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Heat transfer coefficients in an agitated vessel using verticle-tube baffles

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HEAT TRANSFER COEFFICIENTS IN AN AGITATED VESSEL USING VERTICLE-TUBE BAFFLES

BY

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ABSTRACT

Data are presented to show the effect of impeller speed, impeller type, and fluid properties on the forced-convection film coefficient of heat transfer for verticle tubes in a cylindrical vessel. A generalized equation to predict the film coefficient of heat transfer on the outside of verticle tubes in an agitated vessel for both cooling and heating is also presented. This equation may be used to estimate the necessary areas of heat transfer for verticle tubes in an agitated vessel within the degree of accuracy required from most engineering calculations.

This study was the first initiated to develop an equation to predict film coefficients of heat transfer for verticle tubes in an agitated vessel from fluid properties, agitator speed and system dimensions. Dunlop and Rushton (6) have developed a generalized equation to predict film coefficients of heat transfer for verticle tubes in an agitated vessel since this study was initiated. The data of this thesis do not fit the equation proposed by Dunlop and Rushton. This is probably due to the difference in system dimensions and dissimilarity between their equipment and the equipment of this thesis.

In the course of this study it was further determined that for a well baffled system both radial flow and axial flow impellers have the same effect on the film coefficient.

Previous investigators have proposed that the exponent of the viscosity ratio was constant. It was determined in the work of this thesis that this is not true and that the ratio is a function of the fluid properties.

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I. INTRODUCTION

It is often necessary to transfer heat to or from a fluid during mixing operations. Two types of heat transfer surface are in common use. The jacketed vessel, which provides a heat transfer surface around the periphery, and the immersed coil, which is either in the form of a helix or pancake.

The physical dimensions of the heat transfer surface and its location with respect to the mixing unit appreciably affect the rate of heat transfer. To obtain an optimum heat transfer arrangement, sacrifices have to be made in the mixing efficiency.

To obtain the best mixing conditions, it is essential that the flow pattern in the tank be made up of both horizontal and verticle lines of flow. This can be accomplished by the use of baffles. When baffles are placed in jacketed vessels, the high peripheral velocity is decreased with a resultant decrease in heat transfer. Thus for jacketed vessels, it is impossible to obtain optimum heat transfer when obtaining the best mixing conditions.

When helical coils are used, they act partially like a baffle and promote a verticle flow component. In the space between the coil and the tank wall a relative motionless area is likely to result. In this area very little mixing takes place. Due to the high velocity component on the side of the coil facing the mixing unit, good heat transfer is obtained.

The use of verticle heat transfer tubes in a cylindrical mixing vessel produces good fluid flow patterns for mixing as well as excellent heat transfer. The verticle tubes act as baffles as well as providing the necessary heat transfer surface. This type of arrangement circumvents the disadvantage of the jacketed vessel or helical coil.

A. Purpose

The purpose of this thesis was to determine the effect of impeller speed, impeller type, and fluid properties on the forced-convection film coefficient of heat transfer for verticle tubes in a cylindrical agitated vessel. A generalized correlation of the film coefficients with fluid properties and system dimensions was also undertaken, so that the necessary areas of heat transfer surface could be calculated for dimensionally similar systems.

B. Review of Literature

Gordon (1) in 1941, studied the effect of agitation on the film coefficients of heat transfer for a jacketed kettle. Four different liquids, water and three hydrocarbon oils were used. Only heating coefficients were obtained. An attempt to correlate the data in dimensionaless terms was made, but the units chosen were not entirely dimensionaless.

Data for the film coefficients of heat transfer in a kettle with a jacket and helical coil are given by Chilton, Drew, and Jebens (2). A flat-bladed agitator was used in this study. Water, 92% glycerine solution, and two hydrocarbon oils were used as the bulk fluids. Heating film coefficients were obtained on the jacket and cooling film coefficients on the coil. The following equations were derived from this experimental work:

for jacket heating film coefficients;

$$(1)\quad N_{Nu} = 0.36\, N_{Pr}^{1/3}\, N_{Re}^{2/3} \left(\mu_j/\mu\right)^{-0.14}$$

for the coil cooling film coefficients;

$$(2)\quad N_{Nu} = 0.87\, N_{Pr}^{1/3}\, N_{Re}^{0.62} \left(\mu_c/\mu\right)^{-0.14}$$

See Section XI for the nomenclature applying to the above and following equations.

Rushton, Lichtmann, and Mahony (3) in 1948 were the first to investigate verticle heating and cooling banks as a means of transferring heat in agitated vessels. Film coefficients of heat transfer for verticle tubes acting as baffles in addition to heat exchange surface are presented. A four foot diameter cylindrical tank was used as well as two flat bladed mixing turbines, one 12 inches in diameter and the other 16 inches in diameter. The only liquid used was water. It was determined that the optimum position of the impeller for maximum heat transfer was midway between the bottom of the tank and the surface of the bulk liquid. The film coefficients were correlated against the Reynolds number. The following equations were obtained:

For the 16-inch, 6-blade turbine:

$$(3)\ h_{heat} = 0.00285\ L^2 N \rho \mu^{-1}$$

$$(4)\ h_{cool} = 0.00265\ L^2 N \rho \mu^{-1}$$

For the 12-inch, 4-blade turbine:

$$(5)\ h_{heat} = 0.00235 \left(L^2 N \rho / \mu\right)^{0.7}$$

$$(6)\ h_{cool} = 0.00220 \left(L^2 N \rho / \mu\right)^{0.7}$$

Cummings and West (4) have made a study of both jacket and coil film coefficients of heat transfer in an agitated vessel. A jacketed 30-inch diameter, stainless steel dished kettle was used. A helical coil was immersed for cooling, such that the system could be brought to a steady state condition. Six liquids of widely differing physical and thermal properties were used. In addition, data were obtained on two phase systems, two liquid-solid systems and two liquid-liquid systems. No attempt was made to correlate the data from these systems.

Two different types of agitators were used, a retreating blade turbine and a 45° pitched-blade turbine. It was found that when the pitched-blade turbine was used, the film coefficients were on the order of 10% lower then those for the retreating blade turbine. The work of Chilton et al (2), the work of Gordon (1), and the work of Rushton et al (3) were recalculated and correlated with their data. The following generalized correlations for the combined data were reported:

For cooling fluids in a vessel with a single helical coil as the heat transfer surface;

$$(7)\quad \frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{C\mu}{K}\right)^{1/3}} = 1.01\left(L^2 N \rho/\mu\right)^{0.62}$$

For heating fluids in a vessel with a jacket as the heat transfer surface;

$$(8)\quad \frac{\left(\frac{h_j D_K}{K}\right)\left(\frac{\mu_j}{\mu}\right)^{0.14}}{\left(\frac{C\mu}{K}\right)^{1/3}} = 0.40\left(L^2 N \rho/\mu\right)^{2/3}$$

Carrol (5) in 1952 made a series of runs in a one-foot jacketed flat bottomed kettle. An internal helical copper coil was used to obtain steady state conditions. Two different flat bladed agitators were used. Water and glycerine solutions were used as the bulk fluids.

Chilton et al.'s (2) method of correlation of the film coefficients was used. The equations presented are as follows:

For jacket heating:

$$(9)\quad N_{Nu} = 0.6\, N_{Pr}^{1/3}\, N_{Re}^{2/3} \left(\mu_j/\mu\right)^{-0.14}$$

For cooling coil:

$$(10)\quad N_{Nu} = 1.5\, N_{Pr}^{1/3}\, N_{Re}^{0.62} \left(\mu_c/\mu\right)^{-0.14}$$

Dunlap and Rushton (6) in 1953 reported data for individual film coefficients on the mixed-liquid side of heating and cooling tubes. The tubes were placed in a verticle position and acted as baffles as well as heat transfer surface. Three fluids were used, water and two hydrocarbon oils. Data were also obtained to show the temperature and film coefficient distribution over the entire area of the heat transfer surface. Two dimensionally similar systems were studied to check the geometric similitude of the heat-transfer characteristics.

The method of Chilton et al (2) was used to correlate the data. To make the correlation more general, the dimensions of the equipment were incorporated. The authors used the diameter of a tube in a baffle as the length dimension in the Nusselt number. In previous studies of this type, the tank diameter was used. The following relation was obtained.

$$(11)\ \frac{h_c d_o}{K} = 0.09\left(\frac{L}{D_K}\right)^{0.33}\left(\frac{R}{B}\right)^{0.2}\left(\frac{\mu}{\mu_f}\right)^{0.4}\left(\frac{L^2 N \rho}{\mu}\right)^{0.65}\left(\frac{c\mu}{K}\right)^{1/3}$$

The viscosity of the film (μ_f) was determined at its mean temperature. In previous works by other authors, the viscosity of the film was determined at the temperature of the tube wall. This would account for the difference in exponents for the viscosity ratio.

The authors discuss the effect of natural convection on the data and show that Reynolds number cannot be used as a criterion for relative liquid agitation.

Oldshue and Gretton (7) in 1954 measured film coefficients of heat transfer for helical coils in a mixing vessel when using flat-bladed turbine impellers. A four-foot diameter tank was used and the coils so arranged as to provide steady and unsteady-state operations. Data were obtained for baffled conditions. Among the variables investigated were impeller speed, power, ratio of impeller size to tank size, baffle position and tube diameter.

The mean heat-transfer coefficient (ho) at $\mu_f/\mu = 1.0$ was correlated against the other variables and the following relation was obtained:

$$(12)\quad \frac{h_c d}{K} = 0.17\left(\frac{L^2 N \rho}{\mu}\right)^{0.67}\left(\frac{C\mu}{K}\right)^{0.37}\left(\frac{L}{D_K}\right)^{0.1}\left(\frac{d}{D_K}\right)^{0.5}$$

The authors propose that the exponent of the ratio of film viscosity to the bulk viscosity is not constant under varying conditions. The two possible causes are:

a. The value of (e) is a function of fluid properties such as μ, k and c.

b. The value of (e) is dependent upon the value of (μ_f/μ) and thus on the heat flux through the film.

It was further determined that the baffle position had relatively little effect on the heat transfer coefficient.

Pursell (8) has collected all previous published data and with his own data attempted to correlate in general terms the heat transfer on the walls of a jacketed kettle agitated by a paddle type impeller. The following relationship was reported:

$$(13)\quad \frac{h D_K}{K} = 0.112\left(\frac{C\mu}{K}\right)^{0.44}\left(\frac{L^2 N \rho}{\mu}\right)^{0.75}\left(\frac{\mu}{\mu_j}\right)^{0.25}\left(\frac{D_K}{L}\right)^{0.40}\left(\frac{L_W}{L}\right)^{0.13}$$

II. THEORY

A. General

Mixing is usually the most basic step in any processing sequence, but relatively little is known about this unit operation. Quillen (9) has proposed the following definitions for agitation and mixing:

> "Agitation is the creation of a state of activity, motion, or turbulence, apart from any mixing accomplished."

and,

> "Mixing can be defined as the intermingling of two or more dissimilar portions of material, resulting in the attainment of a desired level of uniformity, either physical or chemical, in the final product.

In fluid mixing, the intermingling of one component in the other is thought to be accomplished by two mechanisms, diffusion and fluid motion. Three distinct regions of intermingling have been postulated. They are:

a. Radom mixing of the masses in the main body of the fluid in a state of complete agitation.

b. Molecular diffusion in the laminar flow area adjacent to a wall. Within this film there is no radial mass flow or mixing, and the transfer of momentum occurs by the slow process of diffusion.

c. Transitional region, where both turbulent mixing and molecular diffusion occur.

Turbulence or agitation usually occurs by one or more of the following three ways; skin friction, form separation, or fluid discontinuity. Skin friction and form separation occur when the liquid medium comes into contact with a solid surface, such as a baffle or wall. Fluid discontinuity occurs when two masses of fluid come into contact with each other at different velocities or from different directions.

The rotating impeller or agitator imparts to the liquid this turbulence. The paddle type impeller will cause a minimum of displacement or pumping action and in an unbaffled tank will impart a swirling motion to the liquid. The rotational speed of the swirling liquid approaches that of the paddle and very poor mixing is obtained due to the absence of turbulence. Therefore, for effective mixing with radial flow type impellers, baffles are a necessity.

There are other types of impellers loosely defined as turbines. They are usually bladed and produce a radial flow component, as in the paddle agitator, or a combination of radial flow with axial flow.

The forced-convection film coefficients of heat transfer are a function of the liquid turbulence and the origin of the turbulence. When helical coils are used there is little resistance to fluid rotation, and a swirling motion is imparted to the fluid. In an unbaffled tank this could result in the formation of a vortex at high agitator speeds. When the proper baffling is present, the tendency for a vortex to form is decreased. Furthermore, turbulence is increased due to fluid discontinuity and form separation as discussed previously. Thus, the presence of baffles promote an axial component of flow, which is desired for good mixing, and furthermore results in good heat transfer.

B. Heat Transfer Equations

The basic equations for conductive and convective heat transfer were used to obtain the forced-convection heat transfer coefficients on the outside of the verticle tubes. The amount of heat transferred from the bank of tubes was calculated from the following well known equation.

$$(14) \quad Q = Wc\Delta t$$

The overall coefficient (U) was then determined from the following equation:

$$(15) \quad U = \frac{Q}{A\Delta t_m}$$

The film coefficient on the outside of the tubes was then calculated from the overall coefficient, the resistance of the pipe wall, and the film coefficient for the water flowing inside the verticle tubes. Since clean pipe was used, the dirt film resistance was neglected. The inside film coefficients were calculated from the well known Dittus-Boelter equation (11) which follows:

$$(16)\ h_i = 0.0225 \frac{K}{d_i} \left(\frac{d_i V \rho}{\mu}\right)^{0.8} \left(\frac{C \mu}{K}\right)^{0.4}$$

The exponent of the Prandtl number is 0.3 when the fluid is being cooled.

The outside film coefficients were then calculated from the following equation:

$$(17)\ \frac{1}{h_c} = \frac{1}{U} - \left(\frac{1}{h_i} + \frac{1}{r_w}\right)$$

C. Correlation of Dimensionaless Groups

Heat transfer defies exact mathematical analysis because of the large number of variables and lack of true understanding of the exact mechanism by which heat is transferred. The only recourse is to combine variables in dimensionaless groups, using the concepts of similarity.

The method of Chilton et al (2) was used to group the variables in dimensionaless form. The turbine speed and turbine diameter were incorporated in a mixing Reynolds number $\left(\frac{L^2 N \rho}{\mu}\right)$. The film coefficient of heat transfer was included in the Nusselt number ($\frac{h D_K}{K}$) and the physical properties of the bulk liquid in a Prandtl number ($\frac{c\mu}{K}$).

The following equation has been found to hold for most forced-convection heat transfer systems and this was used to relate the various dimensionless groups.

$$(18) \quad N_{Nu} = X\left(N_{Re}\right)^{a}\left(N_{Pr}\right)^{b}$$

The exponent of the Reynolds number and Prandlt number were obtained by a series of cross plots of the data. When the values of the exponents and coefficients were determined in equation (18), it was found that two equations were obtained, one for the heating baffle and one for the cooling baffle. Since the liquid turbulence is the same around both baffles, the deviation could only be explained by a film property. The major difference in the films around the cooling and heating baffles is the film viscosities. The ratio of the viscosity of the film at the wall temperature to the viscosity of the bulk liquid was introduced into the general equation to obtain one correlation for heating as well as cooling. The generalized equation that was used in correlating all the data of this study is as follows:

$$(19) \quad N_{Nu} = X_1\left(N_{Re}\right)^{a}\left(N_{Pr}\right)^{b}\left(\mu_c/\mu\right)^{e}$$

This equation holds for both cooling and heating film coefficients.

III. EQUIPMENT

A. Discription of Apparatus

The agitating tank is a cylindrical, flat-bottomed tank, 2 feet in diameter and 3 feet high. The agitators were mounted on a shaft which was connected to a ½ H.P. motor. The agitator shaft was located on the verticle center line of the tank. The heating and cooling pipe baffles were made of galvanized iron pipe, ½ inch standard iron pipe size. Four lengths of pipe were connected with standard ell and tee fittings to form one baffle, and were arranged so that water would be able to pass internally through the pipes in series. Each bank of pipe was placed so that it extended seven inches towards the center of the tank, with a 5/8 inch clearance between the tank wall and the heating pipe nearest it.

Thermometer wells of copper tubing were provided for temperature measurements of the incoming and outgoing water. The thermometer wells were situated on top of tee fittings at the entrance and exit of each bank of pipes. Thermometers with a 1° F calibration were used to take all the temperature measurements. The bulk temperature of the liquid in the tank was obtained by suspending a 1° F graduated thermometer at a point midway between the outer edge of the agitator and the tank wall. Rushton, Lichtmann, and Mahoney (3) found that no matter where this thermometer was placed, the reading of the bulk temperature was constant for a given steady state condition. For all the runs a liquid depth of 2 feet was used.

Two different types of agitators of the same diameter, 8 inches, were used. Figures 1 and 2 give detailed information on the agitators. Agitator speeds were varied by using a series of different diameter pulleys attached to the agitator shaft. The speeds were measured by a tachometer, and were varied between 10 and 212 RPM. In all runs the agitator was immersed half way in the fluid.

B. Overall Arrangement

Hot water was obtained by the direct injection of steam into the hot water tank. Hot water was continously circulated between the heating banks and the hot water tank by a centrifugal pump. Hot water off the discharge of the pump was by-passed back to the hot water tank for temperature and flow control.

Hot water was sent to the two baffles located opposite each other. The cold water, supplied from the city main, was sent to the other two sets of baffles. Water entered each bank at the length of pipe closest to the center of the tank. Only the total flow of water going to each set of baffles was measured by calibrated orifices.

The flow through each individual bank was regulated by valves so that identical exit water temperatures were obtained in both heating banks and both cooling banks. By obtaining identical exit temperatures for both heating banks, the flow to each bank was maintained at the same rate. The same situation existed with the cold water banks. Equal outlet temperatures were maintained to insure equal flow rates.

Equipment arrangement and detailed diagrams of the apparatus are attached in Figures, 3, 4 and 5.

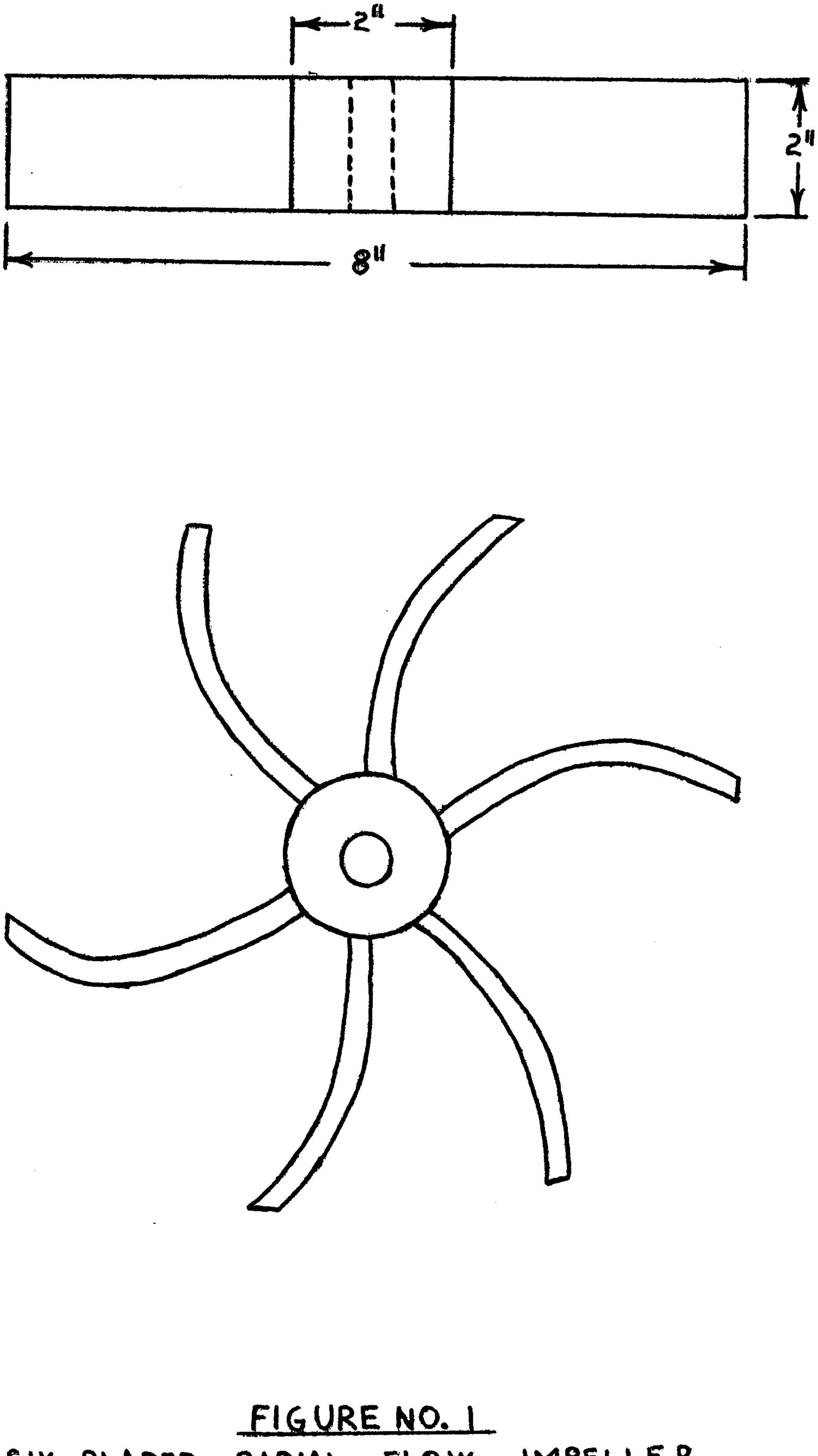


FIGURE NO. 1

SIX BLADED RADIAL FLOW IMPELLER

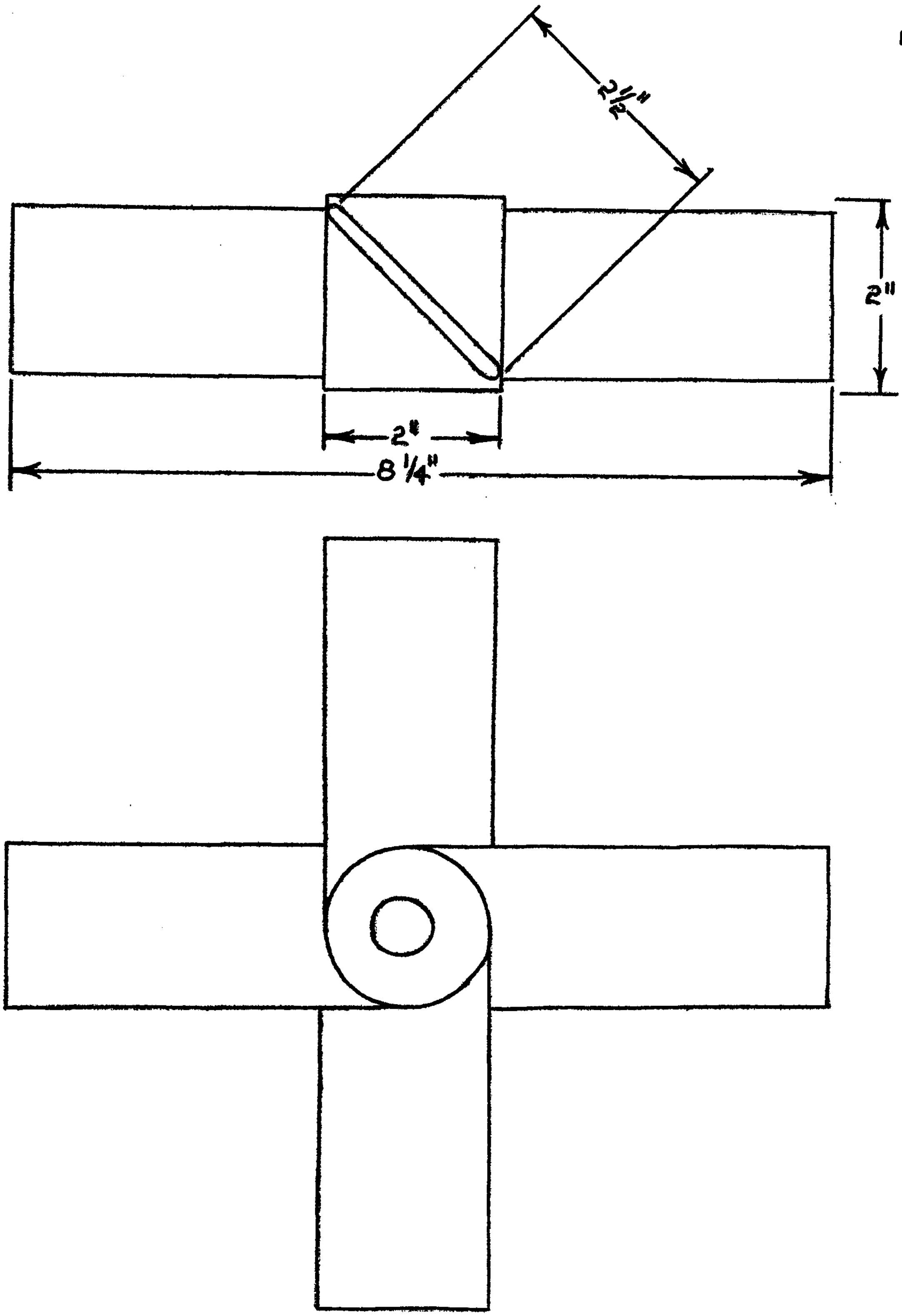


FIGURE NO. 2

FOUR BLADED T-TYPE IMPELLER

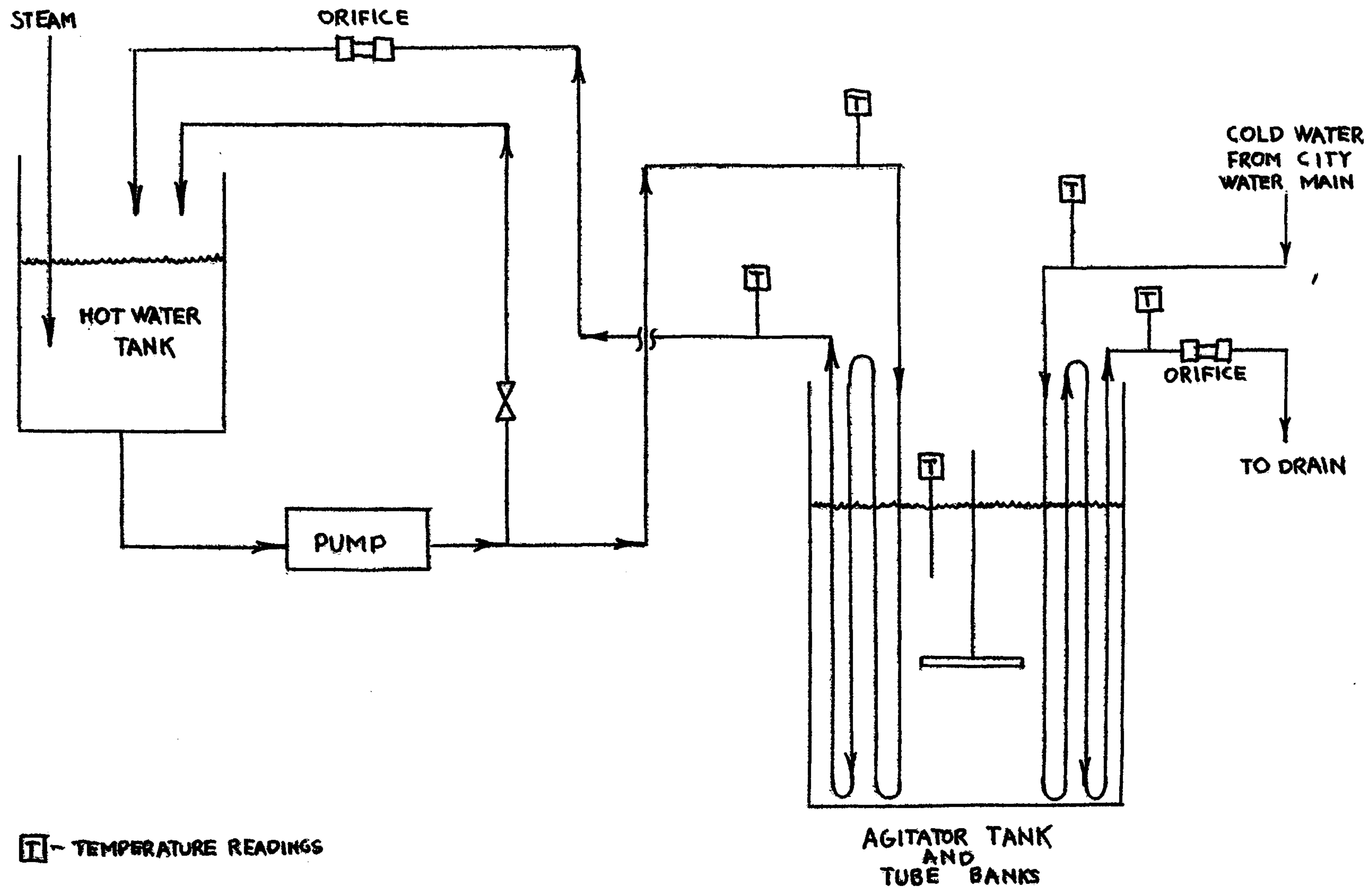


FIGURE 3
OVERALL ARRANGEMENT OF EQUIPMENT

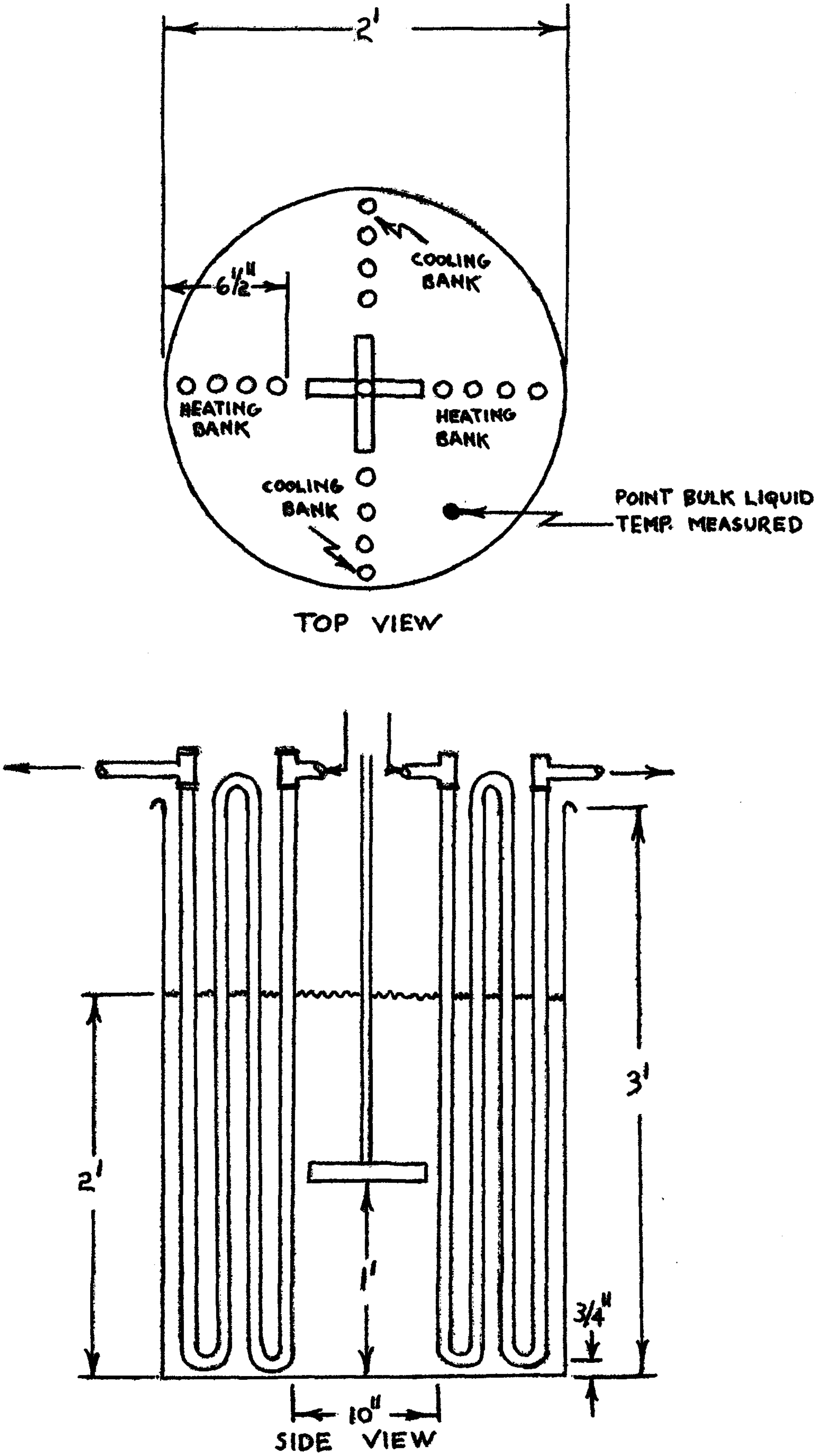
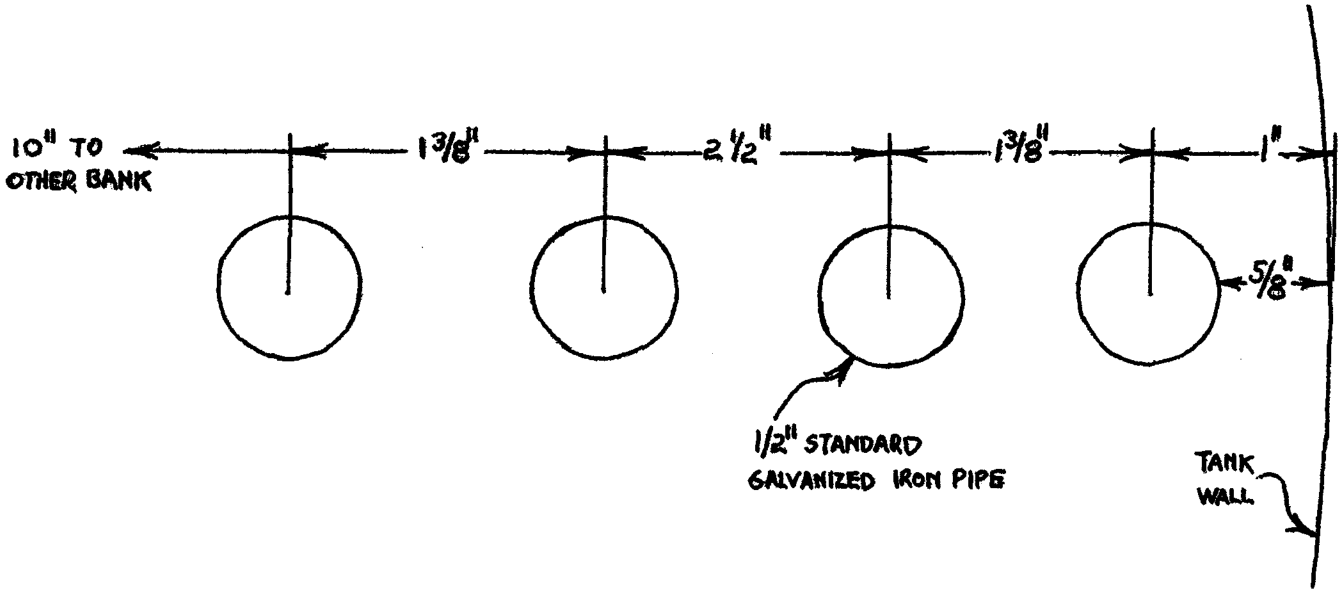
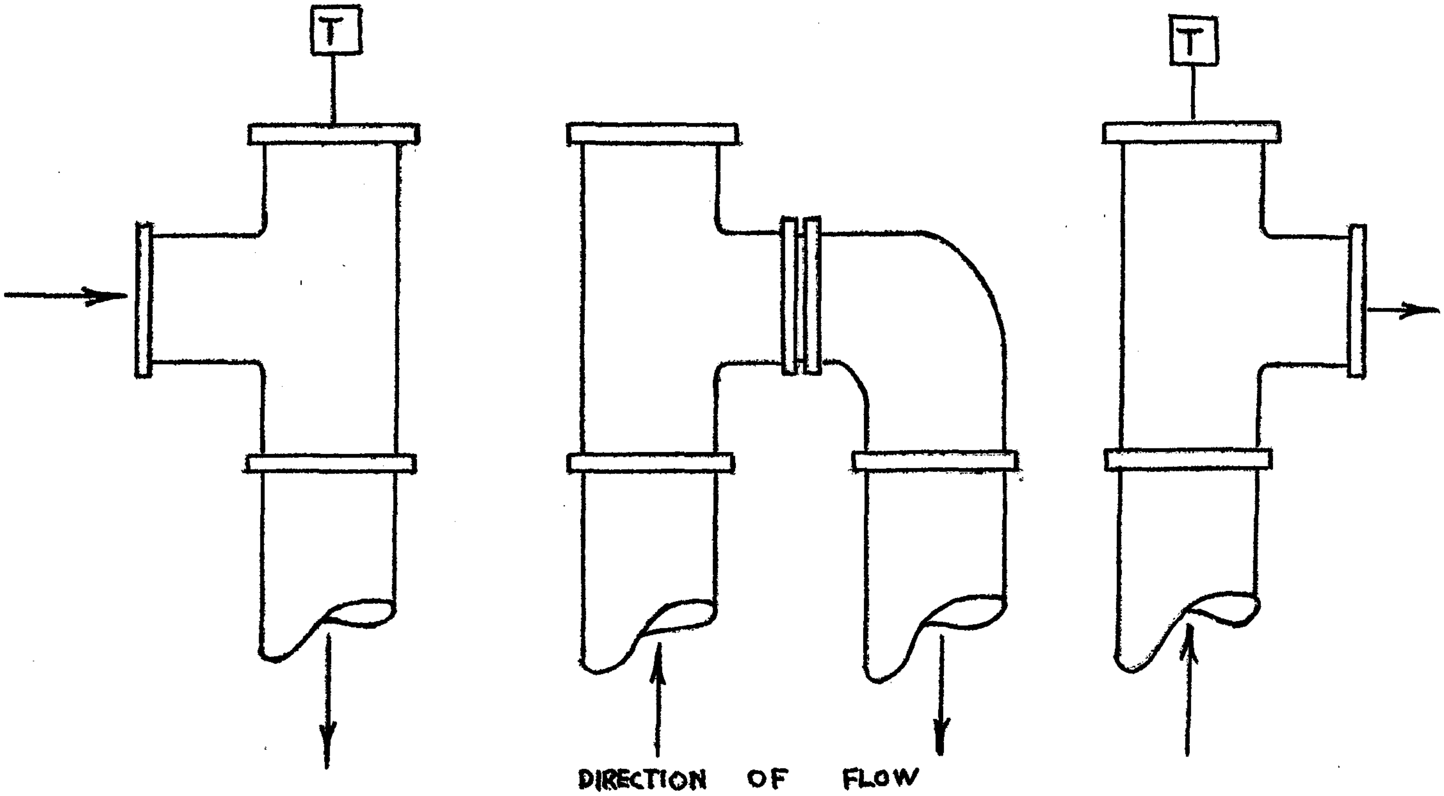


FIGURE NO. 4

DIAGRAM OF AGITATOR TANK AND TUBE BANKS



TOP VIEW



T -TEMPERATURE READINGS

SIDE VIEW

FIGURE NO. 5
ARRANGEMENT OF TUBE BANK

IV. PHYSICAL PROPERTIES OF FLUIDS

A. Water

Water from the city water supply of Newark was used. Viscosity, density, thermal conductivity, and specific heat data were obtained from the International Critical Tables (13).

B. Glycerine Solutions

Four groups of runs were made with the following bulk liquids:

a. Water

b. 14.8 Wt.% glycerine solution

c. 20.8 Wt.% glycerine solution

d. 61.0 Wt.% glycerine solution

The glycerine solutions were made from CP-USP grade glycerine and water. The density of the solutions were obtained on a Westphal balance. The weight percent glycerine in solution was determined from the density by the data of Bosart and Snoddy (14). The viscosity of these solutions were obtained from the data of the Glycerine Research Laboratories (15). The thermal conductivities were obtained from the data of Bates(16) and the specific heats from the International Critical Tables (13).

V. EXPERIMENTAL PROCEDURE

The bulk liquid was charged to a height of two feet above the bottom of the tank. The agitator was immersed midway between the tank bottom and the top liquid level.

Cold water was sent to the two cooling banks and hot water to the two heating banks. The flow to all banks was adjusted gradually until the two hot water banks had the same exit temperature and the two cold water banks had the same exit temperature.

When steady state conditions were reached, the temperatures at the entrance and exit of each bank were recorded by readings taken from 1° F graduated thermometers. The temperature of the bulk liquid was obtained from another 1° F graduated thermometer. Temperatures were taken at different locations and it was found that a uniform temperature existed at all points in the bulk liquid.

The total cold water and hot water flow were read on calibrated orifices. Since the temperature at the exit of each cold water bank was identical, the flow rate to each bank was one-half that obtained at the cold water orifice. The same situation existed with the hot water banks. In arriving at the final calculated data, only the total flow and total heat transfer surface for both cooling and both heating banks were used.

The agitator speed was obtained by a tachometer and checked at frequent intervals throughout each run.

VI. CALCULATIONS AND DATA TABULATION

The procedure used in calculating the film coefficients of heat transfer and the relating of the variables in dimensionaless groups was discussed in Section II, Theory. The experimental data taken in the course of this study will be found in Appendix A. A tabulation of all the calculations will be found in Appendix B.

VII. RESULTS

The following generalized equation has been found to correlate best the data obtained in this study for both heating and cooling:

$$(20)\quad \frac{h_c D_K}{K} = 4.3\left(\frac{L^2 N \rho}{\mu}\right)^{0.47}\left(\frac{C\mu}{K}\right)^{1/3}\left(\frac{\mu_c}{\mu}\right)^{-0.14}$$

This equation was derived from Figure 6 where the data are plotted in dimensionaless groups. This equation may be used to estimate the necessary areas of heat transfer for verticle tubes in agitated vessels that are dimensionally similar to the equipment of this thesis.

This study was the first initiated to develop such a correlation. Since this study was initiated, Dunlop and Rushton (6) have developed a generalized equation for verticle tubes in an agitated vessel. The data of this thesis do not fit this equation, and thus indicate that differences in system dimensions and dissimilarity are an important variable.

As a result of this thesis it was determined that for a well baffled system both radial flow and axial flow impellers have the same effect on the film coefficient. It was further determined that the exponent of the viscosity ratio is not constant and is a function of the fluid properties.

VIII. DISCUSSION

A. Results

Equation (20) will predict the mean forced-convection film coefficient of heat transfer on the outside of verticle tubes in cylindrical agitated vessels for both heating and cooling for tube arrangements similar to that of the equipment of this thesis. This equation may be used to estimate the necessary areas of heat transfer for verticle tubes in an agitated vessel within the degree of accuracy required for most engineering calculations.

In referring to Figure 6, it will be noted that the data points for cooling and heating do not coincide. For Reynolds numbers above 40,000, the data points for heating fall below those for cooling. The reverse is true for Reynolds numbers below 40,000. This occurs when a common exponent of the viscosity ratio (μ_c/μ) is used to correlate all the data.

When the viscosity ratio is not used in correlating the data, i. e., exponent equal to zero, the heating and cooling data points for Reynolds numbers above 40,000 coincide. Those points below Reynolds numbers of 40,000 tend to separate further. This is shown in Figure 7. Actually, two equations would be needed to relate the data, one for heating and one for cooling.

In Figure 8, an exponent of 0.25 for the viscosity ratio has been used. The heating and cooling data points for Reynolds numbers below 40,000 now coincide, but those above 40,000 do not. For Reynolds numbers below 40,000, the only bulk liquid used was the 61.0 wt.% glycerine solution.

When using one exponent for the viscosity ratio, it is impossible to obtain an exact correlation for both heating and cooling. The heating and cooling data points for water, the 14.8 wt.% glycerine solution, and the 20.8 wt.% glycerine solution coincide when the exponent of the viscosity ratio is zero. When the exponent is 0.25, the heating and cooling data points for the 61.0 wt.% glycerine solution coincide. Thus, equation (20) best defines the mean film coefficient of heat transfer for both cooling and heating.

Oldshue and Gretton (7) also noticed that the exponent of the viscosity ratio is not constant under varying conditions. They suggested two possible causes:

a. The value of (e) is a function of fluid properties.

b. The value of (e) is dependent upon the value of (μ_c/μ) and thus on the heat flux through the film.

With this in mind, the data was further analyzed to see what variables affected the exponent of the viscosity ratio. Table I presents the data in a narrow range of agitator speeds (66-82 RPM) for the various bulk fluids. The variable of relative liquid agitation is eliminated by restricting the data to this narrow range. The film coefficients for this data are plotted against the viscosity ratio in Figure 9. It will be noted that the slope of the lines for water and the 20.8 wt.% glycerine solution are zero. This agrees with the data presented in Figure 7, where it was necessary to use an exponent of zero for the viscosity ratio to get both the heating and cooling points to coincide. In Figure 9, the slope for the 61.0 wt.% glycerine solution is 0.25. This agrees with Figure 8, where an exponent of 0.25 for the viscosity ratio was necessary to make the heating and cooling points coincide for this solution.

It is not evident from the limited data of this thesis that the exponent for the viscosity ratio is dependent upon the value of (μ_c/μ) or the heat flux. The data does indicate though, that the exponent is a function of the fluid properties.

Two different agitators were used in this study and are described in the section on equipment. Briefly, one is a six-bladed curved radial type impeller and the other a pitched four-bladed type impeller that gave an axial flow component. No difference was found in the heat-transfer coefficients when using either impeller. Cummings and West (4) had found that when using a pitched blade turbine the heat-transfer coefficients for a helical coil were in the order of 10% lower than when a radial type impeller was used. In their experimental work no baffle was present to change the direction of flow. With a baffled system axial flow components are induced; so it seems reasonable that the type of impeller used with baffling will have little affect on the film coefficient.

B. Limitations

Forced convection heat-transfer coefficients for agitated vessels are only defined by a straight line relationship in a region of well defined turbulence. When turbulence becomes to great, a gas phase may be entrained with the liquid and lower coefficients than would be predicted will be obtained. This usually is accompanied by the formation of a vortex. In the other direction, where low turbulence is encountered, the influence of natural convection becomes great enough to definitely affect the forced-convection coefficients. In this region, higher coefficients than predicted will be obtained.

A few of the data points of this thesis were obtained in the region of low turbulence. The points in question are those in the Reynolds number range of 1600 to 4000. In this range, the influence of natural convection may have been large. The coefficients in this region, as predicted by the generalized correlation, may not be true forced-convection heat-transfer coefficients.

C. Comparison With Literature.

It is difficult to make a direct comparison of this data with that found in the literature. The physical structure of the systems cannot be ignored and are very important in establishing the generalized relationship for the heat-transfer coefficient. Most of the work done in this field has been on helical coils and jacketed vessels. The only other data collected on verticle tubes in agitated vessels were that of Rushton, Lichtmann, and Mahony (3) and that of Dunlap and Rushton (6).

In Figure 10, the data of Rushton, Lichtmann and Mahony (3) are shown with the line representing the generalized equation of this thesis. The system used in their work was geometrically similar to that of this thesis. A four foot diameter tank with four sets of verticle baffles were used. The only liquid studied was water and all the work was conducted at a high Reynolds number range. The data points have a considerable spread and do not fall along the line of the generalized correlation of this thesis.

Dunlap and Rushton (6) used a system similar to that of this thesis, but it differed in the following respects. The tank diameter was four feet. Only two heat transfer banks were used, one heating and one cooling. The dimensions of the banks and tube spacing in the banks were different. Only three tubes were used per bank. Furthermore, flow through the tubes in the banks were in parallel, while in this thesis flow was reversed in successive tubes. The generalized equation presented for their data is as follows:

$$(11)\quad \frac{h_c d_o}{K} = 0.09\left(\frac{L}{D_K}\right)^{0.33}\left(\frac{2}{B}\right)^{0.2}\left(\frac{\mu}{\mu_f}\right)^{0.4}\left(\frac{L^2 N \rho}{\mu}\right)^{0.65}\left(\frac{c\mu}{K}\right)^{1/3}$$

The length dimension in the Nusselt number was represented by the diameter of one of the tubes in a verticle tube bank. The mean temperature of the film was used to determine the film viscosity, μ_f. This accounts for the difference in the exponent of the viscosity ratio (μ_f/μ). For all previous studies the viscosity of the film was determined at the wall temperature. By inserting the tank diameter instead of the tube diameter, and the other system dimensions in equation (11), the following relationship is obtained. This equation is now basically similar to equation (20).

$$(21)\quad \frac{h_c D_K}{K} = 1.56\left(\frac{L^2 N \rho}{\mu}\right)^{0.65}\left(\frac{c\mu}{K}\right)^{1/3}\left(\frac{\mu_c}{\mu}\right)^{-0.14}$$

The graphical representation of this line is presented in Figure 10. The data of this thesis do not fit this line. The differences are probably a reflection of the difference in system dimensions and desimilarity.

Cummings and West (4) have derived the following equation for heating and cooling in helical coils.

$$(7)\quad \frac{h_c D_K}{K} = 1.01\left(\frac{L^2 N \rho}{\mu}\right)^{0.62}\left(\frac{C\mu}{K}\right)^{1/3}\left(\frac{\mu_c}{\mu}\right)^{-0.14}$$

The data of their study is also shown in Figure 10 and agrees within reason with that presented for the data of this thesis.

D. Experimental Error

The estimated precision obtained by fitting the data shown in Figure 6 within a two sigma range is ± 20%. Thus, Equation 20 may be used to estimate the forced-convection film coefficient on the outside of verticle tubes in an agitated vessel with a probable precision of ± 20% in 95% of the cases involved.

TABLE I.

RELATION BETWEEN FILM COEFFICIENT OF HEAT TRANSFER AND VISCOSITY RATIO

RUN	RPM	HEAT. μ_c/μ	COOL. μ_c/μ	h_h	h_c	$\frac{c\mu}{K}$	e	BULK LIQUID
2E	69	0.470	4.10	184	92	38.20	0.25	61.0%Glyc.
5D	82	0.491	3.41	233	135	38.20	0.25	61.0%Glyc.
6D	66	0.502	3.73	191	129	35.80	0.25	61.0%Glyc.
4A	76	0.681	1.560	298	300	4.38	-0-	Water
6A	76	0.640	1.583	255	276	4.40	-0-	Water
22A	74	0.656	1.680	287	294	3.87	-0-	Water
23A	74	0.663	1.725	281	278	3.81	-0-	Water
24A	74	0.643	1.672	301	262	3.98	-0-	Water
1B	74	0.667	1.828	246	259	6.96	-0-	20.8%Glyc.
2B	74	0.652	1.756	234	271	7.25	-0-	20.8%Glyc.
22B	74	0.674	1.803	239	235	6.89	-0-	20.8%Glyc.
23B	74	0.664	1.780	229	236	7.01	-0-	20.8%Glyc.
24B	74	0.638	1.712	240	222	7.31	-0-	20.8%Glyc.

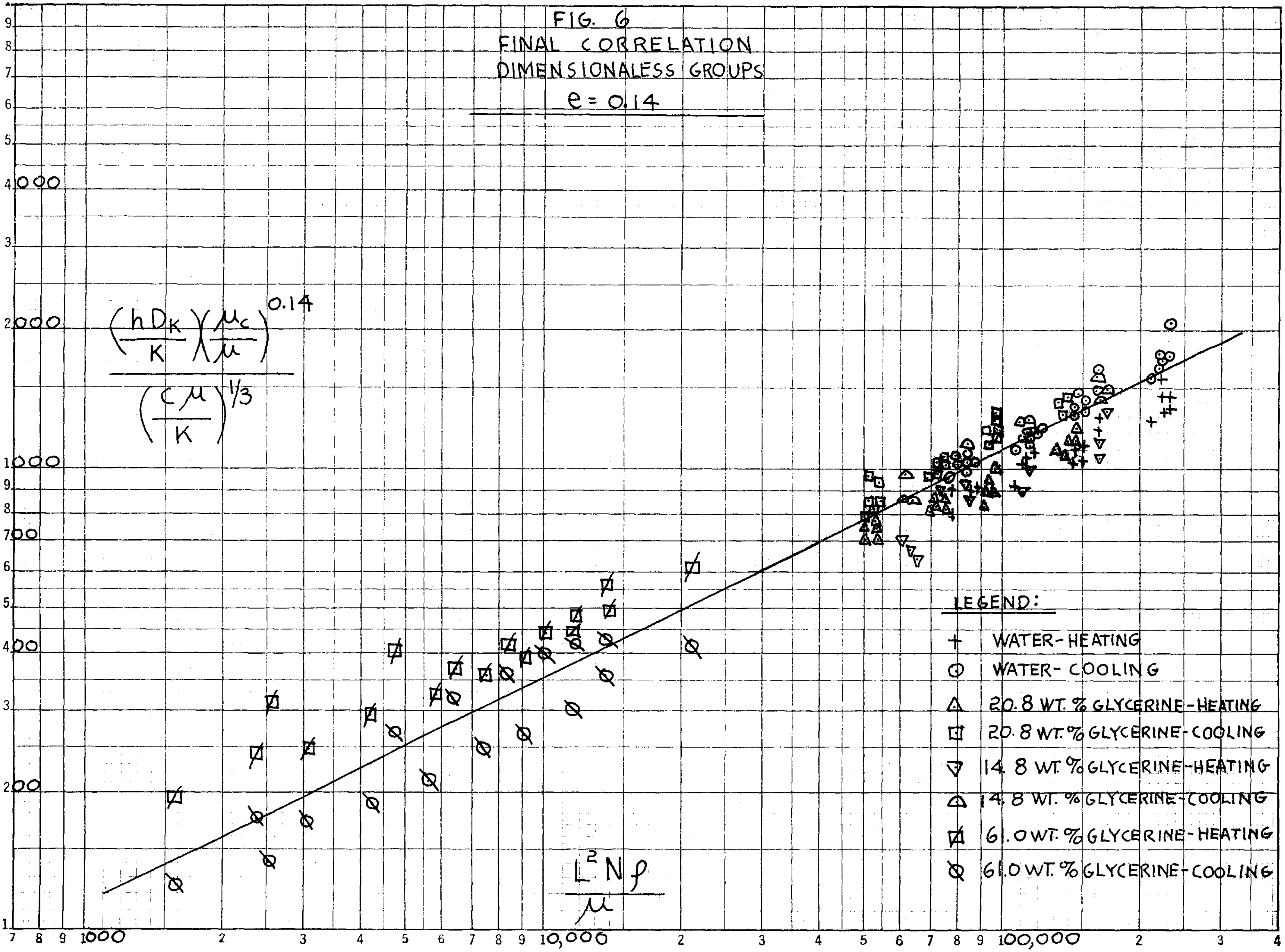


FIG. 6
FINAL CORRELATION
DIMENSIONALESS GROUPS
e = 0.14
$\dfrac{\left(\frac{hD_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{C\mu}{K}\right)^{1/3}}$
$\dfrac{L^2 N \rho}{\mu}$
LEGEND:
WATER-HEATING
WATER-COOLING
20.8 WT. % GLYCERINE-HEATING
20.8 WT. % GLYCERINE-COOLING
14.8 WT. % GLYCERINE-HEATING
14.8 WT. % GLYCERINE-COOLING
61.0 WT. % GLYCERINE-HEATING
61.0 WT. % GLYCERINE-COOLING

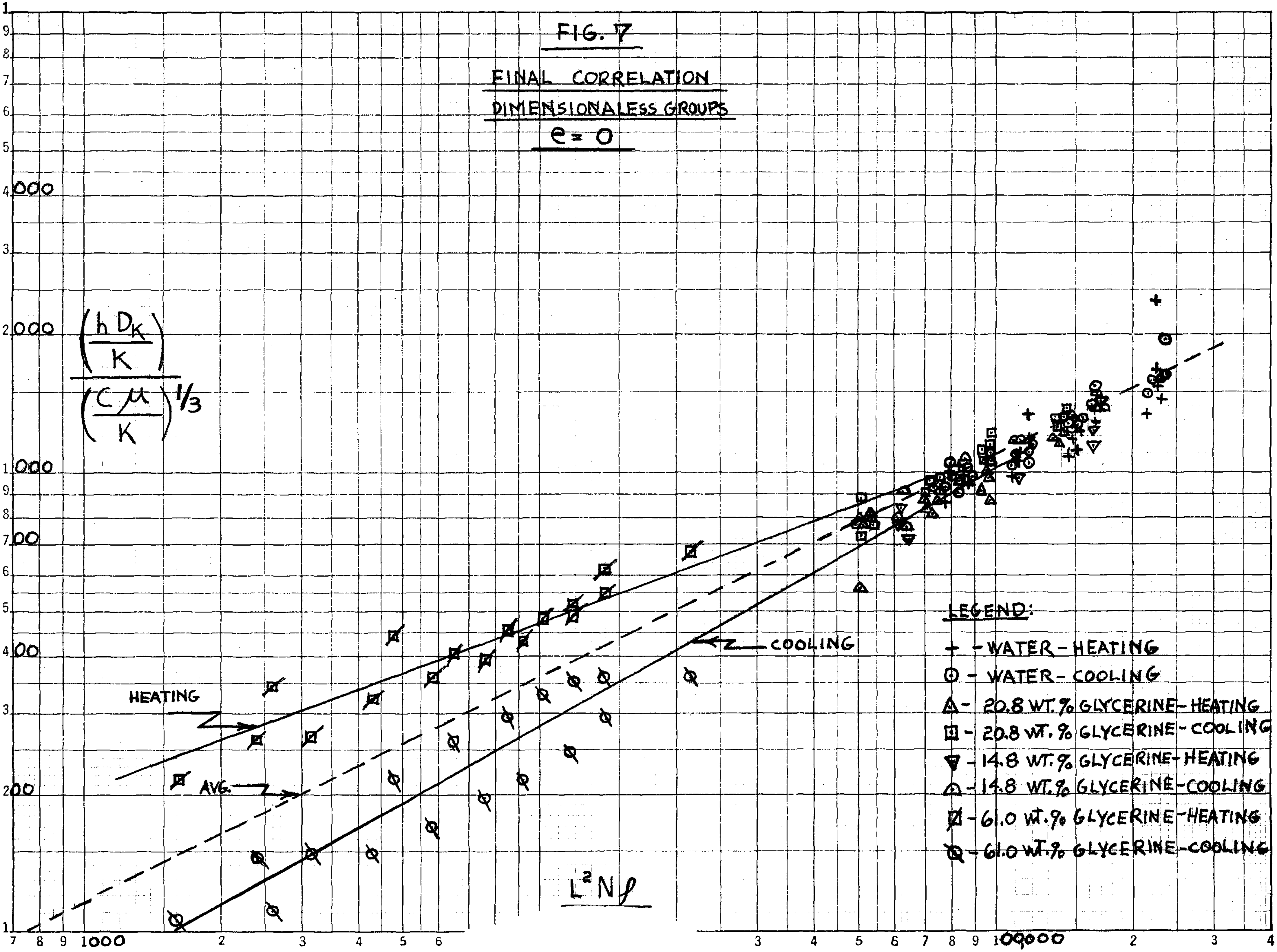


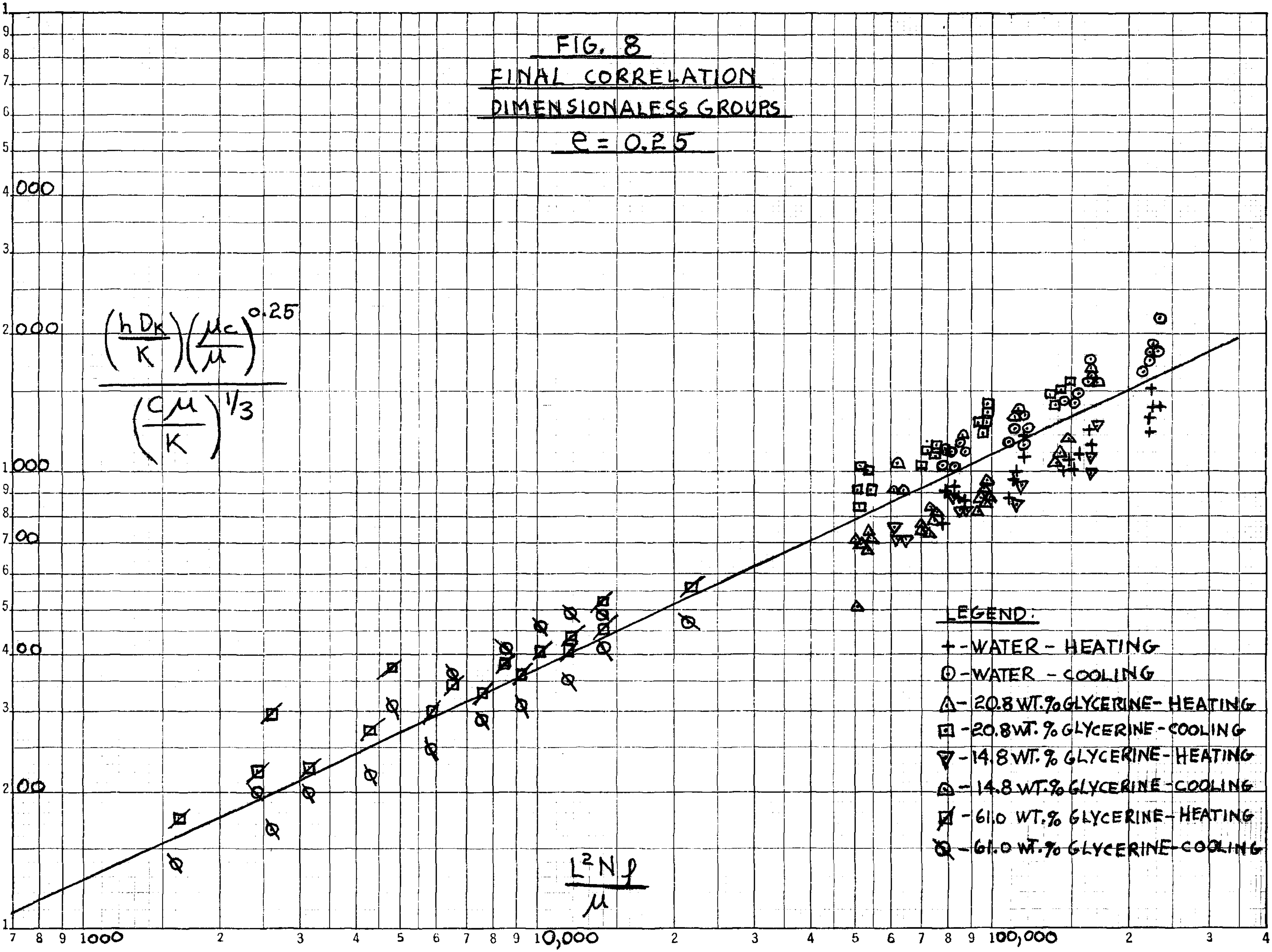
FIG. 7
FINAL CORRELATION
DIMENSIONALESS GROUPS
e = 0
$\dfrac{\left(\dfrac{hD_K}{K}\right)}{\left(\dfrac{C\mu}{K}\right)^{1/3}}$
$L^2N\rho$
HEATING
AVG.
COOLING
LEGEND:
+ - WATER - HEATING
⊙ - WATER - COOLING
20.8 WT. % GLYCERINE - HEATING
20.8 WT. % GLYCERINE - COOLING
14.8 WT. % GLYCERINE - HEATING
14.8 WT. % GLYCERINE - COOLING
61.0 WT. % GLYCERINE - HEATING
61.0 WT. % GLYCERINE - COOLING
1000
10,000

FIG. 8

FINAL CORRELATION

DIMENSIONALESS GROUPS

e = 0.25



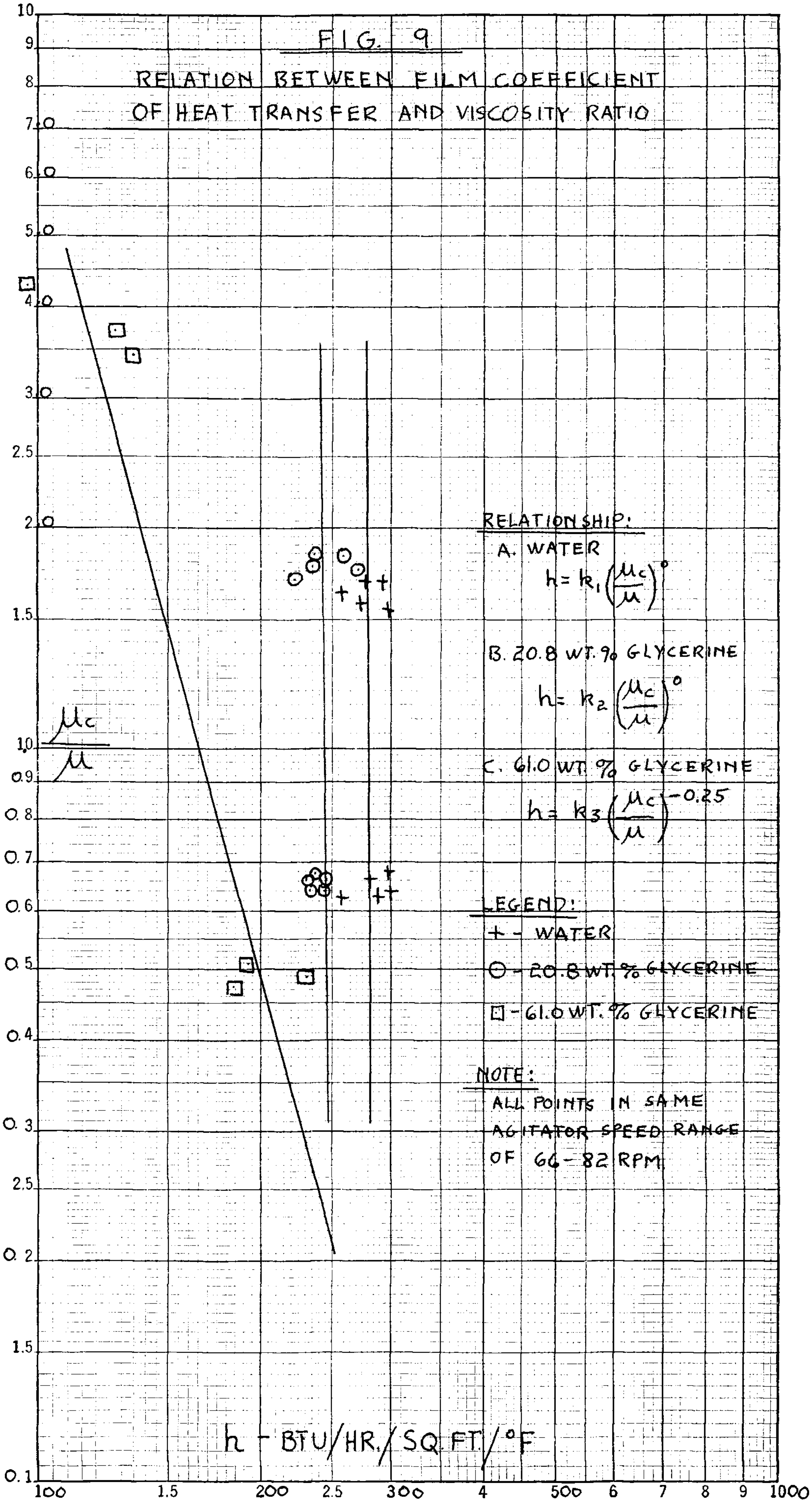


FIG. 9
RELATION BETWEEN FILM COEFFICIENT OF HEAT TRANSFER AND VISCOSITY RATIO
RELATIONSHIP:
A. WATER
$h = k_1 \left(\frac{\mu_c}{\mu}\right)^0$
B. 20.8 WT. % GLYCERINE
$h = k_2 \left(\frac{\mu_c}{\mu}\right)^0$
C. 61.0 WT. % GLYCERINE
$h = k_3 \left(\frac{\mu_c}{\mu}\right)^{-0.25}$
LEGEND:
+ - WATER
○ - 20.8 WT. % GLYCERINE
□ - 61.0 WT. % GLYCERINE
NOTE:
ALL POINTS IN SAME AGITATOR SPEED RANGE OF 66 - 82 RPM
$\frac{\mu_c}{\mu}$
h - BTU/HR./SQ.FT/°F
10
9
8
7.0
6.0
5.0
4.0
3.0
2.5
2.0
1.5
1.0
0.9
0.8
0.7
0.6
0.5
0.4
0.3
2.5
0.2
1.5
0.1
100
1.5
200
2.5
300
4
500
6
7
8
9
1000

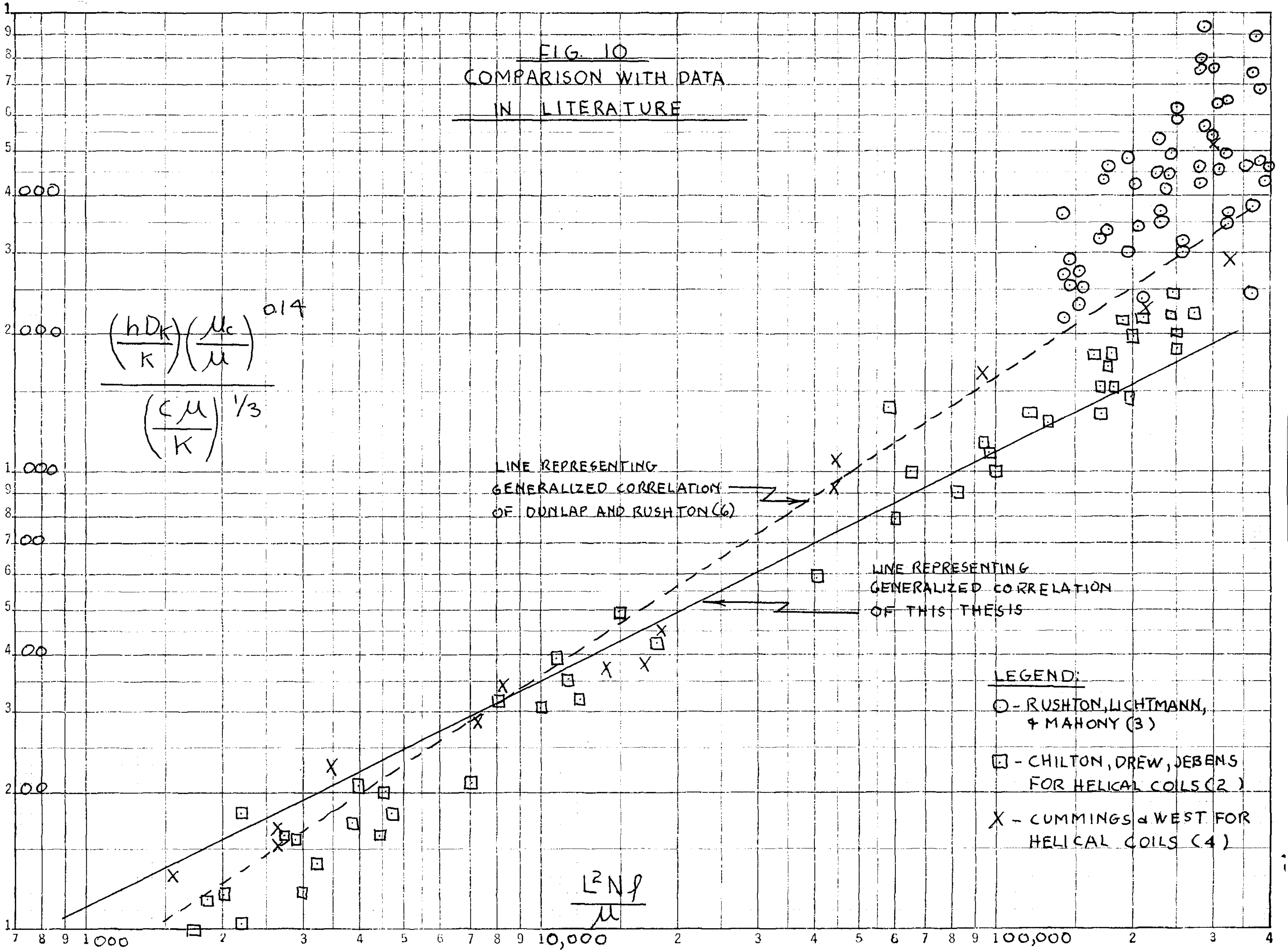


FIG. 10
COMPARISON WITH DATA
IN LITERATURE
$$\frac{\left(\frac{hD_K}{k}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{k}\right)^{1/3}}$$
$\frac{L^2 N \rho}{\mu}$
LINE REPRESENTING
GENERALIZED CORRELATION
OF DUNLAP AND RUSHTON (6)
LINE REPRESENTING
GENERALIZED CORRELATION
OF THIS THESIS
LEGEND:
⊙ - RUSHTON, LICHTMANN, & MAHONY (3)
⊡ - CHILTON, DREW, JEBENS FOR HELICAL COILS (2)
X - CUMMINGS & WEST FOR HELICAL COILS (4)
4,000
2,000
1,000
700
400
200
1000
10,000
100,000

IX. CONCLUSIONS

From the data and calculations of this thesis, it is concluded that:

1) The following generalized equation defines the mean forced-convection heat-transfer coefficient for cooling and heating as it varies with fluid properties and agitator speeds for verticle tubes in an agitated vessel.

$$(20) \quad \frac{h_c D_K}{K} = 4.3\left(\frac{L^2 N \rho}{\mu}\right)^{0.47}\left(\frac{C\mu}{K}\right)^{1/3}\left(\frac{\mu_c}{\mu}\right)^{-0.14}$$

For Reynolds numbers above 40,000 this equation will predict film coefficients for heating that are slightly higher and film coefficients for cooling that are slightly lower than will actually be obtained. The reverse is true for Reynolds numbers below 40,000.

2) The exponent of the viscosity ratio is not constant under varying conditions and it is a function of the fluid properties.

3) For a well baffled system, a radial flow impeller has the same affect on the film coefficients as an impeller with an axial flow component.

4) The data of this thesis do not fit the equation proposed by Dunlap and Rushton (6) for the prediction of film coefficients of heat transfer for verticle tubes in an agitated vessel.

X. RECOMMENDATIONS

1) It is recommended that work be continued on equipment similar to that of this thesis to determine the effect of varying the baffle dimensions on the film coefficient.

2) It is recommended that an investigation be made to determine exactly what variables affect the exponent of the viscosity ratio and how the exponent varies with these functions.

3) It is recommended that work be continued on a dimensionally similar system, but varying tank diameter and impellar diameter and width, to determine the effect on the film coefficient.

4) It is further recommended that the length term in the Nusselt number be resolved as to whether the tank diameter or baffle tube diameter is the correct variable to be used.

XI. NOMENCLATURE

The following dimensions are those used in this paper:

A	Heat transfer area, sq. ft.
a	Exponent of Reynolds No.
B	Number of baffles.
b	Exponent of Prandtl No.
C	Specific heat, BTU/ (lb.) (^{o}F.)
D_k	Tank diameter, ft.
do	Outside diameter of tube in baffle, ft.
di	Inside diameter of tube in baffle, ft.
e	Exponent of viscosity ratio.
hi	Film coefficient of heat transfer inside tubes, B.T.U./(hr.) (sq.ft.) (^{o}F.)
hc	Film coefficient of heat transfer on vessel side for tube or coil, B.T.U./(hr.) (sq.ft.) (^{o}F.)
hj	Film coefficient of heat transfer on vessel side for jacket, B.T.U./(hr.) (sq.ft.) (^{o}F.)
K	Thermal conductivity, B.T.U./(hr.) (sq.ft.)(^{o}F.)/ft.
L	Agitator diameter, ft.
Lw	Width of Agitator, ft.

N	Agitator speed, revolutions/hr.
N_{Nu}	Nusselt No., hD_k/K - dimensionaless.
N_{Pr}	Prandtl No., $c\mu/K$ - dimensionaless.
N_{Re}	Reynolds No., $L^2N\rho/\mu$ - dimensionaless.
Q	Heat transfer rate, B.T.U./hr.
r_w	Resistance of the coil wall, (hr.)(sq.ft.)(°F.)/B.T.U.
U	Overall heat transfer coefficient, B.T.U./(hr.)(sq.ft.) (°F.)
V	Fluid velocity in coil, ft./sec.
W	Fluid rate in tubes, lbs./hr.
X	Empirical constant.
Δt_m	Log mean temperature difference, °F.
Δt	Temperature difference, °F.
μ	Viscosity at bulk temperature, lb./(hr.)(ft.)
μ_c	Viscosity at surface temperature on vessel side of coil or tube, lb./(hr.) (ft.)
μ_f	Viscosity at mean film temperature on vessel side of tube, lb./(hr.) (ft.)
μ_j	Viscosity at surface temperature on vessel side of jacket, lb./(hr.) (ft.)
ρ	Density at bulk temperature, lb./cu. ft.

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APPENDIX A

EXPERIMENTAL DATA

TABLE 2

EXPERIMENTAL DATA - BULK LIQUID-WATER

RUN NO.	AGITATOR TYPE	AGITATOR SPEED-RPM	BULK LIQUID TEMP.-°F	HOT WATER INLET TEMP.-°F	HOT WATER OUTLET TEMP.-°F	HOT WATER FLOW MAN. RD.	COLD WATER INLET TEMP.-°F	COLD WATER OUTLET TEMP.-°F	COLD WATER FLOW MAN. RD.
1A	4 Blade	104	114.0	170.0	153.5	3.00	65.0	74.5	2.60
2A	4 "	104	107.0	162.0	145.0	2.00	64.5	72.5	2.65
3A	4 "	104	105.0	162.0	140.5	1.00	64.2	71.5	2.65
4A	4 "	76	103.5	151.5	139.7	3.00	64.0	70.7	2.65
5A	4 "	76	105.0	157.0	147.3	2.00	64.0	71.2	2.65
6A	4 "	76	103.0	162.8	142.7	1.00	63.5	70.2	2.65
7A	4 "	139	110.0	161.5	146.5	2.80	63.5	73.2	2.65
8A	4 "	139	108.0	161.5	144.5	2.00	64.0	73.0	2.65
9A	4 "	139	105.0	166.5	142.8	1.00	64.0	72.5	2.65
10A	4 "	212	110.0	162.0	145.0	2.80	64.0	76.0	2.65
11A	4 "	212	108.3	163.5	143.5	2.00	64.0	74.5	2.65
12A	4 "	212	104.0	163.0	143.0	1.00	64.0	73.5	2.65
13A	6 Blade	204	110.0	161.5	144.0	2.85	64.5	75.5	2.65
14A	6 "	204	110.5	169.4	147.5	2.00	64.5	75.5	2.65
15A	6 "	204	104.5	169.0	141.5	1.00	64.5	73.5	2.65
16A	6 "	139	118.0	178.0	159.5	2.80	64.5	76.5	2.65
17A	6 "	139	116.5	179.7	158.0	2.00	64.5	75.6	2.65
18A	6 "	139	106.0	170.8	145.0	1.00	64.5	73.0	2.65
19A	6 "	104	112.0	165.0	150.0	2.80	65.0	73.5	2.65
20A	6 "	104	111.7	169.3	151.5	2.00	65.0	73.7	2.65
21A	6 "	104	108.0	167.0	144.5	1.00	64.5	72.5	2.65
22A	6 "	74	114.0	171.0	156.5	2.80	65.0	73.5	2.65
23A	6 "	74	116.0	174.5	157.5	2.00	65.0	73.5	2.65
24A	6 "	74	111.5	174.5	151.0	1.00	64.5	72.0	2.65

TABLE 3

EXPERIMENTAL DATA - BULK LIQUID-20.8 WT. % GLYCERINE. SOLN.

				------HOT WATER	----------	----------	----------COLD WATER	----	----
RUN NO.	AGITATOR TYPE	AGITATOR SPEED-RPM	BULK LIQUID TEMP.-°F	INLET TEMP.-°F	OUTLET TEMP.-°F	FLOW MAN. RD.	INLET TEMP.-°F	OUTLET TEMP.-°F	FLOW MAN. RD.
1B	6 Blade	74	114.0	164.8	153.0	2.80	65.0	72.7	2.65
2B	6 "	74	110.6	161.2	148.2	2.00	65.0	72.6	2.65
3B	6 "	74	109.6	168.0	152.5	1.00	65.0	71.8	2.65
4B	6 "	139	112.0	163.8	149.5	2.75	65.0	74.3	2.65
5B	6 "	139	111.4	165.8	149.5	2.00	65.0	74.2	2.65
6B	6 "	139	107.0	165.5	144.6	1.00	65.0	73.2	2.65
7B	6 "	204	110.2	159.4	144.7	2.75	65.0	74.5	2.65
8B	6 "	204	109.0	162.8	145.0	2.00	65.5	74.5	2.65
9B	6 "	204	106.8	167.8	143.0	1.00	65.7	74.4	2.65
10B	6 "	104	115.0	172.0	158.0	2.75	65.5	74.4	2.65
11B	6 "	104	113.5	165.5	151.0	2.00	65.5	73.8	2.65
12B	6 "	104	107.5	165.8	145.4	1.00	66.3	73.2	2.65
13B	4 Blade	104	113.0	165.0	152.0	2.75	66.0	74.0	2.65
14B	4 "	104	112.0	161.0	148.0	2.00	66.0	74.0	2.65
15B	4 "	104	108.5	163.0	144.4	1.00	66.0	73.2	2.65
16B	4 "	139	110.0	157.7	145.0	2.80	66.0	74.5	2.65
17B	4 "	139	111.0	162.3	147.0	2.00	66.0	74.5	2.65
18B	4 "	139	109.2	169.0	147.0	1.00	66.4	74.4	2.65
19B	4 "	212	111.0	159.5	144.6	2.80	66.4	76.0	2.65
20B	4 "	212	110.0	162.0	144.5	2.00	66.4	75.8	2.65
21B	4 "	212	106.5	163.0	144.5	1.00	66.4	74.8	2.65
22B	4 "	74	114.8	163.0	152.0	2.80	66.5	73.8	2.65
23B	4 "	74	113.5	163.8	151.0	2.00	66.5	73.6	2.65
24B	4 "	74	110.0	164.0	146.3	1.00	67.0	73.2	2.65

BULK LIQUID Sp. Gr. - 1.0500 @ 28°C

TABLE 4

EXPERIMENTAL DATA - BULK LIQUID-14.8 WT.% GLYCERINE SOLN.

				------HOT WATER	---------	---------	---COLD WATER	-----	-----
RUN NO.	AGITATOR TYPE	AGITATOR SPEED-RPM	BULK LIQUID TEMP.-°F	INLET TEMP.-°F	OUTLET TEMP.-°F	FLOW MAN. RD.	INLET TEMP.-°F	OUTLET TEMP.-°F	FLOW MAN. RD.
1C	4 Blade	76	116.0	167.0	156.0	2.65	67.0	74.4	2.65
2C	4 "	76	110.5	165.0	149.3	1.50	67.0	73.7	2.65
3C	4 "	139	114.0	167.0	152.5	2.85	67.2	76.5	2.65
4C	4 "	139	111.0	162.0	145.2	1.50	67.0	75.8	2.65
5C	4 "	212	110.8	158.0	143.0	2.85	67.0	76.7	2.65
6C	4 "	212	108.5	160.2	141.0	1.50	67.0	76.3	2.65
7C	4 "	104	112.0	163.0	150.0	2.85	67.0	75.0	2.65
8C	4 "	104	109.0	162.0	142.2	1.00	66.7	74.2	2.65
9C	6 Blade	204	111.0	159.3	144.7	2.85	68.0	77.5	2.65
10C	6 "	139	112.5	162.0	148.5	2.85	67.8	76.7	2.65
11C	6 "	74	115.5	167.0	155.0	2.85	68.0	75.9	2.65
12C	6 "	104	114.0	165.8	152.5	2.85	67.7	76.3	2.65

BULK LIQUID - Sp. Gr. = 1.0351 @ 30°c.

TABLE 5

EXPERIMENTAL DATA - BULK LIQUID-61.0 WT.% GLYCERINE SOLN.

				------HOT WATER		-----------	----------COLD WATER	-----	
RUN NO.	AGITATOR TYPE	AGITATOR SPEED-RPM	BULK LIQUID TEMP.-°F	INLET TEMP.-°F	OUTLET TEMP.-°F	FLOW LBS./SEC.	INLET TEMP.-°F	OUTLET TEMP. °F	FLOW LBS./SEC.
1D	6 Blade	10.5	107.0	169.2	164.6	50/55	59.7	61.5	100/107
2D	6 "	17.5	117.0	164.0	160.0	50/55	59.4	62.4	100/107
3D	6 "	13.0	119.0	169.0	164.9	50/55	59.3	62.3	100/107
4D	-------	-0-	118.5	168.8	164.4	50/55	59.3	61.7	100/107
5D	6 Blade	82.0	113.0	166.3	156.7	50/61	55.4	60.8	50/45
6D	6 "	66.0	117.4	171.0	162.5	50/61	55.0	60.2	50/45
7D	6 "	46.0	118.7	172.3	164.7	50/61	54.4	59.3	50/45
8D	6 "	55.0	118.7	173.3	165.2	50/61	54.4	59.8	50/45
9D	6 "	34.5	120.3	172.2	165.5	50/61	54.3	58.8	50/45
10D	6 "	25.0	121.3	176.1	168.5	50/61	54.2	58.1	50/45

1E	4 Blade	87.0	110.0	170.4	160.2	50/61	48.0	53.6	50/48
2E	4 "	69.0	113.5	172.0	163.0	50/61	48.0	53.0	50/48
3E	4 "	53.0	114.7	170.8	163.0	50/61	47.8	52.3	50/48
4E	4 "	43.0	115.6	170.2	163.2	50/61	47.6	51.8	50/48
5E	4 "	32.5	117.2	171.4	165.0	50/61	47.6	51.3	50/48
6E	4 "	23.5	118.3	169.5	164.0	50/61	47.4	50.7	50/48
7E	4 "	13.0	123.5	168.8	163.8	50/61	47.3	50.0	50/48
8E	4 "	132.0	111.6	168.0	157.0	50/61	47.3	54.2	50/48

BULK LIQUID - Sp. Gr. = 1.1532 @ 82°F.

APPENDIX B

SUMMARY OF CALCULATIONS

TABLE 6

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - WATER

HEATING

RUN NO.	Q BTU/MIN.	A SQ. FT.	Δtm	U_{heat}	h_i	h_h
1A	705	3.41	47.1	264	1590	386
2A	569	3.41	45.8	218	1241	317
3A	522	3.41	48.3	190	960	287
4A	506	3.41	41.9	214	1402	298
5A	326	3.41	46.9	122	1245	149
6A	492	3.41	49.0	176	971	255
7A	610	3.41	43.6	246	1465	358
8A	567	3.41	44.5	224	1241	331
9A	578	3.41	48.7	209	974	327
10A	692	3.41	42.9	283	1465	442
11A	668	3.41	44.5	264	1253	424
12A	488	3.41	48.4	178	969	257
13A	720	3.41	42.1	302	1458	490
14A	733	3.41	47.2	273	1267	446
15A	671	3.41	49.6	237	984	400
16A	763	3.41	50.3	267	1576	394
17A	726	3.41	51.7	247	1332	372
18A	629	3.41	50.9	218	990	347
19A	618	3.41	45.0	242	1500	347
20A	594	3.41	45.0	232	1280	344
21A	549	3.41	46.8	207	974	323
22A	597	3.41	49.5	212	1535	287
23A	568	3.41	49.4	203	1326	281
24A	573	3.41	50.3	200	1007	301

TABLE 6 (Cont.)

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - WATER

COOLING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δt_m	U_{cool}	h_i	h_o
1A	602	3.41	44.2	240	1567	339
2A	508	3.41	38.6	232	1568	323
3A	466	3.41	36.7	224	1559	308
4A	428	3.41	34.4	219	1552	300
5A	459	3.41	37.5	215	1558	291
6A	428	3.41	36.6	206	1548	276
7A	618	3.41	41.5	263	1565	386
8A	575	3.41	39.4	257	1566	373
9A	542	3.41	36.8	259	1567	378
10A	765	3.41	39.7	339	1579	571
11A	670	3.41	38.9	303	1572	479
12A	606	3.41	35.1	304	1525	485
13A	702	3.41	41.3	299	1576	467
14A	702	3.41	40.1	308	1576	488
15A	574	3.41	35.3	287	1568	440
16A	766	3.41	47.2	290	1580	445
17A	708	3.41	46.4	269	1576	397
18A	542	3.41	37.0	258	1569	375
19A	542	3.41	42.7	223	1571	306
20A	556	3.41	42.0	233	1571	324
21A	518	3.41	39.4	232	1566	322
22A	551	3.41	44.6	217	1571	294
23A	551	3.41	46.6	208	1571	278
24A	487	3.41	43.1	199	1564	262

TABLE 7

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 20.8 WT.% GLYCERINE SOLN.

HEATING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δt_m	U_{heat}	h_i	h_h
1B	480	3.41	44.8	189	1515	246
2B	435	3.41	43.5	176	1273	234
3B	378	3.41	50.2	132	991	171
4B	576	3.41	44.3	229	1491	322
5B	545	3.41	46.0	208	1293	292
6B	510	3.41	47.4	190	990	282
7B	594	3.41	41.5	252	1462	370
8B	596	3.41	44.5	236	1270	354
9B	605	3.41	47.7	223	991	361
10B	563	3.41	49.9	205	1543	274
11B	484	3.41	44.5	191	1293	260
12B	498	3.41	48.0	182	993	265
13B	524	3.41	45.2	204	1398	278
14B	435	3.41	42.1	182	1272	245
15B	454	3.41	44.5	179	988	256
16B	519	3.41	41.2	222	1473	309
17B	512	3.41	43.1	209	1274	297
18B	537	3.41	48.1	196	1006	294
19B	606	3.41	40.7	262	1420	397
20B	586	3.41	42.7	242	1269	368
21B	451	3.41	46.7	170	986	240
22B	446	3.41	42.6	184	1510	239
23B	429	3.41	43.5	173	1293	229
24B	431	3.41	44.7	170	992	240

TABLE 7 (Cont.)

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 20.8 WT.% GLYCERINE SOLN.

COOLING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δtm	U_{cool}	h_i	h_c
1B	491	3.41	45.2	192	1572	259
2B	485	3.41	41.8	204	1572	271
3B	434	3.41	41.0	186	1562	240
4B	593	3.41	42.1	255	1572	369
5B	586	3.41	41.6	248	1572	355
6B	523	3.41	37.9	243	1570	344
7B	606	3.41	40.0	267	1577	396
8B	574	3.41	39.1	258	1580	374
9B	555	3.41	36.3	269	1580	397
10B	567	3.41	46.0	217	1580	294
11B	529	3.41	44.0	211	1572	284
12B	440	3.41	37.5	206	1580	274
13B	510	3.41	42.9	209	1580	280
14B	510	3.41	41.9	214	1580	288
15B	459	3.41	38.9	208	1572	278
16B	541	3.41	39.7	240	1571	339
17B	541	3.41	40.8	234	1571	327
18B	510	3.41	38.7	232	1580	321
19B	612	3.41	39.4	274	1592	408
20B	599	3.41	38.5	274	1593	406
21B	535	3.41	35.8	263	1582	385
22B	465	3.41	44.7	183	1571	235
23B	453	3.41	43.5	184	1580	236
24B	395	3.41	39.6	175	1578	222

TABLE 8

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 14.8 WT.% GLYCERINE SOLN.

HEATING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δt_m	U_{heat}	h_i	h_h
1C	430	3.41	45.5	166	1490	210
2C	454	3.41	46.5	172	1148	232
3C	597	3.41	45.4	232	1498	327
4C	485	3.41	42.4	202	1130	293
5C	619	3.41	39.3	277	1480	426
6C	555	3.41	41.3	236	1112	373
7C	536	3.41	44.2	214	1519	292
8C	484	3.41	42.4	201	980	309
9C	601	3.41	44.4	239	1487	341
10C	555	3.41	42.5	230	1508	323
11C	494	3.41	45.4	192	1550	250
12C	549	3.41	45.1	214	1539	291

TABLE 8 (Cont.)

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 14.8 WT.% GLYCERINE SOLN.

COOLING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δtm	U_{cool}	h_i	h_c
1C	472	3.41	46.0	181	1583	231
2C	427	3.41	40.5	186	1574	240
3C	593	3.41	42.1	246	1600	348
4C	561	3.41	39.9	248	1596	351
5C	619	3.41	39.1	279	1600	418
6C	593	3.41	36.8	292	1591	450
7C	510	3.41	41.4	217	1586	293
8C	479	3.41	38.8	217	1588	293
9C	605	3.41	38.3	278	1604	415
10C	568	3.41	40.2	249	1600	355
11C	504	3.41	43.5	204	1596	270
12C	548	3.41	42.5	227	1596	312

TABLE 9

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 61.0 WT.% GLYCERINE SOLN.

HEATING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δt_m	U_{heat}	h_i	h_h
1D	252	3.22	59.2	79.2	1947	84.0
2D	219	3.22	45.5	89.2	1908	98.5
3D	224	3.22	48.0	87.0	1947	95.6
4D	240	3.22	48.1	93.0	1942	103.0
5D	472	3.22	48.3	183.5	1655	233.0
6D	418	3.22	49.4	157.7	1780	191.0
7D	375	3.22	49.5	140.8	1812	166.5
8D	399	3.22	50.3	148.0	1817	176.0
9D	330	3.22	48.9	125.7	1813	146.0
10D	374	3.22	51.2	136.0	1838	159.3

1E	502	3.22	55.3	169.3	1787	208.5
2E	444	3.22	54.0	153.0	1798	184.0
3E	384	3.22	52.2	137.0	1792	161.3
4E	345	3.22	51.1	125.6	1792	145.7
5E	316	3.22	51.0	115.3	1807	132.0
6E	271	3.22	48.5	104.2	1792	117.7
7E	257	3.22	42.8	107.3	1788	122.0
8E	541	3.22	50.9	199.5	1765	257.0

TABLE 9 (Cont.)

SUMMARY OF CALCULATIONS
DETERMINATION OF FORCED-CONVECTION
FILM COEFFICIENT - 61.0 WT.% GLYCERINE SOLN.

COOLING

RUN NO.	Q BTU/MIN.	A SQ.FT.	Δt_m	U_{cool}	h_i	h_c
1D	102	3.37	48.2	37.7	1309	39.8
2D	168	3.37	59.1	50.6	1312	54.4
3D	168	3.37	60.2	49.6	1312	53.2
4D	135	3.37	60.4	39.8	1309	42.1
5D	360	3.37	55.6	115.2	1477	134.7
6D	347	3.37	59.8	103.3	1470	129.0
7D	326	3.37	61.9	94.0	1463	106.8
8D	360	3.37	61.6	104.0	1477	120.0
9D	300	3.37	63.7	83.8	1463	93.5
10D	260	3.37	65.2	71.0	1456	78.0
1E	350	3.37	64.8	96.2	1332	110.7
2E	313	3.37	68.0	81.9	1324	92.1
3E	281	3.37	69.2	72.3	1332	80.1
4E	263	3.37	70.1	66.6	1323	73.1
5E	231	3.37	71.4	57.6	1320	62.5
6E	206	3.37	72.5	50.7	1318	54.5
7E	162	3.37	77.5	37.3	1317	39.3
8E	431	3.37	67.7	108.5	1330	127.4

TABLE 10

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS HEATING

RUN NO.	RE. NO. $\times 10^{-2}$	$\frac{c\mu}{K}$	$\left(\frac{c\mu}{K}\right)^{1/3}$	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_h D_K}{K}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1A	1198	3.88	1.572	0.672	0.946	0.905	2090	1258	1202	1330
2A	1118	4.21	1.614	0.677	0.947	0.907	1738	1020	975	1076
3A	1096	4.30	1.626	0.691	0.950	0.912	1577	922	886	980
4A	789	4.38	1.637	0.681	0.948	0.908	1642	950	911	1003
5A	800	4.30	1.626	0.660	0.943	0.902	819	----	----	----
6A	785	4.40	1.639	0.640	0.940	0.895	1408	806	768	857
7A	1536	4.06	1.596	0.650	0.941	0.898	1952	1152	1097	1224
8A	1510	4.14	1.606	0.674	0.946	0.910	1810	1068	1026	1127
9A	1466	4.30	1.626	0.669	0.945	0.908	1797	1045	1005	1105
10A	2350	4.06	1.596	0.694	0.950	0.913	2410	1442	1380	1510
11A	2310	4.14	1.605	0.685	0.948	0.910	2320	1372	1313	1445
12A	2210	4.36	1.633	0.647	0.941	0.897	1413	815	777	865
13A	2260	4.05	1.596	0.687	0.949	0.911	2670	1590	1526	1675
14A	2270	3.98	1.585	0.666	0.945	0.904	2430	1450	1387	1532
15A	2140	4.32	1.629	0.637	0.939	0.894	2200	1270	1208	1352
16A	1660	3.73	1.551	0.663	0.945	0.902	2130	1300	1240	1372
17A	1640	3.78	1.558	0.649	0.942	0.898	2015	1220	1163	1292
18A	1480	4.26	1.621	0.660	0.944	0.902	1907	1112	1061	1177
19A	1175	3.98	1.585	0.680	0.947	0.908	1885	1128	1080	1190
20A	1170	3.98	1.585	0.663	0.945	0.902	1870	1112	1063	1180
21A	1130	4.16	1.607	0.664	0.945	0.902	1765	1040	991	1100
22A	851	3.87	1.571	0.656	0.943	0.900	1555	934	890	989
23A	869	3.81	1.562	0.663	0.945	0.902	1522	922	880	975
24A	831	3.98	1.586	0.643	0.940	0.896	1638	972	926	1032

TABLE 10 (Cont.)

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS
COOLING

RUN NO.	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_c D_K}{K}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1A	1.660	1.074	1.135	1835	1255	1325	1167
2A	1.595	1.068	1.124	1770	1172	1232	1093
3A	1.585	1.066	1.122	1692	1110	1169	1042
4A	1.560	1.064	1.117	1654	1075	1130	1020
5A	1.583	1.066	1.121	1600	1050	1102	985
6A	1.550	1.064	1.116	1522	990	1037	930
7A	1.638	1.072	1.131	2110	1417	1496	1322
8A	1.608	1.069	1.126	2040	1360	1423	1252
9A	1.560	1.064	1.117	2080	1365	1430	1280
10A	1.593	1.067	1.123	3110	2090	2190	1950
11A	1.585	1.066	1.122	2620	1740	1830	1632
12A	1.522	1.061	1.111	2570	1670	1750	1572
13A	1.593	1.067	1.123	2545	1705	1792	1595
14A	1.593	1.067	1.123	2660	1780	1886	1680
15A	1.533	1.062	1.113	2420	1580	1656	1487
16A	1.722	1.079	1.146	2405	1670	1777	1550
17A	1.700	1.077	1.142	2150	1490	1576	1380
18A	1.555	1.064	1.117	2060	1355	1420	1270
19A	1.650	1.072	1.113	1662	1127	1167	1048
20A	1.650	1.072	1.113	1760	1190	1237	1110
21A	1.608	1.079	1.126	1765	1185	1236	1100
22A	1.680	1.075	1.138	1592	1090	1153	1012
23A	1.725	1.080	1.146	1505	1043	1105	963
24A	1.672	1.074	1.137	1425	966	1022	898

TABLE 11

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS
HEATING

RUN NO.	RE. NO. $\times 10^{-2}$	$\frac{c\mu}{K}$	$\left(\frac{c\mu}{K}\right)^{1/3}$	$\left(\frac{\mu_c}{\mu}\right)$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_h D_K}{K}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1B	534	6.96	1.910	0.667	0.945	0.904	1563	776	740	819
2B	514	7.25	1.936	0.652	0.942	0.899	1496	728	695	773
3B	508	7.35	1.944	0.645	0.940	0.896	1093	529	504	562
4B	980	7.14	1.926	0.652	0.942	0.899	2050	1001	957	1063
5B	974	7.19	1.930	0.648	0.941	0.896	1860	907	863	862
6B	928	7.58	1.964	0.628	0.937	0.890	1810	863	820	921
7B	1411	7.30	1.940	0.659	0.943	0.902	2365	1150	1100	1220
8B	1392	7.41	1.950	0.641	0.940	0.895	2265	1093	1040	1162
9B	1363	7.61	1.967	0.617	0.935	0.886	2320	1102	1043	1180
10B	758	6.88	1.901	0.634	0.938	0.892	1740	859	816	915
11B	746	7.00	1.912	0.664	0.944	0.903	1653	817	783	865
12B	699	7.55	1.962	0.631	0.938	0.891	1705	815	775	869
13B	741	7.05	1.918	0.660	0.944	0.901	1772	871	833	922
14B	733	7.14	1.925	0.652	0.942	0.899	1560	763	729	811
15B	706	7.46	1.953	0.638	0.939	0.894	1641	789	751	840
16B	960	7.32	1.942	0.667	0.945	0.904	1973	960	920	1014
17B	970	7.22	1.932	0.665	0.944	0.903	1895	927	886	981
18B	951	7.39	1.948	0.663	0.944	0.903	1882	912	873	966
19B	1482	7.23	1.933	0.665	0.944	0.903	2535	1239	1184	1311
20B	1465	7.32	1.942	0.658	0.943	0.901	2350	1142	1090	1210
21B	1412	7.64	1.970	0.634	0.938	0.892	1543	735	697	784
22B	539	6.89	1.902	0.674	0.946	0.905	1517	755	721	797
23B	530	7.01	1.915	0.664	0.944	0.903	1460	720	689	762
24B	511	7.31	1.941	0.638	0.939	0.894	1535	743	708	790

TABLE 11 (Cont.)

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS
COOLING

RUN NO.	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_c D_K}{K}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{C\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{C\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)}{\left(\frac{C\mu}{K}\right)^{1/3}}$
1B	1.828	1.088	1.163	1648	939	1002	862
2B	1.756	1.082	1.151	1732	969	1030	895
3B	1.745	1.081	1.149	1533	853	907	789
4B	1.770	1.083	1.153	2350	1325	1408	1220
5B	1.756	1.082	1.151	2260	1267	1347	1160
6B	1.690	1.076	1.140	2210	1210	1282	1122
7B	1.735	1.080	1.148	2530	1410	1500	1302
8B	1.700	1.077	1.142	2390	1322	1400	1225
9B	1.655	1.073	1.132	2550	1392	1470	1297
10B	1.815	1.087	1.161	1866	1067	1140	981
11B	1.803	1.086	1.159	1807	1027	1093	944
12B	1.677	1.075	1.138	1762	965	1022	898
13B	1.780	1.084	1.155	1784	1010	1075	930
14B	1.756	1.082	1.151	1835	1032	1096	954
15B	1.700	1.077	1.142	1782	983	1042	912
16B	1.722	1.079	1.146	2165	1203	1276	1115
17B	1.745	1.081	1.149	2085	1166	1240	1080
18B	1.700	1.077	1.142	2055	1137	1205	1055
19B	1.700	1.077	1.142	2604	1450	1540	1348
20B	1.690	1.076	1.140	2595	1440	1523	1283
21B	1.644	1.072	1.132	2475	1347	1420	1255
22B	1.803	1.086	1.159	1492	851	909	785
23B	1.780	1.084	1.155	1505	852	906	786
24B	1.712	1.078	1.144	1419	788	836	730

TABLE 12

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS
HEATING

RUN NO.	RE. NO. $\times 10^{-2}$	$\frac{c\mu}{K}$	$\left(\frac{c\mu}{K}\right)^{1/3}$	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_h D_K}{K}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1C	645	5.72	1.787	0.507	0.910	0.844	1274	650	608	713
2C	608	6.12	1.830	0.522	0.913	0.850	1419	707	660	776
3C	1155	5.86	1.802	0.508	0.910	0.844	1987	1003	931	1103
4C	1120	6.07	1.823	0.561	0.922	0.865	1790	905	850	981
5C	1705	6.09	1.825	0.593	0.930	0.878	2600	1326	1252	1425
6C	1657	6.26	1.841	0.600	0.931	0.880	2290	1158	1095	1243
7C	846	6.00	1.817	0.531	0.915	0.854	1780	896	836	980
8C	820	6.22	1.838	0.560	0.922	0.865	1894	950	892	1030
9C	1645	6.07	1.823	0.572	1.925	0.870	2080	1056	993	1140
10C	1137	5.97	1.813	0.556	0.921	0.864	1968	1000	938	1086
11C	625	5.75	1.791	0.504	0.909	0.842	1515	764	713	846
12C	864	5.86	1.802	0.519	0.912	0.849	1770	896	834	983

TABLE 12 (Cont.)

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS
COOLING

RUN NO.	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_c D_K}{K}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{C\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{C\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)}{\left(\frac{C\mu}{K}\right)^{1/3}}$
1C	1.757	1.082	1.151	1400	848	902	783
2C	1.666	1.074	1.136	1468	862	911	802
3C	1.711	1.078	1.144	2115	1266	1345	1172
4C	1.655	1.073	1.134	2145	1263	1333	1177
5C	1.655	1.073	1.134	2555	1504	1588	1400
6C	1.611	1.069	1.127	2760	1602	1690	1500
7C	1.690	1.076	1.140	1786	1057	1120	980
8C	1.645	1.072	1.133	1795	1047	1108	978
9C	1.633	1.071	1.131	2530	1486	1568	1388
10C	1.666	1.074	1.136	2165	1282	1357	1192
11C	1.722	1.079	1.145	1637	987	1046	914
12C	1.700	1.077	1.142	1896	1132	1200	1052

TABLE 13

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS HEATING

RUN NO.	RE. NO. $\times 10^{-2}$	$\frac{c\mu}{K}$	$\left(\frac{c\mu}{K}\right)^{1/3}$	$\left(\frac{\mu_c}{\mu}\right)$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_h D_K}{K}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1D	16.2	42.10	3.48	0.425	0.887	0.807	758	193	176	218
2D	31.4	36.05	3.30	0.522	0.913	0.850	885	245	228	268
3D	24.1	34.70	3.26	0.514	0.911	0.847	858	240	223	263
4D	----	-----	----	-----	-----	-----	---	---	---	---
5D	139.5	38.25	3.37	0.491	0.905	0.837	2100	563	522	624
6D	119.2	35.80	3.29	0.502	0.908	0.842	1716	474	439	521
7D	84.7	34.85	3.27	0.511	0.910	0.845	1495	416	386	457
8D	101.3	34.85	3.27	0.505	0.909	0.843	1582	440	407	484
9D	65.1	34.05	3.24	0.516	0.912	0.847	1310	369	342	404
10D	48.1	33.40	3.22	0.522	0.913	0.850	1428	404	377	444

RUN NO.	RE. NO. $\times 10^{-2}$	$\frac{c\mu}{K}$	$\left(\frac{c\mu}{K}\right)^{1/3}$	$\left(\frac{\mu_c}{\mu}\right)$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_h D_K}{K}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_h D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1E	140.5	40.00	3.42	0.456	0.896	0.822	1881	492	452	550
2E	117.8	38.20	3.37	0.470	0.900	0.828	1657	442	407	492
3E	91.7	37.50	3.35	0.483	0.903	0.834	1454	392	362	434
4E	75.7	36.50	3.32	0.492	0.906	0.838	1312	358	332	395
5E	58.6	35.70	3.29	0.492	0.906	0.838	1187	327	302	361
6E	43.1	35.20	3.28	0.511	0.910	0.845	1056	293	272	322
7E	26.0	32.20	3.18	0.558	0.921	0.864	1093	316	297	344
8E	214.0	40.00	3.42	0.469	0.899	0.826	2315	608	559	676

TABLE 13 (Cont.)

SUMMARY OF CALCULATIONS - DIMENSIONALESS GROUPS COOLING

RUN NO.	$\frac{\mu_c}{\mu}$	$\left(\frac{\mu_c}{\mu}\right)^{0.14}$	$\left(\frac{\mu_c}{\mu}\right)^{0.25}$	$\frac{h_c D_K}{K}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.14}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)\left(\frac{\mu_c}{\mu}\right)^{0.25}}{\left(\frac{c\mu}{K}\right)^{1/3}}$	$\frac{\left(\frac{h_c D_K}{K}\right)}{\left(\frac{c\mu}{K}\right)^{1/3}}$
1D	2.90	1.161	1.305	369	123	138	106
2D	3.38	1.186	1.356	489	170	200	148
3D	3.50	1.191	1.368	478	175	200	146
4D	----	-----	-----	---	---	---	---
5D	3.41	1.188	1.359	1212	426	489	360
6D	3.73	1.202	1.390	1158	423	490	352
7D	3.86	1.208	1.401	959	355	411	293
8D	3.86	1.208	1.401	1078	399	462	330
9D	4.02	1.215	1.416	838	315	366	259
10D	4.10	1.218	1.423	699	264	309	217

1E	3.86	1.208	1.401	999	353	410	292
2E	4.10	1.218	1.423	830	300	351	246
3E	4.24	1.224	1.435	721	264	308	216
4E	4.30	1.227	1.440	658	247	285	198
5E	4.44	1.232	1.452	561	210	248	170
6E	4.57	1.237	1.462	490	185	218	149
7E	5.04	1.254	1.498	352	139	166	111
8E	3.89	1.209	1.405	1148	406	471	360

APPENDIX C

SAMPLE CALCULATION

SAMPLE CALCULATION

RUN 13A: Hot Water

A. DATA:

Agitator Speed, RPM	- 204
Bulk Liquid Temp., ^{o}F	- 110.0
Hot Water Inlet Temp., ^{o}F	- 161.5
Hot Water Outlet Temp., ^{o}F	- 144.0
Hot Water Flow, Lbs./Min.	- 41.2

B. CALCULATION OF Q:

$$Q = Wc\Delta t$$

$$Q = (41.2)(1.00)(161.5 - 144.0)$$

$$Q = 720 \text{ BTU/MIN.}$$

C. CALCULATION OF U:

$$U = \frac{Q}{A\Delta t_m}$$

$$U = \frac{(720)(60)}{(3.41)(42.1)} = 302 \text{ BTU/HR./SQ.FT./}^{\circ}\text{F}$$

D. CALCULATION OF h_i:

$$h_i = 0.0225 \frac{K}{d_i} \left(\frac{d_i V \rho}{\mu}\right)^{0.8} \left(\frac{c\mu}{K}\right)^{0.3}$$

$$h_i = (0.0225)\frac{(0.382)}{(0.0519)}\left[58600\right]^{0.8}\left[\frac{(2.42)(1.00)(0.424)}{(0.382)}\right]^{0.3}$$

$$h_i = (0.1660)(6520)(1.344)$$

$$h_i = 1458 \text{ BTU}/\text{HR.}/\text{SQ. FT.}/{}^\circ\text{F}$$

E. CALCULATION OF h_c:

$$\frac{1}{h_c} = \frac{1}{U} - \left(\frac{1}{h_i} + \frac{1}{r_w}\right)$$

$$\frac{1}{h_c} = \frac{1}{302} - \left(\frac{D_o}{D_i h_i} + \frac{D_o \ell}{D_{av.} K_m}\right)$$

$$\frac{1}{h_c} = 0.00331 - (0.000925 + 0.000348)$$

$$\frac{1}{h_c} = 0.00204$$

$$h_c = 490 \text{ BTU}/\text{HR.}/\text{SQ. FT.}/{}^\circ\text{F}$$

F. CALCULATION OF DIMENSIONALESS GROUPS:

Nusselt No. $= \dfrac{h_c D_K}{K} = \dfrac{(490)(2.0)}{(0.366)} = 2670$

Prandtl No. $= \dfrac{c\mu}{K} = \dfrac{(1.00)(0.616)(2.42)}{0.366} = 4.05$

Re. No. $= \dfrac{L^2 N \rho}{\mu} = \dfrac{(11.01)(204)(61.86)}{(0.616)} = 226000$