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Thermodynamic Properties of Water for Computer Simulation of Power Plants

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Commission

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ABSTRACT

Steam property evaluations may represent a significant portion of the computing time necessary for power system simulations. The iterative nature of the solutions for heat transfer and kinetic equations often requires thousands of steam property evaluations during the execution of a single program. Considerable savings may be realized by simplification of property evaluations.

Empirical equations have been obtained for the thermodynamic properties of water in the region of interest. To maintain thermodynamic consistency, the compressibility factor Z, in terms of pressure and temperature, was obtained by curve fitting, and the enthalpy, entropy, and internal energy were derived by standard relationships. Formulations for heat capacity, saturation temperature as a function of saturation pressure, the specific volume of saturated water as a function of saturation pressure, and specific volume of saturated water as a function of the saturation temperature were determined by curve fitting of independent equations. Derivatives were obtained by differentiation of the appropriate formulations.

Evaporator and superheater components of a liquid metal fast breeder reactor power plant simulator were chosen as test cases for the empirical representations. Results obtained using the empirical equations were comparable to those obtained using tabular values, but significant savings in computational costs were realized. Execution time for the

evaporator program with the empirical forms was approximately 27 percent less than for the program with tables. Execution time for the superheater program was approximately 23 percent less.

TABLE OF CONTENTS

LIST OF ILLUSTRATIONS

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CHAPTER 1

INTRODUCTION

Simulations involving steam power systems have become so complex that thousands of steam property evaluations are often necessary during the execution of a single program. Much of this complexity is due to the iterative nature of the solutions for heat transfer and kinetic equations. For a simulation the system may be divided into a large number of segments or, in the dynamic situation, into a large number of time steps, each requiring a number of iterations and corresponding property evaluations. The evaluation of the steam properties may represent a significant portion of the computing time necessary for the simulation. Therefore, any simplification in property evaluation could result in considerable savings.

Any scheme chosen to represent thermodynamic properties must have acceptable accuracy, be economical, give smooth values, employ convenient independent variables, be easy to compute, and require a minimum of switching from one formulation to another in the various regions of interest. Several methods are presently used. One of the simpler methods is table look-up. This method requires the storage of a large number of points in a regular grid and an interpolation scheme to obtain interstitial values. Often the properties are simply entered for a given range and the tables called whenever a value is needed. Discontinuities are a problem with this method.

In a more sophisticated approach, the properties are evaluated for a certain region or "window" (Campbell and Jenner, 1980) in an independent variable plane. An interpolation scheme is used to obtain the required value from the stored values. If a value is required which lies outside the window, the reference formulation is used. To maintain continuity at the boundaries of the window, a "blending" technique is used between the tabulated results and the reference formulation (Campbell and Jenner, 1980). This method again requires extensive computer storage. The reference formulation may also be very complex and, consequently, expensive to run. Some reference formulations have the additional drawback of requiring subregions with corresponding formulations in each region. These formulations are not necessarily continuous across the boundaries. The 1967 IFC Industrial Formulation has six subregions (Meyer et al., 1967). The Keenan, Keyes, Hill, and Moore formulation (Keenan et al., 1969) does not have the limitation of subregions, but it does have some difficulty in representing saturation line data. The Pollak 1976 formulation does not exhibit the problem with saturation data nor does it require subregions (Campbell and Jenner, 1980; Sato, Uematsu, and Watanabe, 1980). However, the formulation is complex and the evaluations require a large amount of computer time.

Cubic and bicubic splines are also used to compute the required thermodynamic properties. The cubic spline is often used when the functional form of the equations is not known but experimentally determined data pairs are available. If a set of data points is available, the functions may be approximated in the given intervals by cubic polynomials. To maintain smoothness, the first and second derivatives must be continuous at the end point ("knot") for adjacent intervals. In the case of two independent variables, bicubic splines are used over a rectangular domain. Since the saturation line represents a curved boundary, a transformation must be made and curved parts mapped onto rectangles. The knots and coefficients for the polynomials need to be calculated only once (Caldwell and Spragg, 1977). Although cubic and bicubic splines give an accurate and relatively fast method for computing thermodynamic properties, they still do not offer the flexibility necessary in dynamic studies.

A less expensive and more flexible alternative is to determine approximate curve fits for the required properties over a region of interest. Some accuracy may be sacrificed for ease of calculation and computational economy. Curve fitting often requires extensive effort to determine the appropriate empirical equations, but it does save considerable computer time in the actual calculation of the properties. Often independent curve fits are found for the different properties and regions required. For ease of computation and to decrease the possibility of discontinuities, the number of regions should be kept to a minimum. Ideally one equation would suffice for the entire region of interest.

To maintain thermodynamic consistency, it is desirable to determine an empirical representation for a given property in terms of the selected independent variables and then to use this representation to derive the other thermodynamic variables. In this work the compressibility factor Z, in terms of temperature and pressure, was chosen as the basic property and, along with a simple form for the zero pressure heat capacity, was used to derive enthalpy, entropy, internal energy, and heat capacity by using standard thermodynamic relationships. More complex expressions for the zero pressure heat capacity did not yield any improvement in the derived properties.

Several expressions gave satisfactory results for the compressibility and the derived enthalpy, entropy, and internal energy equations. However, the heat capacity proved to be quite troublesome, especially in representing saturation line data at higher pressures. It was decided that a separate empirical formula might be required to represent the heat capacity because of its anomalous behavior, especially near the critical region.

After each separate empirical formula for heat capacity was derived, it was used as a starting point, and a new expression was derived for compressibility which in turn was used to derive the other properties. Several expressions gave reasonable results for the heat capacity, although the errors in representing the heat capacity were uniformly larger than those encountered for the compressibility and the other derived thermodynamic properties. This approach proved unsuccessful since in every case the expressions derived from the new compressibility did not yield satisfactory results for the enthalpy, entropy, and internal energy equations.

As a further check, the best independently derived heat capacity equation was combined with each of two different compressibility equations to derive a pair of equations for each of the other properties. It was decided to adopt the equations derived from the compressibility alone because they are far simpler than the combined forms and the results are almost identical. It was also decided that a separate heat capacity representation would be used.

Independent forms were also determined for the saturation temperature as a function of saturation pressure, the specific volume of saturated water as a function of saturation pressure, and specific volume of saturated water as a function of the saturation temperature. Derivatives were obtained by differentiation of the independent forms or the appropriate equations derived from the compressibility.

Although the independent variables throughout most of this work are pressure and temperature, it is sometimes advantageous to use enthalpy as one of the independent variables. It was decided that it would be preferable to obtain additional forms by curve fitting rather than solve the previously obtained equations iteratively. A function was determined for the saturation temperature in terms of the fluid enthalpy. An expression for temperature in the superheat region as a function of pressure and enthalpy (Agrawal, 1978) considerably increases the flexibility of the previously derived and independently determined equations. Since the temperature can be calculated fairly simply, the independent variables for many of the equations can now be either pressure and temperature or pressure and enthalpy. The feasibility of using this method was checked by comparing the equation for the volume of the gas derived from the compressibility and an independently determined form (Agrawal, 1978) in terms of pressure and enthalpy. The independently determined equation gives only slightly better results than the derived form.

Chapter 2 contains methods for the derivation of the empirical equations, the ranges of applicability, and comparisons with literature values. Two examples of the application of the derived equations are given in Chapter 3. Execution times for programs utilizing the empirical equations and those using tabular values are also compared in Chapter 3.

CHAPTER 2

EMPIRICAL EQUATIONS FOR PROPERTIES OF STEAM AND WATER

A nonlinear least squares program BMDX 85 (Sampson, 1964) was used to obtain the parameters for the various specified functions by stepwise Gauss-Newton iterations. The equations described in this chapter are the ones used in the simulations in Chapter 3. Unless otherwise noted, the range of applicability is 1 bar to 150 bars and 373.15 K to 873.15 K in the superheat region, and 1 bar (372.78 K) to 160 bars (620.47 K) on the saturation line. Temperature (T) is given in K, pressure (P) in bars, specific volume (v) in cm^3/g, enthalpy (h) and internal energy (U) in J/g, and entropy (s) and heat capacity (Cp) in J/g K. The comparisons between calculated and reference values are given as relative percent differences and are for the ranges given above. Additional formulations, ranges, and descriptions may be found in Appendix A.

2-1 The Compressibility Factor Z

The compressibility is given by

$$Z = z_o + \frac{z_1 P}{T} + z_2 P + \frac{z_3 P^2}{T^2} + \frac{z_4 P^2}{T} + z_5 P^2 + \frac{z_6 P^3}{T} + z_7 P^3 + \frac{z_8 P^3}{T^2} + \frac{z_9 P^3}{T^3} \qquad (2.1)$$

where

$z_o = 1.0001$

$z_1 = -5.3391$

$z_2 = 6.1322\text{x}10^{-3}$

$z_3 = -7.4961\text{x}10^{1}$

$z_4 = 2.5547\text{x}10^{-1}$

$z_5 = -2.0580\text{x}10^{-4}$

$z_6 = -8.9010\text{x}10^{-3}$

$z_7 = 4.3505\text{x}10^{-6}$

$z_8 = 6.1847$

$z_9 = -1.4715\text{x}10^{3}$

The differences between the compressibilities calculated by using the given equation and those obtained by using tabulated volumes (Schmidt, 1969) in the well-known relationship, $Z=Pv/RT$, are less than or equal to 1.5%. Generally the differences are less than 1.0%.

2-2 Zero Pressure Heat Capacity

Parameters for the commonly-used form for heat capacity at zero pressure were determined along the zero pressure isobar in the range 373.15 K to 773.15 K. The zero pressure heat capacity is given by

$$Cp^\circ = c_1 + c_2T + c_3T^2 \qquad (2.2)$$

where

$c_1 = 1.7524$

$c_2 = 2.4936\text{x}10^{-4}$

$c_3 = 3.0978\text{x}10^{-7}$

The differences between the calculated and tabulated values (Schmidt, 1969) are less than 0.4% in the range 373.15 K to 873.15 K at zero pressure.

2-3 Saturation Temperature as a Function of Saturation Pressure

The saturation temperature is given by

$$T_{sat} = x_1 + x_2P + x_3P^2 + x_4P^3 + \frac{x_5}{P} + \frac{x_6}{P^2} + \frac{x_7}{P^3} + x_8P^{\frac{1}{2}} + \frac{x_9}{P^{\frac{1}{2}}} \tag{2.3}$$

where

$x_1 = 3.5653\times10^{2}$

$x_2 = -2.0611$

$x_3 = 5.7064\times10^{-3}$

$x_4 = -1.0110\times10^{-5}$

$x_5 = 2.1841$

$x_6 = -9.9885\times10^{-3}$

$x_7 = 2.5304\times10^{-5}$

$x_8 = 3.8845\times10^{1}$

$x_9 = -2.2420\times10^{1}$

When compared to tabulated values (Bain, 1964; Keenan et al., 1969) the calculated saturation temperatures differ by less than 0.5% over the entire saturation line.

2-4 Gas Volume and Derivatives

The volume was derived by using Eq. (2.1) and the relationship

$$v = \frac{ZRT}{P}$$

where $R = 4.619$ bar cm^3/g K.

The volume is given by

$$v = R\left(\frac{z_0 T}{P} + z_1 + z_2 T + \frac{z_3 P}{T} + z_4 P + z_5 PT + z_6 P^2 + z_7 P^2 T + \frac{z_8 P^2}{T} + \frac{z_9 P^2}{T^2}\right) . \qquad (2.4)$$

The differences between the calculated and tabulated volumes (Schmidt, 1969) in the superheat region are less than or equal to 1.5%.

A line in the surface, the saturation line, was checked by entering a saturation pressure, calculating the corresponding saturation temperature using Eq. (2.3), and then calculating the saturation gas volume by using Eq. (2.4). The differences between the calculated gas volumes and the tabulated values on the saturation line (Keenan et al., 1969) are less than 1.5%.

The total derivative of the saturated gas volume was obtained by using the derivatives of Eqs. (2.3) and (2.4) and the following relationship:

$$\left[\frac{dv}{dP}\right]_{sat} = \left(\frac{\partial v}{\partial P}\right)_T + \left(\frac{\partial v}{\partial T}\right)_P \left[\frac{dT}{dP}\right]_{sat}$$

Compared to values calculated by the finite difference method, the disagreement is less than 2.5%. The values obtained in both cases are smooth.

2-5 Enthalpy of the Vapor

The enthalpy was derived by using the following relationships:

$$dh = CpdT + \left[v - T\left(\frac{\partial v}{\partial T}\right)_P\right] dP$$

$$Cp = Cp^\circ - \int_0^P T \left(\frac{\partial^2 v}{\partial T^2}\right)_{P'} dP'$$

The zero pressure heat capacity, Cp°, is given by Eq. (2.2), and the volume and partial derivatives are obtained from section 2-4. The triple point was chosen as the reference point and the values set as follow:

$T_1 = 273.16$ K

$P_1 = 0.006113$ bars

$h_1 = 2501.3$ J/g

The enthalpy equation is

$$\begin{aligned} h = {} & c_1(T-T_1) + \frac{c_2}{2}(T^2-T_1^2) + \frac{c_3}{3}(T^3-T_1^3) \\ & + R_1\left[z_1(P-P_1) + z_3\left(\frac{P^2}{T} - \frac{P_1^2}{T_1}\right) + \frac{z_4}{2}(P^2-P_1^2)\right. \\ & \left. + \frac{z_6}{3}(P^3-P_1^3) + \frac{2}{3} z_8\left(\frac{P^3}{T} - \frac{P_1^3}{T_1}\right) + z_9\left(\frac{P^3}{T^2} - \frac{P_1^3}{T_1^2}\right)\right] + h_1 \end{aligned} \tag{2.5}$$

where $R_1 = 0.4619$ J/g K.

The maximum difference between tabulated (Schmidt, 1969) and calculated values is 2.3% and occurs at 873.15 K and 150 bars. Generally the differences are less than 2.0%. On the saturation line, the differences are equal to or less than 2.0%, whether given temperatures and pressures are used or the saturation temperature is calculated for the given saturation pressure by Eq. (2.3).

The total derivative for enthalpy with respect to pressure on the saturation line was obtained by taking the partial derivatives of Eq. (2.5) for enthalpy and the total derivative of Eq. (2.3) for temperature and applying the following relationship:

$$\left[\frac{dH}{dP}\right]_{sat} = \left(\frac{\partial H}{\partial P}\right)_T + \left(\frac{\partial H}{\partial T}\right)_P \left[\frac{dT}{dP}\right]_{sat}$$

The calculated values for $\left[\frac{dH}{dP}\right]_{sat}$ differ by orders of magnitude from those obtained by the finite difference method when the enthalpy values vary quite slowly over a range. When the variation is essentially zero any small difference between calculated and tabulated values results in a very large relative percent error. McClintock and Silvestri (1968) also warn that if finite difference techniques are applied in too small an interval, errors may result. The comparison is still useful in showing that the derived equation yields smoothly varying values, while the straight line interpolation finite difference method shows oscillations in $\left[\frac{dH}{dP}\right]_{sat}$.

2-6 Entropy of the Vapor

The entropy was derived by using the following relationships:

$$ds = \frac{Cp}{T}\, dT - \left(\frac{\partial v}{\partial T}\right)_P dP$$

$$Cp = Cp^\circ - \int_0^P T \left(\frac{\partial^2 v}{\partial T^2}\right)_{P^1} dP^1$$

The zero pressure heat capacity is given in Eq. (2.2) and the partial derivatives are obtained from section 2-4.

The triple point was again chosen as the reference point and the values set as follow:

$T_1 = 273.16$ K

$P_1 = 0.006113$ bars

$s_1 = 9.1571$ J/g K

The equation for the entropy is given by

$$\begin{aligned} s = {} & c_1 \ln\left(\frac{T}{T_1}\right) + c_2(T-T_1) + \frac{c_3}{2}(T^2-T_1^2) \\ & - R_1 \left[z_o \ln\left(\frac{P}{P_1}\right) + z_2(P-P_1) - \frac{z_3}{2}\left(\frac{P^2}{T^2} - \frac{P_1^2}{T_1^2}\right) \right. \\ & + \frac{z_5}{2}(P^2-P_1^2) + \frac{z_7}{3}(P^3-P_1^3) - \frac{z_8}{3}\left(\frac{P^3}{T^2} - \frac{P_1^3}{T_1^2}\right) \\ & \left. - \frac{2}{3} z_9 \left(\frac{P^3}{T^3} - \frac{P_1^3}{T_1^3}\right)\right] + s_1 \end{aligned} \tag{2.6}$$

where $R_1 = 0.4619$ J/g K.

The differences between calculated and tabulated values (Schmidt, 1969) are less than or equal to 1.7% in both the superheat region and on the saturation line. Generally the differences are less than 1.0%. The results on the saturation line are again almost identical when both temperatures and pressures are given or when saturation temperatures are calculated from saturation pressures by using Eq. (2.3).

2-7 Internal Energy of the Vapor

The internal energy was obtained by using the following relationship:

$$U = h - Pv \tag{2.7}$$

The internal energy is given by

$$\begin{aligned} U = {} & c_1(T-T_1) + \frac{c_2}{2}(T^2-T_1^2) + \frac{c_3}{3}(T^3-T_1^3) \\ & + R_1 \left[z_1(P-P_1) + z_3\left(\frac{P^2}{T} - \frac{P_1^2}{T_1}\right) + \frac{z_4}{2}(P^2-P_1^2) \right. \\ & \left. + \frac{z_6}{3}(P^3-P_1^3) + \frac{2}{3} z_8\left(\frac{P^3}{T} - \frac{P_1^3}{T_1}\right) + z_9\left(\frac{P^3}{T^2} - \frac{P_1^3}{T_1^2}\right) \right] \\ & - R_1 \left[z_0 T + z_1 P + z_2 PT + \frac{z_3 P^2}{T} + z_4 P^2 + z_5 P^2 T \right. \\ & \left. + z_6 P^3 + z_7 P^3 T + \frac{z_8 P^3}{T} + \frac{z_9 P^3}{T^2} \right] + h_1 \end{aligned} \tag{2.8}$$

where

$T_1 = 273.16$ K

$P_1 = 0.006113$ bars

$h_1 = 2501.3$ J/g

$R_1 = 0.4619$ J/g K

If the enthalpy and volume are calculated in another part of the program, one only need apply the appropriate conversion factor to the Pv term and use Eq. (2.7) to obtain the internal energy.

The differences between tabulated (Schmidt, 1969) and calculated values are equal to or less than 2.7%. Generally the differences are less than 1.5%.

The derivative of internal energy with respect to pressure on the saturation line was obtained by using the following expression:

$$\left[\frac{dU}{dP}\right]_{sat} = \left[\frac{dH}{dP}\right]_{sat} - P\left[\frac{dv}{dP}\right]_{sat} - Pv$$

The derivatives and volumes are given in sections 2-4 and 2-5. To make the units consistent, the appropriate conversion factor must be applied to the last two terms in the equation. The results are comparable to those obtained for $\left[\frac{dH}{dP}\right]_{sat}$ in section 2-5.

2-8 Specific Volume of Saturated Water

The volume of saturated water is given by

$$v_f = v_1 + v_2P + v_3P^2 + v_4P^3 + v_5P^4 \tag{2.9}$$

where

$v_1 = 1.0562$

$v_2 = 7.2848\text{x}10^{-3}$

$v_3 = -7.8325\text{x}10^{-5}$

$v_4 = 5.8788\text{x}10^{-7}$

$v_5 = -1.4035\text{x}10^{-9}$

The maximum difference between the tabulated (Bain, 1964) and calculated values is 1.9% and occurs at 1 bar. The differences are generally less than 0.5%.

The derivative of the saturated fluid volume with respect to pressure is obtained by differentiation of Eq. (2.9).

2-9 Enthalpy of Saturated Water

The enthalpy of saturated water is given by

$$h_f = w_1 + w_2P + w_3P^2 + w_4P^3 + w_5P^4 + w_6P^5 + \frac{w_7}{P} + \frac{w_8}{P^2} \qquad (2.10)$$

where

$w_1 = 6.4480\times10^{2}$

$w_2 = 1.8818\times10^{1}$

$w_3 = -2.7564\times10^{-1}$

$w_4 = 2.7955\times10^{-3}$

$w_5 = -1.4470\times10^{-5}$

$w_6 = 2.9450\times10^{-8}$

$w_7 = -4.8854\times10^{2}$

$w_8 = 2.4370\times10^{2}$

The largest differences between tabulated (Bain, 1964) and calculated values occur between 1 and 2 bars, reaching a maximum of 2.7%. Above 7 bars, the differences are less than or equal to 0.3%.

The derivative of the enthalpy of the fluid with respect to pressure is obtained by differentiation of Eq. (2.10).

2-10 Density of Saturated Water

The volume of saturated water is given by

$$v_f = d_1 + d_2T + d_3T^2 + d_4T^3 \qquad (2.11)$$

where

$d_1 = -4.4267$

$d_2 = 3.5980\times10^{-2}$

$d_3 = -8.0055\text{x}10^{-5}$

$d_4 = 6.1163\text{x}10^{-8}$

The density, ω, is given by

$$\omega = \frac{1}{v_f}$$

and has the units of g/cm^3.

The maximum difference between tabulated (Bain, 1964) and calculated volumes is less than 1.9%. The differences are generally less than 1.0%.

2-11 Specific Heat Capacity

The specific heat capacity form was determined by using the zero pressure heat capacity equation in section 2-2 and adding pressure and temperature dependent terms. The coefficients y_1 through y_{11} were determined by curve fitting. The specific heat capacity is given by

$$Cp = c_1 + c_2T + c_3T^2 + (y_1P^3 + y_2PT + y_3PT^2 + y_4PT^3 + y_5P^2T + y_6P^2T^2 + y_7P^2T^3 + y_8P^3T + y_9P^3T^2 + y_{10}P^3T^3)/(T-y_{11}P)^2 \tag{2.12}$$

where

$y_1 = 1.1120\text{x}10^{1}$

$y_2 = 5.2110\text{x}10^{1}$

$y_3 = -1.0089\text{x}10^{-1}$

$y_4 = 4.5463\text{x}10^{-5}$

$y_5 = 4.3768\text{x}10^{-1}$

$y_6 = -1.5544\text{x}10^{-3}$

$y_7 = 1.2808\text{x}10^{-6}$

$y_8 = -4.9183\text{x}10^{-2}$

$y_9 = 7.2632\text{x}10^{-5}$

$y_{10} = -3.5782\text{x}10^{-8}$

$y_{11} = 2.9971$

The differences between tabulated (Schmidt, 1969) and calculated values are less than 12.6% in the superheat region. On the saturation line the maximum difference is 11.4% for given temperatures and pressures. When the saturation temperature is calculated for a given saturation pressure by using Eq. (2.3), the maximum difference is 13.0%. The two methods of calculation yield almost identical results at lower pressures. Slight variations, however, occur at higher pressures.

Although the percent differences obtained for Cp are greater than were hoped for, they are not significantly greater than those in other investigations. Literature values (Schmidt, 1969; Jůza and Mareš, 1980; Woolf, 1980) vary as much as 10% in the higher temperature ranges. Even in the ASME Steam Tables (Meyer et al., 1967) calculated by using the 1967 IFC Formulation for Industrial Use there were local variations of approximately 5% near the boundaries between the subregions. To smooth the tables, graphical spline fits were used in these regions.

2-12 Temperature of Superheated Steam as a Function of Pressure and Enthalpy

In the preceding sections pressure and temperature were used as the independent variables. Additional equations were required for the simulation in which enthalpy was one of the independent variables. The following equation gives additional flexibility to the equations for the superheat region in sections 2-4, 2-6, 2-7, and 2-11. The form of the equation is from Agrawal (1978), but the parameters were determined for the region of interest in this work. The temperature is given by

$$T = a_1 + a_2P + a_3P^2 + a_4h + a_5hP + a_6hP^2 + a_7h^2 + a_8h^2P + a_9h^2P^2 \qquad (2.13)$$

where

$a_1 = -1.1584\text{x}10^3$

$a_2 = 3.5880\text{x}10^1$

$a_3 = -1.2666\text{x}10^{-1}$

$a_4 = 6.3660\text{x}10^{-1}$

$a_5 = -1.9700\text{x}10^{-2}$

$a_6 = 7.1515\text{x}10^{-5}$

$a_7 = -2.3814\text{x}10^{-5}$

$a_8 = 2.7366\text{x}10^{-6}$

$a_9 = -1.0107\text{x}10^{-8}$.

The maximum difference between tabulated (Schmidt, 1969) and calculated values is 1.7%. The differences are generally less than 0.5%.

2-13 Volume of Vapor as a Function of Pressure and Enthalpy

The following equation was adopted from the equation for density (Agrawal, 1978) and the coefficients determined by curve fitting in the region of interest. The volume of the vapor is given by

$$v_g = b_1 + b_2P + \frac{b_3}{P} + b_4h + b_5Ph + \frac{b_6h}{P} \tag{2.14}$$

where

$b_1 = 6.8242$

$b_2 = -6.2259\text{x}10^{-2}$

$b_3 = -4.4119\text{x}10^{3}$

$b_4 = -2.2842\text{x}10^{-3}$

$b_5 = 1.9732\text{x}10^{-5}$

$b_6 = 2.2893$.

The maximum difference between tabulated (Schmidt, 1969; Bain, 1964) and calculated values is 4.3% at 145 bars and 2635.8 J/g. This point is very close to the saturation line and is on the 613.15 K isotherm which is one of the boundaries between subregions in the 1967 IFC Formulation (Rivkin, Aleksandrov, and Kremenevskaya, 1978). The other large differences are along this same isotherm as well as the 623.15 K isotherm. However, the tabulated values agree very closely to the skeletal table values (Rosner and Straub, 1980). The other differences are generally less than 1.0%.

To check the feasibility of using Eq. (2.13) to convert the input from pressure and enthalpy to pressure and temperature, Eq. (2.4) for the gas volume was used as a test. Again the largest variations between the tabulated (Schmidt, 1969) and calculated values are found along the 613.15 K and 623.15 K isotherms. The maximum difference is 5.5%. There is a suggestion that this same trend may be present in the differences in the volumes calculated in section 2-4, although it is far less pronounced. The same trend appears in the calculations of temperature in section 2-12. It would appear that the larger variations are transmitted through the enthalpy. However, the apparent anomaly is not observed between the calculated and tabulated values of enthalpy (section 2-5).

2-14 Temperature of Saturated Water as a Function of Enthalpy

The temperature is given by

$$T_{sat} = f_1 + f_2 h + f_3 h^2 \qquad (2.15)$$

where

$f_1 = 2.5678 \times 10^{2}$

$f_2 = 2.9060 \times 10^{-1}$

$f_3 = -4.1831 \times 10^{-5}$.

The differences between tabulated (Keenan et al., 1969) and calculated values are less than or equal to 0.5% over the enthalpy range corresponding to 373.15 K to 623.15 K on the saturation line.

CHAPTER 3

APPLICATION AND CONCLUSIONS

Two components of a digital simulator developed at the University of Arizona for a liquid metal fast breeder reactor power plant (Hetrick and Sowers, 1978; Hetrick, Shinaishin, and Wait, 1978; Shinaishin, 1976) were chosen as test cases for the empirical equations discussed in Chapter 2. A schematic diagram for the Clinch River Breeder Reactor Plant (CRBRP) is shown in Figure 1. The original tables for water properties in the evaporator and superheater were replaced by the appropriate formulations. Details of the original model may be found in Hetrick et al. (1978) and Shinaishin (1976). Some modifications of the original model as well as a user's manual may be found in Hetrick and Sowers (1978). A computer listing of the evaporator portion of the simulator, with the empirical equations for the thermodynamic properties of water incorporated, may be found in Appendix B. Appendix C contains a computer listing of the superheater portion of the simulator with the appropriate equations for water properties.

The DARE P (Differential Analyzer Replacement, Portable) continuous system simulation language developed at the University of Arizona by Professors Granino Korn and John Wait (Korn and Wait, 1978) was used on a CDC CYBER-175 for the simulations. The DARE P system allows the user a choice of integration rules. The Runge-Kutta-Merson rule was used in this work.

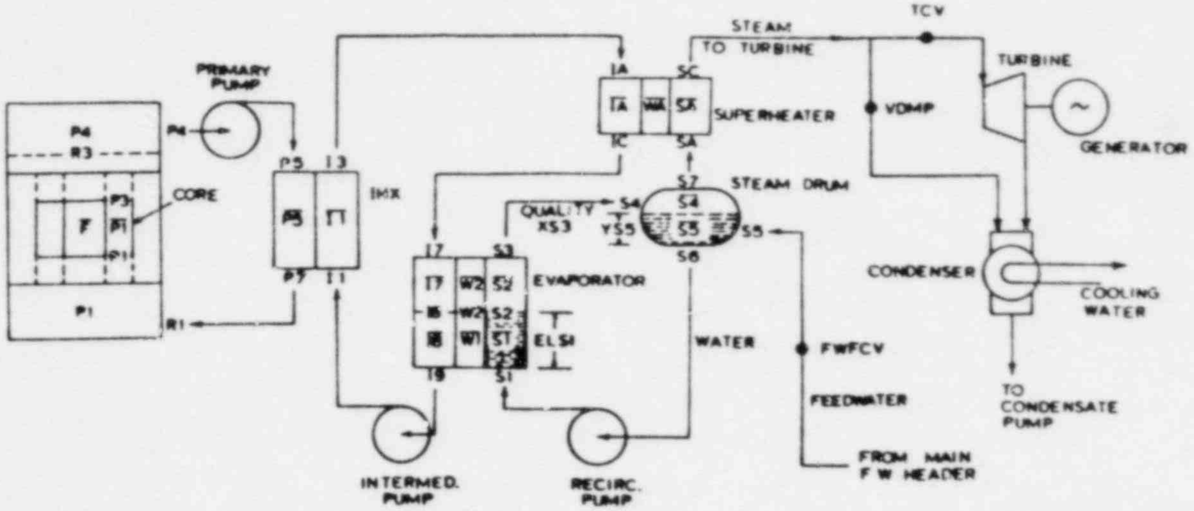

Figure 1. Schematic diagram of the CRBRP (Hetrick and Sowers, 1978).

The results for mild and moderately severe transients obtained by using the programs with the empirical equations were compared to solutions obtained with programs using tabular values. In both cases the transients are artifically introduced into isolated power plant components.

A mild transient for the evaporator was initiated by decreasing the enthalpy of the water being pumped from the steam drum through the recirculation pump to the evaporator. This decrease in enthalpy corresponds to pumping progressively cooler water into the water side of the evaporator, which is a sodium-water heat exchanger. The sodium temperature at the outlet and the saturation temperature of water in the evaporator decrease. As can be seen from the computer listing in Appendix D, the results for the programs using tabulated values and empirical formulas differ by less than 8 percent. Graphical representations of a portion of the data are shown in Figures 2 and 3.

A moderately severe transient was introduced in the superheater (another sodium-water heat exchanger) by decreasing the intermediate sodium flow rate according to the following expression

$$\mathrm{WI} = \mathrm{WI0}\big(1.0-0.002\exp(\mathrm{T})\big) \quad .$$

In some respects this can be considered as a pump coastdown. The water temperature and enthalpy decrease at the superheater exit. The results are shown in Appendix E.

There is a significant difference between the results for the steam specific heat in the two programs. This is not surprising since

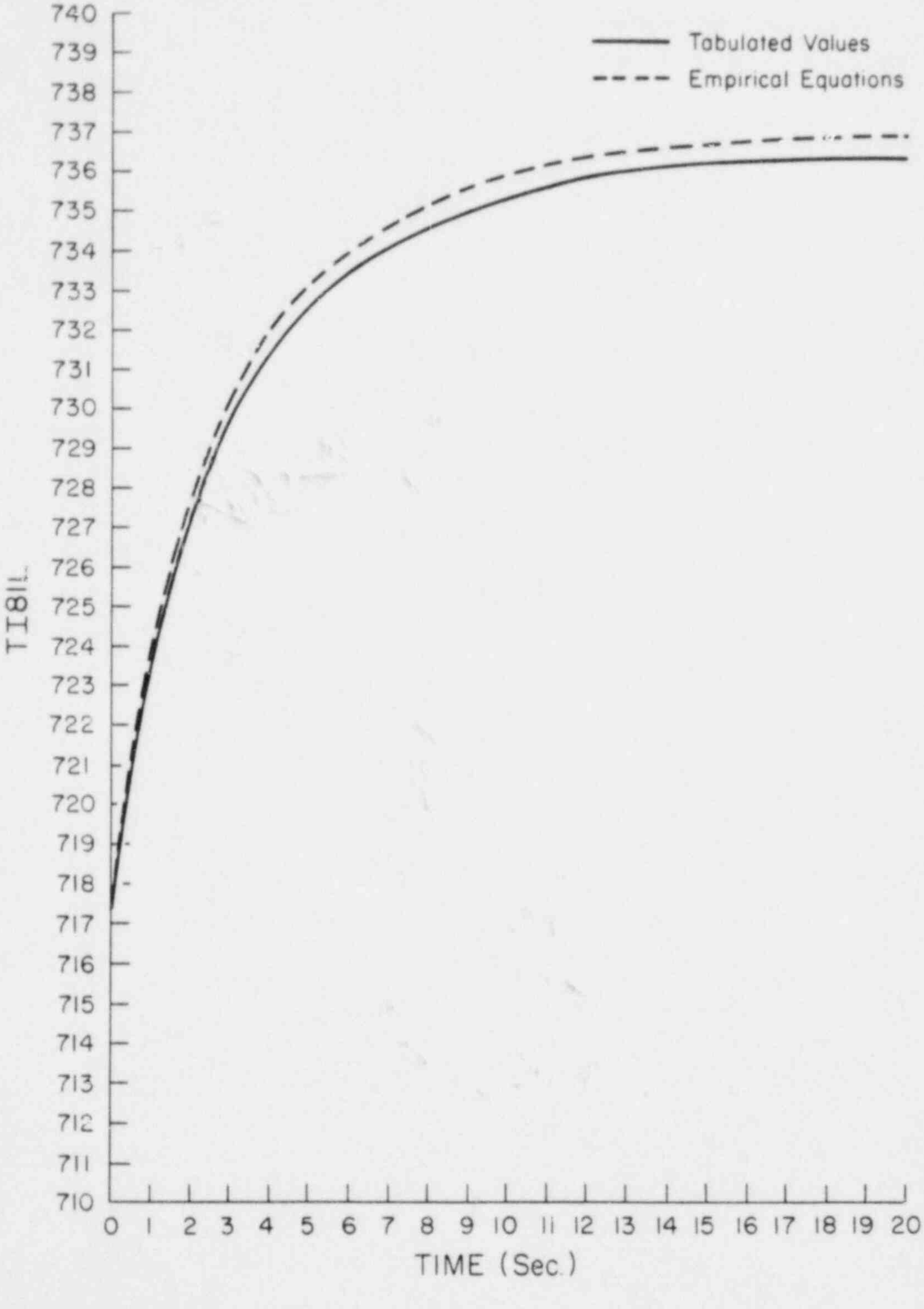

Figure 2. Sodium temperature (°F) inside evaporator following step decrease in recirculation water enthalpy.

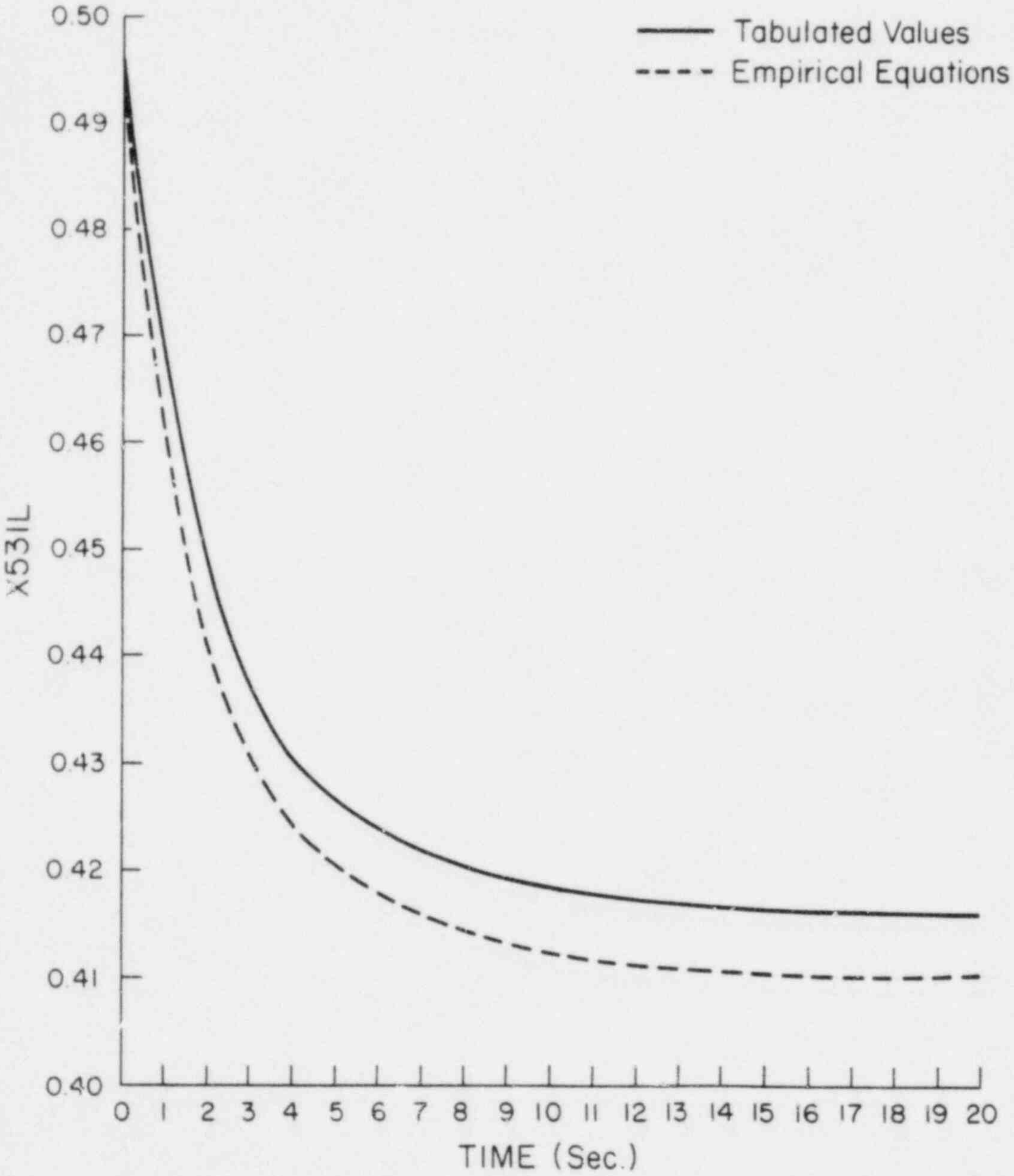

Figure 3. Evaporator outlet steam quality following step decrease in recirculation water enthalpy.

the representation for the specific heat of water, $\left(\frac{\partial h}{\partial T}\right)_P$, in the superheater program using tabular values is the finite difference $\frac{\Delta h}{\Delta T}$. The values for the specific heat obtained by $\frac{\Delta h}{\Delta T}$ may vary as much as 55 percent from the reference values. The values of Cp using the empirical form are much closer to the reference values (see section 2-11).

Figure 4 shows a graphical representation of the differences between the initial temperatures and temperatures at various times during the simulation.

Although there are potential advantages of smoothness arising from replacement of tabular interpolation by smooth empirical functions, the main advantage in using the empirical equations is computational economy. The compilation times are approximately equal, but the execution time for the evaporator program with the empirical representations is approximately 27 percent less than for the program with tables. The superheater program with the empirical formulas requires approximately 23 percent less execution time and approximately 35 percent less compilation time than the one with tables. These reductions in execution and compilation times represent significant savings in computational costs.

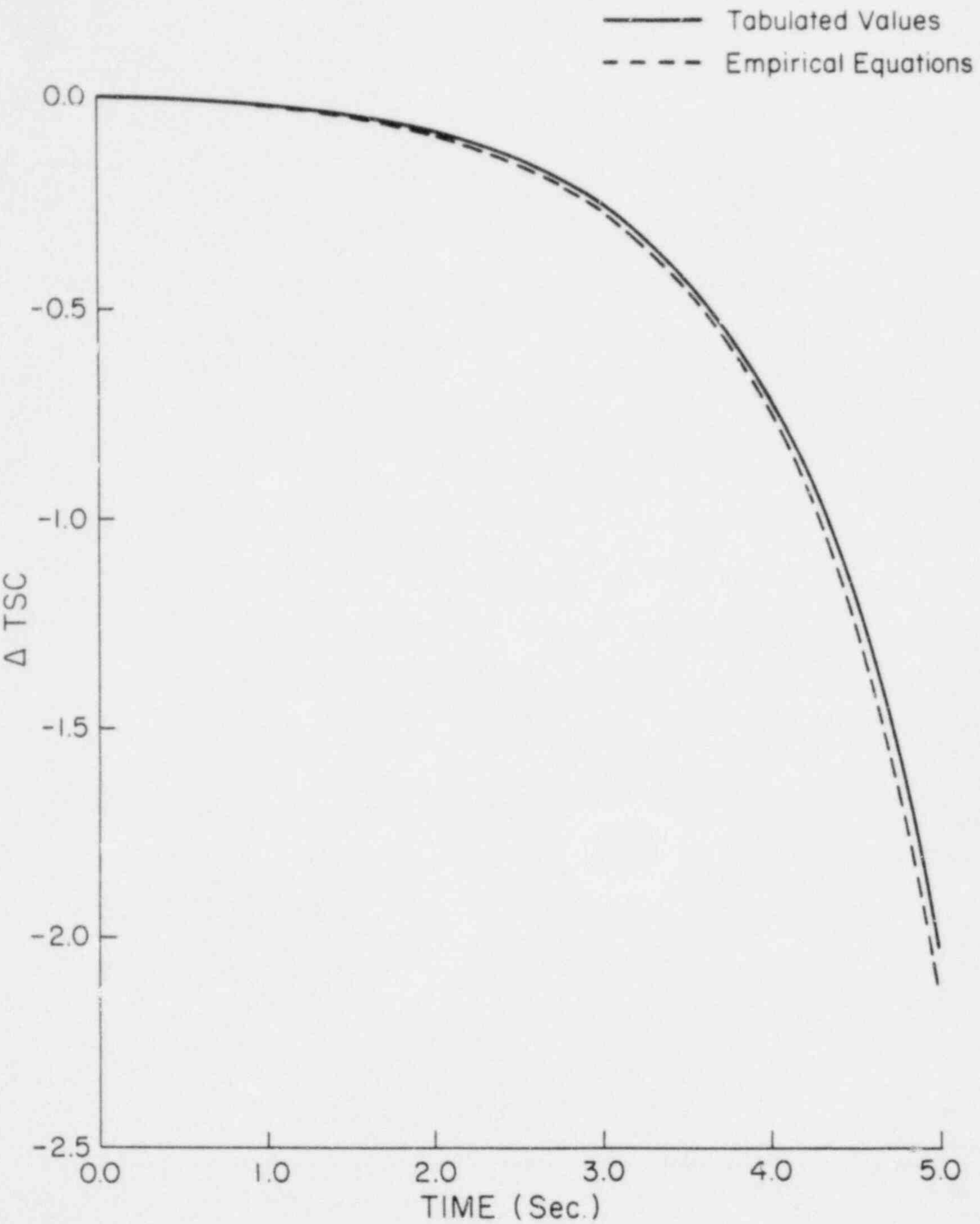

Figure 4. Change in superheater outlet steam temperature (°F) following simulated flow coastdown.

APPENDIX A

ADDITIONAL FORMULATIONS FOR WATER AND STEAM

This appendix contains additional formulations, ranges, and descriptions for thermodynamic properties of water and steam. Unless otherwise noted, the range of applicability is the same as in Chapter 2 and the differences between tabulated and calculated values are again given as relative percent differences. The units are the same as in Chapter 2.

A-1 Compressibility

Five additional pressure and temperature terms were added to the equation in section 2-1. The compressibility is given by

$$Z = z_o + \frac{z_1 P}{T} + z_2 P + \frac{z_3 P^2}{T^2} + \frac{z_4 P^2}{T} + z_5 P^2 + \frac{z_6 P^3}{T} + z_7 P^3$$

$$+ z_8 P^4 + \frac{z_9 P^4}{T} + \frac{z_{10} P^3}{T^3} + \frac{z_{11} P^3}{T^2} + \frac{z_{12} P^4}{T^4} + \frac{z_{13} P^4}{T^3} + \frac{z_{14} P^4}{T^2}$$

where

$z_o = 9.9681 \times 10^{-1}$

$z_1 = -4.9457$

$z_2 = 6.2050 \times 10^{-3}$

$z_3 = -1.6439 \times 10^{2}$

$z_4 = 5.0804 \times 10^{-1}$

$z_5 = -3.9945 \times 10^{-4}$

$z_6 = -1.3509\text{x}10^{-2}$

$z_7 = 7.9215\text{x}10^{-6}$

$z_8 = -5.0734\text{x}10^{-8}$

$z_9 = 1.2179\text{x}10^{-4}$

$z_{10} = -1.3791\text{x}10^{3}$

$z_{11} = 7.5838$

$z_{12} = -9.3142\text{x}10^{3}$

$z_{13} = 5.1827\text{x}10^{1}$

$z_{14} = -1.1555\text{x}10^{-1}$

The differences between the calculated and reference values are less than or equal to 1.4%.

A-2 Heat Capacity at Zero Pressure

The zero pressure heat capacity may be represented by

$$Cp^\circ = c_1 + c_2T + c_3T^2 + c_4T^3$$

where

$c_1 = 1.8232$

$c_2 = -1.4191\text{x}10^{-4}$

$c_3 = 1.0104\text{x}10^{-6}$

$c_4 = -4.0744\text{x}10^{-10}$

In the range 373.15 K to 773.15 K, this commonly-used cubic equation gives results which differ by less than 0.005% from the reference (Schmidt, 1969) values.

A simple two-term equation may also be used to represent the zero pressure heat capacity. In this case

$$Cp^\circ = c_o + c_{o1}T$$

where

$c_o = 1.6772$

$c_{o1} = 5.6899\text{x}10^{-4}$

For this form, the maximum difference between calculated and reference (Schmidt, 1969) values is 0.9%.

A-3 Temperature as a Function of Pressure on the Saturation Line

The following expression is slightly more complex than the one in section 2-3. The saturation temperature is given by

$$T_{sat} = x_1 + x_2P + x_3P^2 + x_4P^3 + x_5P^4 + \frac{x_6}{P} + \frac{x_7}{P^2} + \frac{x_8}{P^3} + \frac{x_9}{P^4} + x_{10}P^{\frac{1}{2}} + \frac{x_{11}}{P^{\frac{1}{2}}} \qquad (A.1)$$

where

$x_1 = 3.5674\text{x}10^2$

$x_2 = -2.4699$

$x_3 = 1.1148\text{x}10^{-2}$

$x_4 = -4.3795\text{x}10^{-5}$

$x_5 = 7.0705\text{x}10^{-8}$

$x_6 = 2.8328$

$x_7 = -2.2207\text{x}10^{-2}$

$x_8 = 1.3199\text{x}10^{-4}$

$x_9 = -3.1097\text{x}10^{-7}$

$x_{10} = 4.0220\text{x}10^{1}$

$x_{11} = -2.4315\text{x}10^{1}$.

The maximum difference between tabulated (Bain, 1964) and calculated values is at 274.15 K and is equal to 0.4%. From 277.15 K to the critical point the differences are less than or equal to 0.2%.

A-4 Steam Volume and Derivatives

The methods of deriving the following equations are identical to the ones used in section 2-4. The volume of the steam is given by

$$v = R\left(\frac{z_0T}{P} + z_1 + z_2T + \frac{z_3P}{T} + z_4P + z_5PT + z_6P^2 + z_7P^2T + z_8P^3T + z_9P^3 + \frac{z_{10}P^2}{T^2} + \frac{z_{11}P^2}{T} + \frac{z_{12}P^3}{T^3} + \frac{z_{13}P^3}{T^2} + \frac{z_{14}P^3}{T}\right) .$$

The constants z_o to z_{14} are given in section A-1.

The differences between the tabulated (Schmidt, 1969) and calculated volumes in the superheat region are less than or equal to 1.4%. The saturation line data were checked by entering a saturation pressure, calculating the corresponding saturation temperature, and then calculating the gas volume. In addition to the temperature and volume equation combination discussed in section 2-4, the three other combinations for the volume equations (section 2-4 and this section) and temperature equations (sections 2-3 and A-3) were checked against tabulated values (Schmidt, 1969). The four combinations give almost identical results between calculated and tabulated values.

The $\left[\frac{dv}{dP}\right]_{sat}$ was obtained by using the temperature equation in section 2-3 and the partial derivatives of the volume in this section. The maximum difference between the calculated $\left[\frac{dv}{dP}\right]_{sat}$ values and those obtained by the finite difference method is 2.8%.

A-5 Enthalpy of Steam

The methods of derivation and the reference values are identical to those of section 2-5. The constants z_o to z_{14} are given in section A-1. The steam enthalpy is given by

$$
\begin{aligned}
h = {} & c_1(T-T_1) + \frac{c_2}{2}(T^2-T_1{}^2) + \frac{c_3}{3}(T^3-T_1{}^3) \\
& + R_1\Bigg[z_1(P-P_1) + z_3\left(\frac{P^2}{T} - \frac{P_1{}^2}{T_1}\right) + \frac{z_4}{2}(P^2-P_1{}^2) \\
& + \frac{z_6}{3}(P^3-P_1{}^3) + \frac{z_9}{4}(P^4-P_1{}^4) + z_{10}\left(\frac{P^3}{T^2} - \frac{P_1{}^3}{T_1{}^2}\right) \\
& + \frac{2}{3}z_{11}\left(\frac{P^3}{T} - \frac{P_1{}^3}{T_1}\right) + z_{12}\left(\frac{P^4}{T^3} - \frac{P_1{}^4}{T_1{}^3}\right) + \frac{3}{4}z_{13}\left(\frac{P^4}{T^2} - \frac{P_1{}^4}{T_1{}^2}\right) \\
& + \frac{z_{14}}{2}\left(\frac{P^4}{T} - \frac{P_1{}^4}{T_1}\right)\Bigg] + h_1 \quad .
\end{aligned}
$$

In the superheat region the differences between tabulated (Schmidt, 1969) and calculated values are less than or equal to 1.6%. On the saturation line, the differences are again less than or equal to 1.6% whether the temperature and pressure are given, or the saturation temperature is calculated by using Eq. (2.3) or (A.1).

The results for $\left[\frac{dH}{dP}\right]_{sat}$ are similar to the ones discussed in section 2-5.

A hybrid form for enthalpy was derived by replacing Cp in the preceding derivation by the independently determined Cp in section 2-11. The resulting equation is

$$\begin{aligned}
h = {} & c_1(T-T_1) + \frac{c_2}{2}(T^2-T_1^2) + \frac{c_3}{3}(T^3-T_1^3) + \left[-\frac{1}{A} + \frac{1}{B}\right][y_1P_1^3] \\
& + \left[-\frac{y_{11}P_1}{A} + \ell nA + \frac{y_{11}P_1}{B} - \ell nB\right][y_2P_1 + y_5P_1^2 + y_8P_1^3] \\
& + \left[T + 2y_{11}P_1\,\ell nA - \frac{y_{11}^2P_1^2}{A} - T_1 - 2y_{11}P_1\ell nB + \frac{y_{11}^2P_1^2}{B}\right] \\
& [y_3P_1 + y_6P_1^2 + y_9P_1^3] + \left[\frac{T^3}{2A} + \frac{3}{2}y_{11}P_1T + 3y_{11}^2P_1^2\ell nA\right. \\
& - \frac{3}{2}\frac{y_{11}^3P_1^3}{A} - \frac{T_1^3}{2B} - \frac{3}{2}y_{11}P_1T_1 - 3y_{11}^2P_1^2\ell nB \\
& \left. + \frac{3}{2}\frac{y_{11}^3P_1^3}{B}\right][y_4P_1 + y_7P_1^2 + y_{10}P_1^3] + R_1\left[z_1(P-P_1)\right. \\
& + \frac{z_3}{T}(P^2-P_1^2) + \frac{z_4}{2}(P^2-P_1^2) + \frac{z_6}{3}(P^3-P_1^3) + \frac{z_9}{4}(P^4-P_1^4) \\
& + \frac{z_{10}}{T^2}(P^3-P_1^3) + \frac{2}{3}\frac{z_{11}}{T}(P^3-P_1^3) + \frac{z_{12}}{T^3}(P^4-P_1^4) \\
& \left. + \frac{3}{4}\frac{z_{13}}{T^2}(P^4-P_1^4) + \frac{1}{2}\frac{z_{14}}{T}(P^4-P_1^4)\right] + h_1
\end{aligned}$$

where

$$A = (T - y_{11}P_1)$$

$$B = (T_1 - y_{11}P_1)$$

and

c_1 to c_3 are given in section 2-2;

y_1 to y_{11} are given in section 2-11.

The results are almost identical to those discussed earlier in this section.

When the Cp in section 2-11 is used to replace the Cp in the derivation in section 2-5, the results are again almost identical to the simpler form derived from the compressibility alone.

A-6 Entropy of Steam

The reference values and the methods of derivation are identical to those of section 2-6. The constants z_o to z_{14} are given in section A-1. The entropy is given by

$$s = c_1 \ln\left(\frac{T}{T_1}\right) + c_2(T-T_1) + \frac{c_3}{2}(T^2-T_1^2) - R_1\Big[z_o \ln\left(\frac{P}{P_1}\right)$$

$$+ z_2(P-P_1) - \frac{z_3}{2}\left(\frac{P^2}{T^2} - \frac{P_1^2}{T_1^2}\right) + \frac{z_5}{2}(P^2-P_1^2) + \frac{z_7}{3}(P^3-P_1^3)$$

$$+ \frac{z_8}{4}(P^4-P_1^4) - \frac{2}{3} z_{10}\left(\frac{P^3}{T^3} - \frac{P_1^3}{T_1^3}\right) - \frac{z_{11}}{3}\left(\frac{P^3}{T^2} - \frac{P_1^3}{T_1^2}\right)$$

$$- \frac{3}{4} z_{12} \left(\frac{P^4}{T^4} - \frac{P_1^4}{T_1^4} \right) - \frac{1}{2} z_{13} \left(\frac{P^4}{T^3} - \frac{P_1^4}{T_1^3} \right)$$

$$- \frac{z_{14}}{4} \left(\frac{P^4}{T^2} - \frac{P_1^4}{T_1^2} \right) \Bigg] + s_1$$

In the superheat region the differences between tabulated (Schmidt, 1969) and calculated values are less than or equal to 1.5%. The results are the same on the saturation line and are not affected by which equation is used for the saturation temperature.

The following combined form was obtained by replacing the Cp in the above derivation by Eq. (2.12). The entropy is given by

$$s = c_1 \ln \left(\frac{T}{T_1} \right) + c_2(T-T_1) + \frac{c_3}{2} (T^2-T_1^2) + \Bigg[- \frac{1}{y_{11}P_1A} - \frac{1}{y_{11}^2P_1^2}$$

$$\ln A + \frac{1}{y_{11}^2P_1^2} \ln T + \frac{1}{y_{11}P_1B} + \frac{1}{y_{11}^2} \ln B - \frac{1}{y_{11}^2P_1^2} \ln T_1 \Bigg]$$

$$[y_1P_1^3] + \left[- \frac{1}{A} + \frac{1}{B} \right] [y_2P_1 + y_5P_1^2 + y_8P_1^3] + \Bigg[- \frac{y_{11}P_1}{A}$$

$$+ \ln A + \frac{y_{11}P_1}{B} - \ln B \Bigg] [y_3P_1 + y_6P_1^2 + y_9P_1^3]$$

$$+ \Bigg[T + 2y_{11}P_1 \ln A - \frac{y_{11}^2P_1^2}{A} - T_1 - 2y_{11}P_1 \ln B$$

$$+ \frac{y_{11}^2P_1^2}{B} \Bigg] [y_4P_1 + y_7P_1^2 + y_{10}P_1^3] - R_1 \Bigg[z_o \ln \left(\frac{P}{P_1} \right)$$

$$+ z_2(P-P_1) - \frac{z_3}{2T^2} (P^2-P_1^2) + \frac{z_5}{2} (P^2-P_1^2) + \frac{z_7}{3} (P^3-P_1^3)$$

$$+ \frac{z_8}{4} (P^4-P_1^4) - \frac{2}{3} \frac{z_{10}}{T^3} (P^3-P_1^3) - \frac{z_{11}}{3T^2} (P^3-P_1^3)$$

$$- \frac{3}{4} \frac{z_{12}}{T^4} (P^4-P_1^4) - \frac{1}{2} \frac{z_{13}}{T^3} (P^4-P_1^4) - \frac{z_{14}}{4T^2} (P^4-P_1^4) \Bigg] + s_1$$

where

$A = (T - y_{11}P_1)$

$B = (T_1 - y_{11}P_1)$

and

c_1 to c_3 are given in section 2-2;

y_1 to y_{11} are given in section 2-11.

The results are identical to those discussed earlier in this section.

When the Cp in section 2-6 is replaced by Eq. (2.12), the results are almost identical to those obtained from the equation derived from the compressibility alone.

A-7 Specific Volume of the Saturated Fluid

A simpler expression than the one in section 2-8 gives better results at higher pressures, but is not as good at lower pressures. The volume of the saturated fluid is given by

$$v_f = v_1 + v_2P + v_3P^2 + v_4P^3$$

where

$v_1 = 1.0650$

$v_2 = 6.0201 \times 10^{-3}$

$v_3 = -3.7424 \times 10^{-5}$

$v_4 = 1.5809 \times 10^{-7}$

The maximum difference between tabulated (Bain, 1964) and calculated values occurs at 1 bar and equals 2.7%. Between 2 bars and 206 bars, the differences are less than or equal to 1.5%.

A-8 Saturation Temperature as a Function of Enthalpy

A cubic expression for saturation temperature as a function of enthalpy yields slightly better results than those obtained in section 2-14. The saturation temperature is represented by

$$T_{sat} = f_1 + f_2h + f_3h^2 + f_4h^3$$

where

$f_1 = 2.7861\text{x}10^{2}$

$f_2 = 2.1508\text{x}10^{-1}$

$f_3 = 3.7188\text{x}10^{-5}$

$f_4 = -2.5489\text{x}10^{-8}$

The differences in the saturation enthalpy range corresponding to 373.15 K to 623.15 K are less than or equal to 0.1%.

APPENDIX B

COMPUTER LISTING FOR THE EVAPORATOR

A glossary of variable names and units may be found in Hetrick and Sowers (1978). Since English units are used in the simulator, the appropriate conversion factors were applied to the equations discussed in Chapter 2.

```
CALL DAREP,*NENEW.
^Z
 $M1
 $D1
*
          HS11I=HS61
        RTIC1L=RHONA(TIC1L)
        TTI71L=703800.0*RTIC1L*ENEV1L/W11L
*
        PROCED  TICI=TIC1L
        IF(T.EQ.0.0)  TICI=TIC1L
        ENDPRO
*
        TI71L=PDELAY(TIC1L,TTI71L,19,TICI)
*
* THE LEFT EVAPORATOR IN THE FIRST LOOP FOLLOWS
*
       PROCED DLS11L,DPS21L,TS61L,TS11L,W11L,W31L,W11LS,W31LS,
      $XS31L,A11L,B11L,C11L,DW11LD,DW31LD,DTW11L,DTW21L
      $,HDS11L,HDS21L
      $,TAS11L,HDW11L,HDW21L,TSAT1L,DETE1L,EL21L,DTI81L,DTI91L
$=HS61,ELS61,ELS62,ELS63,ELS64,HS11I,PA21L,PA41L,ENE1L,ENEV1L,W11LD,
       $W31LD,HFE1L,W11L,EL11L,TW21L,AS11,ELS31,ELS32,YS51L,DOD1L,
       $DPRC1L,FRC1L,F30B1L,TI71L,TI81L,TI91L,TIC1L,TW11L,NDELY1
       CALL EVAP( DLS11L,DPS21L,TS61L,TS11L,W11L,W31L,W11LS,W31LS,
      $XS31L,A11L,B11L,C11L,DW11LD,DW31LD,DTW11L,DTW21L
      $,HDS11L,HDS21L
      $,TAS11L,HDW11L,HDW21L,TSAT1L,DETE1L,EL21L,DTI81L,DTI91L
$,HS61,ELS61,ELS62,ELS63,ELS64,HS11I,PA21L,PA41L,ENE1L,ENEV1L,W11LD,
       $W31LD,HFE1L,W11L,EL11L,TW21L,AS11,ELS31,ELS32,YS51L,DOD1L,
       $DPRC1L,FRC1L,F30B1L,TI71L,TI81L,TI91L,TIC1L,TW11L,NDELY1)
        ENDPRO
*
        EL11L.=DLS11L/3600.0
*
        PA21L.=DPS21L/3600.0
****************************************
*         III-2.  WALL
*****************************************
        TW11L.=DTW11L/3600.0
        TW21L.=DTW21L/3600.0
*****************************
*         III-3.  INTERMEDIATE  SODIUM
*****************************
*
        TI81L.=DTI81L/3600.0
*
        TI91L.=DTI91L/3600.0
       W11LD.=DW11LD
       W31LD.=DW31LD
          STORE,EL11L,PA21L,TW11L,TI81L,TI91L,W11LD,
       $  DLS11L,DPS21L,TS61L,XS31L,DETE1L,TAS1L,TSAT1L,
       $  DTI81L,DTI91L,DTW11L,DTW21L,DW11LD,DW31LD,DT
 $P
 $T1
        RHONA , 6
*       NEXT  IS  THE  INDEPENDENT  VARIABLE  FOLLOWED  BY  THE  DEPENDENT
           392.0      ,56.4
           572.0      ,55.1
           752.0      ,53.6
           932.0      ,52.1
```

```
          1112.0     ,50.5
          1292.0     ,48.9
 $T9
      RD , 3 , 5
*
* THE  FOLLOWING  IS  A  DEFINITION  OF  THE  SECOND
* INDEPENDENT  VARIABLE.
*
                     0.2        ,0.4        ,0.6        ,0.8        ,1.0
*
* NEXT  COMES  THE  FIRST  INDEPENDENT  VARIABLE
* FOLLOWED  BY  THE  DEPENDENT  VARIABLE.
*
          1000.0     ,4.81801  ,7.43707  ,10.6082  ,12.915   ,14.4123
          2000.0     ,2.48938  ,3.21363  ,4.30274  ,4.97874  ,5.35559
          2500.0     ,1.72844  ,2.23131  ,2.6778   ,2.98751  ,2.98751
 $F
       SUBROUTINE TSATF(PAS2,TSAT)
       COMMON/TEMP/TK
       PB=PAS2*0.0689475729
       TK=3.5653E+02+PB*(-2.0611+PB*(5.7064E-03-1.011E-05*PB))
     $ +((2.1841+(-9.9865E-03+2.5304E-05/PB)/PB)/PB)
     $ +3.8845E+01*PB**0.5-2.242E+01/PB**0.5
       TSAT=1.8*TK-459.67
       RETURN
       END
       SUBROUTINE VG(PAS2,VGS2)
       COMMON/TEMP/TK
       PB=PAS2*0.0689475729
       VGAS=4.619*(1.0001*TK/PB-5.3391+6.1322E-03*TK
     $ +PB*(2.5547E-01-2.058E-04*TK-7.4961E+01/TK
     $ +PB*(-8.901E-03+4.3505E-06*TK+6.1847/TK-1.4715E+03/TK**2)))
       VGS2=VGAS*.01601846
       RETURN
       END
       SUBROUTINE HG(PAS2,HGS2)
       COMMON/TEMP/TK
       PB=PAS2*0.0689475729
       HGAS=TK*(1.7524+TK*(1.2468E-04+1.0326E-07*TK))
     $ +0.4619*PB*(-5.3391+PB*(-7.4961E+01/TK+1.27735E-01
     $ +PB*(-2.967E-03+4.12313333/TK-1.4715E+03/TK**2)))
     $ +2.011240092E+03
       HGS2=HGAS*0.429922614
       RETURN
       END
 $F
      SUBROUTINE EVAP( DLS1,DPAS2,TS6,TS1,WS1,WS3,WS1S,WS3S,
     $XS3,A,B,C,DWS1D,DWS3D,DTAW1,DTAW2
     $,HDS1,HDS2
     $,TAS1,HDW1,HDW2,TSAT,DETERM,ELS2,DTI8,DTI9,
     $HS6,ELS61,ELS62,ELS63,ELS64,HS11,PAS2,PAS4,ENE,ENEVAP,WS1D,
     $WS3D,HFE,W1,ELS1,TAW2,AS1,ELS31,ELS32,YS5,DDD,
     $DPRCSF,FRC,F30B,TI7,TI8,TI9,TIC,TAW1,NDELAY)
*    INLET WATER ENTHALPY
        WIL=SATAM(W1,1.0E-03,1.0E 30)
        FS6=8.421922E-3
      ENL=1.0
        TS6=2.534+HS6*(1.216684-4.073711E-04*HS6)
        RHOTS6=62.4279606/(5.59422E-01+TS6*(3.921435E-03
     $  +TS6*(-1.024599E-05+1.048748E-08*TS6)))
      WS1S=ENE*WS1D
```

```
      WS1SL=SATAM(WS1S,1.0E-03,1.0E 30)
      TAUHS1=(1.76744*ELS61+1.576246*ELS62+1.22716*ELS63
     $   +0.441786*ENEVAP*ELS64/ENL)*(ENL*RHOTS6/WS1SL)
       HS1=PDELAY(HS6,TAUHS1,NDELAY,HS1I)
       DHS1=3600.0*(HS6-HS1)/TAUHS1
*
*
       HS2=2.772141E+02+PAS2*(5.576054E-01+PAS2*(-5.633399E-04
     $  +PAS2*(3.939188E-07+PAS2*(-1.405836E-10+1.972743E-14*PAS2
     $  +((-3.046291E+03+2.203981E+04/PAS2)/PAS2)
      HAS1=(HS1+HS2)/2.0
       TS1=2.534+HS1*(1.216684-4.073711E-04*HS1)
       CALL TSATF(PAS2,TSAT)
      TAS1=(TS1+TSAT)/2.0
      HFS2=HS2
       CALL HG(PAS2,HGS2)
       VFS2=0.0169187+PAS2*(8.04558E-06+PAS2*(-5.964295E-09
     $  +PAS2*(3.066501E-12-5.080531E-16*PAS2)))
      RHOFS2=1./VFS2
       CALL VG(PAS2,VGS2)
      RHOGS2=1./VGS2
*
*   INLET WATER FLOW
       WS1=3600.0*ENE*WS1D
*
       WS3=3600.0*ENE*WS3D
*
*
      WS3S=ENE*WS3D
      WS2AD=(WS1D+WS3D)/2.0
*
*          HEAT  TRANSFER  COEFFICIENTS  IN  THE  EVAPORATOR
*
      RES1=225.31059*(WS1/ENE)
      RES2=359.10685*(WS1/ENE)
      HWE=12.817068*(WS1/ENE)**0.8
      HDS1=3.1415926*ENE/(0.0097486015+29.484029/HFE+29.484029/HWE)
      HDS2=3.1415926*ENE/(0.0097486015+29.484029/HFE)
*
      PEE=0.36373052*(WI/ENE)
      HNE=1281.316768+3.4231943*(WI/ENE)**0.8
      HDW1=3.1415926*ENE/(0.0097486015+19.2/HNE)
      HDW2=HDW1
*
      AS1=0.00090347512*ENE
*
*        STEAM PROPERTIES
*
      ELS2=46.0-ELS1
*
      SAI=RHOGS2/RHOFS2
      XS3=HDS2*ELS2*(TAW2-TSAT)/(WS3*(HGS2-HS2))
      ALF3=XS3/(SAI+(1.-SAI)*XS3)
      ALFSI=ALF3*(1.-SAI)
      ALFA2=ALF3*(0.5+ALFSI/6.0+ALFSI*ALFSI/12.)
*
        RHOAS1=62.4279606/(5.594222E-01+TAS1*(3.921435E-03
     $  +TAS1*(-1.024599E-05+1.048748E-08*TAS1)))
      RHOAS2=ALFA2*RHOGS2+(1.-ALFA2)*RHOFS2
      RHOHS2=ALFA2*RHOGS2*HGS2+(1.-ALFA2)*RHOFS2*HFS2
```

```
*
      HS3=HGS2*XS3+HFS2*(1.-XS3)
      RHOS3=ALF3*RHOGS2+(1.-ALF3)*RHOFS2
*
      DRFP2=-(1./VFS2**2)*DVF(PAS2)
      DRGP2=-(1./VGS2**2)*DVG(PAS2)
      DHFP2=DHF(PAS2)
      DHGP2=DHG(PAS2)
      DRHFP2=RHOFS2*DHFP2+HFS2*DRFP2
      DRHGP2=RHOGS2*DHGP2+HGS2*DRGP2
*
      A=ELS2*(ALFA2*(DRHGP2-DRGP2*HFS2)+(1.-ALFA2)*(DRHFP2-DRFP2*HFS2))
      B=ELS1*RHOGS2*DHFP2*(HGS2-HFS2)/(2.0*A*(HS2-HS1))
      C=1.0+2.*B*(RHOAS1-RHOFS2)/(RHOFS2-RHOGS2)
*
*         III-1.  WATER / STEAM
*
      DLS1=(2./C)*(-HDS1*ELS1*(TAW1-TAS1)/(AS1*RHOAS1*(HS2-HS1))
     $   +WS1/(AS1*RHOAS1)+(ELS1/2.0)*DHS1/(HS2-HS1)
     $   +B*(WS1-WS3)/(AS1*(RHOFS2-RHOGS2)))
      DLS1S=DLS1/3600.0
*
      WS2=WS1-AS1*RHOAS1*DLS1
*
      DPAS2=(RHOGS2*(HGS2-HFS2)/(A*(RHOFS2-RHOGS2)))*(-(RHOAS1-RHOFS2)
     $   *DLS1+(WS1-WS3)/AS1)
      DPAS2S=DPAS2/3600.0
*
*         III-2.  WALL
*
      WMHCD=0.073034229*ENE
      TW2=(TAW1*ELS1+TAW2*ELS2)/46.0
      TAI7=(TI7+TI8)/2.0
      TAI8=(TI8+TI9)/2.0
*
      DTAW1=(1./WMHCD)*(-(HDW1+HDS1+WMHCD*DLS1/ELS1)*TAW1
     $   +HDS1*TAS1+HDW1*TAI8)+TW2*DLS1/ELS1
*
      DTAW2=(1./WMHCD)*(-(HDW2+HDS2-WMHCD*DLS1/ELS2)*TAW2
     $   +HDS2*TSAT+HDW2*TAI7)-TW2*DLS1/ELS2
*
*         III-3.  INTERMEDIATE  SODIUM
*
*
      RHOTIC=RHONA(TIC)
      TAUTI7=703800.0*RHOTIC*ENEVAP/WIL
*
*
      TIE=(TI7+TI9)/2
      RHOIE=RHONA(TIE)
      SMHCD=0.0020848433*RHOIE*ENE
      WI8=WI+0.00682079*ENE*RHOIE*DLS1
*
      DTI8=(1./(SMHCD*ELS2))*(-(HDW2*ELS2-SMHCD*DLS1)*TAI7
     $   +HDW2*ELS2*TAW2+0.30566168*(WI*TI7-WI8*TI8))
*
      DTI9=(1./(SMHCD*ELS1))*(-(HDW1*ELS1+SMHCD*DLS1)*TAI8
     $   +HDW1*ELS1*TAW1+0.30566168*(WI8*TI8-WI*TI9))
*
*
*         RECIRCULATION  LOOP  MOMENTUM  EQUATIONS
```

```
*
*
      R=RD(PAS2,XS3)
*
* IN THE NEXT THREE LINES SINCE THE FLOW TC EACH OF THE
*EVAPORATORS IS HALF THE TOTAL FLOW BEFORE THE PIPE SPLIT
* HAS BEEN DIVIDED BY TWO.HENCE THE DIFFERENCE FROM BRENDA.
      TAUWS3=ELS31/(11.42886*ENEVAP)+ELS32*2.0/(68.844553*ENL)
      TAUWS1=ELS61*2.0/(56.9*ENL)+ELS62*2.0/(50.755*ENL)
     $   +ELS6*2.0/(39.515*ENL)+ELS64/(14.22555*ENEVAP)
*
      A11T=ELS1/(32.2*AS1)+ELS2/(64.4*AS1)+TAUWS1
      A12T=ELS2/(64.4*AS1)+0.30020714/AS1+TAUWS3
      A21T=ELS2/(128.8*AS1)
      A22T=ELS2/(128.8*AS1)+0.15010357/AS1+TAUWS3
*
      DETERM=A11T*A22T-A12T*A21T
*
*          FRICTION  COEFFICIENTS
*
      IF(RES1.LT.1000.0)  FS1=8.0/RES1
      IF(RES1.GE.1000.0.AND.RES1.LT.100000.0)  FS1=0.05256/(RES1**0.243)
      IF(RES1.GE.100000.0)  FS1=0.02117/(RES1**0.164)
      FS1A=102426.1*FS1
*
      IF(RES2.LT.1000.0)  FS2=8.0/RES2
      IF(RES2.GE.1000.0.AND.RES2.LT.100000.0)  FS2=0.05256/(RES2**0.243)
      IF(RES2.GE.100000.0)  FS2=0.02117/(RES2**0.164)
      FS2A=112238.73*FS2
*
      DPRC=DPRCSF-FRC*(WS1S*2.0/ENL)**2
*
      HOSPH=ELS1*RHOAS1+ELS2*RHOAS2+10.208333*RHOS3-52.70833*RHOTS6
     $   -(YS5+(DOD-6.0)/2.0)*RHOTS6
      CHMPH=(1./(32.2*AS1*AS1))*(WS3S**2/RHOS3-WS2AD**2/RHOFS2)
      FRPH=(FS1A*ELS1+618.90+(ENE*2.0/ENL)**2*FS6)*(WS1D**2)
     $   +(FS2A*R*ELS2)*(WS2AD**2)+(63259.1+9.66667*FS2A*R)*(WS3D**2)
*
      HSPH34=ELS2*RHOAS2/2.0+7.6666665*RHOS3
      FRPH34=FS2A*R*ELS2*(WS2AD**2)/2.0+(4.833335*FS2A*R
     $   +F308)*(WS3D**2)
      B1T=-(HOSPH+CHMPH-144.0*DPRC+FRPH)
      B2T=144.0*(PAS2-PAS4)-HSPH34-CHMPH/2.0-FRPH34
*
      WS1DD=A22T*B1T/(ENE*DETERM)-A12T*B2T/(ENE*DETERM)
      WS3DD=A11T*B2T/(ENE*DETERM)-A21T*B1T/(ENE*DETERM)
*
      DWS1D=WS1DD
      DWS3D=WS3DD
      RETURN
      END
*
        FUNCTION DHF(PAS)
        DHF=5.578054E-01+PAS*(-1.12666E-03+PAS*(1.181756E-06
     $  +PAS*(-[illegible].623345E-10+9.863715E-14*PAS)))
     $  +((3.046291E+03-4.407962E+04/PAS)/PAS**2)
        RETURN
        END
        FUNCTION DVF(PAS)
        DVF=6.045582E-06+PAS*(-1.192359E-08+PAS*(9.259505E-12
     $  -2.032213E-15*PAS))
```

```
      RETURN
      END
      FUNCTION DVG(PAS)
      COMMON/TEMP/TK
      COMMON/DERIV/DTK
      PB=PAS*0.0689475729
      DTK=-2.0611+1.14128E-02*PB-3.033E-05*PB*PB-2.1841/PB**2
     $ +1.9977E-02/PB**3-7.5912E-05/PB**4+1.94225E+01/PB**0.5
     $ +1.121E+01/PB**1.5
      DVDP=4.619*(-1.0001*TK/PB**2-7.4961E+01/TK+2.5547E-01
     $ -2.058E-04*TK-1.7802E-02*PB+8.701E-06*PB*TK+1.23694E+01*PB/TK
     $ -2.943E+03*PB/TK**2)
      DVDT=4.619*(1.0001/PB+6.1322E-03+7.4961E+01*PB/TK**2
     $ -2.058E-04*PB+4.3505E-06*PB*PB-6.1847*PB*PB/TK**2
     $ +2.943E+03*PB*PB/TK**3)
      DV=DVDP+DVDT*DTK
      DVG=0.0011044342*DV
      RETURN
      END
      FUNCTION DHG(PAS)
      COMMON/TEMP/TK
      COMMON/DERIV/DTK
      PB=PAS*0.0689475729
      DHDP=0.4619*(-5.3391-1.49922E+02*PB/TK+2.5547E-01*PB
     $ -6.901E-03*PB*PB+1.23694E+01*PB*PB/TK-4.4145E+03*PB*PB/TK**2)
      DHDT=1.7524+2.4936E-04*TK+3.0978E-07*TK*TK-0.4619*(-7.4961E+01*
     $ PB*PB/TK**2+4.12313333*PB**3/TK**2-2.943E+03*PB**3/TK**3)
      DH=DHDP+DHDT*DTK
      DHG=DH*0.0296421207
      RETURN
      END
*
END
       TMAX=50.0          ,   DT=1.0E-5          ,   DTMIN=1.0E-8
       NPOINT=51         ,   DTMAX=0.03
       EMAX=1.0E-5        ,   SY(9)=2.0
       SY(8)=3.0          ,   SY(12)=1.0
       NDELY1=8
        HS61=537.34491*0.9   ,AS11=4.1/6.0,ELS64=30.0
       ELS61=15.0,ELS62=100.0,ELS63=75.0,ELS31=70.0,ELS32=10.0
       EL11L=21.97481122
       F3081L=18055.8227292
       PA21L=1854.0,PA41L=1804.6520099
       W11LD=0.40730956  ,   W31LD=0.40730956
       TW11L=652.0908241 ,  TW21L=725.379849991
       TI81L=717.381316   ,   TI91L=651.000
         HFE1L=2543.695070
      ENE1L=4542./6.0,WI1L=3.834E7/6.0
     TIC1L=856.0, ODD1L=7.0,  FRC1L=7.1000730E-5,DPRC1L=175.0
      ENEV1L=1.0
      YS51L=3.0
END
          LIST,TI81L,TI91L,XS31L,TSAT1L
END
```

APPENDIX C

COMPUTER LISTING FOR THE SUPERHEATER

The appropriate conversion factors were used to convert the units of the equations given in Chapter 2 to the English units used in the simulator. A list of the variable names and units for the superheater may be found in Hetrick and Sowers (1978).

```
CALL DAREP,*NENEW.
^Z
 $M1
 $D1
*         II.  SUPERHEATER
*
*    STEAM FLOW RATE
        WSA=3600.0*WSAS
*
*    INLET STEAM ENTHALPY
       WSASL=SATAM(WSAS,1.0E-03,1.0E 30)
       TAUHSA=0.92175*ELS7*ENL*RHOVS4/WSASL
       HSA=PDELAY(HS4,TAUHSA,7,HS4I)
*
*    SODIUM INLET TEMPERATURE
        TIA=PDELAY(TI3,TAUTIA,3,TI3I)
        DTIA=(TI3-TIA)/TAUTIA
*
*    INLET STEAM PRESSURE
       PSA=PSAD
*
*
       FSC1=7.888943E-4*ELSC1/(ELSC1+0.573425*ELSC2*(ENL/ENSC2)**2)
       FSC2=3.506214E-4*ELSC2/(ELSC2+1.74391*ELSC1*(ENSC2/ENL)**2)
*
       TAUWSA=2.339742E-4*ELS7/ENL+3.1533E-4*ELSHT/ENL
      $    +1.3682E-4*ELSC1/ENL+1.095337E-4*ELSC2/ENSC2
       DWSASD=(1./TAUWSA)*(PAS4-PS9+(53.21+YS5)*RHOVS4/144.0
      $-(2.8059027129E-3+FSC1)*(WSAS/ENL)**2-FSC2*(WSAS/ENSC2)**2)
*
       DWSAS=DWSASD
*
       WSAS.=DWSAS
*
       PSAD=PAS4+(1./144.)*(-0.03369228*ELS7*DWSAS/ENL+(53.21+YS5)*RHOVS4
      $    -0.08286330732*(WSAS/ENL)**2)
*
       PSC=PSA-2.229948861E-3*(WSAS/ENL)**2-3.1533E-4*ELSHT*DWSAS/ENL
*
*
       TAI1=(TI1+TI3)/2.0
*         TI3=936.0*(1.0-0.002*EXP(T))
          WI=WIO*(1.0-0.002*EXP(T))
*
       RHOTI3=RHONA(TI3)
*
       WIL=SATAM(WI,1.0E-03,1.0E 30)
       TAUTIA=8474400.0*RHOTI3*ENL/WIL
*
*
          PB=PSC*0.0689475729
          TK=(TSC+459.67)/1.8
          HG=TK*(1.7524+TK*(1.2468E-04+1.0326E-07*TK))
      $   +0.4619*PB*(-5.3391+PB*(-7.4961E+01/TK+1.27735E-01
      $   +PB*(-2.967E-03+4.12313333/TK-1.4715E+03/TK**2)))
      $   +2.011240092E+03
          HSC=HG*0.429922614
*
*
          HCPT=1.7524+TK*(2.4936E-04+3.0978E-07*TK)
      $   +(PB*(TK*(5.211E+01+TK*(-1.0089E-01+4.5463E-05*TK))
```

```
      $  +PB*(TK*(4.3768E-01+TK*(-1.5544E-03+1.2808E-06*TK))
      $  +PB*(TK*(-4.9183E-02+TK*(7.2632E-05-3.5782E-08*TK))
      $  +1.1120E+01)))))/(TK-2.9971*PB)**2
         CPS=0.238845897*HCPT
*
       HASA=(HSA+HSC)/2.0
       PSS=(PSA+PSC)/2.0
*
         VSA=0.109313-6.876095E-05*PSA-1.025008E+03/PSA
      $  +HSA*(-8.510687E-05+5.081824E-08*PSA+1.237127/PSA)
         VG=4.619*(1.0001*TK/PB-5.3391+6.1322E-03*TK
      $  +PB*(2.5547E-01-2.058E-04*TK-7.4961E+01/TK+
      $  PB*(-8.901E-03+4.3505E-06*TK+6.1847/TK
      $  -1.4715E+03/TK**2)))
         VSC=VG*0.01601846
       RHOASA=(1./VSA+1./VSC)/2.0
*
       TAIA=(TIA+TIC)/2.0
       ROTAIA=RHONA(TAIA)
*
*           HEAT  TRANSFER  COEFFICIENTS  IN  THE  SUPERHEATER
*
       PES=0.37604738*(WI/ENS)
       HNS=1222.0763+3.3530778*(WI/ENS)**0.8
       HWA=144.51326*ENS/(.0097486015+19.2/HNS)
*
       RES=577.54204*(WSA/ENS)
       HSS=4.7053166*(WSA/ENS)**0.8
       HSSA=144.51326*ENS/(0.017076813+29.484029/HSS)
*
*
*          II-1.  SUPERHEATED  STEAM
*
       DTSC=(0.0066838/(CPS*RHOASA*ENS))*(HSSA*(TAWA-TASA)
      $  -WSA*(HSC-HSA))
       TSC.=DTSC
*
*
*          II-2.  WALL
*
       DTAWA=(8.2682436E- 5/ENS)*(HWA*(TAIA-TAWA)-HSSA*(TAWA-TASA))
       TAWA.=DTAWA
*
*          II-3.  INTERMEDIATE  SODIUM
*
*
       DTIC=(2.93739E- 3/(ROTAIA*ENS))*(0.3060422401 8*WI*(TIA-TIC)
      $   -HWA*(TAIA-TAWA))
       TIC.=DTIC
         TSA=(-2.54479E+03+PSA*(4.45291-1.083802E-03*PSA))
      $  +HSA*(2.665317+PSA*(-5.686793E-03+1.423368E-06*PSA)
      $   +HSA*(-2.319126E-04+PSA*(1.8374783E-06-4.67899E-10*PSA)))
       TASA=(TSA+TSC)/2.
         RHOVS4=1.0/VS4
         VS4=0.109313-6.876095E-05*PAS4-1.025008E+03/PAS4
      $  +HS4*(-8.510687E-05+5.081824E-08*PAS4+1.237127/PAS4)
$T1
       RHONA , 6
*      NEXT  IS  THE  INDEPENDENT  VARIABLE  FOLLOWED  BY  THE  DEPENDENT
          392.0      ,56.4
          572.0      ,55.1
```

```
          752.0      ,53.6
          932.0      ,52.1
          1112.0     ,50.5
          1292.0     ,49.9
 $F
END
TMAX=5.0,NPOINT=51
WSAS=925.051,ELS7=120.0,ENL=3.0,HS4I=1147.9148,HS4=HS4I
TI3I=936.0,ELSC1=200.0,ELSC2=100.0,ENSC2=2.0
ELSHT=65.0,PS9=1464.69996,PAS4=1804.652,YS5=3.0,WIO=10650.*3600.
TI3=936.0,TSC=917.150,ENS=2271.0,TAWA=863.880
TIC=858.041,TI1=651.0
END
LIST,TSC,HSC,CPS,VSC
END
```

APPENDIX D

RESULTS FOR THE EVAPORATOR

All units are English units. A detailed discussion of the choices for the output as well as the list of variable names may be found in Hetrick and Sowers (1978). The variables in the sample output are summarized below.

1 = minor loop

L = left evaporator

TI81L, TI91L = liquid sodium temperatures (intermediate loop)

XS31L = the quality at node S3 (evaporator outlet)

TSAT1L = the saturation temperature in the evaporator

EVAPORATOR - TABULAR VALUES

```
    TIME          TI81L          TI91L          XS31L          TSAT1L
0.             7.17381E+02    6.51000E+02    4.91363E-01    6.24667E+02
1.00000E+00    7.23297E+02    6.51439E+02    4.62094E-01    6.24517E+02
2.00000E+00    7.27510E+02    6.49338E+02    4.41873E-01    6.24476E+02
3.00000E+00    7.30157E+02    6.46852E+02    4.30338E-01    6.24444E+02
4.00000E+00    7.31874E+02    6.44654E+02    4.24238E-01    6.24420E+02
5.00000E+00    7.33074E+02    6.42662E+02    4.20557E-01    6.24407E+02
6.00000E+00    7.33958E+02    6.41496E+02    4.18003E-01    6.24398E+02
7.00000E+00    7.34629E+02    6.40420E+02    4.16137E-01    6.24391E+02
8.00000E+00    7.35144E+02    6.39587E+02    4.14733E-01    6.24386E+02
9.00000E+00    7.35542E+02    6.38942E+02    4.13653E-01    6.24382E+02
1.00000E+01    7.35850E+02    6.38442E+02    4.12817E-01    6.24378E+02
1.10000E+01    7.36088E+02    6.38054E+02    4.12168E-01    6.24376E+02
1.20000E+01    7.36273E+02    6.37753E+02    4.11664E-01    6.24374E+02
1.30000E+01    7.36416E+02    6.37519E+02    4.11272E-01    6.24373E+02
1.40000E+01    7.36528E+02    6.37337E+02    4.10966E-01    6.24371E+02
1.50000E+01    7.36614E+02    6.37195E+02    4.10727E-01    6.24371E+02
1.60000E+01    7.36681E+02    6.37085E+02    4.10541E-01    6.24370E+02
1.70000E+01    7.36733E+02    6.36999E+02    4.10395E-01    6.24369E+02
1.80000E+01    7.36774E+02    6.36932E+02    4.10281E-01    6.24369E+02
1.90000E+01    7.36806E+02    6.36880E+02    4.10192E-01    6.24369E+02
2.00000E+01    7.36830E+02    6.36839E+02    4.10123E-01    6.24368E+02
2.10000E+01    7.36849E+02    6.36808E+02    4.10069E-01    6.24368E+02
2.20000E+01    7.36864E+02    6.36783E+02    4.10026E-01    6.24368E+02
2.30000E+01    7.36876E+02    6.36764E+02    4.09993E-01    6.24368E+02
2.40000E+01    7.36885E+02    6.36749E+02    4.09967E-01    6.24368E+02
2.50000E+01    7.36892E+02    6.36737E+02    4.09946E-01    6.24368E+02
2.60000E+01    7.36898E+02    6.36728E+02    4.09931E-01    6.24368E+02
2.70000E+01    7.36902E+02    6.36720E+02    4.09918E-01    6.24368E+02
2.80000E+01    7.36905E+02    6.36715E+02    4.09908E-01    6.24368E+02
2.90000E+01    7.36908E+02    6.36710E+02    4.09901E-01    6.24368E+02
3.00000E+01    7.36910E+02    6.36707E+02    4.09895E-01    6.24367E+02
3.10000E+01    7.36912E+02    6.36704E+02    4.09890E-01    6.24367E+02
3.20000E+01    7.36913E+02    6.36702E+02    4.09886E-01    6.24367E+02
3.30000E+01    7.36914E+02    6.36701E+02    4.09884E-01    6.24367E+02
3.40000E+01    7.36915E+02    6.36699E+02    4.09881E-01    6.24367E+02
3.50000E+01    7.36915E+02    6.36698E+02    4.09880E-01    6.24367E+02
3.60000E+01    7.36916E+02    6.36698E+02    4.09878E-01    6.24367E+02
3.70000E+01    7.36916E+02    6.36697E+02    4.09877E-01    6.24367E+02
3.80000E+01    7.36916E+02    6.36696E+02    4.09876E-01    6.24367E+02
3.90000E+01    7.36917E+02    6.36696E+02    4.09876E-01    6.24367E+02
4.00000E+01    7.36917E+02    6.36696E+02    4.09875E-01    6.24367E+02
4.10000E+01    7.36917E+02    6.36696E+02    4.09875E-01    6.24367E+02
4.20000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.30000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.40000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.50000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.60000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.70000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.80000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
4.90000E+01    7.36917E+02    6.36695E+02    4.09874E-01    6.24367E+02
5.00000E+01    7.36917E+02    6.36695E+02    4.09873E-01    6.24367E+02
```

EVAPORATOR - EMPIRICAL EQUATIONS

```
      TIME            T181L           T191L           XS31L           TSAT1L
0.               7.17381E+02     6.51000E+02     4.97253E-01     6.24260E+02
1.00000E+00      7.23107E+02     6.51401E+02     4.68090E-01     6.24120E+02
2.00000E+00      7.27168E+02     6.49315E+02     4.48272E-01     6.24070E+02
3.00000E+00      7.29741E+02     6.46863E+02     4.36811E-01     6.24042E+02
4.00000E+00      7.31419E+02     6.44693E+02     4.30434E-01     6.24021E+02
5.00000E+00      7.32594E+02     6.42937E+02     4.26600E-01     6.24006E+02
6.00000E+00      7.33463E+02     6.41560E+02     4.23992E-01     6.23996E+02
7.00000E+00      7.34126E+02     6.40490E+02     4.22074E-01     6.23989E+02
8.00000E+00      7.34636E+02     6.39661E+02     4.20620E-01     6.23984E+02
9.00000E+00      7.35031E+02     6.39017E+02     4.19506E-01     6.23979E+02
1.00000E+01      7.35338E+02     6.38518E+02     4.18647E-01     6.23976E+02
1.10000E+01      7.35576E+02     6.38130E+02     4.17985E-01     6.23974E+02
1.20000E+01      7.35760E+02     6.37628E+02     4.17464E-01     6.23972E+02
1.30000E+01      7.35903E+02     6.37594E+02     4.17062E-01     6.23970E+02
1.40000E+01      7.36015E+02     6.37412E+02     4.16749E-01     6.23969E+02
1.50000E+01      7.36101E+02     6.37271E+02     4.16505E-01     6.23968E+02
1.60000E+01      7.36169E+02     6.37160E+02     4.16316E-01     6.23967E+02
1.70000E+01      7.36221E+02     6.37074E+02     4.16168E-01     6.23967E+02
1.80000E+01      7.36262E+02     6.37008E+02     4.16052E-01     6.23966E+02
1.90000E+01      7.36293E+02     6.36956E+02     4.15962E-01     6.23966E+02
2.00000E+01      7.36318E+02     6.36915E+02     4.15891E-01     6.23966E+02
2.10000E+01      7.36337E+02     6.36883E+02     4.15837E-01     6.23966E+02
2.20000E+01      7.36352E+02     6.36859E+02     4.15794E-01     6.23965E+02
2.30000E+01      7.36364E+02     6.36839E+02     4.15760E-01     6.23965E+02
2.40000E+01      7.36373E+02     6.36824E+02     4.15734E-01     6.23965E+02
2.50000E+01      7.36380E+02     6.36813E+02     4.15714E-01     6.23965E+02
2.60000E+01      7.36385E+02     6.36804E+02     4.15698E-01     6.23965E+02
2.70000E+01      7.36390E+02     6.36797E+02     4.15685E-01     6.23965E+02
2.80000E+01      7.36393E+02     6.36791E+02     4.15675E-01     6.23965E+02
2.90000E+01      7.36396E+02     6.36787E+02     4.15668E-01     6.23965E+02
3.00000E+01      7.36398E+02     6.36783E+02     4.15662E-01     6.23965E+02
3.10000E+01      7.36399E+02     6.36781E+02     4.15657E-01     6.23965E+02
3.20000E+01      7.36400E+02     6.36779E+02     4.15654E-01     6.23965E+02
3.30000E+01      7.36401E+02     6.36777E+02     4.15651E-01     6.23965E+02
3.40000E+01      7.36402E+02     6.36776E+02     4.15649E-01     6.23965E+02
3.50000E+01      7.36403E+02     6.36775E+02     4.15647E-01     6.23965E+02
3.60000E+01      7.36403E+02     6.36774E+02     4.15645E-01     6.23965E+02
3.70000E+01      7.36404E+02     6.36773E+02     4.15644E-01     6.23965E+02
3.80000E+01      7.36404E+02     6.36773E+02     4.15644E-01     6.23965E+02
3.90000E+01      7.36404E+02     6.36772E+02     4.15643E-01     6.23965E+02
4.00000E+01      7.36404E+02     6.36772E+02     4.15642E-01     6.23965E+02
4.10000E+01      7.36404E+02     6.36772E+02     4.15642E-01     6.23965E+02
4.20000E+01      7.36405E+02     6.36772E+02     4.15642E-01     6.23965E+02
4.30000E+01      7.36405E+02     6.36772E+02     4.15641E-01     6.23965E+02
4.40000E+01      7.36405E+02     6.36772E+02     4.15641E-01     6.23965E+02
4.50000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
4.60000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
4.70000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
4.80000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
4.90000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
5.00000E+01      7.36405E+02     6.36771E+02     4.15641E-01     6.23965E+02
```

APPENDIX E

RESULTS FOR THE SUPERHEATER

A glossary of variable names and units may be found in Hetrick and Sowers (1978). The following variables are used in the sample output:

TSC = the temperature at node SC (superheater exit)
HSC = the enthalpy at node SC
CPS = the steam specific heat in region SA
VSC = the specific volume of superheated steam at node SC

An artificial transient was introduced by decreasing the flow rate. The driving equation is given by

$$WI = WI0\left(1.0-0.002\exp(T)\right)$$

SUPERHEATER - TABULAR VALUES

```
     TIME             TSC            HSC            CPS            VSC
0.               9.06529E+02    1.43165E+03    9.67883E-01    4.80425E-01
1.00000E-01      9.06529E+02    1.43165E+03    9.67883E-01    4.80424E-01
2.00000E-01      9.06528E+02    1.43165E+03    9.67884E-01    4.80424E-01
3.00000E-01      9.06527E+02    1.43165E+03    9.67886E-01    4.80423E-01
4.00000E-01      9.06525E+02    1.43164E+03    9.67887E-01    4.80423E-01
5.00000E-01      9.06523E+02    1.43164E+03    9.67890E-01    4.80422E-01
6.00000E-01      9.06521E+02    1.43164E+03    9.67892E-01    4.80421E-01
7.00000E-01      9.06519E+02    1.43164E+03    9.67895E-01    4.80419E-01
8.00000E-01      9.06516E+02    1.43164E+03    9.67899E-01    4.80418E-01
9.00000E-01      9.06513E+02    1.43164E+03    9.67902E-01    4.80417E-01
1.00000E+00      9.06509E+02    1.43163E+03    9.67907E-01    4.80415E-01
1.10000E+00      9.06505E+02    1.43163E+03    9.67911E-01    4.80413E-01
1.20000E+00      9.06501E+02    1.43163E+03    9.67917E-01    4.80411E-01
1.30000E+00      9.06496E+02    1.43163E+03    9.67922E-01    4.80409E-01
1.40000E+00      9.06491E+02    1.43162E+03    9.67929E-01    4.80406E-01
1.50000E+00      9.06485E+02    1.43162E+03    9.67936E-01    4.80403E-01
1.60000E+00      9.06479E+02    1.43162E+03    9.67943E-01    4.80400E-01
1.70000E+00      9.06472E+02    1.43161E+03    9.67952E-01    4.80397E-01
1.80000E+00      9.06464E+02    1.43161E+03    9.67961E-01    4.80393E-01
1.90000E+00      9.06456E+02    1.43160E+03    9.67972E-01    4.80389E-01
2.00000E+00      9.06446E+02    1.43160E+03    9.67983E-01    4.80384E-01
2.10000E+00      9.06436E+02    1.43159E+03    9.67996E-01    4.80379E-01
2.20000E+00      9.06424E+02    1.43158E+03    9.68010E-01    4.80373E-01
2.30000E+00      9.06412E+02    1.43157E+03    9.68025E-01    4.80367E-01
2.40000E+00      9.06398E+02    1.43157E+03    9.68042E-01    4.80360E-01
2.50000E+00      9.06382E+02    1.43156E+03    9.68061E-01    4.80353E-01
2.60000E+00      9.06365E+02    1.43155E+03    9.68081E-01    4.80345E-01
2.70000E+00      9.06347E+02    1.43154E+03    9.68104E-01    4.80335E-01
2.80000E+00      9.06326E+02    1.43152E+03    9.68129E-01    4.80325E-01
2.90000E+00      9.06303E+02    1.43151E+03    9.68157E-01    4.80314E-01
3.00000E+00      9.06278E+02    1.43149E+03    9.68188E-01    4.80302E-01
3.10000E+00      9.06250E+02    1.43148E+03    9.68222E-01    4.80288E-01
3.20000E+00      9.06219E+02    1.43146E+03    9.68260E-01    4.80273E-01
3.30000E+00      9.06185E+02    1.43144E+03    9.68301E-01    4.80256E-01
3.40000E+00      9.06147E+02    1.43141E+03    9.68347E-01    4.80238E-01
3.50000E+00      9.06105E+02    1.43139E+03    9.68398E-01    4.80217E-01
3.60000E+00      9.06058E+02    1.43136E+03    9.68455E-01    4.80194E-01
3.70000E+00      9.06007E+02    1.43133E+03    9.68517E-01    4.80169E-01
3.80000E+00      9.05950E+02    1.43129E+03    9.68587E-01    4.80141E-01
3.90000E+00      9.05887E+02    1.43125E+03    9.68663E-01    4.80111E-01
4.00000E+00      9.05817E+02    1.43121E+03    9.68749E-01    4.80076E-01
4.10000E+00      9.05739E+02    1.43116E+03    9.68843E-01    4.80038E-01
4.20000E+00      9.05653E+02    1.43111E+03    9.68948E-01    4.79996E-01
4.30000E+00      9.05558E+02    1.43105E+03    9.69065E-01    4.79950E-01
4.40000E+00      9.05452E+02    1.43099E+03    9.69195E-01    4.79898E-01
4.50000E+00      9.05334E+02    1.43091E+03    9.69339E-01    4.79840E-01
4.60000E+00      9.05202E+02    1.43083E+03    9.69500E-01    4.79776E-01
4.70000E+00      9.05056E+02    1.43075E+03    9.69679E-01    4.79704E-01
4.80000E+00      9.04893E+02    1.43065E+03    9.69879E-01    4.79625E-01
4.90000E+00      9.04711E+02    1.43053E+03    9.70102E-01    4.79536E-01
5.00000E+00      9.0[illegible]8E+02    1.43041E+03    9.70352E-01    4.79436E-01
```

SUPERHEATER - EMPIRICAL EQUATIONS

```
    TIME            TSC            HSC            CPS            VSC
0.             9.17150E+02    1.42439E+03    6.10831E-01    4.83280E-01
1.00000E-01    9.17150E+02    1.42439E+03    6.10831E-01    4.83280E-01
2.00000E-01    9.17148E+02    1.42439E+03    6.10831E-01    4.83280E-01
3.00000E-01    9.17147E+02    1.42439E+03    6.10831E-01    4.83279E-01
4.00000E-01    9.17145E+02    1.42439E+03    6.10831E-01    4.83278E-01
5.00000E-01    9.17143E+02    1.42439E+03    6.10831E-01    4.83277E-01
6.00000E-01    9.17141E+02    1.42439E+03    6.10832E-01    4.83276E-01
7.00000E-01    9.17138E+02    1.42438E+03    6.10832E-01    4.83275E-01
8.00000E-01    9.17135E+02    1.42438E+03    6.10832E-01    4.83274E-01
9.00000E-01    9.17132E+02    1.42438E+03    6.10833E-01    4.83272E-01
1.00000E+00    9.17129E+02    1.42438E+03    6.10833E-01    4.83270E-01
1.10000E+00    9.17125E+02    1.42438E+03    6.10834E-01    4.83269E-01
1.20000E+00    9.17120E+02    1.42437E+03    6.10834E-01    4.83267E-01
1.30000E+00    9.17115E+02    1.42437E+03    6.10835E-01    4.83264E-01
1.40000E+00    9.17110E+02    1.42437E+03    6.10836E-01    4.83262E-01
1.50000E+00    9.17104E+02    1.42436E+03    6.10837E-01    4.83259E-01
1.60000E+00    9.17097E+02    1.42436E+03    6.10838E-01    4.83256E-01
1.70000E+00    9.17090E+02    1.42436E+03    6.10839E-01    4.83253E-01
1.80000E+00    9.17082E+02    1.42435E+03    6.10840E-01    4.83249E-01
1.90000E+00    9.17073E+02    1.42434E+03    6.10841E-01    4.83245E-01
2.00000E+00    9.17063E+02    1.42434E+03    6.10842E-01    4.83240E-01
2.10000E+00    9.17053E+02    1.42433E+03    6.10844E-01    4.83235E-01
2.20000E+00    9.17041E+02    1.42433E+03    6.10845E-01    4.83230E-01
2.30000E+00    9.17028E+02    1.42432E+03    6.10847E-01    4.83223E-01
2.40000E+00    9.17013E+02    1.42431E+03    6.10849E-01    4.83217E-01
2.50000E+00    9.16997E+02    1.42430E+03    6.10851E-01    4.83209E-01
2.60000E+00    9.16980E+02    1.42429E+03    6.10853E-01    4.83201E-01
2.70000E+00    9.16960E+02    1.42428E+03    6.10856E-01    4.83192E-01
2.80000E+00    9.16938E+02    1.42426E+03    6.10859E-01    4.83182E-01
2.90000E+00    9.16915E+02    1.42425E+03    6.10862E-01    4.83171E-01
3.00000E+00    9.16888E+02    1.42423E+03    6.10866E-01    4.83159E-01
3.10000E+00    9.16859E+02    1.42422E+03    6.10870E-01    4.83145E-01
3.20000E+00    9.16827E+02    1.42420E+03    6.10874E-01    4.83130E-01
3.30000E+00    9.16792E+02    1.42417E+03    6.10879E-01    4.83114E-01
3.40000E+00    9.16752E+02    1.42415E+03    6.10884E-01    4.83095E-01
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3.60000E+00    9.16660E+02    1.42410E+03    6.10897E-01    4.83053E-01
3.70000E+00    9.16607E+02    1.42406E+03    6.10904E-01    4.83028E-01
3.80000E+00    9.16548E+02    1.42403E+03    6.10912E-01    4.83000E-01
3.90000E+00    9.16482E+02    1.42399E+03    6.10921E-01    4.82970E-01
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5.00000E+00    9.15045E+02    1.42312E+03    6.11123E-01    4.82301E-01
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16. ABSTRACT *(200 words or less)* Steam property evaluations may represent a significant portion of the computing time necessary for power system simulations. The iterative nature of the solutions for heat transfer and kinetic equations often requires thousands of steam property evaluations during the execution of a single program. Considerable savings may be realized by simplification of property evaluations.

Empirical equations have been obtained for the thermodynamic properties of water in the region of interest. To maintain thermodynamic consistency, the compressibility factor Z, in terms of pressure and temperature, was obtained by curve fitting, and the enthalpy, entropy, and internal energy were derived by standard relationships. Formulations for heat capacity, saturation temperature as a function of saturation pressure, the specific volume of saturated water as a function of saturation pressure, and specific volume of saturated water as a function of the saturation temperature were determined by curve fitting of independent equations. Derivatives were obtained by differentiation of the appropriate formulations.

Evaporator and superheater components of a liquid metal fast breeder reactor power plant simulator were chosen as test cases for the empirical representations. Results obtained using the empirical equations were comparable to those obtained using tabular values and required 24% less computing time.

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