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HEAT TRANSFER STUDIES OF PACKED BED REACTORS

By

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HEAT TRANSFER STUDIES OF PACKED BED COAL LIQUEFACTION REACTORS

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ABSTRACT

The effective thermal conductivity (K_e) and wall heat transfer coefficient (H_w) were determined in a packed bed reactor model over a range of gas mass velocities from 0 to 15,000 kg/m² hr, liquid mass velocities from 5,000 to 65,000 kg/m² hr, and pressures that simulate coal liquefaction pressures from 1,000 to 4,000 psig. A correlation for K_e and H_w was developed for the range of study. Several mathematical models for one and two dimensions, and for one and two parameters, were derived to describe two-phase flow in packed beds. Two flow regimes are described. The entrance effects for the low liquid flow regime seem to be limited to a length of about ten diameters, while the entrance effects for the high liquid flow seem to extend through the entire length of the column.

I. INTRODUCTION

Tubular packed bed reactors are often employed for carrying out reactions occurring with porous catalysts. Coal liquefaction reactions are known to be exothermic. In such cases, certain reactor conditions can promote a state of unstable equilibrium. In this state, small changes in conditions, i.e., feed rate or temperature, will cause serious thermal excursions. This could result in the formation of coke and the eventual blockage of the reactor. Thus, any final reactor design must ensure low sensitivity to operating conditions and prevent undesirable operating regimes in pilot or industrial plants.

The objectives of this study were two-fold. The first objective was to obtain effective thermal conductivities (K_e) and wall heat transfer coefficients (H_w) for two-phase flow (nitrogen and water) in a packed bed. This entailed measuring the radial and axial temperature profiles for a wide range of mass velocities and pressures that would simulate coal liquefaction conditions. The primary variables were liquid and gas mass velocities and gas densities. The second objective was to develop a theoretical model for use in reactor design analysis, parameter sensitivity study and scale-up predictions of the temperature profile.

If one compares the heat generated in the reaction with the ability of the system to remove heat, based on single-phase correlations (19), one would conclude that the maximum diameter for a stable packed bed coal liquefaction reactor will be approximately 0.15 meters. However, the heat transfer properties of single-phase flow are definitely different from those of two-phase flow, and cold hydrogen can be added to the system to slow down the reaction. Nevertheless, this illustrates the need for heat transfer data for two-phase flow in packed beds.

II. BACKGROUND INFORMATION

Very little heat transfer research has been conducted in two-phase flow and most of this has been in unpacked bubble columns. A review of bubble column research has been written by Mashelkar (11). However, unpacked bubble columns are not very relevant to two-phase flow in packed beds because the packing creates entrance and wall effects for the entire length of the reactor. The hydrodynamics for two-phase downflow in a packed column have been reported by references 4,5, and 6, and for upflow by references 7,8,9, and 10. Heat transfer for two-phase downflow has been studied by Weekman and Myers (2) and K. Hashimoto, et al. (12). No literature was found on heat transfer for two-phase upflow in packed beds in the flow regimes of interest.

In the paper of Weekman and Myers, heat transfer resistance near the surface is neglected and only the resistance in the inner sections of the packed bed is considered. An interrelating equation for the overall radial effective thermal conductivity of a gas-liquid concurrent packed bed was proposed. However, the heat transfer resistance at the wall is large and should not be neglected. K. Hashimoto, et al. (12) adapted the single-phase flow method of Phillips, et al., which included the resistance at the wall to two-phase flow. This method is similar to the method employed in this paper except that Phillips' method assumes a constant wall temperature. In this system, it was felt that the wall temperature varied significantly with height; and thus the approximation of a constant wall temperature was unacceptable and wall temperature variations had to be taken into account. It should be noted that, in any industrial application, the wall temperature will probably vary significantly along the length of the reactor.

III. EXPERIMENTAL

A. Apparatus

A schematic illustration of the experimental system used in the heat transfer experiments is shown in figure 1. The reactor model is shown in figure 2. The reactor is six feet high, has a three-inch inside diameter, and is surrounded by a cooling water jacket. The column is packed with $1/4 \times 1/4$ -inch ceramic cylinders. Water and nitrogen are heated in a chamber containing six heating rods and packed with copper shot. Above the heating rods, there is a distributor with a single $1/2$ -inch hole to provide uniform mixing of the hot fluid. A barrier against migration of the copper shot into the reactor was provided by a porous plate located just above the distributor. A six-inch calming section filled with $1/4 \times 1/4$ -inch cylinders is situated close to the porous plate. The reactor is constructed of 304 stainless steel. Radial temperature profiles were determined using multiple thermocouples containing five individual type k thermocouples. The spacing of the thermocouple positions was $1/4$ -inch. Six of the multiple thermocouples were situated a foot apart along the height of the bed. The first bed thermocouple is six inches above the entrance. Another multiple thermocouple was placed at the entrance of the bed. At the exit of the bed, two thermocouples measured the average exit temperature. Bed pressures were measured using pressure transducers situated at five different locations along the height of the bed. The pressure drop was generally between 5 and 15 kPa/m, depending on the mass velocities. There is significant error in this measurement due to substantial pressure fluctuations in the bed. The pressure fluctuations were due to the hydrodynamics of two-phase flow, rather than to some experimental artifact such as liquid pumping, since there were no pressure fluctuations for single-phase flow. Nitrogen is supplied from a liquid nitrogen storage tank at 1.0 MPa (150 psig), and gas flows were measured separately by two Foxboro orifice gauges. The bed pressure was controlled by a letdown valve at the top of the bed. After the pressure was reduced, the fluid entered a liquid-gas separator and was vented.

B. Void fraction

The average density of the particles was obtained by counting several hundred pellets, weighing them, and then obtaining their volume by water displacement. The model reactor was then filled with a known weight of pellets. The reactor was sealed, and nitrogen and water were passed through the bed for several hours. At this point, the model reactor was opened and more pellets were added to make up for settling. This procedure was repeated several times until no more settling occurred. Since one knew the density of the pellets, the volume of the model reactor, and the weight of pellets in the model reactor, the void fraction could be calculated. Settling of the bed was found to be very important. The settling procedure would change the void fraction from about 0.35 to 0.25.

C. Operating procedure

The normal operating procedure was to set the liquid flow first. After the entire bed was wet, the desired nitrogen flow rate and the system pressure were established. At this point, the temperature of the incoming fluid was increased to approximately 60° C. Only after the entire system remained constant for at least a half hour were any temperature profiles measured.

IV. ANALYTICAL SOLUTIONS OF THE CONCURRENT TWO-PHASE FLOW ENERGY EQUATION APPLIED TO A PACKED BED (Ref. 15)

A. One dimensional model - no radial dependence - one parameter

The simplest derivation for the solution of the energy equation when variable wall temperature boundary conditions are applied considers only one dimension. The differential equation is derived assuming no radial variation of temperature. The only "parameter" involved is the wall heat transfer coefficient. The energy balance on a differential volume is

$$S(G_g C_{pg} + G_l C_{pl}) dt = h_w (t_w(z) - t(z)) dA \quad (1)$$

The relevant boundary condition is

$$t = t_o \text{ at } z = 0 \quad (2)$$

If a linear variation of the wall temperature is assumed,

$$t_w = t_{wo} - cz \quad (3)$$

the analytical solution is

$$t(z) = t_o \exp(-z/2) + (t_{wo} - cz) (1 - \exp(-z/a)) \quad (4)$$

where

$$a = (R/2h_w) (G_g C_{pg} + G_l C_{pl}) \quad (5)$$

The constant c is the absolute value of the slope of the wall temperature and must be known.

If an exponential variation is assumed for the wall temperature,

$$t_w = t_{wo} e^{-uz} \quad (6)$$

the boundary condition is

$$t = t_o \text{ at } z = 0 \quad (7)$$

and the analytical solution is

$$t(z) = t_o \exp(-z/a) + t_{wo}/(1 + au) (\exp(-uz) - \exp(-z/a)) \quad (8)$$

The constant u is the exponential constant which describes the wall temperature variation and must be known.

B. Two dimensional model - one parameter

To account for radial dependence, the differential equation is modified. However, in this case, the parameter is changed to K_e , the effective thermal conductivity. When conduction is considered only in the radial direction, the energy balance over a differential volume of packed bed is

$$(G_g C_{Pg} + G_l C_{Pl}) \frac{\partial t}{\partial z} = K_e \left(\frac{\partial^2 t}{\partial r^2} + \frac{1}{r} \frac{\partial t}{\partial r} \right) \quad (9)$$

The relevant boundary conditions are

$$t = t_o \text{ at } z = 0$$

$$\frac{\partial t}{\partial r} = 0 \text{ at } r = 0 \quad (10)$$

$$t = t_w \text{ at } r = R$$

Note that the wall temperature is involved only in the application of the boundary conditions.

For a linear wall temperature:

$$t = t_{wo} - cz \quad (11)$$

the analytical solution is

$$t(r, z) = t_{wo} - cz + \sum_{n=1}^{\infty} \frac{(t_o - t_{wo})}{B_n J_1(B_n)} J_0(B_n \rho) \exp(-B_n^2 z/a) \quad (12)$$

B_n is the root of the following equation:

$$J_0(B_n) = 0 \quad (13)$$

and

$$a = \frac{R^2 (G_g C_{Pg} + G_l C_{Pl})}{K_e} \quad (14)$$

and

$$\rho = r/R \quad (15)$$

C. Two dimension model - two parameter model

More accuracy in predicting temperatures is possible by using a two parameter model. For this case K_e , the effective thermal conductivity, and h_w , the wall film coefficient, are used as parameters. The energy balance over a differential volume of the packed bed is identical to equation 9. However, the boundary conditions now become

$$\begin{aligned} t &= t_o \text{ at } z = 0 \\ \frac{\partial t}{\partial r} &= 0 \text{ at } r = 0 \\ h_w (t_w - t) &= K_e \left(\frac{\partial t}{\partial r} \right)_R \text{ at } r = R \end{aligned} \quad (16)$$

For a linear wall temperature, the analytical solution is

$$\begin{aligned} t(r, z) &= t_{wo} - cz - (ca/4) (1 - r^2/R^2 + 2/B_i) \\ &+ 2 Bi (t_o - t_{wo}) \sum_{n=1}^{\infty} \frac{J_0(B_n \rho) \exp(-B_n^2 z/a)}{(B_n^2 + Bi^2) J_0(B_n)} \\ &+ 2 Bi c a \sum_{n=1}^{\infty} \frac{J_0(B_n \rho) \exp(-B_n^2 z/a)}{B_n^2 (B_n^2 + Bi^2) J_0(B_n)} \end{aligned} \quad (17)$$

B_n is the root of the following equation:

$$B_n J_1(B_n) - Bi J_0(B_n) = 0 \quad (18)$$

and a is given by equation 14.

V. COMPUTER PROGRAM DEVELOPMENT (15)

A computer program called TEMP was written to analyze the cold model experiments. It is specific with respect to the geometry and number and location of thermocouples of the experimental cold model. With relatively few changes, it can be generalized to predict the effective conductivity and the wall film coefficient parameters. The main program and the sub-routines will be discussed briefly.

A. MAIN

MAIN controls the program and calls for the subroutines when they are needed. The theoretical temperatures as a function of radial position and height are calculated for a given K_e and h_w .

B. Subroutine BESSEL

Subroutine BESSEL calculates the zeroth Bessel function from the following equations:

$$J_0(B_n \rho) = \sum_{n=1}^{\infty} \frac{(-1)^{n+1} (1/2 B_n \rho)^{2n+2}}{(n+1)! \Gamma(n+2)} \quad (19)$$

and

$$J_0(B_n) = \sum_{n=1}^{\infty} \frac{(-1)^{n+1} (1/2 B_n)^{2n+2}}{(n+1)! \Gamma(n+2)} \quad (20)$$

C. Subroutine ROOT

Subroutine ROOT calculates the first five roots, B_n , of equation 18. The actual values of the roots were taken from Carslaw and Jaeger (17) and fitted with quadratic regression as a function of Bi, the Biot number. For Bi in the range of 1.0 to 4.0, the expression for the first root is:

$$B_n = 0.827023 + 0.496367 Bi + 0.0568818 Bi^2 \quad (21)$$

D. Subroutine OPTEMP

Subroutine OPTEMP is a minimization routine for the two parameters K_e and h_w . It is the search part of minimization. The procedure is based on the work of Nelder and Mead (18). It is an extension of the simplex methods which adapts itself to the local landscape using reflected, expanded, and contracted points to locate the minimum. This program (see reference 16) was chosen because derivatives were not required. Since the method assumes unimodality, different initial values of K_e and h_w may need to be tried.

E. Subroutine FUNC

Subroutine FUNC is the objective function for the search routine in OPTEMP. The function attempts to get the best match of experimental temperatures to calculated temperatures in a least squares sense. All the temperatures, at each radial and axial position, are used simultaneously in the objective function. With minor modifications, the program TEMP can compute the value of K_e and h_w at each axial position using all the radial temperatures. This type of calculation would give the variation of K_e and h_w as a function of height. It would indicate if a significant entrance length effect existed. Calculations at each height are now taking place. The objective function is given by the following equation:

$$F = \sum_{n=1}^{40} (t_{\text{exp}} - t_{\text{calc}})^2 \quad (22)$$

F. Subroutine WTEMP

Subroutine WTEMP uses linear regression to fit the preheater/reactor wall temperatures as a function of axial position. Its major output is an intercept wall temperature at the base of the preheater/reactor and the slope of the wall temperatures.

G. Subroutine RTEMP

Subroutine RTEMP is used to compute the average radial temperatures as a function of axial position using an area average:

$$\bar{T}(z) = \frac{2 \int_0^R r t(r,z) dr}{R^2} \quad (23)$$

H. Subroutine HEAT

Subroutine HEAT calculates the wall temperature of the preheater/reactor. The subroutine first calculates the coolant water bulk temperature. The overall heat transfer coefficient is determined between the coolant water bulk temperature and the preheater/reactor wall temperature using

$$U_i = 1 / \{ (r_o - r_i) A_i / k_{ss} A_{lm} + A_i / h_o A_o \} \quad (24)$$

The coolant water film coefficient is determined from the Dittus Boelter equation.

$$h = 0.023 \frac{k}{D_H} R_e^{0.8} P_r^{0.4} \quad (25)$$

The amount of heat transferred through the preheater/reactor wall is determined by the loss of heat the two phase flow suffers as it rises in the preheater/reactor. The heat transfer equation is simply:

$$Q = U_i A_i \{t_w(z) - t_{lw}(z)\} \quad (26)$$

Since everything is known in this equation except the wall temperature (t_w), the equation is used to calculate t_w . The calculation is repeated over each section of the preheater/reactor between each axial position of temperature measurement.

VI. RESULTS AND CONCLUSIONS

Figure 3 shows a typical radial temperature profile. The experimental temperatures are shown for constant bed height versus radial position. The solid lines are for the theoretical curve fit. As one can see, the curve matches the core temperatures very well and is fairly good near the wall. Figure 4 shows the axial center temperature profile, which agrees closely with the theoretical. Table 1 in the appendix presents a summary of the data. Water flow rates varied from 5,000 to 65,000 kg/m² hr and nitrogen flows from 0 to 15,000 kg/m² hr. Most of the experiments were conducted at 927 kPa (120 psig), which under coal liquefaction conditions corresponds to a gas density of approximately 27,600 kPa (4,000 psig). Some experiments were also conducted at 513 kPa (60 psig) and 306 kPa (30 psig). These gas densities correspond to 13,800 kPa (2,000 psig) and 6,900 kPa (1,000 psig), respectively.

The single-phase liquid K_e and H_e agree very well with the results reported by K. Hashimoto, et al. (12)^w and Weekman and Myers (2). The two-phase results are higher in this case than reported by K. Hashimoto, et al. and by Weekman and Myers. This is to be expected since the differences in packing shape and upflow versus downflow will be much more apparent in two-phase flow than in single-phase flow.

A. Variation with flow conditions

Figure 5 shows the variations of the effective thermal conductivity with gas mass velocities. The trends are the same for low and high liquid mass velocities. Initially, as the flow rate is increased, the effective thermal conductivity increases markedly, then decreases to about 70% - 90% of the maximum value. The trend for liquid flows from approximately 25,000 kg/m² hr to 30,000 kg/m² hr is different. It also increases initially, but upon reaching a maximum, the data becomes so scattered that any decrease is not obvious. A transition between two flow regimes seems to occur at this liquid flow rate. At this flow rate, the bed starts to shift and the settling rate is much faster than at other liquid mass velocities. Pressure fluctuations also change character in this region. Finally, D. Smith (14) has noted a flow transition at slightly less liquid velocity but for pressures near atmospheric.

B. Variation of K_e with liquid flows

Figure 6 shows a typical variation of K_e with liquid flow rate. The trend is nearly to the square root with respect to G_l throughout the range of study and for all the gas flow rates. The data, however, tend to be scattered in the region of 25,000-30,000 kg/m² hr. The dependence of the heat transfer properties on the fluid flow conditions for the region above a liquid flow rate of 30,000 kg/m² hr is probably different from the dependency below 25,000 kg/m² hr, but the quantity and the quality of the data above 30,000 kg/m² hr are not sufficient for a definitive answer.

C. Variation of H_w with gas flows

The effect of gas flow on the wall heat transfer coefficient, shown in figure 7, is different than on the effective thermal conductivity. Initially, the wall heat transfer coefficient increases with the addition of gas. Then additional gas causes no further increases. As with the effective thermal conductivity, the wall heat transfer coefficient shows more scatter in the liquid region of 25,000-30,000 kg/m² hr.

The variation of the wall heat transfer coefficient with liquid mass velocity (figure 8) is essentially the same as that of the effective thermal conductivity. It is nearly to the one-half power of the liquid mass velocity except for some scatter in the liquid region 25,000-30,000 kg/m² hr.

D. Variation of K_e and H_w with bed height

In the initial derivation, K_e and H_w were assumed to be constant throughout the bed. However, one can choose to perform the calculation on various lengths of the reactor. For instance, the calculation may be applied to the region from the entrance to the first thermocouple, then the entrance to the second thermocouple, etc. In this manner, one can obtain an indication of the variation of K_e and H_w as a function of bed height. Figures 9a and 9b show the variation of K_e and H_w for two general types of flow patterns. Within the accuracy of these experiments, K_e and H_w have the same bed height dependence with flows. There seem to be two basic types of behavior. For liquid flows less than 25,000 kg/m² hr, K_e or H_w will start low, increase, and then level off as the bed height increases. This type of behavior indicates a stagnant volume at the bottom of the reactor with the remainder of the bed fairly uniform. The other type of behavior which occurs for liquid flows above 30,000 kg/m² hr does not reach a constant K_e or H_w . This indicates extensive entrance and/or exit effects throughout the entire length of the bed.

E. Correlation of K_e and H_w

The effective thermal conductivity and the wall heat transfer coefficient were correlated for two-phase flow and 927 kPa (120 psig). These correlations are strictly empirical; and cannot be applied to very low gas flows. K_e was found to correlate to the formula:

$$K_e = 2,000 G_1^{0.53} G_n^{-0.11} \quad (27)$$

Figure 10 shows the predicted versus observed values for K_e . Considering the range of conditions, the correlations are fairly good. Equation 27 correlates two-phase flow very well for high gas mass velocities and is

generally within 20% for lower gas mass velocities. H_w was found to correlate to the formula:

$$H_w = 14,000 G_l^{0.62} G_n^{0.04} \quad (28)$$

Figure 11 shows the predicted versus the observed values for H_w . The correlation for H_w indicates an almost negligible dependence of H_w on the gas flow rate. The correlation is not as good as that of K_e , but generally predicts H_w to within 20% over the entire range of interest. While the correlations were developed with only 927 kPa (120 psig) data, the correlations seem to agree well with the 513 kPa (60 psig) and 206 kPa (30 psig) experiments. There seems to be little, if any, dependence of K_e and H_w on pressure in the range of this study.

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² Research Equipment Operator

NOMENCLATURE

A_i = Inside area of pipe/unit length

A_o = Outside area of pipe/unit length

A_{em} = Log mean area of pipe/unit length

$a = (R/2 H_w) (G_g C_{pg} + G_e C_{pe})$

$ai = R^2 (G_g C_{pg} + G_e C_{pe}) K\theta$

Bi = Biot number = $H_w R/K_e$

c = absolute value of slope of the wall temperature

C_p = Heat capacity (Joules/Kg ° C)

D_h = Hydraulic diameter = $4 \left\{ \frac{\text{Area of stream cross section}}{\text{wetted perimeter}} \right\}$

F = Objective function

G = Superficial mass velocity based on the empty tube (Kg/m² hr)

hr = Hour

H_w = Wall heat transfer coefficient (Joules/m² hr ° C)

J_o = Zero order Bessel function of the first kind

K = Thermal conductivity (Joules/m hr ° C)

K_e = Effective thermal conductivity (Joules/m hr ° C)

Pr = prandtl number = $C_p \mu / K$

Q = Heat transferred through the reactor wall (Joules)

R = Radius of bed (m)

r = Radial distance from center of bed (m)

R_e = Reynolds number = $G 2R/\mu$

S = Cross-sectional area (m²)

T = Average radial temperature (° C)

t = temperature (° C)

- u = Exponential constant which describes the wall temperature variation
 U = Overall heat transfer coefficient ($\text{Joules/m}^2 \text{ hr } ^\circ \text{C}$)
 z = Bed height (m)

Subscripts

- g = Gas
 l = Liquid
 n = Nitrogen
 ss = Stainless steel
 w = Wall
 w_o = Wall at entrance

Greek Symbols

- μ = Viscosity (Kg/m hr)
 ρ = r/R

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Figure 1- Heat transfer study system schematic.

Figure 2 — Heat transfer reactor model

12-6-78 L-16350

Figure 3 — Typical radial temperature profile

12-6-78 L-16351

Figure 4 -Axial temperature profile for center of reactor model

12-6-78 L-16352

Figure 5 - Typical variation of the effective thermal conductivity with gas mass velocities at constant liquid mass velocity.

12-6-78 L-16353

Figure 6- Typical variation of the effective thermal conductivity with liquid mass velocity

12-6-78 L-16354

Figure 7 - Typical variation of the wall heat transfer with gas mass velocity at constant liquid flow rate.

12-6-78 L-16355A

Figure 8 — Typical variation of the wall heat transfer coefficient with liquid mass velocity

12-6-78 L-16356A

Figure 9 -Entrance effects on K_e (a) and H_w (b)

12-6-78 L-16357

Figure 10 - The predicted versus observed values of the effectivity thermal conductivity for the correlation $K_e = 1984 G_g^{.53} G_n^{-.11}$

12-6-78 L-16358

Figure II - Predicted versus observed values of the wall heat transfer coefficient for the correlation $H_w = 14,000 G_L^{.62} G_n^{.042}$

12-6-78 L-16359

APPENDIX

TABLE I.- Results of Heat Transfer Studies

Run Number	Pressure K Pa	Inlet Temp ° C	Exit Temp ° C	Liquid Mass Velocity Kg/m ² hr	Gas Mass Velocity Kg/m ² hr	Effective Thermal Conductivity Joules/m hr ° C (X10 ⁻⁵)	Wall Heat Transfer Coefficient Joules/m ² hr ° C (X10 ⁻⁷)
RS-1	927*	58.0	14.3	35,000	7,500	1.75	1.16
RS-2	do.	58.7	15.7	25,000	7,500	1.75	1.24
RS-3	do.	58.7	16.1	20,000	7,500	1.49	1.03
RS-4	do.	53.2	17.8	42,100	7,500	1.99	1.21
RS-5	do.	57.1	11.1	15,000	7,500	1.07	.697
RS-6	do.	63.0	9.7	15,000	7,500	1.25	.931
RS-7	do.	67.9	7.8	10,000	7,500	.868	.618
RS-8	do.	55.0	25.2	45,000	7,500	1.72	.770
RS-9	do.	55.5	18.2	35,000	7,500	1.66	.911
RS-10	do.	58.9	21.9	25,000	7,500	1.28	.788
RS-11	do.	56.3	11.2	15,000	7,500	1.05	.701
RS-12	do.	62.0	9.5	10,000	7,500	.961	.657
RS-13	do.	52.6	11.7	20,000	7,500	1.29	.841
RS-14	do.	66.0	9.3	5,000	7,500	.688	.479
RS-15	do.	59.5	12.8	10,000	5,000	.991	.696
RS-16	do.	65.7	9.0	10,000	7,500	.863	.571
RS-17	do.	62.0	8.5	10,000	10,000	.988	.654
RS-18	do.	61.2	9.6	10,000	12,500	.896	.573
RS-19	do.	64.2	15.5	10,000	2,500	1.32	.904
RS-20	do.	61.7	14.5	15,000	2,500	1.62	1.18
RS-21	do.	60.3	13.5	15,000	5,000	1.31	.876
RS-22	do.	54.8	10.2	15,000	10,000	1.16	.655
RS-23	do.	57.3	10.0	15,000	1,500	1.44	1.03
RS-24	do.	58.4	9.8	10,000	1,500	1.18	.837
RS-25	do.	55.0	11.8	20,000	1,500	1.48	1.00
RS-26	do.	57.5	17.8	20,000	2,500	1.81	1.30
RS-27	do.	57.5	16.8	15,000	-	.631	.467
RS-28	do.	59.1	14.8	10,000	-	.498	.423
RS-29	do.	57.5	9.5	15,000	12,500	1.11	.740
RS-30	do.	62.3	10.3	15,000	15,000	1.09	.801

*(120 psig)

TABLE I.- Results of Heat Transfer Studies-Continued

Run Number	Pres- sure K Pa	Inlet Temp ° C	Exit Temp ° C	Liquid Mass Velocity Kg/m ² hr	Gas Mass Velocity Kg/m ² hr	Effective Thermal Conductivity Joules/m hr ° C (X10 ⁻⁵)	Wall Heat Transfer Coefficient Joules/m ² hr ° C (X10 ⁻⁷)
RS-31	do.	65.0	9.5	10,000	15,000	.918	.630
RS-32	do.	56.0	16.5	20,000	-	.704	.482
RS-33	do.	62.5	21.0	20,000	-	.727	.520
RS-34	do.	61.0	13.0	20,000	5,000	1.44	.905
RS-35	do.	59.5	23.5	20,000	10,000	1.36	.799
RS-36	do.	61.0	11.5	20,000	12,500	1.38	.748
RS-37	do.	60.0	11.5	20,000	15,000	1.32	.748
RS-38	do.	57.5	16.5	25,000	-	.822	.499
RS-39	do.	53.0	18.5	35,000	-	1.02	.538
RS-40	do.	56.0	24.0	45,000	-	1.25	.583
RS-41	do.	62.0	20.0	45,000	1,500	2.38	1.16
RS-42	do.	59.0	13.0	25,000	1,500	1.52	.801
RS-43	do.	63.5	16.5	35,000	1,500	2.13	1.20
RS-44	do.	62.5	12.5	25,000	2,500	1.72	1.19
RS-45	do.	62.5	14.0	25,000	5,000	1.48	.940
RS-46	do.	61.5	14.5	25,000	7,500	1.43	.831
RS-47	do.	63.0	21.0	25,000	10,000	1.54	1.06
RS-48	do.	61.0	14.5	25,000	12,500	1.65	.752
RS-49	do.	63.0	17.0	30,000	2,500	1.69	.959
RS-50	do.	61.0	16.5	30,000	5,000	1.66	1.00
RS-51	do.	58.5	16.5	30,000	7,500	1.61	.990
RS-52	do.	60.5	16.5	30,000	10,000	1.56	1.22
RS-53	do.	60.0	18.0	35,000	2,500	2.25	1.08
RS-54	do.	57.0	17.5	35,000	5,000	1.94	1.06
RS-55	do.	57.0	23.0	30,000	10,000	1.75	.908
RS-56	do.	55.0	23.0	35,000	10,000	1.74	1.31
RS-57	do.	62.5	13.5	25,000	15,000	1.54	1.08
RS-58	do.	62.5	17.5	30,000	15,000	1.44	1.10
RS-59	do.	62.5	15.5	30,000	1,500	1.83	.921
RS-60	do.	58.5	17.0	30,000	12,500	1.57	1.46

TABLE I.- Results of Heat Transfer Studies-Continued

Run Number	Pres- sure K Pa	Inlet Temp ° C	Exit Temp ° C	Liquid Mass Velocity Kg/m ² hr	Gas Mass Velocity Kg/m ² hr	Effective Thermal Conductivity Joules/m hr ° C (X10 ⁻⁵)	Wall Heat Transfer Coefficient Joules/m ² hr ° C (X10 ⁻⁷)
RS-61	do.	56.0	18.5	35,000	12,500	1.70	1.42
RS-62	do.	59.0	29.0	45,000	2,500	1.78	.930
RS-63	do.	57.5	28.5	45,000	5,000	1.81	1.01
RS-64	do.	54.0	23.5	45,000	7,500	1.81	1.23
RS-65	do.	52.5	22.5	45,000	10,000	1.98	1.31
RS-66	do.	58.0	22.5	45,000	12,500	2.47	1.19
RS-67	do.	59.0	20.5	35,000	12,500	2.12	1.54
RS-68	do.	57.5	21.0	35,000	15,000	1.80	1.38
RS-70	do.	56.0	27.0	55,000	1,500	2.73	1.10
RS-71	do.	59.0	32.0	55,000	-	2.37	.983
RS-72	do.	57.0	34.0	65,000	-	2.83	1.38
RS-73	do.	57.5	28.0	55,000	8,000	2.73	2.18
RS-74	do.	61.5	30.0	55,000	5,000	2.40	1.12
RS-75	do.	62.0	24.0	55,000	7,500	2.15	1.82
RS-76	do.	63.1	26.0	55,000	10,000	2.25	1.93
RS-77	do.	61.0	26.5	65,000	10,000	2.70	1.96
RS-78	do.	60.5	32.0	55,000	-	2.02	.723
RS-79	do.	58.5	32.0	65,000	-	2.22	.846
RS-80	do.	59.0	27.5	55,000	12,500	1.94	1.63
RS-81	do.	62.5	30.0	55,000	2,500	3.05	1.65
RS-82	do.	58.0	28.5	65,000	1,500	3.00	1.71
RS-83	do.	58.0	29.5	55,000	-	2.11	.835
RS-84	do.	58.5	28.0	65,000	-	2.51	1.21
RS-85	do.	57.5	25.0	65,000	2,500	3.37	2.26
RS-86	do.	61.5	27.5	65,000	5,000	3.17	2.27
RS-87	do.	57.5	27.5	65,000	7,500	2.58	2.20
RS-88	do.	55.5	27.0	65,000	10,000	2.68	2.30
RS-89	do.	55.5	27.5	65,000	7,500	2.40	2.37
RS-90	do.	58.0	25.0	45,000	1,500	2.79	1.33

TABLE I.- Results of Heat Transfer Studies-Continued

Run Number	Pres- sure K Pa	Inlet Temp ° C	Exit Temp ° C	Liquid Mass Velocity Kg/m ² hr	Gas Mass Velocity Kg/m ² hr	Effective Thermal Conductivity Joules/m hr ° C (X10 ⁻⁵)	Wall Heat Transfer Coefficient Joules/m ² hr ° C (X10 ⁻⁷)
RS-91	do.	56.5	24.0	45,000	15,000	2.00	1.52
RS-92	do.	64.0	20.2	5,000	1,500	1.01	.479
RS-93	do.	65.5	19.5	5,000	2,500	.985	.474
RS-94	do.	63.0	19.0	5,000	5,000	.715	.474
RS-95	do.	62.5	21.5	5,000	7,500	.649	.424
RS-96	do.	65.5	19.0	5,000	10,000	.638	.419
RS-97	do.	68.0	20.0	5,000	-	.280	.243
RS-98	do.	63.0	21.5	5,000	12,500	.591	.331
RS-99	do.	70.5	20.0	5,000	15,000	.609	.358
RS-100	do.	63.0	20.6	20,000	1,000	1.79	1.11
RS-101	do.	65.0	25.0	20,000	750	1.85	1.23
LP-1	514**	60.0	21.5	20,000	-	.718	.555
LP-2	do.	56.0	19.5	20,000	5,000	1.59	1.08
LP-3	do.	62.0	20.0	20,000	1,500	2.17	1.54
LP-4	do.	61.0	20.5	20,000	2,500	1.98	1.44
LP-5	do.	62.0	20.5	20,000	7,500	1.53	1.09
LP-6	do.	57.5	23.0	20,000	10,000	1.47	1.00
LP-7	do.	59.0	21.5	20,000	12,500	1.50	1.00
LP-8	do.	58.5	20.5	20,000	15,000	1.45	.830
LP-9	do.	62.0	17.5	15,000	5,000	1.39	1.02
LP-10	do.	65.5	16.5	10,000	5,000	1.14	.819
LP-14	do.	58.0	18.5	20,000	1,000	2.33	1.70
LP-15	do.	60.0	22.0	20,000	1,000	2.17	1.51
LP-18	do.	64.0	20.5	20,000	750	2.17	1.50
LP-20	do.	63.0	20.5	20,000	500	2.00	1.32
LP-11	207***	56.2	16.0	20,000	5,000	1.53	1.10
LP-12	do.	58.5	18.0	20,000	7,500	1.51	1.09
LP-13	do.	59.5	18.5	15,000	5,000	1.23	.890
LP-16	do.	63.0	19.0	20,000	1,000	2.00	1.34
LP-17	do.	60.0	17.0	20,000	750	2.10	1.38
LP-19	do.	61.0	27.0	20,000	1,500	2.00	1.35
LP-21	do.	62.0	21.0	20,000	500	2.21	1.50

**(60 psig)

*** (30 psig)