

To the Graduate Council:

I am submitting herewith a thesis written by Thanh H. Nguyen entitled "A Computational Code for the Heating of Gas-Liquid Mixtures in a Horizontal Tube." I have examined the final copy of this thesis for form and content and recommend that it be accepted in partial fulfillment of the requirements for the degree of Master of Science, with a major in Chemical Engineering.

George C. Frazier
George C. Frazier, Major Professor

We have read this thesis
and recommend its acceptance:

Fred Weber

John M. Holmes

Accepted for the Council:

Comminkel
The Graduate School

A COMPUTATIONAL CODE FOR THE
HEATING OF GAS-LIQUID MIXTURES
IN A HORIZONTAL TUBE

A Thesis
Presented for the
Master of Science
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ABSTRACT

A simple steady state heat transfer model and computer code have been developed for designing tubular heaters for two-phase (gas-liquid) multicomponent mixtures. Principle assumptions on which the model is based include: approach to local equilibrium between the phases, both the gas and liquid phases are ideal, and the liquid phase is Newtonian. The code utilizes correlations available from the literature for the two-phase pressure drop, the liquid holdup fraction, and the two-phase heat transfer coefficient. The assumption that the liquid is Newtonian can be readily relaxed as correlations become available for various classes of non-Newtonian fluids.

A tentative validation of the model and code has been made by comparing computed results for the air-water system with a limited amount of data found in the literature. Best agreement between computed and experimental results were obtained using the Johnson - Abou Sabe correlations for pressure drop and liquid holdup and by using a two-phase heat transfer coefficient as computed by the Davis - David correlation, but increased by 20%. Heat transfer coefficients in this range are still within the experimental data scatter.

For the air-water system, the calculated pressure drops agreed within 5.7% of the experimental values and the outlet mixture temperatures agreed within one degree Fahrenheit, except for one case, where the difference between calculation and experiment was 17.6°F.

Unfortunately, it is not possible to validate the model and code for two-phase multicomponent systems at this time because of the lack of experimental data. However, an illustrative example for such systems is provided for a gas-binary liquid system (heptane and decane), in order to demonstrate the use of the code and to show that iterative procedures converge properly for tight convergence criteria.

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1. INTRODUCTION

1.1 Statement of the Problem

Fired heaters for the preheat of multiphase, gas-oil-coal feed mixture to a direct coal liquefaction process are important elements in the process train. Heaters of this type have been in use in petroleum refineries for decades, where they are sometimes subject to fouling due to coke and coke-like deposits. The formation of coke in the tube of the fired heater for coal-oil slurries is an even greater problem. Currently, reliable methods have not been proposed to prevent this problem since the causes of coking are not completely understood and the coke formation and deposition are often unpredictable. However, the heating rate of coal liquid slurries, coal type, solvent properties, slurry concentration, slurry residence time, etc. in the heater tube do appear to be factors in determining the extent to which coking occurs⁽¹⁾. Before the effect of these factors on coking propensity are examined, one must be concerned with flow pattern, interphase mass transfer, pressure, liquid holdup, and temperature profile of the mixture flowing in the heater tube. Hence, a study of heat transfer and fluid flow in the heater must be accomplished in order to focus on the further understanding of the coke formation problem. For a more general consideration, methods to compute flow characteristics such as liquid holdup, pressure drop, temperature increase,

etc. of nonisothermal, simultaneous gas-liquid flow in a tube are of interest.

In addition, heat transfer to two-phase flow is very important in several other chemical and heat engineering processes. A large number of investigations have been conducted since the 1930's. Since many different flow patterns may exist in two-phase gas-liquid flow (e.g., bubble flow, annular flow, stratified flow, slug flow, etc.)⁽²⁾, most investigations are restricted within some specified regimes of flow. For this reason, a method which can be broadly applied to several different gas-liquid flow systems is desired. Therefore, to provide more fundamental information for the further study of coke formation and from the requisite of a more general method to find the correlations of heat transfer to two-phase gas-liquid flow, this work is undertaken in order to develop a calculational procedure that fits for these essential needs.

1.2 Objective

The objective of this work is to develop a computer code for use in the design of a tubular heater for a two-phase, multicomponent, gas-liquid system. The mass and energy conservation principles form the basis of this code. However, in view of the fact that heat transfer to flows of this type is complex and depends on factors such as the type of flow (flow regime), pressure drop, and the corresponding gas and liquid holdup, numerous empirical correlation of these

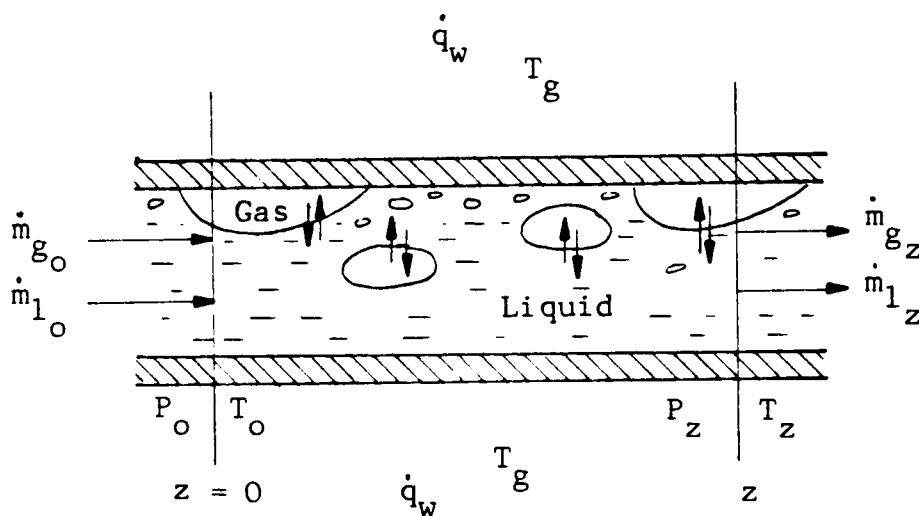
factors are required in order to provide quantitative results.

The code can be applied to the systems which have:

- High gas mass fraction (≥ 0.1) but no burnout (only gas and vapor)
- Four types of flow regime:
 - tt - Both gas and liquid phases are turbulent
 - tv - Liquid phase is turbulent and gas phase is viscous
 - vt - Liquid phase is viscous and gas phase is turbulent
 - vv - Both gas and liquid phases are viscous
- Several different flow patterns such as wave, slug, plug, annular, bubble, stratified, and dispersed flow
- No chemical reaction occurs in the system
- Horizontal, co-current gas-liquid flow

1.3 Introduction to the Problem

Consider a two-phase gas-liquid system flowing in a tube as depicted in Figure 1.1. Let the tube be oriented in the z direction (from $z=0$ to $z=1$) where $\dot{m}_{g_0}$ and $\dot{m}_{l_0}$ are the feed rates of gas and liquid streams at $z=0$, and T_0 and P_0 are temperature and total pressure of these inlet streams, respectively. If the heat flux rate to the system or temperature of the heating medium (e.g., the tube may be heated in a gas-fired furnace) are specified, the problem now is how



- $\dot{m}_g, \dot{m}_l$: Gas and liquid mass flow rates, lb/hr
 P : Total pressure, psia
 T : Fluid temperature, $^{\circ}\text{F}$
 T_g : Heating medium temperature, $^{\circ}\text{F}$
 $\dot{q}_w$: Heat flux rate to the system, $\text{BTU}/\text{ft}^2\text{-hr.}$

Figure 1.1. Two-Phase Gas-Liquid Flow in a Section of the Horizontal Tube

to determine the temperature, pressure, and flow rates of the the outlet streams. Briefly, given the compositions, flow rates, pressure, temperature of the streams at the inlet, and heating medium temperature, how long must the tube be in order to raise the mixture to a desired temperature, and what will be the compositions of the gas and liquid phases?

Typically, in the fired coal slurry preheater, slurry feed rate (oil and coal particles) ranges between 800 to 10,000 lb/hr (or superficial velocity of the slurry is from 0.4 to 3.0 ft/sec), the gas mass fraction ranges between 0.1 to 0.2, average total pressure is 1200 psi, and average outlet heating temperature is 800-1000°F. It is unclear as to which flow pattern or regimes may occur in a coal slurry heater tube, but due to the high viscosity of the liquid (40-60 cp), slug, plug and viscous flows are the most likely to occur.

The heating of a multiphase system in a tube is clearly a complicated process due largely to the range of complex hydrodynamic flow patterns which may exist. A simplified model of this situation is clearly needed which will facilitate the design of heaters of this type. Such a model is proposed in this study. Computational results based on this model are compared with the limited data in the literature.

2. LITERATURE BACKGROUND

2.1 Fired Heaters

In many chemical processes, heat is exchanged between the flowing fluids on different sides of the tube through the tube wall. In this study which is connected to the coke formation in the coal liquefaction process equipment, heat transfer through the tube wall to the flowing mixture is supplied by a fired heater. In the coal liquefaction plant, fired heaters are generally of the single or parallel helical-coil type. Thermocouples for measuring both the fluid and coil skin temperature are located along the coil. The flowing fluid consists of several phases, including gas, liquid and solid, (mineral matter as well as organic). Typical operating conditions of the preheater are summarized in Table 2.1. While flowing down the tube, chemical reactions, dissolution of coal particle, evaporation of the solvent, or temperature increase due to exothermic chemical reactions may take places. The coke deposition is observed increasingly toward the heater outlet. Operating conditions such as an increase in heat flux rate, larger residence time, low gas partial pressure and slow agitation in the flowing mixture tend to favor coke formation increase. Typical operational problems attribute to coke formation and subjects related to fired heater are discussed by Frazier, Lin, and Edwards.⁽¹⁾

Table 2.1. Typical Ranges of Operating Conditions of
Fired Preheater⁽¹⁾

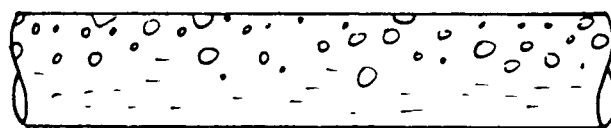
Coil Diameter	1 - 2 inches
Coil Length	400 - 700 ft
Temperature, °F	
Inlet	115 - 155
Outlet	770 - 885
Inlet Pressure, psia	
Total	1720 - 2410
H ₂ Partial Pressure	1460 - 2050
Wt % Coal in Feed Slurry	30 - 50
Flow Rates of Feed Streams, lb/hr	
Coal	246 - 560
Solvent (Oils)	600 - 11000
H ₂ Gas	100 - 570

2.2 Two-Phase Gas-Liquid Flow Correlations

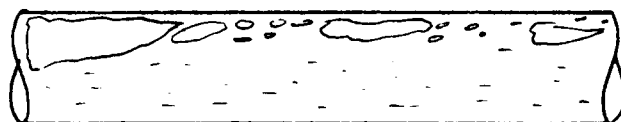
Two-phase gas-liquid flow is very important in many chemical processes and still not yet an area in which theoretical prediction of flow properties is completely developed. A large number of investigations with different ranges of operating conditions have been made. The following correlations for pressure drop, liquid holdup, and heat transfer for two-phase flow in tubes have been developed by numerous investigators.

2.2.1 Flow Patterns

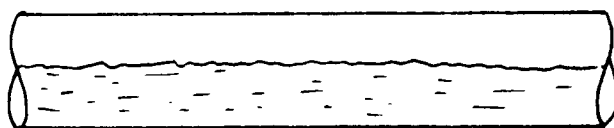
In order to understand how heat is transferred to a two-phase fluid flow system, an accurate description of the flow patterns is necessary. Figure 2.1 illustrates the patterns which may occur in a two-phase horizontal flow. In the bubble flow configuration, bubbles are moving in the upper part of the tube. When the gas flow rate is increased, bubbles coalesce and a plug flow takes place. For low liquid and gas flow rates, a stratified flow appears with a smooth interface. At higher gas rates, waves propagate along the interphase and can reach the top wall of the tube giving rise to a slug flow. Finally at high gas flow rates and low liquid flow rates (high gas mass fraction), an annular flow can exist with the thicker film in the lower part of the tube⁽²⁾. In these types of flow, the main complication is that gravitational force acts on the liquid phase causing it to be displaced towards the bottom of the tube.



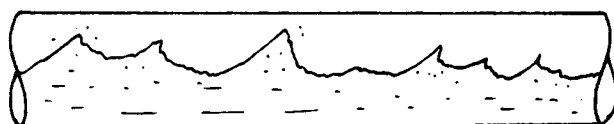
(a)



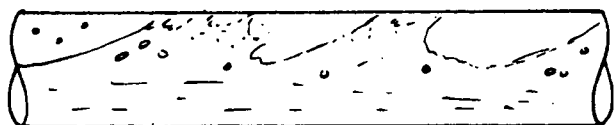
(b)



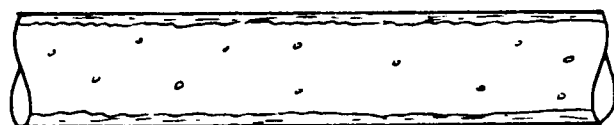
(c)



(d)



(e)



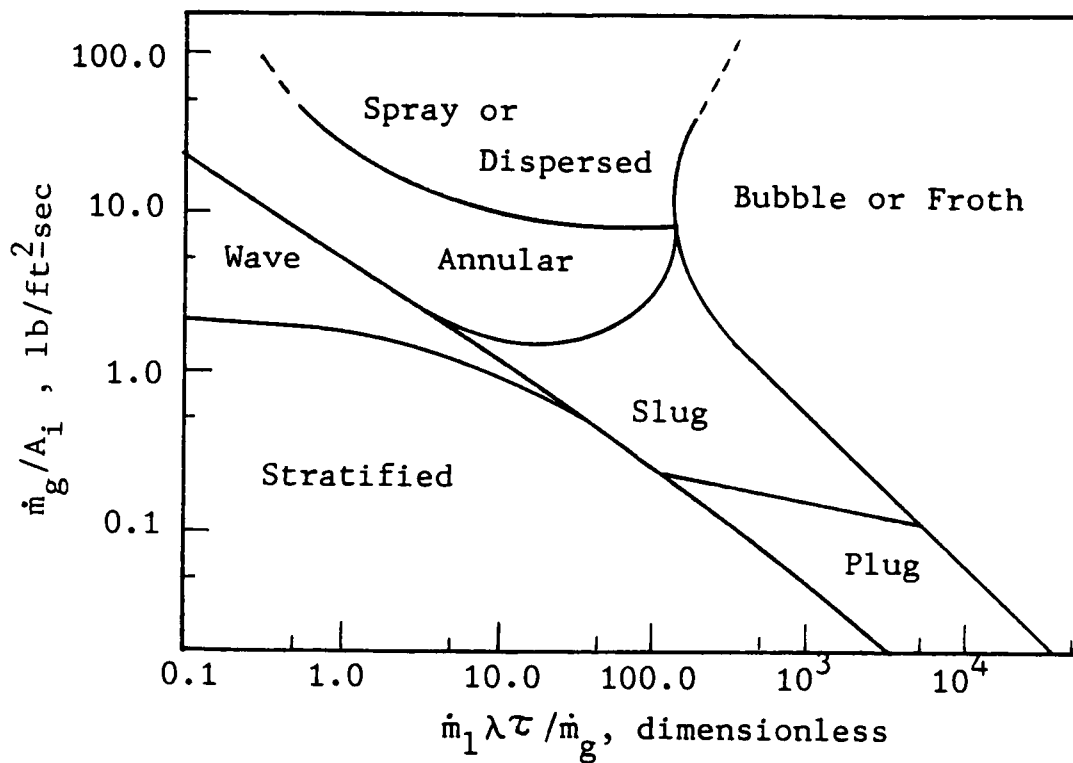
(f)

Figure 2.1. Flow Patterns in Horizontal Gas-Liquid Flows
 (a) Bubble Flow, (b) Plug Flow, (c) Strati-
 fied Flow, (d) Wavy Flow, (e) Slug Flow,
 (f) Annular Flow

More reliable flow pattern information is obtained by visual observation. On the other hand, flow pattern can be deduced by several different methods. The Baker diagram (1958)⁽³⁾ is modified and commonly used in the petroleum industries and condenser design. In this diagram, flow pattern of the mixture is a function of flow rates, density, viscosity, and liquid surface tension. Butterworth and Hewitt⁽⁴⁾ suggested an alternative map that introduced by Schicht (1969), which appears to give better results than the original Baker map. The Baker and Schicht maps and their correlations for flow patterns are presented in Figures 2.2 and 2.3, respectively. However, due to the complexity of the application of these maps in computer programming, another map, proposed by Mandhane (1974)⁽⁵⁾, turns out to be more favorable. In this latter diagram, flow patterns are determined by the superficial velocities of the gas and liquid streams. This diagram is valid for the wide ranges in tube diameter, gas and liquid densities, viscosities, and flow rates. The Mandhane map is illustrated in Figure 2.4.

2.2.2 Liquid Holdup Fraction and Pressure Drop

Four types of flow regime can exist with several patterns during the flow of gas and liquid in a horizontal tube. Liquid holdup, pressure drop, and heat transfer coefficient are dependent variables of these flow characteristics. Lockhart and Martinelli⁽⁶⁾ introduced some empirical correla-



A_i : Tube cross-sectional area, inside diameter basis, ft^2

$\dot{m}_g, \dot{m}_l$: Gas and liquid mass flow rates, lb/sec

λ : $(\rho_g/0.075)(\rho_l/62.3)^{0.5}$

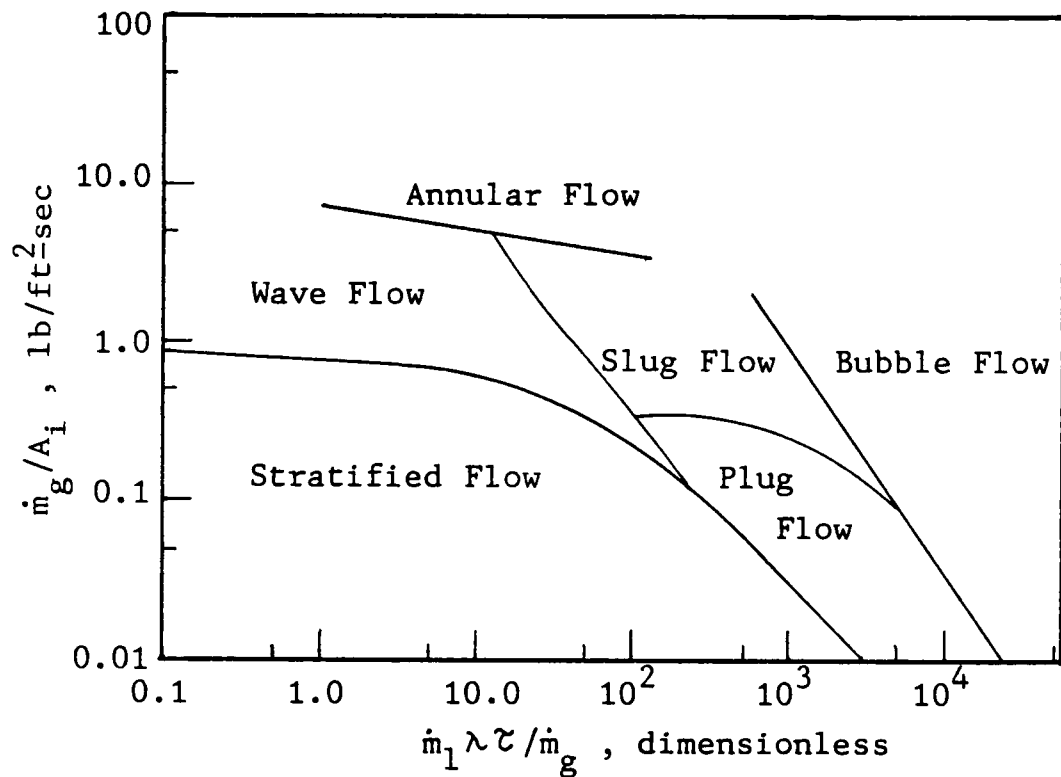
τ : $(73/\sigma)(\mu_l(62.3/\rho_l)^2)^{1/3}$

ρ_g, ρ_l : Gas and liquid densities, lb/ft^3

μ_l : Liquid viscosity, cp

σ : Liquid surface tension, dynes/cm

Figure 2.2. Flow Pattern Diagram Proposed by Baker⁽³⁾



A_i : Tube cross-sectional area, inside diameter basis, ft^2

$\dot{m}_g, \dot{m}_l$: Mass flow rates of gas and liquid phases, lb/sec

λ : $(\rho_g/0.075)(\rho_l/62.3)^{0.5}$

τ : $(73/\sigma)(\mu_l(62.3/\rho_l)^2)^{1/3}$

ρ_g, ρ_l : Gas and liquid densities, lb/ft^3

μ_l : Liquid Viscosity, cp

σ : Liquid surface tension, dynes/cm

Figure 2.3. Flow Pattern Diagram Proposed by Schicht⁽⁴⁾

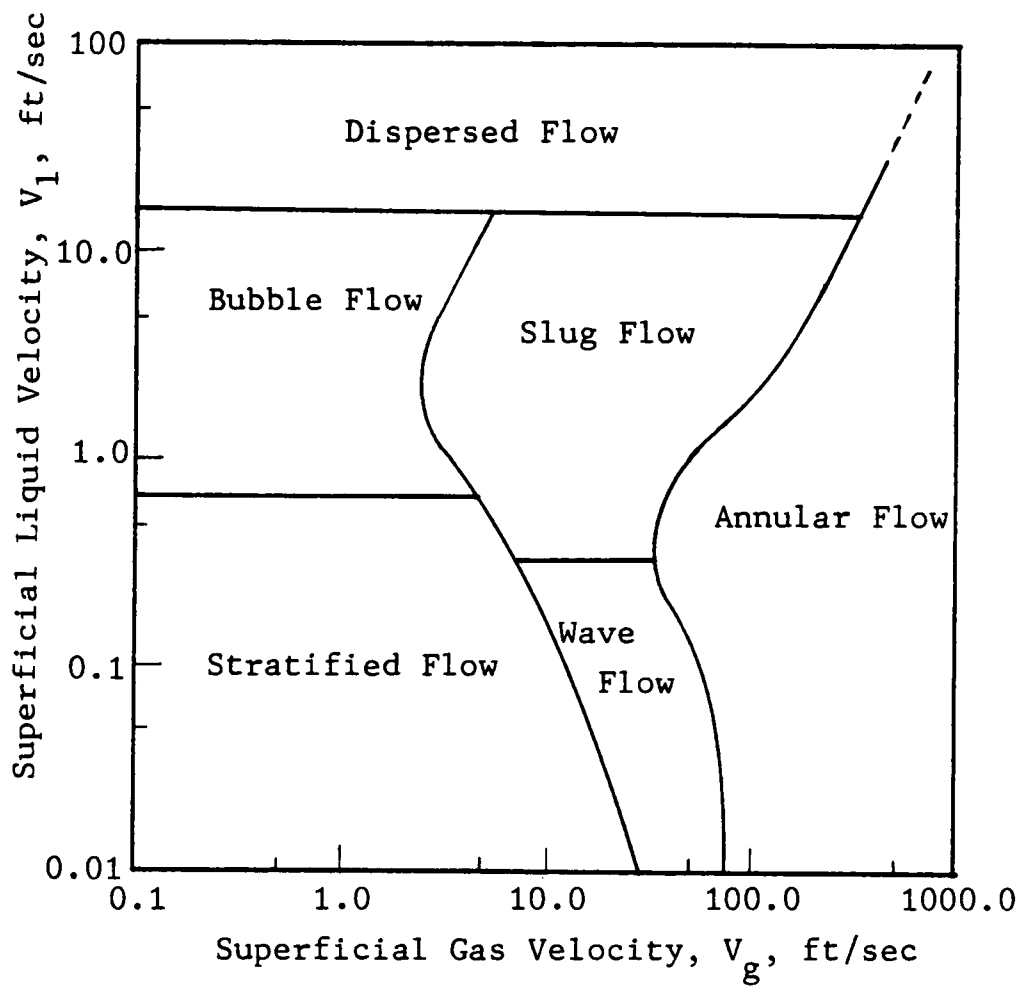


Figure 2.4. Flow Pattern Map Proposed by Mandhane⁽⁵⁾

tions to find liquid holdup and pressure drop based on the investigations which used air and liquids including water, benzene, kerosene, and various oils (these liquids are Newtonian; the experiments were isothermal). Consider a set of three experiments in the same tube as following:

- A two-phase mixture flow with average mass flow rate are $\dot{m}_g$ and $\dot{m}_l$. The frictional pressure drop in a section Δz is denoted by $(\Delta P/\Delta z)_{tp}$.
- A single gas phase flow in the same tube with mass flow rate equal to $\dot{m}_g$. The frictional pressure drop is then denoted by $(\Delta P/\Delta z)_g$.
- A single liquid phase flow in the same tube with mass flow rate equal to $\dot{m}_l$. The frictional pressure drop is then denoted by $(\Delta P/\Delta z)_l$.

The experiments provide the basic data, along with separated measurements of the holdup fraction, which are used to develop the two-phase flow correlations, relating, for example, the two-phase flow pressure drops to the flow rates and pressure drop of the separate phase. The approach used by Lockhart-Martinelli was to define the two parameters ϕ_g and ϕ_l according to:

$$\phi_g^2 = \frac{(\Delta P/\Delta z)_{tp}}{(\Delta P/\Delta z)_g} \quad (2.1)$$

$$\phi_1^2 = \frac{(\Delta P / \Delta z)_{tp}}{(\Delta P / \Delta z)_1} \quad (2.2)$$

and to correlate the experimental data with the parameter defined according to:

$$\chi^2 = \frac{\phi_g^2}{\phi_1^2} = \frac{(\Delta P / \Delta z)_1}{(\Delta P / \Delta z)_g} \quad (2.3)$$

The correlations between ϕ_g and ϕ_1 versus χ are presented in Figure 2.5. These curves can be approximated by the equations⁽²⁾:

$$\phi_1^2 = 1 + \frac{\phi_c}{\chi} + \frac{1}{\chi^2} \quad (2.4)$$

$$\phi_g^2 = 1 + \phi_c + \chi^2 \quad (2.5)$$

where ϕ_c is a constant corresponding to the flow regimes.

With:

$$\phi_c = 20 \text{ for turbulent-turbulent flow}$$

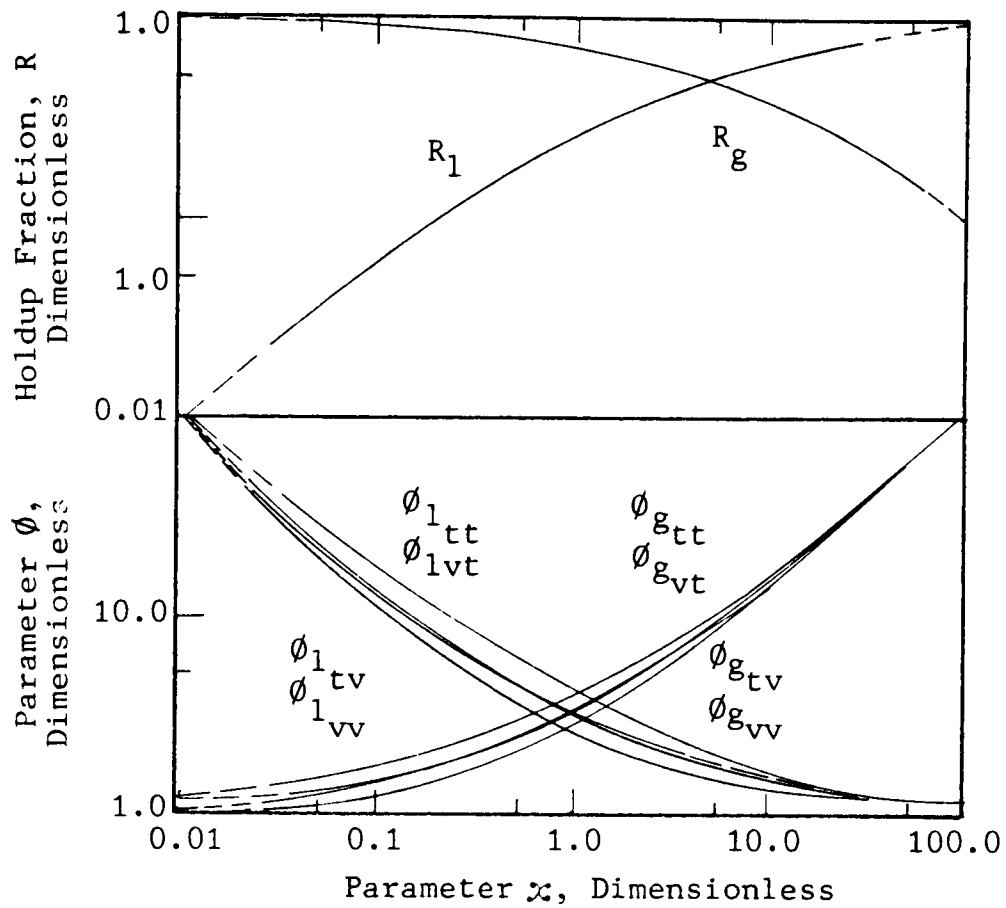
$$\phi_c = 12 \text{ for viscous-turbulent flow}$$

$$\phi_c = 10 \text{ for turbulent-viscous flow}$$

$$\phi_c = 5 \text{ for viscous-viscous flow}$$

Equations to calculate single-phase pressure drop corresponding to flow regimes are:

$$\left(\frac{\Delta P}{\Delta z}\right)_g = \frac{2 \left(\frac{4}{\pi}\right)^{2-m} C_g \mu_g \dot{m}_g^{2-m}}{D_i^{5-m} \rho_g g_c} \quad (2.6)$$



g : Gas Phase

l : Liquid Phase

tt : Turbulent-turbulent regime

vt : Viscous-turbulent regime

tv : Turbulent-viscous regime

vv : Viscous-viscous regime

Figure 2.5. Lockhart-Martinelli Holdup Fraction and Pressure Drop Correlations

$$\left(\frac{\Delta P}{\Delta z}\right)_1 = \frac{2\left(\frac{4}{\pi}\right)^{2-n} C_1 \mu_1 \dot{m}_1^{2-n}}{D_i^{5-n} \rho_1 g_c} \quad (2.7)$$

where C_g , C_1 , m , n : Constants corresponding to flow regimes, given in Table 2.2, dimensionless

g_c : Newton's law proportionality factor,
32.174 ft-lb/lbf-sec²

$\dot{m}_g$, $\dot{m}_1$: Mass flow rates of gas and liquid streams,
lb/hr

ρ_g , ρ_1 : Average gas and liquid density, lb/ft³

μ_g , μ_1 : Average gas and liquid viscosity,
lb/ft-sec

Two-phase pressure drop therefore is computed by Equations (2.1) or (2.2). These pressure drop correlations, according to Fried⁽⁸⁾, can be applied to nonisothermal flow.

Finally, the square root ratio of Equation (2.3) to compute parameter $\mathcal{X}$ is established for each regime as following:

turbulent-turbulent

$$\mathcal{X}_{tt}^2 = \left(\frac{\dot{m}_1}{\dot{m}_g}\right)^{1.8} \left(\frac{\rho_g}{\rho_1}\right) \left(\frac{\mu_1}{\mu_g}\right)^{0.2} \quad (2.8)$$

viscous-turbulent

$$\mathcal{X}_{vt}^2 = \left(\frac{0.046}{16}\right) Re_g^{-.8} \left(\frac{\dot{m}_1}{\dot{m}_g}\right) \left(\frac{\rho_g}{\rho_1}\right) \left(\frac{\mu_1}{\mu_g}\right) \quad (2.9)$$

Table 2.2 . Flow Regime Constants for the Lockhart-Martinelli (6)
Single-Phase Pressure Drop Correlations

Flow Regime Liquid-Gas	Constants			
	m	n	C_g	C_l
Turbulent-Turbulent	0.2	0.2	0.046	0.046
Viscous-Turbulent	0.2	1.0	0.046	16.0
Turbulent-Viscous	1.0	0.2	16.0	0.046
Viscous-Viscous	1.0	1.0	16.0	16.0

turbulent-viscous

$$x_{tv}^2 = \frac{0.046}{16} Re_1^{.8} \left(\frac{\dot{m}_1}{\dot{m}_g} \right) \left(\frac{\rho_g}{\rho_1} \right) \left(\frac{\mu_1}{\mu_g} \right) \quad (2.10)$$

viscous-viscous

$$x_{vv}^2 = \left(\frac{\dot{m}_1}{\dot{m}_g} \right) \left(\frac{\rho_g}{\rho_1} \right) \left(\frac{\mu_1}{\mu_g} \right) \quad (2.11)$$

where $Re_g = 4 \dot{m}_g / \pi D_i \mu_g$ and $Re_1 = 4 \dot{m}_1 / \pi D_i \mu_1$ are Reynold numbers of the gas and liquid phases.

Similar as pressure drop, Lockhart and Martinelli⁽⁶⁾ represented holdup fractions, R_1 , as a function of parameter

This correlation is illustrated in Figure 2.5 (page 16). Liquid holdup correlation in Figure 2.5 can be approximated by the follow equation⁽²⁾:

$$R_1 = \frac{x}{\sqrt{x^2 + 20x + 1}} \quad (2.12)$$

It should be noticed that all four flow regimes have the same correlation for holdup.

By using the same procedure to develop the correlations for holdup fraction and pressure drop of Lockhart and Martinelli, Johnson and Abou-Sabe⁽⁷⁾ developed correlations for turbulent-turbulent, nonisothermal, air-water flows. Contrary to the conclusion of Fried⁽⁸⁾, Johnson and Abou-Sabe showed that pressure drop in a nonisothermal flow system is higher than that in an isothermal system. Comparisions

between Lockhart-Martinelli and Johnson-Abou Sabe correlations are shown in Figure 2.6. In addition, these correlations can be improved if they are used for a more narrow range of conditions. For example, the correlations for pressure drop are slightly different when liquid flow rates change as described in Figure 2.7. Generally, Johnson-Abou Sabe correlations for liquid holdup and pressure drop as shown in Figure 2.6 can be approximated as:

$$R_1 = \frac{x}{(1.20 - 0.002x) \sqrt{x^2 + 20x + 1}} \quad (2.13)$$

and

$$\phi_1^2 = \left[1 + \frac{20}{x} + \frac{1}{x^2}\right] (1.25 - 0.0025x) \quad (2.14)$$

Operating conditions for both investigations of Lockhart-Martinelli and Johnson-Abou Sabe are listed in Table 2.3.

Several other correlations^(2,8,9) are also available in the literature. Usually, these methods are restricted under some specified operating conditions (e.g., range of density, temperature, flow regimes, flow patterns, fluid properties, etc.)

2.2.3 Two-Phase Heat Transfer Coefficient

The heat transfer rate to flowing gas-liquid systems requires a correlation for the two-phase heat transfer coefficient, h_i . Experimentally, this parameter is determined

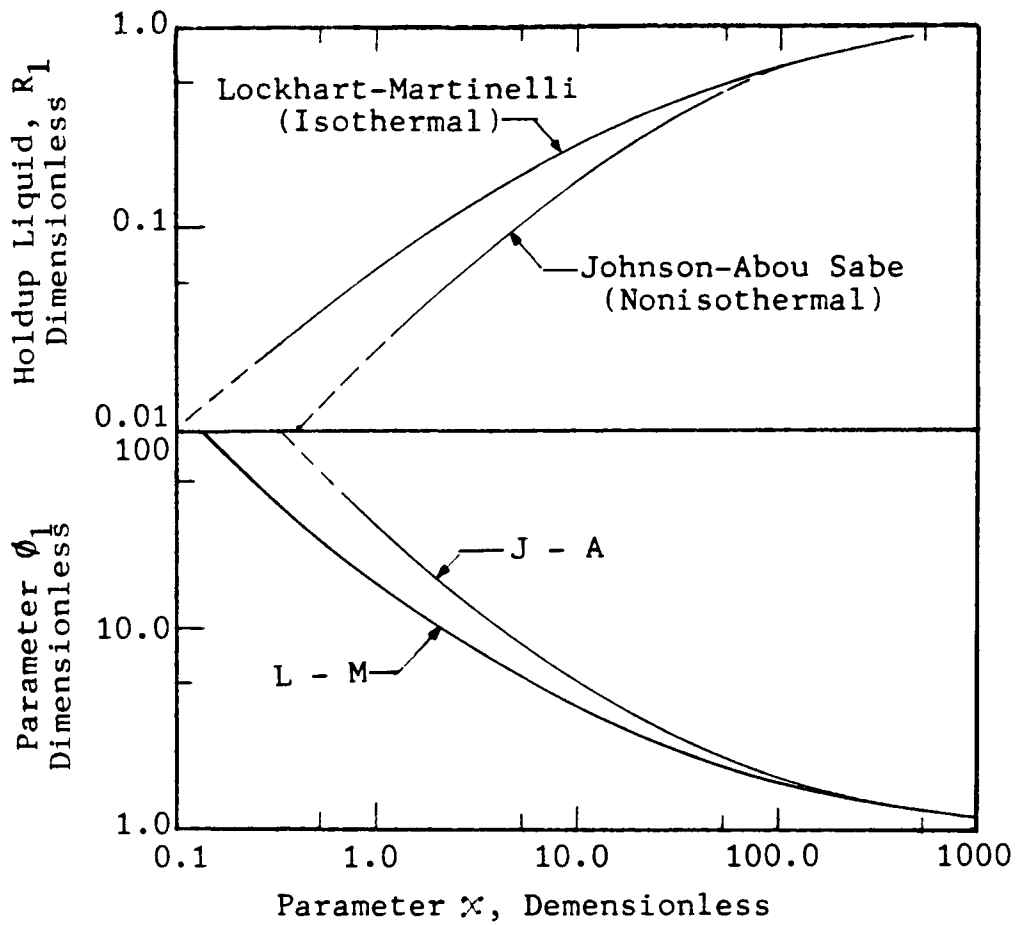


Figure 2.6. Lockhart-Martinelli Correlations for liquid Holdup and Pressure Drop of Turbulent-Turbulent Flow Comparing to Those of Johnson-Abou Sabe

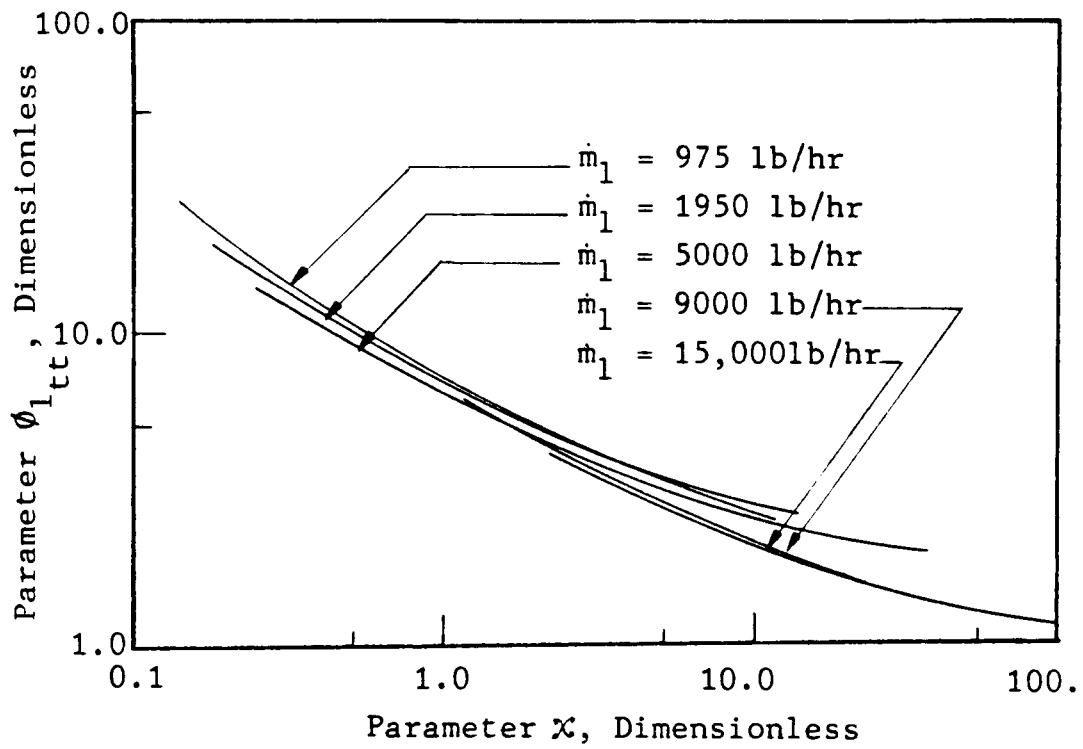


Figure 2.7. Johnson-Abou Sabe Correlation for Pressure Drop for Nonisothermal, Turbulent-Turbulent Air-Water Flows⁽⁷⁾

if local bulk temperature of the fluid, tube wall temperature, and heat flux rate are measured. However, in many chemical process equipment as heaters or heat exchangers, fluid and tube wall temperatures cannot be accurately obtained.

Therefore, the calculation of heat transfer coefficient must depend on the estimation of these quantities. In addition to this difficulty, there are several forms of the heat transfer coefficient correlation related to flow patterns which may be involved in this type of flow. Therefore, to obtain a good prediction of heat transfer coefficient, most investigators restricted their studies to a relatively narrow range of flow conditions.

Davis and David^(10,11) introduced a correlation for calculating the coefficients for the system with high gas mass fractions ($\geq 10\%$). There are two equations corresponding to two flow models. In the homogeneous flow model, both gas and liquid phases have the same velocity, and the model includes the bubble, dispersed, slug, and plug flow patterns. The equation for this model is:

$$h_i = (0.033) \frac{k_l}{D_i} \left(\frac{D_i G_M}{\mu_{tp}} \right) \left(\frac{C_p \mu}{k} \right)_l^{0.4} \quad (2.15)$$

In the separated model or split model, gas and liquid phases have different velocities, and the model includes stratified, annular, and wave flow patterns. Equation is:

$$h_i = (0.06) \frac{k_l}{D_i} \left(\frac{D_i \xi G_M}{\mu_l} \right)^{.87} \left(\frac{\rho_l}{\rho_g} \right)^{.28} \left(\frac{Cp^M}{k} \right)_l^{0.4} \quad (2.16)$$

where:

- Cp_l : Specific heat of liquid, BTU/lb-°F
- D_i : Tube inside diameter, ft
- G_M : Total mass flow rate of the mixture per unit of cross-sectional area, lb/ft²-hr
- h_i : Two-phase heat transfer coefficient, BTU/ft²-hr-°F
- k_l : Thermal conductivity of the liquid, BTU/ft-hr-°F
- ρ_g, ρ_l : Gas and liquid densities, lb/ft³
- μ_g, μ_l : Gas and liquid viscosity, lb/ft-hr
- μ_{tp} : Two-phase fluid viscosity, $\left[\frac{\xi}{\mu_g} + \frac{1-\xi}{\mu_l} \right]^{-1}$, lb/ft-hr
- ξ : Gas mass fraction, $\frac{\dot{m}_g}{\dot{m}_l + \dot{m}_g}$, dimensionless

These two equations are developed from the investigations of cocurrent steam and water flow systems. Average operating conditions for these systems are listed in Table 2.3. Similar as the liquid holdup and pressure drop correlations, there are numerous other correlations (e.g., Berge⁽²⁾, Baker⁽³⁾, Johnson-Abou Sabe⁽⁷⁾, Fried⁽⁸⁾, Kurdika⁽¹²⁾, Groothuis⁽¹³⁾, Hewitt⁽¹⁴⁾, etc.) which have been developed for heat transfer coefficients corresponding to particular flow conditions;

Table 2.3. Operating Conditions of Two-Phase Flow Correlations

Correlations	Systems	D_i Ft	$\dot{m}_l$ lb/hr	$\dot{m}_g$ lb/hr	Avg. Temp. $^{\circ}$ F	Avg. Press. psia	Ref.
L - M	Air and liquids including benzen, kerosene, water, and various oils	.0049 - .085	(*)	(*)	70	16 - 50	(6)
J - A	Air-water Steam-water	.06133 - .0725	1000 - 1500	4.5 - 200	70 - 140	1 - 50	(7)
D - D	Steam-water	(Equiv. Dia.) .0324	40 - 500	12 - 450	70 - 150	25 - 150	(10, 11)

Note: (*) : Not available
 L - M: The Lockhart-Martinelli correlations for liquid holdup and pressure drop
 J - A: The Johnson-Abou Sabe correlations for liquid holdup and pressure drop
 D - D: The Davis-David correlations for heat transfer

these correlations involve one or more empirically determined constants.

2.3 Equilibrium Flash Calculations

When a large amount of heat is added to a gas-liquid flow system, mass transfer between these two phases will be significant. By assuming that the entire system is in equilibrium and local temperature and total pressure are unchanged, the relative amount of gas and liquid at any location along the tube can be estimated by isothermal flash calculation. The application of this method requires K-values of the components to proceed a trial-and-error solution. Rachford-Rice⁽¹⁵⁾ presented a calculational procedure which is illustrated in Table 2.4 when K-values are independent of composition so that local flow rates of the flowing streams corresponding to temperature and pressure are computed. Treybal⁽¹⁶⁾ also established a similar method to apply to the ideal systems. In the latter method, K-values are determined by the ratio of vapor pressure of components to total pressure of the system.

Flash calculations can be applied for air-water flow system. However, a simpler method was developed by Keenan and Keyes⁽¹⁷⁾. By neglecting the amount of air absorbed by water and assuming the ideal gas mixture (for water vapor and air), mole fraction of water vapor in the gas phase is:

Table 2.4. Procedure of Flash Calculations When K-Values Are Independent of Compositions⁽²³⁾

(1) Comp. #	(2) # Moles Entering at z (Mole/hr)	(3) Overall Mole % at z (%)	(4) $\Psi = \frac{F_{gz}}{F_z}$	(5) Mole % in the Gas Phase at z	(6) Mole % in the Liquid Phase at z
1	F_{1z}	Z_{1z}	$\frac{Z_1(1 - K_1)}{1 + \Psi(K_1 - 1)}$	$Y_1 = \frac{Z_1 K_1}{1 + \Psi(K_1 - 1)}$	$X_1 = Y_1/K_1$
2	F_{2z}	Z_{2z}	$\frac{Z_2(1 - K_2)}{1 + \Psi(K_2 - 1)}$	$Y_2 = \frac{Z_2 K_2}{1 + \Psi(K_2 - 1)}$	$X_2 = Y_2/K_2$
3	F_{3z}	Z_{3z}	$\frac{Z_3(1 - K_3)}{1 + \Psi(K_3 - 1)}$	$Y_3 = \frac{Z_3 K_3}{1 + \Psi(K_3 - 1)}$	$X_3 = Y_3/K_3$
⋮	⋮	⋮	⋮	⋮	⋮
j	F_{jz}	Z_{jz}	$\frac{Z_j(1 - K_j)}{1 + \Psi(K_j - 1)}$	$Y_j = \frac{Z_j K_j}{1 + \Psi(K_j - 1)}$	$X_j = Y_j/K_j$
⋮	⋮	⋮	⋮	⋮	⋮
n	F_{nz}	Z_{nz}	$\frac{Z_n(1 - K_n)}{1 + \Psi(K_n - 1)}$	$Y_n = \frac{Z_n K_n}{1 + \Psi(K_n - 1)}$	$X_n = Y_n/K_n$
$F_z = \sum_{j=1}^n F_{jz} \quad 1.0$			$f(\Psi) = \sum_{j=1}^n \frac{Z_j(1 - K_j)}{1 + \Psi(K_j - 1)} = 0$	$\sum_j Y_j = 1.0$	$\sum_j X_j = 1.0$

Note: Columns (2) and (3) are the molar feed rates and compositions of the components in the feed. The iteration is started by guessing a value for Ψ ($0 < \Psi < 1$) which is defined as the ratio of the molar gas flow rate in an equilibrium state at z. New value of Ψ must be guessed until finally $f(\Psi)$ in Column (4) is equal to zero. Use the last value of Ψ to compute Y 's in Column (5) and use Y 's and K -values to find X 's. Molar flow rates of the gas phase, F_{gz} , and of the liquid phase, F_{1z} , are determined by:

$$F_{gz} = \Psi F_z$$

$$F_{1z} = F_z - F_{gz}$$

$$Y_w = \frac{P_w^o}{P_t} = \frac{P_c}{P_t} \exp [\tau(\theta_s)] \quad (2.17)$$

with

$$\tau(\theta_s) = -7.4691 \left(\frac{\theta_c}{\theta_s} - 1 \right) \left[\frac{1 + .67508(\theta_c - \theta_s) + .18746(\theta_c - \theta_s)^3}{1 + 0.81642(\theta_c - \theta_s)} \right] \quad (2.18)$$

where:

P_c : Critical pressure of water, psia

P_t : Total pressure of system, psia

P_w^o : Vapor pressure of water, psia

θ_c : Normalized critical temperature, T_c/T_b :

T_c (K) and T_b (K) are critical and boiling temperature of water

θ_s : Gas phase normalized temperature, T/T_b :

T (K) and T_b (K) are temperature of water and boiling water temperature

2.4 Heat Transfer

Basic equations and properties required for heat transfer calculations such as enthalpy of components, gases, and liquids are available in the literature^(18,19,20). Radiation heat transfer and tube wall temperature estimations are illustrated by McCabe and Smith⁽¹⁸⁾. These latter aspects are detailed in the next sections.

2.5 Physical Properties Data for Illustrative Flow Systems

Several air-water flows and a multicomponent system of hydrogen, nitrogen, heptane, and decane are utilized for demonstration of the code application. Physical properties such as density, viscosity, thermal conductivity, latent heat, specific heat, etc. of the components are required to support the calculations. Fortunately, many of these properties are described as functions of temperature or pressure available in the literature. Other properties, for example K-values and density, are more likely to be presented by diagramming or tabularizing^(21,22) because they are sensitive to both temperature and pressure.

Since data for K-values of hydrogen and nitrogen are not sufficiently available in the literature, following method is frequently employed to estimate the new values. For example, at 250°F and 500 psia, K-values of N₂ and n-decane are 20 and .013 respectively. Their ratio is:

$$K_{N_2}/K_{n-C_{10}} = 20/0.013.$$

At 250°F and 20 psia, K-value of N-decane is .14, K-value of N₂ at this condition is estimated by: $K_{N_2} = (20/0.013)(0.14) = 215$. K-values of hydrogen and nitrogen for some conditions are available in the literature⁽²³⁾. Physical properties data of the components used in illustrative systems are listed in Table C.1 to Table C.6 in Appendix C.

2.6 Literature Survey Data

Some experimental data of air-water flows and a multicomponent, gas-liquid flow are used for demonstration of the code application. In Table 2.5 data of 6 runs of air-water are presented. The first two runs were performed by Johnson and Abou-Sabe⁽⁷⁾. Air and water flowing through a horizontal 15.7 ft. length, 1-in. 16 gage drawn-brass tube were heated by a steam jacket. In the jacket, internal baffling was arranged so that the condensate formed by external heat loss was withdraw separately. Pressure was measured at two ends and at the middle of the tube, and tube wall temperature was observed by means of 24 thermocouples located spirally along the tube. It was inferred that the thermocouple lead wires caused irregular condensate flow and so produced the variation observed. Measurements of tube wall temperature might have a deviation as much as $\pm 10^{\circ}\text{F}$. Temperature or pressure of steam and condensate in the steam jacket were not provided, pressure drop was taken between inlet and outlet, flow patterns were visually observed, liquid holdup was calculated (from flow-patterns factors and effective hydraulic diameters), and heat flux rate was evaluated by assuming saturation of air throughout and considered as a function of single-phase heat transfer coefficients of both phases. A similar experiment was performed in Run No.3⁽²⁴⁾; however a 15 ft. length, 0.736 in.

Table 2.5. Experimental Data of Two-Phase Air-Water Flows in the Horizontal Tubes

Run No.	$\dot{m}_g$		$\dot{m}_l$	Fluid Temp., °F		Total Press. psia		Avg. T_w °F	Avg. R_l	Q/A $\frac{\text{BTU}}{\text{ft}^2\text{-hr}}$	h_l $\frac{\text{BTU}}{\text{ft}^2\text{-hr-}^\circ\text{F}}$	Flow Regime	Flow Pattern	D_l Ft	L Ft	Ref.
	SCFM	lb/hr		Inlet	Outlet	Inlet	Outlet									
1	29.7	133.6	975	64.5	179.6	26.7	20.7	196	.08	60000	(*)	T-T	Ann.	.0725	15.7	(7)
2	44.7	201.1	975	65.0	171.3	30.6	22.9	194	.04	59300	(*)	T-T	Ann.	.0725	15.7	(7)
3	42.5	191.2	1375	75.8	178.0	43.8	28.0	202	.12	63000	(*)	T-T	Ann.	.06133	15.0	(24)
4	30.1	135.4	1000	82.0	183.0	34.0	23.5	(*)	(*)	(*)	1190	T-T	(*)	.06142	15.7	(8)
5	45.1	202.9	1000	81.0	175.0	38.9	24.2	(*)	(*)	(*)	1130	T-T	(*)	.06142	15.7	(8)
6	45.0	202.4	2000	96.0	180.0	48.0	30.4	(*)	(*)	(*)	1740	T-T	(*)	.06142	15.7	(8)

(*) Not available

I.D. copper pipe was used. The last three runs were made by Fried⁽⁸⁾. The tube is a horizontal 0.737 in. I.D. copper with a length of 15.7 ft. Air-water mixtures were also heated by a steam jacket, but both temperature and pressure of steam entering the jacket were not provided. Neither tube wall temperature or flow patterns were recorded, and the two-phase heat transfer coefficient was computed from heat loss of the steam in the jacket. For all six runs in Table 2.5, both air and water streams were turbulent and the flows were separated model. Outlet flow rates were not provided; consequently mass exchanges between phases are not available.

Since experimental data for two-phase, multi-component, gas-liquid flow were not found in the open literature, a system of hydrogen, nitrogen, heptane, and decane is constructed. Operating conditions of this system (e.g., feed rates, tube size, heating medium temperature, etc.) are selected similarly, relatively to those of the air-water flows. Also, due to the difficulty of collecting all data required to support the calculations, some data (for example, K-values in some specified temperature, thermal conductivity of hydrogen and nitrogen in the liquid phase) are empirically estimated. Physical property data for both air-water and multicomponent systems are listed in Appendix C.

3. MODEL DEVELOPMENT

3.1 Assumptions and Restrictions

A mixture of volatile, organic components is heated in the presence of small particles (ground coal) and certain "permanent" gases, such as hydrogen, from an initial temperature to a final temperature in a tubular heater. A certain amount of vaporization will take place during the heating process, so the compositions of the phases will change and the system pressure will decrease due to frictional losses as the materials flow down the tube. The following set of assumptions and restrictions is proposed in order to simplify the problem, yet retain a reasonably good description of the physical process, and in order to determine the set of equations (i.e., match the number of equations with the number of unknowns).

- A. Neglect the presence of the solid particles so that the system becomes a two-phase gas-liquid flow
- B. Steady state
- C. One-dimensional system (eg., flow in the axial z-direction only)
- D. Transport within the tube is by convection only (axial dispersion is neglected)
- E. Negligible chemical reaction (in certain applications, there will be some reaction such as hydrogenation of unsaturated oils and such as cracking of some hydrocarbons)

- F. Equilibrium is approached locally between the gas and liquid phases
- G. The two-phase flow characteristics, such as the pressure drop and the liquid holdup, can be predicted by use of correlations developed from data obtained from cold flow experiments
- H. The liquid phase is essentially ideal so that the heats of mixing are negligible
- I. The tube is located in a horizontal plane
- J. The gas phase is also essentially ideal

The governing conservation equations with the use of these assumptions and restrictions take the forms as outlined in the following sections.

3.2 Mass Balances

Consider two-phase gas-liquid flow in a section of a horizontal tube as shown in Figure 3.1. the mass flow rates of the gas and liquid phases at z are $\dot{m}_g$ and $\dot{m}_l$, respectively. The n components constituting the system are distributed between the phases, such that:

$$\dot{m}_g = \sum_{j=1}^n \dot{m}_{jg} \quad (3.1)$$

$$\dot{m}_l = \sum_{j=1}^n \dot{m}_{jl} \quad (3.2)$$

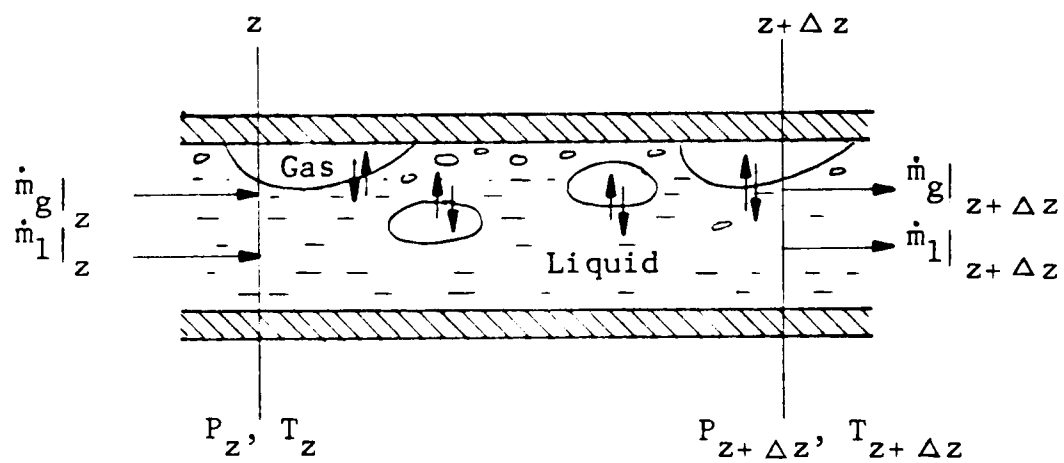


Figure 3.1. Two-Phase Gas-Liquid Flow in a Section of a Horizontal Tube

where $\dot{m}_{g_j}$ and $\dot{m}_{l_j}$ are the mass flow rates of component j in the gas and liquid phases respectively. The total mass exchange rates between phases (by vaporation or dissolution) is $\dot{m}_t$ in an element of tube volume, ΔV . The overall and component mass balances for the two phases may now be developed.

3.2.1 Overall Mass Balances for a Section of the Tube

Refer to Figure 3.1, the liquid phase balances around the elemental volume ΔV of length Δz and cross-sectional area A_i may be written as:

$$\dot{m}_l|_z = \dot{m}_l|_{z+\Delta z} + \dot{m}_t \Delta V_l \quad (3.3)$$

Expressing the liquid volume ΔV_l in term of the liquid holdup fraction, R_l , dividing by Δz and let $\Delta z \rightarrow$ "small", we obtain the following first order differential equation for the overall, local mass balance on the liquid phase:

$$\frac{d\dot{m}_l}{dz} = - \dot{m}_t R_l A_i \quad (3.4)$$

where $\dot{m}_l$: Liquid mass flow rate, lb/hr

$\dot{m}_t$: Local net mass transfer rate between gas and liquid phase per unit liquid volume, lb/hr-ft³

R_l : Liquid holdup, dimensionless, $R_l = \Delta V_l / \Delta V$

A_i : Cross-sectional area of the tube based

on tube inside diameter, ft²

D_i : Tube inside diameter, ft

Similarly, for the gas phase:

$$\frac{d\dot{m}_g}{dz} = \dot{m}_t R_1 A_i \quad (3.5)$$

Equations (3.4) and (3.5) show that the mass lost by the liquid phase is gained by the gas phase. If these two equations are added and the resulting equation integrated, the result is that the total mass flow rate down the tube is constant, consistent with Assumption B above.

3.2.2 Component, Local Mass Balances

By a procedure similar to that above, we obtain the local, component mass balances for the two phases:

$$\frac{d\dot{m}_{jg}}{dz} = \dot{m}_{jt} R_1 A_i \quad j = 1, 2, 3, \dots, n \quad (3.6)$$

$$\frac{d\dot{m}_{jl}}{dz} = -\dot{m}_{jt} R_1 A_i \quad j = 1, 2, 3, \dots, n \quad (3.7)$$

where $\dot{m}_{jg}$, $\dot{m}_{jl}$: Mass flow rate of component j in the gas and liquid phases, lb/hr

$\dot{m}_{jt}$: Local net mass transfer rate of component j between the gas and liquid phase per unit liquid volume, lb/sec-ft³

The problem at this point is that we do not have mass transfer correlations for exchange between phases in

two-phase flow in the tube. As a consequence, the $\dot{m}_{jt}$ are unknown as well as the $\dot{m}_{jg}$ and the $\dot{m}_{jl}$. Equations (3.6) and (3.7) therefore constitute a set of $2n$ equations containing $3n$ unknowns, a situation which is clearly indeterminant. The only way we have of proceeding is to invoke Assumption F, local equilibrium between phases. With this assumption, we can relate the local phase compositions for each component, j , and therefore relate the respective flow rates in each phase. The number of unknown mass flow rates is reduced by n , but as equilibrium is temperature and pressure dependant, two additional unknowns temperature and pressure are introduced. The result is that we have $2n$ equations and $2n+2$ unknowns. A minimum of two additional independent equations must therefore be written. One equation is the enthalpy balance, another is the Lockhart-Martinelli equation for computing total pressure of the system at location z , given the inlet pressure and the flow rates.

3.2.3 Exchange of Mass Between the Phases

Equilibrium data for hydrocarbon systems, for example, are frequently reported in terms of "K-value", defined as the ratio of the equilibrium mole fraction in the gas phase, y , to that in the liquid phase, x , at a given temperature and pressure^(15,20,25). For a component j ,

$$K_j = \frac{y_j}{x_j} \quad (3.8)$$

and if component j is an ideal solution,

$$K_j = \frac{y_j}{x_j} = \frac{P_j^0}{P_t} \quad (3.9)$$

where P_j^0 is vapor pressure and P_t is total pressure of the system. Based on this definition of K-value, flash calculation is employed to compute the stream flow rates for a section of the tube and the procedure is described as following.

Consider a cell from z to $z + \Delta z$ as shown in Figure 3.2. The flow system is steady and no chemical reaction involved. Under these conditions, total number of mole of components entering the cell at z and leaving the cell at $z + \Delta z$ will be the same. Total number of moles in both gas and liquid phases at z , F_z , is give by:

$$F_z = F_{g_z} + F_{l_z} \quad (3.10)$$

$$\text{with } F_{g_z} = \sum_{j=1}^n F_{jg_z} \quad (3.11)$$

$$F_{l_z} = \sum_{j=1}^n F_{jl_z} \quad (3.12)$$

where F_{g_z} , F_{l_z} : Molar flow rates of the gas and liquid phases at z , # mole/hr

F_{jg_z} , F_{jk_z} : Molar flow rates of component j in the

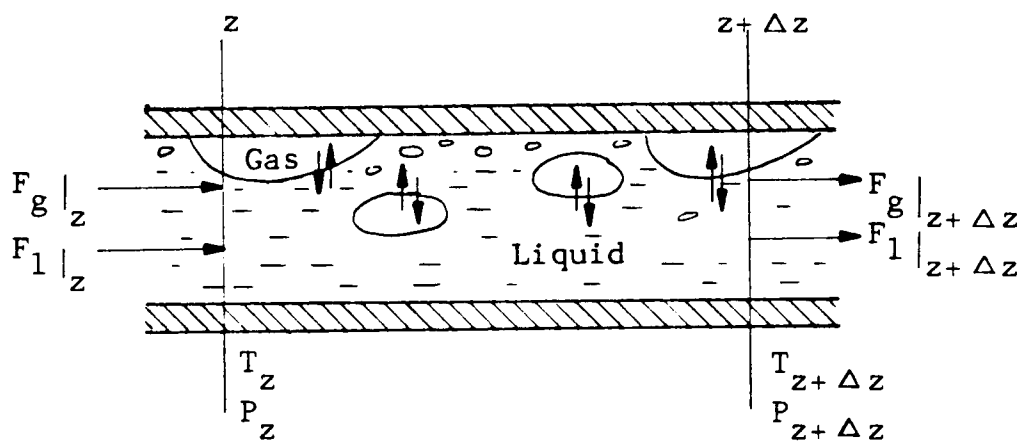


Figure 3.2. Molar Flow Rates of Gas and Liquid Phases for Mass Balance Development

gas and liquid phases at z , # mole/hr
 F_z : Total molar flow rate of the mixture
 at z , # mole/hr

The two-phase flow mixture is assumed to be in equilibrium at z where temperature is T_z and total pressure is P_z , all molar flow rates at z can be determined by flash calculation, a procedure which involves a trial-and-error solution as described step by step in Table 2.4.

With M_j as molecular weight of component j , after mole fractions, x_j and y_j are calculated, we have:

$$\dot{m}_{jg_z} = M_j F_{jl_z} = M_j y_{j_z} F_{g_z} \quad (3.13)$$

and

$$\dot{m}_{jl_z} = M_j F_{jl_z} = M_j x_{j_z} F_{l_z} \quad (3.14)$$

where M_j : Molecular weight of component j ,
 lb/lb mol

$\dot{m}_{jg_z}, \dot{m}_{jl_z}$: Gas mass flow rates of component j in
 the gas and liquid phases at z , lb/hr

x_{j_z}, y_{j_z} : Mole fractions of component j in the
 liquid and gas phases at z , dimension-
 less

The same procedure is applied at $z + \Delta z$ of the cell;
 similarly we obtain:

$$\dot{m}_{jg_{z+\Delta z}} = M_j F_{jg_{z+\Delta z}} = M_j y_{j_{z+\Delta z}} F_{g_{z+\Delta z}} \quad (3.15)$$

and

$$\dot{m}_{j1_{z+\Delta z}} = M_j^F j1_{z+\Delta z} = M_j^y j_{z+\Delta z} F_{1_{z+\Delta z}} \quad (3.16)$$

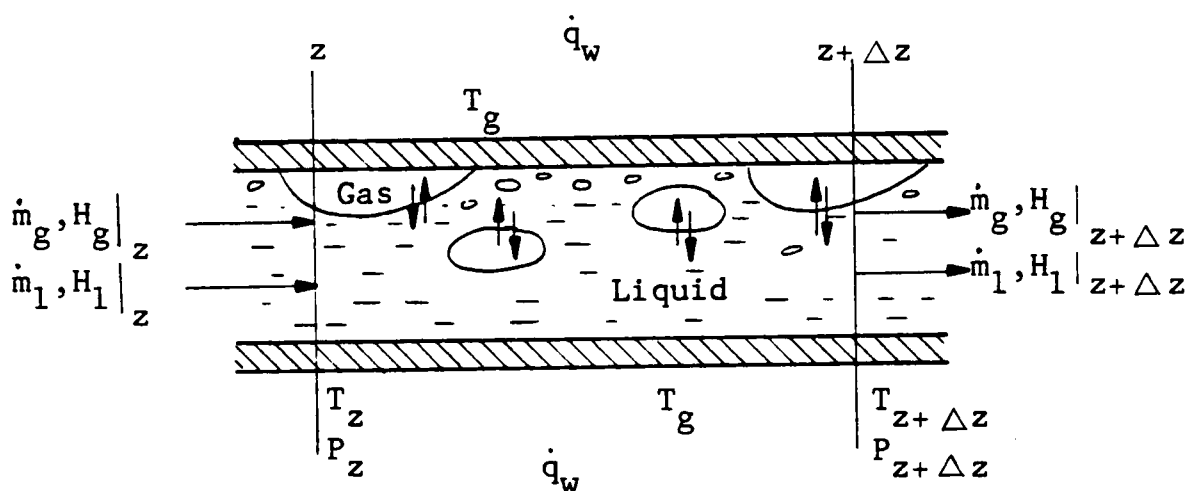
Recall Equations (3.1) to (3.7); total mass flow rates and net mass transfer between phases now can be determined. By repeatedly applying this calculation procedure from cell to cell, mass flow rates of the streams at any location along the tube will be determined.

3.3 Local Enthalpy Balances

After the local flow rates (based on iterating temperature and pressure, details are described in Sections 4.1 and 4.2) are found, enthalpy balances for the flow system can be developed so that a temperature profile of the mixture of the entire tube length will be introduced. Consider a section of the tube from z to $z + \Delta z$ as shown in Figure 3.3 where temperature, pressure, and flow rates of the inlet and outlet streams are known (e.g., after iteration), the local enthalpy balance in the absence of chemical reaction may be written as:

$$\frac{d}{dz} (\dot{m}_l H_l + \dot{m}_g H_g) = D_i \dot{q}_w \quad (3.17)$$

where D_i : Tube inside diameter, ft
 dz : Length of a small tube section, ft
 H_g, H_l : Average enthalpy of the gas and liquid phases, BTU/lb
 $\dot{m}_g, \dot{m}_l$: Mass flow rates of the gas and liquid



T_g : Heating medium temperature, $^{\circ}\text{F}$
 $\dot{q}_w$: Heat flux rate, $\text{BTU}/\text{ft}^2\text{-hr.}$
 H_g, H_l : Enthalpy of the streams, BTU/lb

Figure 3.3. Enthalpy Balance for a Section of the Tube

phases at z and $z + \Delta z$, lb/hr

$\dot{q}_w$: Heat flux through the tube wall,
BTU/ft²-hr

Using Assumption G, heat mixing is negligible, H in Equation (3.17) can be determined as:

$$H_l = \sum_{j=1}^n x'_j H_{jl} \quad (3.18)$$

and

$$H_g = \sum_{j=1}^n y'_j H_{jg} \quad (3.19)$$

x'_j , y'_j , and H_j are mass fractions and enthalpy of component j in the liquid and gas phases respectively.

Using Assumption F (local equilibrium), the only difference between the gas phase and liquid phase enthalpy is the latent heat, λ_j . For a given component j , we have:

$$H_{jg} = H_{jl} + \lambda_j \quad (3.20)$$

Equation (3.19) becomes:

$$H_g = \sum_{j=1}^n y'_j (H_{jl} + \lambda_j) \quad (3.21)$$

where the liquid phase enthalpy of component j , H_{jl} , may be written, neglecting the pressure contribution to the enthalpy as:

$$H_{jl} = H_j^0 + \int_{T_r}^T C_{pj} dT \quad (3.22)$$

or

$$H_{jl} = H_j^0 + \overline{C}_{pj} (T - T_r) \quad (3.23)$$

where

C_{pj} : Specific heat of component j in the liquid phase BTU/lb - °F

$\overline{C}_{pj}$: Average specific heat of component j in the liquid phase, BTU/lb - °F

H_j^0 : Enthalpy of component j in the liquid phase at reference temperature T_r , BTU/lb

T_r : Reference temperature, °F

The amount of evaporation from z to z + Δz is given

by:

$$\Delta \dot{m} = \dot{m}_t \Delta z A_i \quad (3.24)$$

or

$$\Delta \dot{m} = \dot{m}_l \Big|_z - \dot{m}_l \Big|_{z+\Delta z} = \dot{m}_g \Big|_{z+\Delta z} - \dot{m}_g \Big|_z \quad (3.25)$$

where

A_i : Tube inside cross sectional area, ft²

$\dot{m}_g, \dot{m}_l$: Mass flow rates of gas and liquid streams, lb/hr

$\dot{m}_t$: Local net mass transfer rate between gas and liquid phases per unit liquid volume, lb/hr-ft³

$\Delta \dot{m}$: Net mass transfer between gas and liquid phases in a tube section of length Δz , lb/hr

Δz : Length of the section of the tube, ft

Hence, the average latent heat is empirically approximated as:

$$\Delta H_{\text{Latent}} = \frac{\dot{\Delta m}}{2} (\lambda_z + \lambda_{z+\Delta z}) \quad (3.26)$$

From Equations (3.20) and (3.23), the heat added to this evaporated amount is:

$$\Delta H'_g = \Delta \dot{m} \bar{C}_{p1} (T_{z+\Delta z} - T_z) + \Delta H_{\text{Latent}} \quad (3.27)$$

where

$\bar{C}_{p1}$: Average specific heat of the liquid phase, BTU/lb-°F

ΔH_{Latent} : Latent heat given by Equation (3.26)

$\bar{C}_{p1}$ and $\bar{C}_{pg}$ are average specific heat of the liquid and gas streams. These properties are computed from mass fraction and specific heat of the components in each phase.

We have

$$\bar{C}_{pg} = \sum_{j=1}^n \left(\frac{y'_{jz} + y'_{jz+\Delta z}}{2} \right) \bar{C}_{pjg} \quad (3.28)$$

$$\bar{C}_{p1} = \sum_{j=1}^n \left(\frac{x'_{jz} + x'_{jz+\Delta z}}{2} \right) \bar{C}_{pj1} \quad (3.29)$$

The total enthalpy change of the liquid phase is:

$$\Delta H_1 = \dot{m}_1 H_1 \Big|_{z+\Delta z} - \dot{m}_1 H_1 \Big|_z \quad (3.30)$$

Equation (3.30) can be approximated by using average liquid

flow rate and specific heat. We have

$$\Delta H_1 = \dot{m}_1 \bar{C}_{p1} (T_{z+\Delta z} - T_z) \quad (3.31)$$

Without including the amount of evaporated liquid, the average enthalpy difference in the gas phase is:

$$\Delta H_g = \dot{m}_{g_z} (H_{g_{z+\Delta z}} - H_{g_z}) \quad (3.32)$$

By using the average specific heat, Equation (3.32) becomes:

$$\Delta H_g = \dot{m}_{g_z} \bar{C}_{pg} (T_{z+\Delta z} - T_z) \quad (3.33)$$

Total enthalpy change of the system is:

$$\Delta H_{\text{total}} = \Delta H_g + \Delta H'_g + \Delta H_1 \quad (3.34)$$

Use of Equations (3.18) to (3.34) in Equation (3.17) results in an equation in terms of the flow rates and local temperature of the fluid as following:

$$\begin{aligned} \Delta H_{\text{total}} &= \dot{m}_{g_z} \bar{C}_{pg} (T_{z+\Delta z} - T_z) + \Delta \dot{m} \bar{C}_{p1} (T_{z+\Delta z} - T_z) \\ &\quad + \Delta H_{\text{latent}} + \dot{m}_{1_{\text{avg}}} \bar{C}_{p1} (T_{z+\Delta z} - T_z) \\ &= \pi D_i \Delta z \dot{q}_w \end{aligned} \quad (3.35)$$

The heat flux to the fluid, $\dot{q}_w$, may be written in terms of the difference between the temperature of the heating medium, T_g , and the local bulk temperature of the fluid, T_z , as driving force (logarithmic mean temperature difference, ΔT_{LMTD}) and the overall local heat transfer coefficient, U_i , according to:

$$\dot{q}_w = U_i \Delta T_{\text{LMTD}} \quad (3.36)$$

When radiant heat transfer to the tube is considered, with h_c and h_r as convective and radiation heat transfer coefficients and the tube surface assumed to be black surface, $\dot{q}_w$ is given by:

$$\dot{q}_w = (h_c + h_r) (T_g - T_w) \quad (3.37)$$

If the heat flux by radiation is $\dot{q}_r$, then h_r is⁽¹⁸⁾:

$$h_r \equiv \frac{\dot{q}_r}{(T_g - T_w)} \quad (3.38)$$

In terms of the view factor and for gray surfaces $\dot{q}_r$ is given by⁽¹⁸⁾.

$$\dot{q}_r = \sigma F (T_g^4 - T_w^4) \quad (3.39)$$

where

F : Overall interchange factor, dimensionless

σ : Stefan-Boltzmann constant, 0.1713×10^{-8}
BTU/ft²-hr-R⁴

Overall heat transfer coefficient, U_i , in Equation (3.36) is:

$$U_i = \left[\frac{1}{h_i} + \frac{D_i x_w}{\bar{D} k_w} + \frac{D_i}{D_o h_o} \right]^{-1} \quad (3.40)$$

and

$$\Delta T_{LMTD} = \frac{T_{z+\Delta z} - T_z}{\text{LN} \left[\frac{T_g - T_z}{T_g - T_{z+\Delta z}} \right]} \quad (3.41)$$

where

- D_i, D_o : Tube inside and outside diameters, ft
 $\bar{D}$: Log mean diameter, $[D_o - D_i / \text{LN}(D_o - D_i)]$,
ft
 h_i : Two phase fluid heat transfer coefficient,
 $\text{BTU/ft}^2\text{-hr-}^\circ\text{F}$
 h_o : Outside film heat transfer coefficient,
 $\text{BTU/ft}^2\text{-hr-}^\circ\text{F}$
 k_w : Tube wall thermal conductivity,
 $\text{BTU/ft-hr-}^\circ\text{F}$
 T_g : Heating medium temperature, $^\circ\text{F}$
 T_z : Fluid temperature, $^\circ\text{F}$
 x_w : Tube wall thickness, ft

In Equation (3.40) above, h_o is estimated from the heating system outside the tube (fired heater, steam jacket, etc.) or from Equation (3.37) for combination of both h_o and h_r , h_i is calculated by Davis and David^(10,11) correlations (Equations (2.15) and (2.16), pages 23 and 24).

Obviously, the use of Equations (2.15) and (2.16) requires an ascertainment of the flow model, separated or homogeneous, which is directly connected to the study of flow patterns introduced in Section 2.2.1.

3.4 Tube Wall Temperature

In the problems of heat transfer to fluid flows,

tube wall temperature is always of interest because of its connection to the calculation of heat transfer coefficient and evaluation of fluid viscosity at the wall. Usually, tube wall temperature is measured by several thermocouples located along the tube, eventhough deviation of the mean temperature is very large due to the technical and operating problems. The following method is introduced to compute tube wall temperature, T_w , after fluid temperature is found. The estimate of the wall temperature requires a calculation based on the resistance equation given in McCable and Smith⁽¹⁸⁾:

$$\frac{\Delta T}{1/U_i} = \frac{\Delta T_i}{1/h_i} = \frac{\Delta T_w}{(x_w/k_w)(D_i/\bar{D})} \quad (3.42)$$

where

ΔT : Average temperature difference between the heating medium temperature and the fluid temperature, $^{\circ}\text{F}$

ΔT_i : Temperature drop through the inside two-phase fluid, $^{\circ}\text{F}$

ΔT_w : Temperature drop through the wall, $^{\circ}\text{F}$

After the individual resistances [each term in Equation (3.42)] are estimated neglecting the tube wall resistance, $(x_w/k_w)(D_i/\bar{D})$. The total temperature drop, ΔT , can be split into the individual temperature drops by the use of

Equation (3.42). We have:

$$\Delta T_i = \frac{U_i}{h_i} \Delta T \quad (3.43)$$

Substitute U_i from Equation (3.34) and neglecting the wall-resistance terms gives

$$\Delta T_i = \frac{1/h_i}{1/h_i + D_i/D_o h_o} \Delta T \quad (3.44)$$

with

$$\Delta T = T_g - \frac{T_z + T_{z+\Delta z}}{2} \quad (3.45)$$

where T_g is the heating medium temperature, T_z and $T_{z+\Delta z}$ are inlet and outlet stream temperature. The wall temperature. The wall temperature T_w is then obtained by the following equation

$$T_w = T + \Delta T_i \quad (3.46)$$

where T is average fluid temperature given by:

$$T = \frac{T_z + T_{z+\Delta z}}{2} \quad (3.47)$$

3.5 Flow Patterns and Flow Models

Flow patterns are determined by the Mandhane correlations⁽⁵⁾ as shown in Figure 2.4. Superficial velocities of the gas stream is given by:

$$v_g = \frac{\dot{m}_g / 3600}{\rho_g A_i} \quad (3.48)$$

and for the liquid stream:

$$V_1 = \frac{\dot{m}_1 / 3600}{\rho_1 A_i} \quad (3.49)$$

where

- A_i : Tube inside cross-sectional area, ft^2
 $\dot{m}_g, \dot{m}_l$: Mass flow rates of gas and liquid streams,
lb/hr
 V_g, V_l : Superficial velocities of the gas and
liquid stream, ft/sec
 ρ_g, ρ_l : Densities of gas and liquid streams,
lb/ft³

In order to apply the Davis-David^(10,11) correlations, the flow patterns are categorized and classified into one of two flow models. The homogeneous flow model consists of patterns in which the gas and liquid phases have the same velocities such as plug, bubble, and dispersed flows, and the separated flow model consists of patterns in which the gas and liquid phases have different velocities, such as wavy, annular, and stratified flows. Flow patterns may change along the tube due to frictional pressure drop and heat transfer to the fluid. In this case, we have a transition between the flow patterns or flow models.

3.6 Liquid Holdup Fraction and Pressure Drop

Four types of flow regimes, both liquid and gas phases are turbulent, liquid phase is viscous and gas phase is turbulent, liquid phase is turbulent and gas phase is viscous, and both liquid and gas phases are viscous, can exist simultaneously with several flow patterns during the flow of gas and liquid in a horizontal tube. The stream is viscous when its Reynold number is less than 1000 and turbulent when its Renynold number is larger than 2000⁽⁶⁾. For the gas stream in a circular tube, Reynold number is given by:

$$Re_g = \frac{4 \dot{m}_g}{\pi D_i \mu_g} \quad (3.50)$$

and for the liquid phase

$$Re_l = \frac{4 \dot{m}_l}{\pi D_i \mu_l} \quad (3.51)$$

where

D_i : Tube inside diameter, ft

$\dot{m}_g, \dot{m}_l$: Mass flow rates of the gas and liquid streams, lb/hr

Re_g, Re_l : Reynold number of the gas and liquid stream, dimensionless

μ_g, μ_l : Viscosity of the gas and liquid streams, lb/ft-hr

As previously mentioned in Section 2.2.2, single phase pressure drop given by Equations (2.6) and (2.7) corresponding to the flow regimes. Similarly, parameter χ is computed by Equations (2.8) to (2.11). Liquid holdup is then established by Figures 2.5 or 2.6. Finally, two-phase pressure drop is computed by correlations in the same figures and Equations (2.1) and (2.2) as following:

$$\left[\frac{\Delta P}{\Delta z} \right]_{tp} = \phi_g^2 \left[\frac{\Delta P}{\Delta z} \right]_g \quad (3.52)$$

or

$$\left[\frac{\Delta P}{\Delta z} \right]_{tp} = \phi_1^2 \left[\frac{\Delta P}{\Delta z} \right]_1 \quad (3.53)$$

3.7 Model Development Summary

Flow rates, compositions, pressure, and temperature of the outlet streams are calculated by the equations resulted from mass balances, heat balances, and pressure drop correlations. Table 3.1 lists the fundamental equations of the model development. Application of these equations are introduced in the next section.

Table 3.1. Fundamental Equations and Correlations in the Model Development

Subjects	Component j	Gas Phase	Liquid Phase
Mass Balance			
Mass flow rates from flash calculation	$\dot{m}_j = M_j F_j$ (3.13-3.17)	$\dot{m}_g = \sum_{j=1}^n \dot{m}_j$ (3.1)	$\dot{m}_l = \sum_{j=1}^n \dot{m}_j$ (3.2)
Diff. eqt. of mass for small Δz	$\frac{d\dot{m}_j}{dz} = \dot{m}_j R_j A_1$ (3.6-3.7)	$\frac{d\dot{m}_g}{dz} = \dot{m}_g R_g A_1$ (3.5)	$\frac{d\dot{m}_l}{dz} = -\dot{m}_l R_l A_1$ (3.4)
Enthalpy	$H_j = H_{jl} + \lambda_j$ (3.21) $H_{jl} = H_j^0 + C_p(T - T_r)$ (3.24)	$H_g = \sum_{j=1}^n y_j H_j$ (3.20)	$H_l = \sum_{j=1}^n x_j H_j$ (3.19)
Enthalpy Balances			
Latent heat		$\Delta H_{\text{Latent}} = \frac{\Delta \dot{m}}{2} (\lambda_g + \lambda_l)$ (3.27)	
Heat added to evaporated liquid		$\Delta H' = -\Delta \dot{m} \bar{C}_{p_l} (T_{z\Delta z} - T_z) + \Delta H_{\text{Latent}}$ (3.28)	
Enthalpy change in liquid phase		$\Delta H_l = \dot{m}_l \bar{C}_{p_l} (T_{z\Delta z} - T_z)$ (3.32)	
Enthalpy change in gas phase		$\Delta H_g = \dot{m}_g \bar{C}_{p_g} (T_{z\Delta z} - T_z)$ (3.34)	
Overall		$\Delta H_{\text{Total}} = \dot{m}_g \bar{C}_{p_g} (T_{z\Delta z} - T_z) + \Delta \dot{m} \bar{C}_{p_l} (T_{z\Delta z} - T_z) + \Delta H_{\text{Latent}} + \dot{m}_l \bar{C}_{p_l} (T_{z\Delta z} - T_z) - \pi D \Delta z \dot{q}_w$ (3.36)	
Heat Transfer Coefficient			
Overall		$U_l = (1/h_l + D_l x_w / \bar{D} k_w + D_l / D_o h_o)^{-1}$ (3.39)	
Davis-David Expts.		$h_l = .033 (k_l / D_l) (D_l C_{p_l} / \mu_{tp}) (C_p \mu / k)^{-.4}$ (2.15)	
		$h_l = .06 (k_l / D_l) (D_l C_{p_l} / \mu_l)^{.87} (\rho_l / \rho_g)^{.28} (C_p \mu / k_l)^{.4}$ (2.16)	
Liquid Holdup		Calculate parameter X and ΔP_{tp} (Equations (2.8) - (2.11), (2.4) - (2.6))	
Pressure Drop		$(\Delta P / \Delta z)_{tp} = \phi_g^2 (\Delta P / \Delta z)_g$ (3.48) or $(\Delta P / \Delta z)_{tp} = \phi_l^2 (\Delta P / \Delta z)_l$ (3.49)	

4. COMPUTER CODE DEVELOPMENT

4.1 Calculational Procedure

Assume that the tube is divided into a series of k local equilibrium cells, each cell of length $\Delta z = L/k$, where L is the heater tube length. The input to the first cell (at $z=0$) is the set of specified inlet conditions on temperature, pressure, compositions of both streams, and their flow rates. The heating medium is also specified (i.e., film coefficient h_o) and temperature is measured. The tube and its cells are illustrated in Figure 4.1. The heat flux to the cell, $\dot{q}_w$, is determined according to Equation (3.35), page 47. The streams from the cells are "forced" to equilibrium by the flash calculation, the composition being dependent on the temperature and pressure at the end of the cell (at $z + \Delta z$). As these two variables are not known at the outset, we can proceed by guessing a value for each, and then iterating until the material and energy balance equations are satisfied at the outlet from the cell. The problem is one of the double iteration type to achieve an acceptable level of convergence. The computational procedure includes the following steps:

- A. Specify the stream inlet (at $z_{k=0}$) and operating conditions: T_o , P_o , and mass flow rates
- B. Provide the first estimate of the stream temperature leaving the cell equal to temperature of entering streams, $T_{z_k} = T_{z_{k-1}}$

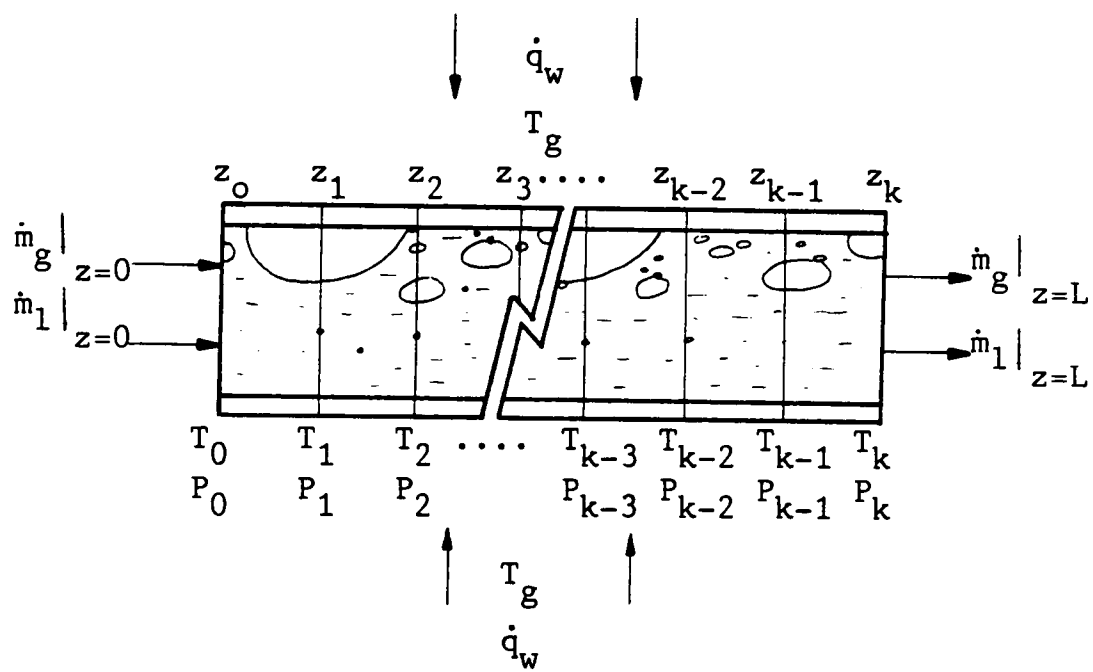


Figure 4.1. Tube and Cells for the Model Development

- C. Provide the first estimate of the stream pressure leaving the cell equal to pressure of the entering streams, $P_{z_k} = P_{z_{k-1}}$
- D. Calculate the flow rates of inlet and outlet streams of the cell by flash calculation [Table 2.4 and Equations (3.10) to (3.16)]. Average flow rates of the cell are taken from the inlet and outlet streams
- E. Find the Reynold number of the streams by Equations (3.50) and (3.51). Therefore flow regime is determined
- F. Using Lockhart-Martinelli correlations, Equations (2.4) to (2.14), (3.52), (3.53) and Figures 2.5 to 2.7 to find liquid holdup and pressure drop along the cell
- G. Compare the estimated pressure drop
- $$(\Delta P_{tp})_{est.} = P_z - (P_{z+\Delta z})_{\text{from Step C}}$$
- and the calculated pressure drop (given by Step F). If the difference between these two values is larger than 0.2% (pressure convergence criteria), estimate another value for pressure (e.g., $P_{new} = (P_{old} + P_{calc.})/2$), and return to Step C.
- H. Find the velocities of the streams which leads to the determination of flow patterns and flow models (the Mandhane flow pattern diagram, Figure 2.4)
- I. Calculate the heat flux to the cell, $\dot{q}_w$, by

Equations (3.36) to (3.38), page 47)

- J. Use $\dot{q}_w$ to find $T_{z+\Delta z}$ by trial-and-error in Equation (3.35), page 47. Compare $T_{z+\Delta z}$ with estimated a outlet streams temperature in Step B. If the difference between these two values is larger than 0.3% (temperature convergence criterion), estimate another value for outlet stream temperature (e.g., $T_{\text{new}} = (T_{\text{old}} + T_{\text{calc.}}) / 2$) and return to Step B.
- K. Find tube wall temperature by Equations (3.42) to (3.45), pages 50 and 51
- L. The conditions of the outlet stream of this cell are the conditions of the inlet stream of the next cell down the tube. Return to Step B and the cell by cell procedure is continued until the specified outlet temperature is reached. The tube length required to produce this temperature is then given by $k\Delta z$ (k is total number of cells in the system), summations of temperature increase, pressure drop, and net mass exchange between plases in each cell are the total changes in the system.

The overall block diagram to present these steps is illustrated in Figure 4.2.

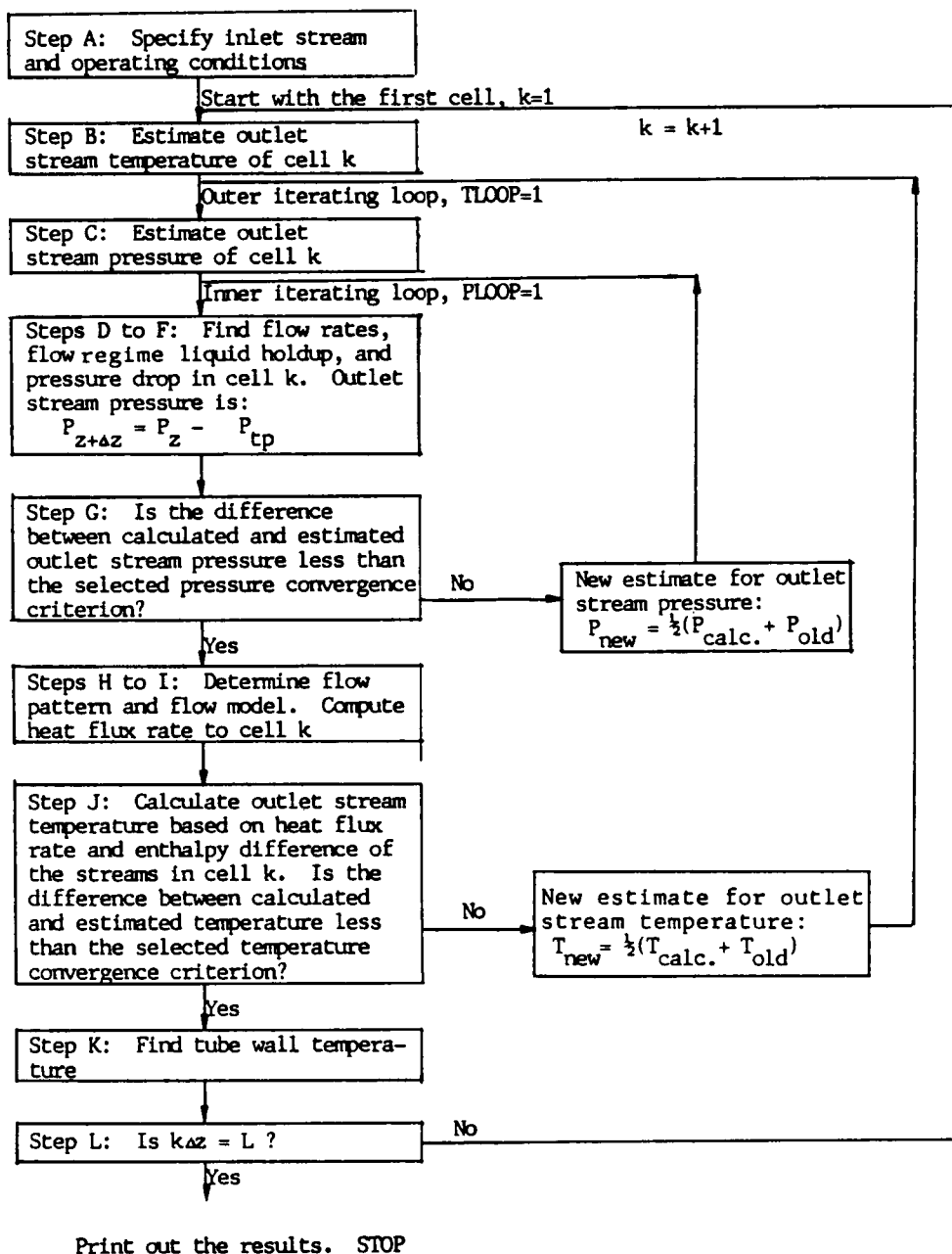


Figure 4.2. Overall Block Diagram of the Calculational Procedure

4.2 Computer Code

Referring to the calculational procedure established, Figure 4.1, FORTRAN IV Programs are developed for the design of a tubular heater for a two-phase, multicomponent, gas-liquid stream. The code is written for the use with the DEC 10 computer at the University of Tennessee, so that modifications may be required for use with other equipment. All equations used in the programs are described in Section 3 and the programs follow the design procedure outlined therein.

4.2.1 Organization

The computer code consists of a main program and several subprograms. The main program organizes and performs the calculations of pressure drop, temperature increase, mass exchange between phases, and flow rates for each cell and the entire system. It has two built-in iterating loops, the inner loop is for pressure computation and the outer loop is for temperature. Information such as K-values, physical property, heat of evaporation, ect. are supplied by the subprograms. Block diagram of the code is illustrated in Figure 4.3, and the programs and their functions are shown by Figure 4.4.

Except the first cell, the conditions of streams leaving one cell will be the conditions of the streams entering the next cell down the tube. Start from the first cell, temperature, pressure, and flow rates of the flow mixture are calculated from cell to cell. Both temperature and

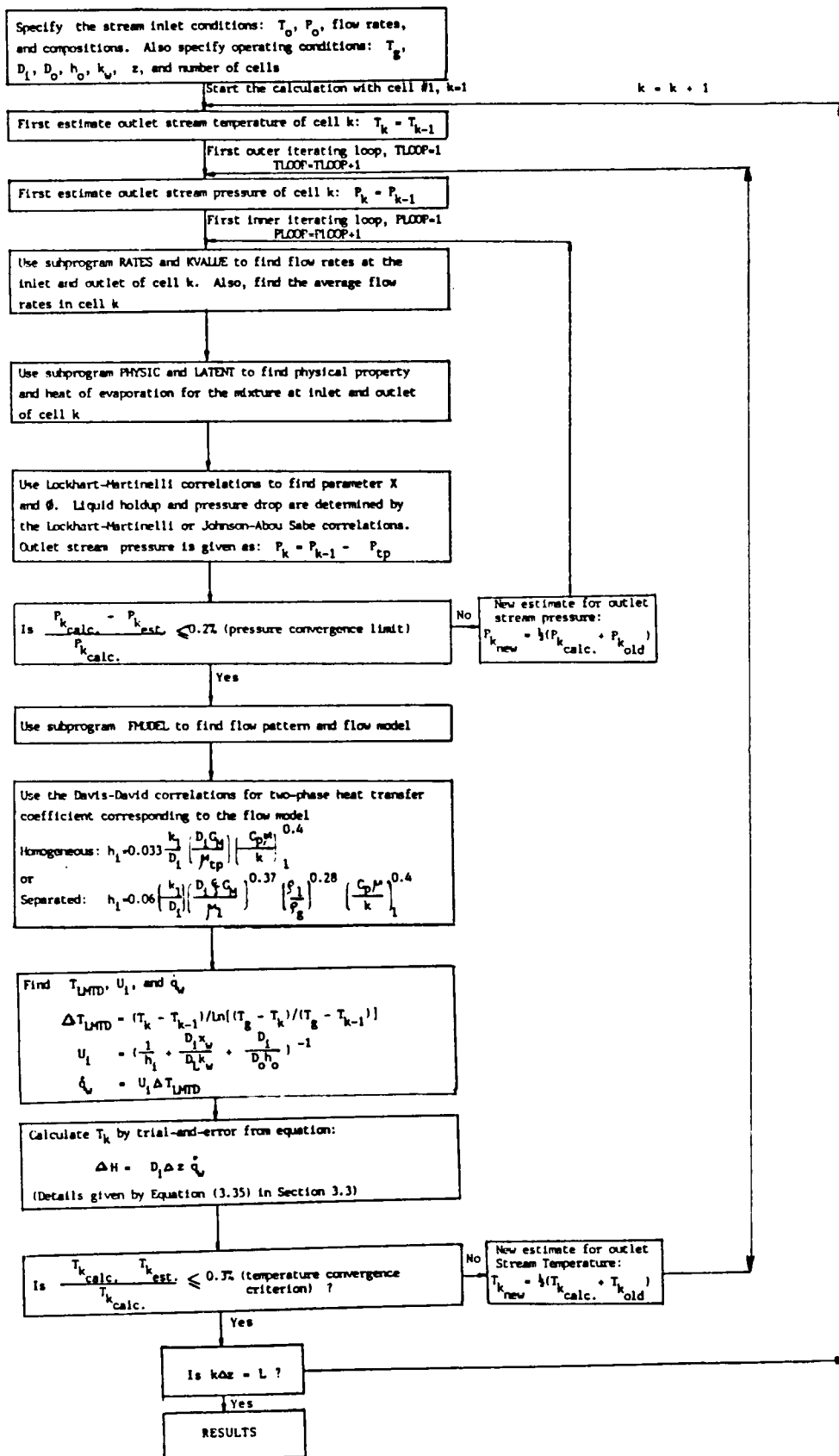


Figure 4.3. Block Diagram of the Computational Procedure

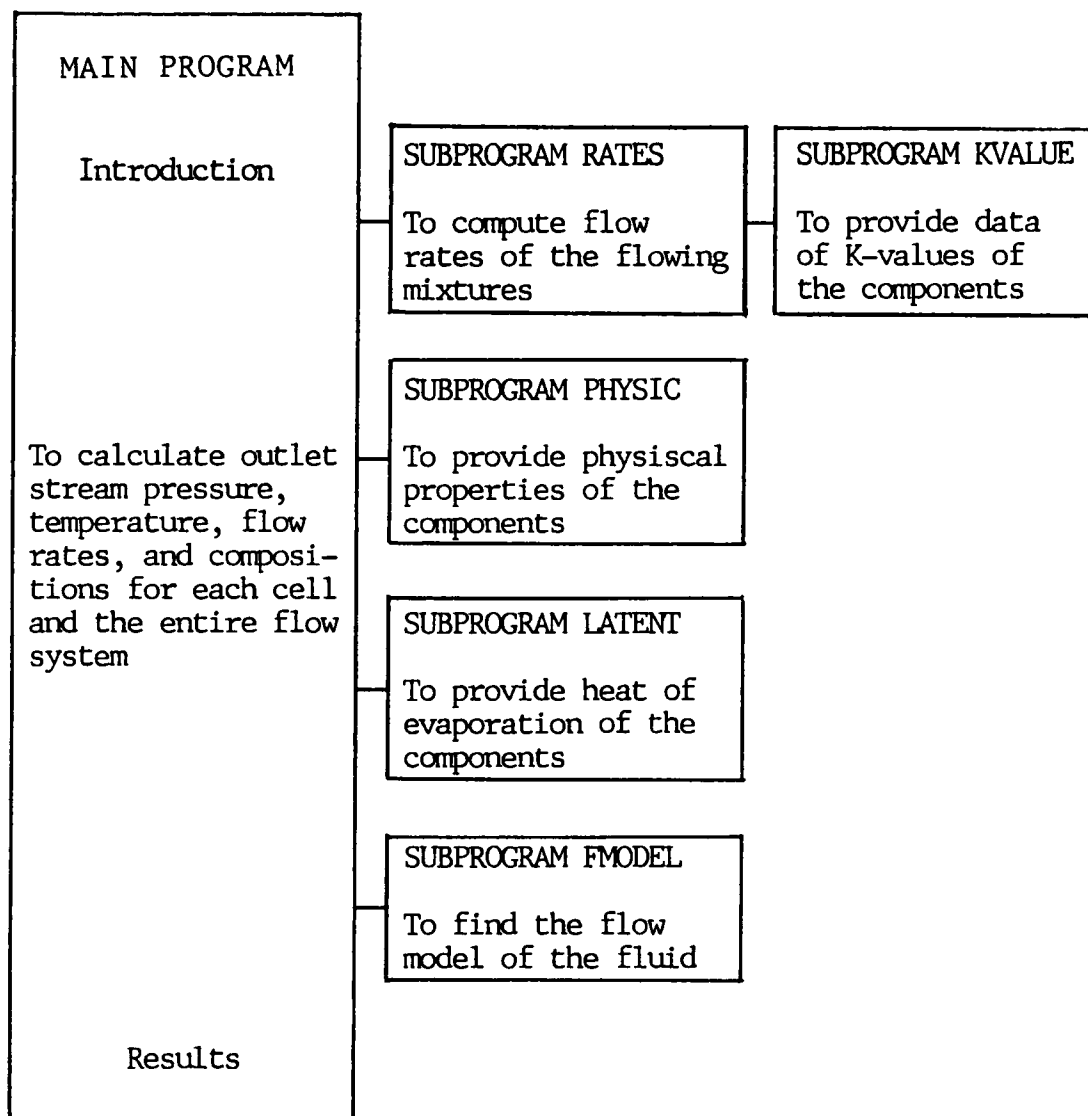


Figure 4.4. Main Program and Subprograms

pressure iterating loops are converged when the differences between the estimated and calculated values are 0.3% and 0.2%, respectively. This may lead to an average error between 1 and 5%. For example, a deviation of 10°F may be imposed on a calculated outlet stream temperature of 200°F . It is desired to keep these criterion as small as we can, even though the results are slightly different when these limits are changed (due to some errors are cancelled out during the accumulating calculation from cell to cell). These problems will be reviewed and illustrated when the code application is demonstrated in the next section.

Several subprograms are developed in order to support the calculation. In subprogram RATES, equilibrium flash calculations are employed. Figure 4.5 illustrates a block diagram of flash calculation procedure with K-values of the components independent of the compositions. K-values are supplied by subprogram KVALUE, which uses interpolation of pressure and temperature to approximate these values. Physical properties such as density, viscosity, specific heat, thermal conductivity, etc. are provided by subprogram PHYSIC. Except some properties which can be obtained by simple expressions (e.g., viscosity, specific heat), the others are obtained by using interpolations of the data corresponding to pressure and temperature. Heat of evaporation for components is given by subprogram LATENT. Depend on how accurate and convenient to supply the data to the calculations, we can

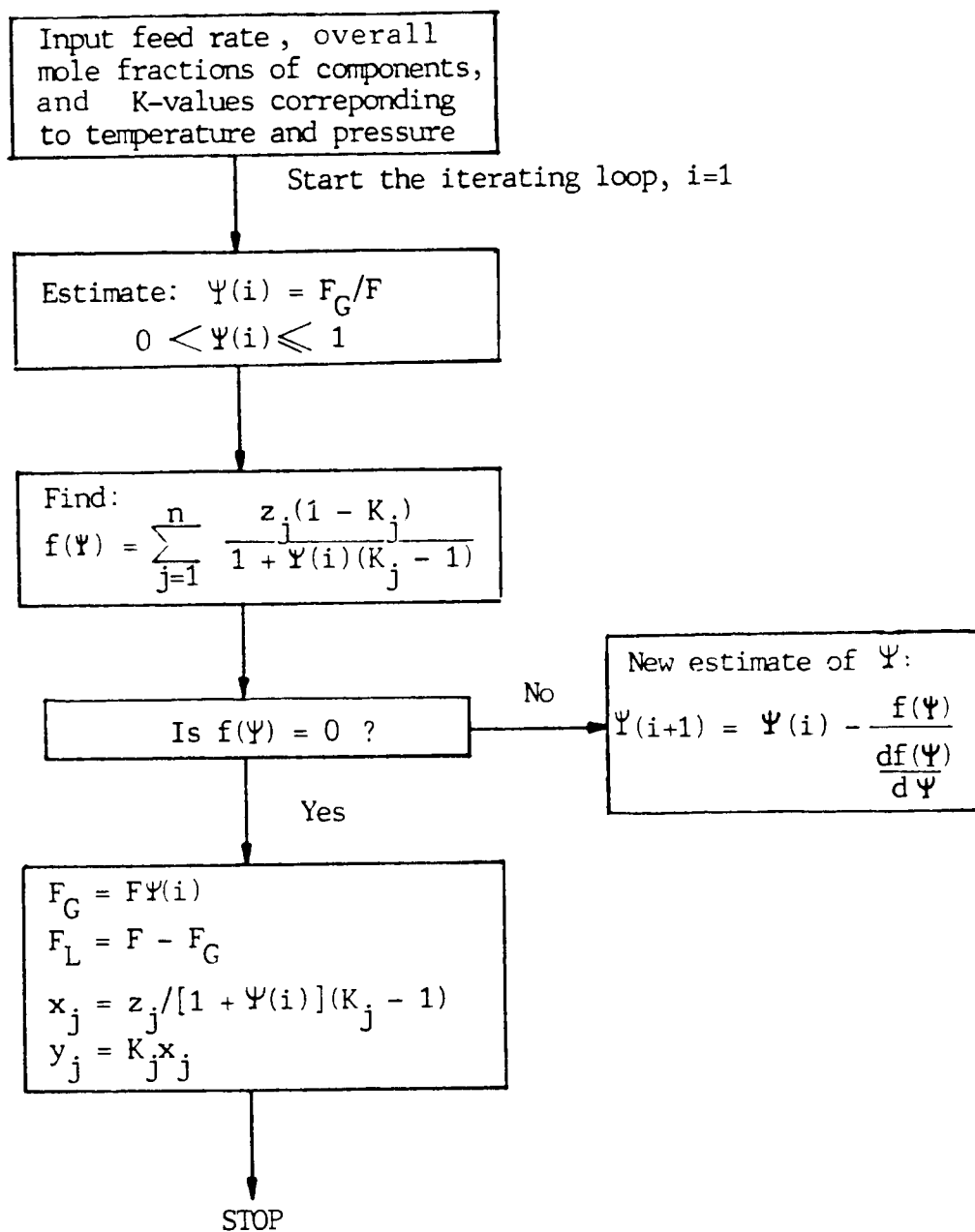


Figure 4.5. Block Diagram of Equilibrium Flash Calculation Procedure (Refer to Table 3.1)

develop subprograms KVALUE, PHYSIC, and LATENT in proper manners.

Subprogram FMODEL is used to determine the flow model by using the Mandhane map. In this program, the curves are divided into small lines and each line is presented by an equation in terms of gas and liquid superficial velocities.

4.2.2 Application

As data are not available the open literature against which the two phase, multicomponent model can be checked, a simplified version is constructed in order to provide results which can be checked against the limited amount of experimental data which has been published for the air-water system. Two programs are written for illustration of the developed calculation procedure. Following the steps as outlined in block diagrams, Figures 4.3 to 4.5, program FLOW1.FOR is utilized for a two-phase, two-component, air-water flow systems, and program FLOW1.FOR is developed for a two-phase, multicomponent, gas-liquid flow system. Details of these programs are provided in Appendix D (for both systems). Discussion of the results and the limits are introduced in the next section.

In program FLOW1.FOR, instead of using flash calculation we use a simple method by assuming ideal gas for the air-water vapor and the correlation of Keenan and Keyes⁽¹⁷⁾ as described by Equation 2.17, page 28. Most of the physical property data are provided by approximated correlations with temperature and pressure. Some others are obtained by interpolation

method. Physical properties of the components and their approximated equations are listed in Table C.1 to Table C.6 in Appendix C. Program FLOW1.FOR is applied to some experimental data available in the literature. Limits of the application are based on the literature used to develop the calculational procedure (e.g., Lockhart-Martinelli, Johnson-Abou Sabe, and Davis-David as shown in Table 2.3, page 25).

Program FLOW2.FOR is developed for a two-phase, multicomponent gas-liquid flow system. The operating conditions are constructed as following:

- Gas feed: 200 lb/hr with 70% of hydrogen and 30% of nitrogen, mass fraction basis
- Liquid feed: 2500 lb/hr with 60% of heptane and 40% of decane, mass fraction basis
- Inlet temperature 120°F and pressure 50 psia
- Heating medium temperature is 230°F, neglect heat transfer by radiation, and outside film heat transfer coefficient $h_o = 2000 \text{ BTU/ft}^2\text{-hr-}^\circ\text{F}$
- Tube is made of brass, tube inside diameter $D_i = 0.0725 \text{ ft}$, outside diameter $D_o = 0.0833 \text{ ft}$

Although the gas mass fraction in the feed ($\phi = 0.074$) is less than 0.1, under equilibrium conditions this ratio will be higher 0.1 due to the high volatility of the liquids.

In both program FLOW1.FOR and FLOW2.FOR, the first estimated temperature and pressure of the outlet streams of each cell are equal to those of inlet streams. Both pressure

and temperature iterating loops are converged when the differences between calculated and estimated values are less than the specified criterion. Details are provided along the constructed program bodies in Appendix D .

5. RESULTS AND INTERPRETATION

5.1 Air-Water Flow System

The first test of the model and computer code is made against the experimental data provided by Johnson-Abou Sabe⁽⁷⁾, Fried⁽⁸⁾, and Ebin⁽²⁴⁾ for the air water system. As not all experimental parameters required for a reasonably complete test of the model were published by these investigators, it is necessary to estimate certain values, using literature as a guide. In 6 runs of air and water in Table 2.5, page 31, air-water mixtures were heated by the steam jackets. However, neither temperature nor pressure of steam in the jackets were given, consequently we have to estimate the steam temperature and outside film heat transfer coefficient, h_o . The effect of h_o is displayed in Figure 5.1 by using Run No. 2 with the tube being divided into 10 cells and the convergence criteria for pressure and temperature iterating loops are 0.2% and 0.3% (these approaches are discussed latter). When h_o increases from 1000 to 3000 BTU/ft²-hr-°F, the range of values reported for dropwise condensation by McCabe and Smith⁽¹⁸⁾, the outlet stream temperature increases from 163 to 173°F. This effect is considered small compared to the wide range of h_o . It is clear that the outside film is not controlling the heat transfer rate and therefore little error is introduced by the use of a value of 2000.0

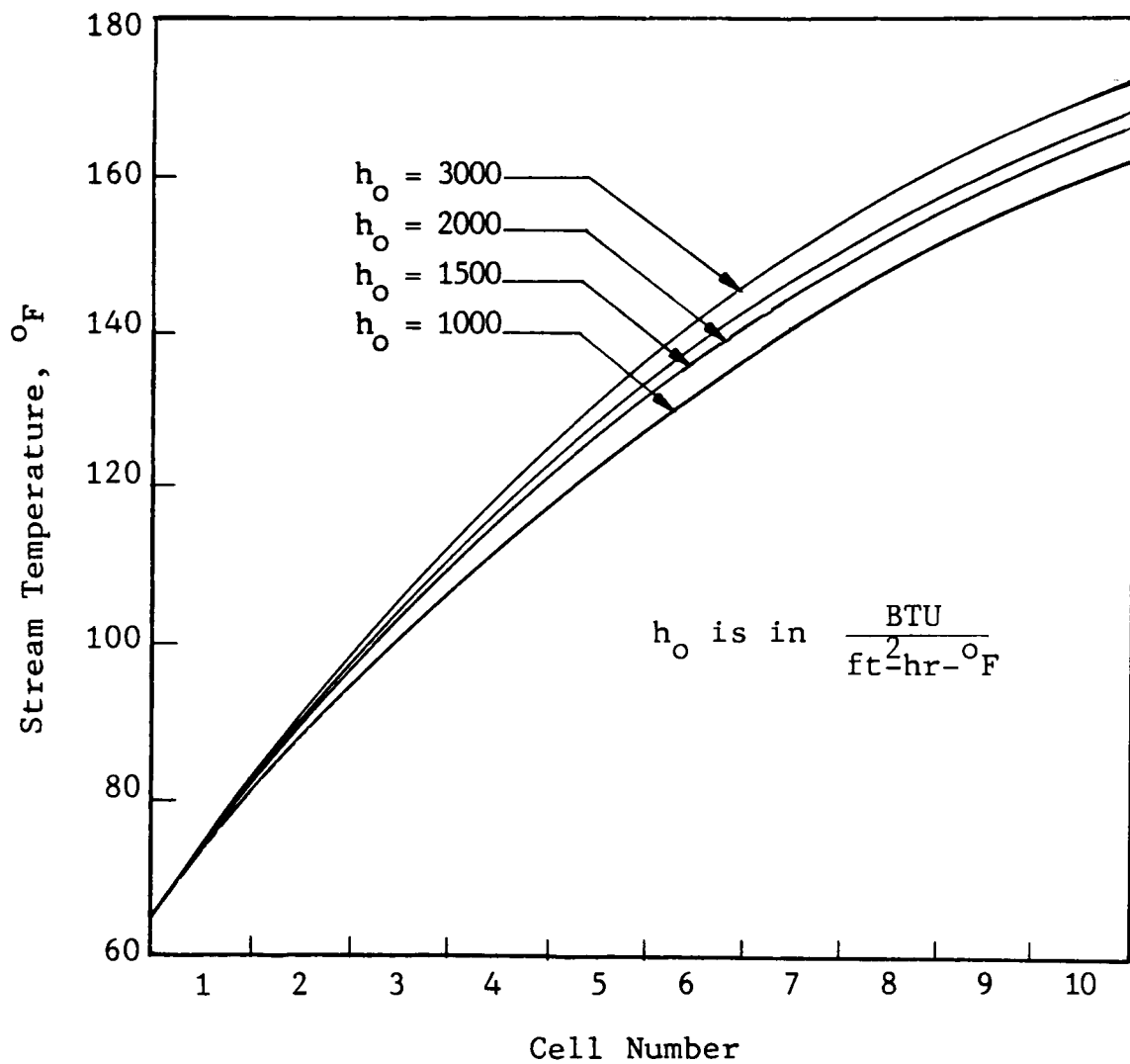


Figure 5.1. Effect of h_o Estimation on the Outlet Stream Temperature of Run No. 2

BTU/ft²-hr-°F. This value of h_o , of course, can be used for the other 5 runs because their operating conditions are similar.

As the steam side film coefficients are relatively large and as the average temperature of the tube walls were measured in some cases, these wall temperatures may be used as a guide in estimating temperature ranges for the steam, T_g , for use in Equations (3.42) to (3.47), pages 50 and 51.

In view of the incompleteness of the experimental data, two approaches are therefore taken in an attempt to test the heat transfer model and corresponding computer code. First, values of the steam temperature are searched for which the procedure air-water outlet temperature which corresponding approximately to the experimental values reported in Table 2.5, page 31, The average tube wall temperatures are computed and compared with the reported experimental wall temperatures. A reasonably close agreement between the computed wall temperature and the measurements would lend credibility to the model.

The second approach used to test the model is to search for steam temperatures which yield average values of the tube wall temperature which corresponds reasonably closely to the wall temperature measurements. In this approach, close agreement between the computed outlet two-phase mixture temperature and the measured values lends credibility to the model.

Another problem encountered here is how long a cell should be in order to obtain acceptable convergence to experimental results. In Figure 5.2, temperature profiles of the mixture along the tube, and in Figure 5.3 outlet mixture temperature of Run No. 2 corresponding to number of cell in tube are indicated (in these calculations, $h_o = 2000$ BTU/ft²-hr-°F, convergence criteria for pressure iteration is .2%, and for temperature is 0.3%). Both Figures 5.2 and 5.3 show that the outlet mixture temperatures (of cell and entire system) reach a consistent degree when number of cell varies from 5 to 20 cells (equivalent to the cell lengths of 0.8 to 3 feet). Out of this range, the results tend to have higher error due to the inaccuracy of taking average of flow properties in a long cell or accumulation of a large number of iterating loops (of pressure and temperature) if the cell is short. Based upon these results, we establish a ten-cell tube (cell length of 1.57ft in this case) as a standard in this series of calculations, relative errors of the other systems comparing to this selected are illustrated in Figure 5.4.

A range is values for convergence criteria for both temperature and pressure were explored. The range explored for temperature is from 30% down to 0.03% and for pressure, from 0.2% down to 0.01%. As can be seen from the results

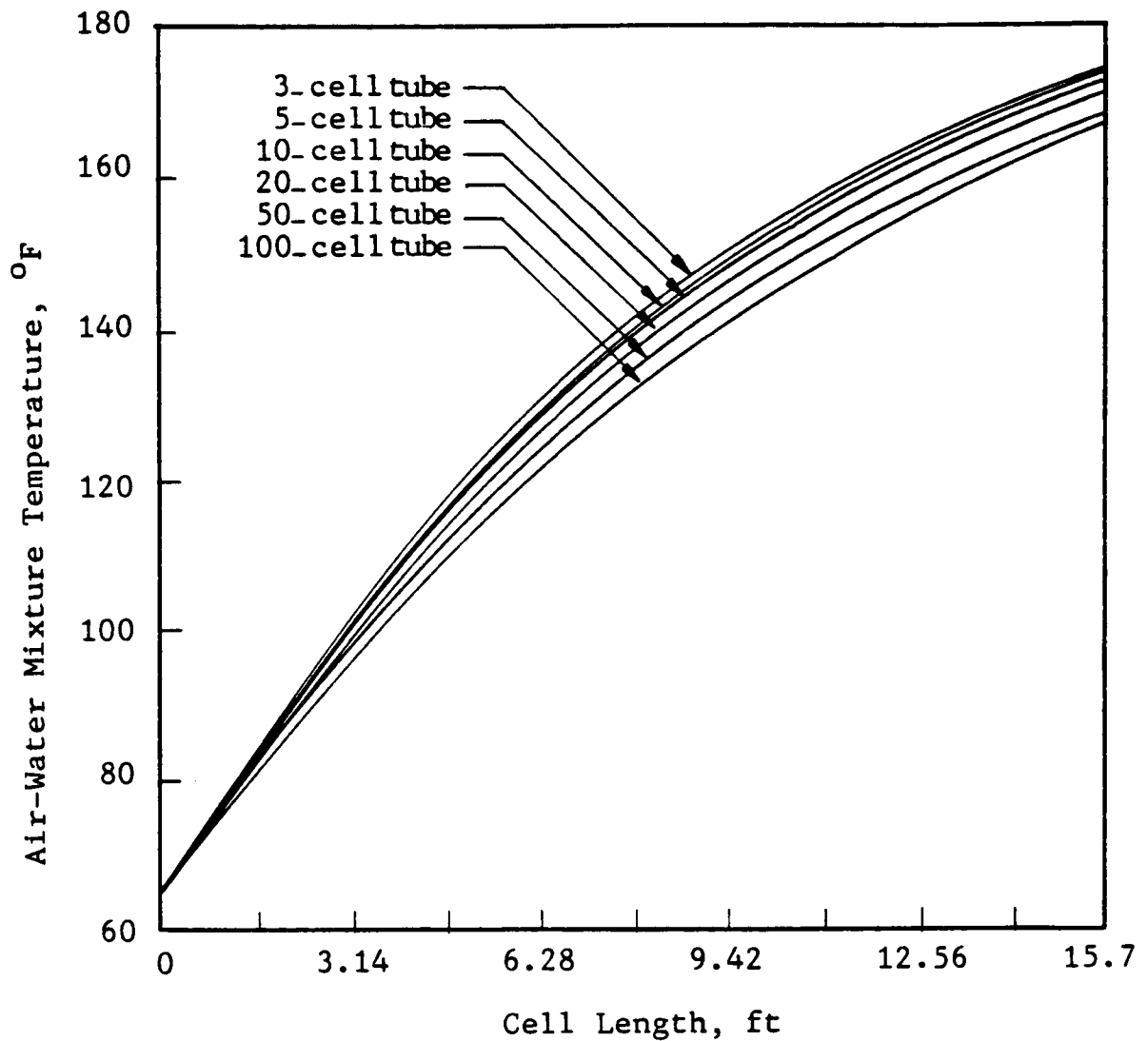


Figure 5.2. Temperature Profiles of the Air-Water Mixture Corresponding to Number of Cells in a Tube of Run No. 2 (Using Johnson and Abou-Sabe Correlations)

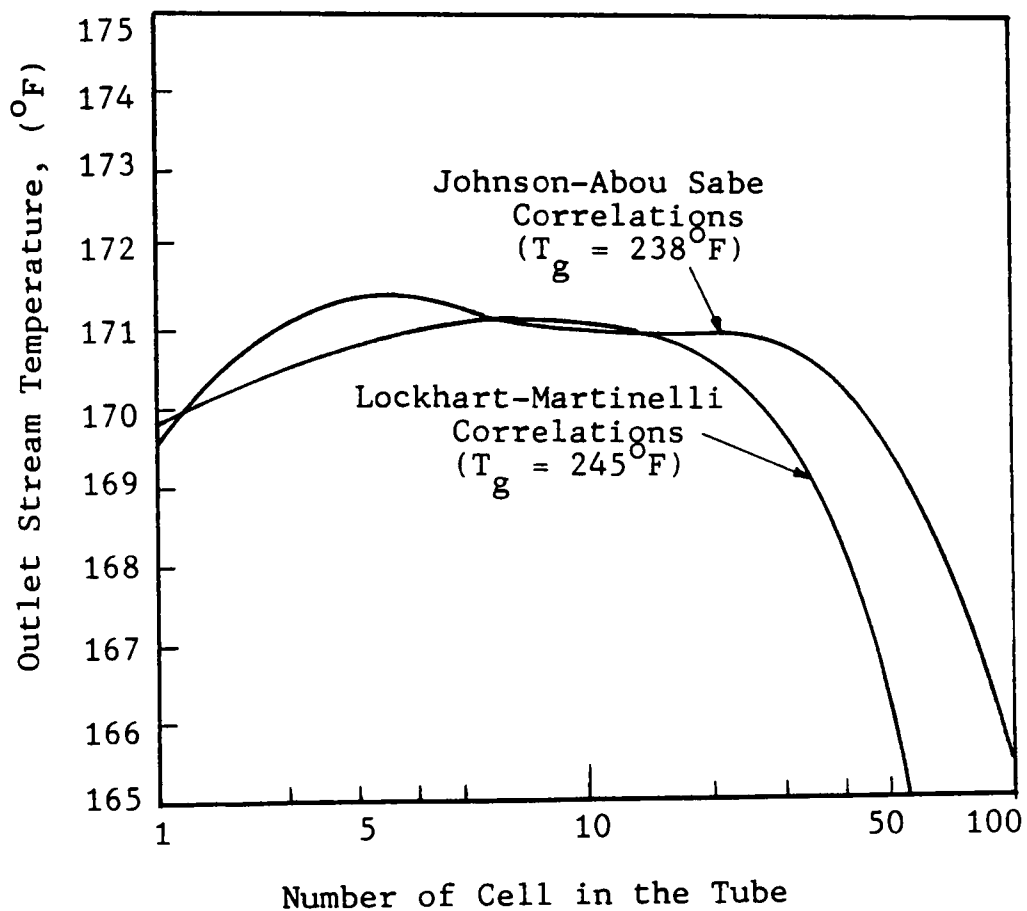


Figure 5.3. Outlet Stream Temperature Corresponding to Number of Cell in the Tube Given by Lockhart-Martinelli and Johnson-Abou Sabe Correlations (Data of Run No. 2)

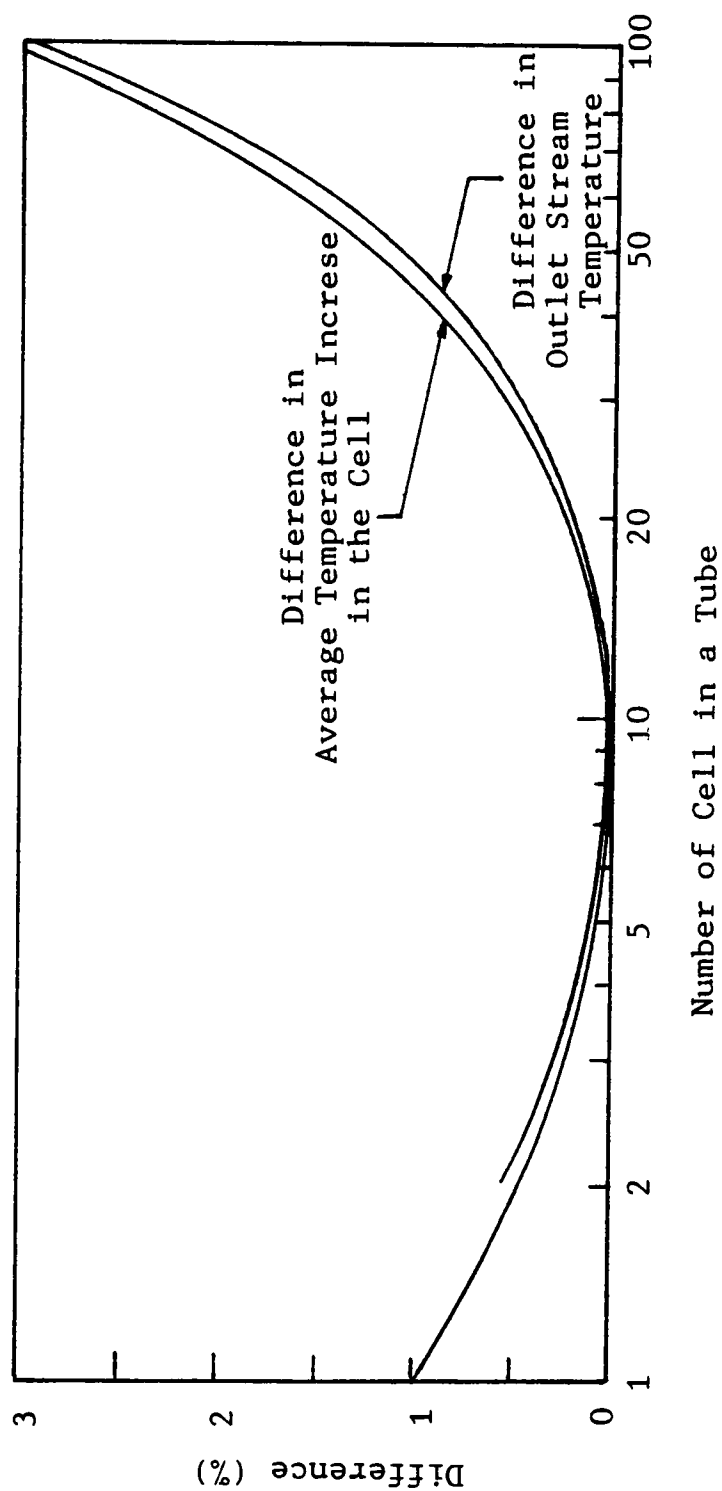


Figure 5.4. Comparisons of Average Temperature Increase in the Cell and Outlet Fluid Temperature by Using the 10-Cell Tube as the Standard.

shown in Figure 5.5, values on temperature convergence in the range of 0.5% - 0.03%, for pressure convergence criteria in the the above range affect the outlet temperature by less than about 2 - 3°F, which is expected to be within the experimental measurment of the data reported in Table 2.5, page 31. Therefore in order to minimize computer time, and still provide adequate accuracy. The following value of the convergence criteria were selected:

Temperature, $C_{ti} = 0.3\%$

Pressure, $C_{pi} = 0.2\%$

From the above discussion, calculations for all 6 runs of air and water are performed (using $h_o = 2000 \text{ BTU/ft}^2\text{-hr-}^\circ\text{F}$, 10-cell tube, 0.2% and 0.3% for pressure and temperature convergence criteria, respectively). Heat transfer by radiation to the outer tube wall in these cases is negligible. Physical property data supporting the calculation are given in Appendix C. In all air-water runs, temperatures of steam are taken as the heating medium temperature T_g . Since both the steam jacket pressure and temperature were not published, values of T_g are estimated for each run and assumed adequate when calculated outlet temperature became equal to the experimental temperature. Both Lockhart-Martinelli and Johnson-Abou Sabe correlations of liquid holdup and pressure drop are used and their results are separately presented in Table 5.1 and Table 5.2. A review of the values in

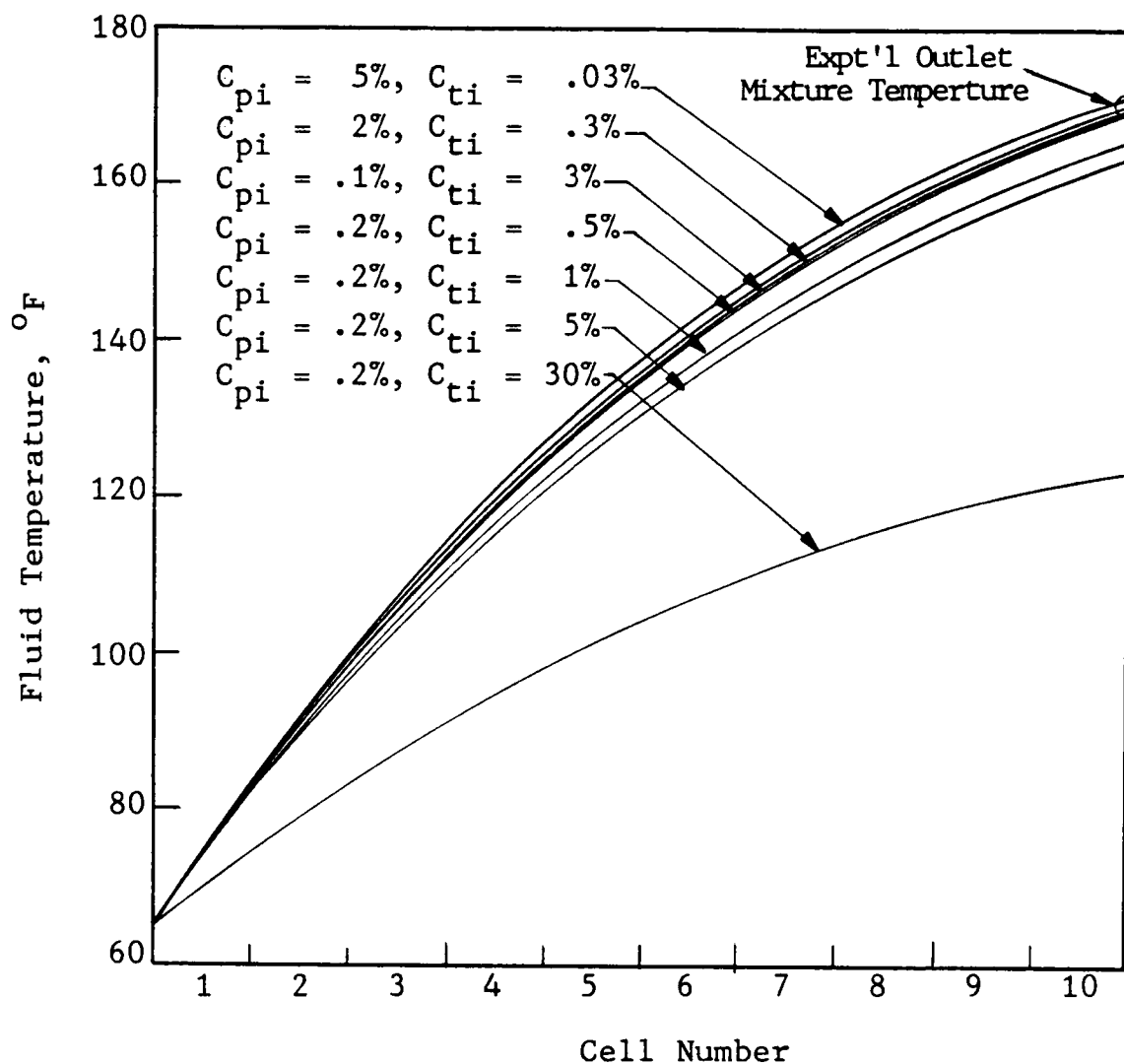


Figure 5.5. Effects of Temperature and Pressure Convergence Criteria on the Calculated Temperature of the Fluid

Table 5.1. Comparison of Experimental and Calculated Results for Air-Water System Based on the Liquid Holdup and Pressure Drop Correlation of Lockhart-Martinelli (6) - Matching Outlet Stream Temperature Calculational Model

Run No.	Inlet Stream		Outlet Stream Conditions						Avg. T_y		Est.	Outlet Flow Rate		R_1		Regime		Pattern		h_l BTU/ft ² -hr-°F
	Temp. °F	Press. Psia	Expt.		Calc.		Expt. °F	Calc. °F	T_g	Water lb/hr		Air lb/hr	Expt.	Calc.	Expt.	Calc.	Expt.	Calc.		
			Temp. °F	Press. Psia	Temp. °F	Press. Psia														
1	64.5	26.7	179.6	20.7	179.0	23.94	196	267.2	286	938.81	169.79	.08	.120	T-T	T-T	Ann.	Sep.	330		
2	65.0	30.6	171.3	22.9	170.6	26.94	194	226.1	244	939.10	237.00	.04	.100	T-T	T-T	Ann.	Sep.	437		
3	75.8	43.8	178.0	28.0	177.9	34.72	202	226.4	250	1344.48	221.72	.12	.137	T-T	T-T	Ann.	Sep.	735		
4	82.0	34.0	183.0	23.5	182.6	28.29	(a)	238.1	256	967.50	167.90	(a)	.128	T-T	T-T	Ann.	Sep.	530		
5	81.0	38.9	175.0	24.2	174.2	30.96	(a)	210.8	228	966.28	236.62	(a)	.109	T-T	T-T	Ann.	Sep.	670		
6	96.0	48.0	180.0	30.4	180.0	35.27	(a)	231.4	260	1966.35	236.05	(a)	.162	T-T	T-T	Ann.	Sep.	755		

(a) Not available

(b) Calculated by the authors

Table 5.2. Comparison of Experimental and Calculated Results for Air-Water System Based on the Liquid Holdup and Pressure Drop Correlation of Johnson-Abou Sabe⁽⁷⁾ - Matching Outlet Stream Temperature Calculational Model

Run No.	Inlet Stream		Outlet Stream Conditions				Avg. T_g		Est.	Outlet Flow Rate		R_L		Regime		Pattern		h_i BTU/ft ² -hr-°F
			Expt.		Calc.		Expt.	Calc.		Water	Air	Expt.	Calc.	Expt.	Calc.	Expt.	Calc.	
	Temp. °F	Press. Psia	Temp. °F	Press. Psia	Temp. °F	Press. Psia	°F	°F	lb/hr	lb/hr								
1	64.5	26.7	179.6	20.7	178.9	21.94	196	255.3	275	934.04	174.56	.08	.097	T-T	T-T	Ann.	Sep.	390
2	65.0	30.6	171.3	22.9	171.3	24.26	194	219.3	238	932.91	243.19	.04	.084	T-T	T-T	Ann.	Sep.	510
3	75.8	43.8	178.0	28.0	178.1	28.44	202	221.8	247	1335.21	230.99	.12	.110	T-T	T-T	Ann.	Sep.	833
4	82.0	34.0	183.0	23.5	182.2	24.29	(a)	231.2	250	960.09	175.31	(a)	.100	T-T	T-T	Ann.	Sep.	600
5	81.0	38.9	175.0	24.2	174.5	25.30	(a)	209.2	228	995.67	247.23	(a)	.088	T-T	T-T	Ann.	Sep.	780
6	96.0	48.0	180.0	30.4	180.8	26.06	(a)	230.4	262	1948.50	253.90	(a)	.130	T-T	T-T	Ann.	Sep.	870

(a) Not available

Table 5.1 shows that calculated pressure drop (given by the differential pressure of the inlet and outlet streams) is higher and liquid holdup is lower than the experimental values. Results computed by the Johnson-Abou Sabe correlations in Table 5.2 are in a better agreement with experiments. Obviously, mass exchanges between phases and changes in fluid properties due to the amount of heat added to the system play an important role in the determination of liquid holdup and pressure drop.

All six air-water runs have the same flow regimes (both gas and liquid are turbulent) and models (separated). The two-phase heat transfer coefficients, h_i , given by the Davis-David correlations^(10,11) show that heat transfer is controlled by the inside film. However, these values are low comparing to the experimental values as shown in Table 2.4. The high values of h_i in the literature may be due to the fact that they are computed from measurement of heat loss of steam in the jacket^(7,8). Similarly, the high heat flux rate of the experiments can be explained by the same manner.

The other experimental measurement we can use to compare with our results is the tube wall temperature. Even with a large deviation in the mean of literature tube wall temperature⁽⁷⁾, the calculated values are still higher by about 20 to 30°F. An examination of the flow rates in

Table 5.1 and Table 5.2 show that the calculated wall temperature tends to be high when both flow rate and gas mass fraction are low (Run No. 1, 4, and 6). As previously mentioned, tube wall temperature is a dependent variable of the inside heat transfer coefficient, and the Davis-David correlations^(10,11) for this coefficient are based on the systems which have high gas mass fraction and flow rates. Higher error in the calculated tube wall temperatures of Runs No. 1, 4, and 6 (comparing to Runs No. 2, 3, and 5) may be due to these flow conditions.

Also, there is some uncertainty in the literature value of the liquid holdup of Run No. 2 (which was calculated by the authors), even though the air rate of Run No. 2 is higher than that of Run No. 1 by a factor of 1.5, liquid holdup of Run No. 2 is unlikely equal to a half the liquid holdup of Run No. 1.

In the other approach we search for a temperature of steam in the jacket such that the average tube wall temperature would have been matched. Calculated results based on this approach are listed in Table 5.3 and Table 5.4. The heating medium temperatures of Run No. 1 and Run No. 2 are very low (hot water instead of steam in the jacket). Even with the assumption of maximum tube wall temperature (from the deviance of $\pm 10^{\circ}\text{F}$)⁽⁷⁾, the calculated outlet fluid temperatures are still much lower than experimental values,

Table 5.3. Comparison of Experimental and Calculated Results for Air-Water System Based on the Liquid Holdup and Pressure Drop Correlations of Lockhart-Martinelli⁽⁶⁾ - Matching Average Tube Wall Temperature Calculational Model

Run No.	Inlet Stream		Outlet Stream Conditions				Avg. T_w		Est. T_g	Outlet Flow Rate		R_1		Regime		Pattern		h_1 BTU/ft ² -hr-°F
	Temp. T_{wF} Psia	Press. P_{wF} Psia	Expt.		Calc.		Expt. T_{wF} °F	Calc. T_{wF} °F		Water lb/hr	Air lb/hr	Expt.	Calc.	Expt.	Calc.	Expt.	Calc.	
			Temp. T_{wF} °F	Press. P_{wF} Psia	Temp. T_{wF} °F	Press. P_{wF} Psia												
1	64.5	26.7	179.6	20.7	141.5	24.20	196	197.0	208	963.40	145.20	.08	.126	T-T	T-T	Ann.	Sep.	292
2	65.0	30.6	171.3	22.9	151.9	27.10	194	194.5	208	954.30	221.80	.04	.106	T-T	T-T	Ann.	Sep.	410
3	75.8	43.8	178.0	28.0	160.3	40.17	202	202.8	220	1359.14	207.06	.12	.143	T-T	T-T	Ann.	Sep.	518

(*) Calculated by the authors

Table 5.4. Comparison of Experimental and Calculated Results for Air-Water System Based on the Liquid Holdup and Pressure Drop Correlations of Johnson-Abou Sabe^(*) - Matching Average Tube Wall Temperature Calculational Model

Run No.	Inlet Stream		Outlet Stream Conditions				Avg. T_w		Est. T_g	Outlet Flow Rate		R_1		Regime		Pattern		h_1 BTU/ft ² -hr-°F
	Temp. °F	Press. Psia	Expt.		Calc.		Expt.	Calc.		Water	Air	Expt.	Calc.	Expt.	Calc.	Expt.	Calc.	
			Temp. °F	Press. Psia	Temp. °F	Press. Psia												
1	64.5	26.7	179.6	20.7	148.2	22.34	196	197.6	210	959.50	149.10	.08	.102	T-T	T-T	Ann.	Sep.	345
2	65.0	30.6	171.3	22.9	156.8	24.45	194	195.0	210	948.01	228.09	.04	.086	T-T	T-T	Ann.	Sep.	480
3	75.8	43.8	178.0	28.0	165.8	37.78	202	201.3	220	1355.30	210.93	.12	.118	T-T	T-T	Ann.	Sep.	611

(*) Calculated by the authors

being as much as 40°F too low using the Lockhart-Martinelli correlations (Table 5.3) and as 30°F too low, using the Johnson-Abou Sabe correlations (Table 5.4).

The consequence is that the model, using the established correlations and corresponding computer code, predicts a longer tube required to heat the two phase mixture to a given outlet temperature than is achieved experimentally, clearly an inconsistency. It is therefore necessary to examine the assumptions in which the model is based in an effort to determine the principal cause of this inconsistency.

Of the assumptions listed in Section 4.1, Assumption F, that local equilibrium is achieved between the gas and liquid phases, is the one assumption under these experimental conditions in which we have least confidence. This assumption implies that there is negligible transfer resistance for heat and mass between phases. Therefore one would expect should this be true, the model would underpredict the tube length required to heat the mixture to a specified outlet temperature, contrary to the actual result. It is therefore unlikely that the short coming in the model is due to Assumption F (local equilibrium).

Another possible source of the discrepancy between the model predictions and the experimental results for the air-water system is that the controlling heat transfer coefficient, the inner two-phase coefficient h_i , is too low.

Davis and David^(10,11) suggested that the error band on their correlations, Equations (2.15) and (2.16), page 24, may be as much as ± 10 to 30%, due to data scatter, which is not surprising in view of the fact that data scatter in establishing single-phase film heat transfer coefficients frequently is as high as $\pm 20^{\circ}\text{F}$. It is therefore of interest to test whether or not an increase of about 20% in the two-phase heat transfer h_i is sufficient to produce better agreement between the model results and experimental results. An increase in h_i of this magnitude is still within the error limits of the correlations, Equations (2.15) and (2.16), as suggested by Davis and David. Calculations were therefore repeated for Run No. 1 to 3, Table 2.5, using $h'_i = 1.2h_i$, where h_i is computed according to Equations (2.15) and (2.16). The results are compared with the experimental data in Table 2.5, page 31, using the Johnson-Abou Sabe holdup and pressure drop correlations, and in Figure 5.6 for Run No. 2. These calculations were performed using only the Johnson-Abou Sabe correlations, as these equations have generally provided predictions in closer agreement with the experimental measurements than have been achieved by use of the Lockhart-Martinelli correlations. The approach taken in these calculations was to search for a steam temperature such that the computed average tube wall temperature, T_w , agreed approximately with the average measured wall temperatures.

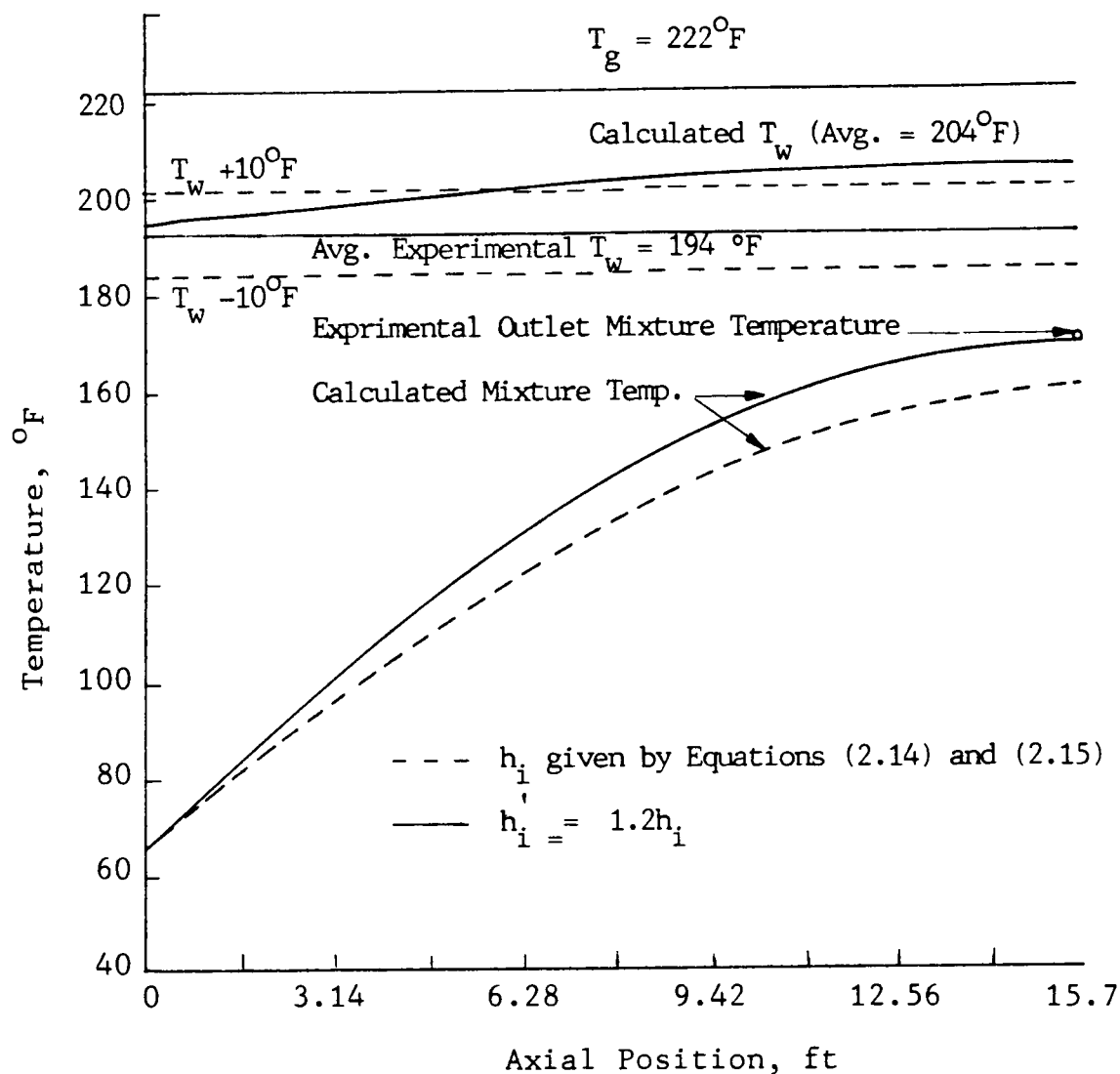


Figure 5.6. Mixture Temperature of Run No. 2 for Air-Water Flows Computed by the Johnson-Abou Sabe Correlations for Liquid Holdup and Pressure Drop-Matching Tube Wall Temperature Model (h_i Given by the Davis-David Correlation Comparing to $h'_i = 1.2 h_i$)

On comparing the calculated results and the experimental measurements, Table 5.5, one sees reasonable good agreement between the outlet steam temperature for Runs NO. 2 and 3 and for the outlet pressure in all these runs. The maximum error between calculated and measured outlet pressure is about 5.7% for Run No. 2. The largest discrepancy between calculated and experimental values is for the mixture outlet temperature for Run No. 1, the calculated value of 162°F being 17.6°F lower than the experimental value of 179.6°F . This difference may not be due entirely to weakness in the model (flow rate and gas mass fraction ranges of the Davis-David correlation), as the experimental value of 179.6°F seems inconsistently high, when the experimentally determined average tube wall temperature are compared. For example, the wall temperature of 194°F for Run No. 1 is only two degrees higher than for Run No. 2, yet the mixture outlet temperature is 8.6°F higher for this run than for Run No. 2.

Figure 5.6 shows the axial temperature profiles for Run No. 2. Two stream temperature profiles are provided, one using h_i as computed from Equations (2.15) and (2.16), the other computed using $h'_i = 1.2h_i$. Here one sees that the stream temperature approaches to within 1°F of the experimentally determined temperature at the tube outlet for this run when the higher value of the inner film coefficient is

Table 5.5. Comparison of Experimental and Calculated Results for Air-Water Systems Based on the Liquid Holdup and Pressure Drop Correlations of Johnson-Abou Sabe (7) - Matching Average Tube Wall Temperature Model and h_i (Given by the Davis-David(10,11) Correlations) Increases by a Factor of 1.2

Run No.	Avg. Gas Mass Fraction	Outlet Stream Temp., °F		Outlet Stream Press., psia		Avg. T_w , °F		T_g °F	
		Expt.	Calc.	Expt.	Calc.	Expt.	Calc.	Expt.	Calc.
1	.1275	179.6	162	20.7	21.5	196 ± 10	206.0	222	
2	.1830	171.0	170	22.9	24.2	194 ± 10	203.7	222	
3	.1284	178.0	178	28.0	28.3	202 ± 10	207.3	232	

Note: Run No.1 has low flow rates

used, in contrast to the result for the case when the lower value of h_i is used. Also, the calculated tube wall temperature generally falls within the error band on the measured average tube wall temperatures. These results lead to the conclusion that the heat transfer model developed in this work provides results in reasonably good agreement with experimental data for the air water system, when the Johnson-Abou Sabe pressure drop and holdup correlations and the Davis-David two-phase heat transfer correlations modified by a factor of 1.2 are used for the film coefficient h_i , this value is within the error band on h_i .

The further details of the flowing mixtures corresponding to the axial positions on the tube are illustrated by using Run No.2 of air-water flow. The liquid holdup and pressure drop correlations of Lockhart-Martinelli and Johnson-Abou Sabe are used in both models of searching for the steam jacket temperatures in order to match the outlet stream temperatures and the average tube wall temperatures (h_i increases by a factor of 1.2 for the latter model). Comparisons and results of these calculations are presented in Appendix B. Also, some results of Run No.1 and Run No.3 with their corresponding modified factors are shown in this appendix.

5.2 Gas-Hydrocarbon System

In order to progress to a further application of the developed procedure to the flow of gas and slurry in the

preheater in coal liquefaction process, several modification, must be established (e.g., correlations for non-Newtonian liquids, coal properties, holdup, pressure drop, flow patterns, flow regimes, etc.). Due to the insufficient data for such systems available in the open literature, a complex system of multicomponent, high volatility, gas-hydrocarbon liquid system is instituted in the view of restrictions and assumptions described in Section 3.1. Intentionally, several conditions such as flow rates, heating medium temperature, h_o , tube sizes, cell length, etc. are chosen from the examinations of air-water flows developed earlier. Details of this complex system as inlet flow rates, compositions, temperature, pressure, and operating conditions as tube size, jacket steam temperature, etc. are given in Table 5.6. With physical property data of the components are provided in Appendix C, pressure and temperature convergence criteria are 0.2% and 0.3%, respectively, calculated results of this multicomponent system are presented in Table 5.7 and Table 5.8 corresponding to the pressure drop and liquid holdup correlations of Lockhart-Martinelli and Johnson-Abou Sabe. More details of this calculation are presented in Appendix D.

The gas mass fraction in the feed is low (7.4%), however when equilibrium is presumed, a large amount of liquid becomes vapor in the gas phase, and consequently average gas mass fraction of the first cell rises up to

Table 5.6. Inlet Stream and Operating Conditions for the Two-Phase, Multicomponent,
Gas-Liquid Flow

Component No.	Component	Feed Rate, lb/hr		
		Gas	Liquid	
1	Hydrogen	140	0	Inlet Stream Temperature 80°F Inlet Stream Pressure 60 psia Cell Length 1.57 ft Number of Cell 10 cells Tube Inside Diameter .87 inch Tube Outside Diameter 1.0 inch Tube is Made of Brass Tube Wall Thermal Conductivity 64 BTU/ft-hr-°F Heating Medium Temperature 230°F Outside Film Heat Transfer Coefficient 2000 BTU/ft ² -hr-°F
2	Nitrogen	60	0	
3	Heptane	0	1500	
4	Decane	0	1000	
Total		200	2500	

Table 5.7. Result of the Multicomponent System Based on the Lockhart-Martinelli (6) Correlations for Liquid Holdup and Pressure Drop

Cell No.	Temp. °F	Outlet Fluid		Avg. V_{TP} ft/Sec	h_i $\frac{\text{ft}^2\text{-hr}-\text{F}}{\text{FTO}}$	U_i $\frac{\text{FTO}}{\text{ft}^2\text{-hr}-\text{F}}$	Avg. T_{OK}	Avg. R_1	Gas Mass Fraction	Flow Regime	Flow Pattern (*)
		Press. psia	m_1 lb/hr								
1	102.7	58.96	2184.23	415.77	2252	1044	161.4	.1638	.1422	T-T	Ann.
2	120.8	57.77	2079.57	520.43	2805	1150	165.0	.1460	.1800	T-T	Ann.
3	135.6	56.41	1974.34	625.66	3355	1232	169.6	.1295	.2204	T-T	Ann.
4	147.0	54.86	1866.71	733.29	3877	1296	174.3	.1160	.2613	T-T	Ann.
5	156.1	53.15	1760.18	839.82	4366	1346	178.6	.1040	.3025	T-T	Ann.
6	164.1	51.26	1660.85	939.15	4813	1386	182.7	.0939	.3421	T-T	Ann.
7	170.4	49.21	1564.49	1035.51	5230	1420	186.4	.0854	.3797	T-T	Ann.
8	176.2	47.01	1476.54	1123.46	5630	1446	190.0	.0780	.4152	T-T	Ann.
9	180.9	44.65	1392.87	1207.13	6008	1470	192.8	.0715	.4482	T-T	Ann.
10	185.2	42.15	1314.81	1285.19	6370	1490	195.5	.0660	.4793	T-T	Ann.

(*) Computed from superficial velocities of gas and liquid phases and the Mandhane Map

Table 5.8. Result of the Multicomponent System Based on the Johnson-Abou Sabe (7) Correlations for Liquid Holdup and

Pressure Drop												
Cell No.	Outlet Fluid				Avg. V_{TP} ft/sec	h_i $\frac{ft}{hr} \sqrt{\frac{lb}{ft^2}}$	U_i $\frac{ft}{hr} \sqrt{\frac{lb}{ft^2}}$	Avg. T_{OF}	Avg. R_L	Gas Mass Fraction	Flow Regime	Flow Pattern (*)
	Temp. O_F	Press. psia	$\dot{m}_l$ lb/hr	$\dot{m}_g$ lb/hr								
1	103.9	58.37	2175.24	424.76	28.12	2636	1220	156.3	.1360	.1440	T-T	Ann.
2	122.3	56.47	2061.55	538.45	31.80	3307	1225	161.0	.1193	.1850	T-T	Ann.
3	137.3	54.26	1949.06	650.94	35.85	3960	1305	166.6	.1050	.2290	T-T	Ann.
4	148.1	51.74	1829.62	770.38	40.18	4567	1365	172.0	.0929	.2733	T-T	Ann.
5	156.5	48.90	1710.87	889.13	44.85	5130	1411	176.3	.0823	.3191	T-T	Ann.
6	163.4	45.73	1596.66	1003.34	49.90	5648	1447	180.2	.0733	.3640	T-T	Ann.
7	169.3	42.24	1489.11	1110.89	55.30	6134	1477	183.7	.0655	.4070	T-T	Ann.
8	174.2	38.42	1385.98	1214.02	61.20	6590	1502	186.8	.0587	.4471	T-T	Ann.
9	177.8	34.24	1282.37	1317.63	67.80	7024	1525	189.3	.0524	.4870	T-T	Ann.
10	180.6	29.66	1180.55	1419.45	75.61	7442	1543	191.2	.0470	.5264	T-T	Ann.

(*) Computed from superficial velocities of gas and liquid phases and the Marchane map

14.2% and 14.4% as show in Tables 5.7 and 5.8. The average two-phase heat transfer coefficient for this system is between 4000 - 5000 BTU/ft²-hr-°F, and the average overall heat transfer coefficient, U_i , is between 1200 to 1300 BTU/ft²-hr-°F. In this calculation, obviously outside film coefficient h_o (2000 BTU/ft²-hr-°F) is in stronger controlling the heat transfer rate. The fact that we have high value of heat transfer coefficient is due to the fluid properties and high velocity (average 20 - 80 ft/sec) of the turbulent-turbulent, separated flow of the mixture. Since both heptane and decane are highly volatile liquids, the increase of mass in the gas phase down the tube attributes a high decrease in liquid holdup fraction. Plots of total pressure, temperature, and liquid holdup of the mixture corresponding to tube position are illustrated in Figure 5.7 and Figure 5.8.

Similar as air-water system, the Lockhart-Martinelli correlations for liquid holdup and pressure drop result in a lower pressure drop and liquid holdup fraction in the system comparing to those given by the Johnson-Abou Sabe correlations. Figure 5.7 indicates a good agreement in the fluid temperatures given by these correlations due to the fact that outside tube film is controlling the heat transfer rate. Total pressure of the mixture corresponding to the axial tube positions is shown in Figure 5.8.(a), pressure tends to decrease faster down the tube due to more liquid

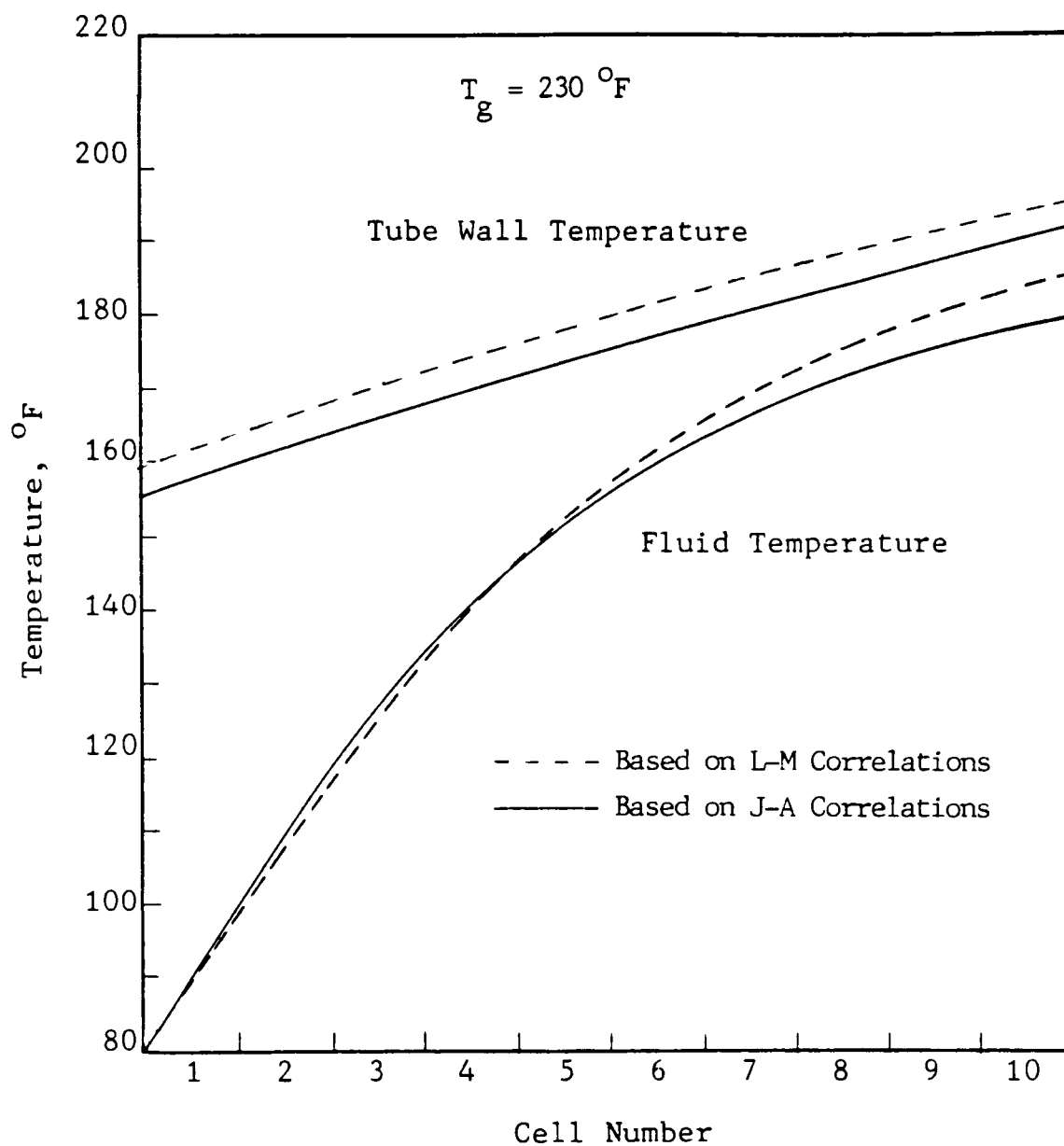


Figure 5.7. Temperature Profile of the Multi-Component Mixture Corresponding to the Cell Number

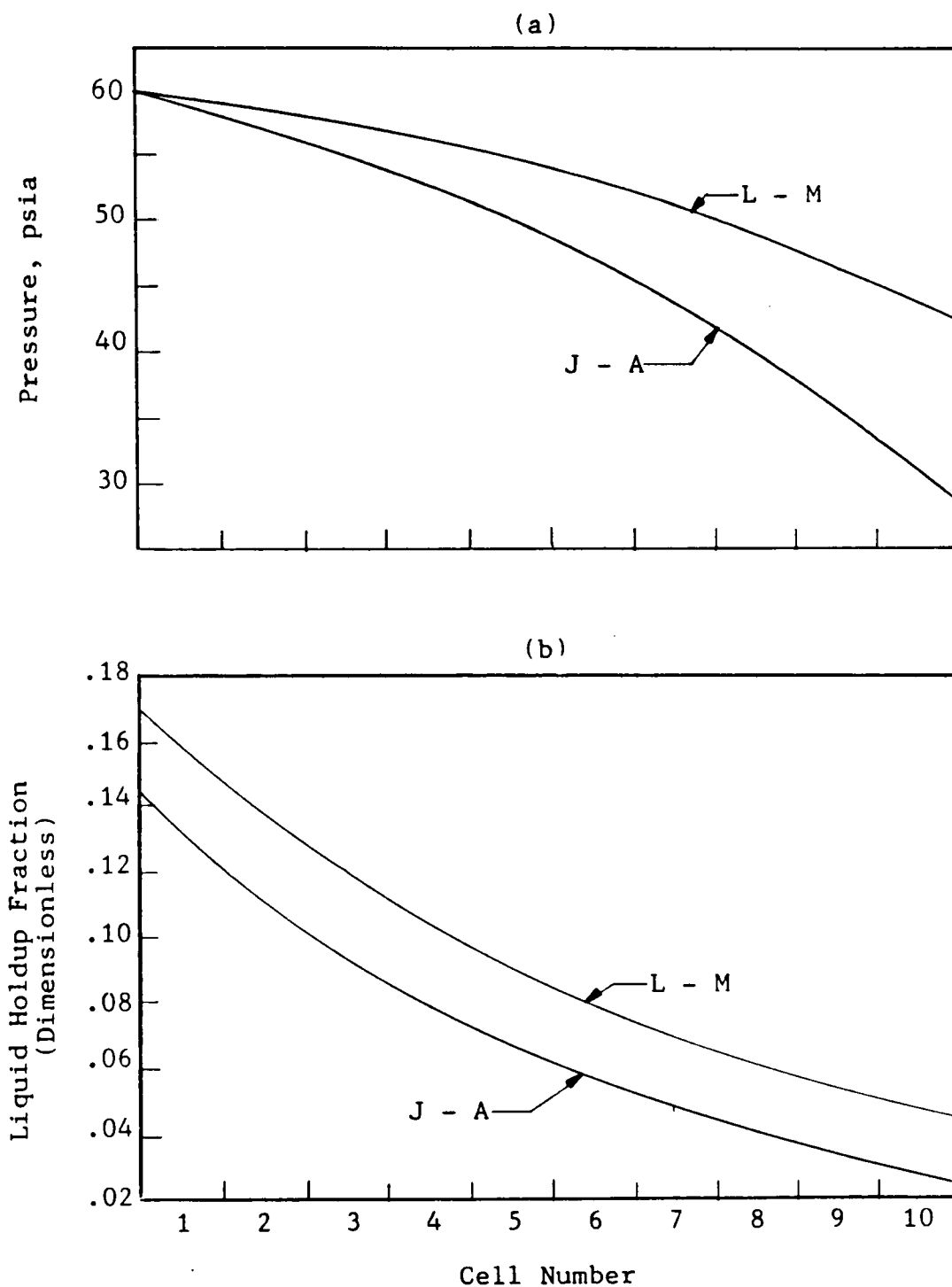


Figure 5.8 Pressure and Liquid Holdup Profiles for the Multicomponent Mixture
(a) Pressure Profile for the Multicomponent Mixture, (b) Liquid Holdup Profile for the Multicomponent Mixture

becomes vapor. Liquid holdup is illustrated in Figure 5.8(b). Because the system consists of highly volatile liquids (C_7H_{16} and $C_{10}H_{22}$), liquid holdup resulted from these correlations is significantly different and fastly decreased from the very beginning of the first cell.

Since mass transfer between phases in this system is very important, and due to the fact that heat transfer rate is controlling by the outside film heat transfer coefficient, it is expected that this calculation results in a shorter tube. Verification of this conclusion can be performed by comparing to the experimental data. Unfortunately, such data are not currently available in the literature. Also, for this multi-component, gas-liquid flow, the calculation with $h'_i = 1.2 h_i$ (or other modification) is not necessary here since we cannot perform any comparison between the calculated results to the experimental values. Evaluation of the assumptions such as equilibrium between phases, ideal gas and liquid, etc. using in this calculation can be established, and the calculational procedure can be modified once such comparison becomes available.

6. SUMMARY AND CONCLUSIONS

A steady state heat transfer model, a calculational procedure, and a computer code have been developed for the design of a horizontal tubular heater for two-phase, multi-component, gas-liquid flow. The code determines the length of the tube required to produce a specified fluid outlet temperature, for given values of the inlet phase flow rates, their compositions, and the temperature of the heating medium (e.g., steam temperature, or temperature of the gases in a fired preheater). The code employs established correlations in the literature for calculating the pressure drop, the liquid holdup, and the two-phase heat transfer film coefficient. Equilibrium ratios (K-values) are required for each of the volatile components in the liquid phase as a function of temperature and pressure. Principle assumptions on which the model is based include:

- . Equilibrium is approached locally between the gas and liquid phases
- . Both liquid and gas phases are ideal
- . Liquid phase is Newtonian

The model and code have been at least partially validated by comparing computed values with a limited quantity of experimental data in the literature for the air-water system. A validation for multicomponent systems has

not been made at this time because of the lack of experimental data for comparison. However, the use of the code has been illustrated for a binary, hydrocarbon liquid mixture (heptane and decane) - gas (hydrogen and nitrogen) system. the code can be readily extended to apply to non-Newtonian liquid-gas systems when correlations become available for the pressure drop, liquid holdup, and heat transfer coefficients for these systems.

The following conclusions are made, based on the results of this study:

1. The computer codes are successful for both the air-water system and the gas-liquid hydrocarbon system in that the iteration procedures converge for a relatively wide range of values of the specified convergence parameters.
2. The Johnson - Abou Sabe two-phase liquid holdup and pressure drop correlations yield results which are in better agreement with the experimental data for the air-water system than those of Lockhart - Martinelli. This result is not unexpected, since the Lockhart and Martinelli correlations are based on isothermal flow.
3. The Davis and David two-phase heat transfer correlation gives values for the film coefficient which appear to be about 20% too low. This result,

however, is not unreasonable in view of the fact that the stated error range on the correlation is 20 to 30%.

4. Reasonably good agreement between calculated results and experimental results are obtained for the air-water system when the Johnson-Abou Sabe two-phase flow correlations are incorporated in the computer code and when the Davis-David two-phase heat transfer coefficient correlation is increased by 20%. With the code modified in this way, calculated values of the pressure drop agree within 5.7% of experimental values, the calculated two-phase fluid outlet temperatures agree within one degree Fahrenheit (except for one run where the calculated value is 17.6°F too low), and the calculated and experimental average tube wall temperatures are within about 3 to 5°F of one another in all three experimental runs for which these data are available. The extent of the agreement between the computed results and experiment for this system is therefore considered sufficient to provide at least a tentative validation of the heat transfer model and calculation of procedure.

5. The literature contain insufficient data for complete validation of the model for both binary and multicomponent systems at this time.

In view of the fact that the two-phase flow heat transfer model presented here has not been fully validated at this time, caution must be exercised in its use, especially outside the range of the experimental conditions of the available data. These ranges are summarized as follows:

Inlet Stream

Temperature	65 - 96°F
Pressure	27 - 48 psia

Outlet Stream

Temperature	171 - 183°F
Pressure	21 - 31 psia

Flow Rates

Gas	134 - 203 lb/hr (32,500-64,500 $\frac{\text{lb}}{\text{hr-ft}^2}$)
Liquid	975 - 2000 lb/hr (115000-635000 $\frac{\text{lb}}{\text{hr-ft}^2}$)

Liquid Holdup Fraction (estimated):

0.04 - 0.12

Finally, the experimental correlations which gave best results (the Johnson-Abou Sabe pressure drop and liquid holdup correlations, and the modified Davis-David heat transfer correlations) were developed from a very limited data base.

The air-water-steam system provides the sole data for the Johnson-Abou Sabe correlation; the Davis-David correlation is based only on data for the steam-water system. Although this latter correlation has been reported to be valid for a limited number of other systems, it is not known to what extent these correlations apply in general.

7. RECOMMENDATIONS FOR FUTURE WORK

The limited amount of data in the literature for the air-water system provide a tentative validation of the model and corresponding computer code developed in this work.

It is recommended that additional experimental data be obtained in order to provide a rigorous test of the model over a wider range of the operating conditions such as gas and liquid flow rates and liquid holdup fraction. It is suggested that experiments be performed on two systems. Additional data for the air-water system shall be of value in testing the two-phase heat transfer correlation of Davis and David. However this system probably is not adequate for testing one of the fundamental assumptions, that of local gas-liquid equilibrium, used in this model. The reason for this is that the mass transfer from one phase to the other is somewhat limited. Heat transfer data for a gas-multi-component liquid system would be more appropriate for this test, where at least one of the components has a high relative volatility in the accessible experimental temperature range. Candidate systems may be found in a homologous series of hydrocarbons, although precautions must be taken to ensure safety when working with such systems.

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APPENDIXES

APPENDIX A

LIST OF SYMBOLS AND SUBSCRIPTS

LIST OF SYMBOLS

A_i	Tube inside cross-sectional area, ft^2
C_g, C_l	Constants depend on phase and flow-type, dimensionless
C_p	Specific heat, $\text{BTU/lb-}^\circ\text{F}$
C_{pi}	Pressure convergence criterion, .2%, dimensionless
$\bar{C}_p$	Average specific heat, $\text{BTU/lb-}^\circ\text{F}$
C_{ti}	Temperature convergence criterion, .3%, dimensionless
D_i	Tube inside diameter, ft
D_o	Tube outside diameter, ft
$\bar{D}$	Logarithmic mean diameter, ft
F	Total molar feed rate, mole/hr
F_g	Molar flow rate of the gas phase in a equilib- rium mixture, mole/hr
F_l	Molar flow rate of the liquid phase in an equilibrium mixture, mole/hr
G_M	Total mass flux rate, lb/hr-ft^2
g_c	Newton's law proportionality factor, $32.174 \text{ lb-ft/lbf-sec}^2$
H	Enthalpy, BTU/lb
H_j	Enthalpy of component j, BTU/lb
H_j^0	Enthalpy of component j at reference state, BTU/lb

h_i	Two-phase heat transfer coefficient, BTU/ft ² -hr-°F
h_o	Outside film heat transfer coefficient, BTU/ft ² -hr-°F
h_r	Radiation heat transfer coefficient, BTU/ft ² -hr-°F
K_j	K value of component j, dimensionless
k	Total number of cells in the tube
k_l	Thermal conductivity of the liquid, BTU/ft-hr-°F
k_w	Thermal conductivity of the tube wall, BTU/ft-hr-°F
L	Tube length, ft
M_j	Molecular weight of component j, lb/lb mole
m	Flow type constant, dimensionless
$\dot{m}$	Mass flow rate, lb/hr
$\dot{m}_t$	Average net mass exchange between phase per unit liquid volume, lb/ft ³ -hr
n	Flow type constant, dimensionless
P, P_t	Total pressure, psia
P_j^o	Vapor pressure of component j, psia
P_{r1}	Prandtl number of the liquid phase, dimensionless
$\dot{q}_w$	Heat flux rate to the system, BTU/hr-ft ²
Re	Reynold number, dimensionless

R_g	Gas holdup fraction, dimensionless
R_l	Liquid holdup fraction, dimensionless
T_g	Temperature of the heating medium, $^{\circ}\text{F}$
T_r	Reference temperature, $^{\circ}\text{F}$
T_w	Tube wall temperature, $^{\circ}\text{F}$
T_z	Local fluid temperature, $^{\circ}\text{F}$
U_i	Overall heat transfer coefficient, inside diameter basis, $\text{BTU}/\text{ft}^2\text{-hr-}^{\circ}\text{F}$
V_g	Superficial velocity of the gas phase, ft/hr
V_l	Superficial velocity of the liquid phase, ft/hr
$\mathcal{X}$	Lockhart and Martinelli parameter, dimensionless
x_j	Mole fraction of component j in the liquid phase, dimensionless
x'_j	Mass fraction of component j in the liquid phase, dimensionless
x_w	Tube wall thickness, ft
Y_j	Mole fraction of component j in the gas phase, dimensionless
Y'_j	Mass fraction of component j in the gas phase, dimensionless
Z	Cartesian coordinate direction, dimensionless
z	Cartesian index along the tube length, dimensionless

z_j	Overall mole fraction of component j in the feed, dimensionless
Greek Symbols	
ΔH	Enthalpy difference between inlet and outlet streams BTU/hr
ΔH_{total}	Total enthalpy difference between inlet and outlet streams, BTU/hr
$\Delta H'_g$	Total heat added to the evaporated amount of liquid between inlet and outlet, BTU/hr
$\dot{m}$	Total mass exchange between phase, lb/hr
ΔP	Pressure drop in a section (or entire) of the tube, psi
ΔP_{tp}	Two-phase flow pressure drop in a section (or entire) of the tube, psi
$(\Delta P/\Delta z)_g$	Single gas flow pressure drop per unit length of the tube, psi/ft
$(\Delta P/\Delta z)_l$	Single liquid flow pressure drop per unit length of the tube, psi/ft
$(\Delta P/\Delta z)_{\text{tp}}$	Two-phase flow pressure drop per unit length of the tube, psi/ft
ΔT_{LMTD}	Logarithmic mean temperature difference, °F
ΔV	Elemental volume, ft ³
Δz	Sectional length of the tube, ft
ϵ_w	Emissivity of the surface, dimensionless
λ	Latent heat, BTU/lb

μ	Viscosity, lb/ft-hr
μ_{tp}	Two-phase fluid viscosity, lb/ft-hr
ρ	Density, lb/ft ³
σ	Square root ratio of two-phase flow pressure drop to single-phase flow pressure drop per unit length of the tube, dimensionless
σ_c	Constant of the flow type, dimensionless
ξ	Gas mass fraction, dimensionless
Ψ	Molar fraction of gas flow rate in the feed to the total feed, dimensionless

SUBSCRIPTS

g	Gas phase
j	Index for component j
l	Liquid phase
t-t	Both liquid and gas phases are turbulent
t-v	Liquid phase is turbulent and gas phase is viscous
v-t	Liquid phase is viscous and gas phase is turbulent
v-v	Both liquid and gas phases are viscous
w	Tube wall
z	Index of location on the tube
0, 1, 2,...	Location index along the tube
1, 2, 3,...n	Component index
1, 2, 3,...k	Cell index

APPENDIX B

DETAILS OF CALCULATED
RESULTS OF RUN NO.2
FOR AIR-WATER SYSTEM

DETAILS OF CALCULATED RESULTS
OF RUN NO.2 FOR AIR-WATER SYSTEM

Using both Lockhart-Martinelli and Johnson-Abou Sabe correlations for liquid holdup and pressure drop, results of Run No. 2 obtained by searching for steam jacket temperature to match the outlet mixture temperature to the experimental values are listed in Table B.1. Comparing these values to the experimental measurements, obviously the calculation based on the Johnson-Abou Sabe correlations results in a better agreement. However, average tube wall temperatures are still too high comparing to the experimental measurements. When the increase of inner film heat transfer coefficient by a factor of 1.2 is employed, Table B.2, the calculated tube wall temperature, outlet stream temperature, and pressure drop appear to be in a much better agreement with the experimental measurements. Figures B.1 and B.3 present the mixture temperatures, tube wall temperatures, and steam jacket temperatures corresponding to axial tube positions for different calculational models. The verification of the calculational procedure developed in this study can be well established here if the literature data are provided. In Figures B.2 and B.4, liquid holdup, total mixture pressure, and heat transfer coefficient are plotted against axial tube position. Even the tube wall

temperatures can be as much different as 20°F (Figure B.3), the mixture pressure and liquid holdup given by the Johnson-Abou Sabe correlations are consistent as shown in Figure B.4. This is due to the fact that mass exchange between phase is not very significant in this flow system.

Figure B.5 and Figure B.6 show the results of Run No.1 and Run No.3 of air-water flow listed in Table 2.5. The Johnson-Abou Sabe correlations for liquid holdup and pressure drop are used in these calculations. In order to obtain a good agreement between calculated and experimental values, modified factor for the heat transfer coefficient is 1.7 for Run No.1 and 1.1 for Run No.3. This may be due to the fact that both gas mass fraction and total flow rate of Run No.1 are low. Based on these results, the modification of two-phase heat transfer coefficient given by Davis and David seems to be functioned by the total mass flow rate and gas mass fraction of the system.

Table B.1. Calculated Result of Run No.2 of Air-Water Flow Based on the Correlations
for Liquid Holdup and Pressure Drop of Lockhart-Martinelli ⁽⁶⁾ -
Searching for Steam Temperature to Match the Outlet Stream Temperature
Calculational Model

Cell Number	Outlet Stream Temp. Of	Press. psia	R ₁	$\frac{h_i}{G_{TO}} \frac{lb}{ft^2-hr-Of}$	$\frac{U_i}{G_{TO}} \frac{lb}{ft^2-hr-Of}$	$\dot{m}_i$ lb/hr	$\dot{m}_o$ lb/hr
1	81.43	30.29	.1164	332.17	283.73	972.80	203.30
2	96.88	29.98	.1135	362.23	305.39	971.34	204.76
3	110.83	29.66	.1108	390.35	325.13	969.32	206.78
4	123.44	29.33	.1080	416.07	342.78	966.63	209.47
5	134.36	28.99	.1050	438.28	357.72	963.37	212.73
6	143.98	28.63	.1018	456.94	370.05	959.48	216.62
7	152.04	28.23	.0987	472.99	380.51	955.16	220.94
8	159.12	27.82	.0955	486.77	389.37	950.28	255.82
9	165.28	27.39	.0930	502.07	399.10	944.89	231.21
10	170.58	26.94	.0904	515.93	407.81	939.10	237.00

Table B.2. Calculated Result of Run No.2 of Air-Water Flow Based on the Correlations
for Liquid Holdup and Pressure Drop of Johnson-Abou Sabe (7) -
Searching for Steam Temperature to Match the Outlet Stream Temperature
Calculational Model

Cell Number	Outlet Stream Temp. °F	Press. psia	R_1	h_i $\frac{\text{ft}^2\text{-hr-}^\circ\text{F}}{\text{Btu}}$	U_i $\frac{\text{Btu}}{\text{ft}^2\text{-hr-}^\circ\text{F}}$	$\dot{m}_i$ lb/hr	$\dot{m}_q$ lb/hr
1	82.99	30.08	.0969	388.98	324.18	972.67	203.43
2	99.62	29.55	.0942	426.59	349.89	970.45	205.15
3	114.33	29.01	.0913	459.86	371.96	968.54	207.56
4	127.23	28.44	.0882	487.54	389.87	965.34	210.76
5	138.12	27.85	.0850	510.55	404.44	961.46	214.64
6	147.43	27.22	.0819	529.67	416.35	956.85	219.65
7	155.07	26.52	.0788	545.36	425.98	957.71	224.39
8	161.57	25.80	.0764	563.08	436.71	945.91	230.19
9	167.00	25.05	.0740	578.85	446.15	939.55	236.55
10	171.29	24.26	.0717	592.59	454.16	932.91	243.19

Table B.3. Calculated Result of Run No.2 of Air-Water Flow Based on the Correlations
for Liquid Holdup and Pressure Drop of Johnson-Abou Sabe⁽⁷⁾ -
Searching for Steam Temperature to Match the Tube Wall Temperature
Calculational Model and h_i Increases by a Factor of 1.2

Cell Number	Outlet Stream Temp. °F	Press. psia	R_1	h_i $\frac{\text{Btu}}{\text{ft}^2\text{-hr-}^\circ\text{F}}$	U_i $\frac{\text{Btu}}{\text{ft}^2\text{-hr-}^\circ\text{F}}$	$\dot{m}_i$ lb/hr	$\dot{m}_g$ lb/hr
1	84.14	30.08	.0968	468.16	377.37	972.58	203.52
2	101.29	29.55	.0940	515.69	407.66	970.74	205.36
3	116.32	29.01	.0910	556.92	433.00	968.14	207.96
4	129.03	28.44	.0879	590.48	453.02	964.82	211.28
5	139.55	27.84	.0847	617.36	468.68	960.89	215.21
6	148.38	27.21	.0816	639.17	481.14	965.35	219.75
7	155.53	26.51	.0786	656.57	490.93	951.38	224.72
8	161.52	25.79	.0763	676.16	501.81	945.93	230.17
9	166.49	25.03	.0741	693.28	511.17	940.06	236.04
10	170.38	24.24	.0719	707.92	519.08	934.03	242.07

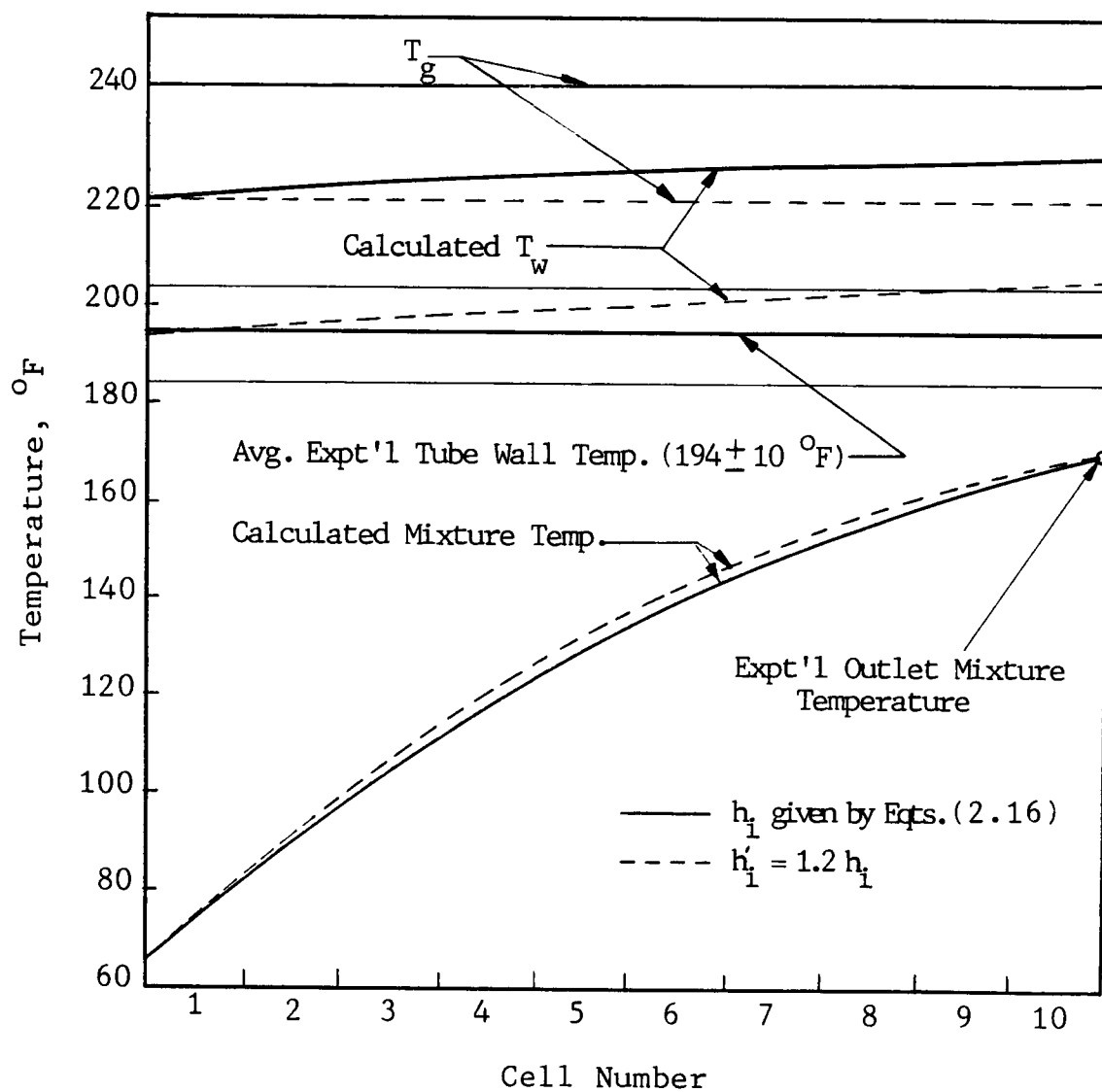


Figure B.1. Fluid and Tube Wall Temperature Profiles of Run No. 2 Based on the Lockhart-Martinelli Correlations ($T_g = 244^\circ\text{F}$)

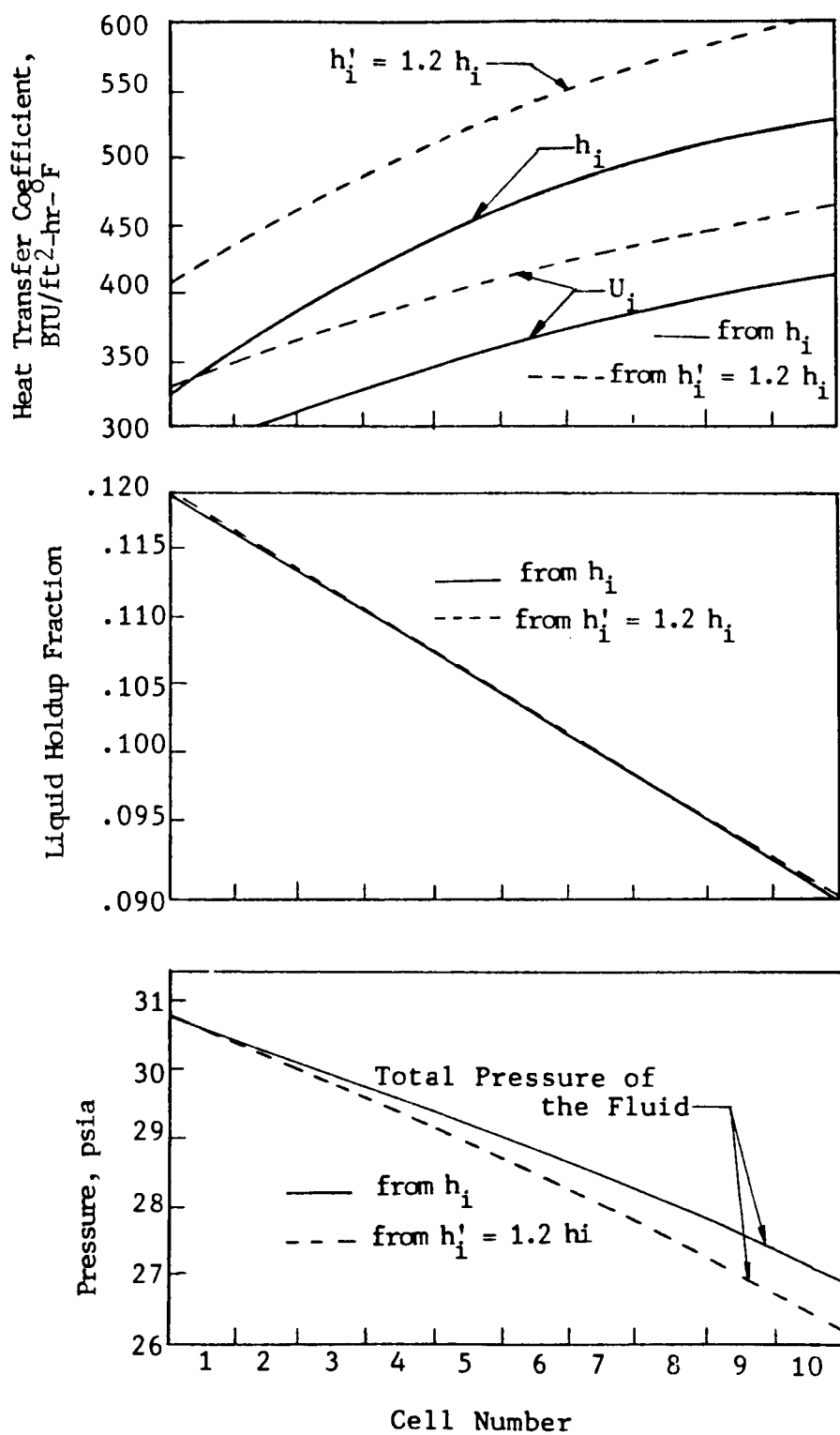


Figure B.2. Heat Transfer Coefficients, Liquid Holdup Fractions, and Stream Pressure of Run No. 2 Based on the Lockhart-Martinelli Correlations

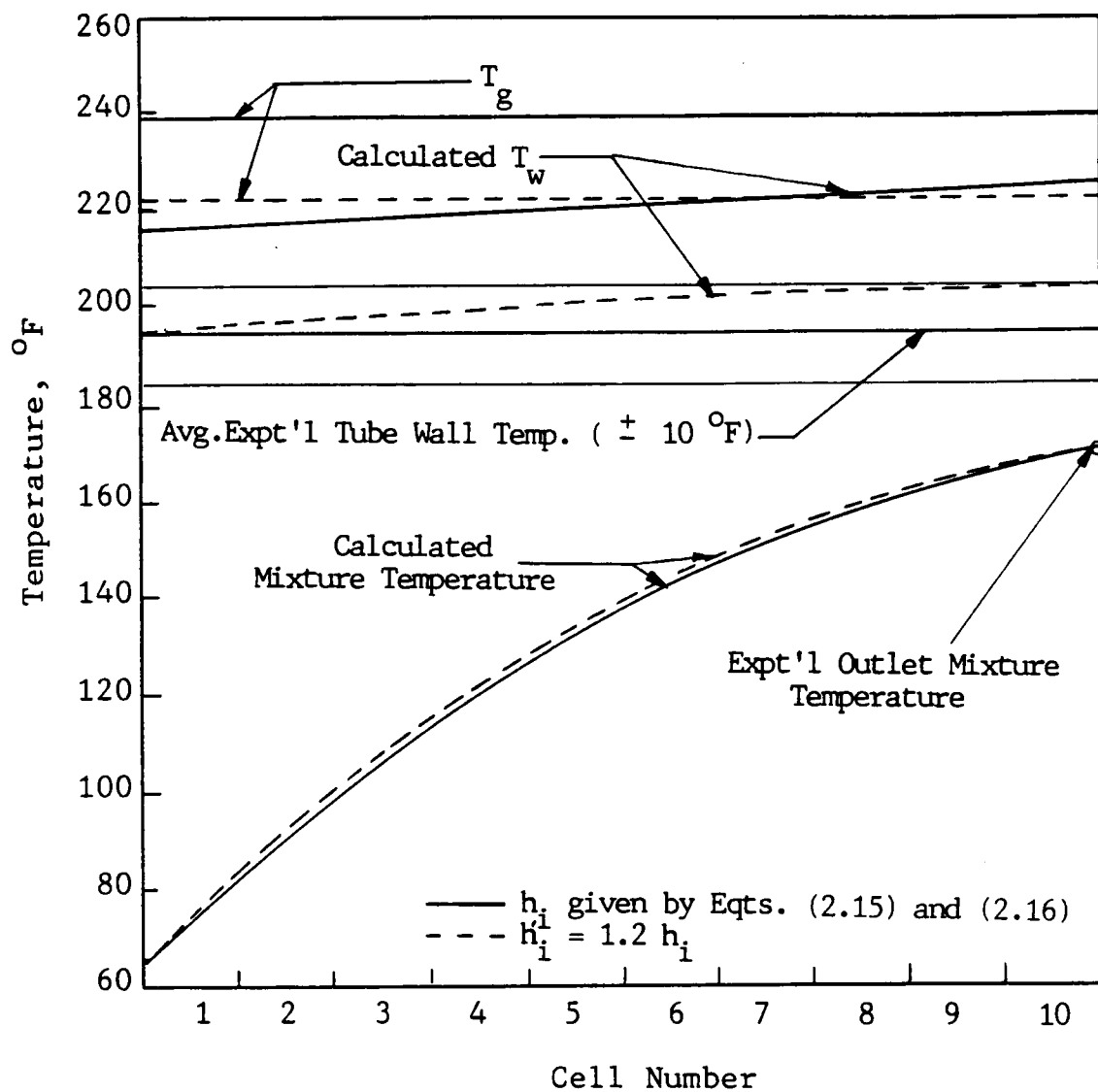


Figure B.3. Fluid and Tube Wall Temperature Profiles of Run No. 2 Based on the Johnson-Abou Sabe⁽⁷⁾ Correlations ($T_g = 238^\circ\text{F}$)

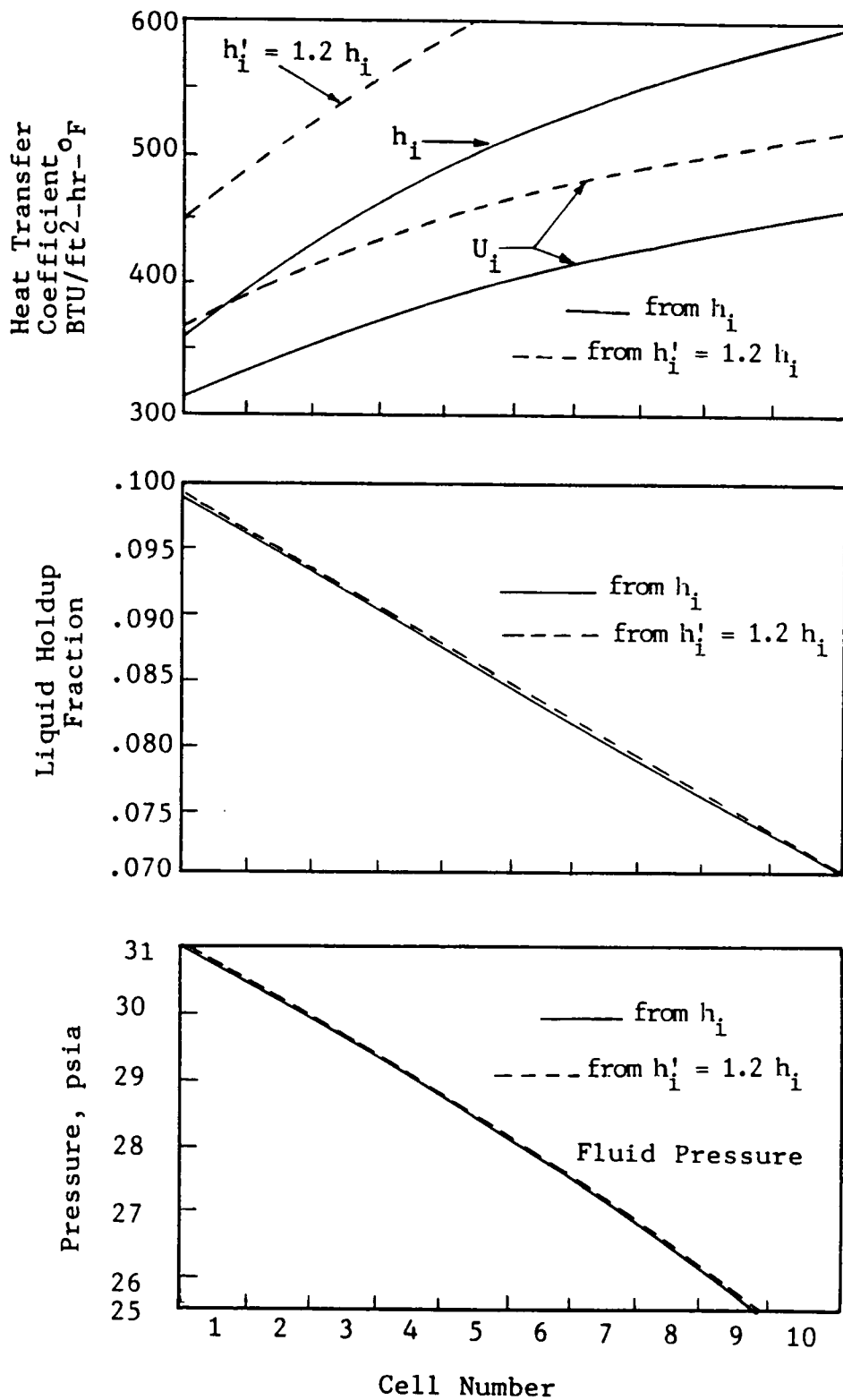


Figure B.4. Heat Transfer Coefficients, Liquid Holdup Fractions, and Stream Pressure of Run No. 2 Based on the Johnson-Abou Sabe Correlations

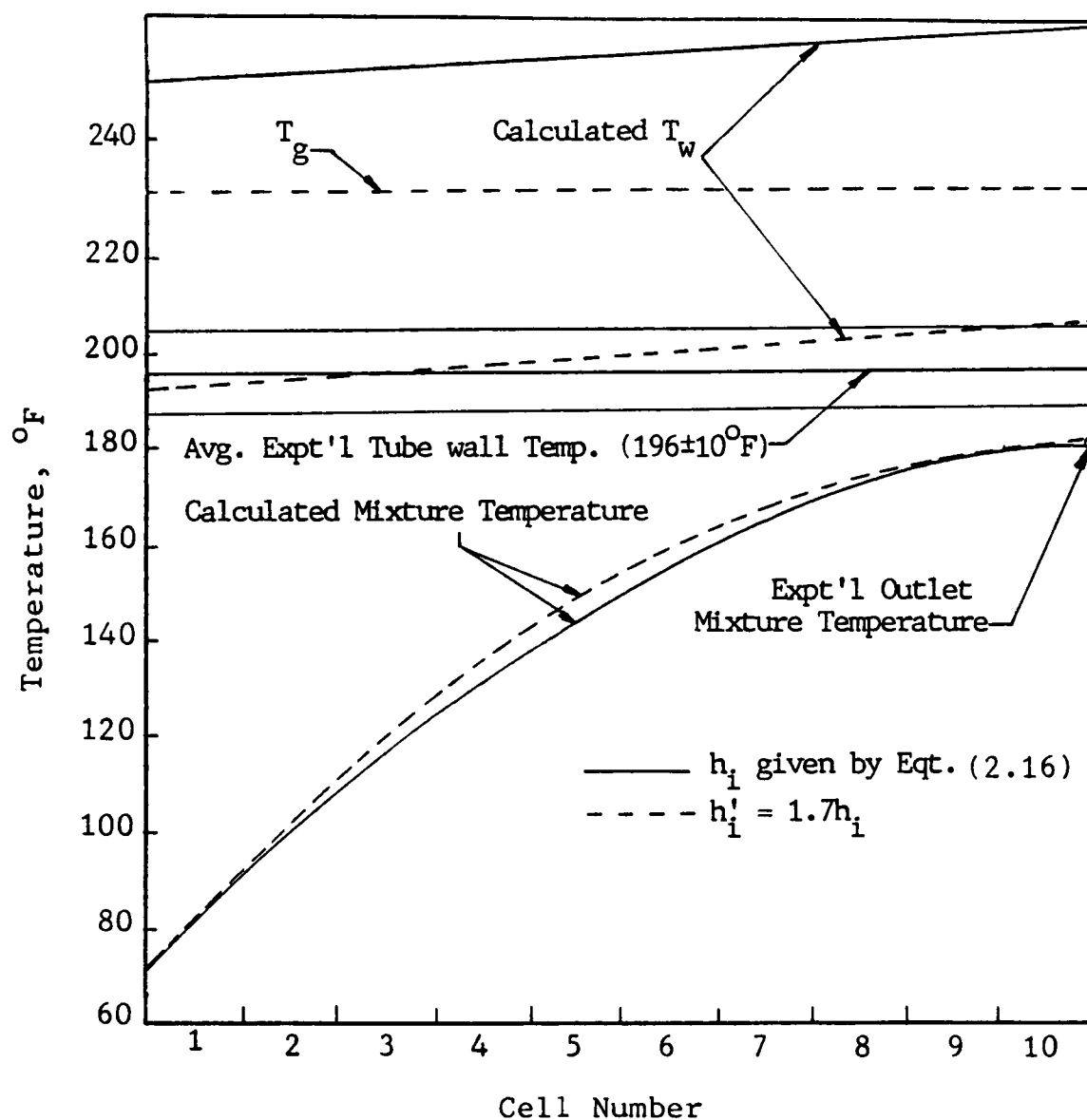


Figure B.5. Mixture and Tube Wall Temperatures of Run No. 1 of Air-Water System Based on the Liquid Holdup and Pressure Drop of Johnson-Abou Sabe Correlations ($T_g = 275^\circ\text{F}$ for the Calculation with h_i)

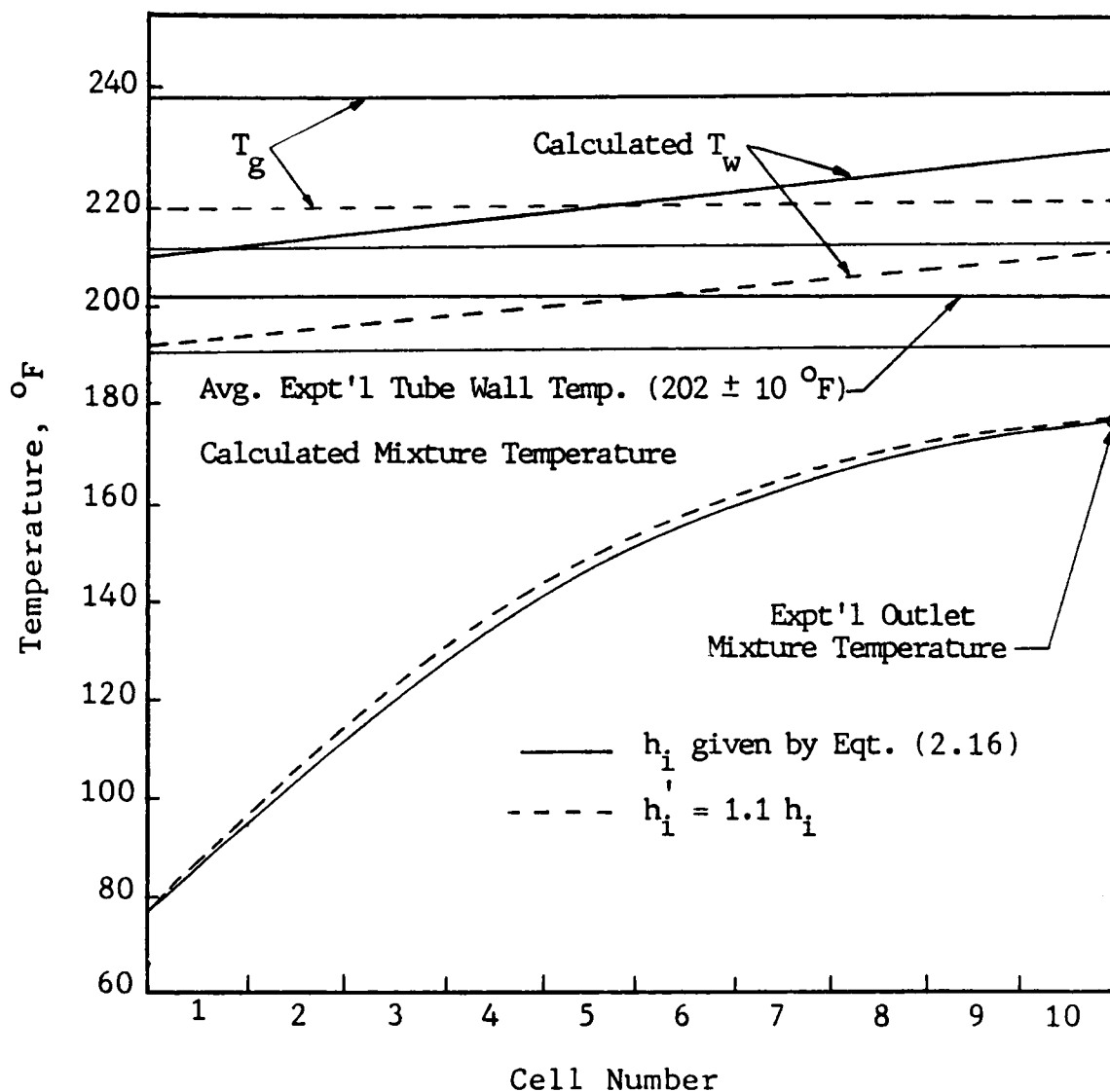


Figure B.6. Mixture and Tube Wall Temperature of Run No. 3 of Air-Water System Based on the Liquid Holdup and Pressure Drop of Johnson-Abou Sabe Correlations

APPENDIX C

PHYSICAL PROPERTIES DATA FOR
USE IN THE COMPUTER CODES

Table C.1. Density of Liquids, lb/ft³

Press. psia	Comp.	Temperature, °F					Reference
		80	150	200	250	300	
14.7	Water	62.22	61.19	60.11	58.80	57.31	(26)
	H ₂	3.91	2.08	1.56	1.25	1.04	(25)
	N ₂	15.61	8.32	6.24	5.00	4.16	(25)
	C ₇ H ₁₆	42.32	40.20	38.70	36.96	34.96	(22)
	C ₁₀ H ₂₂	45.31	43.46	42.1	40.68	39.25	(22)
30	Water	62.22	61.19	60.11	58.80	57.31	(26)
	H ₂	3.99	2.12	1.60	1.28	1.06	(25)
	N ₂	15.95	8.50	6.38	5.11	4.25	(25)
	C ₇ H ₁₆	42.33	40.21	38.70	37.06	34.98	(22)
	C ₁₀ H ₂₂	45.49	45.50	42.17	40.72	39.29	(22)
100	Water	62.72	61.19	60.11	58.80	57.31	(26)
	H ₂	4.38	2.33	1.78	1.41	1.17	(25)
	N ₂	17.50	9.33	7.02	5.62	4.68	(25)
	C ₇ H ₁₆	42.35	40.23	38.72	37.02	35.05	(22)
	C ₁₀ H ₂₂	45.53	43.68	42.32	40.90	39.47	(22)

Note: Hydrogen and nitrogen liquid densities are roughly estimated

Table C.2. Density of Gases and Vapors, lb/ft³

Press. psia	Comp.	Temperature, °F					Reference
		80	150	200	250	300	
14.7	Air	.0735	.0648	.0602	.056	.051	(21)
	Water	.0016	.0100	.0297	.0352	.033	(26)
	H ₂	.000348	.000308	.0042	.0039	.0036	(27)
	N ₂	.071134	.063	.0582	.0544	.0505	(27)
	C ₇ H ₁₆	.2186	.2024	.1862	.17	.16	(22)
	C ₁₀ H ₂₂	.2405	.2226	.2050	.187	.176	(22)
30	Air	.225	.19	.165	.140	.115	(21)
	Water	.0016	.01	.0297	.071	.068	(26)
	H ₂	.01008	.0231	.0079	.0074	.0068	(27)
	N ₂	.1050	.12843	.0840	.0935	.1030	(27)
	C ₇ H ₁₆	.492	.4554	.4189	.3825	.360	(22)
	C ₁₀ H ₂₂	.5410	.5009	.4609	.4208	.400	(22)
100	Air	.5083	.435	.383	.331	.279	(21)
	Water	.0016	.010	.0297	.071	.155	(26)
	H ₂	.0347	.0307	.0284	.0264	.0247	(27)
	N ₂	.4844	.4280	.3858	.3645	.3431	(27)
	C ₇ H ₁₆	1.107	1.025	.9426	.8606	.730	(22)
	C ₁₀ H ₂₂	1.218	1.130	1.037	.9467	.904	(22)

Note: Water is assumed to be saturated

Table C.3. Approximated Correlations for Specific Heat of the
Components

Component	Phase	Specific Heat, BTU/lb-°F	Reference
Air	Gas	$.25 + .00004(T - 50)$	(18)
Water	Vapor	$.40 + .00008(T - 50)$	(18)
	Liquid	$1.0 - .0004(T - 50)$	(18)
Hydrogen	Gas	3.5 for 50 T 300°F	(18)
	Liquid	7 for 50 T 300°F	(25)
Nitrogen	Gas	$.25 + .00002(T - 50)$	(18)
	Liquid	$.25 + .00002(T - 50)$	(25)
Heptane	Vapor	$.248 + .00039(T - 50)$	(22)
	Liquid	$.375 + .00058(T - 50)$	(22)
Decane	Vapor	$.210 + .00014(T - 50)$	(22)
	Liquid	$.400 + .00058(T - 50)$	(22)

Note: T is in Fahrenheit. Liquids of hydrogen and nitrogen are roughly estimated

Table C.4. Approximated Correlations for Thermal Conductivity of
Components in the Liquid Phase

Component	Thermal Conductivity, BTU/ft-hr-°F	Reference
Water	$0.333 + .000252(T - 50)$	(18)
Hydrogen	$3167.2 + 4.0383(T - 80.6)$	(25)
Nitrogen	$448.3 + .65382(T - 80.6)$	(25)
Heptane	$1.39 - .000741(T - 86)$	(22)
Decane	$1.45 - .000556(T - 86)$	(22)

Note: Rough estimate for hydrogen and nitrogen values

Table C.5. Approximated Correlations for Viscosities of the Components

Component	Phase	Viscosity, cp	Reference
Air	Gas	$.017 + .000024(T - 50)$	(18)
	Vapor	$.0095 + .000018(T - 50)$	(18)
Water	Liquid	$1.27 - .00456(T - 50)$	(18)
	Gas	$.0085 + .0001042(T - 60)$	(18)
Hydrogen	Liquid	$.0010 + .0000004167(T - 60)$	(25)
	Gas	$.1070 + .000025(T - 60)$	(18)
Nitrogen	Liquid	$.0020 + .001042(T - 60)$	(25)
	Vapor	$.0600 + .001042(T - 60)$	(22)
Heptane	Liquid	$\text{Exp}(436.73(1/T(^{\circ}\text{K}) - 1/232.53))$	(22)
	Vapor	$.0500 + .00008333(T - 60)$	(22)
Decane	Liquid	$\text{Exp}(558.61(1/T(^{\circ}\text{K}) - 1/277.8))$	(22)

Note: T is in Fahrenheit. Liquids of hydrogen and nitrogen are roughly estimated

Table C.6. Approximated Correlations for Latent Heat

Components	Latent Heat, BTU/lb	Reference
Water	$910 + .552(T - 80)$	(26)
H ₂	$1800 + 3.50(T - 80)$	(29)
N ₂	$237 + .264(T - 80)$	(29)
C ₇ H ₁₆	$162 + .2364(T - 80)$	(22)
C ₁₀ H ₂₂	$162 + .1955(T - 80)$	(22)

Note: T is in Fahrenheit

Table C.7. K-Values of Components (20) in the Multicomponent System

Press. psia	Comp.	Temperature, °F											
		80	100	120	140	160	180	200	220	240	260	280	300
14.7	H ₂	566	920	1486	2266	3255	4529	6369	8492	11323	14863	16808	24771
	N ₂	246	400	646	985	1415	1969	2769	3692	4923	6462	8308	10770
	C ₇ H ₁₆	.08	.13	.21	.32	.46	.64	.90	1.2	1.6	2.1	2.7	3.5
	C ₁₀ H ₂₂	.0018	.0037	.0072	.0130	.024	.038	.06	.1	.155	.22	.33	.46
20	H ₂	482	788	1244	2156	2736	3647	4975	6881	9120	12020	16578	19895
	N ₂	214	350	553	958	1216	1621	2211	3058	4053	5342	7368	8842
	C ₇ H ₁₆	.058	.095	.150	.260	.33	.44	.60	.83	1.1	1.45	2.00	2.4
	C ₁₀ H ₂₂	.0015	.0028	.0052	.0100	.0156	.027	.042	.070	.11	.153	.240	.34
25	H ₂	376	660	1021	1415	2200	3063	4085	5500	7230	9429	11785	15715
	N ₂	171	300	464	643	1000	1393	1857	2500	3286	4286	5357	7143
	C ₇ H ₁₆	.0480	.084	.130	.18	.28	.39	.52	.70	.92	1.2	1.5	2.0
	C ₁₀ H ₂₂	.0013	.0025	.0048	.0084	.015	.024	.038	.060	.090	.13	.19	.28
30	H ₂	346	559	845	1228	1765	2458	3225	4377	5760	7372	9215	12285
	N ₂	161	250	393	571	821	1143	1500	2036	2679	3429	4286	5714
	C ₇ H ₁₆	.045	.070	.11	.16	.23	.32	.42	.57	.75	.96	1.2	1.6
	C ₁₀ H ₂₂	.0012	.0022	.0042	.0075	.013	.020	.032	.05	.074	.110	.16	.24
40	H ₂	254	546	615	869	1380	1810	2316	3259	4345	5069	6590	8545
	N ₂	121	200	293	414	655	862	1103	1551	2069	2414	3138	4069
	C ₇ H ₁₆	.035	.058	.085	.12	.19	.25	.32	.45	.60	.70	.91	1.18
	C ₁₀ H ₂₂	.0011	.0018	.0034	.0058	.011	.017	.026	.040	.062	.080	.13	.18
60	H ₂	193	308	478	677	923	1308	1769	2306	3075	3690	4766	5920
	N ₂	94	150	233	330	450	638	863	1125	1500	1800	2325	2888
	C ₇ H ₁₆	.025	.040	.062	.088	.120	.170	.23	.30	.40	.48	.62	.78
	C ₁₀ H ₂₂	.001	.0015	.0027	.0046	.0074	.0120	.018	.028	.040	.06	.09	.12
100	H ₂	129	200	314	458	614	850	1142	1500	1858	2642	3286	3786
	N ₂	64	100	157	229	307	429	571	750	929	1312	1643	1893
	C ₇ H ₁₆	.018	.028	.044	.064	.086	.120	.160	.210	.26	.37	.46	.53
	C ₁₀ H ₂₂	.0009	.0012	.0020	.0034	.0055	.0090	.014	.02	.03	.045	.065	.08

Note: Most of H₂ and N₂ K-values are estimated by relative correlations with the other components.

APPENDIX D

COMPUTER CODES

Programs FLOW1.FOR and FLOW2.FOR are developed for the air-water and multicomponent gas-liquid flows, respectively. Some important symbols in these programs are listed below:

ACROSS	Cross-Sectional Area of the Tube
AGD	Average gas density
AGMFR	Average gas mass flow rate
AGV	Average gas viscosity
AHEAT	Heat transfer surface
AIRR	Air feed rate
ALD	Average liquid density
ALMFR	Average liquid mass flow rate
ALV	Average liquid viscosity
CCFP	Pressure convergence criterion
CCFT	Temperature convergence criterion
CPL	Specific heat of the liquid phase
DI	Tube inside diameter
DLM	Log mean diameter
DLIQ	Density of the liquid phase
DML	Mass transfer between phase
DO	Tube outside diameter
DTP	Two-phase fluid density
DVAP	Density of the gas phase
DZ	Cell length
EXCHM	Mass exchange between phases

N, N-1	Number of cell
NC	Number of component
P	Total pressure of the mixture
PHIL	Parameter ϕ
PLOOP	Pressure loop code
PI	Inlet stream pressure
PO	Outlet stream pressure
PRL	Prandtl number
PX	Parameter α
P1	Inlet stream pressure
REGAS	Reynold number of the gas phase
RELIQ	Reynold number of the liquid phase
RL	Liquid holdup fraction
SLPPD	Single liquid phase pressure drop
SPHG	Specific heat of the gas phase
SPHL	Specific heat of the liquid phase
T	Temperature
TCL	Thermal conductivity of the liquid
TG	Heating medium temperature
TI	Inlet stream temperature
TLM	Log mean temperature difference
TLOOP	Temperature loop code
TO	Outlet stream temperature
TPDROP	Two-phase pressure drop
TW	Tube wall temperature
T1	Inlet stream temperature

GASDEN	Density of the gas phase
GASFRA	Gas mass fraction
GAVIS	Viscosity of the gas phase
GF	Gas feed rate
GFRATE	Gas flow rate
GL	Liquid feed rate or superficial liquid velocity
GV	Superficial gas velocity
HEAT	Heat added to the system
HFLUX	Heat flux rate
HI	Two-phase heat transfer coefficient
HO	Outside film heat transfer coefficient
KLIQ	Thermal conductivity of the liquid phase
KV	K-values
KW	Thermal conductivity of the tube wall
LAMDA	Heat of vaporization
LFRATE	Flow rate of the liquid phase
LIQDEN	Density of the liquid phase
LPDROP	Single liquid phase pressure drop
LTHEAT	Latent heat
LVIS	Viscosity of the liquid phase
MGI	Inlet gas flow rate
MGO	Outlet gas flow rate
MLI	Inlet liquid flow rate
MLO	Outlet liquid flow rate

UG	Gas phase velocity
UI	Overall heat transfer coefficient, tube inside diameter basis
UL	Liquid phase velocity
UTP	Two-phase fluid velocity
VTP	Two-phase fluid viscosity
WATERR	Water feed rate
XP	Molar fraction in the liquid phase
XW	Tube wall thickness
XX	Mass fraction in the liquid phase
YP	Molar fraction in the gas phase
YY	Mass fraction in the gas phase

D.1 Computer Code for the Air-Water System

Program FLOW1.FOR is developed for the air-water flow. Run No.2 of air-water in Table 2.5 is used for illustration of the application of FLOW1.FOR. Interpretation are provided along the program structure. In this program the Johnson-Abou Sabe correlations for liquid holdup and pressure drop are used, and the calculation model is to match the outlet stream temperature without modification of the two-phase heat transfer coefficient given by Davis and David.

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00010 *****PROGRAM FLOW1.FOR*****
00020 *
00030 *
00040 *
00050 *
00060 *
00070 *
00080 *
00090 *
00100 *
00110 *
00120 *****MAIN PROGRAM*****
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00150 *
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00170 *
00180 *
00190 *
00200 *
00210 *
00220 *****
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      REAL MFRL1,MFRL2,MFRV1,MFRV2,LATENT,KW,MYA1,MYA2,MYW1,MYW2,
      # LPD1,LPD2,LPD3,L,KLIQ,LAMDA1,LAMDA2,LTH
C-----SPECIFIED INLET STREAM AND OPERATING CONDITIONS
C-----HEATING MEDIUM TEMPERATURE, OUTSIDE FILM HEAT TRANSFER
C-----COEFFICIENT, THERMAL CONDUCTIVITY OF TUBE WALL, AND TUBE
C-----OUTSIDE AND INSIDE DIAMETERS
      DATA TG,HC,KW,DC,DI/208.0,2000.,64.,.08333,.07250/
C-----FEED RATES OF AIR & WATER AND INLET TEMP. & PRESSURE
      DATA AIRR,WATERR,T1,P1/201.100,0975.,65.0,30.600/
C-----CELL LENGTH AND NUMBER OF CELLS
      DATA DZ,N/1.57,10/
C-----PRINT OUT THE SPECIFIED INFORMATION GIVEN ABOVE
      AIRRW = AIRR/(60.*.074965)
      WRITE(21,2)
00380      2
00390      FORMAT(2X,'*****INLET CONDITIONS*****')
00400      WRITE(21,4) WATERR
00410      4
00420      FORMAT(//,2X,'WATER FEED RATE = ',F9.2,' LBS/HR')
00430      WRITE(21,6) AIRR,AIRRW
00440      6
00450      FORMAT(2X,'AIR FEED RATE= ',F9.2,' LBS/HR, ( OR ',F8.2,
      # ' S.C.F.M.)')
00460      WRITE(21,8) T1
00470      8
00480      FORMAT(2X,'INLET STREAM TEMPERATURE = ',F9.2,' (F)')
00490      WRITE(21,10) P1
00500      10
00510      FORMAT(2X,'INLET STREAM PRESSURE = ',F9.2,' (PSIA)')
00520      WRITE(21,12) TG
00530      12
00540      FORMAT(2X,'JACKET STEAM TEMPERATURE = ',F9.2,' (F)')
00550      WRITE(21,14) HO
00560      14
00570      FORMAT(2X,'OUTSIDE FILM HEAT TRANSFER COEFFICIENT = ',
      # F10.2,' (BTU/SCFT-HR-F)')
00580      WRITE(22,20)
00590      WRITE(21,20)
00600      WRITE(23,20)
00610      20
00620      FORMAT(//,2X,'*****CALCULATED RESULTS*****')
00630      WRITE(22,21)
00640      21
00650      FORMAT(1X,'CELL # TW FLOW MODEL ')
00660      WRITE(21,25)

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00610 25      FORMAT(//,1X,'CELL # MLI      MGI      MLO      MGO      TI
00620      # PI      TO      PO      RL      HI      UI      LATENT
00630      #      C/A')
00640      WRITE(21,27)
00650 27      FORMAT(2X,'      LB/HR      LB/HR      LB/HR      LB/HR      F      PSIA
00660      # F      PSIA      BTU/SQFT-HR-F BTU/SQFT-HR-F BTU/HR
00670      #      BTU/SQFT')
00680      WRITE(23,28)
00690 28      FORMAT(2X,'CELL #      DML      DTP      VTP      UL      UG
00700      #      UTP      TLM      ALD      AGD      ALV      AGV')
00710 C-----CHANGE UNIT (TEMP. IN KELVIN AND PRESSURE IN ATM.)
00720      TK1      = ((T1-32.)/1.8)+273.0
00730      PA1      = P1/14.7
00740      STW      = 0.0
00750      SKLC     = 0.0
00760 C-----START THE CALCULATION FROM CELL #1 TO CELL #N
00770      DO 600 K = 1,N
00780 C-----FIRST ESTIMATE FOR OUTLET STREAM TEMPERATURE AND PRESSURE
00790 C-----OF CELL #K
00800      TK2      = TK1
00810      PA2      = PA1
00820 C-----START THE TEMPERATURE ITERATING LOOP (OUTER LOOP)
00830      K2       = 1
00840 C-----START THE PRESSURE ITERATING LOOP (INNER LOOP)
00850 29      K1      = 1
00860 C-----FEED RATES IN LB/SEC
00870      ARATE     = AIRR/3600.0
00880      WRATE     = WATERR/3600.0
00890 C-----USE SUBPROGRAM RATES TO DETERMINE THE COMPOSITIONS AND FLOW
00900 C-----RATES IN CELL #K
00910 30      CALL RATES(TK1,TK2,PA1,PA2,WRATE,ARATE,MFRL1,MFRL2,MFRV1,MFRV2,
00920      # MYA1,MYA2,MYW1,MYW2,MV1,MV2)
00930 C-----
00940 C-----AMOUNT OF EVAPORATED LIQUID IN CELL #K
00950      CML      = 3600.*(MFRL1-MFRL2)
00960 C-----AVERAGE FLOW RATES OF THE STREAMS IN CELL #K, LB/SEC
00970      ALMFR    = (MFRL1+MFRL2)/2.
00980      AGMFR    = (MFRV1+MFRV2)/2.
00990 C-----USE SUBPROGRAM PHYSIC TO DETERMINE THE PHYSICAL PROPERTIES OF
01000 C-----THE STREAMS
01010      CALL PHYSIC(TK1,TK2,PA1,PA2,MYA1,MYA2,MYW1,MYW2,AGD,ALD,AGVE,
01020      # ALVE,KLIQ,CPA1,CPA2,CPW1,CPW2,CPWV1,CPWV2,MV1,MV2)
01030 C-----
01040 C-----AVERAGE LIQUID SPECIFIC HEAT (BTU/LB-F)
01050      CPL      = 0.5*(CPW1+CPW2)
01060 C-----REYNOLD NUMBER OF THE GAS AND LIQUID PHASES
01070      REGAS    = 4.0*AGMFR/(3.1415927*DI*AGVE)
01080      RELIQ    = 4.0*ALMFR/(3.1415927*DI*ALVE)
01090 C-----FIND THE FLOW MECHANISMS OF THE SYSTEM
01100      IF(REGAS-2000.0) 32,32,33
01110 32      IF(RELIQ-2000.0) 34,34,35
01120 33      IF(RELIQ-2000.0) 36,36,37
01130 C-----FIND THE LOCKHART-MARTINELLI      PARAMETER X CORRESPONDING TO THE
01140 C-----FLOW MECHANISMS
01150 34      MECHAN  = 1
01160      PX1      = (ALMFR/AGMFR)*(AGD/ALD)
01170      PX2      = ALVE/AGVE
01180      PX       = (PX1*PX2)**0.5
01190 C-----PARAMETER "PHI" OF THE LIQUID PHASE
01200      PHIL     = (1.0+5.0/PX+1.0/(PX**2.0))**.5

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01210      CCNSTN = 1.0
01220      GO TO 35
01230 35      MECHAN = 2
01240      PX1 = .002875*(RELIC**0.8)*(AGC/ALD)
01250      PX2 = (ALMFR/AGMFR)*(ALVF/AGVE)
01260      PX = (PX1*PX2)**0.5
01270 C-----PARAMETER "PHI" OF THE LIQUID PHASE
01280      PHIL = (1.0+10.0/PX+1.0/(PX**2.0))**.5
01290      CCNSTN = 0.2
01300      GC TO 35
01310 36      MECHAN = 3
01320      PX1 = 347.826087*(REGAS**-C.8)*(AGD/ALD)
01330      PX2 = (ALMFR/AGMFR)*(ALVE/AGVE)
01340      PX = (PX1*PX2)**0.5
01350 C-----PARAMETER "PHI" OF THE LIQUID PHASE
01360      PHIL = (1.0+12.0/PX+1.0/(PX**2.0))**.5
01370      CCNSTN = 1.0
01380      GC TO 35
01390 37      MECHAN = 4
01400      PX1 = ((ALMFR/AGMFR)**1.8)*(AGD/ALD)
01410      PX2 = (ALVE/AGVE)**0.2
01420      PX = (PX1*PX2)**0.5
01430 C-----PARAMETER "PHI" OF THE LIQUID PHASE
01440 C-----CORRECTION FACTOR OF JOHNSON AND ABCU-SABE
01450      AJF = 1.25-.0025*PX
01460      FJA = 1.2-.002*PX
01470      PHIL = ((1.+(20./PX)+(1./(PX**2.0))**.5)*AJF
01480      CCNSTN = 0.2
01490 C-----LOCKHART'S AND MARTINELLI CORRELATION FOR LIQUID HOLDUP
01500 C-----FRACTION
01510 39      RLC = (PX/((PX**2.0)+20.*PX+1.0)**.5)/FJA
01520 C-----LOCKHART AND MARTINELLI CORRELATIONS FOR TWO-PHASE PRESSURE
01530 C-----DROP
01540      LPD1 = 0.092*((4.0/3.1415927)**(2.0-CCNSTN))
01550      LPD2 = (ALVE**CCNSTN)*(ALMFR**(2.0-CCNSTN))
01560      LPD3 = (DI**(5.0-CCNSTN))*ALD*32.174
01570 C-----SINGLE LIQUID PHASE PRESSURE DROP PER UNIT LENGTH OF THE TUBE,
01580 C-----PSI/FT
01590      SLPPD = (LPD1+LPD2)/(LPD3*144.0)
01600 C-----TWO-PHASE PRESSURE DROP IN CELL #K, PSI
01610      TPDROP = (PHIL**2.0)*SLPPD*CZ
01620 C-----CALCULATED OUTLET STREAM PRESSURE, ATM
01630      P2N = PA1-(TPDROP/14.7)
01640 C-----ERROR OF PRESSURE ESTIMATION
01650      CCFP = ABS(P2N-PA2)/PA2
01660 C-----PRESSURE CONVERGENCE CRITERION
01670      IF(CCFP.LE.0.002) GC TO 40
01680 C-----NEW ESTIMATE FOR OUTLET STREAM PRESSURE
01690      PA2 = (PA2+P2N)/2.0
01700      K1 = K1+1
01710      IF(K1.EC.100) GC TO 625
01720 C-----RETURN FOR ANOTHER PRESSURE ITERATING LOOP
01730      GO TO 30
01740 C-----USE SUBPROGRAM LATENT TO FIND LATENT HEAT OF THE STREAMS
01750 40      CALL LATENT(TK1,TK2,PA1,PA2,LAMDA1,LAMDA2)
01760 C-----
01770 C-----AVERAGE LATENT HEAT
01780      LTH = .5*(LAMDA1+LAMDA2)*DML
01790 C-----AVERAGE GAS MASS FRACTION IN CELL #K
01800      GASFRA = AGMFR/(AGMFR+ALMFR)

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01810 C-----CROSS-SECTIONAL AREA OF THE TUBE, SQFT
01820 ACROSS = ((DI/2.0)**2.0)*3.1415927
01830 C-----SUPERFICIAL VELOCITIES OF THE STREAMS, FT/SEC
01840 GV = AGMFR/(ACROSS*AGD)
01850 GL = ALMFR/(ACROSS*ALD)
01860 C-----SUBPROGRAM FMCDEL TO DETERMINE THE FLOW MODEL IN THE CELL
01870 CALL FMCDEL(GV,GL,TYPE)
01880 C-----
01890 C-----VELOCITY OF THE STREAMS, FT/SEC
01900 LG = GV/(1.0-RLC)
01910 UL = GL/RLC
01920 C-----AVERAGE TWO-PHASE STREAM VELOCITY, FT/SEC
01930 UTP = 1.0/((GASFRA/UG)+((1.0-GASFRA)/UL))
01940 C-----AVERAGE TWO-PHASE STREAM DENSITY, LB/CUFT
01950 DTP = 1.0/((GASFRA/AGD)+((1.0-GASFRA)/ALD))
01960 C-----AVERAGE TWO-PHASE STREAM VISCOSITY, LB/FT-SEC
01970 AVTP = 1.0/((GASFRA/AGVE)+((1.0-GASFRA)/ALVE))
01980 IF(TYPE.EQ.1.0) GO TO 100
01990 C-----TWO-PHASE HEAT TRANSFER COEFFICIENT GIVEN BY DAVIS-DAVID
02000 C-----CORRELATION
02010 C-----FOR A SEPARATED FLOW MODEL
02020 HH11 = 0.06*(KLIC/DI)*((ALD/AGD)**0.28)
02030 HH12 = (DI*GASFRA*UTP*CTP/ALVE)**0.87
02040 PRL1 = (CPL*ALVE*3600.0/KLIC)**0.4
02050 C-----TWO-PHASE HEAT TRANSFER COEFFICIENT, HI, BTU/SQFT-HR-F
02060 HI = HH11*HH12*PRL1
02070 GC TO 200
02080 C-----FOR A HOMOGENEOUS FLOW MODEL
02090 ILO HH21 = (KLIC/DI)*0.033
02100 HH22 = (DI*UTP*DTP/AVTP)**0.87
02110 PRL2 = (CPL*ALVE*3600.0/KLIC)**0.4
02120 C-----TWO-PHASE HEAT TRANSFER COEFFICIENT, H, BTU/SQFT-HR-F
02130 HI = HH21*HH22*PRL2
02140 C-----HEAT FLUX AREA OF CELL #K
02150 200 AHEAT = CI*DZ*3.1415927
02160 C-----TEMPERATURE IN FAHRENHEIT
02170 TB1 = ((TK1-273.)*1.8)+32.0
02180 TB2 = ((TK2-273.)*1.8)+32.0
02190 IF(K2.EQ.1) GGTC 250
02200 C-----LOG MEAN TEMPERATURE DIFFERENCE, F
02210 IF(TB1.EQ.TG) GC TO 300
02220 TLM = (TB2-TB1)/(ALOG((TG-TB1)/(TG-TB2)))
02230 GC TO 300
02240 C-----FIRST ESTIMATION OF LOG MEAN TEMPERATURE DIFFERENCE, F
02250 250 TLM = TG-TB1
02260 C-----LOG MEAN DIAMETER, FT
02270 300 DLM = (DC-DI)/(ALOG(DO/DI))
02280 C-----TUBE WALL THICKNESS, FT
02290 Xw = (DC-DI)/2.0
02300 C-----OVERALL HEAT TRANSFER COEFFICIENT, TUBE INSIDE DIAMETER BASIS.
02310 C-----BTU/SQFT-HR-F
02320 UI1 = (1.0/HI)+(DI/(DC*HC))
02330 UI2 = (DI/DLM)*(Xw/Kw)
02340 UI = 1.0/(UI1+UI2)
02350 C-----TOTAL HEAT ADDED TO CELL #K, BTU/HR
02360 HEAT = UI*AHEAT*TLM
02370 C-----HEAT FLUX RATE TO CELL #K, BTU/SQFT-HR
02380 HFLUX = HEAT/AHEAT
02390 C-----OUTLET STREAM TEMPERATURE CALCULATION BASED ON ENTHALPY
02400 C-----DIFFERENCE OF INLET AND OUTLET STREAMS AND TOTAL HEAT ADDED

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02410 C-----TC CELL #K
02420 400 T2N = TB1
02430 460 FF1 = HEAT-LTH-(0.5*(CPA1+CPA2)*ARATE*3600.*(T2N-TB1))
02440 FF2 = .5*(CPW1+CPW2)*(T2N-TB1)*3600.*(0.5*(WV1+WV2)+ALMFR)
02450 DFF1 = .5*(CPA1+CPA2)*ARATE*3600.
02460 DFF2 = .5*(CPW1+CPW2)*3600.*(0.5*(WV1+WV2)+ALMFR)
02470 DFF = -DFF1-DFF2
02480 T2NO = T2N
02490 C-----CALCULATED OUTLET STREAM TEMPERATURE, F
02500 T2N = T2N-((FF1-FF2)/DFF)
02510 IF((ABS(T2N-T2NO)/T2NO).LE.C.COC1) GO TO 470
02520 GO TO 460
02530 C-----ERROR OF TEMPERATURE ESTIMATION
02540 470 CCFT = ABS(T2N-TB2)/TB2
02550 C-----TEMPERATURE CONVERGENCE CRITERION
02560 IF(CCFT.LE.0.003) GO TO 500
02570 C-----NEW ESTIMATE FOR OUTLET STREAM TEMPERATURE, F
02580 TB2 = (T2N+TB2)/2.0
02590 475 TK2 = ((TB2-32.0)/1.8)+273.0
02600 K2 = K2+1
02610 IF(K2.GE.200) GO TO 650
02620 C-----RETURN FOR ANOTHER TEMPERATURE ITERATING LOOP
02630 GO TO 25
02640 C-----RESULTS PRINTED OUT
02650 500 PSIA1 = PA1*14.7
02660 PSIA2 = PA2*14.7
02670 Z11 = AVTP*3600.
02680 Z22 = UG*3600.
02690 Z33 = UL*3600.
02700 Z44 = LIP*3600.
02710 Z55 = ALVE*3600.
02720 Z66 = AGVE*3600.
02730 FOF1 = MFR1*3600.
02740 FOF2 = MFR2*3600.
02750 FOA1 = MFRV1*3600.
02760 FOA2 = MFRV2*3600.
02770 TFA = C.5*(TB1+TB2)
02780 C-----TUBE WALL TEMPERATURE
02790 TW = TFA+(TG-TFA)*(((1./FI)/((1./HI)+((DI/DO)*(1./HO))))
02800 IF(TYPE.EQ.2.0) GO TO 515
02810 WRITE(22,510) K,TW
02820 510 FORMAT(2X,I3,2X,F8.3,3X,'HOMOGENEOUS FLOW MODEL')
02830 GO TO 525
02840 515 WRITE(22,517) K,TW
02850 517 FORMAT(2X,I3,2X,F8.3,3X,'SEPARATED FLOW MODEL')
02860 525 WRITE(21,530) K,FOF1,FOA1,FOF2,FOA2,TB1,PSIA1,TB2,PSIA2,
02870 # RLC,HI,UI,LTH,HFLUX
02880 530 FORMAT(2X,I3,4F8.2,4F8.2,2X,F5.4,3F12.2,4X,F12.2)
02890 WRITE(23,540) K,DML,DTP,Z11,Z22,Z33,Z44,TLM,AGO,ALD,Z55,Z66
02900 540 FORMAT(2X,I3,3X,F7.4,2F7.4,3F10.2,F8.2,4F9.4)
02910 560 TK1 = TK2
02920 PA1 = PA2
02930 STW = STW+TW
02940 SKLC = SKLC+RLC
02950 600 CONTINUE
02960 GO TO 655
02970 C-----ERROR MESSAGE
02980 625 WRITE(21,620)
02990 620 FORMAT(2X,'TOO MANY PRESSURE LOCPS !!!!!!!!!!!!!!!!!!!')
03000 GO TO 655

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C3010 650 WRITE(21,651)
C3020 651 FORMAT(2X,'TOC MANY TEMPERATURE LOOPS !!!!!!!!!!!!!!!')
C3030 695 RLC = SRLC/N
C3040 Th = STh/N
C3050 WRITE(23,700) RLC
C3060 700 FORMAT(//,2X,'AVERAGE LIQUID FCLCUP IN THE TUBE = ',F9.5)
C3070 WRITE(22,750) Th
C3080 750 FORMAT(//,2X,'AVERAGE TUBE WALL TEMPERATURE = ',F8.2,' (F)')
C3090 STOP
C3100 C-----END OF THE MAIN PROGRAM
C3110 END
C3120 C
C3130 C
C3140 *****SUBPROGRAM RATES*****
C3150 *
C3160 * IN THIS SUBPROGRAM FLOW RATES OF GAS AND LIQUID PHASES *
C3170 * AT THE ENDS OF THE CELL ARE DETERMINED BY FLASH CALCULATIONS. *
C3180 * THE APPLICATION OF THIS CALCULATION REQUIRES THE DATA OF *
C3190 * K-VALUES OF THE COMPONENTS AS INDEPENDENT VARIABLES OF THE *
C3200 * COMPOSITIONS. *
C3210 *
C3220 *****
C3230 C
C3240 C
C3250 SUBROUTINE RATES(TK1,TK2,PA1,PA2,WRATE,ARATE,MFRL1,MFRL2,
C3260 # MFRV1,MFRV2,MYA1,MYA2,MYW1,MYW2,WV1,WV2)
C3270 REAL MFRL1,MFRL2,MFRV1,MFRV2,NM,LR1,LR2,MYA1,MYA2,MYW1,MYW2
C3280 C-----USE SUBPROGRAM KVALUE TO CALCULATE COMPOSITIONS IN THE STREAMS
C3290 CALL KVALUE(TK1,TK2,TAU1,TAU2)
C3300 C-----
C3310 C-----MOLE FRACTIONS OF THE GAS STREAMS AT INLET AND OUTLET OF CELL *K
C3320 Y1h = (217.6/PA1)*(EXP(TAU1))
C3330 Y2h = (217.6/PA2)*(EXP(TAU2))
C3340 Y1A = 1.0-Y1h
C3350 Y2A = 1.0-Y2h
C3360 C-----TOTAL NUMBER OF MOLES IN THE FEED
C3370 NM = (WRATE/18.0)+(ARATE/28.96)
C3380 C-----TOTAL NUMBER OF MOLE IN THE GAS PHASE AT INLET AND OUTLET
C3390 VR1 = (ARATE/28.96)/Y1A
C3400 VR2 = (ARATE/28.96)/Y2A
C3410 C-----TOTAL NUMBER OF MOLE IN THE LIQUID PHASE AT INLET AND OUTLET
C3420 LR1 = NM-VR1
C3430 LR2 = NM-VR2
C3440 C-----MASS FLOW RATES OF THE STREAMS AT INLET AND OUTLET, LB/SEC
C3450 MFRV1 = Y1h*18.*VR1+Y1A*28.96*VR1
C3460 MFRV2 = Y2h*18.*VR2+Y2A*28.96*VR2
C3470 MFRL1 = LR1*18.0
C3480 MFRL2 = LR2*18.0
C3490 C-----WATER VAPOR AT INLET AND OUTLET, LB/SEC
C3500 WV1 = Y1h*18.*VR1
C3510 WV2 = Y2h*18.*VR2
C3520 C-----MASS FRACTION IN THE STREAMS AT INLET AND OUTLET
C3530 MYA1 = Y1A*28.96*VR1/MFRV1
C3540 MYA2 = Y2A*28.96*VR2/MFRV2
C3550 MYW1 = 1.0-MYA1
C3560 MYW2 = 1.0-MYA2
C3570 C-----RETURN TO THE MAIN PROGRAM
C3580 RETURN
C3590 END
C3600 C

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03610 C
03620 *****SUBPROGRAM KVALUE*****
03630 *
03640 *          THIS IS A DATA SUBPROGRAM. INTERPOLATION IS APPLIED TO*
03650 *          DETERMINE K-VALUES OF THE COMPONENTS AT APPROXIMATE TEMPERATURE*
03660 *          AND PRESSURE
03670 *
03680 *****
03690 C
03700 C
03710 SUBROUTINE KVALUE(TK1,TK2,TAU1,TAU2)
03720 C
03730 C
03740 C-----FRAZIER'S AND CHANG'S CORRELATIONS TO FIND COMPOSITION IN
03750 C-----AN AIR-WATER SYSTEM
03760 C-----NORMALIZED CRITICAL TEMPERATURE
03770 PC = (374.3+273.)/(100.+273.)
03780 C-----NORMALIZED TEMPERATURE OF INLET AND OUTLET STREAMS
03790 PS1 = TK1/(100.+273.)
03800 PS2 = TK2/(100.+273.)
03810 C-----FUNCTION TAU
03820 TAU1 = -7.4691*((PC/PS1)-1.)/(1.+81642*(PC-PS1))
03830 TAU2 = 1.+67508*(PC-PS1)+.18746*((PC-PS1)**3.)
03840 TAU3 = -7.4691*((PC/PS2)-1.)/(1.+81642*(PC-PS2))
03850 TAU4 = 1.+67508*(PC-PS2)+.18746*((PC-PS2)**3.)
03860 TAU1 = TAU1*TAU2
03870 TAU2 = TAU3*TAU4
03880 C-----RETURN TO SUBPROGRAM RATE
03890 RETURN
03900 END
03910 C
03920 C
03930 *****SUBPROGRAM PHYSIC*****
03940 *
03950 *          THIS SUBPROGRAM IS USED TO FIND THE PHYSICAL CONSTANTS *
03960 *          OF THE COMPONENTS AT APPROXIMATE TEMPERATURE AND PRESSURE *
03970 *
03980 *****
03990 C
04000 C
04010 SUBROUTINE PHYSIC(TK1,TK2,PA1,PA2,MYA1,MYA2,MYW1,MYW2,AGD,ALI),
04020 # AGVE,ALVE,KLIC,CPA1,CPA2,CPW1,CPW2,CPhV1,CPhV2)
04030 C
04040 C
04050 REAL NMW1,NMW2,MFRV1,MFRV2,MFRL1,SVVW(76),SVLW(76),TP(76),
04060 # LAMDA2,MFRL2,MYA1,MYA2,MYW1,MYW2,LRL,LR2,KLIQ1,KLIQ2,KLIC,
04070 # TK(11),TLC(11),SH(5,9),AP(5),AT(9),DENS(5,9),TCL(11),APSH1(9),
04080 # APSH2(9),APDEN1(9),APDEN2(9)
04090 C-----PRESSURE RANGE FOR DENSITY AND SPECIFIC HEAT OF AIR, ATM
04100 DATA AP/1.,2.,3.,5.,7./
04110 C-----TEMPERATURE RANGE FOR DENS.&SPECIFIC HEAT OF AIR, K
04120 DATA AT/280.,290.,300.,320.,340.,360.,380.,400.,450./
04130 C-----DATA FOR AIR SPECIFIC HEAT, BTU/LB-F
04140 DATA SH/29.11,29.17,29.22,29.34,29.45,29.12,29.17,29.22,29.32,
04150 # 29.42,29.13,29.18,29.22,29.31,29.40,29.16,29.20,29.24,29.32,
04160 # 29.40,29.19,29.22,29.26,29.33,29.40,29.22,29.25,29.28,29.34,
04170 # 29.41,29.25,29.28,29.31,29.36,29.42,29.28,29.31,29.34,29.38,
04180 # 29.43,29.36,29.38,29.39,29.43,29.47/
04190 C-----DATA FOR AIR DENSITY, LB/CUFT
04200 DATA DENS/22965.,11477.,7648.,4584.,3271.,23788.,11889.,7923.,

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04210      # 4750.,3391.,24610.,12302.,8199.,4916.,3410.,26255.,13126.,
04220      # 8749.,5248.,3748.,27900.,13949.,9299.,5580.,3986.,29543.,
04230      # 14773.,5849.,5911.,4223.,31187.,15596.,10398.,6241.,4459.,
04240      # 32830.,16418.,10947.,6571.,4695.,36937.,18474.,12319.,7396.,
04250      # 5286./
04260      C-----DATA FOR WATER VAPOR DENSITY, LB/CUFT
04270      DATA SVVW/1207.6,1129.2,1056.5,989.1,926.5,868.4,814.3,
04280      # 764.1,717.4,673.9,633.3,595.5,560.3,527.5,456.8,468.1,441.3,
04290      # 416.3,392.9,370.9,350.4,331.1,313.1,296.18,280.3,265.39,
04300      # 251.38,238.22,225.85,241.21,203.36,192.95,183.24,174.09,
04310      # 165.47,157.33,149.66,142.41,135.57,129.11,123.00,117.22,
04320      # 111.76,106.59,101.70,97.07,92.68,88.52,84.57,80.83,77.29,73.92,
04330      # 70.72,67.68,64.80,62.06,59.45,56.97,54.61,52.36,50.22,48.189,
04340      # 46.249,44.40,42.638,40.957,39.354,37.824,36.364,34.97,33.639,
04350      # 32.395,31.151,30.015,28.878,27.839/
04360      C-----DATA FOR WATER DENSITY, LB/CUFT
04370      DATA SVLW/.016033,.016036,.016039,.016043,.016046,.016050,
04380      # .016054,.016058,.016063,.016067,.016072,.016077,.016082,
04390      # .016087,.016093,.016099,.016105,.016111,.016117,.016123,
04400      # .016130,.016137,.016144,.016151,.016158,.016165,.016173,
04410      # .016180,.016188,.016196,.016204,.016213,.016221,.016229,
04420      # .016238,.016247,.016256,.016265,.016274,.016284,.016293,
04430      # .016303,.016313,.016322,.016332,.016343,.016353,.016363,
04440      # .016374,.016384,.016395,.016406,.016417,.016428,.016440,
04450      # .016451,.016463,.016474,.016486,.016498,.016510,.016522,
04460      # .016534,.016547,.016559,.016572,.016585,.016598,.016611,
04470      # .016624,.016637,.016651,.016664,.016678,.016691,.016705/
04480      C-----TEMPERATURE RANGE FOR WATER THERMAL CONDUCTIVITY DATA, K
04490      DATA TK/288.56,294.11,299.67,305.22,310.78,321.89,333.00,
04500      # 344.11,355.22,366.33,377.44/
04510      C-----THERMAL CONDUCTIVITY DATA FOR WATER, BTU/FT-HR-F
04520      DATA TCL/.34,.346,.352,.358,.362,.371,.378,.384,.388,.392,.394/
04530      C-----TEMPERATURE RANGE FOR WATER VAPOR AND WATER DENSITY DATA
04540      TP(1) = 60.0
04550      DO 10 I = 1,75
04560      TP(I+1) = TP(I)+2.0
04570      10      CCNTINUE
04580      DO 30 I = 1,76
04590      IP(I) = ((TP(I)-32.0)/1.8)+273.0
04600      30      CCNTINUE
04610      C-----INTERPOLATION PROCESSES
04620      DO 60 I = 1,76
04630      IF(TP(I)-TK1) 60,60,50
04640      50      TPTIE = (TP(I)-TP(I-1))/(TP(I)-TK1)
04650      SVLW1 = SVLW(I)-(SVLW(I)-SVLW(I-1))/TPTIE
04660      SVVW1 = SVVW(I)-(SVVW(I)-SVVW(I-1))/TPTIE
04670      GO TO 70
04680      60      CCNTINUE
04690      DO 90 I = 1,76
04700      IF(TP(I)-TK2) 90,90,80
04710      80      TPTIE = (TP(I)-TP(I-1))/(TP(I)-TK2)
04720      SVLW2 = SVLW(I)-(SVLW(I)-SVLW(I-1))/TPTIE
04730      SVVW2 = SVVW(I)-(SVVW(I)-SVVW(I-1))/TPTIE
04740      GO TO 100
04750      90      CCNTINUE
04760      DO 105 I = 1,5
04770      IF(AP(I)-PA1) 105,105,95
04780      95      PITIE = (AP(I)-AP(I-1))/(AP(I)-PA1)
04790      DO 98 J = 1,9
04800      APSH1(J) = SH(I,J)-(SH(I,J)-SH(I-1,J))/PITIE

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04810      APDEN1(J)= DENS(I,J)-(DENS(I,J)-DENS(I-1,J))/P1TIE
04820 98      CCNTINUE
04830      GO TO 110
04840 1C5     CCNTINUE
04850 110     DO 200 I = 1,5
04860      IF(AT(I)-TK1) 200,200,120
04870 120     T1TIE = (AT(I)-AT(I-1))/(AT(I)-TK1)
04880      CCAIR1 = APSH1(I)-(APSH1(I)-APSH1(I-1))/T1TIE
04890      DCAIR1 = APDEN1(I)-(APDEN1(I)-APDEN1(I-1))/T1TIE
04900      GO TO 210
04910 200     CCNTINUE
04920 210     DO 300 I = 1,5
04930      IF(AP(I)-PA2) 300,300,220
04940 220     P2TIE = (AP(I)-AP(I-1))/(AP(I)-PA2)
04950      DC 230 J = 1,9
04960      APSH2(J) = SH(I,J)-(SH(I,J)-SH(I-1,J))/P2TIE
04970      APDEN2(J) = DENS(I,J)-(DENS(I,J)-DENS(I-1,J))/P2TIE
04980 230     CCNTINUE
04990      GO TO 310
05000 300     CCNTINUE
05010 C-----RETURN TO THE MAIN PROGRAM
05020 310     DO 400 I = 1,9
05030      IF(AT(I)-TK2) 400,400,320
05040 320     T2TIE = (AT(I)-AT(I-1))/(AT(I)-TK2)
05050      CCAIR2 = APSH2(I)-(APSH2(I)-APSH2(I-1))/T2TIE
05060      DCAIR2 = APDEN2(I)-(APDEN2(I)-APDEN2(I-1))/T2TIE
05070      GO TO 410
05080 400     CCNTINUE
05090 410     DC 500 I = 1,11
05100      IF(TK(I)-TK1) 500,500,450
05110 450     TKTIE1 = (TK(I)-TK(I-1))/(TK(I)-TK1)
05120      KLIQ1 = TCL(I)-(TCL(I)-TCL(I-1))/TKTIE1
05130      GO TO 510
05140      CCNTINUE
05150 510     DO 600 I = 1,11
05160      IF(TK(I)-TK2) 600,600,550
05170 550     TKTIE2 = (TK(I)-TK(I-1))/(TK(I)-TK2)
05180      KLIQ2 = TCL(I)-(TCL(I)-TCL(I-1))/TKTIE2
05190      GO TO 610
05200 600     CCNTINUE
05210 C-----DENSITY OF WATER VAPOR, LB/CUFT
05220 610     DVW1 = 1.0/SVW1
05230     DVW2 = 1.0/SVW2
05240 C-----DENSITY OF WATER, LB/CUFT
05250     DLW1 = 1.0/SVLW1
05260     DLW2 = 1.0/SVLW2
05270 C-----SPECIFIC HEAT OF AIR, BTU/LB-F
05280     CAIR1 = CCAIR1/(28.96*4.187)
05290     CAIR2 = CCAIR2/(28.96*4.187)
05300 C-----DENSITY OF AIR, LB/CUFT
05310     DAIR1 = 1810.237203/CAIR1
05320     DAIR2 = 1810.237203/DCAIR2
05330 C-----EQUATIONS TO ESTIMATE SPECIFIC HEAT OF WATER VAPOR AND WATER,
05340 C-----BTU/LB-F
05350     SHLR = (1.06-0.99)/(371.89-288.56)
05360     CPW1 = ((TK1-288.56)*SHLR)+0.99
05370     CPW2 = ((TK2-288.56)*SHLR)+0.99
05380     CPWV1 = ((.5-.45)/(700.-283.))*TK1+.45
05390     CPWV2 = ((.5-.45)/(700.-283.))*TK1+.45
05400 C-----EQUATIONS TO ESTIMATE VISCOSITY

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05410      WVVR      = (0.0125-0.0093)/(371.89-288.56)
05420      WVVIS1   = ((TK1-288.56)*WVVR)+0.0093
05430      WVVIS2   = ((TK2-288.56)*WVVR)+0.0093
05440      AVR      = (0.021-0.0175)/(371.89-288.56)
05450      AVIS1     = ((TK1-288.56)*AVR)+0.0175
05460      AVIS2     = ((TK2-288.56)*AVR)+0.0175
05470      ChVR     = (0.445-0.45)/(371.89-288.56)
05480      ChV1      = ((TK1-288.56)*ChVR)+0.445
05490      ChV2      = ((TK2-288.56)*ChVR)+0.445
05500      WLVIS1    = 10.0**((658.25*((1./TK1)-(1./283.16))))
05510      WLVIS2    = 10.0**((658.25*((1./TK2)-(1./283.16))))
05520      AGD1      = MYA1*DAIR1+MYW1*DVW1
05530      AGD2      = MYA2*DAIR2+MYW2*DVW2
05540      C-----AVERAGE DENSITY OF THE GAS PHASE, LB/CUFT
05550      AGD        = 0.5*(AGD1+AGD2)
05560      C-----AVERAGE DENSITY OF THE LIQUID PHASE, LB/CUFT
05570      ALD        = 0.5*(DLW1+DLW2)
05580      AGV1       = MYA1*AVIS1+MYW1*WVVIS1
05590      AGV2       = MYA2*AVIS2+MYW2*WVVIS2
05600      C-----AVERAGE VISCOSITY OF THE GAS PHASE, LB/FT-SEC
05610      AGVE       = 0.5*(AGV1+AGV2)*C.CCC67197
05620      C-----AVERAGE VISCOSITY OF THE LIQUID PHASE, LB/FT-SEC
05630      ALVE       = 0.5*(WLVIS1+WLVIS2)*C.00067197
05640      C-----AVERAGE THERMAL CONDUCTIVITY OF THE LIQUID PHASE, BTU/FT-H-F
05650      KLIQ       = .5*(KLIQ1+KLIQ2)
05660      C-----RETURN TO THE MAIN PROGRAM
05670      RETURN
05680      END
05690      C
05700      C
05710      *****SUBPROGRAM LATENT*****
05720      *
05730      *           THIS SUBPKGGRAM IS USED TO FIND THE HEAT OF EVAPORATION*
05740      *           OF THE COMPONENTS AT APPROXIMATE TEMPERATURE AND PRESSURE *
05750      *
05760      *****
05770      C
05780      C
05790      SUBROUTINE LATENT(TK1,TK2,PA1,PA2,LAMDA1,LAMDA2)
05800      C
05810      C
05820      REAL LAMDA1,LAMDA2,LAMDA(76),TP(76)
05830      C-----DATA OF LATENT HEAT, BTU/LB
05840      DATA LAMDA/1059.57,1058.5,1057.4,1056.3,1055.2,1054.0,1052.9,
05850      # 1051.8,1050.7,1049.5,1048.4,1047.3,1046.1,1045.0,1043.9,1042.7,
05860      # 1041.6,1040.5,1039.3,1038.2,1037.1,1035.9,1034.8,1033.6,1032.5,
05870      # 1031.4,1030.2,1029.1,1027.9,1026.8,1025.6,1024.5,1023.3,1022.2,
05880      # 1021.0,1019.8,1018.7,1017.5,1016.4,1015.2,1014.0,1012.9,1011.7,
05890      # 1010.5,1009.3,1008.2,1007.0,1005.8,1004.6,1003.4,1002.2,1001.0,
05900      # 999.8,998.6,997.4,996.2,995.0,993.8,992.6,991.4,990.2,989.0,
05910      # 987.8,986.5,985.3,984.1,982.8,981.6,980.4,979.1,977.9,976.6,
05920      # 975.4,974.1,972.8,971.5/
05930      C-----TEMPERATURE RANGE
05940      TP(1)        = 60.0
05950      DO 10 I = 1,75
05960      TP(I+1)      = TP(I)+2.0
05970      10 CONTINUE
05980      DO 30 I = 1,76
05990      TP(I)        = ((TP(I)-32.0)/1.8)+273.0
06000      30 CONTINUE

```

```

06010 C-----INTERPOLATION PROCESSES
06020      DO 60 I = 1,76
06030      IF(TP(I)-TK1) 60,60,50
06040      50      TPTIE = (TP(I)-TP(I-1))/(TP(I)-TK1)
06050      LAMDA1 = LAMDA(I)-(LAMDA(I)-LAMDA(I-1))/TPTIE
06060      GO TO 70
06070      60      CCNTINUE
06080      70      DO 90 I = 1,76
06090      IF(TP(I)-TK2) 90,90,80
06100      80      TPTIE = (TP(I)-TP(I-1))/(TP(I)-TK2)
06110      LAMDA2 = LAMDA(I)-(LAMDA(I)-LAMDA(I-1))/TPTIE
06120      GO TO 100
06130      90      CCNTINUE
06140 C-----RETURN TO THE MAIN PROGRAM
06150      100      RETURN
06160      END
06170 C
06180 C
06190 *****SUBPROGRAM FMODEL*****
06200 *
06210 *      THIS SUBPROGRAM IS USED TO FIND THE FLOW MODEL BASED ON*
06220 *      THE FLOW-PATTERN MAP OF MANHANCE ET AL. THERE ARE TWO FLOW *
06230 *      MODEL:
06240 *      HOMOGENEOUS MODEL WHICH INCLUDES SLUG, DISPERSEC,AND *
06250 *      BUBBLE FLOWS.
06260 *      SEPARATED MODEL WHICH INCLUDES STRATIFIED, WAVE, AND *
06270 *      ANNULAR FLCWS.
06280 *
06290 *****
06300      SUBROUTINE FMODEL(GV,GL,TYPE)
06310 C
06320 C
06330 C-----SUPERFICIAL VELOCITY, M/SEC
06340      VSG = GV*0.3048
06350      VSL = GL*0.3048
06360 C-----SIMPLIFIED MANHANE'S FLOW PATTERN MAP
06370      IF(VSG.LE.1.8) GO TO 50
06380      IF(VSG.LE.2.3) GO TO 60
06390      IF(VSG.LE.12.0) GO TO 70
06400      IF(VSG.LE.90.0) GO TO 80
06410      50      IF(VSL.LE.0.5) GO TO 300
06420      GO TO 200
06430      60      HCRZ = -0.3*VSG+1.04
06440      IF(VSL.LE.HCRZ) GO TO 300
06450      GO TO 200
06460      70      IF(VSL.LE.0.35) GO TO 300
06470      GO TO 200
06480      80      HCRZ = 0.11*VSG+0.097
06490      IF(VSL.LE.HCRZ) GO TO 300
06500      200      TYPE = 1.0
06510      GO TO 400
06520      300      TYPE = 2.0
06530 C-----RETURN TO THE MAIN PROGRAM
06540      400      RETURN
06550      END

```

J.F. JORDAN

```

WATER FLOW RATE = 975.00 LBS/HO
WATER FLOW RATE = 201.10 LBS/HO, ( 00
INLET STEAM TEMPERATURE = 65.00 (F)
INLET STEAM PRESSURE = 70.00 (PSIA)
JACKET STEAM TEMPERATURE = 73.00 (F)
JACKET STEAM PRESSURE = 70.00 (PSIA)
OUTSIDE FILM TRANSFER COEFFICIENT = 2000.00 (BTU/FOOT-HO-F)

```

*****CIN ATFG 2651175*****

[illegible]

AVERAGE LIQUID MOLALITY IN THE TUBS = 0.09394

CELL #	TM	FLY MODEL
1	216.245	SEPARATED FLY MODEL
2	215.910	SEPARATED FLY MODEL
3	216.198	SEPARATED FLY MODEL
4	217.449	SEPARATED FLY MODEL
5	215.063	SEPARATED FLY MODEL
6	219.168	SEPARATED FLY MODEL
7	217.366	SEPARATED FLY MODEL
8	222.153	SEPARATED FLY MODEL
9	224.121	SEPARATED FLY MODEL
10	223.137	SEPARATED FLY MODEL

D.2 Computer Code for the Multicomponent, Gas-Liquid System

Program FLOW2.FOR is developed for the multicomponent, gas-hydrocarbon liquids system. Operating conditions of this run are listed in Table 5.6 (page 91). Interpretation are provided along the code structure. The results given by this program are based on the Johnson-Abou Sabe correlations for liquid holdup and pressure drop.

```

00010 *****PROGRAM FLOW2.FOR*****
00020 *
00030 *
00040 *
00050 *           INTRODUCTION
00060 *           THIS PROGRAM IS DEVELOPED FOR USE IN THE DESIGN OF
00070 *           THE TUBULAR HEATER FOR A TWO-PHASE, MULTICOMPONENTS, GAS-LIQUID*
00080 *           STREAM. BRIEFLY, GIVEN THE COMPOSITIONS OF THE INLET STREAMS *
00090 *           AND THE HEAT FLUX RATE TO THE FLOW SYSTEM, HOW LONG MUST *
00100 *           THE TUBE BE IN ORDER TO RAISE THE OUTLET STREAM TO A DESIRED *
00110 *           TEMPERATURE.
00120 *
00130 *****MAIN PROGRAM*****
00140 *
00150 *           THE MAIN PROGRAM CONSIST OF TWO ITERATING LOOPS, THE
00160 *           INNER LOOP AND THE OUTER LOOP, TO DETERMINE PRESSURE AND
00170 *           TEMPERATURE OF THE OUTLET STREAM.
00180 *           THE CALCULATION OF PRESSURE DROP IS BASED ON LOCKHART
00190 *           AND MARTINELLI METHOD. THE ITERATION OF THE PRESSURE LOOP IS
00200 *           CONVERGED WHEN THE DIFFERENCE BETWEEN THE ESTIMATED AND
00210 *           CALCULATED PRESSURE DROPS IS LESS THAN 2%.
00220 *           THE OUTER LOOP IS THE TEMPERATURE ITERATING LOOP. THE
00230 *           CONVERGENCE OF THIS LOOP IS BASED ON THE CONSERVATION OF ENERGY.*
00240 *
00250 *****
00260 *
00270 *
00280 *           INTEGER PLOOP,TLOOP,TYPE
00290 *           REAL LFR,P(50),T(50,100),XP(4),YP(4),VG(4),VL(4),CPL(4),CPG(4),
00300 *           # DVAP(4),DLIC(4),TCL(4),LFRATE(50),GFRATE(50),GASDEN(50),TZ(50),
00310 *           # LIQDEN(50),KLIQ(50),GASVIS(50),LIQVIS(50),SPHG(50),SPHL(50),
00320 *           # REGAS(50),RELIQ(50),LAMDA(4),LTHEAT,LPD1,LPD2,LPD3,SHV(50),
00330 *           # LPURCP,HGV(50),Kw,CCFT,LPD4,YV(4)
00340 C-----SPECIFIED INFORMATION
00350 C-----TUBE WALL THERMAL CONDUCTIVITY,OUTSIDE TUBE HEAT TRANSFER
00360 C-----COEFFICIENT, AND HEATING MEDIUM TEMPERATURE
00370 DATA Kw,HQ,TG/64.0,2000.0,230.0/
00380 C-----INLET TEMPERATURE AND PRESSURE
00390 DATA T(1,1),P(1)/80.0,60.0/
00400 C-----NUMBER OF COMPONENTS, INSIDE AND OUTSIDE TUBE DIAMETERS
00410 DATA NC,DI,DO/4,0.0725,0.083333/
00420 C-----CELL LENGTH, NUMBER OF CELL-1
00430 DATA DZ,N/1.57,11/
00440 II = 1
00450 C-----STARTING VALUES FOR THE SUMMATION OF PRESSURE DROP AND
00460 C-----INTERPHASE MASS TRANSFER
00470 DELTAP = 0.0
00480 TEXCHM = C.C
00490 SRLC = 0.0
00500 STW = C.C
00510 C-----PRINT OUT THE SPECIFIED INFORMATION
00520 WRITE(23,1)
00530 1 FORMAT(2X,'SPECIFIED CONDITIONS AND INFORMATION')
00540 WRITE(23,2) NC
00550 2 FORMAT(2X,'NUMBER OF COMPONENTS IN THE SYSTEM = ',I2)
00560 WRITE(23,4) DI,DO,DZ
00570 4 FORMAT(2X,'INSIDE TUBE DIAMETER = ',F7.5,' (FT), OUTSIDE TUBE
00580 # DIAMETER = ',F7.5,' (FT), CELL LENGTH = ',F7.2,' (FT)')
00590 NUMC = N-1
00600 WRITE(23,5) NUMC

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00610 5      FORMAT(2X,'NUMBER OF CELLS IN THE TUBE =',I3)
00620      WRITE(23,6) T(1,1),P(1)
00630 6      FORMAT(2X,'INPUT STREAM TEMPERATURE = ',F7.2,' (F), INPUT STREAM
00640      #      PRESSURE = ',F7.2,' (PSIA)')
00650      WRITE(23,3) TG
00660 8      FORMAT(2X,'AVERAGE TEMPERATURE OF THE HEATING MEDIUM = ',F7.2,
00670      #      ' (F)')
00680      WRITE(23,10) HG
00690 10     FORMAT(2X,'OUTSIDE TUBE HEAT TRANSFER COEFFICIENT = ',F10.2,
00700      #      ' (BTU/SQFT-HR)')
00710      WRITE(23,12) KW
00720 12     FORMAT(2X,'TUBE WALL THERMAL CONDUCTIVITY = ',F6.2,
00730      #      ' (BTU/FT-H-F)')
00740 30     WRITE(23,31) I1
00750 31     FORMAT(//,2X,'LOCATION ON THE TUBE : Z(',I2,')')
00760      IF(I1.GT.1) GO TO 35
00770      WRITE(23,32) I1,T(I1,1),I1,P(I1)
00780 32     FORMAT(2X,'TEMPERATURE AT Z(',I2,') = ',F7.2,' (F), AND
00790      #      PRESSURE AT Z(',I2,') = ',F6.2,' (PSIA)')
00800 35     TLOCP = 1
00810 38     PLOCP = 1
00820 40     TEMP = T(I1,TLOCP)
00830      PRESS = P(I1)
00840 C-----USE SUBPROGRAM RATES TO DETERMINE THE COMPOSITION OF THE ENDS
00850 C-----OF THE CELL.
00860      CALL RATES(NG,I1,TEMP,PRESS,LFR,GFR,XP,YP)
00870 C-----
00880 C-----MASS FLOW RATE OF THE GAS PHASE (LBS/HR)
00890      GFRATE(I1) = GFR
00900 C-----MASS FLOW RATE OF THE LIQUID PHASE (LBS/HR)
00910      LFRATE(I1) = LFR
00920 C-----USE SUBPROGRAM PHYSIC TO DETERMINE THE PHYSICAL PROPERTIES OF
00930 C-----THE COMPONENTS.
00940      CALL PHYSIC(NG,TEMP,PRESS,VG,VL,CPG,CPL,DVAP,DLIQ,TCL)
00950 C-----
00960 C-----FIND THE PHYSICAL PROPERTIES OF THE STREAM AT THE ENDS OF THE
00970 C-----CELL BY USING THE MASS FRACTIONS OF THE COMPONENTS
00980      LIQVIS(I1) = 0.C
00990      GASVIS(I1) = 0.C
01000      GASDEN(I1) = 0.0
01010      LIQDEN(I1) = 0.C
01020      KLIQ(I1) = 0.C
01030      SPHG(I1) = 0.0
01040      SPHL(I1) = 0.C
01050      DC 50 I3 = 1,NG
01060      GASDEN(I1) = GASDEN(I1)+DVAP(I3)*YP(I3)
01070      LIQDEN(I1) = LIQDEN(I1)+DLIQ(I3)*XP(I3)
01080      KLIQ(I1) = KLIQ(I1)+TCL(I3)*XP(I3)
01090      GASVIS(I1) = GASVIS(I1)+VG(I3)*YP(I3)
01100      LIQVIS(I1) = LIQVIS(I1)+VL(I3)*XP(I3)
01110      SPHG(I1) = SPHG(I1)+CPG(I3)*YP(I3)
01120      SPHL(I1) = SPHL(I1)+CPL(I3)*XP(I3)
01130 5C      CONTINUE
01140 C-----REYNOLD NUMBER OF THE GAS PHASE
01150      REGAS(I1) = 4.0*GFRATE(I1)/(7.6*GASVIS(I1))
01160 C-----REYNOLD NUMBER OF THE LIQUID PHASE
01170      RELIQ(I1) = 4.0*LFRATE(I1)/(7.6*LIQVIS(I1))
01180 C-----SUBPROGRAM LATENT TO DETERMINE LATENT HEAT OF THE COMPONENTS
01190      CALL LATENT(NG,TEMP,LAMDA)
01200 C-----

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01210      HCV(II) = 0.0
01220      DO 95 I2 = 1,AC
01230      HCV(II) = XP(I3)*LAMDA(I3)+HCV(II)
01240  95      CCNTINUE
01250      IF(II.EC.1) GC TO 394
01260  C-----AVERAGE GAS DENSITY IN THE CELL (LB/CUFT)
01270      AGD = C.5*(GASDEN(II)+GASDEN(II-1))
01280  C-----AVERAGE LIQUID DENSITY IN THE CELL (LB/CUFT)
01290      ALD = C.5*(LIQDEN(II)+LIQDEN(II-1))
01300  C-----AVERAGE LIQUID THERMAL CONDUCTIVITY IN THE CELL (BTU/FT-HR-F)
01310      AKLIQ = C.5*(KLIQ(II)+KLIQ(II-1))
01320  C-----AVERAGE GAS VISCOSITY IN THE CELL (CP)
01330      AGVIS = C.5*(GASVIS(II)+GASVIS(II-1))
01340  C-----AVERAGE LIQUID VISCOSITY IN THE CELL (CP)
01350      ALVIS = C.5*(LIQVIS(II)+LIQVIS(II-1))
01360  C-----AVERAGE SPECIFIC HEAT OF GAS IN THE CELL (BTU/SQFT-HR-F)
01370      ASPHG = C.5*(SPHG(II)+SPHG(II-1))
01380  C-----AVERAGE SPECIFIC HEAT OF LIQUID IN THE CELL (BTU/SQFT-HR-F)
01390      ASPHL = C.5*(SPHL(II)+SPHL(II-1))
01400  C-----AVERAGE GAS MASS FLOW RATE IN THE CELL (LB/HR)
01410      AGFR = C.5*(GFRATE(II)+GFRATE(II-1))
01420  C-----AVERAGE LIQUID MASS FLOW RATE IN THE CELL (LB/HR)
01430      ALFR = C.5*(LFRATE(II)+LFRATE(II-1))
01440  C-----INTERPHASE MASS TRANSFER IN THE CELL (LB/HR)
01450      EXCHM = LFRATE(II-1)-LFRATE(II)
01460  C-----AVERAGE REYNOLD NUMBER OF THE GAS PHASE IN THE CELL
01470      AREGAS = C.5*(REGAS(II)+REGAS(II-1))
01480  C-----AVERAGE REYNOLD NUMBER OF THE LIQUID PHASE IN THE CELL
01490      ARELIQ = C.5*(RELIQ(II)+RELIQ(II-1))
01500  C-----GAS MASS FRACTION
01510      GASFRA = AGFR/(AGFR+ALFR)
01520  C-----FIND THE FLOW MECHANISM OF THE SYSTEM
01530  120      IF(AREGAS-2000.0) 124,134,135
01540  134      IF(ARELIQ-2000.0) 140,140,142
01550  135      IF(ARELIQ-2000.0) 144,144,146
01560  C-----LUCKHART AND MARTINELLI PARAMETER X CORRESPONDING TO FLOW
01570  C-----MECHANISMS
01580  140      MECHAN = 1.0
01590      PX1 = (ALFR/AGFR)*(AGD/ALD)
01600      PX2 = ALVIS/AGVIS
01610      PX = (PX1*PX2)**0.5
01620  C-----PARAMETER "PHI" OF THE LIQUID PHASE
01630      PHIL = (1.0+5.0/PX+1.0/(PX**2.0))**.5
01640      CCNSTN = 1.0
01650      GC TO 149
01660  142      MECHAN = 2
01670      PX1 = .002875*(ARELIQ**0.8)*(AGD/ALD)
01680      PX2 = (ALFR/AGFR)*(ALVIS/AGVIS)
01690      PX = (PX1*PX2)**0.5
01700  C-----PARAMETER "PHI" OF THE LIQUID PHASE
01710      PHIL = (1.0+10.0/PX+1.0/(PX**2.0))**.5
01720      CCNSTN = 0.2
01730      GC TO 149
01740  144      MECHAN = 3
01750      PX1 = 347.826087*(AREGAS**0.8)*(AGD/ALD)
01760      PX2 = (ALFR/AGFR)*(ALVIS/AGVIS)
01770      PX = (PX1*PX2)**0.5
01780  C-----PARAMETER "PHI" OF THE LIQUID PHASE
01790      PHIL = ((1.0+12.0/PX+1.0/(PX**2.0))**.5)
01800      CCNSTN = 1.0

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01810      GO TO 149
01820 146     MECHAN = 4
01830      PX1 = ((ALFR/AGFR)**1.8)*(AGD/ALD)
01840      PX2 = (ALVIS/AGVIS)**0.2
01850      PX = (PX1*PX2)**0.5
01860 C-----PARAMETER "PHI" OF THE LIQUID PHASE
01870 C-----CORRECTION FACTORS BASED ON JOHNSON AND ABOU-SABE
01880      AJF = 1.25-0.0025*PX
01890      FJA = 1.2-.002*PX
01900      PHIL = ((1.C+20./PX+1./((PX**2.))**0.5)*AJF
01910      CONSTN = 0.2
01920 C-----AVERAGE LIQUID HOLDUP FRACTION
01930 149     RLC = (PX/((PX**2.)+20.*PX+1.))**.5)/FJA
01940 C-----DETERMINE PRESSURE DROP IN THE CELL BY LOCKHART AND MARTINELLI
01950 C-----CORRELATIONS
01960      LPD1 = 0.092*((4./3.1415927)**(2.0-CONSTN))
01970      LPD2 = ((ALVIS*.00067197)**CONSTN)
01980      LPD3 = (ALFR/3600.))**(2.0-CONSTN)
01990      LPD4 = (DI**(5.0-CONSTN))*ALD*32.174
02000      LPDRCF = (LPD1*LPD2*LPD3)/(LPD4*144.0)
02010      TPDRCP = (PHIL**2.0)*LPDRCF*DZ
02020 C-----COMPARE ESTIMATED AND CALCULATED OUTLET STREAM PRESSURE
02030      P2N = P(II-1)-TPDRCP
02040      CCFP = ABS(P2N-P(II))/P(II)
02050 C-----PRESSURE CONVERGENCE CRIT.
02060      IF(CCFP.LE.0.002) GO TO 160
02070      P(II) = (P(II)+P2N)/2.0
02080      PLCCP = PLCCP+1
02090      IF(PLCCP.EQ.30) GO TO 500
02100 C-----RETURN FOR A NEW PRESSURE ITERATING LOOP
02110      GO TO 4C
02120 160     P(II) = P2N
02130 C-----CROSS-SECTIONAL AREA OF THE TUBE (SQFT)
02140      ACROSS = ((DI/2)**2.0)*3.1415927
02150 C-----SUPERFICIAL GAS VELOCITY, (FT/HR)
02160      GV = AGFR/(ACROSS*AGD)
02170 C-----SUPERFICIAL LIQUID VELOCITY, (FT/HR)
02180      GL = ALFR/(ACROSS*ALD)
02190 C-----SUBPROGRAM FMDEL TO FIND FLOW MODEL
02200      CALL FMDEL(GV,GL,TYPE)
02210 C-----
02220 C-----GAS VELOCITY BASED ON HOLDUP FRACTION (FT/HR)
02230      UG = GV/(1.-RLC)
02240 C-----LIQUID VELOCITY BASED ON HOLDUP FRACTION (FT/HR)
02250      UL = GL/RLC
02260 C-----TWO-PHASE STREAM VELOCITY (FT/HR)
02270      UTP = 1.0/((GASFRA/UG)+((1.-GASFRA)/UL))
02280 C-----TWO-PHASE STREAM DENSITY (LB/CUFT)
02290      DTP = 1.0/((GASFRA/AGD)+((1.-GASFRA)/ALD))
02300 C-----TWO-PHASE STREAM VISCOSITY (CP)
02310      VTP = 1.0/((GASFRA/AGVIS)+((1.-GASFRA)/ALVIS))
02320 C-----LATENT HEAT
02330      LTHEAT = .5*(HOV(II)+HOV(II-1))*(LFRATE(II-1)-LFRATE(II))
02340 C-----DAVIS AND DAVID CORRELATIONS TO DETERMINE TWO-PHASE HEAT
02350 C-----TRANSFER COEFFICIENT
02360      IF(TYPE.EQ.2) GO TO 300
02370 C-----TWO-PHASE GAS-LIQUID HEAT TRANSFER COEFFICIENT (BTU/SQFT-HR-F)
02380 C-----FOR HOMOLOGEOUS FLOW MODEL
02390      H1 = (AKLIQ/DI)*0.033
02400      H2 = ((DI*UTP*DTP)/(VTP*2.419092))**0.87

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02410      PRL      = ((ASPHL*ALVIS*2.419092)/AKLIQ)**0.4
02420      HI       = F1*H2*PRL
02430      GO TO 320
02440  C-----GAS MASS FRACTION GREATER THAN 0.1
02450  200      IF(TYPE.EQ.2) GO TO 300
02460  C-----TWO-PHASE GAS-LIQUID HEAT TRANSFER COEFFICIENT (BTU/SQFT-HR-F)
02470  C-----FOR HOMOGENEOUS FLOW MODEL
02480      HI       = (AKLIQ/DI)*0.033
02490      H2       = ((DI*UIP*UIP)/(VTP*2.419092))**0.87
02500      PRL      = ((ASPHL*ALVIS*2.419092)/AKLIQ)**0.4
02510      HI       = H1*H2*PRL
02520      GO TO 320
02530  C-----FOR SEPARATED FLOW MODEL
02540  C-----TWO-PHASE GAS-LIQUID HEAT TRANSFER COEFFICIENT (BTU/SQFT-HR-F)
02550  300      HI       = .06*(AKLIQ/DI)*((ALD/AGJ)**.28)
02560      H2       = ((DI*GASFRA*UIP*UIP)/(ALVIS*2.419092))**0.87
02570      PRL      = ((ASPHL*ALVIS*2.419092)/AKLIQ)**0.4
02580      HI       = H1*H2*PRL
02590  C-----TOTAL HEAT FLUX AREA OF THE CELL (SQFT)
02600  320      AHEAT  = DI*DI*3.1415927
02610      IF(TLOOP.EQ.1) GO TO 330
02620  C-----LOG MEAN TEMPERATURE DIFFERENCE (F)
02630      TLM1     = T(II,TLOOP)-TSTART
02640      IF(TLM1.EQ.0.0) GO TO 335
02650      TLM2     = ALCG((TG-TSTART)/(TG-T(II,TLOOP)))
02660      TLM      = TLM1/TLM2
02670      GO TO 340
02680  C-----LOG MEAN DIAMETER OF THE TUBE (FT)
02690  330      TLM    = TG-TSTART
02700      GO TO 340
02710  335      TLM    = 0.0
02720  340      DLM    = (DC-DI)/(ALOG(DC/DI))
02730  C-----TUBE WALL THICKNESS (FT)
02740      XW       = (DC-DI)/2.0
02750  C-----OVERALL HEAT TRANSFER COEFFICIENT BASED ON TUBE INSIDE DIAMETER
02760  C----- (BTU/SQFT-HR-F)
02770      UI       = 1./((1./HI)+(DI/(DC*XW))+((DI*XW)/(DLM*XW)))
02780  C-----TOTAL HEAT REQUIRED TO RAISE THE STREAM TEMPERATURE TO THE
02790  C-----OUTLET STREAM TEMPERATURE.
02800      HEATFL    = UI*TLM
02810      HEAT       = HEATFL*AHEAT
02820  C-----ADJUST ESTIMATED OUTLET STREAM TEMPERATURE BY HEAT BALANCES OF
02830  C-----THE SYSTEM
02840      IJ        = II-1
02850      T2N       = TSTART
02860  350      FF1 = HEAT-LHEAT
02870      FF2 = SPHG(II)*GFRATE(II)*T2N-SPHG(II-1)*GFRATE(II-1)*TSTART
02880      FF3 = 0.5*(SPHL(II)+SPHL(II-1))*(T2N-TSTART)
02890      FF4 = 0.5*(LFRATE(II)+LFRATE(II-1))*LFRATE(II-1)-LFRATE(II)
02900      FF5 = FF3*FF4
02910      DF1 = -SPHG(II)*GFRATE(II)
02920      DF2 = -0.5*(SPHL(II)+SPHL(II-1))*FF4
02930      T2NO  = T2N
02940      T2N    = T2N-((FF1-FF2-FF5)/(CF1+DF2))
02950      IF(ABS((T2N-T2NO)/T2NO).LE.0.0001) GO TO 390
02960      GO TO 350
02970  C-----TEMPERATURE CONVERGENCE CRITERION
02980  390      CCFT    = ABS(T2N-T(II,TLOOP))/T(II,TLOOP)
02990      IF(CCFT.LE.0.003) GO TO 393
03000      TLOOP     = TLOOP + 1

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03010      IF(TLOOP.GE.50) GO TO 600
03020      T(II,TLOOP)= (T(II,TLOOP-1)+T2N)/2.0
03030      GC TO 38
03040      C-----PRINT OUT THE RESULTS
03050      C-----INLET STREAM TEMPERATURE (F) AND PRESSURE (PSIA) OF THE NEXT
03060      C-----CELL
03070      C-----DETERMINE TEMPERATURE INCREASE THROUGH THE CELL
03080      393      T(II,TLOOP)= T2N
03090      DTCELL = T(II,TLOOP)-TSTART
03100      C-----TUBE WALL TEMPERATURE
03110      TFA      = 0.5*(T(II,TLOOP)+TSTART)
03120      TW        = TFA+(TG-TFA)*(((1./HI)/((1./HI)+((DI/DO)*(1./HO))))
03130      394      TSTART = T(II,TLOOP)
03140      PSTART = P(II)
03150      IF(II.EQ.1) GC TO 396
03160      WRITE(23,395) II,TSTART,II,PSTART
03170      395      FORMAT(2X,'TEMPERATURE AT Z(',I2,') = ',F7.2,' (F), AND
03180      # PRESSURE AT Z(',I2,') = ',F6.2,' (PSIA)')
03190      396      WRITE(23,401) II
03200      401      FORMAT(2X,'SYSTEM AT Z(',I2,') IS ASSUMED TO BE IN EQUILIBRIUM
03210      # STATE')
03220      DO 403 I5=1,NC
03230      WRITE(23,402) I5,YP(I5),XP(I5)
03240      402      FORMAT(2X,'MASS FRACTION OF COMPONENT ',I1,' IN THE GAS
03250      # PHASE = ',F6.5,' IN THE LIQUID PHASE = ',F6.5)
03260      403      CCNTINUE
03270      DO 406 I6=1,NC
03280      WRITE(23,405) I6,DVAP(I6),DLIQ(I6)
03290      405      FORMAT(2X,'DENSITY OF COMPONENT ',I1,' IN THE GAS PHASE = ',
03300      # F8.5,' IN THE LIQUID PHASE = ',F8.5,' (LB/CUFT)')
03310      406      CCNTINUE
03320      DO 409 I7=1,NC
03330      WRITE(23,408) I7,TCL(I7)
03340      408      FORMAT(2X,'THERMAL CONDUCTIVITY OF COMPONENT ',I1,' IN THE
03350      # LIQUID PHASE = ',F9.4,' (BTU/FT-H-F)')
03360      409      CCNTINUE
03370      DO 411 I8=1,NC
03380      WRITE(23,410) I8,VG(I8),VL(I8)
03390      410      FORMAT(2X,'VISCOSITY OF COMPONENT ',I2,' IN THE GAS PHASE = ',
03400      # F7.4,' AND IN THE LIQUID PHASE = ',F7.4,' (CP)')
03410      411      CCNTINUE
03420      DO 414 I9=1,NC
03430      WRITE(23,413) I9,CPG(I9),CPL(I9)
03440      413      FORMAT(2X,'SPECIFIC HEAT OF COMPONENT ',I1,' IN THE GAS
03450      # PHASE = ',F6.3,' IN THE LIQUID PHASE = ',F6.3,' (BTU/LB-F)')
03460      414      CCNTINUE
03470      WRITE(23,415) GASDEN(II),LIQDEN(II)
03480      415      FORMAT(2X,'AVERAGE DENSITY OF GAS = ',F8.5,' LIQUID = ',
03490      # F6.3,' (LB/CUFT)')
03500      WRITE(23,417) GASVIS(II),LIQVIS(II)
03510      417      FORMAT(2X,'AVERAGE VISCOSITY OF GAS = ',F8.5,' LIQUID = ',
03520      # F8.5,' (CP)')
03530      WRITE(23,419) KLIQ(II)
03540      419      FORMAT(2X,'AVERAGE THERMAL CONDUCTIVITY OF LIQUID = ',F8.5,
03550      # ' (BTU/FT-H-F)')
03560      WRITE(23,421) SPHG(II),SPHL(II)
03570      421      FORMAT(2X,'AVERAGE SPECIFIC HEAT OF GAS = ',F6.3,' AND OF
03580      # LIQUID = ',F6.3,' (BTU/LB-F)')
03590      WRITE(23,423) REGAS(II),RELIQ(II)
03600      423      FORMAT(2X,'AVERAGE REYNOLDS NUMBER OF GAS = ',F10.2,' AND

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03610      # LIQUID = ',F10.2)
03620      IF(II.EC.1) GO TO 495
03630      WRITE(23,424) IJ,IJ,II
03640 424      FORMAT(//,2X,'CELL #',I3,' FROM Z(',I2,') TO Z(',I2,')')
03650      WRITE(23,425) AGD,ALD,IJ
03660 425      FORMAT(2X,'AVERAGE DENSITY OF GAS = ',F7.5,' AND LIQUID = ',
03670      # F7.3,' CF CELL # ',I2,' (LB/CUFT)')
03680      WRITE(23,427) AGVIS,ALVIS,IJ
03690 427      FORMAT(2X,'AVERAGE VISCOSITY OF GAS = ',F8.5,' AND LIQUID =
03700      # ',F8.5,' CF CELL # ',I2,' (CP)')
03710      WRITE(23,429) IJ,AKLIQ
03720 429      FORMAT(2X,'AVERAGE THERMAL CONDUCTIVITY OF CELL # ',I2,' = ',
03730      # F7.5,' (BTU/FT-H-F)')
03740      WRITE(23,431) ASPHG, ASPHL,IJ
03750 431      FORMAT(2X,'AVERAGE SPECIFIC HEAT OF GAS = ',F6.3,' AND
03760      # LIQUID = ',F6.3,' OF CELL # ',I2,' (BTU/LB-F)')
03770      WRITE(23,432) AGFR,ALFR,IJ
03780 432      FORMAT(2X,'AVERAGE GAS MASS FLOW RATE = ',F9.2,5X,' AND LIQUID
03790      # MASS FLOW RATE = ',F9.2,' OF CELL # ',I2,' (LB/HR)')
03800      WRITE(23,433) AREGAS,ARELIQ,IJ
03810 433      FORMAT(2X,'AVERAGE REYNOLD NUMBER OF GAS = ',F10.2,' AND LIQUID
03820      # = ',F10.2,' OF CELL # ',I2)
03830      WRITE(23,434) IJ,GASFRA
03840 434      FORMAT(2X,'GAS MASS FRACTION IN CELL # ',I2,' = ',F7.4)
03850      IF(II.EC.1) GO TO 495
03860      WRITE(23,435) LPDRCP
03870 435      FORMAT(2X,'SINGLE LIQUID PHASE PRESSURE DROP = ',F8.6,
03880      # ' (PSI/FT)')
03890      WRITE(23,437) IJ,P(II-1),P(II)
03900 437      FORMAT(2X,'PRESSURE OF THE STREAM ENTERS CELL #',I3,' = ',F6.2,
03910      # ' AND EXITS = ',F6.2,' (PSIA)')
03920      WRITE(23,439) IJ,TPDRCP
03930 439      FORMAT(2X,'TWO-PHASE PRESSURE DROP IN CELL # ',I2,' = ',F8.6,
03940      # ' (PSI/CELL LENGTH)')
03950      CCFP = CCFT * 100.0
03960      WRITE(23,441) IJ,CCFP,PLLOOP
03970 441      FORMAT(2X,'ERROR OF PRESSURE DROP IN CELL # ',I2,' = ',F6.2,
03980      # ' %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = ',I3)
03990      WRITE(23,442) RLC
04000 442      FORMAT(2X,'LIQUID HOLDUP FRACTION = ',F9.4)
04010      TOLD = T(II,TLCOP)-DTCELL
04020      WRITE(23,443) IJ,TOLD,T(II,TLCOP)
04030 443      FORMAT(2X,'TEMPERATURE OF THE STREAM ENTERING CELL # ',I2,' = ',
04040      # F6.2,' AND LEAVING ',F6.2,' (F)')
04050      WRITE(23,445) IJ,DTCELL
04060 445      FORMAT(2X,'TEMPERATURE INCREASED IN CELL # ',I2,' = ',F6.2,
04070      # ' (F)')
04080      CCFT= CCFT*100.0
04090      WRITE(23,447) IJ,CCFT,TLOOP
04100 447      FORMAT(2X,'ERROR OF TEMPERATURE ITERATING IN CELL # ',I2,' = ',
04110      # F6.2,' %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = ',I3)
04120      WRITE(23,449) IJ,EXCHM
04130 449      FORMAT(2X,'MASS TRANSFERRED BETWEEN PHASES IN CELL # ',I2,
04140      # ' = ',F5.3,' (LB/HR)')
04150      IF(MECHAN-2) 450,452,454
04160 450      WRITE(23,451)
04170 451      FORMAT(2X,'VISCOUS-VISCOUS FLOW MECHANISM')
04180      GO TO 460
04190 452      WRITE(23,453)
04200 453      FORMAT(2X,'TURBULENT-VISCOUS FLOW MECHANISM')

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04210      GO TO 460
04220 454      IF(MECHAN.EQ.4) GO TO 458
04230      WRITE(23,455)
04240 455      FORMAT(2X,'VISCOUS-TURBULENT FLOW MECHANISM')
04250      GO TO 460
04260 458      WRITE(23,459)
04270 459      FORMAT(2X,'TURBULENT-TURBULENT FLOW MECHANISM')
04280 460      IF(TYPE.EQ.2) GO TO 462
04290      WRITE(23,461)
04300 461      FORMAT(2X,'HOMOGENEOUS FLOW MODEL')
04310      GO TO 470
04320 462      WRITE(23,463)
04330 463      FORMAT(2X,'SEPERATED FLOW MODEL')
04340 470      WRITE(23,472) UG,UL,UTP
04350 472      FORMAT(2X,'VELOCITY OF THE GAS PHASE = ',F9.2,' OF THE LIQUID
# PHASE = ',F9.2,' AND OF THE TWO-PHASE FLOW = ',F9.2,' (FT/HR)')
04370      WRITE(23,484) DIP,VTP
04380 484      FORMAT(2X,'TWO-PHASE FLUID DENSITY = ',F7.3,' (LB/CU. FT),
# TWO-PHASE FLUID VISCOSITY = ',F7.3,' (CP)')
04390      WRITE(23,485) TW,AHEAT
04410 485      FORMAT(2X,'TUBE WALL TEMPERATURE = ',F8.2,' (F)   HEAT FLUX
# AREA OF THE CELL = ',F7.2,' (SQ. FT)')
04420      WRITE(23,486) TLM,DLM
04440 486      FORMAT(2X,'LOG MEAN TEMPERATURE DIFFERENCE = ',F7.2,' (F)   LOG
# MEAN DIAMETER = ',F7.2,' (FT)')
04460      WRITE(23,487) HI,UI
04470 487      FORMAT(2X,'TWO-PHASE HEAT TRANSFER COEFFICIENT = ',F7.2,'
# UI BASED ON TUBE INSIDE DIAMETER = ',F7.2,' (BTU/SQFT-HR-F)')
04480      WRITE(23,488) HEATFL
04490      FORMAT(2X,'CALCULATED HEAT FLUX TO THE CELL = ',F10.2,
# ' (BTU/SQFT-HR)')
04500      WRITE(23,448) IJ,HEAT
04530 448      FORMAT(2X,'CALCULATED HEAT REQUIRED IN CELL # ',I2,' = ',F9.2,
# ' (BTU/HR)')
04550      C-----TOTAL PRESSURE DROP ALONG THE TUBE (PSIA)
04560 495      CELTAP = CELTAP+TPDROP
04570      C-----TOTAL MASS TRANSFERRED BETWEEN PHASES (LB/HR)
04580      TEXCHM = TEXCHM + EXCHM
04590      C-----TOTAL HEAT ADDED TO THE ENTIRE SYSTEM (BTU/HR)
04600      SUMQ = SUMQ+HEAT
04610      T(II+1,1)= T(II,TLOGP)
04620      P(II+1) = P(II)
04630      TZ(II) = T(II,TLOGP)
04640      II = II+1
04650      STW = STW+TW
04660      SRLC = SRLC+RLC
04670      IF(II.LE.N) GO TO 30
04680      GO TO 670
04690      C-----PRINT OUT THE ERROR IN THE CALCULATION
04700 500      WRITE(23,550)
04710 550      FORMAT(2X,'NO CONVERGENCE FOR THE PRESSURE LOOP')
04720      GO TO 700
04730 600      WRITE(23,650)
04740 650      FORMAT(2X,'NO CONVERGENCE FOR THE TEMPERATURE LOOP')
04750      GO TO 700
04760      C-----TOTAL TUBE LENGTH (FT)
04770 670      TUBEL = DZ*(N-1)
04780      WRITE(23,671) TUBEL
04790 671      FORMAT(2X,'TOTAL TUBE LENGTH = ',F7.2,' (FT)')
04800      WRITE(23,674) SUMQ

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04810 674   FORMAT(2X,'TOTAL HEAT ADDED TO THE ENTIRE SYSTEM = ',F10.2,
04820      #   ' (BTU/HR)')
04830      DELTAT = TSTART-T(1,1)
04840      WRITE(23,675) DELTAP
04850 675   FORMAT(2X,'TOTAL PRESSURE DROP ALONG THE TUBE = ',F8.3,
04860      #   ' (PSI/TUBE LENGTH) ')
04870      WRITE(23,678) DELTAT
04880 678   FORMAT(2X,'TOTAL TEMPERATURE INCREASE = ',F8.3,' (F)')
04890      WRITE(23,690) TEXCHM
04900 690   FORMAT(2X,'TOTAL MASS TRANSFERRED BETWEEN PHASES IN THE ENTIRE
04910      #   SYSTEM = ',F7.2,' (LB/HR)')
04920      WRITE(23,680)
04930 680   FORMAT(2X,'COMPLETE CALCULATION')
04940      WRITE(23,695)
04950 695   FORMAT(2X,'TEMPERATURE      PRESSURE      GAS MASS FLOW RATE
04960      #   LIQUID MASS FLOW RATE')
04970 700   DO 800 J2=1,N
04980      WRITE(23,720) TZ(J2),P(J2),GFRATE(J2),LFRATE(J2)
04990 720   FORMAT(2X,F9.2,5X,F9.2,11X,F9.2,11X,F9.2)
05000 800   CONTINUE
05010      STW = STW/(N-1)
05020      SRLC = SRLC/(N-1)
05030      WRITE(23,820) STW
05040 820   FORMAT(//,2X,'AVERAGE TUBE WALL TEMPERATURE = ',F8.2,' (F)')
05050      WRITE(23,850) SRLC
05060 850   FORMAT(2X,'AVERAGE LIQUID HOLDUP FRACTION = ',F10.4)
05070      STOP
05080 C-----END OF THE MAIN PROGRAM
05090      END
05100      *
05110      *
05120 *****SUBPROGRAM RATES*****
05130      *
05140      *           IN THIS SUBPROGRAM FLOW RATES OF GAS AND LIQUID PHASES *
05150      *           AT THE ENDS OF THE CELL ARE DETERMINED BY FLASH CALCULATIONS. *
05160      *           THE APPLICATION OF THIS CALCULATION REQUIRES THE DATA OF *
05170      *           K-VALUES OF THE COMPONENTS AS INDEPENDENT VARIABLES OF THE *
05180      *           COMPOSITIONS. *
05190      *
05200 *****
05210      *
05220      *
05230      SUBROUTINE RATES(NC,II,TEMP,PRESS,LFR,GFR,XP,YP)
05240      *
05250      *
05260      REAL MW,K,L,LF,LMF,LFR
05270      DIMENSION YY(4),XX(4),Y(4,50),X(4,50),Z(4),XP(4),YP(4),K(4),
05280      #   MW(4),L(50),G(50),FC(4),PHI(50)
05290 C-----MOL. WT. OF THE COMPONENTS (LB/LBMOLE)
05300      DATA MW/2.,28.,100.,142./
05310 C-----MASS FLOW RATES OF THE FEED (LB/HR)
05320      DATA GF,LF/200.,2400./
05330 C-----MASS FRACTIONS IN THE LIQUID FEED
05340      DATA XX/0.,0.,.6,.4/
05350 C-----MASS FRACTIONS IN THE GAS FEED
05360      DATA YY/.7,.3,0.,0./
05370 C-----PRINT OUT THE COMPOSITIONS IN THE FEED
05380      IF(II.GE.2) GO TO 9
05390      DO 8 IO = 1,NC
05400      WRITE(23,5) IO,YY(IO),XX(IO)

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05410 5      FORMAT(2X,'MASS FRACTION OF COMPONENT ',I1,' IN THE GAS FEED
05420      #      = ',F4.3,' IN THE LIQUID FEED = ',F4.3)
05430 8      CCNTINUE
05440 C-----FIND OVERALL MOLE FRACTIONS IN THE FEED
05450 9      FEED = 0.0
05460      DO 10 I1 = 1,NC
05470          FC(I1) = (YY(I1)*GF+XX(I1)*LF)/MW(I1)
05480          FEED = FEED+FC(I1)
05490 10     CCNTINUE
05500      DO 20 I2 = 1,NC
05510          Z(I2) = FC(I2)/FEED
05520          IF(I1.GE.2) GO TO 20
05530          WRITE(23,15) I2,Z(I2)
05540 15     FORMAT(2X,'OVERALL MOLE FRACTION OF COMPONENTS ',I1,' IN THE
05550      #      FEED = ',F6.5)
05560 20     CCNTINUE
05570 C-----USE SUBPROGRAM KVALUE TO FIND K-VALUES OF THE COMPONENTS
05580      CALL KVALUE(NC,TEMP,PRESS,K)
05590 C-----
05600 C-----START FLASH CALCULATION
05610      J = 1
05620      PHI(J) = 0.5
05630 30     SUMF = 0.0
05640      DSUMF = 0.0
05650      DO 40 I4 = 1,NC
05660          FPHI = Z(I4)*(1.-K(I4))/(1.+PHI(J)*(K(I4)-1.))
05670          DFPHI = Z(I4)*((1.-K(I4))**2.)/(1.+PHI(J)*(K(I4)-1.))**2.
05680          SUMF = SUMF+FPHI
05690          DSUMF = DSUMF+DFPHI
05700 40     CONTINUE
05710 C-----ERROR FOR THE CONVERGENCE OF THE FLASH CALCULATIONS
05720      IF(ABS(SUMF).LE.0.0001) GO TO 50
05730      J = J+1
05740      PHI(J) = PHI(J-1)-(SUMF/DSUMF)
05750      ERRPHI = ABS(PHI(J)-PHI(J-1))/PHI(J)
05760      IF(ERRPHI.LE.0.0001) GO TO 50
05770      GO TO 30
05780 50     DO 60 I5 = 1,NC
05790 C-----MOLE FRACTIONS IN THE LIQUID PHASE AT THE END OF THE CELL
05800          X(I5,I1)=Z(I5)/(1.+PHI(J)*(K(I5)-1.))
05810 C-----MOLE FRACTIONS IN THE GAS PHASE AT THE END OF THE CELL
05820          Y(I5,I1)=K(I5)*X(I5,I1)
05830 60     CCNTINUE
05840 C-----GAS MOLAR FLOW RATE
05850      G(I1) = PHI(J)*FEED
05860 C-----LIQUID MOLAR FLOW RATE
05870      L(I1) = FEED-G(I1)
05880      GFR = 0.0
05890      LFR = 0.0
05900      DO 70 I6 = 1,NC
05910 C-----GAS MASS FLOW RATE AT THE END OF THE CELL (LB/HR)
05920          GFR = GFR+Y(I6,I1)*G(I1)*MW(I6)
05930 C-----LIQUID MASS FLOW RATE AT THE END OF THE CELL (LB/HR)
05940          LFR = LFR+X(I6,I1)*L(I1)*MW(I6)
05950 70     CONTINUE
05960      DO 80 I7 = 1,NC
05970 C-----MASS FRACTIONS IN THE GAS PHASE AT THE END OF THE CELL
05980          YP(I7) = Y(I7,I1)*G(I1)*MW(I7)/GFR
05990 C-----MASS FRACTIONS IN THE LIQUID PHASE AT THE END OF THE CELL
06000          XP(I7) = X(I7,I1)*L(I1)*MW(I7)/LFR

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06010      80          CONTINUE
06020      C-----RETURN TO THE MAIN PROGRAM
06030          RETURN
06040          END
06050      *
06060      *
06070      *****SUBPROGRAM KVALUE*****
06080      *
06090      *          THIS IS A DATA SUBPROGRAM. INTERPOLATION IS APPLIED TO*
06100      *          DETERMINE K-VALUES OF THE COMPONENTS AT APPROXIMATE TEMPERATURE*
06110      *          AND PRESSURE
06120      *
06130      *****
06140      *
06150      *
06160          SUBROUTINE KVALUE(NC,TEMP,PRESS,K)
06170      *
06180      *
06190          REAL KV(4,7,12),RP(7),RT(12),PK(4,12),K(4)
06200      C-----DATA SUPPORTED FOR THE CALCULATION OF K-VALUES
06210      C-----PRESSURE RANGE OF THE DATA OF K-VALUES
06220          DATA RP/14.,20.,25.,30.,40.,60.,100./
06230      C-----TEMPERATURE RANGE OF THE DATA OF K-VALUES
06240          DATA RT/80.,100.,120.,140.,160.,180.,200.,220.,240.,260.,280.,
06250      #      300./
06260      C-----DATA OF K-VALUES OF THE COMPONENTS
06270          DATA KV/566.,246.,.08.,.0018,482.,214.,.058.,.00145,376.,171.,.048,
06280      #      .0013,346.,161.,.045.,.0012,254.,121.,.035.,.0011,193.,94.,.025,
06290      #      .001,1296.,64.3.,.018.,.0009,920.,400.,.13.,.0037,788.,350.,.095,
06300      #      .0028,660.,300.,.084.,.0025,559.,250.,.07.,.0022,546.,200.,.058,
06310      #      .0018,308.,150.,.04.,.0015,200.,100.,.028.,.0012,1486.,646.,.21,
06320      #      .0072,1244.,553.,.15.,.0052,1021.,464.,.13.,.0048,845.,393.,.11,
06330      #      .0042,615.,293.,.085.,.0034,478.,233.,.062.,.0027,314.,157.,.044,
06340      #      .002,2246.,958.,.32.,.013,2156.,958.,.26.,.01,1415.,643.,.18,
06350      #      .0084,1228.,571.,.16.,.0075,869.,414.,.12.,.0058,677.,330.,.088,
06360      #      .0046,458.,229.,.064.,.0034,3255.,1415.,.46.,.024,2736.,1216.,.33,
06370      #      .0156,2200.,1000.,.28.,.015,1765.,821.,.23.,.013,1380.,655.,.19,
06380      #      .011,923.,450.,.12.,.0074,614.,307.,.086.,.0055,4529.,1969.,.64,
06390      #      .038,3647.,1621.,.44.,.027,3065.,1393.,.39.,.024,2458.,1143.,.32,
06400      #      .02,1810.,862.,.25.,.017,1308.,638.,.17.,.012,858.,429.,.12,
06410      #      .009,6369.,2769.,.9.,.06,4975.,2211.,.6.,.042,4085.,1857.,.52,
06420      #      .038,3225.,1500.,.42.,.032,2316.,1103.,.32.,.026,1769.,863.,.23,
06430      #      .018,1142.,571.,.16.,.014,8492.,3692.,.12.,.1,6881.,3058.,.83.,.07,
06440      #      5500.,2500.,.7.,.06,4377.,2036.,.57.,.05,3259.,1551.,.45.,.04,
06450      #      2306.,1125.,.3.,.028,1500.,750.,.21.,.02,11323.,4923.,1.6.,.155,
06460      #      9120.,4053.,1.1.,.11,7230.,3286.,.92.,.09,5760.,2679.,.75.,.074,
06470      #      4345.,2069.,.6.,.062,3075.,1500.,.4.,.04,1875.,929.,.26.,.03,14863,
06480      #      6462.,2.1.,.22,12020.,5342.,1.45.,.153,9429.,4286.,1.2.,.13,7372.,
06490      #      3429.,.56.,.11,5069.,2414.,.7.,.08,3690.,1800.,.48.,.06,2642.,
06500      #      1321.,.37.,.045,16808.,8308.,2.7.,.33,16578.,7368.,2.,.24,11785.,
06510      #      5357.,1.5.,.19,9215.,4286.,1.2.,.16,6590.,3138.,.91.,.13,4766.,
06520      #      2325.,.62.,.09,3286.,1643.,.46.,.065,24771.,10770.,3.5.,.46,
06530      #      19895,8642.,2.4.,.34,15715.,7143.,2.,.28,12285.,5714.,1.6.,.24,
06540      #      8545.,4069.,1.18.,.18,5920.,2888.,.77.,.12,3786.,1893.,.53.,.08/
06550      C-----INTERPOLATION TO FIND K-VALUES
06560          DO 200 J1= 1,7
06570              IF(RP(J1)-PRESS) 200,200,10
06580      10          PTIE = (RP(J1)-RP(J1-1))/(RP(J1)-PRESS)
06590              DO 100 J2= 1,12
06600                  DO 20 J3= 1,NC

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C661C      PK(J3,J2) = KV(J3,J1,J2)-(KV(J3,J1,J2)-KV(J3,J1-1,J2))/PTIE
C662C      2C      CONTINUE
C663C      100      CONTINUE
C664C      GO TO 300
C665C      200      CONTINUE
C666C      300      DO 400 J4=1,12
C667C      IF(RT(J4)-TEMP) 400,400,310
C668C      310      TTIE = (RT(J4)-RT(J4-1))/(RT(J4)-TEMP)
C669C      DO 350 J5=1,NC
C670C      C-----K-VALUES OF COMPONENTS AT APPROXIMATE TEMPERATURE AND PRESSURE
C671C      K(J5) = PK(J5,J4)-(PK(J5,J4)-PK(J5,J4-1))/TTIE
C672C      350      CONTINUE
C673C      GO TO 450
C674C      400      CONTINUE
C675C      C-----RETURN TO SUBPROGRAM RATES
C676C      450      RETURN
C677C      END
C678C      *
C679C      *
C680C      *****SUBPROGRAM PHYSIC*****
C681C      *
C682C      *      THIS SUBPROGRAM IS USED TO FIND THE PHYSICAL CONSTANTS *
C683C      *      OF THE COMPONENTS AT APPROXIMATE TEMPERATURE AND PRESSURE *
C684C      *
C685C      *****
C686C      *
C687C      *
C688C      SUBROUTINE PHYSIC(NC,TEMP,PRESS,VG,VL,CPG,CPL,DVAP,DLIQ,TCL)
C689C      *
C690C      *
C691C      DIMENSION TCL(4),VG(4),VL(4),CPL(4),CPG(4),DVAP(4),DLIQ(4),
C692C      # DV(4,3,5),DL(4,3,5),PP(3),TT(5),PDV(4,5),PDL(4,5)
C693C      C-----DATA SUPPLIED
C694C      C-----PRESSURE RANGE OF THE DATA
C695C      DATA PP/14.7,30.0,100.0/
C696C      C-----TEMPERATURE RANGE OF THE DATA
C697C      DATA TT/80.,150.,200.,250.,300./
C698C      C-----DATA OF VAPOR DENSITIES OF THE COMPONENTS
C699C      DATA DV/.000348,.071134,.2186,.2405,.01008,.105,.492,.541,
C700C      # .03466,.4844,1.107,1.2177,.003081,.06295,.2024,.2226,
C701C      # .02307,.12843,.4554,.50001,.030694,.42802,1.025,1.13,
C702C      # .004182,.05818,.1862,.2050,.0079,.084,.41895,.4609,.02838,
C703C      # .3858,.9426,1.037,.00389,.05435,.17,.187,.00735,.0935,.3825,
C704C      # .4208,.02638,.3645,.8606,.9467,.003632,.05051,.16,.176,.00686,
C705C      # .103,.36,.4,.024649,.3431,1.3,1.43/
C706C      C-----DATA OF LIQUID DENSITIES OF THE COMPONENTS
C707C      DATA DL/3.91,15.61,42.32,45.31,3.99,15.95,42.33,45.49,4.38,
C708C      # 17.5,42.35,45.53,2.08,8.32,40.2,43.46,2.12,8.5,40.21,43.5,
C709C      # 2.33,9.33,40.23,43.68,1.56,6.24,38.7,42.1,1.6,6.38,38.7,
C710C      # 42.17,1.78,7.02,38.72,42.32,1.25,5.36,96,40.68,1.28,5.11,
C711C      # 37.06,40.72,1.41,5.62,37.02,40.9,1.04,4.16,34.96,39.25,
C712C      # 1.06,4.25,34.58,39.29,1.17,4.68,35.05,39.47/
C713C      C-----INTERPOLATION TO FIND DENSITIES OF THE COMPONENTS
C714C      DO 100 J1 = 1,3
C715C      IF(PP(J1)-PRESS) 100,100,10
C716C      10      PTIE = (PP(J1)-PP(J1-1))/(PP(J1)-PRESS)
C717C      DO 50 J2 = 1,5
C718C      DO 40 J3=1,NC
C719C      PJV(J3,J2)=DV(J3,J1,J2)-(DV(J3,J1,J2)-DV(J3,J1-1,J2))/PTIE
C720C      PDL(J3,J2)=DL(J3,J1,J2)-(DL(J3,J1,J2)-DL(J3,J1-1,J2))/PTIE

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07210 40      CONTINUE
07220 50      CCNTINLE
07230      GO TO 200
07240 100     LCNTINUE
07250 200     CC 400 J4=1,5
07260      IF (TT(J4)-TEMP) 400,400,310
07270 310     TTIE = (TT(J4)-TT(J4-1))/(TT(J4)-TEMP)
07280      DO 350 J5=1,NC
07290      DVAP(J5)=PDV(J5,J4)-(PDV(J5,J4)-PDV(J5,J4-1))/TTIE
07300      DLIQ(J5)=PDL(J5,J4)-(PDL(J5,J4)-PDL(J5,J4-1))/TTIE
07310 350     CCNTINLE
07320      GO TO 500
07330 400     CCNTINUE
07340 C-----EQUATIONS TO ESTIMATE THERMAL CONDUCTIVITIES OF THE COMPONENTS
07350 C-----IN THE LIQUID PHASE (BTU/FT-HR-F)
07360 500     TCL(1) = 3167.2+(TEMP-80.6)*4.0383
07370      TCL(2) = 448.30+(TEMP-80.6)*.65382
07380      TCL(3) = 1.3900+(TEMP-80.6)*(-0.0007407407)
07390      TCL(4) = 1.4500+(TEMP-80.6)*(-0.0005555556)
07400 C-----EQUATIONS TO ESTIMATE VISCOSITIES OF THE COMPONENTS IN THE GAS
07410 C-----PHASE (CP)
07420      VG(1) = 0.00850+(TEMP-60.0)*0.000104167
07430      VG(2) = 0.0170+(TEMP-60.0)*0.000025
07440      VG(3) = 0.0600+(TEMP-60.0)*0.001042
07450      VG(4) = 0.0500+(TEMP-60.0)*0.00008333
07460 C-----EQUATIONS TO ESTIMATE VISCOSITY OF THE COMPONENTS IN THE LIQUID
07470 C-----PHASE (CP)
07480      VL(1) = 0.0010+(TEMP-60.0)*0.0000004167
07490      VL(2) = 0.0020+(TEMP-60.0)*0.001042
07500      TK = ((TEMP-32.0)/1.8)+273.0
07510      VL(3) = 10.0*(436.73*((1./TK)-(1./232.53)))
07520      VL(4) = 10.0*(558.61*((1./TK)-(1./277.8)))
07530 C-----EQUATIONS TO ESTIMATED SPECIFIC HEAT OF THE COMPONENTS IN THE
07540 C-----GAS PHASE (BTU/LB-F)
07550      CPG(1) = 3.5
07560      CPG(2) = .25+(TEMP-50.0)*0.00002
07570      CPG(3) = .2475+(TEMP-50.0)*0.00039
07580      CPG(4) = .21+(TEMP-50.0)*0.00014
07590 C-----EQUATIONS TO ESTIMATED SPECIFIC HEAT OF THE COMPONENTS IN THE
07600 C-----LIQUID PHASE (BTU/LB-F)
07610      CPL(1) = 7.0
07620      CPL(2) = 0.5+(TEMP-50.0)*0.0004
07630      CPL(3) = 0.375+(TEMP-50.0)*0.00058
07640      CPL(4) = 0.40+(TEMP-50.0)*0.00058
07650 C-----RETURN TO THE MAIN PROGRAM
07660      RETURN
07670      END
07680 *
07690 *
07700 *****SUBPROGRAM LATENT*****
07710 *
07720 *      THIS SUBPROGRAM IS USED TO FIND THE HEAT OF EVAPORATION*
07730 *      OF THE COMPONENTS AT APPROXIMATE TEMPERATURE AND PRESSURE *
07740 *
07750 *****
07760 *
07770 *
07780      SUBROUTINE LATENT(NC,TEMP,LAMDA)
07790 *
07800 *

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07810      REAL LAMDA(4)
07820      C-----EQUATIONS TO ESTIMATED LATENT HEAT OF THE COMPONENTS
07830      LAMDA(1)=1800.+(TEMP-80.)*3.5
07840      LAMDA(2)=237.+(TEMP-80.)*.264
07850      LAMDA(3)=162.+(TEMP-80.)*.2364
07860      LAMDA(4)=162.+(TEMP-80.)*.1955
07870      C-----RETURN TO THE MAIN PROGRAM
07880      RETURN
07890      END
07900      *
07910      *
07920      *****SUBPROGRAM FMODEL*****
07930      *
07940      *          THIS SUBPROGRAM IS USED TO FIND THE FLOW MODEL BASED ON*
07950      *          THE FLOW-PATTERN MAP OF MANDHANE ET AL.  THERE ARE TWO FLOW *
07960      *          MODEL:
07970      *          HOMOGENEOUS MODEL WHICH INCLUDES SLUG, DISPERSED, AND
07980      *          BUBBLE FLOWS.
07990      *          SEPARATED MODEL WHICH INCLUDES STRATIFIED, WAVE, AND
08000      *          ANNULAR FLOWS.
08010      *
08020      *****
08030      *
08040      *
08050      SUBROUTINE FMODEL(VG,VL,TYPE)
08060      *
08070      *
08080      INTEGER TYPE
08090      C-----SUPERFICIAL GAS VELOCITY (M/SEC)
08100      VGG      = (VG*0.3048)/3600.0
08110      C-----SUPERFICIAL LIQUID VELOCITY (M/SEC)
08120      VLL      = (VL*0.3048)/3600.
08130      C-----SIMPLIFIED FLOW-PATTERN MAP OF MANDHANE ET AL
08140      IF(VGG.LE.1.8) GO TO 50
08150      IF(VGG.LE.2.3) GO TO 60
08160      IF(VGG.LE.12.) GO TO 70
08170      IF(VGG.LE.90.) GO TO 80
08180      50      IF(VLL.LE.0.5) GO TO 300
08190      GO TO 200
08200      60      HCRZ      = -0.3*VGG+1.04
08210      IF(VLL.LE.HCRZ) GO TO 300
08220      GO TO 200
08230      70      IF(VLL.LE..35) GO TO 300
08240      GO TO 200
08250      80      HCRZ      = 0.11*VGG+0.097
08260      IF(VLL.LE.HCRZ) GO TO 300
08270      200      TYPE      = 1
08280      GO TO 400
08290      300      TYPE      = 2
08300      C-----RETURN TO THE MAIN PROGRAM
08310      400      RETURN
08320      END

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(Results of the Multicomponent System)

SPECIFIED CONDITIONS AND INFORMATION
 NUMBER OF COMPONENTS IN THE SYSTEM = 4
 INSIDE TUBE DIAMETER = .6250 (FT), OUTSIDE TUBE DIAMETER = .6833 (FT), CELL LENGTH = 1.57 (FT)
 NUMBER OF CELLS IN THE TUBE = 10
 INPUT STREAM TEMPERATURE = 80.00 (F), INPUT STREAM PRESSURE = 60.00 (PSIA)
 AVERAGE TEMPERATURE OF THE HEATING MEDIUM = 230.00 (F)
 OUTSIDE TUBE HEAT TRANSFER COEFFICIENT = 2000.00 (BTU/SQFT-HR)
 TUBE WALL THERMAL CONDUCTIVITY = 64.00 (BTU/FT-H-F)

LOCATION ON THE TUBE : Z (1)
 TEMPERATURE AT Z (1) = 80.00 (F), AND PRESSURE AT Z (1) = 60.00 (PSIA)
 MASS FRACTION OF COMPONENT 1 IN THE GAS FEED = .700 IN THE LIQUID FEED = .000
 MASS FRACTION OF COMPONENT 2 IN THE GAS FEED = .300 IN THE LIQUID FEED = .000
 MASS FRACTION OF COMPONENT 3 IN THE GAS FEED = .000 IN THE LIQUID FEED = .600
 MASS FRACTION OF COMPONENT 4 IN THE GAS FEED = .000 IN THE LIQUID FEED = .400
 OVERALL MOLE FRACTION OF COMPONENTS 1 IN THE FEED = .75024
 OVERALL MOLE FRACTION OF COMPONENTS 2 IN THE FEED = .02297
 OVERALL MOLE FRACTION OF COMPONENTS 3 IN THE FEED = .15434
 OVERALL MOLE FRACTION OF COMPONENTS 4 IN THE FEED = .07246
 SYSTEM AT Z (1) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = .43102 IN THE LIQUID PHASE = .00009
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = .18479 IN THE LIQUID PHASE = .00008
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = .37259 IN THE LIQUID PHASE = .57563
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = .01080 IN THE LIQUID PHASE = .42021
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.22061 IN THE LIQUID PHASE = 4.15714 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.26763 IN THE LIQUID PHASE = 16.61429 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.75557 IN THE LIQUID PHASE = 42.31857 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.83101 IN THE LIQUID PHASE = 45.50714 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3164.7770 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 447.9077 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3904 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4503 (BTU/FT-H-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0106 AND IN THE LIQUID PHASE = 0.0010 (CPI)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0175 AND IN THE LIQUID PHASE = 0.0228 (CPI)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.0809 AND IN THE LIQUID PHASE = 0.3795 (CPI)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0517 AND IN THE LIQUID PHASE = 0.7133 (CPI)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.251 IN THE LIQUID PHASE = 0.512 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.259 IN THE LIQUID PHASE = 0.392 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.214 IN THE LIQUID PHASE = 0.417 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.34845 LIQUID = 43.665 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.03448 LIQUID = 0.51971 (CPI)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.72502 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.657 AND OF LIQUID = 3.403 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 4427.92 AND LIQUID = 2305.19

LOCATION ON THE TUBE : Z (2)
 TEMPERATURE AT Z (2) = 103.85 (F), AND PRESSURE AT Z (2) = 58.37 (PSIA)
 SYSTEM AT Z (2) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = .22936 IN THE LIQUID PHASE = .00005
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = .14104 IN THE LIQUID PHASE = .00004
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = .51417 IN THE LIQUID PHASE = .53159
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = .01543 IN THE LIQUID PHASE = .43932
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02212 IN THE LIQUID PHASE = 3.44881 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.25576 IN THE LIQUID PHASE = 13.95138 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.72253 IN THE LIQUID PHASE = 41.61876 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.75447 IN THE LIQUID PHASE = 44.35024 (LB/CUFT)

THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3200.0944 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 403.4372 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3729 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4371 (BTU/FT-HR-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0131 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0181 AND IN THE LIQUID PHASE = 0.0476 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1056 AND IN THE LIQUID PHASE = 0.3294 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0536 AND IN THE LIQUID PHASE = 0.5952 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.206 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.251 IN THE LIQUID PHASE = 0.522 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.268 IN THE LIQUID PHASE = 0.406 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.218 IN THE LIQUID PHASE = 0.431 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.42718 LIQUID = 41.032 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.00197 LIQUID = 0.44588 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.56587 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.330 AND OF LIQUID = 0.417 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 3607.48 AND LIQUID = 2567.64

CELL # 1 FROM Z(1) TO Z(2)
 AVERAGE DENSITY OF GAS = 0.36801 AND LIQUID = 43.34% OF CELL # 1 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.05023 AND LIQUID = 3.48275 OF CELL # 1 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 1 = 1.64744 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.493 AND LIQUID = 0.410 OF CELL # 1 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 374.26 AND LIQUID MASS FLOW RATE = 2225.74 OF CELL # 1 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 4017.70 AND LIQUID = 2436.41 OF CELL # 1
 GAS MASS FRACTION IN CELL # 1 = 0.1435
 SINGLE LIQUID PHASE PRESSURE DROP = 0.17640 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 1 = 60.00 AND EXITS = 58.37 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 1 = 1.629655 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 1 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID HOLDUP FRACTION = 0.1358
 TEMPERATURE OF THE STREAM ENTERING CELL # 1 = 80.00 AND LEAVING 103.85 (F)
 TEMPERATURE INCREASED IN CELL # 1 = 23.85 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 1 = 0.09 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 4
 MASS TRANSFERED BETWEEN PHASES IN CELL # 1 = 101.012 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 27052.13 OF THE LIQUID PHASE = 91609.87 AND OF THE TWO-PHASE FLOW = 101245.24 (FT/HR)
 TWO-PHASE FLUID DENSITY = 2.559 (LB/CU. FT), TWO-PHASE FLUID VISCOSITY = 0.216 (CP)
 TUBE WALL TEMPERATURE = 156.25 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 137.78 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 2635.54 BTU/(SQ.FT-HR) BASED ON TUBE INSIDE DIAMETER = 1119.44 (BTU/SQ.FT-HR-F)
 CALCULATED HEAT FLUX TO THE CELL = 154235.17 (BTU/SQ.FT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 1 = 55154.86 (BTU/HR)

LOCATION ON THE TUBE : Z(3)
 TEMPERATURE AT Z(3) = 122.28 (F), AND PRESSURE AT Z(3) = 56.47 (PSIA)
 SYSTEM AT Z(3) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = 0.25982 IN THE LIQUID PHASE = 0.0003
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = 0.11130 IN THE LIQUID PHASE = 0.0003
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = 0.60612 IN THE LIQUID PHASE = 0.54019
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = 0.22276 IN THE LIQUID PHASE = 0.45575
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02336 IN THE LIQUID PHASE = 2.96440 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.24442 IN THE LIQUID PHASE = 11.86197 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.69208 IN THE LIQUID PHASE = 41.05433 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.76146 IN THE LIQUID PHASE = 44.33264 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3335.8837 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 475.6167 (BTU/FT-HR-F)

THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3591 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4268 (BTU/FT-H-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0130 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0186 AND IN THE LIQUID PHASE = 0.0070 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1250 AND IN THE LIQUID PHASE = 0.2572 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0552 AND IN THE LIQUID PHASE = 0.5218 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.251 IN THE LIQUID PHASE = 0.329 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.276 IN THE LIQUID PHASE = 0.417 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.220 IN THE LIQUID PHASE = 0.442 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.47009 LIQUID = 42.560 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.08258 LIQUID = 0.40046 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.50455 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.109 AND OF LIQUID = 0.429 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 3416.23 AND LIQUID = 2707.28

CELL # 2 FROM Z(2) TO Z(3)
 AVERAGE DENSITY OF GAS = .44863 AND LIQUID = 42.756 OF CELL # 2 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.07247 AND LIQUID = 0.42317 (BTU/FT-H-F)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 2 = 1.53721 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.220 AND LIQUID = 0.423 OF CELL # 2 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 481.86 AND LIQUID MASS FLOW RATE = 2118.32 OF CELL # 2 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 3511.86 AND LIQUID = 2638.46 OF CELL # 2
 GAS MASS FRACTION IN CELL # 2 = 0.1553
 SINGLE LIQUID PHASE PRESSURE DROP = .015920 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 2 = 58.37 AND EXITS = 56.47 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 2 = 1.03734 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 2 = 3.02 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID HOLDUP FRACTION = 0.1193
 TEMPERATURE OF THE STREAM ENTERING CELL # 2 = 103.85 AND LEAVING 122.28 (F)
 TEMPERATURE INCREASED IN CELL # 2 = 18.44 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 2 = 0.07 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 3
 MASS TRANSFERED BETWEEN PHASES IN CELL # 2 = 113.333 (LB/HR)
 TURBULENCE-TURBULENCE FLOW MECHANISM
 SEPARATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 295420.62 OF THE LIQUID PHASE = 100471.19 AND OF THE TWO-PHASE FLOW = 114462.33 (FT/HR)
 TWO-PHASE FLUID DENSITY = 2.315 (LB/CU. FT), TWO-PHASE FLUID VISCOSITY = 0.223 (CP)
 TUBE WALL TEMPERATURE = 101.01 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 116.85 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 3307.83 BTU/(SQ-FT-HR)
 CALCULATED HEAT FLUX TO THE CELL = 142915.25 (BTU/SQ-FT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 2 = 51105.32 (BTU/HR)

LOCATION ON THE TUBE : Z(4)
 TEMPERATURE AT Z(4) = 137.23 (F), AND PRESSURE AT Z(4) = 54.26 (PSIA)
 SYSTEM AT Z(4) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = .21494 IN THE LIQUID PHASE = .00002
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = .09209 IN THE LIQUID PHASE = .00002
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = .64212 IN THE LIQUID PHASE = .51767
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = .03386 IN THE LIQUID PHASE = .48229
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02442 IN THE LIQUID PHASE = 2.54258 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.23309 IN THE LIQUID PHASE = 10.16143 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.62234 IN THE LIQUID PHASE = 40.50071 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.72393 IN THE LIQUID PHASE = 43.91384 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3396.2845 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 465.3995 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3480 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4185 (BTU/FT-H-F)

VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0169 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0189 AND IN THE LIQUID PHASE = 0.0026 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1406 AND IN THE LIQUID PHASE = 0.2747 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0564 AND IN THE LIQUID PHASE = 0.4723 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.252 IN THE LIQUID PHASE = 0.535 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.287 IN THE LIQUID PHASE = 0.426 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.222 IN THE LIQUID PHASE = 0.451 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.48767 LIQUID = 42.197 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.10012 LIQUID = 0.37009 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.40727 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.969 AND OF LIQUID = 0.438 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 3422.98 AND LIQUID = 2771.54

CELL # 3 FROM Z(3) TO Z(4)

AVERAGE DENSITY OF GAS = 0.47888 AND LIQUID = 42.378 OF CELL # 3 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.09155 AND LIQUID = 0.38527 OF CELL # 3 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 3 = 1.48591 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 1.039 AND LIQUID = 0.433 OF CELL # 3 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 594.87 AND LIQUID MASS FLOW RATE = 2005.13 OF CELL # 3 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 3419.61 AND LIQUID = 2740.41 OF CELL # 3
 GAS MASS FRACTION IN CELL # 3 = 0.2288
 SINGLE LIQUID PHASE PRESSURE DROP = 0.014291 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 3 = 56.47 AND EXITS = 54.26 (PSIA)
 ENROR OF PRESSURE DROP IN CELL # 3 = 2.207229 (PSI/CELL LENGTH)
 TWC-PHASE PRESSURE DROP IN CELL # 3 = 0.012, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID HOLDUP FRACTION = 0.1050
 TEMPERATURE OF THE STREAM ENTERING CELL # 3 = 122.28 AND LEAVING 137.23 (F)
 TEMPERATURE INCREASED IN CELL # 3 = 14.95 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 3 = 0.074, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 3
 MASS TRANSFERED BETWEEN PHASES IN CELL # 3 = 112.557 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPERATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 336215.19 OF THE LIQUID PHASE = 109135.70 AND OF THE TWC-PHASE FLOW = 129082.89 (FT/HR)
 TWC-PHASE FLUID DENSITY = 2.016 (LB/CU. FT), TWC-PHASE FLUID VISCOSITY = 0.222 (CP)
 TUBE WALL TEMPERATURE = 166.57 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LCC MEAN TEMPERATURE DIFFERENCE = 100.01 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWC-PHASE HEAT TRANSFER COEFFICIENT = 3961.36 BT/SCFT-HR
 CALCULATED HEAT FLUX TO THE CELL = 130501.35 (BTU/SCFT-HR)
 CALCULATED HEAT ACQUIRED IN CELL # 3 = 46666.21 (BTU/HR)

LOCATION ON THE TUBE : Z(5)

TEMPERATURE AT Z(5) = 148.03 (F), AND PRESSURE AT Z(5) = 51.74 (PSIA)
 SYSTEM AT Z(5) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = 0.1868 IN THE LIQUID PHASE = 0.03002
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = 0.0764 IN THE LIQUID PHASE = 0.0002
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = 0.70074 IN THE LIQUID PHASE = 0.99199
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = 0.03973 IN THE LIQUID PHASE = 0.50758
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02521 IN THE LIQUID PHASE = 2.24309 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.22149 IN THE LIQUID PHASE = 8.98835 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.63379 IN THE LIQUID PHASE = 40.27993 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.65729 IN THE LIQUID PHASE = 43.61435 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3438.5619 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 492.2995 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.1402 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4126 (BTU/FT-H-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0177 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0152 AND IN THE LIQUID PHASE = 0.0036 (CP)

VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1510 AND IN THE LIQUID PHASE = 0.2008 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0573 AND IN THE LIQUID PHASE = 0.4414 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 1.000 IN THE LIQUID PHASE = 0.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.252 IN THE LIQUID PHASE = 0.534 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.286 IN THE LIQUID PHASE = 0.432 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.224 IN THE LIQUID PHASE = 0.457 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.49265 LIQUID = 41.973 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.11320 LIQUID = 0.35254 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.44313 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.865 AND OF LIQUID = 0.445 (BTU/LB-F)
 AVERAGE REYNOLD NUMBER OF GAS = 3581.83 AND LIQUID = 2731.44

CELL # 4 FROM Z1 4) TO Z1 5)
 AVERAGE DENSITY OF GAS = 0.49066 AND LIQUID = 42.085 OF CELL # 4 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.10666 AND LIQUID = 0.36131 OF CELL # 4 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 4 = 1.4520 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.917 AND LIQUID = 0.441 OF CELL # 4 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 710.78 AND LIQUID MASS FLOW RATE = 1889.22 OF CELL # 4 (LB/HR)
 AVERAGE REYNOLD NUMBER OF GAS = 3502.41 AND LIQUID = 2751.49 OF CELL # 4
 GAS MASS FRACTION IN CELL # 4 = 0.2734
 SINGLE LIQUID PHASE PRESSURE DROP = 0.12765 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 4 = 54.26 AND EXITS = 51.74 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 4 = 2.520446 (PSI/CELL LENGTH)
 ENTER OF PRESSURE DROP IN CELL # 4 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID HOLDUP FRACTION = 0.0029
 TEMPERATURE OF THE STREAM ENTERING CELL # 4 = 137.23 AND LEAVING 140.08 (F)
 TEMPERATURE INCREASED IN CELL # 4 = 10.85 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 4 = 0.12 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
 MASS TRANSFERED BETWEEN PHASES IN CELL # 4 = 119.252 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 386331.41 OF THE LIQUID PHASE = 117080.01 AND OF THE TWO-PHASE FLOW = 144656.09 (FT/HR)
 TWO-PHASE FLUID DENSITY = 1.741 (LB/CC. FT), TWO-PHASE FLUID VISCOSITY = 0.215 (CP)
 TUBE WALL TEMPERATURE = 171.90 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 87.33 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 4567.54 .01 BASED ON TUBE INSIDE DIAMETER = 1364.60 (BTU/SQFT-HR-F)
 CALCULATED HEAT FLUX TO THE CELL = 119167.01 (BTU/SQFT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 4 = 42613.15 (BTU/HR)

LOCATION ON THE TUBE = Z1 6)
 TEMPERATURE AT Z1 6) = 156.40 (F), AND PRESSURE AT Z1 6) = 48.90 (PSIA)
 SYSTEM AT Z1 6) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = 0.13743 IN THE LIQUID PHASE = 0.0001
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = 0.6746 IN THE LIQUID PHASE = 0.0001
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = 0.7267 IN THE LIQUID PHASE = 0.46408
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = 0.4655 IN THE LIQUID PHASE = 0.53589
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02369 IN THE LIQUID PHASE = 2.11194 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.02391 IN THE LIQUID PHASE = 8.45786 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.66313 IN THE LIQUID PHASE = 40.03032 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.66345 IN THE LIQUID PHASE = 43.38457 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3472.2063 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 497.6821 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3341 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4080 (BTU/FT-HR-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0185 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0194 AND IN THE LIQUID PHASE = 0.1022 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1002 AND IN THE LIQUID PHASE = 0.2506 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0350 AND IN THE LIQUID PHASE = 0.4155 (CP)

SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.252 IN THE LIQUID PHASE = 0.542 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.209 IN THE LIQUID PHASE = 0.437 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.225 IN THE LIQUID PHASE = 0.462 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.40751 LIQUID = 41.027 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.12341 LIQUID = 0.34110 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.42684 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.769 AND OF LIQUID = 0.450 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 3792.04 AND LIQUID = 2039.65

CELL # 5 FROM Z(5) TO Z(6)
 AVERAGE DENSITY OF GAS = .49076 AND LIQUID = 41.900 OF CELL # 5 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.11031 AND LIQUID = 0.34082 OF CELL # 5 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 5 = 1.43499 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.827 AND LIQUID = 0.447 OF CELL # 5 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 329.78 AND LIQUID MASS FLOW RATE = 1770.22 OF CELL # 5 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 3686.54 AND LIQUID = 2685.04 OF CELL # 5
 GAS MASS FRACTION IN CELL # 5 = 0.2151
 SINGLE LIQUID PHASE PRESSURE DROP = .011311 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 5 = 51.74 AND EXITS = 48.90 (PSIA)
 INC-PHASE PRESSURE DROP IN CELL # 5 = 2.841947 (PSI/CELL LENGTH)
 PERCENT OF PRESSURE DROP IN CELL # 5 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID HOLDUP FRACTION = 0.0023
 TEMPERATURE OF THE STREAM ENTERING CELL # 5 = 148.08 AND LEAVING 150.46 (F)
 TEMPERATURE INCREASED IN CELL # 5 = 8.38 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 5 = 0.21 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
 MASS TRANSFERED BETWEEN PHASES IN CELL # 5 = 110.754 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 44303.02 OF THE LIQUID PHASE = 124295.01 AND OF THE TWO-PHASE FLOW = 161477.65 (FT/HR)
 INC-PHASE FLUID DENSITY = 1.560 (LB/CU. FT), INC-PHASE FLUID VISCOSITY = 0.215 (CP)
 TUBE WALL TEMPERATURE = 176.22 (F) HEAT FLX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 77.83 (F) LOG MEAN DIAMETER = 3.08 (FT)
 INC-PHASE HEAT TRANSFER COEFFICIENT = 5125.46 BTU/SCFT-HR
 CALCULATED HEAT FLX TO THE CELL = 109756.83 (BTU/SCFT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 5 = 39262.44 (BTU/HR)

LOCATION ON THE TUBE : Z(7)
 TEMPERATURE AT Z(7) = 163.34 (F), AND PRESSURE AT Z(7) = 45.73 (PSIA)
 SYSTEM AT Z(7) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = .13950 IN THE LIQUID PHASE = .00001
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = .05978 IN THE LIQUID PHASE = .00001
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = .74245 IN THE LIQUID PHASE = .43531
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = .05327 IN THE LIQUID PHASE = .56467
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.02161 IN THE LIQUID PHASE = 2.03093 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.18437 IN THE LIQUID PHASE = 8.12727 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.57125 IN THE LIQUID PHASE = 39.32408 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.62028 IN THE LIQUID PHASE = 43.19481 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3499.6639 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 502.1275 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3290 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4043 (BTU/FT-H-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0192 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0146 AND IN THE LIQUID PHASE = 0.1093 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1673 AND IN THE LIQUID PHASE = 0.2427 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0366 AND IN THE LIQUID PHASE = 0.4026 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.200 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.252 IN THE LIQUID PHASE = 0.542 (BTU/LB-F)

SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.202 IN THE LIQUID PHASE = 0.440 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.226 IN THE LIQUID PHASE = 0.465 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.7477 LIQUID = 41.727 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.1144 LIQUID = 0.34297 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.41545 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.733 AND OF LIQUID = 0.455 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 4017.97 AND LIQUID = 2523.64

CELL # 6 FROM Z(6) TO Z(7)
 AVERAGE DENSITY OF GAS = 0.48134 AND LIQUID = 41.777 OF CELL # 6 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.12743 AND LIQUID = 0.33703 OF CELL # 6 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 6 = 1.42115 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.761 AND LIQUID = 0.452 OF CELL # 6 (BTU/LB-F)
 AVERAGE MASS FLOW RATE = 946.29 AND LIQUID MASS FLOW RATE = 1653.71 OF CELL # 6 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 3905.00 AND LIQUID = 2581.74 OF CELL # 6
 GAS MASS FRACTION IN CELL # 6 = 0.3640
 SINGLE LIQUID PHASE PRESSURE DROP = -0.05979 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 6 = 48.90 AND EXITS = 45.73 (PSIA)
 TAC-PHASE PRESSURE DROP IN CELL # 6 = 3.16643 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 6 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID FLOCC FRACTION = 0.0733
 TEMPERATURE OF THE STREAM ENTERING CELL # 6 = 156.46 AND LEAVING 163.34 (F)
 TEMPERATURE INCREASED IN CELL # 6 = 6.89 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 6 = 0.25 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
 MASS TRANSFERED BETWEEN PHASES IN CELL # 6 = 114.270 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL

VELOCITY OF THE GAS PHASE = 513869.41 OF THE LIQUID PHASE = 130872.08 AND OF THE TWO-PHASE FLOW = 179589.08 (FT/HR)
 TAC-PHASE FLUID DENSITY = 1.296 (LB/CU-FT), INC-PHASE FLUID VISCOSITY = 0.211 (CP)
 TUBE WALL TEMPERATURE = 180.17 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ-FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 70.26 (F) LOG MEAN DIAMETER = 0.08 (FT)
 INC-PHASE HEAT TRANSFER COEFFICIENT = 5043.90 BTU/(SQ-FT-HR) BASED ON TUBE INSIDE DIAMETER = 1447.38 (BTU/FT-HR-F)
 CALCULATED HEAT FLUX TO THE CELL = 101651.27 (BTU/SCFT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 6 = 36363.57 (BTU/HR)

LOCATION ON THE TUBE : Z(8)
 TEMPERATURE AT Z(8) = 169.25 (F), AND PRESSURE AT Z(8) = 42.24 (PSIA)
 SYSTEM AT Z(8) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = -12601 IN THE LIQUID PHASE = -00001
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = -05400 IN THE LIQUID PHASE = -00001
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = -75087 IN THE LIQUID PHASE = -40700
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = -06933 IN THE LIQUID PHASE = -59298
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.01949 IN THE LIQUID PHASE = 1.95710 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.16400 IN THE LIQUID PHASE = 7.92681 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.53605 IN THE LIQUID PHASE = 39.03967 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.55168 IN THE LIQUID PHASE = 43.02405 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3524.1850 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 506.0982 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3245 (BTU/FT-HR-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.4009 (BTU/FT-HR-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0159 AND IN THE LIQUID PHASE = 0.0010 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0197 AND IN THE LIQUID PHASE = 0.1156 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1736 AND IN THE LIQUID PHASE = 0.2355 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0591 AND IN THE LIQUID PHASE = 0.3884 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.222 IN THE LIQUID PHASE = 0.548 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.254 IN THE LIQUID PHASE = 0.444 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.227 IN THE LIQUID PHASE = 0.469 (BTU/LB-F)

AVERAGE DENSITY OF GAS = 0.45023 LIQUID = 41.646 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.13756 LIQUID = 0.32633 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.40710 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.671 AND OF LIQUID = 0.459 (BTU/LB-F)
 AVERAGE REYNOLDS NUMBER OF GAS = 4238.17 AND LIQUID = 2401.57

CELL # 7 FROM Z1 7) TO Z1 8)
 AVERAGE DENSITY OF GAS = 0.46550 AND LIQUID = 41.686 OF CELL # 7 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.13470 AND LIQUID = 0.32965 OF CELL # 7 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 7 = 1.41128 (BTU/FT-H-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.712 AND LIQUID = 0.457 OF CELL # 7 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 1057.19 AND LIQUID MASS FLOW RATE = 1542.61 OF CELL # 7 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 4128.07 AND LIQUID = 2462.60 OF CELL # 7

GAS MASS FRACTION IN CELL # 7 = 0.4066
 SINGLE LIQUID PHASE PRESSURE DROP = 0.008787 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 7 = 45.73 AND EXITS = 42.24 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 7 = 3.491342 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 7 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID FLOCCUP FRACTION = 0.0655
 TEMPERATURE OF THE STREAM ENTERING CELL # 7 = 163.34 AND LEAVING 169.29 (F)
 TEMPERATURE INCREASED IN CELL # 7 = 5.95 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 7 = 0.17 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
 MASS TRANSFERED BETWEEN PHASES IN CELL # 7 = 107.525 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL

VELOCITY OF THE GAS PHASE = 588074.53 OF THE LIQUID PHASE = 136922.04 AND OF THE TWO-PHASE FLOW = 199025.25 (FT/HR)
 TWO-PHASE FLUID DENSITY = 1.120 (LB/CU. FT), TWO-PHASE FLUID VISCOSITY = 0.208 (CP)
 TUBE WALL TEMPERATURE = 183.69 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 63.75 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 6134.46 BTU/(SQ-FT-HR)
 CALCULATED HEAT FLUX TO THE CELL = 94235.07 (BTU/SQ-FT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 7 = 33657.69 (BTU/HR)

LOCATION ON THE TUBE = Z1 9)
 TEMPERATURE AT Z1 9) = 174.21 (F), AND PRESSURE AT Z1 9) = 38.42 (PSIA)
 SYSTEM AT Z1 9) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = 0.11530 IN THE LIQUID PHASE = 0.00001
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = 0.6941 IN THE LIQUID PHASE = 0.0001
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = 0.75489 IN THE LIQUID PHASE = 0.37772
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = 0.08040 IN THE LIQUID PHASE = 0.62226
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.61745 IN THE LIQUID PHASE = 1.99420 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.14326 IN THE LIQUID PHASE = 7.57240 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.50372 IN THE LIQUID PHASE = 39.48646 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.55386 IN THE LIQUID PHASE = 42.88226 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3544.2621 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 509.3481 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3208 (BTU/FT-H-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.3981 (BTU/FT-H-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0204 AND IN THE LIQUID PHASE = 0.0016 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0198 AND IN THE LIQUID PHASE = 0.1208 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1788 AND IN THE LIQUID PHASE = 0.2307 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0595 AND IN THE LIQUID PHASE = 0.3773 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.300 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.252 IN THE LIQUID PHASE = 0.550 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.256 IN THE LIQUID PHASE = 0.447 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.227 IN THE LIQUID PHASE = 0.472 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.47398 LIQUID = 41.600 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.14100 LIQUID = 0.32192 (CP)

AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.4604W (BTU/FT-F) (F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.658 AND OF LIQUID = 0.403 (BTU/LB-F)
 AVERAGE METACAL NUMBER OF GAS = 1406.8c AND LIQUID = 2265.84

CELL # 8 FROM Z(10) TO Z(19)
 AVERAGE DENSITY OF GAS = 0.4506 AND LIQUID = 41.623 OF CELL # 8 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.14051 AND LIQUID = 0.12412 OF CELL # 8 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 8 = 1.40377 (BTU/FT-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.674 AND LIQUID = 0.461 OF CELL # 8 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 1162.54 AND LIQUID MASS FLOW RATE = 1437.46 OF CELL # 8 (LB/HR)
 AVERAGE METACAL NUMBER OF GAS = 4352.53 AND LIQUID = 2333.70 OF CELL # 8
 GAS MASS FRACTION IN CELL # 8 = 0.4471
 SINGLE LIQUID PHASE PRESSURE DROP = 0.007722 (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 8 = 12.24 AND EXITS = 38.42 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 8 = 3.82365 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 8 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID FLOW FRACTION = 0.0567
 TEMPERATURE OF THE STREAM ENTERING CELL # 8 = 169.29 AND LEAVING 174.21 (F)
 TEMPERATURE INCREASED IN CELL # 8 = 4.92 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 8 = 0.14 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
 MASS TRANSFERED BETWEEN PHASES IN CELL # 8 = 103.168 (LB/HR)
 TURBULENT-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL
 VELOCITY OF THE GAS PHASE = 672100.18 OF THE LIQUID PHASE = 142632.50 AND OF THE TWO-PHASE FLOW = 220196.92 (FT/HR)
 TWO-PHASE FLUID DENSITY = 0.582 (LB/CL. FT), TWO-PHASE FLUID VISCOSITY = 0.205 (CP)
 TUBE WALL TEMPERATURE = 186.81 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG P-RAN TEMPERATURE DIFFERENCE = 50.34 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 6551.21 (BTU/SCFT-MR)
 CALCULATED HEAT FLUX TO THE CELL = 0.7640.84 (BTU/SCFT-MR)
 CALCULATED HEAT REQUIRED IN CELL # 8 = 31341.79 (BTU/HR)

LOCATION ON THE TUBE : Z(10)
 TEMPERATURE AT Z(10) = 177.77 (F), AND PRESSURE AT Z(10) = 34.24 (PSIA)
 SYSTEM AT Z(10) IS ASSUMED TO BE IN EQUILIBRIUM STATE
 MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = 0.10624 IN THE LIQUID PHASE = 0.0001
 MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = 0.0553 IN THE LIQUID PHASE = 0.0001
 MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = 0.75672 IN THE LIQUID PHASE = 0.34539
 MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = 0.09150 IN THE LIQUID PHASE = 0.65460
 DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.01501 IN THE LIQUID PHASE = 1.84562 (LB/CUFT)
 DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.12218 IN THE LIQUID PHASE = 7.37755 (LB/CUFT)
 DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.46827 IN THE LIQUID PHASE = 39.38047 (LB/CUFT)
 DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.51491 IN THE LIQUID PHASE = 42.77818 (LB/CUFT)
 THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3554.5428 (BTU/FT-F)
 THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 511.6603 (BTU/FT-F)
 THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3182 (BTU/FT-F)
 THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.3962 (BTU/FT-F)
 VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0207 AND IN THE LIQUID PHASE = 0.0013 (CP)
 VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0159 AND IN THE LIQUID PHASE = 0.1244 (CP)
 VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1824 AND IN THE LIQUID PHASE = 0.2270 (CP)
 VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0556 AND IN THE LIQUID PHASE = 0.3697 (CP)
 SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.253 IN THE LIQUID PHASE = 0.551 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.257 IN THE LIQUID PHASE = 0.449 (BTU/LB-F)
 SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.279 IN THE LIQUID PHASE = 0.474 (BTU/LB-F)
 AVERAGE DENSITY OF GAS = 0.4466 LIQUID = 41.623 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.1406 LIQUID = 0.1241 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.40377 (BTU/FT-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.674 AND OF LIQUID = 0.461 (BTU/LB-F)

AVERAGE REYNOLDS NUMBER OF GAS = 4729.15 AND LIQUID = 2100.25

CELL # 9 FROM Z(1) TO Z(11)
AVERAGE DENSITY OF GAS = .42128 AND LIQUID = .41002 OF CELL # 9 (LB/CUFT)
AVERAGE VISCOSITY OF GAS = 0.14485 AND LIQUID = 0.32118 (BTU/FT-H-F)
AVERAGE THERMAL CONDUCTIVITY OF CELL # 9 = 1.37781 (BTU/FT-H-F)
AVERAGE SPECIFIC HEAT OF GAS = 0.643 AND LIQUID = 0.464 OF CELL # 9 (BTU/LB-F)
AVERAGE GAS MASS FLOW RATE = 1705.03 AND LIQUID MASS FLOW RATE = 1334.12 OF CELL # 9 (LB/HR)
AVERAGE REYNOLDS NUMBER OF GAS = 4598.03 AND LIQUID = 2186.04 OF CELL # 9
GAS MASS FRACTION IN CELL # 9 = 0.4065
SINGLE LIQUID PHASE PRESSURE DROP = .036743 (PSI/FT)
PRESSURE OF THE STREAM ENTERS CELL # 9 = 38.42 AND EXITS = 34.24 (PSIA)
TWO-PHASE PRESSURE DROP IN CELL # 9 = 4.17772 (PSI/CELL LENGTH)
ERROR OF PRESSURE DROP IN CELL # 9 = 0.01 %, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
LIQUID HOLDUP FRACTION = 0.0524
TEMPERATURE OF THE STREAM ENTERING CELL # 9 = 174.21 AND LEAVING 177.77 (F)
TEMPERATURE INCREASED IN CELL # 9 = 3.56 (F)
ERROR OF TEMPERATURE ITERATING IN CELL # 9 = 0.15 %, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 5
MASS TRANSFERED BETWEEN PHASES IN CELL # 9 = 103.515 (LB/HR)
TURBULENT-TURBULENT FLOW MECHANISM
SEPARATED FLOW MODEL

VELOCITY OF THE GAS PHASE = 768154.80 OF THE LIQUID PHASE = 148149.91 AND OF THE TWO-PHASE FLOW = 244059.43 (FT/HR)
TWO-PHASE FLUID DENSITY = 0.850 (LB/CU-FT), TWO-PHASE FLUID VISCOSITY = 0.202 (CPI)
TUBE WALL TEMPERATURE = 139.31 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
LOG MEAN TEMPERATURE DIFFERENCE = 54.12 (F) LOG MEAN DIAMETER = 0.08 (FT)
TWO-PHASE HEAT TRANSFER COEFFICIENT = 7024.74 BTU/SCFT-HR
CALCULATED HEAT FLUX TO THE CELL = 82478.40 (BTU/SCFT-HR)
CALCULATED HEAT REQUIRED IN CELL # 9 = 29453.60 (BTU/HR)

LOCATION ON THE TUBE : Z(11)
TEMPERATURE AT Z(11) = 180.84 (F), AND PRESSURE AT Z(11) = 29.66 (PSIA)
SYSTEM AT Z(11) IS ASSUMED TO BE IN EQUILIBRIUM STATE
MASS FRACTION OF COMPONENT 1 IN THE GAS PHASE = .09862 IN THE LIQUID PHASE = .03001
MASS FRACTION OF COMPONENT 2 IN THE GAS PHASE = .34226 IN THE LIQUID PHASE = .03000
MASS FRACTION OF COMPONENT 3 IN THE GAS PHASE = .75489 IN THE LIQUID PHASE = .31210
MASS FRACTION OF COMPONENT 4 IN THE GAS PHASE = .10422 IN THE LIQUID PHASE = .68785
DENSITY OF COMPONENT 1 IN THE GAS PHASE = 0.01364 IN THE LIQUID PHASE = 1.00453 (LB/CUFT)
DENSITY OF COMPONENT 2 IN THE GAS PHASE = 0.10062 IN THE LIQUID PHASE = 7.21430 (LB/CUFT)
DENSITY OF COMPONENT 3 IN THE GAS PHASE = 0.42793 IN THE LIQUID PHASE = 39.29644 (LB/CUFT)
DENSITY OF COMPONENT 4 IN THE GAS PHASE = 0.47340 IN THE LIQUID PHASE = 42.65411 (LB/CUFT)
THERMAL CONDUCTIVITY OF COMPONENT 1 IN THE LIQUID PHASE = 3589.6066 (BTU/FT-H-F)
THERMAL CONDUCTIVITY OF COMPONENT 2 IN THE LIQUID PHASE = 513.4515 (BTU/FT-H-F)
THERMAL CONDUCTIVITY OF COMPONENT 3 IN THE LIQUID PHASE = 1.3162 (BTU/FT-H-F)
THERMAL CONDUCTIVITY OF COMPONENT 4 IN THE LIQUID PHASE = 1.3546 (BTU/FT-H-F)
VISCOSITY OF COMPONENT 1 IN THE GAS PHASE = 0.0210 AND IN THE LIQUID PHASE = 0.0011 (CPI)
VISCOSITY OF COMPONENT 2 IN THE GAS PHASE = 0.0206 AND IN THE LIQUID PHASE = 0.1273 (CPI)
VISCOSITY OF COMPONENT 3 IN THE GAS PHASE = 0.1853 AND IN THE LIQUID PHASE = 0.2243 (CPI)
VISCOSITY OF COMPONENT 4 IN THE GAS PHASE = 0.0600 AND IN THE LIQUID PHASE = 0.3640 (CPI)
SPECIFIC HEAT OF COMPONENT 1 IN THE GAS PHASE = 3.500 IN THE LIQUID PHASE = 7.000 (BTU/LB-F)
SPECIFIC HEAT OF COMPONENT 2 IN THE GAS PHASE = 0.253 IN THE LIQUID PHASE = 0.552 (BTU/LB-F)
SPECIFIC HEAT OF COMPONENT 3 IN THE GAS PHASE = 0.258 IN THE LIQUID PHASE = 0.451 (BTU/LB-F)
SPECIFIC HEAT OF COMPONENT 4 IN THE GAS PHASE = 0.228 IN THE LIQUID PHASE = 0.476 (BTU/LB-F)
AVERAGE DENSITY OF GAS = 0.27766 LIQUID = 41.033 (LB/CUFT)
AVERAGE VISCOSITY OF GAS = 0.14903 LIQUID = 0.32641 (CPI)
AVERAGE THERMAL CONDUCTIVITY OF LIQUID = 1.35158 (BTU/FT-H-F)
AVERAGE SPECIFIC HEAT OF GAS = 0.605 AND OF LIQUID = 0.458 (BTU/LB-F)
AVERAGE REYNOLDS NUMBER OF GAS = 5012.31 AND LIQUID = 1935.15

CELL # 10 FROM Z(10) TO Z(11)
 AVERAGE DENSITY OF GAS = .39317 AND LIQUID = 41.619 OF CELL # 10 (LB/CUFT)
 AVERAGE VISCOSITY OF GAS = 0.14785 AND LIQUID = 0.32042 OF CELL # 10 (CP)
 AVERAGE THERMAL CONDUCTIVITY OF CELL # 10 = 1.39337 (BTU/FT-HR-F)
 AVERAGE SPECIFIC HEAT OF GAS = 0.617 AND LIQUID = 0.467 OF CELL # 10 (BTU/LB-F)
 AVERAGE GAS MASS FLOW RATE = 1368.57 AND LIQUID MASS FLOW RATE = 1211.43 OF CELL # 10 (LB/HR)
 AVERAGE REYNOLDS NUMBER OF GAS = 4870.75 AND LIQUID = 2022.69 OF CELL # 10
 GAS MASS FRACTION IN CELL # 10 = 0.5264
 SINGLE LIQUID PHASE PRESSURE DROP = .60583? (PSI/FT)
 PRESSURE OF THE STREAM ENTERS CELL # 10 = 34.24 AND EXITS = 29.66 (PSIA)
 TWO-PHASE PRESSURE DROP IN CELL # 10 = 4.578308 (PSI/CELL LENGTH)
 ERROR OF PRESSURE DROP IN CELL # 10 = 0.014, TOTAL NUMBER OF PRESSURE ITERATING LOOP = 1
 LIQUID FLOWUP FRACTION = 0.0466
 TEMPERATURE OF THE STREAM ENTERING CELL # 10 = 177.77 AND LEAVING 180.64 (F)
 TEMPERATURE INCREASED IN CELL # 10 = 2.87 (F)
 ERROR OF TEMPERATURE ITERATING IN CELL # 10 = 0.22%, TOTAL NUMBER OF TEMPERATURE ITERATING LOOP = 7
 MASS TRANSFERRED BETWEEN PHASES IN CELL # 10 = 101.863 (LB/HR)
 TURBULENCE-TURBULENT FLOW MECHANISM
 SEPARATED FLOW MODEL
 VISCOSITY OF THE GAS PHASE = 884378.35 OF THE LIQUID PHASE = 153856.18 AND OF THE TWO-PHASE FLOW = 272215.55 (FT/HR)
 TWO-PHASE FLUID DENSITY = 0.741 (LB/CC. FT), TWO-PHASE FLUID VISCOSITY = 0.198 (CP)
 TUBE WALL TEMPERATURE = 191.14 (F) HEAT FLUX AREA OF THE CELL = 0.36 (SQ. FT)
 LOG MEAN TEMPERATURE DIFFERENCE = 50.58 (F) LOG MEAN DIAMETER = 0.08 (FT)
 TWO-PHASE HEAT TRANSFER COEFFICIENT = 7441.56 UI BASED ON TUBE INSIDE DIAMETER = 1542.61 (BTU/SQFT-HR-F)
 CALCULATED HEAT FLUX TO THE CELL = 78645.75 (BTU/SQFT-HR)
 CALCULATED HEAT REQUIRED IN CELL # 10 = 28123.07 (BTU/HR)
 TOTAL TUBE LENGTH = 15.70 (FT)
 TOTAL HEAT ADDED TO THE ENTIRE SYSTEM = 393821.91 (BTU/HR)
 TOTAL PRESSURE DROP ALONG THE TUBE = 30.341 (PSI/TUBE LENGTH)
 TOTAL TEMPERATURE INCREASE = 100.642 (F)
 TOTAL MASS TRANSFERRED BETWEEN PHASES IN THE ENTIRE SYSTEM = 1095.75 (LB/HR)
 COMPLETE CALCULATION

TEMPERATURE	PRESSURE	GAS MASS FLOW RATE	LIQUID MASS FLOW RATE
60.00	323.75	2276.25	
103.85	58.37	424.76	2175.24
122.28	56.47	538.60	2041.40
137.23	54.26	651.15	1948.85
148.08	51.74	770.41	1829.59
156.46	48.90	889.16	1710.84
163.34	45.72	1003.43	1596.57
169.25	42.24	1110.55	1489.05
174.21	38.42	1214.12	1385.88
177.77	34.24	1317.64	1282.36
180.64	29.66	1419.50	1183.50

AVERAGE TUBE WALL TEMPERATURE = 176.22 (F)
 AVERAGE LIQUID FLOWUP FRACTION = 0.0822

VITA

Thanh Hoang Nguyen was born in Binh Dinh, Viet-Nam on February 20, 1956. He attended elementary school in Sao Mai, DaNang and was graduated from Dac-Lo High School, SaiGon in June 1973. The following September he entered the University of SaiGon with a major of chemical engineering. He left Viet-nam and came to The United States in March 1977.

He entered the University of Tennessee, Knoxville in January 1978, worked for International Fertilizer Development Center-Muscle Shoals, Alabama from 1979 to 1981 during the Engineering Cooperative Program, and received a Bachelor of Science degree in Chemical Engineering in August 1982. He re-entered the Graduate School of the University of Tennessee in September 1982, worked for Professional Analysis Inc. in Oak Ridge, Tennessee from April 1984 to December 1984 as a research engineer, and received a Master of Science degree in Chemical Engineering in March 1985.