CACHE—An Extended BASIC Program Which Computes the Performance of Shell and Tube Heat Exchangers

J. R. Tallackson

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CACHE — AN EXTENDED BASIC PROGRAM WHICH COMPUTES THE
PERFORMANCE OF SHELL AND TUBE HEAT EXCHANGERS

J. R. Tallackson

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ABSTRACT

An extended BASIC program, CACHE, has been written to calculate steady state heat exchange rates in the core auxiliary heat exchangers, (CAHE), designed to remove afterheat from High-Temperature Gas-Cooled Reactors (HTGR). Computationally, these are unbaffled counterflow shell and tube heat exchangers. The computational method is straightforward. The exchanger is subdivided into a user-selected number of lengthwise segments, heat exchange in each segment is calculated in sequence and summed. The program takes the temperature dependencies of all thermal conductivities, viscosities and heat capacities into account providing these are expressed algebraically. CACHE is easily adapted to compute steady state heat exchange rates in any unbaffled counterflow exchanger. As now used, CACHE calculates heat removal by liquid weight from high-temperature helium and helium mixed with nitrogen, oxygen and carbon monoxide.

A second program, FULTN, is described. FULTN computes the geometrical parameters required as input to CACHE. As reported herein, FULTN computes the internal dimensions of the Fulton Station CAHE. The two programs are chained to operate as one.

Complete user information is supplied. The basic equations, variable lists, annotated program lists, and sample outputs with explanatory notes are included.

KEYWORDS: heat transfer - heat exchangers - spiral tubes - cross flow - helium - programming - helical tubes - afterheat removal - counterflow - gas cooling - reactor safety - afterheat removal - basic programming - HTGR.

CACHE — AN EXTENDED BASIC PROGRAM WHICH COMPUTES THE
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Introduction

The CACHE program was written, specifically, to determine the steady-state performance of the Core Auxiliary Heat Exchangers¹ (CAHE) in High-Temperature Gas-Cooled Reactors (HTGR). These gas-to-water exchangers, with spirally wound water tube bundles, are of somewhat unconventional design. Gas flow outside the tubes is essentially axial and unidirectional. It is assumed that the internal geometry is uniform along the axis; hence, the computational model may be represented by the familiar conventional concentric shell and tube device, Fig. 1.

Fig. 1. Model of typical counterflow heat exchanger

CACHE, listed and described with detail in Appendix A, takes into account the temperature dependencies of the fluid properties and the tube material. Fluid and material properties and film coefficients of heat transfer are specified by equations, not by tabulations. The user must specify the inlet water temperature and also specify or estimate the exit gas temperature. Extensive, detailed information printout capability, optional with the user, is available. CACHE now requires that it be preceded by a program which is used to develop and store the detailed internal dimensions of the CAHE and, in addition, provides a means for the user to input the flow parameters. This program is described in Appendix B.

The CACHE program can be used to obtain heat transfer calculations for any conventional counterflow heat exchange assuming that the user specifies the fluid and material properties, heat transfer correlations and cold end temperatures. Modifications enabling the user to specify hot end temperatures will, in all likelihood, be minor, involving sign changes to obtain subtractions instead of additions.

The program is not presented as being thoroughly optimized. It has, however, become an effective design tool that has produced results required to evaluate CAHE capabilities.

Heat Transfer Computations

A. Nomenclature and Symbols

C_p	= specific heat at constant pressure	Btu/lb-°F
k	= thermal conductivity	Btu/hr ft ² -°F/ft
μ	= viscosity	lb/ft-hr
W	= total flow rate	lb/hr
A_f	= cross-sectional area for flow	ft ²
D_e	= equivalent flow diameter	ft
	$(D_e)_1$ = I. Diameter of tubing	
	$(D_e)_3$ = O. Diameter of tubing	
G	= mass flow rate = $\frac{W}{A_f}$	lb/ft ² -hr
C	= leading coefficient in equations for N_{Nu}	dimensionless
m	= exponent of N_{Re} in equations for N_{Nu}	dimensionless
N_{Re}	= Reynolds number = $\frac{D_e G}{\mu}$	dimensionless
N_{Pr}	= Prandtl number = $\frac{C \mu}{k}$	dimensionless
N_{Nu}	= Nusselt number = $\frac{h D_e}{k}$	dimensionless
A_1	= water side tube surface area	ft ²
A_2	= logarithmic mean tube wall heat transfer area	ft ²
A_3	= gas side tube surface area	ft ²
h_1	= local water side film coefficient of heat transfer	Btu/hr ft ² -°F
h_2	= local tube wall coefficient of heat transfer	Btu/hr ft ² -°F
h_3	= local gas side film coefficient of heat transfer	Btu/hr ft ² -°F
U	= overall heat transfer coefficient	Btu/hr ft ² -°F
T_1	= bulk temperature of lower temperature fluid, (water)	°F
T_2	= average temperature of the tube wall	°F
T_3	= bulk temperature of higher temp. fluid, gas	°F

B. The Heat Transfer Coefficients

The film coefficients for both helium and water are determined, as usual, by evaluating correlating equations for the Nusselt No. as a function of the Reynolds No. and the Prandtl No. The general form of these Nusselt No. equations is:

$$N_{Nu} = C N_{Re}^m N_{Pr}^n \quad (1)$$

The water side film coefficient, h_1 , is computed from the familiar and well established Dittus-Boelter type equation for turbulent flow in straight pipes, viz;

$$(N_{Nu})_1 = 0.023 (N_{Re})_1^{0.8} (N_{Pr})_1^{0.4} \quad (2)$$

The spiral tubes would be expected to produce slightly higher water side film coefficients than straight tubes. However, the gas side film coefficient controls heat transfer in these systems; therefore, any small improvement obtained with spiral flow is not significant.

The axes of the tubes in these heat exchangers are very nearly normal to the direction of gas flow; therefore, equations for the Nusselt number developed for cross flow of gases over banks of parallel tubes apply. In the CACHE program the Nusselt number is calculated using

$$(N_{Nu})_3 = F_n C_3 (N_{Re})_3^m (N_{Pr})_3^{0.333} \quad (3)$$

F_n = a factor, dimensionless, determined by the number of tube rows counted in the direction of gas flow (Ref. 2).

C_3 and m = constants depending on tube spacing and whether tubes are "in-line" or "staggered" (Ref. 3).

N_{Re} and N_{Pr} evaluated at the mean film temperature of the gas (Ref. 3).

The evaluation of C_3 and m is based primarily on the original study by Grimison³ based on experiments conducted by Hoge⁴ and Pierson⁵ who studied heat exchange in banks of straight tubes 10 rows deep in the direction of gas flow. Sanders and Hoppes,⁶ using Grimison's tabulated values of C and m augmented by more recent European data, derived polynomial fitting equations which have been used to obtain the values of C_3 and m in Eq. (3).

Grimison's tabulation³ and Sanders' polynomial fitting equations both include the assumption that the gas side Prandtl number is constant or nearly so and has a value of 0.69. In CACHE, the Prandtl numbers are considered to be variable and are computed directly each time the Nusselt numbers are calculated. Therefore, the leading constant, C_3 , is obtained by multiplying the corresponding constant, C'_3 , derived from Sanders' polynomials by 1.13. This is illustrated by equating the two methods of computing the Nusselt number for air or similar gases having a constant Prandtl number of 0.69. The expression on the left side below is Eq. (3); that on the right is from Grimison.³

$$F_n C_3 (N_{Re})_3^m (N_{Pr})_3^{0.33} \equiv C_3^! (N_{Re})_3^m$$

in which

$$N_{Pr} = 0.69$$

$$F_n = 1.0 \text{ (for 10 tube bank depth)}$$

$$C_3^! = \text{leading constant, from Ref. 3 or Ref. 6.}$$

therefore

$$C_3 = C_3^! / (N_{Pr})^{0.33} = C_3^! / (.69)^{0.33}$$

$$C_3 = 1.13 C_3^!$$

Because Eq. (3) is for straight tubes, 10 tube rows in depth, it is not directly applicable to CAHE tube geometry. The available and widely accepted heat transfer correlations for cross-flow in tube banks are for arrays of tubes with absolutely uniform tube-to-tube spacing throughout the array. Such is not the case in a CAHE tube bank composed of spirally wound tubes in the annular space between coaxial cylindrical shells. This is true even when the tubes are coiled so that the average density or space occupancy of the tubes is uniform and is made uniform by coiling the tubes so that:

- (1) The tubes in any set of coils having a common helix diameter are arranged to have a uniform tube-to-adjacent-tube spacing measured axially in the direction of gas flow.
- (2) All sets of spiral coils, regardless of helix diameter, are wound to have identical spacing per (1) above.
- (3) The radial separation, tube-to-tube, of the helical sets of coils is uniform.

In other words, vertical sections by cutting planes which contain the shell axis, and therefore, the common axis of the spiral tube sets, would show that the tube-to-tube geometry is characterized by a degree of local randomness, even though the average tube occupancy in any relatively small region is uniform. Moreover, there is substantial incentive to design the tube array with a degree of random non-uniformity to prevent the development of regular, organized vortex shedding patterns capable of producing severe tube vibration.

The Nusselt number equation, (Eq. 3), is based on the assumption that the tube array is absolutely regular and that the tube-to-tube arrangement is either rectangular or triangular. The random aspect of spiral arrays has been taken into account by assuming that a fraction of the tubes are arranged in a triangular (staggered) pattern and the remainder are pitched rectangularly (in-line). The fractional division between the two types of arrays is now somewhat arbitrary and, currently, it is

assumed for HTGR CAHEs that 25% of the total tubes form a staggered array. In CACHE the gas side Nusselt number is, therefore, determined by a weighted average of the Nusselt numbers for in-line and staggered arrays, thus:

$$(N_{Nu})_{CAHE} = 0.25 [N_{Nu}]_{stag} + 0.75 [N_{Nu}]_{in-line} \quad (4)$$

$[N_{Nu}]_{stag}$ and $[N_{Nu}]_{in-line}$ are calculated with Eq. (3). C_3 in this equation is for arrays ten-tube rows deep counted in the direction of gas flow. A typical CAHE will be 90 to 100 rows of tubes in depth. F_n takes the effect of this difference into account.

The heat transfer coefficient across the tubing walls is given by

$$h_2 = \frac{k_2}{R_o - R_i} \quad , \quad \text{Btu/hr ft}^2\text{-}^\circ\text{F/ft} \quad (5)$$

R_o, R_i = outside and inside radii of the tubing, ft.

To summarize, the heat transfer coefficients, h , across the water film, the gas film, and the tube wall are calculated with Equations (2), (4) and (5), respectively. Because the thermal transport properties⁷ of water, tubing, and gases are temperature dependent the heat transfer coefficients are evaluated locally at a multiplicity of points representing small regions within the heat exchanger. From the values of h so obtained, overall coefficients U_1 and U_3 are computed, thus:

$$\frac{1}{U_1} = \frac{1}{h_1} + \frac{A_1}{A_2} \left(\frac{1}{h_2} \right) + \frac{A_1}{A_3} \left(\frac{1}{h_3} \right) \quad (6)$$

$$\frac{1}{U_3} = \frac{A_3}{A_1} \left(\frac{1}{h_1} \right) + \frac{A_3}{A_2} \left(\frac{1}{h_2} \right) + \frac{1}{h_3} \quad (7)$$

In these equations,

U_1 = overall heat transfer coefficient, helium to water, referenced to A_1 , the water side surface area; Btu/hr ft²-°F/ft

U_3 = overall heat transfer coefficient, helium to water, referenced to A_3 , the helium side surface area; Btu/hr ft²-°F/ft

The total heat transferred is,

$$Q = U_1 A_1 \Delta T \quad , \quad \text{Btu/hr} \quad (8)$$

or

$$Q = U_3 A_3 \Delta T \quad , \quad \text{Btu/hr} \quad (9)$$

$$\Delta T = T_3 - T_1 \quad , \quad ^\circ\text{F}$$

From heat balance considerations:

$$Q_1 = W_1 (C_p)_1 \Delta T_1 \quad (10)$$

$$Q_3 = -W_3 (C_p)_3 \Delta T_3 \quad (11)$$

ΔT_1 and ΔT_3 are bulk temperature changes in the water and gas, respectively, as they traverse the region in the heat exchanger with surface areas A_1 and A_3 .

C. The Computational Method

C.1 The Model and Assumptions

The heat exchanger is subdivided into an arbitrary number of incremental lengths, or segments, Fig. 2. The heat transferred in each segment is calculated, iteratively, until the temperature dependent heat flux, Eq. (8) or (9), agrees satisfactorily with a heat balance computation, Eq. (10). CACHE* requires that the computations begin at the low temperature end and proceed in the direction of increasing temperatures.

It is assumed that the fluid and material properties within any short segment in the exchanger are well represented by evaluating them at an average temperature at a midplane through the segment normal to the heat exchanger axis. It is also assumed that, in any segment, the driving force for heat transfer is the arithmetic mean temperature difference of the bulk fluids at the segment midplane; that is, in a short segment, the error produced by using the arithmetic mean instead of the log mean temperature difference is negligible. Segment length must be chosen accordingly. This is a matter of user judgment.

C.2 The Step-by-Step Procedure

First, the number of incremental lengths is decided upon. For a given design in a particular design, there will exist an

*The geometrical and flow parameters are calculated or established and then filed by a preceding program chained to the CACHE program. For calculations involving the Fulton Generating Station CAHE, this program is currently named FULTN and is described in Appendix B.

Fig. 2. Diagram of the model used to calculate temperatures and heat transfer rates in an HTGR Core Auxiliary Heat Exchanger

- (a) Temperature symbols, subscripts omitted, are as they appear in CACHE
- (b) Relevant line numbers in CACHE are in parentheses

optimum band for the number of segments above which computational time increases and no meaningful improvement in computational accuracy is produced and below which the results may be without sufficient accuracy or detail. It can be demonstrated that an increase in segment length with the consequent decrease in the number of increments to be considered by the computation does not necessarily reduce the total time required to complete a computation. Longer segments may require more iterations to complete a computation within a segment.

Boundary temperatures are established, for computational purposes, at the low temperature end of the heat exchanger. Two cases are permitted; these are:

Case A. The user is able to specify the correct inlet water temperature and the correct outlet gas temperature.

Case B. The user is able to specify the correct inlet water temperature and the correct inlet gas temperature.

The Case A calculation is the simpler. The computation merely proceeds, segment by segment, from the cold end to the hot end of the exchanger. Recent CAHE calculations have been for Case B conditions. In these circumstances, the program completes a Case A computation using an estimated gas outlet temperature as an initial boundary, resets this gas outlet temperature in the direction that produces a more nearly correct gas inlet temperature and repeats a Case A computation based on the reset gas outlet temperature. This process is iterated until the calculated gas inlet temperature is, within a reasonable user specified tolerance, equal to the required gas inlet temperature.

Figure 2, the diagram of the computational model, shows the locations and symbols of the temperature parameters used in CACHE. The subscript designates the segment number and does not appear in the CACHE list, Appendix A, as a subscript.

To begin a computation the user must establish boundary temperatures in the first segment and initialize computed temperatures with educated guesses. The required temperatures are:

- (T1) bulk water temperature at the segment midplane
- (T2) bulk water temperature entering the segment
- (T3) bulk gas temperature at the segment midplane
- (T4) bulk gas temperature leaving the segment
- (T6) average tube wall temperature at the segment midplane
- (T7) gas side tube surface temperature at the segment midplane
- (T8) water side tube surface temperature at the segment midplane
- (T9) mean gas film temperature at the segment midplane
- (P4) inlet gas temperature (at hot end of the CAHE)
- (P2) (P2) = (T2), see above.

The equations or instructions used in CACHE to establish boundary and initial temperatures, Fig. 2, are listed below. CACHE symbols and equations are shown wherever applicable as well as their counterparts listed in Section A.

$$(T1)_1 = (T2)_1 + 260/(2 * MO)$$

The maximum increase in CAHE water temperature from the inlet at $(T2)_1$ to the outlet is about 260°F. The above equation linearly interpolates this temperature rise at the midplane of the first segment. MO is the number of segments.

$(T2)_1 = P2$; an established system parameter; an INPUT P2 request to the user each time the program is run.

$$(T3)_1 = (T4)_1 + 1050/(2 * MO)$$

A typical value for the temperature change of helium, inlet to outlet, was taken to be 1050°F. This equation for $(T3)_1$ linearly interpolates this temperature change at the midplane of the first segment. MO is the number of segments.

$(T4)_1$: Case A; an established system parameter.

Case B; any reasonable estimate by the user.

$(T4)_1$ is an INPUT T4 request to the user each time the program is run.

$(T6)_1 = (T7)_1$, see $(T7)_1$ below.

$$(T7)_1 = (T3)_1 - 0.75 * [(T3)_1 - (T2)_1]$$

This equation is based on the assumption that approximately 75% of the temperature difference, gas to water, is across the gas film and neglects the difference between the exit and midplane water temperatures.

$(T8)_1 = (T7)_1$, see $(T7)_1$ above.

$$(T9)_1 = 0.5 * [(T3)_1 + (T7)_1]$$

This is the mean gas film temperature.

P4 ; P4 is an INPUT P4 request by the program each time the program is run. For Case A runs any reasonable value is used since its only function is to maintain program operation; i.e., it is not used in the calculations. For Case B runs it is a system parameter.

Only the system parameters need be specified precisely; these are $(T2)_1$, and either $(T4)_1$ or P4 depending on whether the calculation is for Case A or Case B.

Having established an initial, average difference in the bulk temperatures of the two fluids,

$$\Delta T \equiv (T3)_1 - (T1)_1$$

the program calculates the gas, tube wall, and water properties dependent on $(T1)_1$, $(T3)_1$, $(T6)_1$ and $(T9)_1$. The heat transfer coefficients, h_1 , h_2 and h_3 and U_3 are computed with Eqs. (2), (5), (4) and (7), respectively. A first approximation of the heat transferred in the first segment is, from Eq. (9)

$$Q'_1 = Q(1) = U_3 A_3 \Delta T = U_3 A_3 [(T3)_1 - (T1)_1]$$

$$Q(1) = \text{heat transferred, Btu/hr}$$

$$(h_1)_1 = H1 = \text{water side film heat transfer coefficient, Btu/hr ft}^2\text{-}^\circ\text{F-ft}$$

$$(h_2)_1 = H2 = \text{tube wall heat transfer coefficient, Btu/hr ft}^2\text{-}^\circ\text{F/ft}$$

$$(h_3)_1 = H3 = \text{gas side film heat transfer coefficient, ft}^2$$

$$A_1 = A1 = \text{segment surface area, water side, ft}^2$$

$$A_3 = A3 = \text{segment surface area, gas side, ft}^2$$

$$A_2 = A2 = \text{segment tube wall log mean area}$$

$$A_2 = \frac{(A_3 - A_1)}{\log (A_3/A_1)}$$

This approximation of the heat transfer rate, $Q(1)$, is used in heat balance computations with Eqs. (10) and (11) to compute temperature changes in the gas, tube wall, and the water. These temperature changes are used to readjust or reset all temperatures, except $(T2)_1$ and $(T4)_1$, which were used to initiate the computation. $Q(1)$ is retained in storage for use in the next computation of heat transfer in the first segment.

The equations used to revise the first segment temperatures are:

$$(T1)_1 = (T2)_1 + 0.5*Q(1)/(W1*S1) \quad (12)$$

$$(T3)_1 = (T4)_1 + 0.5*Q(1)/(W3*S3) \quad (13)$$

$$(T7)_1 = (T3)_1 - Q(1)/(A3*H3) \quad (14)$$

$$(T8)_1 = (T1)_1 + Q(1)/(A1*H1) \quad (15)$$

$$(T6)_1 = 0.5*[(T7)_1 + (T8)_1] \quad (16)$$

$$(T9)_1 = 0.5*[(T3)_1 + (T7)_1] \quad (17)$$

$Q(Y)$ = heat transfer rate in the segment
 Y = number of the iteration in the segment

$T1 = T_1$, typical, all temperatures; e.g., $T3 = T_3$ etc.

$W1 = W_1$ = total water flow rate, lb/hr

$W3 = W_3$ = total gas flow rate, lb/hr

$S1 = (C_p)_1$ = heat capacity, water, Btu/lb-°F

$S3 = (C_p)_3$ = heat capacity, gas, Btu/lb-°F

Equations (12) through (17) are in the order they are used in CACHE; therefore, the temperatures, $(T3)_1$ and $(T1)_1$, on the right hand side of Eqs. (14) and (15) are the revised values computed with Eqs. (13) and (12). Similarly, $(T7)_1$ and $(T8)_1$ in Eqs. (16) and (17) are the revised temperatures calculated with Eqs. (14) and (15).

Having established a new set of temperatures with Eqs. (12) through (17), a revised heat transfer rate, $Q(2)$ in the 1st segment, is calculated as described. This second evaluation, $Q(2)$ is compared with the first to determine if the following inequality is satisfied:

$$\frac{Q(Y+1)}{Q(Y)} - 1 = \frac{Q(2)}{Q(1)} - 1 \geq \epsilon ; \text{ line 4620 in CACHE} \quad (18)$$

ϵ = acceptable error tolerance* established by user.

Unless the user has been unusually clairvoyant and has made extremely accurate first estimates of the first segment temperatures, the specified tolerance will not be achieved with the completion of the first iteration. $Q(2)$ is used to revise the temperatures with Eqs. (12) to (17) incl. and is stored to be used in Eq. (18) after the 2nd iteration. This iterative procedure continues until Eq. (18) is satisfied.

The computation now advances to the second segment. The procedure is identical to that just described for the first segment but with one exception; the establishment of initial temperatures in the second and in all additional segments is accomplished by the program. The equations, lines 6400 through 6550 in CACHE, in Appendix A, which establish approximate initial temperatures in all segments but the first are:

$$(T2)_{n+1} = (T2)_n + Q(Y)_n / (W1 * S1)_n \quad (19)$$

$$(T4)_{n+1} = (T4)_n + Q(Y)_n / (W3 * S3)_n \quad (20)$$

*The error tolerance, ϵ , in current use for CAHE calculations, is 0.001.

$$(T1)_{n+1} = (T1)_n + Q8/(W1*S1) \quad (21)$$

$$(T3)_{n+1} = (T3)_n + Q8/(W3*S3) \quad (22)$$

$$(T8)_{n+1} = (T1)_{n+1} + Q8/(A1*H1) \quad (23)$$

$$(T7)_{n+1} = (T3)_{n+1} - Q8/(A3*H3) \quad (24)$$

$$(T6)_{n+1} = [(T7)_{n+1} + (T8)_{n+1}]/2 \quad (25)$$

$$(T9)_{n+1} = [(T3)_{n+1} + (T7)_{n+1}]/2 \quad (26)$$

$$Q8 = Q(Y)_n * (1 + A3 * (U3)_n * ((1/(W3*S3)) - (1/(W1*S1)))) \quad \text{See footnote (a)}$$

Subscripts n and $(n+1)$ refer to the n th and $(n+1)$ th segment. Eqs. (19) and (20) are written with the assumption that fluid properties and the heat transfer rate are uniform within each small segment and, therefore, the axial rates of change of gas and water temperatures within the segment are uniform. Eqs. (21) through (24) attempt to take into account the nonlinear variation of fluid temperatures which has the effect of increasing the rate of heat transfer as the computation advances toward the hot end of the CAHE.

Heat transfer rates in all the remaining segments are computed as in segment number two just discussed. Segment heat rates are summed, segment by segment, as the calculation proceeds. When the heat transfer rate in the last segment has been computed a Case A calculation, based on specified cold end temperatures, $(T2)_1$ and $(T4)_1$, is complete.

A Case B computation for which the inlet water and inlet gas temperatures are specified is accomplished by a repetitive sequence of Case A computations which converge to the specified gas inlet temperature. At the start of the run the user is required to provide a very crude estimate of the gas outlet temperature, $(T4)_1$, associated with this specified inlet temperature.

The run is initiated as for a Case A computation except the user responds to the "INPUT P4" instruction with the desired value of gas inlet temperature. The first Case A computation is made with the user estimated gas outlet temperature, $(T4)$, and on completion, arrives at an incorrect value for the required gas inlet temperature P4. The estimated gas outlet temperature is corrected with this relation:

$$[(T4)_1]_{j+1} = [(T4)_1]_j + R2 \quad (27)$$

Subscripts j and $(j+1)$ designate the number of the Case A run (loop) in which the value of $T4$ will be or has been used.

(a) This equation for $Q8$ takes into account the non-linear variation of the heat transfer rate.

$j = 1, 2, \dots, N$; N = number of loops required for convergence to P_4 .

The equation for R_2 , the additive correction to the outlet gas temperature, is:

$$R_2 = \frac{(P_4 - E_4)(1 - C_8)}{(C_6 - C_8)} \quad (28)$$

in which

P_4 = the specified gas inlet temperature, °F

E_4 = the gas inlet temperature computed with the j th computation

$C_6 = \frac{(\Delta T)_{\text{hot end}}}{(\Delta T)_{\text{cold end}}}$, from the j th computation

$C_6 = (E_4 - E_2)/(J_4 - P_2)$

$C_8 = W_3(C_p)_3 / W_1(C_p)_1 = W_3 S_3 / W_1 S_1$

The diagram, Fig. 3, shows the locations of these temperatures.

Fig. 3. Successive Case A axial temperature profiles in a CAHE during convergence of a Case B calculation

When the absolute value of the difference between the computed and specified gas inlet temperatures is sufficiently small, the Case B computation is considered complete. The size of the difference or error is at the user's disposal. For CAHE calculations this difference has been specified to be 1.0°F or less.

Extensive printout is provided by CACHE. If the operator so desires, these results may be printed after the computation in each segment is completed.*

1. The thermal transport properties of the gas and tubing at the segment midplane temperatures.
2. The Reynolds and Prandtl numbers, at the segment midplane, of the gas and water.
3. The gas film heat transfer coefficients, $h_{in-line}$ and h_{stag} , Eq. (3) and H3, their weighted average; the tube wall and water film heat transfer coefficients, H2 and H1. These are at segment midplane temperatures.
4. The overall heat transfer coefficients, U, referenced to water and gas side surface areas.
5. The heat transfer rate within the segment and the total heat transferred by all segments previously computed; i.e., if the user has divided the CAHE into 20 segments, the printout for the 10th segment includes the total heat transferred by the first 10 segments.
6. The bulk gas and water temperatures entering, at midplane, and leaving the segment and the mean gas film temperature at the segment midplane.
7. Segment averaged values of the bulk water and gas temperatures, mean gas film temperature, gas side Reynolds and Nusselt numbers, gas film heat transfer coefficient, the gas film viscosity and thermal conductivity. These quantities, computed at the segment midplanes, are summed, segment by segment, as the computation proceeds and arithmetically averaged by dividing by the total segments in which heat transfer has been computed; e.g., a 10th segment printout will include the arithmetic averages of 10 computations of the above quantities calculated at the segment midplanes.
8. The segment number, the location of the segment midplane as a fraction of the overall length of the exchanger, the total length of all segments in which heat transfer rates have been computed, and the number of iterations required to satisfactorily complete the calculation for the particular segment.

*In general, if the run is on-line with teletype printout, user patience is sorely tried if more than 3 or 4 segments are selected for printout.

On completion of a Case A run, the final printout contains these data:

1. The temperature differences, $(\Delta T)_h$ and $(\Delta T)_c$, bulk gas to bulk water, at the hot and cold ends of the CAHE.
2. The log mean temperature difference in the CAHE defined to be:

$$\bar{T}_{\log \text{ mean}} = \frac{(\Delta T)_h - (\Delta T)_c}{\log(\Delta T)_h - \log(\Delta T)_c} \quad (29)$$

In the strict sense this equation is only correct when the thermal transport properties of the materials in the system are temperature independent.

3. Average values of the water film coefficient, H1; the tube wall coefficient, H2; and the gas film coefficient, H3. These are arithmetic averages of the midplane heat transfer coefficients in each segment.
4. An overall heat transfer coefficient, referenced to the gas side surface area and based on the log mean temperature difference, i.e.:

$$\bar{U}_{\log \text{ mean}} = R7 = Q_{\text{total}} / (A_{\text{eff}} \bar{T}_{\log \text{ mean}}) \quad (30)$$

$Q_{\text{total}} = Q0$, the total calculated heat exchange rate

$A_{\text{eff}} = A$, the total gas side surface area credited with heat transfer

5. An overall heat transfer coefficient, referenced to the effective gas side surface area, and based on the arithmetic average of the overall temperature differences, gas to water, in each segment.

$$\bar{U}' = Q0 \cdot M0 / (A \cdot (\phi3 - \phi1)) = Q_{\text{tot}} / \left((A_{\text{eff}})^{(1/N)} \sum_{j=1}^N (\Delta T)_j \right) \quad (31)$$

$N = M0 = \text{number of segments}$

6. An overall heat transfer coefficient, referenced to the gas side surface area and based on the arithmetic average of the overall coefficients computed at each segment.

$$\bar{U} = U7/M0 = (1/N) \sum_{j=1}^N U_j \quad (32)$$

7. An average overall temperature difference based on $\bar{U}$, (Eq. 32), the total heat transfer rate, and the effective area, viz:

$$\bar{T} = Q_0 M_0 / (A \bar{U}) = Q_{\text{tot}} / \bar{U} A_{\text{eff}}$$

8. The effective gas side surface area, A .
9. The total number of iterations required to complete the computation.
10. The arithmetic average of the water side Reynolds numbers computed at each segment.

Case B runs, consisting of a sequence of Case A runs, provide similar printout. As now written and listed in Appendix A, the results of the first and last Case A run in the series are printed. In between these runs, as the computation is converging, the revised values of the gas outlet temperature, as computed by Eq. (27) and Eq. (28), are printed after each Case A run as is the percent error in the calculated gas outlet temperature after the previous run. The total number of Case A runs and the grand total segment iterations required to obtain convergence are printed.

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APPENDIX A

Block Diagram, Symbols, Program List, and
Typical Output by the CACHE Program

Figure A-1 is a block diagram showing only the bare computational procedure described in the main section of this report. It is not intended to serve as a finely detailed map of the program. The list and output that follow are thoroughly annotated and are intended to provide the instructions and explanations that may be needed by a user.

Fig. A-1. Block diagram of CACHE program

LIST OF SYMBOLS IN THE CACHE PROGRAM

If (FULTN) appears with the symbol, the variable or parameter is established with the FULTN program (Appendix B), stored in the file EXCH, and transmitted to CACHE with a "CHAIN CACHE" statement.

- A, (FULTN); "Effective" gas side heat transfer area of the heat exchanger tubes, ft^2 .
- A0, (FULTN); Total gas side surface area of the helically wound portions of all heat exchanger tubes, ft^2 .
- A1, (FULTN); Effective water side heat transfer area of all the heat exchanger tubes in an axial length segment, (See Fig. 2), ft^2 .
- A2, (FULTN); Effective tube wall heat transfer area of all the heat exchanger tubes in an axial length segment, (See Fig. 2), ft^2 .
- A3, (FULTN); Effective gas side heat transfer area of all the heat exchanger tubes in an axial length segment, (See Fig. 2), ft^2 .
- A7, (FULTN); Total transverse area for water flow, ft^2 .
- A9, (FULTN); Total transverse area for gas flow, ft^2 .

- B; A counter used to schedule the printout of Case B computations.
- B7; Inverse of the total heat capacity per hour of the cooling water; $1/(W1*S1)$, $1/((\text{lb/hr}) * (\text{Btu/lb-}^\circ\text{F}))$.
- B9; Inverse of the total heat capacity per hour of the gas; $1/(W3*S3)$, $1/((\text{lb/hr}) * (\text{Btu/lb-}^\circ\text{F}))$.

- C1; Leading constant in the equation for the water film Nusselt number, dimensionless.
- C3; Leading constant in the equation for the gas film Nusselt number for gas flow across in-line tube arrays, dimensionless.
- C5; Leading constant in the equation for the gas film Nusselt number for gas flow across staggered tube arrays, dimensionless.

- C6; A ratio of temperatures in the equation, (Eq. 28 and Fig. 3), used to reset the gas outlet temperature in a Case B calculation, refer to line 1188, dimensionless.
- C8; The ratio of the heat capacities per hour of the gas flow to the water flow; used in Eq. 28, (refer to line 1187), dimensionless.
- D0, (FULTN); Inside diameter of the heat exchanger tubing, in.
- D1, (FULTN); Inside diameter of the heat exchanger tubing, ft.
- D2, (FULTN); Outside diameter of the heat exchanger tubing, in.
- E, (FULTN); Fraction of total tube area considered to be effective.
- E2; Temperature of the water at the outlet end of a segment, Fig. 2, °F.
- E4; Temperature of the gas at the inlet end of a segment, Fig. 2, °F.
- E9; Convergence criterion for the allowable error in the final value of the inlet gas temperature in a Case B computation, Fig. 2, °F.
- F0; Tube bank factor, (F_n in Eq. 3), dimensionless.
- F3; Ratio, raised to the $M4^{th}$ power, of the bulk gas absolute temperature to the outside tube wall absolute temperature, dimensionless. NOTE, this factor, shown equal to 1 in the CACHE list, is not applicable to current CAHE computations with CACHE, dimensionless.
- F5; Gas film Nusselt number for staggered tube arrays, Eq. 4, dimensionless.
- F9; Gas film Nusselt number for in-line tube arrays, Eq. 2, dimensionless.
- G1; Mass flow rate of water, lb/ft^2-hr
- G3; Mass flow rate of gas, lb/ft^2-hr
- H1; Heat transfer coefficient of the water film, $Btu/hr\ ft^2-°F$

- H2; Heat transfer coefficient of the water film, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- H3; Heat transfer coefficient of the gas film, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- H7; Sum, segment-by-segment, of the calculated values of the water film heat transfer coefficients, H_1 , $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- H8; Sum, segment-by-segment, of the calculated values of the tube wall heat transfer coefficients, H_2 , $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- H9; Sum, segment-by-segment of the calculated values of the gas film heat transfer coefficients, H_3 , $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- I0; The molecular weight of the gas or mixture of gases.
- I1; The molecular weight of the gas or gases that dilute the helium, (assumed to be air with molecular weight = 29.97).
- I2; The molecular weight of pure helium ($I_2 = 4.00$).
- I4; A computed variable; a function of I0, I1, and I2, viz., line 1540; $I_4 = I_1(I_0 - I_2)$.
- I5; A computed variable; a function of I0, I1, and I2, viz., line 1545; $I_5 = I_1(I_0 - I_1)$.
- I6; A computed variable; a function of I0, I1, and I2, viz., line 1550; $I_6 = I_0(I_1 - I_2)$.
- I9; The difference between the computed and specified gas inlet temperatures in a Case B computation expressed as a percent error referred to the specified value.
- J4; The temperature of the gas leaving the heat exchanger; a fixed parameter during a case A run, $^\circ\text{F}$.
- K1; Thermal conductivity of liquid water, $\text{Btu/hr ft}^2\text{-}^\circ\text{F/ft}$
- K2; Thermal conductivity of tubing steel, $\text{Btu/hr ft}^2\text{-}^\circ\text{F-ft}$
- K7; Thermal conductivity of nitrogen-helium gas mixtures as a function of mixture molecular weight and at a temperature of 1061°Rankine (601°F); Ref. 7, $\text{Btu/hr ft}^2\text{-}^\circ\text{F/ft}$
- K8; A term in an equation for K7 (See above).
- K9; A term in an equation for K7 (See above). NOTE: $K_7 = K_8 + K_9$

- L0, (FULTN); The axial length (conceptual) of the heat exchanger; i.e., the length of the tube helices measured on the axial centerline of the heat exchanger, ft.
- L1, (FULTN); The axial length of a segment of the heat exchanger, Fig. 2, ft.
- L5, (FULTN); The conceptual or idealized length of the helically coiled section of a typical tube measured on the axial centerline of the tube, ft.
- L9, (FULTN); The wall thickness of the heat exchanger tubes, ft.
- M0, (FULTN); The number of axial length segments into which the heat exchanger is subdivided for computation; a number established by the user based on judgment and experience.
- M1; The exponent of the Reynolds number in the equation for the water film Nusselt number, Eq. 2.
- M3; The exponent of the Reynolds number in the equation for the gas film Nusselt number for gases flowing across in-line tube arrays, Eq. 3.
- M4; An exponent applied to the ratio of the absolute temperature of the bulk gas to the absolute temperature of the outer surface of the tube. NOTE: This variable not used in current CAHE computations.
- M5; The exponent of the Reynolds number in the equation for the Nusselt number for gases flowing across staggered tube banks, Eq. 3.
- M8; The exponent applied to the absolute temperature ratio in the equation⁷ for K3, line 3440, the thermal conductivity of helium-air (or nitrogen) gas mixtures.
- M9; The mole fraction of air in the helium-air gas mixture.
- N0, (FULTN); The total number of tubes in the heat exchanger.
- N1; The exponent of the Prandtl number in the equation for the water film Nusselt number, Eq. 2, dimensionless.
- N3; The exponent of the Prandtl number in the equations for the gas film Nusselt number, dimensionless.
- N7; The water side Reynolds number, dimensionless.
- N9; The gas side Reynolds number, dimensionless.

- ϕ_1 ; Sum, segment-by-segment, of the water temperatures, T_1 , at the transverse midplane through each segment, Fig. 2, $^{\circ}\text{F}$.
- ϕ_3 ; Sum, segment-by-segment, of the bulk temperatures of the gas, T_3 , at the transverse midplane through each segment, Fig. 2, $^{\circ}\text{F}$.
- ϕ_9 ; Sum, segment-by-segment, of the mean film temperature of the gas, T_9 , at the transverse midplane through each segment, Fig. 2, $^{\circ}\text{F}$.
- P_1 ; The Prandtl number of the water, dimensionless.
- P_2 ; The temperature of the water at the inlet of the heat exchanger; an "INPUT" request to the user each time the program is run, $^{\circ}\text{F}$.
- P_3 ; The Prandtl number of the gas, dimensionless.
- P_4 ; The temperature of the gas at the inlet of the heat exchanger; an "INPUT" request to the user each time the program is run, $^{\circ}\text{F}$.
- P_9 ; The absolute value of the difference between the specified and computed values of the gas inlet temperature in a Case B computation, $^{\circ}\text{F}$.
- $Q(Y)$; The heat transfer rate in the Yth segment, ($Y = 1, 2, 3 \dots M_0$), Btu/hr. NOTE: $Y = n$ in Fig. 2.
- Q_0 ; The sum of the heat transfer rates, $Q(Y)$, taken segment-by-segment as a computation proceeds; when $Y = M_0$, Q_0 is the total heat transfer rate by the exchanger, Btu/hr.
- Q_8 ; An estimated, (extrapolated) heat transfer rate used in Eqs. 21 to 24, inclusive, to establish initial temperature estimates in all segments but the first, Btu/hr.
- R_2 ; An additive correction applied to the outlet gas temperature, T_4 , ($J_4 = T_4$), to obtain convergence to the specified gas inlet temperature, P_4 , in a Case B computation. Refer to Figs. 2 and 3, and Eq. 28, $^{\circ}\text{F}$.
- R_5 ; The overall temperature difference, gas to water, at the cold end of the heat exchanger, $^{\circ}\text{F}$.
- R_6 ; The overall temperature difference, gas to water, at the hot end of the heat exchanger, $^{\circ}\text{F}$.

- R7; The logarithmic mean temperature difference, gas to water, based on R5 and R6, computed with Eq. 29, $^{\circ}\text{F}$.
- R8; The convergence criterion in a Case B computation; expressed as the maximum acceptable difference between the computed and specified gas inlet temperatures, $^{\circ}\text{F}$.
- S1; The specific heat, C_p , of liquid water⁷ at the bulk temperature, T_1 , Btu/lb- $^{\circ}\text{F}$.
- S3; The specific heat at constant pressure, C_p , of the gas⁷ at the bulk temperature, T_3 , Btu/lb- $^{\circ}\text{F}$.
- S5; The specific heat at constant pressure, C_p , of pure air⁷ at the bulk temperature, T_3 , Btu/lb- $^{\circ}\text{F}$.
- S6; The specific heat at constant pressure, C_p , of pure air⁷ at the mean gas film temperature, T_9 , Btu/lb- $^{\circ}\text{F}$.
- S7; The specific heat at constant pressure, C_p , of the gas⁷ at the mean gas film temperature, T_9 , Btu/lb- $^{\circ}\text{F}$.
- S9; The sum, taken segment-by-segment, as the computation proceeds, of the specific heats, S3, of the gas at the transverse midplane of each segment, Btu/lb- $^{\circ}\text{F}$.
- T1; The temperature of the bulk water at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.
- T2; The temperature of the bulk water entering a segment, Fig. 2, $^{\circ}\text{F}$.
- T3; The temperature of the bulk gas at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.
- T4; The temperature of the bulk gas leaving a segment, Fig. 2, $^{\circ}\text{F}$.
- T6; The average temperature of the tube wall at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.
- T7; The temperature of the outer surface of the tube wall, (the gas side), at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.
- T8; The temperature of the inner surface of the tube wall, (the water side), at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.
- T9; The mean temperature of the gas film at the transverse midplane through a segment, Fig. 2, $^{\circ}\text{F}$.

- U1; The overall heat transfer coefficient, gas to water, at a transverse midplane through a segment and referenced to the water side surface area, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- U3; The overall heat transfer coefficient, gas to water, at a transverse midplane through a segment and referenced to the gas side surface area, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- U4; An average value of the overall heat transfer coefficient for the entire heat exchanger based on the logarithmic mean average temperature difference, R7, and the effective gas side surface area. Eq. 29, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- U5; The sum, taken segment-by-segment as a computation proceeds, of the overall heat transfer coefficients, U1, referenced to the water side surface area, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- U7; The sum, taken segment-by-segment as a computation proceeds, of the overall heat transfer coefficients, U3, referenced to the gas side surface area, $\text{Btu/hr ft}^2\text{-}^\circ\text{F}$
- V0; The sum, taken segment-by-segment as the computation proceeds, of the gas viscosities, V3, within each segment, lb/ft-hr .
- V1; The viscosity of liquid water at the bulk temperature, T1, and at the transverse midplane through a segment, lb/ft-hr .
- V3; The viscosity of the gas at the mean gas film temperature, T9, and at the transverse midplane through a segment, lb/ft-hr .
NOTE: This definition does not apply to V3 as written on line 3475.
- W1, (FULTN); Total water flow rate in the heat exchanger, lb/hr .
- W3, (FULTN); Total gas flow rate in the heat exchanger, lb/hr .
- X; A loop counter; the value of X is the number of the segment undergoing computation, i.e., $X = 1, 2, 3, \dots M0$ with segment 1 at the low temperature end of the heat exchanger, Fig. 2.
- X1; A loop counter; the value of X1 is the number of the Case A computation during a Case B calculation.
- X5; The fraction of the tube array that is considered to be staggered.
- X9; The fraction of the tube array that is considered to be in-line.

- Y; A loop counter; the value of Y is the iteration number of the computation of heat transfer in the segment undergoing computation.
- Y1; The sum, taken segment-by-segment as the computation proceeds, of the number of iterated calculations, Y, within each segment. NOTE: The function of Y1 has been superseded by Y6.
- Y3; The leading coefficient, C3 in Eq. 3, for the gas film Nusselt number for flow across in-line tube arrays.
- Y5; The leading coefficient, C3 in Eq. 3, for the gas film Nusselt number for flow across staggered tube arrays.
- Y6; Identical to Y1.
- Y7; An "arrangement factor" in an alternative equation proposed by Grimison³ for the gas film Nusselt number for flow across in-line tube arrays. NOTE: This alternative approach has been deleted by setting $Y7 = 1$.
- Y9; An "arrangement factor" in an alternative equation proposed by Grimison³ for the gas film Nusselt number for flow across staggered tube arrays. NOTE: This alternative approach has been deleted by setting $Y9 = 1$.
- Z; A user established number (line 200) available for scheduling the printout of computations within particular segments, (see Z1 below).
- Z1; A number, line 3040, whose value is a function of the segment number, X, undergoing calculation, the segment length, L1, and Z, thus: $Z1 = X * L1/Z$
NOTE: In versions of CACHE other than that listed herein, the segment computations have been printed when Z1 is an integer, i.e. when $Z1 = \text{INT}(Z1)$. This procedure is best applied when L1 is an integer, a multiple, or a sub-multiple of an integer. The CACHE list that follows by-passes this procedure.
- Z7; The sum, taken segment-by-segment as the computation proceeds, of the water side Reynolds numbers, N7, calculated at the transverse midplane through the segments, dimensionless.
- Z8; The elapsed time after reactor shutdown; an "INPUT Z9" request to the user, hours.
- Z9; The sum, taken segment-by-segment, of the gas side Reynolds numbers, N9, calculated at the transverse midplane through the segments, dimensionless.

LIST OF CACHE USED TO CALCULATE THE POST-SHUTDOWN PERFORMANCE
OF A CORE AUXILIARY HEAT EXCHANGER (CONCEPTUAL DESIGN)
IN THE FULTON GENERATING STATION

CACHE

17:16

19-MAY-75

```

18 PRINT
20 PRINT "BEGIN CACHE"
25 PRINT
30 PRINT " GAS = HELIUM OR HELIUM MIXED WITH N2, O2, AND CO."
31 PRINT
35 PRINT "INPUT: Z8= TIME AFTER SHUTDOWN"
40 PRINT
45 INPUT Z8
50 PRINT
55 PRINT "HRS AFTER S'DOWN=" Z8
60 PRINT

```

```

100 FILES FXCH
120 READ #1,A0,A1,A2,A3, D1,D2,D3,L0,L1,L9,A7,A9,N0
125 READ #1,M0,W1,W3,G1,G3,A
127 READ #1, E
135 PRINT "GAS SIDE: TOT AREA="A0,"EFFEC AREA="A
150 PRINT "MASS FLOW RATES, H2O AND GAS =G1 AND G3"
155 PRINT "G1=" G1, "G3="G3
157 PRINT
160 PRINT "TOT H2O FLOW ="W1, "TOT GAS FLOW ="W3, "LBS/HR"
180 PRINT

```

Printout to
identify the
calculation.

Reads file EXCH.
Refer to FULTN
program, Appendix B.

CACHE List (continued)

185 PRINT	
200 Z = 3	See footnote (a)
210 DIM Q(25)	
395 B = 1	See footnote (b)
400 M1 = .8	M1 and N1 are exponents of Reynolds and Prandtl numbers, respectively in Eq. (2), for the water side Nusselt number.
405 N1 = .4	
410 M3 = .743	M3 and M5 are exponents of gas side Reynolds number in Nusselt number equations for in-line and staggered tubes.
412 N3 = .3333	
415 M5 = .567	
420 Y3 = .046	
422 Y7 = 1	N3 is the exponent of the gas side Prandtl No.
425 Y5 = .532	Y3 and Y5 are leading coefficients in Nusselt number equations for in-line and staggered tubes, respectively from Sanders. ⁶
427 Y9 = 1	Y7 and Y9 are arrangement factors in an alternative equation suggested by Grimison. ³ If set equal to 1 they are ineffectual.

30

^a_Z, line 200, has been used in other versions of CACHE to schedule the printout of individual segment computations. This mode of printout selection does not apply in this listed version. Line 210 provides a maximum of 25 iterations per segment to obtain convergence. Typically, 3 to 4 are required for CAHE calculations.

^b_B, line 395, is a counter used to specify the number of printouts in a Case B computation. Refer to lines 1440, 1442, 6048, 9000, and 9020. In this version of CACHE the first and last Case A iterations are printed.

CACHE List (continued)

430 M4 = .14
435 C1 = .023
440 F0 = 1.07
445 C3 = F0*Y3*Y7
450 C5 = F0*Y5*Y9
460 X9 = .75
465 X5 = 1-X9

M4 is an exponent used in certain equations for the gas film Nusselt number. It is not applicable in this version of CACHE.
F0 is the tube bank depth factor reported by Kays, London and Lo², (F0 = F_n in Eq. 3).
C1 is the leading coefficient in the Dittus-Boelter equation for the water film heat transfer coefficient, Eq. 2.
C3 and C5 are the leading coefficients for Grimison's equations for in-line and staggered tube banks, respectively.
X9 and X5 are the tube array fractions considered to be "in-line" and staggered, respectively. (Refer to Eq. 4.)

CACHE List (continued)

```
1160 PRINT "INPUT: P2 = INLET H2O TEMP, ESTAB. PARAM."
1165 INPUT P2
1166 PRINT
1167 PRINT "P2 = "P2
1168 PRINT
1170 PRINT "INPUT: P4 = INLET GAS TEMP., ESTAB. PARAM." See footnote (a)
1171 PRINT
1172 INPUT P4
1173 PRINT
1175 PRINT "P4 = "P4
1176 PRINT
```

^aP4 is an "ESTABLISHED PARAMETER" only for Case B situations. If the user is making a Case A run in which the inlet water and outlet gas temperatures are the governing conditions, any reasonable number will suffice for P4. A value for P4 must be provided if the program is to run.

CACHE List (continued)

```

1180 FOR X1=1 TO 20
1182 REM "                INITIALIZE T2"
1183 T2 = P2
1185 IF X1 = 1 THEN 1200
1187 C8 = (W3*S3)/(W1*S1)
1188 C6 = (E4 - E2)/(J4 - P2)
1190 R2 = (P4 - E4)*(1 - C8)/(C6 - C8)
1192 T4 = J4 + R2
1197 GO TO 1215
1199 PRINT
1200 PRINT "INPUT: T4 = INITIAL EST. GAS OUTLET TEMP."
1202 PRINT
1204 INPUT T4
1206 PRINT
1215 PRINT "T4 = "T4
1216 J4 = T4
1217 PRINT
1260 T3 = T4 + 1050/(2*M0)
1280 T1 = T2 + 260/(2*M0)
1300 T7 = T3 - .75*(T3-T2)
1320 T8 = T7
1340 T6 = T7
1350 T9 = .5*(T3 + T7)

```

See footnote (a)

See footnote (b)

T4 is a system parameter in a Case A calculation. In a Case B computation, the better the estimate the fewer iterations.

These are initial, first trial estimated temperatures in the first segment.

33

^a Outermost loop counter, iterates the Case A runs in a Case B computation.

^b In a Case B computation, lines 1187 to 1192 reset the value of T4, the outlet gas temperature, after the first and all subsequent Case A runs until convergence is obtained.

CACHE List (continued)

1360 F3 = ((T3+460)/(T7+460))*M4	If F3 = 1, it is ineffectual; see lines 1802 and 1810.
1380 F3 = 1	
1400 Q0 = 0	See footnote (a)
1440 IF X1<>B THEN 2640	
1442 IF X1>1 THEN 1840	
1450 PRINT "FRAC IN-LINE="X9, "FRAC STAG="X5	
1455 PRINT "TUBE BANK FACTOR="FO	See footnote (b)
1460 PRINT	
1465 PRINT "GAS SIDE N RE EXPON:", "STAG="M5, "IN LINE="M3	
1470 PRINT "LEAD COEFS:", "IN-LINE ="Y3, "STAG ="Y5	
1480 PRINT "ARRANGE FACTS:", " ", "IN-LINE="Y7, "STAGG ="Y9	
1485 PRINT	

^aLine 1400 initializes Q0, the sum, segment by segment, of the heat transfer rates computed in each segment.

^bLines 1450 to 1480 identify the constants in the gas side Nusselt No. equations calculated for in-line and staggered tube arrays.

CACHE List (continued)

```
1500 PRINT " INPUT: I0 = MØL WT, MIXT"
```

```
1505 INPUT I0
```

```
1507 PRINT
```

```
1510 PRINT "    MØL WT MIXT =" I0
```

```
1515 PRINT
```

```
1525 I1 = 28.97
```

```
1530 I2 = 4
```

```
1535 PRINT " MØL WTS:" , "AIR =" I1, "HEL =" I2
```

```
1540 I4 = I1*(I0 - I2)
```

```
1545 I5 = I2*(I0 - I1)
```

```
1550 I6 = I0*(I1 - I2)
```

```
1570 M9 = (I0-I2)/(I1-I2)
```

```
1580 M8 = .66
```

```
1584 K8 = .03991/(.27716 + M9)
```

```
1586 K9 = -.003196 + .07014*M9 - .15767*(M9^2) + .08612*(M9^3) See footnote (a)
```

```
1590 K7 = K8 + K9
```

```
1600 PRINT "MØL FRAC AIR=" M9, "WT FRAC AIR =" I4/I6
```

```
1620 PRINT
```

Identified the gas composition and associated parameters required to evaluate the thermal transport properties (Ref. 7).

^aLines 1580 through 1590 evaluate K9, the thermal conductivity at 601°F of helium mixed with air or air-like gases, as a function of the mole fraction of air, M9. See Ref. 7.

CACHE List (continued)

```

1800 PRINT "NUS NO,H20="C1"* (N,RE) + "M1"* (N,PR) + "N1
1801 PRINT
1802 PRINT "NUS NO, GAS="C5"* (N,RE) + "M5"* (N,PR) + "N3"*F3"
1804 PRINT " ", "FOR STAG TUBES"
1806 PRINT
1810 PRINT "NUS NO, GAS="C3"* (N,RE) + "M3"* (N,PR) + "N3"*F3"
1812 PRINT " ", "FOR IN-LINE TUBES"
1814 IF F3 = 1 THEN 1826
1815 PRINT " ", "F3 NOT EQUAL TO 1"
1818 GO TO 1828
1820 PRINT
1826 PRINT " ", "F3 SET EQUAL TO 1"
1828 PRINT
1830 PRINT "GAS MASS FLOW RATE, G3=G,MAX="G3"LBS/(FT+2-HR)"
1840 PRINT "CUMULATIVE ITERATIONS WITHIN SEGMENTS ="Y6
2070 PRINT

```

See footnote (a)

See line 1380

See Footnote (b)

^aIdentifies, numerically, the Nusselt number equations which are used to calculate the film coefficients. Refer also to lines 400 through 465.

^bLine 1840 enables the user to follow the iterative process during printouts.

CACHE List (continued)

```
2500 PRINT "-TEMPS, LOCATIONS, HT RATES, REY AND PR NOS, COEFS"
2510 PRINT
2520 PRINT
2530 PRINT "SEG NØ", "SUM SEGS", "LEN FRAC", "NØ ITER"
2540 PRINT " ", "N,RE,H2Ø", "N,PR,H2Ø", "N,RE,GAS", "N,PR,GAS"
2550 PRINT "COEFS:", "H,H2Ø", "H,TUBE", "H,GAS", "U,GAS SIDE"
2560 PRINT
2570 PRINT "DEG,F", "H2Ø IN", "H2Ø AVG", "H2Ø ØUT"
2580 PRINT "DEG F", "GAS ØUT", "GAS AVG", "GAS IN"
2590 PRINT "DEG F", "T,WALL,H2Ø", "T,WALL,GAS", "T,TUBE"

2600 PRINT
2610 PRINT " ", "Q,SEG", "Q,TØT"
2620 PRINT
2630 PRINT
```

The above, lines 2500 through 2610, prints the heading that precedes the printout of segment data produced by the calculation. This is repeated once each time a Case A calculation is printed.

CACHE List (continued)

<u>2640 PRINT</u>		
<u>2800 H7 = 0</u>	Initial values of the segment-by-segment summation	of H1, H2, and H3, respectively, lines 6004-6008.
<u>2805 H8 = 0</u>		
<u>2810 H9 = 0</u>		
<u>2815 U5 = 0</u>	Initial values of the segment-by-segment summation	of U1 and U3, lines 6012 and 6016.
<u>2820 U7 = 0</u>		
<u>2825 Z7 = 0</u>	Initial values of the segment-by-segment summation	of N7 and N9, lines 6020 and 6024.
<u>2830 Z9 = 0</u>		
<u>2840 V0 = 0</u>	Initial values of the segment-by-segment summation	of V3 and K3, lines 6028 and 6030.
<u>2845 K0 = 0</u>		
<u>2850 N0 = 0</u>	Initial values of the segment-by-segment summation	of: (1) gas side Nusselt No., No., line 6032; midplane water temperature, $\emptyset 1$, line 6038; gas heat capacity at bulk temperature, S9, line 6010; gas bulk and film temperatures $\emptyset 3$ and $\emptyset 9$, lines 6040 and 6042; segment iterations, Y1, line 6000.
<u>2900 $\emptyset 1 = 0$</u>		
<u>2905 S9=0</u>		
<u>2910 $\emptyset 3 = 0$</u>		
<u>2920 $\emptyset 9 = 0$</u>		
<u>3020 Y1 = 0</u>		

Average values of the more important or interesting results are computed by summing the results calculated in each segment and taking the simple arithmetic average. These averages are computed continuously as the computation proceeds, segment-to-segment. The sample printout that follows this list provides detailed information.

Averages are printed in accordance with lines 6310 through 6320, and 8480 and 8800.

CACHE List (continued)

3030 FOR X = 1 TO NO	
3040 Z1 = X*L1/Z	See footnote (a)
3050 FOR Y = 1 TO 100	
3400 K1 = .31 + 6.91E-4*T1 - 1.48E-6*T1 ² + 5.26E-10*T1 ³	
3405 IF X/NO > .75 THEN 3418	
3410 K2 = 22.21 - 2.76E-3*T6	
3414 GO TO 3425	See footnote (b)
3418 K2 = 16.7	
3425 IF IO = 4 THEN 3450	
3440 K3 = K7*(((T9+460)/1061) ^{M8})	See footnote (c)
3445 GO TO 3460	
3450 K3 = 1.0221E-3*((T9 + 460) ^{.71})	
3460 V1 = 378.4/(T1 ^{.181})	See footnote (a)

^aIn line 3030; X = segment number and is the counter that advances the computation from segment to segment. In line 3050; Y is the counter controlling the innermost computational loop that iterates the heat rate-heat balance calculations in each segment until convergence is obtained. In a typical CACHE calculation the maximum value for Y is 3 to 4. In lines 3400 and 3460; K1 and V1 are the thermal conductivity and viscosity of liquid water.

^bIn lines 3410 and 3418, K2 is the thermal conductivity of the tubing in the Fulton Generating Station CAHE. The higher alloy steel in the hot end, 25% of the CAHE, has a temperature independent conductivity of 16.7 Btu/hr ft²-°F/ft, line 3418. The conductivity of the remaining 75% is evaluated by line 3410, (Ref. 7).

^cIf the gas is pure helium the temperature dependent conductivity is evaluated by line 3450; if the gas is a mixture line 3440 applies. K7 and M8 are computed with lines 1580 to 1590.

CACHE List (continued)

```

3465 IF I0 = 4 THEN 3490
3470 IF I0 > 11 THEN 3480
3475 V3 = (M9/.28)*5.942E-4*((T9+460)↑.71)
3476 V3 = V3 + ((.28-M9)/.28)*5.89E-4*((T9+460)↑.7) See footnote (a)
3478 G0 T0 3500
3480 V3 = 5.942E-4*((T9 + 460)↑.71)
3485 G0 T0 3500
3490 V3 = 5.89E-4*((T9+460)↑.7)
3500 S1 = 1.035 - 4.55E-4*T1 + 1.38E-6*T1↑2 S1 = specific heat of liquid water
3510 S5 = .2304 + 3.7672E-5*T3 - 4.86E-9*(T3↑2)
3520 S3 = (1/I6)*(S5*I4 - 1.242*I5) See footnote (b)
3530 S6 = .2304 + 3.7672E-5*T9 - 4.86E-9*(T9↑2)
3540 S7 = (1/I6)*(S6*I4 - 1.242*I5)

```

^aIf the gas is pure helium, (I0 = 4), gas viscosity is calculated by line 3490; if the gas is a mixture having a molecular weight above 4 to 11 ($4 < I0 < 11$) the viscosity is evaluated with lines 3475 and 3476 in which M9 = mole fraction of air, line 1570; for gas mixtures with molecular weights above 11 ($I0 > 11$) the viscosity is calculated by line 3480. These equations are from Ref. 7.

^bS5 and S6, lines 3510 and 3530 are equations for the heat capacity of pure air at the segment midplane bulk temperature, T3, and the mean film temperature, T9, respectively; S3 and S7, lines 3520 and 3540 evaluate gas heat capacity at the bulk and mean film temperatures, respectively. The heat capacity of helium is temperature independent and is equal to 1.242 Btu/lb-°F. These equations are developed in Ref. 7.

CACHE List (continued)

3580	$N7 = D1 \cdot G1 / V1$	
3600	$N9 = D3 \cdot G3 / V3$	
3620	$P1 = S1 \cdot V1 / K1$	
3640	$P3 = S7 \cdot V3 / K3$	See footnote (a)
3650	$F5 = C5 \cdot (N9 + M5) \cdot (P3 + N3)$	
3660	$F7 = C1 \cdot (N7 + M1) \cdot (P1 + N1)$	
3680	$F9 = C3 \cdot (N9 + M3) \cdot (P3 + N3) \cdot F3$	
3700	$H1 = K1 \cdot F7 / D1$	
3720	$H2 = K2 / L9$	
3740	$H3 = (X9 \cdot F9 + X5 \cdot F5) \cdot K3 / D3$	H1, H2, and H3 are water film, tube wall and gas film heat transfer coefficients.
3760	$B7 = 1 / (W1 \cdot S1)$	
3780	$B9 = 1 / (W3 \cdot S3)$	
3800	$U8 = (A3 / (A1 \cdot H1)) + (A3 / (A2 \cdot H2)) + (1 / H3)$	
3820	$U3 = 1 / U8$	
3840	$U9 = (1 / H1) + (A1 / (A2 \cdot H2)) + (A1 / (A3 \cdot H3))$	U1 and U3 are the overall heat transfer coefficients referenced to water side and gas side surface areas.
3860	$U1 = 1 / U9$	
3880	$Q(Y) = U3 \cdot A3 \cdot (T3 - T1)$	See footnote (b)

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^aN7 and N9, lines 3580 and 3600, are water and gas Reynolds numbers, respectively; P1 and P3, lines 3620 and 3640, are water and gas Prandtl numbers, respectively; F7, line 3660, is the water side Nusselt number; F5 and F9, lines 3650 and 3680 are gas side Nusselt numbers, for staggered and in-line tube arrays, respectively.

^b $Q(Y)$, line 3880, is the heat transferred in the X^{th} segment (see line 3030) after the Y^{th} iteration, (line 3050), in that segment.

CACHE List (continued)

4600 IF Y = 1 THEN 4720	See footnote (a)
4620 E9 = ABS((Q(Y)/Q(Y-1))-1)	
4700 IF E9<= .001 THEN 6000	Convergence test, see footnote (b)
4720 T1 = T2 + .5*Q(Y)*B7	
4740 T3 = T4 + .5*Q(Y)*B9	
4760 T7 = T3 - Q(Y)/(A3*H3)	
4780 T8 = T1 + Q(Y)/(A1*H1)	See footnote (b)
4800 T6 = .5*(T7+T8)	
4810 T9 = .5*(T3 + T7)	
4820 NEXT Y	

^aThe convergence test is by-passed when Y = 1 since this test compares two successive computations of Q(Y) and, with Y = 1, only one is available.

^bThese equations, lines 4720 through 4810 reset the segment temperatures, after the Yth computation, for use in the (Y + 1)th, line 4820, computation. A segment computation is considered complete when the ratio of two successive computations of the heat transfer rate differ from 1.0 by E9 or less.

CACHE List (continued)

This section of CACHE takes sums calculated within each segment during the computation and when convergence is obtained in each segment. The sums except as noted in footnote (c) are used to calculate averages after the last segment ($X = M0$) has been computed.

6000	$Y1 = Y1 + Y$
6002	$Y6 = Y6 + Y$
6004	$H7 = H7 + H1$
6006	$H8 = H8 + H2$
6008	$H9 = H9 + H3$
6010	$S9 = S9 + S3$
6012	$U5 = U5 + U1$
6016	$U7 = U7 + U3$
6020	$Z7 = Z7 + N7$
6024	$Z9 = Z9 + N9$
6028	$V0 = V0 + V3$
6030	$K0 = K0 + K3$
6032	$N0 = N0 + F9 * X9 + F5 * X5$
6036	$Q0 = Q0 + Q(Y)$
6038	$\theta 1 = \theta 1 + T1$
6040	$\theta 3 = \theta 3 + T3$
6042	$\theta 9 = \theta 9 + T9$

Y1 and Y6 count the iterations.

Sums of the water film, tube wall, and gas film heat transfer coefficients.

Sums the heat capacity, C_p , of the gas at bulk temperature.

Sums the overall heat transfer coefficients referenced to the water and gas side surface areas.

Sums the water and gas Reynolds numbers.

Sums the viscosity and thermal conductivity of the gas at mean film temperature.

N0 is the sum of the gas film Nusselt numbers.

See footnote (a).

Sums the segment midplane bulk water temperatures, T1 and the midplane bulk gas and mean gas film temperatures, T3 and T9.

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^aQ0 is the sum of the individual heat transfer rates calculated in each segment up to and including the X^{th} segment.

CACHE List (continued)

6044 E4 = T4 + Q(Y)*B9

See footnote (a)

6046 E2 = T2 + Q(Y)*B7

6048 IF X1<>B THEN 6380

6050 IF X = 1 THEN 6140

6055 IF X = 12 THEN 6140

6060 IF X = M0 THEN 6140

6090 G0 T0 6120

6100 IF Z1 = INT(Z1) THEN 6140

6120 G0 T0 6380

This section determines when, during a computation, the calculated results of individual segment computations are printed. Segment printout is specified by lines 6140 through 6340. See footnote (b).

44

^aE4 and E2, lines 6044 and 6046, are the inlet gas temperature and the outlet water temperature calculated after the segment computation has converged. The calculation assumes that the temperature variations are linear within the segment. E4 and E2 are printed, lines 6220 and 6240. B9 and B7 are defined by lines 3760 and 3780.

^bThis version of CACHE specifies segment printout for the first and last segments of: (1) a Case A run or, (2) the first and last Case A runs of the set comprising a Case B run.

CACHE List (continued)

This section of CACHE specifies parameters that are included in a segment printout. The sample printout in the next section of this appendix offers additional explanation.

6140 PRINT X, X*L1, L1*(X-.5)/L0, Y	See footnote (a)
6142 PRINT	
6143 PRINT "K,TUBE =" K2	Thermal transport properties of the tubing and the gas. See footnote (b).
6145 PRINT "C SUBP:", "BULK="S3, "FILM="S7	
6147 PRINT "K,GAS="K3,"VIS,GAS="V3	
6150 PRINT	
6160 PRINT "O DIM NOS", N7, P1, N9, P3	Reynolds and Prandtl numbers for water and gas films and the gas side Nusselt numbers.
6165 PRINT "N,NU,GAS ="F9*X9 + F5*X5	

^aLine 6140 specifies printout of the segment number, X; the sum of the lengths of all, 1 through X, segments iterated, (X*L1); the midplane location of the Xth segment relative (normalized) to the total length, L0, of the heat exchanger; and, the number of iterations required to obtain convergence in the segment computation.

^bBecause the gas side heat transfer coefficient is very strongly controlling the variation of water properties in these calculations is relatively unimportant and not included in the printout.

CACHE List (continued)

This section of CACHE specifies parameters that are included in a segment printout. The sample printout in the next section of this appendix offers additional explanation.

```
6180 PRINT "CØEFS=", H1, H2, H3, "U3="U3
```

```
6185 PRINT
```

Heat transfer coefficients, see footnote (a).

```
6190 PRINT "H,IN-LINE ="F9*K3/D3
```

```
6192 PRINT "H,STAGGERED="F5*K3/D3
```

```
6200 PRINT
```

```
6220 PRINT "DEG F=",T2,T1,E2
```

Segment temperatures at each end and at the midplane. Refer to footnote (b) and Fig. 2.

```
6240 PRINT "DEG F=",T4,T3,E4
```

```
6250 PRINT "DEG F=", T8, T7, T6
```

^aH1, H2, and H3 are the water film, tube wall, and gas film heat transfer coefficients within the segment. U3 is the overall coefficient referenced to the gas side surface area, A3.

^bT2, T1, and E2 are water temperatures; inlet, midplane, and outlet, respectively. T4, T3, and E4 are gas temperatures; outlet, midplane, and inlet, respectively. T8, T7, and T6 are midplane tube wall temperatures; inner surface, average, and outer surface, respectively.

CACHE List (continued)

6255 PRINT	
6260 PRINT "T,FILM="	.5*(T3+T7)
6262 PRINT	
6265 PRINT "AVG FLUID TEMPS:"	"BULK H2O ="
6270 PRINT "BULK GAS ="	03/X, "GAS FILM ="
6280 PRINT	
6300 PRINT "BTU/HR="	Q(Y), Q0
6305 PRINT	
6310 PRINT "N,RE,GAS,AVG ="	Z9/X, "N,NU,AVG ="
6312 PRINT	
6315 PRINT "GAS SIDE:"	"H,AVG ="
6320 PRINT " "	"V,AVG ="
6322 PRINT	
6340 PRINT "U,H2O="	U1, "Q,H2O="

U1*A1*(T3-T1)

This prints the mean gas film temperature at the segment midplane. See footnote (a).

These are average values of the indicated parameters within the region of the heat exchanger consisting of segments 1 to X, incl.; i.e., when X = M0 these are averages for the entire unit.

See footnote (b).

^aThe gas properties that determine the film coefficient of heat transfer are evaluated at the mean film temperatures. This number is useful to verify correct program operation.

^bLine 6340, printout of U1 and "Q, H₂O" refer to the segment only, these are not averages. This line was added to check operation.

CACHE List (continued)

```

6360 PRINT
6370 PRINT
6380 IF X=M0 THEN 6555
6400 Q8 = Q(Y)*(1 + A3*U3*(B9-B7))
6420 T2 = T2 + Q(Y)*B7
6440 T4 = T4 + Q(Y)*B9
6460 T1 = T1 + Q8*B7
6480 T3 = T3 + Q8*B9
6500 T8 = T1 + Q8/(A1*H1)
6520 T7 = T3 - Q8/(A3*H3)
6540 T6 = (T7 + T8)/2
6550 T9 = .5*(T3+T7)
6555 NEXT X
6905 PRINT
7900 IF X1<>B THEN 8820
8000 R6 = E4 - E2
8010 R5 = J4 - P2
8020 R7 = (R6-R5)/LOG(R6/R5)

```

These are the equations for estimating the initial temperatures in segment (X + 1) after the calculations for segment X have converged.

NOTE: X = n in Fig. 2.

See footnote (a)

See footnote (b)

^aIf X1<>B, line 7900, the printout of average parameters after a Case A iteration during a Case B run is by-passed. The value of B is programmed by lines 8960 through 9020.

^bR6 and R5, lines 8000 and 8010, are the overall temperature differences, bulk gas to bulk water, at the hot end and cold end of the heat exchanger. R7, line 8020, is the log mean temperature difference in the heat exchanger. In strict terms this equation for R7 implies that all fluid and material properties are temperature independent.

CACHE List (continued)

8200 PRINT	
8220 PRINT "TEMP DIF. AT COLD END ="R5	
8240 PRINT "TEMP DIF AT HOT END ="R6	Refer to footnote (b) on the preceding page.
8260 PRINT	
8280 PRINT "LOG MEAN DELTA T ="R7	
<hr/>	
8300 PRINT	
8320 PRINT	
8340 PRINT "GRAND TOTAL INTER SEGMENT ITER. ="Y6	See footnote (a)
8360 PRINT	
8380 PRINT "TOTAL ENTIRE LENGTH ITER. ="X1	See footnote (b)
8400 PRINT	
8420 PRINT "TOTAL SEGMENT ITER. ="X1*M0	See footnote (c)
8440 PRINT	

^aY6, line 8340, is the grand total of all segment iterations required to complete a computation.

^bX1, line 8380, is the number of Case A loops required to complete a Case B computation.

^cLine 8420 causes the product of the number of segments, M0, and X1, the number of times that a heat transfer rate in each segment was computed, to be printed.

These print statements enable the user to follow the course of the computations as they proceed.

CACHE List (continued)

8460 U4 = Q0/(A*R7)	Refer to Eq. (30).
8480 PRINT "MISC AVGS:"	
8500 PRINT "H7 ="H7/M0, "H8 ="H8/M0, "H9 =" H9/M0	Average values of the water film, tube wall and gas film heat transfer coefficients.
8520 PRINT	
8540 PRINT "U,AVG="U4	
8560 PRINT " ", "REF T0 GAS SIDE AND LOG MEAN DELTA T"	See line 8460.
8580 PRINT	
8600 PRINT "U BAR="Q0*M0/(A*(03-01))	
8620 PRINT " ", "BASED ON Q TOT AND SEG AVG DELTA T"	Refer to Eq. (31).
8640 PRINT	
8660 PRINT "U,AVG OF SEG U'S =" U7/M0	
8680 PRINT	Refer to Eq. (32).
8700 PRINT "DELTA T,AVG="Q0*M0/(A*U7)	
8720 PRINT " ", "BASED ON AVG OF SEG U'S"	
8740 PRINT	
8760 PRINT "GAS PROP EVAL AT MEAN FILM TEMP"	
8780 PRINT	
8800 PRINT "N,RE,H20,AVG =" Z7/M0	
8810 PRINT "EFFEC GAS SIDE AREA =" A	

CACHE List (continued)

```
8820 R8 = 1
8840 P9 = ABS(P4-E4)
8860 I9 = (P9/P4)*100
8880 PRINT
8900 PRINT "% ERROR IN INLET GAS TEMP ="I9
8940 IF P9<=R8 THEN 9000
8960 G0 T0 9040
8980 PRINT
9000 IF B=X1 THEN 9080
9020 B = X1 + 1
9040 NEXT X1
9060 PRINT
9080 PRINT "END OF CACHE"
9100 PRINT
9200 PRINT
9300 PRINT
9999 END
```

See footnote (a)

The programming of B schedules the printout during a Case B run. B appears in line numbers 395, 1180, 6048, 7900, 9000, and 9020.

In this version of CACHE the results of the first and final Case A runs are printed.

51

^aLines 8820 through 8940 state the convergence criterion for a Case B computation; i.e., R8 and P9, lines 8820, 8840, and 8940 require that the computed value of gas inlet temperature, E4, equals the specified value, P4, (line 1170), within $\pm 1^\circ\text{F}$. Line 8900 enables the user to follow the course of convergence during program operation.

Printout of a typical CACHE run to calculate CAHE performance in cooling a gas mixture.

CACHE

17:54

19-MAY-75

BEGIN CACHE

GAS = HELIUM OR HELIUM MIXED WITH N₂, O₂, AND CO.

INPUT: Z8= TIME AFTER SHUTDOWN

?5.3964

HRS AFTER S'DOWN= 5.3964

GAS SIDE: TOT AREA= 2749 EFFEC AREA= 2598.3

MASS FLOW RATES, H₂O AND GAS =G1 AND G3

G1= 2507160 G3= 11885.7

TOT H₂O FLOW = 1050200

TOT GAS FLOW = 175860

LBS/HR

INPUT: P2 = INLET H₂O TEMP, ESTAB. PARAM.

?119.4

Both Case A and Case B conditions require accurate knowledge of inlet water temperature.

P2 = 119.4

INPUT: P4 = INLET GAS TEMP., ESTAB. PARAM.

?1110

In this Case B example the user has assumed that the CAHE is removing heat from hot gas entering at 1110°F.

P4 = 1110

INPUT: T4 = INITIAL EST. GAS OUTLET TEMP. - See Note.

?410

Note: This value of T₄, 401°F, is a crude, off-the-cuff estimate. However, the printout after the first iteration will give the calculated performance characteristics of the CAHE when the discharge gas temperature is 410°F, i.e., a Case A computation in which the gas outlet temperature is established to be 410°F will be complete after the first iteration.

T4 = 410

CACHE printout - cont'd

These data identify the gas composition, the Nusselt number equations, for both staggered and in-line tube arrangements, the in-line fraction, and the mass flow rate of the gas.

```

FRAC IN-LINE= 0.75          FRAC STAG= 0.25
TUBE BANK FACTOR= 1.07

GAS SIDE N RE EXPON:      STAG= 0.567    IN LINE= 0.743
LEAD COEFS:   IN-LINE = 4.60000E-2      STAG = 0.532
ARRANGE FACTS:      IN-LINE= 1      STAGG = 1

INPUT: IO = MOL WT, MIXT
?10.72

MOL WT MIXT = 10.72

MOL WTS:      AIR = 28.97    HEL = 4
MOL FRAC AIR= 0.269123      WT FRAC AIR = 0.727285

NUS NO,H2O= 2.30000E-2 *(N,RE)+ 0.8 *(N,PR)+ 0.4

NUS NO, GAS= 0.56924 *(N,RE)+ 0.567 *(N,PR)+ 0.3333 *F3
              FOR STAG TUBES

NUS NO, GAS= 4.92200E-2 *(N,RE)+ 0.743 *(N,PR)+ 0.3333 *F3
              FOR IN-LINE TUBES
              F3 SET EQUAL TO 1

GAS MASS FLOW RATE, G3=G,MAX= 11885.7 LBS/(FT2-HR)
CUMULATIVE ITERATIONS WITHIN SEGMENTS = 0

```

The above printout specified by lines 1450 through 1840 in CACHE.

CACHE printout - cont'd

This and the next 2 pages show the printouts for the first, twelfth and last segments (Nos. 1, 12, and 20) as computed during the first Case A iteration. Explanatory notes follow the printout for segment number 12.

-TEMPS, LOCATIONS, HT RATES, REY AND PR NOS, COEFS

SEG NO	SUM SEGS N,RE,H2O H,H2O	LEN FRAC N,PR,H2O H,TUBE	NO ITER N,RE,GAS H,GAS	N,PR,GAS U,GAS SIDE
DEG F	H2O IN	H2O AVG	H2O OUT	
DEG F	GAS OUT	GAS AVG	GAS IN	
DEG F	T,WALL,H2O	T,WALL,GAS	T,TUBE	
	Q,SEG	Q,TOT		
1	0.635	2.50000E-2	3	
K,TUBE = 21.8515				
C SUBP:	BULK= 0.517164		FILM= 0.513581	
K,GAS= 6.20771E-2		VIS,GAS= 6.43218E-2		
O DIM NOS	126420.	3.54912	15398.8	0.532152
N,NU,GAS = 65.9579				
COEFS=	2573.82	2622.18	49.1338	U3= 47.0284
H,IN-LINE = 38.3923				
H,STAGGERED= 81.3581				
DEG F=	119.4	120.272	121.144	
DEG F=	410	420.072	430.14	
DEG F=	127.12	132.693	129.907	
T,FILM= 276.383				
AVG FLUID TEMPS:		BULK H2O = 120.272		
BULK GAS = 420.072		GAS FILM = 276.383		
BTU/HR=	1.83168E+6	1.83168E+6		
N,RE,GAS,AVG = 15398.8		N,NU,AVG = 65.9579		
GAS SIDE:	H,AVG = 49.1338	K,AVG = 6.20771E-2		
	V,AVG = 6.43218E-2			
U,H2O= 58.7855		Q,H2O= 1.83168E+6		

CACHE printout - cont'd

Results of computations for a typical CAHE segment (Segment 12), first Case A iteration.

12	7.62	0.575	3	
K, TUBE = 21.7477				
C SUBP:	BULK= 0.524693		FILM= 0.518099	
K, GAS= 7.18284E-2		VIS, GAS= 7.52469E-2		
O DIM NOS	162176.	2.69433	13163.	0.542757
N, NU, GAS = 59.7732				
COEFS=	2882.13	2609.73	51.521	U3= 49.332
H, IN-LINE = 39.7965				
H, STAGGERED= 86.6945				
DEG F=	146.692	148.511	150.33	
DEG F=	722.729	743.388	764.047	
DEG F=	161.239	173.745	167.492	
T, FILM= 458.566				
AVG FLUID TEMPS:		BULK H2O = 132.831		
BULK GAS = 564.263		GAS FILM = 357.716		
BTU/HR=	3.81255E+6	3.24456E+7		
N, RE, GAS, AVG = 14337.2		N, NU, AVG = 63.0455		
GAS SIDE:	H, AVG = 50.2281	K, AVG = 6.64853E-2		
	V, AVG = 6.92516E-2			
U, H2O= 61.6651		Q, H2O= 3.81255E+6		

From the above:

First line. The total length of the first 12 segments is 7.62 ft., the normalized distance of the midplane of segment 12 from the cold end of the CAHE is 0.575 (total CAHE length set equal to 1.00); the computation of segment 12 converged after 3 iterations.

Material and fluid properties at midplane of segment 12.

k, tube wall; K2 = 21.7 Btu/hr ft²-°F/ft
 Cp, gas at bulk temp.; S3 = 0.525 Btu/lb-°F
 Cp, gas at film temp.; S7 = 0.518 Btu/lb-°F
 k, gas at film temp.; K3 = 0.0718 Btu/hr ft²-°F/ft
 μ, gas at film temp.; V3 = 0.0752 lb/ft-hr

CACHE printout - cont'd

Explanatory notes for results printed after 12th segment computation.

Dimensionless numbers and heat transfer coefficients at the 12th segment midplane.

Reynolds No., water, $N_7 = 162,000$; Prandtl No., water, $P_1 = 2.69$

Reynolds No., gas, $N_9 = 13,200$; Prandtl No., gas, $P_3 = 0.543$

Nusselt No., gas, $F_9 \times X_9 + F_5 \times X_5 = 59.8$

Heat transfer coefficients; Btu/hr ft²-°F

h, water film, $H_1 = 2882$; h, tube wall, $H_2 = 2610$

h, gas film, $H_3 = 51.5$; see (a) below

U, overall, referenced to gas side area,

U, overall, referenced to water side area, $U_1 = 61.7$

h, gas film, 100% in-line tubes, $F_9 \times K_3/D_3 = 39.8$

h, gas film, 100% staggered tubes, $F_5 \times K_3/D_3 = 86.7$

(a) From Eq.(4):

$$h, \text{ gas film} = 0.75 h_{\text{in-line}} + 0.25 h_{\text{stag}} = 51.5$$

Temperatures, °F, in the 12th segment

Refer to Fig. 2, (n = 12)

Water: Inlet, $T_2 = 146.7$; midplane, $T_1 = 148.5$; outlet, $E_2 = 150.3$

Gas: Outlet, $T_4 = 722.7$; midplane, $T_3 = 743.4$; inlet, $E_4 = 764.0$

Tube wall at the midplane:

inner surface, $T_8 = 161.2$, outer surface, $T_7 = 173.7$

mean, $T_6 = 167.5$

Mean gas film, $T_9 = .5 \times (T_3 + T_7) = 458.6$

Averaged temperatures, heat transfer coefficients and parameters

These are arithmetic averages obtained by summing the values computed in segments 1 to X and dividing by X; X = 12.

Temperatures: Bulk water, $O_1/X = 132.8^\circ\text{F}$

Bulk gas, $O_3/X = 564.3^\circ\text{F}$

Gas film, $O_9/X = 357.7^\circ\text{F}$

CACHE printout - cont'd

Reynolds No., gas, $Z9/X = 14,300$

Nusselt No., gas, $NO/X = 63.0$

$\bar{h}$, gas film; $H9/X = 50.2$, Btu/hr $\text{ft}^2\text{-}^\circ\text{F}$

$\bar{k}$, gas film; $KO/X = 0.0665$, Btu/hr $\text{ft}^2\text{-}^\circ\text{F}/\text{ft}$

$\bar{\mu}$, gas film; $VO/X = 0.0693$, lb/ft-hr

Heat transfer rates, Btu/hr

Heat transferred within segment 12, $Q(Y) = 3.813 \times 10^6$

Heat transferred in segments 1 to 12, incl., $Q0 = 3.245 \times 10^7$

CACHE printout - cont'd

This page contains the printout of the first set of calculated results for the 20th and last segment in the CAHE after the first Case A run of the series.

20	12.7	0.975	3	
K, TUBE = 16.7				
C SUBP:	BULK= 0.533713	FILM= 0.523916		
K, GAS= 8.41927E-2	VIS, GAS= 8.92596E-2			
O DIM NOS	213475.	1.99895	11096.6	0.555446
N, NU, GAS = 53.7384				
CØEFS=	3264.19	2004.	54.2927	U3= 51.6572
H, IN-LINE = 41.4059				
H, STAGGERED= 92.9529				
DEG F=	184.243	187.424	190.606	
DEG F=	1145.84	1181.38	1216.91	
DEG F=	207.085	235.596	221.34	
T, FILM= 708.485				
AVG FLUID TEMPS:		BULK H2O = 147.151		
BULK GAS = 726.738		GAS FILM = 449.908		
BTU/HR=	6.67045E+6	74646337		
N, RE, GAS, AVG = 13411.2		N, NU, AVG = 60.415		
GAS SIDE:	H, AVG = 51.3359	K, AVG = 7.12206E-2		
	V, AVG = 7.45887E-2			
U, H2O= 64.5715		Q, H2O= 6.67045E+6		

NOTE: Had the user specified a Case A computation based on system operation requiring the gas outlet temperature to be 410°F, (the "INPUT: T4 ..." instruction) this and the next page show the final results, i.e.:

Total heat transferred by the CAHE = 7.46×10^7 Btu/hr

Gas inlet temperature = 1217°F

Water outlet temperature = 221°F

Average gas film heat transfer coefficient in the CAHE = 51.3

CACHE printout - cont'd

Printout shown on this page is specified by lines
8220 through 8810 in CACHE.

TEMP DIF. AT COLD END = 290.6
TEMP DIF AT HOT END = 1026.31

LOG MEAN DELTA T = 583.073

GRAND TOTAL INTER SEGMENT ITER.= 60

TOTAL ENTIRE LENGTH ITER. = 1 This number, (1), states that the
first Case A iteration has been
completed.

TOTAL SEGMENT ITER. = 20

MISC AVGS:

H7 = 2861.2 H8 = 2461.9 H9 = 51.3359

U,AVG= 49.2716

REF TO GAS SIDE AND LOG MEAN DELTA T

U BAR= 49.5679

BASED ON Q TOT AND SEG AVG DELTA T

U,AVG OF SEG U'S = 49.0688

DELTA T,AVG= 585.483

BASED ON AVG OF SEG U'S

GAS PROP EVAL AT MEAN FILM TEMP

N,RE,H2O,AVG = 160742.

EFFEC GAS SIDE AREA = 2598.3

From the above:

The average heat transfer coefficients in the CAHE, based on the averaged
values computed in 20 length segments are;

h, water film = 2861, h, tube wall = 2462, h, gas film = 51.3

U, overall = 49.07. The alternate values of U, 49.27 and 49.57
are calculated with the equations listed at lines 8540 and 8600
in CACHE.

CACHE printout - cont'd

These computations are printed to enable the user to follow the convergence to the specified gas inlet temperature, P4.

% ERROR IN INLET GAS TEMP = 9.63166

T4 = 381.721

The initial estimate, by the user, of T4, the outlet gas temperature for an inlet gas temperature of 1110°F ("Input P4") was 410.

% ERROR IN INLET GAS TEMP = 0.565522

T4 = 383.393

% ERROR IN INLET GAS TEMP = 3.40642E-2

T4 = 383.292

CUMULATIVE ITERATIONS WITHIN SEGMENTS = 180

The first line tells the user that his initial estimate of the outlet gas temperature, T4 = 410°F, produced a computed gas inlet temperature, too high by 9.63%, i.e., the preceeding printout for segment 20 shows a value of 1216.9°F instead of 1110°F as specified. The program reset T4 (lines 1187 through 1192) to be 381.721°F and repeated the calculation. This process was repeated two more times and the final Case A computation was based on a gas outlet temperature of 383.292°F. The printout for this final run follows. Segment 12 is omitted in this report.

CACHE printout - cont'd

This is the start of the final Case A computation. The explanatory notes on the 3rd page of this printout apply. The segment 12 printout is not included.

-TEMPS, LOCATIONS, HT RATES, REY AND PR NOS, COEFS

SEG NO	SUM SEGS	LEN FRAC	NO ITER	
COEFS:	N,RE,H2O H,H2O	N,PR,H2O H,TUBE	N,RE,GAS H,GAS	N,PR,GAS U,GAS SIDE
DEG,F	H2O IN	H2O AVG	H2O OUT	
DEG F	GAS OUT	GAS AVG	GAS IN	
DEG F	T,WALL,H2O	T,WALL,GAS	T,TUBE	
	Q,SEG	Q,TOT		
1	0.635	2.50000E-2	3	
K,TUBE = 21.8543				
C SUBP: BULK= 0.516486 FILM= 0.513212				
K,GAS= 6.12682E-2 VIS,GAS= 0.063421				
O DIM NOS	126316.	3.55235	15617.5	0.531245
N,NU,GAS =	66.5465			
COEFS=	2572.86	2622.52	48.9262	U3= 46.8379
H,IN-LINE = 38.2694				
H,STAGGERED= 80.8965				
DEG F=	119.4	120.189	120.977	
DEG F=	383.292	392.413	401.529	
DEG F=	126.384	131.371	128.877	
T,FILM= 261.892				
AVG FLUID TEMPS:				
BULK GAS = 392.413 BULK H2O = 120.189				
GAS FILM = 261.892				
BTU/HR=	1.65647E+6	1.65647E+6		
N,RE,GAS,AVG = 15617.5 N,NU,AVG = 66.5465				
GAS SIDE:	H,AVG = 48.9262	K,AVG = 6.12682E-2		
	V,AVG = 0.063421			
U,H2O= 58.5474 Q,H2O= 1.65647E+6				

CACHE printout - cont'd

This page is the printout of the last segment calculated, (segment 20), during the 4th and final Case A run. Convergence was obtained.

20	12.7	0.975	3	
K,TUBE = 16.7				
C SUBP:	BULK= 0.531709		FILM= 0.522617	
K,GAS= 8.14359E-2		VIS,GAS= 8.61210E-2		
O DIM NOS	204204.	2.09635	11501.	0.552684
N,NU,GAS = 54.9472				
CØEFS=	3199.46	2004.	53.6961	U3= 51.0966
H,IN-LINE = 41.0613				
H,STAGGERED= 91.6003				
DEG F=	177.666	180.509	183.352	
DEG F=	1046.25	1078.11	1109.98	
DEG F=	198.427	223.896	211.161	
T,FILM= 651.005				
AVG FLUID TEMPS:		BULK H2O = 144.387		
BULK GAS = 669.134		GAS FILM = 418.462		
BTU/HR=	5.95850E+6	6.70425E+7		
N,RE,GAS,AVG = 13729.		N,NU,AVG = 61.3157		
GAS SIDE:	H,AVG = 50.9581	K,AVG = 6.96066E-2		
	V,AVG = 7.27695E-2			
U,H2O= 63.8708		Q,H2O= 5.95850E+6		

Using the operating parameters, heat transfer correlations and gas and material properties specified within the program this printout shows the calculated heat transfer by the CAHE to be 6.70×10^7 Btu/hr. The specified gas inlet temperature was 1110°F; the computed value was 1109.98°F.

CACHE printout - cont'd

This page contains the average results of the final computation.
The notes and comments for the similar printout on the preceding pages are generally applicable.

TEMP DIF. AT COLD END = 263.892
TEMP DIF AT HOT END = 926.625

LOG MEAN DELTA T = 527.65

GRAND TOTAL INTER SEGMENT ITER.= 240

TOTAL ENTIRE LENGTH ITER. = 4

TOTAL SEGMENT ITER. = 80

MISC AVGS:

H7 = 2833.36 H8 = 2462.71 H9 = 50.9581

U,AVG= 48.9007

REF TO GAS SIDE AND LOG MEAN DELTA T

U BAR= 49.1712

BASED ON Q TOT AND SEG AVG DELTA T

U,AVG OF SEG U'S = 48.7146

DELTA T,AVG= 529.665

BASED ON AVG OF SEG U'S

GAS PROP EVAL AT MEAN FILM TEMP

N,RE,H2O,AVG = 157133.

EFFEC GAS SIDE AREA = 2598.3

Z ERROR IN INLET GAS TEMP = 2.04825E-3
END OF CACHE

TIME: 10.06 SECS.

This is the total time required
by the ORNL PDP-10 computer to
make this Case B computation.

APPENDIX B

The FULTN Program

General Description

The FULTN program is used to develop the dimensions and flow parameters required by CACHE to calculate CAHE performance in the Fulton Generation Station.⁸ A nearly similar program, CONFIG, has been used for Summit Station CAHE performance.

Originally, these programs were part of CACHE. As CACHE evolved and expanded, it became advantageous to develop separate programs to describe the different CAHE designs being studied. A file named EXCH is used to store the parameters developed by FULTN and required by CACHE. The two programs are linked with a "CHAIN CACHE" instruction, line 5000 in FULTN.

Some of the dimensional parameters produced by FULTN are for a conceptual and idealized heat exchanger and may not agree precisely with the dimensions of the CAHE specified by the designers. For example, the Fulton Station CAHE contains 120 tubes in 9 sets of helical rings. This number permits a very logical arrangement for piping and headers (4x5x6) so that flow distribution problems are minimized. However, an array of 120 tubes cannot be conveniently arranged in 9 sets of helical rings such that the tubes are absolutely uniformly distributed in the CAHE. In this case the departure from uniformity is negligible. Again, FULTN computes tube spacings based on a specified heat exchanger length, not on a coil helix length. The axial length of the helix may be less than the specified exchanger length in order to accommodate lead-in tubes. Within reasonable limits the length of the tube helices is a designer's option and, because of practical considerations, may depart from a value based on a stated heat exchanger length. The dimensional parameters which are arbitrarily at the disposal of the designer and, therefore, are supplied by the user are noted in the list of FULTN in this appendix and in the list of symbols.

FULTN also computes mass flow rates and the user is required to "INPUT" gas and water flow rates. Provision is included to note the effect of a reduction in water flow rate from the specified value. A similar provision allows the user to specify an "effective surface area."

As a convenience to the user a section of the program is allotted to the computation of afterheat rates and integrated afterheat. This enables the user to compare the computation of CAHE heat transfer with the afterheat rate. The equation used for afterheat computation is from Ref. 1.

Two printout formats are available, a "total print" and a "reduced print." The choice must be specified by the user in response to an "INPUT" request during program operation. Once the dimensions have been established and printed, there is little need to repeat them each time a run is made; the "reduced print" instruction reduces on-line time since it limits printout to those parameters which are necessary and useful for identifying the conditions that govern the particular computation. A substantial amount of the dimensional data which is computed and available with the "total print" option is not required for input to the CACHE program. It serves merely to provide a general description of the conceptual design of the CAHE.

The list and typical printouts, augmented by the list of symbols, are largely self-explanatory. These are included in this appendix.

Dimensions of a Conceptual or Idealized CAHE

The original version of FULTN was based on an ideal design with these general characteristics:

1. The tubes are uniformly arranged and distributed within the annular volume enclosed by coaxial cylindrical shells.
2. The "length" of the CAHE is defined to be the axial length of the tube helices (measured on the axial centerline of the CAHE).
3. The effective surface area for heat transfer is limited to the helically coiled portion of the tubes with an estimated reduction for tube support hardware.

The remainder of this section outlines the relatively simple geometrical relationships used in FULTN to establish the dimensional parameters required to calculate heat exchange rates with CACHE.

Figure B-1 is a view with the line of sight parallel to the heat exchanger axis. The requirement that the tubes be uniformly distributed dictates that the radial distance, S_T , between any pair of adjacent helical sets of tubes be a constant. S_7 in this diagram is equivalent to the dimension, S_T , defined by Grimison³ for parallel straight tube heat exchangers. For those CAHE calculations in which the value of S_T was supplied by the designer, the latter value was chosen in order to specify the equation for the Nusselt number.

Figure B-2 is a diagram of a typical, conceptual header and the start of a ring of helical tubes. For uniform tube occupancy in the heat exchanger, the dimension, P_7 , must be constant for all the headers regardless of the helix diameter of the set of tubes. The dimension S_8 is assumed to be equivalent to S_T as defined by Grimison³ for in-line tubes in straight tube heat exchangers. The designer's specification for S_8

was used in preference to the ideal value on Fig. B-2. It is assumed that all tubes are of identical length and that all tube helices are coiled with the same helix angle, B9.

Figures B-1 and B-2, taken separately, may leave the impression that, neglecting the slight slope and curvature of the tubes, the tube array in a CAHE is entirely comparable and similar to an in-line array of straight tubes. Such is not the case. Whereas adjacent tubes in the same helical set are in-line, adjacent tubes in adjoining sets or rings are likely not to be in line when viewed in a direction normal to the common axis of the tube helices. In short, the tube array is neither wholly in-line nor wholly staggered. The designers have assumed that the array is 75% in-line, 25% staggered. This has been considered reasonable and acceptable.

Figure B-3 is an elevation view of a typical tube spiral showing the symbols used to designate the dimensions.

The equations which follow were developed to initiate an idealized, conceptual design of a typical CAHE. If complete design dimensions are available to the program user, they are not required. This has seldom been the case and they are useful for calculating averaged dimensions and to estimate the configuration required to attain a specified performance.

Equations and Symbols for the Dimensions of a Conceptual Helically Tubed Heat Exchanger

Where the symbols are defined they are shown as they appear on Figs. B-1, B-2, and B-3 and, also (entirely upper case) as they are used in the FULTN program.

1. The transverse area between the inner and outer heat exchanger shells, Fig. B-1.

$$A_{ann} = A5 = \pi (R_{os}^2 - R_{is}^2) \quad (B-1a)$$

R_{os} = X9, the inside diameter of the outer shell.

R_{is} = X8, the outside diameter of the inner shell.

A_{ann} = the area bounded by R_{is} and R_{os}

If the tube-to-shell clearances are made equal to the tube-to-tube clearances, then

$$A_{ann} = A5 = \pi [(R_M + R_1) (R_M - R_1 + 2S_T - d)] \quad (B-1b)$$

ORNL DWG 75-11390

Fig. B-1. Plan view diagram of a helically-tubed Core Auxiliary Heat Exchanger (CAHE)

ORNL-DWG 75-11385

Fig. B-2. Vertical, developed section through a typical inlet header and the start of a helical set or ring of tubes

ORNL DWG 75-11388

Fig. B-3. A typical helically wound tube in the j th tube ring

R_m = X7, radius, measured to the tube bore centerline, of the outermost ring or set of tubes.

R_1 = X6, radius, measured to the tube bore centerline, of the innermost ring or set of tubes.

S_T = S7, radial (transverse) tube spacing; the distance separating adjacent sets or rings of tubes measured from tube bore centerlines.

d = D2, the outside diameter of the tubing.

2. Projected area of the tubing, Fig. B-1.

$$A_{\text{proj}} = A8 = \pi d M (R_1 + R_m) \quad (\text{B-2})$$

A_{proj} , the area of all tube rings projected on a plane normal to the axial centerline of the heat exchanger.

R_1 , the radius of the innermost set of helical tubes measured to the tube bore centerlines.

R_m , the radius of the outermost set of helical tubes measured to the tube bore centerlines.

M , the number of sets or rings of helical tubes.
 $M = 4$ in Fig. B-1.

d , the outside diameter of the tubing.

3. Flow area for gas, Fig. B-1.

$$A_{\text{gf}} = A9 = A_{\text{ann}} - A_{\text{proj}} - A_{\text{ts}} \quad (\text{B-3})$$

$$A9 = A5 - A8 - B$$

A_{gf} , the unobstructed area for gas flow through the heat exchanger. This is the "see through" area for an observer whose line of sight is parallel to the axis of the heat exchanger and is the area that determines the mass flow rate.

A_{ann} and A_{proj} are defined in (1) and (2) above.

$A_{\text{ts}} = B$, the area of the tube support structure projected on a plane normal to the axial centerline of the heat exchanger.

4. Longitudinal tube spacing, Fig. B-2.

$$S_L = S_8 = P_c \tan \alpha \quad (B-4)$$

S_L , the distance, in a direction parallel to the heat exchanger axis and measured from the tube bore centerlines, of adjacent tubes in the same ring or set of helical tubes.

$\alpha = B_9$; the helix angle, Fig. B-2 or B-3.

$$\alpha = \sin^{-1} \left(\frac{L_h}{L_t} \right)$$

$L_h = L_0$, the overall length, measured parallel to the axis of the tube helix, of the helically coiled section of any tube in the heat exchanger. This is the conceptual length of the exchanger.

$L_t = L_5$, the total length, measured along the tube bore, of the helically coiled section of any tube in the heat exchanger.

$P_c = P_7$, the circular pitch of the header holes or the downcomers at the start (or end) of the helically wound sections, Fig. B-2.

NOTE: Uniform tube occupancy in the heat exchanger requires that S_L and S_T (Eq. B-1 and Fig. B-1) be uniform; therefore, P_c and α are constant for all tube rings.

5. Normal tube spacing (normal pitch).

$$S_\alpha = S_9 = P_c \sin \alpha \quad (B-5)$$

$$S_\alpha = \frac{P_c L_h}{L_t}$$

S_α , the shortest distance, Fig. B-2, between the tube bore centerlines of adjacent tubes in the same tube ring or set of tubes.

The parameters on the right hand side of Eq. (B-5) are defined for Eq. (B-4) above.

6. Surface area covered by the tube support structure.

The support structure is considered to be equivalent to a series of relatively thin, flat plates of length, L_h , the length of the heat exchanger and width, $R_{os} - R_{is}$, the radial dimension of the annulus, Fig. B-1. The plates are disposed radially in the annular space and the tubes pass through close fitting holes punched or drilled in the plates. The projected area, of the plates, A_{ts} , refer to Eq. (B-3), is given by

$$A_{ts} = B = w (R_{os} - R_{is})$$

$$R_{os} = X9; R_{is} = X8$$

and

$$w = \frac{A_{ts}}{R_{os} - R_{is}}, \text{ the sum of the plate thicknesses}$$

Obviously the support structure model can be considered as a single plate of thickness w . When a tube passes through a hole in the plate the loss of surface area exposed to the gas stream is

$$\begin{aligned} (\text{Area lost per} &= A_{\ell t} = \pi d w \\ \text{tube per hole}) \end{aligned}$$

$d = D2$, the outside diameter of the tubing.

Consider any one of the sets or rings of helically wound tubes; the number of holes required for passage of all tubes through the plate, Fig. B-2.

$$\text{holes per ring} = \frac{L_h}{S_L} + 1$$

and if there are M sets or rings the total number of holes in the plate is

$$N_t = M \left(\frac{L_h}{S_L} + 1 \right)$$

$$M = J9$$

$L_h = L0$, overall conceptual length of the heat exchanger.

$S_L = S8$, longitudinal tube pitch, Eq. (B-4).

The total tube area shielded by the support structure, A_{sh} , is

$$A_{sh} = B3 = N_t A_{\ell t}$$

$$A_{sh} = \pi d w M \left(\frac{L_h}{S_L} + 1 \right) \quad (B-6a)$$

Using the symbols, but not the format, in FULTN, Eq. (B-6a) is written, line 900

$$B3 = B*J9* \left(\frac{12*L0}{S8} + 1 \right) \left(\frac{\pi*D2}{12*(X9-X8)} \right) \quad (B-6b)$$

The symbols have been defined in the preceding text.

This model of the support structure, albeit wholly impracticable was developed because the only information available has been a numerical value for B.

Table B-1 contains the dimensions and flow parameters of the CAHE whose performance at 5.4 hours after a Design Basis Depressurization Accident (DBDA) has been presented. The list and printout by FULTN that follow are based on these dimensions and preceded the CACHE computation in Appendix A.

Table B-1

The Fulton Generating Station Core Auxiliary Heat Exchanger (CAHE).
Dimensions and Operating Parameters Used in the FULTN and CACHE
Programs Listed in this Report

The symbols and a relevant
line number in FULTN
are in parentheses

A. Overall Dimensions

1. Length, (L0, 485)	12.7 ft.
2. Outside radius, inner shell; (X8, 780)	1.042 ft.
3. Inside radius, outer shell; (X9, 820)	3.000 ft.
4. Void volume; (gas space); (G7, 1040)	258.5 ft. ³
5. Tubing volume; (G8, 1020)	57.3 ft. ³
6. Total volume; (G5, 1000)	315.8 ft. ³

B. Tubing

1. Outside diameter; (D2, 500)	0.0833 ft. = 1.00 in.
2. Wall thickness; (W8, 505)	0.00833 ft. = 0.100 in.
3. Inside diameter; (D0, 540)	0.0667 ft. = 0.80 in.
4. Length, uncoiled; (L5, 600)	87.5 ft.
5. Surface areas	
a. Inner surface	0.2094 ft ² /ft
b. Log mean tube wall	0.2346 ft ² /ft
c. Outer surface	0.2618 ft ² /ft
6. Material	
a. Upper 75% of length; 1/2% Cr-1/2% Mo steel	
b. Lower 25% of length; 2 1/4% Cr-1% Mo steel	

C. Tube Bundle

1. Total number of tubes; (N0, 520)	120
2. Total surface areas	
a. Inner, H ₂ O side	2199 ft ²
b. Log mean tube wall	2464 ft ²
c. Outer, gas side (A0, 350)	2749 ft ²

These areas, C.2 above, are based on 100% of the areas of 120 tubes specified to be 87.5 ft. in length.

3. "Effective" surface areas	
a. Outer, gas, side; (A, 920)	2598 ft ²

This is an "expected" value specified by the designers. The inner surface and tube wall areas were reduced proportionately to obtain their "effective" values. This value, 2598 ft² is 96.65% of the total area and is programmed with the "INPUT E" request (E = 0.9665), lines 250 and 260.

4. Total number of helical sets or rings of tubes; (J9, 620)	9
5. Tube-to-tube spacing	
a. Radius, innermost ring; (X6, 760)	1.150 ft = 13.805 in.
b. Radius, outermost ring; (X7, 800)	2.891 ft = 34.693 in.
c. Radial distance, ring to-ring; (S7, 660)	0.2175 ft = 2.61 in.
d. Vertical distance, tube-to-tube, in the same helical set; (S8, 720)	0.139 ft = 1.67 in.

The above dimensions are measured from tube bore centerlines.

The above dimensions, 5 (a) to (d), incl., are idealized and subject to modification by the designer. The vertical distance, 5(d), was stated to be 1.5 in. and this value was used when it influenced the computations.

D. Flow Areas

a. Annular area; (A5, 840)	24.866 ft ²
b. Projected area of tube helices on a plane normal to the heat exchanger axis; (A8, 860)	9.523 ft ²
c. Projected area of the tube support structure (B, 280)	0.547 ft ²
d. Unobstructed flow area for gas; (A9, 920)	14.796 ft ²
e. Flow area for water (A7, 1480)	0.419 ft ²

E. Operational parameters

a. Water flow rate	1,050,200 lb/hr
b. Gas flow rate	175,860 lb/hr
c. Reactor power, before shutdown	3150 Mw(t)

These data are supplied by the user in response to "INPUT" requests during program operation. Refer to lines 1080, 1100, 1220, 1240, and 1280.

LIST OF SYMBOLS IN THE FULTN PROGRAM

If (CACHE) appears with the symbol the variable or parameter is stored in the file EXCH and transmitted to CACHE with the "CHAIN CACHE" statement.

- A, (CACHE); "Effective" gas side heat transfer area of the heat exchanger tube, ft^2 , (line 920).
- A0, (CACHE); Total gas side surface area of the helically wound portions of all heat exchanger tubes, ft^2 , (line 350).
- A1, (CACHE); Effective water side heat transfer area of all the heat exchanger tubes in an axial length segment, (see Fig. B-2), ft^2 , (line 980).
- A2, (CACHE); Effective tube wall heat transfer area of all the heat exchanger tubes in an axial length segment, (see Fig. B-2), ft^2 , (line 980).
- A3, (CACHE); Effective gas side heat transfer area of all the heat exchanger tubes in an axial length segment, (see Fig. B-2), ft^2 , (line 960).
- A5; Transverse area of the annular space between the inner and outer shells, ft^2 , (line 840).
- A7, (CACHE); Total transverse area for water flow, ft^2 , (line 1480).
- A8; Projected area, on a plane normal to the axial centerline of the heat exchanger, of the helical sets or rings of tubes, ft^2 , (line 860).
- A9, (CACHE); Total transverse area for gas flow, ft^2 , (line 880).
- B; Projected area, normal to the axial centerline of the heat exchanger, of the tube support structure, ft^2 , (line 280).
- B3; The reduction in useful outer surface tube area affected by the tube support structure, ft^2 , (line 900).
- B9; The helix angle of the spirally coiled tubes, radians, (line 700).

- D0; Inside diameter of the heat exchanger tubing, inches, (line 540).
- D1, (CACHE); Inside diameter of the heat exchanger tubing, feet, (line 560).
- D2, (CACHE); Outside diameter of the heat exchanger tubing, inches, (line 500).
- D3, (CACHE); Equivalent diameter in the Reynolds number for gas flow, ft, (lines 3000 to 3080, inclusive). NOTE: In the following list of FULTN the equivalent diameter is defined by lines 2880 and 3080.
- E, (CACHE); Fraction of total tube area considered to be effective, (lines 250, 260).
- E1; Effective fraction of the nominal, specified water flow rate, lbs/hr, (lines 265, 270). NOTE: Differs from 1.00 only when the user desires to determine the effect of changing the water flow rate, (lines 265, 270).
- G1, (CACHE); Mass flow rate of water, $\text{lb/ft}^2\text{-hr}$, (line 1500).
- G3, (CACHE); Mass flow rate of gas, $\text{lb/ft}^2\text{-hr}$, (line 1520).
- G5; Total volume of the heat exchanger; the volume between the inner and outer shells, ft^3 , (line 1000).
- G6; The free flow volume, ft^3 , (line 1060).
- G7; The total void volume, the volume between the heat exchanger shells, G5, less the volume occupied by tubes, G8, ft^3 , (line 1020).
- G8; The total volume occupied by the heat exchanger tubes, ft^3 , (line 1020).
- I; Integrated afterheat, Mw-hr, (line 1345).
- I1; Integrated afterheat, Btu, (line 1350).
- J9; The number of rings or sets of helically coiled tubes.
- L0, (CACHE); The axial length (conceptual) of the heat exchanger; i.e., the length of the tube helices measured on the axial center-line of the heat exchanger, ft, (line 485).

- L1, (CACHE); The axial length of a segment of the heat exchanger, Fig. B-2, ft, (line 490).
- L5, (CACHE); The conceptual or idealized length of the helically coiled section of a typical tube measured on the axial centerline of the tube, ft, (line 600).
- L9, (CACHE); The wall thickness of the heat exchanger tubes, ft, (line 520).
- M0, (CACHE); The number of axial length segments into which the heat exchanger is subdivided for computation; an arbitrary number established by the user, (line 300).
- N0, (CACHE); The total number of tubes in the heat exchanger, (line 580).
- N9; An integer selected during program operation by the user, (N9 = 1, 2, or 3). The value of N9 determines the equation that defines the gas side equivalent diameter, D3. In the present version of FULTN this selection mode is by-passed, (lines 2840-3080).
- P7; The circumferential spacing, (header pitch), of the lead-in tubes, Fig. B-2, (line 640), inches.
- S7; The radial spacing (pitch), of the tubes in adjacent tube circles or rings measured from the axial centerlines of the tubes, ft, (line 660). NOTE: This is the transverse tube spacing S_T as defined by Grimison³.
- S8; The vertical spacing (pitch), of adjacent tubes in the same tube circle or ring of tubes measured from the tube centerlines and in a direction of the heat exchanger, inches, (line 720). NOTE: This distance may be taken to be equivalent to the longitudinal tube spacing S_L , as defined by Grimison³.
- S9; The minimum spacing, (normal pitch) of adjacent tubes in the same tube circle or ring of tubes, measured from the tube centerlines and in a direction normal to these centerlines, inches, (line 740).
- T; Elapsed time after shutdown, seconds, (line 1300).
- T1; Elapsed time after shutdown, hours, (lines 1240 and 1280).
- W0; The specified water flow rate, lb/hr, (lines 1080, 1100).

W1, (CACHE); Effective water flow rate, lb/hr, (line 1160).

W3, (CACHE); Gas flow rate, lb/hr, (lines 1080, 1100).

W8; Tube wall thickness, inches, (line 505).

X6; Radius of the tube bore centerlines of the innermost tube set or ring of tubes, ft, (line 760).

X7; Radius of tube bore centerlines of the outermost tube set or ring of tubes, ft, (line 800).

X8; Radius of the outer surface of the inner heat exchanger shell, ft, (line 780).

X9; Radius of the inner surface of the outer heat exchanger shell, ft, (line 820).

Y7; An intermediate variable, the sine of the tube coil helix angle, (line 680).

Z; An integer, $Z = 1$ or 2 , selected during program operation by the user. The value of Z determines the amount, of printout, total or reduced, by FULTN, (lines 1540 to 1600, inclusive).

FULTN - A PROGRAM FOR OBTAINING DIMENSIONS OF AND FLOW RATES
IN THE FULTON GENERATING STATION CORE AUXILIARY HEAT EXCHANGER

FULTN

14:07

01-JUL-75

```

100 FILE #1,"EXCH"
120 SCRATCH #1
250 PRINT " INPUT SERVICEABLE TUBE FRACTION"
260 INPUT E
265 PRINT " INPUT, EFFEC FRACTION OF 100% H2O FLOW"
270 INPUT E1
280 B = .547
300 M0 = 20
350 A0 = 2749
450 REM "A0 = TOTAL SURF AREA OF TUBES, GAS SIDE"
485 L0 = 12.7
490 L1 = L0/M0
500 D2 = 1
505 W8 = .1
520 L9 = W8/12
540 D0 = D2 - 2*W8
560 D1 = D0/12
580 N0 = 120
600 L5 = A0/(3.14159*(D2/12)*N0)
620 J9 = 9
640 P7 = 11.4
660 S7 = 2.611/12
680 Y7 = L0/L5
700 B9 = ATN(Y7/SQR(1-Y7^2))
720 S8 = P7*(Y7/SQR(1-Y7^2))
740 S9 = P7*Y7

```

Notes:

- (1) M0, line 300, is the number of segments used to make the computation by CACHE, (M0 = N in Fig. 2).
- (2) The design dimensions of the Fulton Station CAHE are established by lines 280, 350, 485, 500, 505, 580, 620, 640, 660, 760, 780, and 800.

FULTN, (Continued)

```

760 X6 = 13.805/12
780 X8 = 12.5/12
800 X7 = X6 + S7*(J9-1)
820 X9 = 3
840 A5 = 3.14159*(X9+2-X8+2)
860 A8 = 3.14159*J9*(D2/12)*(X6+X7)
880 A9 = A5 - A8 - B
900 B3 = B*J9*((12*L0/S8) + 1)*(3.14159*(D2/12))/(X9 - X8)
920 A = E*(A0-B3)
940 A1 = (D0/D2)*A/M0
960 A3 = A/M0
980 A2 = (A3-A1)/LOG(A3/A1)
1000 G5 = A5*L0
1020 G8 = .7854*((D2/12)+2)*L5*N0
1040 G7 = G5-G8
1060 G6 = A9*L0
1080 PRINT " INPUT, 100%, (MAX), TOT EFFEC FLOWS, W0 AND W3"
1100 INPUT W0, W3
1120 PRINT " MAX H2O ="W0"LB/HR", "MAX GAS ="W3 "LB/HR"
1140 PRINT " ", "WITH HT EXCHGR 100% EFFEC"
1160 W1 = E1*W0
1180 PRINT "EFFEC H2O FLOW ="W1"LB/HR"
1200 PRINT
1220 PRINT "INPUT, OPERATING THERMAL POWER = Q0 ,MW"
1240 PRINT "INPUT, ELAPSED TIME AFTER S'DOWN=T1, HRS"
1260 PRINT
1280 INPUT Q0, T1
1300 T = 3600*T1
1320 Q = .128*Q0/(T+.261)
1340 Q1 = 3.41214E6*Q
1345 I = ((.1732/3600)*Q0)*(T+.739)
1350 I1 = 3.41214E6*I
1360 PRINT "NORMALIZED A'HEAT RATE="Q/Q0
1380 PRINT

```

FULTN, (Continued)

```

1400 PRINT "100% A'HEAT =" Q"MW =" Q1 "BTU/HR"
1420 PRINT "110% A'HEAT =" 1.1*Q"MW =" 1.1*Q1 "BTU/HR"
1440 PRINT "120% A'HEAT =" 1.2*Q"MW =" 1.2*Q1 "BTU/HR"
1460 PRINT
1465 PRINT "100% INTEG HEAT=" I "MW-HR=" I1 "BTU"
1470 PRINT "110% INTEG HEAT =" 1.1*I "MW-HR =" 1.1*I1 "BTU"
1475 PRINT "120% INTEG HEAT =" 1.2*I "MW-HR =" 1.2*I1 "BTU"
1477 PRINT
1480 A7 = .7854*D1+2*N0
1500 G1 = W1/A7
1520 G3 = W3/A9
1540 PRINT "INPUT Z,PRINTOUT SELECT"
1560 PRINT "IF Z = 1 --REDUCED PRINT"
1580 PRINT "IF Z = 2-- TOTAL PRINT"
1600 INPUT Z
1620 IF Z = 1 THEN 2480
1640 PRINT "TUBE I.D.=" D0"IN.", "O.D.=" D2"IN."
1660 PRINT "TUBE LEN =" L5 "FT", "TOT TUBES =" N0
1680 PRINT "EXCHNGR LEN,FT =" L0
1700 PRINT
1720 PRINT "TOTAL TUBE SURF AREAS:"
1740 PRINT " ", "A,H20=" A0*D0/D2, "A,HE=" A0
1760 PRINT " LOG MEAN TUBE AREA =" A0*(1-(D0/D2))/LOG(D2/D0)
1780 PRINT " EFFEC GAS SIDE AREA FRAC =" E
1800 PRINT " EFFFC GAS SIDE SURF ARFA=" A
1820 PRINT
1840 PRINT " PROJ AREA TUBE SUPPORTS =" B" SQ FT"
1860 PRINT
1880 PRINT "ANN O.D. =" 2*X9"FT=" 24*X9"IN."
1900 PRINT "ANN I.D. =" 2*X8"FT=" 24*X8"IN."
1920 PRINT
1940 PRINT "HELIX ANGLE,ALPHA =" B9*180/3.14159
1960 PRINT "HEADER PITCH =" P7 "IN."
1980 PRINT "RADIAL PITCH =" S7 "FT=" S7*12 "IN."
2000 PRINT "VERT PITCH =" S8 "IN.=" S8/12 "FT"
2020 PRINT "NORM PITCH,(S,ALPHA) =" S9"IN.=" S9/12 "FT"
2040 PRINT
2060 PRINT "INNER RING RAD =" X6 "FT"
2080 PRINT "OUTER RING RAD =" X7 "FT"
2100 PRINT
2120 PRINT "NO TUBE CIRCS=" J9
2140 PRINT "ANN AREA =" A5"SQ FT"
2160 PRINT "PROJ AREA,TUBE RINGS =" A8 "SQ FT"
2180 PRINT " ", "DIFF =" A5-A8 "SQ FT"
2200 PRINT "SURF AREA RED. BY TUBE SUPPORTS=" B3"SQ FT"
2220 PRINT "100% EXPOSED TUBE SURF =" A0 - B3" SQ FT"
2260 PRINT

```

FULTN, (Continued)

```

2280 PRINT "TOT TUBE VOL=" G8 "FT+3"
2300 PRINT "TOT VOID VOL =" G7"FT+3"
2320 PRINT "TOT EXCHGR VOL=" G5 "FT+3"
2340 PRINT
2360 PRINT "FREE FLOW VOL=" G6 "FT+3"
2380 PRINT "TOT FLOW RATES,LBS/HR"
2400 PRINT " ", "H2O="W1, "GAS="W3
2420 PRINT
2440 PRINT "FLOW AREAS, SQ FT"
2460 PRINT " ", "H2O="A7, "GAS, UNOBSTRUCTED="A9
2480 PRINT "MASS FLOW RATES, LBS/SQ FT-HR"
2500 PRINT " ", "H2O="G1, "GAS="G3
2520 PRINT
2540 PRINT "TOT GAS SIDE TUBE AREA ="A0 "SQ FT"
2560 PRINT "EFFEC GAS SIDE AREA ="A
2580 PRINT
2600 PRINT "EFFEC SEG AREAS:"
2620 PRINT "A1="A1, "A2=" A2, "A3="A3
2640 PRINT
2660 PRINT "SEG LEN, FT=" L1
2680 PRINT "NO SEGS =" M0
2700 PRINT
2720 PRINT "SEG NO 1 AT LOW TEMP END"
2740 PRINT
2760 PRINT
2780 G0 T0 3080
2800 PRINT "INPUT HEL SIDE EQUIV DIA SPEC = N9"
2820 PRINT
2840 PRINT "IF N9 = 1, EQ DIA = 4*(FREE FLOW VOL)/(TUBE SURF)"
2860 PRINT "IF N9 = 2, EQ DIA = 4*(TOT VOID VOL)/(TUBE SURF)"
2880 PRINT "IF N9 = 3, EQ DIA = TUBE O. DIA"
2900 PRINT
2920 PRINT "INPUT N9 =1 OR 2 OR 3, SEE ABOVE"
2940 INPUT N9
2960 IF N9 = 2 THEN 3040
2980 IF N9 = 3 THEN 3080
3000 D3 = 4*G6/A0
3020 G0 T0 3120
3040 D3 = 4*G7/A0
3060 G0 T0 3120
3080 D3 = D2/12
3100 G0 T0 3140
3120 PRINT "N9=" N9 "SEE ABOVE"
3140 PRINT "EQ DIAS, FT:", "H2O SIDE ="D1, "GAS SIDE ="D3
3160 WRITE #1, A0, A1, A2, A3, D1, D2, D3, L0, L1, L9, A7, A9, N0
3180 WRITE #1, M0, W1, W3, G1, G3, A
3200 WRITE #1, E
5000 CHAIN CACHE
9999 END

```

TYPICAL TOTAL PRINTOUT^a BY FULTN

FULTN

18:23

23-JUN-75

```

INPUT SERVICEABLE TUBE FRACTION
?.9665
INPUT, EFFEC FRACTION OF 100% H2O FLOW
?1
INPUT, 100%, (MAX), TOT EFFEC FLOWS, W0 AND W3
?1.0502E6, 175860
MAX H2O = 1050200 LB/HR      MAX GAS = 175860 LB/HR
      WITH HT EXCHGR 100% EFFEC
EFFEC H2O FLOW = 1050200 LB/HR

INPUT, OPERATING THERMAL POWER = 00 ,MW
INPUT, ELAPSED TIME AFTER S'DOWN=T1, HRS

?3150, 5.3964
NORMALIZED A'HEAT RATE= 9.72604E-3

100% A'HEAT = 30.637 MW = 1.04538E+8 BTU/HR
110% A'HEAT = 33.7007 MW = 1.14992E+8 BTU/HR
120% A'HEAT = 36.7644 MW = 1.25445E+8 BTU/HR

100% INTEG HEAT= 223.712 MW-HR= 7.63335E+8 BTU
110% INTEG HEAT = 246.083 MW-HR = 8.39669E+8 BTU
120% INTEG HEAT = 268.454 MW-HR = 9.16002E+8 BTU
INPUT Z, PRINTOUT SELECT
IF Z = 1 --REDUCED PRINT
IF Z = 2-- TOTAL PRINT
?2

TUBE I.D.= 0.8 IN.      O.D.= 1 IN.
TUBE LEN = 87.5035 FT   TOT TUBES = 120
EXCHNGR LEN, FT = 12.7

TOTAL TUBE SURF AREAS:
      A, H2O= 2199.2 A, HE= 2749
LOG MEAN TUBE AREA = 2463.88
EFFEC GAS SIDE AREA FRAC = 0.9665
EFFEC GAS SIDE SURF AREA= 2598.3

PROJ AREA TUBE SUPPORTS = 0.547 SQ FT

ANN O.D. = 6 FT= 72 IN.
ANN I.D. = 2.08333 FT= 25. IN.

```

(a) The dimensions of this heat exchanger are in Table B-1.

PRINTOUT BY FULTN, (Continued)

HELIX ANGLE, ALPHA = 8.34523
 HEADER PITCH = 11.4 IN.
 RADIAL PITCH = 0.217583 FT= 2.611 IN.

VERT PITCH = 1.67227 IN.= 0.139356 FT
 NORM PITCH, (S, ALPHA) = 1.65456 IN.= 0.13788 FT

INNER RING RAD = 1.15042 FT
 OUTER RING RAD = 2.89108 FT

NØ TUBE CIRCS= 9
 ANN AREA = 24.8655 SQ FT
 PROJ AREA, TUBE RINGS = 9.52255 SQ FT
 DIFF = 15.3429 SQ FT
 SURF AREA RED. BY TUBE SUPPORTS= 60.6359 SQ FT
 100% EXPOSED TUBE SURF = 2688.36 SQ FT

TØT TUBE VOL= 57.271 FT³
 TØT VOID VOL = 258.52 FT³
 TØT EXCHGR VOL= 315.791 FT³

FREE FLOW VOL= 187.908 FT³
 TØT FLOW RATES, LBS/HR
 H2O= 1050200 GAS= 175860

FLOW AREAS, SQ FT
 H2O= 0.41888 GAS, UNØBSTRUCTED= 14.7959
 MASS FLOW RATES, LBS/SQ FT-HR
 H2O= 2.50716E+6 GAS= 11885.7

TØT GAS SIDE TUBE AREA = 2749 SQ FT
 EFFEC GAS SIDE AREA = 2598.3

EFFEC SEG AREAS:
 A1= 103.932 A2= 116.441 A3= 129.915

SEG LEN, FT= 0.635
 NØ SEGS = 20

SEG NØ 1 AT LOW TEMP END

EQ DIAS, FT: H2O SIDE= 6.66667E-2 HEL SIDE= 8.33333E-2

TYPICAL REDUCED PRINTOUT BY FULTN

FULTN

18:27

23-JUN-75

INPUT SERVICEABLE TUBE FRACTION

?.9665

INPUT, EFFEC FRACTION OF 100% H2O FLOW

?1

INPUT, 100%, (MAX), TOT EFFEC FLOWS, W0 AND W3

?1.0502E6

?175860

MAX H2O = 1050200 LB/HR MAX GAS = 175860 LB/HR
WITH HT EXCHGR 100% EFFEC

EFFEC H2O FLOW = 1050200 LB/HR

INPUT, OPERATING THERMAL POWER = 00 ,MW

INPUT, ELAPSED TIME AFTER S'DOWN=T1, HRS

?3150, 5.3964

NORMALIZED A'HEAT RATE= 9.72604E-3

100% A'HEAT = 30.637 MW = 1.04538E+8 BTU/HR

110% A'HEAT = 33.7007 MW = 1.14992E+8 BTU/HR

120% A'HEAT = 36.7644 MW = 1.25445E+8 BTU/HR

100% INTEG HEAT= 223.712 MW-HR= 7.63335E+8 BTU

110% INTEG HEAT = 246.083 MW-HR = 8.39669E+8 BTU

120% INTEG HEAT = 268.454 MW-HR = 9.16002E+8 BTU

INPUT Z, PRINTOUT SELECT

IF Z = 1 --REDUCED PRINT

IF Z = 2-- TOTAL PRINT

?1

MASS FLOW RATES, LBS/SQ FT-HR

H2O= 2.50716E+6

GAS= 11885.7

TOT GAS SIDE TUBE AREA = 2749 SQ FT

EFFEC GAS SIDE AREA = 2598.3

EFFEC SEG AREAS:

A1= 103.932 A2= 116.441 A3= 129.915

SEG LEN, FT= 0.635

NO SEGS = 20

SEG NO 1 AT LOW TEMP END

EQ DIAS, FT: H2O SIDE= 6.66667E-2

HEL SIDE= 8.33333E-2

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8. Fulton Generating Station, Units 1 and 2, Preliminary Safety Analysis Report (PSAR), Philadelphia Electric Co.

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