WAESD
TR-83-1011

(NASA-CR-174442) SOLAR THERMOCHEMICAL PROCESS INTERFACE STUDY Final Report (Westinghouse Electric Corp.) 125 p
HC A06/MF A01
CSCL 10A
N85-19527
Unclas
G3/44 18697

SOLAR THERMOCHEMICAL PROCESS INTERFACE STUDY

FINAL REPORT

Prepared For
JET PROPULSION LABORATORY
CALIFORNIA INSTITUTE OF TECHNOLOGY
Under Contract JPL 956510

FEBRUARY 1984

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Report No.	2. Government Accession No.	3. Recipient's Catalog No.

4. Title and Subtitle Solar Thermochemical Process Interface Study	5. Report Date 18 November 1983 6. Performing Organization Code
7. Author(s) Westinghouse Electric Corporation	8. Performing Organization Report No. WAESD-TR-83-1011 10. Work Unit No.
9. Performing Organization Name and Address Westinghouse Electric Corporation Advanced Energy Systems Division P.O. Box 10864 Pittsburgh, PA 15236	11. Contract or Grant No. JPL 956510 13. Type of Report and Period Covered Final Report
12. Sponsoring Agency Name and Address Assistant Secretary for Solar U.S. Department of Energy Washington, D.C.	14. Sponsoring Agency Code

15. Supplementary Notes

16. Abstract

The results reported herein describe the design and analyses of a subsystem of a hydrogen production process. The process is based on solar driven thermochemical reactions. The subject subsystem receives sulfuric acid of 60% concentration at 100°C, 1 Atm pressure. The acid is further concentrated, vaporized, and decomposed (at a rate of 122 g moles/sec H_2SO_4) into SO_2, O_2, and water. The produce stream is cooled to 100°C.

Three subsystem options, each being driven by direct solar energy, were designed and analyzed. The results were compared with a prior study case in which solar energy was provided indirectly through a helium loop.

17. Key Words (Suggested by Author(s)) Solar Hydrogen Fuels Chemical	18. Distribution Statement

19. Security Classif. (of this report) Unclassified	20. Security Classif. (of this page) Unclassified	21. No. of Pages 119	22. Price*

* For sale by the National Technical Information Service, Springfield, Virginia 22161

.ASA-C-168 (Rev. 10-75)

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1.0 SUMMARY

Four alternative subsystems to decompose 122 g moles/sec of sulfur trioxide (SO_3) were studied. Three cases were based on direct application of solar energy. The fourth case was a subsystem driven by solar energy through an intermediate helium loop. Energy from thermal storage was utilized in acid concentration in all cases and in acid vaporization in one case.

The study directives (requirements and guidelines) were organized to respond the contractural work statements, and reviewed with customer.

Flowsheets and process streams were prepared and analyzed for each case. Design concept parameters including composition, temperature, pressure, enthalpy and entropy, were established. Mass and energy balances were prepared. Operational modes were described for each case. A procedure for scale down of the process design to a potential test hardware size of 10 MW solar input (approximately 1/10 scale) was described.

Conceptual designs of the solar receivers were prepared and other components were sized. These were used in the preparation of cost estimates, selection of materials of construction, and assessment of fabrication and construction methods. Energy losses from major equipment components (solar receivers) were assessed. Both state-of-art and growth potential designs were established. The state-of-the-technology and development requirements were assessed. The subsystem driven through the intermediate helium loop was a modification (in size) of a system previously studied.[1] This case was utilized as a reference to which the three directly driven cases were compared.

All requirements of the contract work scope were successfully completed. Three subsystem concepts were examined and each appears to be technically feasible. Each exhibits a good to excellent probability of being successfully developed. However, considerable research and development will be required in any of the cases.

Case 1 (Direct solar acid vaporization and decomposition) subsystem appears to be the leading contender for further consideration, based on the available data and information. This conclusion should be reconsidered when additional system data, as well as further examination of the receiver/tower/ heliostat field interface, is available.

The above noted interface and further refinement of the receivers (to increase performance and reduce cost) represent areas where further effort is recommended. Stress analysis studies of the silicon carbide vaporizer and cyclic testing of helicoflex seals (for use in subsystem heat exchangers) are also recommended areas for further work that should be carried out to define and minimize component development risk.

Implementation of these recommendations for future work would provide input for further refinement of the development requirements discussed herein.

Given the technical scope and resources of the contract, detailed economic evaluation was not possible. The equipment cost estimation was based on the conceptual design of the equipment. The heliostat field, the solar tower and the plant piping (requires plant layout) were not included in the economic evaluation. Further refinement of the economic evaluation is necessary once more information becomes available.

1. Westinghouse Electric Corporation, "Solar Thermal Hydrogen Production Process", Final Report, DOE-ET-20608-1, dated December 1982.

2.0 INTRODUCTION AND APPROACH

The purpose of the study reported herein was to examine the interfacing of a solar tower with a thermochemical process based on the decomposition of sulfuric acid. This section of the report addresses: 2.1 - Study Work Scope, 2.2 - Relationship of the Subprocess Studied to the Balance of System, and 2.3 - Technical Approach.

2.1 Study Work Scope

The objective of the study was to identify an optimum configuration of receivers, thermal storage and heat exchangers, to maximize process efficiency, and to minimize product costs.

The work scope was divided into four tasks, as follows:

Task 1 - Preliminary Engineering Design
Task 2 - Design Concept Analysis
Task 3 - Technical Management/Meetings
Task 4 - Documentation

The principal technical effort was carried out under Tasks 1 and 2.

2.1.1 Preliminary Engineering Design

The general study and design requirements and guidelines were established in the contract and defined in the Kick-off Meeting. Four different subprocess configurations were selected. Variations included options in methods of applying the solar energy including: direct solar energy into process components (receivers), indirect solar energy through an intermediate loop and through thermal storage, and recovery of energy from decomposed product. In the direct solar energy cases, different combinations of solar receivers were considered for decomposition, vaporization and thermal storage. The available energy from decomposed product was established in each case.

2.1.2 Design Concept Analysis

Process stream data including composition, temperature, pressure, enthalpy and entropy were established. Mass and energy balances were calculated. Energy requirements and losses were established for major components. A scale down procedure was described. Operational modes, for each case, were defined. Major components were configured. Materials were selected. Cost estimates were prepared. Fabrication and construction approaches were examined. State-of-the-technology and development requirements were assessed.

2.1.3 Technical Management/Meetings

In addition to the Kick-off Meeting (see 2.1.1), several informal project review meetings were attended by Customer and contractor personnel. Significant contributions to the success of the technical effort resulted from these meetings. The Customer along with his Consultant, provided technical input as well as direction in connection with these discussions. The contractor's preliminary designs were reviewed at several stages. A final project review meeting was held.

2.1.4 Documentation

Minutes of each meeting were prepared. Monthly financial reports, including brief statements of technical progress, were prepared. This final report was prepared.

2.2 Relationship of the Subprocess Studied to the Balance of the System

The production of hydrogen with solar energy appears to have an important potential in the longer term national energy picture. Hydrogen can be produced from water by dissociation so that there is no question of the adequacy of supply. Hydrogen is a clean, storeable, transportable energy medium that could serve a very wide variety of energy applications on land, at sea, and in space. Its production using the inexhaustible solar energy source could completely resolve any question of resources.

The subprocess studied in this effort is an important element of at least two processes for hydrogen production. One of these processes is a total thermochemical system in which solar energy is utilized as heat at all steps in the process. The other is a hybrid electrochemical/thermochemical system. In this latter case, solar energy is utilized in two ways. For the electrochemical subprocess. electricity is generated by solar power and then used to drive the electrolyzer. For the thermochemical subprocess, the solar energy is used as heat.

Both of these hydrogen production processes consume sulfur dioxide (SO_2) and produce sulfuric acid (H_2SO_4).

The overall hydrogen production system concept is indicated, functionally, in Figure 2-1. The scope of the subprocess, the subject of this contract, is indicated by dashed lines. This subprocess applies heat to the sulfuric acid resulting in its decomposition (in the presence of a catalyst):

$$H_2SO_4 \xrightarrow{\text{heat}} 1/2O_2 + SO_2 + H_2O$$

Hence, the system cycle is completed. Balance of plant items (such as towers and plant instrumentation and control) were excluded except for interface considerations.

2.3 Technical Approach

This study focused on the equipment in an overall solar powered hydrogen production plant that provides the capability to absorb solar energy into the process fluids for driving the electrochemical process for producing hydrogen. The equipment included were the sulfuric acid concentrator, acid vaporizer, decomposer, intermediate heat exchanger and storage tanks.

These configurations were initially defined, Figures 3-1, 3-4 and 3-7, which considered alternatives for using solar power, directly, for the thermal need of the sulfur cycle process. A fourth alternative, Figure 3-9 was defined in which the solar power was indirectly coupled to the process fluids. In first configuration (Figure 3-1) the acid vaporizer and decomposer would be directly powered by solar input; the concentrator would receive power from a solar thermal storage system. A second configuration (Figure 3-2) would use the decomposer exit fluid, regeneratively, to augment the power required to vaporize the acid. A third configuration would use power from thermal storage to vaporize the acid. The fourth configuration used a helium loop to absorb the solar power in thermal receivers and to deliver the power to the acid vaporizer and decomposer. Energy balances and state points were calculated for each of these configurations. Heat exchanger duties were adjusted to most efficiently use the energy within the boundary of the subprocesses considered for this study. The operational modes and complexity of operation were defined for each configuration.

Design configurations were identified for the solar powered acid vaporizer and decomposer process tubing based on process thermal and reaction limitations and on material property limitations. Using results from previous solar receiver studies, receiver concepts were defined which would enclose the tube bundles. Each receiver was sized for the thermal needs consistent with the four configuration alternatives. The balance of the exchangers were designed based on selecting appropriate existing heat exchanger concepts and scaling to the thermal demand.

Cost estimates were developed for each of the heat exchangers based on vendor information and engineering estimates from manufacturers of the process equipment. Fabrication and assembly techniques and problem areas were defined for the acid vaporizer and decomposer. Areas that require technology development were identified. Areas considered ranged in context from design verification, through fabrication size scale up, to materials development. The final step in this study was to select a preferred configuration. Proper selection requires consideration of trade offs within the entire solar powered hydrogen plant.

However, for this study the sub process configurations were ranked by using a comparison matrix technique in which the above described design information were weighted and were assigned relative numercial values for each configuration. Figure of Merits were then calculated for the four configurations.

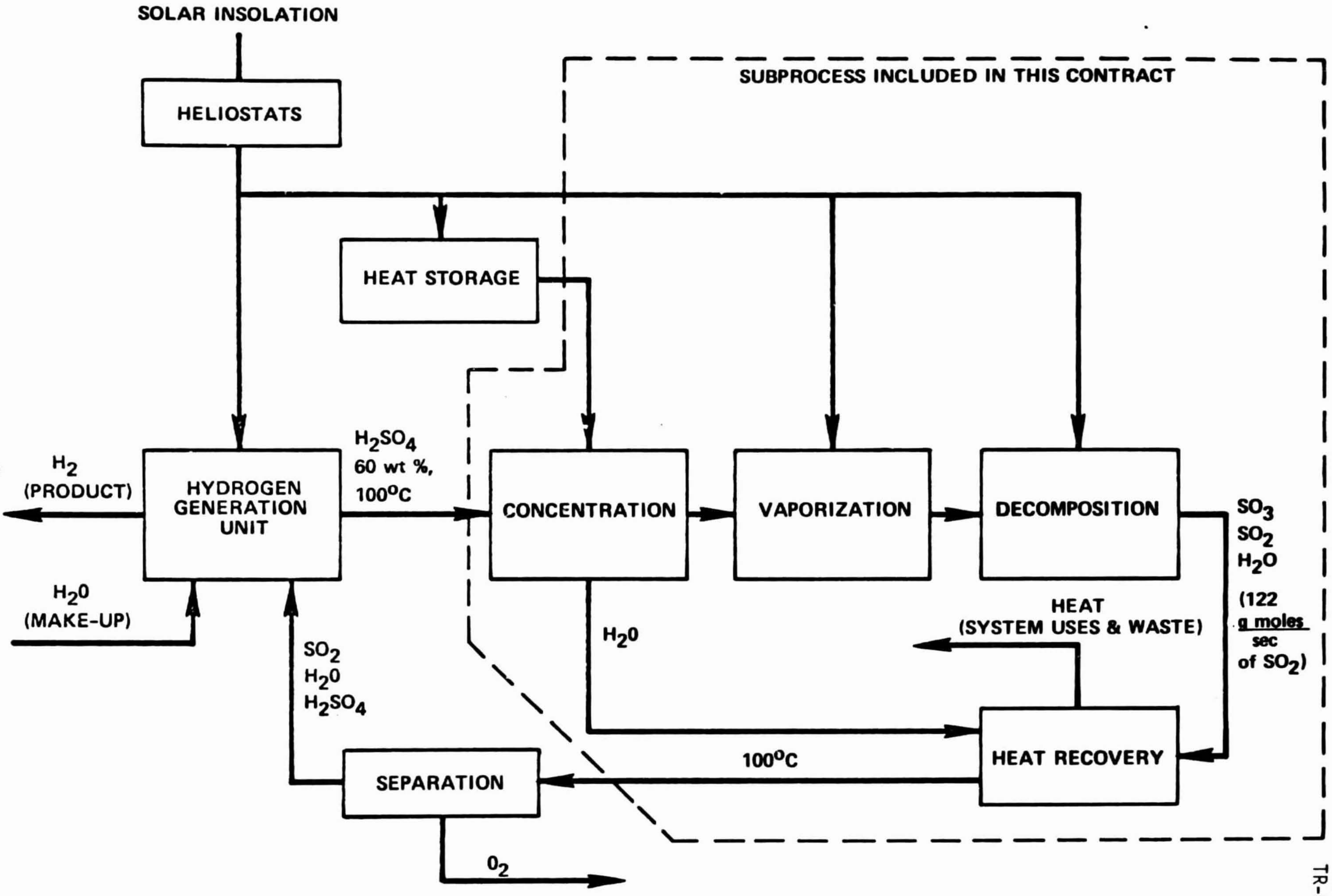

FIGURE 2-1. OVERALL HYDROGEN PRODUCTION SYSTEM CONCEPT

3.0 SYSTEM STUDIES

3.1 CASES STUDIED

Four cases were studied. The first three cases are the coupling of the sulfuric acid concentration, vaporization and decomposition processes directly to solar heat sources. The fourth case is the use of helium as a heat transport medium to decouple the acid decomposition process from the solar heat source. In each of the first three cases there is a separate solar receiver for the high temperature acid decomposer and the thermal energy storage. However, the thermal energy required for the acid vaporization is supplied: 1) from a separate solar receiver for Case 1; 2) from internal heat exchange with process streams for Case 2; 3) from thermal energy storage for Case 3.

3.2 SYSTEM REQUIREMENTS/ASSUMPTIONS

The design requirements/assumptions are as follows:

- The plant is to be operated 24 hours a day, but the solar receivers are to be operated 9 hours a day. The solar flux is assumed to be constant during this 9 hour period.
- Plant capacity: decomposing 122 g-mole per second of SO_3.
- Maximum process fluid temperature: 900°C.
- Feed acid conditions: 60 weight percent H_2SO_4 solution at 100°C and one atmosphere pressure.
- Heat exchanger ΔT: 40°C minimum.
- The decomposed hot gases will be returned to a temperature of 100°C with heat exchange equipment.
- The balance of the plant is not specified and is beyond the scope of this study. Therefore, the thermal storage (that also provides some thermal energy to the balance of the plant) will not be designed.
- Nominal system pressure: 7.5 atmospheres.
- All the thermodynamic data used in this study are derived from References 3-1, 3-2, 3-3 and 3-4.

3.3 CASE DESCRIPTION

3.3.1 CASE 1 DESCRIPTION

3.3.1.1 SYSTEM ANALYSES

The process flow sheet and the state points for daytime operation are shown in Figure 3-1 and Table 3-1 respectively. Figure 3-2 is the heating curve of the feed acid showing temperature versus enthalpy. Sixty weight percent (21.6 mol %) of H_2SO_4 solution at 100°C and 1 atmosphere is fed to an acid concentrator. The heat required for the acid concentration, 11.86 MW_t, is provided from a molten salt thermal storage device.

During the acid concentration process the vapor is continuously separated and condensed in a heat exchanger (HX-2) where 8.657 MW thermal energy can be recovered. The acid is concentrated to 84.5 weight percent (50 mol %) and is sent to an acid accumulator. Approximately 0.92 mole percent of H_2SO_4 is lost to vapor phase during acid concentration. The acid concentrator is operated 24 hours a day. During the nine hour daytime operation, the hot acid is pumped to 7.5 atmospheres and is sent to a receiver/vaporizer (SR-2) where solar energy supplies the necessary heat, 32.325 MW_t, for the acid vaporization. The acid vapor leaving SR-2 then enters a receiver/decomposer where 37.566 MW_t is required to decompose the acid vapor into H_2O, SO_3, SO_2 and O_2. The high temperature gases are then cooled to 100°C with a heat exchanger where 44.72 MW of thermal energy can be recovered. The state points for the nightime operation are shown in Table 3-2. During nighttime operation, only the acid concentrator and the HX-2 are operating and the acid accumulator is charged. The heat recovered from HX-1 and HX-2 is available to be utilized elsewhere in the plant.

The energy balance for Case 1, is summarized in Table 3-3 and illustrated in Figure 3-3.

In Tables 3-1, 3-2, 3-9 and 3-13 H_T^o and S^o are the assigned enthalpies and absolute entropies, respectively. Values of S^o were taken from Reference

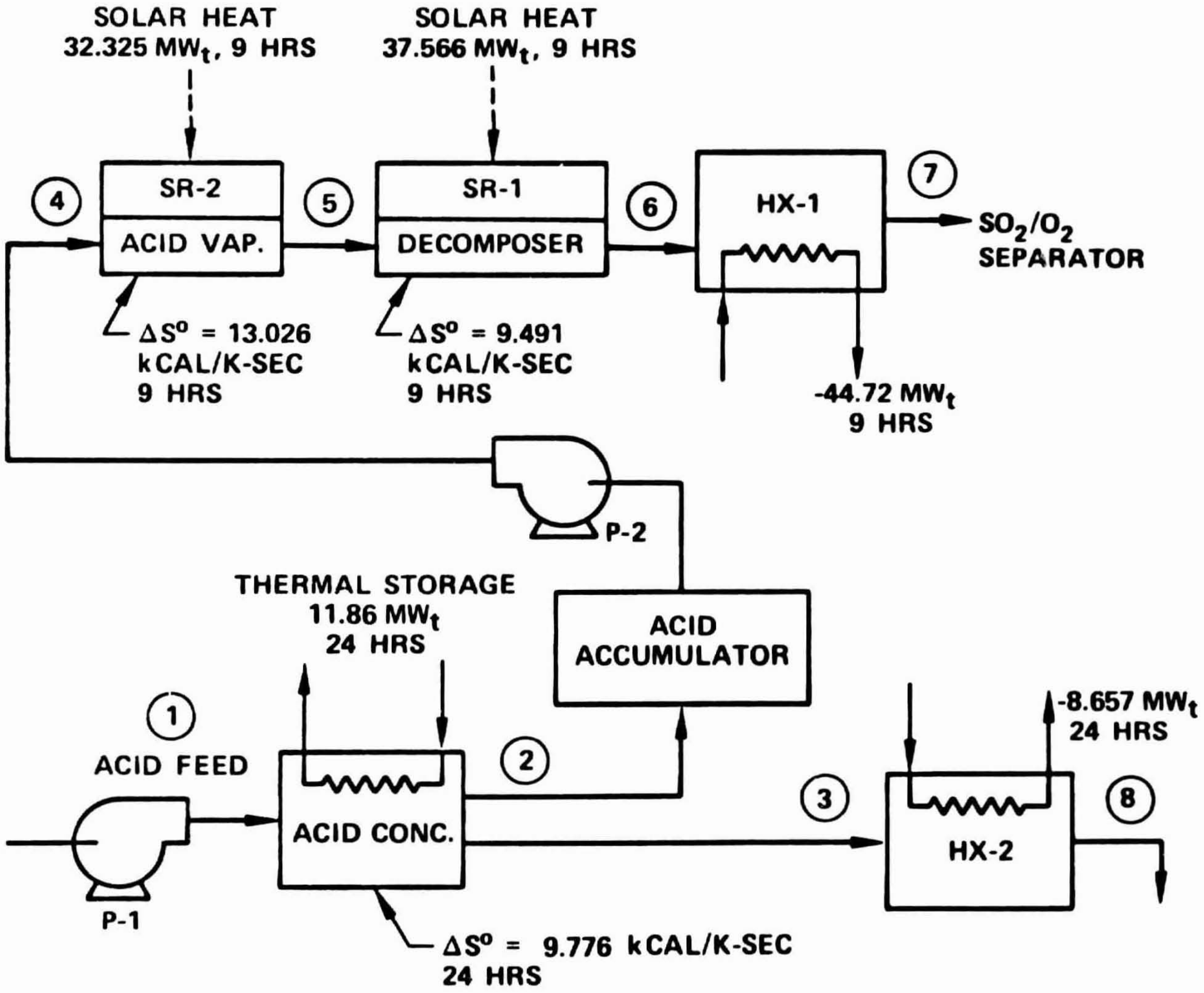

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FIGURE 3-1. PROCESS FLOW SHEET FOR CASE 1 (DAYTIME OPERATION)

TABLE 3-1. STATE POINTS FOR CASE 1
(DAYTIME OPERATION)

Stream No.	1	2	3	4	5	6	7	8
Temp., K	373	498.5	498.5	498.5	687.5	1173	373	373
Pressure, kPa	135.8	101.4	101.4	784.3	760.5	657.0	637.0	98.6
Flow Rate, $\frac{\text{g-Mol}}{\text{sec.}}$								
H_2SO_4	66.632	66.019	0.613	176.052	131.68		54.052	0.613
H_2O	241.876	66.019	175.857	176.052	220.136	352.104	298.052	175.857
SO_3					44.084	54.052		
SO_2						122.0	122.0	
O_2						61.0	61.0	
H^o_T, kCAL/SEC	-29883	-17071	-9980	-45524	-37803	-28804	-39512	-12048
S^o, kCAL/K-SEC	4.126	5.18	8.722	13.815	26.841	36.332	19.05	3.665

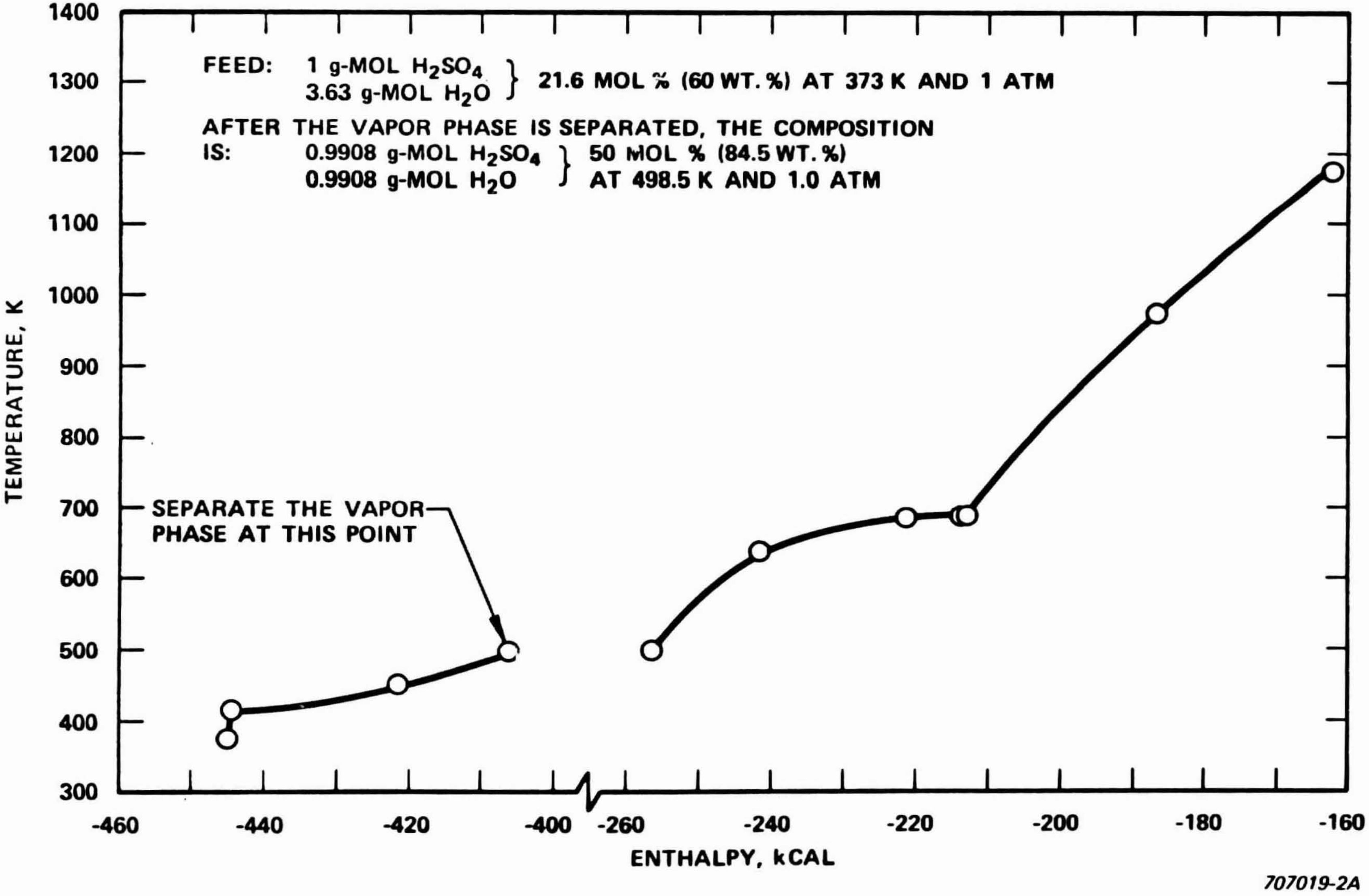


FIGURE 3-2. HEATING CURVE FOR 60 WEIGHT PERCENT H_2SO_4 SOLUTION

TABLE 3-2. STATE POINTS FOR CASE 1
(NIGHT-TIME OPERATION)

Stream Number	1	2	3	4	5	6	7	8
Temperature, K	373	498.5	498.5					373
Pressure, kPa	135.8	101.4	101.4					98.6
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$								
H_2SO_4	66.632	66.019	0.613					0.613
H_2O	241.876	66.019	175.857					175.857
SO_3								
SO_2								
O_2								
H^0_T, kCAL/SEC	-29883	-17071	-9980					-12048
S_o, kCAL/K-SEC	4.126	5.18	8.722					3.665

3-1. For each species, heats of formation were combined with sensible heats to give assigned enthalpies H_T^o. By definition, $H_T^o \equiv H_{298.15}^o + (H_T^o - H_{298.15}^o)$. We have arbitrarily assumed $H_{298.15}^o = (\Delta H_f^o)_{298.15}$. Therefore $H_T^o = (\Delta H_f^o)_{298.15} + (H_T^o - H_{298.15}^o)$. For elements, $(\Delta H_f^o)_{298.15} = H_{298.15}^o = 0$, which gives a common base with Reference 3-1.

3.3.1.2 PLANT OPERATING MODES

The plant operating modes must accommodate the variations in solar thermal input and the daily plant cycling. There are four basic operating modes for Case 1, defined as follows:

MAJOR CHARACTERISTICS

Mode 1:

- SR-1 supplying heat to decomposer
- SR-2 supplying heat to acid vaporizer
- A separate solar receiver supplying heat to thermal storage
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging and discharging at the same time
- Supplemental electric power (from grid) required

Mode 2:

- SR-1 Standby
- SR-2 Standby
- The receiver for thermal storage standby
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging
- Supplemental electric power required

Mode 3:

- All receivers operating to or from standby
- Supplemental electric power required

Mode 4:

- All subsystems shutdown

The individual operating modes of the various subsystems during the four plant operating modes are shown in Table 3-4.

3.3.1.3 PLANT OPERATION AND CONTROL REQUIREMENTS

Typical daily operation, assuming a clear day and design solar insolation levels, is as follows: The day begins in Mode 2. As the sun rises above the horizon, heliostats are commanded to acquire the sun and begin tracking. During this time the receivers are gradually brought up to temperature. When the intensity of solar radiation increases to a preset fraction, probably 50% of the design value, the receivers are brought on line and the plant switched into Mode 3. Assuming a day of design insolation, the solar intensity will continue to increase and the plant will enter Mode 1. Sufficient energy will be supplied to the thermal storage system to provide the energy charge for the required plant operation during the night. Following the solar peak at noon, insolation levels begin to decline and the plant moves successively back through Modes 3 and 2.

Whenever the solar flux on the receiver exceeds the receiver design capacity, selected heliostats are defocused to maintain a constant peak flux. For the purpose of this study, the design heat flux is assumed to be maintained for 9 hours. This period corresponds to operation in Mode 1. Mode 3 is a transition mode between Mode 1 and Mode 2. During cloud cover periods resulting in reduced solar flux, the acid feed to the acid vaporizer has to be reduced accordingly. Otherwise the acid in the liquid state would corrode the metal tubes in the following decomposer. Thus strict control of the acid flow rate to the acid vaporizer is extremely important.

TABLE 3-3. SUMMARY OF ENERGY BALANCE FOR CASE 1 (ADJUSTED TO 9 HOUR BASIS)

Total Energy Input:

Acid Concentrator --------------- 11.86 x 24/9 = 31.626 MW_t
Acid Vaporizer ------------------ 32.325 MW_t
Decomposer ---------------------- 37.566 MW_t

Total --------------------------- 101.517 MW_t

Heat Rejected:

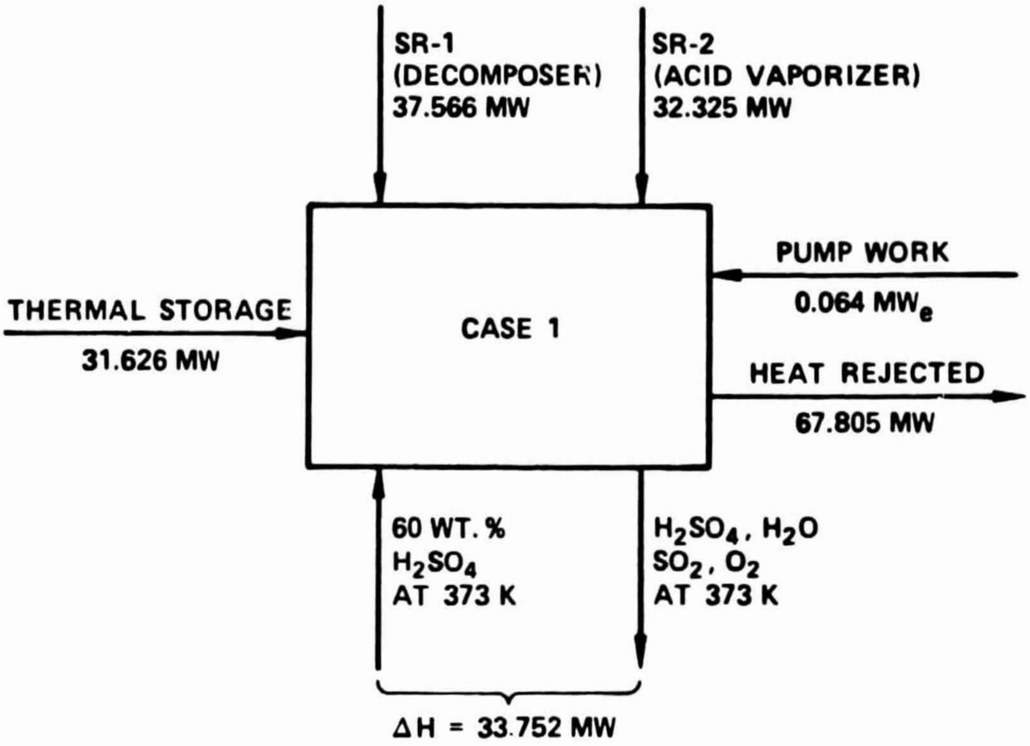
Heat Exchanger- 1 --------------- -44.72 MW_t
Heat Exchanger - 2 -------------- -8.657 x 24/9 = -23.085 MW_t

Total --------------------------- -67.805 MW_t

Pumping Power Requirements

For Acid Feed (P-1) ------------- 0.83 x 24/9 = 2.21 kWe
Acid to Vaporizer (P-2) --------- 56.3 kWe
For HX-1 Cooling Water ---------- 1.8 x 24/9 = 4.80 kWe
For HX-2 Cooling Water ---------- 0.95 kWe

Total --------------------------- 64.26 kWe



VALUES SHOWN ARE ADJUSTED TO 9 HOUR OPERATING BASIS

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FIGURE 3-3. DIAGRAM OF ENERGY BALANCE FOR CASE 1

TABLE 3-4. SUBSYSTEM OPERATING MODES FOR CASE 1

Subsystem	Mode 1	Mode 2	Mode 3	Mode 4
SR-1	0	S	0	S
SR-2	0	S	0	S
Thermal Storage	C,D	D	C,D	S
Acid Concentrator	0	0	0	S
Acid Accumulator	C,D	C	C,D	S
HX-1	0	S	0	S
HX-2	0	0	0	0
Aux. Electric Power	Yes	Yes	Yes	No

NOTE: 0 = Operating
S = Standby
C = Charging
D = Discharging

This acid flow rate must be controlled by (feedback from) the pressure and temperature instrumentation at the outlet of the acid vaporizer. Using data from these instruments, the control system would be designed to maintain the acid vaporizer outlet condition at saturation or, preferably, slightly superheated.

Since the acid vaporizer includes a ceramic material (silicon carbide) it is desirable to maintain the acid vaporizer at a certain minimum temperature at night to minimize the shock of thermal cycling due to heat-up and cool-down operations. The means to maintain the acid vaporizer temperature at night requires further study.

3.3.2 CASE 2 DESCRIPTION

3.3.2.1 SYSTEM ANALYSES

The process flow sheet is shown in Figure 3-4 and the state points in Table 3-5 for the daytime operation. Table 3-6 shows the state points for the night-time operation. As distinguished from Case 1, the majority of the heat (26.25 MW_t) for the acid vaporization is recuperated from the decomposed high temperature stream leaving the decomposer (SR-1). Figure 3-5 shows the heating and cooling curves for 50 mole percent (84.5 Wt. %) H_2SO_4 solution by plotting temperature versus enthalpy. The heating curve starts at the conditions of the acid leaving the acid accumulator (A), and ends at the conditions of the decomposed gases leaving the decomposer (B). The cooling curve then takes the decomposed gas stream from 900°C to 100°C (C). The amount of heat rejected from the decomposer that can be utilized as a heat source for acid vaporization is also shown in Figure 3-5 (D-E). Note that this is the energy represented by the portion (B-F) of the Heat Rejected curve. Due to the pinch point limit in the heat exchanger, the recuperated heat is insufficient for all the acid vaporization. Thus a small solar receiver/vaporizer is required to complete the final acid vaporization, some 6.076 MW thermal energy. Since a large quantity of heat has been recuperated in the acid vaporizer-1, the heat load in HX-1 for Case 2 is significantly smaller than Case 1. The rest of the system is the same as Case 1. The summary of the energy balance is shown in Table 3-7 and Figure 3-6.

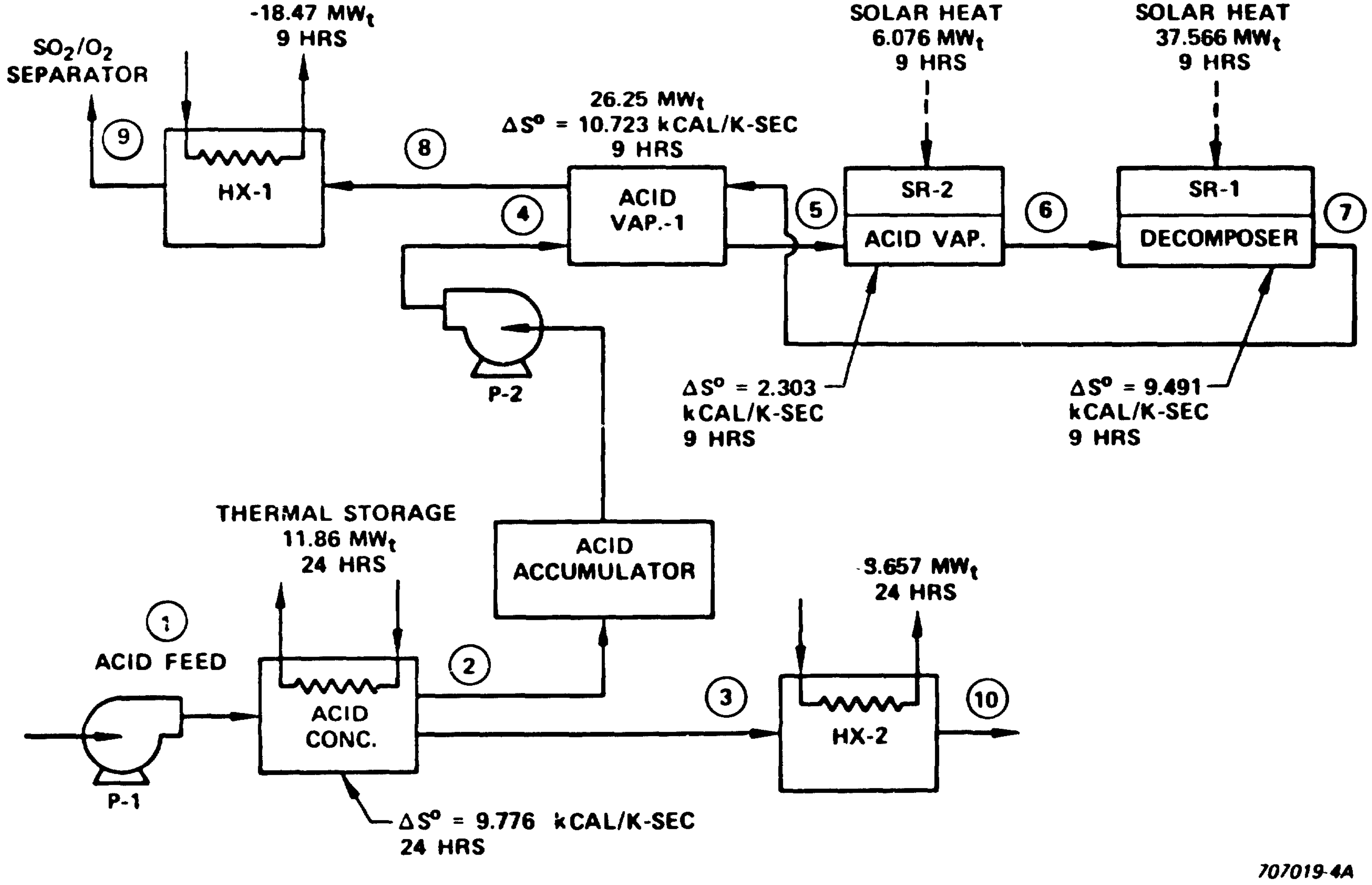

FIGURE 3-4. PROCESS FLOW SHEET FOR CASE 2 (DAYTIME OPERATION)

TABLE 3-5. STATE POINTS FOR CASE 2
(Daytime Operation)

Stream No.	1	2	3	4	5	6	7	8	9	10
Temp., K	373	498.5	498.5	498.5	682.4	687.5	1173	577.2	373	373
Pressure, kPa	135.8	101.4	101.4	800.9	778.8	760.5	657.0	637.0	617.9	98.6
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$										
H_2SO_4	66.632	66.019	0.613	176.052	147.899	131.968		54.052	54.052	0.613
H_2O	244.876	66.019	175.857	176.052	204.198	220.136	352.104	298.052	298.052	175.857
SO_3					28.149	44.084	54.052			
SO_2							122.0	122.0	122.0	
O_2							61.0	61.0	61.0	
H_T^o kCAL/SEC	-29883	-17071	-9980	-45524	-39254	-37803	-28804	-35074	-39512	-12048
S^o, kCAL/K-SEC	4.126	5.18	8.722	13.815	24.538	26.841	36.332	28.686	19.05	3.665

TABLE 3-6. STATE POINTS FOR CASE 2
(Daytime Operation)

Stream No.	1	2	3	4	5	6	7	8	9	10
Temp., K	373	498.5	498.5							373
Pressure, kPa	135.8	101.4	101.4							98.6
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$										
H_2SO_4	66.632	66.019	0.613							0.613
H_2O	244.876	66.019	175.857							175.857
SO_3										
SO_2										
O_2										
H^o_T kCAL/SEC	-29883	-17071	-9980							-12048
S^o, kCAL/K-SEC	4.126	5.18	8.722							3.665

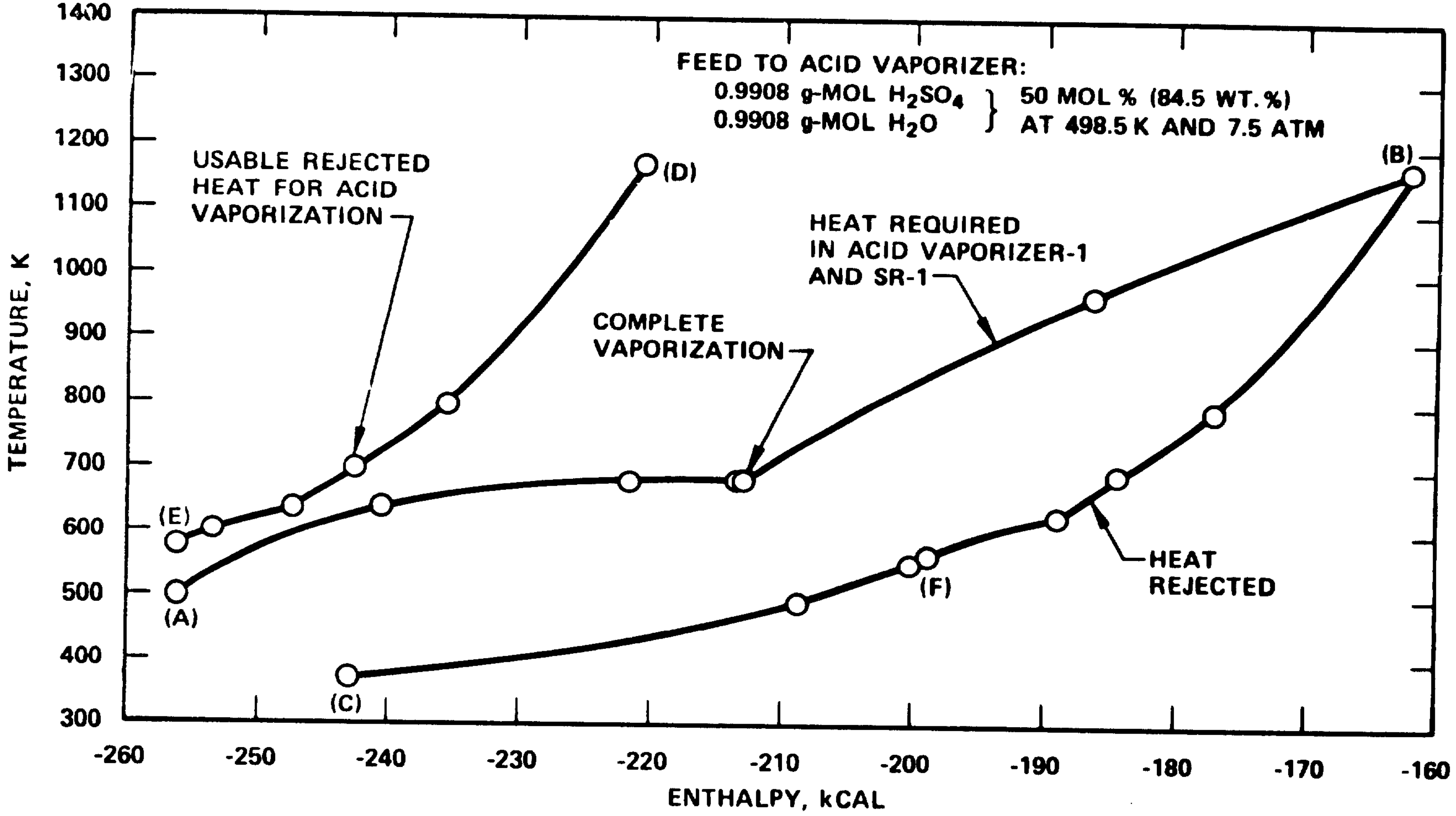

FIGURE 3-5. HEATING AND COOLING CURVES FOR 50 MOLE PERCENT (84.7 WT.%)

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TABLE 3-7. SUMMARY OF ENERGY BALANCE FOR CASE 2 (ADJUSTED TO 9 HOUR BASIS)

Total Energy Input: (During 9 hour Day-time)

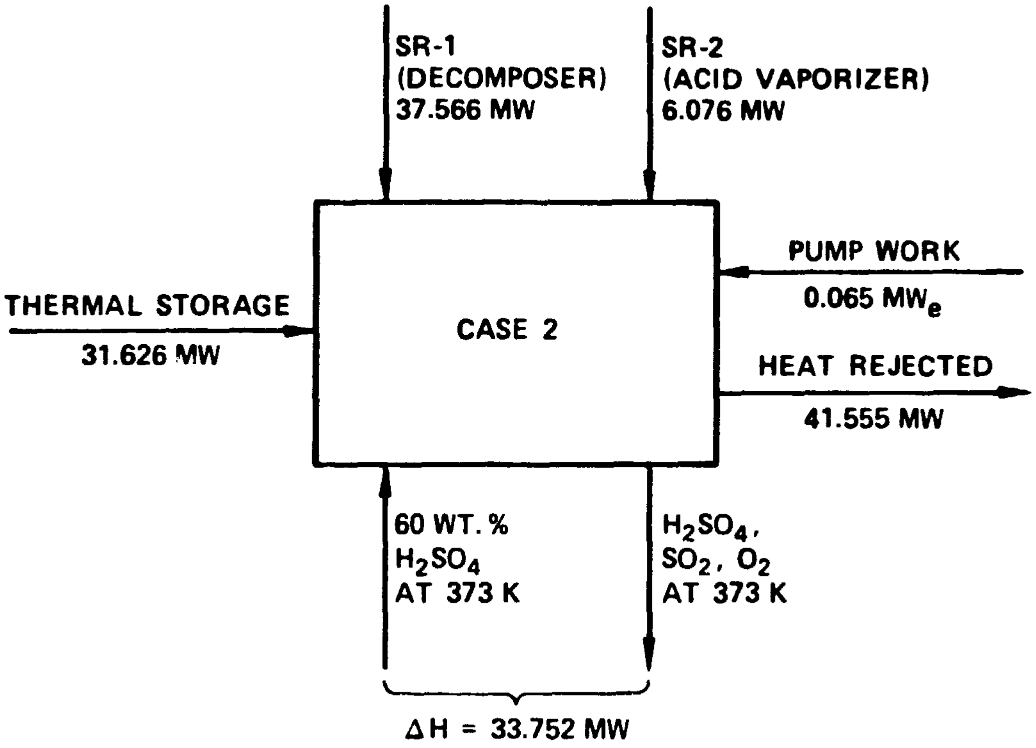
Acid Concentrator ------11.86 x 24/9 = 31.626 MW_t
Acid Vaporizer-1 (Use Recuperated Heat)
Acid Vaporizer-2--------------------------- 6.076 MW_t
Decomposer -------------------------------- 37.566 MW_t
Total ------------------------------------- 75.268 MW_t

Heat Rejected:

Heat Exchanger-1 ------------------------ -18.47 MW_t
Heat Exchanger-2 --------- -8.657 x 24/9 = -23.085
Total ----------------------------------- -41.555

Pumping Power Requirements:

For Acid Feed (P-1) --------- 0.83 x 24/9 = 2.21 kWe
For Acid to Vaporizer (P-2) ---------------- 56.3 kWe
For HX-1 Cooling Water -------- 2.03 x 24/9 = 5.41 kWe
For HX-2 Cooling Water---------------------- 0.95 kWe
Total -------------------------------------- 64.87 kWe



VALUES SHOWN ARE ADJUSTED TO 9 HOUR OPERATING BASIS

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FIGURE 3-6. DIAGRAM OF ENERGY BALANCE OF CASE 2

3.3.2.2 PLANT OPERATING MODES

There are four basic operating modes for Case 2, defined as follows:

MAJOR CHARACTERISTICS

Mode 1:

- SR-1 supplying heat to decomposer
- SR-2 supplying heat to acid vaporizer-2
- Decomposed hot gases from SR-1 supplying heat to acid vaporizer-1
- A separate solar receiver supplying heat to thermal storage
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging and discharging at the same time
- Supplemental electric power required

Mode 2:

- SR-1 Standby
- SR-2 Standby
- The receiver for thermal storage standby
- Acid vaporizer-1 standby
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging
- Supplemental electric power required

Mode 3:

- All receivers operating to or from standby
- Acid vaporizer-1 operating to or from standby
- Supplemental electric power required

Mode 4:

- All subsystems shutdown

The individual operating modes of the various subsystems during the four plant operating modes are shown in Table 3-8.

TABLE 3-8. SUBSYSTEM OPERATING MODES FOR CASE 2

Subsystems	Mode 1	Mode 2	Mode 3	Mode 4
SR-1	0	S	0	S
SR-2 (AV-2)	0	S	0	S
AV-1	0	S	0	S
Thermal Storage	C,D	D,	C,D	S
Acid Concentrator	0	0	0	S
Acid Accumulator	C,D	C	C,D	S
HX-1	0	S	0	S
HX-2	0	0	0	S
Aux. Electric Power	Yes	Yes	Yes	No

NOTE: 0 = Operating
S = Standby
C = Charging
D = Discharging

3.3.2.3 PLANT OPERATION AND CONTROL REQUIREMENTS

The plant operation for Case 2 is the same as Case 1 described in 3.3.1.3. The control requirements are also the same as Case 1 except that the acid flow rate to the acid vaporizer-1 must be controlled by (feedback from) the pressure and temperature at the outlet of the last acid vaporizer (SR-2).

3.3.3 CASE 3 DESCRIPTION

3.3.3.1 SYSTEM ANALYSES

Case 3 is similar to Case 1 except that the thermal energy required for the acid vaporization comes from thermal energy storage. Figure 3-7 shows the process flow sheets for daytime operation. The state points are shown in Tables 3-9 and 3-10 for the daytime and the nightime operation, respectively. The summary of the energy balance is shown in Table 3-11 and Figure 3-8.

3.3.3.2 PLANT OPERATING MODES

There are four basic operating modes for Case 3, defined as follows:

MAJOR CHARACTERISTICS

Mode 1:

- SR-1 supplying heat to decomposer
- A separate solar receiver supplying heat to thermal storage
- Acid vaporizer drawing heat from thermal storage
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging and discharging at the same time
- Supplemental electric power required

Mode 2:

- SR-1 Standby
- The receiver for thermostorage standby
- Acid vaporizer standby
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging
- Supplemental electric power required.

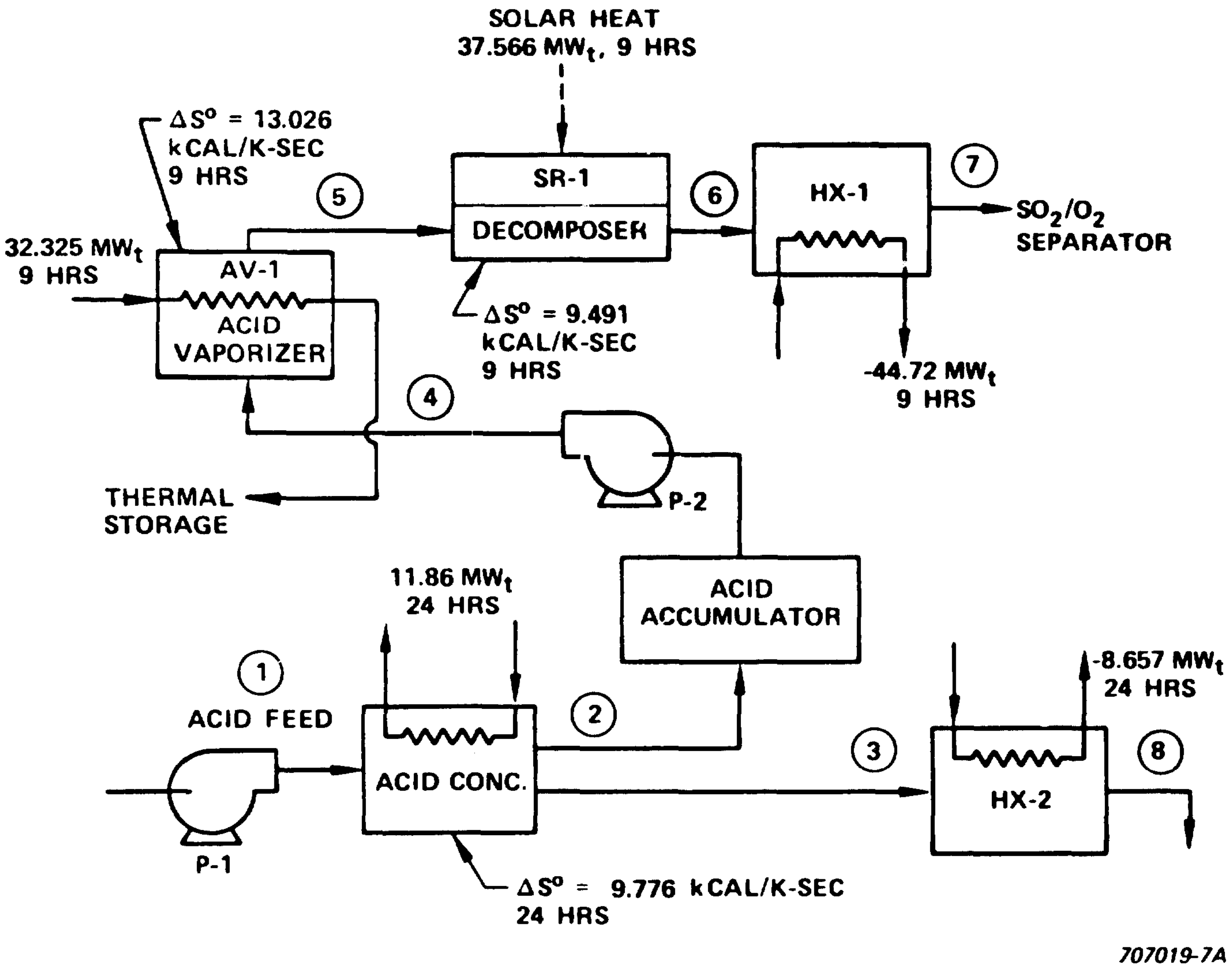


FIGURE 3-7. PROCESS FLOW SHEET FOR CASE 3 (DAYTIME OPERATION)

TABLE 3-9. STATE POINTS FOR CASE 3

(DAYTIME OPERATION)

Stream No.	1	2	3	4	5	6	7	8
Temp., K	373	498.5	499.5	498.5	687.5	1173	373	373
Pressure, kPa	135.8	101.4	101.4	797.8	760.5	657.0	637.0	
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$								
H_2SO_4	66.632	66.019	0.613	176.052	131.968		54.052	0.613
H_2O	241.876	66.019	175.857	176.053	220.136	352.014	298.052	175.857
SO_3					44.084	54.052		
SO_2						122.0	122.0	
O_2						61.0	61.0	
H^o_T, kCAL/SEC	-29883	-17071	-9980	-45524	-37803	-28804	-39512	12048
S^o, kCAL/K-SEC	4.126	5.18	8.722	13.815	26.841	36.332	19.05	3.665

TABLE 3-10. STATE POINTS FOR CASE 3
(DAY TIME OPERATION)

Stream Number	1	2	3	4	5	6	7	8
Temp., K	373	498.5	498.5					373
Pressure, kPa	135.8	101.4	101.4					98.6
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$								
H_2SO_4	66.632	66.019	0.613					0.613
H_2O	241.876	66.019	175.857					175.857
SO_3								
SO_2								
O_2								
H_T^o, kCAL/SEC	-29883	-17071	-9980					-12048
S^o, kCAL/K-SEC	4.126	5.18	8.722					3.665

TABLE 3-11. SUMMARY OF ENERGY BALANCE FOR CASE 3
(ADJUSTED TO 9 HOUR BASIS)

Total Energy Input: (During 9 hour Day-time)

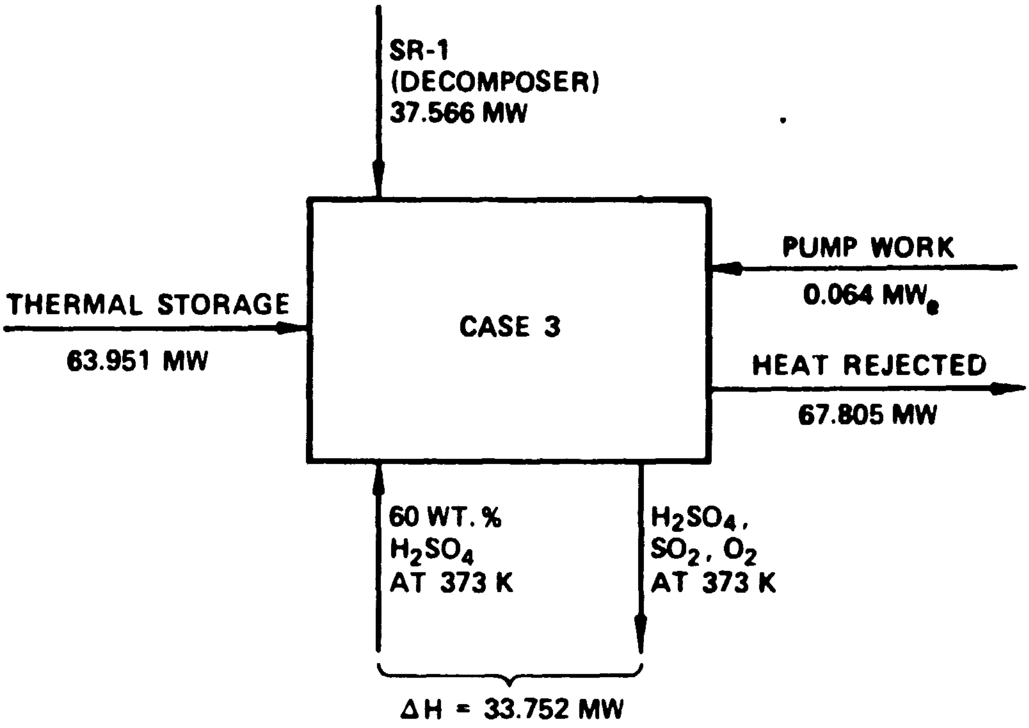
Acid Concentrator ------------ 11.86 x 24/9 = 31.626 MWt
Acid Vaporizer ---------------------------- 32.325 MWt
Decomposer -------------------------------- 37.566 MWt
Total ------------------------------------- 101.517 MWt

Heat Rejected:

Heat Exchanger-1 ------------------------- -44.72 MWt
Heat Exchanger-2 ----------- -8.657 x 24/9 = -23.085 MWt
Total ------------------------------------ -67.805 MWt

Pumping Power Requirements:

For Acid Feed (P-1) --------- 0.83 x 24/9 = 2.21 kWe
For Acid to Vaporizer (P-2) -------------- 56.3 kWe
For HX-1 Cooling Water ------- 1.8 x 24/9 = 4.80 kWe
For HX-2 Cooling Water ------------------- 0.95 kWe
Total ------------------------------------ 64.26 kWe



VALUES SHOWN ARE ADJUSTED TO 9 HOUR OPERATING BASIS

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FIGURE 3-8. DIAGRAM OF ENERGY BALANCE FOR CASE 3

Mode 3:

- All receivers operating to or from standby
- Supplemental electric power required

Mode 4:

- Subsystem shutdown

The individual operating modes of the various subsystems during the four plant operating modes are shown in Table 3-12.

3.3.3.3 PLANT OPERATION AND CONTROL REQUIREMENTS

The plant operation and control requirements for Case 3 are the same as Case 1.

3.3.4 CASE 4 DESCRIPTION

3.3.4.1 SYSTEM ANALYSES

Unlike the previous three cases, Case 4 uses helium as a heat transport medium to decouple the acid decomposition process from the solar heat source. The thermal energy required for the final stage of the acid vaporization and the acid decomposition is provided through a helium loop that receives solar energy from a solar receiver. The rest of the system is the same as Case 2. Although the electric motor driven helium circulator consumes 4338 kW electricity, the energy is largely (subject to equipment efficiency) recovered as heat in the helium. Figure 3-9 shows the process flow sheet for the daytime operation. Table 3-13 and 3-14 are the state points for the daytime and the night-time operation, respectively. The energy balance is summarized in Table 3-15 and Figure 3-10.

TABLE 3-12. SUBSYSTEM OPERATING MODES FOR CASE 3

Subsystem	Mode 1	Mode 2	Mode 3	Mode 4
SR-1	0	S	0	S
Acid Vaporizer	0	S	0	S
Thermal Storage	C,D	D	C,D	S
Acid Concentrator	0	0	0	S
Acid Accumulator	C,D	C	C,D	S
HX-1	0	S	0	S
HX-2	0	0	0	S
Aux. Electric Power	Yes	Yes	Yes	No

NOTE: 0 = Operating
S = Standby
C = Charging
D = Discharging

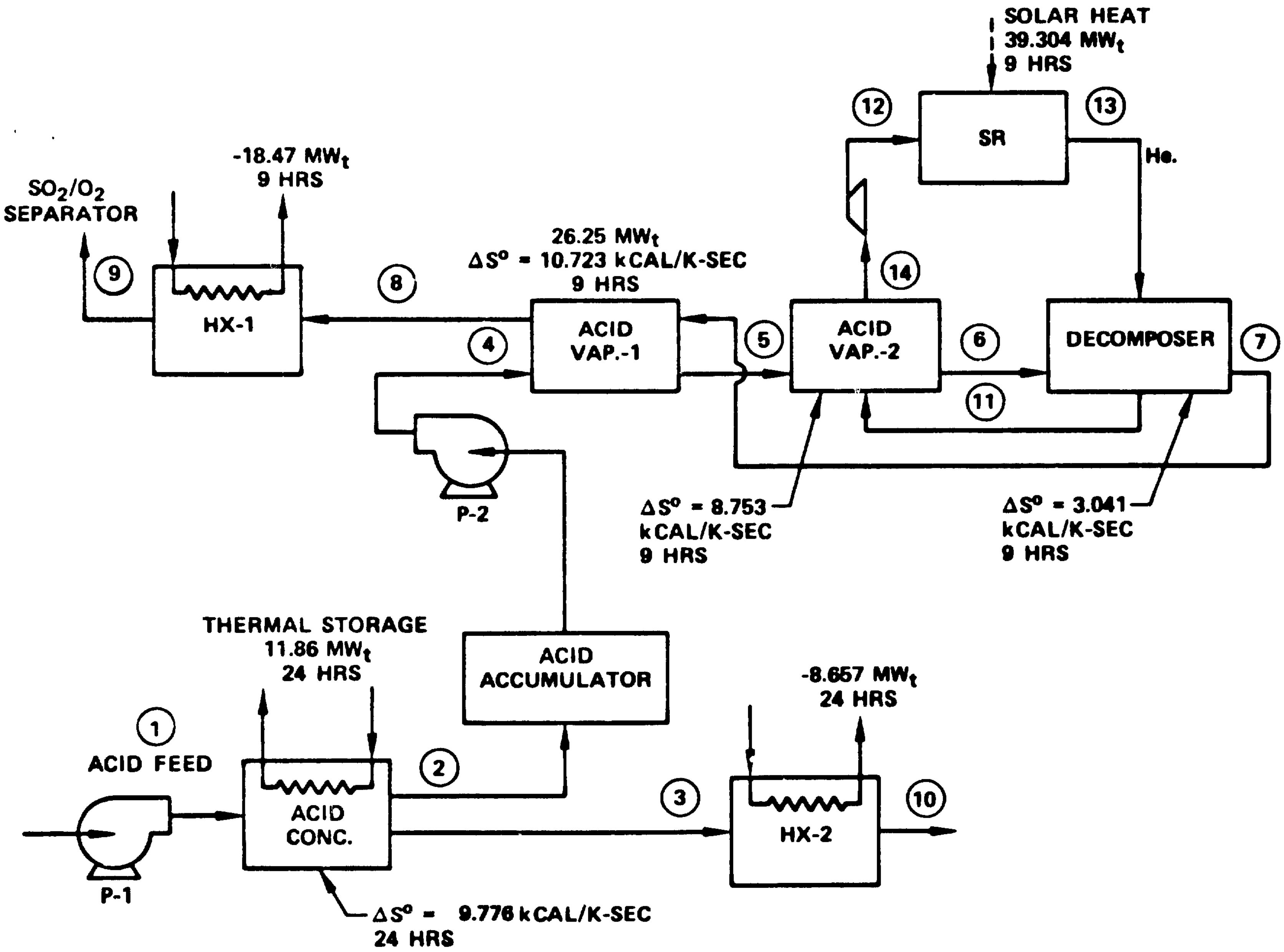


FIGURE 3-9. PROCESS FLOW SHEET FOR CASE 4

TABLE 3-13. STATE POINTS FOR CASE 4

(DAY-TIME OPERATION)

Stream No.	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Temp., K	373	498.5	498.5	498.5	682.4	687.5	1173	577.2	373	373	854.1	835.8	1249.7	790.2
Pressure, kPa	135.8	101.4	101.4	800.9	778.8	760.5	577.8	560.5	543.6	98.6	766.3	847.7	766.9	760.5
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$														
H_2SO_4	66.632	66.019	0.613	176.052	147.899	131.968		54.052	54.052	0.613				
H_2O	241.876	66.019	175.857	176.052	204.198	220.136	352.104	298.052	298.052	175.857				
SO_3					28.149	44.084	54.052							
SO_2							122.0	122.0	122.0					
O_2							61.0	61.0	61.0					
He											4566.909	4566.909	4566.909	4566.909
H^o_T, kCAL/SEC	-29883	-17071	-9980	-45524	-39254	-37803	-28804	-35074	-39512	-12048				
S^o, kCAL/K-SEC	4.126	5.18	8.722	13.815	24.538	26.841	36.332	28.868	19.05	3.665				

TABLE 3-14. STATE POINTS FOR CASE 4

(NITE-TIME OPERATION)

Stream Number	1	2	3	4	5	6	7	8	9	10	11	12	13	14
Temperature, K	373	498.5	498.5							373				
Pressure, kPa	135.8	101.4	101.4							98.6				
Flow Rate, $\frac{\text{g-Mol}}{\text{Sec}}$														
H_2SO_4	66.632	66.019	0.613							0.613				
H_2O	241.876	66.019	175.857							175.857				
SO_3														
SO_2														
O_2														
H^o_T, kCAL/SEC	-29883	-17071	-9980							-12048				
S^o, kCAL/K-SEC	4.126	5.18	8.722							3.665				

TABLE 3-15. SUMMARY OF ENERGY BALANCE FOR CASE 4
(Adjusted to 9 Hour Basis)

Total Energy Input (During 9 hr. Daytime)

Acid Concentrator -- 1.86 x 24/9 = 31.636 MW_t
Acid Vaporizer -1 (Use recuperated heat)
Acid Vaporizer -2 Heat from helium Loop------ 39.304 MW_t
Total -- 70.93 MW_t

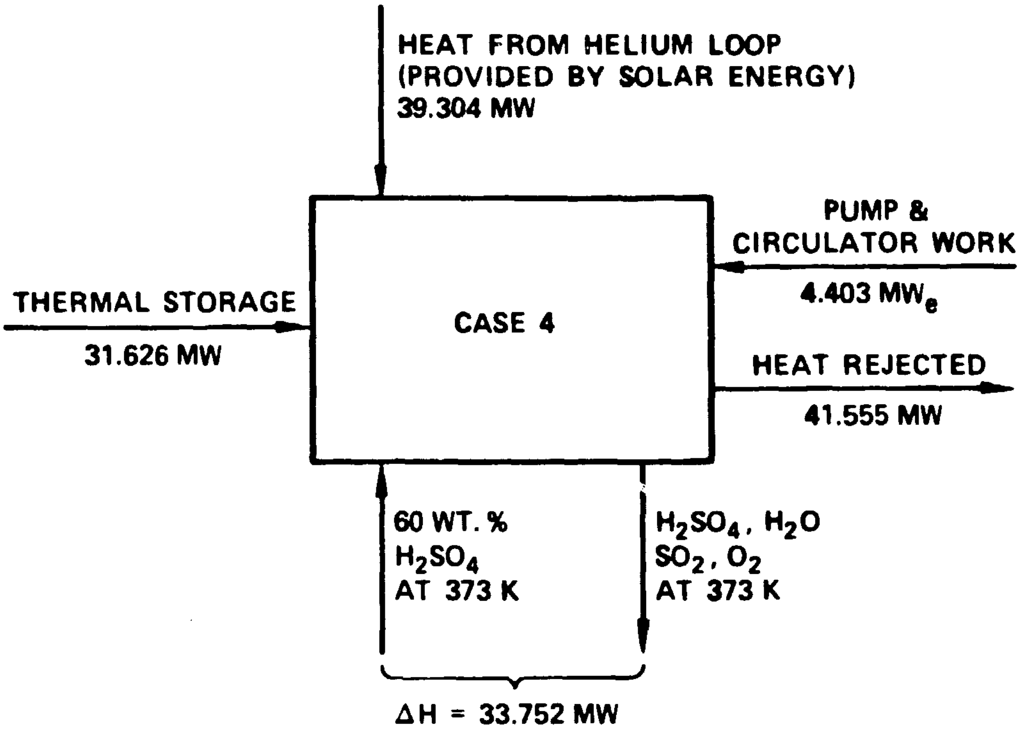
Heat Rejected:

Heat Exchanger - 1 ---------------------------- -18.47 MW_t
Heat Exchanger - 2 ---------------------------- -8.657 x 24/9 = -23.085 MW_t
Total -- -41.555 MW_t

Pumping and Circulating Power Requirements:

For acid feed (P-1) -------------- 0.83 x 24/9 = 2.21 kW_e
For acid to vaporizer (P-2) ------------------- = 56.3 kW_e
For HX-1 Cooling Water ------------ 2.03 x 24/9 = 5.41 kW_e
For HX-2 Cooling Water ------------------------- 0.95 kW_e
Helium Circulator ----------------------------- 4338.0 kW_e*
Total -- 4402.87 kW_e

*4338 kW_e will raise the helium temperature so that the energy input to the solar receiver for helium heating can be reduced by 4338 kW_t.



VALUES SHOWN ARE ADJUSTED TO 9 HOUR OPERATING BASIS

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FIGURE 3-10. DIAGRAM OF ENERGY BALANCE FOR CASE 4

3.3.4.2 PLANT OPERATING MODES

There are four basic operating modes for Case 4, defined as follows:

MAJOR CHARACTERISTICS

MODE 1:

- SR suppling heat to helium loop.
- Helium loop supplying heat to decomposer.
- Helium loop supplying heat to acid vaporizer-2.
- Decomposed hot gas from decomposer supplying heat to acid vaporizer-1.
- A separate solar receiver supplying heat to thermal storage.
- Acid concentrator drawing heat from thermal storage.
- Acid accumulator charging and discharging at the same time.
- Supplemental electric power required.

Mode 2:

- SR standby
- Decomposer standby
- Acid vaporizer-1 standby
- Acid vaporizer-2 standby
- The receiver for thermal storage standby
- Acid concentrator drawing heat from thermal storage
- Acid accumulator charging
- Supplemental electric power required

Mode 3:

- All receivers operating to or from standby
- Decomposer operating to or from standby
- Acid vaporizer-1 operating to or from standby
- Acid vaporizer-2 operating to or from standby
- Supplemental electric power required

Mode 4:

- All subsystems shutdown

The individual operating modes of the various subsystems during the four plant operating modes are shown in Table 3-16.

3.3.4.3 PLANT OPERATION AND CONTROL REQUIREMENTS

The plant operation and control requirements are the same as for Case 2. The acid flow rate to the acid vaporizer-1 must be controlled by (feedback from) the pressure and temperature at the outlet of the last acid vaporizer (AV-2).

3.4 REFERENCES

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3-2 H. Lennartz, "Experimental Investigation of the Vapor Liquid Phase Equilibrium of the System $H_2O-H_2SO_4$," Doctoral thesis submission at RWTH Aachen (University of Aachen), March 31, 1980. We acknowledge the advance availability of this data to us by Professors H. Hartmann and K. F. Knoche of the University of Aachen prior to publication.

3-3 J. Helmig, "Computer Print-OUt of P,T,X data on the $H_2O-H_2SO_4$ System," Doctoral thesis at University of Aachen, West Germany, October, 1981. Experimental data are based on (3-2) above.

3-4 R. W. Werner and F. L. Ribe, "Synfuels from Fusion Using the Tandem Mirror Reactor and a Thermochemical Cycle to Produce Hydrogen," Lawrence Livermore Laboratory, November, 1982.

TABLE 3-16
SUBSYSTEM OPERATING MODES FOR CASE 4

Subsystems	Mode 1	Mode 2	Mode 3	Mode 4
SR	0	S	0	S
Decomposer	0 RTS	0	S	
AV-1	0	S	0	S
AV-2	0	S	0	S
Thermal Storage	C, D	TD	C, D	S
Acid Concentrator	0	0	0	TS
Acid Accumulator	C, D	C	C, D	TS
HX-1	0	S	0	S
HX-2	0	0	0	S
Aux. Electric Power	yes	yes	yes	no

4.0 EQUIPMENT CONCEPTUAL DESIGN

The design, sizing and engineering evaluation of major pieces of equipment are reported in this section. Subsections, 4.1 through 4.4, provide results for the four concepts discussed under System Studies, Section 3.0.

4.1 CASE #1

The equipment discussed in this subsection meet the requirements of the subsystem described in Section 3.3.1. The study results are presented in two subsections: 4.1.1 - Equipment Design and 4.1.2 - Equipment Cost Estimates.

4.1.1 Equipment Design

Each of the major pieces of equipment, that make up the subsystem of Case #1, are treated in the following subsections:

4.1.1.1 - Receiver/Decomposer (SR-1), 4.1.1.2 - Receiver/ Vaporizer (SR-2),
4.1.1.3 - Acid Concentration, 4.1.1.4 Heat Exhanger-1 (HX-1),
4.1.1.5 - Heat Exchanger-2 (HX-2), 4.1.1.6 - Acid Accumulator,
4.1.1.7 - Low Temperature Acid Pump, 4.1.1.8 - High Temperature Acid Pump.

4.1.1.1 Receiver/Decomposer (SR-1)

This acid decomposer is a direct solar radiant cavity type receiver. Tower mounted, it will receive insolation from a field of heliostats. The study results are presented below in two subsections: 4.1.1.1.1 - Design Analyses and 4.1.1.1.2 Design and Fabrication.

4.1.1.1.1 Design Analyses

The solar heat flux design limit is based on temperature limiting criteria due to material property considerations. The acid decomposer is divided into two separate sections: a preheat section and a catalytic reaction section. In the preheat section the saturated acid vapor is heated from 777.7°F (414.3°C) to 1292°F (700°C) and decomposed from H_2SO_4 into H_2O and SO_3. The tubes contain a packed bed of ceramic pellets. The total flow rate is 162,216 lb/hr

(20.439 kg/sec). In the catalytic reaction section SO_3 is decomposed into SO_2 and O_2 at the maximum process fluid temperature of 1652°F (900°C). The tubes are packed with aluminum oxide catalyst pellets to accelerate the rate of sulfur trioxide decomposition. Since the process fluid in the decomposer is in the gaseous phase, it is assumed that the inside film coefficient (hi) is constant. For the heat transfer from the tube wall to the process gas stream, the Beek correlation, Reference 4-1, was employed for the effective film coefficient, h, with:

$$h = \frac{k_g}{D_p} (2.58\ Re_p^{1/3}\ Pr^{1/3} + 0.094\ Re_p^{0.8}\ Pr^{0.4})$$

where

k_g = thermal conductivity of process gas
D_p = catalyst particle diameter
Re_p = particle Reynolds number based on superficial gas velocity
P_r = Prandtl number of process gas

For the pressure drop calculations, the Hicks correlations, Reference 4-2, were employed:

$$F = 3.4 \cdot \frac{(1 - \varepsilon)^{1.2}}{\varepsilon^3} Re_p^{-0.2}$$

$$-\frac{\Delta P}{L} = \frac{1}{\Psi\ D_p} \cdot \frac{1}{2} \rho_g\ \mu_o^2 \cdot 4\ (\frac{F}{g_c})$$

where

L = tube length, ft
Re_p = particle Reynolds number based on superficial gas velocity
μ_o = average velocity, ft/sec
ρ_g = fluid density, lb/ft^3
D_p = catalyst particle diameter, ft

ε = voidage (fractional free volume)
($\varepsilon = 0.45$ was used in the analyses)
Ψ = shape factor of the solid
($\Psi = 0.8745$ for cylindrical shape)

ANALYSES ON CATALYTIC SECTION

The design criteria are: a) to obtain 90% equilibrium conversion from SO_3 to SO_2 and O_2; b) the maximum tube wall temperature ∿ 1700°F (927°C); c) the maximum pressure drop in the tube section is approximately 15 psi. Based on these criteria, the net heat flux to the tubes is varied for the studies. Once the net heat flux to the tube is established, the required tube surface area is defined as follows:

A_t = Required tube surface area (m^2)
$A_t = Q/q = \pi\ D_o\ L\ N$

where

A_t = Required tube surface area (m^2)
Q = Thermal requirement (kW) = 18.128 kW for the catalyst section
q = Permissible net heat flux to tubes (kW/M^2)
D_o = Tube outside diameter (m)
L = Tube length (m)
N = Number of tubes

It is further assumed that the tubes are spaced one diameter apart or two diameters from center to center so that approximately one-half of the solar flux is reflected on the back side of the tube after passing through the one diameter space between tubes.

The first task is to select a tube diameter. To select a tube diameter, the pressure drop inside the tube (the most important factor) is plotted against the tube diameter in Figure 4-1 for a fixed tube length of 7 ft (2.1336 m) and an average heat flux of 14582 Btu/hr-ft^2 (46 kW/m^2). It is apparent that a tube diameter of 2.0 inches is indicated to meet the pressure drop criterion of

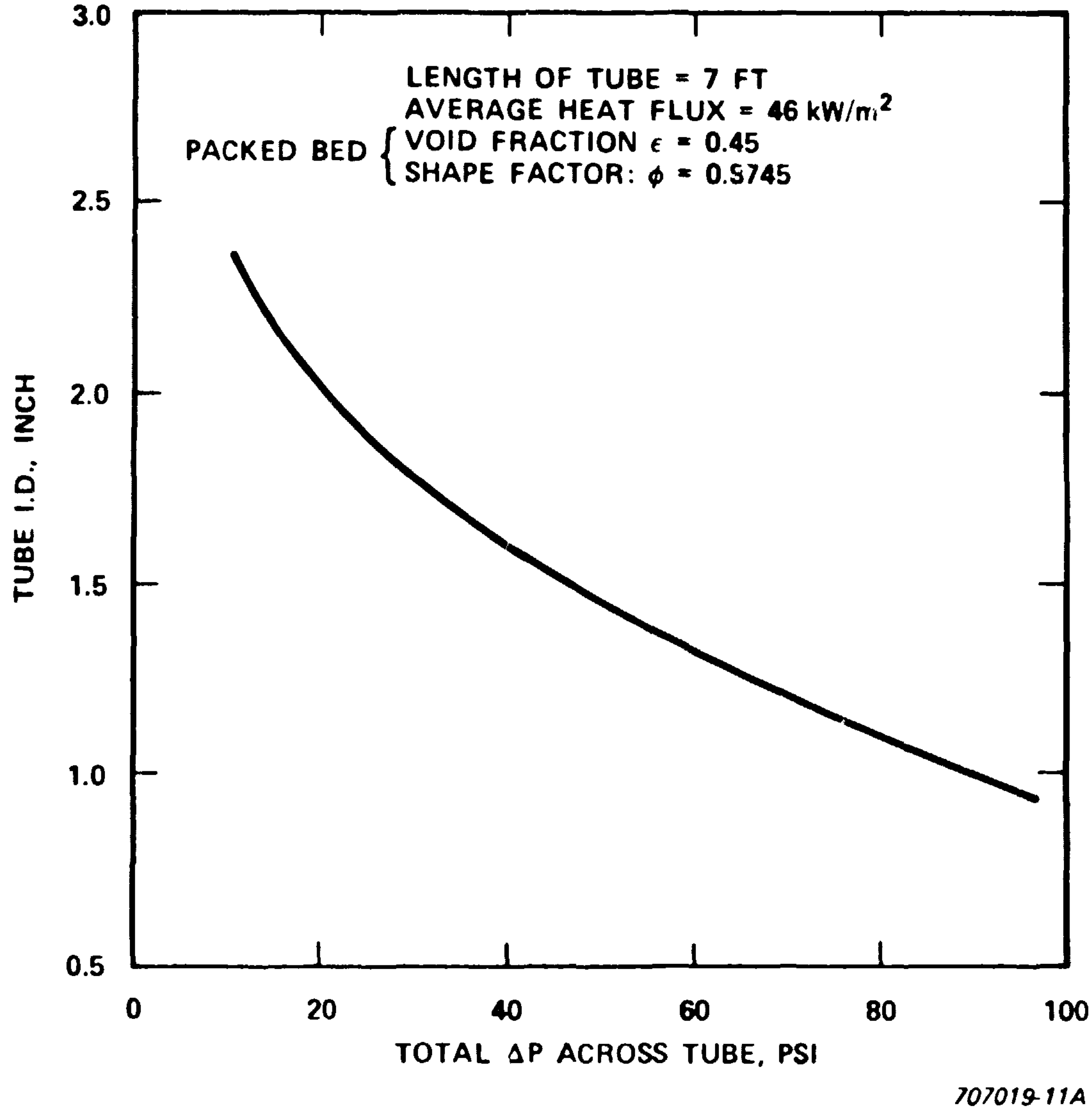

FIGURE 4-1. PRESSURE DROP VS. TUBE DIAMETER

15 psi. Thus a tube inside diameter of 2.0 inches was selected for subsequent studies. Figure 4-2 shows the pressure drop versus inside film coefficient for the cases of average heat flux at 35 kW/m^2 and 40 kW/m^2. As can be seen, the pressure drop is approximately doubled when the film coefficient is increased by 15 Btu/hr-ft^2-°F. Figure 4-3 shows inside film coefficient versus average temperature drop from tube inner wall to process fluid for the cases with average heat flux at 35 kW/m^2 and 40 kW/m^2. Figure 4-4 shows average heat flux versus inside film coefficient and average temperature drop from tube inner wall to process fluid with a fixed tube length at 6 ft. The summary of the analyses is also shown in Table 4-1. In all the analyses, it was conservatively assumed that 69% conversion of SO_3 to SO_2 and O_2 was attained at the maximum process fluid temperature of 900°C (1652°F). By carefully reviewing Table 4-1, a case with a minimum average ΔT (63.8°C) and a reasonable ΔP (11.32 psi) was selected as the reference design configuration. The reference design has 1281 tubes with 2.067 inches tube inside diameter and 0.154 inch wall, and each tube is 7 feet in length.

A computer program, Reference 4-3, was then used to calculate a more accurate and detailed performance for reference design. A rate equation was included in this computer program to calculate actual SO_3 conversion in each section of the tube longitudinally. The performance data are as follows:

Flow rate per tube = 126.6 lb/hr
Heat absorbed per tube = 48809.9 Btu/hr
Average heat flux = 12880.3 Btu/hr-ft^2 (40.63 kW/m^2)
Inlet heat flux = 20274.3 Btu/hr-ft^2 (63.96 kW/m^2)
Outlet heat flux = 6051.4 Btu/hr-ft^2 (19.09 kW/m^2)
Maximum process fluid temperature = 900.4°C (1652.7°F)
Maximum inside tube wall temperature = 930.7°C (1707°F)
Maximum outside tube wall temperature = 933.4°C (1712°F)
SO_3 equilibrium conversion = 78.2 mole %
SO_3 actual conversion = 78.0 mole %
Pressure drop = 11.32 psi

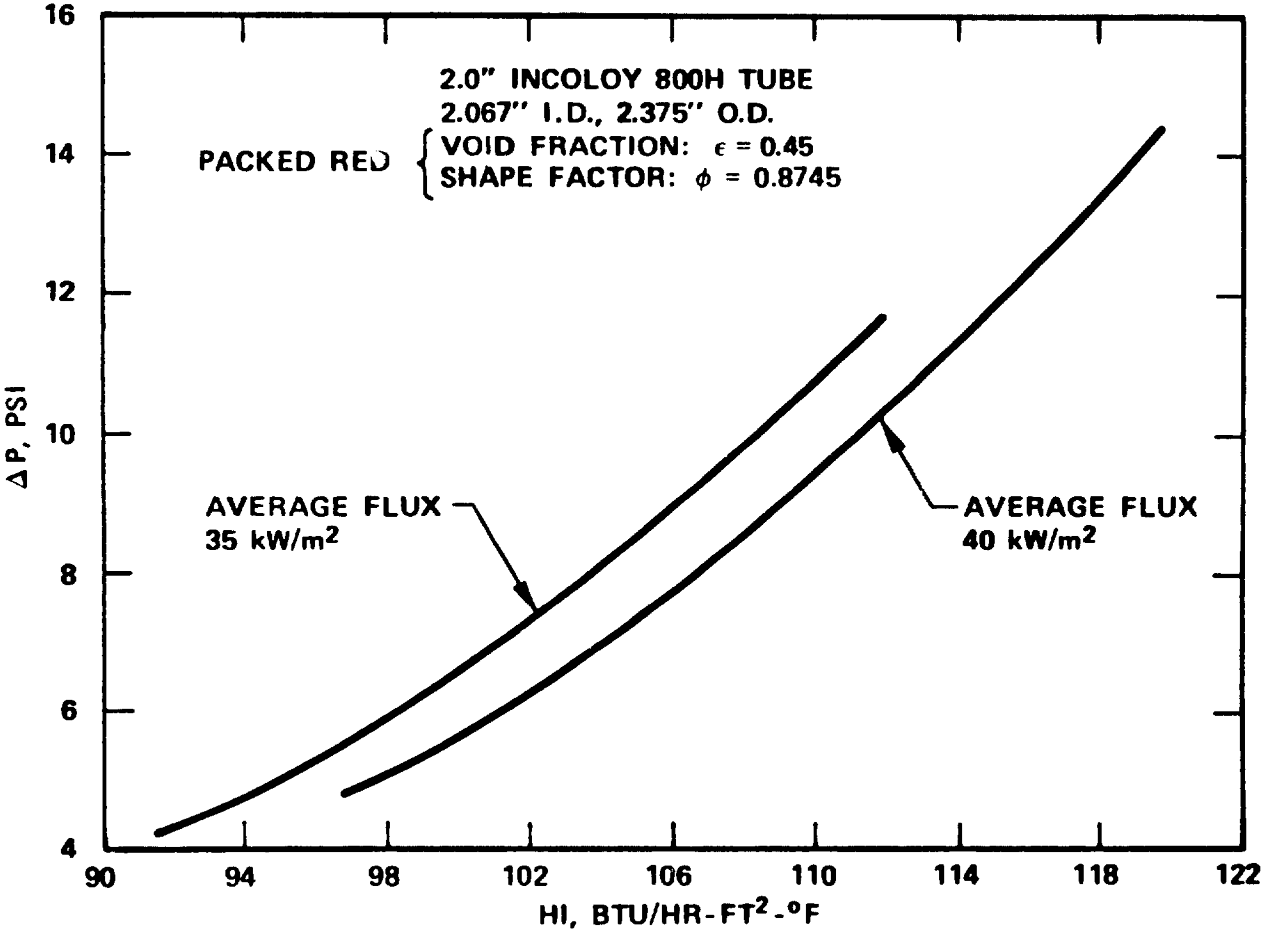

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FIGURE 4-2. PRESSURE DROP VERSUS INSIDE FILM COEFFICIENT

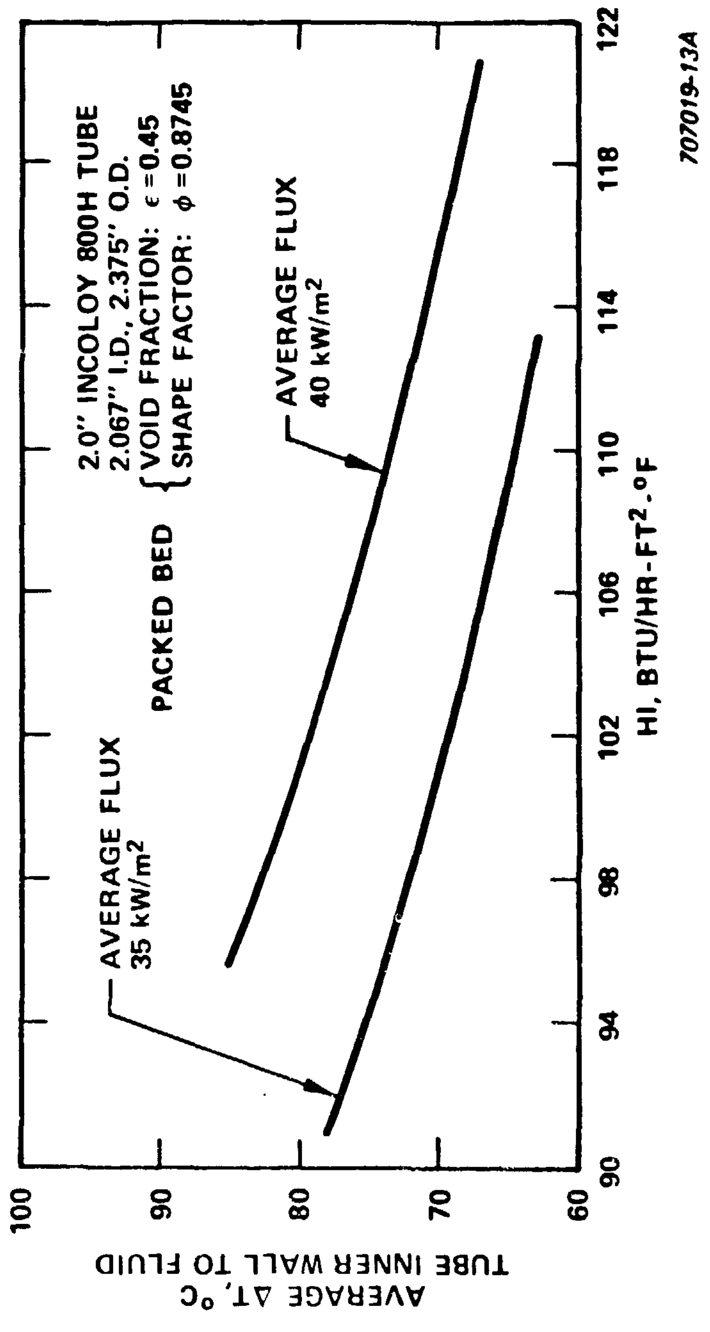

FIGURE 4-3. INSIDE FILM COEFFICIENT VERSUS TEMPERATURE DROP

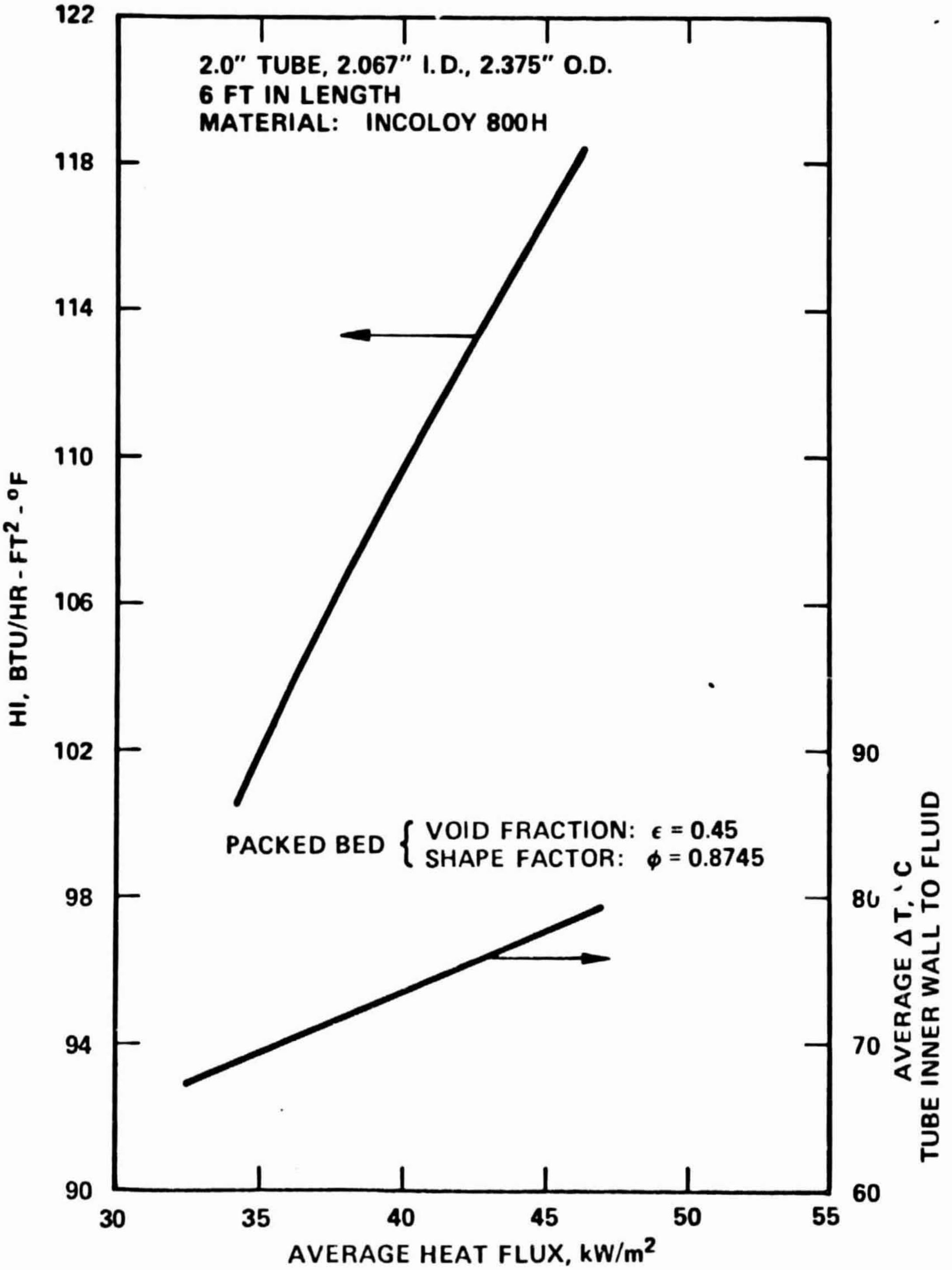

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FIGURE 4-4. AVERAGE HEAT FLUX VERSUS FILM COEFFICIENT AND TEMPERATURE DROP

TABLE 4-1. SUMMARY OF DECOMPOSER ANALYSES (CATALYST SECTION)

Fix the Average Heat Flux at 35 kW/m^2

L(ft)	N_{tube}	hi (Btu/hr-ft^2°F)	ΔP(Psi)	ΔT_{avg} (°C)	W (lb/hr-Tube)
5	1794	92.5	4.41	76.5	90.4
6	1494	102	7.36	69.4	108.6
7	1281	111	11.32	63.8	126.6

Fix the Average Heat Flux at 40 kW/m^2

5	1570	99.3	5.6	81.4	103.3
6	1308	109.7	9.3	73.7	124.0
7	1121	119.5	14.4	67.7	144.7

Fix the Tube Length at 6 ft.

Avg. Flux (kW/m^2)	N_{tube}	hi (Btu/hr-ft^2-°F)	ΔP(Psi)	ΔT_{avg}(°C)	W (lb/hr-tube)
35	1494	102	7.36	69.4	108.6
40	1308	109.7	9.3	73.7	124.0
45	1163	117	11.53	77.8	139.5

This calculation indicates the conservatism in the initial assumption that 69% of SO_3 would be converted to SO_2 and O_2 at 900°C. The computer program predicted that almost 78% of SO_3 would be converted at 900°C. To attain 69% of SO_3 conversion, the required tube length is 5 feet 3 inches and the maximum process fluid temperature is 850°C.

The maximum tube inside and outside temperature are 898°C (1648°F) and 902.3°C (1656°F), respectively. The pressure drop in the tube section is 8.5 Psi for the tube length of 5 feet 3 inches. The results suggested that there was some leeway as far as maximum metal temperature and pressure drop are concerned for a more aggressive design, thus a more optimistic design based on higher heat flux is warranted. After analyzing several configurations, an optimistic design configuration was selected as follows: 648 tubes with 2.5 inches inside diameter and 0.1875 inch wall, and 6 feet in tube length. Again the computer program was used to calculate the performance. The results are shown in the following:

Flow rate per tube = 250.3 lb/hr
Heat abosrbed per tube = 81449 Btu/hr
Average heat flux = 20732 Btu/hr-ft^2 (65.3 kW/m^2)
Inlet heat flux = 27180 Btu/hr-ft^2 (85.74 kW/m^2)
Outlet heat flux = 15390 Btu/hr-ft^2 (48.55 kW/m^2)
Maximum process fluid temperature = 846.4°C (1555.5°F)
Maximum tube inside wall temperature = 911.6°C (1673°F)
Maximum tube outside wall temperature = 919.9°C (1688°F)
SO_3 equilibrium conversion = 70.1% mole %
SO_3 actual conversion = 69.1 mole %
Pressure drop = 16.66 Psi

The increase in pumping power due to the higher pressure drop is not significant.

ANALYSES ON PREHEAT SECTION

The approach in analyzing the preheat section is the same as the catalytic section except that it is simpler due to the absence of catalytic reaction. The thermal energy absorbed in the preheat section is 19.438 MW_t. The reference design configuration for the preheat section is 1281 tubes with 2.067 inches inside diameter and 0.154 inch wall and 41.28 inches in length. The performance data were calculated as follows:

Flow rate per tube = 126.6 lb/hr
Heat absorbed per tube = 53657 Btu/hr
Average heat flux = 28827 Btu/hr-ft^2 (90.94 kW/m^2)
Inlet heat flux = 40504.4 Btu/hr-ft^2 (127.78 kW/m^2)
Outlet heat flux = 19644.2 Btu/hr-ft^2 (61.97 kW/m^2)
Maximum tube inside wall temperature = 807.1°C (1484.8°F)
Maximum tube outside wall temperature = 815.9°C (1500.6°F)
Pressure drop = 3.7 Psi

An optimistic configuration for the preheat section was designed to match the optimistic configuration of the catalytic section. The configuration is 648 tubes with 2.5 inches inside diameter and 0.1875 inch wall, and 65 inches in length. The performance data were calculated as follows:

Flow rate per tube = 250.3 lb/hr
Heat absorbed per tube = 103096 Btu/hr
Average heat flux = 29036.5 Btu/hr-ft^2 (91.6 kW/m^2)
Inlet heat flux = 35029 Btu/hr-ft^2 (110.5 kW/m^2)
Outlet heat flux = 22841 Btu/hr-ft^2 (72.05 kW/m^2)
Maximum tube inside wall temperature = 798°C (1468°F)
Maximum tube outside wall temperature = 810°C (1490°F)
Pressure drop = 8.9 Psi

The impact of increased pressure drop on pumping power may be offset by the reduced cost of the equipment.

For a conservative approach, we have not selected the optimistic configuration as the reference configuration due to: a) there is some uncertainity in the rate equation; b) the radial temperature distribution in the catalyst bed is not clear especially in the case of the larger tube diameter.

The efficiency of this solar receiver operated under the given conditions was calculated to be approximately 70%.

4.1.1.1.2 DESIGN AND FABRICATION

The key features of the decomposer include:

- A cylindrical, bottom aperture, cavity type receiver.
- Process tubing arranged along the internal circumference of the receiver to permit uniform distribution of heat flux on all the tubes.
- Separate banks of tubing, connected through manifolds, for preheating and for catalytic conversion of the acid into gases.
- All manifolding located external to the receiver.

These features are discussed below:

Receiver:

The main function of the receiver is to convert the incident solar energy into process heat. Figures 4-5 and 4-6 show details of the receiver construction. The receiver is cylindrical in shape with internal dimensions of 40 feet diameter and 50 feet in height. The roof is conical in shape. The inside of the receiver is insulated using a combination of blanket-type insulation and a refractory lining. The surfaces of the receiver around the aperture will also be insulated to protect the steel structure from any incident solar flux.

The preheater and catalyst tubing penetrate the receiver and are uniformly spaced along the inner circumference. The tubes are connected to manifolds outside the receiver. The penetrations in the external shell of the receiver for the process tubing are sealed to prevent convection heat losses as shown in

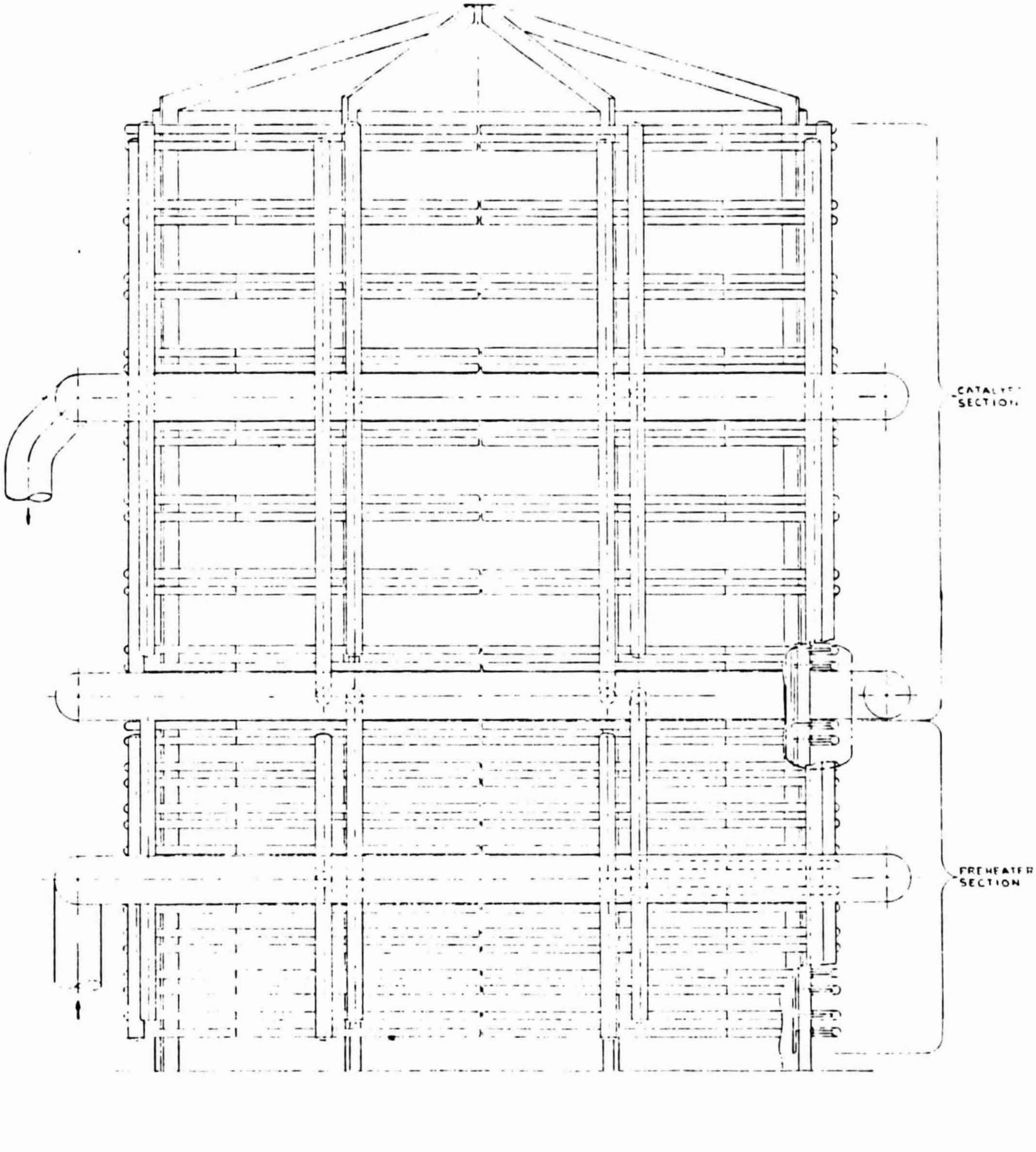

FIGURE 4-5. RECEIVER FOR DECOMPOSER--EXTERNAL VIEW

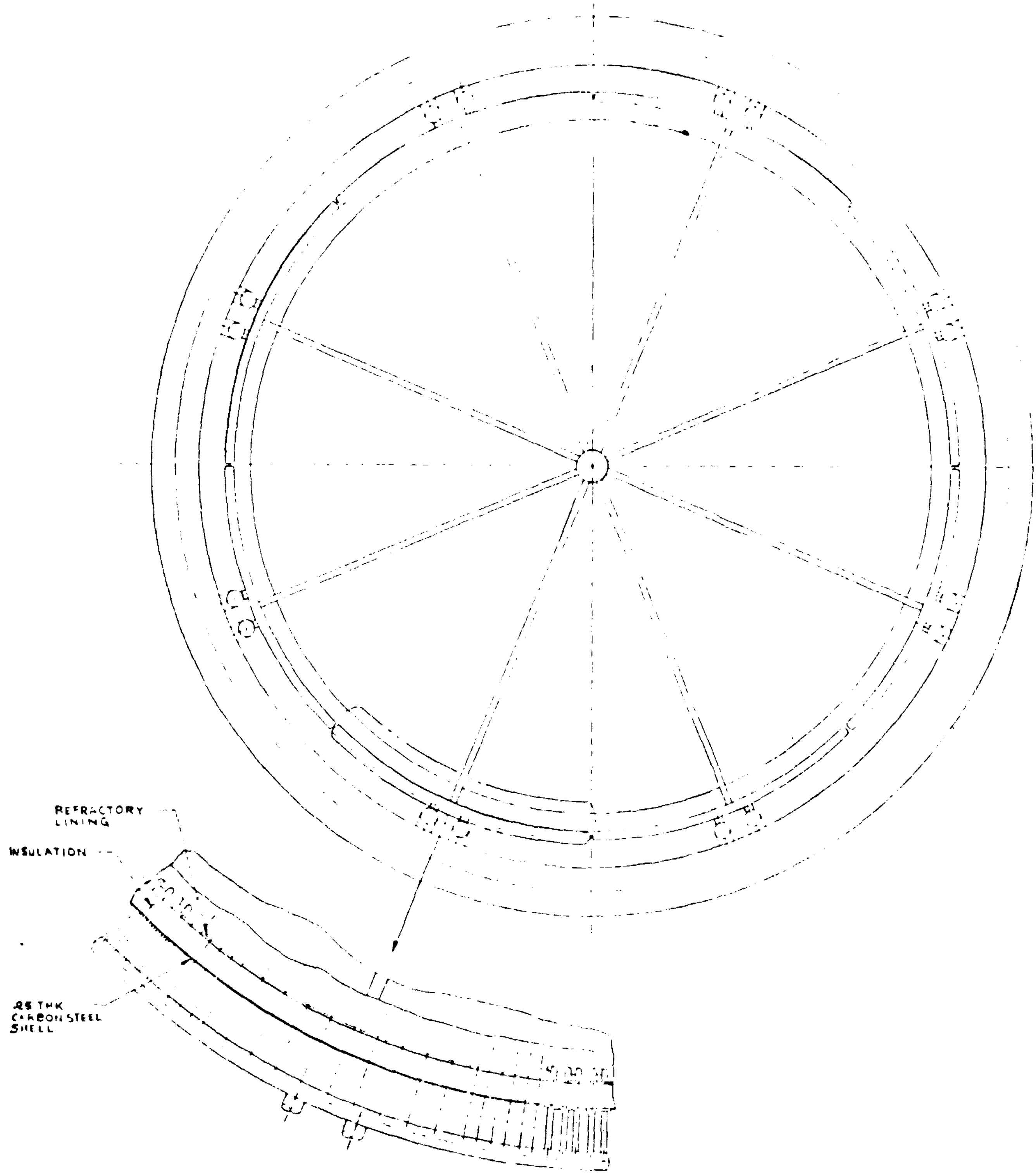

FIGURE 4-6. RECEIVER FOR DECOMPOSER--PLAN AND EXTERNAL VIEW

Figure 4-7. The receiver is designed so that the structural steel framework, the insulation, the process tubing and manifolds can be preassembled in eight panels at the ground level and then installed at the top of the receiver tower.

Process Tubing

The preheater section of the process tubing is arranged in eight circumferential rows in the lower portion of the receiver. Two-inch, Schedule 40, Incoloy 800 H pipe is used for the preheater tubing. Each row has 160 tubes so that a total of 1280 tubes is provided for the preheater section. The external manifolding of these tubes is designed to provide eight vertical banks of preheater tubing, as shown in Figure 4-5. Each preheater tube has a U-shape with unequal legs as shown in Figure 4-8. The total active length of each tube is 3 ft 6 in.

The catalyst section of the process tubing is arranged in eight circumferential rows above the preheater section. The total number of tubes, tube size and tube material are the same as for the preheater section. However, the total active length of each tube is 7 ft.

The catalyst tubing has 3/4 inch nozzles for the filling and removal of the catalyst as shown in Figure 4-8. In addition the top of the longer leg is provided with a perforated barrier to contain the catalyst in the tubing.

Manifolding

All manifolding is located on the outside of the receiver as shown in Figures 4-5 and 4-6 and will be supported off of the structural steel members of the receiver. The vertical manifolds are arranged to provide eight banks each of the preheater and the catalyst tubing. The main riser and downcomer manifolds will be routed through the spider supporting the receiver in order to minimize optical interference.

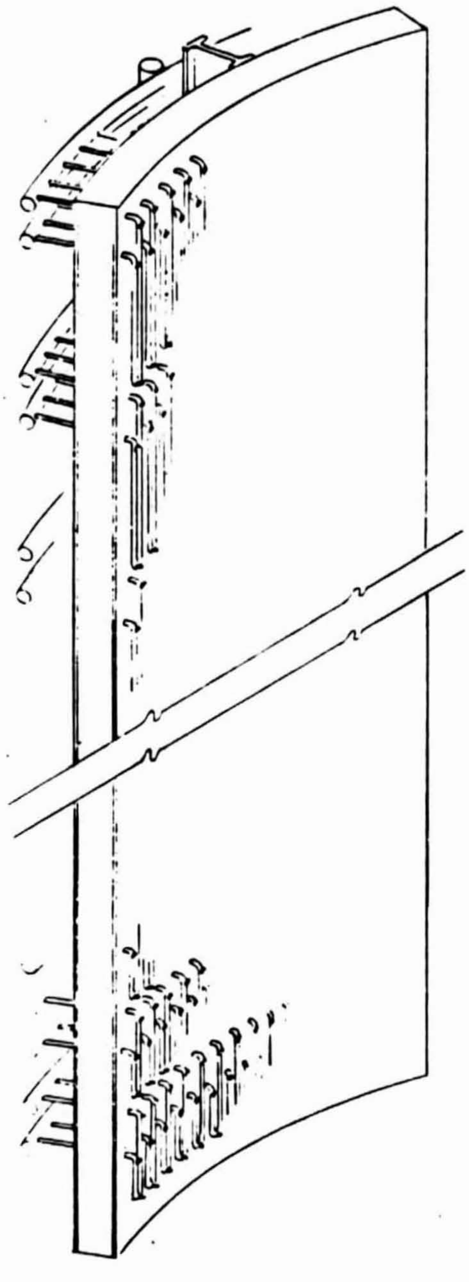

FIGURE 4-7. DECOMPOSER SUB ASSEMBLY

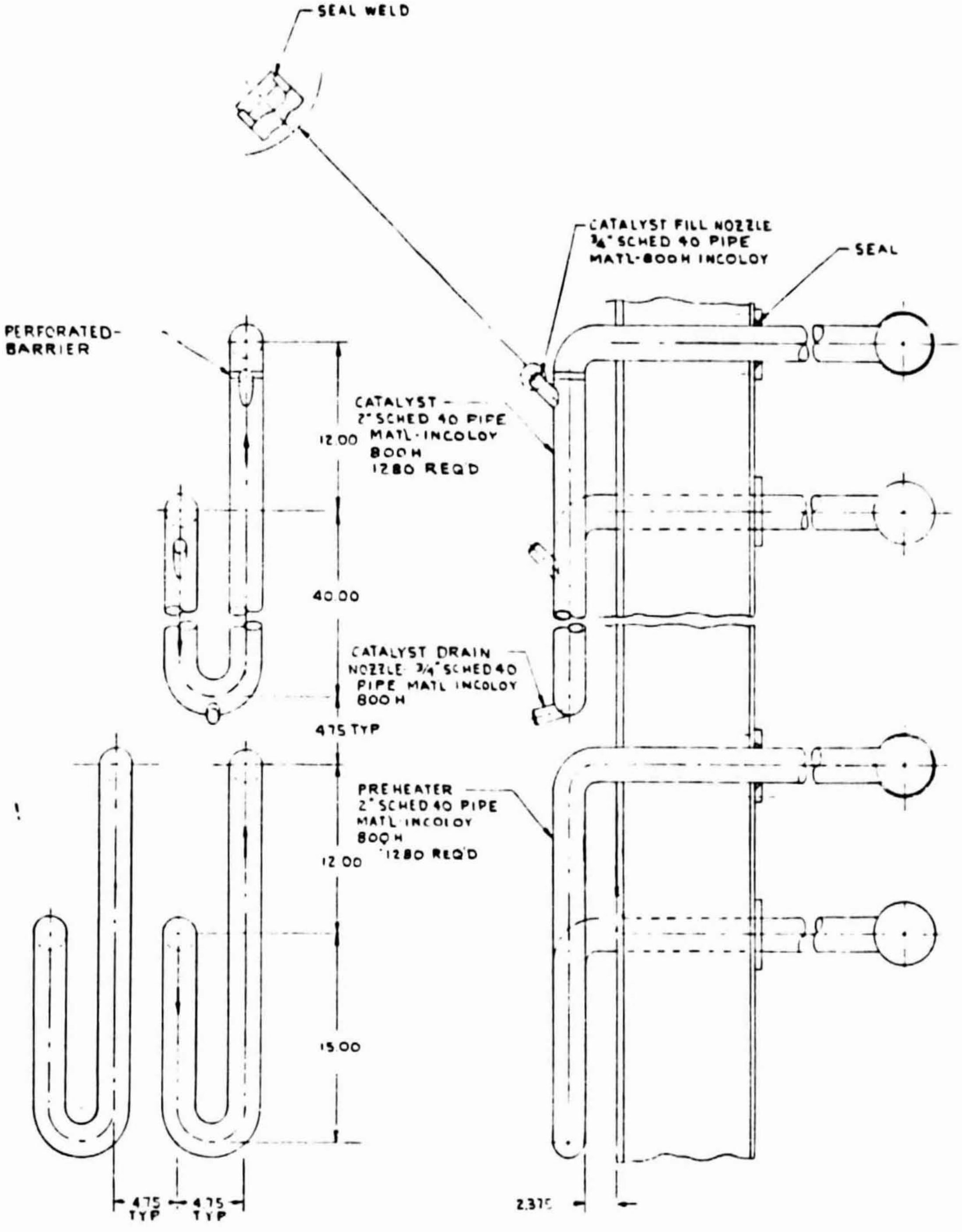

FIGURE 4-8. DECOMPOSER PREHEATER AND CATALYST TUBING

Materials

Incoloy 800 H was identified as the material for the tubes. A potential limitation is the Incoloy 800 H cannot presently be ASME Code certified for temperature above 815°C (1500°F). As alternate material, Hastelloy X, is presently approved by the Code for use at the maximum anticipated service temperatures of the components addressed in this report. Hastelloy X also has the advantage of higher allowable stress at elevated temperatures. Hence the potential use of Hastelloy X should be investigated in the next phase of the design effort. The possibility of obtaining an ASME Code case ruling for the use of Incoloy 800 H at temperature higher than 815°C should also be explored. The fabrication of corrosion resistant alloy such as Incoloy 800 or Hastelloy X is within the state-of-the-art and will be consistent with the ASME Code.

4.1.1.2 RECEIVER/VAPORIZER (SR-2)

4.1.1.2.1 DESIGN ANALYSES

This acid vaporizer requires 32.325 MW_t to change a 50 mole percent of sulfuric acid solution from a liquid state at 498.5 K and 7.5 atmospheres to a condition of saturated vapor at 687.5 K with a total flow rate of 162,216 lb/hr (20.439 kg/sec). The transport properties were established at the inlet, the saturated liquid state, the 65% quality state and the outlet with saturated vapor. Tube wall temperatures were calculated as a function of net heat flux to the tubes for inlet and outlet conditions. The inlet condition thermal analysis is a prediction of tube temperatures from the inlet to the saturated liquid point. Beyond the saturated liquid point and up to qualities of about 70% the film coefficient increases due to boiling such that tube wall temperatures will be less than those predicted between the inlet and the saturated liquid point. Beyond the location of 70% quality, the film coefficient decreases and approaches a value close to the state of saturated vapor at the outlet. This is the location of minimum film coefficient and, therefore, it will locally have the highest tube wall temperatures. Once a permissible net heat flux to the tubes is established, the required tube area is defined by:

$$A_t = Q/q = \pi\ D_o NL$$

where:

Q = thermal requirement (KW)
q = permissible net heat flux to tubes (KW/m^2)
D_o = tube outside diameter (m)
L = tube length (m)
N = number of tubes

It is further assumed that the tubes are spaced one diameter apart or two diameters from center to center so that approximately one-half of the solar flux is reflected on the back side of the tube after passing through the one diameter space between tubes.

The incoming nominal solar flux requirement is then approximately:

$$q_s = \frac{\pi}{2}(q) + q_r + q_a$$

where:

q_r = reradiation from the tubewalls to the atmosphere
q_a = heat flux absorbed by the reflective surface behind the tubes
q = heat flux absorbed by the tubes

Tube Surface Temperature

The tube outside surface temperatures were computed as a function of heat flux absorbed by the tube (q) from the following equation.

$$T_s = q\ [\frac{D_o}{h\ D_i} + \frac{D_o}{2_K} \ln\ (D_o/D_i)] + T_f$$

where:

T_f = fluid temperature (°F)
K = Tube wall conductivity (Btu/hr-ft-°F)

K = (67.48 - .0303 T), 400 < T < 1600°F

D_o = tube outside diameter (ft)

D_i = tube inside diameter (ft)

$h = \frac{k}{D_i} (.023) (Re)^{.8} (Pr)^{.4}$ (Btu/hr-ft^2-°F)

$Re = \frac{4 \dot{m}}{\pi D_i \mu}$

$\dot{m}$ = flowrate in tube (lb/hr) = 161890/N

N = number of tubes

μ = Fluid viscosity (lb/hr-ft)

Pr = Prandl Number

k = Fluid thermal conductivity (Btu/hr-ft-°F)

Figure 4-9 is a plot of the tube outside surface temperature as a function of heat flux absorbed by the tube. Increasing the number of tubes will decrease the flow per tube, decrease the film coefficient, increase the tube wall temperature and decrease the tube length.

Design Selections

A heat flux on the tube surface of 200 kW/m^2 corresponds to an incident solar flux of about 350 kW/m^2 when back side reflector absorption, tube wall reradiation and the $\pi/2$ factor are considered. With this flux as the basis for the selection criteria:

A_t = 32,325 kW/200 kW/m^2 = 161.6 m^2 (1,739 ft^2)

$A_t = \pi D_o NL$

If D_o = 1.0 in and N = 240 tubes are the size and quantity selections, then L = 27.7 feet is the tube length requirement. The fluid states and temperatures are described over this length as follows:

The acid solution remains in liquid phase for about the first 5.5 feet of flow until it becomes saturated liquid. For the next 17 feet of flow the liquid boils to about a 65% state of quality. For the last 5.2 feet, the heat

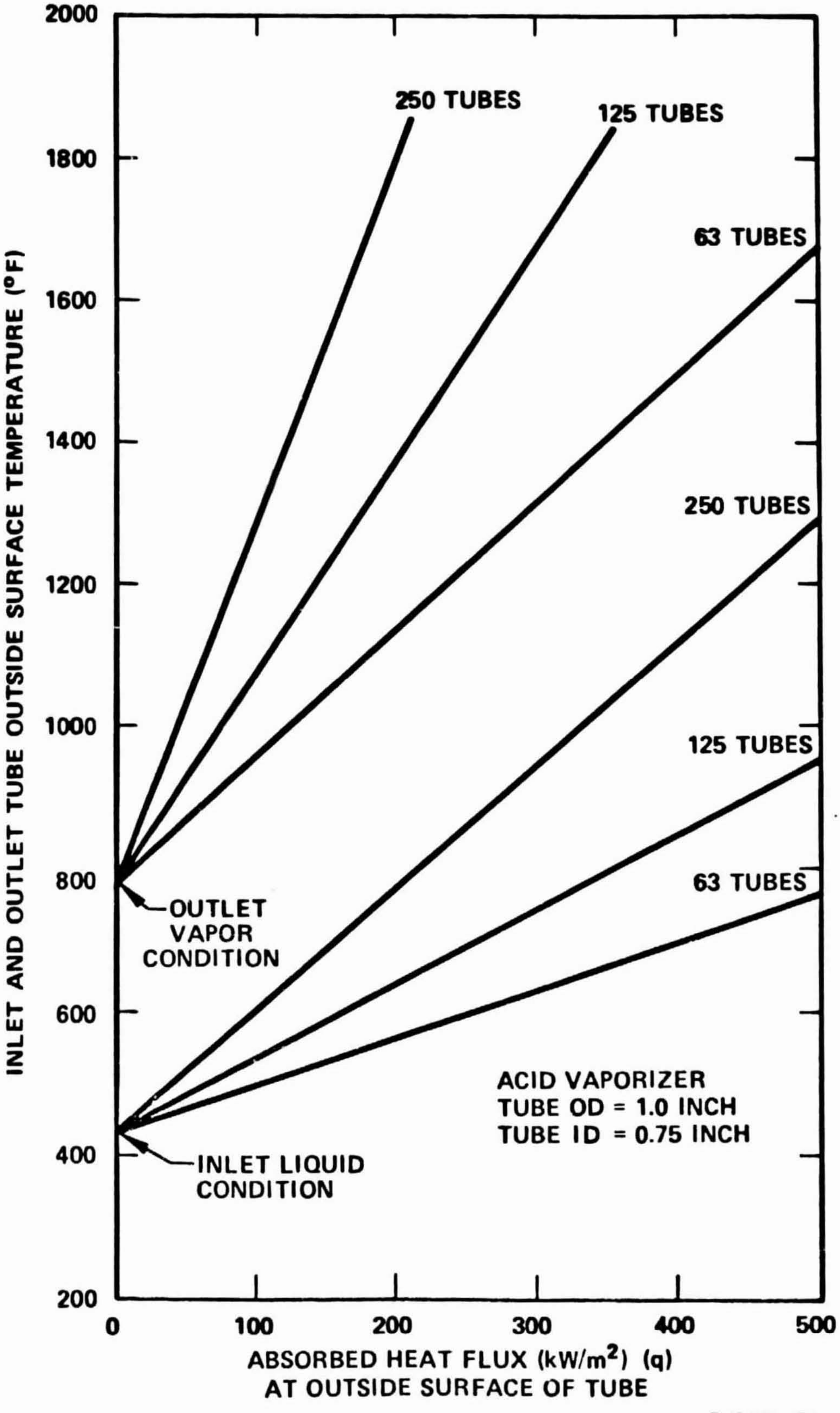

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FIGURE 4-9. TUBE TEMPERATURE VERSUS HEAT FLUX exchanger transforms the liquid from the 65% quality state to the saturated vapor state at the outlet. At the inlet the tube wall temperature is at about 750°F which is about 220°F hotter than the fluid. The film coefficient will not change significantly up to the saturated liquid state (5.5 feet beyond the entrance). There the tube wall temperature will be about 800°F which is about 200°F above the saturated liquid temperature. Over the next 17 feet the boiling will enchance heat transfer, so at the 65% quality state which is at 682.4 K (769°F), the tube wall temperature will be somewhere between 800 and 900°F. Over the last 5.2 feet of the tube, quality will rapidly increase to the saturated vapor state at the outlet, and Figure 4-9 indicates the tube wall temperatures will rapidly approach the 1700°F level over the last couple of feet near the outlet of the heat exchanger. The last 5.2 feet of heat exchanger is the same as that required for the Case 2 heat exchanger where the heating requirement is only 6.076 MW_t.

Tube Presure Drop

There are four fairly well defined state points along the tube. These are at:

X = 0 ft, liquid at 438°F (inlet)
X = 5.53 ft, saturated liquid at 603°F
X = 22.5 ft, 65% quality at 769°F
X = 27.7 ft, saturated vapor at 778°F (outlet)

The pressure drop per unit length of flow at these locations is computed with the following equation.

$$\frac{\Delta P}{L} = \left(\frac{f}{D_i}\right) \frac{\dot{m}^2}{2\ g_c\ \rho\ A^2\ (144)} \quad \text{(psi/ft)}$$

where:

f = friction factor for smooth wall tubing at correct Reynolds number
D_i = tube inside diameter (ft)

g_c = 32.2 (gravitational constant)
$\dot{m}$ = tube flowrate (lb/sec)
ρ = fluid density (lb/ft^3)
A = tube flow area (ft^2)

The state point calculation results are:

Variable/Feet	0	5.53	22.5	27.7
T, °F	438	603	769	778
ρ, lb/ft^3	73.47	58.96	0.387	0.428
μ, lb/hr-ft	1.309	0.931	*	0.2718
N_{Re}	10489	14748	*	50518
f	0.03	0.027	*	0.02
$\Delta P/L$, Psi/ft	0.0026	0.0029	0.16	0.30

*At 22.5 feet, the quality is 65%. A two phase multiplier from Reference 4-6 was used for the calculations.

The internal tube pressure drop for the Case 1 acid vaporizer with the 27.7 ft length is computed by conservatively assuming a linear variation between the above four state points.

$$\Delta P_{tube} = (.00275)(5.53)+.081(17)+.23(5.2) = 2.56 \text{ psi}$$

The tube entry and exist losses are assumed to have coefficients of .5 and 1.0 respectively.

$$\Delta P_e = K_e \frac{\dot{m}^2}{2g_c \, \rho A^2 \, (144)}$$

ΔP_{entry} = .0027 psi
ΔP_{exit} = .95 psi

Combining the internal tube friction losses with the entry and exist losses give the total tube pressure drops as:

$\Delta P = .0027 + 2.56 + .95 = 3.51$ psi

Optimistic Design Selections

The possibility of reducing the vaporizer cost through R&D was considered. The conclusion reached is that opportunity for cost reduction may exist for the Case 1 acid vaporizer, but Case 2 offers little (if any) potential.

The Case 1 acid vaporizer with a 27.7 foot length has much less reradiation as a percentage of its length, because only about the last two feet reach the higher temperatures shown on Figure 4-9 due to the lower convective heat transfer coefficient for the fluid approaching the saturated vapor phase. This design as an absolute upper limit could be locally heated to temperatures based upon acceptable structural limits for the tube material. For the sake of establishing this heat flux limit, it is assumed that the design can be configured so that reradiation from tube walls near the end of the exchanger is for the most part recaptured and not lost to the atmosphere. An extreme upper temperature limit for the Silicon Carbide is assumed here to be 2700°F. With the vapor phase heat transfer coefficient and with the assumption that there is no lost reradiation, a tube heat flux of 390 kW/m^2 will produce a tube wall temperature of 2700°F. With this heat flux, the metal temperature at the state point defined by 65% quality is only about 1050°F which is about the same as that at the inlet. It is only over the last couple of feet of the heat exchanger that the wall temperature will rise above 2000°F. A design change that can take place if the permissible heat flux to a tube is increased to 400 kW/m^2 is a case where only one-half as many tubes would be required. Figure 4-9 shows that the tube wall temperatures could only reach about 1950°F with 120 tubes and a heat flux of 400 kW/m^2. This would double the flowrate per tube which would increase the tube pressure drop by a factor of four. The above three design possibilities for Case 1 are listed for comparison where all are based on a 1.0 inch OD and .125 inch wall thickness.

Tube No.	Tube Length (ft)	Tube Heat Flux (kW/m^2)	Wall Temp. Inlet (°F)	Wall Temp. Outlet (°F)	Tube Pressure Drop (psi)
240	27.7	200	780	1800	3.51
240	14.2	390	1100	2700	2.23
120	27.7	400	850	1950	14.0

Higher average heat fluxes are conceivable with schemes of flux profiling near the exit or partial shading near the exit which go beyond the scope of this effort.

The selected reference design configuration is 240 one inch O.D. silicon carbide tubes with 27.7 feet in tube length. The heat flux is 200 kW/m^2. The configuration for the optimistic design is 120 one inch O.D. silicon carbide tubes and 27.7 feet in tube length, but the heat flux is raised to 400 kW/m^2. The efficiency of the solar receiver operated under the given conditions was estimated to be 80 to 85%.

4.1.1.2.2 DESIGN AND FABRICATION

The key features of the acid vaporizer are:

- A stepped panel, external type receiver
- Silicon carbide process tubing installed in modular arrays
- All manifolding located behind the receiver panel

These features are discussed below:

Receiver

Figures 4-10 through 4-13 provide details of the receiver construction. A stepped panel configuration is used to provide optimum use of the incident solar energy. The stepped configuration also provides a means for support of the process tubing and protects the tube-to-manifold connections from high

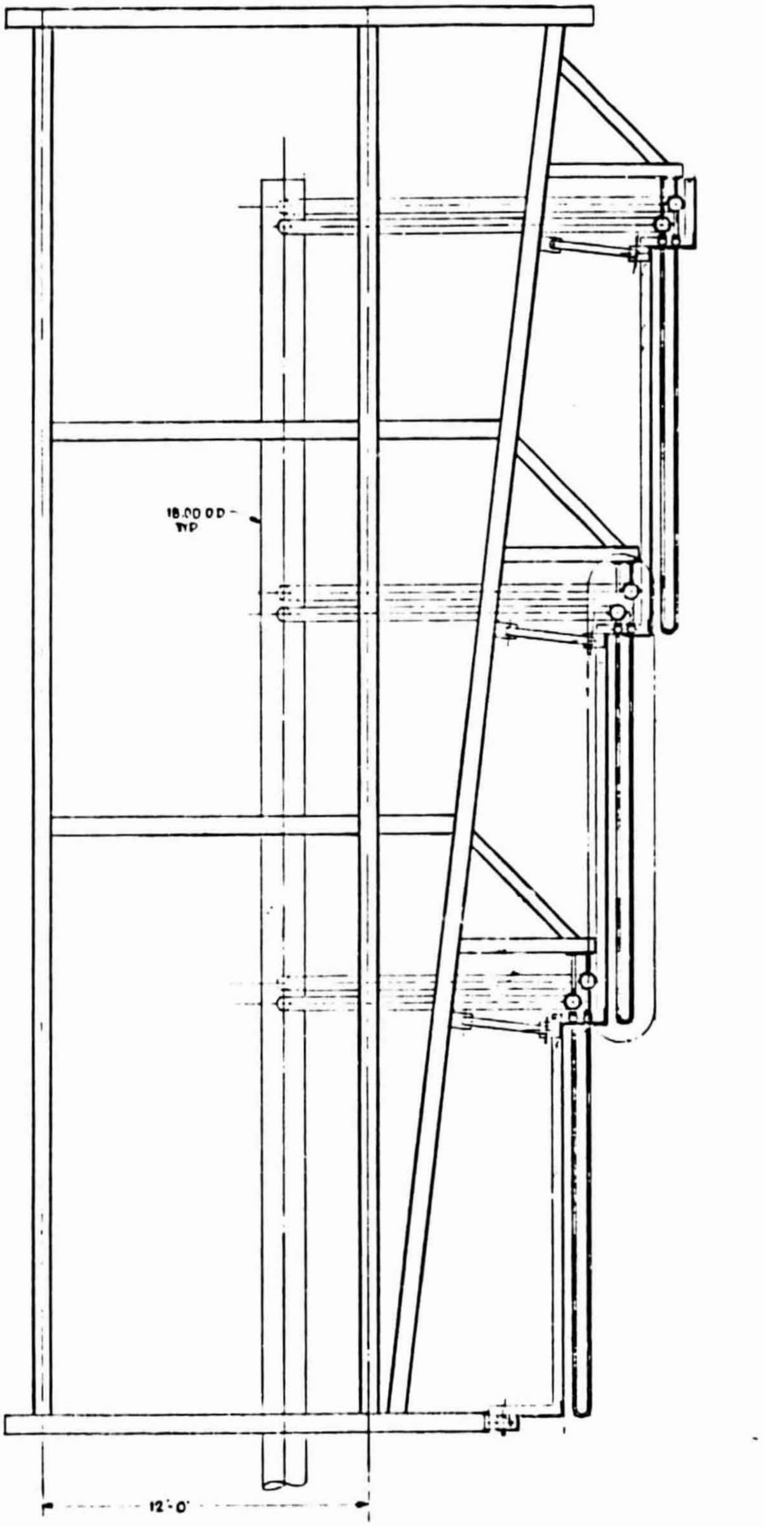

FIGURE 4-10. ACID VAPORIZER--SECTIONAL VIEW

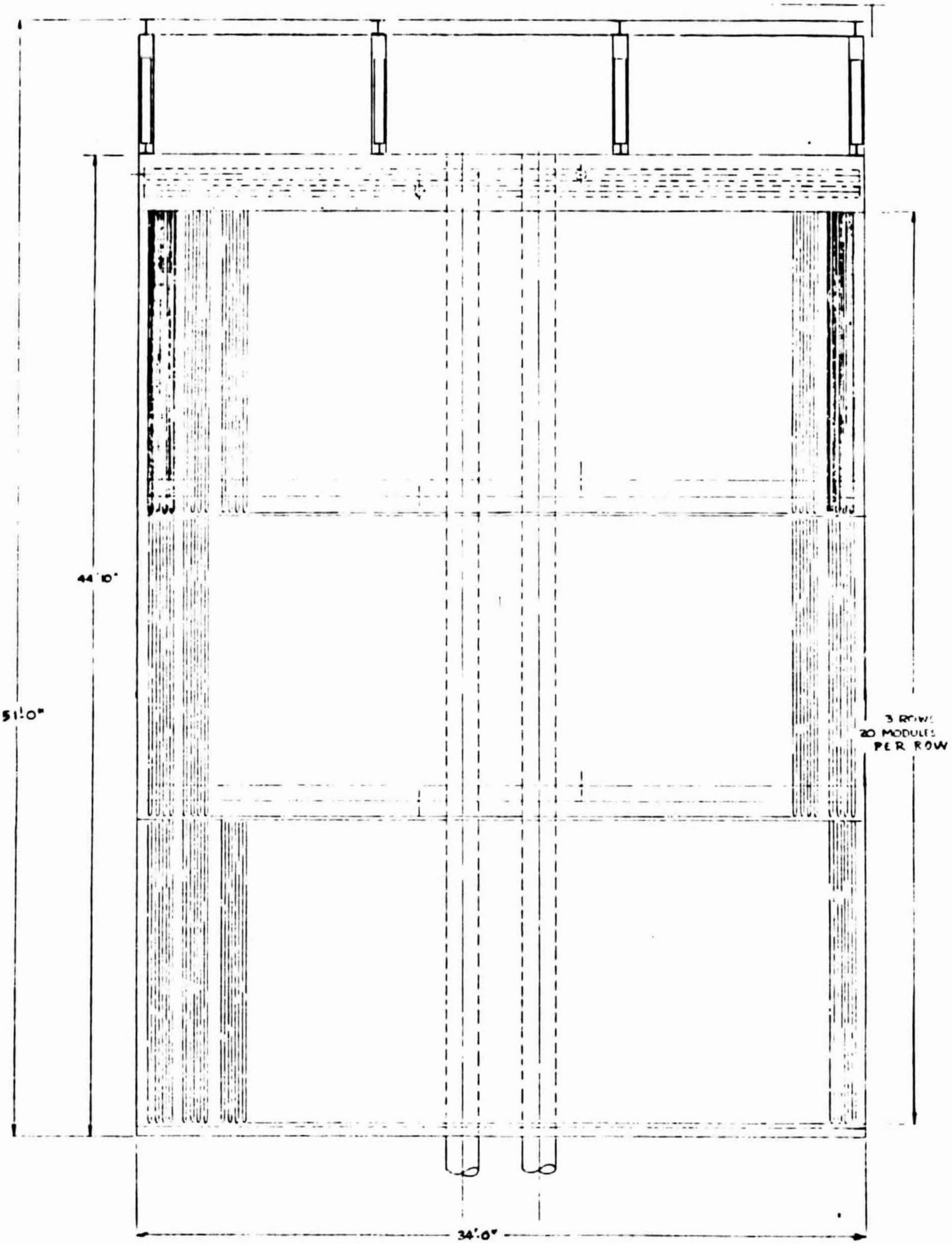

FIGURE 4-11. ACID VAPORIZER--FRONT ELEVATION

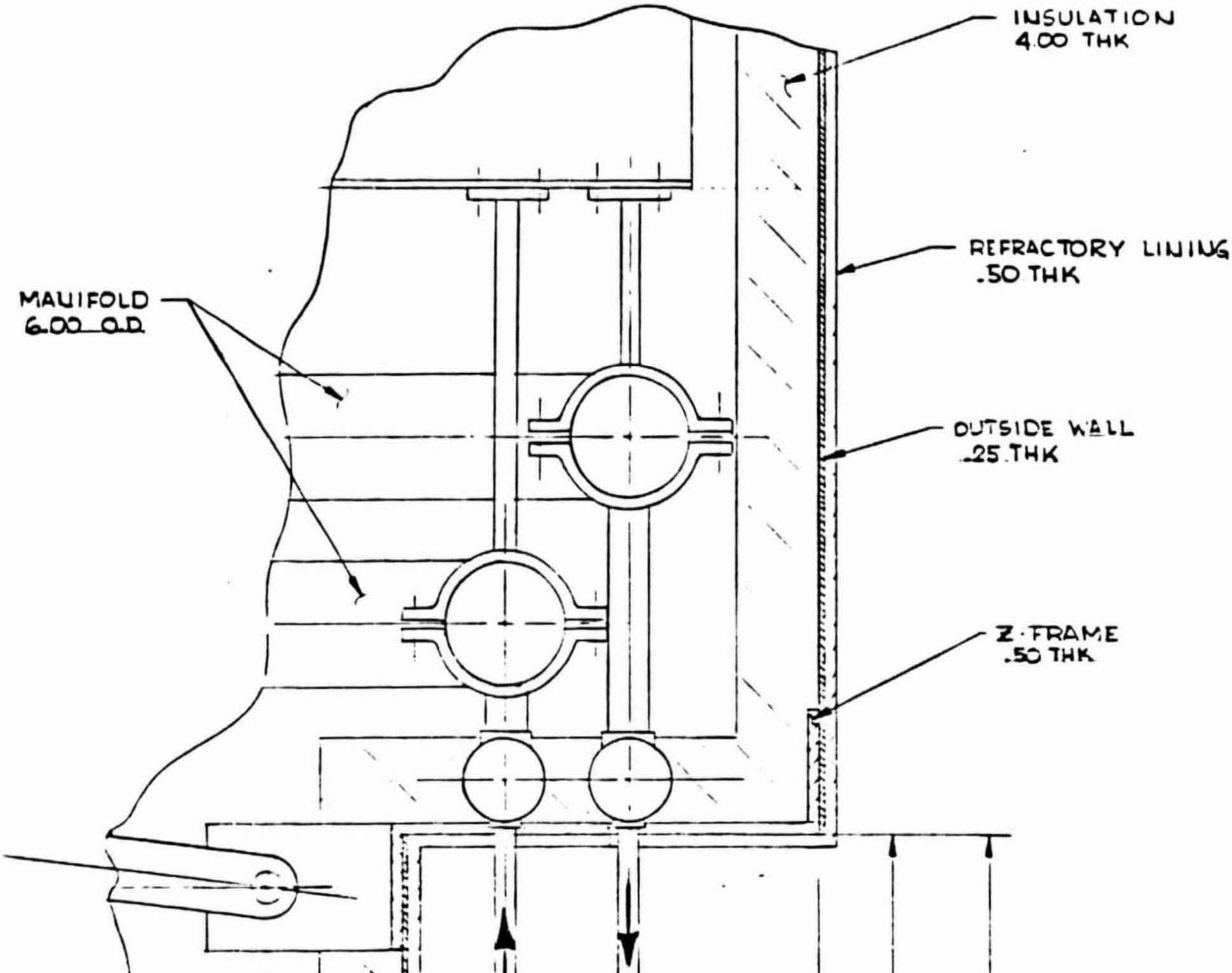

FIGURE 4-12. ACID VAPORIZER--SECTIONAL VIEW

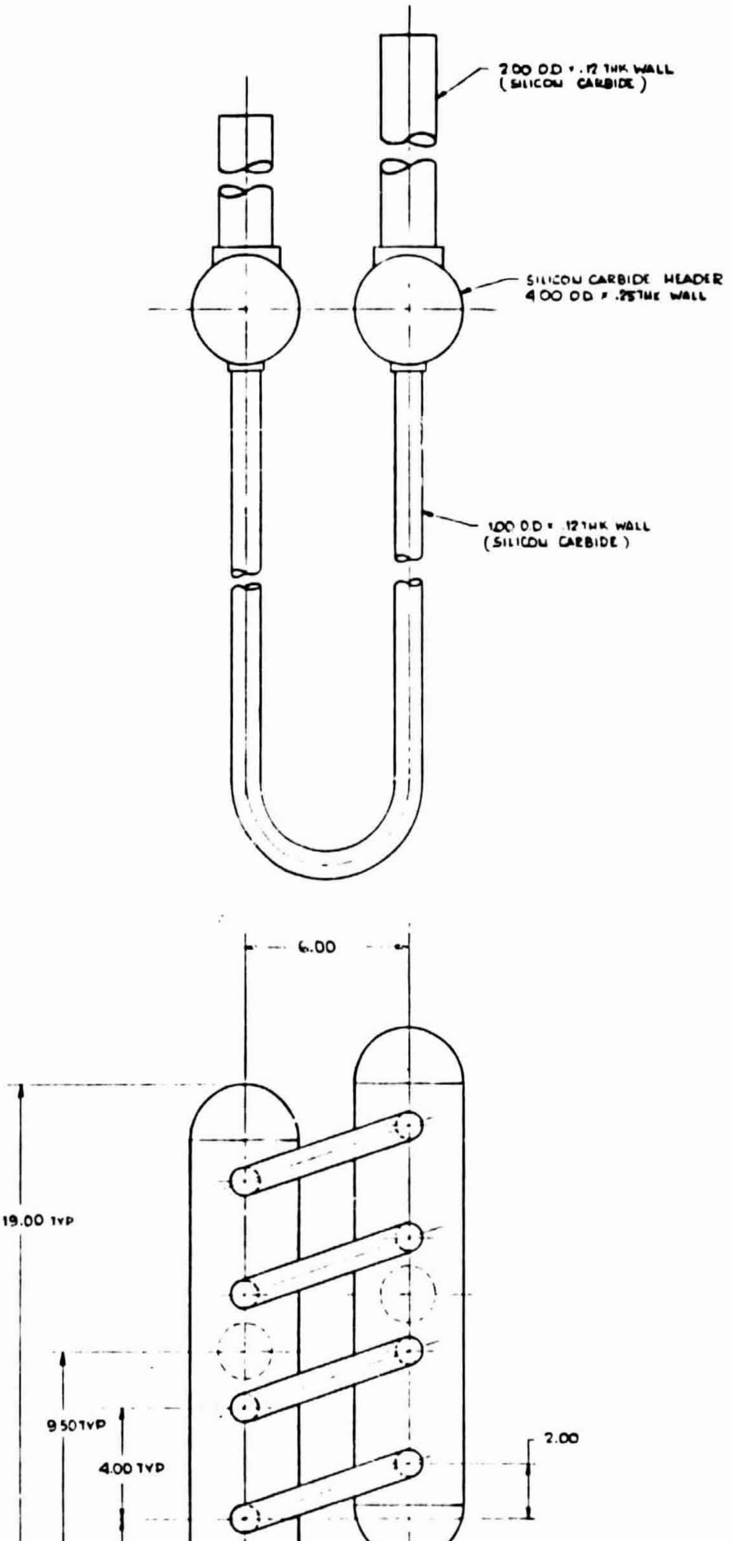

FIGURE 4.13. ACID VAPORIZER--SILICON CARBIDE MODULE

temperatures resulting from direct incident solar flux. The panel, made from carbon steel plate, is supported by a structural steel frame that is attached to the top of the receiver tower. The front face of the panel is protected by a refractory insulation lining. Additional blanket type insulation is provided on the back of the panel. The process tubing is suspended from the stepped sections in front of the panel. The structural framework for the receiver panel also serves to support the manifolding.

Process Tubing

The process tubing is made of sintered alpha silicon carbide. Each tube is 1 inch O.D. x 0.12 inch wall x 28 feet long, bent into a U-shape. Four of these tubes are connected to a pair of silicon carbide headers to form a single module, as shown in Figure 4-13. The inlet and outlet tube stub for the module are also of silicon carbide.

This type of modular construction has the following advantages:

- The all silicon carbide module can be completely fabricated in the shop with resultant assurance of high quality construction.
- Replacements are accomplished easily by removing an entire module.
- U-tube construction permits free thermal expansion of the tubes and minimizes thermal stresses.
- Minimizes silicon carbide-to-metal joints which have to be made in the field.

A total of 60 modules is arranged in 3 rows of 20 modules on the receiver panel. Tube supports may be required to minimize vibration of the long U-tubes. These are not shown in the Figures.

Manifolding

The inlet manifolds to the process tubing will be carbon steel pipe lined with PTFE (Polytetrafluoroethylene). The outlet manifolds will be of Incoloy 800 H. Specially designed seals and clamping devices will be used between the metal manifolds and the silicon carbide header connections. All manifolding will be located behind the receiver panel and will be insulated.

Materials

The heat exchanger section of the acid vaporizer will be made from sintered alpha phase silicon carbide. This grade of silicon carbide is preferred to other grades for the following reasons:

- It has higher flexural strength than two-phase silicon carbide materials. Although it is a brittle material, it contains fewer and smaller flaws; evidence of this is the observation that hot isostatic pressing does not improve its mechanical properties.
- Below 1500°C, sub-critical flaws do not grow to become critical flaws since the fracture strength has been observed to be independent of loading duration.
- Proof testing at room temperature to develop stresses slightly (about 10 percent) greater than those to be encountered in service will eliminate by destruction those components containing critical flaws. A small margin in stress is satisfactory because the strength of the material increases at higher temperatures.
- Complex assemblies can be built up from simple pre-sintered shapes by joining them together with "green" pressed components and then sintering the assembly. The bond line between components joined in this manner cannot be detected metallographically.

Alpha phase silicon carbide tubing and manifolds are used in the areas of components that will be exposed to hot (over 260°C) concentrated sulpheric acid vapor, in order to assure chemical stability. Any metal components that will be required in these areas, such as gaskets and seals, will be gold-plated for protection against the acid.

FABRICATION METHODS FOR THE SILICON CARBIDE COMPONENTS

The fabrication of silicon carbide components is currently limited to relatively simple shapes and small sizes. The silicon carbide components that are simple in shape can be formed by:

- Uniaxial dry pressing, at about 20,000 psi
- Isopressing, at about 20,000 psi isostatic pressure
- Extrusion
- Slip casting, in a plastic mold.

The slip casting method also allows the production of fairly complex, relatively thin-walled shapes in large quantities. It is a less expensive process than precision casting and is suitable for large shapes. However, the slip-casting process is not as precise as precision casting.

More complex shapes can be formed by:

- Mandrel coating, the reverse of slip casting, on a porous mandrel.
- Thixotropic casting, into a rubber mold which is pulled away after the thixotrope has gelled.
- Plastic forming, using a resin vehicle and conventional plastic molding techniques.

The shapes formed by any of the above techniques can be machined in the "green" state to incorporate features or precise dimensions prior to sintering.

Several techniques are currently available for shop joining of silicon carbide to silicon carbide. "Green" compacts can be joined together by using silicon carbide slurry or a proprietary cement (General Motors Corp.). After sintering such a joint in compression, the joint is not detectable. Another method of assembly, possible when the components are coaxial at the joint, is by heat shrinking. The inner component is formed and sintered. The outer component is formed and fitted loosely over the inner component. The assembly is then sintered, with the shrinkage of the "green" component providing a clamping force across the bond. Joints made in this manner cannot be detected metallographically and the second sintering operation has no deleterious effect on the previously sintered component. Either or both of these techniques could be used for the assembly of the silicon carbide modules and for the silicon carbide tube to tube sheet joints.

Shop assembly of silicon carbide to metal joints can be performed using the heat shrink method. In this case, the method of assembly depends on the difference in the coefficient of thermal expansion between metal and silicon carbide. As silicon carbide has a coefficient of thermal expansion that is one-third that of steel, it can be heat shrunk into a metal sleeve. The major disadvantage of heat shrinking is the cost of holding tight tolerances on both the metal and silicon carbide at the interface (i.e. to control the interference stresses). However, these costs can be significantly reduced by dimensional matching in production.

The method of making silicon carbide to metal joints that is presently preferred is by using mechanical seals that are clamped between the mating parts. The sealing technique is to clamp a gasket ring between two flat flanges, one metal and the other ceramic. The gasket is a metal spring-energized ring consisting of a garter spring wound with initial tension and a toroidal wrapper sheet of U-shaped cross section, the opening of the U being on the outer circumference and, therefore, not exposed to corrosive products. The spring material will be Incoloy X750, with a gold plated liner. The location of the gasket and the external metal clamping devices will be chosen to minimize bending stresses. The metal spring-energized ring configuration allows the U-shaped liner to bend to permit differential radial displacement of the sealing lines and the garter spring rolls about its toroidal axis to permit the same motion. The metal spring-energized ring seal has been used extensively in nuclear applications abroad.

4.1.1.3 ACID CONCENTRATOR

4.1.1.3.1 DESIGN ANALYSES

The acid concentrator was analyzed as a counter flow heat exchanger. The heat source is from a solar thermal storage device with a heat transport fluid assumed to be molten salt consisting of 60% sodium nitrate and 40% potassium nitrate by weight. The properties of the heat transport fluid are listed in Table 4-2 (Reference 4-5).

TABLE 4-2. HEAT TRANSPORT FLUID (MOLTEN SALT) PROPERTIES

T,		ρ,		c_p,		$\mu \times 10^3$,		K,			$\beta \times 10^4$	
°C	(°F)	$\frac{kg}{m^3}$	(lb/ft³)	$\frac{J}{kg\text{-}°C}$	$(\frac{Btu}{lb\text{-}°F})$	Pa-s	$(\frac{lb}{ft\text{-}s})$	$\frac{W}{m\text{-}k}$	$(\frac{Btu}{h\text{-}ft\text{-}°F})$	Pr	$\frac{1}{°C}$	$(\frac{1}{°F})$
Liquid												
260	(500)	1928.6	(120.4)	1553.3	(0.371)	4.00	(2.69)	0.398	(0.23)	15.62	3.4	(1.89)
316	(600)	1888.6	(117.9)	1553.3	(0.371)	2.80	(1.88)	0.398	(0.23)	10.92	3.4	(1.91)
371	(700)	1848.5	(115.4)	1553.3	(0.371)	2.10	(1.41)	0.398	(0.23)	8.01	3.5	(1.96)
427	(800)	1819.7	(113.6)	1553.3	(0.371)	1.63	(1.10)	0.398	(0.23)	6.45	3.6	(2.00)
482	(900)	1789.3	(111.7)	1553.3	(0.371)	1.30	(0.880)	0.398	(0.23)	5.66	3.7	(2.04)
538	(1000)	1741.2	(108.7)	1553.3	(0.371)	1.07	(0.721)	0.398	(0.23)	3.91	3.7	(2.08)
Solid												
37	(100)	1922.2	(120)	1553.3	(0.371)			0.363	(0.21)			
93	(200)	1922.2	(120)	1553.3	(0.371)			0.363	(0.21)			

For a counter flow heat exchange analysis the log mean temperature approach was used. The overall heat transfer results in air expression are as follows:

$$UA = \frac{q}{\text{log mean } \Delta T}$$

where

$$\text{log mean } \Delta T = \frac{\Delta T_1 - \Delta T_2}{\ln \Delta T_1/\Delta T_2}$$

ΔT_1 = entrance ΔT between two fluid streams.
ΔT_2 = exit ΔT between two fluid streams.
q = heat source total requirements.
U = overall heat transfer coefficient.
A = required heat transfer area.

The overall heat transfer coefficient was based on the inner diameter of the tube so that;

$$U_i = \frac{1}{\frac{1}{h_i} + \frac{r_i}{K \ln r_o/r_i} + \frac{1}{r_o/r_i \; h_o}}$$

where;
h_i = tube side heat transfer coefficient.
h_o = shell side heat transfer coefficient.
$r_i/K \ln r_o/r_i$ = resistance across the tube wall.

In general, the heat transfer coefficients were based on the Dittus-Boelter relationship:

$$h = 0.023 \frac{K}{D} (Re)^{0.8} (Pr)^{0.4}$$

For the boiling regime the heat transfer was evaluated by using a boiling correlation (Reference 4-6):

$$\frac{h}{h_{1o}} = 3.5 \left(\frac{1}{X_{tt}}\right)^{0.5}$$

where:

$$h_{1o} = 0.023 \frac{K_1}{D} \left(\frac{GD}{\mu_1}\right)^{0.8} (Pr_1)^{0.4}$$

$$\frac{1}{X_{tt}} = \left(\frac{x}{1-x}\right)^{0.9} \left(\frac{V_v}{V_1}\right)^{0.5} \left(\frac{\mu_v}{\mu_1}\right)^{0.1}$$

x = quality

V = specific volume

μ = viscosity

v = vapor

1 = liquid

The pressure drop across the tubes included the entrance, friction (both single and two phases) and exit two phase losses in a tube. The shell side included the same losses along with baffle pressure losses. For two phase analysis a Martinelli approach (Reference 4-6) was used based on the fluid quality in the tube.

TWO PHASE PRESSURE DROP

A Martinelli approach was used to estimate the two phase pressure losses inside of the one inch outside diameter silicon carbide tubes. The approach gave two phase multipliers Φ^2_{Lo} of 225 for the acid concentrator and 100 for the acid vaporizer, respectively. These were used in estimating the overall pressure losses. The Moody friction losses were multiplied by the Φ^2_{Lo} factors to account for boiling in tubes.

DESIGN SELECTIONS

Table 4-3 shows the parametric results as a function of number of tubes for the acid concentrator. From the results in Table 4-3, a preferred design configuration was selected and shown in Figure 4-14. The acid concentrator consists of 300 one-inch tubes. Each tube is 34.2 feet in length, constructed in U-shape. The diameter of the shell is 3 feet.

4.1.1.3.2 DESIGN AND FABRICATION

The acid concentrator consists of 300 U-shape silicon carbide tubes. Each tube is 1.0 inch O.D. with 0.125 inch wall and 35 feet in total length. The tubes are arranged in a triangular pitch array with a pitch to diameter ratio of 1.4. The shell is 3 feet in diameter. The material of the tube sheet is silicon carbide. The material of the shell and the upper head is an alloy steel that contains 2-1/4% chromium and 1% molybdenum. However, the lower head should be fabricated with either glasteel (fusing glass to steel) or carbon steel with PTFE (polytetrafluoroethylene) molded lining because it would contact the two phase acid at 225°C. The fabrication of the silicon carbide parts is similar to the receiver/vaporizer (SR-2) discussed in the Section 4.1.1.2.2. The fabrication of the metal parts should comply with the requirements of the ASME Code.

4.1.1.4 HEAT EXCHANGER-1 (HX-1)

The HX-1 is an assumed piece of equipment to complete the process cycle. It brings down the decomposed high temperature gaseous stream from 900°C to 100°C. In this cool down process, the undecomposed SO_3 is recombined with the water vapor first and then condensed to form 49.7 weight percent sulfuric acid solution. The SO_2 and O_2 still remain in a gas phase. The energy removed in the HX-1 is 44.72 MW_t, which can be utilized as process heat for the balance of the process that is not within the scope of this study.

A counter current flow shell and tube heat exchanger was selected with the hot acid in the tube side and the cooling water in the shell side. The flow rate of the cooling water is 1,387,133 lb/hr (174.78 kg/sec). The overall heat

TABLE 4-3

SUMMARY OF THE ANALYSES FOR THE ACID CONCENTRATOR
(SIZE OF TUBE: 1.0" O.D., 0.75" I.D.; P/D = 1.4)

No. Tubes	Length L, ft	Mean Dia. D, ft	Pressure Drop ΔP, psi	*Overall Heat Trans. Coefficient U, Btu/hr-ft^2-°F
50	22.2	1.2	200	172/308
100	22.8	1.68	49.2	102/191
200	18.9	2.42	12.4	60/115
300	17.1	2.97	5.0	44/85
500	15.2	3.82	2.0	29/57
1000	13.1	4.64	.5	17/33

*Two values for each U. One is for the liquid region and the other is for the two phase region.

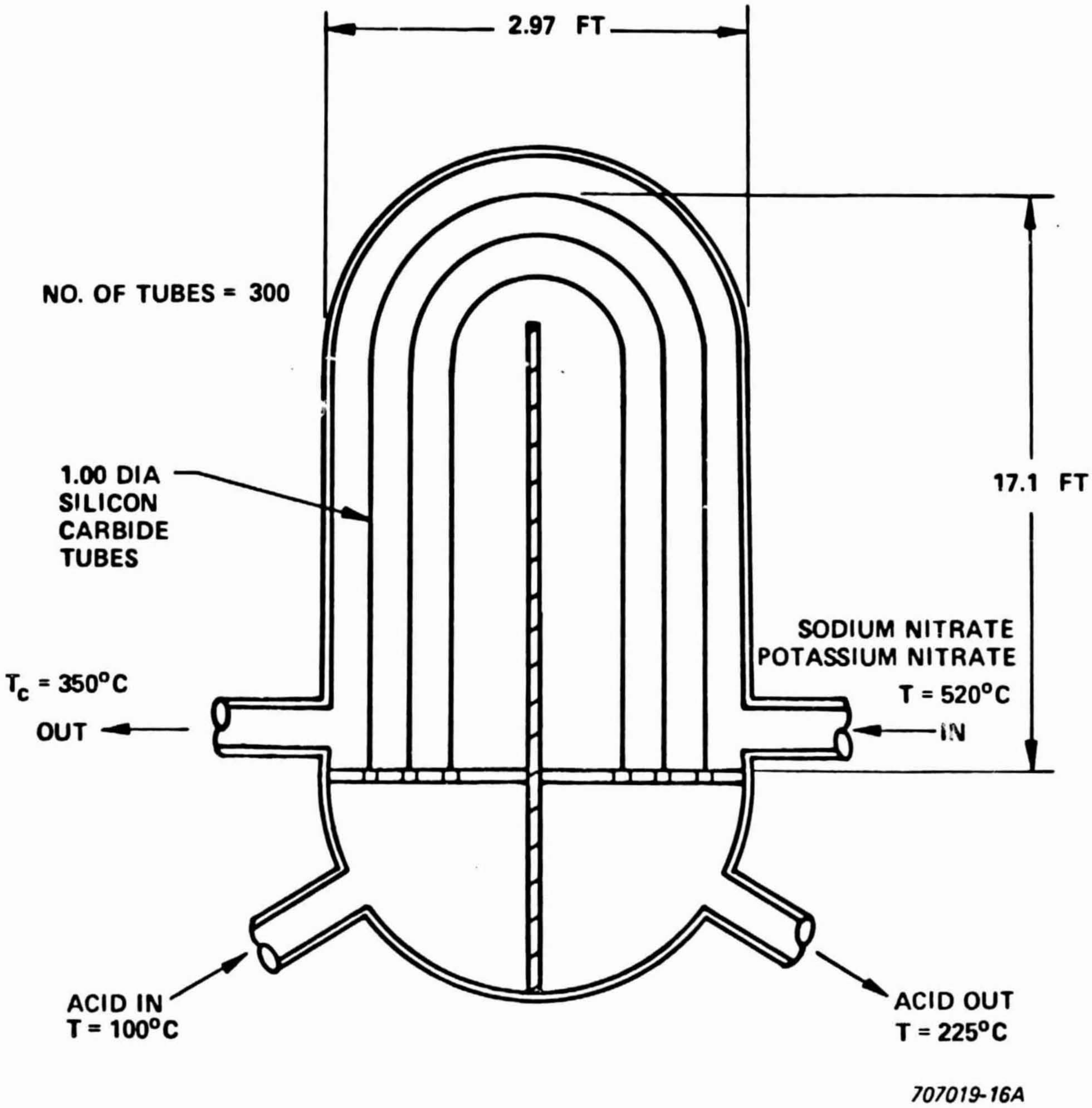

FIGURE 4-14. ACID CONCENTRATOR–SOLAR STORAGE FLUID

transfer coefficient (U_0) was calculated to be 200 Btu/ft^2-Hr-°F with a log-mean temperature drop of 537°F.

The selected design is 678 straight silicon carbide tubes with 1.0 inch O.D. and 0.75 inch I.D. Each tube is 8 feet in length. The tubes are arranged in a triangle pitch array with a center to center distance of 1.36 inches. The shell is 3.1 feet in diameter with 0.25 inch in wall thickness. Due to the high temperature corrosive fluid in the tube side, silicon carbide was selected as the tube material, but the material of the shell can be carbon steel. The material for the hot end tube-sheet and head is Incoloy 800, and for the cold end tube-sheet and head is Hastelloy B. The critical component which is the key to proper functioning of the reactor is the radial Helicoflex seals. Helicoflex seals should be used to seal between the silicon carbide tubes and the metal tube-sheets to handle the difference in thermal expansion between silicon carbide and metal. Helicoflex seals are all-metal flexible seals consisting of a toroidal coil spring core, closely wound with initial tension, surrounded by a seal member ("liner") of softer, plastically deformable metal; a gap between the edges of the liner may be positioned anywhere around the minor circumference of the torus, or the edges may be overlapped to isolate the spring completely from the process fluid. The materials chosen for these seals is Inconel X-750 springs, for resistance to creep at high temperature and stress, with gold liners, for corrosion resistance in the process fluid. The tube-to-tube sheet seals are installed with radial deformation (caused by pressing the tapered ends of the tubes axially into the annular seals), with the gaps in the liners open into the acid stream. Installation of Helicoflex seals in this manner is approved by the manufacturer and has resulted in successful application. Some axial sliding of the tube or mandrel in the liner is necessary at initial installation, and additional sliding probably will occur in service because of pressure and temperature excursions. These seals are expected to be used in other components for sealing between silicon carbide tube and tube-sheet. The fabrication of the metal parts should be performed to the requirements of the ASME Code.

4.1.1.5 HEAT EXCHANGER-2 (HX-2)

The HX-2 is to condense the vapor leaving the acid concentrator to 100°C. The composition of the vapor is steam carring 1.8% sulfuric acid by weight. The heat load is 8.657 MW_t (29,537,684 Btu/hr). A shell and tube heat exchanger was selected for the design with condensing vapor in the tube side and cooling water in the shell side. Tantalum was selected for the tube material due to the corrosive nature of the fluid. The material for the shell is carbon steel. The tube side flow rate is 9916.84 lb/hr (1.249 kg/sec). The cooling water flow rate is 738,442 lb/hr (93 kg/sec). The calculated overall heat transfer coefficient (Uo) was 350 Btu/ft^2-hr-°F, and the log-mean temperature drop was 201°F. The selected design configuration is 230 tubes with 1.0 inch O.D. and 0.75 inch I.D. Each tube is 7 feet in length. The tubes are arranged in a triangle pitch array with a center to center distance of 1.36 inches. The shell is 1.8 feet in diameter with 0.125 inch wall thickness. The hot end head should be constructed with tantalum or carbon steel with PTFE molded lining, and the cold end head could be constructed with Hastelloy B.

4.1.1.6 ACID ACCUMULATOR

Since the acid vaporization/decomposition system is operated during the daytime only, the concentrated acid released from the acid concentrator that is operated continuously must be put into storage during the night. The acid accumulator serves as a storage tank. The required minimum capacity of the tank is 10,674 cubic feet (79,840 gallons). For practical purpose, the tank was sized at 20% over the minimum capacity. To store the corrosive high temperature acid, extensive efforts were made to find a proper material for the tank, and several tank suppliers were consulted. A tank made with Glasteel was slected. Fusing glass to steel produces a composite material with good structural strength. Moreover at the given acid conditions, Glasteel is essentially inert. Since the capacity of the largest tank available is 24,000 gallons, four tanks are required. These four tanks should be installed in series so that only one acid pump would be required.

4.1.1.7 LOW TEMPERATURE ACID PUMP (P-1)

This pump was sized to deliver 86,453 lb/hr (10.85 kg/sec) of sulfuric acid at 100°C and 10 psig. Due to the corrosive nature of sulfuric acid at this temperature high silicon iron, Durichlor 51, which contains 14.5% silicon and 2% chromium was selected as the material of the pump.

4.1.1.8 HIGH TEMPERATURE ACID PUMP (P-2)

This pump is to receive the sulfuric acid at 225°C and 1 atmosphere pressure and to deliver it to the top of the solar tower at 7.5 atmosphere. The required discharge head of the pump could be up to 300 Psig. Although Durichlor 51 can resist corrosion of sulfuric acid at this temperature level, it cannot withstand the demanded pressure. Therefore, the casing of the pump should be fabricated with metal with teflon or silicon carbide molded lining. It is, therefore, concluded that the pump is not commercially available at the present time and that will require some development.

4.1.2 EQUIPMENT COST ESTIMATES

The cost estimates were based on one of the following sources or the combination of them: a) vendor's quote; b) results from past published studies; c) equipment cost estimating method from literature; d) Westinghouse own experiences; e) discussion with vendors. The costs were all normalized to mid-1983 dollars for consistency. The equipment costs for the Case 1 are summarized in the following Table:

Equipment	Cost (Mid-1983 $)
Acid Concentrator	889,300
Acid Accumulator	364,000
Acid Vaporizer (SR-2)	935,900
Decomposer (SR-1)	1,909,100
Heat Exchanger (HX-1)	785,000
Heat Exchanger (HX-2)	93,800
Cold Acid Pump (P-1)	6,500
Hot Acid Pump (P-2)	15,000
Pump for HX-1 Cooling Water	2,500
Pump for HX-2 Cooling Water	1,400
Total	5,002,500

The value of the total cost is for the process equipment only. It does not include the piping because the plant layout is not within the scope of the study. Also, the cost of each solar receiver does not include the cost of the foundation and the tower itself.

For the optimistic design, the acid vaporizer (SR-2) cost would be $501,500 in comparison with $935,900 for the reference design, and the decomposer cost would be $1,234,200 compared to $1,909,100. Thus, the total major equipment cost might be reduced to $3,893,200 instead of $5,002,500.

4.2 CASE #2

4.2.1 EQUIPMENT DESIGN

As most of the process equipment in Case 1 and Case 2 are identical, only the equipment which are different from Case 1 will be described in this section. Those equipment are: a) acid vaporizer-1 (AV-1); b) Receiver/vaporizer (SR-2); c) Heat exchanger-1 (HX-1).

4.2.1.1 ACID VAPORIZER-1 (AV-1)

The AV-1 was analyzed as a counter flow heat exchanger. The thermal energy required in the AV-1, 26.25 MW, is recuperated from the decomposed high temperature stream leaving SR-1. In the analyses the total number of tubes was varied and the required tube lengths were calculated. The analysis included the various heat transfer regimes of heating, boiling and film boiling both inside and outside of the one inch diameter silicon carbide tubes. The pressure drop across the tube lengths considered both single and two phase flow. The stream leaves the AV-1 as a two phase mixture with 65% quality.

The method of analysis is identical to Section 4.1.1.3.1, and the results of the study are summarized in Table 4-4. The design configuration was selected as follows: 1000 tubes, 1.0 inch O.D., 0.75 inch I.D. Each tube is 40 feet in length and is constructed in U-Shape. The tubes are arranged in a triangular pitch with a pitch to diameter ratio of 1.4. The material for both the shell and tube must be silicon carbide due to the hostile environment. The configuration is illustrated in Figure 4-15. The total pressure drop is 3.2 Psi.

TABLE 4-4. SUMMARY OF THE ANALYSES FOR AV-1
(Size of Tube = 1.0" O.D., 0.75" I.D., P/D = 1.4)

No. Tubes	L, ft	Dm, ft	ΔP, psi**	U, NB/B*
500	25.5	3.82	12.8	51/96
1000	19.9	5.42	3.2	30/56
2000	17.2	7.66	0.84	17/33
3000	15.8	9.38	0.38	12/24

*NB = Nonboiling
B = Boiling
D_m = Diameter of Shell
**Pressure Drop Inside Silicon Carbide Tubes

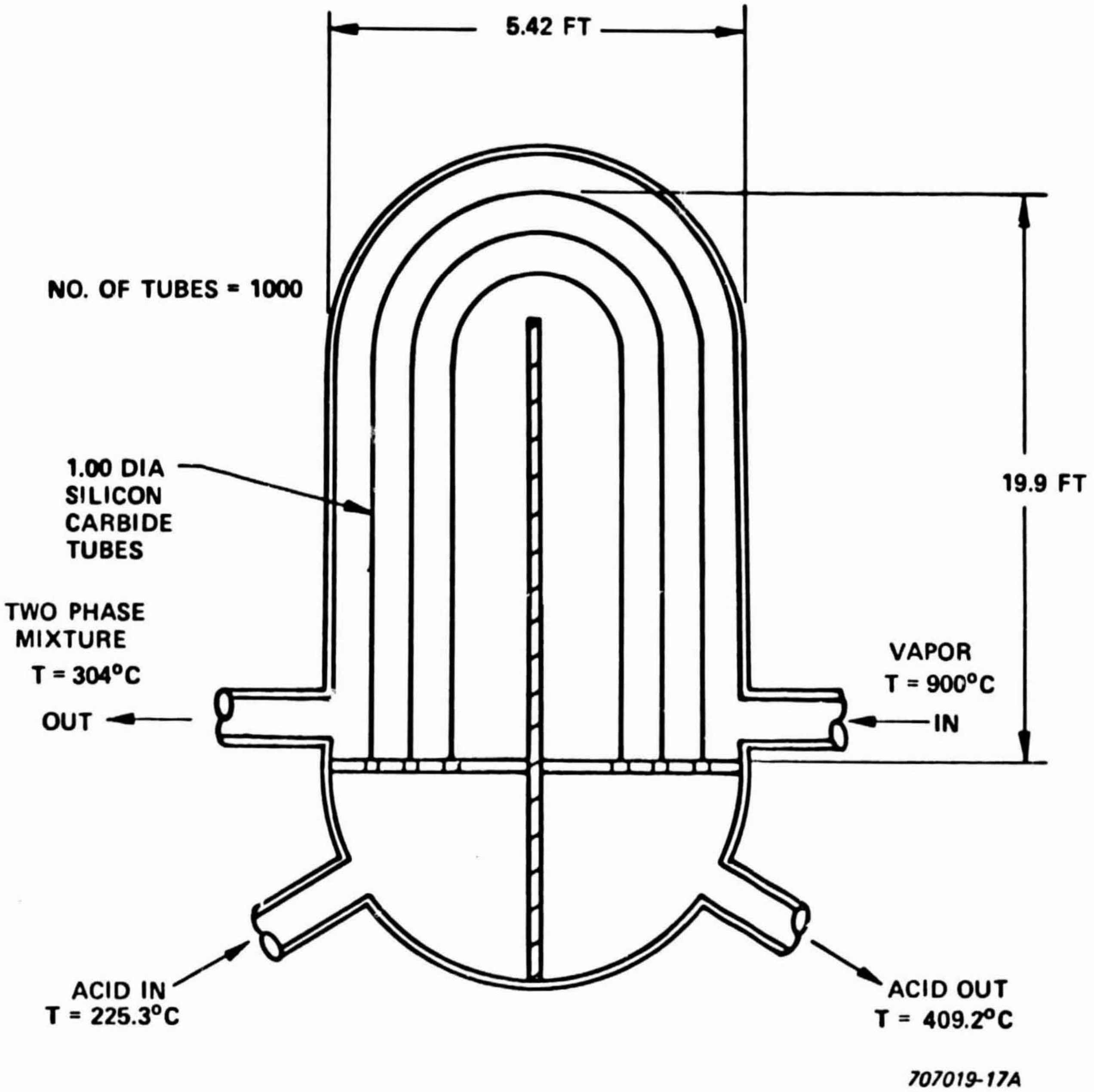

FIGURE 4-15. ACID VAPORIZER (AV-1) – RECUPERATED HEAT SOURCE

The technique of fabricating the silicon carbide components is the same as the SR-2 discussed in the Section 4.1.1.2.2.

4.2.1.2 ACID VAPORIZER (SR-2)

The purpose of the SR-2 is to provide the final stage of vaporization with direct solar heat for the two phase mixture leaving the AV-1. It requires only 6.076 MW_t to change that same flow of acid solution to the condition of saturated vapor at 414.3°C and 7.5 atmospheres, because the acid is preheated in the counterflow heat exchanger (AV-1) to a quality of about 65% at 409.2°C and 7.5 atmospheres. Due to flowrate and outlet condition similarity, the SR-2 design is identical to the outlet end section on the SR-2 design of Case 1 described in the Section 4.1.2. The selected design configuration for the SR-2 of the Case 2 is 240 silicon carbide tubes, 1.0 inch O.D. with 0.125 inch wall thickness and 5.2 feet in length. The heat flux is 200 kW/m^2. The maximum tube wall temperature could reach 1750°F and the total tube pressure drop is 2.66 Psi. The 200 kW/m^2 is a realistic upper limit for the heat flux on this case because a large fraction of the tube length is at a temperature of 1600°F or higher resulting in severe reradiation and energy loss. The receiver efficiency was estimated to be 80 to 85%.

4.2.1.3 HEAT EXCHANGER-1 (HX-1)

The HX-1 for this case is the same as that of the Case 1 described in 4.1.1.4, except that heat load is 18.47 MW instead of 44.72 MW. It is a countercurrent flow heat exchanger. The selected design configuration is 600 silicon carbide tubes, 1.0 inch O.D. with 0.125 inch wall thickness and 8 feet in length. The diameter of the shell is approximately 3 feet. The fabricating method is the same as the HX-1 discussed in the Section 4.1.1.4 except that this heat exchanger is a scale-down version.

4.2.2 EQUIPMENT COST ESTIMATES

The equipment costs for the Case 2 are summarized as follows:

Equipment	Cost (Mid-1983 $)
Acid Concentrator	889,300
Acid Accumulator	364,000
Acid Vaporizer (AV-1)	3,131,600
Acid Vaporizer (SR-2)	208,000
Decomposer (SR-1)	1,909,100
Heat Exchanger (HX-1)	703,200
Heat Exchanger (HX-2)	93,800
Cold Acid Pump (P-1)	6,500
Hot Acid Pump (P-2)	15,000
Pump for HX-1 Cooling Water	2,500
Pump for HX-2 Cooling Water	1,400
Total	7,324,400

For the optimistic design, the cost of the decomposer would be $1,234,200. Thus the total major equipment cost might be reduced to $6,649,500.

4.3 CASE #3

4.3.1 EQUIPMENT DESIGN

All the process equipment for the Case 1 and the Case 3 are identical except the acid vaporizer. The acid vaporizer (AV-1) in the Case 3 utilizes the heat from the thermal storage for the acid vaporization in comparison with the direct solar heat for the Case 1. The design was analyzed as a counter flow heat exchanger with the acid inside the silicon carbide tubes. The shell side fluid from the thermal storage is the molton salt described in Section 4.1.1.3.1. The material of the shell is an alloy steel contains 2-1/4% chromium and 1% molybdenum. However, the tube sheet and the lower head must be made with silicon carbide because it would be in contact with the acid at the temperature range between 225°C and 415°C.

In the analysis the total number of tubes was varied and the required tube lengths were calculated. The analysis included the various heat transfer regimes of heating, boiling and film boiling inside of the one inch silicon carbide tubes. The results of the analysis are summarized in Table 4.5.

TABLE 4-5

ACID VAPORIZER - THERMAL STORAGE HEAT SOURCE

(T_{in} = 500°C; T_{ont} - 350°C; P/D - 1.4)

Silicon Carbide Tube, 1.0: O.D., 0.125" Wall Thickness

No. Tubes	L, ft	Dm, ft	ΔP, psi**	U, NB/B*
500	33.7	3.82	29.6	63/177
1000	28.1	5.42	7.5	36/106
2000	23.9	7.66	1.9	21/62
3000	21.8	9.38	0.8	15/45

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*NB = Nonboiling

B = Boiling

Dm = Diameter of Shell

** Pressure Drop Inside Silicon Carbide Tubes

The selected design is shown in Figure 4-16. It consists of 1000 tubes constructed in U-shape. Each tube is 57 feet in total length. The technique of fabricating the silicon carbide components is the same as that for the SR-2 discussed in Section 4.1.1.2.2. The fabrication of the metal components should be performed to the requirements of the ASME Code.

4.3.2 EQUIPMENT COST ESTIMATES

The equipment costs for the Case 3 are summarized in the following:

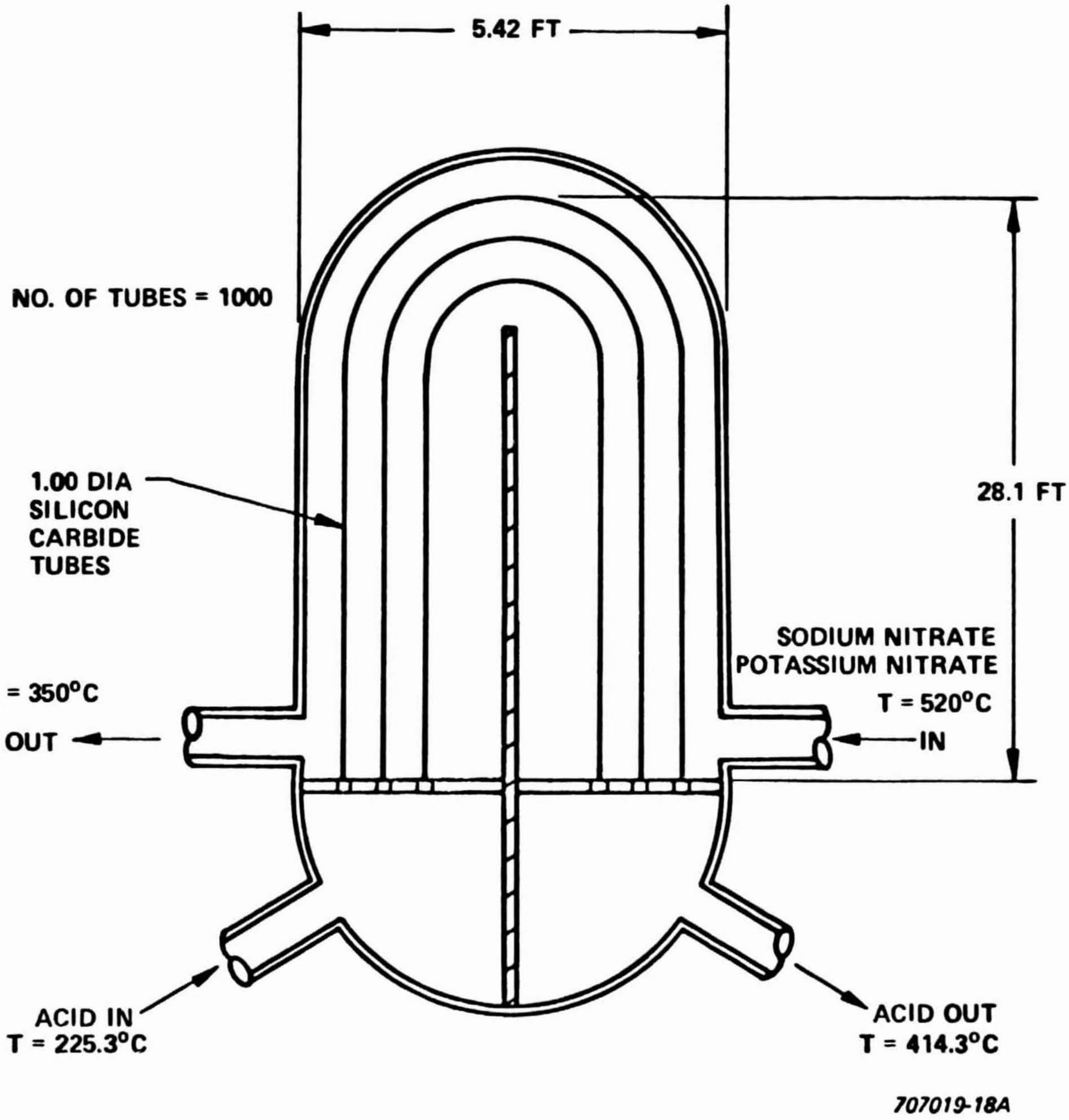

FIGURE 4-16. ACID VAPORIZER (AV-1) – THERMAL STORAGE HEAT SOURCE

EQUIPMENT	COST (MID-1983$
ACID CONCENTRATOR	889,300
ACID ACCUMULATOR	364,000
ACID VAPORIZER (AV-1)	4,000,600
DECOMPOSER (SR-1)	1,909,100
HEAT EXCHANGER (HX-1)	785,000
HEAT EXCHANGER (HX-2)	93,800
COLD ACID PUMP (P-1)	6,500
HOT ACID PUMP (P-2)	15,000
PUMP FOR HX-1 COOLING WATER	2,500
PUMP FOR HX-2 COOLING WATER	1,400
TOTAL:	8,067,200

For the optimistic design, the decomposer cost could be reduced to \$1,234,200 as in the previous two cases. Thus the total major equipment cost would be \$7,392,300.

4.4 CASE NUMBER 4

4.4.1 EQUIPMENT DESIGN

The equipment for the Case 4 are the same as the Case 2 except that the acid vaporizer (AV-2) and the decomposer are powered with an indirect solar heat source using helium as a heat transport medium.

4.4.1.1 ACID VAPORIZER-2 (AV-2)

The design was analyzed as a counter flow heat exchanger with the acid in the tube side and the helium in the shell side. The energy requirement in the AV-2 is 6.076 mW. The helium enters the AV-2 at 580°C (1078°F) and leaves at 517°C (963°F) with a flow rate of 144,981 lb/hr (18.267 kg/sec). The selected design configuration is 250 silicon carbide tubes, 1.0 inch O.D. with 0.125 inch wall and 13 feet in length. The calculated overall heat transfer coefficient is 100 Btu/hr-ft^2-°F and a log-mean temperature difference of 243°F (470°C). The calculated pressure drop is 0.85 psi for the helium side and 2.74 psi for the acid side. The straight tubes are arranged in a triangular pitch array with a center to center distance of 1.36 inches. The shell is constructed of stainless steel and is 2 feet in diameter. The material of the shell can be 316 SS. However, the material of the hot end head and tube-sheet must be Incoloy 800H due to contacting the acid vapor at 415°C. The cold end head and the tube-sheet must be silicon carbide because they would contact with the two phase acid at 409°C. The fabrication of the silicon carbide components is the same as the SR-2 described in Section 4.1.1.2.2. For the seals between the tubes and the tube-sheet, Helicoflex seals should be used as discussed in Section 4.1.1.4.

4.4.1.2 DECOMPOSER

The decomposer is essentially a scaled up version of the decomposer described in Reference 4-7. It was designed as a countercurrent flow heat exchanger with the acid in the shell side and the helium in the tube side. The energy requirement in the decomposer is 37.566 MW_t. The helium enters the decomposer at 976°C (1789°F) and leaves at 580°C (1078°F) with a flow rate of 144,981 lb/hr (18.267 kg/sec). The decomposer is divided into two sections. One is the preheat section and the other is the catalytic section. The acid side of the preheat section is packed with ceramic pellets to enhance the heat transfer. The acid side of the catalytic section is packed with 0.12 inch diameter Fe_2O_3 catalyst pellets to accelerate the SO_3 decomposition reaction. In order to avoid excessive pressure drop on the acid side and still maintain a reasonable heat transfer coefficient, a large number of tubes is required. The material of both shell and tube is Incoloy 800 H (possibly Hastelloy X). The selected design configuration is 9280 tubes 0.75 inch O.D. and 0.5 inch I.D. The lengths are 4.5 feet and 5.5 feet for the preheat section and the catalytic section, respectively. The tubes are arranged in a square pitch with a distance of center to center being 1.0 inch. The calculated pressure drop is 26.5 Psi for the acid side and 0.1 Psi for the helium side. The design and fabrication of this decomposer should be consistent with the requirements of the ASME Code.

4.4.1.3 SOLAR RECEIVER/HELIUM HEATER (SR)

The SR is to employ direct solar energy to heat the helium from 563°C to 976°C at a flow rate of 144.981 lb/hr (18.267 kg/sec). The net thermal energy absorbed by the helium is 39.3 MW. In spite of inert nature of helium, silicon carbide was selected as the tube material due to operating temperature limitations. To design a higher heat flux and enhance the heat transfer, longer and smaller numbers of tubes were selected for the design at the expense of higher pressure drop. Although the helium circulator consumes a significant amount of electricity, the majority of the energy would be recovered as thermal energy to increase the temperture of the helium. The selected design configuration is as follows: 854 silicon carbide tubes, 1.0 inch O.D., 0.125 inch wall thickness, 19 feet in length and constructed in U-shape. The design heat flux is 100 kW/m^2. The maximum tube wall temperature could reach 1823°F. The calculated pressure drop is 11.7 Psi. The construction of this SR is the same as the SR-2 described in the Section 4.1.1.2.2.

4.4.1.4 HELIUM CIRCULATOR

The helium circulator is to circulate the high temperature helium through the decomposer, acid vaporizer (AV-2) and the solar helium heater (SR) at a flow rate of 144.981 lb/hr (18.267 kg/sec) and a total pressure drop if 12.65 Psi. It consumes 4,338 kW electricity, however, the energy is recovered as heat in the helium. Although the pressurized high temperature (1045°F) circulator is not a commercially available item, there is no foreseeable technical difficulty in making this circulator.

4.4.2 EQUIPMENT COST ESTIMATES

The equipment costs for the Case 4 are summarized as follows:

EQUIPMENT	COST (Mid-1983 $)
Acid Concentrator	889,300
Acid Accumulator	364,000
Acid Vaporizer (AV-1)	3,131,600
Acid Vaporizer (AV-2)	523,300
Decomposer	2,307,200
Receiver/Helium Heater (SR)	2,088,700
Heat Exchanger (HX-1)	703,200
Heat Exchanger (HX-2)	93,800
Cold Acid Pump (P-1)	6,500
Hot Acid Pump (P-2)	15,000
Pump for HX-1 Cooling Water	2,500
Pump for HX-2 Cooling Water	1,400
Helium Circulator	434,000
Total	10,560,500

The cost is highest for this case and must be evaluated in comparison with the values obtained by decoupling the thermochemical processes from the solar heat source.

For easier comparison, the equipment costs and the costing methods are summarized in Table 4-6 for all four cases.

TABLE 4-6. EQUIPMENT COST SUMMARY FOR FOUR CASES

EQUIPMENT	CASE #1		CASE #2		CASE #3		CASE #4	
	COST (Mid-1983 $)	*ESTIMATING METHOD	COST (Mid-1983 $)	*ESTIMATING METHOD	COST (Mid-1983 $)	*ESTIMATING METHOD	COST (Mid-1983 $)	*ESTIMATING METHOD
ACID CONCENTRATOR	889,300	C and E	889,300	C and E	889,300	C and E	889,300	C and E
ACID ACCUMULATOR	364,000	A	364,000	A	364,000	A	364,000	A
ACID VAPORIZER (SR-2)	935,900	A and D	208,000	A and D				
ACID VAPORIZER (AV-1)			3,131,600	C and E	4,000,600	C and E	3,131,600	C and E
ACID VAPORIZER (AV-2)							523,300	C and E
DECOMPOSER (SR-1)	1,909,100	B and C	1.909,100	B and C	1,909,100	B and C		
DECOMPOSER							2,307,200	
RECEIVER/HELIUM HEATER (SR)							2,088,700	
HEAT EXCHANGER (HX-1)	785,000	C and E	703,200	C and E	785,000	C and E	703,200	C and E
HEAT EXCHANGER (HX-2)	93,800	C	93,800	C	93,800	C	93,800	C
COLD ACID PUMP (P-1)	6,500	A	6,500	A	6,500	A	6,500	A
HC ACID PUMP (P-2)	15,000	A	15,000	A	15,000	A	15,000	A
HELIUM CIRCULATOR							434,000	C
PUMP FOR HX-1 COOLING WATER	2,500	C	2,500	C	2,500	C	2,500	C
PUMP FOR HX-2 COOLING WATER	1,400	C	1,400	C	1,400	C	1,400	C
TOTAL	5,002,500		7,324,400		8,067,200		10,560,500	

*A = Vendor's Quote
C = Cost Estimating Method from Literatures
E = Discussion with Vendors
B = Using Information from Past Published Studies
D = Westinghouse Own Experience

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5.0 SCALE DOWN PROCEDURE

In the course of developing the engineering data and information for the design of full-size equipment, properly scaled down prototypical designs can be a cost effective experimental approach. The key elements of a generalized scaledown procedure, appropriate to the components and subsystem under study, are presented herein.

Many process equipment items, including heat exchangers, pumps, and piping can usually be designed for the prototype without any special attention to scaling parameters. Factors in equipment scale-down and design for some of the process equipment are shown in Table 5-1.

TABLE 5.1

FACTORS IN EQUIPMENT SCALE-DOWN AND DESIGN

Type of Equipment	Major Variables for Operational Design (Other Than Flow Rate)	Major Variables Characterizing Size or Capacity
Continuous Reactors	• Reaction rate • Equilibrium state	• Flow rate • Residence time
Evaporators	• Latent heat of vaporization • Temperature	• Flow rate • Heat-transfer area
Tube-and Shell Heat Exchanger	• Temperatures • Viscosities • Thermal conductivities	• Flow rate • Heat-transfer area
Centrifugal Pumps	• Discharge head	• Flow rate • Power input • Impeller diameter

Based on the indicated characterizing variables, those equipment mentioned above can be scaled down to a ratio of at least 10:1.

Scale considerations for other process equipment are discussed as follows:

- Acid accumulator: Scale factor should be based on the total volume of the acid that has to be accumulated in the operation of the prototype.
- Thermal Storage: Scale factor should be based on the demand of the prototype thermal power from the storage, in kW-hrs.
- Solar Receiver/Vaporizer: Since the heat exchanger tubes are a modular type construction, the ideal scale-down would be based on the thermal power requirements to determine the numbers of modules required. Maintaining the same heat flux, tube size, flow rate per tube, an identical fluid dynamic conditions could be maintained. Identical fluid dynamic conditions will assure the same heat and mass transfer rates for both smaller and larger units. However, in order to maintain a practical ratio of height/width for the receiver configuration, the module arrangement, tube size, tube length and heat flux may have to be changed for the smaller unit. As a result, compromises may be required in the overall heat transfer coefficient and the maximum tube wall temperature. Every effort should be made to minimize such compromises.
- Solar Receiver/Reactor: The reactor tubes are also a modular type construction. Therefore, the scale-down should be based on the thermal power requirements for the reactors so that identical fluid dynamic conditions can be maintained. The use of identical fluid dynamic conditions will assure the same heat and mass transfer rates between catalyst and reacting gas, and the same pressure drop for both smaller and larger units. In addition, using the same size and make of commercial catalyst to run at the same rate will achieve not only similarity but identity in all sizes for the catalyst - gas system. In case the heat flux or the reactor tube modular arrangement has to be changed for the smaller plant (to maintain a desirable ratio of height to diameter for the receiver configuration) the overall heat transfer coefficient and the maximum tube wall temperature may have to recalculated. If this is necessary, the pressure, temperature, mass velocity, Reynolds number, catalyst particle diameter and gaseous space velocity should be maintained as closely as possible to the desired full-scale equipment values.

6.0 DEVELOPMENT REQUIREMENTS

Although detailed design of the process equipment is not within the scope of the studies, based on current technology and material considerations a preliminary configuration design as well as sizing of the equipment has been carried out. In addition, each piece of equipment was rated in terms of its developmental status and a risk measure using a equipment application rating matrix. This matrix assigns equipment into four different classifications:

A - Established Technology
B - Near Term
C - Developmental
D - Speculative

Table 6-1 illustrates the matrix and elaborates on these cataegories. Table 6-2 shows the summary of the rating for each major item of equipment in each case.

The development requirements in the area of design, materials and fabrication and assembly of actual components are described in Section 6.1, 6.2 and 6.3. No major difficulty is anticipated in performing any of these activities and in successfully resolving problem areas within the schedular constraints of the STPIS program.

6.1 DESIGN

- The design of heat exchanger systems made of silicon carbide is more complex than the design of conventional metal components and systems because silicon carbide, a ceramic, is brittle. The joints, particularly at structural discontinuities and the support systems require sophisticated analyses and design refinements (Reference 6-1) to minimize strains. Design data on mechanical properties derived from testing of fabricated specimens are not presently available and are required for determining safe stress levels. Similarly, knowledge of in-service strains generated on tubing must be obtained.

TABLE 6-1

EQUIPMENT APPLICATION RATING CRITERIA

APPLICATION RATING	EQUIPMENT STATUS EQUIPMENT STATUS	PERFORMANCE DATA BASE	EXTRAPOLATION FROM PERFORMANCE DATA BASE	WORK REQUIRED	R&D PROGRAM CHARACTERISTICS	
					RATIONALE	SUCCESS PROBABILITY
Established (A)	Firm selections can be made. Equipment is commercially available in form required.	Sufficient	None	Minimal, routine applications engineering	Not Applicable	Not Applicable
Near Term (B)	A number of equipment candidates are identified. Candidates are commercial or near commercial.	Incomplete	Short extrapolations from existing data base are involved.	Confirmatory testing and minimal R & D	Straight-forward	Virtually Certain
Developmental (C)	Equipment not previously designed, but engineering data base exists for design.	Incomplete; important gaps exist.	Large extrapolations from existing data required.	Considerable R & D is required.	A credible rationale exists. Alternative avenues are evident.	Good to excellent.
Speculative (D)	Equipment not previously designed with major materials, design or manufacturing uncertainties.	Sparse or Absent	Highly speculative or not possible.	Extensive R & D is required.	Rationale is not clear, or requires a breakthrough or serendipity.	Fair to Poor.

TABLE 6-2. RATING OF THE EQUIPMENT

	Case 1	Case 2	Case 3	Case 4
Acid Conc.	C	C	C	C
Acid Accuml.	A	A	A	A
SR-1	B	B	B	NA
SR-2	C	C	C	NA
AV-1	NA	C	C	C
Decomposer	NA	NA	NA	B
AV-2	NA	NA	NA	C
SR	NA	NA	NA	C
HX-1	C	C	C	C
HX-2	B	B	B	B
Cold Acid Pump	A	A	A	A
Hot Acid Pump	B	B	A	A
Helium Circulator	NA	NA	NA	B
Molton Salt Pump	A	A	A	A

- The design of an adequate support systems to avoid excessive vibrational strains requires development. Tube spacers may be required for the acid vaporizer and acid concentrator silicon carbide tubing. A satisfactory design of such tube spacers needs to be developed.

- The acid concentrator will have a temperature differential of about 250°C across the silicon carbide tube sheet thickness. A thermal liner will be required to minimize the effect of this temperature differential and the design of an effective liner and supports has to be developed.

- Very little data is available on the performance of silicon carbide to silicon carbide and silicon carbide to metal joints and seals under thermal cycling conditions. Information in these areas need to be developed through testing.

6.2 MATERIALS AND FABRICATION

- The forming and fabrication of silicon carbide components are presently limited to small shapes and sizes. Technology for manufacturing long and large diameter tubes and spiral shapes such as S and U bends needs to be developed.

- Large diameter (> 6 inch) Incoloy 800 H tubing or piping and fittings are not readily available at the present time. Alternate methods, such as the use of Incoloy lined pipe or fabricated (welded) Incoloy pipe may require development.

- Techniques for the field joining of silicon carbide to silicon carbide and silicon carbide to metal are not presently available. The development of these techniques would permit greater design flexibility.

- More information needs to be developed in regard to the durability of silicon carbide tubes under service conditions. The effect of static and cyclic fatigue on the strength of silicon carbide needs to be understood more fully. Similarly crack growth data under service conditions and erosion-corrosion mechanisms on the surface of the silicon carbide have not yet been evaluated (Reference 6-1).

6.3 HIGH TEMPERATURE ACID PUMP (P-2)

As discussed in the Section 4.1.1.8 in order to withstand 300 Psig pressure, the metal casing of the pump has to be lined with either PTFE teflon or silicon carbide. The temperature at which PTFE is able to resist the acid corrosion is drastically reduced with increased pressure level. The manufacture's test data indicated that PTFE could not withstand sulfuric acid corrosion at 225°C and 300 Psig. Thus a better teflon to resist acid corrosion at higher pressure and temperature needs to be developed. Silicon carbide could withstand acid corrosion at high pressure and temperature, but it may pose difficulty in making a molded lining to metal. The immediate solution would be to use the low temperature acid pump described in the Section 4.1.7 to pump the acid to the top of the solar receiver tower, and locating the acid accumulator on the tower. As a result, the casing of the high temperature pump only has to withstand a pressure of 95 Psig so that a commercially available pump made with Durichlor 51, from Duriron Company can serve the purpose. On the other hand the low temperature acid pump would require a higher discharge pressure head (205 Psig), thus its casing must be made with Hastelloy B instead of Durichlor 51.

6.4 REFERENCES

6-1 H. W. Carpenter and J. Campbell, Jr., "High-Temperature, Coal-Fired Combustor with Ceramic Heat Exchangers for CCGT Systems," ASME 80-GT-155.

7.0 COMPARISON OF DESIGN ALTERNATIVES

The four system concepts, described in Section 3, were evaluated against the commonly accepted criteria: a) Efficiency, b) Cost, and c) Risk. The selection of a preferred configuration for further evaluation and development, as presented herein, is limited to consideration of that portion of the total system under study. It will be necessary to consider, at some future time, any impacts of the remainder of the system when the information becomes available.

In the area of efficiency: within the system under study, only the energy requirements per unit of SO_2 produced can be addressed. The energy requirements include the thermal energy and electrical energy. As presented in Section 3.0 for a same quantity of SO_2 production, Case 2 requires least amount of energy input, then Case 4, then Cases 1 and 3.

In the area of cost: only the equipment cost within the scope of the study was evaluated. As indicated in Section 4.0, the equipment cost for Case 1 is the lowest, then in the order of Case 2, Case 3 and Case 4. The equipment cost for Case 4 is twice as much as that of Case 1 mainly because decoupling the solar heat source requires an additional high temperature heat exchanger and a high temperature helium circulator.

In the area of risk: considerations were given to development requirements with chance of success, and operational concern including safety and ease of operation. The evaluation of the development requirements were based on the results presented in Section 6.0, and the operational concern was based on engineering judgement.

In view of the high equipment cost for Case 4, it was ruled out for further consideration despite its least operational concern. Case 3 has also been ruled out because it is worse then Cases 1 and 2 in the areas of efficiency and cost. Qualitatively, we feel that Case 1 is better than Case 2 within the scope of the study. However, to have a better comparison the study should be expanded to cover the heliostat field.

Efforts were made to develop a methodology for quantitative comparison to evaluate the four system concepts. The evaluation process is described in Section 7.1. The evaluation criteria are discussed in Section 7.2. This subsection also presents the selected numerical weighting factors. The resulting evaluation is presented in Section 7.3. Although each individual may have somewhat different opinion in assigning the weighting factors, we believe that this methodology, in fact, is a good and fair numerical evaluation process. When the scope of the study is expanded and detailed design information is available the methodology can be applied to perform further evaluation.

7.1 EVALUATION PROCESS

For the selection of a preferred configuration for development, a numerical evaluation process was used. Basic evaluation criteria, and how well each candidate system met these criteria, were considered. The process was based on classical systems optimization techniques, and used a combination/comparison of numerical values representing the "standing" or relative worth of each configuration in each criterion, and on an overall basis, to aid in final selection of the most appropriate configuration for further development.

Simply stated, the process used a "figure-of-merit" comparison to delineate the most worth candidate configuration. The "figure-of-merit" (FOM) for the N^{th} configuration (FOM(N)) was defined as:

$$FOM(N) = W_1 C_{1N} + W_2 C_{2N} + \dots + W_{I-1} C_{(I-1)N} + W_I C_{IN}$$

$$FOM(N) = \sum_{K=1}^{I} W_K C_{KN}$$

where

W_K = Numerical weighting factor for K^{th} criterion in find criterion set, single-valued for each K

C_{KN} = Normalized numerical value representing the worth of each configuration N in each criterion area K; thus, a single discrete value of C must be derivable for each pair (K, N)

The set of K criteria and their respective weighting factors W_K were derived through an interactive process in which technical personnel familiar with the systems in general were participants.

The most worthy configuration was defined as that configuration which, when evaluated with respect to the specified set, resulted in the highest figure-of-merit.

7.2 EVALUATION CRITERIA

As noted previously, experienced technical and management personnel interacted to delineate appropriate criteria for the evaluation, and to assign weighting factors according to the perceived importance of the criteria to the choice of a best configuration. The appropriate criteria were selected so that they represented the three major issues areas which are generally common to the evaluation of nearly all complex systems, that is, the areas of:

- Efficiency (Energy Requirements)
- Cost
- Risk

Classical systems engineering evaluation techniques always prescribe the utilization of evaluation criteria from all three areas in order to achieve a balanced viewpoint with regard to the specific system being evaluated.

Within the scope of this contract study, the following four criteria were selected for the evaluation process:

In The Efficiency Area

- Energy input per unit of SO_2 produced.

In The Cost Area

- Capital Cost of the major equipment items. Piping, solar tower and heliostat costs were not considered as these are beyond the scope of this study.

In The Risk Area

- Development requirements.
- Operational concern.

The set of four criteria chosen limits the complexity of the selection process but provides full coverage with respect to the main objective.

7.2.1 NUMERICAL EVALUATION

The most difficult feature of the evaluation process is the conversion of comparative worth within a criterion into a numerical value. Optimally, this can be done by specifying a given quantifiable characteristic of the configuration under study which represents the whole of each criterion. If a single quantitative value cannot be defined, then a set of such values, each characteristic of one or more of fully representative dimensions or measures within each criterion, and for each configuration. It is clear that each criterion must have measures defined within it so that the final result of judging the worth of a configuration is a numerical value. These values must be arrived at in a fashion which assigns the highest value to the configuration of the highest worth, if the evaluation is to be consistent. Hence, the definition of measures and valuations within each criterion must be attended to with this simple relationship in mind.

The following subsection defines the measures and valuation process used for each criterion.

CRITERION 1 - ENERGY INPUT PER UNIT OF SO_2 PRODUCED

This criterion can be virtually completely represented by a directly calculable value characteristic for each configuration--total energy input divided by total moles of SO_2 produced. Since this contract study is only a sub-system of the solar thermochemical hydrogen production process and the useful product from this sub-system is the SO_2, therefore, the total energy input per unit of SO_2 produced is a measure of efficiency. Since the Case #2 required least energy input, it was assigned a highest value, 1.0, and the rest of the three cases were compared with the Case #2 by dividing the energy input for the Case #2 by the energy input for each case. To normalize C_{1N}, the value from each case was divided by the summation of the values from the four cases.

CRITERION 2 - EQUIPMENT COST

The equipment costs are shown in the Section 4.0 of this report, which do not include piping, solar tower and the solar collectors (heliostats). The case #1 (lowest cost) was given a highest value, 1.0, then the other three cases were compared with the Case #1 to obtain a relative value (fraction of the value of the Case 1). Then the value for each case was divided by the summation of the values from the four cases to obtain the normalized C_{2N}.

CRITERION 3 - DEVELOPMENT REQUIREMENTS

Table 6-2 in the Section 6.0 classifies individual pieces of equipment in terms of their developmental status. The matrix assigns equipment into four different development classifications, and these classifications may be used as a rough estimate for the probability of failure of each piece of equipment. For instance, it may be assumed that failure of any one of the major components of a system results in a similar consequence for each configuration considered, that is, failure of the entire program. Thus, the risk is a function of the probability of failure, and may be evaluated numerically (i.e., on a relative basis) by comparing the relative development status of the major components in each candidate configuration, given an acceptable "mapping" of the alphabetic classifications into numerical values.

The specific method of application used for evaluation is described below:

a) The list of major components for each candidate PTU configuration is drawn up. Specific equipment related to the solar heat source systems is included to permit an estimate of the total system failure rate.

b) The development status of each component is assigned by reference to the equipment application rating matrix.

c) The alphabetic status indicator is converted to a numerical value by use of the arbitrary mapping

A = 10
B = 7
C = 4
D = 1

d) For each configuration, the assigned values for all components are summed directly.

e) The resulting sum is normalized by division by the number [10 x (total number of components considered)]. This results in a value equal to or less than 1.0 for the Risk Criterion Value, designated C_{3N} for the N^{th} configuration.

CRITERION 4 - OPERATION CONCERN

Particular emphases have been put on safety and environmental considerations and social effects. However, the lack of detailed and operating experience precludes the use of a numerical measure. To deal with this type of criterion, a value of rating was given to each case based on best judgement with a scale of 0 to 10 (10 is the least operational concern). Then each value was divided by the summation of four values to normalize C_{4N}.

7.2.2. WEIGHTING FACTOR

Each criterion was assigned a weighting factor which related to its perceived importance in the development program. These factors were chosen on a subjective scale of 0-10, with 10 representing a criterion of highest importance. The weighting of each criterion was as follows:

Criterion	Weighting	Normalized Factor
Energy input per unit of SO_2 produced	9	0.3
Equipment Costs	9	0.3
Development Requirements	6	0.2
Operational Concern	6	0.2

The weighting factors were then normalized, and the final form of figure-of-merit (FOM) became for the Nth configuration
$\text{FOM (N)} = 0.3\ C_{1N} + 0.3\ C_{2N} + 0.2\ C_{3N} + 0.2\ C_{4N}$

7.3 CALCULATION OF EVALUATION PROCESS

7.3.1 CALCULATION OF C_{1N}'S

The calculation of C_{1N} is described in the Section 7.2.1 and is based on the total energy input shown in the Section 3.3 of this report.

Case No.	Total Energy Input (kW)	Weighting	Normalized Factor
1	101,640	0.741	0.2177
2	75,324	1.000	0.2939
3	101,581	0.741	0.2177
4	81,775*	0.921	0.2707
		3.403	1.000

*Since the helium circulator uses a significant amount of electricity, a 40% efficiency for the conversion from thermal energy to electricity is assumed.

7.3.2 CALCULATION OF C_{2N}'S

The calculation of C_{2N} is described in the section 7.2.1 and is based on the cost summary shown in the Section 4.0 of this report

Case No.	Equipment Cost ($)	Weighting	Normalized Factor
1	5,002,500	1.000	0.3601
2	7,324,400	0.683	0.2459
3	8,067,200	0.620	0.2233
4	10,560,500	0.474	0.1707
		2.777	1.0000

7.3.3 CALCULATION OF C_{3N}'S

The calculation of C_{3N} is described in the Section 7.2.1 and is based on the equipment rating shown in Table 6-2.

Case No.	Weighting	Normalized Factor
1	0.700	0.2550
2	0.670	0.2441
3	0.700	0.2550
4	0.675	0.2459
	2.745	1.0000

7.3.4 CALCULATION OF C_{4N}'S

The calculation of C_{4N} is described in the Section 7.2.1 previously.

Case No.	Weighting	Normalized Factor
1	6	0.2069
2	6	0.2069
3	8	0.2759
4	9	0.3103
	29	1.0000

$$\text{FOM} = 0.3\ (C_{1N} + C_{2N}) + 0.2\ (C_{3N} + C_{4N})$$

Case No.	C_{1N}	C_{2N}	C_{3N}	C_{4N}	FOM	Ranking
1	0.2177	0.3601	0.2550	0.2069	0.26572	1
2	0.2939	0.2459	0.2441	0.2069	0.25214	2
3	0.2177	0.2233	0.2550	0.2759	0.23848	4
4	0.2707	0.1707	0.2459	0.3103	0.24366	3

Within the scope of this contract study, the Case #1 receives the highest ranking.

8.0 NEW TECHNOLOGY

Under this contract study, one innovative idea, a stepped panel concept for external type solar receiver, was identified. The concept was invented by B. R. Nair and A. R. Jones, and was described in this report on Pages 69 and 70. It was also reported verbally at the final review meeting with J.P.L. and D.O.E. on October 29, 1983.

9.0 RECOMMENDATIONS FOR FURTHER WORK

The efforts reported herein have clarified the mechanical, fluid, and energy interfaces of the vaporization and decomposition subsystem. Important conceptual design and cost information has been developed for the major equipment items. Appropriate directions for continued effort were also identified.

The energy requirements relate directly to the interfaces with the heliostat field and the solar tower. Potentially valuable refinements of the solar receivers, for vaporization and decomposition, may be achievable through further study of these interfaces. One objective should be to refine the comparison between the study cases through inclusion of tower and heliostat field costs. A second objective should be to refine the desirable range of receiver design and development values, such as heat flux on the receiver tubes, in terms of overall economics. A third objective should be cost reduction of the receivers, especially the vaporizer since it is to be constructed of silicon carbide. A fourth objective should be cost reduction of the heliostat field through study aimed at improving receiver efficiency. An example of an approach to this last objective is to seek to optimize the arrangement of decomposer receiver, tower and related portion of the field so that the cavity opening (which is a major factor in heat loss) is minimized.

In the design study of the decomposer, a flat radial temperature profile in the catalyst bed was assumed. Preliminary analysis indicated that the assumption was valid for the values of the heat flux and average reaction rate used in the design. However, further study will be required if higher than the designed heat flux is used, and the design of the decomposer should be refined based on the considerations of heat flux, reaction rate, pressure drop and heat transfer.

An approach to cost reduction of the vaporizer should be to explore the cost effecitveness of design variations as a function of heat flux on the receiver tubes. If the results indicate that higher heat flux (compared to design assumptions used to date) would be economically attractive, analytical work and modified design concepts to allow such heat fluxes should be pursued.

An important design consideration, in components to be constructed of silicon carbide, is the potential for thermal stresses that place the relatively brittle silicon carbide in tension. In the prior work, the design approach has been to minimize the risk of such thermal stresses by using U-tube configurations in the vaporizer/receiver. However, should the need for increased heat flux arise, the risk would be intensified. It may also prove significant as off-design conditions, such as startup, are explored. These areas of risk should be explored through thermal and stress analyses. It is recommended that finite element techniques are used to calculate stresses due to pressure, differential expansion and thermal loading at steady state conditions and thermal transient conditions for major silicon carbide components.

The success of the heat exchangers which require construction using silicon carbide or silicon carbide and metal, is largely dependent on the approach to joining the tubes to tube sheets. The previous work by Westinghouse has led to the recommendation that Helicoflex seals be considered. Under a previous solar hydrogen contract between the U. S. Department of Energy and Westinghouse, a test rig was designed by Westinghouse and all components of the test rig and the test articles were procured. The test rig can be used to demonstrate the performance of a single Helicoflex seal identical to those in the silicon carbide heat exchangers in design, materials, installation, pressure loading, temperature, fluid environment, and axial displacement. A mini-computer controlled testing machine which can be used in conjunction with the test rig is available at Westinghouse Advanced Energy Systems Division. It could be made available for future cyclic testing. It is strongly recommended that the Helicoflex seal be tested in the test rig to verify the seal concept.

10.0 CONCLUSIONS

1. All requirements of the Contract Work Scope were completed.

2. Each of the selected subsystems appear to be technically feasible with a good to excellent probability of successful development.

3. Each of the selected subsystems contains major equipment items requiring considerable research and development.

4. The Case 1 (Direct solar acid vaporization and decomposition) subsystem was judged to have the highest--most favorable--rating. It should be noted that the evaluation was necessarily limited to considerations within the subsystem (i.e., the work scope).

5. The comparison of design alternatives should be reconsidered when the total system data become available. The cost and efficiency criteria may be significantly impacted when the solar tower and heliostat fields are considered, since the solar energy requirement varies from 70 to 101 MW.

6. The interface between receivers, tower and heliostat field has an important bearing on the evaluation of the cases as well as on the performance and cost of the receivers and should be explored in future work. The results obtained in the study of this interface should then be factored into refined receiver designs, including stress analyses and cost estimates.

7. Several heat exchangers,considered in the study, require silicon carbide material. Previous Westinghouse efforts have identified helicoflex seals as a promising technique for joining tubes to tube sheets. Cyclic testing is recommended.