HR 1100 PSIA 120° F
25-MILE
8" SCH 40 (C.S.) 4" INSULATION
485,600 LBS/HR 875 PSIA 600° F
485,600 LBS/HR 75 PSIA 500° F
P-2 T-60 RETURN PUMP 1000 GPM 690 B.H.P.
U-TUBES 1126 3/4" O.D. ON 15/16" Δ PITCH TEMA TYPE CFU TOTAL HEAT RELEASED 92.5 MBH
BFW 233,000 LBS/HR 425 PSIA 60° F
E-2 BFW EXCHANGER (3) SHELL 24'-0" T-T 3'-1" I.D.
446° F 410 PSIA
NOTE: MBH = 10^6 BTU/HR

Charging Section

The kiln exit gases will be diverted to the Therminol 60 heater (E-1) before entering the high temperature electrostatic precipitator. The exit gases enter the heater at 1300^{o}F and cool to 600^{o}F, releasing about 134.0 x 10^{6} Btu/hr of thermal energy which is absorbed by Therminol 60 (T-60). The oil in turn is heated from 60^{o}F to 600^{o}F. The heater, E-1, is a counter-current exchanger with the gases on the shellside and T-60 on the tubeside. The shell is single pass while the tubeside has multiple passes. Based on the operating conditions an overall heat transfer coefficient of 22.4 Btu/Ft^{2} Hr ^{o}F was used. The T-60 flowrate is controlled to prevent the oil film temperature from exceeding 650^{o}F. Figure 2.2-5 shows the temperature and enthalpy relationship for the heater, E-1.

Transmission Section

The heated T-60 oil at 600^{o}F is pumped 25 miles to the user end via an insulated above ground pipeline. The 8-inch pipeline was designed for a T-60 flowrate of 485,600 lbs/hr at an average temperature of 550^{o}F for the insulated forward line and an average of 60^{o}F for the uninsulated return line. The velocity was between 6 and 7.5 ft/sec for both directions, and at this velocity the transmission time for the forward direction is about five hours.

As with the Therminol 66 system, an aboveground pipe, a 4^{o}F/mile temperature drop and a 4-inch thick insulation (calcium silicate) was specified. Thus, after 25-mile transmission the therminol fluid arrives at the user end at 500^{o}F.

Discharging Section

Two heat recovery options were studied and the most efficient one, based on total heat recovered, was chosen. The options are described below:

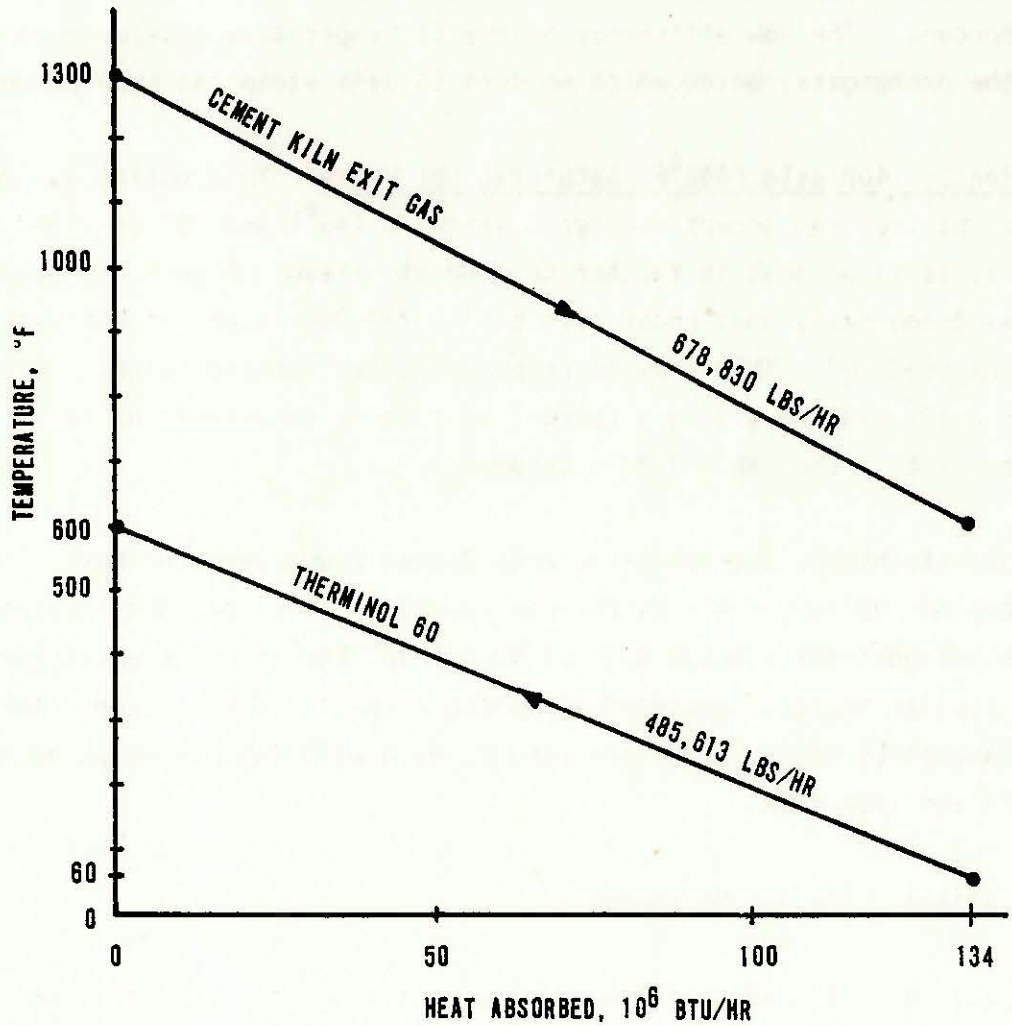


FIGURE 2.2-5
TEMPERATURE-ENTHALPY DIAGRAM
CHARGING SECTION, T-60 SYSTEM

Option 1 - 50 psia steam: This option assumes the user requires 50 psia saturated steam (281^{o}F). Figure 2.2-6(a) shows graphically that this product can be produced at a flowrate of 56,000 lbs/hr, corresponding to 64.3 x 10^6 Btu/hr heat recovery. As indicated in Table 2.2-7, column 1, the thermal efficiency for this option is 35 percent. The low efficiency is due to temperature crossover point in the exchangers, below which no more 50 psia steam can be produced.

Option 2 - 400 psia (446^{o}F) Saturated Hot Water: This option assumes that the user can accept saturated water at 446^{o}F and 400 psia and use it as is or heat it further to generate steam. Figure 2.2-6(b) shows graphically that under this option 233,000 lbs/hr of hot water can be produced. This flow corresponds to heat absorption of 92.5 x 10^6 Btu/hr, giving a thermal efficiency of between 55 to 61 percent. (See Table 2.2-7, Column 2).

Due to its higher thermal efficiency Option 2 was selected for conceptual design of the discharging section. Thus the BFW exchangers were designed for a total duty of 92.5 x 10^6 Btu/hr. The exchangers are similar to those designed in Option 3 for the T-66 system. Three exchangers (E-2A, 2B & 2C) are needed, each with two passes on both shell and tube side.

Operational Life and Maintenance

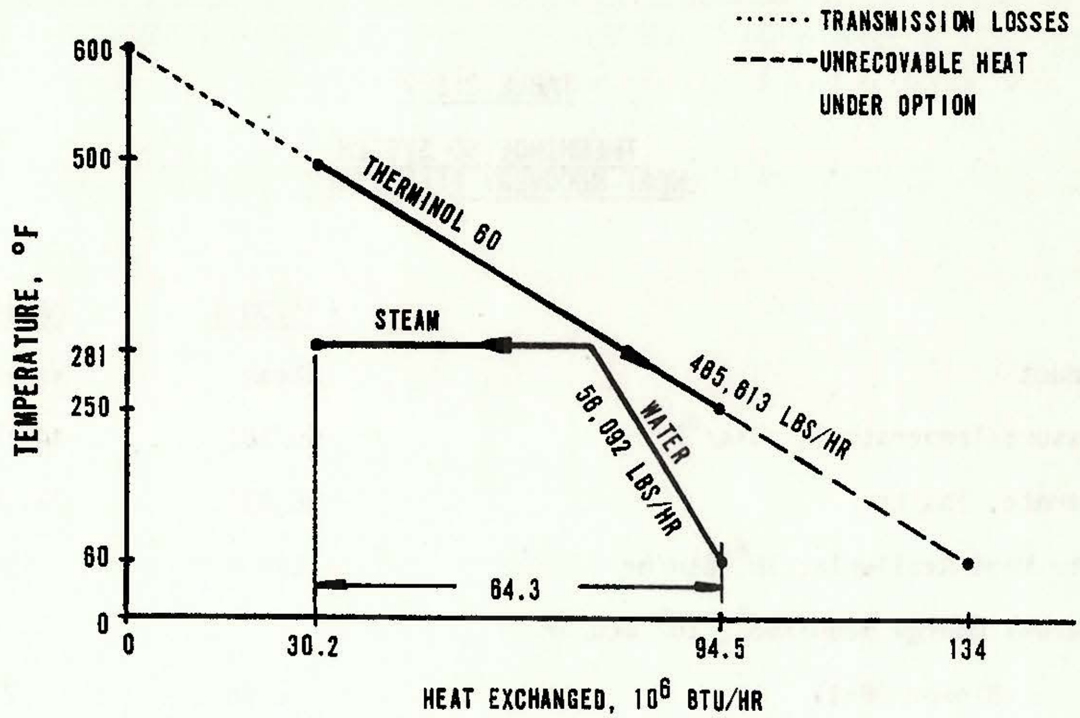
Operational life and maintenance will be similar to that for T-66. (See Section 2.2.2).

Environmental & Safety Factors

Similar to T-66 case.

System Disadvantages, Limitations, Institutional Barriers

Same as for T-66 system, Section 2.2.2.



(a) OPTION 1: 50 PSI STEAM HEAT RECOVERED 64.3 X 10^6 BTU/HR

(b) OPTION 2: 400 PSI (445° F) WATER HEAT RECOVERED 92.45 X 10^6 BTU/HR

FIGURE 2.2-6
HEAT RECOVERY OPTIONS
DISCHARGING SECTION, T-60 SYSTEM

TABLE 2.2-7

THERMINOL 60 SYSTEM
HEAT RECOVERY EFFICIENCY

	OPTION 1	OPTION 2[a]
Product	Steam	Water
Pressure/Temperature, psia/^{o}F	50/281	400/445
Flowrate, lbs/hr	56,092	233,095
Waste Heat Available, 10^6 Btu/hr	134.0	134.0
External Energy Required[b], 10^6 Btu/hr		
Blower (B-1)	5.68	7.12
Pump (P-1)	5.94	5.94
Pump (P-2)	5.34	5.94
Total	16.96	19.00
Useful Heat Recovered, 10^6 Btu/hr	64.3	92.5
Efficiency of Heat Recovery, %		
Definition 1[c]	42.6	60.5
Definition 2[d]	35.3	54.9

----

[a] OPTION 2 was selected for conceptual design.

[b] Thermal energy to electric energy conversion efficiency of 30% is assumed.

[c] Definition 1 = (Useful Heat x 100)/(Waste Heat + Ext. Energy).

[d] Definition 2 = (Useful Heat-Ext. Energy) x 100/(Waste Heat).

Equipment Specification and Cost

These were not developed in complete detail, but sufficiently to obtain manufacturer's estimate. Major equipment specifications are given below followed by Table 2.2-8 giving the manufacturer's bare equipment quote.

2.2.4 Pressurized Hot Water System

This system represents a commercially available alternative to the Sulfuric Acid/Water CRS used for thermal storage. The flowsheet is given in Figure 2.2-7. The waste heat is saturated steam at 120 psig and 350^{o}F from a fuel cell installation. It is used to heat water from 140 to 325^{o}F (corresponding to a pressure of 80 psig). The water is stored in pressurized tanks and is used to supply heat to a commercial or residential hot water heating system during the time when the fuel cell is not operating.

The storage system is assumed to be in close proximity to the fuel cell generator and the recovered energy user. It is used for storage application only with no provision for transport.

There are six horizontal pressurized water storage tanks (12' ID x 61' long) holding approximately 50,000 gallons each. At the start of the endothermic mode, five tanks are filled with water at 140^{o}F; the sixth tank is empty. Water is pumped from one of the tanks, heated in the steam condenser E-1, and returned to the empty tank. The originally filled tank is now empty. This operation is continued for all the tanks. At the end of the cycle, five tanks are filled with water at 325^{o}F and 80 psig, and the sixth tank is empty.

The tanks are designed for 100 psig. Two inches of fiberglass insulation is assumed. This reduces the heat loss to 9 x 10^{6} Btu/day from all tanks.

TABLE 2.2-8

THERMINOL 60 SYSTEM
EQUIPMENT COST

Equipment	Qty.	Due to Capacity $	Due to Energy Storage $	Source
Exhaust Blower, B-1	1	121,000	-	Buffalo Forge
Therminol Heater, E-1	1	243,000	-	Zurn
BFW Heat Exchangers, E-2	3	525,000	-	Goulds
Feed Pump, P-1	1	41,000	-	RECO
Return Pump, P-2	1	41,000	-	Goulds
Therminol Surge Tank, T-1	2	-	94,000	YUBA
Therminol Surge Tank, T-2	1	-	47,000	RECO
		971,000	141,000	

Total Equipment Cost = $971,000 + $141,000 = $1,112,000

NOTE: All costs are on a mid-1979 basis.

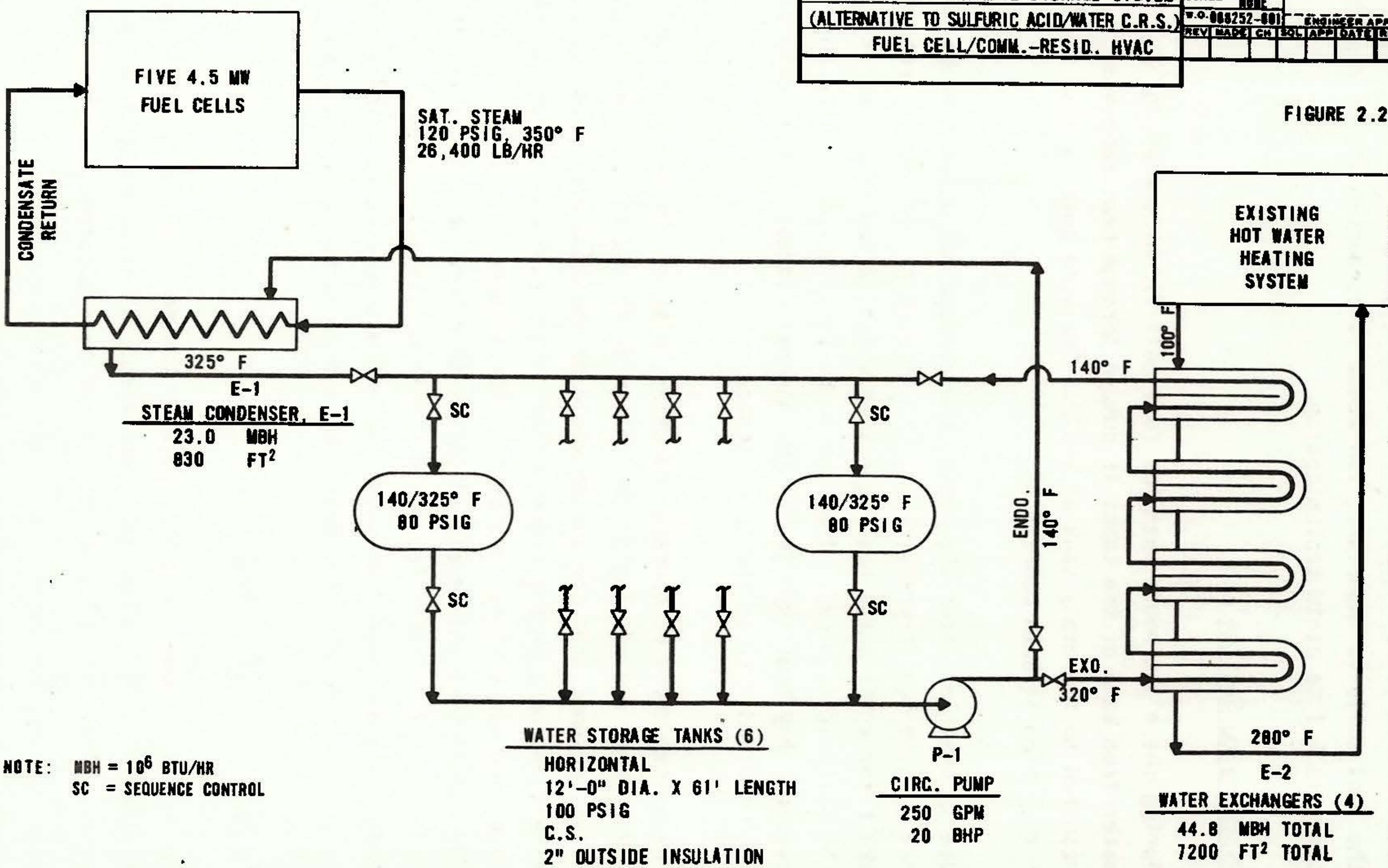
EPRI/SANDIA LABORATORIES
CONCEPTUAL PROCESS FLOW SHEET
HOT WATER THERMAL STORAGE SYSTEM
(ALTERNATIVE TO SULFURIC ACID/WATER C.R.S.)
FUEL CELL/COMM.-RESID. HVAC

MADE	CHKD	DRAWING NO.	SH. NO.	REV
WGD 5-8-79		SS-		

GILBERT ASSOCIATES, INC.
ENGINEERS AND CONSULTANTS
READING, PA.

SCALE NONE
W.O. 088252-001

FIGURE 2.2-7



The simplest design would involve condensing steam directly in the water filled tanks. However, since the quality of water returned to the fuel cells is important, the steam is condensed inside copper tubes in E-1 to avoid contamination.

Discharging Section

During the eight hours when the fuel cell is not operating, the hot water from each of the tanks is pumped through heat exchanger E-2 and returned to the empty tank at $140^{o}F$. The same pump is used for water circulation in both modes.

Due to heat loss from the tanks, the average hot water temperature is $320^{o}F$. The heat transferred in E-2 is used to heat water from 100 to $280^{o}F$ for a hot water heating system which is the same user as for the sulfuric acid CRS. There would be fired heaters in the hot water system. Any heat supplied by the thermal storage system would reduce the required duty on the fired heater.

The system could be operated manually, with an operator controlling the opening and closing of the valves at the tanks. In this study, it was assumed that the system would be operated with motor operated valves, using a sequence controller. Each tank would have duplicate level and temperature indicators. The sequence controller would use the temperature and level indicators to determine what valves to open and shut. The controller would continually check that both temperature indicators and level indicators agreed within a certain tolerance. If not, personnel at the heat source or user would be notified.

Heat Recovery Efficiency

Table 2.2-9 shows the energy balance around this system. Because the system is cyclic rather than continuous, it is shown on a daily basis. There is 367×10^6 Btu/day of heat absorbed from the fuel

TABLE 2.2-9

PRESSURIZED HOT WATER STORAGE SYSTEM HEAT RECOVERY EFFICIENCY

	10^6 Btu/Day
Waste Heat Absorbed	367
Heat Loss from Tanks	9
Power for Circulating Pump[a]	3
Heat Delivered	355
Efficiency of Heat Recovery, %	
Definition 1[b]	95.9
Definition 2[c]	95.9

Note:

[a] Thermal energy to electric energy conversion efficiency of 30% is assumed.

[b] Definition 1 = (Useful Heat x 100)/(Waste Heat + Ext. Energy)

[c] Definition 2 = (Useful Heat-Ext. Energy) x 100/(Waste Heat)

cell steam. Heat loss from the tanks is 9×10^6 Btu/day and electric power for the circulating pump is equivalent to 3×10^6 Btu/day. The energy recovery efficiency is 95.9%.

Operational Life and Maintenance Requirements

There would be no items requiring high maintenance in the system. Periodic checking of the condition of the water with respect to corrosion inhibitor concentration would be required. Occasional maintenance on the circulating pump would be required.

Environmental and Safety Errors

There would be no significant environmental or safety problems with this system.

System Disadvantages, Limitations, Institutional Barriers

This study assumes an adjacent siting for the waste heat source and the user. In an actual case, there may be a sizeable distance between the source and user. A requirement for piping would increase the cost of the recovered energy.

It is assumed that the entire quantity of waste heat absorbed from the source can be used by the heating system. In an actual case, this may not be the case during the summer months. Any unused storage capacity during part of the year would lead to higher unit costs for recovered heat.

Equipment Specifications and Cost

Descriptions of the major equipment for the pressurized hot water storage system are given in Appendix D-6. Equipment specifications were not developed in complete detail, but only sufficiently to allow obtaining budgetary cost estimates.

Table 2.2-10 presents the manufacturer's bare equipment costs.

TABLE 2.2-10

PRESSURIZED HOT WATER SYSTEM
EQUIPMENT COST

Equipment	Qty.	Total Equipment Cost: Due to Capacity $	Total Equipment Cost: Due to Energy Storage $	Source
Steam Condenser, E-1	1	25,000	-	YUBA
Water Storage Tanks, T-1	6	-	420,000	RECO
Circulation Pump, P-1	2	7,000	-	GOULDS
Water Exchangers, E-2	4	100,000	-	YUBA
		132,000	420,000	

Total Equipment Cost = \$132,000 + \$420,000 = \$552,000

NOTE: All costs are on a mid-1979 basis.

SECTION 3.0

ECONOMIC ANALYSIS OF THERMAL ENERGY STORAGE/TRANSPORTATION SYSTEMS

This section presents the economic analysis of the three CRSs and three alternative thermal energy storage/transportation systems described in Section 2.0. For each of these six systems, the total capital requirement, total operating cost, and life cycle energy cost are first developed at the base case conditions (design capacity, 25-mile transmission, and 30-year life). In the subsequent analysis, the sensitivity of costs with respect to changes in system capacity, transmission distance, and plant life is investigated to assess the effect of these parameters on the life cycle cost.

3.1 BASE CASE ECONOMICS

The total capital requirements for a system includes all capital necessary to complete the entire project. The total capital requirements for the systems in this study were estimated using the EPRI Technical Assessment Guide [3-1] and the "percentage of delivered-equipment cost method" [3-2]. The former was also used to calculate both the operating costs and life cycle energy costs.

3.1.1 <u>Capital and Operating Costs</u>

As defined in the EPRI Guide, the items comprising the total capital requirements (TCR) for a complete plant are:

- o Total Plant Investment (TPI)
- o Prepaid Royalties
- o Start-up costs
- o Inventory capital
- o Initial chemical and catalyst charge
- o Allowance for Funds During Construction (AFDC)
- o Land

The total operating costs (TOC) is the sum of the fixed operating costs and the variable operating costs. The fixed operating costs comprise operating labor, maintenance labor, and overhead charges. The variable operating costs include fuel, water, chemicals, waste disposal, etc.

Total Capital Requirements (TCR)

Since four systems out of the six studied in this phase involve pipeline transmission, the TCR is subdivided into plant capital requirement and transmission capital requirement. This segregation was made to show the impact of transmission cost on the economics of life cycle energy cost. The other two systems, being on-site thermal storage, do not involve any pipeline transmission and hence do not reflect this cost.

The plant capital requirement includes all costs in both the charging and discharging sections along with the costs of the compressors and pumps required for transmission. The transmission capital requirement includes the cost of either the buried or above-ground pipeline as the case may be, and the insulation if required.

Plant Capital Requirement: The bare equipment costs for major equipment were obtained either directly from vendors or from GAI's in-house information. For those equipment resized in Phase II, the costs were estimated by applying scale-up factors suggested by Guthrie (see Appendix B) or as per GAI in-house information to vendor quotations obtained during Phase I study.

The total direct cost was obtained by multiplying the bare equipment cost by factors to compensate for equipment erection labor, field installation labor, and field material costs. The process capital (PC) was then obtained by adding 75% (field indirect) of the sum of equipment erection labor and field installation labor to the total direct cost.

The total investment (TI) was then arrived by the summation of the process capital and percentage of process capital to reflect indirect costs such as general facilities (GF), engineering and home office fees, project contingency and process contingency.

Finally, the plant capital requirement was derived by the summation of the total investment and a percentage for royalty allowance, start-up costs, inventory capital, initial catalysts and chemicals, allowance for funds during construction (AFDC) and the land costs. It should be added that the percentages or factors applied vary for different systems. They were obtained from the EPRI Guide and GAI in-house information (See Appendix C).

Transmission Capital Requirement: The process capital (PC) for the pipeline and insulation was determined by the GAI Cost Engineering Department. It includes the pipeline material, erection, excavation, back-fill, rented equipment for underground installation, coating/wrapping materials and labor, support and protection saddles for aboveground installation, insulation as required.

The final transmission capital requirement was derived by similar procedures as for plant capital requirements explained earlier.

Total Operating Costs (TOC)

Fixed Operating Cost: This cost is again subdivided into the two groups - plant and transmission. These costs are calculated for the first year of operation only. The fixed operating cost is a function of operating labor, maintenance labor, maintenance material, administration and support labor.

Variable Operating Cost: This cost is also based on the first year requirements. It is a function of the water and electricity requirements as well as the chemicals and catalysts required. Listed below are some variables and their unit costs used in this study.

Electricity, \$/kWh	0.03
Water \$/1000 Gals.	
Cooling Water	0.10
Make-up Water	0.43
Chemicals and Catalysts	
Methanation catalyst, $\$/ft^3$	135.00
Reforming catalyst, $\$/ft^3$	157.00
Cyclohexane Dehydrogenation catalyst, $\$/ft^3$	300.00
Benzene Hydrogenation catalyst, $\$/ft^3$	300.00
Natural gas, $\$/10^6$ Btu	2.05
Benzene, \$/Gal	1.00
Cyclohexane, \$/Gal	1.10
Sulfuric acid (100%), \$/ton	52.50
Therminol 66, \$/Gal	8.51
Therminol 60, \$/Gal	7.45

Based upon the above procedures, the Total Capital Requirement (TCR) and the Total Operating Costs (TOC) for the three CRSs and their alternatives are presented in Table 3.1-1 thru Table 3.1-6.

3.1.2 Life Cycle Cost

The life cycle energy costs (LCEC) is determined for the complete project (which includes transmission where applicable). As defined in the EPRI Guide, the life cycle energy cost is a function of total capital requirement (TCR), total operating costs (TOC), levelized annual capacity factor (CF; assumed to be 1 in this study) and annual plant capacity (G). The capital and operating costs are both levelized over the entire plant life, assumed to be 30 years. Thus, the life cycle energy cost is defined by the following equation:

$$\text{LCEC } (\$/10^6 \text{ Btu}) = \frac{0.1463 \text{ TCR} + 1.935 \text{ TOC}}{(\text{CF})\ (\text{G})}$$

TABLE 3.1-1

CAPITAL REQUIREMENT AND OPERATING COST METHANE/SYNGAS CRS

Base Case

Heat Available, 10^6 Btu/hr	83.1 for 16 hr/day, 6 days/wk
Heat Recovered, 10^6 Btu/hr	29.3 for 16 hr/day, 6 days/wk
Transmission Distance, Miles	25 (Underground)

Capital Requirement (10^6 \$)	Plant	Transmission Pipeline
Process Capital (PC)	3.731	9.051
General Facilities (GF)	0.560	-
Eng. and Home Office Fees	0.373	0.905
Subtotal	4.664	9.956
Project Contingency	0.700	1.493
Process Contingency	0.448	0.453
Total Investment (TI)	5.812	11.902
Royalty Allowance	0.019	-
Start-up Cost	0.180	0.060
Inventory Capital	0.058	-
Initial Catalysts/Chemicals	0.078	-
Allowance for Funds During Construction (AFDC)	0.457	0.952
Land	0.040	0.200
Capital Requirement	6.644	13.114
Fixed Operating Cost (10^6 \$/yr)		
Operating Labor	0.2838	-
Maintenance Labor	0.0631	0.0250
Maintenance Material	0.0946	0.0375
Administrative and Support Labor	0.0991	0.0075
Fixed O&M (1st Yr.)	0.5406	0.0700
Variable Operating Cost (10^6 \$/yr)		
Electricity	0.1757	-
Water	0.0002	-
Catalyst	0.0364	-
Other	0.0086	-
Variable O&M (1st Yr.)	0.2209	-

Summary

Total Capital Requirement (TCR) =	\$ 19.758 Million
Total Operating Cost (TOC) =	\$ 0.8315 Million/Yr

TABLE 3.1-2

CAPITAL REQUIREMENT AND OPERATING COST
BENZENE/CYCLOHEXANE CRS

Base Case		
Heat Availble, 10^6 Btu/hr	133.7 Continuous for 347 days/yr.	
Heat Recovered, 10^6 Btu/hr	92.6 Continuous for 347 days/yr.	
Transmission Distance, Miles	25 (Underground)	
Capital Requirement (10^6 \$)	**Plant**	**Transmission Pipeline**
Process Capital (PC)	3.823	10.712
General Facilities (GF)	0.573	-
Eng. and Home Office Fees	0.382	1.071
Subtotal	4.778	11.783
Project Contingency	0.717	1.767
Process Contingency	0.573	0.536
Total Investment (TI)	6.068	14.086
Royalty Allowance	0.019	-
Start-up Cost	0.240	0.070
Inventory Capital	0.061	-
Initial Catalysts/Chemicals	0.453	-
Allowance for Funds During Construction (AFDC)	0.485	1.127
Land	0.040	0.200
Capital Requirement	7.366	15.483
Fixed Operating Cost (10^6 \$/yr)		
Operating Labor	0.4687	-
Maintenance Labor	0.0646	0.0250
Maintenance Material	0.0969	0.0375
Administrative and Support Labor	0.1600	0.0075
Fixed O&M (1st Yr.)	0.7902	0.0700
Variable Operating Cost (10^6 \$/yr)		
Electricity	0.5720	-
Water	0.0444	-
Catalyst	0.0774	-
Other	0.0631	-
Variable O&M (1st Yr.)	0.7569	-

Summary

Total Capital Requirement (TCR) = \$ 22.849 Million
Total Operating Cost (TOC) = \$ 1.6171 Million/Yr

TABLE 3.1-3

CAPITAL REQUIREMENT AND OPERATING COST
SULFURIC ACID/WATER CRS

Base Case

Heat Available, 10^6 Btu/hr	23.0 for 16 hr/day, 7 days/wk
Heat Recovered, 10^6 Btu/hr	11.0 for 8 hr/day, 7 days/wk
Transmission Distance, Miles	0 (Storage at site)

Capital Requirement (10^6 \$)	Plant
Process Capital (PC)	1.187
General Facilities (GF)	0.237
Eng. and Home Office Fees	0.178
Subtotal	1.602
Project Contingency	0.240
Process Contingency	0.119
Total Investment (TI)	1.961
Royalty Allowance	0.006
Start-up Cost	0.051
Inventory Capital	0.020
Initial Catalysts/Chemicals	0.009
Allowance for Funds During	0.157
Construction (AFDC)	0.040
Land	0.020
Capital Requirement	2.224

Fixed Operating Cost (10^6 \$/yr)	
Operating Labor	0.0300
Maintenance Labor	0.0204
Maintenance Material	0.0306
Administrative and Support Labor	0.0151
Fixed O&M (1st Yr.)	0.0961

Variable Operating Cost (10^6 \$/yr)	
Electricity	0.0029
Water	0.0402
Catalyst	-
Other	-
Variable O&M (1st Yr.)	0.0431

Summary

Total Capital Requirement (TCR) = \$ 2.224 Million
Total Operating Cost (TOC) = \$ 0.1392 Million/Yr.

TABLE 3.1-4

CAPITAL REQUIREMENT AND OPERATNG COST THERMINOL 66 SYSTEM

Base Case

Heat Available, 10^6 Btu/hr	93.3 for 16 hr/day, 6 days/wk
Heat Recovered, 10^6 Btu/hr	66.5 for 16 hr/day, 6 days/wk
Transmission Distance, Miles	25 (Aboveground)

Capital Requirement (10^6 \$)	Plant	Transmission	
		Insulation	Pipeline
Process Capital (PC)	3.380	3.272	9.730
General Facilities (GF)	0.507	-	-
Eng. and Home Office Fees	0.338	-	0.937
Subtotal	4.225	3.272	10.703
Project Contingency	0.634	0.491	1.605
Process Contingency	0.169	0.164	0.486
Total Investment (TI)	5.028	3.927	12.794
Royalty Allowance	-	-	-
Start-up Cost	0.140	-	0.064
Inventory Capital	0.050	-	-
Initial Catalysts/Chemicals	6.561	-	-
Allowance for Funds During Construction (AFDC)	0.402	0.314	1.024
Land	0.040	-	0.200
Capital Requirement	12.221	4.241	14.082

Fixed Operating Cost (10^6 \$/yr)	Plant	Pipeline
Operating Labor	0.0600	-
Maintenance Labor	0.0571	0.0250
Maintenance Material	0.0857	0.0375
Administrative and Support Labor	0.0351	0.0075
Fixed O&M (1st Yr.)	0.2379	0.0700

Variable Operating Cost (10^6 \$/yr)	Plant	Pipeline
Electricity	0.2470	-
Water	-	-
Catalyst	-	-
Other	-	-
Variable O&M (1st Yr.)	0.2470	-

Summary

Total Capital Requirement (TCR) = \$ 30.544 Million

Total Operating Cost (TOC) = \$ 0.5549 Million/Yr

TABLE 3.1-5

CAPITAL REQUIREMENT AND OPERATING COST
THERMINOL 60 SYSTEM

Base Case

Heat Available, 10^6 Btu/hr	133.7	Continuous for 347 days/yr
Heat Recovered, 10^6 Btu/hr	92.5	Continuous for 347 days/yr
Transmission Distance, Miles	25	(Aboveground)

Capital Requirement (10^6 \$)	Plant	Transmission	
		Insulation	Pipeline
Process Capital (PC)	4.782	3.272	9.730
General Facilities (GF)	0.574	-	-
Eng. and Home Office Fees	0.383	-	0.973
Subtotal	4.782	3.272	10.703
Project Contingency	0.717	0.491	1.605
Process Contingency	0.191	0.164	0.486
Total Investment (TI)	5.690	3.927	12.794
Royalty Allowance	-	-	-
Start-up Cost	0.170	-	0.064
Inventory Capital	0.057	-	-
Initial Catalysts/Chemicals	5.602	-	-
Allowance for Funds During Construction (AFDC)	0.455	0.314	1.024
Land	0.040	-	0.200
Capital Requirement	12.014	4.241	14.082

Fixed Operating Cost (10^6 \$/yr)	Plant	Transmission
Operating Labor	0.0600	-
Maintenance Labor	0.0646	0.0250
Maintenance Material	0.0970	0.0375
Administrative and Support Labor	0.0374	0.0075
Fixed O&M (1st Yr.)	0.2590	0.0700

Variable Operating Cost (10^6 \$/yr)	Plant	Transmission
Electricity	0.4172	-
Water	-	-
Catalyst	-	-
Other	-	-
Variable O&M (1st Yr.)	0.4172	-

Summary

Total Capital Requirement (TCR) = \$ 30.337 Million

Total Operating Cost (TOC) = \$ 0.7462 Million/Yr

TABLE 3.1-6

CAPITAL REQUIREMENT AND OPERATING COST
HOT WATER SYSTEM

Base Case	
Heat Available, 10^6 Btu/hr	23.0 for 16 hr/day, 7 days/wk
Heat Recovered, 10^6 Btu/hr	44.8 for 8 hr/day, 7 days/wk
Transmission Distance Miles	0 (Storage on site)
Capital Requirement (10^6 \$)	**Plant**
Process Capital (PC)	1.848
General Facilities (GF)	0.370
Eng. and Home Office Fees	0.277
Subtotal	2.495
Project Contingency	0.374
Process Contingency	0.092
Total Investment (TI)	2.961
Royalty Allowance	-
Start-up Cost	0.078
Inventory Capital	0.030
Initial Catalysts/Chemicals	-
Allowance for Funds During Construction (AFDC)	0.237
Land	0.020
Capital Requirement	3.326
Fixed Operating Cost (10^6 \$/yr)	
Operating Labor	0.0300
Maintenance Labor	0.0290
Maintenance Material	0.0434
Administrative and Support Labor	0.0177
Fixed O&M (1st Yr.)	0.1201
Variable Operating Cost (10^6 \$/yr)	
Electricity	0.0987
Water	-
Catalyst	-
Other	-
Variable O&M (1st Yr.)	0.0987

Summary

Total Capital Requirement (TCR) = \$ 3.326 Million
Total Operating Cost (TOC) = \$ 0.2188 Million/Yr

The levelized fixed charge rate of 0.1463 for the TCR is based on a 30-year book life, a 20-year tax life, and flow through accounting. It is derived as follows:

Total return (weighted cost of capital)	10.00%
Book depreciation (sinking fund over 30 years)	0.61%
Allowance for retirement dispersion	0.56%
Levelized annual income tax	4.70%
Levelized annual accelerated depreciation factor	<2.47%>
Levelized annual investment tax credit at 4%	<0.77%>
Property taxes, insurance, etc.	2.00%
	14.63%

The following values are assumed for the base parameters used in financial calculations:

Debt/equity ratio	50/50
Debt cost, %/year	8
Preferred stock, %	15
Preferred stock cost, %/yr	8.5
Common stock, %	35
Common stock cost, %/yr	13.5
Weighted cost of capital, %/yr	10
Federal and state income tax, %	50
Property taxes and insurance, %/yr	2
Investment tax credit, %	4

The levelized factor of 1.935 applied to the total first year operating cost (TOC) is based on the following:

Inflation rate for labor, materials and consumables, %/yr	6.0
Real escalation for labor, materials and consumables, %/yr	0.2

Based on the above equation and TCR and TOC as shown in Tables 3.1-1 through 6, the life cycle costs for the six systems were calculated. Table 3.1-7 presents these life cycle costs along with their TCR and TOC. The table also shows the annual heat recovered, thermal efficiency and transmission distance for each system.

As seen from Table 3.1-7, among the three CRSs considered the Benzene/Cyclohexane CRS presents the most economical life cycle cost at $8.40 per million Btu with the Methane/Syngas system being the most expensive at $30.74 per million Btu. The more favorable economics of the Benzene/Cyclohexane CRS is attributed to its relatively high capacity combined with its high thermal efficiency at 25 miles transmission distance.

Among the three respective alternatives considered, the pressurized hot water system (alternative to the Sulfuric Acid/Water CRS) is the most economical at $7.14 per million Btu, but the T-60 system (alternative to Benzene/Cyclohexane CRS) at $7.63 per million Btu is not far behind. Of course, the Hot Water (HW) is an on-site thermal storage system whereas the T-60 system delivers heat to the user 25 miles from the heat source. In both the CRSs and their alternatives, the Methane/Syngas and its alternative, T-66 system, are the most expensive systems.

When the CRSs are contrasted with the alternatives, each alternative system is more economical than its corresponding CRS. The T-66 and HW system both have life cycle costs nearly 45-60 percent better than Methane/Syngas CRS and Sulfuric Acid/Water CRS, respectively. Between Benzene/Cyclohexane CRS and its alternative T-60 system, the difference is much smaller. The B/C CRS life cycle cost is only about 10 percent higher than that of its alternative ($8.40 versus $7.63 per million Btu). For a detailed comparison between a CRS and its alternative, see Sections 4.0 and 5.0.

TABLE 3.1-7

ECONOMICS SUMMARY FOR BASE CASES

	CRS			Alternatives		
	M/S	B/C	SA/W	T-66	T-60	HW
Annual Heat Recov., 10^9 Btu	147	771	32	333	770	130
Thermal Efficiency, %[(a)]						
Definition 1	30.4	57.0	23.7	60.3	60.5	95.9
Definition 2	19.4	47.0	23.2	53.0	54.9	95.9
Transmission Distance, Miles	25	25	0	25	25	0
Capital Requirements, 10^6 $						
Plant	6.644	7.366	2.244	12.221	12.014	3.326
Transmission	13.114	15.483	0	18.323	18.323	0
Total	19.758	22.849	2.244	30.544	30.337	3.326
Total Oper. Cost (1st Yr), 10^6 $/Yr	0.8315	1.6171	0.1392	0.5549	0.7462	0.2188
Life Cycle Cost, $/$10^6$ Btu[(b)]	30.74	8.40	18.67	16.59	7.63	7.14

(a) Definition 1 = (Useful Heat x 100)/(Waste Heat + External Energy)
Definition 2 = (Useful Heat - External Energy) x 100/Waste Heat

(b) Based on a 30-year plant life

3.2 COST SENSITIVITY ANALYSIS

In Section 3.1, the capital, operating, and life cycle costs of CRSs and alternative systems were developed at the base case conditions. In this section the sensitivity of these costs to changes in system capacity, transmission distance, and plant life is discussed.

3.2.1 Effect of System Capacity

Capital costs for capacities different than the base case were arrived at by starting with the base case equipment costs. The cost of each piece of major equipment was adjusted to the new capacity using the exponents (or scaling factors) listed in Appendix B. The following formula was used.:

$$(\text{Cost, new cap.}) = (\text{Cost, base cap.}) \left(\frac{\text{new cap.}}{\text{base cap.}}\right)^{\text{exp.}}$$

In those instances when the equipment size was considered the maximum available or desirable, multiple units were used. After the total equipment cost was obtained for the new case, total plant/transmission investment was calculated using the same factors used for the base case. The operating costs and life cycle costs were calculated in the same manner as for the base case.

The effect of system capacity was analyzed at four relative capacity levels; namely, 0.5, 1(base), 2, and 5. Table 3.2-1 presents capital and operating costs at these four levels for all six systems. Major results are discussed below:

- In the two transportation CRSs, (M/S and B/C), plant capital cost increases more rapidly than transmission capital cost as plant capacity increases. At five times base capacity, the plant capital requirement is over 50 percent of the total capital requirements in comparison to only 34 percent for the base case.

TABLE 3.2-1

SENSITIVITY ANALYSIS: CAPITAL REQUIREMENT AND OPERATING COST

CRS		Relative System Capacity 0.5		1.0		2.0		5.0	
		Capital	Operating	Capital	Operating	Capital	Operating	Capital	Operating
M/S	Plant	3.945	0.4915	6.644	0.7615	10.921	1.1928	24.732	2.4171
	Trans.	9.543	0.0700	13.114	0.0700	19.524	0.0700	25.314	0.0700
	Total	13.488	0.5615	19.758	0.8315	30.445	1.2628	50.046	2.4871
B/C	Plant	3.891	0.9304	7.366	1.5471	14.206	2.6346	32.262	5.6404
	Trans.	11.382	0.0700	15.483	0.0700	20.620	0.0700	28.629	0.0700
	Total	15.273	1.0004	22.849	1.6171	34.826	2.7046	60.891	5.7104
SA/W	Plant	1.405	0.096	2.244	0.1392	3.480	0.2142	6.374	0.4183
	Trans.	0	0	0	0	0	0	0	0
	Total	1.405	0.096	2.244	0.1392	3.480	0.2142	6.374	0.4183
Alternatives									
T-66	Plant	6.266	0.2878	12.221	0.4849	20.397	0.8562	46.254	1.9437
	Trans.	17.567	0.070	18.323	0.0700	22.411	0.0700	28.185	0.0700
	Total	23.833	0.3578	30.544	0.5549	42.808	0.9262	74.439	2.0137
T-60	Plant	7.088	0.3950	12.014	0.6762	24.192	1.2398	59.954	2.9413
	Trans.	17.175	0.0700	18.323	0.0700	25.217	0.0700	34.701	0.0700
	Total	24.263	0.4650	30.337	0.7462	49.409	1.3098	94.655	3.0113
HW	Plant	1.705	0.1338	3.326	0.2188	5.867	0.3938	13.410	0.8930
	Trans.	0	0	0	0	0	0	0	0
	Total	1.705	0.1338	3.326	0.2188	5.867	0.3938	13.410	0.8930

Note: Capital requirements are in 10^6 \$ and operating cost in 10^6 \$/Yr.
All other parameters are the same as base case.

- In the two transportation alternatives (T-66 and T-60), increase in plant capital requirement with an increase in system capacity is even more pronounced. For the lowest system capacity, the plant capital requirement is only between 26 and 29 percent of the total capital requirement, but increases to about 62 percent at five times base capacity. This rapid increase is contributed mostly by the increase in the cost of expensive transfer fluid, T-66 and T-60.

- For the four transportation systems, the operating costs for the transmission section is constant irrespective of the plant capacity. This is due to the fact that the compressor or pumps required for the 25-mile transmission is considered a part of the plant section and, hence, the operating costs of these equipment are reflected in the plant operating costs.

The life cycle costs, as determined using these capital and operating costs, are summarized in Column (a) of Table 3.2-2. Table 3.2-3 presents the percentage breakdown, by elements, of the life cycle costs for the various capacities studied. Life cycle costs are shown graphically in Figure 3.2-1 for CRSs and Figure 3.2-2 for alternative systems. Summarized below are major results with discussions.

- For each system, the life cycle cost decreases significantly with an increase in system capacity.

- The three CRSs all show approximately the same slope of life cycle costs versus capacity.

- The life cycle costs for CRSs is higher than its corresponding alternative system, with an exception of the T-60 system. At 50 percent capacity level, the T-60 system shows a slightly higher life cycle cost than the B/C CRS. This is largely due to the high cost associated with thick insulation required for the T-60 system.

TABLE 3.2-2

SENSITIVITY ANALYSIS - LIFE CYCLE COST ($\$/10^6$ BTU)

	(a) Capacity (Relative Factor)				(b) Distance (miles)				(c) Life (Yrs.)	
	0.5	1	2	5	0	10	25	100	20	30
CRS										
Methane/Syngas	41.72	30.74	23.51	16.55	-	22.33	30.74	75.69	30.47	30.74
Benz./Cyclohex.	10.82	8.40	6.70	5.18	-	6.39	8.40	20.26	8.11	8.40
S.A./Water	24.45	18.67	14.43	10.89	18.67	-	-	-	18.15	18.67
Alternatives										
T-66	25.06	16.59	12.08	8.87	2.87	8.48	16.59	-	17.01	16.59
T-60	11.56	7.63	6.34	4.90	1.86	3.97	7.63	-	7.74	7.63
Hot Water	7.85	7.14	6.25	5.70	7.14	-	-	-	6.91	7.14

Note: Underlined figures are the base case costs for each sensitivity parameter (relative capacity, transmission distance, and plant life).

All other parameters are the same as base case.

TABLE 3.2-3

LIFE CYCLE COST
PERCENTAGE BREAKDOWN BY ELEMENTS

		Relative System Capacity											
CRS		0.5			1.0 (Base)			2.0			5.0		
		Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total
M/S	Plant	19	31	50	22	32	54	23	34	57	30	38	68
	Trans.	46	4	50	43	3	46	41	2	43	31	1	32
	Total	65	35	100	65	35	100	64	36	100	61	39	100
B/C	Plant	14	43	57	17	46	63	20	50	70	24	57	78
	Trans.	40	3	43	35	2	37	29	1	30	21	1	22
	Total	54	46	100	52	48	100	49	51	100	45	58	100
SA/W	Plant	53	47	100	55	45	100	55	45	100	54	46	100
	Trans.	-	-	-	-	-	-	-	-	-	-	-	-
	Total	53	47	100	55	45	100	55	45	100	54	46	100
Alternatives													
T-66	Plant	22	13	35	32	17	49	37	21	58	46	25	71
	Trans.	62	3	65	48	3	51	40	2	42	28	1	29
	Total	84	16	100	80	20	100	77	23	100	74	26	100
T-60	Plant	23	17	40	30	22	52	36	25	61	45	28	73
	Trans.	56	4	60	46	2	48	38	1	39	26	1	27
	Total	79	21	100	76	24	100	74	26	100	71	29	100
HW	Plant	49	51	100	53	47	100	53	47	100	53	47	100
	Trans.	-	-	-	-	-	-	-	-	-	-	-	-
	Total	49	51	100	53	47	100	53	47	100	53	47	100

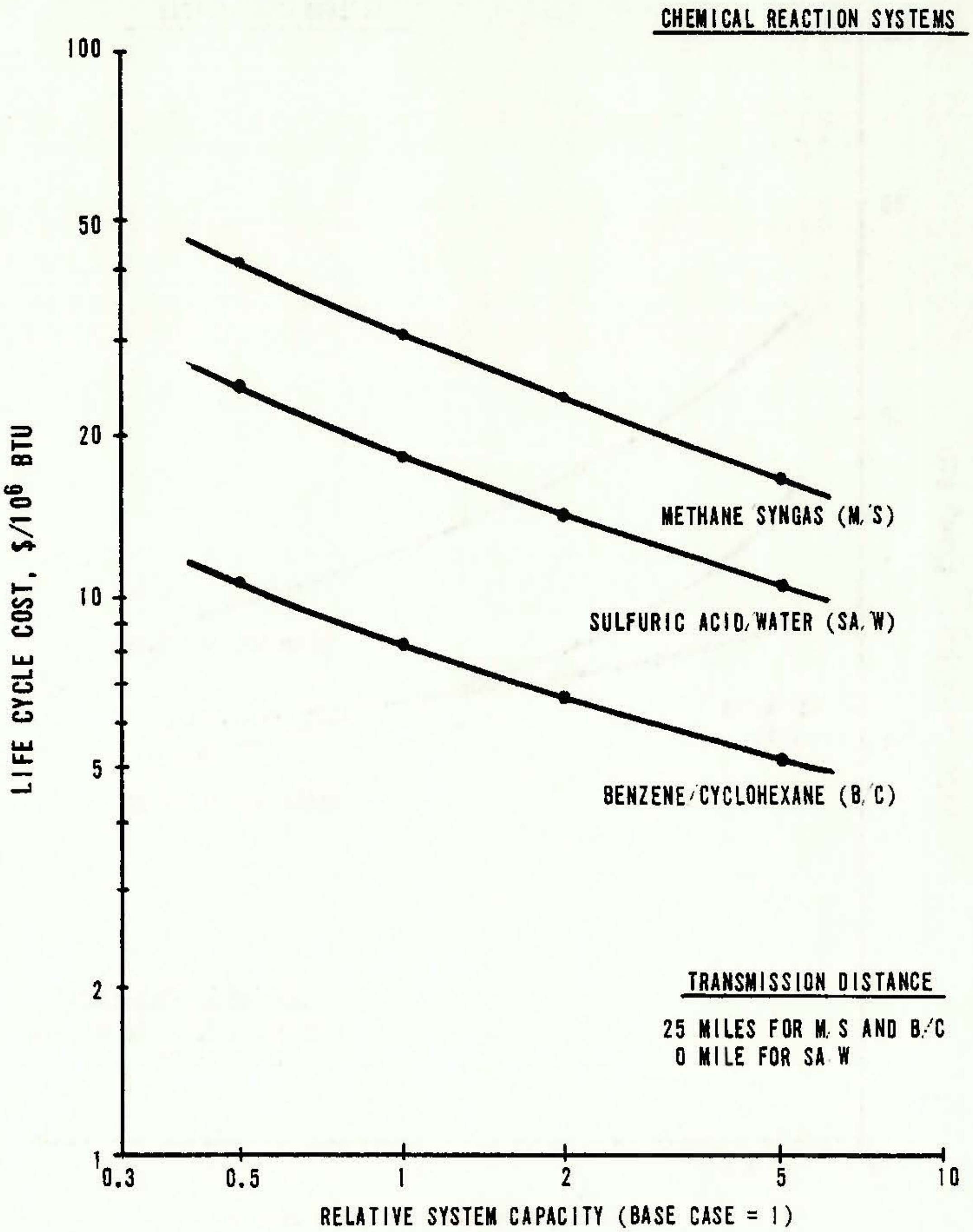


FIGURE 3.2-1
LIFE CYCLE COST VS. SYSTEM CAPACITY: CRSs

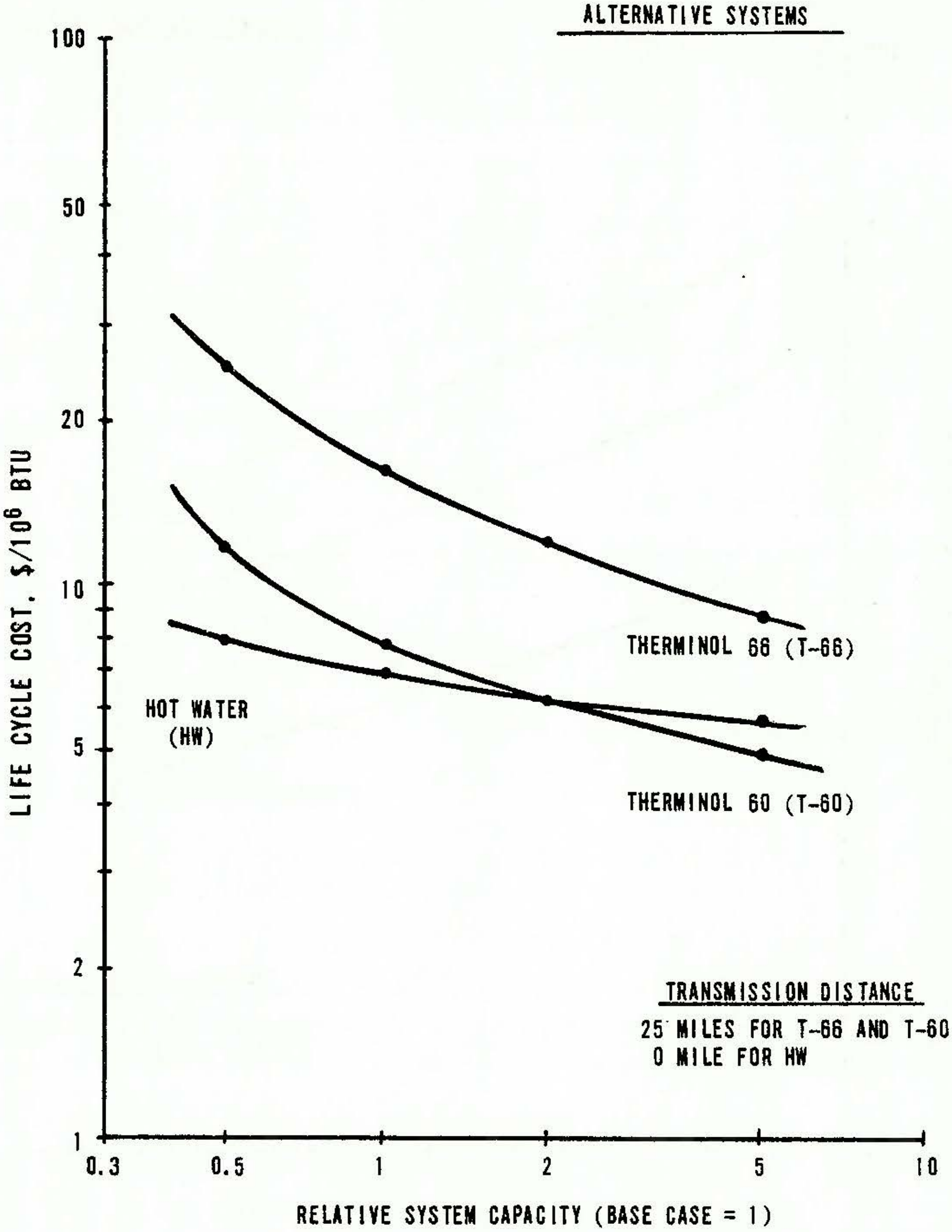


FIGURE 3.2-2

LIFE CYCLE COST VS. SYSTEM CAPACITY: ALTERNATIVES

- Among the three CRSs considered, the B/C CRS shows the most economical life cycle costs which range from $5.18 to $10.82 per million Btu. The SA/W CRS ranks second economically ($10.89 to $24.45) and the M/S CRS is the most expensive system ($16.55 to $41.72). The relatively favorable economics of the B/C CRS is attributable to its high capacity combined with its higher thermal efficiency.

- Among the three corresponding alternatives considered, the HW system is the most economical at $5.70 to $7.85 per million Btu. The T-60 system is the next at $4.90 to $11.56 and the T-66 system the most expensive at $8.87 to $25.06 per million Btu.

- In the two CRS transportation systems, 45-65% of the life cycle cost is attributable to the capital requirement (plant + transmission). In the two alternative transportation systems, this percentage is about 71-84%. The difference is due to the fact that each CRS has relatively low capital requirements, but relatively high operating costs than its corresponding alternative system.

3.2.2 Effect of Transmission Distance

For the two CRSs involving thermal transport, the transmission distance studied were 10 miles, 25 miles (base case), and 100 miles. The 0 mile (on-site) case was excluded since operational complexity makes them unlikely candidates for on-site applications.

For the two transportation alternatives, the transmission distances studied were 0 mile (or on-site), 10 miles, and 25 miles (base case). The 100 mile transmission case was not analyzed, since excessive heat loss through the pipeline over this distance would result in very low capacity or heat delivery at the user end.

The capital cost for transmission pipelines was assumed to be linear with distance. In other words, a 100 mile line would cost four times the cost of a 25 mile line. For the two CRSs at 100 miles, however, pumping stations at 50 miles were added, and this added to the transmission cost.

The effect of transmission distance on various capital requirements and operating costs, discussed below, is presented in Table 3.2-4. Table 3.2-5 presents the percentage breakdown, by elements, of the life cycle costs at the various transmission distances studied.

- For the two CRSs, the plant capital requirement is more or less constant for transmission distances between 10 miles to 100 miles. The marginal increase is due to the increased compressor/pumping cost at longer distance. The transmission capital requirement increases linearly with transmission distance. The transmission capital requirement for the 100 mile transmission systems includes the cost of intermediate pumping/compressing stations required. The 10 mile and 25 mile cases do not require these stations.

- For the two alternative systems, the plant capital requirements vary significantly with transmission distance. The variation in plant capital requirements at different transmission distances is mainly due to the high inventory costs of the heat transfer fluids (T-66 and T-60). For example, the cost of Therminol 66 required in the base case (25 mile transmission) is \$6.561 million, \$2.637 million for 10 mile transmission and only \$0.213 million for 0 mile transmission. The transmission capital requirement increases linearly with distance.

- The plant operating costs for the CRSs are higher than their alternative systems. This is due to the fact that the CRS is a more complex plant and needs more operating labor than the highly automated alternative system. The transmission operating costs are about constant for distances up to 25 miles. These costs

TABLE 3.2-4

SENSITIVITY ANALYSIS: CAPITAL REQUIREMENT AND OPERATING COST

		Transmission Distance, Miles							
CRS		0		10		25 (Base)		100	
		Capital	Operating	Capital	Operating	Capital	Operating	Capital	Operating
M/S	Plant	-	-	6.519	0.7329	6.644	0.7615	6.886	0.9666
	Trans.	-	-	5.246	0.0700	13.114	0.0700	52.940	0.2473
	Total	-	-	11.765	0.8029	19.758	0.8315	59.826	1.2139
B/C	Plant	-	-	7.158	1.4487	7.366	1.5471	8.034	2.0209
	Trans.	-	-	6.194	0.0700	15.483	0.0700	63.724	0.6210
	Total	-	-	13.352	1.5187	22.849	1.6171	71.758	2.6419
Alternatives									
T-66	Plant	5.367	0.2332	7.897	0.4131	12.221	0.4849	-	-
	Trans.	0	0	7.305	0.0700	18.323	0.0700	-	-
	Total	5.367	0.2332	15.202	0.4831	30.544	0.5549	-	-
T-60	Plant	7.198	0.4368	8.436	0.5257	12.014	0.6762	-	-
	Trans.	0	0	7.305	0.0700	18.323	0.0700	-	-
	Total	7.198	0.4368	15.741	0.5957	30.337	0.7462		

Note: Capital requirements are in 10^6 \$ and operating costs in 10^6 \$/Yr.
All other parameters are the same as base case.

TABLE 3.2-5

LIFE CYCLE COST
PERCENTAGE BREAKDOWN BY ELEMENTS

		Transmission Distance, Miles											
CRS		0.5			10			25			100		
		Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total	Cap.	Op.	Total
M/S	Plant	-	-	-	29	43	72	22	32	54	9	17	26
	Trans.	-	-	-	23	5	28	43	3	46	68	6	74
	Total	-	-	-	52	48	100	65	35	100	77	23	100
B/C	Plant	-	-	-	21	58	79	17	46	63	8	25	33
	Trans.	-	-	-	19	2	21	35	2	37	60	7	67
	Total	-	-	-	40	60	100	52	48	100	68	32	100
Alternatives													
T-66	Plant	64	36	100	37	25	62	32	17	49	-	-	-
	Trans.	-	-	-	34	4	38	48	3	51	-	-	-
	Total	64	36	100	71	29	100	80	20	100	-	-	-
T-60	Plant	55	45	100	36	29	65	30	22	52	-	-	-
	Trans.	-	-	-	31	4	35	46	2	48	-	-	-
	Total	55	45	100	67	33	100	76	24	100	-	-	-

increase rapidly for the 100 mile transmission in the CRSs. The increase is primarily due to the operating costs attributable to the intermediate pumping/compression stations.

Figures 3.2-3 and 3.2-4 present the life cycle costs versus transmission distance for the two CRSs and two alternatives. Highlights are summarized below:

- For CRSs, the life cycle cost increases with increasing transmission distance. Since the plant capital requirement as well as operating costs is about constant, the increase in life cycle cost is mainly due to transmission distance.

- The percentage of life cycle cost attributable to the transmission capital requirement increases with increasing transmission distance. For example, this percentage is 23% at 10 miles and 68% at 100 miles for the M/S CRS.

- In the two alternative systems, both the rate of heat delivery (10^6 Btu/hr) and the external energy requirement vary strongly with transmission distance (Appendix B). As a result, the thermal efficiency decreases significantly with increasing distance:

		Thermal Efficiency, % [(a)]		
	Distance, Miles	0	10	25
T-66	Definition 1	88.2	74.3	60.3
	Definition 2	87.2	72.1	53.0
T-60	Definition 1	86.4	71.6	60.5
	Definition 2	85.6	69.0	54.9

[(a)] Definition 1 = (Useful Heat x 100)/(Waste Heat + External Energy)
Definition 2 = (Useful Heat - External Energy) x 100/Waste Heat

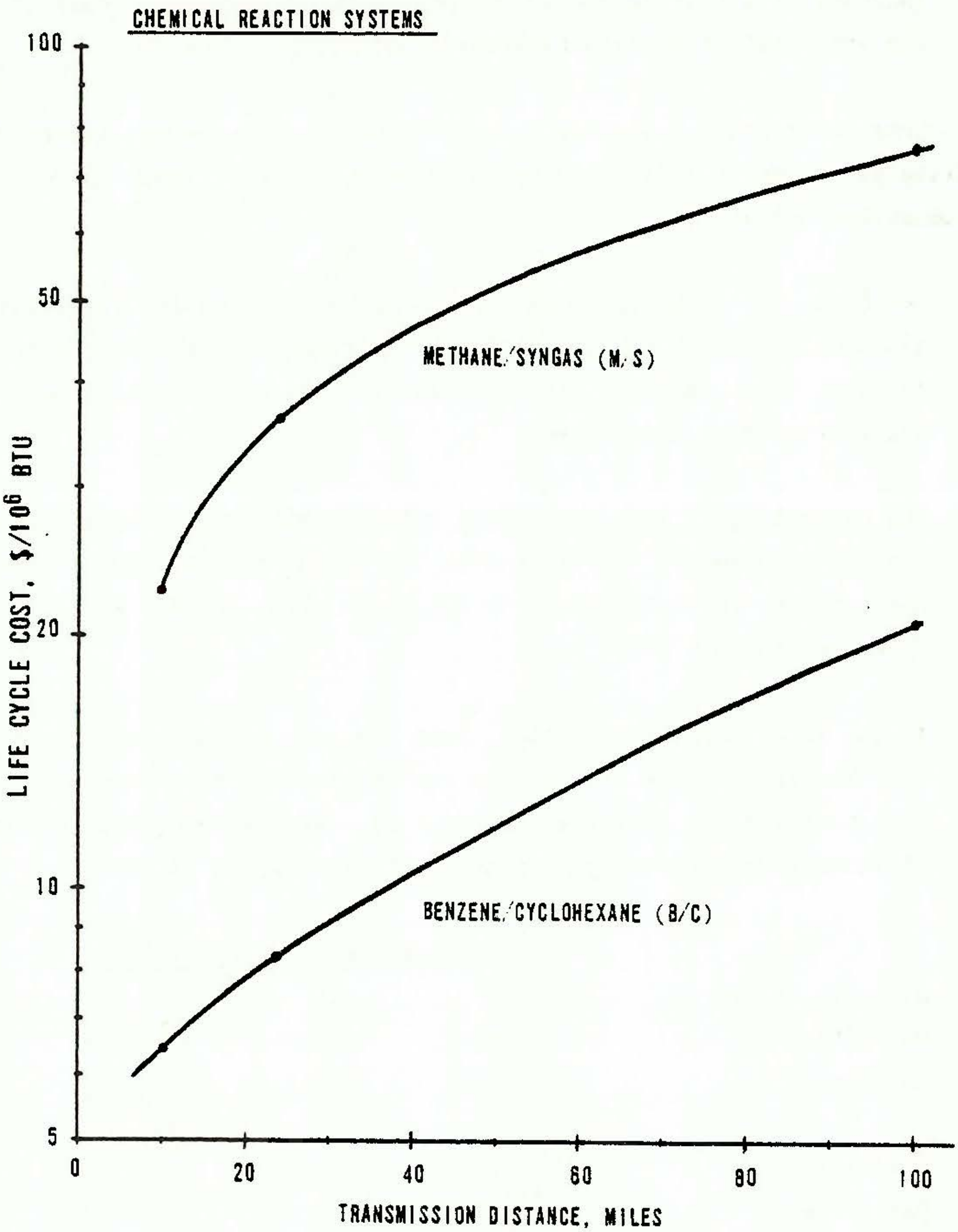


FIGURE 3.2-3
LIFE CYCLE COST VS. TRANSMISSION DISTANCE: CRSs

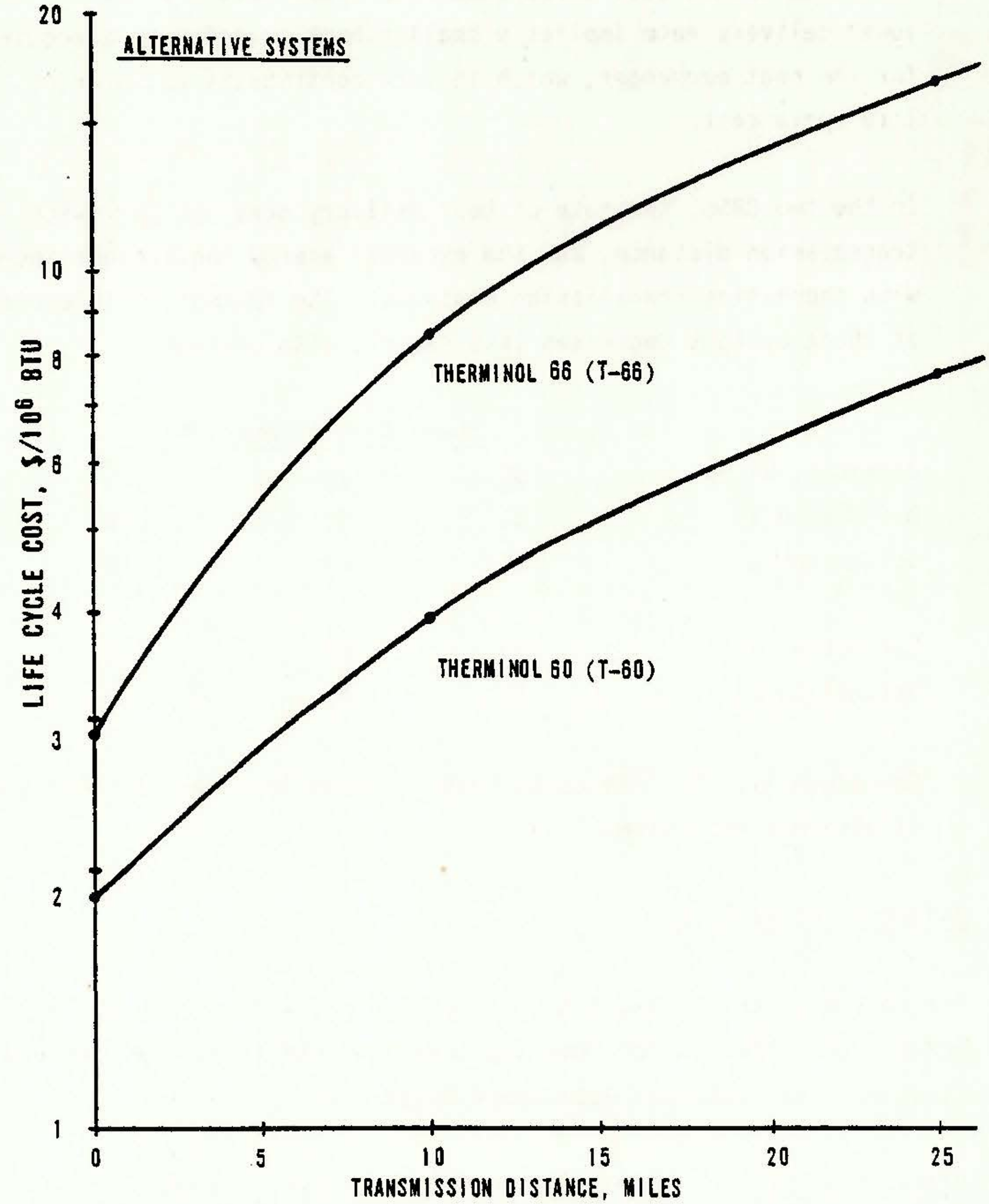


FIGURE 3.2-4

LIFE CYCLE COST VS. TRANSMISSION DISTANCE: ALTERNATIVES

The life cycle cost increases with a decrease in the thermal efficiency. However, this effect is partly offset by a reduction in the equipment cost associated with lower delivery rates. A lower delivery rate implies a smaller heat transfer area requirement for the heat exchanger, which in turn contributes to lowering the life cycle cost.

In the two CRSs, the rate of heat delivery does not vary with transmission distance, but the external energy requirement increases with increasing transmission distance. The thermal efficiencies of these systems decreases less rapidly with distance:

		Thermal Efficiency, %		
	Distance, Miles	10	25	100
M/S	Definition 1	31	30	29
	Definition 2	21	19	15
B/C	Definition 1	58	57	50
	Definition 2	50	47	30

Consequently, the life cycle cost increase less rapidly than those of alternative systems.

3.2.3 Effect of Plant Life

The cost sensitivity was analyzed for two plant lives, 20 and 30 years. The fixed charge rate and levelized O&M factors as derived from the EPRI Technical Assessment Guide are:

Plant Life, Yr.	Fixed Charge Rate	Levelized O&M Factor
20	0.1559	1.6593
30	0.1463	1.9350

The life cycle costs were determined using these factors and are shown in Table 3.2-2, Column(c). As can be seen, the plant life does not have a significant and consistent effect on life cycle cost. The T-66 and T-60 systems show a slight decrease in life cycle cost with plant life, whereas the other systems show a small increase with plant life. This is largely due to the counterbalancing variations in the fixed charge rate and O&M cost factor. The slight decrease in life cycle cost with plant life in the T-66 and T-60 systems is due to the relatively high capital requirements and low operating costs than their corresponding CRSs.

3.2.4 Other Factors

For the base case Benzene/Cyclohexane CRS, the effect of hydrogen loss on the life cycle cost was studied. It was assumed that any hydrogen loss from compressors, valve seals, reactor seals, and/or exhcanger seals would be made up by adding an equivalent quantity of cyclohexane to the CRS, generating make-up hydrogen by the following reaction in the dehydrogenation reactor:

$$C_6H_{12} \rightarrow C_6H_6 + 3H_2$$

This reaction would produce excess benzene which would have to be disposed of on a periodic basis.

It was determined that the rate of hydrogen loss under typical process conditions (at a temperature of 100°F and pressure of 365 psia) could be as high as 210 lb/(ft^2) (sec). Daily hydrogen loss from a 1 square millimeter hole, for example, would therefore be equivalent to 345 gallons of cyclohexane loss, or about 0.24% of cyclohexane inventory (144,000 gallons) in the CRS.

The life cycle energy cost (LCEC) can thus be expressed as a function of equivalent cyclohexane loss/day (percent of inventory = x), depending on the method of benzene disposal:

- o LCEC, ($\$/10^6$ Btu) = 8.40 + 1.45 x,
  if no credit is given to the excess benzene produced, or

- o LCEC, ($\$/10^6$ Btu) = 8.40 + 0.79 x,
  if 50¢/gal. credit is given to the excess benzene produced.

Assuming a daily cyclohexane loss of 0.1%, the life cycle cost would increase from \$8.40 to between \$8.48 and \$8.55 per 10^6 Btu.

SECTION 4.0
COMPARISON OF CRS WITH ALTERNATIVE TECHNOLOGY

This section compares each of the three CRSs with its commercially available alternative system. The comparison covers four categories -- process conditions, process efficiency, economics, and the advantages/disadvantages.

4.1 METHANE/SYNGAS CRS VS. THERMINOL 66 (T-66)

Table 4.1-1 summarizes the technical and economic results of the two processes.

Process Summary: The CRS requires a high temperature waste heat source (1520^{o}F) to achieve favorable chemical equilibrium. The T-66 system has no such restraint and could be designed for a lower temperature heat source if necessary. The CRS can utilize only a fraction of the waste heat available. More than half of the waste heat absorbed by the CRS is lost via air cooling of the reactor effluent, since there is no use for this heat. The T-66 system does not incur this loss and can utilize most of the heat from the heat source. However, it suffers a substantial loss in the transmission pipeline, proportional to the distance.

At the user end, the CRS produces high pressure steam while the T-66 produces only hot water. The difference in the quality of product is due to the temperature level and method of releasing heat. In the CRS, the exothermic heat of reactor is released at relatively constant high temperature (550-760^{o}F), while the T-66 system delivers sensible heat as it cools (550-140^{o}F).

Process Efficiency: Both systems require external electrical energy to operate pumps, compressors, and blower. Using a thermal energy to electric energy conversion efficiency of 30%, the CRS requires 13.2 x 10^6 Btu/hr external energy, while the alternative system requires 17.2 x 10^6 Btu/hr.

TABLE 4.1-1

SYSTEM COMPARISON: METHANE/SYNGAS VS. THERMINOL 66

	CRS METHANE/SYNGAS	ALTERNATIVE THERMINOL 66
I. Process Summary		
Charging Section		
Heat source	Municipal Incinerator Flue Gas	Municipal Incinerator Flue Gas
Operating period	16 hrs/day; 6 days/wk	16 hrs/day; 6 days/wk
Flow rate, 10^6 lbs/hr	0.27	0.27
Temp., ^oF, In/Out	1700/667	1700/440
Press., psia, In/Out	Atmospheric	Atmospheric
Heat absorption method	Endothermic Reaction (Steam Reforming)	Sensible Heat (Absorption)
Heat absorption medium	Methane/Steam	Therminol 66
Flow rate, 10^6 lbs/hr	0.04	0.31
Temp., ^oF, In/Out	900/1520	60/650
Press., psia, In/Out	620/588	700/570
Heat absorbed, 10^6 Btu/hr	83.1	93.3
Transmission Section		
Distance, miles	25	25
Pipeline	Underground	Aboveground
Pipe diameter		
Forward line, in., (No.)	8 (1)	8 (1)
Return line, in., (No.)	8 (1)	8 (1)
Discharging Section		
Heat release method	Exothermic Reaction (Methanation)	Sensible Heat (Release)
Heat release medium	Syngas	Therminol 66
Flow rate, 10^6 lbs/hr	0.03	0.31
Temp., ^oF, In/Out	500/760	550/120
Press., psia, In/Out	593/588	70/20
Heat released, 10^6 Btu/hr	29.3	66.5
User end	Existing Utility	Existing Utility
Heat absorption medium	BFW	BFW
Product	Steam	Hot Water
Operating Period	16/hrs/day; 6 days/wk	16 hrs/day; 6 days/wk
Flow rate, 10^3 lbs/hr	23.0	150.3
Temp., ^oF, In/Out	200/486	60/486
Press, psia, In/Out	593/588	650/645

TABLE 4.1-1 (Cont'd)

	CRS METHANE/SYNGAS	ALTERNATIVE THERMINOL 66
II. Process Efficiency Summary		
Heat absorbed, 10^6 Btu/hr	83.1	93.3
Heat released, 10^6 Btu/hr	29.3	66.5
Ext. energy req'd[a], 10^6 Btu/hr	13.2	17.2
Eff. of heat recovery, %		
Definition 1[b]	30.4	60.2
Definition 2[c]	19.4	52.8
III. Economic Summary		
Plant life	30	30
Capital costs, 10^6 \$		
Charging & discharging section	6.644	12.221
Pipeline	13.114	14.082
Insulation	-	4.241
Total	19.758	30.544
Operating costs (1st yr.), 10^6 \$	0.830	0.555
Life cycle cost, \$/10^6 Btu	30.74	16.59

----

[a] Thermal energy to electric energy conversion efficiency of 30% assumed

[b] Definition 1 = (Heat released x 100)/(Heat absorbed + Ext. energy required)

[c] Definition 2 = (Heat released-Ext. energy required) x 100/(Heat absorbed)

By definition 2 (Table 4.1-1), the CRS efficiency is 19 percent while that of the T-66 system is 53 percent. (Of course, this is based strictly on efficiency definition and does not consider the quality of heat.) The efficiency of the T-66 system is higher because the loss of heat in the transmission pipe is not as great as the heat discarded by the CRS at the endothermic section.

Economic Summary: The T-66 system is more cpaital intensive, requiring a total of \$30.5 million versus \$21.2 million for the CRS. The single, most expensive item in the T-66 system, besides the pipeline, was the cost of the heat transfer oil itself. At \$8.51/gal, the total cost of T-66 required is \$6,561,000. The CRS requires only \$78,000 for initial catalyst and chemicals.

Though the conventional system requires higher initial capital investment, it has lower operation costs than the CRS. This is primarily due to the fact that fewer and less complex equipment is involved in the T-66 system, whereas the CRS is a more complicated system which requires additional manpower and higher level of maintenance effort.

Overall, because of lower heat losses discussed previously, the T-66 system delivers more than twice as much heat as the CRS. Consequently, the life cycle cost for unit heat delivery is \$16.59 per million Btu delivered compared to \$30.74 per million Btu delivered for the Methane/Syngas CRS.

Advantages/Disadvantages: The major advantage of the Methane/Syngas CRS over the conventional T-66 system is its insensitivity to ambient temperature and transmission distance. Since the heat "carrier" is delivered to the user end at ambient temperatures, there is no substantial heat loss during transmission and the same quantity of heat (exothermic heat of reaction) can be recovered irrespective of the distance. This is not possible with the T-66 system in which heat losses through the pipewall can be significant and hence requires expensive insulation. Thus the heat delivered by the T-66 system is sensitive to ambient temperature and transmission distance. The second advantage of the CRS system is the quality of heat delivered at the user end. The Methane/Syngas system can deliver high temperatures, up to about 750°F, at any distance, whereas the T-66 system cannot achieve such high temperatures. Due to the characteristics of Therminol 66, the maximum temperature the oil can be heated to is 650°F. Thus, depending on the insulation, ambient temperature, and transmission distance, the maximum temperature at the user end can be much less than 650°F.

The CRS also has an operational advantage over the T-66 system. Since the heat source operates only 16 hours a day and six days a week, both the heat source and charging sections will be started and shut down at the same time. As soon as the dehyrogenation reactor in the Charging Section operates, the dehydrogenation reactor in the Discharging Section will experience the flow of chemicals and hence go into operation, releasing heat to the user. Thus there is no time lag between the heat source and the user end. This operational characteristic of the CRS exists irrespective of the distance between the two ends. Similarly, there is no time lag during shutdown. On the other hand, there is always a time lag in the T-66 system. Since heat is "carried" by the transfer oil, the heated oil has to travel the whole distance between the heat source and the user end. In the base case designed in this study, the user end will receive heat only about seven hours after the heat source is operated. Also, even if the heat source is shut down, the heat transfer oil will still have to be pumped for the next seven hours until all the heated oil reaches the user end. Hence, even though the heat source was available only 16 hours a day, the plant will have to be operated 23 hours or essentially all day.

The T-66 system cannot operate at ambient temperatures below 15°F but the CRS can operate below 15°F because no liquid transmission is involved.

The CRS requires a high temperature to operate (1520^{o}F); whereas, the T-66 system could operate with waste heat sources of about 700^{o}F. This reduces the range of application of the CRS.

The CRS is not proven in operation, even on a pilot plant scale. Except for the long transmission line, the T-66 system is commercially available and reliably operated in industry today.

4.2 BENZENE/CYCLOHEXANE CRS VS. THERMINOL 60 (T-60)

Table 4.2-1 summarizes the technical and economic results of the two systems.

<u>Process Summary</u>: The heat source is exit gas at 1300^{o}F from a cement kiln that is operated continuously. The exit gas passes through a high temperature precipitator before being vented to atmosphere via a chimney. The precipitator operates at about 600^{o}F and hence the gases existing the reactor in the CRS or the T-60 heater should be at this temperature. Therefore, the heat available for both the CRS and the alternative is 134 x 10^{6} Btu/hr.

In the CRS the catalytic reaction of cyclohexane dehydrogenation takes place above 500^{o}F, while the Therminol 60 oil absorbs sensible heat. The hydrogen, benzene and some unreacted cyclohexane from the CRS reactor exit at 820^{o}F; are cooled and pumped to the user end at ambient temperatures. In the alternative system the T-60 oil is heated to 600^{o}F and then pumped at the elevated temperature to the user end. Since the products from the CRS are two phases, liquid (benzene) and gas (hydrogen), two pipelines are needed for the forward line. An 8-inch diameter pipeline facilitates hydrogen flows and a 5-in diameter for the liquid. The return line is a 5-inch diameter since only liquid (cyclohexane) is returned. In the alternative system only liquid (T-60) is pumped in both directions. An 8-inch insulated line is required for the forward pipeline while an uninsulated 8-inch pipeline suffices for the return pipeline.

TABLE 4.2-1

SYSTEM COMPARISON: BENZENE/CYCLOHEXANE VS. THERMINOL 60

	CRS BENZENE/CYLOHEXANE	ALTERNATIVE THERMINOL 60
I. Process Summary		
Charging Section		
Heat source	Cement Kiln Exit Gas	Cement Kiln Exit Gas
Operating period	Continuous	Continuous
Flow rate, 10^6 lbs/hr	0.68	0.68
Temp., ^oF, In/Out	1300/600	1300/600
Press., psia, In/Out	Atmospheric	Atmospheric
Heat absorption method	Endothermic Reaction (Dehydrogenation)	Sensible Heat (Absorption)
Heat absorption medium	Cyclohexane	Therminol 60
Flow rate, 10^6 lbs/hr	0.14	0.49
Temp., ^oF, In/Out	500/820	60/600
Press, psia, In/Out	410/390	900/875
Heat absorbed, 10^6 Btu/hr	134.0	134.0
Transmission Section		
Distance, miles	25	25
Pipeline	Underground	Aboveground
Pipe diameter		
Forward line, in., (No.)	8 (1), 5 (1)	8 (1)
Return line, in., (No.)	5 (1)	8 (1)
Discharging Section		
Heat release method	Exothermic Reaction (Hydrogenation)	Sensible Heat (Release)
Heat release medium	Benzene & H_2	Therminol 60
Flow rate, 10^6 lbs/hr	0.14	0.49
Temp., ^oF, In/Out	575/700	500/120
Press., psia, In/Out	355/340	75/50
Heat released, 10^6 Btu/hr	92.5	92.5
User end	Process Steam or Existing Utility	Existing Utility
Heat absorption medium	BFW	BFW
Product	Steam	Hot Water
Operating Period	Continuous	Continuous
Flow rate, 10^3 lbs/hr	89.4	233.0
Temp., ^oF, In/Out	200/446	60/446
Press., psia, In/Out	410/400	425/410

TABLE 4.2-1 (Cont'd)

	CRS BENZENE/CYCLOHEXANE	ALTERNATIVE THERMINOL 60
II. Process Efficiency Summary		
Heat absorbed, 10^6 Btu/hr	134.0	134.0
Heat released, 10^6 Btu/hr	92.5	92.5
Ext. energy req'd[a], 10^6 Btu/hr	29.8	19.0
Eff. of heat recovery, %		
Definition 1[b]	56.5	60.5
Definition 2[c]	46.9	54.9
III. Economic Summary		
Plant life	30	30
Capital costs, 10^6 $		
Charging & discharging section	7.366	12.014
Pipeline	15.483	14.082
Insulation	-	4.241
Total	22.849	30.277
Operating costs (1st yr.), 10^6 $	1.62	0.75
Life cycle cost, $/$10^6$	8.40	7.63

----

[a] Thermal energy to electric energy conversion efficiency of 30% assumed

[b] Definition 1 = (Heat released x 100)/(Heat absorbed + Ext. energy required)

[c] Definition 2 = (Heat released-Ext. energy required) x 100/(Heat absorbed)

At the user end, the CRS delivers heat via the catalytic exothermic reaction of benzene hydrogenation, producing cyclohexane. The reaction takes place between 575°F and 700°F. In the alternative system, the T-60 experiences about a 100°F temperature drop over the 25-mile transmission and hence is available at only about 500°F at the user end. The temperature drop is due to the heat losses through the pipe walls, even with a 4-inch insulation. As shown in Table 4.1-2, both systems deliver the same quantity of heat to the user end, 92.5 x 10^6 Btu/hr. But, due to the method of heat delivery, the CRS produces 400 psia saturated steam (446°F) whereas the alternative T-60 system can produce only 400 psia saturated hot water at 446°F.

Process Efficiency: Both the CRS and T-60 system absorb and deliver the same quantity of heat, but the CRS requires 29.8 x 10^6 Btu/hr external energy, versus only 19.0 x 10^6 Btu/hr for the T-60 system to operate pumps, compressors, and blowers. Due to this difference, the overall efficiency of heat recovery (definition 2) for the CRS is 46.9% compared to 54.9% for the T-60 system.

Economic Summary: The alternative T-60 system requires a total capital investment of $30.3 million versus only $22.9 million for the CRS. Once again, the high investment for the T-60 system is due to the high cost of the heat transfer oil T-60 ($5.63 million) and the necessity to insulate the forward pipeline (insulation cost = $4.24 million). The CRS requires only about $453,000 for the chemicals and does not require any insulation of pipes. Due to the simplicity of the T-60 system, operating costs are less than half those of the CRS ($0.75 million vs. $1.62 million). The lower operating cost and slightly higher heat recovery overcome the higher capital charge, and the life cycle cost for the alternative is about $7.63 per million Btu delivered vs. $8.40 for the CRS. It should be noted, however, that the CRS delivers the heat in the form of 400 psi saturated steam (446°F) while the alternative T-60 system delivers heat as saturated hot water at 400 psia and 446°F.

Advantages/Disadvantages: The major advantage Benzene/Cyclohexane CRS has over the alternative is its capability to deliver the same quantity and quality of heat, independent of the transmission distance and ambient temperatures. The T-60 system is sensitive to transmission distance and ambient temperatures. In the present study, the T-60 oil temperatures drops about 4^{o}F per mile of transmission. Hence, beyond a certain transmission distance, the T-60 system would not be able to deliver heat at a specified temperature.

Another advantage of the Benzene/Cyclohexane CRS is that it does not have a time lag between the heat source and the user end as the T-60 system would have. This may not be a major problem in the present design because the heat source is available continuously, and hence there are no regular shutdowns or start-ups.

The CRS cannot operate at as low an ambient temperature as the T-60 system. The pipeline must be kept above 44^{o}F (freezing point of cyclohexane). The T-60 system has no real restraint since the freezing point of T-60 fluid is -60^{o}F.

The major disadvantage of the CRS is that it is not commercially proven. There are uncertainties - reactor design, hydrogen losses, ability to control process, catalyst life and side reactions.

4.3 SULFURIC ACID/WATER CRS VS. PRESSURIZED HOT WATER (HW)

Table 4.3-1 summarizes the primary technical and economic results of these two processes.

Process Summary: In both systems the same amount of heat, 22.95 x 10^{6} Btu/hr, is absorbed (during the first 16 hours of operation). In the CRS, the heat is absorbed to concentrate the acid solution by driving out the water in the form of steam. The steam, produced

TABLE 4.3-1

SYSTEM COMPARISON: SULFURIC ACID/WATER VS. PRESSURIZED HOT WATER

	CRS SULFURIC ACID/WATER	ALTERNATIVE PRESSURIZED HOT WATER
I. Process Summary		
Charging Section		
Heat source	Fuel Cell Reject Steam	Fuel Cell Reject Steam
Operating period	16 hrs/day; 7 days/wk	16 hrs/day; 7 days/wk
Flow rate, 10^3 lbs/hr	26.4	26.4
Temp., ^oF, In/Out	350/350	350/350
Press, psia, In/Out	120/120	120/120
Heat absorption method	Sensible Heat and Heat of Concentration	Sensible Heat (Absorption)
Heat absorption medium	50% Sulfuric Acid	Pressurized Hot Water
Flow rate, 10^3 lbs/hr	39.3	122.0
Temp., ^oF, In/Out	150/320	140/325
Press., psia, In/Out	30/20	85/80
Heat absorbed, 10^6 Btu/hr	23.0	23.0
Discharging Section		
Heat release method	Sensible Heat and Heat of Dilution	Sensible Heat (Release)
Heat release medium	87% Sulfuric Acid	Pressurized Hot Water
Flow rate, 10^3 lbs/hr	78.5	244.0
Temp., ^oF, In/Out	385/150	320/140
Press., psia, In/Out	20/15	85/80
Heat released, 10^6 Btu/hr	11.0	44.8
User end	District Heating	District Heating
Heat absorption medium	Heating Water	Heating Water
Product	Hot Water	Hot Water
Operating period	Next 8 hrs. of the day	Next 8 hrs. of the day
Flow rate, 10^3 lbs/hr	60.5	247.3
Temp., ^oF, In/Out	100/280	100/280
Press, psia, In/Out	65/50	65/50

TABLE 4.3-1 (Cont'd)

	CRS SULFURIC ACID/WATER	ALTERNATIVE PRESSURIZED HOT WATER
II. Process Efficiency Summary		
Heat absorbed, 10^6 Btu/cycle	367.2	367.2
Heat released, 10^6 Btu/cycle	87.7	355.0
Ext. energy req'd[a], 10^6 Btu/hr	2.4	3.0
Eff. of heat recovery, %		
Definition 1[b]	23.7	95.9
Definition 2[c]	23.2	95.9
III. Economic Summary		
Plant life, years	30	30
Capital costs, 10^6 $	2.24	3.33
Operating costs (1st yr.), 10^6 $	0.14	0.23
Life cycle costs, $/$10^6$ Btu	18.67	7.14

----

[a] Thermal energy to electric energy conversion efficiency of 30% assumed

[b] Definition 1 = (Heat released x 100)/(Heat absorbed + Ext. energy required)

[c] Definition 2 = (Heat released-Ext. energy required) x 100/(Heat absorbed)

during the concentration of the acid solution, is condensed at just below boiling point of water and stored at atmospheric pressure. It is here that most of the inefficiency occurs since about 75% of absorbed heat is rejected. (The heat cannot be used for other purposes since the operation of the fuel cell supplies ample heat.)

In the alternative system, all the heat is transferred from source to the water as sensible heat only. There is practically no heat rejection except the heat losses during storage. The heated water is stored at $325^{o}F$ (80 psig).

During the eight hours when the fuel cell is not operating, the CRS releases the stored heat as heat of mixing, whereas the hot water system releases sensible heat. The temprature of the sulfuric acid solution during dilution reaches about $385^{o}F$ which is higher than the original source temperature ($350^{o}F$). This heats about 127 gpm of the district heating feedwater from $100^{o}F$ to $280^{o}F$, corresponding to about 11×10^{6} Btu/hr. In the alternative system, about 500 gpm of the district heating water is heated from $100^{o}F$ to $280^{o}F$, corresponding to 44.8×10^{6} Btu/hr.

Process Efficiency: The CRS requires 2.4×10^{6} Btu/hr of external energy to drive pumps and blower while the alternative system requires 3.0×10^{6} But/hr of external energy. Based on definition 1 (Table 4.1-3), the CRS has a thermal efficiency of 23.7 percent versus 95.9 percent for the alternative system and, based on definition 2, the CRS has an efficiency of 23.2 percent versus 95.9 percent for the alternative system. As was pointed out earlier, this vast difference is primarily due to the heat lost in condensing the steam produced during the charging cycle in the CRS. About 75 percent of the heat absorbed is lost in this cooling process, hence the low efficiency.

Economic Summary: Both systems have relatively low capital investment requirements, due basically to the lack of long distance transmission. The alternative system's capital and operating requirements are 50 percent higher than those for the CRS, but due to its higher efficiency, the life cycle cost is much lower. For a plant life of 30 years, the life cycle cost for the CRS is $18.67 per million Btu delivered, compared to only $7.14 per million Btu for the alternative system.

Advantages/Disadvantages: The Sulfuric Acid/Water CRS does not compare well with its alternative, Pressurized Hot Water system, as a thermal storage system. Its only advantage over the alternative system is its capability to deliver heat at a slightly higher temperature level than the source temperature. The hot water system, besides being more efficient, is also much simpler, its technology commercially developed and available. The CRS, due to the acid, requires special alloys and rubber-lined tanks, whereas the alternative system uses the more conventional and inexpensive metals.

SECTION 5.0
CRS MARKET POTENTIAL

In order to be commercially attractive, a CRS must at least meet two criteria. First, the cost for recovered energy must be lower than the cost of supplying energy by burning fuel. Second, the CRS must recover heat at a lower cost than would be possible with a conventionally available system, or it must recover heat in a more useful form.

Average costs for various fuels on a mid-1979 basis are given below: (5-1)

	Delivered Cost, $\$/10^6$ Btu	
	Industrial	Utility
Natural Gas	2.00	1.55
Residual Fuel Oil	2.46	2.24
Coal	N.A.	1.18

To be useful in this study, current fuel costs must be translated to life cycle cost over a thirty-year plant life. This can be done by multiplying current costs by 1.935, to account for an assumed annual 6.0% inflation/ 0.2% escalation. The average efficiency of industrial furnaces is assumed to be 80%. Purchased fuel costs must be divided by 0.80 to express costs on a usable energy basis. With these two adjustments, the cost of energy by burning fuel would range from $\$2.90\text{-}6.00/10^6$ Btu for industrial and electric utility users, depending on fuel type.

If it is assumed that natural gas continues to approach fuel oil prices, then the cost of energy (excluding coal and nuclear) would be in the $\$5.46\text{-}6.00/10^6$ Btu range. This latter cost is used to gauge the attractiveness of waste heat recovery via a CRS.

The future cost of fuel is subject to uncertainty, but it is generally concluded that fuel costs will escalate in relation to other commodities. If it is assumed that fuel costs will escalate at 2.4% per year relative to other costs, then the multiplier is 2.606 to convert current costs to a levelized 30 year cost [5-2]. In this case, fuel costs would be \$7.30-8.00/10^6 Btu. It should be noted that there is no capital cost considered for burning the fuel; an existing facility is assumed.

Table 5-1 gives the costs for energy storage or transport for each CRS and its commercially available alternative. Costs are shown for the range of capacities and distances discussed in Section 3.2.

METHANE/SYNGAS CRS (Transport)

Figure 5-1 shows the cost of recovering, transporting, and delivering waste heat with the CRS versus transmission distance. Also shown in a similar curve for the commercially available alternative (using T-66 organic fluid). The cross-hatched bars represent the cost of obtaining energy by burning purchased fuel. The lower cost is assuming 0.2%/year escalation of fuel costs; the upper cost is based on 2.4%/year escalation.

These results indicate that at the base capacity the CRS is not competitive with purchased fuel even assuming 2.4%/year escalation of fuel costs.

Furthermore, even if the CRS has a capacity of five times the base case, it would still not be competitive.

The CRS is more economical than the alternative system only above 65 miles transmission distance. Below 65 miles, the CRS would be more useful if the user required steam rather than hot water and the waste heat was available about 1700^{o}F.

In summary, it does not appear that there will be significant market potential for the Methane/Syngas CRS.

TABLE 5-1

LIFE CYCLE COST ($/10^6 BTU)

	Capacity (Relative Factor)				Distance (Miles)			
	0.5	1.0	2.0	5.0	0	10	25	100
CRS								
Methane/Syngas	41.72	30.74	23.51	16.55	-	22.33	30.74	75.69
Benz./Cyclohex.	10.82	8.40	6.70	5.18	-	6.39	8.40	20.26
S.A./Water	24.45	18.67	14.43	10.89	18.67	-	-	-
Alternatives								
T-66	25.06	16.59	12.08	8.87	2.87	8.48	16.59	-
T-60	11.56	7.63	6.34	4.90	1.86	3.97	7.63	-
Hot Water	7.85	7.14	6.25	5.79	7.14	-	-	-

Notes: 1. Plant life is 30 years.

2. 6% inflation and 0.2% escalation assumed during 30-year life.

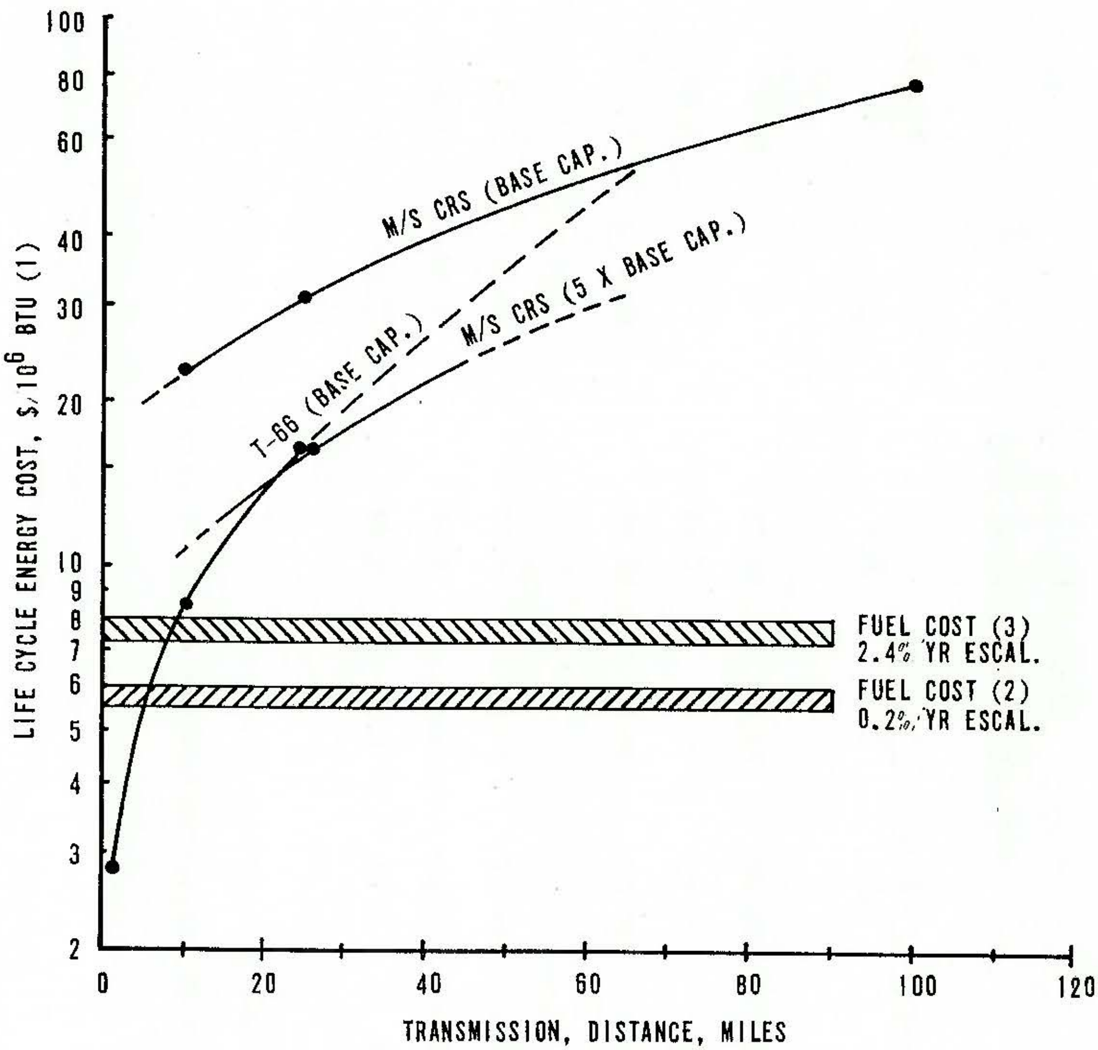


FIGURE 5-1
METHANE/SYNGAS CRS
ECONOMIC COMPARISON WITH FUEL COST AND WITH ITS COMMERCIALLY AVAILABLE ALTERNATIVE SYSTEM

NOTES:

1. LEVELIZED OVER 30 YEARS, ASSUMING 6% INFLATION & 0.2% ESCALATION OF FUEL COSTS.
2. LEVELIZED OVER 30 YEARS (80% EFFICIENCY OF USE), ASSUMING 6% INFLATION & 0.2% ESCALATION OF FUEL COSTS.
3. LEVELIZED OVER 30 YEARS (80% EFFICIENCY OF USE), ASSUMING 6% INFLATION & 2.4% ESCALATION OF FUEL COSTS.

BENZENE/CYCLOHEXANE CRS (Transport)

Figure 5-2 shows the cost of recovering, transporting, and delivering waste heat with the CRS versus transmission distance. Also shown is a similar curve for the commercially available alternative (using T-60 organic fluid) and the cost of obtaining energy from burning purchased fuel.

At the base capacity, the CRS is not really competitive with purchased fuel unless there is appreciable escalation of fuel costs. If fuel costs should escalate at 2.4% per year relative to other costs, then the Benzene/Cyclohexane CRS would be competitive at transmission distances below about 20 miles.

If a larger capacity CRS application is assumed, then the economy of scale would make the CRS more economical. At a capacity of twice the base case, it would be economical below about 17 miles. Assuming 2.4%/year fuel cost escalation, it would be economical below about 34 miles. The curve for five times base capacity is shown, but this case represents a very large size and is not considered representative of waste heat sources available today.

Below about 30 miles, the alternative system is more economical. However, the T-60 system cannot deliver steam, but only hot water. The alternative system would be preferred in cases where the user requires sensible heat delivery, rather than steam.

In summary, the Benzene/Cyclohexane CRS has a possibility of economic application. It is the most economical CRS studied. However, the application would have to meet the following criteria:

1. Waste Heat source is above 1000^oF
2. Waste heat source is essentially continuous
3. Waste heat source is $>150 \times 10^6$ Btu/hr
4. Transporting distance is less than about 25 miles

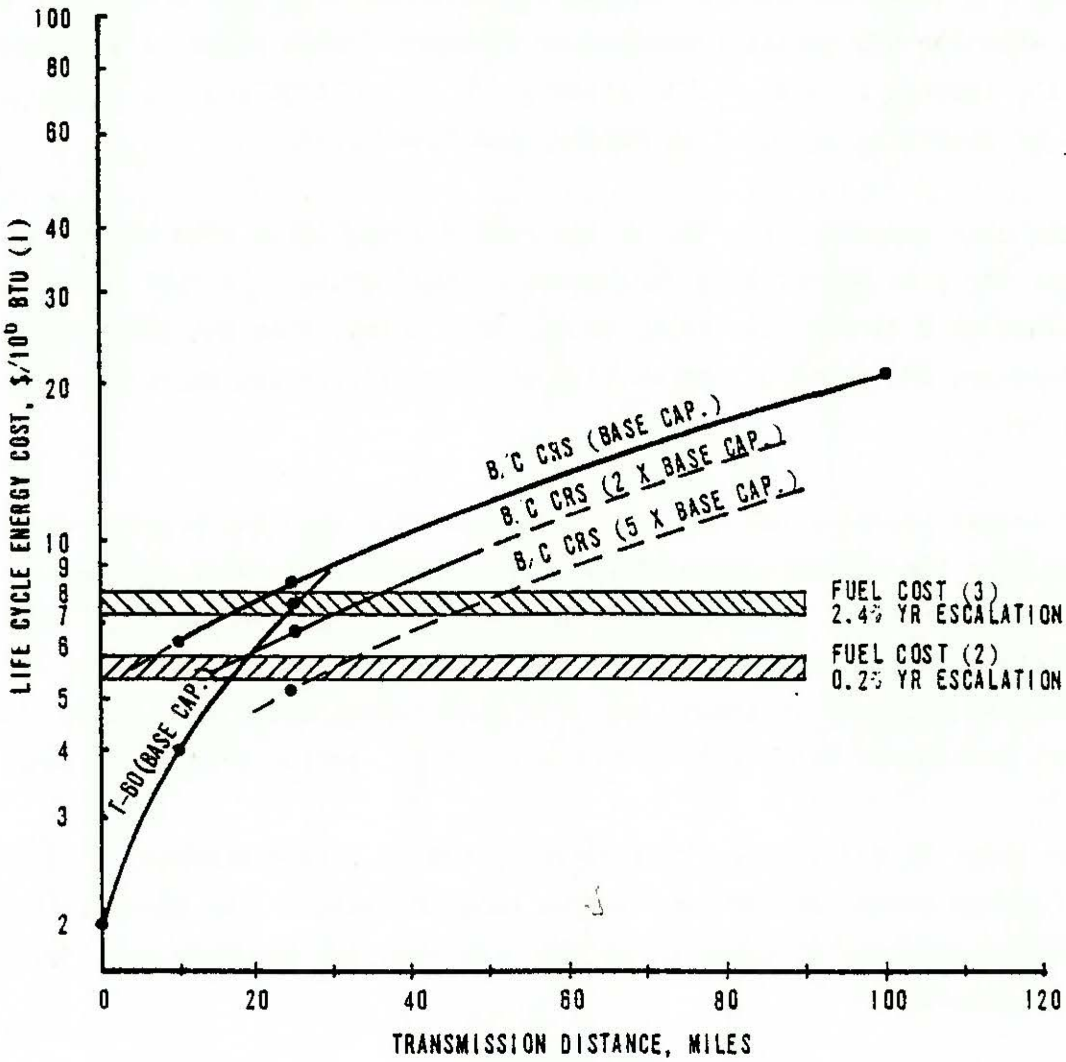


FIGURE 5-2
BENZENE/CYCLOHEXANE CRS
ECONOMIC COMPARISON WITH FUEL COSTS AND WITH ITS
COMMERCIALLY AVAILABLE ALTERNATIVE SYSTEM

NOTES:

1. LEVELIZED OVER 30 YEARS, ASSUMING 6% INFLATION & 0.2% ESCALATION OF FUEL COSTS.
2. LEVELIZED OVER 30 YEARS (80% EFFICIENCY OF USE), ASSUMING 6% INFLATION & 0.2% ESCALATION OF FUEL COSTS.
3. LEVELIZED OVER 30 YEARS (80% EFFICIENCY OF USE), ASSUMING 6% INFLATION & 2.4% ESCALATION OF FUEL COSTS.

5. User requires steam rather than hot water
6. Temperature inside transmission pipe is kept above 44^{o}F (freezing point of cyclohexane).

Because of all the above qualifications, it is estimated that the market potential for the Benzene/Cyclohexane CRS is rather limited. The need for joint action between owner of waste heat source and energy user on a pioneeing venture would be a deterrent to use of CRS. This means that the cost for delivered energy would have to be well below fuel costs before a project would be implemented.

SULFURIC ACID/WATER CRS (Storage)

Reference to Table 5-1 shows that the cost of storing energy with the CRS is more than double that with the alternative hot water system over a wide capacity range. Both systems deliver heat at approximately the same temperature. Therefore, there does not appear to be any potential market for the Sulfuric Acid/Water CRS.

SECTION 6.0
CONCLUSIONS AND RECOMMENDATIONS

SPECIFIC CONCLUSIONS ON CRSs STUDIED

The Methane/Syngas CRS is not expected to offer economical application for heat recovery and transport within the 1980-2000 period. Its efficiency is low and it is restricted to a very high temperature heat source.

The Benzene/Cyclohexane CRS is the most promising of those studied. But, for heat recovery and transport, it is not competitive versus fuels (natural gas or oil) at current prices nor against a commercially available alternative system. However, it can delivery high pressure steam, which the alternative system cannot do. This CRS has the potential for becoming economical in selected applications within the 1980-2000 period, should fuel costs escalate considerably above current levels.

The Sulfuric Acid/Water CRS was found to offer no great advantage over a pressurized hot water system for thermal storage. It can deliver heat at a temperature slightly higher than the waste heat source, which the conventional system cannot do. However, its much lower heat recovery efficiency results in higher costs than with the conventional system. It does not appear that this CRS will find use in thermal storage.

GENERAL CONCLUSIONS ON CRS

The concept of CRS for thermal energy storage or transport has been shown to be technically feasible in several individual applications. It can deliver useful heat over a considerable distance. In general, CRS appears to have the best potential when coupled with a large, continuous heat source for transport at 10 to 30 mile distances.

None of the CRSs studied are economical versus fuels at present prices. Should fuel costs escalate at a relatively high rate within the 1980-2000 period, some CRS applications might prove economically competitive.

Limited availability of heat sources at a sufficiently high temperature to drive the endothermic reaction would hinder commercial application. High temperature waste heat would be recovered on-site whenever possible. Only when efficient on-site use cannot be practiced would CRS be needed. Also working against CRS use is the requirement for a joint project between two locations - source and user. The CRS would have to be more than marginally economical before a project would be undertaken.

RECOMMENDATIONS

- The Methane/Syngas CRS might be considered for future thermal transport applications at much larger capacities than in present study.

- Since the Benzene/Cyclohexane CRS appears to have some potential for future applications, further definition of sources of waste heat available to drive endothermic reaction would be useful. Work on defining reactor designs and catalyst development might also be useful.

SECTION 7.0
REFERENCES

2-1 Gilbert Associates, Inc., An Assessment of the Use of Chemical Reaction Systems in Electric Utility Applications, Phase I, Final Report prepared for EPRI, Contract RP 1086-1, December, 1978.

2-2 General Electric Co., Closed Loop Chemical Systems for Energy Storage and Transmission (Chemical Heat Pipe), Final Report prepared for U.S. Dept. of Energy, ERDA Contract Ey-76-C-02-2676, Schenectady, New York, February, 1978.

2-3 Mathtech, Inc., An Analysis of the Application of Fuel Cells in Dual Energy Use Systems, Prepared for EPRI, November, 1978.

2-4 Rocket Research Corp., Sulfuric Acid/Water Chemical Energy Storage System, Report 76-R-520 prepared for U.S. ERDA, Redmond, Washington, May, 1976.

3-1 Electric Power Research Institute, The EPRI Technical Assessment Guide, Palo Alto, California, June, 1978.

3-2 Guthrie, K.M., "Capital Cost Estimating", Chemical Engineering, March 24, 1969, p. 114-142.

5-1 U. S. Department of Energy, Monthly Energy Review, March, 1979.

5-2 Electric Power Research Institute, The EPRI Technical Assessment Guide, Palo Alto, California, June, 1978.

APPENDIX A

INSULATION SELECTION CRITERIA

Insulation generally should meet the following four criteria:

1. The material must be compatible with the chemicals being processed. It should be essentially inert regarding flash-point depression and similar properties, when chemicals may contact the insulation.

2. Both the insulation and its barrier system should resist mechanical abuse over the life of the plant.

3. The barrier system has to be effective, allowing little or no water leakage. Penetration of the insulation by water seriously diminishes its insulating properties until it dries out.

4. Installation of the insulating material on valves and pipe fittings should be as easy as possible. Normally, the cost of insulating a valve is about three times that of insulating a similar length of straight pipe.

Since in the T-66 and T-60 system the insulation is for outdoor and aboveground use, it should be designed to withstand abuse from weather, man and animal. The temperature range encountered in this study is 0^{o}-650^{o}F.

The most important type of insulation in this range is fiber-reinforced hydrous calcium silicate. Insulation type used in this service range include:

1. Calcium silicate
2. Diatomaceous silica
3. Cellular glass (to about 850^{o}F)

4. Glass fiber bonded with high-temperature binders
5. Magnesium carbonate with asbestos or other fibers and binders (to about $600^{\circ}F$)
6. Rock wood or mineral-derived fibers.
7. Expanded perlite with binders.
8. Metal-sheet reflective systems.

Selection of material in this temperature range is much more strongly dictated by the value of thermal conductivity than it is in the lower temperature ranges. But other factors, such as mechanical properties, forms available and installation cost must also be taken into consideration.

Tables A-1 and 2 list the representative pipe insulations commerically available and the secondary properties of these insulations.

TABLE A-1

REPRESENTATIVE PIPE INSULATIONS+

Insulation Type	Temperature range, °F	Conductivity, k Btu/(h)(ft^2)(°F/in.)	Density lb/ft^3	Applications
Urethane foam	-300 to 300	0.11 to 0.14	1.6 to 3.0	Hot and cold piping
Cellular glass blocks	-350 to 500	0.20 to 0.75	7.0 to 9.5	Tanks and piping
Fiberglass blanket for wrapping	-120 to 550	0.15 to 0.54	0.60 to 3.0	Piping and pipe fittings
Fiberglass preformed shapes	-60 to 450	0.22 to 0.38	0.60 to 3.0	Hot and cold piping
Fiberglass mats	150 to 700	0.21 to 0.38	0.60 to 3.0	Piping and pipe fittings
Elastomeric pre-formed shapes and tape	-40 to 220	0.25 to 0.27	4.5 to 6.0	Piping and pipe fittings
Fiberglass with vapor barrier jacket	20 to 150	0.20 to 0.31	0.65 to 2.0	Refrigerant lines, dual-temperature lines, chilled-water lines, fuel-oil piping
Fiberglass without vapor barrier jacket	to 500	0.20 to 0.31	1.5 to 3.0	Hot piping
Cellular glass blocks and boards	70 to 900	0.20 to 9.75	7.0 to 9.5	Hot piping
Urethan foam blocks and boards	200 to 300	0.11 to 0.14	1.5 to 4.0	Hot piping

TABLE A-1 (CONT'D)

REPRESENTATIVE PIPE INSULATIONS+

Insulation Type	Temperature range, °F	Conductivity, k Btu/(h)(ft^2)(°F/in.)	Density lb/ft^3	Applications
Mineral-fiber preformed shapes	to 1,200	0.24 to 0.63	8.0 to 10.0	Hot piping
Mineral-fiber blankets	to 1,400	0.26 to 5.6	8.0	Hot piping
Fiberglass field applied jacket for exposed lines	500 to 800	0.21 to 0.55	2.4 to 6.0	Hot piping
Mineral-wool blocks	850 to 1,800	0.36 to 0.90	11.0 to 18.0	Hot piping
Calcium silicate blocks	1,200 to 1,800	0.33 to 0.72	10.0 to 14.0	Hot piping

+ These data reproduced from "Cost-Effective Thermal Insulation," by Harrison and Pelanne, Chemical Engineering, December 19, 1977.

TABLE A-2

SECONDARY PROPERTIES OF REPRESENTATIVE INSULATION+

Property	Calcium Silicate	Mineral Fiber	Fiberglass	Elastomeric, Closed Cell	Cellular Glass	Urethane Foam
Alkalinity, pH	8 to 12.5	7 to 9	7.5 to 9	-	7 to 8	6.5 to 7.5
Capillarity	Will wick	Will wick	Neg.	None	Neg.	None
Flame-spred index	-	Less than 25	Less than 25	25 to 75	-	75 or less
Smoke-density index	-	Less than 50	Less than 50	150 to 490	-	450 or less
Non-combustible	N.C.	Some N.C.	-	-	N.C.	-
Compressive strength lb/in^2 @ % deformation	100 to 250 @ 5%	1 to 18 @ 10%	0.02 to 3.4 @ 10%	2 to 6 @ 25%	100 @ 5%	16 to 100 @ 5%
Specific heat, Btu/(lb)(^{o}F)	0.20 to 0.28	0.22	0.20	0.19 to 0.27	0.20	0.23 to 0.25

+These data reproduced from "Cost-Effective Thermal Insulation", by Harrison and Pelanne, Chemical Engineering, December 19, 1977.

APPENDIX B

BASIS FOR COST SENSITIVITY ANALYSIS

The basis for cost sensitivity analysis performed in Section 3.2 is summarized in this appendix.

B-1 Effect of System Capacity

Capital Requirement

1. The size and number of major equipment items, as well as the scaling factors (or exponents) used in cost estimation, are listed in Tables B-1 through B-6.

2. For transportation CRSs or alternatives, the pipeline is sized so that the pressure drop remains the same as that in the base case. For Therminol-66 and Therminol-60 systems, insulation thickness is such that heat losses are $4^{o}F$/mile. (See Tables B-1 through B-6).

Operating Cost

1. The operating labor requirement for the plant is determined using a scaling factor of 0.35. An allowance of $70,000/yr (the same as the base case) is assumed for the maintenance labor, material, and support labor for the transmission section.

2. The operating requirement for electricity, water, catalysts, and chemicals of the plant are proportional to the system capacity.

Life Cycle Cost

For each system, the equation given in Section 3.1.2 is used to calculate the life cycle cost. The annual heat delivered is proportional to the system capacity.

TABLE B-1

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR

METHANE/SYNGAS CRS

I. Plant		Relative System Capacity				
		0.5	1 (Base)	2	5	
Equip. No.	Equipment	No. of Units[a]				Scaling Factor
R-1	Ref. reactor	1	1	1	2	0.65
E-1	Waste heat boiler	1	1	2	5	0.65
E-2	Flue gas cooler	1	1	1	2	0.80
C-1	Exhaust gas blower	1	1	1	2	0.82
E-3	Preheater 1	1	1	1	2	0.65
E-4	Preheater 2	1	1	1	1	0.65
E-5	Air-fin cooler	1	1	2	5	0.80
D-2, -4	Separators	1	1	1	1	0.65
P-1 to -3	Pumps	1	1	1	1	0.52
D-3, -5	K.O. drums	1	1	1	1	0.65
C-2	Compressor	1	1	1	1	0.82
R-2	Methan. reactor	1	1	1	1	0.65
D-6	Steam drum	1	1	1	1	0.60
E-7	Methan. preheater	1	1	1	2	0.65
H-1	Start-up heater	1	1	1	1	0.82
II. Transmission						
	Pipeline length, miles	25	25	25	25	
	Size, inches (no.)	6(2)	8(2)	10(2)	16(2)	

(a) Number listed indicates operating units.

TABLE B-2

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR

BENZENE/CYCLOHEXANE CRS

I. Plant		Relative System Capacity				
		0.5	1 (Base)	2	5	
Equip. No.	Equipment	No. of Units[a]				Scaling Factor
R-1	Reactor	2	2	4	10	0.65
TK-1, -2	Cyclohexane tank	1	1	1	1	0.63
P-1 to -4	Pumps	1	1	1	1	0.52
C-2 to -5	Compressors	1	1	1	1	0.82
D-1 to -4	V-L separators	1	1	1	1	0.65
E-1, -4	Preheaters	1	1	1	1	0.65
E-2	Air cooler	1	1	1	5	0.65
E-3	Trim cooler	1	1	1	1	0.80
R-2	Reactor	1	1	1	1	0.65
E-4	Feed preheater	1	1	1	1	0.65
E-5	Final cooler	1	1	1	1	0.65
H-1	Fired heater	1	1	1	1	0.85
D-5	Steam drum	1	1	1	1	0.60
C-1	Exhaust blower	1	1	1	2	0.82
II. Transmission						
	Pipeline length, miles	25	25	25	25	
	Pipe size, inches (no.)					
	Hydrogen	6(1)	8(1)	10(1)	16(1)	
	Benzene	4(1)	5(1)	8(1)	10(1)	
	Cyclohexane	4(1)	5(1)	8(1)	10(1)	

(a) Number listed indicates operating units.

TABLE B-3

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR
SULFURIC ACID/WATER CRS

Plant

Equipment No.	Equipment	Relative System Capacity 0.5	1 (Base)	2	5	Scaling Factor
		No. of Units[a]				
TK-1 to -3	Tanks	1	1	1	1	0.63
P-1 to -7	Pumps	1	1	1	1	0.52
E-1 to -3	Heat exchangers	1	1	1	1	0.65
D-1	Separator	1	1	1	1	0.65
E-5	Cooler	1	1	1	1	0.65
-	Evaporator package	1	1	1	1	0.75
D-2	Drum	1	1	1	1	0.60

(a) Number listed indicates operating units.

TABLE B-4

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR

THERMINOL 66 SYSTEM

I. Plant		Relative System Capacity				
		0.5	1 (Base)	2	5	
Equip. No.	Equipment	No. of Units[a]				Scaling Factor
B-1	Exhaust blower	1	1	1	1	0.82
E-1	Therminol heater	1	1	1	1	0.85
E-2	BFW heat exchanger	1	2	4	10	0.65
P-1, -2	Pumps	1	1	1	1	0.52
T-1	Therminol surge tank	1	2	2	3	0.63
T-2	Therminol surge tank	1	2	2	2	0.63
II. Transmission						
	Pipeline length, miles	25	25	25	25	
	Pipe size, inches (no.)	5(2)	8(2)	10(2)	16(2)	
	Insulation thick, inches (no.)	6½(1)	4(1)	3½(1)	2(1)	

(a) Number listed indicates operating units.

TABLE B-5

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR

THERMINOL 60 SYSTEM

I. Plant

Equip. No.	Equipment	Relative System Capacity 0.5	1 (Base)	2	5	Scaling Factor
		No. of Units[a]				
B-1	Exhaust blower	1	1	1	2	0.82
E-1	Therminol heater	1	1	1	1	0.85
E-2	BFW heat exchanger	1	3	6	15	0.65
P-1, -2	Pumps	1	1	1	2	0.52
T-1	Therminol surge tank	1	2	2	3	0.63
T-2	Therminol surge tank	1	1	1	2	0.63

II. Transmission

		0.5	1 (Base)	2	5	
	Pipeline length, miles	25	25	25	25	
	Pipe size, inches (no.)	5(2)	8(2)	12(2)	20(2)	
	Insulation thick, inches (no.)	6½(1)	4(1)	3(1)	2(1)	

----

[a] Number listed indicates operating units.

TABLE B-6

NUMBER OF MAJOR EQUIPMENT AND SCALING FACTOR

HOT WATER SYSTEM

Plant Equip. No.	Equipment	Relative System Capacity 0.5	1 (Base)	2	5	Scaling Factor
		No. of Units[a]				
E-1	Steam condenser	1	1	1	1	0.65
T-1	Water storage tank	3	6	11	26	0.63
P-1	Circulation pump	2	2	2	2	0.52
E-2	Water exchanger	2	4	4	8	0.65

[a] Number listed indicates operating units.

B-2 Effect of Transmission Distance

Capital Requirement

1. The majority of the plant equipment items remains the same; however, some compressors and pumps do vary in capacity and number with the transmission distance.

   Table B-7 lists the items which are resized with changes in transmission distance. The equipment cost is then determined by using the scaling factors listed in Table B-1, B-2, B-4, and B-5.

2. For transportation less than 25 miles, the capital requirement for the pipeline and insulation is proportional to the transmission distance. For 100 mile transportation, the transmission capital requirement includes the intermediate compressor or pumping stations.

3. The process capital for the compressor and pumping stations is estimated in an analogous manner as for the plant equipment.

Operating Cost

1. The external energy requirement for each case is adjusted using the data shown in Table B-7.

2. The costs associated with maintenance material, maintenance labor, and support labor for the compressor or pumping stations are estimated in a similar manner as that for the plant. An allowance of $30,000/yr is assumed for the operating labor at these stations.

Life Cycle Cost

The life cycle cost is calculated using the equation given in Section 3.1.2. However, for T-66 and T-60 systems, the heat delivered varies with the transmission distance:

TABLE B-7

DESIGN CHANGES ASSOCIATED WITH TRANSMISSION DISTANCES
(PUMPS AND COMPRESSORS ONLY)

Equipment	0	10	25	100
	Transmission Distance, Miles — BHP (No. of Units)			
M/S CRS				
Plant				
Compressor, C-2	-	52(1)	98(1)	178(1)
Compressor, C-3	-	148(1)	255(1)	513(1)
Transmission				
Compressor, C-2	-	-	-	178(1)
Compressor, C-3	-	-	-	513(1)
B/C CRS				
Plant				
Compressor, C-2	-	140(1)	400(1)	820(1)
Pump, P-2	-	-	85(1)	85(1)
Pump, P-4	-	16(1)	145(1)	145(1)
Transmission				
Compressor, C-2	-	-	-	820(1)
Pump, P-2	-	-	-	180(3)
Pump, P-4	-	-	-	230(3)
T-66 System				
Plant				
Pump, P-1	120(1)	240(1)	467(1)	-
Pump, P-2	-	195(1)	554(1)	-
T-60 System				
Plant				
Pump, P-1	100(1)	310(1)	694(1)	-
Pump, P-2	-	285(1)	690(1)	-

Transmission Distance, miles	Heat Delivered, 10^6 Btu/Hr		
	0	10	25
T-66	86.1	74.4	66.6
T-60	122.7	104.6	92.5

For M/S or B/C CRS, the heat delivered is the same as that of the base case.

APPENDIX C

FACTORS USED FOR CAPITAL COST ESTIMATE

Direct Cost

As described in Section 3.1, the total direct capital was estimated by a method known as the "percentage of delivered equipment cost method." In this method, direct cost items such as equipment erection labor, field installation labor and field material costs are estimated as percentages of the bare equipment cost. The percentages or factors used in making an estimation of this type should be determined on the basis of the type of process involved, design complexity, required materials of construction, past experience, and other items dependent on the particular system under consideration.

The percentages for equipment erection labor used in this study are presented in Table C-1. Similarly, the percentages used for estimating field installation labor and field material costs are given in Table C-2.

Indirect and Other Costs

As described in Section 3.1, the total investment (TI) was obtained by the summation of the process capital (PC) and indirect costs which account for such items as general facilities (GF), engineering and home office fees (EH), project contingency and process contingency. The capital requirement for plant or transmission section was then arrived at by summing the total investment with other cost items such as royalty allowance, start-up costs, inventory capital, initial catalysts and chemicals, allowance for fund during construction (AFDC), and the land costs. The factors or percentages used to estimate all these indirect and other cost items are presented in Table C-3.

TABLE C-1

LABOR FACTOR FOR EQUIPMENT ERECTION

Equipment	Labor for Erection % of Bare Equip. Cost
Blower	30
Compressor	25
Pump	25
Evaporator	20
Fired Heater	20
Heat Exchanger	20
Reactor	20
Tank	20
Boiler	15
Preheater	15
Vapor/Liquid Separator	15
Drum	10

TABLE C-2

MATERIAL AND LABOR FACTORS FOR FIELD INSTALLATION

Item	Field Material Factor (% of Bare Equip. Cost) M/S	B/C	SA/W	T-66	T-60	HW	Field Labor Factor (% of Field Material Cost) All Systems
Foundations	6	6	6	6	6	7	133
Structures	5	6	6	5	5	6	100
Building[(a)]	4	5	5	4	4	5	125
Insulation	5	2	2	3	3	2	150
Instrumentation	8	5	10	12	12	10	60
Electrical	6	9	9	8	8	9	75
Piping	35	35	35	40	40	35	70
Painting	0.4	0.5	0.5	0.4	0.4	0.5	300
Miscellaneous	4	6	6	5	5	6	80
Total	73.4	74.5	79.5	83.4	83.4	80.5	

(a) Blowers and coolers (for M/S CRS), and tanks (for SA/W CRS) are installed outside.

TABLE C-3

FACTORS FOR INDIRECT AND OTHER COST ITEMS

Items	Plant Section						Transmission Section	
	M/S	B/C	SA/W	T-66	T-60	HW	Pipeline	Insulation
INDIRECT								
General Facilities (GF) % of PC	15	15	20	15	15	20	-	-
Eng. and Home Office Fees (EH), % of PC	10	10	15	10	10	15	10	-
Project Contingency % of (PC+GF+EH)	15	15	15	15	15	15	15	15
Process Contingency % of PC	12	15	10	5	5	5	5	5
OTHERS								
Royalty Allowance % of PC	0.5	0.5	0.5	-	-	-	-	-
Start-up Costs								
Op. Cost, Month	1	1	1	1	1	1	(a)	-
% of TI	2	2	2	2	2	2	0.5	-
Inventory Capital % of TI	1	1	1	1	1	1	-	-
Initial Cat. & Chem.	← As Required →						-	-
AFDC % of TI	8	8	8	8	8	8	8	8
Land, acres[b]	8	8	4	8	8	4	(a)	-

---

[a] Allowances are given in Tables 3.1-1, 3.1-2, 3.1-4, and 3.1-5.

[b] Land cost is based on $5,000 per acre.

APPENDIX D-1

METHANE/SYNGAS CRS-EQUIPMENT SPECIFICATIONS

Methane Reforming Reactor, R-1

Type: Tubular reactor with reactant gases inside tubes. Catalyst pellets of Ni on alumina inside tubes. Incinerator flue gas outside tubes supplies heat of reaction.

Shell:	
Fluid	- $N_2/CO_2/O_2/H_2O$
Temperature, Inlet/Outlet	- 1700/1100°F
Pressure, Inlet/Outlet	- 14.7/14.0 psia
Material of Construction	- 18-8 stainless steel
Diameter	- 7' 9"
Length T-T	- 38'
Thickness	- 1/8"
Tubes:	
Fluid	- $CH_4/CO_2/H_2/H_2O$
Temperature Inlet/Outlet	- 900/1520°F
Pressure Inlet/Outlet	- 620/588 psia
Material of Construction	- 25% Cr/20% Ni SS (HK-40)
Number of Tubes	- 1700
Pitch	- 1.885" square
Tube Size	- 1" Sch. 160
Heat Transfer:	
Duty	- 49.8×10^6 Btu/hr
Overall U	- 11.2 Btu/hr-ft^2-°F
LMTD	- 200°F
Surface Area	- 22,200 ft.^{2}

Waste Heat Boiler, E-1

Fluid	-	$N_2/CO_2/O_2/H_2O$
Temperature Inlet/Outlet	-	1100/667°F
Pressure, Inlet/Outlet	-	14.0/13.5 psia
Steam Pressure/Temp.	-	625 psia/490°F
Duty	-	33.3×10^6 Btu/hr
Steam Rate	-	32,140 lbs/hr
Type of Boiler	-	Tubular heat exchanger
Material	-	Carbon steel

Air-Fin Cooler, E-2

Fluid	-	$N_2/CO_2/O_2/H_2O$
Temperature In/Out	-	667/400°F
Pressure, In/Out	-	13.5/13.0 psia
Air Temperature, In/Out	-	80/120°F
Duty	-	19.3×10^6 Btu/hr
Overall U	-	10 Btu/hr-ft^2-°F
LMTD	-	404°F
Surface Area (Bare Tube Basis)	-	4,780 ft.^{2}
Material	-	Stainless Steel

Exhaust Gas Blower, C-1

Type	-	Centrifugal
Fluid	-	$N_2/O_2/CO_2/H_2O$
Capacity	-	113,000 ACFM @ Suction
Temperature	-	400°F
Suction Pressure	-	13.0 psia
Discharge Pressure	-	14.8 psia
Material of Const.	-	Carbon steel
Power/Driver	-	1060 BHP/Motor

Feed-Product Exchanger 1, E-3

	Shell	Tubes
Fluid	Feed - $CH_4/CO_2/H_2/H_2O$	Effluent - $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	486/900^oF	1520/1187^oF
Pressure, In/Out	625/620 psia	588/584 psia
Material of Const.	Carbon steel	Stainless steel
Duty	10.1×10^6 Btu/hr	
Overall U	14.6 Btu/hr-ft^2-^oF	
LMTD	190^oF	
Surface Area	3,640 ft.^{2}	

Feed-Product Exchanger 2, E-4

	Shell	Tubes
Fluid	Feed - $CH_4/CO_2/H_2/H_2O$	Effluent - $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	70/486^oF	1187/1090^oF
Pressure, In/Out	630/625 psia	584/580 psia
Material of Const.	Carbon steel	Stainless steel
Duty	2.9×10^6 Btu/hr	
Overall U	14 Btu/hr-ft^2-^oF	
LMTD	850^oF	
Surface Area	240 ft.^{2}	

Air-Fin Cooler 1, E-5

Fluid	- $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	- 1090/600^oF
Pressure, In/Out	- 580/577 psia
Air Temperature, In/Out	- 80/120^oF
Material of Const.	- Low alloy steel
Duty	- 13.7×10^6 Btu/hr
Overall U	- 70 BTU/hr-ft^2-^oF
LMTD	- 702
Surface Area (Bare Tube Basis)	- 280 ft^2

<u>Air-Fin Cooler 2</u>, E-6		
Fluid	-	$CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	-	600/200^oF
Pressure, In/Out	-	577/573 psia
Air Temperature, In/Out	-	80/120^oF
Material of Const.	-	Carbon steel
Duty	-	33.3×10^6 Btu/hr
Overall U	-	110 Btu/hr ft^2-^oF
LMTD	-	230^oF
Surface Area (Bare Tube Basis)	-	1,320 ft^2
<u>Vapor-Liquid Separator</u>, D-2		
Type	-	Vertical
Vapor	-	$CH_4/CO/CO_2/H_2/H_2O$
Liquid	-	Water
Diameter	-	4' - 0"
Length, T-T	-	14' - 0"
Temperature	-	200^oF
Pressure	-	573 psia
Material of Const.	-	Carbon steel
<u>Water Circulation Pump</u>, P-1		
Type	-	Centrifugal
Fluid	-	Water
Capacity	-	47.6 GPM
Temperature	-	200^oF
Suction Pressure	-	573 psia
Discharge Pressure	-	625 psia
Power/Driver	-	2.5 BHP/Motor
<u>Knock-Out Drum</u>, D-3		
Type	-	Vertical
Vapor	-	$CH_4/CO_2/H_2/H_2O$
Liquid	-	Water

Knock-Out Drum, D-3 (Cont'd)

Diameter	-	3' - 0"
Length, T-T	-	10' - 0"
Temperature	-	$70^{o}F$
Pressure	-	480 psia
Material of Const.	-	Carbon steel

Condensate Pump, P-2

Type	-	Centrifugal
Fluid	-	Water
Capacity	-	20 GPM
Temperature	-	$70^{o}F$
Suction Pressure	-	480 psia
Discharge Pressure	-	625 psia
Power/Driver	-	8.5 BHP/Motor

Reformer Feed Gas Compressor, C-2

Type	-	Single Stage Reciprocating
Fluid	-	$CH_4/CO_2/H_2/H_2O$
Capacity	-	140 ACFM @ Suction
Suction Temperature	-	$70^{o}F$
Suction Pressure	-	480 psia
Discharge Pressure	-	630 psia
Drive	-	98 BHP/Motor

Metahanation Reactor, R-2

Type: Tubular reactor with reactant gas inside tubes. Catalyst pellets of Ni on alumina inside tubes. Boiling water outside tubes removes heat of reaction.

Shell:

Fluid	- Water
Temperature	- 486^{o}F
Pressure	- 600 psia
Material of Const.	- Carbon steel
Diameter	- 5' - 4"
Length, T-T	- 18' - 0"
Thickness	- 2.1"

Tubes:

Fluid	- $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	- 550/760^{o}F
Pressure, In/Out	- 593/588 psia
Material of Const.	- 2% Cr-0.5% Mo
Number of Tubes	- 1,348
Pitch	- 1.25" triangular
Tube Size	- 3/4" Sch. 40

Heat Transfer:

Rate	- 29.3 x 10^6 Btu/hr
Overall U	- 40.0 Btu/hr-ft^{2}-^{o}F
LMTD	- 144^{o}F
Surface Area Required	- 5090 ft.^{2}

Methanation Preheater, E-7

	Shell	Tubes
Fluid	Effluent - $CH_4/CO_2/H_2/H_2O$	Feed - $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	760/398^{o}F	109/550^{o}F
Pressure, In/Out	588/580 psia	598/593 psia
Material of Const.	Carbon steel	Carbon steel
Duty	12.2 x 10^6 Btu/hr	
Overall U	100 Btu/hr-ft^{2}-^{o}F	
LMTD	248^{o}F	
Surface Area	492 Ft.^{2}	

Water Separator, D-4

Type	- Vertical
Vapor	- $CH_4/CO_2/H_2/H_2O$
Liquid	- Water
Diameter	- 3' - 0"
Length, T-T	- 6' - 0"
Material of Const.	- Carbon steel

Water Recirculation Pump, P-3

Type	- Centrifugal
Fluid	- Water
Capacity	- 18 GPM
Temperature	- 398^oF
Suction Pressure	- 580 psia
Discharge Pressure	- 598 psia
Material of Const.	- Carbon steel
Power/Driver	- 7.5 BHP/Motor

Start-Up Heater, H-1

Fluid	- $CH_4/CO/CO_2/H_2/H_2O$
Temperature, In/Out	- $109/550^oF$
Pressure, In/Out	- 618/598 psia
Process Duty	- 13.5×10^6 Btu/hr

Knock-Out Drum, D-5

Type	- Vertical
Vapor	- $CH_4/CO/CO_2/H_2/H_2O$
Liquid	- Water
Diameter	- 2' - 6"
Length, T-T	- 6' - 0"
Material of Const.	- Carbon steel

Methanation Feed Gas Compressor, C-3

Type	-	Single Stage Reciprocating
Fluid	-	$CH_4/CO/CO_2/H_2/H_2O$
Capacity	-	390 ACFM @ Suction
Suction Temperature	-	70°F
Suction Pressure	-	483 psia
Discharge Pressure	-	618 psia
Material of Const.	-	Carbon steel
Power/Driver	-	255 BHP/Motor

Steam Drum, D-6

Type	-	Vertical
Vapor	-	H_2O
Liquid	-	Water
Diameter	-	3' - 6"
Length, T-T	-	14' - 0"
Material of Const.	-	Carbon steel

APPENDIX D-2

BENZENE/CYCLOHEXANE CRS - EQUIPMENT SPECIFICATIONS

Dehydrogenation Reactors, R-1A and R-1B

Type: Tubular reactor with reactant gases inside tubes. Catalyst pellets of Ni on alumina inside tubes. Hot exit gas from cement kiln outside tubes supplies heat of reaction.

Shell:		
Fluid	-	$N_2/O_2/CO_2/H_2O$
Temp. Inlet/Outlet	-	1300/600^{o}F
Pressure Inlet/Outlet	-	14.5/13.3 psia
Material of Const.	-	9% Cr
Diameter	-	7' - 0"
Length T-T	-	32' - 0"
Thickness	-	1/8"
Tubes:		
Fluid	-	Benzene/cyclohexane/H_2
Temp. Inlet/Outlet	-	500^{o}F/820^{o}F
Pressure Inlet/Outlet	-	410/390 psia
Material of Const.	-	5 Cr - 0.5 Mo
Number of Tubes	-	2400
Pitch	-	1.45" square
Tube Size	-	7/8" BWG 16
Heat Transfer:		
Rate	-	67.0 x 10^6 Btu/hr
Overall U	-	15.4 Btu/hr ft.^{2} ^{o}F
LMTD	-	242^{o}F
Surface Area Required	-	17,940 ft.^{2}

Cyclohexane Surge Tank, TK-1

Fluid	- 85% cyclohexane/15% benzene
Capacity	- 10,000 gallons
Temperature	- $70^{o}F$
Pressure	- Atm.
Material of Const.	- Carbon steel

Cyclohexane Feed Pump, P-1

Type	- Centrifugal
Fluid	- 85% cyclohexane/15% benzene
Capacity	- 340 GPM
SP. Gr. @ F.T.	- 0.78
Temperature	- $70^{o}F$
Suction Pressure	- 16 psia
Discharge Pressure	- 415 psia
Material of Const.	- Carbon steel
Power/Driver	- 120 BHP/Motor

Benzene Transfer Pump, P-2

Type	- Centrifugal
Fluid	- 85% benzene/15% cyclohexane
Capacity	- 305 GPM
SP. Gr. @ F.T.	- 0.83
Temperature	- $100^{o}F$
Suction Pressure	- 380 psia
Discharge Pressure	- 720 psia
Material of Const.	- Carbon steel
Power/Driver	- 85 BHP/Motor

Exhaust Gas Blower, C-1

Type	- Centrifugal
Fluid	- $N_2/O_2/CO_2/H_2O$
Capacity	- 233,000 ACFM @ Suction
Temperature	- $450^{o}F$

Exhaust Gas Blower, C-1

Suction Pressure	- 13.5 psia
Discharge Pressure	- 14.8 psia
Power/Driver	- 1510 BHP/Motor

Hydrogen Compressor, C-2

Type	- Centrifugal
Fluid	- Hydrogen
Capacity	- 900 ACFM @ suction
Suction Temperature	- $100^{o}F$
Suction Pressure	- 380 psia
Discharge Pressure	- 500 psia
Power/Driver	- 400 BHP/Motor

Recovered H_2 Compressor, C-3

Type	- Two-stage reciprocating, inter-cooler
Fluid	- Hydrogen
Suction Capacity	- 28 ACFM
Suction Temperature	- $70^{o}F$
Suction Pressure	- 30 psia
Discharge Pressure	- 390 psia
Power/Driver	- 17 BHP/Motor

Cyclohexane Preheater, E-1

	Shell	Tubes
Fluid	Feed-cyclohexane	Effluent-benzene, H_2
Temperature, In/Out	$70/500^{o}F$	$820/375^{o}F$
Pressure, In/Out	415/410 psia	390/385 psia
Material of Const.	Carbon steel	0.5% Mo steel
Duty	38.0×10^6 Btu/hr	
Overall U	140 Btu/hr ft.^{2} ^oF	
LMTD	$305^{o}F$	
Surface Area	900 ft.^{2}	

Air Cooler, E-2

Fluid	- Effluent-benzene, H_2
Temperature, In/Out	- 375/150^oF
Pressure, In/Out	- 385/382 psia
Air Temperature. In/Out	- 80/120^oF
Material of Const.	- Carbon steel
Duty	- 34.2 x 10^6 Btu/hr
Overall U	- 70 Btu/hr ft.^{2} ^oF
LMTD	- 132^oF
Surface Area (bare tube basis)	- 3700 ft.^{2}
Fan Drives	- 200 H.P.

Trim Cooler, E-3

	Shell	Tubes
Fluid	Cooling Water	Effluent-benzene, H_2
Temperature, In/Out	80/110^oF	150/100^oF
Pressure, In/Out		382/380 psia
Material of Const.	carbon steel	carbon steel
Duty	4.9 x 10^6 Btu/hr	
Overall U	141 Btu/hr-ft^2-^oF	
LMTD	29^oF	
Surface Area	1200 ft.^{2}	

Vapor-Liquid Separator, D-1

Type	- Vertical
Vapor	- Hydrogen, some benzene
Liquid	- Benzene
Inside Diameter	- 4' - 0"
Length, T-T	- 14' - 0"
Temperature	- 100^oF
Pressure	- 380 psia
Material of Const.	- Carbon steel

Vapor-Liquid Separator, D-2

Type	- Vertical
Vapor	- Hydrogen
Liquid	- Cyclohexane
Inside Diameter	- 4' - 0"
Length, T-T	- 15' - 0"
Temperature	- $70^{o}F$
Pressure	- 30 psia
Material of Const.	- Carbon steel

Hydrogeneration Reactor, R-2

Type: Tubular reactor with reactant gases inside tubes. Catalyst pellets of Ni on alumina inside tubes. Boiling water at 400 psia outside of tubes removes heat of reaction.

Shell:

Fluid	- Water
Temperature	- $445^{o}F$
Pressure	- 400 psia
Material of Const.	- Carbon steel
Diameter	- 6' - 0"
Length T-T	- 7' - 0"
Thickness	- 1.6"

Tubes:

Fluid	- Benzene/cyclohexane/H_2
Temperature Inlet/Outlet	- $575/700^{o}F$
Pressure Inlet/Outlet	- 355/340 psia
Material of Const.	- 0.5% Mo steel
Number of Tubes	- 3300
Pitch	- 1.25" triangular
Tube Size	- 7/8" BWG 16

Heat Transfer:		
Rate	-	92.6×10^6 Btu/hr
Overall U	-	119 Btu/hr ft.^{2} ^oF
LMTD	-	250^oF
Surface Area Required	-	5210 ft.^{2}

<u>Benzene Surge Tank, TK-2</u>

Fluid	-	85% benzene/15% cyclohexane
Capacity	-	10,000 gallons
Temperature	-	70^oF
Pressure	-	Atm.
Material of Const.	-	Carbon steel

<u>Benzene Feed Pump, P-3</u>

Type	-	Centrifugal
Fluid	-	85% benzene/15% cyclohexane
Capacity	-	305 GPM
Temperature	-	70^oF
Suction Pressure	-	16 psia
Discharge Pressure	-	475 psia
Driver	-	Motor, 180 H.P.

<u>Cyclohexane Transfer Pump, P-4</u>

Type	-	Centrifugal
Fluid	-	85% cyclohexane/15% benzene
Capacity	-	340 GPM
Temperature	-	100^oF
Suction Pressure	-	325 psia
Discharge Pressure	-	840 psia
Power/Driver	-	145 BHP/Motor

Recovered H_2, Compressor, C-4

Type	- Two-stage reciprocating, inter-cooler
Fluid	- Hydrogen
Capacity	- 33 ACFM
Suction Temperature	- 70°F
Suction Pressure	- 30 psia
Discharge Pressure	- 365 psia
Power/Driver	- 22 BHP/Motor

Recycle H_2 Compressor, C-5

Type	- Centrifugal
Fluid	- Hydrogen
Capacity	- 1050 ACFM
Suction Temperature	- 100°F
Suction Pressure	- 325 psia
Discharge Pressure	- 365 psia
Power/Driver	- 270 BHP/Motor

Feed Preheater, E-4

	Shell	Tubes
Fluid	Feed-benzene, H_2	Effluent-cyclo., H_2
Temperature, In/Out	80/550°F	700/180°F
Pressure, In/Out	365/360 psia	340/330 psia
Material of Const.	Carbon steel	0.5% Mo. steel
Duty	66.7×10^6 Btu/hr	
Overall U	167 Btu/hr-ft^2 °F	
LMTD	123°F	
Surface Area	3250 ft^2	

Final Cooler, E-5

	Shell	Tubes
Fluid	Cooling Water	Effluent-cyclo., H_2
Temperature, In/Out	80/110°F	180/100°F

Final Cooler, E-5 (Cont'd)

	Shell	Tubes
Pressure, In/Out	50/45 psig	330/325 psig
Material of Const.	Carbon steel	Carbon steel
Duty	7.9×10^6 Btu/hr	
Overall U	140 Btu/hr. ft^2 ^oF	
LMTD	36^oF	
Surface Area	1550 $ft.^2$	

Vapor-Liquid Separator, D-3

Type	- Vertical
Vapor	- Hydrogen, some benzene
Liquid	- Benzene
Inside diameter	- 4' - 0"
Length, T-T	- 14' - 0"
Temperature	- 70^oF
Pressure	- 30 psia
Material of Const.	- Carbon steel

Vapor-Liquid Separator, D-4

Type	- Vertical
Vapor	- Hydrogen, some cyclohexane
Liquid	- Cyclohexane
Inside diameter	- 4' - 0"
Length, T-T	- 15' - 0"
Temperature	- 100^oF
Pressure	- 325 psia
Material of Const.	- Carbon steel

Steam Drum, D-5

Type	- Horizontal
Inside Diameter	- 4' - 0"
Length, T-T	- 15' - 0"

Steam Drum, D-5 (Cont'd)

Temperature	- 500^oF
Pressure	- 500 psig
Material of Const.	- Carbon steel

Fired Heater, H-1

Fluid	- Hydrogen, benzene vapor
Process duty	- 2.8×10^6
Temperature, In/Out	- $550/650^oF$
Pressure, In/Out	- 360/355 psia

APPENDIX D-3

SULFURIC ACID/WATER CRS - EQUIPMENT SPECIFICATIONS

Storage Tank, TK-1

Fluid	-	50 Wt. % sulfuric acid solution
Capacity	-	57,000 gals.
Temperature	-	$150^{o}F$
Pressure	-	15 psia
Material	-	Carbon steel, rubber lined with 1/4" triflex

Storage Tank, TK-2

Fluid	-	Water
Capacity	-	34,000 gals.
Temperature	-	$200^{o}F$
Pressure	-	15 psia
Material	-	Carbon steel, rubber lined with 1/4" triflex

Storage Tank, TK-3

Fluid	-	87 Wt. % sulfuric acid solution
Capacity	-	27,100 gals.
Temperature	-	$320^{o}F$
Pressure	-	15 psia
Material	-	Carbon steel with plastic lining

Pump, P-1

Type	-	Centrifugal
Fluid	-	50 Wt. % sulfuric acid solution
Capacity	-	57.7 GPM
Temperature	-	$150^{o}F$
Sp. Gr.	-	1.36 @ P.T., 1.4 @ $60^{o}F$
Pressure	-	30 psia (discharge), 15 psia (suction)
Material	-	Duriron or equivalent
Driver	-	Motor, 2.0 H.P.

Pump, P-2

Type	-	Centrifugal
Fluid	-	87 Wt. % sulfuric acid solution
Capacity	-	27.2 GPM
Temperature	-	$320^{o}F$
Sp. Gr.	-	1.66 @ P.T., 1.8 @ $60^{o}F$
Pressure	-	30 psia (discharge) 1 psia (suction)
Material	-	Tantalum or equivalent
Driver	-	Motor, 2.0 H.P.

Pump, P-3

Type	-	Centrifugal
Fluid	-	Water
Capacity	-	15 GPM
Temperature	-	102°F
Sp. Gr.	-	0.99 @ P.T.
Discharge Pressure	-	17 psia (discharge), 1 psia (suction)
Material	-	Duriron
Driver	-	Motor, 1.0 H.P.

Pump 4, P-4

Type	-	Centrifugal
Fluid	-	Water
Capacity	-	69.4 GPM
Temperature	-	$200^{o}F$
Sp. Gr.	-	0.96 @ P.T.
Pressure	-	35 psia (discharge), 15 psia (suction)
Material	-	Duriron
Driver	-	Motor, 2.0 H.P.

Pump, P-5

Type	- Centrifugal
Fluid	- 87% sulfuric acid solution
Capacity	- 54.3 GPM
Temperature	- 320^{o}F
Sp. Gr.	- 1.66 @ P.T., 1.8 @ 60^{o}F
Pressure	- 35 psia (discharge), 15 psia (suction)
Material	- Tantalum or equivalent
Driver	- Motor, 2.0 H.P.

Vacuum Pump, P-6

Type	- Reciprocating
Fluid	- Water vapor/air
Capacity	- 95 ACFM
Temperature	- 102^{o}F
Pressure	- 15 psia (discharge), 1 psia (suction)
Driver	- Motor, 3.0 H.P.

Hot Water Pump, P-7

Fluid	- Water
Capacity	- 127 GPM
Temperature	- 100^{o}F
Sp. Gr.	- 0.99 @ P.T.
Pressure	- 60 psig (discharge), 40 psig (suction)
Material	- Ductile iron
Driver	- Motor, 5.0 H.P.

Heat Exchanger, E-1

Shell Side

Fluid	-	Water
Capacity	-	127 GPM
Temp. In/Out	-	231/280^{o}F
Pressure, In/Out	-	53/50 psig
Material	-	Carbon steel

Tube Side

Fluid	-	79 Wt. % sulfuric acid solution
Capacity	-	62.1 GPM
Temp., In/Out	-	385/260^{o}F
Pressure, In/Out	-	19/14 psig
Material	-	Hastelloy

Performance

Duty	-	2.96 x 10^{6} Btu/hr
LMTD	-	58.7^{o}F
Overall U	-	241 Btu/hr ^{o}F ft.^{2}
Surface Area	-	209 ft.^{2}

Heat Exchanger, E-2

Shell Side

Fluid	-	Water
Capacity	-	127 GPM
Temp. In/Out	-	179/231^{o}F
Pressure, In/Out	-	56/53 psig
Material	-	Carbon steel

Tube Side

Fluid	-	68 Wt. % sulfuric acid solution
Capacity	-	76.9 GPM
Temp. In/Out	-	316/210^{o}F
Pressure, In/Out	-	14/7 psig
Material	-	Hastelloy

Performance

Duty	- 3.17×10^6 Btu/hr
LMTD	- 53.3^oF
Overall U	- 222 Btu/hr ^oF ft.^{2}
Surface Area	- 268 ft.^{2}

Heat Exchanger, E-3

Shell Side

Fluid	- Water
Capacity	- 127 GPM
Temp. In/Out	- $100/179^oF$
Pressure, In/Out	- 59/56 psig
Material	- Carbon steel

Tube Side

Fluid	- 50 Wt. % sulfuric acid solution
Capacity	- 113 GPM
Temp. In/Out	- $246/150^oF$
Pressure, In/Out	- 7/1 psig
Material	- Hastelloy

Performance

Duty	- 4.83×10^6 Btu/hr
LMTD	- 59.4^oF
Overall U	- 213 Btu/hr ^oF ft.^{2}
Surface Area	- 382 ft.^{2}

Heater, E-4

Shell Side

Fluid	- Steam (condensing)
Capacity	- 15,430 lb/hr
Temp. In/Out	- $350/350^oF$
Pressure	- 120 psig
Material	- Carbon steel

Tube Side

Fluid	- 50 Wt. % sulfuric acid solution
Capacity	- 57.7 GPM
Temp. In/Out	- 150/320^{o}F
Pressure	- 28/21 psig
Material	- Hastelloy

Performance

Duty	- 13.43 x 10^{6} Btu/hr
LMTD	- 39^{o}F
Overall U	- 490 Btu/hr ^{o}F ft.^{2}
Surface Area	- 702 ft.^{2}

Liquid-Vapor Separator, D-1

Type	- Vertical, liquid/vapor
Diameter	- 3.5'
Length	- 11' T-T
Temperature	- 320^{o}F
Pressure	- 20 psia
Material	- Carbon steel, plastic lining

Cooler, E-5

Shell Side

Fluid	- Steam
Capacity	- 9,061 lb/hr (93% condensed)
Temp, In/Out	- 320/220^{o}F
Pressure	- 17 psia
Material	- Hastelloy or equivalent

Tube Side

Fluid	- Water
Capacity	- 573 GPM
Temperature	- 80/110^{o}F
Pressure	- 20 psia
Material	- Carbon steel

Performance

Duty	- 8.61×10^6 Btu/hr
LMTD	- 116°F
Overall U	- 318 Btu/hr °F ft.^{2}
Surface Area	- 233 ft.^{2}

Evaporator Package

Type	- Forced circulation, submerged heating elements
Fluid	- 65-87% sulfuric acid solution
Temperature	- 320°F
Pressure	- 1 psia
Material	- Hastelloy
Calandria	- Duty, 9.52×10^6 Btu/hr
	- Surface Area, 927 ft.^{2}
	- Medium, 350°F and 120 psig steam
Condenser	- Duty, 8.67×10^6 Btu/hr
	- Surface Area, 1645 ft.^{2}
	- Cooling Water, 80-95°F
Recirculating Pump	- Capacity, 1670 GPM
	- Motor, 10 H.P.

Drum, D-2

Fluid	- Water
Diameter	- 2.5'
Length	- 5' T-T
Temperature	- 102°F
Pressure	- 1 psia
Material	- Carbon steel, rubber-lined

APPENDIX D-4

THERMINOL 66 SYSTEM - EQUIPMENT SPECIFICATIONS

1. Therminol 66 Heater, E-1

	SHELLSIDE	TUBESIDE
Fluid	Flue Gas	Therminol 66
Temperature, In/Out, ^oF	1700/445	60/650
Pressure, In/Out, psia	14.7/14.2	700/600
Dimension, ft.	13L x 13W x 20H	1" O.D. tube BWG 10
Passes	2	Multiple
Flowrate, @ 60^oF & 14.7 psia	60,800 CFM (267,585 lbs/hr)	82.8 CFM (311,430 lbs/hr)
Duty, 10^6 Btu/hr	Pass 1 = 75.2 Pass 2 = 18.1	
Material of Construction	Carbon steel with refractory lined	Stainless Steel
Overall U, Btu/hr ft^2 ^oF	Pass 1 = 12.5; Pass 2 = 6.0	
LMTD, ^oF	Pass 1 = 738; Pass 2 = 442	
Surface Area, $ft.^2$	Pass 1 = 8344; Pass 2 = 6825	

Note: Gas flows transverse to tubes

2. BFW Heat Exchanger E-2A, E-2B

	SHELLSIDE	TUBESIDE
Fluid	Therminol 66	B.F.W.
Temperature, In/Out, ^oF	550/120	60/500
Pressure, In/Out, psia	70/20	650/645
Material of Construction	1/8" Carbon Steel	S.S.·3/4" O.D. BWG 16
(Total 2 Exchangers)	15/16" pitch; U-tubes, 1126/shell	
Overall U	100 Btu/hr $ft.^2$ ^oF	
LMTD	62^oF	
Surface Area	10,800 $ft.^2$	
Duty	66.5 x 10^6 Btu/hr	
Dimensions-Each Exchanger (2)	37" ID by 24 ft. T-T	

3. Blower B-1

Type	Axial
Fluid	Flue Gas
Capacity	110,000 ACFM
Temperature	450^oF
Suction Pressure	14.3 psia
Discharge Pressure	14.8 psia
Material of Construction	Carbon steel
Driver	375 motor H.P.

4. T-66 Feed Pump P-1 & P-2

Type	Centrifugal
Fluid Pumped	Therminol 66 (Manufactured by Monsanto)
Capacity	700 GPM
Sp. Gravity @ 50^oF	1.02 for Pump #1
@ 150^oF	0.98 for Pump #2
Temperature	P-1 @ 60-100^oF P-2 @ 60-200^oF
Suction Pressure	17 psia
Discharge Pressure, psia	P-1 @ 700 P-2 @ 750
Motor H.P.	P-1 @ 500 P-2 @ 600
Material of Construction	Carbon steel

Note: Viscosity	@0^oF	@50^oF	@100^oF	@200^oF
	150,000 lbs/hr.ft	617 lbs/hr.ft	67.8 lbs/hr.ft	10.1 lbs/hr.ft

5. Therminol-66 Surge Tanks T-1 & T-2

Fluid	Therminol 66
Dimensions	10' I.D. x 36' L
Temperature	$80^{o}F$
Pressure	30 psia
Material of Construction	Carbon steel

APPENDIX D-5

THERMINOL 60 SYSTEM - EQUIPMENT SPECIFICATIONS

1. Therminol-60 Surge Tank, T-1, T-2

Fluid	Therminol 60
Volume (each)	3885 $ft.^3$
Temperature	80^oF
Pressure	30 psia
Material of Construction	Carbon steel

2. T-60 Feed Pump, P-1, P-2

Type	Centrifugal
Fluid Pumped	Therminol 60 (Manufactured by Monsanto)
Capacity, GPM	1200 for P-1 & 1000 for P-2
Sp. Gravity @ 50^oF	1.90 for Pump #1
@ 500^oF	0.96 for Pump #2
Temperature	P-1 @ 60^o-100^oF P-2 @ 60^o-200^oF
Suction Pressure	17 psia
Discharge Pressure	P-1 @ 900 psia P-2 @ 1100 psia
Motor H.P.	P-1 @ 700 P-2 @ 700
Material of Construction	Carbon steel

Note: Viscosity	@ 0^oF	50^oF	100^oF	200^oF
	32	11.7	6.3	4.0
	lbs/hr ft.	lbs/hr ft.	lbs/hr ft	lbs/hr ft.

3. Blower, B-1

Type	Axial or Centrifugal
Fluid	Kiln exit gas
Capacity	300,000 ACFM
Temperature	600^{o}F
Suction Pressure	14.3 psia
Discharge Pressure	14.8 psia
Material of Construction	Carbon Steel
Driver, Motor H.P.	850

4. BFW Heat Exchanger, E-2A, B, C

	Shellside	Tubeside
Fluid	Therminol 60	B.F.W.
Temperature, In/Out, ^{o}F	500/120	60/445
Pressure, In/Out, Psia	70/20	425/410
Material of Construction (Total 3 Exchangers)	1/8" Carbon Steel S.S. 3/4" O.D. BWG $\frac{15}{16}$" on triangle pitch; U-tubes, 1126/shell	
Overall U	100 Btu/hr ft^2 ^{o}F	
LMTD	58^{o}F	
Surface Area	16,000 ft^2	
Duty	66.5 x 10^6 Btu/hr	
Dimensions: Each Exchanger (3 Exchangers)	37" I.D. by 24' T-T	

5. T-60 Heater, E-1

	Shellside	Tubeside
Fluid	Cement kiln exit gas	Therminol 60
Temperature, In/Out, ^{o}F	1300/600	60/600
Pressure, In/Out, Psia	14.5/14.0	900/875
Dimension, ft	13L x 6.5W x 20H	1" O.D. tube BWG 10

Passes	1	Multiple
Flow rate, CFM @ 60^{o}F	136,600	130 CFM
& 14.7 psia (lbs/hr)	(678,830)	(485,613)
Duty	134×10^{6} Btu/hr	
Material of Construction	Carbon steel with refractory lined	Stainless steel
Overall U	22.5 Btu/hr ft^{2} ^{o}F	
LMTD	617^{o}F	
Surface Area	9,702 ft^{2}	

Note: Gas flows transverse to tubes.

APPENDIX D-6

PRESSURIZED HOT WATER SYSTEM - EQUIPMENT SPECIFICATIONS

Steam Condenser, E-1

	Shell	Tubes
Fluid	Water	Steam
Temperature, In/Out	140/325^{o}F	350/350^{o}F
Design Pressure	100 psig	150 psig
Material of Construction	Carbon Steel	Copper or Brass
Duty	23.0×10^{6} Btu/hr	
Overall U	286 Btu/hr ft.^{2} ^{o}F	
LMTD	87^{o}F	
Surface area	920 ft.^{2}	

Water Storage Vessels

Type	Horizontal
Diameter	12' - 0"
Length	61' - 0"
Temperature	325^{o}F
Design pressure	100 psig
Material of Construction	Carbon Steel
Insulation	2" fiberglass (k=0.4 Btu/hr.ft.^{o}F)

Circulating Pump, P-1 A&B

Type	Centrifugal
Fluid	Water
Flow rate @ FT	250 GPM
Temperature	140^{o}F
Suction pressure	0 psig
Discharge pressure	90 psig

Exchanger, E-2

	Shell	Tubes
Fluid	Water	Water
Temperature, In/Out	100/280^{o}F	320/140^{o}F
Design pressure	50 psig	100 psig
Material of Construction	Carbon steel	Carbon steel
Duty	44.8 x 10^6 Btu/hr	
Overall U	210 Btu/hr ft.^{2} ^{o}F	
LMTD	40^{o}F	
Surface area	7,200 ft.^{2}	

INITIAL DISTRIBUTION

UNLIMITED RELEASE

U.S. Department of Energy (7)
Division of Energy Storage Systems
Washington, D.C. 20545
Attn: J. Gahimer (5)
G. Pezdirtz
J. H. Swisher

Electric Power Research Institute (300)
P.O. Box 10412
Palo Alto, CA 94304
Attn: B. Mehta

Solar Energy Research Institute
1536 Cole Blvd.
Golden, CO 80401
Attn: C. Wyman

Oak Ridge National Laboratory
Oak Ridge, TN 37830
Attn: R. Graves

NASA-Lewis Research Center (2)
Cleveland, OH 44101
Attn: J. Calogeras
W. J. Masica

C. J. Swet
Route 4-Box 258
Mt. Airy, MD 21771

Rocket Research Company
York Center
Redmond, WA 98052
Attn: R. D. Smith

General Electric
Corporate Research and Development
Schenectady, NY 12301
Attn: J. W. Flock

Gilbert Associates, Inc. (8)
P.O. Box 1498
Reading, PA 19603
Attn: H. Chen

Australian National University
Department of Engineering Physics
Canberra, A.C.T. 2600
Australia
Attn: P. O. Carden

D. M. Haaland, 5823
R. W. Mar, 8313
J. J. Iannucci, 8326
W. G. Wilson, 8451
J. D. Fish, 8452 (5)
T. T. Bramlette, 8453
F. J. Cupps, 8265/Technical Library Processes Division, 3141 (27)
Technical Library Processes Division, 3141 (2)
Library and Security Classification Division, 8266-2 (3)

Org.	Bldg.	Name	Rec'd by*	Org.	Bldg.	Name	Rec'd by*

*Recipient must initial on classified documents.