ORNL-TM-4289

A DUAL-PURPOSE NUCLEAR DESALTING PLANT USING A LOW-TEMPERATURE EVAPORATOR

John E. Jones, Jr.

This document has been reviewed and is determined to be APPROVED FOR PUBLIC RELEASE.
Name/Title: Leesa Laymance, ORNL TIO
Date: 6/27/2016

MASTER

OAK RIDGE NATIONAL LABORATORY
OPERATED BY UNION CARBIDE CORPORATION • FOR THE U.S. ATOMIC ENERGY COMMISSION

ORNL-TM-4289

Contract No. W-7405-eng-26

A DUAL-PURPOSE NUCLEAR DESALTING PLANT USING A LOW-TEMPERATURE EVAPORATOR

John E. Jones, Jr.

Contributors

L. C. Fuller
E. C. Hise
S. A. Reed
R. J. Kedl
J. V. Wilson

SEPTEMBER 1973

NOTICE

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

OAK RIDGE NATIONAL LABORATORY
Oak Ridge, Tennessee 37830
operated by
UNION CARBIDE CORPORATION
for the
U.S. ATOMIC ENERGY COMMISSION

MASTER

DISTRIBUTION OF THIS DOCUMENT IS UNLIMITED

CONTENTS

	Page
ABSTRACT	1
INTRODUCTION	1
SUMMARY	2
CONCLUSIONS AND RECOMMENDATIONS	5
HEAT COST	6
MATERIALS EVALUATION	8
HEAT TRANSFER TESTS	9
FEED TREATMENT	12
PROCESS DESCRIPTION	15
PRELIMINARY OPTIMIZATION STUDY OF LOW-TEMPERATURE EVAPORATOR	15
REFERENCE PLANT DESIGN	16
COST ESTIMATES	26
REFERENCES	30
APPENDIX A	
APPENDIX B	

A DUAL-PURPOSE NUCLEAR DESALTING PLANT USING A LOW-TEMPERATURE EVAPORATOR

John E. Jones, Jr.

ABSTRACT

A low-temperature vertical tube evaporator is evaluated and compared with a conventional high temperature VTE-MSF. In dual-purpose plants, low-temperature evaporators appear to have a favorable economic advantage because of lower heat cost, lower cost materials, such as aluminum, and lower feed treatment cost.

The reference design is an all-aluminum 50 Mgd forward feed vertical tube evaporator with a single stage of flash feed heating per effect. Maximum brine temperature is 170°F. Water cost for the reference design is 41.3¢/1000 gallons compared with 49.4¢/1000 gallons for the high-temperature VTE-MSF which has a steel shell and uses copper-nickel heat transfer tubes. The major portion of the cost savings is in the cost of operation (heat, power, chemical, operating) where the low-temperature plant costs less than 20¢/1000 gallons to operate.

INTRODUCTION

Most of the development effort on dual-purpose nuclear desalting plants has been directed toward plants using high-temperature evaporators. Primarily, the interest in high-temperature evaporators was a carryover from single-purpose evaporator plants where the heat cost was quite high and also independent of maximum evaporator temperature. Under this circumstance it is desirable to develop the highest-temperature evaporator possible (~260°F maximum brine temperature with acid feed treatment).

In a dual-purpose plant it is not necessarily true that the high-temperature plant will yield the most economical water cost. On the contrary, there are some very good reasons why lower-temperature evaporators may be more economical. First, lower cost materials such as aluminum alloy heat transfer surface and concrete shells appear to be satisfactory for low-temperature operation. Second, significantly lower heat costs can be expected from dual-purpose plants at these lower temperatures. Third, chemical feed treatment costs are substantially reduced at low-temperature conditions. And finally, power plant turbines with exhaust conditions up to about 179°F are being developed for use with dry cooling towers. These

same turbines would be suitable for use with low-temperature evaporators at an ocean site.

Most people buy their water for about 10¢/ton for municipal and industrial use. Some farmers get water for irrigation for less than 1¢/ton. These traditional water prices have tended to set the economic goals for producing fresh water from the sea. The development of any system that will process materials for a few cents per ton is a real technical challenge. Distillation is the best developed and appears at this time to have the most potential for producing fresh water from seawater. But to do this on a large scale at low cost will require the use of dual-purpose plants — i.e., plants that produce both electricity and desalted water — and it will be necessary to bring about further economic improvements in the distillation process. The low-temperature water plant is felt to be a most promising avenue toward significant improvement.

SUMMARY

A schematic flow sheet of the low-temperature evaporator is shown in Fig. 1. Basically, it is a forward feed vertical tube evaporator with a single stage of flash feed heating per effect.

From the standpoint of cost, the dual-purpose plant using a low-temperature evaporator looks very promising. Figure 2 summarizes water cost from the low-temperature plant in comparison with a high-temperature VTE-MSF dual-purpose plant. Costs are based on a 50-Mgd size plant in each case. The low-temperature evaporator is all aluminum construction, while the high-temperature plant has a steel shell and uses copper-nickel heat transfer tubes.

In order to effectively utilize lower cost heat, the low-temperature evaporator requires much more heat transfer surface area. Fortunately, the application of low-cost aluminum alloy heat transfer surface and the simpler, single-stage flash feed heater per effect design offset the capital cost increase which would normally result. The net effect is a slight savings in cost for the low-temperature plant. The major cost saving is in the cost of operation (heat, power, chemical, operating) where the low-temperature plant costs less than 20¢/1000 gallons to

ORNL DWG 73-8383

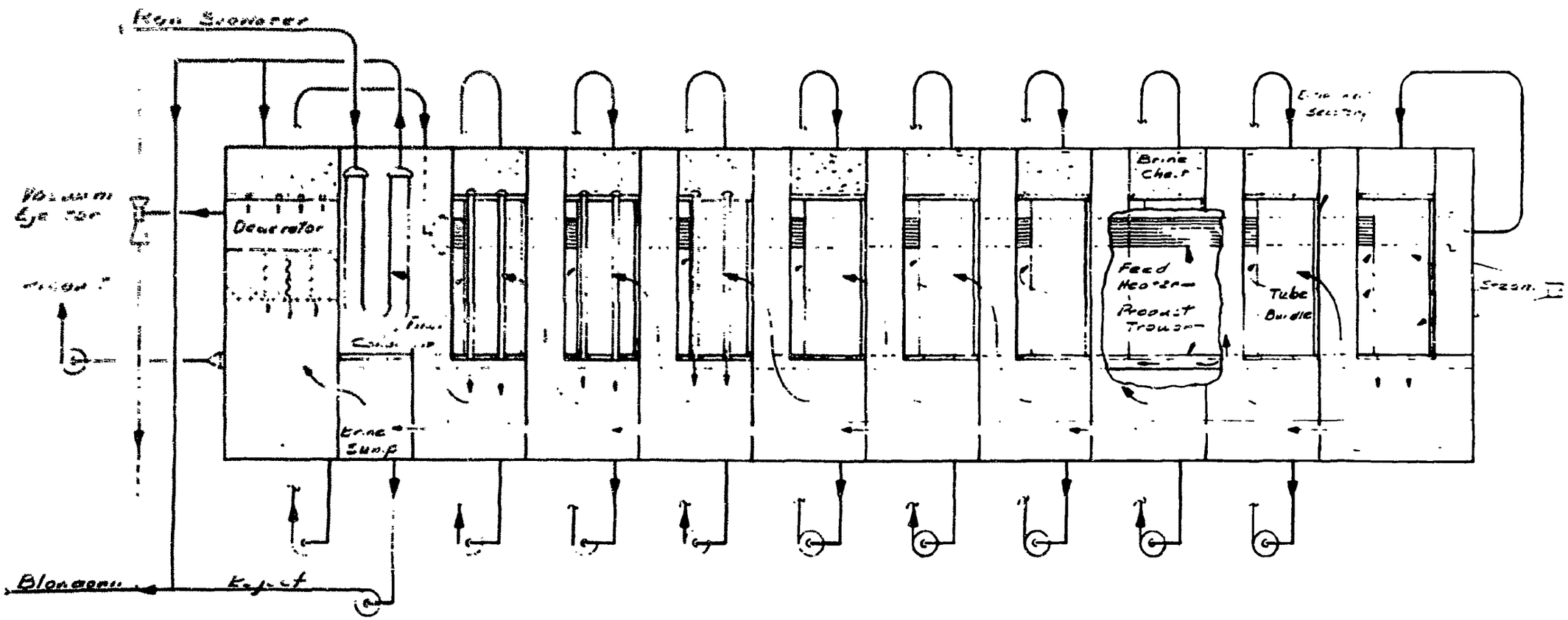


Fig. 1. Schematic flow sheet of low-temperature evaporator.

ORNL-DWG 73-4378

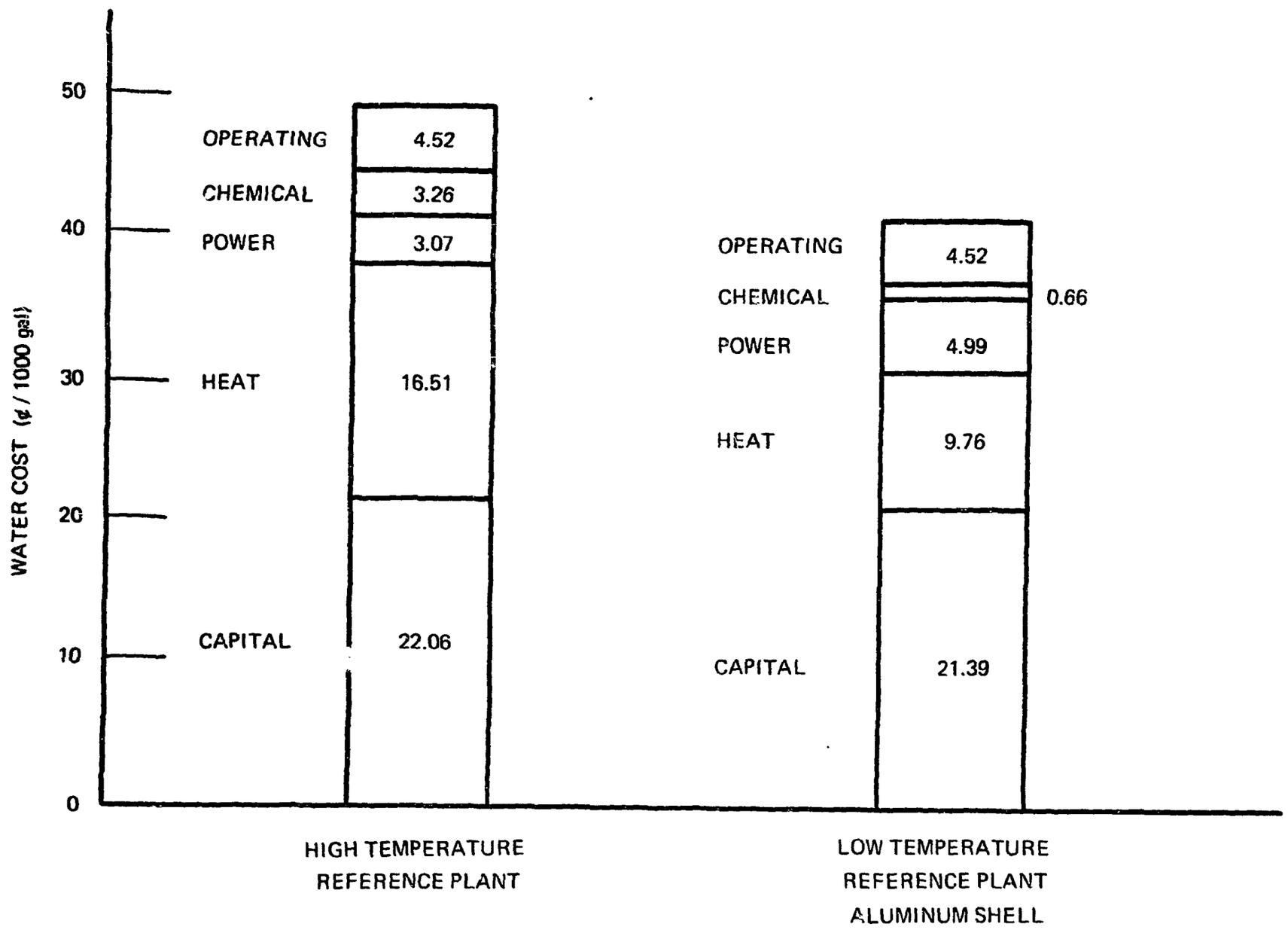


Fig. 2. Water cost from low-temperature plant vs water cost from high-temperature VTE-MSF dual-purpose plant.

operate. This is especially important in the early stages of development because capital cost is often written off for demonstration or prototype plants, but the cost of operation must be recovered from the sale of the product water.

One area where additional costs may be incurred in either plant is in additional coupling equipment such as an isolation loop for protection from possible radioactive contamination or auxiliary condensers to provide a heat sink for the power plant during evaporator outages. These components, if required, would be somewhat more expensive for the low-temperature plant because of its higher steam flow.

CONCLUSIONS AND RECOMMENDATIONS

It is concluded that there is significant potential for additional economic improvement in desalted water cost through the application of low-temperature aluminum evaporators for dual-purpose plants. Realizing this potential depends on further development of aluminum alloy heat transfer tubing, demonstration of the effectiveness of reduced acid feed treatment at low temperature, and demonstration of the process design.

The long-term impact of a shift in emphasis to low-temperature aluminum evaporators appears favorable. Aluminum is not only a less expensive material than copper, but there are also much larger proven reserves so that aluminum cost is less likely to be affected by material shortages. The low-temperature plant has a factor of 5 less chemicals and no copper in its effluent, resulting in less environmental impact. The lower energy cost component for the low-temperature plant may become even more significant in the future because of rapidly increasing energy costs.

It is recommended that the next logical step in development of the low-temperature evaporator is to design, construct, and operate a small low-temperature evaporator pilot plant in the size range of 20,000 Gpd. Several development needs could be served by such a facility:

a) Process Design – the process design is unique in concept and would have to be demonstrated at some small-scale facility before prototype plants could be considered.

b) Materials — the pilot plant would provide a good opportunity to test an all-aluminum design.

c) Heat Transfer — additional heat transfer tests of aluminum alloy vertical tubes are needed at low-temperature.

d) Feed Treatment — the alternate feed treatment methods could be tested and evaluated.

HEAT COST

The cost of supplying process steam from the exhaust of a low-pressure turbine has been developed as a function of exhaust temperature. Data presented in this study is based on the use of a pressurized water reactor plant with all costs in terms of 1972 dollars. Heat costs from other power sources and for other time periods have been developed and are presented in Appendix A.

The unit cost of process steam presented in Fig. 3 is based on the incremental cost to the power plant of providing process heat. In developing this cost it is assumed that the evaporator would provide a reliable heat sink for the power plant. Any additional equipment required to assure a reliable heat sink, such as auxiliary condensers, if required, would have to be charged to the evaporator.

For the prototype 50-Mgd low-temperature plant it is anticipated that auxiliary condensers would be required for the entire heat load. This would add about 0.4¢/1000 gallons to the water cost. For a large-scale low-temperature plant with four or more trains, auxiliary condensers should only be necessary for about one-half the plant capacity and the impact on water cost would be reduced to about 0.2¢/1000 gallons.

The actual price of process heat for a specific site is a matter for negotiation between the participating utilities and cannot be predicted. It could be higher or lower than the cost derived here but would most likely be somewhat higher to include a profit margin.

Table 1 presents an illustration of the process heat cost allocation procedure. First, the total annual cost for the power-only plant is determined. The unit cost for power must then be such that the total annual revenue from power is equal to the total annual cost. The same

ORNL–DWG 73–4306

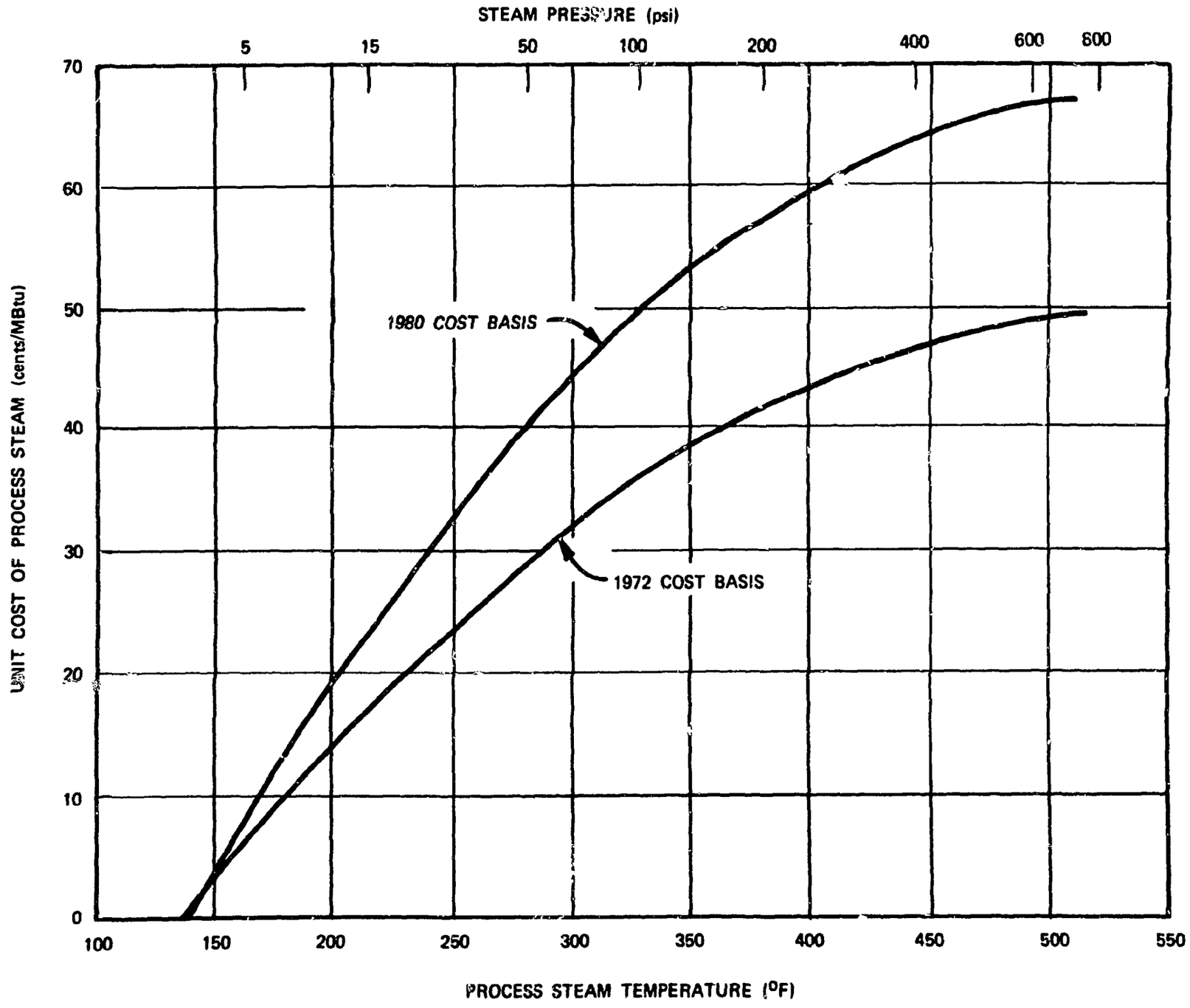


Fig. 3. Unit cost of process steam.

Table 1. Illustration of Process Heat Cost Allocation Procedure

	Power-Only Case	Dual-Purpose Case
Gross generator output, MW(e)	1168	938
Process heat temperature, °F		180
Nuclear island cost, $	166,187,000	166,187,000
Turbine-generator plant cost, $	121,247,000	105,214,000
Heat removal system cost, $	27,165,000	0
Total capital cost, $	314,599,000	271,401,000
Annual capital cost, $	44,044,000	37,996,000
Annual fuel cycle cost, $	17,250,000	17,250,000
Annual O&M cost, $	4,575,000	3,809,000
Total annual cost, $	65,879,000	59,065,000
Total power cost, mills/kWhr	7.85	7.85
Total annual revenue from power, $	65,879,000	52,906,000
Required annual revenue from process heat, $		6,158,000
Process heat, MBtu/hr		8,381
Process heat cost, ¢/MBtu		10.2

unit cost of power is applied to the dual-purpose plant but a lower annual revenue from power results because less power is produced. The total annual cost is somewhat less because the heat removal system cost is eliminated and the turbine-generator and annual O&M costs are less since the turbine-generator is smaller. The difference between total annual cost and total annual revenue from power is the required annual revenue from process heat. From this, the unit cost of process heat is derived.

MATERIALS EVALUATION

The need for cheaper and more plentiful heat transfer materials for use in evaporators is obvious. The cost of copper has increased substantially in recent years and, more important, the available resources may

not be adequate to support a large desalting industry. Aluminum appears to be a satisfactory alternative material, especially for low-temperature applications.

The low-temperature evaporator concept was discussed with both ALCOA and Reynolds Metals. A report of these meetings is included in Appendix B, along with the questionnaire which was forwarded to them in advance. Both companies indicated a greater confidence in the application of aluminum in evaporators now than they did during earlier discussions in 1970. The final letter from ALCOA includes specific alloy recommendations and cost data. Based on this recommendation, our estimates include $1.55/lb for aluminum tube bundles and $1.50/lb for aluminum structure and shell.

The aluminum companies quite naturally recommend and prefer all-aluminum construction. However, both companies agreed that concrete shell construction is compatible with aluminum. Methods for handling concrete to aluminum transitions and pipe penetrations were discussed.

Both aluminum companies recommend strongly against carbon steel shells or carbon steel components in the system. Carbon steel corrosion products not only foul the system but also set up corrosion cells on the aluminum surfaces. They recommend stainless steel where nonaluminum alloys must be used.

There are suitable and proven aluminum alloys available for each class of service — extrusion, structural fabrication, etc. Specific recommendations include 5054-O for shell material, 6061-T6 for tube sheets and structural shapes, and 3003, 5052, or 6063 for tubing.

HEAT TRANSFER TESTS

Experiments were conducted to determine the overall heat transfer characteristics in an axially-fluted, aluminum tube operating in the VTE (Vertical Tube Evaporator) mode in downflow while evaporating seawater. Overall heat transfer coefficients were also measured for a similar aluminum-brass tube, to use as a comparison. The emphasis during these tests was on heat transfer performance below 212°F, and data were obtained as low as 103°F.

In computing the overall heat transfer coefficient, the temperature of both the steam side and brine side were determined from absolute pressure measurements. In the case of the brine side this measurement was made in the lower brine chest. The brine side temperature must therefore be corrected for boiling point elevation (BPE). The BPE correction was based on the mean salt concentration on the brine side of the tube. The ΔT was also measured directly with quartz frequency thermometers installed in the loop, with the measured values agreeing well with the computed ΔT. However, primary reliance has been placed on the temperature difference derived from pressure measurements. The temperature difference corrected for BPE, and the rubber band outside surface area were used to compute the overall heat transfer coefficient which appears on the left hand ordinate of Fig. 4. The right hand ordinate is based on the total developed surface area is about 1.3 times the rubber band area, therefore, the right hand ordinate is equal to the left hand ordinate divided by 1.3.

The results of these experiments are shown in Fig. 4, where the overall heat transfer coefficient is plotted against the brine evaporating temperature. The numbers to the left and right of each data point are the rounded off ΔT (°F) and feed subcooling (°F), respectively. Also shown on this plot is a dashed line representing the results of an earlier test of an identical aluminum-brass tube (designated tube No. 115), and under similar test conditions.

The following observations can be made from this figure:

1. The overall heat transfer coefficient for the recently tested aluminum-brass tube (tube No. 115B) are 10–25% greater than the earlier tests of an identical tube (tube No. 115). There are two possible reasons for this. First, for the earlier tests, the "as received" tube appeared to be clean and was installed in the loop without further cleaning. For the recent tests, the tube appeared to be oily and dusty and was cleaned inside and out with soap and water, prior to installing in the loop. Second, the original tube (tube No. 115) was in the loop a considerably longer time than the recent tube and may have fouled more. At 210°F the difference between these curves, expressed as a fouling resistance, is about 0.00005 hr-ft^2-°F/Btu. The dissolved oxygen concentration in the makeup water for the steam generator was measured to be less than 50 ppb.

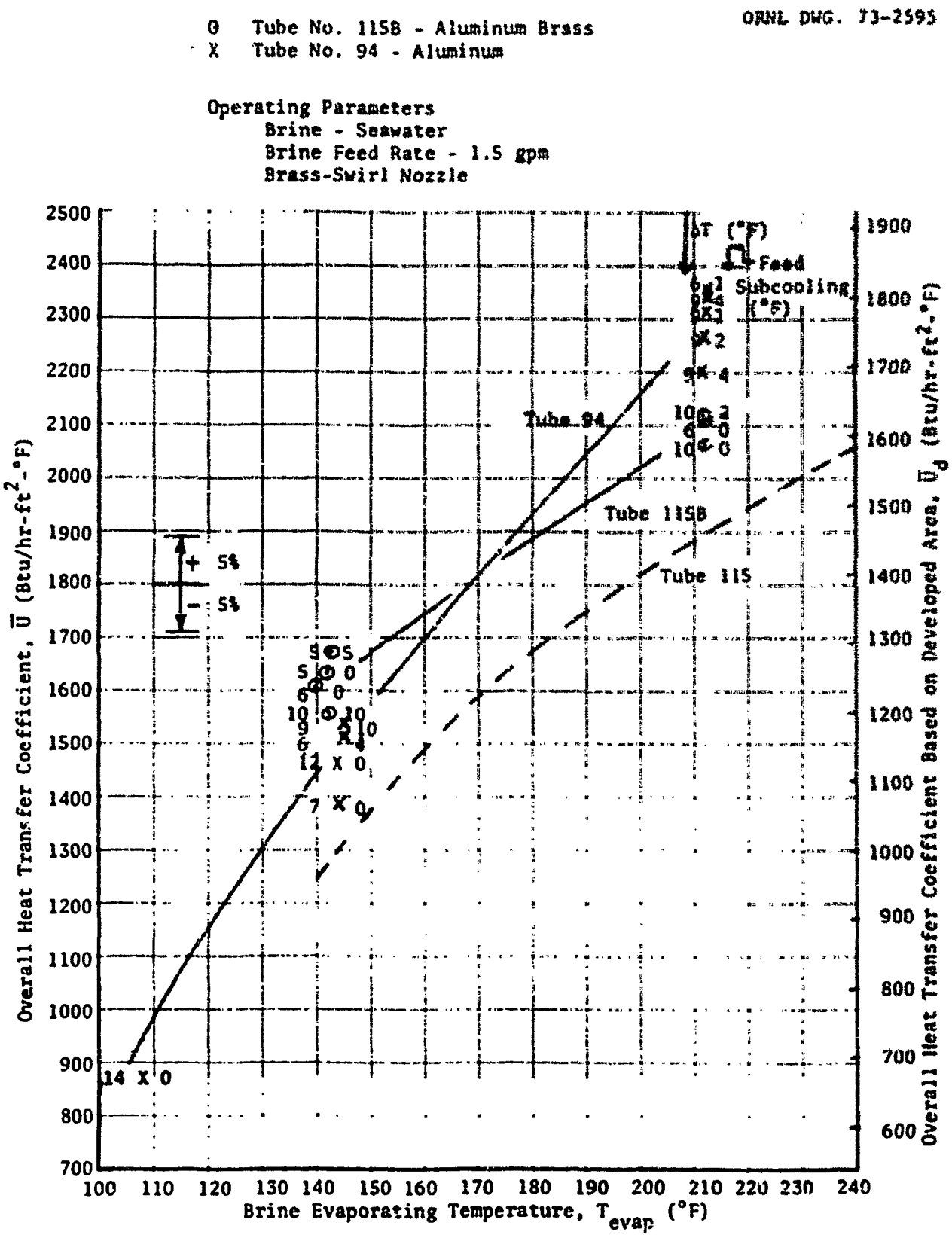


Fig. 4. Heat transfer characteristics of aluminum and aluminum-brass tubes evaporating seawater in VTE mode as function of evaporating temperature.

However, the dissolved oxygen concentration in the brine is known to be much higher and during one test at 212°F was measured to be about 880 ppb.

2. Note the error bands of plus and minus 5% on the side of the figure, which is the estimated precision of the loop. Note also that each cluster of data points, discriminated by tube and evaporating temperature, fall within this band. Since each cluster represents a range of ΔT from 5 to 11°F and of feed subcooling from 0 to 10°F, it has been concluded that within the precision of the loop the overall heat transfer coefficient (for the tubes and solutions tested) is not a significant function of ΔT and feed subcooling.

3. In general, the conclusions are that the same heat transfer correlations may be used for aluminum tubing as has been used in the past for copper alloy tubing. The slight differences in performance are within the expected error band.

FEED TREATMENT

For high-temperature evaporators (~250°F), a stoichiometric quantity of acid must be added to prevent alkaline scale due to accelerated precipitation at temperatures above ~185 to 190°F. For low-temperature evaporators (up to ~170°F), one should need only maintain the pH below the threshold at which precipitation would occur. The limiting values are shown on Fig. 5 for the temperature of interest. The values at 170°F are estimated based on data at 140°F and 212°F. For a concentration ratio of 2 to 3, a pH of about 7.3 would be adequate. Figure 6 shows the pH of aerated seawater as a function of acid addition. Normal acid treatment requires the addition of 110 to 115 ppm H_2SO_4. For the low-temperature evaporator it appears that 20 to 25 ppm would be sufficient to keep the pH below the threshold at which precipitation could occur. A factor of 5 less acid is used compared with conventional high-temperature plants.

Polyphosphate treatment has been used successfully in MSF evaporators up to about 180°F maximum temperature. This feed treatment method allows a slow buildup of "soft" scale which can normally be removed by acid cleaning every two or three months. A reduction in heat transfer performance does occur during the buildup period. This treatment method has not been tested on vertical tube evaporators with enhanced surfaces.

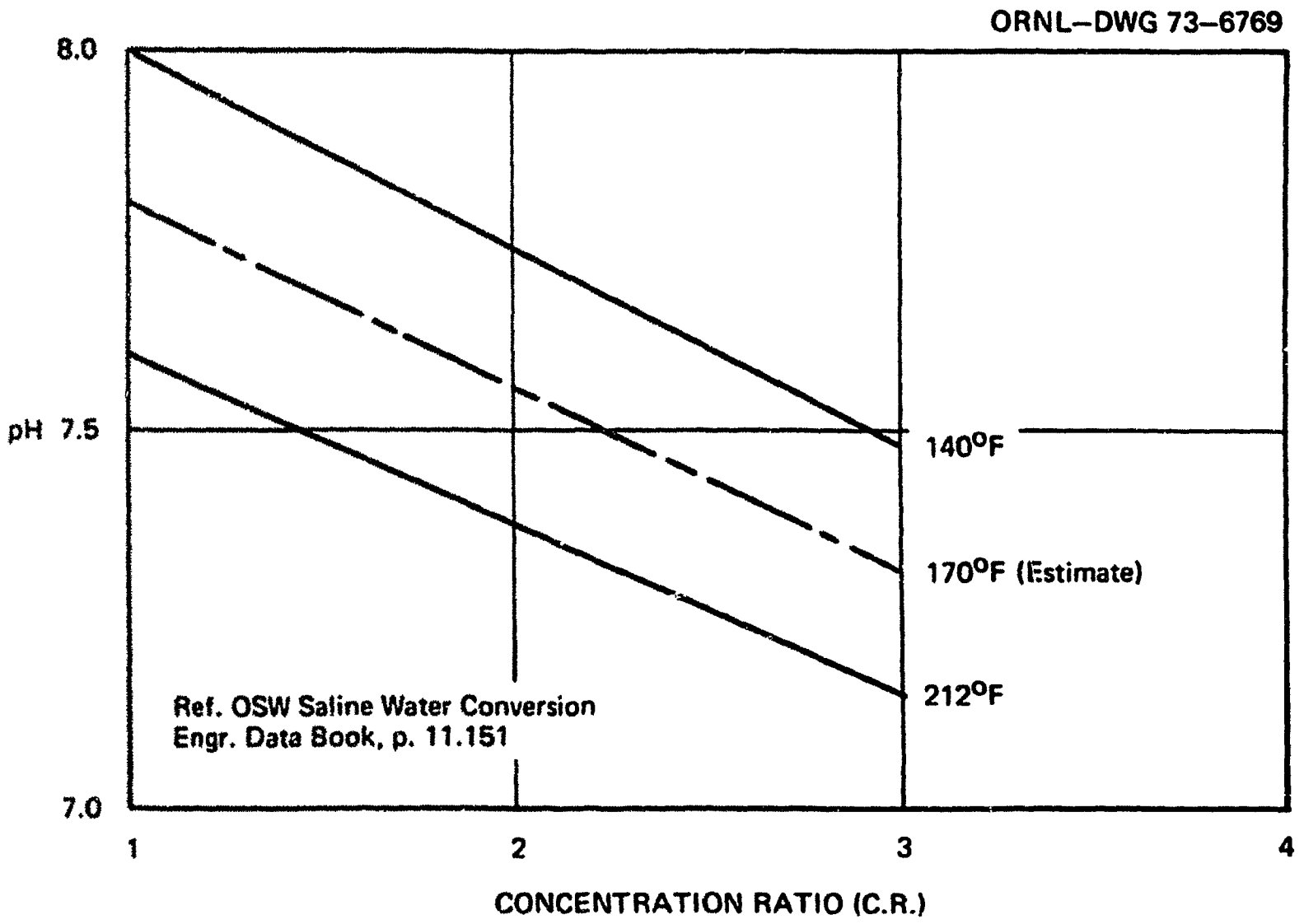


Fig. 5. Effect of concentration ratio on pH at which alkaline scale ($CaCO_3$) will precipitate from seawater.

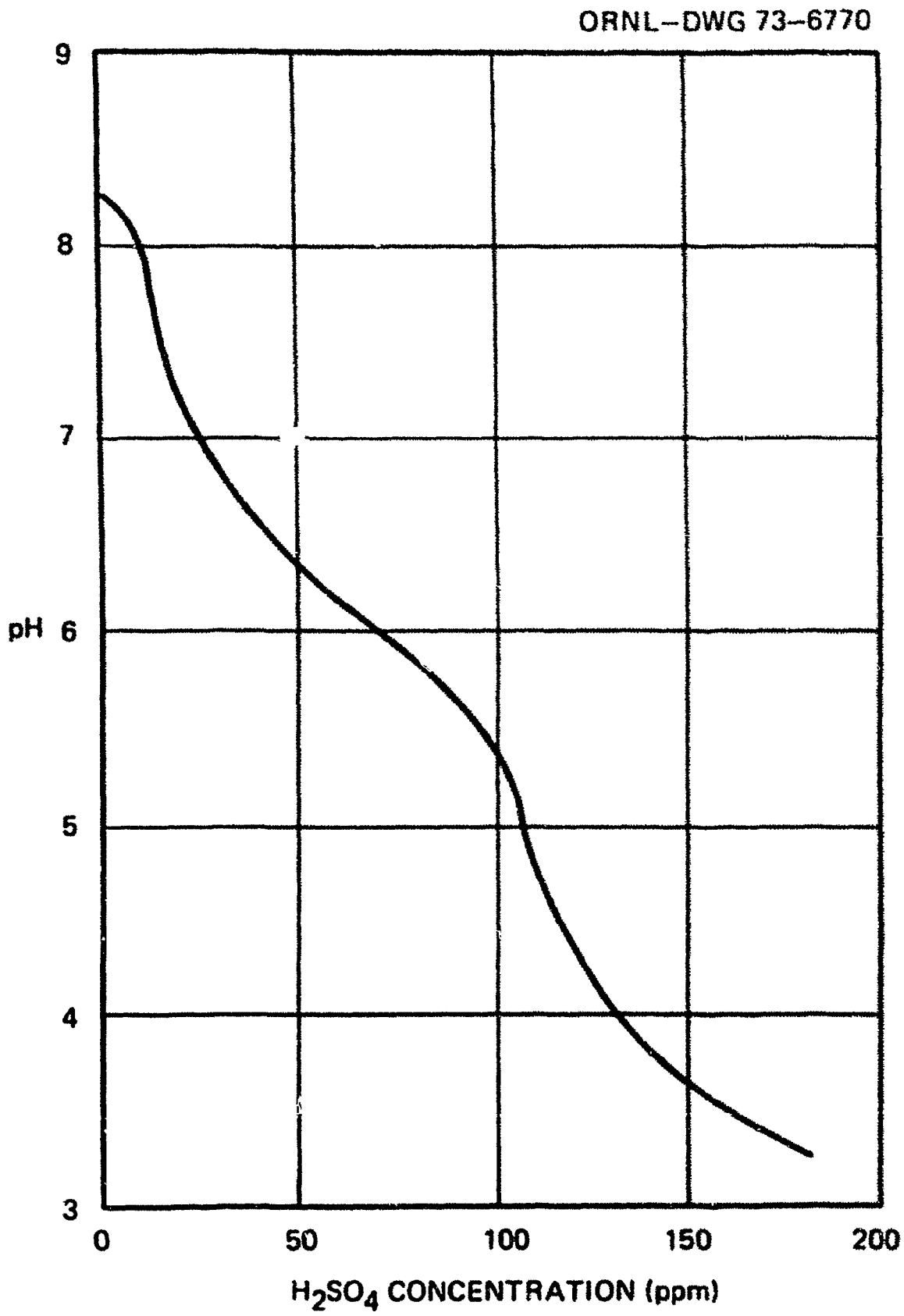


Fig. 6. pH of seawater as a function of H_2SO_4 concentration.

For the present cost estimates the partial acid feed treatment method has been assumed.

PROCESS DESCRIPTION

The process selected for this study is a forward feed, multiple effect, vertical tube evaporator with a single stage of horizontal tube flash feed heating per effect, sump to sump flashing of the brine feed, and recirculation to attain adequate effect feed.

A very brief study of a forward-backward feed VTE plant indicated no apparent economic advantage over the more conventional forward feed plant.

Figure 1 illustrates the process flow sheet developed for the low-temperature plant. The seawater passes first through the final condenser and part of the seawater is rejected to the sea. The remainder passes through the deaerator after acid addition and is then pumped to the feed heater bundle. The water is further heated as it flows through the feed heater tubes to the high-temperature end of the plant where it is fed to the first effect. In the tubes of the first effect, part of the brine evaporates as it is heated by steam from the power plant. Brine discharges from the tubes to the brine sump and flashes as it flows to the second effect brine sump. In the second effect brine is pumped from the sump to the brine chest to feed the effect tubes. Vapor from the first effect provides the heat source for the second effect. A portion of the vapor (about 15%) passes completely through the vertical tube bundle and supplies heat for the adjacent horizontal tube feed heater bundle. Venting takes place in the feed heater bundle. This process is continued down through the plant with vapor from the last effect going to the final condenser and the concentrated brine being rejected to the ocean. Condensed vapor from the vertical tubes and feed heaters is collected in a product trough which passes through the plant under the feed heater bundle.

PRELIMINARY OPTIMIZATION STUDY OF LOW-TEMPERATURE EVAPORATOR

The range of maximum evaporating temperature considered for the low-temperature evaporator was 125 to 170°F. A previous study[1] explored the possibility of a single-effect, very low-temperature evaporator (below

125°F). Above about 170°F there appears to be a change in technology required – i.e., (1) partial acid or phosphate feed treatments are no longer satisfactory, (2) the class of turbines designed for dry cooling towers do not operate above this temperature, and (3) aluminum alloy heat transfer surface is generally considered to be more reliable in this low-temperature range. The parametric survey was extended to 250°F in order to evaluate the future economic potential of aluminum heat transfer surface at higher temperatures. In the parametric survey at least three numbers of effects were evaluated at each of seven temperature levels. The results of this survey are shown on Fig. 7. This parametric study was based on preliminary estimates of material, capital, and operating costs so that the total costs do not represent an accurate absolute value but the relative cost comparison is reasonably accurate. Because of the importance of heat cost the costs are broken into heat cost, all other costs, and total costs. The break in the curves at 190°F results from higher feed treatment costs. As a result of this survey, the nine-effect, 170°F maximum temperature plant was selected for more detailed study.

REFERENCE PLANT DESIGN

Conceptual design sketches (Figs. 8 through 14) were prepared for the nine-effect, 170°F maximum temperature plant in sufficient detail to prepare a cost estimate for the plant. Within the limited scope of the the study it was not possible to develop the design in more detail. Table 2 lists design data for the evaporator plant.

There are two unique features of the design which merit description. First, the vertical tube bundle for each effect is composed of 10-ft wide units which may be shop fabricated and transported to the site as an assembly. All effect bundles are essentially identical so that the 25-Mgd train is composed of 108 of these identical vertical tube units. This should lead to a substantial savings in fabrication cost.

Second, a horizontal tube feed heater bundle which runs the length of the plant is located in the compartment between each pair of 10-ft wide vertical tube units. Product from the vertical tubes and the feed heater is collected in a trough under the feed heater bundle. Brine

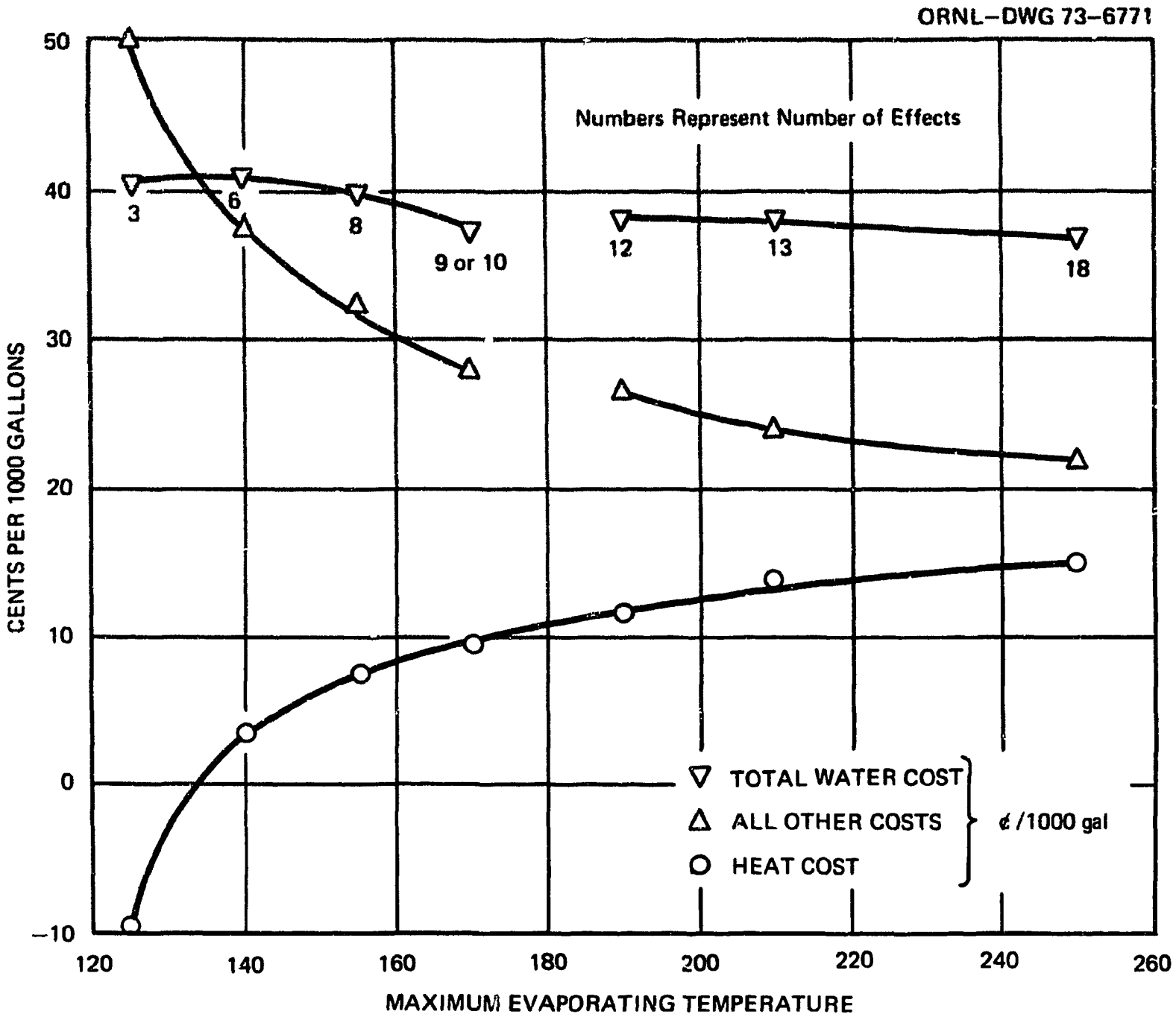


Fig. 7. Results of optimization study.

ORNL DWG 73-8384

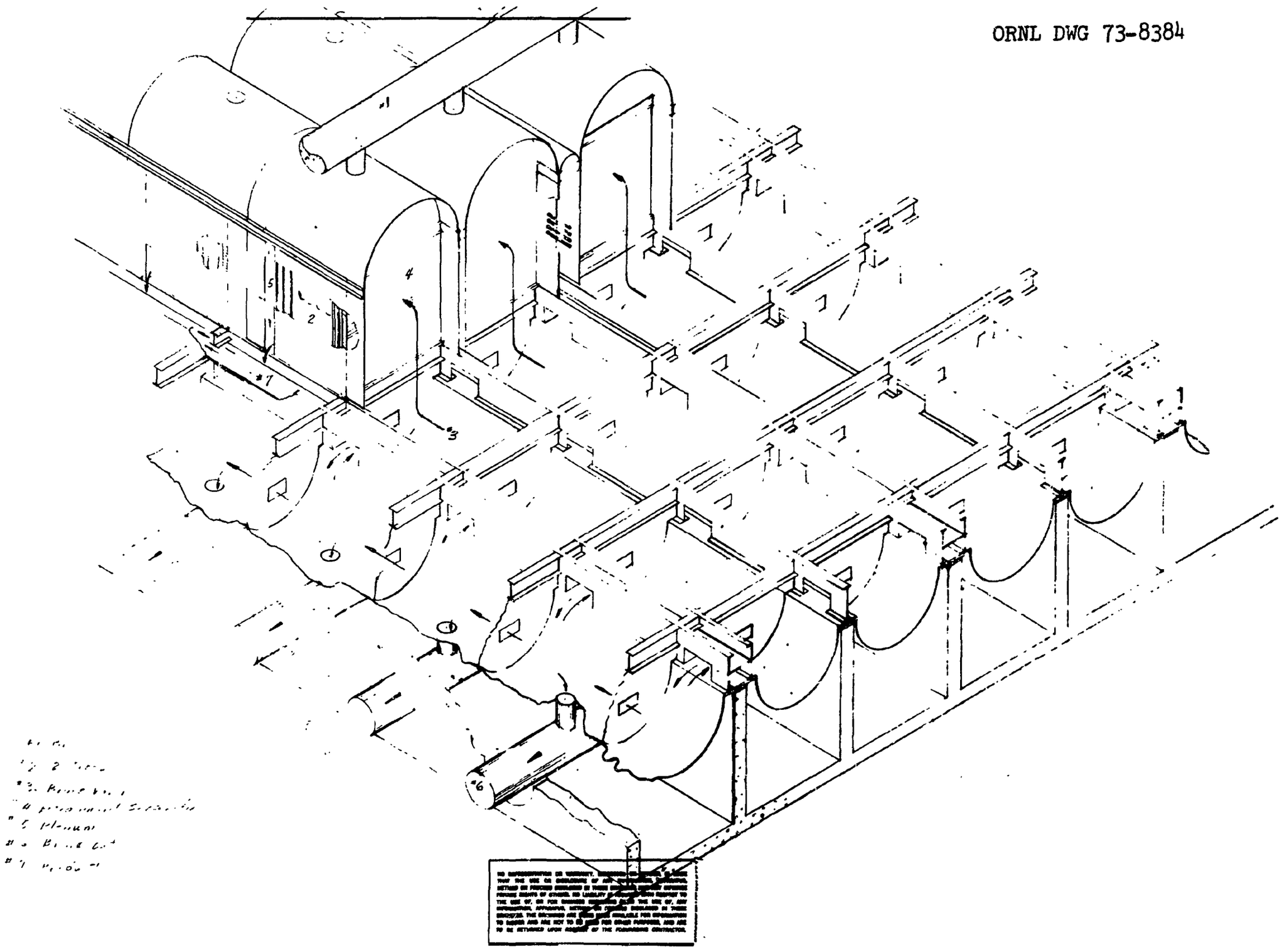


Fig. 8. Partial isometric of low-temperature evaporator.

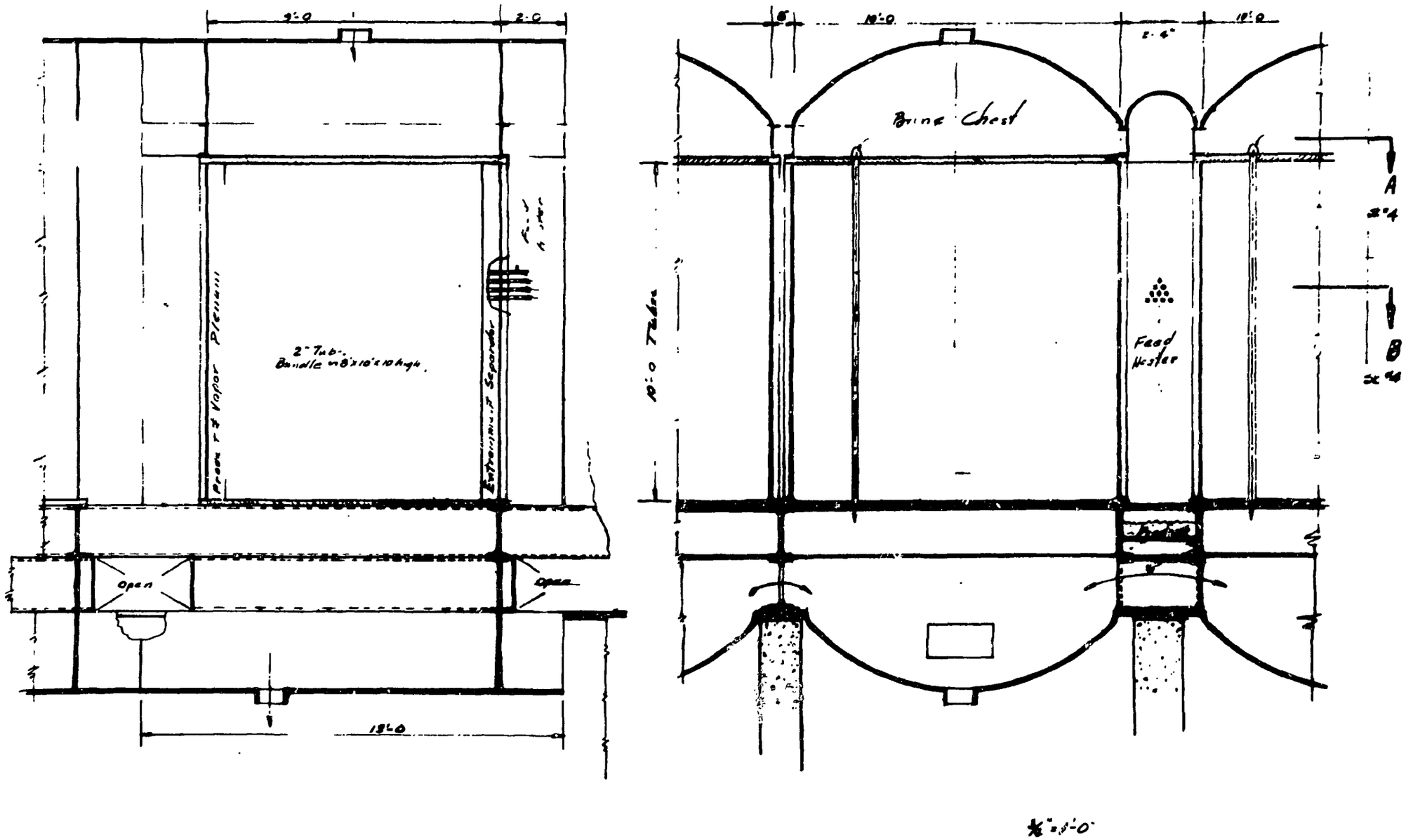


Fig. 9. Typical section through tube bundle.

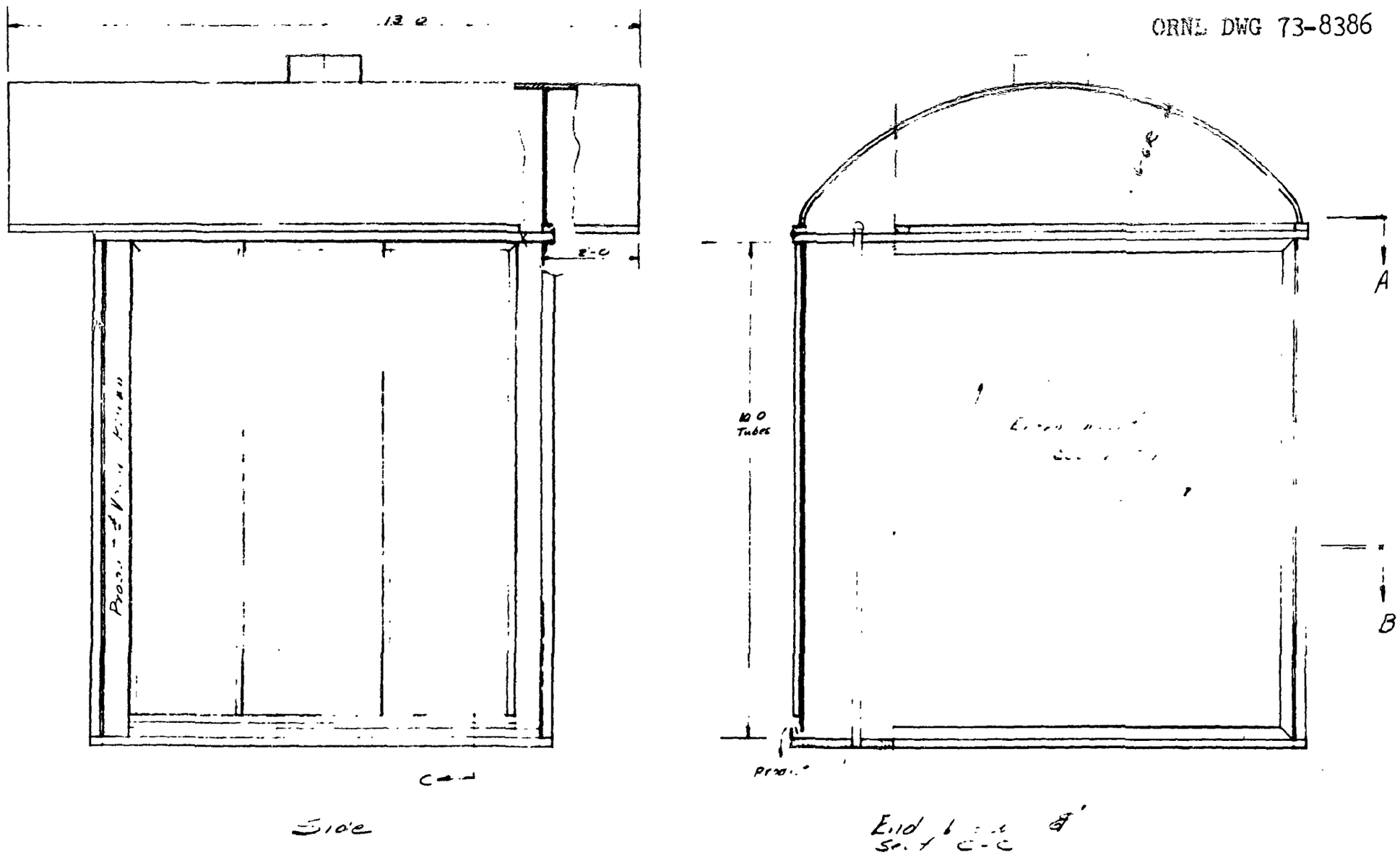


Fig. 10. Typical tube bundle unit - elevation.

ORNL DWG 73-8387

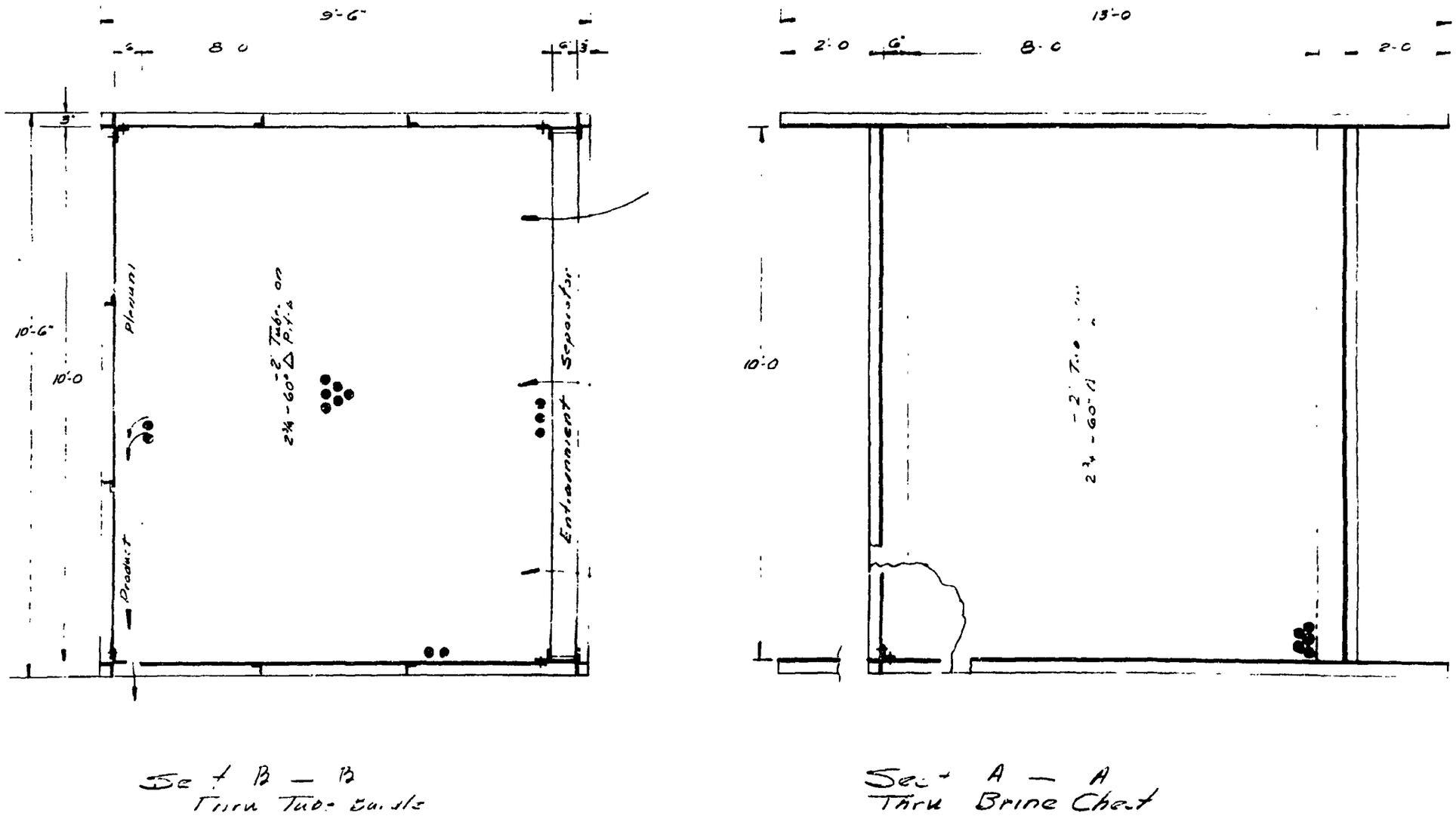


Fig. 11. Typical tube bundle unit, plan view, sections A and B.

ORNL DWG 73-8388

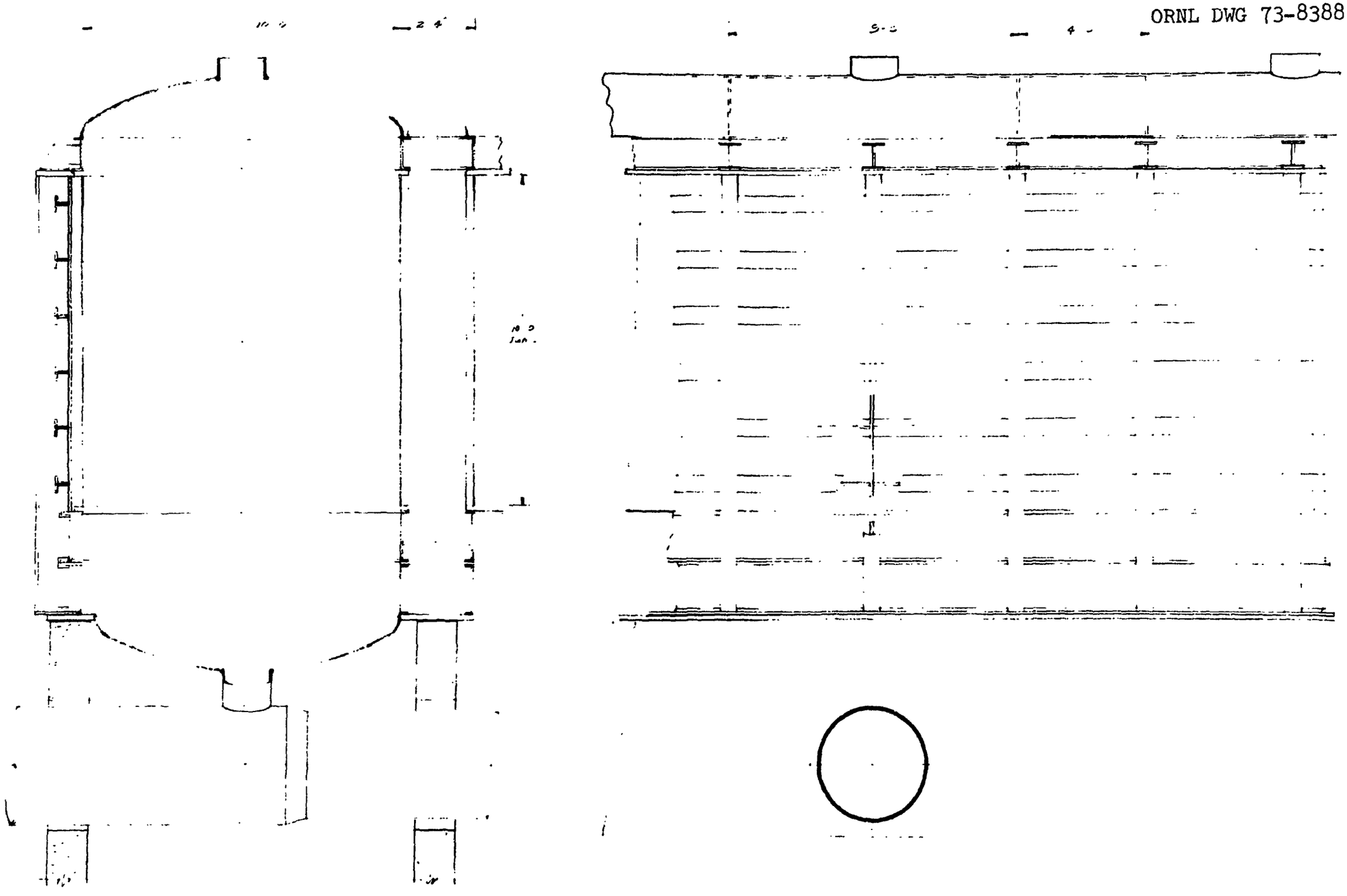


Fig. 12. Outside wall, typical section and side view.

ORNL DWG 73-8389

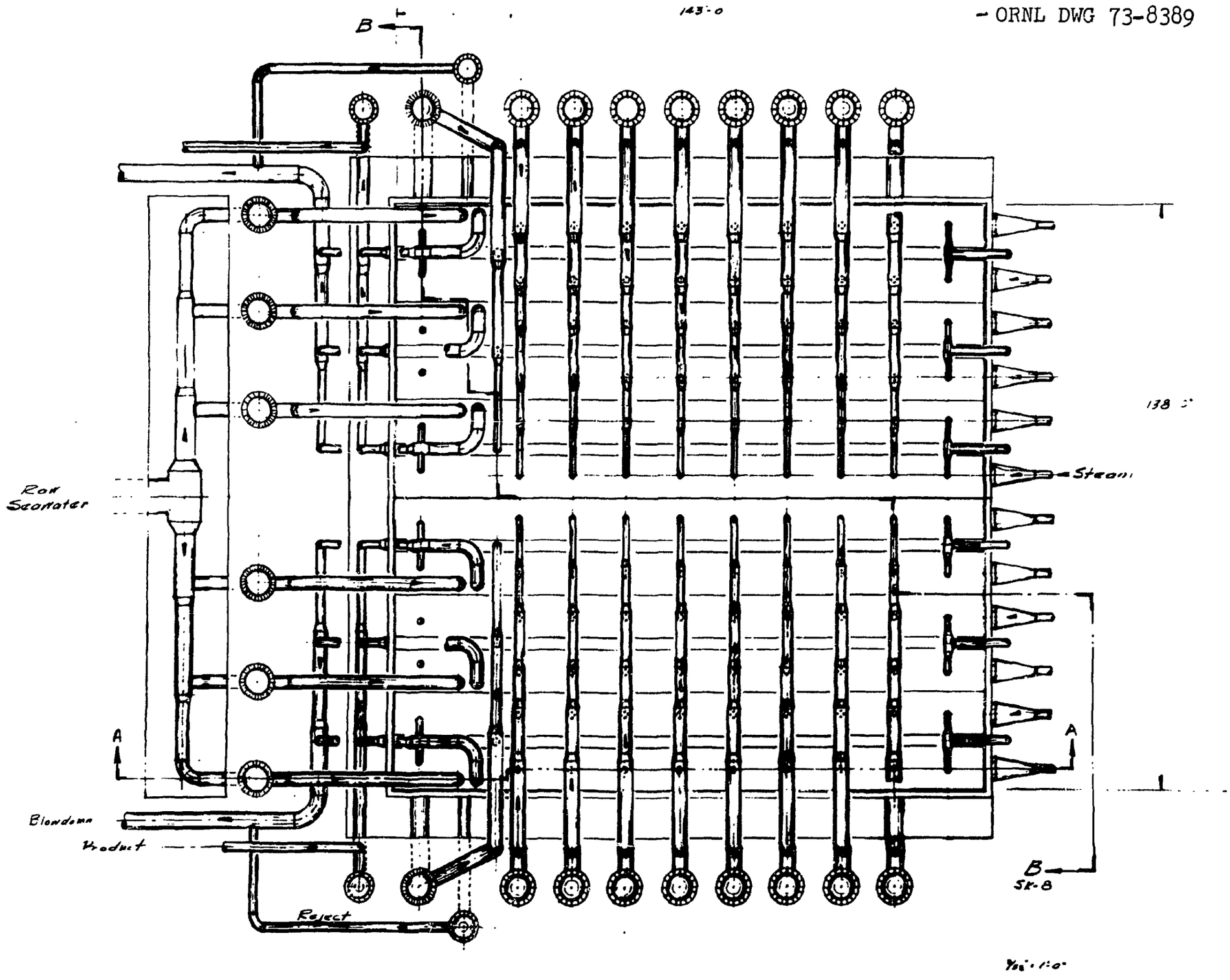
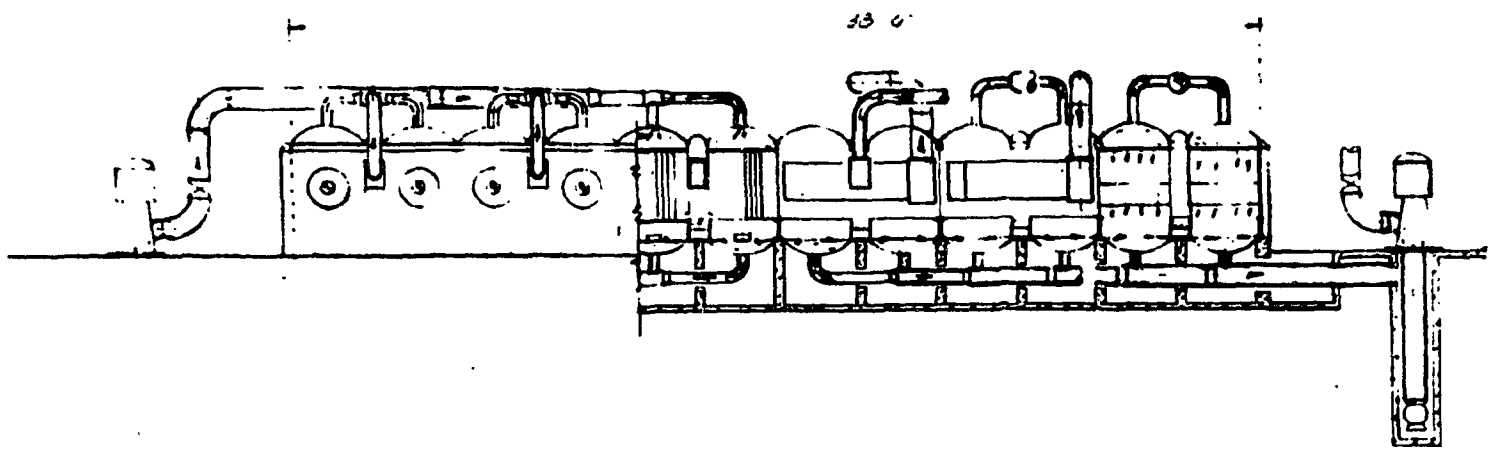
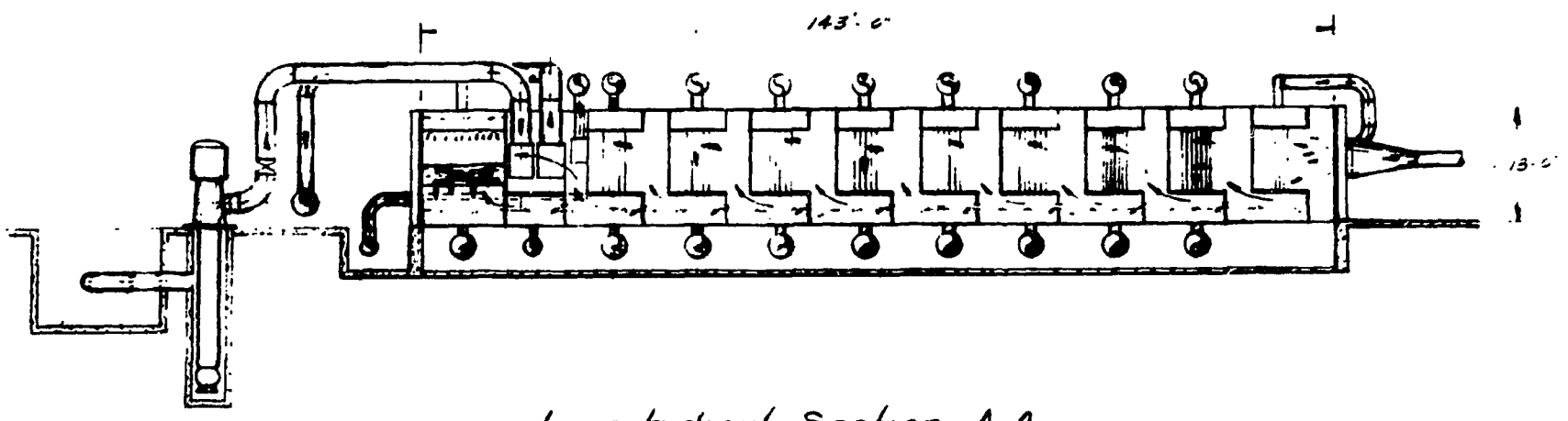


Fig. 13. Piping schematic.

ORNL DWG 73-8390



Transverse Section B-B



Longitudinal Section A-A

Fig. 14. Illustration of liquid and vapor flow paths.

Table 2. Low-Temperature Evaporator Reference Design Data

Train size, Mgd	25
Performance ratio, lb/1000 Btu	7.6
Concentration ratio	2
Number of effects	9
Number of MSF feed heater stages	9
Maximum brine temperature, °F	170
Steam required, lb/hr	1,150,000
Steam temperature, °F	175
Power required, MW	6.5
VTE data	
Tube material	Aluminum 5052
Tube type	Double fluted
Tube OD, in.	2
Tube wall thickness, in.	0.050
Tube length, ft	10
Number of tubes	238,650
Surface area, ft^2	1,590,000
Average overall $\bar{U}$, Btu/hr·ft^2·°F	1,024
Feed heater data	
Tube material	Aluminum 5052
Tube type	Spirally indented
Tube OD, in.	0.75
Tube wall thickness, in.	0.050
Tube length, ft	160
Number of tubes	6,060
Surface area, ft^2	195,500
Average overall $\bar{U}$, Btu/hr·ft^2·°F	684

flashes from sump to sump, providing a single stage of flashing per effect. The steam supply to the feed heater bundle passes through the two adjacent vertical tube units and through a plenum at the back of the vertical tube bundle to the feed heater. The total effect venting takes place in the feed heater bundle, thus eliminating duplication of venting. The single stage per effect is reasonably efficient because of low effect temperature difference of the design.

The large independent flash chamber generally associated with the VTE-MSF design is eliminated, resulting in a very substantial reduction in shell volume.

COST ESTIMATES

Table 3 shows the direct capital cost comparison with a conventional high-temperature VTE-MSF plant. The high-temperature VTE-MSF plant is based on the Envirogenics Company VTE/MSF Plant Design.[2] However, more conservative assumptions have been made in material selection, fouling factors, and heat transfer coefficients than those presented in the Envirogenics report. The modifications are consistent with those made for other studies at ORNL. The vertical tubes are aluminum brass rather than CDA alloy 194 (98 Cu-2 FE) and the final condenser is titanium rather than 70-30 cupro nickel. Fouling factors used for the feed heater and final condenser are 0.0003 and 0.0005 respectively rather than the Envirogenics estimates of 0.00015 and 0.0003. The ORNL design curve was used for vertical tube heat transfer coefficient rather than the Envirogenics estimate which is about 5% higher.

The cost estimates for each plant are based on ORNL evaporator design computer codes and are compared with consistent economic ground rules. Three shell costs are presented for the low-temperature evaporator. The steel and aluminum shell costs are based on the design layouts in this report. The conventional design steel shell is based on a design similar to the high-temperature plant where multiple stages per effect require a large independent MSF feed heater chamber. The low-temperature design with one stage per effect allows for a smaller, simpler shell design than the conventional high-temperature VTE-MSF. The reason for

Table 3. Direct Capital Cost Comparison

	High-Temperature Reference Plant 260°F – 15 Effects 33 Stages – 50 Mgd* $\$10^6$	Low-Temperature Reference Plant 170°F – 9 Effects 9 Stages – 50 Mgd* $\$10^6$
Effect tubes and sheets	4.55	5.36
Feed heater tubes and sheets	1.85	0.43
Vertical evaporator and feed heater shell	11.70	7.03 Steel 7.83 Aluminum 12.28 Conventional shell design – steel
Pumps and motors	2.58	4.20
Seawater intake	2.45	3.60
Valves and piping	1.74	3.02
Chemical capital	1.37	0.82
Instruments	1.03	0.61
Electrical	0.75	0.93
Deaerator	0.18	0.18
Condenser	0.54	0.67
Brine heater	0.17	0.21
Sites, buildings, and cranes	0.77	0.86
Total	29.68	27.92 Steel shell 28.72 Aluminum shell 33.17 Conventional design steel shell

*25-Mgd trains.

including the conventional design shell for the low-temperature plant is to show how much of the economic benefit is related to the shell design. The total indirect cost for each plant is 59% of the direct cost.

Although a concrete shell was not evaluated for this study, previous studies have shown the shell cost for concrete to be about 70 to 75% that of an equivalent steel shell. Using the 75% value, one could roughly estimate a concrete shell cost at $\$5.27 \times 10^6$ for this plant. This would

result in a total water cost estimate of 39.37¢/1000 gallons. Comparable values for steel and aluminum shell are presented in Table 4.

Table 4. Summary of Evaporator Plant Costs

	High-Temperature Reference Plant* 260°F – 15 Effects 33 Stages – 50 Mgd	Low-Temperature Reference Plant 170°F – 9 Effects 9 Stages – 50 Mgd		
		Steel Shell	Al. Shell	Conventional Design – Steel Shell
Direct capital cost, $\$10^6$	29.68	27.92	28.72	33.17
Total capital cost, $\$10^6$ (1.59 x direct)	47.19	44.39	45.66	52.74
Fixed charge rate, %	7	7	7	7
Annual cost – capital, $\$10^6$	3.30	3.11	3.20	3.69
Annual cost – operating, $\$10^6$	4.09	2.98	2.98	2.98
Total annual cost, $\$10^6$	7.39	6.09	6.18	6.67
Thousand gal/yr, 10^6	14.96	14.96	14.96	14.96
Water cost, ¢/1000 gal	49.40	40.71	41.31	44.59

*Based on Envirogenics 200-Mgd Conceptual Design Study.

Table 5 shows the annual operating cost for the high- and low-temperature plants. The three areas where differences appear are in chemical consumption, heat, and power. The required acid feed treatment for the low-temperature plant is about 20% that of the high-temperature plant. Heat cost is derived from the heat cost curve shown in Fig. 3 and power cost is based on 8 mills/kWhr. The low-temperature plant requires considerably more pumping power.

Table 5. Annual Operating Cost of Evaporator Plants

	High-Temperature Reference Plant 50 Mgd ($)	Low-Temperature Reference Plant 50 Mgd ($)
Operating labor	159,000	159,000
Maintenance labor	111,000	111,000
General and administrative	141,000	141,000
Operating supplies and maintenance materials	85,000	85,000
Chemical consumption	488,000	98,000
Other costs (interim replacement)	180,000	180,000
Heat	2,470,000	1,460,000
Power	460,000	747,000
Total	4,094,000	2,981,000

A summary of evaporator plant costs is shown in Table 4. Water cost for the high-temperature plant is about 20% more than the low-temperature plant with either steel or aluminum shell. The aluminum shell plant, although slightly higher in cost than the steel shell, is considered the best design because of problems of compatibility with aluminum tubes and a steel shell.

REFERENCES

1. John E. Jones, Jr., and Truman D. Anderson, A Proposed Dual-Purpose Nuclear Desalting Plant Using Single-Effect Distillation, ORNL-TM-2964, May 1970.
2. Envirogenics Company, 200 MGD Desalination Plant and VTE/MSF Module Conceptual Design, Volumes I through IV, Report No. F1593, November 1970.

APPENDIX A

Process Heat Cost as a Function of Temperature for Nuclear and Fossil Dual-Purpose Plants

L. C. Fuller

OBJECTIVE

The objective of this five-part study was to estimate the cost of process steam from various heat sources and at various process temperatures. The first four parts provide cost using exhaust steam from a low-pressure turbine and relate those costs to exhaust temperature. Part 1 covered a turbine plant cycle suitable for a pressurized-water reactor with the entire low-pressure turbine exhaust steam flow to be used for process heating. Part 2 also covered a PWR cycle, but it differed from Part 1 in that two of the double-flow low-pressure turbines condensed steam while one double-flow unit supplied steam to process. Part 3 covered a turbine plant cycle suitable for a high temperature gas reactor with the entire low-pressure turbine exhaust steam flow to process heating. Part 4 covered a turbine plant cycle for a subcritical-pressure coal-fired plant with the entire low-pressure turbine exhaust steam flow to process heating. Part 5 covered a dual-purpose plant requiring steam at three pressures (throttle, crossover, and low-pressure turbine extraction) for a pressurized-water reactor, and was intended to model a hypothetical chemical plant process steam supply.

For each of Parts 1 through 4, three existing ORNL computer codes were used where applicable and one short bookkeeping code was written to determine process heat cost and produce tables of results. Part 5 was done with a combination of existing ORNL codes and by hand calculation.

PART 1. PRESSURIZED WATER REACTOR – FULL FLOW TO PROCESS

Calculational Procedure

Three existing computer codes and one short "bookkeeping" code were used to obtain process heat cost. Briefly, ORCENT[1] was used to make a heat balance for each cycle; CONCEPT II (Ref. 2) was used to determine breakdown of capital cost to nuclear island, turbine-generator plant, and heat removal system; and ORCOST[3] was used to estimate turbine-generator plant cost for various back pressures. A PROCESS HEAT COST code was used to combine the results from the above three codes, add in operation and

BLANK PAGE

maintenance costs and fuel-cycle costs (derived from Ref. 4 output), and finally calculate process heat cost.

Reference Power-Only Cycle

Figure 1 process flow sheet shows a single-purpose electric power plant with turbine throttle steam conditions of 515.3°F, 780 psia (dry and saturated). Gross generator output is 1168 MW(e) for the 1800 rpm tandem-compound turbine-generator. There are two turbine sections in series consisting of a double-flow high-pressure section and a six-flow low-pressure section (three double-flow low-pressure sections). For the reference cycle steam is condensed at a pressure of 2 in. HgA. This pressure results from the assumption of once-through cooling.

Three moisture removal stages are included in the low-pressure turbine. An external moisture separator (assumed 100% effective) and two stages of steam reheat each with 25°F terminal temperature difference are included between the high- and low-pressure sections of the turbine. The cycle employs seven stages of feed water heating with three stages supplied from the high-pressure section of the turbine and four stages supplied from the low-pressure section. For simplicity in this analysis, all feed water heaters are of the closed type with 5°F terminal temperature difference, flashed drains, and a 10°F approach in the drain cooler. The turbine driven feed water pump is located before feed water heater No. 1.

Dual-Purpose Cycles

For the dual-purpose cycles, steam is condensed at progressively higher pressures with decreasing plant electrical output. The condensing temperatures and pressures for the eight dual-purpose cases are shown below. As single stages are removed from the low-pressure turbine, the total number of feed water heating stages is reduced, as are the number of moisture removal stages. These are also listed below:

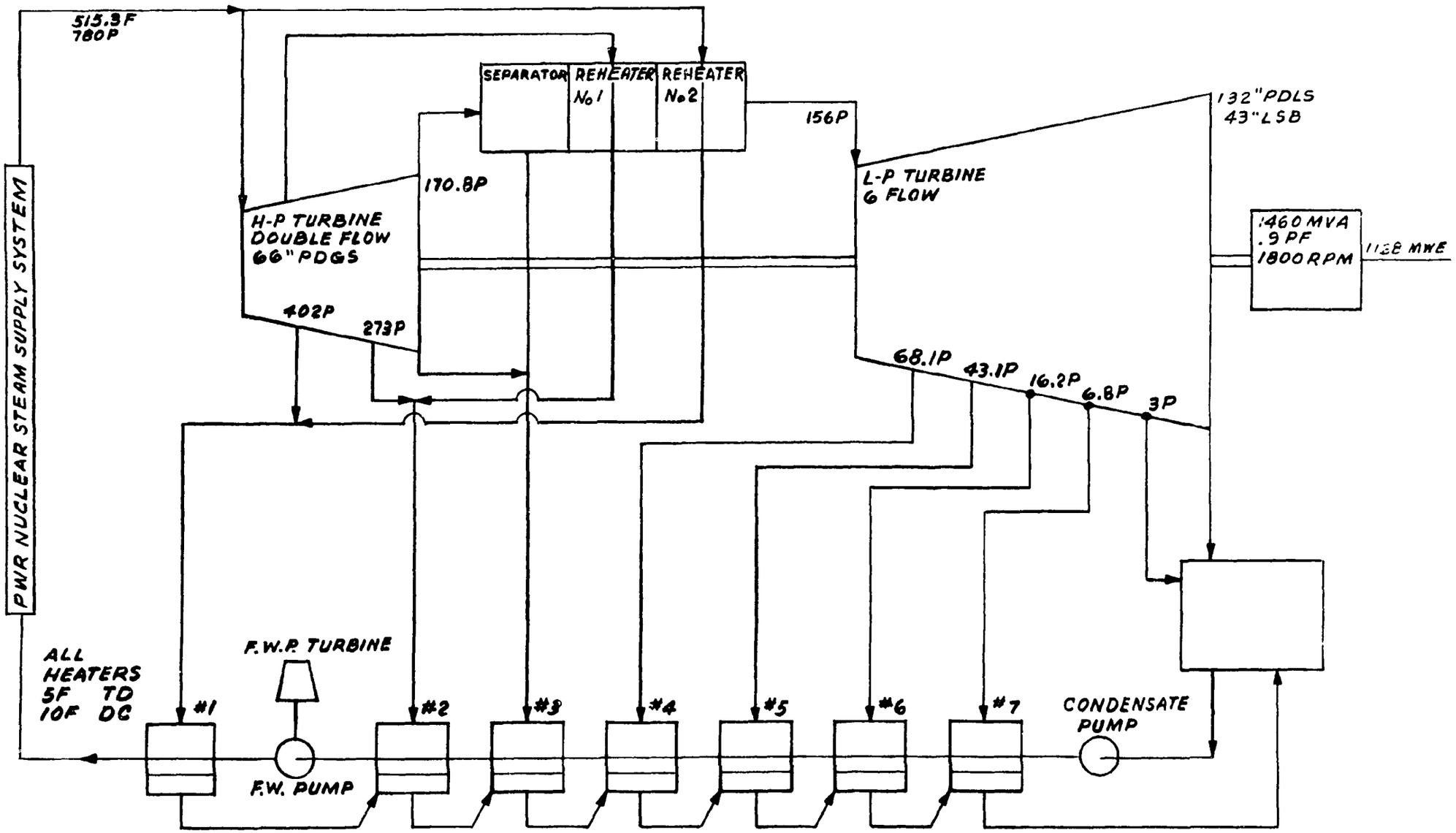


Fig. 1. Process flowsheet for reference power-only cycle.

Temperature °F	Pressure	Feed Water Heaters	Moisture Removal Stages
120	3.45 in. HgA	7	2
140	5.88 in. HgA	6	1
160	9.65 in. HgA	6	1
180	7.51 psia	5	1
200	11.53 psia	5	1
220	17.19 psia	5	none
240	24.97 psia	4	none
267	40.00 psia	4	none

Results

Tabular results for the full-flow PWR cases are shown in Tables 1 and 2 for 1972 and 1978 operation respectively. Figure 2 shows gross generator output as a function of steam temperature. Figure 3 shows the quantity of process heat in millions of Btu/hr as a function of steam temperature. Figure 4 is a plot of process heat cost in cents per million Btu as a function of steam temperature for both 1972 and 1978 operation.

Calculations

Cycle Calculations

ORCENT was used to make heat balance calculations for Parts 1 and 2 of this study. ORCENT is a computer program for saturated and low superheat steam turbine cycle analysis. The program performs maximum guaranteed rating calculations for typical light-water reactor power plants.

Nine ORCENT heat balances were made at 3423 MW(t) to determine gross generator output, turbine exhaust steam flow, feed pump turbine steam flow, feed pump turbine exhaust enthalpy, condenser outlet enthalpy, feed water heater drain flow (flashed to the condenser), heater drain enthalpy, and low-pressure turbine used energy end point.

Economic Calculations

In order to determine the fractions of total plant capital cost attributable to nuclear island, turbine-generator plant, and heat removal

Table 1

PRESSURIZED WATER REACTOR 3423 MWT
FULL BACK PRESSURE CASES
THROTTLE STEAM CONDITIONS 515.3 F 780 PSIA DRY AND SATURATED
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1972.0
YEAR START OF CONSTRUCTION 1965.5

GROSS GEN. OUTPUT, MWE	1168.	1136.	1062.	1001.	938.	882.	823.	760.	675.
GROSS EFFICIENCY	0.3412	0.3319	0.3103	0.2924	0.2740	0.2577	0.2404	0.2220	0.1972
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	8246000.	8538000.	8713000.	8845000.	9072000.	9226000.	9387000.	9553000.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	220.	240.	267.
NUCLEAR ISLAND COST, $	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.
T-G PLANT COST, $	98970640.	97236400.	93151168.	89759632.	86213968.	82976608.	79508000.	76039408.	71376064.
HEAT REM. SYS. COST, $	22174640.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	256799760.	232890880.	228805648.	225414112.	221868448.	218631088.	215162480.	211693888.	207030544.
ANNUAL CAP. COST, $	35951952.	32604720.	32032784.	31557968.	31061568.	30608336.	30122736.	29637136.	28984272.
ANNUAL FUEL COST, $	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.
ANNUAL O&M COST, $	3831698.	3280563.	3246576.	3218560.	3189625.	3163906.	3136808.	3107874.	3068835.
TOTAL ANN. COST, $	54665952.	50767584.	50161664.	49658832.	49133488.	48654544.	48141840.	47627312.	46935408.
POWER COST, MILLS/KWHR	6.52	6.52	6.52	6.52	6.52	6.52	6.52	6.52	6.52
REVENUE FROM POWER, $/YR	54665968.	53168272.	49704832.	46849824.	43901232.	41280256.	38518880.	35570288.	31592032.
REQD REV. PROC. HT.$/YR	0.	-2400688.	456832.	2809008.	5232256.	7374288.	9622960.	12057024.	15343376.
PROC. HEAT, MILL. BTU/HR	0.	7742.	7997.	8209.	8381.	8623.	8828.	9052.	9337.
HEAT COST, CTS/MILL. BTU	0.0	-4.3	0.8	4.8	8.7	11.9	15.2	18.5	22.9

Table 2

PRESSURIZED WATER REACTOR 3423 MWT
FULL BACK PRESSURE CASES
THROTTLE STEAM CONDITIONS 515.3 F 780 PSIA DRY AND SATURATED
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1978.0
YEAR START OF CONSTRUCTION 1971.5

GROSS GEN. OUTPUT, MWE	1168.	1136.	1062.	1001.	938.	882.	823.	760.	675.
GROSS EFFICIENCY	0.3412	0.3319	0.3103	0.2924	0.2740	0.2577	0.2404	0.2220	0.1972
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	8246000.	8538000.	8713000.	8845000.	9072000.	9226000.	9387000.	9553000.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	220.	240.	267.
NUCLEAR ISLAND COST, $	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.
T-G PLANT COST, $	121246768.	119050016.	113962752.	109684768.	105214176.	102208048.	96812464.	92495984.	86599360.
HEAT REM. SYS. COST, $	27165664.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	314599424.	285237248.	280150016.	275872000.	271401216.	268395312.	262959728.	258683248.	252786624.
ANNUAL CAP. COST, $	44043904.	39933200.	39220992.	38622080.	37996160.	37575328.	36819952.	36215648.	35390112.
ANNUAL FUEL COST, $	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.
ANNUAL O&M COST, $	4575241.	3917158.	3876576.	3843123.	3808574.	3777863.	3745508.	3710958.	3664344.
TOTAL ANN. COST, $	65879152.	61110368.	60357584.	59725216.	59064736.	58613200.	57825472.	57186608.	56314464.
POWER COST, MILLS/KWHR	7.85	7.85	7.85	7.85	7.85	7.85	7.85	7.85	7.85
REVENUE FROM POWER, $/YR	65879152.	64074272.	59900416.	56459792.	52906384.	49747808.	46419984.	42866576.	38072288.
REQD REV. PROC. HT.$/YR	0.	-2963904.	457168.	3265424.	6158352.	8865392.	11405488.	14320032.	18242176.
PROC. HEAT, MILL. BTU/HR	0.	7742.	7997.	8209.	8381.	8623.	8828.	9052.	9337.
HEAT COST, CTS/MILL. BTU	0.0	-5.3	0.8	5.5	10.2	14.3	18.0	22.0	27.2

ORNL DWG. 73-2903

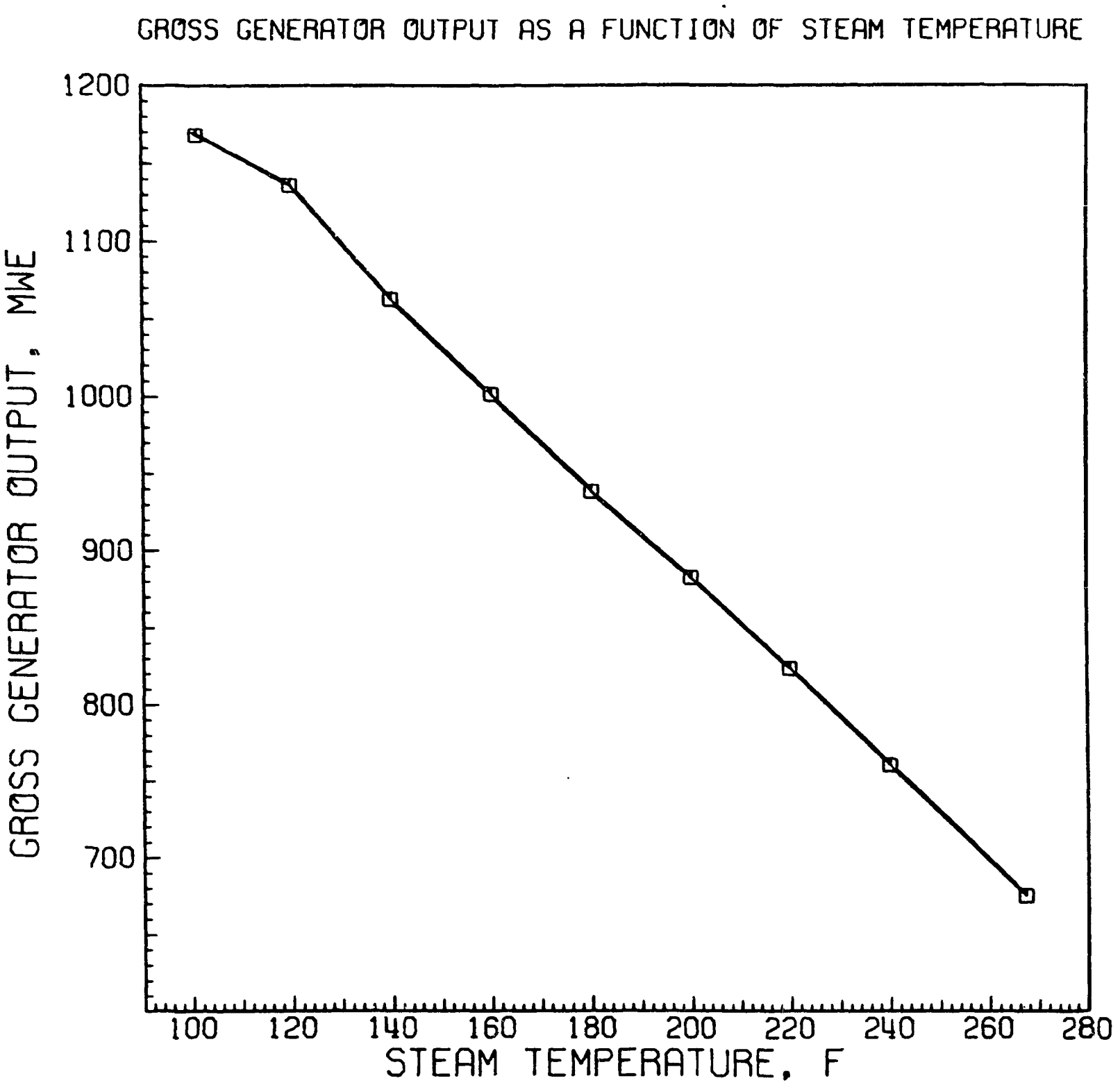


Fig. 2. Gross generator output as a function of steam temperature for PWR full flow cases.

ORNL DWG. 73-2904

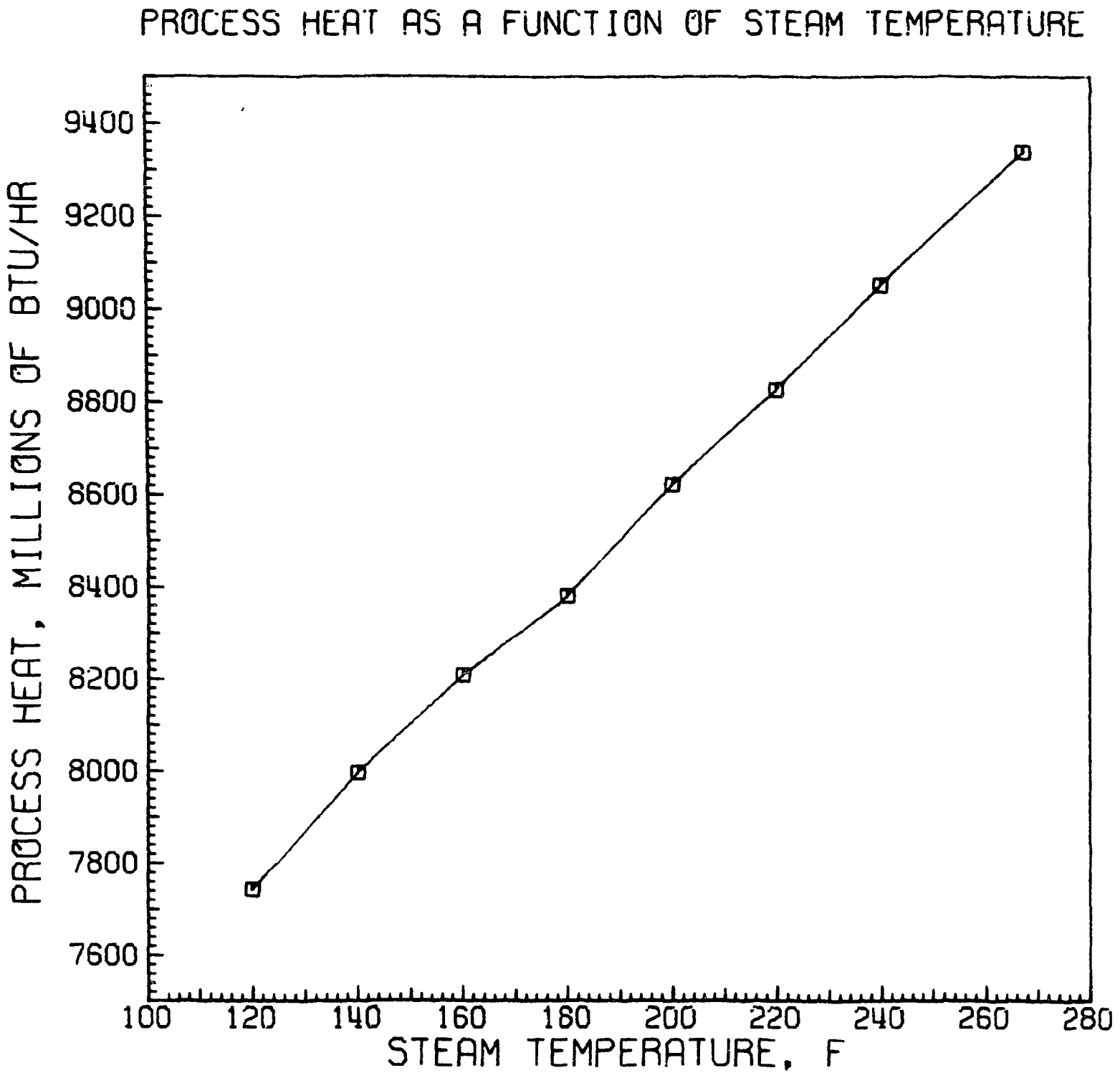


Fig. 3. Process heat as a function of steam temperature for PWR full flow cases.

ORNL DWG. 73-2905

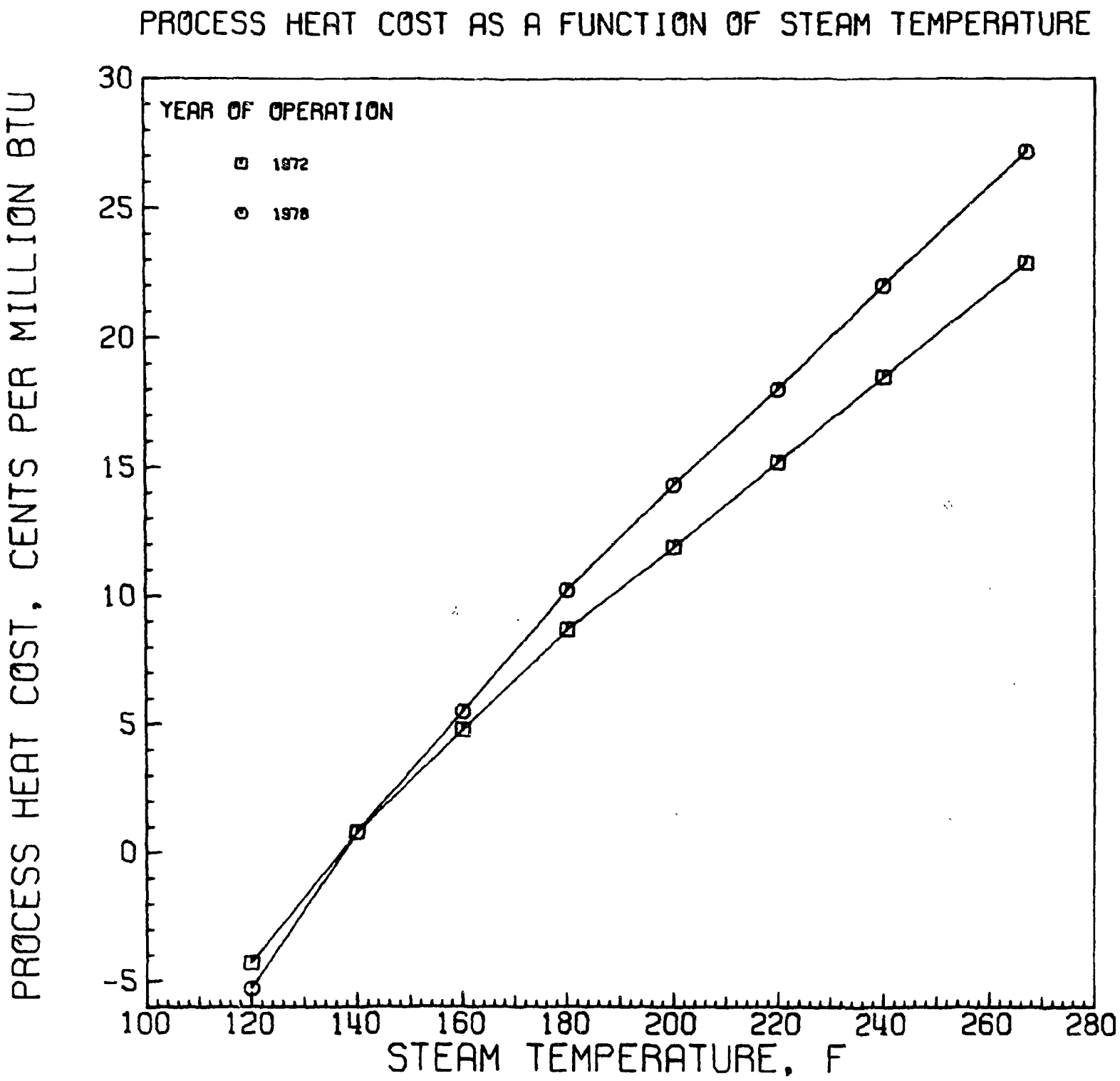


Fig. 4. Process heat cost as a function of temperature for PWR full flow cases.

system, the CONCEPT II code was used. Runs were made at 1168 and 675 MW(e), the highest and lowest gross generator output obtained from ORCENT heat balances with 3423 MW(t) delivered to the turbine-generator plant.

CONCEPT II provides conceptual capital cost estimates for nuclear and fossil-fueled plants as a function of plant type, size, location, and date of operation. CONCEPT II contains a detailed cost model for a PWR plant and can be used to break down costs to the four-digit level using the USAEC accounting system described in NUS-531 (Ref. 5). Using average values from the two runs, nuclear island capital costs were taken to be 52.83%, turbine-generator plant capital costs 38.54%, and heat removal system capital costs 8.63%.

The runs assumed PWRs constructed in Los Angeles with 1972 start of construction and 1978 commercial operation. Forty-hour work week for construction labor was assumed. Interest during construction was assumed to be 7% with no escalation during the construction period.

CONCEPT II was used rather than ORCOST since the four-digit account level is required to make any meaningful breakdown of nuclear island, turbine-generator plant, and heat removal system capital cost fractions. ORCOST does not break down costs to the four-digit level.

It was assumed in this study that turbine-generator plant capital cost was a function of gross electrical output and was independent of low-pressure turbine back pressure. A turbine condensing steam at 2 in. HgA and a turbine with 40 psia back pressure would have the same capital cost if the gross electrical output were held constant.

Using the above assumption, turbine-generator plant capital costs were determined for nine plant sizes, 1168 MW(e) through 675 MW(e), at two different years of commercial operation, 1972 and 1978, using the ORCOST code.

Results from ORCOST and CONCEPT II are in good agreement. The assumptions used for the CONCEPT II runs held true for ORCOST runs. All other input data not specified corresponds to the default values built into ORCOST.

The PROCESS HEAT COST code does little more than bookkeeping taking the results of ORCENT, CONCEPT II, and ORCOST, digesting them and calculating process heat cost. Runs were made for two different years of

commercial operation, 1972 and 1978, and nine steam pressures from 2 in. HgA through 40 psia. These runs required as input the results of nine ORCENT runs, two CONCEPT II runs, 18 ORCOST runs, and points from a plot derived from an ORSAC run.

Operating and maintenance cost for the base case (power-only) using 1971.5 dollars were estimated and input as OM. O&M costs were estimated with reference to an internal ORNL report entitled "Operating and Maintenance Cost Estimates for Steam Electric Power Plants." O&M costs were escalated at 3% per year compounded annually from 1971.5 to the year of commercial operation.

A plant capacity factor of 0.82 was assumed and input through the namelist as CF. This is consistent with the value used in the ORSAC run from which fuel cycle costs were obtained.

The fixed charge rate for private financing was input as 14%. A trial run at 15% to determine the sensitivity to this variable showed an increase in process heat cost of about one cent per million Btu for the highest back pressure cases, much smaller amounts for low temperature cases.

The breakdown of O&M costs between nuclear island, turbine-generator plant, and heat removal system for the base case was assumed to be 72%, 14%, and 14%, respectively. For the full-flow cases in which steam was sold, heat removal system O&M costs were assumed to be zero. Turbine-generator system O&M costs were assumed to vary with gross electrical output. A change in this assumption from 72, 14, 14, to 70, 10, 20 for nuclear island, turbine-generator plant, and heat removal system, respectively, resulted in a decrease of about 1 1/4 cents per million Btu in process heat cost.

Fuel cycle costs in 1970 dollars were taken from Fig. 5. Figure 5 was derived from the ORNL computer code ORSAC.[4] The minimum in Fig. 5 results from the assumptions used in ORSAC for uranium and plutonium price as a function of time. The turnup in the curve past 1978 represents an increase in real price and not inflation. No inflation is included in the curve. All values are expressed in constant dollars. Costs are levelized over an assumed 30-year plant life. The initial fuel loading

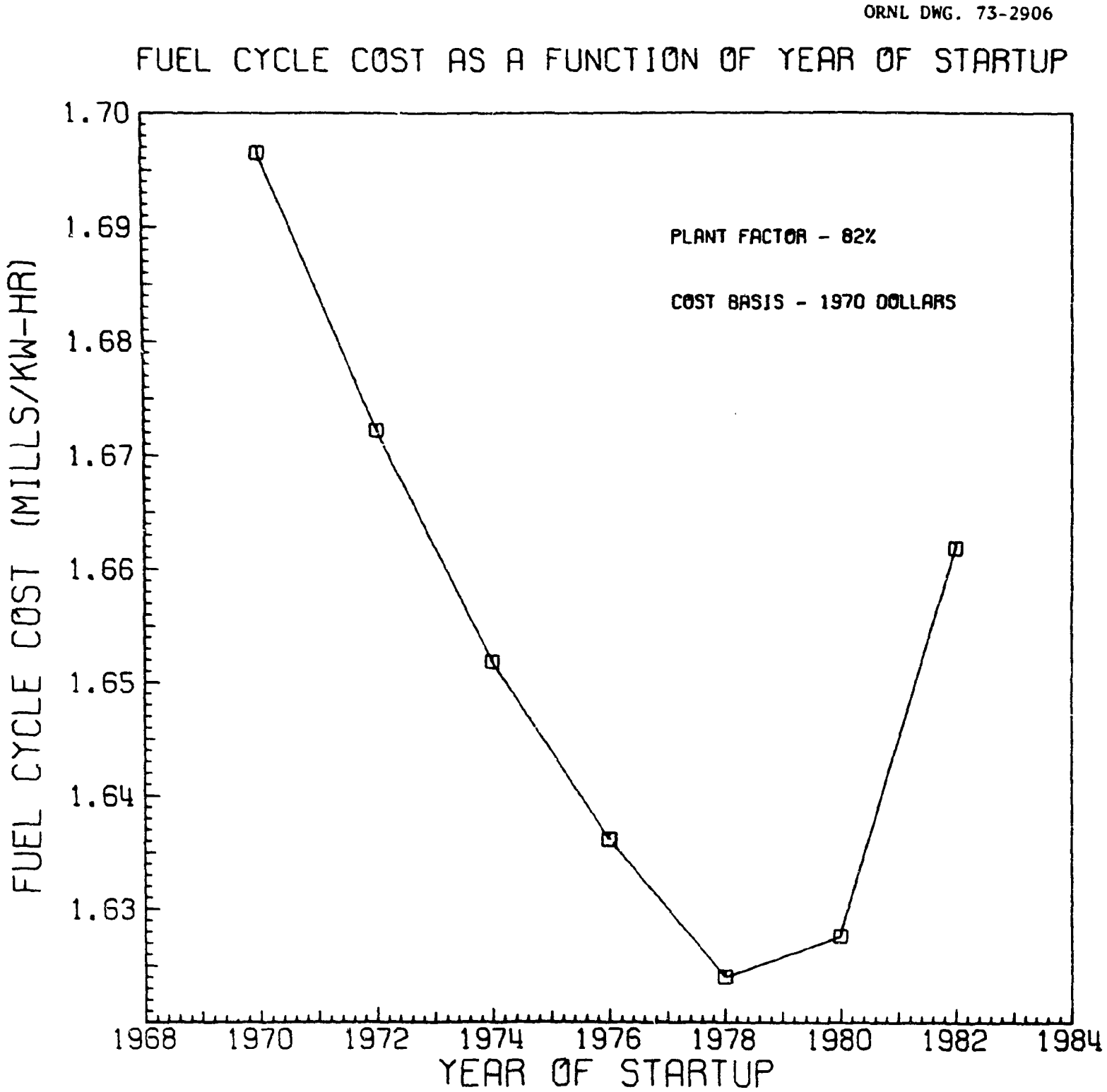


Fig. 5. Fuel cycle cost as a function of year of startup for PWR.

is included. Fuel cycle costs in PROCESS HEAT COST were escalated at 3% per year compounded annually from 1970 to the year of commercial operation.

Total plant capital costs for nine power plants at two years of commercial operation were read into PROCESS HEAT COST. Since we assumed no escalation during design and construction, capital costs at start of construction and commercial operation are the same. Nuclear island cost for all cases (at a given year of operation) was assumed to be independent of process temperature and therefore was determined as NI (nuclear island fraction) times total plant capital cost.

The design and construction period was assumed to be 6 1/2 years for all cases.

Nine cases were calculated first for 1972 operation followed by nine 1978 cases. Turbine-generator plant capital cost was calculated by using the turbine-generator cost factor, TG, times the total capital cost of a power-only plant having the same gross electrical output as each of these dual-purpose cases. This assumes turbine-generator plant cost for dual-purpose or power-only cost the same at a given electrical output. This assumption was recommended by General Electric for previous studies.

The heat removal system cost was calculated for the base (power-only) case but was not included in total capital cost for any other cases. It was assumed that the process heat customer supplies the heat removal system. It was further assumed that condensate is returned from the process use at the normal inlet temperature for the feed water heating train (saturation temperature at process pressure). Base case heat removal cost was calculated using the heat removal cost factor, HR, times the total capital cost of a power-only plant.

Next, fuel cycle costs were converted to an annual basis. For the base case O&M costs were converted to an annual basis and escalated at 3% per year to the year of operation. For the process heat cases the portion of the O&M cost attributable to the heat removal system was subtracted from the base case total, and turbine-generator plant O&M was scaled down in proportion to electrical output.

Total capital cost was then determined by adding nuclear island, turbine-generator plant, and heat removal system capital costs. Total

capital cost was then converted to an annual basis by using the assumed fixed charge rate. The total annual cost was then the sum of the capital, fuel cycle, and O&M costs.

For the power-only case (base case) the cost of power in mills/kWhr was determined by dividing total annual cost by electrical output per year. For the power-only case, the total annual revenue from power was assumed to be equal to the total annual cost. This same value was assumed for power generated in the back pressure cases. In each of the back pressure cases, total annual revenue from power was power cost in mills/kWhr from the base case times electrical output per year.

The required annual revenue from process heat was taken as the difference between total annual cost and total annual revenue from power. A heat balance on the heat rejection system was then made. This allowed determination of process heat available and process heat cost. Process heat cost was taken to be the required annual revenue from process heat divided by the annual process heat available. The same plant factor was used for process heat as for electric power. Note that the process heat costs calculated are the minimum that a power company could charge and still sell power at its cost of production in a power-only station. If profits were included, higher process heat costs would have been calculated.

PART 2. PRESSURIZED-WATER REACTOR — PROCESS FLOW FROM 1 OF 3 L-P UNITS

Calculational Procedure

Results from Part 1 heat balances were used to make the major heat balance calculations for Part 2. The base case (power-only) is the same for Parts 1 and 2. In Part 2, two of the double-flow low-pressure turbines condense steam at 2 in. HgA while one double-flow low-pressure turbine exhausts steam to process. Thus the amount of process steam to be sold for Part 2 is one-third of the amount of process steam to be sold for Part 1. The ORCENT code is not set up to calculate a heat balance on a plant with one cylinder condensing steam at [illegible] in. HgA

with two cylinders condensing steam at 2 in. HgA. Plant electrical output can be closely estimated as two-thirds of the electrical output for a full-flow 2 in. HgA plant plus one-third of the electrical output for a full-flow 3.45 in. HgA plant Gross efficiency is then plant electrical output divided by thermal input [3423 Mw(t)].

Nuclear island capital cost is unchanged from Part 1. Turbine-generator plant capital cost was determined by ORCOST runs at calculated plant electrical output as described above. Turbine-generator plant capital cost is 38.54% of the total capital cost from the ORCOST run. Since the runs in Part 2 have higher electrical output and lower process steam flow, turbine-generator plant costs were generally higher than Part 1.

In Part 1 the heat removal system cost was calculated for the base (power-only) case but was not included in total capital cost for process heat cases. The process heat customer was assumed to supply the heat removal system. In Part 2 it was assumed that the process heat customer supplies one-third of the heat removal system since his process steam represents one-third of the low-pressure turbine exhaust steam flow.

The remainder of the calculations largely follow Part 1. Some additional flows and enthalpies from the ORCENT heat balances are input to PROCESS HEAT COST in order to determine total heat to process. It is assumed that the process uses exhaust steam from the feed pump turbine and flashed drains from the feedwater heaters as well as low-pressure turbine exhaust steam flow. As in Part 1, calculations are made for 1972 and 1978 operation at nine steam pressures from 2 in. HgA through 40 psia.

Operation and maintenance costs for the nuclear island are the same as Part 1. For the turbine-generator plant, O&M costs are assumed to vary with plant electrical output as in Part 1. O&M cost for the heat removal system is reduced by one-third from Part 1 since the process heat user is assumed to provide O&M for the process steam equipment. Fuel-cycle costs were taken from Part 1. All other calculations follow the method of Part 1.

Results

Tabular results for the PWR cases using one of three L-P units for process steam are shown in Tables 3 and 4. Figure 6 shows process heat cost as a function of steam temperature for both 1972 and 1978 operation. Cost as a function of temperature is almost identical to Part 1 with a more reasonable process heat supply in Part 2.

PART 3. HIGH TEMPERATURE GAS REACTOR – FULL FLOW TO PROCESS

Calculational Procedure

We do not have available a computer code that can be used directly for fossil-fueled steam cycle calculations. For the HTGR and coal-fired cases of Parts 3 and 4 a cycle was assumed with 1000°F, 2415 psia throttle conditions and one stage of primary reheat. ORCENT could easily be modified to give a heat balance of much greater accuracy than hand calculation through the revision of a limited number of subroutines. While fossil calculations could be done by hand using the methods of Ref. 6, this would be very time consuming.

The ORCENT MAIN program was modified as were subroutines FWHPAR, GSTAGE, RESULT, SGROUP, DATAIN, PRDROP, FIG167 and FIG8. The method used to obtain the FOSSIL heat balance consists of making hand calculations using the method of Ref. 6 and then applying modifiers to the stage group efficiencies obtained by the method of ORCENT. We therefore force ORCENT to calculate a stage group efficiency which is correct for the fossil conditions.

Subroutine FIG167 was modified to include low-pressure turbine exhaust loss constants for a 31-in. last stage blade 3600 rpm turbine since the fossil cycle assumes a three-section tandem compound six-flow 31-in. LSB 3600 rpm turbine.

The methods of Ref. 6 do not include a separate noncondensing stage group end point correction for exhaust loss as done in ORCENT but rather include the loss in the stage group efficiency calculation. For consistency, subroutine FIG8 was modified to set the exhaust loss equal to zero

Table 3

PRESSURIZED WATER REACTOR 3423 MWT
1 OF 3 L-P SECTIONS PROVIDING PROCESS HEAT
THROTTLE STEAM CONDITIONS 515.3 F 780 PSIA DRY AND SATURATED
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1972.0
YEAR START OF CONSTRUCTION 1965.5

GROSS GEN. OUTPUT, MWE	1168.	1157.	1133.	1112.	1091.	1073.	1053.	1032.	1004.
GROSS EFFICIENCY	0.3412	0.3381	0.3309	0.3250	0.3188	0.3134	0.3076	0.3015	0.2932
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	2748663.	2845996.	2904330.	2948329.	3023996.	3075329.	3128996.	3184329.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	220.	240.	267.
NUCLEAR ISLAND COST, $	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.	135654480.
T-G PLANT COST, $	98970640.	98354064.	97043696.	95926048.	94769840.	93767792.	92650144.	91493936.	89952336.
HEAT REM. SYS. COST, $	22174640.	14783100.	14783100.	14783100.	14783100.	14783100.	14783100.	14783100.	14783100.
TOTAL CAP. COST, $	256799760.	248791632.	247481264.	246363616.	245207408.	244205360.	243087712.	241931504.	240389904.
ANNUAL CAP. COST, $	35951952.	34830816.	34647360.	34490896.	34329024.	34188736.	34032272.	33870400.	33654576.
ANNUAL FUEL COST, $	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.	14882313.
ANNUAL O&M COST, $	3831698.	3647986.	3636657.	3627318.	3617673.	3609100.	3600068.	3590423.	3577410.
TOTAL ANN. COST, $	54665952.	53361104.	53166320.	53000512.	52828992.	52680128.	52514640.	52343120.	52114288.
POWER COST, MILLS/KWHR	6.52	6.52	6.52	6.52	6.52	6.52	6.52	6.52	6.52
REVENUE FROM POWER, $/YR	54665920.	54166704.	53012224.	52060560.	51077696.	50204032.	49283600.	48300752.	46974624.
REQD REV. PROC. HT.$/YR	0.	-805600.	154096.	939952.	1751296.	2476096.	3231040.	4042368.	5139664.
PROC. HEAT, MILL. BTU/HR	0.	2581.	2666.	2736.	2794.	2874.	2943.	3017.	3112.
HEAT COST, CTS/MILL. BTU	0.0	-4.3	0.8	4.8	8.7	12.0	15.3	18.7	23.0

Table 4

PRESSURIZED WATER REACTOR 3423 MWT
1 OF 3 L-P SECTIONS PROVIDING PROCESS HEAT
THROTTLE STEAM CONDITIONS 515.3 F 780 PSIA DRY AND SATURATED
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1978.0
YEAR START OF CONSTRUCTION 1971.5

GROSS GEN. OUTPUT, MWE	1168.	1157.	1133.	1112.	1091.	1073.	1053.	1032.	1004.
GROSS EFFICIENCY	0.3412	0.3381	0.3309	0.3250	0.3188	0.3134	0.3076	0.3015	0.2932
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	2748663.	2845996.	2904330.	2948329.	3023996.	3075329.	3128996.	3184329.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	220.	240.	267.
NUCLEAR ISLAND COST, $	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.	166187264.
T-G PLANT COST, $	121246768.	120514544.	118857312.	117431344.	115966800.	114733536.	113346064.	111881520.	109916032.
HEAT REM. SYS. COST, $	27165664.	18110448.	18110448.	18110448.	18110448.	18110448.	18110448.	18110448.	18110448.
TOTAL CAP. COST, $	314599424.	304811776.	303154688.	301728768.	300264192.	299030784.	297643520.	296178944.	294213376.
ANNUAL CAP. COST, $	44043904.	42673648.	42441648.	42242016.	42036976.	41864304.	41670080.	41465040.	41189872.
ANNUAL FUEL COST, $	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.	17260016.
ANNUAL O&M COST, $	4575241.	4355879.	4342352.	4331201.	4319685.	4309448.	4298663.	4287146.	4271608.
TOTAL ANN. COST, $	65879152.	64289536.	64044016.	63833232.	63616672.	63433760.	63228752.	63012192.	62721488.
POWER COST, MILLS/KWHR	7.85	7.85	7.85	7.85	7.85	7.85	7.85	7.85	7.85
REVENUE FROM POWER, $/YR	65879152.	65277520.	63886240.	62739376.	61554896.	60502032.	59392768.	58208288.	56610192.
REQD REV. PROC. HT.$/YR	0.	-987984.	157776.	1093856.	2061776.	2931728.	3835984.	4803904.	6111296.
PROC. HEAT, MILL. BTU/HR	0.	2581.	2666.	2736.	2794.	2874.	2943.	3017.	3112.
HEAT COST, CTS/MILL. BTU	0.0	-5.3	0.8	5.6	10.3	14.2	18.1	22.2	27.3

ORNL DWG. 73-2907

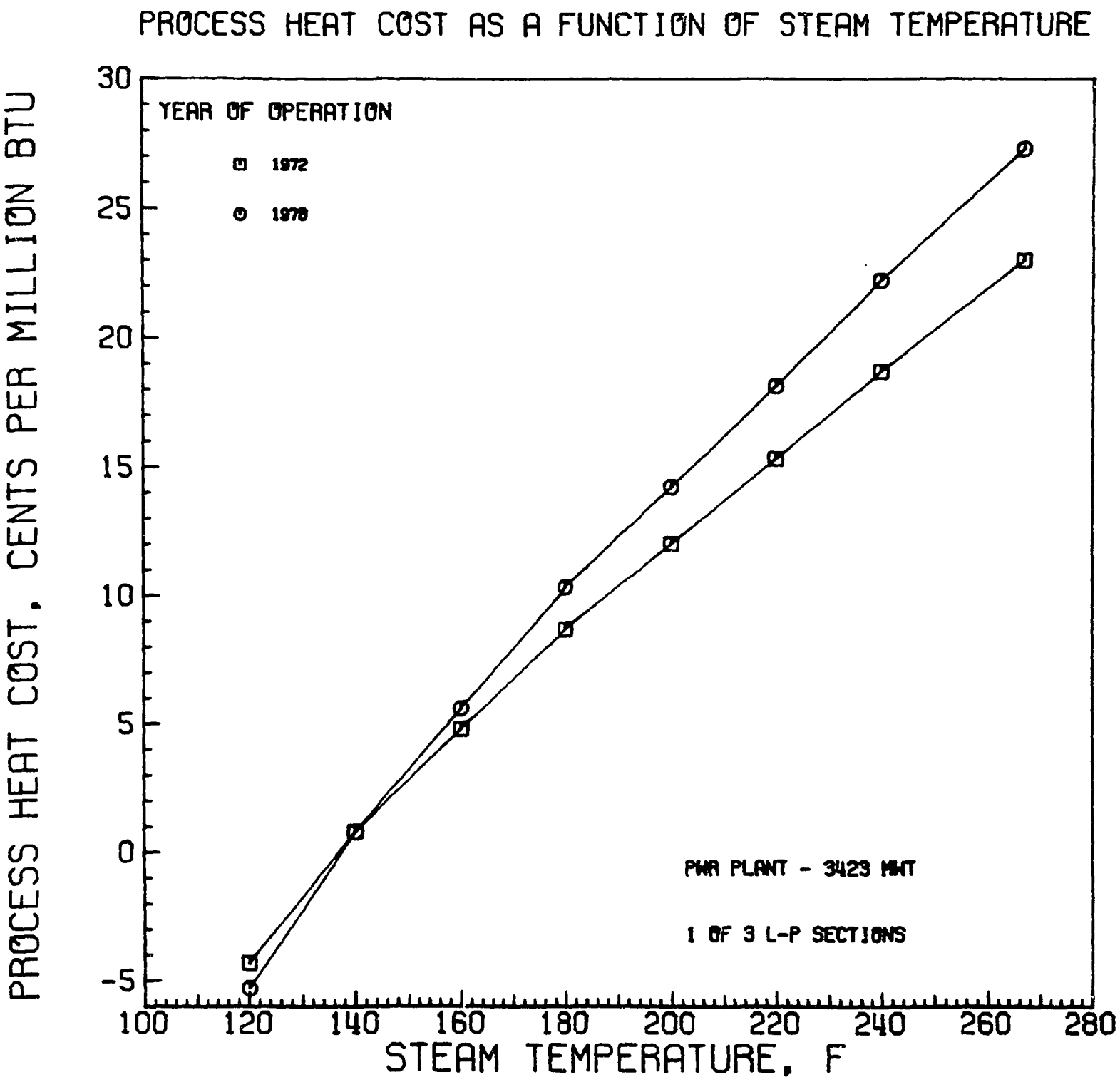


Fig. 6. Process heat cost as a function of temperature for PWR 1 of 3 cases.

and the effect was included in the stage group efficiency correction factor. The intermediate system pressure drops in subroutine PRDROP were not yet revised but would not include significant error in the heat balance as is.

Nine FOSSIL heat balances were made at 2687 MW(t) to obtain required enthalpies and flows for PROCESS HEAT COST. At 2687 MW(t) the power-only case provides 1168 MW(e) gross generator output to match the PWR power-only case. The cycle calculations for the coal-fired plant used the same results from FOSSIL.

The economic calculations for the HTGR plant follow the outline of the full-flow PWR cases. They differ only in constants and how they are obtained. In order to determine the fractions of total plant capital cost attributable to nuclear island, turbine-generator plant, and heat removal system, data from Gulf General Atomic on advantage vs disadvantage of HTGR vs PWR was used. Nuclear island capital costs were taken to be higher (60.59%) for the HTGR, while turbine-generator plant (33.674%) and heat removal system (5.736%) capital costs were assumed lower.

As in Part 1, turbine-generator plant capital cost was assumed to be a function of gross electrical output and independent of low-pressure turbine back-pressure. ORCOST runs for 842 through 1168 MW(e) gross were made and turbine-generator plant capital cost was assumed to be 33.7% of total capital cost.

Operation and maintenance costs for the HTGR cases were assumed to be 10% lower than PWR cases. Fuel cycle costs were derived from an ORSAC run for 1978 operation. Fuel cycle costs were assumed to be the same for 1972 operation when both were expressed in 1970 dollars. Costs in 1970 dollars were escalated at 3% per year compounded annually to the year of commercial operation. Design and construction period remained unchanged at 6 1/2 years.

Results

Tabular results for the HTGR cases are shown in Tables 5 and 6. Figure 7 shows process heat cost as a function of steam temperature for both 1972 and 1978 operation. The irregularity in the curves from 120

Table 5

HIGH TEMPERATURE GAS REACTOR 2687 MWT
THROTTLE STEAM CONDITIONS 1000. F 2415. PSIA
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1972.0
YEAR START OF CONSTRUCTION 1965.5

GROSS GEN. OUTPUT, MWE	1168.	1151.	1096.	1072.	1028.	987.	946.	903.	842.
GROSS EFFICIENCY	0.4347	0.4283	0.4079	0.3989	0.3826	0.3675	0.3521	0.3362	0.3133
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	5197528.	5295324.	5398983.	5505457.	5596903.	5693503.	5795833.	5921610.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	251.	306.	386.
NUCLEAR ISLAND COST, $	156140384.	156140384.	156140384.	156140384.	156140384.	156140384.	156140384.	156140384.	156140384.
T-G PLANT COST, $	86777856.	86003360.	83410464.	82265552.	80177776.	78191008.	76204224.	74082768.	71018432.
HEAT REM. SYS. COST, $	14781667.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	257699904.	242143744.	239550848.	238405936.	236318160.	234331392.	232344608.	230123152.	227158816.
ANNUAL CAP. COST, $	36077984.	33900112.	33537104.	33376816.	33084528.	32806384.	32528240.	32231232.	31802224.
ANNUAL FUEL COST, $	14828904.	14828904.	14828904.	14828904.	14828904.	14828904.	14828904.	14828904.	14828904.
ANNUAL O&M COST, $	3448528.	2958601.	2935989.	2925980.	2907849.	2891094.	2874009.	2856300.	2830955.
TOTAL ANN. COST, $	54355408.	51687600.	51301984.	51131680.	50821264.	50526368.	50231136.	49916416.	49462064.
POWER COST, MILLS/KWHR	6.48	6.48	6.48	6.48	6.48	6.48	6.48	6.48	6.48
REVENUE FROM POWER, $/YR	54355408.	53552400.	51006688.	49879840.	47838464.	45952144.	44028576.	42034784.	39181392.
REQD REV. PROC. HT.$/YR	0.	-1864800.	295296.	1251840.	2982800.	4574224.	6202560.	7881632.	10280672.
PROC. HEAT, MILL. BTU/HR	0.	5183.	5359.	5445.	5597.	5736.	5881.	6029.	6239.
HEAT COST, CTS/MILL. BTU	0.0	-5.0	0.8	3.2	7.4	11.1	14.7	18.2	22.9

Table 6

HIGH TEMPERATURE GAS REACTOR 2687 MWT
THROTTLE STEAM CONDITIONS 1000. F 2415. PSIA
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1978.C
YEAR START OF CONSTRUCTION 1971.5

GROSS GEN. OUTPUT, MWE	1168.	1151.	1096.	1072.	1028.	987.	946.	903.	842.
GROSS EFFICIENCY	0.4347	0.4283	0.4079	0.3989	0.3826	0.3675	0.3521	0.3362	0.3133
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	5197528.	5295324.	5398983.	5505457.	5596903.	5693503.	5795833.	5921610.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	251.	306.	386.
NUCLEAR ISLAND COST, $	191464384.	191464384.	191464384.	191464384.	191464384.	191464384.	191464384.	191464384.	191464384.
T-G PLANT COST, $	106409808.	105399584.	102166864.	100752576.	98092256.	95634128.	93108592.	90414672.	86575776.
HEAT REM. SYS. COST, $	18125744.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	315999744.	296863744.	293631232.	292216832.	289556480.	287098368.	284572928.	281879040.	278040064.
ANNUAL CAP. COST, $	44239952.	41560912.	41108368.	40910352.	40537904.	40193760.	39840208.	39463056.	38925600.
ANNUAL FUEL COST, $	17706384.	17706384.	17706384.	17706384.	17706384.	17706384.	17706384.	17706384.	17706384.
ANNUAL O&M COST, $	4117716.	3532718.	3505719.	3493768.	3472118.	3452112.	3431711.	3410565.	3380300.
TOTAL ANN. COST, $	66064048.	62800000.	62320464.	62110496.	61716400.	61352256.	60978288.	60580000.	60012272.
POWER COST, MILLS/KWHR	7.87	7.87	7.87	7.87	7.87	7.87	7.87	7.87	7.87
REVENUE FROM POWER, $/YR	66064048.	65088096.	61993968.	60624400.	58143312.	55850656.	53512720.	51089456.	47621440.
REQD REV. PROC. HT.$/YR	0.	-2288096.	326496.	1486096.	3573088.	5501600.	7465568.	9490544.	12390832.
PROC. HEAT, MILL. BTU/HR	0.	5183.	5359.	5445.	5597.	5736.	5881.	6029.	6230.
HEAT COST, CTS/MILL. BTU	0.0	-6.1	0.8	3.8	8.9	13.4	17.7	21.9	27.6

ORNL DWG. 73-2908

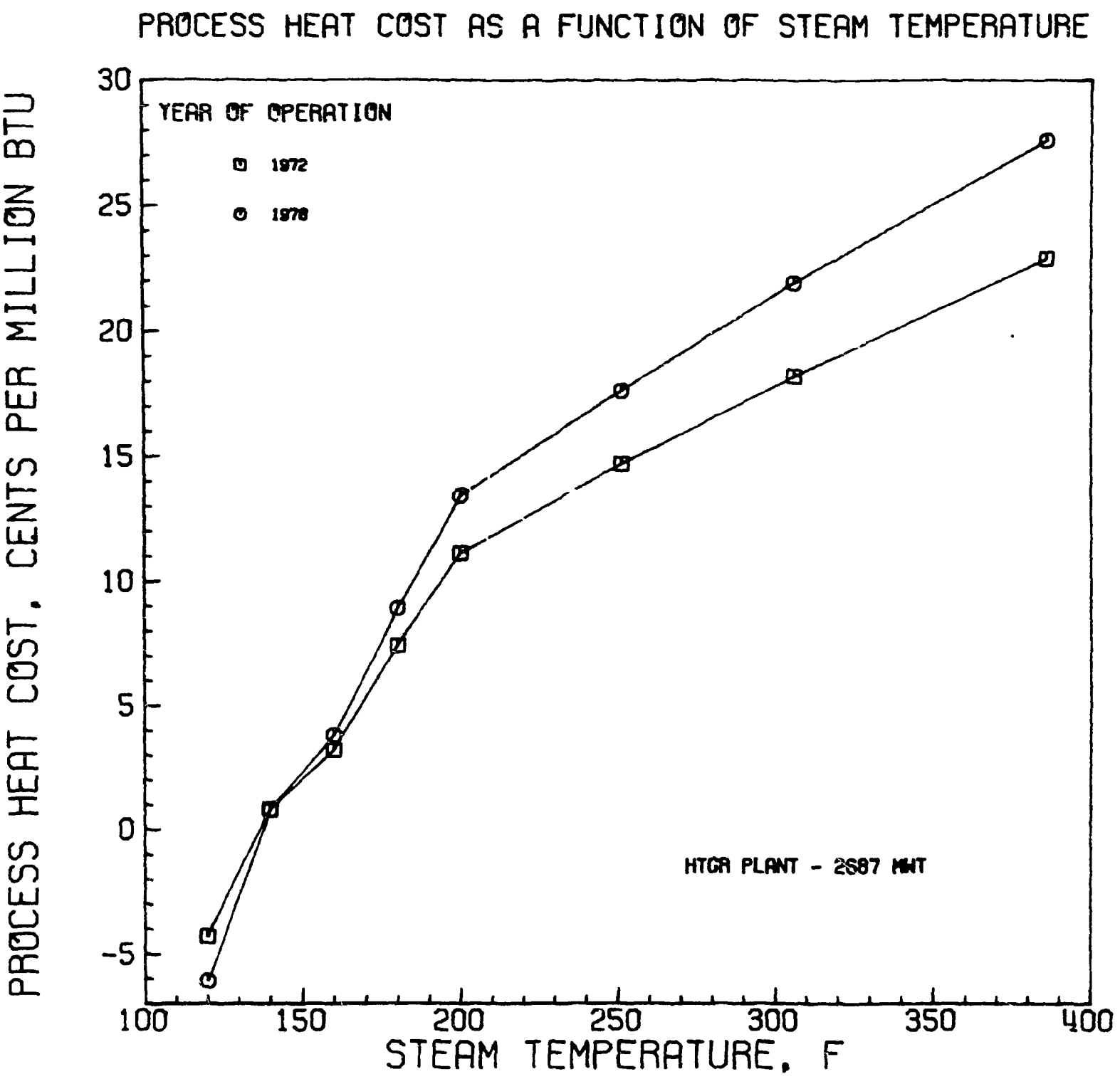


Fig. 7. Process heat cost as a function of temperature for HTGR full flow cases.

to 200°F results from the choices in the feedwater heating cycle resulting in lower than average efficiency for the 160 and 180°F points relative to the 140°F point. Funds are not available to make the necessary changes. The break in the curves at 200°F arises from the transition to superheated exhaust steam. A plot of cost vs process pressure would be smooth at 200°F while the temperature plot has a break in it. The same holds true for the coal-fired cases.

PART 4. COAL-FIRED FOSSIL PLANT – FULL FLOW TO PROCESS

Calculational Procedure

The coal-fired cases use the output of FOSSIL (the modified ORCENT deck) as run for Part 4, the HTGR section. It is assumed that the coal-fired plant operates with the same 2415 psia, 1000°F/1000°F cycle previously described. PROCESS HEAT COST is modified only to input fuel cost in cents per million Btu rather than as a fuel cycle cost in mills/kW-hr. Boiler plant capital costs are taken to be 58.96%, turbine-generator plant 35%, and heat removal system 6.03%. These factors were determined by CONCEPT II runs as for the PWR cases. O&M costs for the coal-fired cases were assumed to be about 25% lower than the PWR cases or 16% lower than the HTGR cases. Seventy percent of the O&M costs were charged to the boiler plant, 15% to the turbine-generator plant, and 15% to the heat removal system for the power-only case. Turbine-generator system O&M costs were again assumed to vary with electrical output. For the cases in which steam was sold, heat removal system O&M costs were assumed to be zero. The design and construction period was assumed to be five years.

Results

Tabular results for the coal-fired cases are shown in Tables 7 and 8. Figure 8 shows process heat cost as a function of steam temperature for both 1972 and 1978 operation. As mentioned in Part 3, the irregularity in the curves in the lower range of temperatures results from the choice of feedwater heating cycle and would have been revised had funds been available.

Table 7

COAL-FIRED PLANT 2687 MWT
THROTTLE STEAM CONDITIONS 1000. F 2415. PSIA
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1972.0
YEAR START OF CONSTRUCTION 1967.0

GROSS GEN. OUTPUT, MWE	1168.	1151.	1096.	1072.	1028.	987.	946.	903.	842.
GROSS EFFICIENCY	0.4347	0.4283	0.4079	0.3989	0.3826	0.3675	0.3521	0.3362	0.3133
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	5197528.	5295324.	5398983.	5505457.	5596903.	5693503.	5795833.	5921610.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	251.	306.	386.
BOILER PLANT COST, $	144454432.	144454432.	144454432.	144454432.	144454432.	144454432.	144454432.	144454432.	144454432.
T-G PLANT COST, $	85762224.	84817088.	81771664.	80441456.	77956112.	75645776.	73300448.	70850096.	67314576.
HEAT REM. SYS. COST, $	14783299.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	244999952.	229271520.	226226096.	224895888.	222410544.	220100208.	217754880.	215304528.	211769008.
ANNUAL CAP. COST, $	34299984.	32098000.	31671648.	31485408.	31137472.	30814016.	30485680.	30142624.	29647648.
ANNUAL FUEL COST, $	34934624.	34934624.	34934624.	34934624.	34934624.	34934624.	34934624.	34934624.	34934624.
ANNUAL O&M COST, $	2895061.	2454386.	2434047.	2425045.	2408736.	2393665.	2378297.	2362369.	2339572.
TOTAL ANN. COST, $	72129664.	69487008.	69040304.	68845072.	68480832.	68142304.	67798592.	67439616.	66921840.
POWER COST, MILLS/KWHR	8.60	8.60	8.60	8.60	8.60	8.60	8.60	8.60	8.60
REVENUE FROM POWER, $/YR	72129664.	71064048.	67685904.	66190560.	63481696.	60978512.	58425952.	55780176.	51993744.
REOD REV. PROC. HT.$/YR	0.	-1577040.	1354400.	2654512.	4999136.	7163792.	9372640.	11659440.	14928096.
PROC. HEAT, MILL. BTU/HR	0.	5183.	5359.	5445.	5597.	5736.	5881.	6029.	6239.
HEAT COST, CTS/MILL. BTU	0.0	-4.2	3.5	6.8	12.4	17.4	22.2	26.9	33.3

Table 8

COAL-FIRED PLANT 2687 MWT
THROTTLE STEAM CONDITIONS 1000. F 2415. PSIA
LOS ANGELES
7 PERCENT INTEREST
FIXED CHARGE RATE (PRIVATE FINANCING) 14 PERCENT
NO ESCALATION DURING DESIGN AND CONSTRUCTION
NO COOLING TOWER
.82 CAPACITY FACTOR

YEAR OF OPERATION 1978.0
YEAR START OF CONSTRUCTION 1973.0

GROSS GEN. OUTPUT, MWE	1168.	1151.	1096.	1072.	1028.	987.	946.	903.	842.
GROSS EFFICIENCY	0.4347	0.4283	0.4079	0.3989	0.3826	0.3675	0.3521	0.3362	0.3133
PROCESS PRESSURE, IN. HGA	2.00	3.45	5.88	9.65					
PSIA					7.51	11.53	17.19	24.97	40.00
STEAM TO PROCESS, #/HR	0.	5197528.	5295324.	5398983.	5505457.	5596903.	5693503.	5795833.	5921610.
PROCESS TEMP.	101.	120.	140.	160.	180.	200.	251.	306.	386.
BOILER PLANT COST, $	178769584.	178769584.	178769584.	178769584.	178769584.	178769584.	178769584.	178769584.	178769584.
T-G PLANT COST, $	106135120.	104979936.	101129392.	99449136.	96333712.	93428288.	90487904.	87407456.	82961824.
HEAT REM. SYS. COST, $	18295072.	0.	0.	0.	0.	0.	0.	0.	0.
TOTAL CAP. COST, $	303199744.	283749376.	279898880.	278218496.	275103232.	272197632.	269257472.	266177040.	261731408.
ANNUAL CAP. COST, $	42447952.	39724912.	39185840.	38950576.	38514448.	38107664.	37696032.	37264768.	36642384.
ANNUAL FUEL COST, $	41713536.	41713536.	41713536.	41713536.	41713536.	41713536.	41713536.	41713536.	41713536.
ANNUAL O&M COST, $	3456848.	2930660.	2906375.	2895625.	2876151.	2858157.	2839807.	2820787.	2793567.
TOTAL ANN. COST, $	87618336.	84369104.	83805744.	83559728.	83104128.	82679344.	82249360.	81799088.	81149472.
POWER COST, MILLS/KWHR	10.44	10.44	10.44	10.44	10.44	10.44	10.44	10.44	10.44
REVENUE FROM POWER, $/YR	87618304.	86323904.	82220336.	80403904.	77113312.	74072640.	70971968.	67758048.	63158560.
REQD REV. PROC. HT.$/YR	0.	-1954800.	1585408.	3155824.	5990816.	8606704.	11277392.	14041040.	17990912.
PROC. HEAT, MILL. BTU/HR	0.	5183.	5359.	5445.	5597.	5736.	5881.	6029.	6239.
HEAT COST, CTS/MILL. BTU	0.0	-5.3	4.1	8.1	14.9	20.9	26.7	32.4	40.1

ORNL DWG. 73-2909

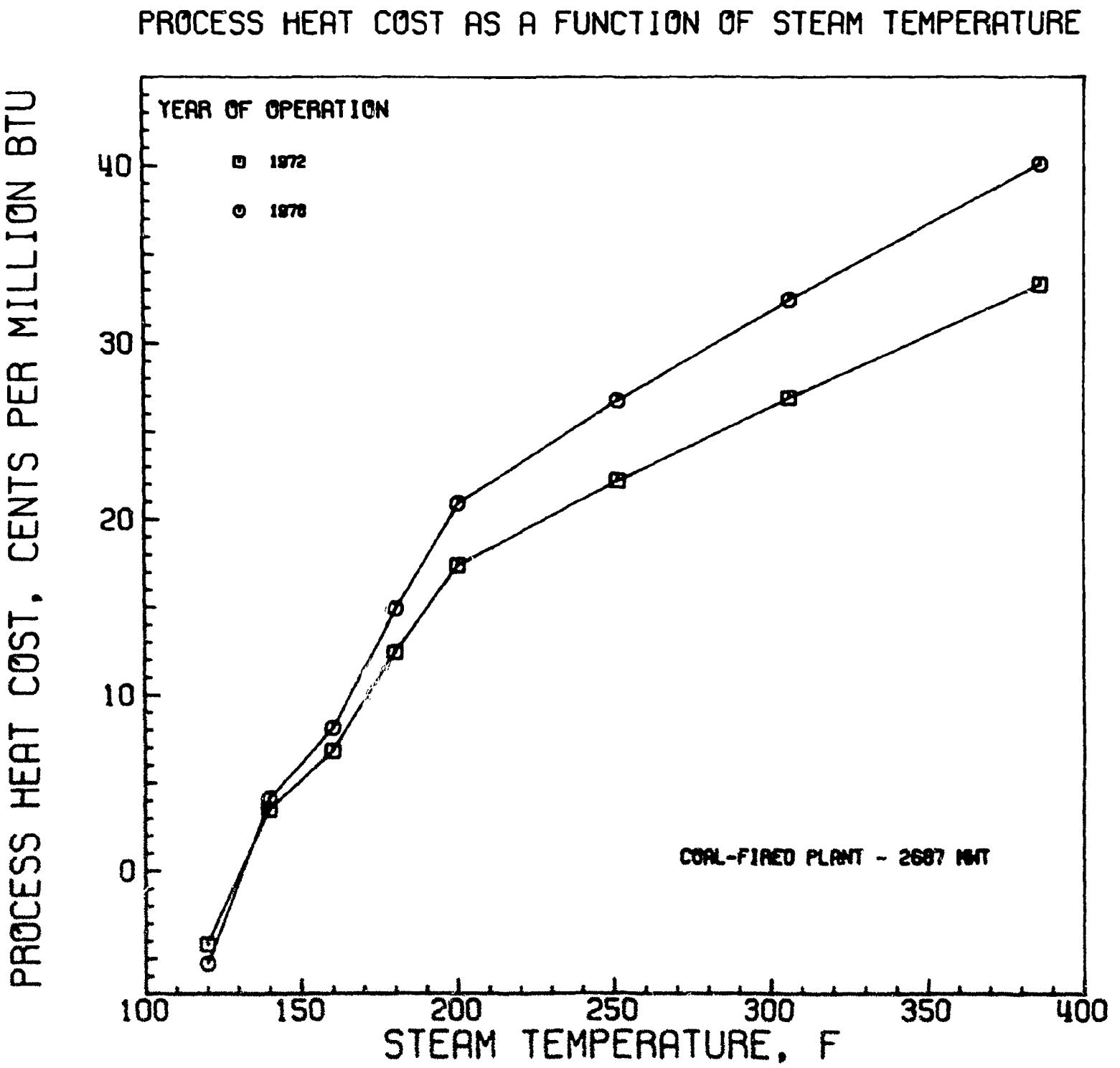


Fig. 8. Process heat cost as a function of temperature for coal-fired full flow cases.

PART 5. PRESSURIZED-WATER REACTOR – PROCESS STEAM AT THREE PRESSURES

Cycle Description

A dual-purpose plant was assumed to require process steam at three pressure levels. Figure 9 shows the three processes tied into a PWR nuclear steam supply system as previously analyzed in Part 1 with:

400,000 #/hr of throttle steam at 780 psia to Process 1

3,000,000 #/hr of crossover steam at 170.8 psia to Process 2

1,600,000 #/hr of L-P turbine extraction steam at 43.1 psia to Process 3

There was no attempt to match the pressures and flows of process steam to known chemical processes. This was an example to establish the method of calculation.

Calculation Procedure

First it was necessary to determine enough of a heat balance to obtain gross electrical output. This was done by superposition of numerous ORCENT heat balances and some hand calculation. Four steam systems were superposed as follows:

1. throttle steam to process 1 returned as condensate,
2. a turbine-generator cycle expanding only to the exhaust pressure of the H-P turbine in Fig. 9,
3. a turbine-generator cycle expanding to the L-P turbine extraction pressure for process 3 in Fig. 9,
4. a condensing turbine-generator cycle expanding to 2 in. HgA.

There are some difficulties in superposing the above when consideration is given to handling of feedwater heater drains in the individual cycles, handling of feed pump turbine exhausts from individual cycles, variation of efficiency of stage groups in individual cycles as volume flow is altered, etc.

Since this cycle was intended only to be used as an example of the methods of superposition of heat balances, economic calculations were made by hand rather than writing another PROCESS HEAT COST code for this

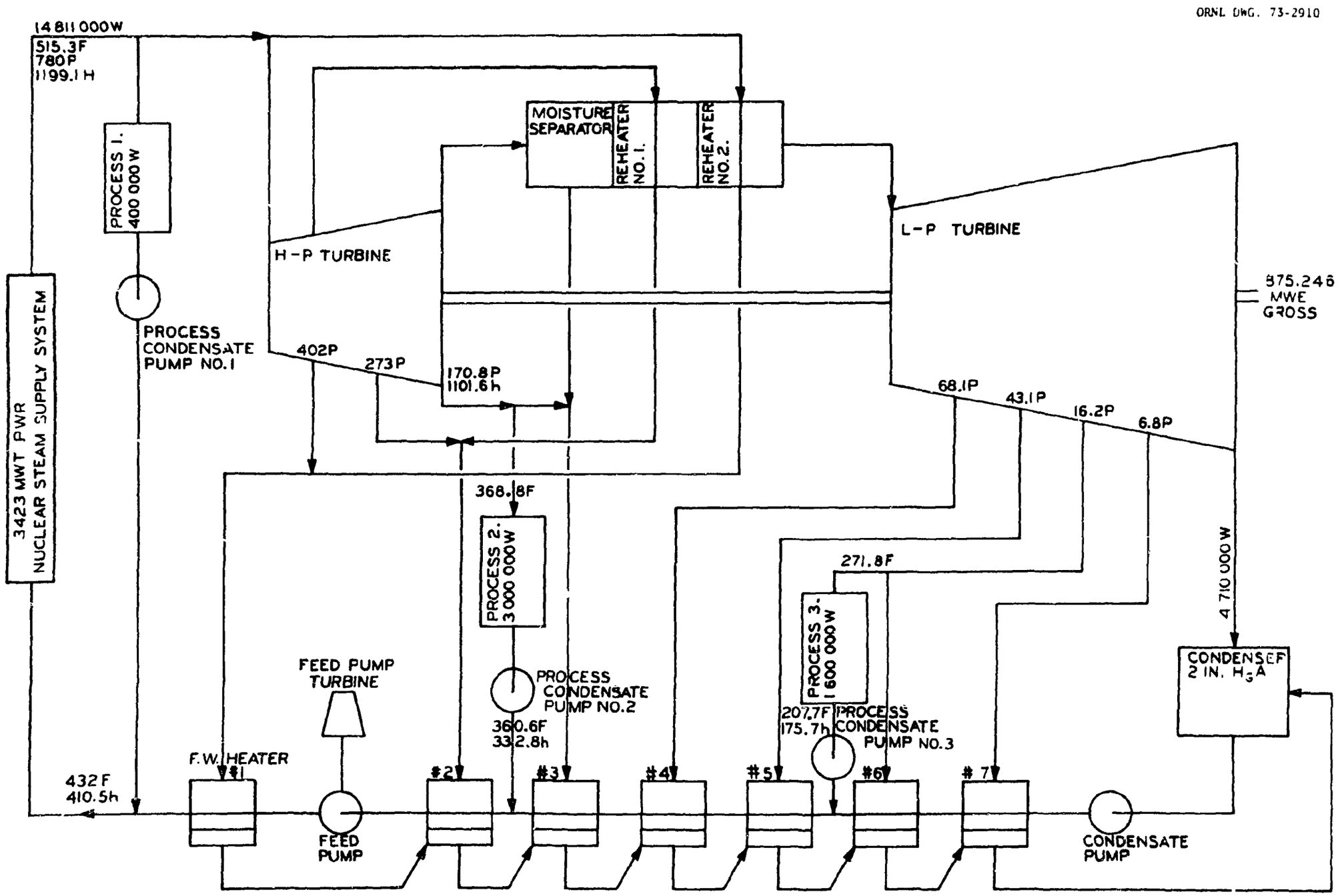


Fig. 9. Process flowsheet for PWR providing process steam at three pressures.

section. The methods of calculation will therefore be described in a bit more detail than in other sections.

The assumptions, ground rules, and methods of calculation are as follows:

The reference power-only case for Part 5 is identical to the power-only cases in Tables 2 and 4 for the PWR full-flow and one-of-three cases. It is repeated in Table 9 for reference. The division between nuclear island, turbine-generator plant, and heat removal system capital costs for the reference power-only case is therefore identical to Parts 1 and 2 as determined by the use of the CONCEPT II code.

Reactor power was assumed to be 3423 MW(t) as in Parts 1 and 2.

Throttle steam conditions are 780 psia, 515.3°F, dry and saturated.

Crossover conditions are 170.8 psia and 368.8°F.

Plant location is assumed to be the Los Angeles area.

Interest during construction is 7%.

There is no escalation during construction. Operation is planned for early 1978 and is expressed in terms of mid-1971 dollars to be consistent with Parts 1 through 4.

There is no cooling tower. Once-through cooling at 2 in. HgA is assumed.

The capacity factor for both power and process heat is 0.82.

O&M costs were assumed to be 0.45 mills/kWhr in mid 1971 and escalated at 3% per year to early 1978.

For the base case (power-only) operation and maintenance costs are assumed to divide 72, 14, and 14% for the nuclear island, turbine-generator plant, and heat removal system. For the turbine-generator plant, including process steam, the factor of 14% is reduced by the ratio (dual-purpose plant electrical output/power-only plant electrical output = 10.49%). Heat removal O&M factor of 14% is reduced by the ratio (condenser heat rejection for the dual-purpose plant/condenser heat rejection power-only plant = 8.11%).

Fixed charge rate (private financing) was assumed to be 14%.

Fuel cycle costs were determined from the results of an ORSAC run in 1970 dollars and escalated at 3% per year.

Table 9

	Power-Only	Dual-Purpose
Gross generator output, MW(e)	1168	875.2
Nuclear island cost, $	166187000	166187000
T-G plant cost, $	121247000	100666000
Heat removal system cost, $	27165000	15731000
Total capital cost, $	314599000	282584000
Annual capital cost, $	44044000	39562000
Annual fuel cycle cost, $	17250000	17250000
Annual O&M cost, $	4575000	4145000
Total annual cost, $	65879000	60957000
Total power cost, mills/kWhr	7.85	
Total annual revenue from power, $	65879000	49367000
Required annual revenue from process heat, $		11590000
Process heat, millions of Btu/hr		4077
Process heat cost, cents per million Btu		39.6

Design and construction period was assumed to be 6.5 years.

Power was valued at 7.85 mills/kWhr for 1978 operation.

Required annual revenue from process heat was taken as the difference between total annual cost and total annual revenue from power.

Process heat cost was taken to be the required annual revenue from process heat divided by the annual process heat available.

ORCOST was run at 875.246 MW in order to find turbine-generator plant cost for the chemical plant. Turbine-generator plant cost was assumed to be 0.3854 times total capital cost from ORCOST. This presumes that the cost of a back pressure turbine or condensing turbine are the same at a given rating. This further presumes that novelty of turbine design is not a factor in cost. The large amounts of steam extracted at the cross-over and in the low-pressure cylinders will result in a turbine that is more cylindrical than conical near the turbine exhaust.

Process steam is returned to the feedwater heating stream at considerably subcooled conditions, particularly that taken at throttle conditions. Throttle steam is sent to process at 515.3°F and returned to the feedwater stream at 432°F or subcooled by 83.3°F. Crossover and L-P turbine extraction steam are returned 8.2 and 9.8°F subcooled respectively. While it might be desirable in an actual cycle to alter these conditions, it was necessary in order to eliminate much hand calculation to accept the subcooling.

In order to superpose heat balances from various ORCENT runs it is necessary that the state points on a turbine expansion lines coincide. For this to be true, the volume flows must match since efficiency is a function of volume flow. This required that all ORCENT runs be made at the same high throttle steam flow of about 14,174,000 lb/hr and using double-flow high-pressure sections and six-flow low-pressure sections regardless of the power developed. These assumptions are used only in the determination of heat balance figures for the cycles to be superposed and have no bearing on the actual final turbine selection that would be used.

Results

Process heat cost was estimated to be 39.6 cents per million Btu assuming 4077 million Btu per hour to process heat. Calculated results are listed in Table 9. A PWR power plant of the same reactor size is also listed in Table 9 to show comparative costs and to establish the value of electrical power to be used for the dual-purpose case.

DISCUSSION AND CONCLUSIONS

This report presents preliminary results of a study of process heat cost as a function of temperature. None of the cycles were optimized. While the results might be useful for comparative purposes, much remained to be done when funds ran out.

Had time permitted comparisons between the various tabular results, assumptions might have been revised to give more typical or average cost

information. All that can really be said for these tables are that they were calculated using consistent methods. Certain things would be changed at this point if the study were to go on. HTGR and coal-fired cases would both have been worked for one-of-three L-P units in addition to full flow. The PWR full flow case delivers 9337 million Btu/hr at maximum back pressure against 3112 for the one-of-three case. The HTGR and coal-fired cases use 6239 million Btu/hr at maximum back pressure. It seems now that 2080 million Btu/hr for one-third of the HTGR or coal-fired cases would have been a more appropriate process heat supply to a potential user.

None of the cycles were optimized. Certainly a four-flow L-P turbine should have been looked at.

No cooling tower was included. This might substantially change costs of process heat since it would substantially change power cost.

Plant-capacity factor was set at 82% since the ORSAC runs used for fuel-cycle costs were made at 82%. It would have been interesting to vary plant-capacity factor and see the effect on cost.

The sensitivity of cost to such things as construction periods of 6.5 years for nuclear and 5.0 for fossil plants would have been interesting to determine.

Auxiliary power requirements over those required for feed pump turbine have been ignored to this point. There was no attempt as yet to factor in the effect of circulator drives in the HTGR cycle. Consequently cycle calculations for the coal and HTGR cases are identical. This does not yield a realistic comparison if one is interested in net powers.

Fuel cycle and O&M costs were escalated at 3% per year while capital costs were not escalated during construction. This inconsistency should be corrected for future work.

The study was run for 1972 and 1978 operation. It is now a matter of history for 1972. The study should perhaps be advanced to 1974 and 1980.

All the cases in Parts 1 through 4 have one common parameter. Gross generator output for the reference power-only case is 1168 MW(e). Something had to be held constant and that was chosen. If reactor power was

held constant at 3423 MW(t) as for the PWR cases, the electrical output would have been 1488 MW(e). This would be too high. I suspect that the amount of power sold has a sizable influence on cost of process heat. The full flow PWR case with 40 psia back pressure sold 675 MW(e) while the corresponding one-of-three case sold 1004 MW(e) or about 50% more.

It would have been useful to develop a table listing comparative costs for Parts 1 through 4. In the absence of such a table, a comparison of some 1972 costs are as follows:

Boiler plant cost for coal is considerably higher than nuclear island cost for Parts 1 and 2 (PWR) due to the inclusion of coal handling and storage costs. This is true even though the thermal output for the coal-fired cases is lower than PWR thermal output. HTGR nuclear island cost is considerably above coal-fired boiler plant cost at the same thermal output.

For the same electrical output (power-only) the PWR cases (Parts 1 and 2) had the highest turbine-generator plant costs with the HTGR turbine-generator plant cost 12% lower and coal-fired 13% lower than PWR.

Heat removal system costs for the power-only cases were as expected. Coal and HTGR were one-third lower than the PWR case.

The HTGR total capital cost (power-only) was less than 0.5% higher than PWR power-only while the coal-fired case was about 5% lower than PWR.

Fuel-cycle costs didn't differ much between PWR and HTGR but coal costs were very high. Coal costs were about 2.3 times PWR fuel costs using 50 cents/million Btu for 1970 coal cost.

O&M costs were not a significant part of the total costs and therefore differences were relatively unimportant. HTGR costs were about 10% lower than PWR for power-only cases. Coal costs were about 25% lower for power-only cases.

Total annual costs of HTGR and PWR plants differed by less than 0.5% while coal-fired was 30% higher than PWR largely due to fuel cost.

Cost of nuclear electric power was about 6.5 mills compared with 8.6 for coal. Similarly process heat cost about 23 cents/million Btu for nuclear at the highest back pressure (40 psia) and 33 cents/million Btu for coal.

REFERENCES

1. H. I. Bowers, ORCENT: A Digital Computer Program for Saturated and Low Superheat Steam Turbine Cycle Analysis, ORNL-TM-2395 (January 1969).
2. H. I. Bowers, R. C. DeLozier, L. D. Reynolds, and B. E. Srite, CONCEPT – Computerized Conceptual Cost Estimates for Steam-Electric Power Plants – Phase II User's Manual, ORNL-4809 (to be published).
3. L. C. Fuller, C. A. Sweet, and H. I. Bowers, ORCOST – A Computer Code for Summary Capital Cost Estimates of Steam-Electric Power Plants – User's Manual, ORNL-TM-3743 (September 1972).
4. ORSAC: The Oak Ridge Systems Analysis Code – User's Manual, ORNL-TM-3223.
5. NUS Corporation, Guide for Economic Evaluation of Nuclear Reactor Plant Designs, NUS-531 (January 1969).
6. R. C. Spencer, K. C. Cotton, and C. N. Cannon, A Method for Predicting the Performance of Steam Turbine-Generators ... 16,500 kW and Larger, GER-2007A (1962).

APPENDIX B

CORRESPONDENCE WITH ALUMINUM COMPANIES

OAK RIDGE NATIONAL LABORATORY
OPERATED BY
UNION CARBIDE CORPORATION
NUCLEAR DIVISION

POST OFFICE BOX Y
OAK RIDGE, TENNESSEE 37830

September 27, 1972

Dr. E. T. Wanderer, Manager
Machinery & Equipment Application
Engineering Division
Aluminum Company of America
New Kensington, PA 15068

Dear Dr. Wanderer:

We are currently embarking on a study, funded jointly by the Atomic Energy Commission and the Office of Saline Water, which incorporates conceptual design and cost studies of a low temperature (120 to 180°F max) multieffect seawater evaporator. This program is the outgrowth of our earlier study of a single effect still constructed of aluminum alloys which we discussed with you in July 1970. Our current study will consider a plant utilizing all-aluminum construction and one with aluminum heat transfer surface and concrete shells.

Since it has been over two years since our last discussions we again seek your advice and recommendations for utilizing aluminum alloys for seawater distillation service. Some of the particular areas of interest are listed on the attached sheets. We would sincerely appreciate meeting with you and other members of your staff at New Kensington to discuss these items. Because our designs will hinge heavily on your recommendations we would like to meet with you sometime early in October, commensurate with your schedule.

Very truly yours,

S. A. Reed
Nuclear Desalination Program

SAR: saw

cc: D. C. Jacobs
E. C. Hise
✓J. E. Jones, Jr.
I. Spiewak

OAK RIDGE NATIONAL LABORATORY
OPERATED BY
UNION CARBIDE CORPORATION
NUCLEAR DIVISION

POST OFFICE BOX Y
OAK RIDGE, TENNESSEE 37830

September 27, 1972

Dr. Eric F. Barkman, Director
Metallurgical Research Division
Reynolds Metals Company
Fourth & Canal Streets
Richmond, VA 23218

Dear Dr. Barkman:

We are currently embarking on a study, funded jointly by the Atomic Energy Commission and the Office of Saline Water, which incorporates conceptual design and cost studies of a low temperature (120 to 180°F max) multieffect seawater evaporator. This program is the outgrowth of our earlier study of a single effect still constructed of aluminum alloys which we discussed with you in July 1970. Our current study will consider a plant utilizing all-aluminum construction and one with aluminum heat transfer surface and concrete shells.

Since it has been over two years since our last discussions we again seek your advice and recommendations for utilizing aluminum alloys for seawater distillation service. Some of the particular areas of interest are listed on the attached sheets. We would sincerely appreciate meeting with you and other members of your staff at Richmond to discuss these items. Because our designs will hinge heavily on your recommendations we would like to meet with you sometime early in October, commensurate with your schedule.

Very truly yours,

S. A. Reed
Nuclear Desalination Program

SAR:sw

Attachment

Low Temperature Evaporator Study

The present multieffect low temperature evaporator design investigation is an outgrowth of an earlier vertical tube still (VTS) study and proposes to operate in a temperature range intermediate between the VTS and the conventional multieffect vertical tube evaporator. The incentives for operating in this temperature range are:

1. Lower heat cost at lower temperature
2. Lower cost of chemical treatment of the seawater feed
3. Reduced materials problems

We further propose to examine the economics of a simplified process which accomplishes the regenerative feed heating in the evaporator surface and thus requires no separate feed heater—condenser type bundles. This reduces the plant almost to a shell or series of shells containing a series of evaporator bundles.

Hopefully these new design freedoms will permit the use of low cost-high performance materials; shop fabrication of standardized bundles of evaporator surface of a conventional or unconventional type; and the installation of those bundles in a very simple shell containing all the necessary flow channels except those for pumping the brine feed to the effects.

Specific Areas of Interest

1. Recommended alloys for fluted heat transfer tubing, tube sheets, vessels and internals.
2. Advantages and limitations of shop fabricated evaporator surface modules.
3. Could one fabricate a unit—say 10 ft x 10 ft x 10 ft—complete with brine chest, entrainment separator and mounting flange? We envision a square bundle with steam entry along one face only and vertical entrainment separators on that face.
4. Is steam impingement erosion on the first few tube rows a problem?
5. Is an estimated cost in $\$/ft^2$ of some such prefabricated unit possible?
6. May we obtain copies of tubing cost estimates which were prepared for OSW-A/E contractors for the 200-Mgd plant design studies?

7. Would it be possible and reasonable to shop fabricate penetrations to be cast in the concrete shell so that pipe could be later installed through the penetration and sealed in the field?
8. Would it be advantageous to prefabricate brine piping for field installation by welding or mechanical joining?
9. Recommendations for installing and connecting pump, valves, instruments. Is stainless steel the best material for pump and valves?
10. In relation to brine distribution devices, is ceramic or aluminum recommended? Is impingement on aluminum tubes a problem? What is the recommended method for retaining the nozzles?
11. Have you any design heat transfer coefficient recommendations based on experimental data or operating results?
12. Do you recommend heavy metal ion traps? If so what are best designs-costs?
13. Compatibility of Al-alloys and cement; means of making transition joints, etc.
14. We have E. D. Verink's papers summarizing results of Freeport, TX loop testing. Are these test data being used for alloy recommendations?
15. Is all-aluminum construction preferred over aluminum-concrete construction?

INTRA-LABORATORY CORRESPONDENCE

OAK RIDGE NATIONAL LABORATORY

November 8, 1972

TO: I. Spiewak

FROM: E. C. Hise

SUBJECT: Trip To Reynolds Metals Company, Richmond, Virginia, Oct. 17, 1972

Persons Present: Ray Lindberg, Materials Res. Director, and D. H. Fauth, Product Development

The discussions followed generally the summary and agenda attached.

1. The corrosion tests and the pilot plant operation indicate that there are several alloys in both the structural and extrudable classes that are completely adequate to the service. These alloys can be mixed to construct a plant as long as good practice is followed. Given a design, Reynolds would specify alloys and techniques.

2., 3., & 5 The same advantages of shop fabrication obtain for aluminum as for steel and conversely field fabrication of aluminum is no more difficult than steel. Shop fabrication of unit bundles is quite feasible. Aluminum barge fabricators were mentioned off handedly. A preliminary estimating rule of $1.00/pound for fabricated equipment was given.

4. They have observed no steam or brine impingement damage per se. One horizontal tube bundle has shown some damage which they are convinced is chemical damage from the amines in the boiler plant steam.

6. Reynolds did not submit a tubing cost estimate for the OSW-A/E contractors. However, they will respond to a cost inquiry.

7. & 8. Yes, penetrations and piping can be prefabricated in the same fashion that one would use for steel.

9. Wherever some metal other than aluminum must be used, they recommend stainless steel. Carbon steel does, of course, corrode and those corrosion porducts not only foul but also set up corrosion cells. Stainless steel pumps and valves can be flanged to aluminum pipe with a "donut anode" filler piece which serves as a large sacrificial anode. Small valves and thermal wells can be aluminum.

10. Both ceramic and aluminum nozzles are compatible. They prefer aluminum because of some unfortunate breakages. Brine inpingement has not been noted to be a problem. A perforated plate hold down is used.

11. The pilot plant is primarily a materials test unit and has not been instrumented adequately for heat transfer information. They have made measurements with the available instrumentation which indicated U = 1600 at 210° and 1350 at 170° evaporating.

12. The pilot plant has had for three years one heavy metal ion trap on the raw seawater feed prior to the holding tank. It is clad aluminum tubing constructed like a heat exchanger. They pay little or no attention to it other than to observe that it corrodes and the plant doesn't. They remarked that the pH in the deaerator is ideal for maximum copper scavenging and that aluminum deaerator packing might serve two purposes if the flowsheet could be suitably arranged.

13. Aluminum and concrete is a design problem that can be handled by observation of the rules. The steel re-bar is more troublesome than the concrete but in any event appropriate insulation is required.

15. An aluminum manufacturer prefers aluminum over concrete for obvious reasons. However, the two can be combined satisfactorily.

Incidental remarks and comments.

Because the tubing is formed by an extrusion process, it is just as easy for the internal and external fluting to be of different patterns if desired.

The final report on the pilot plant should be available from Callahan, OSW. Contract No. 14-30-2669, dated April 1972.

Reynolds is agreeable to fabricating experimental quantities of heat transfer surface at cost.

They emphasized the importance of being careful with raw seawater, dirt and plugging. This is of concern in the heat reject section of an evaporator where large quantities of water are required and treatment is uneconomical. A low temperature low P.R. aggravates the problem.

E. C. Hise

ECH:job
Attachment
cc: T. D. Anderson
J. T. Callahan, OSW
D. M. Eissenberg
R. P. Hammond
J. W. Hill
J. E. Jones
S. A. Reed

INTRA-LABORATORY CORRESPONDENCE

OAK RIDGE NATIONAL LABORATORY

November 8, 1972

TO: I. Spiewak

FROM: E. C. Hise

SUBJECT: Trip To Alcoa Technical Center, Pennsylvania, October 31, 1972

Persons Present: Gary C. Blaze) Application Engineering Div.,
E. T. Wanderer) New Kensington, Pa.

J. L. Brandt)
E. H. Hollingsworth) Chemical Metallurgy Div., Alcoa Research Center
R. A. Bonewitz)

The discussions generally followed the agenda attached.

1. There are suitable and proven alloys available for each class of service-extrusion, structural fabrication, etc. Specific alloys mentioned are 1100, 3003, 3004, 5050, 5052, and 6063. These can be mixed with appropriate care.

2., 3, & 5. Units such as the proposed heat transfer modules would conventionally be shop fabricated. The mix of shapes would average 75¢/lb and the fabricated cost would be ∿$1.50/lb. Alcoa indicated a willingness to cost estimate such a unit if supplied with adequate drawings.

4. Experience with aluminum tubed power plant condensers indicates that dry steam up to sonic velocity will not erode the tubes. Entrained liquid very definitely will. Baffles are used to knock down condensate.

6. Alcoa Technical Center did not supply cost estimates for tubing. They will quote on the basis of a drawing of the desired tube and a statement of quantity.

7., 13, & 15. Concrete to aluminum transitions can be satisfactorily made if appropriate precautions are observed. Pipe penetrations and studs should be coated before being placed in the concrete. Alcoa has published several papers on aluminum-concrete compatibility.[1,2]

---

[1]C. J. Walton, et. al., The Compatibility of Aluminum with Alkaline Building Products (Aluminum Company of America), 1957.

[2]F. L. McGeary, Performance of Aluminum in Chloride-Containing Concrete (Aluminum Company of America), 1965.

9. Pumps, valves, etc., should be stainless steel and the connections to aluminum piping should be with either insulated spool pieces or with sacrificial aluminum donuts. Both systems have been satisfactorily demonstrated.

10. Alcoa has no experience with brine distributors for vertical tubes.

11. Alcoa has made no heat transfer measurements of vertical tubes.

12. It is necessary to remove copper and heavy metals from the feed and desirable to monitor the effectiveness of the removal system.

14. Alcoa is, of course, following the work of Verink and using his data along with their own.

15. Aluminum-concrete construction is much preferred over aluminum-steel construction because of the deleterious effects of any iron corrosion products in the liquid or vapor portions of the plant and because of the extreme difficulty of preventing such corrosion products from forming and circulating. All aluminum is preferred over aluminum-concrete because of the rather large difference in thermal expansion coefficients. They felt that aluminum construction could be competitive or nearly so with the cost of adequately protected steel or concrete. They pointed out that aluminum is competing with epoxy coated steel in certain applications. They offered to cost estimate an aluminum structure if adequate drawings were supplied.

General Comments and Observations

The tests and experience of the last several years led Alcoa to believe that aluminum can be a completely satisfactory material for seawater evaporator service.

They evidenced a willingness to provide some engineering assistance and cost estimating, especially as related to large production.

E. C. Hise

ECH:job
Attachment
cc: T. D. Anderson
J. T. Callahan, OSW
D. M. Eissenberg
R. P. Hammond
J. W. Hill
J. E. Jones
S. A. Reed

[illegible]

March 30, 1973

Mr. E. C. Hise
Oak Ridge National Laboratory
P. O. Box "Y"
Oak Ridge, Tennessee 37830

Dear Mr. Hise:

As we recently discussed, enclosed are four copies of our drawing, G-100081-AE, which shows a rough design we have worked up for an aluminum evaporator shell for your modular plant design. As you can see, we have been unable to incorporate any special design features taking advantage of aluminum's special extrudability which might differ from steel. In the case of part 12, it appears this could be made from two T-shaped extrusions in aluminum as compared to three parts for steel and the result would be a saving of at least one weld. Considering our very superficial approach to the design, I would like to suggest that your designers also determine what the wall thicknesses should be for the aluminum heads, side walls, etc. As you know, our main reason to look at the design was in hopes that we might be able to propose some unique features representative of aluminum's workability characteristics over steel. Unfortunately, we were not able to implement any such features.

Regarding metal cost, all the aluminum plate materials have a selling price of approximately $.46 per pound. The 9" channels and tees have a price of about $.49 per pound. The 24" extrusions would have a selling price of about $.85 per pound. The tubes would be approximately $.54 per pound.

After reviewing some of our own experiences and talking to a few fabricators of shell and tube aluminum heat exchangers, we estimate the cost of fabrication to be about $.80 to $1.10 per pound of aluminum. The metal cost must then be added to this.

The alloys we would recommend are 5454-0 for shell material, 6061-T6 for tube sheets and 9" extruded structural shapes,

6063-T6 for the 24" extruded structural shapes, and 3003, 5052, or 6063 for the tubing.

I hope this information will be helpful to your program. Mr. E. T. Wanderer is planning a visit to Oak Ridge on April 9 regarding the uranium enrichment program and I suggest you contact Ed if you would like to meet with him while he is there. Ed has indicated he will be on a fairly tight schedule but perhaps a meeting can be worked out if it is needed.

Very truly yours,

Gary C. Blaze
Application Engineering Division

GCB/oar

Enclosure

cc: Mr. D. C. Jacobs
Alcoa
P. O. Box 128
Alcoa, Tennessee 37701

Mr. E. T. Wanderer, Alcoa, New Kensington