ANALYSIS OF MASS TRANSFER PROCESSES IN GEOTHERMAL POWER
CYCLES UTILIZING DIRECT CONTACT HEAT EXCHANGE

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REPORT OF WORK

September 21, 1978 to September 30, 1979

AC03-78ET27145

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Prepared for
Department of Energy
Division of Geothermal Energy

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ABSTRACT

A computer program was developed which calculates the isobutane content of the spent brine and the liquid-vapor distribution of carbon dioxide and hydrogen sulfide throughout the components of a geothermal power plant using direct contact heat exchange. The program model assumes separate boiler and preheater vessels, with the preheater being a spray tower. The condenser model is a horizontal tube surface condenser with condensation on the outside.

The program was written in Fortran language. The Fortran source deck consists of 976 cards. The program utilizes 320K for compilation and 72K for execution on an IBM 370/3031.

Sample cases were run which illustrate the effects of salt concentration in the brine and isobutane-to-brine ratio on isobutane and non-condensable gas content of the spent brine.

Introduction

A typical geothermal fluid at the well-head consists of a hot brine liquid phase, steam and noncondensable gases such as carbon dioxide and hydrogen sulfide. If this stream is mixed with a working fluid such as isobutane in a direct contact heat exchanger, the prediction of the compositions of the effluent streams from the heat exchanger is a challenging problem in non-isothermal mass transfer and gas-liquid reactions. Furthermore, water and non-condensable species will be carried with the working fluid through all other parts of the power cycle.

Geothermal fluids include noncondensable gases of varying amounts. Sometimes these noncondensibles make up a significant part of the vapor phase at the well head. For example, well flow tests by San Diego Gas and Electric near Niland gave the following flow rates at the well head:

liquid	324,000 lb/hr
steam	64,000
noncondensibles	12,000

Of the noncondensibles, CO_2 is often the main component with lower concentrations of H_2S , H_2 , CH_4 , N_2 , H_e , Ar, NH_3 and other gases. Analyses of noncondensable gases at several sites are:

% By Weight

Gas	Cerro Prieto	Geysers	Salton Sea	East Mesa
CO ₂	75.5	81.9	97.5	96.6
H ₂ S	19.0	5.6	0.2	0.05
CH ₄	-	4.9	0.3	2.9
NH ₃	-	4.9	2.0	0.4
H ₂	-	1.4	-	-
N ₂	-	1.3	-	-
He, Ar, Ne	0.95	-	-	-
Air	4.35	-	-	-

The presence of noncondensable gases generally seriously affects the performance of a power plant, regardless of the type of power cycle chosen. If the steam is used directly in a turbine, the gases will accompany the steam through the turbine into the condenser where they must be ejected to the atmosphere with a resulting penalty to the power output of the plant. In a binary power cycle with a conventional surface heat exchangers to heat the working fluid, the presence of the gases tends to diminish the heat transfer coefficient on the brine side.

If a direct contact heat exchanger is chosen for the transfer of heat from the geothermal fluid to the working fluid, the behavior of the noncondensable gases and their effects on the performance of the power cycle are much more difficult to predict. The following phenomena must be investigated: (Refer to Figure 1).

1. The geothermal fluid entering the direct contact heat exchanger is in contact with boiling working fluid. The working

Figure 1. Flowsheet of Power Cycle with Direct Contact Heat Exchange.

fluid vapor will strip some of the dissolved noncondensable gases from the geothermal brine. The equilibrium solubilities of the noncondensable gases in the brine may depend on the extents of several chemical reactions involving carbonate, bicarbonate, sulfide and hydrogen ion concentrations, as well as the boiler pressure and brine temperature distribution.

2. The gases passing through the turbine into the condenser will be a mixture of working fluid vapor, steam and noncondensable gases. As condensation begins two liquid phases will be produced: an aqueous liquid and an organic liquid. Noncondensable gases will dissolve in the liquids to an extent depending on condenser temperatures, pressure and mass transfer characteristics.

3. A gas-liquid separator follows the condenser so that noncondensibles may be vented from the system. Some working fluid vapor will go with the gases, and some gases will remain dissolved in the two liquid phases.

4. A liquid-liquid separator will recover the working fluid for recycle to the preheater and boiler. Some dissolved noncondensibles will remain in the working fluid.

5. In the preheater the working fluid will be contacted with brine. Some working fluid will dissolve in the spent brine and dissolved noncondensibles may be transferred between the liquids.

It is desirable that the noncondensibles remain with the brine phase so that the amount going to the condenser is small.

General Approach

The program requires the following information as input data:

1. for the geothermal brine entering the boiler -
 - temperature
 - flowrate
 - total dissolved solids
 - concentration of dissolved carbon dioxide and hydrogen sulfide
2. for the boiler - exit vapor temperature
 - degrees of superheat
3. for the condenser (horizontal - tube, surface type) -
 - tube diameter and length
 - numbers of vertical and horizontal rows
 - pitch - to - diameter ratio
 - condensing temperature
4. for the preheater (spray tower) -
 - tower diameter
 - drop size
 - increment height
5. for the cycle -
 - isobutane flow rate
6. temporary trial values
 - concentrations of carbon dioxide and hydrogen sulfide in isobutane leaving preheater

The program first carries out the boiler calculations using the temporary trial values of item 6 for the noncondensibles. The boiler

heat exchange is determined by the flow rate of isobutane to be vaporized plus the amount of superheat. The boiler subprogram finds the compositions of the boiler effluent streams.

The boiler effluent vapor stream is the feed stream to the condenser. The condenser subprogram finds the compositions of the condensate and vapor streams at the last row of tubes in the boiler. The vapor stream is assumed to be sent to an isobutane recovery operation, while the condensate is sent to the preheater.

The composition and temperature of the brine stream to the preheater were previously determined by the boiler subprogram. The composition of the isobutane stream to the preheater was determined by the condenser subprogram. The preheater subprogram finds the properties of the preheater effluent streams and then the trial values of item 6 are checked with the calculated ones. If the calculated concentrations of carbon dioxide and hydrogen sulfide are different from the assumed values by more than five percent the calculations for the entire cycle starting with the boiler subprogram are repeated with the new values. Iteration is continued until the five percent convergence is attained. A Flowchart is shown in Figure 2.

BOILER MODEL

The engineering of direct contact boiling, which involves the interaction of two liquid phases and a vapor phase, is very complex and is in a rather primitive state of development. In cases where superheating occurs, the following extremes in flow regions can occur. At the end of the exchanger where vaporization is just beginning and not much vapor is yet present, vapor bubbles are shed by the boiling organic and superheated by contact with brine as they rise. At the

Figure 2. Overall Flowchart of Computer Program

hot end of the exchanger, where boiling is nearly complete and the organic exists mostly as vapor, the system consists of many water drops and a few liquid organic drops in a large volume of vapor. Something less than complete axial mixing is required for the boiler to exhibit these axial variations.

All of the studies reported to date have carried out complete vaporization, so that heat transfer coefficients are averaged over all of the various axial states (where there is not complete axial mixing). The various studies reported in the literature exhibit various degrees of axial mixing and superheat of the effluent vapor.

Direct contact boiling studies with isobutane, which appears to be the favored working fluid for most situations, was reported by Suratt and Hart of DSS Engineers, Inc. (1976). The boiler was a cylindrical vessel six feet tall and six inches ID. Hot brine was fed near the top through a spray nozzle and exited from the bottom. Saturated isobutane liquid entered near the bottom through two tangential nozzles and rose counter-currently, leaving the surface of the brine near the midpoint of the boiler, and leaving as vapor through demister mesh at the top. Volumetric coefficients U_v ranged from about 8,000 to 17,000 Btu/hr ft³ °F. Mass flow ratios of isobutane-to-brine ranged from 0.24 to 1.17 but had little effect on values of U_v . Superheating of the effluent vapor by as much as 30°F occurred. Coefficients U_v were influenced most strongly by pressure, and were correlated by the following equation: (1)

$$U_v = 6500 \left[\frac{P_{\text{Sat}}}{P_{\text{Crit}}} \right]^{-0.56} \quad (1)$$

where P_{Sat} is the saturation pressure and P_{Crit} is the critical pressure of the working fluid. Equation (1) was tested over a range of $P_{\text{Sat}}/P_{\text{Crit}}$ of 0.34 to 0.60. On the basis that equation (1) was developed using the kind of working fluid and contactor that are likely to be used in the field, equation (1) was used in the boiler model.

Experimental measurements of mass transfer coefficients in direct contact boilers are not available. In the absence of data, mass transfer coefficients may be estimated using the Colburn analogy between heat and mass transfer:

$$\frac{k\bar{M}}{G} N^{2/3}_{\text{Sc}} = \frac{h}{C_p G} N^{2/3}_{\text{pr}} \quad (2)$$

The driving force associated with k is the mole fraction concentration difference.

The coefficients in equation (2) are individual phase coefficients, while the only experimental coefficient available is an overall coefficient, U_v , incorporating resistances to transfer in both phases as well as the transfer area per unit volume of mixed phases. In boiling it seems quite likely that the principal resistance is in the aqueous phase, since the vapor is formed in the working fluid phase and the resistance should be quite low there. Similarly, the principal resistance to mass transfer of working fluid into the aqueous phase lies in the aqueous phase, since there is no transfer resistance in the working fluid phase. Hence for mass transfer of working fluid into the aqueous phase we can take

$$k_x a = \frac{1}{C_p M} \cdot \left[\frac{N_{pr}}{N_{sc}} \right]^{2/3} U_v \quad (3)$$

where all properties are evaluated in the aqueous phase, and $k_x a$ is a volumetric mass transfer coefficient.

Mass transfer coefficients for the transfer of carbon dioxide and hydrogen sulfide were also estimated with equation (3), using appropriate Schmidt numbers. The argument that the brine-side mass transfer resistance is controlling for the gases depends on two factors: (1) the molecular diffusivities of the gases in isobutane are higher by a factor of about eight than those in brine, and the Schmidt numbers are lower by a factor of about thirty; thus $k_x a$ is much higher on the isobutane side, (2) the ratios of equilibrium solubilities of the gases in isobutane to those in the brine are of the order of fifty; thus the influence of the coefficient for the isobutane side on the overall coefficient is reduced by this ratio as shown in equation (4):

$$\frac{1}{K_b} = \frac{1}{\frac{H_b}{H_i} k_i} + \frac{1}{k_b} \quad (4)$$

Here the subscripts b and i denote the brine and isobutane phases and H_b/H_i is the equilibrium solubility ratio of gas in-isobutane to gas-in-brine.

The boiler model is based on the schematic shown in Figure 3. The superheating is assumed to take place in the vapor continuous region. Heat transfer coefficients in the superheating region are much lower than those in the boiling region. Volumetric coefficients ranged from

FIGURE 3. SCHEMATIC OF BOILER

0 to ~1080 averaging 160 Btu/hr ft³ °F in the DSS experiments at Wrightsville Beach. Since the superheating region does not strongly influence the mass transfer calculations, the average value is used by the model.

The calculations begin by dividing the boiler into a number of axial increments starting at the bottom of the boiler, with the superheating region being a single last increment. Each increment except the last accounts for an equal fraction of the isobutane vaporization. An overall heat balance gives the brine exit temperature from the boiler. Starting with the first increment at the bottom of the boiler, heat balances are made for each increment providing the brine temperature profile and log-mean temperatures differences. Volumetric heat transfer coefficients are obtained from equation (1) for all increments except the superheating one. The volume of each increment and total volume of the boiler are then obtained from equation (5):

for the j th increment

$$q_j = U_v \cdot V_j \cdot (\Delta t)_{lmj} \quad (5)$$

Calculations for the transfer of isobutane into the brine are started at the top of the boiler, where the concentration of isobutane in the brine is known to be zero.

At the average brine temperature for each increment a value of the volumetric mass transfer coefficient, $k_x a$, is calculated using the Colburn analogy between heat and mass transfer. The rate of mass transfer for an increment j is calculated by

$$R_j = (k_x a)_j V_j (x_j^* - \bar{x}_j) \quad (6)$$

A mass balance for increment j can be written:

$$G(x_{j+1} - x_j) = R_j \quad (7)$$

Defining $\bar{x}_j = (x_{j+1} + x_j)/2$ and combining Equations (6) and (7), an expression for x_j can be obtained.

$$x_j = \{x_{j+1} [G - (k_x a V/2)_j] + x_j^* (k_x a V)_j\} / [G + (k_x a V/2)_j] \quad (8)$$

Since the equilibrium solubility depends only on temperature for a given brine, trial-and-error calculations are not required for the isobutane. Values for the isobutane content of the brine for each increment are calculated from equation (8), starting at the top of the boiler and working down.

Calculations for the transport of the noncondensibles between phases are somewhat more complex. This calculation requires a trial and error procedure because the equilibrium solubilities depend on concentrations in the isobutane, which are unknown for the vapor effluent stream at the top of the tower, while the concentrations in the brine effluent are unknown at the bottom of the tower. In the algorithm chosen, the noncondensable concentrations in the effluent vapor are assumed; an overall material balance then gives the concentrations in the effluent brine and incremental calculations up the tower finally result in the effluent vapor concentrations, which are compared to the assumed values. Direct substitution and iteration continue until the agreement is within three percent for each gas. Calculations for each gas are done in sequence. The gas concentrations are low in all phases so that their interaction is negligible.

GAS SOLUBILITY AND GAS-LIQUID REACTIONS IN THE BRINE

Carbon dioxide, upon being dissolved in water or an aqueous solution containing carbonate or bicarbonate ion, might undergo chemical reactions in the liquid phase as follows:

The thermodynamic equilibrium condition for equation (9) may be expressed through a Henry's law constant:

$$H_{\text{CO}_2} = \frac{P_{\text{CO}_2}}{[\text{CO}_2]} \quad (12)$$

For reactions (10) and (11), the equilibrium condition is expressed through the reaction equilibrium constant:

$$K_{\text{CO}_2} = \frac{[\text{H}^+][\text{HCO}_3^-]}{[\text{CO}_2]} \quad (13)$$

$$K_{\text{HCO}_3^-} = \frac{[\text{H}^+][\text{CO}_3^{=}] }{[\text{HCO}_3^-]} \quad (14)$$

Analogous reactions and equilibrium relationships may be written for other soluble gases such as hydrogen sulfide and ammonia.

From equations (9) through (14) it is seen that the solubility of CO_2 may depend on the reaction kinetics and equilibrium values of the three reactions taking place. However, the solubility of CO_2 is affected by the presence of other constituents in the solution which provides any species common to these reactions. In the case of

geothermal brines, the behavior of the liquid phase reactions (10) and (11) is controlled by the presence of silicic acid. Thus Helgeson (1969) shows that for temperatures below about 200°C CO_2 (ag) is the dominant species of dissolved carbon dioxide below pH 6 to 7, HCO_3^- is dominant between pH 6 to 7 and 10 to 11, and at higher pH's $\text{CO}_3^{=}$ is dominant. But the dissociation of silicic acid

buffers the brine. At temperatures of 100 to 150°C, the dissociation constant for reaction (15) is in the range of 10^{-9} to $10^{-9.5}$. Hence for a brine which contains a significant amount of dissolved silica and which is undergoing a process which tends to drive the carbon dioxide out of solution (such as passing through the boiler), the pH will not rise very much because the dissociation of silicic acid will tend to maintain the pH at around 9. Thus the existence of significant concentrations of bicarbonate or carbonate ions is prevented. Similarly, the conversion of dissolved hydrogen sulfide to bisulfide ion is inhibited by the presence of silicic acid.

CONDENSER MODEL

The type of condenser chosen was a horizontal tube surface condenser. This design can accommodate large vapor flows with low pressure drop. The model assumes that the vapor moves down through the tube bundle without large-scale mixing currents in the direction of flow. The condenser is divided into increments in the direction of vapor flow, and flow rates and concentrations in both phases are adjusted for each increment. Mass transfer coefficients are calculated

for vapor and condensate phases for each increment and combined into an overall mass transfer coefficient.

Bernhardt et al. (1971) made photographic studies of the condensation of vapor mixtures of water plus an immiscible organic liquid. Among the organics tested were Freon 112, Freon 113, perchlorethylene and p-xylene. In every case the organic condensate formed a continuous film on the condensing surface, with water droplets as the discontinuous phase. Various correlations for the film coefficient were tested. The best turned out to be

$$h = h_1 V_1 + h_2 V_2 \quad (16)$$

where h_1 and h_2 are the calculated coefficients for pure water and pure organic from the Nusselt equation; and V_1 and V_2 are the volume fractions in the condensate.

In the systems of interest, the incoming vapor consists of only two or three percent water, a lesser amount of noncondensibles, and the rest isobutane or pentane. Hence the condensate film is an organic film with a few water drops floating in it, and the condensing film coefficient is essentially equal to that for pure organic.

Condensing coefficients are calculated from the Nusselt equation:

$$h = 0.951 \left(\frac{k_c^3 p^2 g}{\mu} \cdot \frac{LN^{2/3}}{\dot{m}_t} \right)^{1/3} \quad (17)$$

Vapor coefficients are calculated from the correlations of Pierson (1937) for crossflow of gases over tube banks. Mass transfer coefficients were estimated from the heat transfer coefficients through the Colburn analogy between heat and mass transfer.

For calculational purposes the tube bundle is divided into a number of increments in the direction of flow. For each increment a material balance, equilibrium and rate equations are combined to find the compositions of liquid and vapor leaving each increment (Figure 4). A material balance for one of the components for increment j can be written as:

$$x_j = (G_o y_o - G_j y_j) / L_j \quad (18)$$

The equilibrium relationship is written as

$$\bar{y}_j^* = k_{eq} (x_j + x_{j-1}) / 2 \quad (19)$$

The rate equations can be written as

$$R_j = K_y A [(y_j + y_{j-1}) / 2 - \bar{y}_j^*] \quad (20)$$

and

$$R_j = G_{j-1} y_{j-1} - G_j y_j \quad (21)$$

Equations (18) through (21) can be combined and solved for y_j as

$$y_j = \frac{(G_{j-1} y_{j-1} / K_y A) - (y_{j-1} / 2) + (K x_{j-1} / 2) + (K G_o y_o / 2 L_j)}{(G_j / K_y A) + 1/2 + (K G_j / 2 L_j)} \quad (22)$$

Equation (22) permits the compositions for the j -th increment to be calculated from those of the $j-1$ th increment. A particular equation is required for the first increment, since x_o does not exist. The equation for $j=1$ is

$$y_1 = \frac{y_o [(G_o / K_y A) - 1/2 + (K G_o / L_1)]}{1/2 + (K G_1 / L_1) + (G_1 / K_y A)} \quad (23)$$

Figure 4. INCREMENT SCHEME FOR CONDENSER

PREHEATER MODEL

The preheater is assumed to be a spray tower, in which the liquid isobutane from the condenser enters the bottom of the tower and passes through a nozzle which breaks it into droplets. The droplets of isobutane (discontinuous phase) rise upward through the brine (continuous phase) which is flowing downward. This countercurrent flow is possible because the liquid isobutane is less dense than the brine.

The preheater model is much more elaborate and precise than the boiler or condenser models because spray towers are much better understood. The hydrodynamics of drop flow has been studied much longer and more extensively than the flow in direct contact boilers, and is probably simpler. The extensive chemical engineering literature on solvent extraction in spray towers provides mass transfer correlations for various drop hydrodynamic conditions.

The computer calculations are divided into two phases. The first phase uses the Letan-Kehat (1968) model to predict a temperature profile in which the computer calculates the temperature of both brine and organic phases at each increment of height. When the desired dispersed phase temperature is reached, the computer begins to calculate down the column a mass transfer profile for each increment.

- (1) An energy balance around the column is made assuming no heat loss and the brine output temperature is found.
- (2) Computer calculates average physical property data and checks for flooding (Johnson and Minard, 1952).

The program calculates average hold up using Zen's correlation (Lapidus, Elgin, 1957).

$$V_s = \frac{U_c}{\phi} + \frac{U_D}{1-\phi} \quad (24)$$

The relative slip velocity can be taken as constant (Lapidus, Elgin, 1957).

$$V_s = \text{Constant} \quad (25)$$

We know that when the velocity of the continuous phase is zero

$$\frac{U_d}{1-\phi} = V_T \quad U_c = 0 \quad \text{and when} \quad \frac{U_c}{\phi} = V_T \quad U_D = 0$$

(Free Fall) (Transport)

Therefore, at any intermediate $V_c/\phi + V_d/(1-\phi)$ between zero and the transport velocity, the slip velocity of fluid past the particle is constant and equal to the free fall or transport velocity. The particle knows the movement of only the fluid and not the walls and does not know whether it is moving relative to the latter or not. (Lapidus, Elgin, 1957).

Therefore, we can calculate $V_s = V_t$.

- (3) The slip velocity is found according to the terminal drop velocity.

Klee and Treybal (1956) recommend two correlations for calculating terminal drop velocity in liquid-liquid systems. The two correlations estimate terminal drop velocity for two regions within the tower. A critical drop diameter determines the transition from Region I to Region II.

- (4) The critical drop diameter is obtained from the following:

$$d' = 0.33 \rho_c^{-0.14} \Delta \rho^{0.43} \mu_c^{0.30} \sigma^{0.24} \quad (26)$$

- (5) The best terminal velocity equation is chosen from critical drop diameter from the following:

$$V_{T(I)} = 38.3 \rho_c^{-0.45} \Delta \rho \mu_c^{-0.11} d_p^{0.70} \quad d_p < d' \quad (27)$$

$$V_{T(II)} = 17.6 \rho_c^{-0.55} \Delta\rho^{0.10} \mu_c^{0.18} \sigma \quad d_p > d' \quad (28)$$

- (6) v_T and the resulting holdup are calculated
- (7) The average interfacial area is computed (Lapidus and Elgin, 1957).

$$a = \frac{6(1-\phi)}{d_p} \quad (29)$$

- (8) Program calculates volume of each increment based on height and diameter.

Once preliminary calculations are complete, the computer initiates steps to calculate the column temperature profile using the Letan-Kehat (1968) model.

- (9) Program picks Letan-Kehat wake parameters m , M based on average hold up for column.
- (10) Program calculates temperature at top of zone A, the wake formation zone.
- (11) Program redefines variables and begins incrementing given height of column at bottom of zone C, the wake shedding zone

The Letan-Kehat (1968) indicates a region within the column in which the temperatures for both continuous and dispersed phases are constant (zone B).

- (12) Program iterates until temperature of organic meets preset output temperature.
- (13) Program stores temperature of continuous phase and dispersed phase at each increment.

Once the absolute error between calculated organic and desired

temperature meets a certain criteria, the mass transfer sequence begins.

Heertjes and deNie (1971) suggest mass transfer correlation for calculating Sherwood numbers for the continuous phase based on whether the drop is rigid, circulating, or oscillating. The model assumes drops less than 1mm in diameter are rigid, (Sideman, 1966). Drops 1mm to 4mm are assumed to be circulating. Drops greater than 4mm in diameter are oscillating.

Heertjes and deNie (1971) suggest the following mass transfer correlations:

$$N_{Sh} = 2 + 0.76 (N_{Re})^{1/2} (N_{Sc})^{1/3} \quad d_p < 1\text{mm} \quad (30)$$

$$N_{Sh} = 0.6 (N_{Re})^{1/2} (N_{Sc})^{1/2} \quad 1\text{mm} < d_p < 4\text{mm} \quad (31)$$

$$N_{Sh} = 50 + (.0085 - N_{Re} N_{Sc}^{0.7}) \quad d_p > 4\text{mm} \quad (32)$$

The effect of drop wake is included in both of the above correlations. Since H_2S and CO_2 are soluble in both brine and isobutane, internal mass transfer coefficients are needed to calculate an overall mass transfer coefficient. R. Von Berg (1971) suggests the following coefficients for three internal mass transfer mechanisms.

$$K_d = -\frac{r}{3t} \ln \left[1 - (1 - \exp(-\frac{\pi^2 D_{AB} t}{r^2}))^{1/2} \right] \quad (33)$$

For rigid drops

$$K_d = -\frac{r}{3t} \ln \left[1 - (1 - \exp(-\frac{\pi^2 2.25 D_{AB} t}{r^2}))^{1/2} \right] \quad (34)$$

For circulating drops

$$K_d = 0.00375 \mu \left(\frac{\mu_c}{\mu_c + \mu_d} \right) \quad (35)$$

For oscillating drops

(14) Program recalls the temperature at the top and bottom of each increment.

(15) The average liquid diffusivity for each increment is calculated.

The following (Wilke-Chang, 1955) correlation is used to determine liquid diffusivity:

$$D_{AB} = 7.4 (10^{-8}) \frac{(\psi_A^M)^{1/2} T}{\mu V_A^{0.6}} \quad (36)$$

(16) The average mass transfer coefficient is calculated for low mass transfer rates from the appropriate Sherwood number.

Isobutane is slightly soluble in brine. Henry's law relation holds for calculating the equilibrium concentration at the interface of each increment.

(17) The saturation pressure for isobutane vapor at temperatures for top and bottom of increment are found.

(18) The Henry's law constant at top and bottom of each increment is calculated.

(19) Program computes equilibrium concentration at top and bottom of each increment.

The program has now computed all preliminary mass transfer information.

A material balance around the increment is given by the following

(Johnson and Fleming, 1953):

$$R_j = U_c S (C_{z+1} - C_z) \quad (37)$$

The amount of isobutane transferred to the brine is given by the following relation. (Johnson and Fleming, 1953):

$$R_j = K_o aV (\Delta C)_{LM} \quad (38)$$

where

$$(\Delta C)_{LM} = \frac{(C_{z+1}^* - C_{z+1}) - (C_z^* - C_z)}{\ln \frac{(C_{z+1}^* - C_{z+1})}{(C_z^* - C_z)}} \quad (39)$$

- (20) Given the input concentration of isobutane in brine to the preheater the program solves (Eq.37) and (Eq.38) by trial and error for the isobutane concentration leaving the increment.
- (21) Program stores the results and recalls the temperature at top of next increment.
- (22) The program terminates when all temperature increments have been used.

Sample Cases

The boiler and preheater subprograms were checked out by simulating an experimental run made by DSS at Wrightsville Beach. This run, designated Test C-9-18, was made with separate boiler and preheater vessels and good heat balances were obtained. The run conditions were as follows:

dissolved solids in brine	- none
flow rate of brine	- 2139 lb/hr
flow rate of isobutane	- 1793 lb/hr
brine inlet temperature to boiler	- 336°F
isobutane saturation temperature	- 220°F
vapor exit temperature from boiler	- 237°F

After verifying the results of the subprograms, the entire computer program was run to investigate the rate of convergence for noncondensable gas concentrations as explained in Figure 2. The following additional input data were selected arbitrarily:

CO ₂ content of inlet brine	- 2×10^{-4} mole fraction
H ₂ S content of inlet brine	- 1×10^{-4} mole fraction
condenser tube diameter	- 1 inch
condenser tube length	- 28 ft
pitch-to-diameter ratio	- 3
number of condenser tubes	- 100
condensate saturation temperature	- 130°F

Temporary values for the CO₂ and H₂S content of the isobutane liquid stream leaving the preheater and entering the boiler were taken as 1×10^{-4} and 2×10^{-4} mole fraction, respectively. The convergence by direct iteration. as follows:

Concentrations in Saturated Isobutane, Mole Fraction

	<u>CO₂</u>	<u>H₂S</u>
initial value	0.000100	0.000200
iteration 1	0.000116	0.000074
iteration 2	0.000121	0.000065
iteration 3	0.000119	0.000061
iteration 4	0.000118	0.000059

Therefore, four or five iterations of the program appear to be sufficient, even when the initial guess is quite far off (as it was for H₂S in the example case).

The computer program was used to investigate the effects of dissolved solids concentration and isobutane-to-brine ratio on the isobutane and non-condensable gas concentrations in the spent brine. The effects of dissolved solids concentration are presented in Table 1. The temperature of the spent brine decreases by 6°F as the solids content increases from 0 to 10 percent. This effect is caused by the lower heat capacity of salt than water. Isobutane, carbon dioxide and hydrogen sulfide concentrations all decrease as the solids content increases because of their decreased solubilities.

The effects of flow ratio are much greater, as shown in Table 2. The brine flow was held at the original value, while higher and lower isobutane flows were investigated. The isobutane and gas concentrations were found to be surprisingly sensitive to flow ratio. Decreasing the flowrate of isobutane by 20 percent decreased the isobutane loss by a factor of $\sim^3$ and increases the rate of gas removal from the system by a factor of $\sim^{10}$ for carbon dioxide and $\sim^3$ for hydrogen sulfide. Of course, the power recovered from the brine also decreases by 20 percent.

The program is now available for the investigation of the many additional variables of interest.

TABLE 1. EFFECTS OF DISSOLVED SOLIDS CONCENTRATION

	WEIGHT % DISSOLVED SOLIDS				
	<u>0</u>	<u>2.5</u>	<u>5.0</u>	<u>7.5</u>	<u>10.0</u>
Temperature of spent brine, °F	201	198	197	196	195
Isobutane concentration in spent brine, ppm	99.2	102	99.5	93.2	90.4
CO ₂ concentration in spent brine, ppm	98.3	85.7	83.5	77.8	75.0
H ₂ S concentration in spent brine, ppm	83.2	78.7	77.2	72.5	72.4

TABLE 2. EFFECTS OF ISOBUTANE/BRINE FLOW RATIO

	<u>FLOWRATE OF ISOBUTANE, LBS/HOUR</u>		
	<u>1600</u>	<u>1793</u>	<u>2000</u>
Flow ratio, isobutane/ brine	0.741	0.830	0.926
Preheater height, feet	2.9	4.0	6.1
Spent brine temperature, °F	212	197	181
Isobutane concentration in spent brine, ppm	56	99	142
CO ₂ concentration in spent brine, ppm	212	83	11
H ₂ S concentration in spent brine, ppm	97	77	33

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NOTATION

A	= gas - liquid interfacial area, L^{-1}
a	= liquid - liquid interfacial area, L^{-1}
C^*	= equilibrium concentration, moles/ L^3
C_p	= heat capacity, Btu/lb $^{\circ}$ F
$[CO_2]$	= concentration of CO_2 in liquid phase
D	= liquid diffusivity, L^2/T
d'	= critical drop diameter at transition from Region I to Region II, cm.
d_p	= spherical drop diameter, in
G	= Mass velocity, lb/hr ft 2
g	= gravitational constant
G_o	= molar flow rate of vapor into the condenser, moles/T
G_j	= molar flow rate of vapor leaving increment j , moles/T
h	= convection heat transfer coefficient, $\frac{Btu}{hr\ ft^2\ ^{\circ}F}$
H_o	= Henry's law constant defined by eq (12)
K_c	= overall mass transfer coefficient, moles/TL 3
$(k_x a)_j$	= mass transfer coefficient, lb-moles/hr ft 3 .
k	= mass transfer coefficient, moles/hr ft 2
k_c	= thermal conductivity
k_{eq}	= vapor-liquid equilibrium constant
L	= tube length
L_j	= molar flow rate of condensate leaving increment j , moles/t
M_A	= molecular weight of solute
M	= ratio of wake to drop volume, (dimensionless)
m	= volume of wake element shed per volume of drop and unit length column (cm $^{-1}$)
$\bar{m}$	= average molecular weight of phase

- $\dot{m}_t$ = condensate flow rate for a vertical row
- N_{Sh} = Sherwood number (dimensionless)
- N_{Sc} = Schmidt number
- N_{Pr} = Prandtl number
- N = number of tubes in a vertical row
- P_{CO_2} = partial pressure of CO_2 in the gas phase, atm
- q_j = heat transferred from brine in j th increment, Btu/hr
- r = drop radius, L
- N_{Re} = Reynolds number (dimensionless)
- R_j = rate of mass transfer in j th increment, lb-moles/hr.
- S = tower cross sectional area, L^2
- t = time
- (ΔT) = log mean temperature difference, $^{\circ}F$
- T = temperature,
- U = superficial velocity L/T
- U_v = volumetric heat transfer coefficient, $Btu/hr\ ft^3^{\circ}F$
- V_A = molal volume of solute at normal boiling point, $cm^3\ g\text{-mole}^{-1}$
- V_T = terminal drop velocity, L/T
- V_j = volume of j th increment, ft^3
- x_j^* = equilibrium solubility, mole fraction
- $\bar{x}_j$ = average bulk concentration in brine, mole fraction
- x_j = mole fraction of a component in the liquid phase leaving increment j
- y_0 = mole fraction of component in the entering vapor

Notation (Cont'd)

y_j = mole fraction of component in vapor leaving increment j
 $\bar{y}_j^*$ = average equilibrium gas phase composition
 for the component for increment j

GREEK

$\Delta\rho$ = difference in densities of dispersed and continuous liquids
 ρ = density, g/cc.
 ϕ = void fraction
 $1-\phi$ = hold up
 σ = interfacial tension, dynes/cm
 ψ = "association parameter" (Wilke-Chang, 1955)
 μ = viscosity

SUBSCRIPTS

C = continuous (brine) phase
 D = dispersed (organic) phase
 i = inlet to column or increment
 L_M = log mean difference
 O = outlet from column or increment
 Z = height of increment
 I = Region I for terminal velocity (Klee, Treybal, 1956)
 II = Region II for terminal velocity (Klee, Treybal, 1956)

APPENDIX I

FORTRAN LISTING OF
COMPUTER PROGRAM

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    DIMENSION TF(50),TB(50),TFZONA(1),NTCP(1),H2CH2S(50), C4T(1),
1  H2DCO2(50),H2GIC4(50),H2SP(1),CO2P(1),H2ST(1),CO2T(1)
    DIMENSION DLTLM(100),VBOIL(100),AKXA(100),AKXAV(100),AKXAVG(100)
    DIMENSION VAPOR(100),FLT(100)
    DIMENSION TBOILK(100)
    REAL NB,NF,NBTOP,NB80T,NFTOP,NFBOT,MF,MB,MCLCC2,KTIC4,KBIC4,
1  IC4LEV,
1  IC4IN,
1  KH2SD,KTDH,KCO2D,KTDC2,KCCO2,KOH2S,KBCO2,KBDF,
1  KCO2B,KOBCO2,KORH2S,KH2SB,
1  KIC4,KTCCO2,KBCO2,KCO2,KTH2S,KBH2S,KH2S,M,IC4BOT
1  ,IC4CUT
    COMMON RHOB,NB,RHOF,SIGMAF,VPIC4,FIC4,MOLCC2,HG,HH2S,HCO2,
1  VPH2O,DATC4,DACC2,DAH2S,PPCO2,CO2ATM,PRCO2,TRCO2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
    COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2BO,XH2SBO,NBOIL
    COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
    COMMON TBIN,FLBR,FLIC4
    COMMON PRBOIL,TSAT
    COMMON TBDIA,TBLGTH,RPD,TCOND,PRCOND,YCO2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCCND,GCCND(100)
    COMMON NVERT,NHORIZ
    READ(5,5) WPB,TBIN,FLBR,FLIC4,TSUPER,NBOIL
5  FORMAT(5F8.1,I3)
    READ(5,6) TVOUT,XCO2IO,XH2SIO,XCO2BO,XH2SBO,YCO2B,YH2SB
6  FORMAT(F8.1,6F10.6)
C  CALCULATE SATURATION TEMP FOR IC4 -DEGREES F
    TSAT=TVOUT-TSUPER
C  HVI IS HEAT OF VAPORIZATION OF IC4 BTU/LB
    HVI=EXP(5.86-0.00637*TSAT)
    HEAT=HVI*FLIC4
    SUPERH=0.65*FLIC4*TSUPER
    TBOUT=TBIN-(HEAT+SUPERH)/FLBR
C  CALCULATE BOILER PRESSURE PRBOIL PSIA
    VPH2O=10.**(0.0251*((TVOUT-80.)*0.83)-0.2951)
    VPIC4=26.356*EXP(0.0112*TSAT)
    PRBOIL=VPH2O+VPIC4
C  CALCULATE TEMP PROFILE. INCREMENT 1 AT BOTTOM
C  ALL SUPERHEAT IN VAPOR-CONTINUOUS REGION AT TOP
    N=NBOIL
    UV=6500.*(529.06/VPIC4)**0.56
    QINC=HEAT/N
    DLTB=QINC/FLBR
    TBOIL(1)=TBOUT
    K=N+1
    DO 10 I=2,K
10  TBOIL(I)=TBOIL(I-1)+DLTB
    DO 20 I=1,N
    DLTLM(I)=(TBOIL(I+1)-TBOIL(I))/ALOG((TBOIL(I+1)-TSAT)/(TBOIL(I)-
1TSAT))
20  VBOIL(I)=QINC/(UV*DLTLM(I))
    DLTLM(N+1)=((TBIN-TVOUT)-(TBOIL(N+1)-TSAT))/ALOG((TBIN-TVOUT)/
1(TBOIL(N+1)-TSAT))
    VBOIL(N+1)=SUPERH/(500.*DLTLM(N+1))
C  CALCULATE IC4 CONCENTRATION IN BRINE STARTING WITH N+1
C  WPB IS WT PERCENT SALT IN BRINE
    GBR=FLBR/18.
    DO 30 I=1,N
    PRNO=0.02+353./TBOIL(I)
    SCNO=(22400./TBOIL(I))-44.9

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      AKXA(I)=(UV/18.)*(PRND/SCND)**0.667
30  AKXAV(I)=AKXA(I)*VBOIL(I)
      AKXA(N+1)=(500./18.)*(PRND/SCND)**0.667
      AKXAV(N+1)=AKXA(N+1)*VBOIL(N+1)
      XISO(N+2)=0.
      READ(5,7) TBOIA,TBLGTH,NVERT,NHORIZ,RPC,NCCND
7   FORMAT(F8.6,F8.1,I5,I5,F6.1,I5)
      READ(5,8) TCOND
8   FORMAT(F8.1)
      ITER=6
39  DO 40 J=1,K
      I=N+2-J
      TBOILK(I)=(TBOIL(I)+460.)/1.8
      HIC4=EXP(-0.61425+7033./TBOILK(I)+15.95*ALOG(TBOILK(I)/1000.))+
1WPB*(0.0958+55.3/TBOILK(I)+0.295*ALOG(TBOILK(I)/1000.))
      XEQISO=(HIC4*PRBOIL/14.7)*18./58.F+6
40  XISO(I)=(XISO(I+1)*(GBR-AKXAV(I)/2.)+XEQISO*AKXAV(I))/(GBR+AKXA
1V(I)/2.)
C   CALCULATE HENRYS CONSTANTS FOR CO2 AND H2S IN IC4
      IF(TSAT.LT.160.) HCO2I=40.+0.35*TSAT
      IF(TSAT.GE.160.) HCO2I=96.
      IF(TSAT.GE.220.) HCO2I=170.-0.333*TSAT
      HCO2I=HCO2I*14.7
      HH2SI=7.2+0.214*TSAT
      HH2SI=HH2SI*14.7
C   CALCULATE MASS TRANSFER COEFFICIENTS AKXAVG FOR CO2 AND H2S
      DO 50 I=1,K
50  AKXAVG(I)=AKXAV(I)*1.5
C   IC4 LIQUID FLOWRATE IS FLI(I)
      DELEL=FLIC4/(N*58.)
      DO 60 I=1,N
      FLI(I)=(FLIC4/58.)-I*DELEL
50  VAPOR(I)=I*DELEL
      DO 70 I=1,K
      T(I)=TBOIL(I)
70  CONTINUE
      CALL GAS
150 XCO2B(1)=(FLIC4*(XCO2IO-YCO2B)/58.+FLBR*XCO2BO/18.)/(FLBR/18.)
      YCO2(1)=(AKXAVG(1)*XCO2B(1)+FLIC4*XCO2IO/58.)/(AKXAVG(1)*PRBOIL/
1HCO2B(1)+FLI(1)*PRBOIL/HCO2I+VAPOR(1))
      IF(YCO2(1).LT.0.)YCO2(1)=1.E-6
      N=NBOIL
      DO 80 I=2,N
      XCO2I(I-1)=PRBOIL*YCO2(I-1)/HCO2I
      ABC=XCO2I(I-1)
      IF(ABC.LT.0.)XCO2I(I-1)=1.E-6
      XCO2B(I)=(FLBR*XCO2B(I-1)/18.+AKXAVG(I-1)*(XCO2B(I-1)-PRBOIL
1*YCO2(I-1)/HCO2B(I-1)))/(FLBR/18.)
80  YCO2(I)=(AKXAVG(I)*XCO2B(I)+FLI(I)*XCO2I(I-1)+VAPOR(I-1)*YCO2(I
1-1))/(AKXAVG(I)*PRBOIL/HCO2B(I)+FLI(I)*PRBOIL/HCO2I+VAPOR(I))
      XCO2B(N+1)=(FLBR*XCO2B(N)/18.+AKXAVG(N)*(XCO2B(N)-PRBOIL*YCO2(N)/
1HCO2B(N+1)))/(FLBR/18.)
      YCO2(N+1)=(AKXAVG(N+1)*XCO2B(N+1)+VAPOR(N)*YCO2(N))/(AKXAVG(N+1)
1*PRBOIL/HCO2B(N+1)+VAPOR(N))
      IF(YCO2(N+1).GT.YCO2B) GO TO 110
      GO TO 120
110 YCO2B=YCO2B*1.03
      GO TO 150
120 CONTINUE
155 XH2SB(1)=(FLIC4*(XH2SIO-YH2SB)/58.+FLBR*XH2SBO/18.)/(FLBR/18.)
      YH2S(1)=(AKXAVG(1)*XH2SB(1)+FLIC4*XH2SIO/58.)/(AKXAVG(1)*PRBOIL/

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1 HH2SB(1)+FLI(1)*PRBOIL/HH2SI+VAPOR(1))
  IF(YH2S(1).LT.0.)YH2S(1)=1.E-6
  DO 85 I=2,N
    XH2SI(I-1)=PRBOIL*YH2S(I-1)/HH2SI
    ABC=XH2SI(I-1)
    IF(ABC.LT.0.)XH2SI(I-1)=1.E-6
    XH2SB(I)=(FLBR*XH2SB(I-1)/18.+AKXAVG(I-1)*(XH2SB(I-1)-PRBOIL*
1 YH2S(I-1)/HH2SB(I-1)))/(FLBR/18.)
85 YH2S(I)=(AKXAVG(I)*XH2SB(I)+FLI(I)*XH2SI(I-1)+VAPOR(I-1)*YH2S(I-1
1 ))/(AKXAVG(I)*PRBOIL/HH2SB(I)+FLI(I)*PRBOIL/HH2SI+VAPOR(I))
    XH2SB(N+1)=(FLBR*XH2SB(N)/18.+AKXAVG(N)*(XH2SB(N)-PRBOIL*YH2S(N)/
1 HH2SB(N+1)))/(FLBR/18.)
    YH2S(N+1)=(AKXAVG(N+1)*XH2SB(N+1)+VAPOR(N)*YH2S(N))/(AKXAVG(N+1
1 *PRBOIL/HH2SB(N+1)+VAPOR(N))
    IF(YH2S(N+1).GT.YH2SB) GO TO 115
    GO TO 125
115 YH2SB=YH2SB*1.03
    GO TO 155
125 CONTINUE
    CALL CNDSER
    CALL OUTPUT
C TOWER DIAMETER; IN.
  DIA=6.065
C INCREMENT HEIGHT; IN.
  Z=1.0
  MB=FLBR
  MF=FLIC4
  TFBOT=TCOND
  TFTOP=TSAT
  TBTOP=TBOIL(1)
C AVG. IC4 DROP SIZE; IN
  DP=.15
C ALL CONC. ARE GIVEN IN PPM
  TOTALP=PRBOIL
  CNB=XISO(1)*58.0*1.0E6/18.0
  CBNB=CNB
  H2SA=XH2SB(1)*34.0*1.0E6/18.0
  H2SB=H2SA
  CC2A=XCO2B(1)*44.0*1.0E6/18.0
  CC2B=CC2A
  HSICB=XH2SC(NCOND)
  CCICB=XCO2C(NCOND)
  GB=MB*(((WPB/100.)/58.12+(1.-(WPB/100.)/18.02)))/(
1 3.142*(DIA/12.)**2)/4)
C CALCULATE TOWER ENERGY BALANCE; ASSUME NO HEAT LOSS
C ENTHALPY OF BRINE DEPENDS ON DISSOLVED SOLIDS
  IF(WPB.LE.10.) HBTOP=(1.01-.03*(WPB**.55))*(TBTOP-32.)
  IF(WPB.GT.10.) HBTOP=(1.01-.03*(WPB**.61))*(TBTOP-26.)
C CALCULATE ENTHALPY OF ORGANIC EXIT
  HFTOP=-69.74+(.8463*TFTOP)-(2.32E-3*TFTOP**2)+(7.213E-6)
1 *(TFTOP**3)
C COMPUTE INCOMING ORGANIC ENTHALPY
  HFBOT=-69.74+(.8463*TFBOT)-(2.32E-3*TFBOT**2)+(7.213E-6*
1 TFBOT**3)
C COMPUTE BOTTOM BRINE ENTHALPY
  HBBOT=HBTOP-(MF/MB)*(HFTOP-HFBOT)
C COMPUTE BRINE EXIT TEMP.
  IF(WPB.LE.10.) TBBOT=HBBOT/(1.01-.03*WPB**.55)+32.
  IF(WPB.GT.10.) TBBOT=HBBOT/(1.01-.03*WPB**.61)+26.
C CHECK FOR FLOODING, USING AVG. PHYSICAL PROPERTIES
  NBTOP=.4142*EXP(.29385*EXP(TBTOP/100.))-6.0971)*WPB**

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1 (1.8-.0076*(TBTOP-100.))+(.0115*((TBTOP-100.))**
1 .7881))
C COMPUTE AVG. VISCOSITY OF BRINE
NB3OT=.4132*EXP(.29385*EXP(TBBOT/100.)-6.0971)*WPB**
1 (1.8-.0076*(TBBOT-100.))+(.0115*((TBBOT-100.))**
1 .7881))
AVGNB=(NBBOT+NB3OT)/2.
VISIC4=.115-1.25E-3*((TFTOP-32.)/1.8-50.)
VIS=2.42*VISIC4
C COMPUTE BRINE DENSITY TOWER ENTRANCE
ROBTOP=62.+.46*WPB-.00265*((TBTOP-100.))**1.4166)
C COMPUTE BRINE DENSITY TOWER EXIT
ROBBOT=62.+.46*WPB-.00265*((TBBOT-100.))**1.4166)
C COMPUTE AVG. BRINE DENSITY
AVGROB=(ROBBOT+ROBTOP)/2.
SPGRB=AVGROB/62.
C COMPUTE ORGANIC DENSITY TOWER ENTRANCE
IF(TFTOP.GE.100.)ROFTOP=33.4-.003676*((TFTOP-100.)
1 **1.5835)
IF(TFBOT.GE.100.)ROFBOT=33.4-.003676*((TFBOT-100.))**1.5835)
IF(TFTOP.LT.100.)ROFTOP=33.4
IF(TFBOT.LT.100.)ROFBOT=33.4
AVGROF=(ROFTOP+ROFBOT)/2.
VELOF=MF/(AVGROF*3.142*((DIA/12.))**2)/4.)
SPGRF=AVGROF/62.
C COMPUTE CHANGE IN SPECIFIC GRAVITY
DELSPG=ABS(SPGRF-SPGRB)
DIVSPG=SPGRF/SPGRB
C COMPUTE FLOODING FLOW RATE
CVELOB=((1.8*DP**(.056))*((1./AVGNB**(.075))*DIVSPG**(.5))
1 *VELOF+(47.*((1./AVGNB**(.07))*DELSPG**(.14))*((1./AVGROB**
1 (.5))))**2
C CALCULATE VEL. OF BRINE
VELOB=MB/(AVGROB*(3.142*((DIA/12.))**2)/4.))
C FLOODING?
IF(VELOB.LT.CVELOB) GO TO 21
IF(VELOB.GE.CVELOB) GO TO 15
C WRITE AN ERROR MESSAGE
15 PRINT 17,CVELOB
17 FORMAT(1X,'THE TOWER REJECTS ORGANIC',1X,'CVELOB=',F8.2)
STOP
21 CONTINUE
C FIND TERMINAL VEL,HOLDUP,INTERFACIAL AREA,INCREMENT VOL.
C NOTE INTERCONVERSION BETWEEN METRIC AND ENGLISH UNITS
RHOBM=.016018*AVGROB
RHOFM=.016018*AVGROF
DELRCM=ABS(RHOBM-RHOFM)
C COMPUTE ORGANIC SURFACE TENSION USING AVG. TEMP.
SIGMAF=63.35+.375*WPR-.1035*TBTOP+.05*TFBOT
C COMPUTE CRITICAL DROP DIA.
CDPM=.33*((1./RHOBM**(.14))*((1./DELRCM**(.43))*((AVGNB/100.)
1 **(.3))*(SIGMAF**(.24)))
C CONVERT GIVEN DROP DIA. TO METRIC UNITS
DPM=2.54*DP
C PICK BEST TERMINAL VEL. EQ. FROM DROP DIA.
IF(DPM.LT.CDPM)VTM=38.3*((1./RHOBM**(.45))*DELRCM**(.58)
1 *((1./((AVGNB/100.))**(.11))*((DPM**(.7)))
IF(DPM.GE.CDPM)VTM=17.6*((1./RHOBM**(.55))*DELRCM**(.28)
1 *((AVGNB/100.))**(.1))*(SIGMAF**(.18)))
C USE ZEN'S EQ. TO CALCULATE HOLDUP
VT=VTM/30.48

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VF=VELOF/3600.
VB=VELOB/3600.
EPSLON=((VT+VB-VF)+SQRT((VT+VB-VF)**2-4.*VT*VB))/
1 (2.*VT)
C COMPUTE HOLDUP
H=1.-EPSLON
C COMPUTE INTERFACIAL AREA
A=5.*H/(DP/12.)
C COMPUTE VOL. OF AN INCREMENT
VOLINC=3.142*(Z/12.)*((DIA/12.)**2)/4.
C WRITE TERMINAL VELOCITY, HOLD UP, INTERFACIAL AREA
C LETAN-KEHAT TEMP. PROFILE SEGMENT
C SELECT PROPER WAKE PARAMETERS
IF(H.LT. .06)M=1.0
IF(H.GE. .06)M=.83
IF(H.LT. .06.OR.H.GT..4)SM=.023
IF(H.GE..06.OR.H.LE..4)SM=.03
C COMPUTE TEMP. TOP OF ZONE A
C ASSIGN STORAGE CAPACITY FOR EACH INCREMENT OF TOWER HT.
C COMPUTE AVG. HEAT CAPACITY FOR BRINE AND ORGANIC
CPFTOP=.656+.002*((TFTOP-32.)/1.8-70.)
CPFBOT=.656+.002*((TFBOT-32.)/1.8-70.)
AVGCPF=(CPFTOP+CPFBOT)/2.0
AVGCPB=1.01-.0298*(WPB**.6152)
C ADDITIONAL LETAN-KEHAT PARAMETERS
SR=(AVGCPF*AVGROF)/(AVGCPB*AVGROB)
CAPR=VELOF/VELOB
P=(1./CAPR)+M
C FIRST GUESS EXIT CONC. GUESS IN IC4FOR H2S CO2
TH2SA=((HSICB *ROFTOP/34.04)/1.E6)
TCO2A=((COICB*RCFTOP/44)/1.E6)
C COMPUTE TEMP. AT TOP ZONE A FOR ORGANIC AND BRINE
TFZONA(1)=(TFBOT-TBBOT)*EXP(-M/SR)+TEBOT
PRINT 48,TFZONA(1)
48 FORMAT(1X,'TEMP. OF ORGANIC AT TOP ZONE A=',F10.2)
C ADDITIONAL LETAN-KEHAT PARAMETERS
ALPHA1=(-SM/2.)*((1./M+1./SR-1./P)+SQRT(
1 (1./M+1./SR+1./P)**2.-(4./(M*SR))))
ALPHA2=(-SM/2.)*((1./M+1./SR-1./P)-SQRT((1./M
1 +1./SR+1./P)**2.-(4./(M*SR))))
S=(ALPHA1+(SM/SR)-((SR*SM)/(P*M))*(EXP(M/SR)-1.))
1 /(ALPHA2-ALPHA1)
C BEGIN ADDING INCREMENTS OF HEIGHT
C THE INCREMENTS ARE FOR ZONE C FROM LETAN-KEHAT(1968)
N=0
ZNEW=0.0
ZM=2.54*Z
90 ZNEW=ZNEW+ZM
N=N+1
TF(N)=(((SM/SR)*(((1.+S)/ALPHA1)*(1.-EXP(ALPHA1*
1 ZNEW)))-(S/ALPHA2)*(1.-EXP(ALPHA2*ZNEW)))+1.)
1 *EXP(-M/SR))*(TFBOT-TBBOT)+TEBOT
TB(N)=(((SM/SR)*(((1.+S)/ALPHA1)*(1.-
1 ((SR/SM)*ALPHA1+1.)*EXP(ALPHA1*ZNEW))
1 -(S/ALPHA2)*(1-(SR/SM*ALPHA2+1.))*
1 EXP(ALPHA2*ZNEW)))+1.)*EXP(-M/SR))
1 *(TFBOT-TBBOT)+TEBOT
C SENTINEL TO STOP ITERATION
IF(TFTOP-TF(N).GT.2.E-4)GO TO 90
C APPROX. THE TOTAL TOWER HT. IN CENTIMETERS
TOWER=ZNEW+(42.-(H*100.))+35.

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PRINT 399,TOWER
399  FORMAT(20X,'TOTAL TOWER HT.=' ,F10.2)
      NTOP(1)=N
C MASS TRANSFER SEQUENCE
C COMPUTE SOLUBILITY CONC. IC4 IN BRINE
C ADD CONVERSIONS TO PROGRAM;CHANGE PPM TO #MOL/CU.FT.,#MOL/HR
C CONC. H2S IN BRINE
      H2STOP=(H2SA*ROBTOP/34.04)/1.E6
C CONC. CO2 IN BRINE
      CO2TCP=(CO2A*ROBTOP/44.)/1.E6
C CONC. IC4 IN BRINE
      CCTOP=(CNB*ROBTOP/58.12)/1.E6
C THE UNITS OF ABOVE CONVERSIONS=#MOL/CU FT.
C CONVERT THE SAME IN PUT PARAMETERS TO #MOL/HR
      H2STP=(H2SB*MB/34.02)/1.E6
      CO2TP=(CO2B*MB/44.)/1.E6
      IC4BR=(CBNB*MB/58.12)/1.E6
C CONVERT INPUT FOR H2S,CO2 IN IC4 #MOL/HR
      H2SP(1)=H2STP
      CO2P(1)=CO2TP
      H2SIC4=(HSICB*MF/34.02)/1.E6
      CO2IC4=(COICB*MF/44.)/1.E6
      IC4IN=MF/58.12
      DO 804 K=1,15
      N=NTOP(1)
      TCO2=TCO2A
      TH2S=TH2SA
      H2STOP=(H2SA*ROBTOP/34.02)/1.E6
      CO2TCP=(CO2A*ROBTOP/44.0)/1.E6
      CCTOP=(CNB*ROBTOP/58.12)/1.E6
C HENRY'S LAW FOR IC4 IN BRINE (PPM/ATM)
      HIC4=EXP(-.61425+1000.*((7.0329/((TBTOP-32.)
1 /1.8+273.))+15.946*ALOG(((TBTOP-32.)/1.8+
1 273.)/1000.))+WPB*(.095826+1000.*
1 (.055289/((TBTOP-32.)/1.8+273.))+.29564*
1 ALOG(((TBTOP-32.)/1.8+273.)/1000.)))
C COMPUTE HENRY'S LAW CONSTANT FOR CO2 IN BRINE
      MOLCO2=WPB*0.17094
      IF(MOLCO2.GE.2.)HCO2=62.*((TBTOP-32.)/1.8)+1300.
      IF(MOLCO2.LT.2..AND.MOLCO2.GE.1.)HCO2=51.*((TBTOP-32.)/1.8)
1 +1000.
      IF(MOLCO2.LT.1..AND.MOLCO2.GE..5)HCO2=
1 46.*((TBTOP-32.)/1.8)+900.
      IF(MOLCO2.LT..5)HCO2=42.*((TBTOP-32.)/1.8)+800.
C COMPUTE HENRY'S LAW FOR H2S IN BRINE
      IF(TBTOP.LT.212.)HG=11.6*((TBTOP-32.)/1.8)+420.
      IF(TBTOP.GE.212.)HG=5.6*((TBTOP-32.)/1.8)+1020.
C CORRECT HENRY'S LAW FOR NaCl EFFECT
      HH2S=10.**(.1*MOLCO2)*HG
C COMPUTE SATURATION PRESSURE FOR IC4,H2O
      TVPIC4=26.356*EXP(.0112*TFTOP)
      TVPH2O=10.**(((TBTOP-80.))**.83)*.0251-.2951)
C COMPUTE HENRY'S LAW FOR H2S,CO2 IN IC4 AT TOP OF TOWER
C HENRY'S LAW FOR H2S IN IC4
      IF(TFTOP.GT.160.)HH2SC4=.217*TFTOP+6.333
      IF(TFTOP.LE.160.)HH2SC4=.35*TFTOP-15.
C COMPUTE HENRY'S LAW FOR CO2 IN IC4
      IF(TFTOP.LT.120.)HCO2C4=.6*TFTOP+14.
      IF(TFTOP.LT.140..AND.TFTOP.GE.120.)HCO2C4=.3*TFTOP+50.
      IF(TFTOP.LT.160..AND.TFTOP.GE.140.)HCO2C4=.2*TFTOP+64.
      IF(TFTOP.LT.180..AND.TFTOP.GE.160.)HCO2C4=.1*TFTOP+80.

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IF(TFTOP.LT.200..AND.TFTOP.GE.180.)HCC2C4=98.
IF(TFTOP.GE.200.)HCC2C4=118.-.1*TFTOP
C THE ABOVE EQ. IS GOOD ONLY TO 230 F
C COMPUTE PARTIAL PRESSURE H2O AT TOP OF TOWER
PAH2O=10.**(.0251*((TBTOP-80.)**.83)-.2951)
CESTIMATE DIFFUSIVITIES FOR CO2,H2O,H2S IN BRINE
TDAIC4=((7.4E-8)*((18.02*2.6)**.5)*((TBTOP-32.)/1.8+273.))
1 /((96.2**.6)*NBTC) **3.875
TDACC2=((6.577E-8*((TBTOP-32.)/1.8+273.))/NBTC) **3.875
TDAH2S=((5.403E-8*((TBTOP-32.)/1.8+273.))/NBTC) **3.875
C DEFINE RE, SC NO.; PICK SHERWOOD NO.
REDROP=((DP/12.)*VT*3600*AVGROB)/(AVGNB*2.42)
RE=(DIA/12.)*VELCB*ROBTOP/(NBTC*2.42)
SCH2S=(NBTC*2.42)/(ROBTOP*TDAH2S)
C
SCC2=(NBTC*2.42)/(ROBTOP*TDACC2)
C
SCIC4=(NBTC*2.42)/(ROBTOP*TDAIC4)
IF(DPM.LT..1)SH=2.+.76*(RE**.5)*(SCIC4**(1./3.))
IF(DPM.GE..1)SH=.6*((DIA/12.)*VELCB/TDAIC4)**.5)
IF(REDROP.GT.200..AND.DPM.GE..4)SH=50.+.0085*RE*(SCIC4**
1 .7)
IF(DPM.LT..1)SHC2=2.+.76*(RE**.5)*(SCC2**(1./3.))
IF(DPM.GE..1)SHC2=.6*((DIA/12.)*VELCB/TDACC2)**.5)
IF(REDROP.GT.200..AND.DPM.GE..4)SHC2=50.+.0085*RE*(SCC
1 C2**.7)
IF(DPM.LT..1)SHH2S=2.+.76*(RE**.5)*(SCH2S**(1./3.))
IF(DPM.GE..1)SHH2S=.6*((DIA/12.)*VELCB/TDAH2S)**.5)
IF(REDROP.GT.200..AND.DPM.GE..4)SHH2S=50.+.0085*RE*(SCH2
1 S**.7)
C INTERVAL MASS TRANSFER COEFF. FOR CO2
DCO2=7.4E-8*((58.12)**.5)/(34.**.6))*((TFTOP-32.)/1.8
1 +273.)/VISIC4
TIME=ZM/VTM
IF(DPM.LE..1)KTDC2=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1 (1.-EXP(-(3.142**2.)*DCO2*TIME/((DPM/2.）**2.))))
IF(DPM.GT..1)KTDC2=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1 (1.-EXP(-2.25*(3.142**2.)*DCO2*TIME/((DPM/2.）**2.))))
IF(DPM.GT..4.AND.REDROP.GT.200.)KTDC2=.00375*VTM*
1 (NBTC/(NBTC+VISIC4))
C INTERVAL MASS TRANSFER COEFF. H2S IN IC4
DH2S=7.4E-8*((58.12**.5)*(32.2**.6))*((TFTOP-32.)/1.8+
1 273.)/VISIC4
IF(DPM.LT..1)KTDH=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1 (1.-EXP(-(3.142**2.)*DH2S*TIME/((DPM/2.）**2.))))
IF(DPM.GT..1)KTDH=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1 (1.-EXP(-2.25*(3.142**2.)*DH2S*TIME/((DPM/2.）**2.))))
IF(DPM.GT..4.AND.REDROP.GT.200.)KTDH=.00375*VTM*
1 (NBTC/(NBTC+VISIC4))
DENIC4=ROBTOP/58.12
KH2SD=(KTDH*DENIC4/((2.54**2)*144.))*3600
KCO2D=(KTDC2*DENIC4/((2.54**2)*144.))*3600
C COMPUTE MOLAR DENSITY
CDEN=ROBTOP*((WPB/100.)/58.12)+(1.-(WPB/100.))/18.02)
C COMPUTE SOLUBILITY CONC. FOR CO2,IC4,H2S
TEIC4=(HIC4*TVPIC4/14.7)*ROBTOP/(1.E6*58.12)
C SEGMENT NEEDED FOR SEVERAL CO2 CALCULATIONS
C COMPUTE AVG. MASS TRANSFER COEFFICIENT
KTIC4=(SH*TDAIC4*CDEN)/(DIA/12.)
KTCC2=(SHC2*TDACC2*CDEN)/(DIA/12.)
KTH2S=(SHH2S*TDAH2S*CDEN)/(DIA/12.)

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      KCCO2=1./((1./KTCO2+(1./((HCO2/HCCO2C4)*KCCO2)))
      KOH2S=1./((1./KTH2S+(1./((HH2S/HH2SC4)*KH2SD)))
C EQUILIBRIUM CONC. FOR H2S, CO2 IN BRINE AT TOP INCREMENT
      TEQCO2=(HCO2C4/HCO2)*TCO2
      TEQH2S=(HH2SC4/HH2S)*TH2S
C BEGIN ITERATION DOWN THE TOWER FOR MASS TRANSFER
      200   TBC=TB(N)
           TFC=TF(N)
C CALL PROPERTY ROUTINE
      CALL PROP(TBC,TFC,TOTALP)
C RECOMPUTE SOL. CONC. FOR IC4 AT BOTTOM
      250   BEQIC4=(HIC4*VPIC4/14.7)*RHOB/(1.E6*58.12)
C COMPUTE SOL. CONC. FOR H2S IN BRINE
C REDEFIN RE&SC NO.; PICK SHERWOOD NO.
      RE=((DIA/12.)*VELOB*RHOB)/(NB*2.42)
      SCIC4=(NB*2.42)/(DAIC4*RHOB)
      SCH2S=(NB*2.42)/(DAH2S*RHOB)
      SCCO2=(NB*2.42)/(DACO2*RHOB)
      IF(DPM.LT..1)SH=2.+ .76*((RE**.5)*(SCIC4**(1./3.)))
      IF(DPM.GE..1)SH=.6*((DIA/12.)*VELOB/DAIC4)**.5)
      IF(REDROP.GT.200..AND.DPM.GE..4)SH=50.+ .0085*RE*
1      (SCIC4**.7)
      IF(DPM.LT..1)SHCO2=2.+ .76*((RE**.5)*(SCCO2**(1./3.)))
      IF(DPM.GE..1)SHCO2=.6*((DIA/12.)*VELOB/DACO2)**.5)
      IF(REDROP.GT.200..AND.DPM.GE..3)SHCO2=50.+ .0085*RE*
1      (SCCO2**.7)
      IF(DPM.GE..1)SHH2S=.6*((DIA/12.)*VELOB/DAH2S)**.5)
      IF(DPM.LT..1)SHH2S=2.+ .76*((RE**.5)*(SCH2S**(1./3.)))
      IF(REDROP.GT.200..AND.DPM.GE..4)SHH2S=50.+ .0085*RE
1      *(SCH2S**.7)
C CALCULATE NEW MOLAR DENSITY
      CDEN=RHOB*((WPB/100.)/58.5)+((1.-(WPB/100.))/18.02))
C CALCULATE NEW MASS TRANSFER COEFF.'S
C MASS TR. COEF. AT BOTTOM OF INCREMENT
      KBIC4=(SH*DAIC4*CDEN)/(DIA/12.)
      KBCO2=(SHCO2*DACO2*CDEN)/(DIA/12.)
      KBH2S=(SHH2S*DAH2S*CDEN)/(DIA/12.)
C INTERNAL MASS TR. COEF. FOR CO2 IN IC4
      IF(DPM.LE..1)KBDC2=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1      (1.-EXP(-(3.142**2.)*DCO2*TIME/((DPM/2.)**2.))))
      IF(DPM.GT..1)KBDC2=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1      (1.-EXP((-2.25*(3.142**2.)*DCO2*TIME/((DPM/2.)**2.))))
      IF(DPM.GT..4.AND.REDROP.GT.200.)KBDC2=.00375*VTM
1      *(NB/(NB+VISIC4))
C MASS TR. COEFF. FOR H2S IN IC4
      IF(DPM.LT..1)KBDH=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1      (1.-EXP(-(3.142**2.)*DH2S*TIME/((DPM/2.)**2.))))
      IF(DPM.GE..1)KBDH=-(DPM/(6.*TIME))*ALCG(1.-SQRT
1      (1.-EXP(-2.25*(3.142**2.)*DH2S*TIME/((DPM/2.)**2.))))
      IF(DPM.GT..4.AND.REDROP.GT.200.)KBDH=.00375*VTM*
1      (NB/(NB+VISIC4))
      DENIC4=RHOF/58.12
      KH2SD=(KBDH*DENIC4/((2.54**2)*144.))*3600
      KCO2B=(KBDC2*DENIC4/((2.54**2)*144.))*3600
      KOCO2=1./((1./KBCO2+(1./((HCO2/HCCO2C4)*KCO2B)))
      KOH2S=1./((1./KBH2S+(1./((HH2S/HH2SC4)*KH2SD)))
C OVERALL MASS TR. COEF.
      KIC4=(KBIC4+KTIC4)/2.
      KCO2=(KOCO2+KCCO2)/2.
      KH2S=(KOBH2S+KCH2S)/2.
C CALL SUBROUTINES LOG MEAN CONC.

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      IF (CO2TOP.LT.1.E-6) H2OCO2(N)=CO2TOP
      IF (CO2TOP.LT.1.E-6) GO TO 256
      CALL CO2(KCO2,A,VOLINC,VELOB,DIA,TEQCC2,BEQCC2,CO2TOP
1      ,TCO2,MF,MB)
C COMPLETE MATERIAL BALANCE BEFORE NEXT INCREMENT
255 IF (H2STOP.LT.1.E-6) H2OH2S(N)=H2STOP
      IF (H2STOP.LT.1.E-6) GO TO 257
      CALL H2S(KH2S,A,VOLINC,VELOB,DIA,TEQH2S,BEQH2S,H2STOP
1      ,TH2S,MF,MB)
C COMPLETE MATERIAL BALANCE BEFORE NEXT INCREMENT
C REDEFINE H2S,CO2, AND IC4 VARIABLES
      CO2TOP=CO2BOT
      TCO2=TCO2B
      TEQCC2=BEQCC2
      KCCO2=KORCO2
      H2STOP=H2SBOT
      TH2S=TH2SB
      TEQH2S=BEQH2S
      KOH2S=KOBH2S
257 CALL CC(KIC4,A,VOLINC,VELOB,DIA,TEGIC4,BEGIC4,CCTOP)
      CCTOP=IC4BOT
      TEQIC4=BEQIC4
      KTIC4=KBIC4
      IF (N.LT.1) GO TO 505
      IF (CO2TOP.LT.1.E-6) GO TO 258
      H2OCO2(N)=CO2BOT
258 IF (H2STOP.LT.1.E-6) GO TO 259
      H2OH2S(N)=H2SBOT
259 H2OIC4(N)=IC4BOT
      IF (N.GE.1) GO TO 506
505 CO2T(1)=CO2BOT
      IF (CO2TOP.LT.1.E-6) CO2T(1)=CO2TOP
      H2ST(1)=H2SBOT
      IF (H2STOP.LT.1.E-6) H2ST(1)=H2STOP
      C4T(1)=IC4BOT
506 N=N-1
C RECALL NEXT N
      IF (N) 550,400,500
400 TBC=TBOT
      TFC=TFZONA(1)
      CALL PROP(TBC,TFC,TOTALP)
C IF THE ABOVE N IS POSITIVE GO AROUND THE LOOP AGAIN
      GO TO 250
500 GO TO 200
C DETERMINE THE AMOUNT IC4,H2S,CO2 AT TOWER EXIT, (# MOL/HR)
C
550 H2SLEV=H2STOP*MB/RHOB
      CO2LEV=CO2TOP*MB/RHOB
      IC4LEV=CCTOP*MB/RHOB
C MASS BALANCE FOR H2S,CO2 IN IC4 EXIT STREAM
C
      H2SOUT=TH2SA/ROFTOP*MF
      IF (ABS((H2STP+H2SIC4)-(H2SLEV+H2SOUT)).LT.1.E-5) GO TO 703
      IF ((H2STP+H2SIC4)-(H2SLEV+H2SOUT)) 650,703,680
650 CONTINUE
680 TH2SA=(H2STP+H2SIC4-H2SLEV)*ROFTOP/MF
702 CONTINUE
703 CONH2S=H2SOUT*ROFTOP/MF
      CO2OUT=TCO2A/ROFTOP*MF
      IF (ABS((CO2TP+CO2IC4)-(CO2LEV+CO2OUT)).LT.1.E-5) GO TO 805
      IF ((CO2TP+CO2IC4)-(CO2LEV+CO2OUT)) 685,804,695

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685     CONTINUE
695     TCO2A=(CO2TP+CO2IC4-CO2LEV)*RQFTCP/MF
804     CONTINUE
805     CONTINUE
        CONCO2=CO2OUT*RQFTCP/MF
        IC4OUT=(IC4IN+IC4BR-IC4LEV)
C MASS BALANCE GIVES LB/HR IC4 ENTERING THE BOILER
    PRINT 901,TH2SA,TCO2A
    901     FORMAT(1X,'# MOLE/CU FT.H2S,CO2 IN IC4 TO BOILER=',1P5E16.6)
C COMPUTE MOL. FRACTION H2S,CO2 IN TOWER EXIT
C MOL. FRACTION H2S IN IC4 EXIT STREAM
C
    XHS=H2SOUT/(H2SOUT+IC4OUT+CO2OUT)
C MOL. FRACTION CO2 IN IC4 EXIT STREAM
C
    XCO=CO2OUT/(CO2OUT+IC4OUT+H2SOUT)
C MOL. FRACTION IC4 IN EXIT STREAM
C
    XC4=IC4OUT/(CO2OUT+H2SOUT+IC4OUT)
C DETERMINE PARTIAL PRESSURE OF IC4,H2S, AND CO2
    PACO2=(HCO2C4*XCO)*14.7
    PAH2S=(HH2SC4*XH2)*14.7
    PAIC4=(VPIC4*XIC4)
C SUM THE PARTIAL PRESSURES OF H2O,IC4,H2S,CO2
    SUMP4=PACO2+PAIC4+PAH2S+PAH2O
    PRINT 696
    696     FORMAT(20X,'CONCENTRATIONS OF IC4,CO2,H2S IN BRINE(MOL FRACT)')
    PRINT 697
    697     FORMAT(20X,'TB(F)',10X,'TF(F)',12X,'IC4',12X,'CO2',12X,'H2S')
        N=NTCP(1)
    698     H2OIC4(N)=H2OIC4(N)/CDEN
        H2OCO2(N)=H2OCO2(N)/CDEN
        H2OH2S(N)=H2OH2S(N)/CDEN
    PRINT 699,TB(N),TF(N),H2OIC4(N),H2OCO2(N),H2OH2S(N)
    699     FORMAT(10X,1P5E16.4)
        N=N-1
        IF(N.GT.1) GO TO 698
        C4T(1)=C4T(1)/CDEN
        CO2T(1)=CO2T(1)/CDEN
        H2ST(1)=H2ST(1)/CDEN
    PRINT 499,TBBOT,TFZONA(1),C4T(1),CO2T(1),H2ST(1)
    499     FORMAT(10X,1P5E16.4)
        IF(CONCO2.LT.0.) CONCO2=1.E-6
        XCO2IO=CONCO2*58.0/RHOF
        IF(CONH2S.LT.0.) CONH2S=1.E-6
        XH2SIO=CONH2S*58.0/RHOF
        ITER=ITER-1
        PRBOIL=TOTALP
        IF(ITER.GT.0) GO TO 39
C SUM THE PARTIAL PRESSURES OF H2O,IC4,H2S,CO2
C CHECK TO SEE IF SUM OF PARTIAL PRESSURES EXCEED TOTAL PRESSURE
    IF(SUMP4.GT.TOTALP)GO TO 1000
    IF(SUMP4.LE.TOTALP)GO TO 2000
1000 PRINT 151
    151     FORMAT(1X,'THE COLUMN WILL NOT OPERATE UNDER GIVEN CONDITIONS
    1 . INPUT VALUES ARE CHANGING DUE TO GAS COMING OUT OF SOLUTION')
2000     STOP
        END
C PHYSICAL PROPERTY SUBROUTINE
    SUBROUTINE PROP(TBC,TFC,TOTALP)
        COMMON RHOB,NB,RHOF,SIGMAF,VPIC4,FIC4,MOLCC2,

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1  HG,HH2S,HCO2,VPH2O,DAIC4,DACO2,DAH2S,PFCC2,CC2ATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2BO,XH2SBO,NBOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBOIA,TBLGTH,RPD,TCOND,PRCOND,YCO2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCOND,GCOND(100)
COMMON NVERT,NHORIZ
REAL NB
C BRINE DENSITY
RHOB=62.+46*WPB-.00265*((TBC-100.))**1.4166)
C BRINE VISCOSITY
NB=.4131*EXP(.29385*EXP(TBC/100.)-6.0971)*WPB**((1.8-
1 .0076*(TBC-100.))+.6818-(.0115*((TBC-100.))**.7881))
C IC4 DENSITY
RHOF=33.4-.003676*((TFC-100.))**1.5835)
C SURFACE TENSION IC4
SIGMAF=63.35+(.375*WPB)-(.1035*TBC)+(.05*TFC)
C VAPOR PRESSURE IC4
VPIC4=26.356*EXP(.0112*TFC)
C HENRY'S LAW IC4
HIC4=EXP(-.61425+1000.*((7.0329/((TBC-32.)/1.8+273.))
1 +15.946*ALOG(((TBC-32.)/1.8+273.)/1000.))+WPB*(.09562+1000.*
1 (.055289/((TBC-32.)/1.8+273.))+.29564*ALOG(((TBC-32.)
1 /1.8+273.)/1000.)))
MOLCO2=WPB*.17094
C ESTIMATE HENRY'S LAW CONSTANT FOR H2S IN BRINE
C PICK HENRY'S LAW CONSTANT ACCORDING TO TEMP.
IF(TBC.LE.212.)HG=11.6*((TBC-32.)/1.8)+420.
IF(TBC.GT.212.)HG=5.6*((TBC-32.)/1.8)+1020.
C CORRECT HENRY'S LAW FOR NAOL EFFECT
HH2S=10.**(.1*MOLCO2)*HG
C HENRY'S LAW CONSTANT CO2
IF(MOLCO2.GT.2.)HCO2=62.*((TBC-32.)/1.8)+1300
IF(MOLCO2.LT.2..AND.MOLCO2.GT.1.)FCC2=51.*((TBC-32.)/1.8)
1 +1000.
IF(MOLCO2.LT..5)HCO2=42.*((TBC-32.)/1.8)+800.
IF(MOLCO2.LT.1..AND.MOLCO2.GT..5)FCC2=46.*((TBC-32.)/1.8)
1 +900.
C VAPOR PRESSURE H2O
VPH2O=10.**(.0251*((TBC-80.))**.83)-.2951)
C DIFFUSIVITY IC4, H2S, CO2
DAIC4=((7.4E-8*((18.02*2.6)**.5)*((TBC-32.)/1.8+273.))/
1 ((96.2**.6)*NB))*(3.875))
DACO2=((6.577E-8*((TBC-32.)/1.8+273.))/NB)*3.875
DAH2S=((5.403E-8*((TBC-32.)/1.8+273.))/NB)*3.875
C REDUCED PRESSURE, TEMP. CO2,H2S
PPCO2=TOTALP-VPIC4-VPH2O
IF(PPCO2.LE.0.)PPCO2=TOTALP*1.0E-6
CO2ATM=PPCO2/14.7
PRCO2=CO2ATM/72.9
TRCO2=((TBC-32.)/1.8+273.)/304.2
C CALCULATE REDUCED PROPERTIES FOR H2S
H2SATM=PPCO2/14.7
PRH2S=H2SATM/54.
TRH2S=((TFC-32.)/1.8+273.)/(100.3+273.)
RETURN
END
C SUBROUTINE FOR TRIAL ERROR CAL. IC4 EXIT CCNC.

```

```

SUBROUTINE CC(KIC4,A,VOLINC,VELOB,DIA,TEQIC4,BEQIC4,CC TOP)
REAL IC4BOT,KIC4
COMMON RHOB,NB,RHCF,SIGMAF,VPIC4,HIC4,MCLCC2,HG,HH2S,HCO2,
1 VPH2O,DAIC4,DACC2,DAH2S,PPCO2,CO2ATM,PRCC2,TRCC2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2BO,XH2SBO,NBOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBDIA,TBLGTH,RPD,TCO2ND,PRCO2C,YCC2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCCO2,GCCO2(100)
COMMON NVERT,NHORIZ

```

```

AIC4(X)=(KIC4*A*VOLINC)/(VELOB*3.142*((DIA/12.)**2.)/4.)
1 *(((BEQIC4-X)-(TEQIC4-CC TOP))/ALOG((BEQIC4-X)/(TEQIC4-
1 CCTOP)))+CCTOP
D=CC TOP
B=BEQIC4

```

C SET UP LOOP FOR TRIAL AND ERROR CALCULATION

```

CM=(D+B)/2.
XR=AIC4(CM)
DO 50 I=1,100
XL=AIC4(XR)
IF(AIC4(XR)-AIC4(XL))20,30,40
20 XR=XL
40 XR=XL
50 CONTINUE
30 IC4BOT=XL
RETURN
END

```

C SUBROUTINE FOR CAL. CO2 IN BRINE

```

SUBROUTINE CO2(KCO2,A,VOLINC,VELOB,DIA,TEQCO2,BEQCO2,CO2TOP
1,TCC2,ME,MB)
REAL KCO2
COMMON RHOB,NB,RHCF,SIGMAF,VPIC4,HIC4,MCLCC2,HG,HH2S,HCO2,
1 VPH2O,DAIC4,DACC2,DAH2S,PPCO2,CO2ATM,PRCC2,TRCC2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2BO,XH2SBO,NBOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBDIA,TBLGTH,RPD,TCO2ND,PRCO2C,YCC2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCCO2,GCCO2(100)
COMMON NVERT,NHORIZ
ACO2(X)=(KCO2*A*VOLINC)/(VELOB*3.142*((DIA/12.)**2.)/4.)*
1 (((BEQCO2-X)-(TEQCO2-CO2TOP))/ALOG((BEQCO2-X)/(TEQCO2-CO2TOP)))
1 +CO2 TOP
BL=CO2 TOP
BR=BEQCO2
CB=(BL+BR)/2.
XB=ACO2(CB)
DO 50 I=1,10
XL1=ACO2(XB)
IF(ACO2(XB)-ACO2(XL1))20,30,40
20 XB=XL1
40 XB=XL1
50 CONTINUE
30 CO2BOT=XL1
RETURN
END

```

C SUBROUTINE FOR CAL. H2S IN BRINE

```

SUBROUTINE H2S(KH2S,A,VOLINC,VELCB,DIA,TEQH2S,BEQH2S,H2STOP
1,TH2S,MF,MB)
REAL KH2S
COMMON RHOB,NB,RHOF,SIGMAF,VPIC4,FIC4,MOLCO2,HG,HH2S,HCO2,
1 VP420,DAIC4,DACC2,DAH2S,PPCO2,CC2ATM,PRCO2,TPCC2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CC2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2B0,XH2SB0,NBOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBDIA,TBLGTH,RPD,TCOND,PRCOND,YCC2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCCND,GCCND(100)
COMMON NVERT,NHORIZ
BH2S(X)=(KH2S*A*VOLINC)/(VELCB*3.142*((DIA/12.)*2.)/4.)*
1 (((BEQH2S-X)-(TEQH2S-H2STOP))/ALCG((BEQH2S-X)/(TEQH2S-H2STOP)))
1 +H2STOP
DL=H2STOP
DR=BEQH2S
DB=(DL+DR)/2.
XH=BH2S(DB)
DO 50 I=1,100
XL2=BH2S(XH)
IF(BH2S(XH)-BH2S(XL2))20,30,40
20 XH=XL2
40 XH=XL2
50 CONTINUE
30 H2SBOT=XL2
XCO2IO=XCO
XH2SIO=XHS
RETURN
END

```

SUBROUTINE GAS

C SUBROUTINE GAS CALCULATES HENRYS CONSTANTS FOR CO2 AND H2S IN BRINE

```

COMMON RHOB,NB,RHOF,SIGMAF,VPIC4,FIC4,MOLCO2,HG,HH2S,HCO2,
1 VPH20,DAIC4,DACC2,DAH2S,PPCO2,CO2ATM,PRCC2,TPCC2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCC2B0,XH2SB0,NBOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBDIA,TBLGTH,RPD,TCOND,PRCOND,YCC2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCCND,GCCND(100)
COMMON NVERT,NHORIZ

```

C CALCULATE BRINE DENSITY AND SALT MOLARITY

```

N=NBOIL
K=N+1
DO 100 I=1,K
IF(T(I).LT.100.) RHOW=1.0
IF(T(I).GE.100.) RHOW=1.04-T(I)*4.E-04
RHOB=RHOW+0.0073*WPB
SALTM=(WPB/585.)*RHOB

```

C HCO2W IS HENRYS CONSTANT FOR CO2 IN WATER

```

IF(T(I).LT.212.) HCO2W=-77.+24.4*T(I)
IF(T(I).GE.212.) HCO2W=1560.+16.7*T(I)
IF(T(I).GE.300.) HCO2W=6600.
IF(T(I).GE.390.) HCO2W=12200.-14.4*T(I)
IF(T(I).LT.212.) HH2SW=123.+6.87*T(I)
IF(T(I).GE.212.) HH2SW=921.+3.11*T(I)

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```

      IF(T(I).GE.302.) HH2SW=1390.+1.55*T(I)
      IF(T(I).GE.392.) HH2SW=2435.-1.11*T(I)
      IF(T(I).GE.482.) HH2SW=3400.-3.11*T(I)
C     SCCO2 IS SALTING COEFFICIENT
C     HCO23 IS HENRYS CONSTANT FOR CO2 IN BRINE
      IF(T(I).LT.250.) SCCO2=0.078
      IF(T(I).GE.250.) SCCO2=0.073
      IF(T(I).GE.350.) SCCO2=0.088
      IF(T(I).GE.450.) SCCO2=0.124
      IF(T(I).GE.530.) SCCO2=0.168
      SOH2S=0.10
      HCO23(I)=HCO2W*(10.**((SCCO2*SALTM))*14.7
100 HH2SB(I)=HH2SW*(10.**((SOH2S*SALTM))*14.7
      RETURN
      END
      SUBROUTINE CNDSER
      COMMON RHO8,NB,RHOF,SIGMAF,VPIC4,HIC4,MOLCC2,HG,HH2S,HCO2,
1    VPH2O,DAIC4,DACC2,DAH2S,PPCO2,CO2ATM,PRCO2,TRCO2,H2SATM,
1    PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
      COMMON T(100),WPR,HCO2B(100),HH2SB(100),XCO2BC,XH2SBO,NBOIL
      COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1    (100),XCO2I(100),XH2SI(100)
      COMMON TBIN,FLBR,FLIC4
      COMMON PRBOIL,TSAT
      COMMON TBDIA,TBLGTH,RPD,TCOND,PRCOND,YCC2C(100),YH2SC(100),
1    XCO2C(100),XH2SC(100),CONDL(100),NCCND,GCCND(100)
      COMMON NVERT,NHORIZ
C     TCOND IS CONDENSATION TEMPERATURE FOR ISOBUTANE
      TBDIA=TBDIA/12.
      VPH2O=10.**((0.0251*((TCOND-80.))*0.83)-C.2951)
      VPIC4=26.356*EXP(0.0112*TCOND)
      PRCOND=VPH2O+VPIC4
C     CONDENSER IS HORIZONTAL-TUBE SURFACE TYPE
C     CALCULATE SHELL-SIDE REYNOLDS NUMBER
      N=NBOIL
      SPACE=TBDIA*RPD-TBDIA
      GVAPOR=FLIC4/(SPACE*TBLGTH*NHORIZ)
C     VISCOSITY OF ISOBUTANE VAPOR IS VISC4V
      VISC4V=0.0067+1.2E-5*TCOND
      REVAP=TBDIA*GVAPOR/VISC4V
C     VAPOR-SIDE COEFFICIENTS FOR STAGGERED TUBES
C     DEFAULT FOR PITCH/DIA RATIO (RPD) = 3
      ANU=21.7+0.00583*REVAP
      IF(PDR.EQ.2.) ANU=ANU*0.937
      IF(PDR.EQ.1.5) ANU=ANU*0.875
      TKVAP=0.0082+0.000022*TCOND
      HVAP=ANU*TKVAP/TBDIA
      DIFVAP=0.0107*(((TCOND+460.)/590.))**1.5)
      RHOVAP=1.20*PRCOND*590./((110.*(TCOND+460.))
C     CALCULATE VAPOR-SIDE MASS TRANSFER COEFFICIENT
      SCVAP=VISC4V/(RHOVAP*DIFVAP)
      CPVAP=0.357+0.0006*TCOND
C     SCVAP IS SCHMIDT NUMBER. PRVAP IS PRANDTL NUMBER.
      PRVAP=VISC4V*CPVAP/TKVAP
      AKY=(HVAP/(CPVAP*58.))*((PRVAP/SCVAP)**0.667)
C     CALCULATE LIQUID-SIDE COEFFICIENTS FOR TUBES
      GROUP=4.21E8-1.23E6*TCOND
      EMDOT=FLIC4/NHORIZ
      HLIQ=0.951*(GROUP*TBLGTH*10./EMDOT)**0.333
      VISC4L=(0.115-0.0025*((TCOND-32.)/1.8-50.))*2.42
      PRLIQ=VISC4L*0.667/0.0552

```

```

DIFLIQ=0.00065*((TCOND+460.)/590.)*(0.3/VISC4L)
SCLIQ=VISC4L/(32.*DIFLIQ)
AKX=(HLIQ/38.7)*((PRLIQ/SCLIQ)*0.667)
IF(TCOND.LT.160.) HCO2I=(40.+0.35*TCOND)*14.7
HH2SI=(7.2+0.214*TCOND)*14.7
EQKCO2=HCO2I/PRCOND
EQKH2S=HH2SI/PRCOND
CKYCO2=1./((1./AKY+EQKCO2)/AKX)
CKYH2S=1./((1./AKY+EQKH2S)/AKX)
AREA=(INVERT*NHORIZ/NCOND)*3.1416*TBDIA*TBLGTH
AKYCO2=CKYCO2*AREA
AKYH2S=CKYH2S*AREA
CONDL(1)=(FLIC4/58.)/NCOND
GCCND(1)=(FLIC4/58.)-CONDL(1)
GOYO=YCO2(N+1)*FLIC4/58.
GOYHS=YH2S(N+1)*FLIC4/58.
YCO2C(1)=YCO2(N+1)*(FLIC4/(58.*AKYCO2)-0.5+EQKCO2*FLIC4/(58.*
1CONDL(1)))/(0.5+EQKCO2*GCCND(1)/CONDL(1)+GCCND(1)/AKYCO2)
YH2SC(1)=YH2S(N+1)*(FLIC4/(58.*AKYH2S)-0.5+EQKH2S*FLIC4/(58.*
1CONDL(1)))/(0.5+EQKH2S*GCCND(1)/CONDL(1)+GCCND(1)/AKYH2S)
XCO2C(1)=(GOYO-YCO2C(1)*GCCND(1))/CONDL(1)
IF(XCO2C(1).LT.1.E-6)XCO2C(1)=1.E-6
XH2SC(1)=(GOYHS-YH2SC(1)*GCCND(1))/CONDL(1)
IF(XH2SC(1).LT.0.)XH2SC(1)=1.E-6
DO 1000 I=2,NCOND
CONDL(I)=I*CONDL(1)
GCCND(I)=(FLIC4/58.)-CONDL(I)
GCCND(NCOND)=GCCND(1)/1000.
CONDL(NCOND)=(FLIC4/58.0)-GCCND(NCOND)
REVAP=TBDIA*58.*GCCND(I-1)/(VISC4V*SPACE*TBLGTH*NHORIZ)
ANU=21.7+0.00583*REVAP
IF(PDR.EQ.2.) ANU=ANU*0.937
IF(PDR.EQ.1.5) ANU=ANU*0.875
HVAP=ANU*TKVAP/TBDIA
AKY=(HVAP/(CPVAP*58.))*((PRVAP/SCVAP)*0.667)
CKYCO2=1./((1./AKY+EQKCO2)/AKX)
CKYH2S=1./((1./AKY+EQKH2S)/AKX)
AKYCO2=CKYCO2*AREA
AKYH2S=CKYH2S*AREA
YCO2C(I)=(GCCND(I-1)*YCO2C(I-1)/AKYCO2-YCO2C(I-1)/2.+EQKCO2*
1XCO2C(I-1)/2.+EQKCO2*GOYO/(2.*CONDL(I)))/(GCCND(I)/AKYCO2+0.5+
1EQKCO2*GCCND(I)/(2.*CONDL(I)))
XCO2C(I)=(GOYO-YCO2C(I)*GCCND(I))/CONDL(I)
YH2SC(I)=(GCCND(I-1)*YH2SC(I-1)/AKYH2S-YH2SC(I-1)/2.+EQKH2S*
1XH2SC(I-1)/2.+EQKH2S*GOYHS/(2.*CONDL(I)))/(GCCND(I)/AKYH2S+0.5+
1EQKH2S*GCCND(I)/(2.*CONDL(I)))
1000 XH2SC(I)=(GOYHS-YH2SC(I)*GCCND(I))/CONDL(I)
RETURN
END
SUBROUTINE OUTPUT
COMMON RHOB,NB,RHOF,SIGMAF,VPIC4,HIC4,MOLCC2,HG,HH2S,HCO2,
1 VPH2O,DAIC4,DACO2,DAH2S,PPCO2,CO2ATM,PRCC2,TRCO2,H2SATM,
1PRH2S,TRH2S,IC4BOT,CO2BOT,H2SBOT
COMMON T(100),WPB,HCO2B(100),HH2SB(100),XCO2BO,XH2SBO,ABOIL
COMMON TBOIL(100),XISO(100),XCO2B(100),XH2SB(100),YCO2(100),YH2S
1(100),XCO2I(100),XH2SI(100)
COMMON TBIN,FLBR,FLIC4
COMMON PRBOIL,TSAT
COMMON TBDIA,TBLGTH,RPD,TCOND,PRCOND,YCO2C(100),YH2SC(100),
1XCO2C(100),XH2SC(100),CONDL(100),NCOND,GCCND(100)
COMMON INVERT,NHORIZ

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```

N=NBOIL
WRITE(6,300) WPB
300 FORMAT(1H0,'WEIGHT PERCENT SALT IN BRINE=',5X,F4.1)
WRITE(6,310) XCO2BO
310 FORMAT(1H0,'MOLE FRACTION CO2 IN INLET BRINE=',5X,F10.6)
WRITE(6,320) XH2SBO
320 FORMAT(1H0,'MOLE FRACTION H2S IN INLET BRINE=',5X,F10.6)
WRITE(6,330) TBIN
330 FORMAT(1H0,'BRINE INLET TEMPERATURE(F)',5X,F8.1)
WRITE(6,331) TSAT
331 FORMAT(1H0,'BOILER SATURATION TEMP(F) = ',5X,F8.1)
WRITE(6,332) PRBOIL
332 FORMAT(1H0,'BOILER PRESSURE(PSIA) = ',5X,F8.1)
WRITE(6,340) N
340 FORMAT(1H0,'NUMBER OF INCREMENTS IN BOILER=',I4)
WRITE(6,350)
350 FORMAT(1H0,'BRINE TEMP(F)      BRINE ISC      BRINE CO2      BRINE
1H2S      VAPOR CO2      VAPOR H2S      ISC CO2      ISC H2S')
C XCO2I(N),XCO2I(N+1),XH2SI(N),XH2SI(N+1) DO NOT EXIST-NO LIQUID PHASE
XCO2I(N)=99.
XCO2I(N+1)=99.
XH2SI(N)=99.
XH2SI(N+1)=99.
K=N+1
DO 360 I=1,K
WRITE(6,370) TBOIL(I),XISO(I),XCO2B(I),XH2SP(I),YCC2(I),YH2S(I),
1XCO2I(I),XH2SI(I)
370 FORMAT(1H0,3X,F6.1,10X,F8.6,6X,F8.6,6X,F8.6,6X,F8.6,6X,F8.6,6X,
1F8.6,6X,F8.6)
360 CONTINUE
TBDIA=TB DIA*12.0
WRITE(6,371) TBDIA
371 FORMAT(1H0,'CONDENSER TUBE DIA(INCHES) = ',5X,F8.6)
WRITE(6,372) TBLGTH
372 FORMAT(1H0,'CONDENSER TUBE LENGTH(FEET) = ',5X,F8.1)
WRITE(6,373) RPD
373 FORMAT(1H0,'PITCH-TO-DIAMETER RATIO = ',5X,F6.1)
NTUBE=NVERT*NHORIZ
WRITE(6,379) NTUBE
379 FORMAT(1H0,'NUMBER OF TUBES=',5X,I8)
WRITE(6,374) TCOND
374 FORMAT(1H0,'CONDENSATE TEMP(F) = ',5X,F8.1)
WRITE(6,375) PRCOND
375 FORMAT(1H0,'CONDENSER PRESSURE (PSIA) = ',5X,F8.1)
WRITE(6,376)
376 FORMAT(1H0,'COND(MOLS/HR)      VAPOR CO2      VAPOR H2S      COND CO2
1      COND H2S      VAPOR(MOLS/HR)')
DO 377 I=1,NCOND
WRITE(6,378) CONDL(I),YCO2C(I),YH2SC(I),XCO2C(I),XH2SC(I),GCOND(I)
378 FORMAT(1H0,3X,F6.2,10X,F8.6,6X,F8.6,6X,F8.6,6X,F8.6,6X,F6.2)
377 CONTINUE
RETURN
END

```

APPENDIX 2

COMPUTER OUTPUT FOR
SAMPLE CASE

WEIGHT PERCENT SALT IN BRINE= 0.0

MOLE FRACTION CO2 IN INLET BRINE= 0.000200

MOLE FRACTION H2S IN INLET BRINE= 0.000100

BRINE INLET TEMPERATURE(F) 336.0

BOILER SATURATION TEMP(F) = 220.0

BOILER PRESSURE(PSIA) = 333.3

NUMBER OF INCREMENTS IN BOILER= 10

BRINE TEMP(F)	BRINE ISO	BRINE CO2	BRINE H2S	VAPOR CO2	VAPOR H2S	ISO CO2	ISO H2S
254.5	0.000009	0.000174	0.000104	0.000384	0.000443	0.000090	0.000185
261.6	0.000008	0.000179	0.000107	0.000329	0.000379	0.000077	0.000158
268.8	0.000007	0.000183	0.000110	0.000291	0.000330	0.000068	0.000138
276.1	0.000006	0.000187	0.000112	0.000264	0.000292	0.000062	0.000122
283.3	0.000005	0.000191	0.000114	0.000242	0.000262	0.000057	0.000109
290.5	0.000004	0.000194	0.000116	0.000225	0.000237	0.000053	0.000099
297.8	0.000003	0.000198	0.000118	0.000211	0.000217	0.000050	0.000090
305.0	0.000003	0.000201	0.000120	0.000200	0.000200	0.000047	0.000083
312.3	0.000002	0.000204	0.000122	0.000190	0.000185	0.000044	0.000077
319.5	0.000001	0.000206	0.000123	0.000181	0.000173	*****	*****
326.7	0.000001	0.000209	0.000125	0.000195	0.000181	*****	*****

SAMPLE BOILER OUTPUT

CONDENSER TUBE DIA(INCHES) = 1.000000
 CONDENSER TUBE LENGTH(FEET) = 28.0
 PITCH-TO-DIAMETER RATIO = 3.0
 NUMBER OF TUBES= 100
 CONDENSATE TEMP(F) = 130.0
 CONDENSER PRESSURE (PSIA) = 115.3

COND(MOLS/HR)	VAPOR CO2	VAPOR H2S	COND CO2	COND H2S	VAPOR(MOLS/HR)
3.09	0.000610	0.000186	0.000051	0.000035	27.82
6.18	0.000678	0.000202	0.000060	0.000040	24.73
9.27	0.000764	0.000223	0.000064	0.000043	21.64
12.37	0.000874	0.000250	0.000075	0.000048	18.55
15.46	0.001023	0.000285	0.000085	0.000053	15.46
18.55	0.001233	0.000332	0.000102	0.000061	12.37
21.64	0.001555	0.000400	0.000125	0.000071	9.27
24.73	0.002116	0.000507	0.000164	0.000085	6.18
27.82	0.003364	0.000710	0.000242	0.000109	3.09
30.89	0.009268	0.001342	0.000546	0.000168	0.03

SAMPLE CONDENSER OUTPUT

TEMP. OF ORGANIC AT TOP ZONE A= 195.41
 TOTAL TOWER HT.= 111.05
 # MOLE/CU FT. H₂S, CO₂ IN IC₄ TO BOILER= 8.147067E-05 2.333200E-04
 CONCENTRATIONS OF IC₄, CO₂, H₂S IN BRINE (MOL FRACT)

TR(F)	TF(F)	IC ₄	CO ₂	H ₂ S
2.2508E+02	2.2035E+02	1.0032E-05	1.6184E-04	9.9402E-05
2.2406E+02	2.1925E+02	1.1575E-05	1.4975E-04	9.6443E-05
2.2301E+02	2.1814E+02	1.3053E-05	1.3868E-04	9.3600E-05
2.2195E+02	2.1701E+02	1.4467E-05	1.2853E-04	9.0869E-05
2.2088E+02	2.1586E+02	1.5822E-05	1.1921E-04	8.8244E-05
2.1979E+02	2.1469E+02	1.7118E-05	1.1066E-04	8.5720E-05
2.1868E+02	2.1351E+02	1.8360E-05	1.0279E-04	8.3293E-05
2.1755E+02	2.1231E+02	1.9548E-05	9.5552E-05	8.0957E-05
2.1641E+02	2.1109E+02	2.0684E-05	8.8891E-05	7.8710E-05
2.1525E+02	2.0985E+02	2.1772E-05	8.2755E-05	7.6547E-05
2.1407E+02	2.0860E+02	2.2813E-05	7.7099E-05	7.4464E-05
2.1287E+02	2.0734E+02	2.3809E-05	7.1881E-05	7.2458E-05
2.1165E+02	2.0605E+02	2.4761E-05	6.7064E-05	7.0525E-05
2.1042E+02	2.0475E+02	2.5672E-05	6.2614E-05	6.8662E-05
2.0916E+02	2.0344E+02	2.6543E-05	5.8500E-05	6.6866E-05
2.0789E+02	2.0212E+02	2.7375E-05	5.4695E-05	6.5135E-05
2.0659E+02	2.0078E+02	2.8171E-05	5.1172E-05	6.3465E-05
2.0528E+02	1.9944E+02	2.8932E-05	4.7910E-05	6.1854E-05
2.0394E+02	1.9809E+02	2.9660E-05	4.4886E-05	6.0300E-05
2.0119E+02	1.9541E+02	3.1021E-05	3.9478E-05	5.7351E-05

SAMPLE PREHEATER OUTPUT