K-1897

PRESSURE DROP AND HEAT TRANSFER CHARACTERISTICS OF REFRIGERANT-114 BOILING IN A HORIZONTAL TUBE

W. R. Brock
Special Projects Department
Gaseous Diffusion Development Division

MASTER

October 1977

UNION CARBIDE

OAK RIDGE GASEOUS DIFFUSION PLANT
OAK RIDGE, TENNESSEE

prepared for the **U.S. ENERGY RESEARCH AND DEVELOPMENT ADMINISTRATION**
under **U.S. GOVERNMENT** *Contract W-7405 eng 26*

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*This report was prepared as a thesis presented for the Master of Science Degree at the University of Tennessee, Knoxville in August 1977.

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ABSTRACT

Research was undertaken to study the two-phase pressure drop and heat transfer characteristics of R-114 flowing in a horizontal tube. Flow regime studies were also performed as a part of this effort. Boiling temperature was varied from 100 to 250°F and mass flux was varied from 0.12 to 3.4 x 10^6 lbm/hr-sq ft.

No single method for predicting flow regime was found to be accurate over the range of conditions studied. The method proposed by Knowles was the best available for determining the stratified and intermittent flow regimes, and the analysis of Quandt was the most reliable in predicting the annular flow regime.

Four correlations gave promising results when compared to the experimental, frictional pressure drop data. The Martinelli-Nelson correlation agreed fairly well with all of the data obtained. The degree of data scatter was, however, fairly large. Two correlations proposed by Chisolm which include a mass flux effect gave slightly better results when compared with the experimental data. The correlation proposed by Baroczy, generally, gave good results when applied to the medium-to-high mass flux data.

The experimental heat transfer coefficient data could not be predicted accurately by a single correlation. The original Chen correlation offered the best agreement of any correlation considered. The degree of agreement between experimental data and correlation was improved by applying the original Chen correlation to the low mass flux data and Jallouk's modification to the remaining data.

Recommendations for further study are also given.

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LIST OF SYMBOLS

b,c	Exponents used in Equation (2-20).
B	Parameter defined by Chisholm [30], Table II-2.
Bo	Boiling number, ϕ/Gh_{fg}.
c_p	Specific heat at constant pressure.
C	A constant given by Equation (2-10).
Ca	Constant proposed by Gilmour [41], Equation (2-24).
C_{sf}	Empirical constant suggested by Rohsenow [36].
C_2	A constant defined by Chisholm [29], Equation (2-11).
$\bar{C}$	A parameter defined by Chisholm [29], Equation (2-14).
D	Diameter.
f	Fanning friction factor.
F	Multiplication factor.
f_1, f_2	Functions, Equation (2-19).
F_D	Parameter in Taitel and Dukler's [14] Map.
g	Acceleration due to gravity.
G	Mass flux based on entire tube cross sectional area.
Go	Mass flux parameter used by Gilmour [41], Equation 2-24.
h	Heat transfer coefficient, enthalpy.
h_{fg}	Latent heat of vaporization.
J	Mechanical equivalent of heat.
k	Thermal conductivity.
K_D	Parameter in Taitel and Dukler's [14] Map.
K_F	Load factor, Equation (2-33).
L	Length, usually tube length.
n	Exponent used by Chisholm [30], Equation (2-16).
N_{Fr}	Froude number.

N_{Nu}	Nusselt number.
N_{Pr}	Prandtl number.
N_{Re}	Reynolds number.
N_{We}	Weber number.
p	Pressure.
p_r	Reduced pressure.
Q_{loss}	Heat loss per zone.
R	Hold-up.
S	Suppression Factor.
T	Temperature.
T_D	Parameter in Taitel and Dukler's [14] Map.
T_o	A parameter defined by Chisholm [29], Equation (2-15).
v	Specific volume.
V	Superficial velocity.
x	Quality. Defined in the saturated region as the fraction of the total flow by weight which is vapor. In the subcooled region, it is $(h - h_f)/h_{fg}$. In the superheated region, it is $(h - h_g)/h_{fg})$.
x_{avg}	Average quality.
X	Martinelli parameter.
z	Coordinate in the downstream direction.
Γ	Parameter defined by Chisholm [30], Equation (2-17).
Δp_s	Difference between saturation pressure based on heater temperature and that based on the saturation temperature, Equation (2-21).
ΔT_s	Heating surface temperature less saturation temperature, Equation (2-18).
ε	Chawla's [28] two-phase flow parameter.
λ	A parameter defined by Chisholm [29], Equation (2-11).
μ	Dynamic viscosity.
ρ	Density.

σ	Surface tension.
σ'	Standard deviation.
ν	Kinematic viscosity.
ϕ	Heat flux.
ϕ_B	Heat flux due to boiling effects.
ϕ_C	Heat flux due to forced convection.
Φ	Square root of two-phase multiplier.
Ψ	A parameter defined by Chisholm [29], Equation (2-13).

Subscripts

F	Frictional component.
f	Liquid.
fa	Liquid alone.
fo	Liquid only.
g	Vapor.
ga	Vapor alone.
go	Vapor only.
i	Inlet.
o	Outlet.
PB	Pool boiling.
S	Slip.
tp	Two-phase.
tt	Turbulent-turbulent case.

CHAPTER I

INTRODUCTION AND STATEMENT OF PROBLEM

A. INTRODUCTION

It has long been recognized that the transfer of heat to a boiling or evaporating fluid is a very efficient process. This is due to the fact that large heat fluxes can be transferred through small areas with fairly small temperature differences. This helps reduce the size and cost of heat exchange equipment. As a result of these facts, the boiling process is widely used in industry today.

Although there are two basic types of boiling, pool boiling and flow boiling, flow boiling will be the subject of this study. Consequently, discussions for the most part, will pertain to this type of boiling, and pool boiling will be discussed only when applicable to flow boiling analyses.

Water is definitely the fluid that has been most studied in boiling experiments; however, in recent years, emphasis has been placed on boiling refrigerants. This was brought about for at least three reasons: (1) The refrigeration and air conditioning industries, who use refrigerants exclusively, need the capability of predicting flow regime, boiling coefficients, and pressure drop for a variety of refrigerants over a wide range of conditions. (2) Many industrial concerns must use refrigerants because they employ processes that are incompatible with water. Many of these installations have such large cooling requirements that the need to predict refrigerant boiling characteristics under a wide range of conditions is mandatory.

(3) Many refrigerants, when subjected to moderate pressures, boil at temperatures below ambient. Thus, it is possible to study their boiling characteristics without constructing elaborate and costly experimental facilities. With suitable modeling techniques, data obtained in this manner can be useful in predicting the boiling characteristics of other fluids.

Since the author is presently employed at the Oak Ridge Gaseous Diffusion Plant, the main scope of this effort will be directed toward providing information that is applicable to this type of cooling system which employs Refrigerant-114 (R-114). On the other hand, data will be taken over a wide enough range of conditions to be useful in other applications.

In the cost-conscious society of today, it is very important that equipment perform reliably at a minimum cost. Thus, in order to truly optimize, the designer of a system or component must be able to accurately predict both performance and cost. This is necessarily true of the engineer who endeavors to design a boiler for industry or for any other concern. He must be able to accurately predict cost, but even before that, it is essential that he be able to calculate the following parameters for a given boiler duty:

1. Pressure drop and
2. Heat transfer coefficients.

With this information in hand, the engineer can *size* the boiler to transfer the required amount of heat. At this point, total cost can be estimated; however, it should be emphasized that the cost is only as accurate as the boiler *size* upon which it is based.

In order to assure that boiler performance is determined accurately, one must compute the parameters listed above for the two-phase fluid that is employed. The two-phase pressure drop is needed to determine the required pumping power. The heat transfer coefficient of the two-phase fluid is related to the amount of heat that can be transferred and to the tube wall temperature.

It is unfortunate, but the analysis of two-phase systems is much more complex than the analysis of their single-phase counterparts. The presence of both a liquid and a vapor phase causes the necessary equations and boundary conditions that adequately describe the situation to become numerous and complicated. Many empirical relationships can be found in the literature which attempt to predict pressure drop and heat transfer characteristics for two-phase fluids; however, they are nearly always based on data taken under specific conditions. Thus, it can be stated that few generalized relationships exist today which allow the engineer to determine the performance of a two-phase system with confidence. Consequently, there is a definite need for boiling studies of this nature. The results will serve to extend the available boiling data on refrigerants, as well as provide specific information for R-114 boiling in a horizontal tube. This latter piece of information will, undoubtedly, prove to be very helpful in development efforts at installations, such as the United States Gaseous Diffusion Plants, which actually use R-114.

B. STATEMENT OF PROBLEM

Because of reasons discussed in the previous section, this program to experimentally and analytically study the heat transfer and pressure-drop characteristics of R-114 boiling in a horizontal tube was initiated. This program required effort in the following three areas:

1. Flow Regime Studies - The flow regimes encountered in a two-phase mixture of R-114 flowing in a horizontal tube with heat transfer were studied. This was accomplished by visually observing the flow through a sight glass positioned at the exit end of a copper tube. Videotape pictures were also employed as an aid in clarifying some of the flow regimes. Observations were made over a range of quality, mass flux, heat flux, and boiling temperature. These ranges are defined in Table I-1. The results were compared to information available in the literature.
2. Two-Phase Pressure Drop Studies - Pressure drop was experimentally measured at 2-ft intervals along the test section. Thus, it was possible to compare the resulting data with the well-known Martinelli-Nelson [1] and Homogeneous [2] models, as well as other equations derived for individual flow regimes. The mass flux range given in Table I-1 is also applicable here.
3. Heat Transfer Coefficient Studies - Two-phase heat transfer coefficients were obtained for the range of heat flux, mass flux, and temperature stated in Table I-1. The data obtained were compared to predictions obtained from correlations available in the literature.

Table I-1

RANGE OF VARIABLES STUDIED

Parameter	Range
Quality	0.0 - 1.0
Mass Flux, lbm/hr-sq ft	0.1×10^6 - 3.4×10^6
Heat Flux, Btu/hr-sq ft-°F	5,000 - 16,000
Boiling Temperature, °F	100 - 250

CHAPTER II

LITERATURE SURVEY

A. FLOW REGIMES

The analysis of any two-phase system is greatly simplified when the flow pattern or flow regime that exists can be accurately predicted from certain available parameters. In fact, it may be stated that knowledge of the flow regime present in a two-phase system is at least as important as knowing whether the flow is turbulent or laminar in a single-phase system.

Many experimenters have identified the flow regimes occurring in cocurrent, two-phase flow in horizontal tubes. The generally accepted flow regimes are those reported by Alves [3], based on results from adiabatic air-water and air-oil systems. These different flow regimes are depicted schematically in Figure II-1 and discussed below:

Bubbly Flow. The bubbly flow regime is characterized by vapor bubbles which tend to travel in the upper portion of the tube. The vapor bubbles travel at essentially the same velocity as the liquid.

Plug Flow. The plug flow regime is characterized by alternate plugs of liquid and vapor which travel along the upper portion of the tube.

Stratified Flow. The stratified flow regime is characterized by liquid flowing along the bottom of the tube with the vapor flowing on top of the liquid. The interface between the liquid and vapor is relatively smooth.

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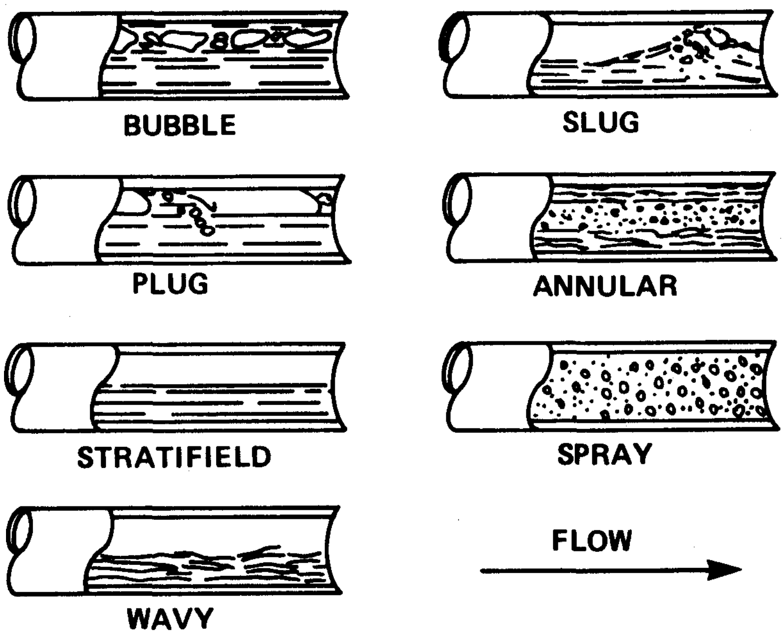

FIGURE II-1. FLOW REGIMES OBSERVED DURING HORIZONTAL TWO-PHASE FLOW IN A TUBE, ALVES [3].

Wavy Flow. The wavy flow regime is similar to the stratified flow regime. However, the vapor travels at a higher velocity than the liquid and the interface becomes disturbed by waves traveling in the direction of flow.

Slug Flow. The slug flow regime occurs at higher vapor velocities than wavy flow. A wave is periodically picked up by the rapidly moving vapor so that a frothy slug is formed. This forthy slug passes through the pipe at a much higher velocity than the liquid velocity.

Annular Flow. The annular flow regime is characterized by a thin film of flowing liquid which moves down the inside wall of the tube. The vapor and some entrained liquid flow at a high velocity in the central core. Alves [3] indicated that it was difficult, in some cases, to differentiate between annular flows and bubbly flows because in both cases many fine bubbles were dispersed throughout the liquid near the inside tube wall. Alves [3] also presented a flow map which he was able to generate from observations made and data taken during this investigation. This map is shown in Figure II-2.

Baker [4] prepared a flow map based on the data of Jenkins [5], Gazely [6], Avles [7], and Kosterin [8]. This map, shown in Figure II-3, is probably the most widely used vehicle today for the prediction of flow regimes in horizontal two-phase flow. As the figure implies, Baker [4] considered the flow regime to be a function of

$$\frac{x\ G}{\left(\frac{\rho_g}{0.075} \cdot \frac{\rho_f}{62.3}\right)^{1/2}}$$

and

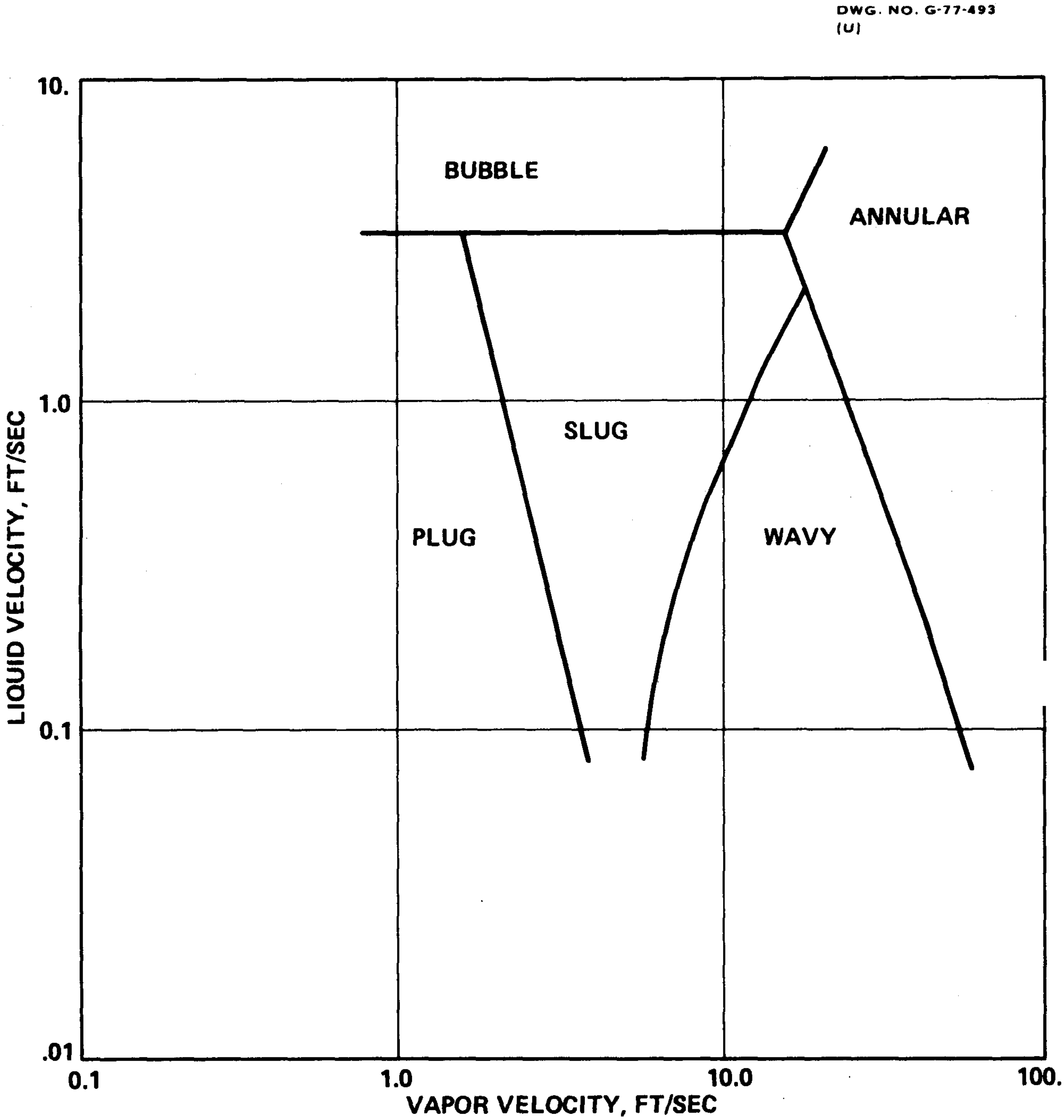

FIGURE II-2. FLOW REGIME MAP, ALVES [3].

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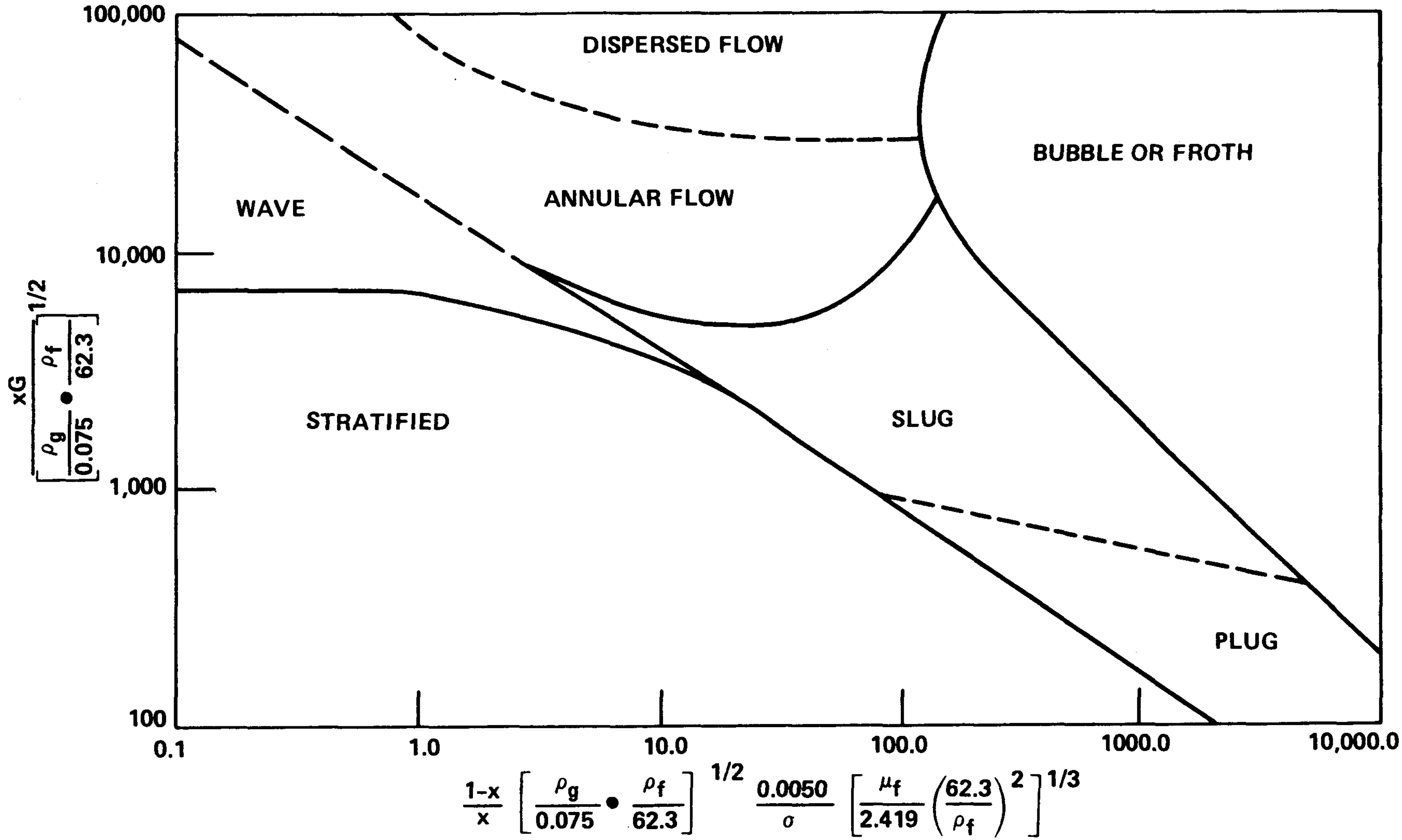

FIGURE II-3. FLOW REGIME MAP, BAKER [4].

$$\frac{1-x}{x}\left(\frac{\rho_g}{0.075}\cdot\frac{\rho_f}{62.3}\right)^{1/2}\left(\frac{0.0050}{\sigma}\right)\left[\frac{\mu_f}{2.419}\left(\frac{62.3}{\rho_f}\right)^2\right]^{1/3}$$

where

x is the vapor quality;

G is the mass flux based on the entire tube cross-sectional area, lbm/hr-sq ft;

ρ_g is the vapor density, lbm/cu ft;

ρ_f is the liquid density, lbm/cu ft;

σ is the surface tension, lbf/ft; and

μ_f is the liquid viscosity, lbm/ft-hr.

Quandt [9] studied vapor-liquid flow in ducts under conditions controlled by pressure gradient, gravity, and surface tension. He was able to establish that bubbly and annular flow were controlled by pressure gradient, and that plug, slug, wave, and stratified flows were controlled by gravitational forces. Quandt [9] also concluded that the magnitude of the Froude and Weber numbers could be used to predict the flow regime present. The Quandt [9] map is shown in Figure II-4.

Knowles [10, 11], who found certain inaccuracies associated with the Baker [4] map, developed another map which is shown in Figure II-5. Knowles [10, 11] suggested the following coordinates:

$$N_{We_{tp}} = \frac{\left(\rho_f\ V_f^2\ D\ R_f\right)^{1/2}}{\sigma} + \frac{\left[\rho_g\ V_s^2\ D\ (1-R_f)^{1/2}\right]}{\sigma} \tag{2-1}$$

and

$$N_{Re_{tp}} = W_t\ R_f^2/D\mu_{tp} \tag{2-2}$$

where

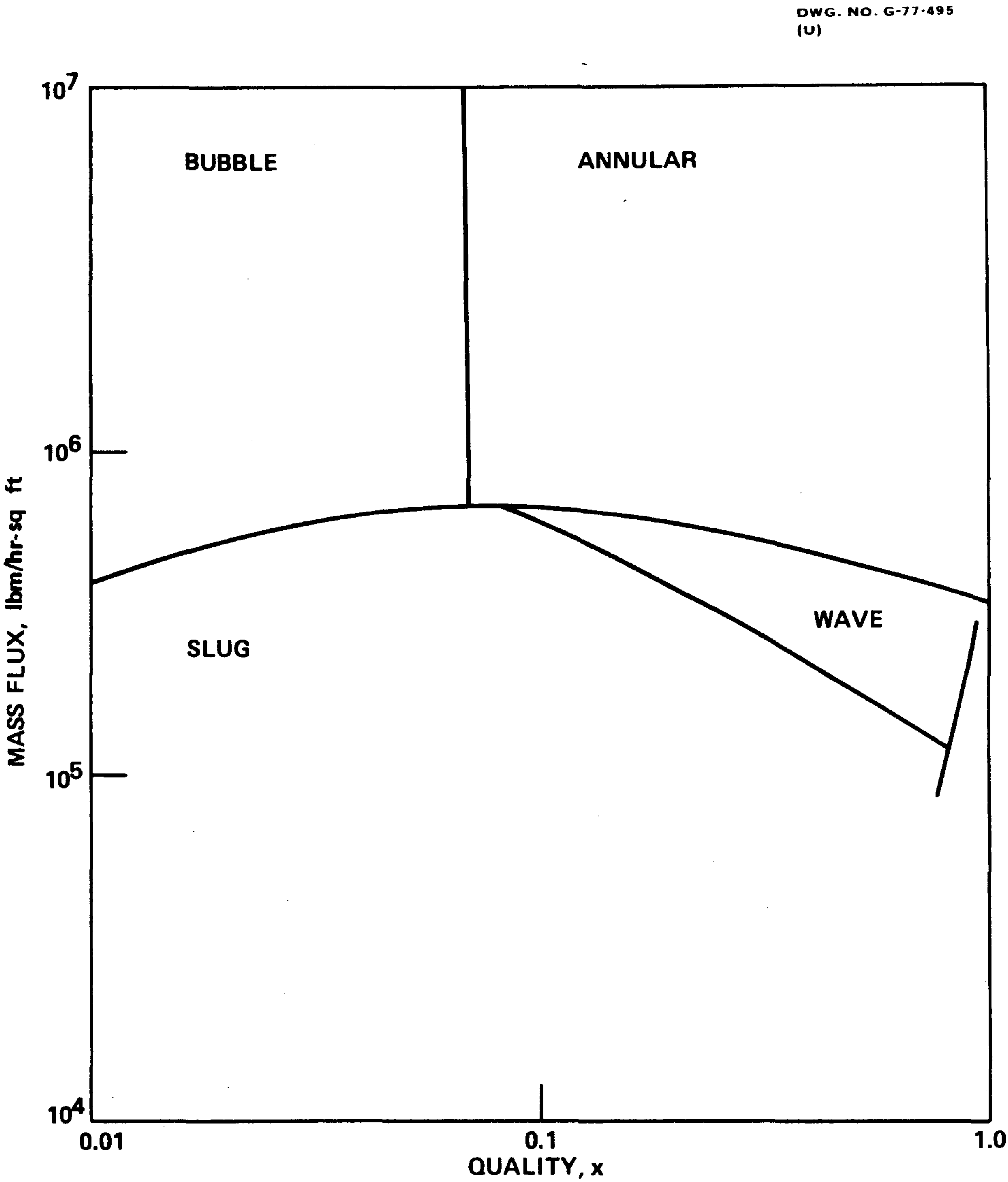

FIGURE II-4. FLOW REGIME MAP, QUANDT [9].

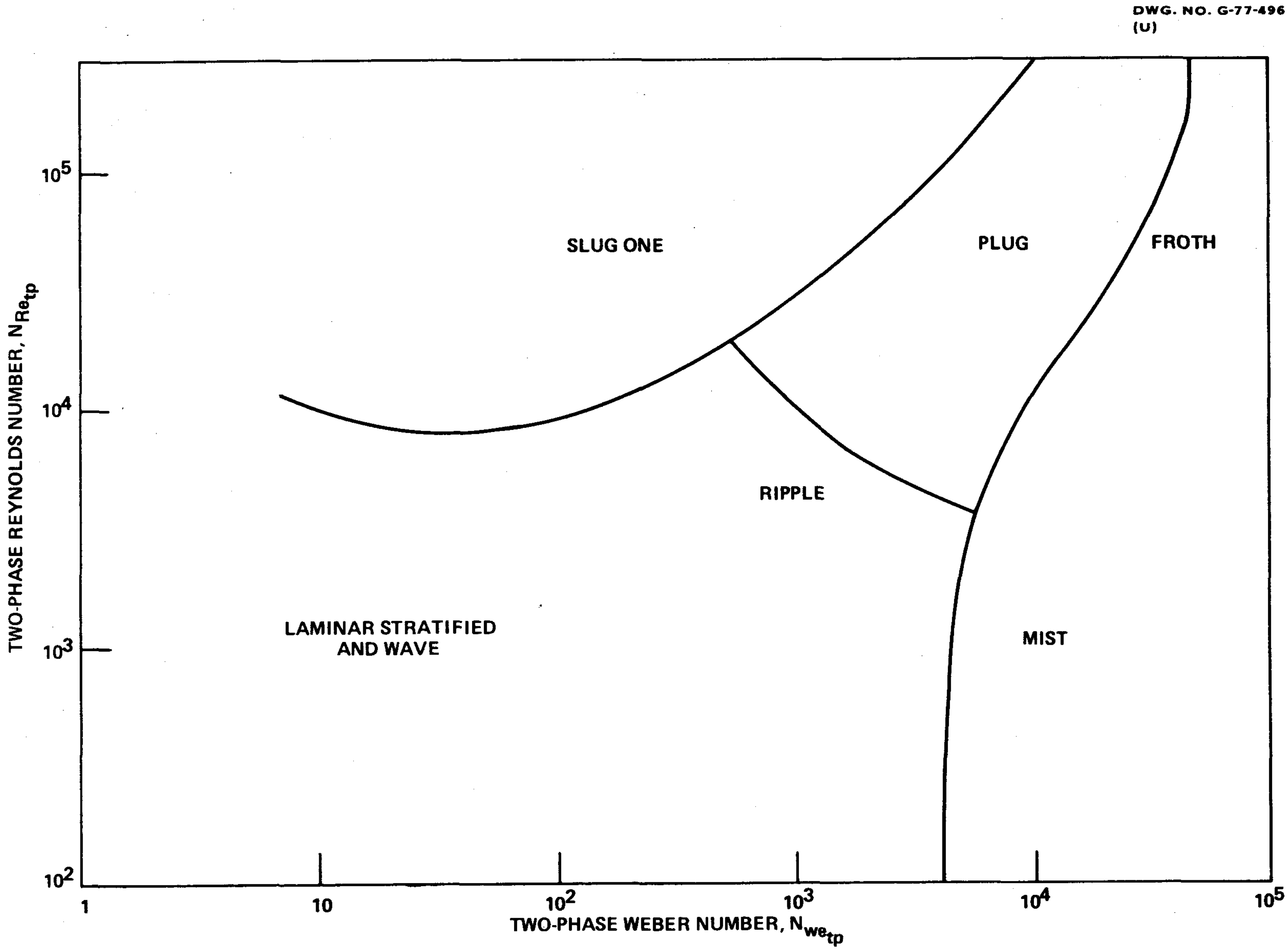

FIGURE II-5. FLOW REGIME MAP, KNOWLES [10, 11]

$N_{We_{tp}}$ is the two-phase Weber number;

$N_{Re_{tp}}$ is the two-phase Reynolds number;

ρ_f is the liquid density, lbm/cu ft;

ρ_g is the vapor density, lbm/cu ft;

μ_{tp} is the two-phase viscosity, cp;

σ is the surface tension, dynes/sq cm;

D is the tube diameter, ft;

V_f is the liquid velocity, ft/sec;

V_g is the vapor velocity, ft/sec;

V_s is the slip velocity, ft/sec;

R_f is the liquid holdup; and

W_t is a total flow rate, lbm/hr.

Knowles [10, 11] defined two-phase viscosity, μ_{tp} as

$$\mu_{tp} = \left(\mu_f\right)^{R_f} \left(\mu_g\right)^{1-R_f} \tag{2-3}$$

where

μ_f is the liquid viscosity, cp; and

μ_g is the vapor viscosity, cp.

He also defined the slip velocity V_s as

$$V_s = V_g - V_f. \tag{2-4}$$

It should be pointed out that Knowles' [10, 11] correlation is inconsistent with Dukler, et al. [12] similarity analysis, and that any extrapolations beyond the recommended ranges ($1 \leq N_{We_{tp}} < 10^5$ and $10^3 \leq W_{Re_{tp}} < 10^6$) should be carefully considered. An obvious difference is in the definition of two-phase viscosity. DeGance and Atherton [13] provided an excellent technical review of both Baker's [4] and Knowles' [10, 11] work. It should also be noted that the appropriate conversion factors must be applied to Equations (2-1) and (2-2) for them to be dimensionless.

A recent publication by Taitel and Dukler [14] discussed another generalized flow map. They considered five basic flow regimes: smooth stratified, wavy stratified, intermittent (slug and plug), annular, and bubble. Starting with the condition of stratified flow, equations for the transition between the various flow regimes were developed. The Martinelli [1] parameter, X, was employed as the abscissa in their map, as shown in Figure II-6 and Table II-1. Taitel and Dukler [14] compared their map with a large quantity of flow regime data for adiabatic air-water systems. Their results were quite good.

The above discussion has been concerned with adiabatic test results. Unfortunately, only a limited amount of work has been done on the analysis of flow regimes encountered in horizontal, two-phase flow with heat transfer. There is some question as to the applicability of flow maps developed under adiabatic conditions to diabatic two-phase systems. Several investigators [15-18] have studied this phenomenon with conflicting results.

Berenson and Stone [19] conducted tests with Refrigerant-113 (R-113) boiling in a horizontal Pyrex tube. Through high-speed photography they were able to observe the bubbly, intermittent, annular, and mist flow regimes. Apparently, the stratified flow regime was never observed. Unfortunately, Berenson and Stone [19] did not compare their observed flow regimes with any of the available flow maps.

Johnson and Abou-Sabe [20] also studied the flow regimes existing in a 0.870-in. horizontal tube with an imposed heat flux. For a

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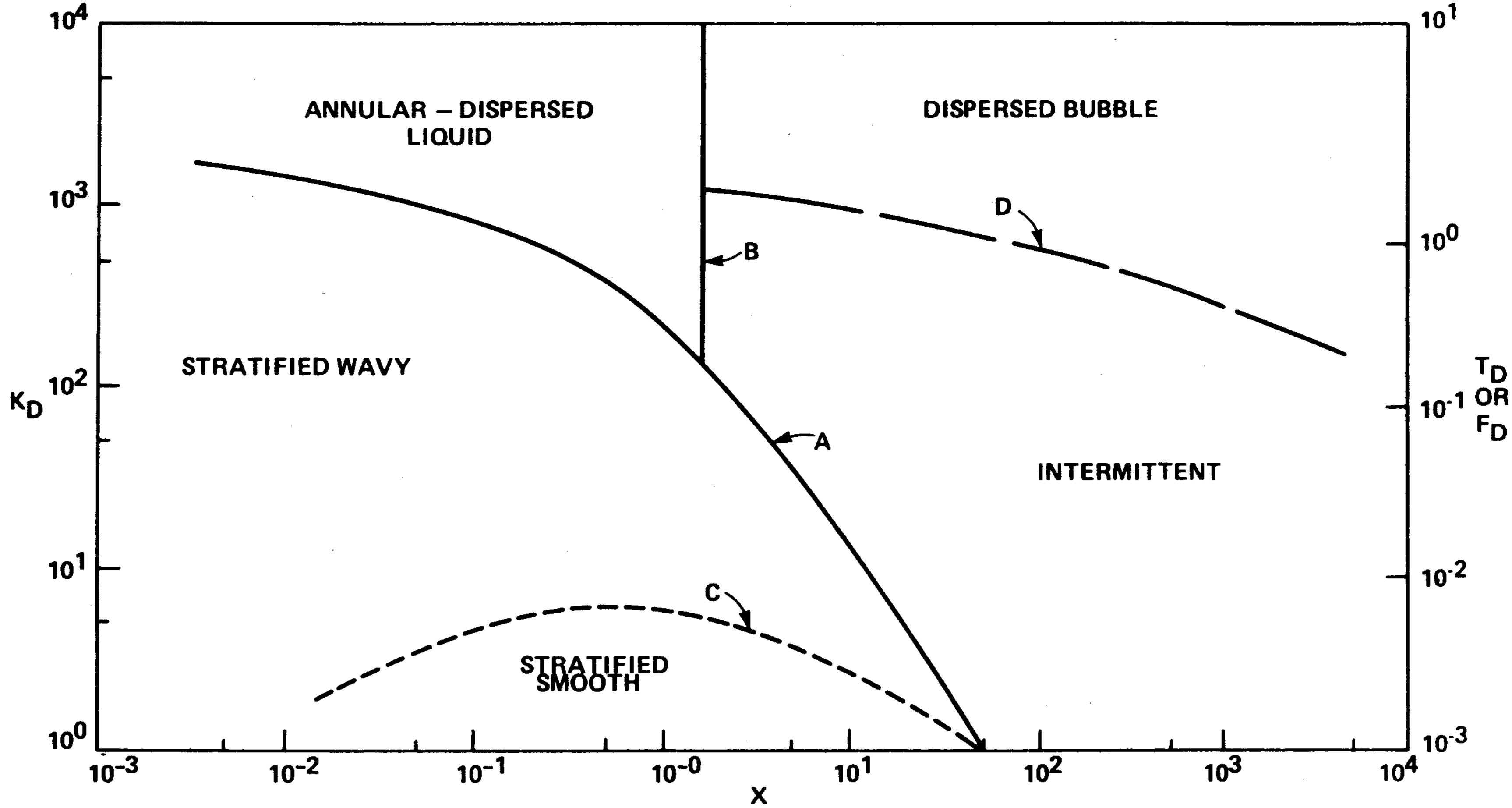

FIGURE II-6. FLOW REGIME MAP, TAITEL AND DUKLER [14].

Table II-1

DEFINITION OF PARAMETERS USED IN TAITEL AND DUKLER'S [14] FLOW MAP FOR HORIZONTAL FLOW

Curve:	A and B	C	D
Coordinates:	F_D vs X	K_D vs X	T_D vs X

where:

$$T_D = \left[\frac{\left(\frac{dp_F}{dz}\right)_{fa}}{(\rho_f - \rho_g)\, g} \right]^{1/2}$$

$$F_D = \sqrt{\frac{\rho_g}{\rho_f - \rho_g}}\ \frac{V_{ga}}{\sqrt{Dg}}$$

$$K_D = \left[\frac{\rho_g\, V_{ga}^2\, V_{fa}}{(\rho_f - \rho_g)\, g\, \nu_f} \right]^{1/2}$$

$$X = \left[\frac{(dp_F/dz)_{fa}}{(dp_F/dz)_{ga}} \right]^{1/2}$$

NOTE:

z is the coordinate in the downstream direction,

g is the acceleration due to gravity, and

ν_f is the liquid kinematic viscosity.

All other parameters were defined previously.

range of air and water flow rates, they were able to generate a flow map which incorporated their own work with work done by others. This map is shown in Figure II-7. The map is not of a generalized nature which makes it of limited value. When Johnson and Abou-Sabe [20] did their work, no generalized flow map like that of Baker [4] was in existence; thus, comparison with available results was limited to that which was done in their publication.

Bergles, et al. [21] conducted tests with boiling water in a horizontal stainless steel tube using an electric probe located at the end of a heated section. Their data indicated that boiling water flow regimes cannot be predicted by conventional adiabatic flow regime plots.

Other investigators have studied this topic, and Hosler [22] provides an excellent review of the more important work done through 1966.

In conclusion, it may be stated that the flow regimes existing in horizontal two-phase flow have been studied by many individuals. Unfortunately, most of the results pertain to adiabatic systems and are not of a generalized nature. Only the results of Baker [4], Quandt [9], and Knowles [10, 11] have possible application to refrigerant systems. Results from diabatic tests are meager, but they do provide sufficient evidence that adiabatic flow maps may not be applicable to systems with heat transfer. The location of the flow regime observation section may also have an effect on the results obtained from a diabatic system, i.e., is the observation section heated or adiabatic? Other variables, such as tube size, also play an important part.

It is likely that no single flow map can be developed which will accurately predict flow regimes under a wide variety of conditions for

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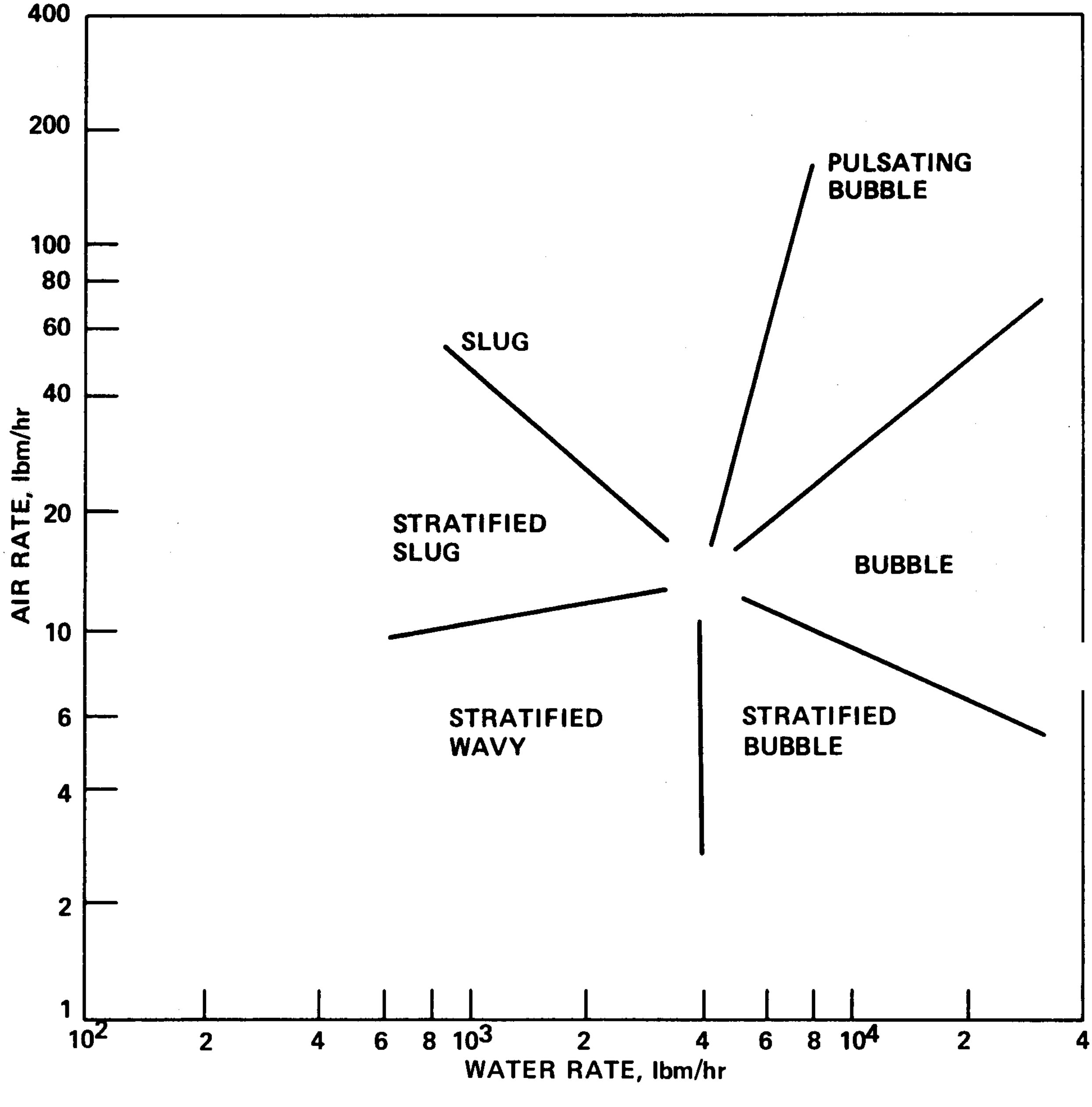

FIGURE II-7. FLOW REGIME MAP, JOHNSON AND ABOU-SABE [20].

a large number of working fluids; however, it does seem plausible that a map can be developed which will greatly reduce the uncertainties that exist as far as refrigerants are concerned.

B. PRESSURE DROP

The ability to predict pressure drop in a two-phase system is very important, since the pumping cost associated with a heat exchanger can be greatly influenced by this parameter. Several individuals have proposed models based on their own experimental data which attempt such a prediction. The following text considers some of the more well-known and widely used correlations based on refrigerant and nonrefrigerant data. The discussion is centered around frictional pressure loss, although losses due to other effects are mentioned as needed.

One of the first attempts made at predicting two-phase pressure drop resulted in a model proposed by Lockhart and Martinelli [23]. The Lockhart-Martinelli [23] or L-M model has been successful in correlating a substantial amount of two-phase, two-component data for isothermal flow. The model was based on separated flow theory, and four basic flow regimes were defined. These flow regimes were designated viscous-viscous, viscous-turbulent, turbulent-viscous, and turbulent-turbulent, depending on the Reynolds number calculated for each individual phase when considered to pass alone through the tube or channel. Thus, for the turbulent-turbulent case, both phases would have Reynolds numbers in excess of 2,000 when flowing alone. L-M [23] postulated that the frictional pressure drop of the two phases would be equal, regardless of flow pattern details. This, of course, is true for only horizontal flow with negligible pressure loss due to acceleration. They correlated their results with the Martinelli parameter, X, which was defined as

$$X^2 = \left(\frac{dp_F}{dz}\right)_{fa} \left(\frac{dp_F}{dz}\right)_{ga} \quad (2\text{-}5)$$

For the turbulent-turbulent case it can be shown that

$$X_{tt} = \left(\frac{1-x}{x}\right)^{0.9} \left(\frac{\rho_g}{\rho_f}\right)^{0.5} \left(\frac{\mu_f}{\mu_g}\right)^{0.1} \quad (2\text{-}6)$$

The L-M [23] model is based on data taken at essentially atmospheric pressure, and it includes no provision for interaction between the two phases. These can be serious limitations under certain conditions, as pointed out by several investigators, including Baroczy [24] and Collier [2]. Despite limitations, the L-M [23] model has provided satisfactory results in many cases as shown by Dukler, et al. [25].

Martinelli and Nelson [1], considering only the turbulent-turbulent case, extended the L-M model by including a pressure dependence and allowing for pressure losses due to acceleration and gravitational effects. The M-N [1] correlation is applicable to forced convective, two-phase flow in any orientation when both liquid and vapor flow are turbulent.

The Homogeneous model, as discussed by Collier [2], is another correlation which attempts to predict two-phase pressure drop. In this model, the two phases are considered to flow as a single fluid which possesses mean fluid properties. Thus, the two phases flow with the same velocity so that slip between the phases is disregarded.

Generally speaking, when comparing the Homogeneous [2] and M-N [1] models, it is reasonable to say that the Homogeneous model is probably more accurate at high flow rate, low quality conditions where bubbly or dispersed flow probably exists. On the other hand, in the annular flow regime at low flows and high qualities, the M-N [1] model is most likely superior. For other conditions, actual pressure losses probably lie between predictions from the two models. This so-called mass flux effect has been observed by many investigators, such as Muscettola [26] and Sher and Green [27] who studied steam-water flow in a vertical, rectangular channel. Their data are plotted in Figure II-8. In the figure, the parameter Φ^2_{fo} is plotted against vapor quality, x. The two-phase frictional multiplier, Φ^2_{fo}, is defined as the ratio of the two-phase frictional pressure gradient to the calculated single-phase frictional pressure gradient that would exist if *only* liquid were flowing. This parameter is similar to the Martinelli multiplier for the turbulent-turbulent case, Φ^2_{ftt}. The difference is that Φ^2_{ftt} is defined as the ratio of the two-phase frictional pressure gradient to the calculated single-phase frictional pressure gradient based on the liquid flowing *alone*. It can be shown that the two parameters are related by

$$\Phi^2_{fo} = \Phi^2_{ftt}\ (1-x)^{1.8} \tag{2-7}$$

As shown by Figure II-8, the Homogeneous [2] model is more accurate for flow rates above about 1.5 x 10^6 lbm/hr-sq ft and the M-N model [1] is superior for flows less than this.

Baroczy [24] was one of the first people to address the mass flux effect and provide a possible solution. He proposed a correlation

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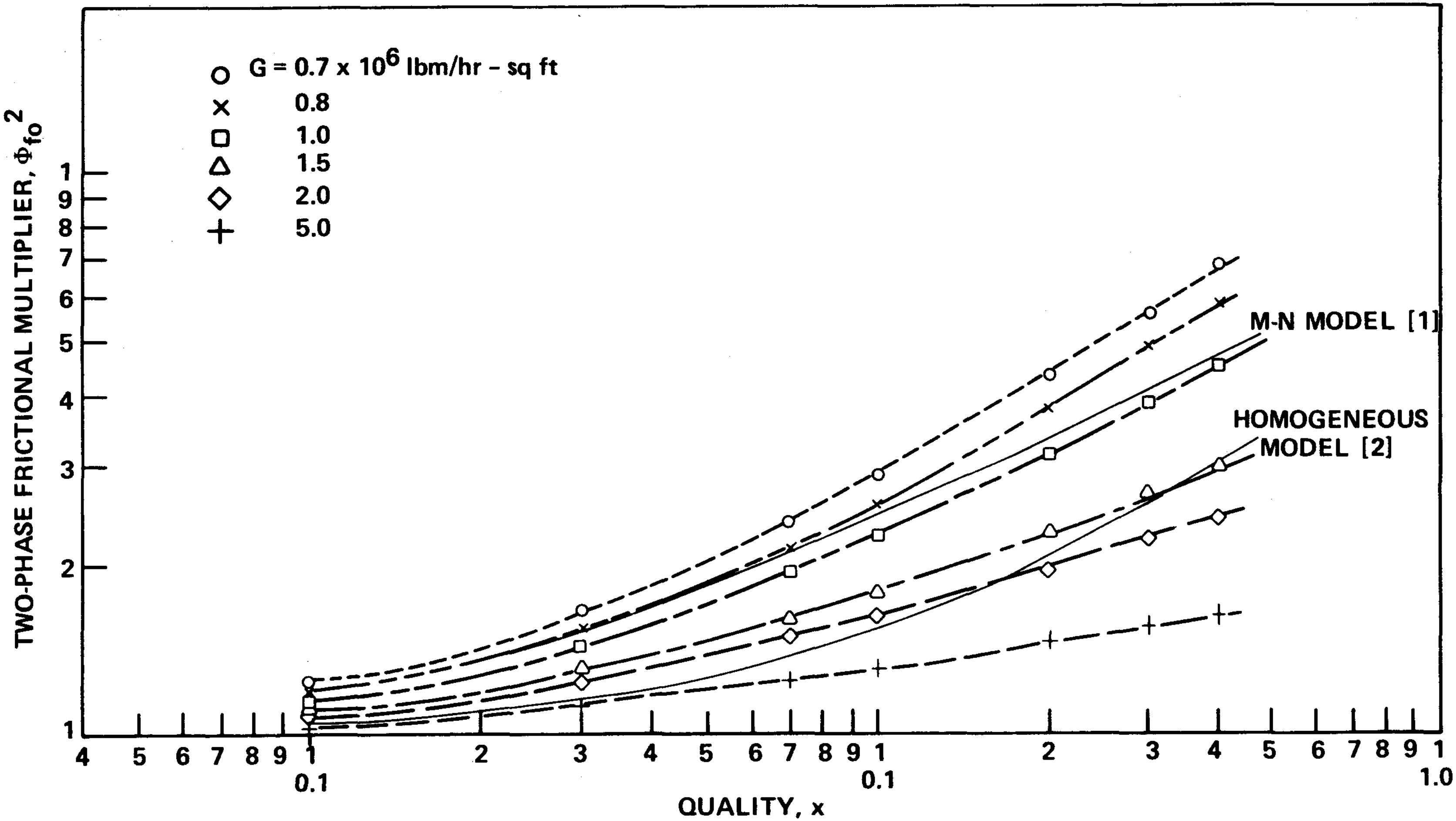

FIGURE II-8. DATA OBTAINED BY SHER AND GREEN [27].

which considered fluid properties, vapor quality, and mass velocity. Through this correlation he was able to predict two-phase frictional pressure drop for both one- and two-component flow. Baroczy [24] modified the M-N model [1] by introducing a property index $(\mu_f/\mu_g)^{0.2}/(\rho_f/\rho_g)$ to account for pressure dependence. The two-phase frictional multiplier, Φ^2_{fo}, was plotted versus the property index for a mass flux of 1.0 x 10^6 lbm/hr-sq ft with quality as a parameter. A second plot was constructed which could be used to correct the two-phase frictional multiplier, Φ^2_{fo}, for mass fluxes of 0.25 x 10^6, 0.5 x 10^6, 2.0 x 10^6, 3.0 x 10^6 lbm/hr-sq ft. In comparing his correlation with two-phase pressure drop data for several different fluids and fluid combinations, Baroczy [24] achieved good agreement. One possible limitation of Baroczy's [24] method is that some of the graphical results would be difficult to computerize.

Chawla [28] also studied two-phase pressure drop in horizontal tubes. From studies with Refrigerant-11 (R-11), at relatively low temperatures, he proposed a correlation for calculating two-phase frictional pressure gradient which included flow pattern influence, roughness of the liquid-vapor interface, and momentum exchange between the two phases. Chawla [28] defined a two-phase flow parameter, ε, shown for varying degrees of pipe roughness, k_o/D, in Figure II-9, and proposed that this parameter would account for the above-mentioned effects. Chawla [28] developed the following equation, which incorporates ε, for

$$\left(\frac{dp_{tp}}{dz}\right)_F = \left[\frac{0.3164}{(GD/\mu_g)^{0.25}}\right]\left(\frac{G^2\ x^{1.75}}{2D\ \rho_g}\right)\left[1 + \frac{1-x}{x\varepsilon(\rho_f/\rho_g)}\right]^{2.375} \qquad (2\text{-}8)$$

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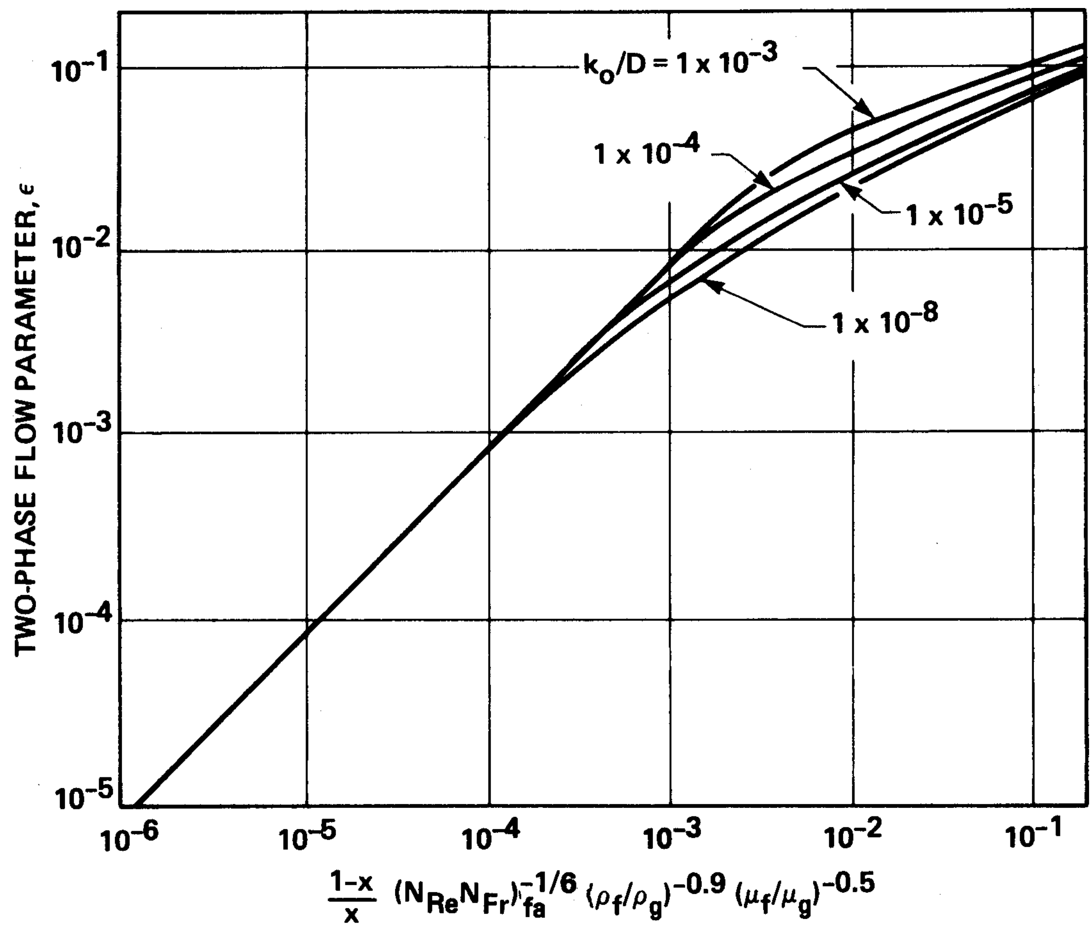

FIGURE II-9. GENERAL CORRELATION OF THE TWO-PHASE FLOW PARAMETER, ϵ, CHAWLA [28] the two-phase frictional pressure gradient. Chawla [28] achieved good results when applying this equation to low-temperature (∿10°C) R-11 data.

In recent years Chisholm [29, 30] has published a large number of papers on two-phase pressure drop. His basic approach has been to modify the L-M model [23] to account for mass flux effects and momentum transfer between the two phases. L-M [23] were able to develop the following expression for Φ^2_{ftt}, the two-phase frictional multiplier:

$$\Phi^2_{ftt} = 1 + \frac{C}{X} + \frac{1}{X^2} \tag{2-9}$$

where C was a constant that depended on flow regime. In a recent publication, Chisholm [29] suggested that

$$C = [\lambda + (C_2 - \lambda)\ (v_{fg}/v_f)^{0.5}]\ [(v_g/v_f)^{0.5} + (v_f/v_g)^{0.5}] \tag{2-10}$$

where

$$\lambda = 0.75,$$
$$C_2 = 1.47 \times 10^6/G, \tag{2-11}$$

and v_{fg} is the latent specific volume, cu ft/lbm. If $C_2 > 4.0$, then $C_2 = 4.0$. At mass velocities exceeding 1.47×10^6 lbm/hr-sq ft, Chisolm [29] suggested that

$$\Phi^2_{ftt} = \left(1 + \frac{\bar{C}}{X} + \frac{1}{X^2}\right)\psi \tag{2-12}$$

where

$$\psi = \left(1 + \frac{C}{To} + \frac{1}{To^2}\right) \Big/ \left(1 + \frac{\bar{C}}{To} + \frac{1}{To^2}\right) \tag{2-13}$$

$$\bar{C} = \left(\frac{\rho_f}{\rho_g}\right)^{0.5} + \left(\frac{\rho_g}{\rho_f}\right)^{0.5} \tag{2-14}$$

and

$$To = \left(\frac{x}{1-x}\right)^{0.9} \left(\frac{\mu_f}{\mu_g}\right)^{0.1} \left(\frac{\rho_g}{\rho_f}\right)^{0.5} \tag{2-15}$$

In comparing this correlation with that of Baroczy [24], Chisholm [29] obtained reasonably good results.

In his most recent paper, Chisholm [30] proposed the following relationship for Φ^2_{fo}:

$$\Phi^2_{fo} = 1 + (\Gamma^2-1) \left[B\ x^{(2-n)/2}\ (1-x)^{(2-n)/2} + x^{(2-n)}\right] \tag{2-16}$$

For n = 0.2 (smooth tubes)

$$\Gamma = \left(\frac{\rho_f}{\rho_g}\right)^{0.5} \left(\frac{\mu_g}{\mu_f}\right)^{0.1} \tag{2-17}$$

which is the reciprocal of the square root of the property index proposed by Baroczy [24]. The recommended values of B as a function of G and Γ are given in Table II-2.

Other investigators have measured pressure drop during two-phase flow. Thom [31] published a correlation very similar to the M-N model [1]. He based his correlation on data taken at the University of Cambridge for a high-pressure steam-water system. Hatch and Jacobs [32] studied two-phase pressure drop for R-11 both with and without heat transfer. In comparing these data with the M-N model [1], they noted that the M-N model overpredicted their data by as much as 40%. Altman, Norris, and Staub [33] measured pressure drops for Refrigerant-22 (R-22) evaporating in horizontal tubes. They compared their results with the M-N model [1] and concluded that the M-N model [1] provided consistently conservative predictions. Additional pressure drop data for R-22 were obtained by Johnston and Chaddock [34], who also obtained

Table II-2

VALUES FOR THE PARAMETER B IN CHISHOLM'S [30]
PRESSURE DROP CORRELATION

Γ	G(lbm/hr-sq ft)	B
	$\leq 0.369 \times 10^6$	4.8
≤ 9.5	$0.369 \times 10^6 < G < 1.401 \times 10^6$	$1.769 \times 10^6/G$
	$\geq 1.401 \times 10^6$	$1{,}493.4/G^{0.5}$
$9.5 < \Gamma < 28$	$\leq 0.442 \times 10^6$	$14{,}119/(\Gamma G^{0.5})$
	$> 0.442 \times 10^6$	$21/\Gamma$
≥ 28		$0.4073 \times 10^6/(\Gamma^2 G^{0.5})$

results for Refrigerant-12 (R-12). These data were collected at low temperatures (subzero). Comparing both their data and Altman, Norris, and Staub's [33] data with the M-N model [1], Johnston and Chaddock [34] obtained the same conclusions as Altman, Norris, and Staub [33]. However, they did remark that the M-N model [1] provided the best accuracy of any general correlation. Blatt and Adt [35] also studied two-phase pressure drop with refrigerants. Boiling R-11 and R-113 in a horizontal, 0.25-in.-ID tube, they collected data at saturation pressures ranging from 7 to 10 psia. Blatt and Adt [35] reported an average deviation of 30% when comparing their results with the M-N model [1].

It is clear from the above discussion that the M-N model [1] is a widely used correlation today, especially for refrigerant systems. Most results show that this correlation gives conservative predictions--sometimes by as much as 40%. The correlation proposed by Chawla [28] was developed with refrigerants in mind. However, since it is based on low-temperature data, it is probably of limited value. Chisholm's [29, 30] correlations show promise, but they have not been used to analyze a significant amount of refrigerant data. Baroczy's [24] correlation, although usually reliable, is difficult to use because the graphical results upon which it is based are not suitable for computerization. In short, additional work needs to be done beginning with perhaps, Chisholm's [30] most recent correlation if a reliable pressure-drop correlation for refrigerants is to be obtained.

C. HEAT TRANSFER COEFFICIENTS

Since the late 1940's, the prediction of two-phase heat transfer coefficients under forced convection conditions has been studied extensively by many individuals. Much of the work has been conducted with steam-water systems; however, a significant amount of data does exist which resulted from refrigerant studies. The more reliable correlations for flow boiling appear to be based, in part, on pool boiling correlations. For this reason, a discussion of some of the work that has been done in both areas follows. During the course of the discussion, it will, hopefully, become apparent that two-phase, forced, convective heat transfer is a very complex subject, and the development of a single correlation which will accurately predict heat transfer coefficients under all possible conditions is very unlikely.

Pool Boiling

Pool boiling occurs when a heating surface, such as a flat plate or wire, is submerged in a pool of liquid at saturation temperature. When heat is supplied to the saturated liquid in the absence of external agitation, saturated pool boiling results. A pool boiling coefficient, h_{PB}, can be defined, based on the temperature difference between the heating surface and the liquid saturation temperature, ΔT_s, and the heat flux supplied by the heater, ϕ_{PB}. This relationship is given by Equation (2-12) below:

$$h_{PB} = \frac{\phi_{PB}}{\Delta T_s} \tag{2-18}$$

Rohsenow [36] developed one of the early pool boiling correlations. He started with the postulation that:

$$N_{Nu} = f_1(N_{Re})\ f_2\ (N_{Pr}) \tag{2-19}$$

where f_1 and f_2 indicate functional dependencies. By using the bubble departure diameter as the significant length in the Nussett and Reynolds numbers and evaluating all properties at the liquid saturation temperature, Rohsenow [36] was able to show that:

$$\frac{c_{pf}\ \Delta T_s}{h_{fg}} = C_{sf}\left(\frac{\phi_{PB}}{\mu_f h_{fg}}\sqrt{\frac{\sigma}{g(\rho_f - \rho_g)}}\right)^b (N_{Pr_f})^c \tag{2-20}$$

where h_{fg} is the latent heat of vaporization, and μ_f is the liquid viscosity. He was able to successfully correlate a large quantity of pool boiling data with values of 0.33 and 1.7 for b and c, respectively. The empirical constant, C_{sf}, is a complex quantity which depends on the heating surface-fluid combination, as well as other things. A brief listing of C_{sf} values for various fluid combinations is given in Kreith [37].

Forster and Zuber [38] proposed another pool boiling correlation based on an analysis of bubble dynamics. Using the same basic postulation that Rohsenow [36] employed, the following equation was obtained:

$$h_{PB} = 0.1743 \left(\frac{\rho_f^{0.49} \, c_{pf}^{0.45} \, k_f^{0.79}}{h_{fg}^{0.24} \, \rho_g^{0.24} \, \sigma^{0.5} \, \mu_f^{0.29}} \right) (\Delta T_s)^{0.24} \, (\Delta p_s)^{0.75} \qquad (2\text{-}21)$$

where σ is the surface tension and Δp_s is the difference between the saturation pressure based on the heater temperature and that based on the saturation temperature. Forster and Zuber [38] developed their correlation using maximum heat flux data for ethanol, n-pentane, benzene, and water. Since it is based on maximum heat flux data, the range of applicability is somewhat questionable. In the expression, the liquid properties are evaluated at the heater temperature and the vapor properties are evaluated at the saturation temperature.

Two Russian correlations have also appeared in the literature. Kutateladze [39] presented the following relationship:

$$h_{PB} = 0.0007 \, (N_{Pr})^{0.35} \left(\frac{\phi_{PB} \, \rho}{h_{fg} \, \rho_g \nu_f} \right)^{0.7} \frac{1}{\sqrt{\sigma}} \, (\rho_f - \rho_g)^{-0.2} \, k_f \qquad (2\text{-}22)$$

where ν_f is the kinematic viscosity, μ_f/ρ_f, of the liquid: Borishanskiy and Minchenko [40] proposed the same basic correlation with a different coefficient and a different exponent on the Prandtl number. Their equation was:

$$h_{PB} = 0.00087\ (N_{Pr})^{0.70} \left(\frac{\phi_{PB}\ \rho}{h_{fg}\ \rho_g \nu_f}\right)^{0.7} \frac{1}{\sqrt{\sigma}} (\rho_f - \rho_g)^{-0.2}\ k_f. \qquad (2\text{-}23)$$

Gilmour [41] also proposed a pool boiling correlation. He used a somewhat innovative approach by defining a nucleate, boiling Colburn j-factor as a function of Reynolds number, Prandtl number, and another dimensionless group which included pressure and surface tension effects. His correlation took the form:

$$\left(\frac{h_{PB}}{c_{pf}\ G_o}\right)^a \left(\frac{\mu_f c_{pf}}{k_f}\right)^b \left(\frac{\rho_f\ \sigma}{P^2}\right)^c = \left(\frac{Ca}{DG_o/\mu_f}\right)^d \qquad (2\text{-}24)$$

where G_o represents the mass flux of liquid which returns to the heating surface as an equal volume of vapor leaves. Ca is an empirical constant which depends on heater surface material, etc. Using a large quantity of pool boiling data, Gilmour [41] proposed the following values for the empirical quantities:

$$\left.\begin{array}{l} a = 1 \\ b = 0.6 \\ c = 0.425 \\ d = 0.3 \end{array}\right\} \qquad (2\text{-}25)$$

He also suggested values for Ca which are listed in Table II-3 below:

Table II-3

VALUES OF THE CONSTANT, Ca, AS PROPOSED BY GILMOUR [41]

Ca	Description of Surface
0.001	Copper and Steel
0.00059	Stainless Steel or Chromium Nickel Alloys
0.0004	Polished Surfaces

As Jallouk [42] points out, very little work has been done to measure pool boiling coefficients for refrigerant systems. Most of the work that has been done was not correlated. In his thesis, Jallouk [42] analyzed a significant quantity of refrigerant pool boiling data and concluded that the Forster-Zuber [38] and Rohsenow [36] correlations were probably the most reliable. He also achieved success with the Borishanskiy-Minchenko [40] correlation but questioned its validity, since it had not been used extensively.

Flow Boiling

Although heat transfer correlations for two-phase fluids flowing inside tubes or ducts have been proposed by many people, probably the most widely used flow boiling correlation is that proposed by Chen [43]. Using the Forster-Zuber [38] pool boiling correlation in conjunction with a Dittus-Boelter type equation, Chen [43] was able to develop the following equation for flow boiling heat transfer coefficients:

$$h_{tp} = 0.023 \left(N_{Re}\right)_{fa}^{0.8} \left(N_{Pr}\right)_{f}^{0.4} \left(\frac{k_f}{D}\right) F$$

$$+ 0.174 \left(\frac{k_f^{0.79}\, c_{pf}^{0.45}\, \rho_f^{0.49}}{h_{fg}^{0.24}\, \rho_g^{0.24}\, \sigma^{0.5}\, \mu_f^{0.29}}\right) \left(\Delta T_s\right)^{0.24} \left(\Delta p_s\right)^{0.75} S \quad (2\text{-}26)$$

where F and S are Chen's [43] multiplication factor and suppression factor, respectively.

In arriving at this correlation, Chen [43] assumed that the contribution of both the nucleate boiling and two-phase convective coefficients were additive. The first part of Equation (2-26) represents the two-phase convective component, while the second portion represents the nucleate boiling contribution. The F-factor was defined as the ratio of the two-phase Reynolds number to the liquid Reynolds number based on the liquid alone. Chen [43] reasoned that F was strictly a flow parameter, and through a momentum transfer analogy, he was able to express F as a function of X_{tt}, the Martinelli parameter. The suppression factor, S, was included to account for differences between bubble growth during pool boiling and bubble growth during flow boiling. As Chen [43] pointed out, the temperature profiles near the wall in a forced convection system are generally steeper than in a corresponding pool boiling case. The bubble growth rate in this forced convection system would be smaller than the comparable pool boiling case, since the bubbles would tend to protrude into a lower temperature region which would suppress their growth rate. This effect prompted Chen [43] to define the suppression factor, S, as:

$$S = \left(\frac{T_{\text{fluid around bubble}} - T_{\text{saturation}}}{T_{\text{wall}} - T_{\text{saturation}}}\right)^{0.99} \qquad (2\text{-}27)$$

The suppression factor was correlated with the two-phase Reynolds number so that F and S are interrelated variables. Chen [43] described the iterative procedure which was used to determine their values from experimental data in his paper.

The Chen [43] correlation has been applied to a large number of vastly different two-phase systems and has fared remarkably well. The applicability to refrigerant systems is somewhat questionable and additional work is needed in this area; however, most experimenters have concluded that the Chen [43] correlation provides conservative results when used to predict heat transfer coefficients for boiling refrigerants. Since it is basically an annular flow model, the Chen [43] correlation would probably be inaccurate when applied to purely stratified and wavy flows.

Cheshire and Sterling [44] proposed one of the early correlations for predicting refrigerant, two-phase heat transfer coefficients. Working with R-114 in a natural circulation, vertical tube evaporator, they were able to correlate 92% of their data to within ±15% with the expression:

$$h_{tp} = \frac{1.093 \times 10^{-11}}{\mu_f^{0.6}} \left[\frac{(T + 460)\Delta T}{\sigma}\right]^2 \tag{2-28}$$

It is interesting to note that this expression is independent of such important quantities as mass velocity and vapor quality. This probably makes the Cheshire and Sterling [44] correlation applicable in a narrow range of operating conditions.

Baker, Touloukian, and Hawkins [45] studied flow boiling in a 0.545-in.-ID copper tube. Based on data taken with R-12 and methyl chloride as working fluids, they presented correlations developed by modifying the relationship proposed by Rohsenow [36] for pool boiling. For R-12 with temperature differentials less than 12°F and flow rates less than 120 lbm/hr, they proposed:

$$\frac{c_{pf}\,\Delta T}{h_{fg}} = 0.0125 \left[\frac{\phi}{\mu_f\, h_{fg}} \sqrt{\frac{\sigma}{g(\rho_f - \rho_g)}}\right]^{1/3} (N_{Pr})_f^{1.7} \left(\frac{x_o}{x_i}\right)^{-0.042} \tag{2-29}$$

In this expression, x_i and x_o are the inlet and outlet vapor qualities, respectively. All of the methyl chloride data and all R-12 data with temperature differentials in excess of 12°F were correlated by:

$$\frac{c_{pf}\ \Delta T}{h_{fg}} = Ci \left[\frac{\phi}{\mu_f\ h_{fg}} \sqrt{\frac{\sigma}{g(\rho_f - \rho_g)}}\right]^{3/4} (N_{Pr})_f^{1.7} \left(\frac{x_o}{x_i}\right)^{-0.042}$$

$$\left[\frac{GD}{\mu_f} (1 - x_{avg}) \frac{\rho_g}{\rho_f}\right]^{-1/3} \tag{2-30}$$

where

$Ci = 0.038$ for methyl chloride,

$Ci = 0.073$ for R-12, and

$x_{avg} = (x_i + x_o)/2.$

Baker, Touloukian, and Hawkins [45] were able to correlate the applicable data to within ±10% with Equation (2-29) and to within ±15% with Equation (2-30).

Dengler and Addoms [46] studied local heat transfer coefficients during the forced circulation boiling of water in a vertical tube. They correlated their data in terms of the Martinelli parameter for the turbulent-turbulent case, X_{tt}. Eighty-five percent of their data were correlated to within ±12% by the expression:

$$\frac{h_{tp}}{h_{fo}} = 3.5 \left(\frac{1}{X_{tt}}\right)^{0.5} \tag{2-31}$$

They claimed that this expression was applicable for all values of $1/X_{tt}$ between 0.25 and 70.0.

Altman, Norris, and Staub [33] boiled R-22 in a 0.343-in.-ID, horizontal tube and correlated their own data, as well as the data of

Baker, Touloukian, and Hawkins [45]. By refining the correlation proposed by Pierre [47], they were able to show that for exit qualities less than 0.80

$$\frac{h_{tp} D}{k_f} = 0.0225 \left(N_{Re_{fo}}^{2} K_F\right)^{0.375} \tag{2-32}$$

when

$$2.5 \times 10^{10} < \left(N_{Re_{fo}}^{2} K_F\right) < 1.5 \times 10^{12}.$$

In this expression K_F is called the load factor and is given by:

$$K_F = \frac{J \Delta x\ h_{fg}}{L} \tag{2-33}$$

where Δx is the quality change from inlet to outlet and J is the mechanical equivalent of heat.

The original correlation of Pierre [47] considered two cases: (1) incomplete evaporation with outlet qualities well below 90%, and (2) complete evaporation with an exit superheat of 11°F. For the first case, Pierre [47] proposed that:

$$\frac{h_{tp} D}{k_f} = 0.0009 \left(N_{Re_{fo}}^{2} K_F\right)^{0.5} \tag{2-34}$$

For the second case, he proposed that:

$$\frac{h_{tp} D}{k_f} = 0.0082 \left(N_{Re_{fo}}^{2} K_F\right)^{0.4} \tag{2-35}$$

In both of these equations the parameter $N_{Re_{fo}}^{2} K_F$ must lie between 10^9 and 0.7×10^{12} exclusively.

The chief limitation in both Pierre's [47] and Altman, Norris, and Staub's [33] analyses is that the influence of local quality is not included. Many investigators have shown that local two-phase heat transfer coefficients can be very strongly dependent on local quality, especially at high mass fluxes.

Davis and David [48] studying steam-water mixtures in a horizontal, rectangular duct developed two correlations. Taking a slip ratio approach, they proposed the following separated-annular flow model which was accurate in the high mass velocity region:

$$\frac{h_{tp}D}{k_f} = 0.060 \left(\frac{\rho_f}{\rho_g}\right)^{0.28} \left(\frac{DxG}{\mu_f}\right)^{0.87} \left(N_{Pr}\right)_f^{0.4} . \tag{2-36}$$

They also proposed a Homogeneous model given by:

$$\frac{h_{tp}D}{k_f} = 0.033 \left(\frac{DG}{\mu_{tp}}\right)^{0.87} \left(N_{Pr}\right)_f^{0.4}$$

where the two-phase viscosity, μ_{tp}, is determined from:

$$\frac{1}{\mu_{tp}} = \frac{x}{\mu_g} + \frac{(1 - x)}{\mu_f} \tag{2-37}$$

Both of these equations predicted their experimental data to within ±15%.

Another correlation for flow boiling heat transfer coefficients was devised by Schrock and Grossman [49], who collected and analyzed data for water boiling in a vertical tube. By utilizing a dimensionless analysis procedure, they were able to show that the flow boiling heat transfer coefficient was influenced by the so-called boiling number, Bo, and the Martinelli parameter, X_{tt}. At sufficiently high values of the boiling number, the local heat transfer coefficient becomes independent of X_{tt}, and at sufficiently low values of Bo, the local heat-transfer coefficient is a function of X_{tt} only. The expression proposed by Shrock and Grossman [49] is given below:

$$\frac{h_{tp}D/k_f}{\left(N_{Re}\right)_{fa}^{0.8}\left(N_{Pr}\right)_f^{1/3}} = 170\left[Bo + 1.50 \times 10^{-4}\left(\frac{1}{X_{tt}}\right)^{2/3}\right] \tag{2-38}$$

where Bo = $\phi/(G \cdot h_{fg})$. Schrock and Grossman [49] stated that this correlation predicted their data to within ±35%.

Blatt and Adt [35] attempted to separate the effects of nucleate boiling and forced convection in their studies with two-phase R-11 and R-113 flowing in a 0.25-in.-ID horizontal tube. They assumed that forced convection effects and bubble motion effects were additive so that:

$$\phi = \phi_B + \phi_C \tag{2-39}$$

where ϕ_B represents the portion of the heat flux that is due to boiling effects and ϕ_C represents that portion which is due to forced convection. By calculating ϕ_C from the well-known Dittus-Boelter equation, Blatt and Adt [35] were able to obtain experimental values for ϕ_B from Equation (2-39). They were successful in correlating these data by using Rohsenow's [36] equation. They established their own experimental values for C_{sf} from independent pool boiling tests. These values are given in Table II-4.

It should be noted that Blatt and Adt's [35] analyses differ from that of Baker, Touloukian, and Hawkins [45], who also correlated their data with the Rohsenow [36] equation, along with some additional variables. Baker, Touloukian, and Hawkins [45] did not attempt to account for forced-convection effects. Thus, their correlation is probably more limited in application than is that of Blatt and Adt [35].

Table II-4

VALUES OF C_{sf} AS DETERMINED BY BLATT AND ADT [35]

Surface - Fluid Combination	C_{sf}
Stainless Steel - R-11	0.006
Copper - R-11	0.010
Stainless Steel - R-113	0.004
Copper - R-113	0.007

Johnston and Chaddock [34] modified the coefficients in Pierre's [47] correlation and obtained good agreement with their average heat transfer coefficient data for R-12 and R-22 boiling in a 0.46-in.-ID horizontal copper tube. Unfortunately, all of their data were taken at low saturation temperatures and low mass fluxes. The equations that they proposed are given below:

$$\frac{h_{tp} D}{k_f} = 0.000634 \left(N_{Re_{fo}}^{2} K_F \right)^{0.5} \tag{2-40}$$

for R-12 and

$$\frac{h_{tp} D}{k_f} = 0.000711 \left(N_{Re_{fo}}^{2} K_F \right)^{0.5} \tag{2-41}$$

for R-22. Both of these relationships are valid when

$$10^9 < N_{Re_{fo}}^{2} K_F < 7 \times 10^{10}$$

An additional flow boiling correlation was proposed by Lavin and Young [50], who considered R-12 and R-22 in both vertical and horizontal tubes. For the annular flow regime, they proposed two relations - one for horizontal flow and one for vertical flow. In the horizontal orientation, Lavin and Young [50] suggested that

$$\frac{h_{tp}}{h_{fa}} = 6.59 \left(\frac{\phi}{G\ h_{fg}}\right)^{-0.1} \left(\frac{1 + x}{1 - x}\right)^{1.16} \tag{2-42}$$

In this expression h_{fa}, the heat transfer coefficient for the liquid alone, is determined from the Seider-Tate relationship with a prefix of 0.023. It is interesting to note that Lavin and Young [50] obtained significantly higher boiling coefficients for horizontal flow than for vertical flow. They explained this by considering the annular flow regime for both orientations. In vertical flow, the film thickness near the wall of a round tube is essentially uniform circumferentially at any one cross section. This is not the case in horizontal flow where, due to surface tension effects, wave crests, and gravity, the liquid film is thicker along the bottom of the tube and thinner over the remaining part. This results in a lower mean film thickness over a large part of the tube than in the vertical case which yields the higher heat transfer coefficients.

In an American Society of Heating, Refrigerating, and Air-Conditioning Engineers (ASHRAE)-supported effort, Gouse and Coumou [51] studied two-phase heat transfer in horizontal glass tubes using R-113 as the working fluid. Their relatively low heat flux and boiling temperature data were compared to many of the existing boiling correlations, including Chen [43], Dengler and Addoms [46], and Guerrieri and Talty [52]. The best results were obtained with the Guerrieri and Talty correlation; however, the Chen [43] correlation, with the nucleate boiling component omitted, also agreed fairly well with the experimental data, especially at the higher vapor qualities.

In a companion effort to Gouse and Coumou [51], Chaddock and Noerager [53] also studied refrigerant boiling in an ASHRAE-sponsored program. They obtained data with R-12 boiling in a horizontal, stainless steel tube. Their data were obtained at a saturation temperature of approximately 54°F for variations in heat flux from about 2,000 Btu/hr-sq ft to 11,000 Btu/hr-sq ft. The results obtained were strongly flow regime-dependent; however, Chaddock and Noerager [53] were able to establish that both heat and mass fluxes have a significant effect on two-phase heat transfer coefficients in horizontal tubes. Their experimental data agreed well with modified forms of the Dengler-Addoms [46] and Guerrieri-Talty [52] correlations. Agreement with the Chen [43] correlation was observed to be poor.

Other ASHRAE-sponsored efforts have furnished two-phase flow heat transfer data. Anderson, Rich, and Geary [54] collected and analyzed data on R-22 boiling in a horizontal tube heated by water flowing in an annulus. When comparing their data with existing correlations, they obtained poor agreement in most cases, especially when attempting to correct for possible nucleation effects. Average heat transfer coefficients for the entire tube length were correlated with fair agreement by Pierre's [47] correlations. An extension to the work of Gouse and Coumou [51] was performed by Gouse and Dickson [55]. The glass test sections were heated by a transparent electric resistance film deposited on the inside tube surface. In Gouse and Coumou's [51] effort, this film was deposited on the outside tube surface. The data acquired during the more recent investigation were compared with available correlations as before; however, some additional ones were also considered. Unfortunately, the inability to reproduce results previously obtained under similar operating conditions with the first test section make Gouse and Dickson's [55] data somewhat questionable.

A very extensive investigation of two-phase heat transfer to refrigerants in horizontal tubes has been carried out by Chawla [56] whose work on two-phase pressure drop was discussed earlier. He used the two-phase flow parameter, ε, in his correlation for heat transfer coefficients. This parameter is identical to the one defined in his pressure drop studies (see Figure II-9, Page 52) to account for flow pattern influence and momentum exchange between the phases. Based on experimental data taken with R-11, Chawla [56] proposed that:

$$\frac{h_{tp}D}{k_f}\left\{1 - \left(1 + \left[\frac{1-x}{x\varepsilon\left(\frac{\rho_f}{\rho_g}\right)}\right]\right)^{-1/2}\right\} = 0.0066\left(N_{Re_{fa}}\, N_{Fr_{fa}}\right)^{0.475}\left(\frac{x}{1-x}\right)$$

$$\left(\frac{\rho_f}{\rho_g}\right)^{0.30}\left(\frac{\mu_f}{\mu_g}\right)^{0.80}\left(N_{Re_{fa}}\right)^{0.35}\left(N_{Pr_f}\right)^{0.42} \qquad (2\text{-}43)$$

for

$$N_{Re_{fa}} \cdot N_{Fr_{fa}} < 109$$

and

$$\frac{h_{tp}D}{k_f}\left\{1 - \left(1 + \left[\frac{1-x}{x\varepsilon\left(\frac{\rho_f}{\rho_g}\right)}\right]\right)^{-1/2}\right\} = 0.015\left(N_{Re_{Fr}}\, N_{Fr_{fa}}\right)^{0.30}\left(\frac{x}{1-x}\right)$$

$$\left(\frac{\rho_f}{\rho_g}\right)^{0.30}\left(\frac{\mu_f}{\mu_g}\right)^{0.80}\left(N_{Re_{fa}}\right)^{0.35}\left(N_{Pr_f}\right)^{0.42} \qquad (2\text{-}44)$$

for

$$N_{Re_{fa}} \cdot N_{Fr_{fa}} > 109.$$

Good agreement was obtained when Chawla [56] applied his correlation to data taken with various fluids.

In another ASHRAE-sponsored effort, Riedle and Purcupile [57] and Riedle, Purcupile, and Schmidt [58] recently studied both horizontal and vertical boiling with R-11, R-12, and R-13 in stainless steel tubes. In the first publication, Riedle and Purcupile [57] presented a model based on wall superheat and claimed that it was applicable in the fully developed nucleate boiling regime. This relation was modified to include a pressure dependency in the second publication and predicted the low quality, low mass flux data to within ±30%.

Jallouk [42] recently completed a very extensive study of refrigerant, two-phase heat transfer coefficients. He compared his own experimental R-114 data, as well as experimental data from other sources, with all of the major correlations available in the literature today. For two-phase flow with R-114 in a vertical tube at mass fluxes below 1.7×10^6 lbm/hr-sq ft, Jallouk [42] suggested that:

$$h_{tp} = 0.023 \left(N_{Re}\right)_{fa}^{0.8} \left(N_{Pr}\right)_{f}^{0.4} \left(\frac{k_f}{D}\right) F + 0.3802 \left(\frac{k_f^{0.79}\, c_{pf}^{0.45}\, \rho_f^{0.49}}{h_{fg}^{0.24}\, \rho_g^{0.24}\, \sigma^{0.5}\, \mu_f^{0.29}}\right) \left(\Delta T_s\right)^{0.24} \left(\Delta p_s\right)^{0.75} S \tag{2-45}$$

or

$$h_{tp} = 0.023 \left(N_{Re}\right)_{fa}^{0.8} \left(N_{Pr}\right)_{f}^{0.4} \left(\frac{k_f}{D}\right) F + \left[\frac{k_f^{5.1}\, (\rho_f - \rho_g)^{0.5}\, g^{0.5}}{(0.0047)^3\, c_{pf}^{2.1}\, f^{4.1}\, \sigma^{0.5}\, h_{fg}^{2}\, g_c^{0.5}}\right] \Delta T^2\, S \tag{2-46}$$

where

$$S = -0.2074 \left(p_r\right)^{-0.0899} \ln \left(N_{Re_{fa}} F^{1.25}\right)$$

$$+ 2.85568 \left(p_r\right)^{-0.06203} \quad (2\text{-}47)$$

and

$$F = (\Phi_{ftt})^{0.89} . \quad (2\text{-}48)$$

Equation (2-45) is a modified form of the original Chen [43] correlation with a different coefficient for the nucleate boiling component and improved relationships for S and F. Equation (2-46) is also a modified form of the Chen [43] correlation; however, the Rohsenow [36] pool boiling correlation with C_{sf} equal to 0.0047 is used to determine the nucleate boiling component rather than the Forster-Zuber [38] correlation. Jallouk [42] was able to correlate a wide range of R-114 vertical, boiling data to within ±30% with either of these relationships

CHAPTER III

EXPERIMENTAL INVESTIGATION

A. DESCRIPTION OF SYSTEM

All of the experimental data taken during this effort were obtained from the Experimental Heat Transfer Facility (EHTF) located in the K-1401 building at the Oak Ridge Gaseous Diffusion Plant. Since the facility is discussed in some detail in Appendix A, this section will deal primarily with the horizontal test section and the associated instrumentation.

A schematic drawing of the horizontal test section is shown in Figure III-1. The test section was divided into 10 zones which were separated by 1/2-in.-wide grooves. Each zone was individually heated and controlled. The grooves were required in order to minimize axial conduction. They were filled with epoxy which was capable of withstanding temperatures up to 500°F.

Wall temperatures were measured at eight locations in each zone by Chromel-Alumel thermocouples. These thermocouples were embedded in the tube wall according to the procedure in Appendix A. Each wall thermocouple was approximately 0.1 in. from the inside tube surface. The instream refrigerant temperatures at the center of each zone were measured by sheathed Chromel-Alumel thermocouples. Additional temperatures were also measured in each zone. Figure III-2 is a schematic drawing which shows the location of these thermocouples, as well as other pertinent information.

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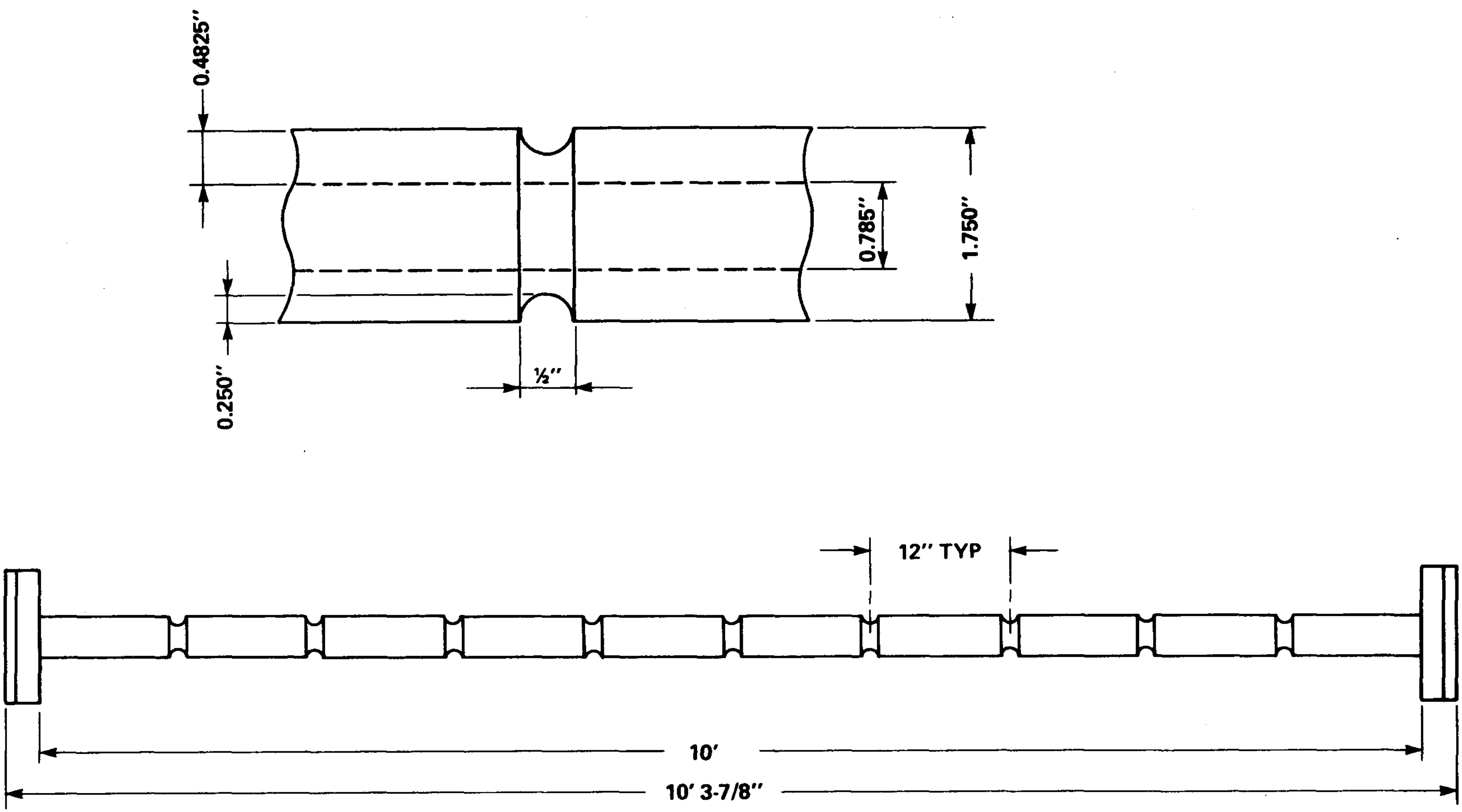

FIGURE III-1. A SCHEMATIC DRAWING OF THE HORIZONTAL TEST SECTION.

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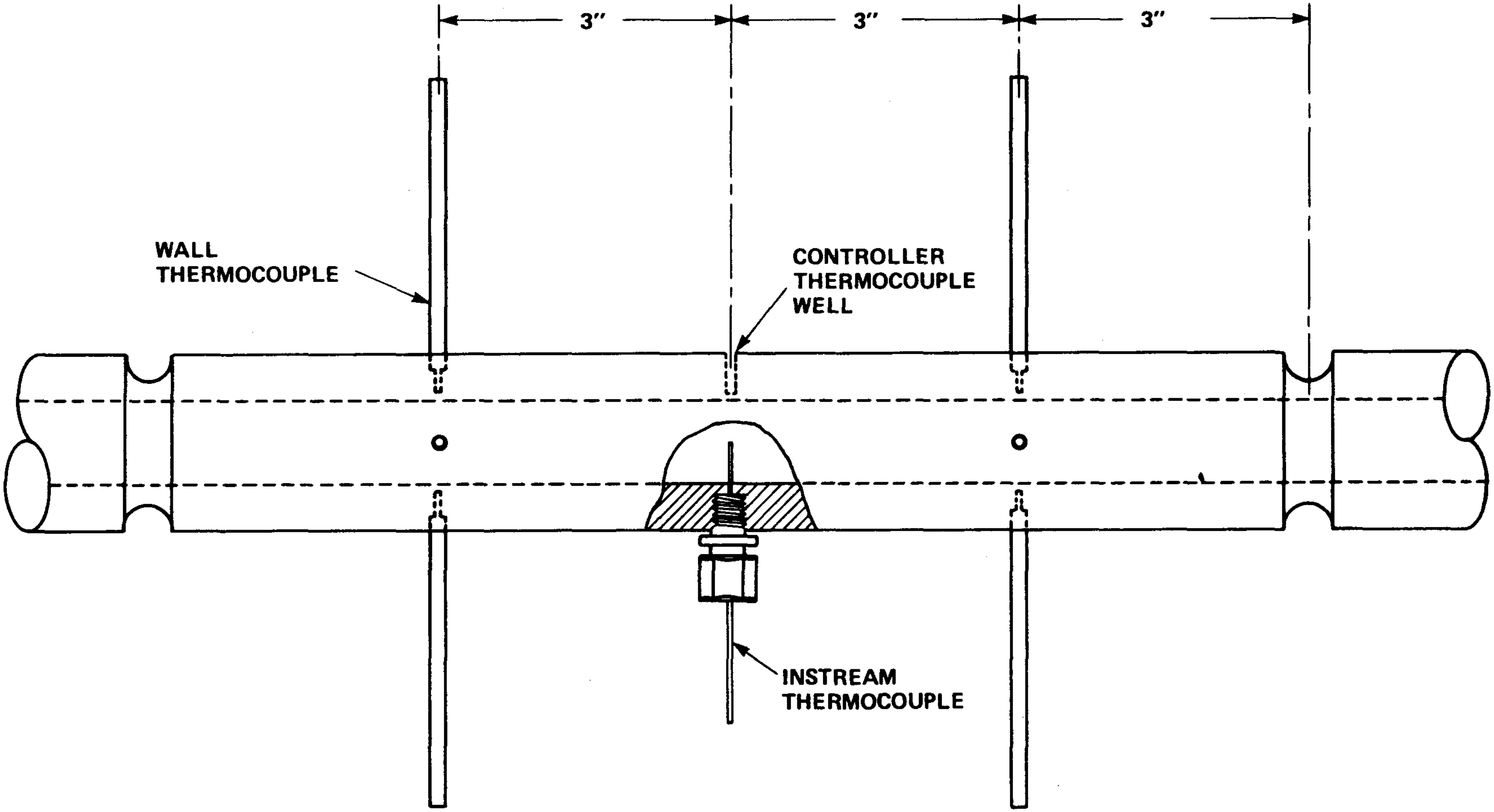

FIGURE III-2. A SCHEMATIC DRAWING SHOWING THERMOCOUPLE LOCATIONS.

A Foxboro D/P cell was used to detect the pressure loss incurred by the working fluid. Pressure taps were installed at 2-ft increments, as shown in Figure III-3. The absolute pressure was sensed at the test section inlet.

A flow visualization section was located at the outlet of the test section. The details of this assembly are shown in Figure III-4.

B. DATA ACQUISITION PROCEDURES

Two different data acquisition procedures were employed during this investigation, depending on the type of data desired.

Experimental Heat Transfer Coefficient and Pressure Drop Data

Data of this type were generally obtained during the same experimental run. In both the natural and forced circulation modes, the desired boiling temperature and heat flux were set and held constant, while the refrigerant mass flux was varied from 0.12×10^6 lbm/hr-sq ft to 3.4×10^6 lbm/hr-sq ft.

After the mass flux had been varied over this range, the heat flux was increased making sure that a critical heat flux condition did not exist. In most cases the heat flux was increased up to about 16,000 Btu/hr-sq ft. This completed all runs of this type for a particular boiling temperature. A different boiling temperature was then selected and the above steps were repeated.

The following data were recorded during these runs:

1. Barometric pressure,
2. Pressure drop at 2-ft intervals along the test section,
3. Gage pressure at the test section inlet,

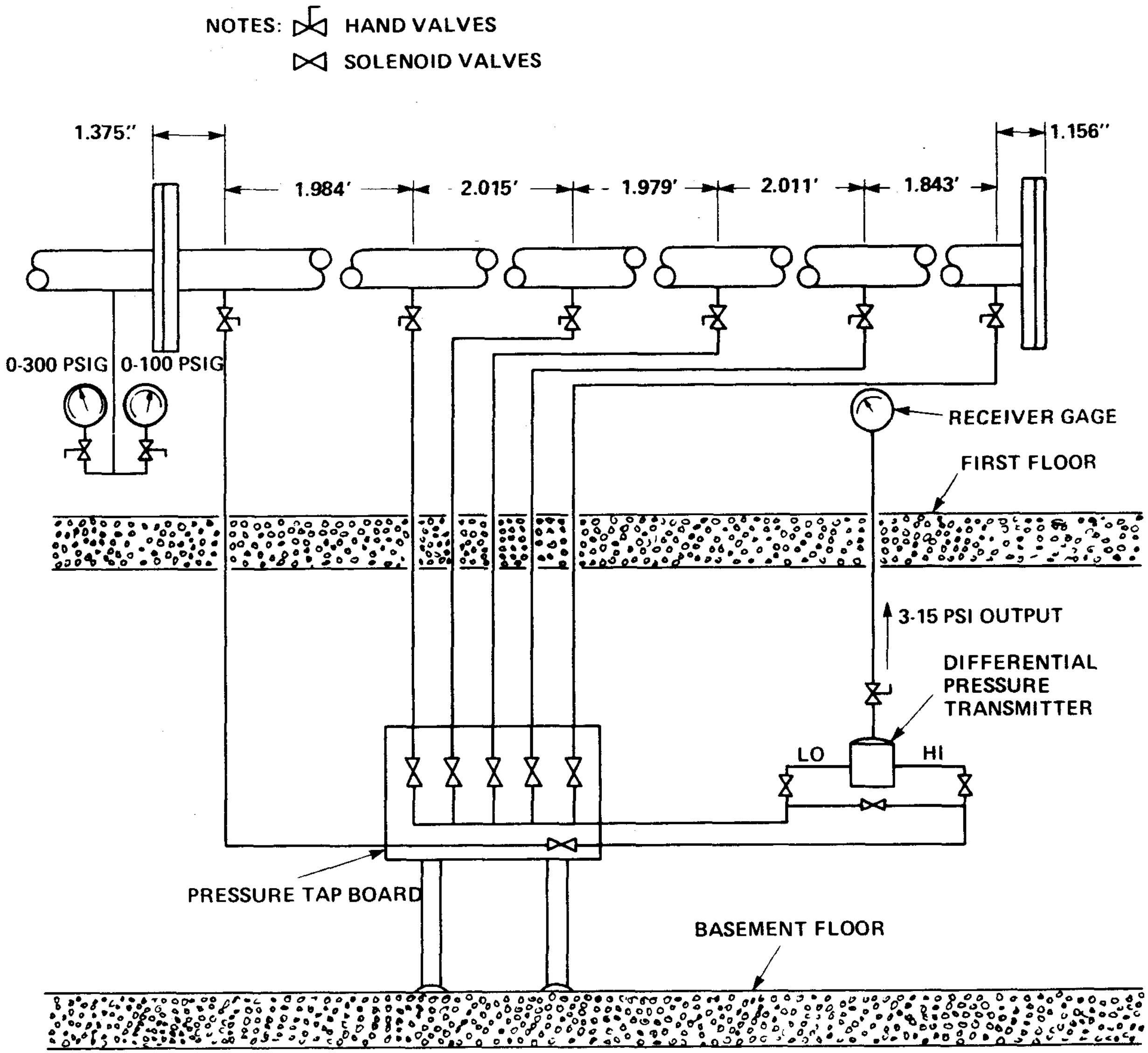

FIGURE III-3. THE PRESSURE DROP MEASURING SYSTEM.

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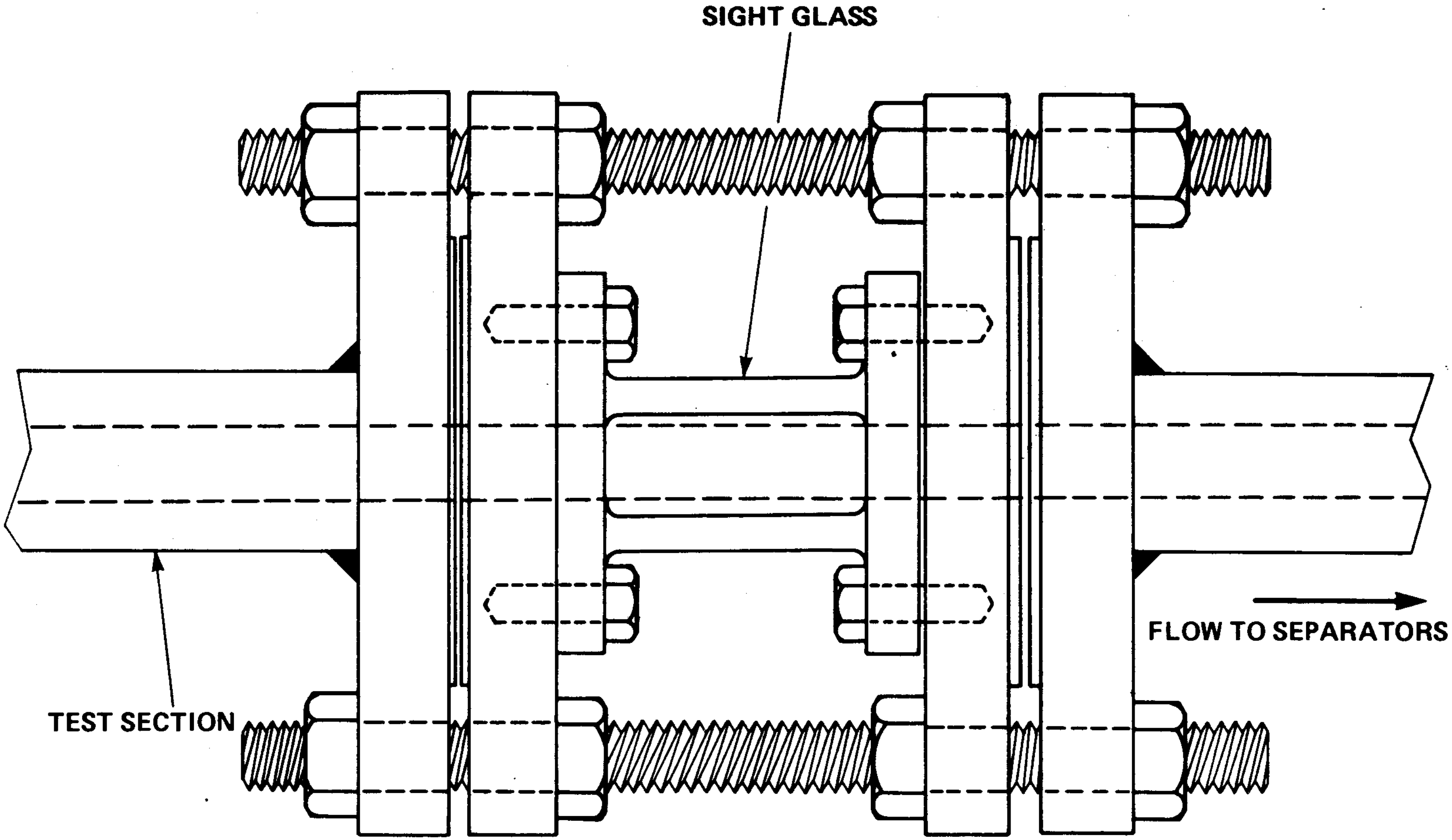

FIGURE III-4. VISUAL FLOW INDICATOR ASSEMBLY.

4. Wall temperatures in each zone,
5. Instream temperature in each zone,
6. Fluid temperature in the pressure tap lines,
7. Gross heat input to each zone,
8. Heater sheath temperature in each zone,
9. Inlet flow to the test section,
10. Entrained liquid flow,
11. Condensate flow,
12. Water flow to condenser, and
13. Certain system temperatures required in connection with the above parameters.

Flow Regime Data

Numerous experimental runs were conducted for the purpose of obtaining the required data to construct a flow map which could be compared to other such maps found in the literature. To obtain data of this type, the runs were conducted in much the same fashion as those discussed above. Experimental pressure drop data were normally not acquired during these runs. As an aid in distinguishing the different flow regimes that existed, videotape pictures were taken of the flow pattern in the flow visualization section. The videotapes resulting from these efforts could be viewed at slow speed so that the flow regimes were easier to identify.

CHAPTER IV

DISCUSSION OF EXPERIMENTAL RESULTS

A. FLOW REGIMES

Introduction

The experimental data related to flow regimes were acquired for the range of parameters given by Table IV-1. A total of 165 runs of this nature were made during the course of this investigation. A tabulation of these data can be found in Appendix C.

Table IV-1

RANGE OF PARAMETERS FOR FLOW REGIME STUDIES

Parameter	Range
Boiling Temperature, ^{o}F	100 - 250
Mass Flux, lbm/hr-sq ft x 10^{-6}	0.12 - 3.4
Quality	0.00 - 1.00

Several different naming conventions descriptive of horizontal, two-phase flow are present in the literature. The following convention, which is essentially equivalent to that recommended by Dukler and Taitel [14], will be used in this investigation:

1. Bubbly flow,
2. Stratified - smooth flow,
3. Stratified - wavy flow,

4. Intermittent flow, and

5. Annular flow.

In all of the videotape pictures obtained during this investigation, one or more of the above flow regimes were observed.

Bubbly Flow. This flow regime was characterized by the presence of small bubbles traveling in the upper portion of the tube. It was observed only at high flow and low heat flux conditions.

Stratified - Smooth Flow. The stratified - smooth flow regime was characterized by liquid flowing in the bottom portion of the tube and vapor at essentially the same velocity in the top portion. The interface between the phases was relatively smooth; however, ripples usually existed. During many of the runs when this flow regime was observed, a thin liquid film could be discerned along the top of the flow visualization section. The resolution of the videotape pictures was not sufficient to detect this film, but it could be observed with the naked eye.

Stratified - Wavy Flow. This flow regime was similar to the stratified - smooth regime, except that the interface was more disturbed by the vapor which traveled at a higher velocity than did the liquid. The occurrence of the thin liquid film mentioned above was even more prevalent when this flow regime was observed.

Intermittent Flow. This flow regime is actually a combination of Alves [3] plug and slug flow regimes. In this investigation, it is more indicative of the slug flow regime, since plug flow was rarely observed. Intermittent flow was therefore characterized, in most cases, by waves being picked up by rapidly moving vapor forming a frothy slug which passed through the tube at a large velocity relative to the liquid velocity. This flow regime was observed most often.

Annular Flow. The annular flow regime was characterized by a thin film of liquid which flowed down the inside wall of the tube. In most cases, many small bubbles were dispersed throughout the liquid film. The central core may have included some entrained liquid; however, this could not be determined from the videotape pictures.

Results

Over 150 runs yielding experimental flow regime data were analyzed during the course of this work. A tabulation of the results is given in Appendix C.

Experimental flow regime maps were constructed by computer from the data for boiling temperatures of 100°F, 150°F, 200°F, and 250°F. These maps are shown in Figures IV-1 to IV-4. The largest area on each of the maps is obviously the intermittent flow regime. The following general comments can be made concerning these maps, as well as observations made with both the naked eye and from the videotape pictures:

1. The occurrence of the bubbly flow regime was a rarity. It was extensively observed on only one occasion during this study. The mass flux was 3.4×10^6 lbm/hr-sq ft and the saturation temperature was 100°F. The only other times bubbly

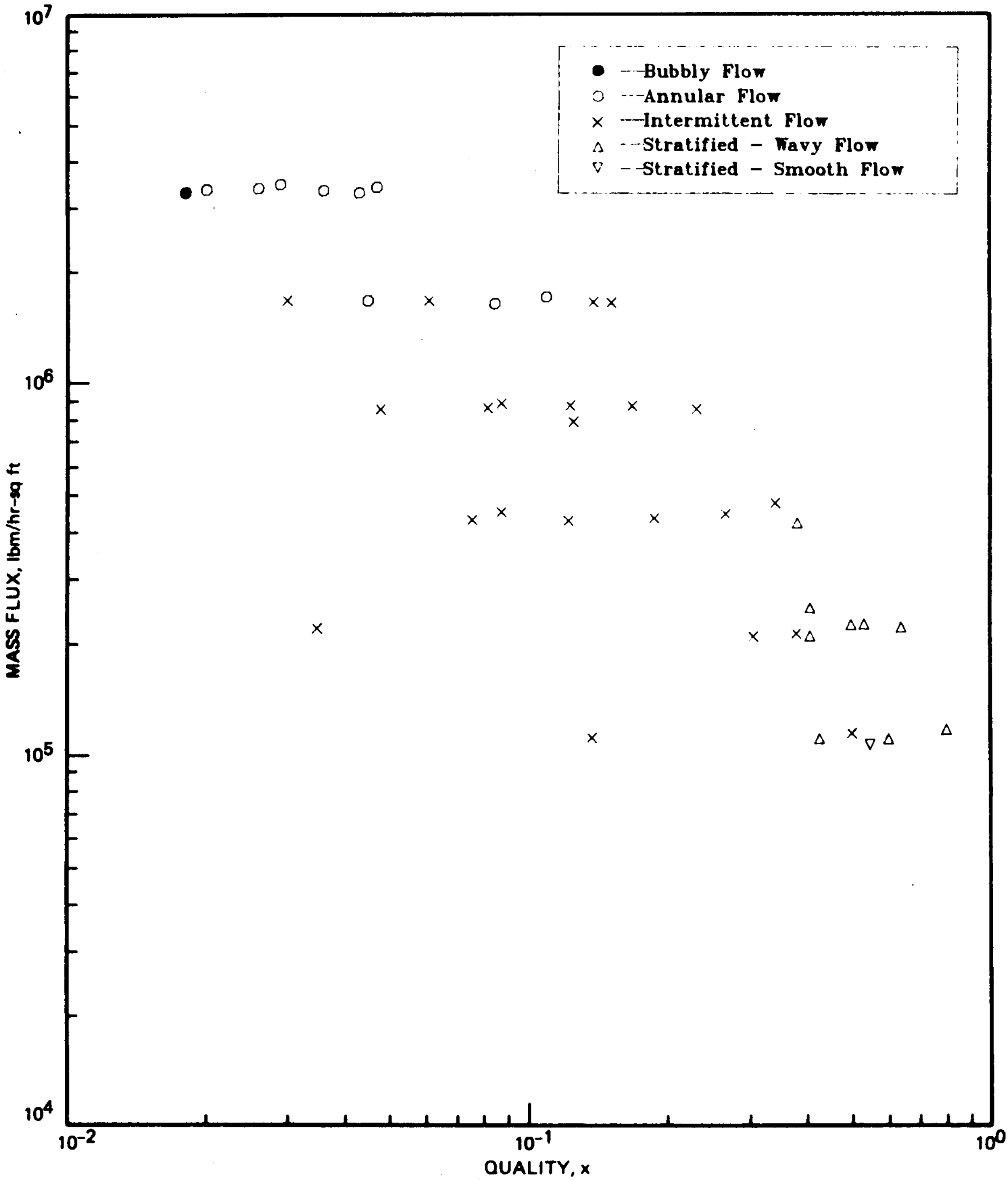

FIGURE IV-1. EXPERIMENTAL FLOW MAP, 100°F.

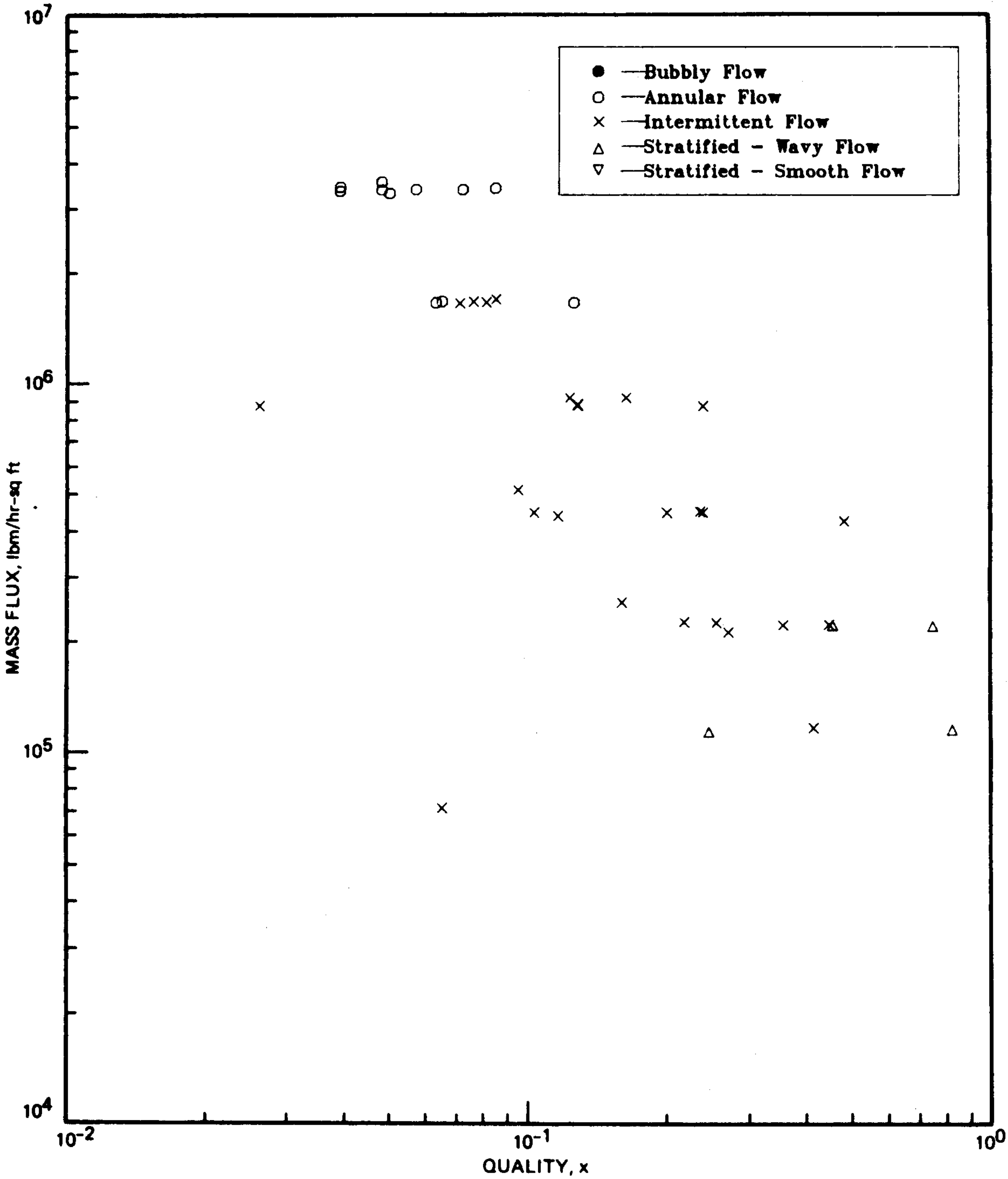

FIGURE IV-2. EXPERIMENTAL FLOW MAP, 150°F.

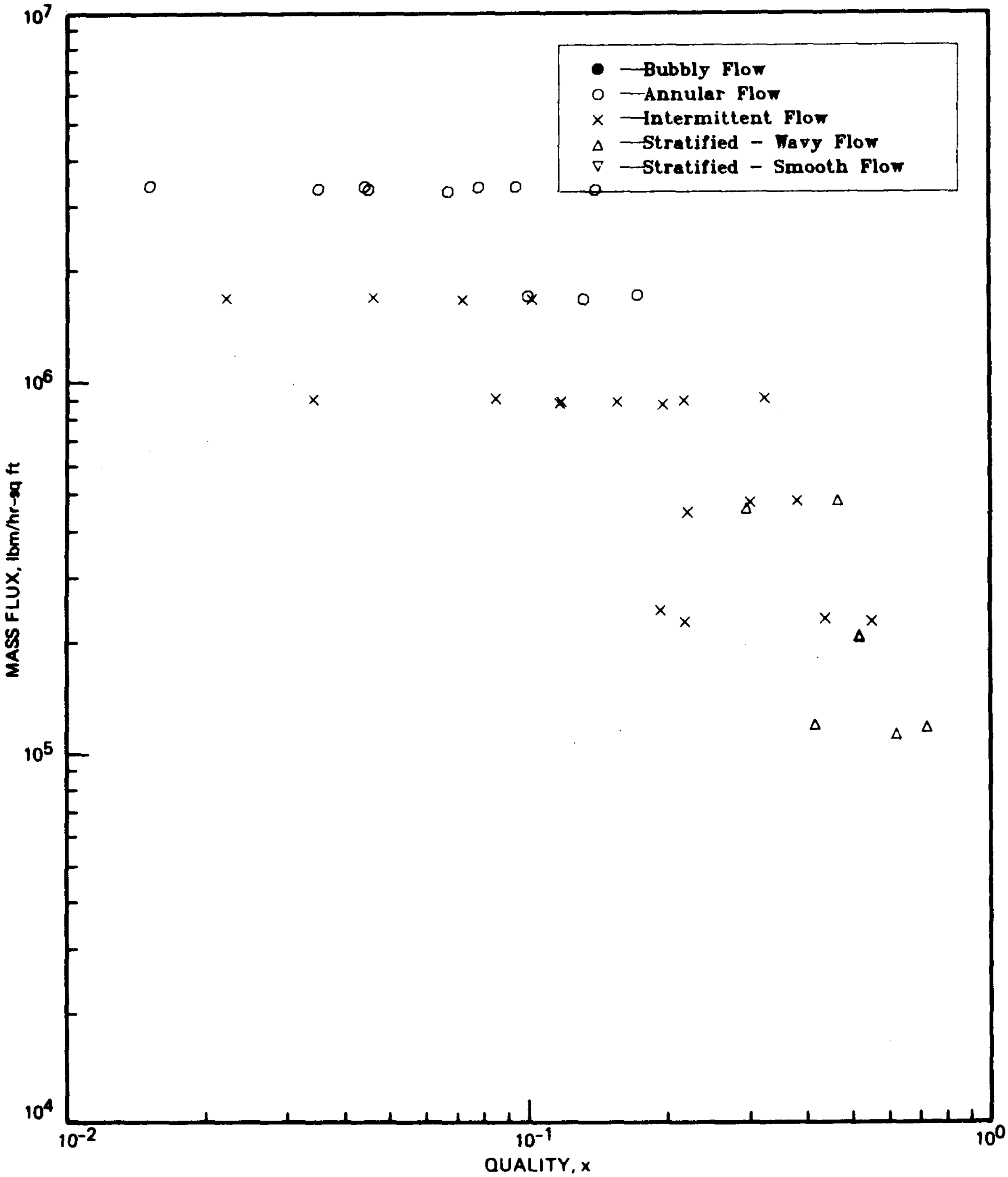

FIGURE IV-3. EXPERIMENTAL FLOW MAP, 200°F.

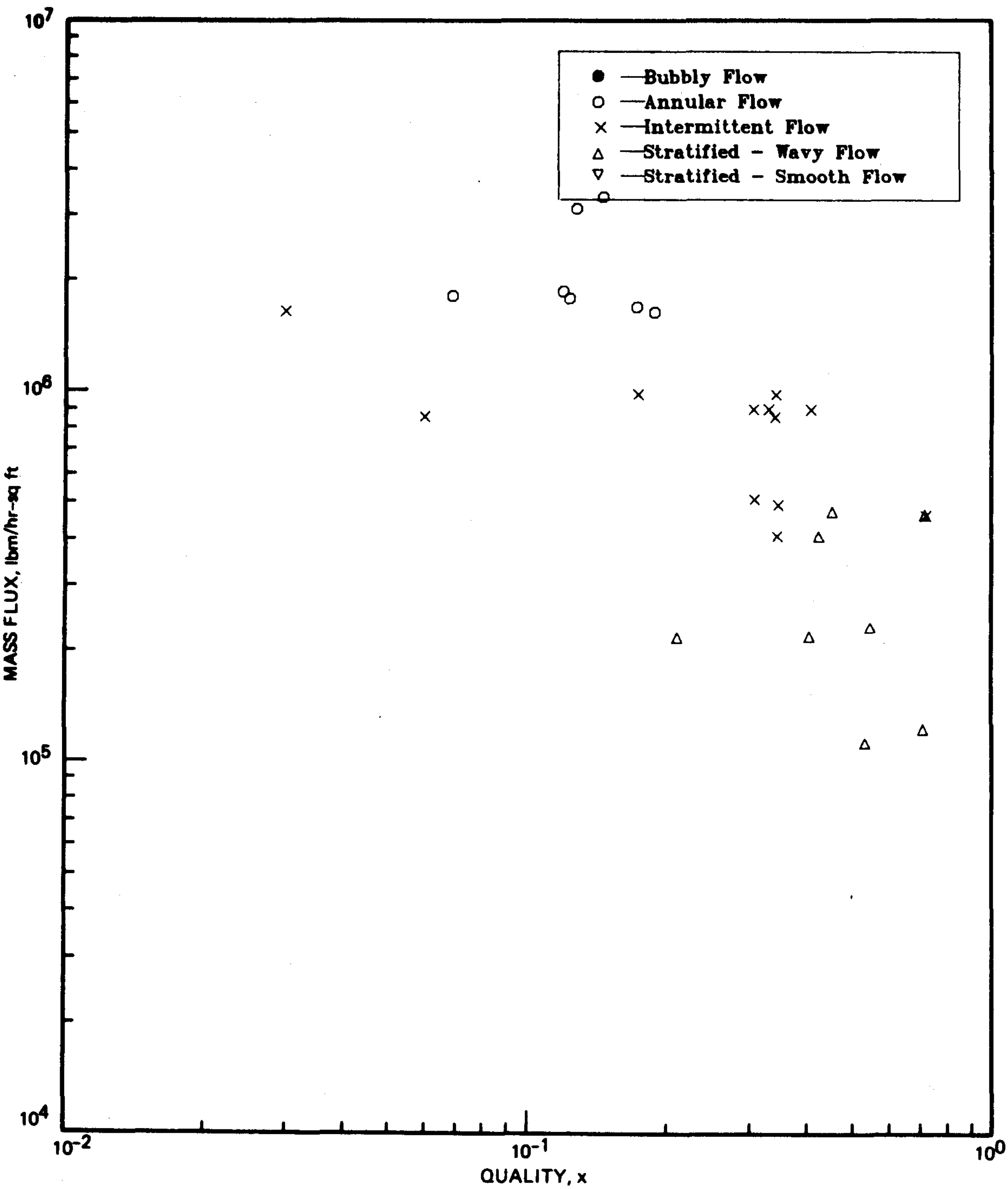

FIGURE IV-4. EXPERIMENTAL FLOW MAP, 250°F.

flow existed were immediately after heat had been applied to the test section. For a period of a few minutes, bubbly flow did exist in the sight glass; however, the regime soon changed to one of the other patterns.

2. As the mass flux was increased at the same saturation temperature, the size of the intermittent flow regime was reduced. Stratified flow also became less prevalent.

3. As the saturation temperature was increased at the same mass flux, the stratified flow regime tended to become less pronounced.

4. In a technical sense, purely stratified flow was rarely observed. As stated earlier in many of the runs where stratified-smooth or stratified-wavy flow appeared to exist in the sight glass, a very thin film of liquid could be observed with the naked eye at the top and sides of the sight glass. The circumferential wall temperature readings at a single cross section taken during this type of run tended to verify the presence of this film. These temperatures were essentially constant which would imply that a liquid film was present. Purely stratified flow occurred only at the lowest mass and heat fluxes.

Comparison with Existing Flow Maps

Several generalized flow maps were discussed in a previous section. The experimental data obtained during this investigation will now be compared to some of these maps.

Baker [4] Flow Map. The results obtained by applying the Baker [4] map are shown in Figures IV-5 to IV-8. Generally speaking, the results were poor, especially at the higher saturation temperatures. The intermittent flow regime was predicted with fair accuracy; however, the remaining flow regimes were predicted very poorly. It should be noted that several investigators, including Knowles [10, 11] and Taitel and Dukler [14], also obtained poor results when applying the Baker [4] map to their data.

Quandt [9] Flow Map. Considerably better results were obtained with Quandt's [9] flow map, as Figures IV-9 to IV-12 would indicate. This map does a satisfactory job predicting the intermittent, and annular flow regimes. Unfortunately, it does not do very well in predicting the remaining regimes. The Quandt [9] map could probably be used with a fair amount of confidence when applied to the annular and intermittent flow regimes.

Knowles [10, 11] Flow Map. Several problems of an analytical nature were encountered while attempting to apply this flow map to the present set of data. Although Knowles' work is discussed in several references [10, 11], a consistent set of equations could not be found. Thus, Knowles, himself, was contacted and from discussions with him as well as additional conversations with Eaton, his former co-worker, the correct set of equations was obtained.

As Figures IV-13 through IV-16 indicate, Knowles [10, 11] method is very accurate in predicting the stratified-smooth, stratified-wavy,

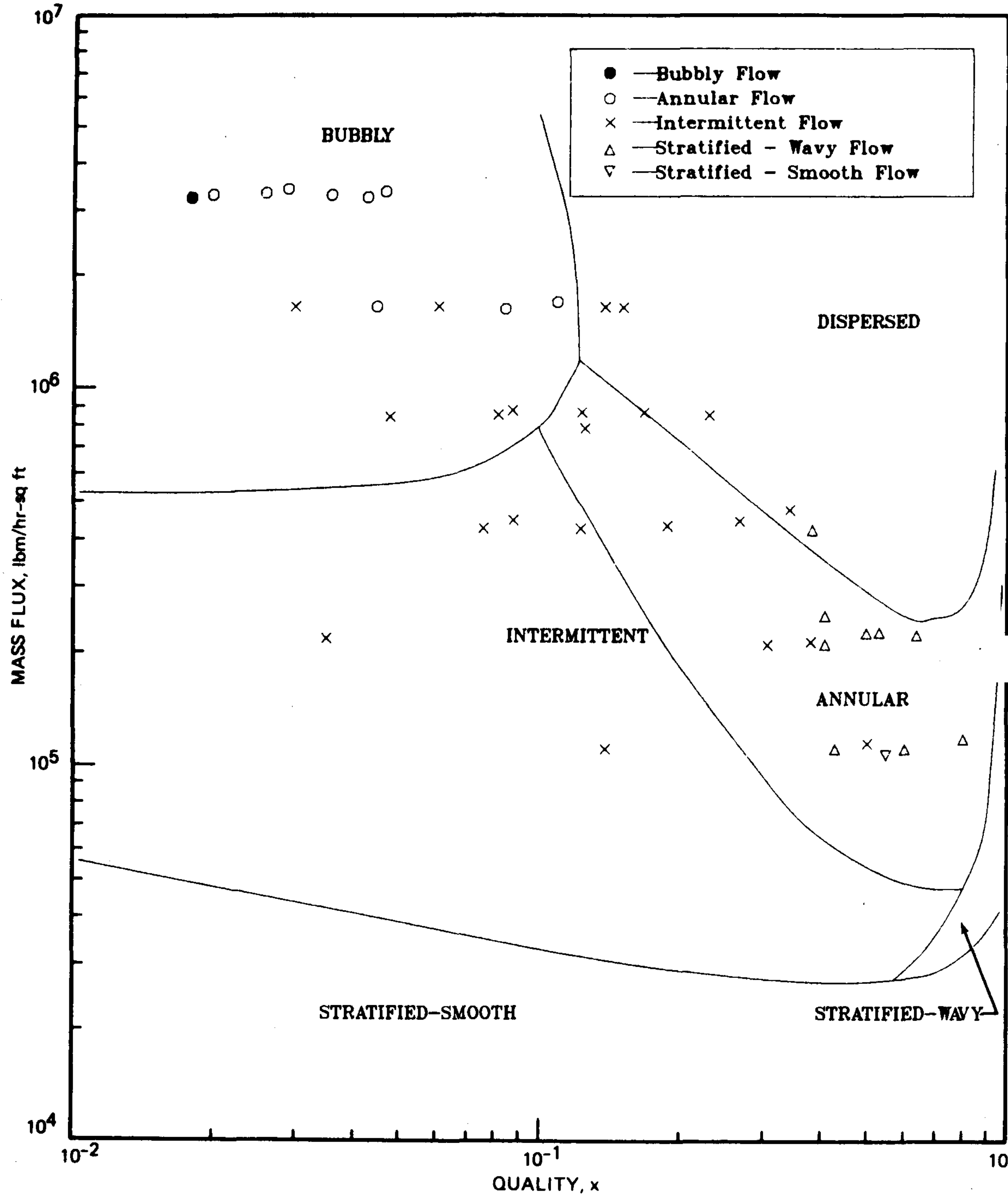

FIGURE IV-5. COMPARISON OF EXPERIMENTAL DATA WITH BAKER [4] MAP, 100°F.

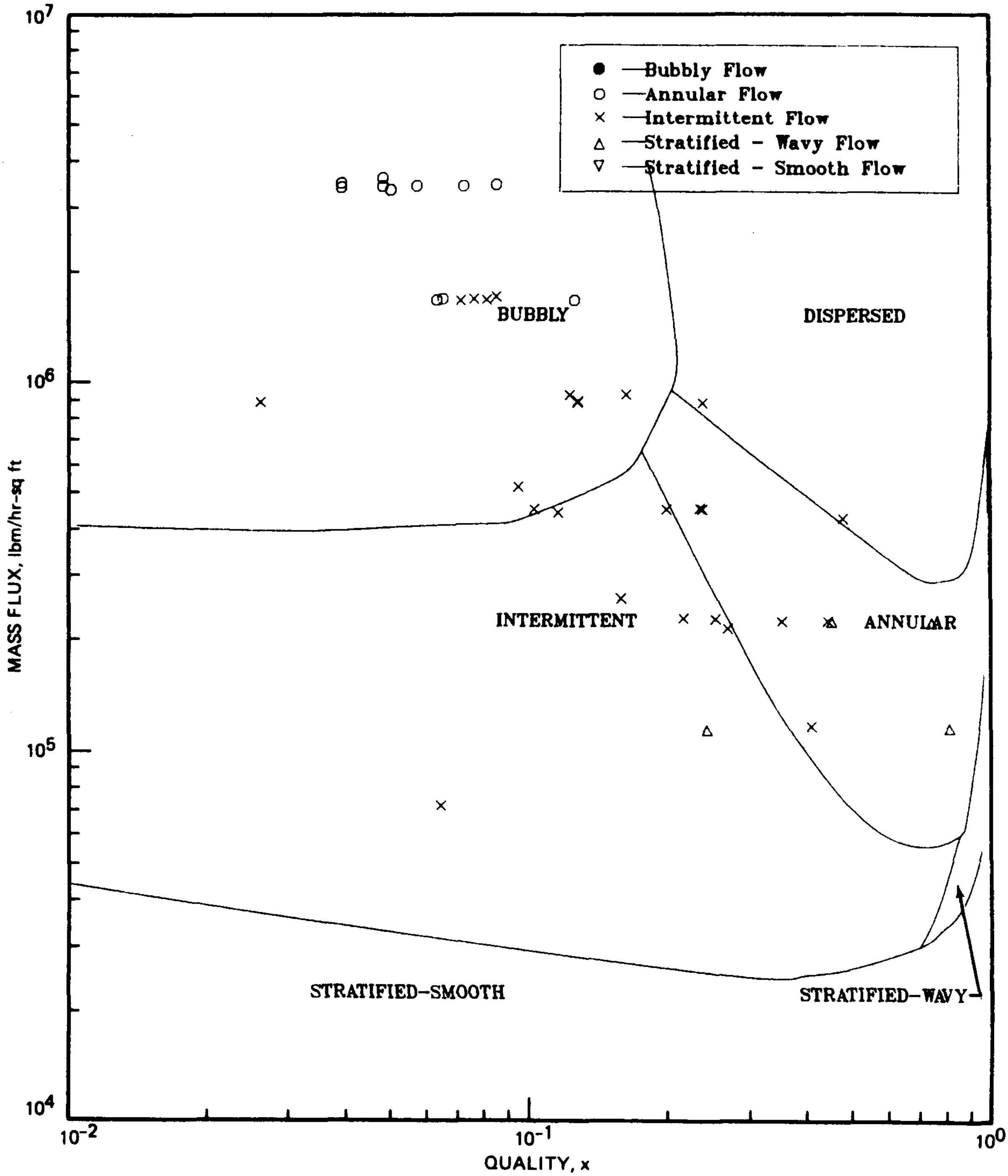

FIGURE IV-6. COMPARISON OF EXPERIMENTAL DATA WITH BAKER [4] MAP, 150°F.

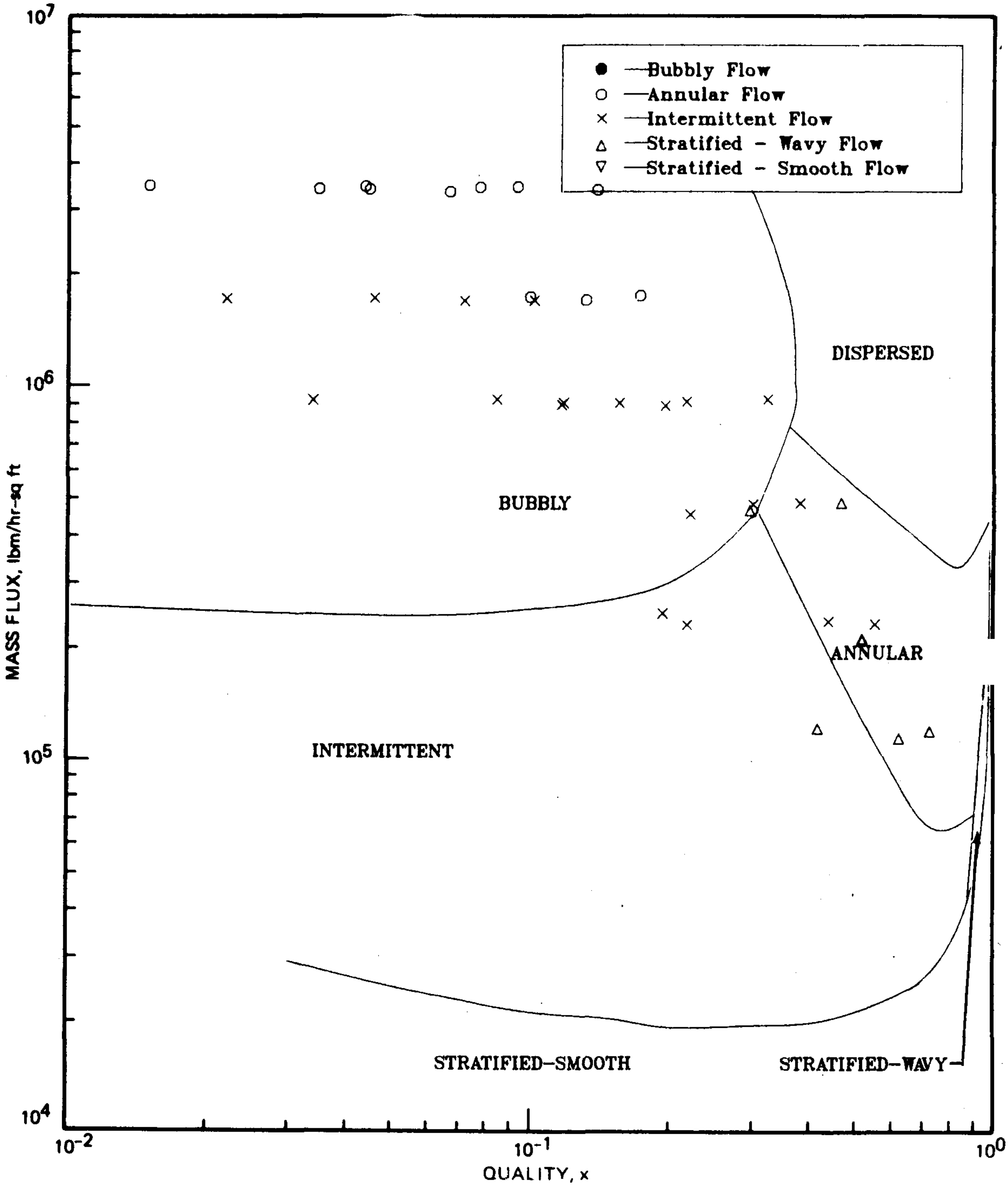

FIGURE IV-7. COMPARISON OF EXPERIMENTAL DATA WITH BAKER [4] MAP, 200°F.

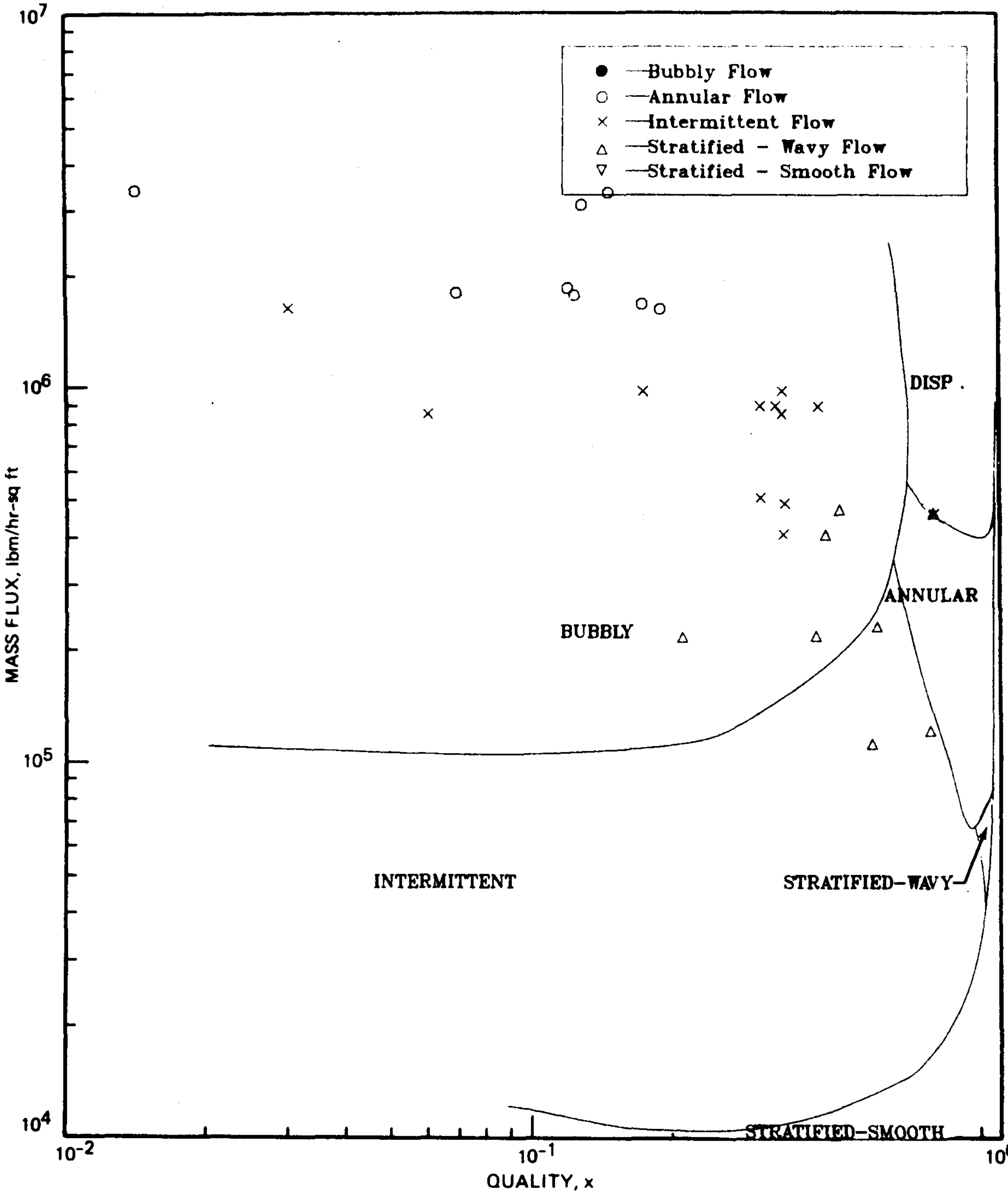

FIGURE IV-8. COMPARISON OF EXPERIMENTAL DATA WITH BAKER [4] MAP, 250°F.

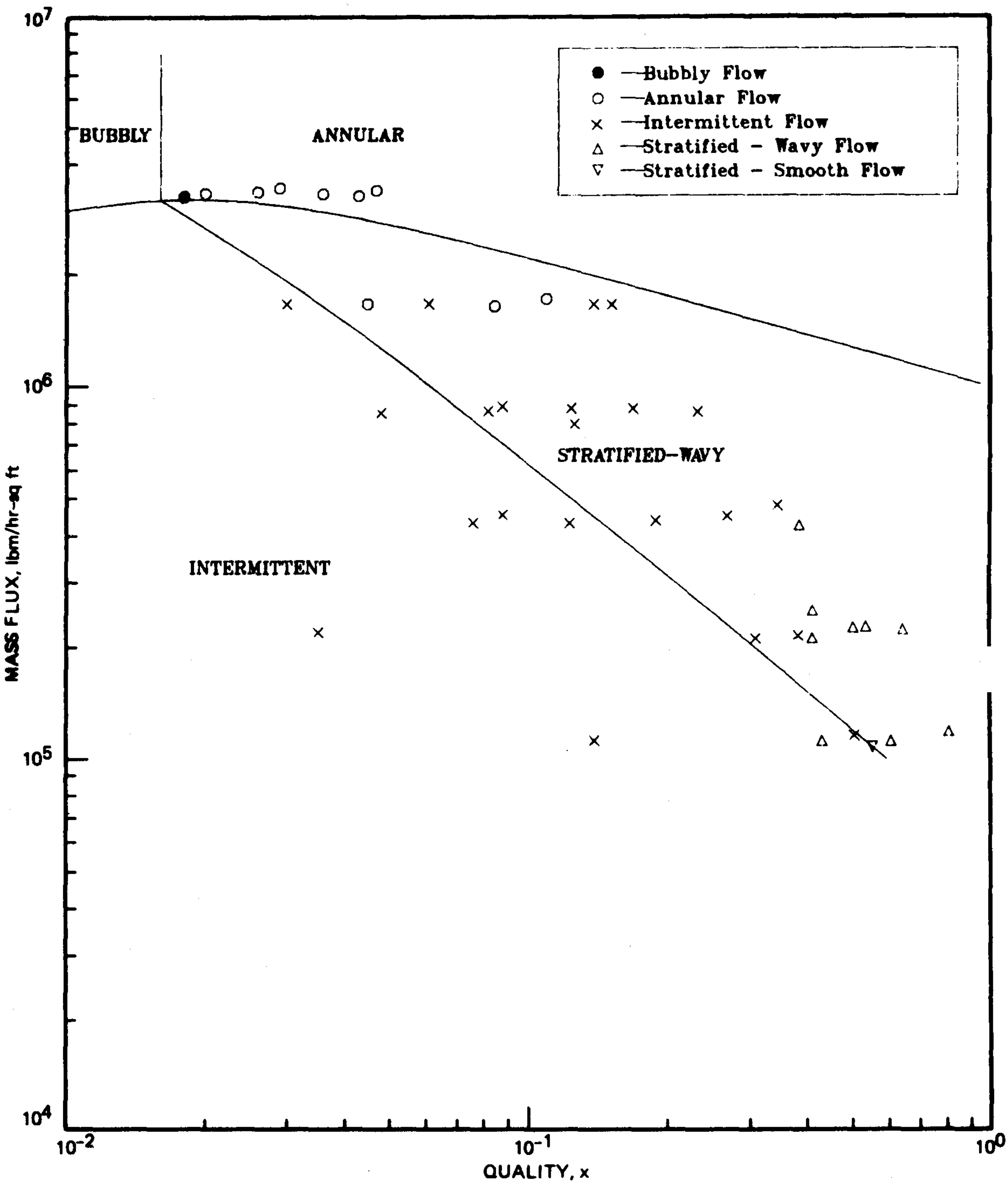

FIGURE IV-9. COMPARISON OF EXPERIMENTAL DATA WITH QUANDT [9] MAP, 100°F.

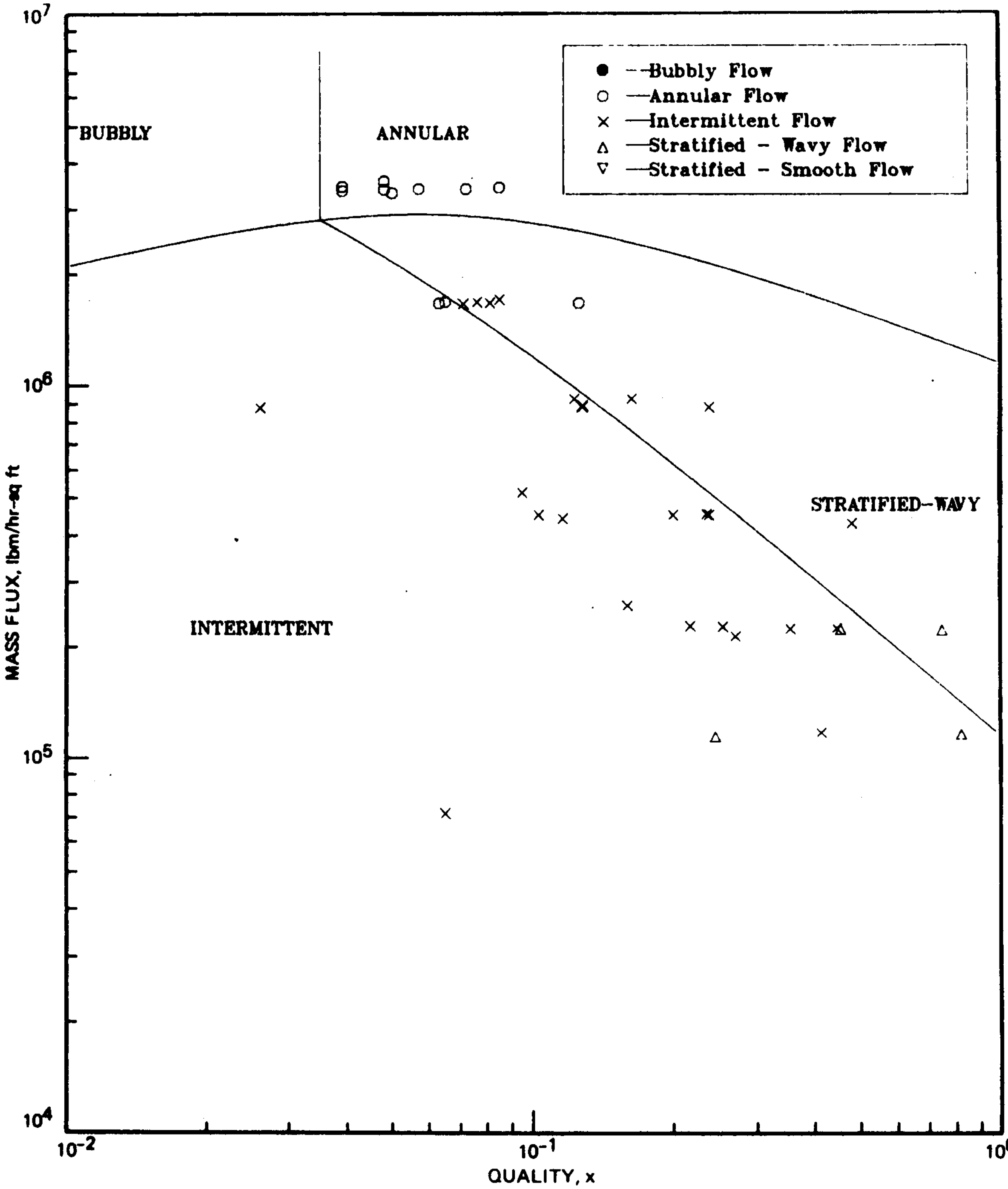

FIGURE IV-10. COMPARISON OF EXPERIMENTAL DATA WITH QUANDT [9] MAP, 150°F.

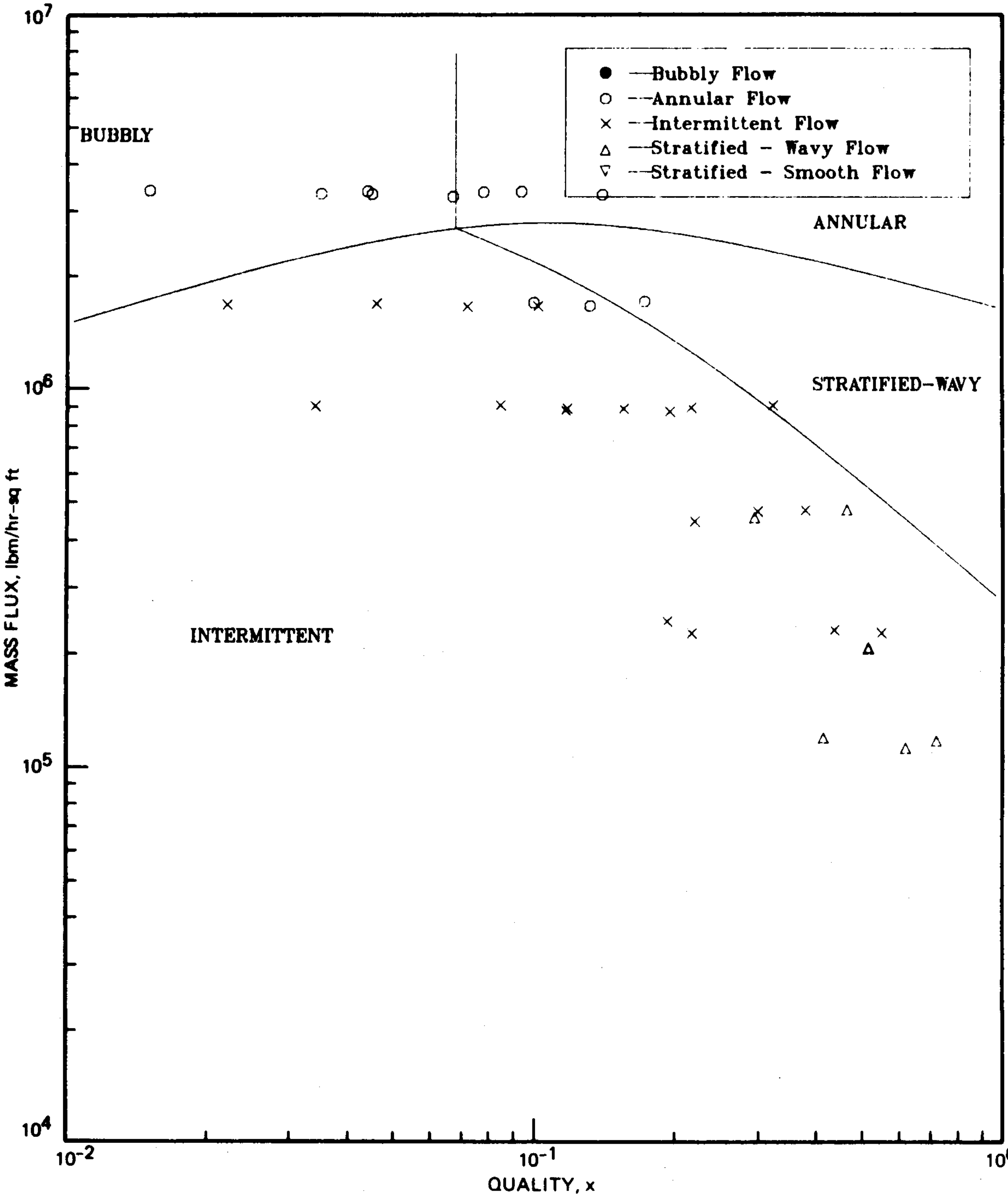

FIGURE IV-11. COMPARISON OF EXPERIMENTAL DATA WITH QUANDT [9] MAP, 200°F.

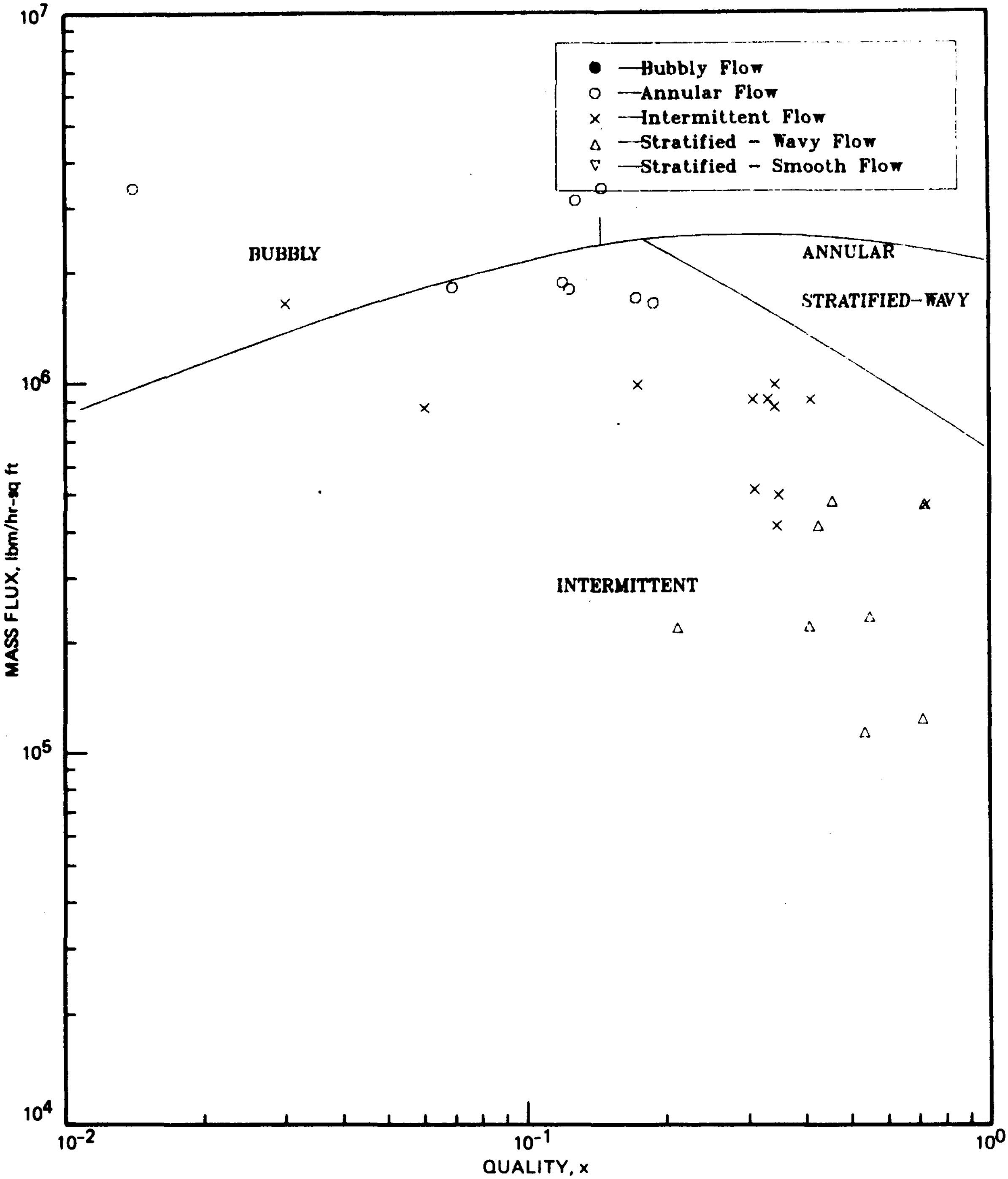

FIGURE IV-12. COMPARISON OF EXPERIMENTAL DATA WITH QUANDT [9] MAP, 250°F.

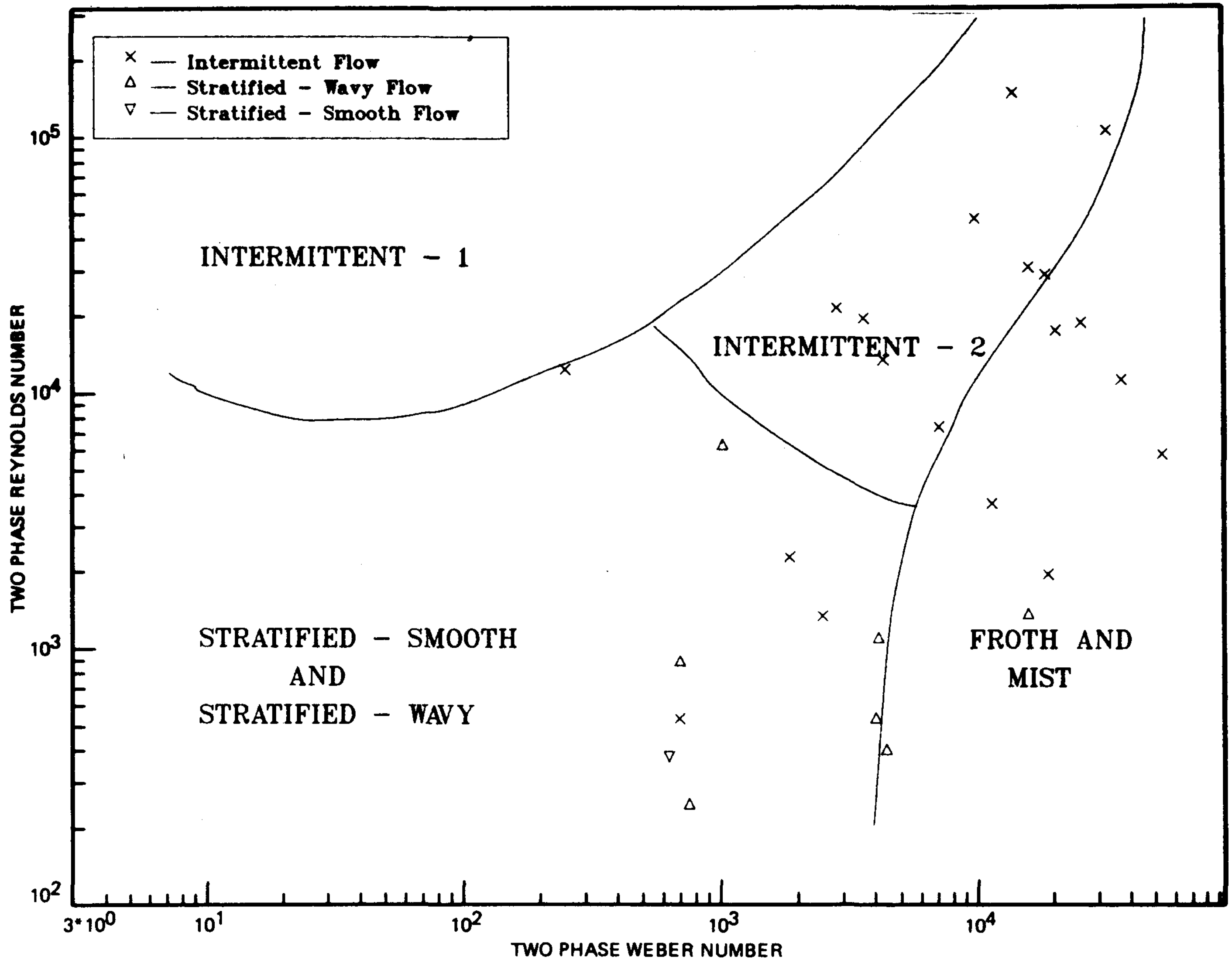

FIGURE IV-13. COMPARISON OF EXPERIMEN _ DATA WITH THE KNOWLES [10, 11] MAP, 100°F.

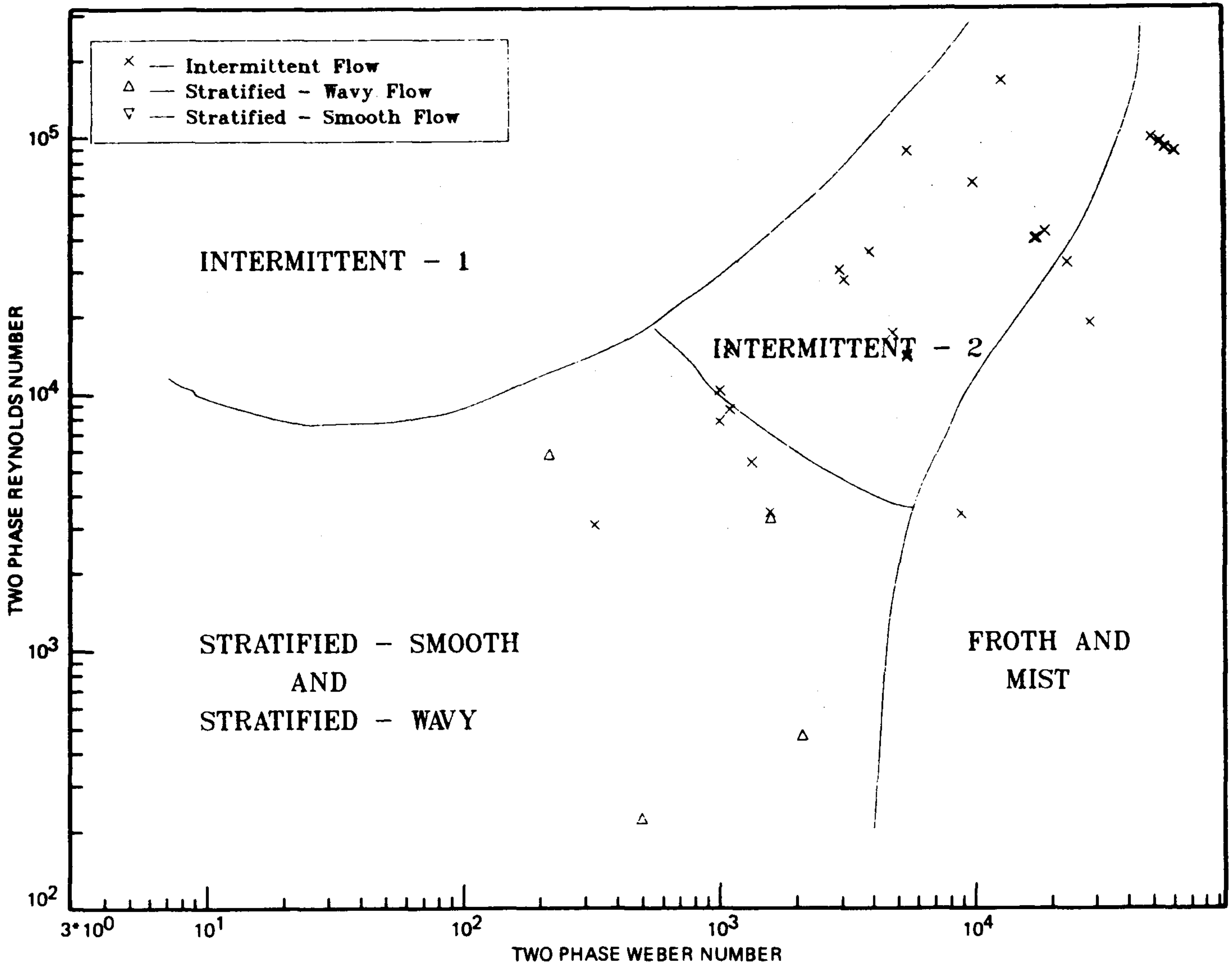

FIGURE IV-14. COMPARISON OF EXPERIMENTAL DATA WITH THE KNOWLES [10, 11] MAP, 150°F.

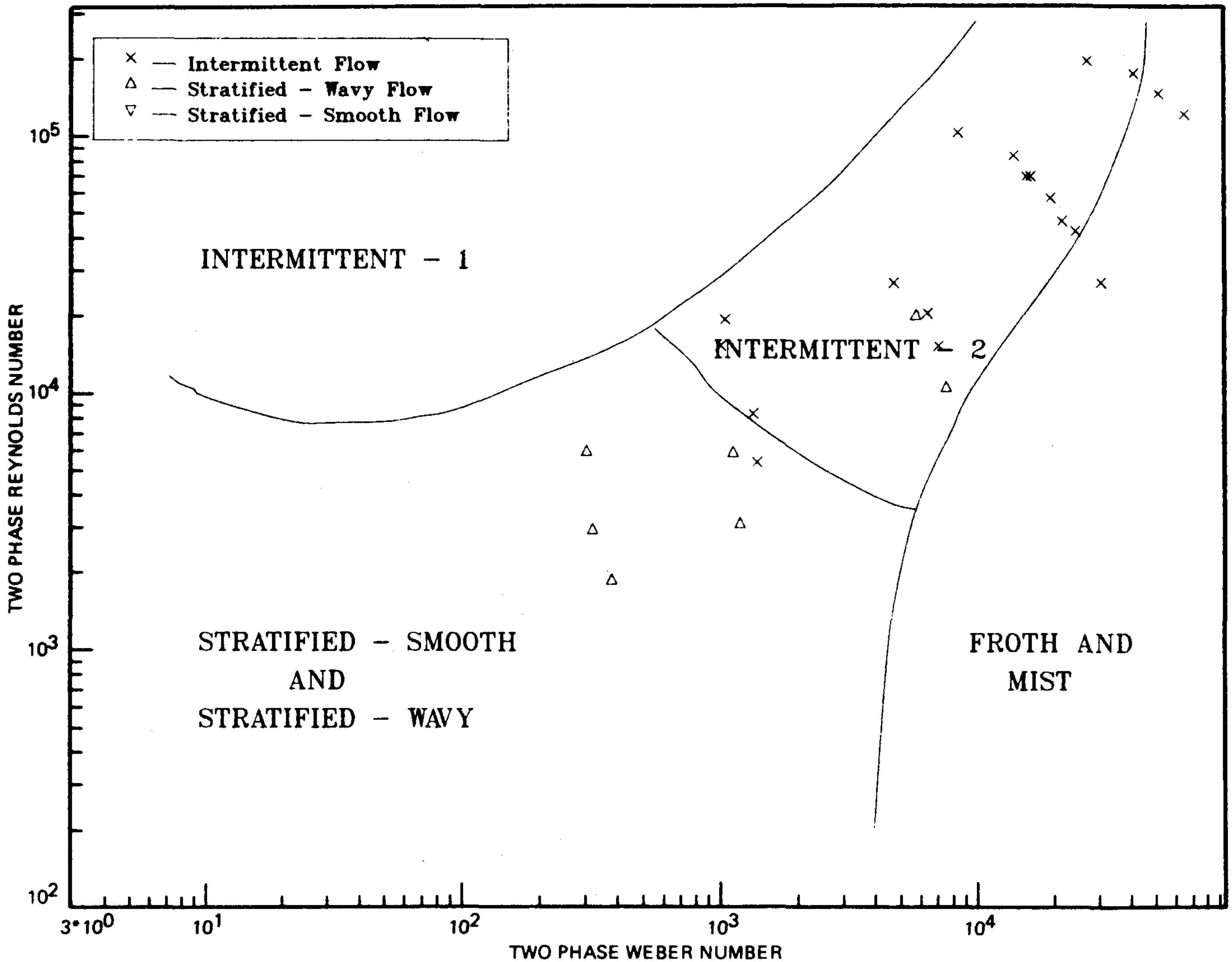

FIGURE IV-15. COMPARISON OF EXPERIMENT DATA WITH THE KNOWLES [10, 11] MAP, 200°F.

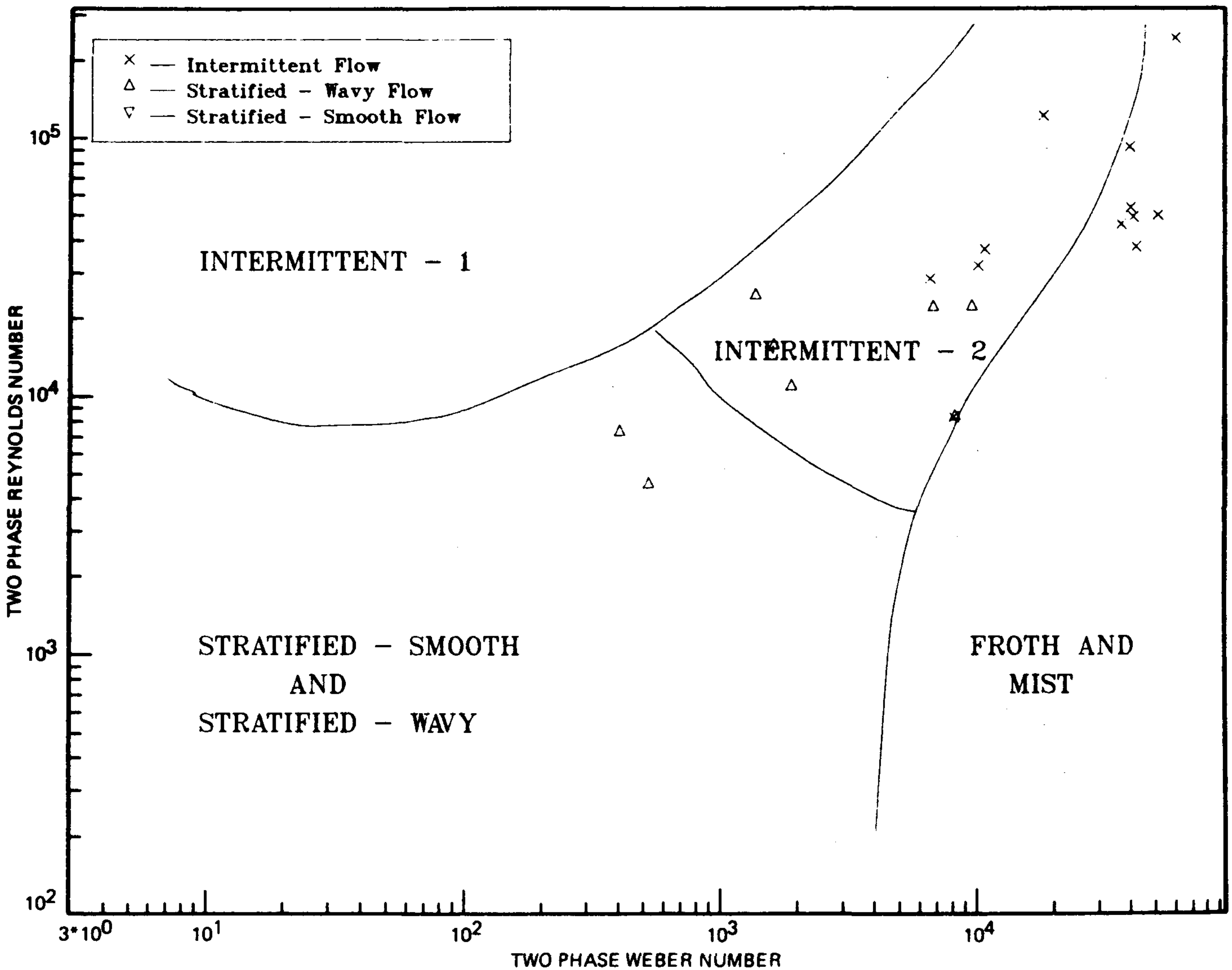

FIGURE IV-16. COMPARISON OF EXPERIMENTAL DATA WITH THE KNOWLES [10, 11] MAP, 250°F.

and intermittent flow regime data at saturation temperatures of 100°F, 150°F, and 200°F. Since Knowles [10, 11] attempted to distinghish between the plug and slug flow regimes, the regimes Intermittent-1 and Intermittent-2 are shown in the figure. No data were obtained which corresponded to the Intermittent-1 regime. The results obtained at 250°F were only fair, Figure IV-16. This was most likely due to the fact that one of the *influence numbers* in Eaton's [10, 11] hold-up correlation, upon which Knowles' work is based, was beyond the applicable range. Even with this short-coming Knowles' [10, 11] map is still the best available for predicting the stratified-smooth, stratified-wavy, and intermittent flow regimes. Degance and Atherton [13] cautioned against applying this correlation beyond the recommended ranges for $N_{Re_{tp}}$ and $N_{We_{tp}}$; however, none of the data obtained during this study were outside the recommended ranges.

From the results discussed above, it would seem that accurate R-114 flow regime prediction can be obtained by combining Quandt's [9] analysis and Knowles' [10, 11] predictive procedures. For instance, Knowles' [10, 11] map could be used initially to determine if the flow regime is stratified or intermittent. If this comparison yields affirmative results, then it may be concluded that either stratified or intermittent flow should actually exist. On the other hand, should the comparison with Knowles' [10, 11] map yield negative results, then the Quandt [9] map should be used to predict the existence of the annular flow regime. Quandt's [9] analysis should also provide an accurate prediction of the intermittent flow regime.

B. PRESSURE DROP

Introduction

Over 300 experimental data points reflecting two-phase pressure drop in a horizontal tube were obtained during this program. These data are tabulated in Appendix C. Data were obtained by operating the system in both the natural and forced circulation modes. Since no differences in results could be distinguished, both the forced circulation data and natural circulation data were considered together.

In the most general case of fluid flow in a tube, total pressure loss is comprised of three components:

1. The loss due to acceleration of the fluid caused by density changes,

2. Gravitational losses resulting from changes in elevation, and

3. Frictional losses due to an irreversible interaction between the fluid and the tube wall.

Two-phase, one-component flow in a horizontal tube when heat is continuously added to the flowing fluid is a special case of the above, since gravitational losses are nonexistent. Using the basic Homogeneous [2] and Martinelli-Nelson [1] models, expressions were developed for calculating two-phase pressure losses due to friction and acceleration. In order to perform the necessary integrations over the tube length, each 1-ft portion of the test section was divided into 12 increments. Physical properties were assumed constant in any one increment.

By calculating both components of the total pressure loss, it was possible to compare the experimental, frictional pressure drop to the calculated, frictional pressure drop for the vast majority of experimental data obtained.

Results

The computer plotted results, obtained by comparing the Homogeneous [2] and Martinelli-Nelson [1] models to the experimental data, are shown in Figures IV-17 to IV-24. In these figures Φ_{ftt} is plotted versus $\sqrt{X_{tt}}$, since this type of graph provides the most basic and meaningful comparison. Instrumentation problems associated with the third pressure tap made it necessary to obtain the pressure drop between the second and fourth taps.

Homogeneous Model [2]. The basis of the Homogeneous model is that both the liquid and vapor are assumed to move with the same velocity. This assumption greatly simplifies calculations, since the void fraction is easily determined.

Figures IV-17 to IV-20 show the results obtained by applying the Homogeneous [2] model. In most cases, it can be seen that the frictional pressure loss measured experimentally exceeded that calculated from the model. This was not too surprising since most of the flow regimes observed were inhomogeneous in nature.

Martinelli-Nelson [1] Model. The results obtained from applying the M-N model on an equal reduced pressure basis are shown in Figures IV-21 to IV-24. In these graphs, Φ_{ftt} is also plotted versus $\sqrt{X_{tt}}$.

From a careful examination of the figures, the M-N [1] model would seem to offer better agreement with the experimental data than the Homogeneous [2] model at all the saturation temperatures considered

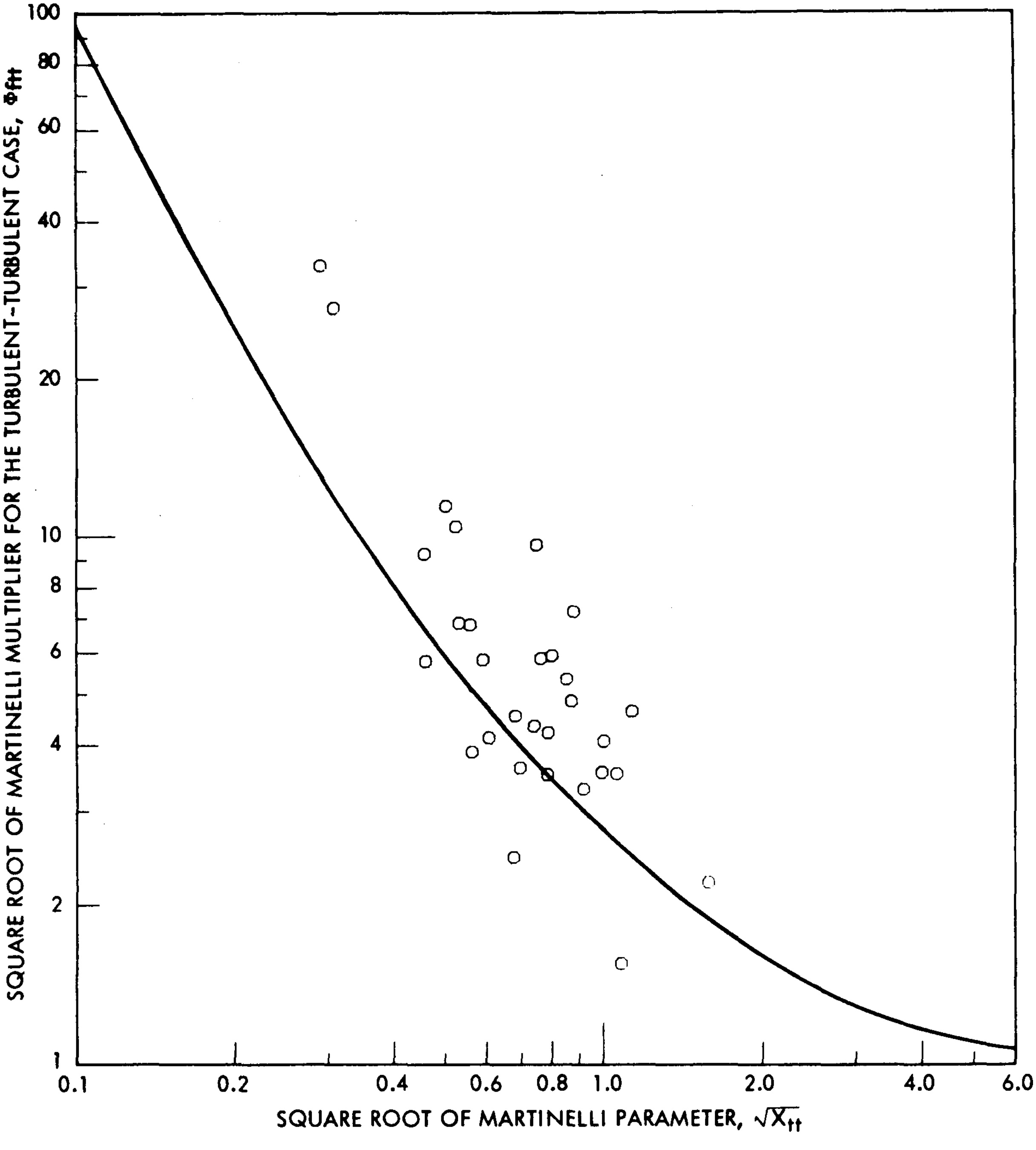

FIGURE IV-17. COMPARISON OF EXPERIMENTAL DATA WITH HOMOGENEOUS [2] MODEL, 100°F

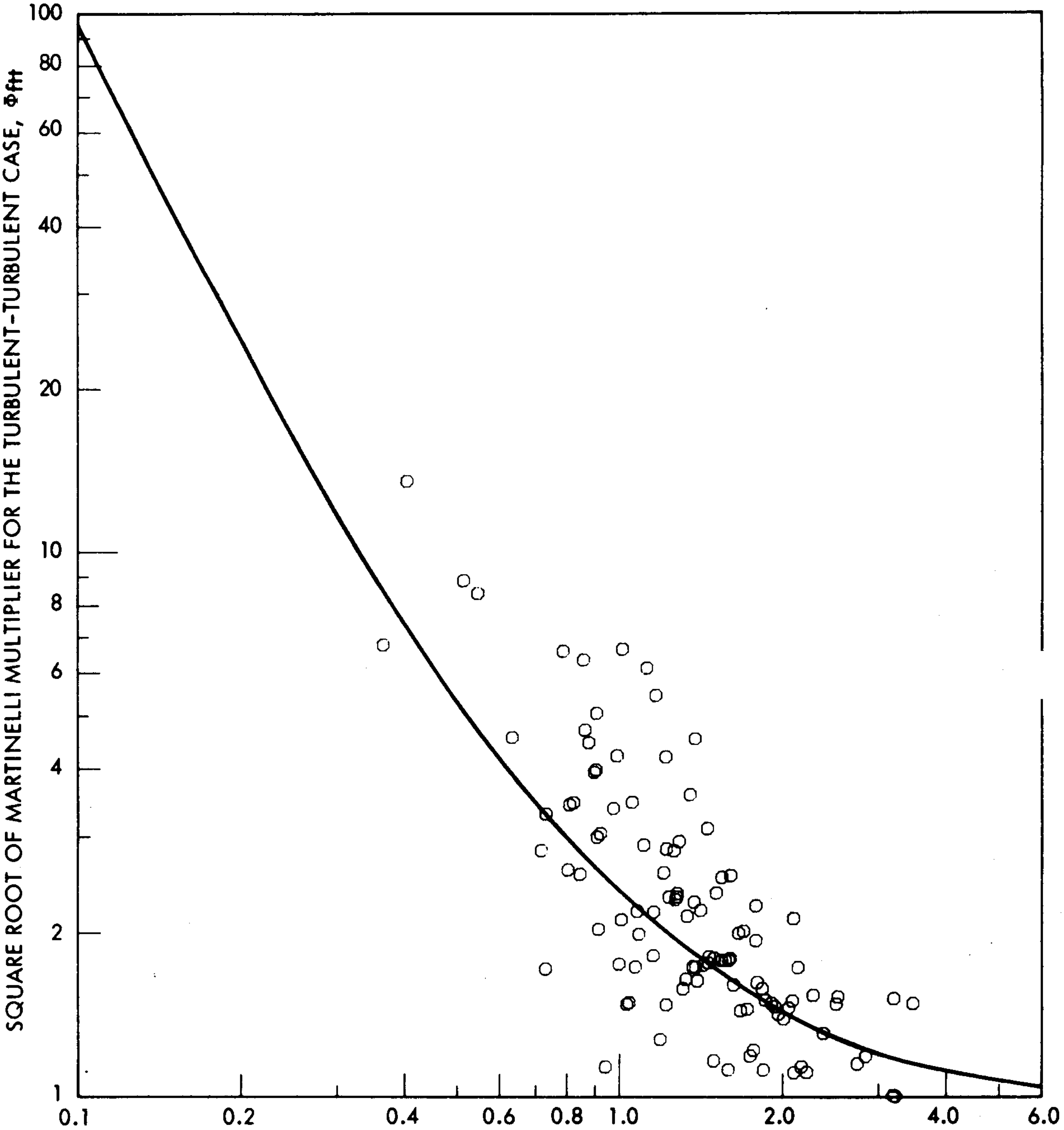

FIGURE IV-18. COMPARISON OF EXPERIMENTAL DATA WITH HOMOGENEOUS [2] MODEL, 150°F

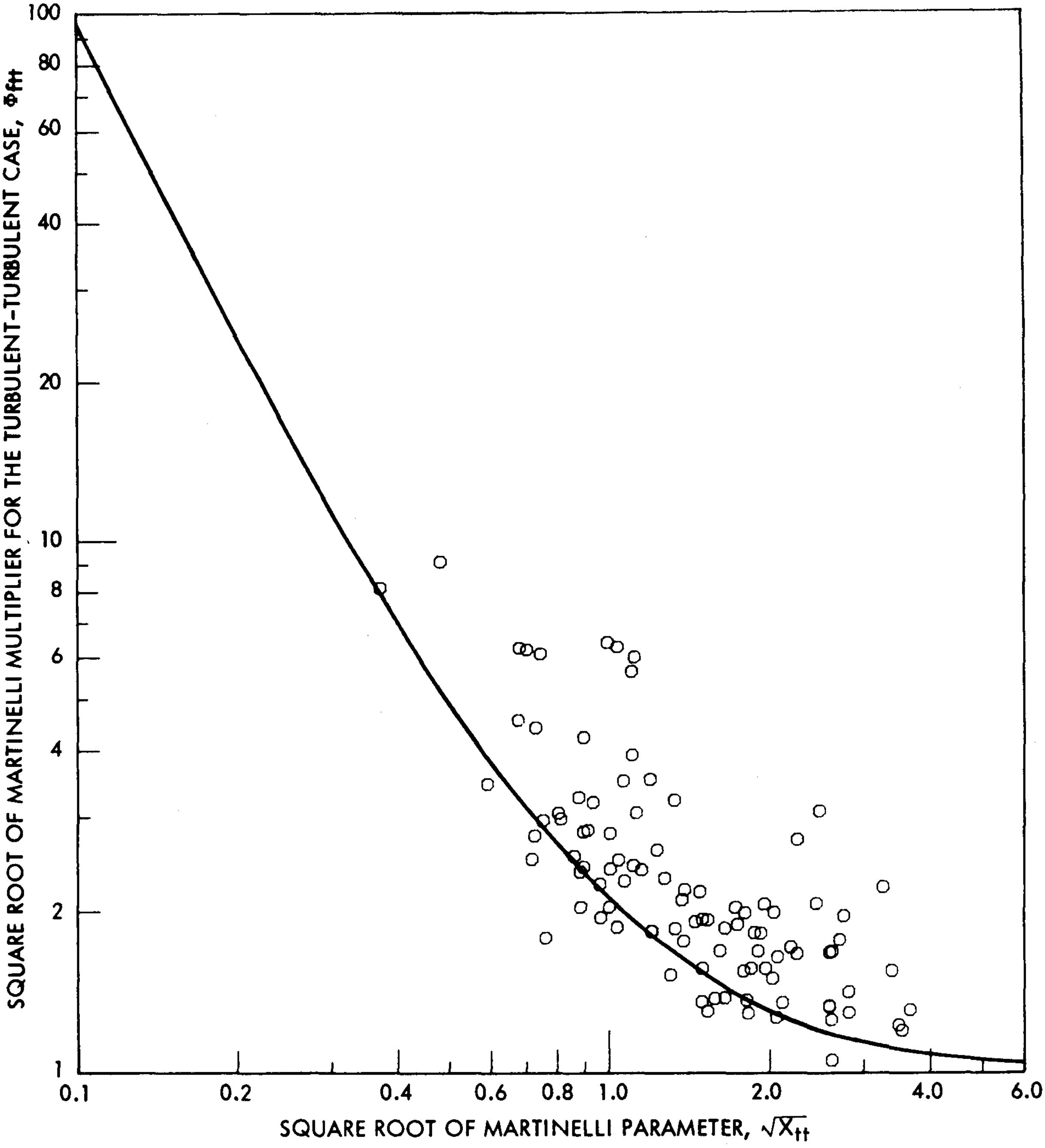

FIGURE IV-19. COMPARISON OF EXPERIMENTAL DATA WITH HOMOGENEOUS [2] MODEL, 200°F

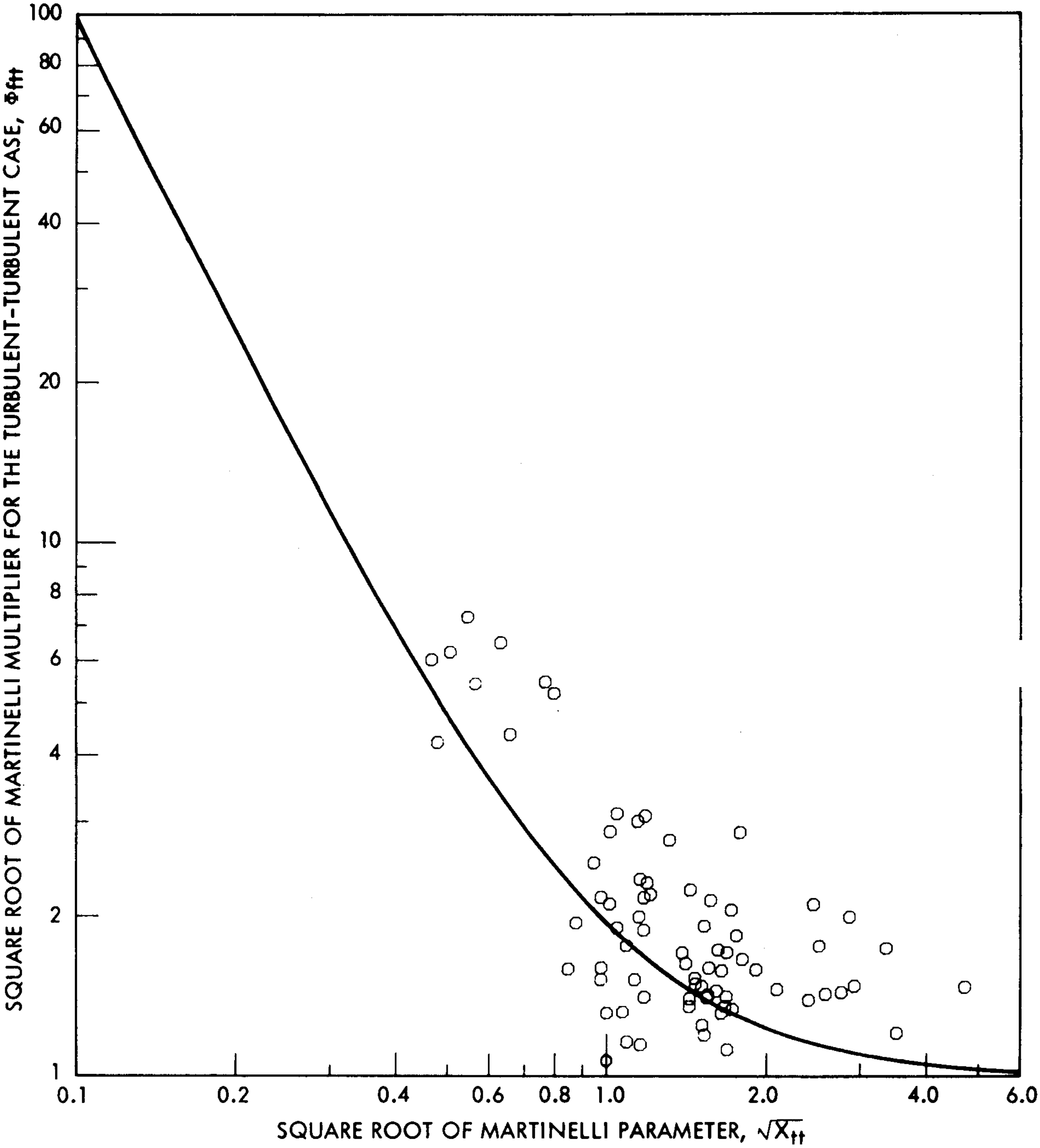

FIGURE IV-20. COMPARISON OF EXPERIMENTAL DATA WITH HOMOGENEOUS [2] MODEL, 250°F

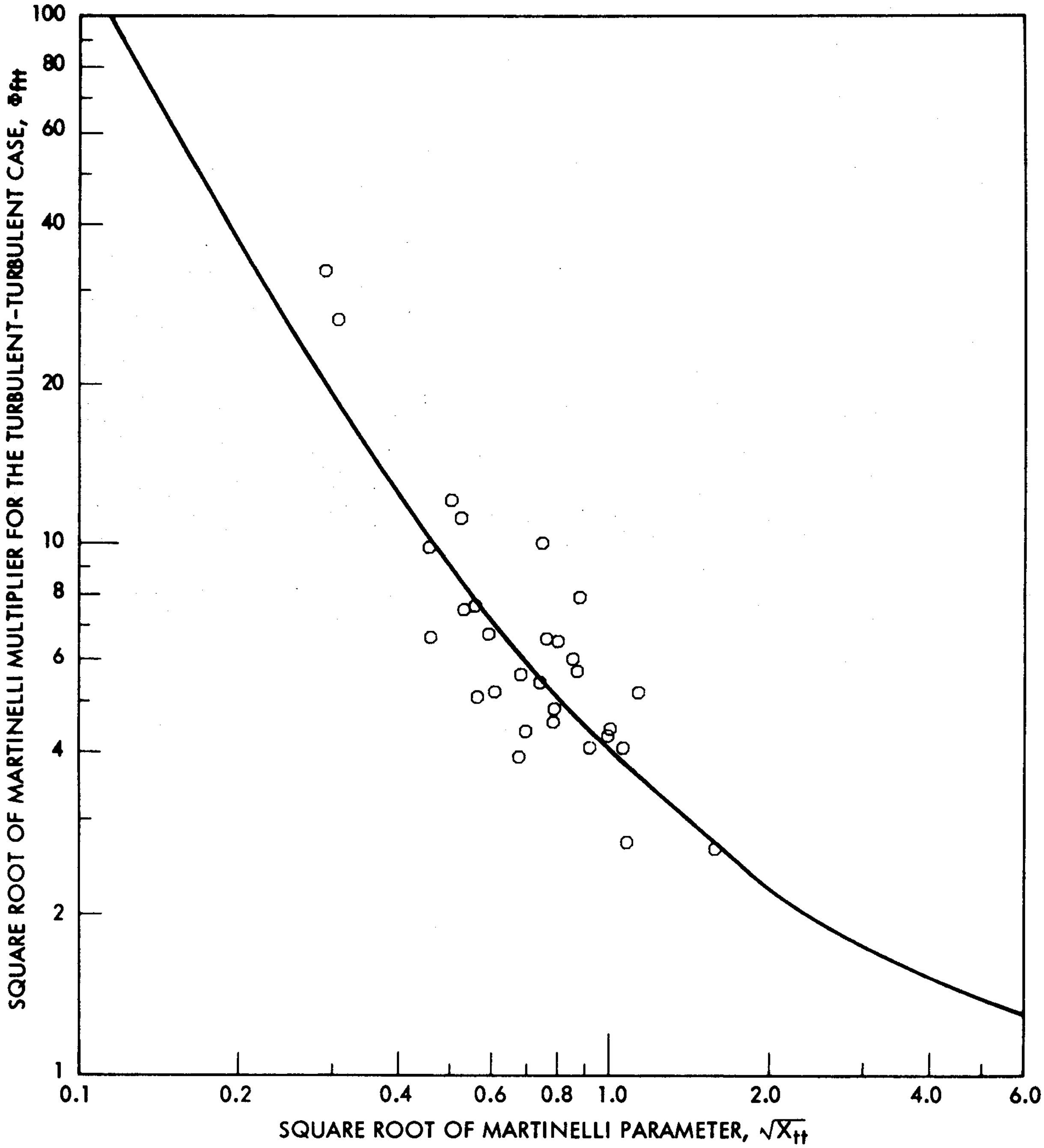

FIGURE IV-21. COMPARISON OF EXPERIMENTAL DATA WITH MARTINELLI-NELSON [1] MODEL, 100°F

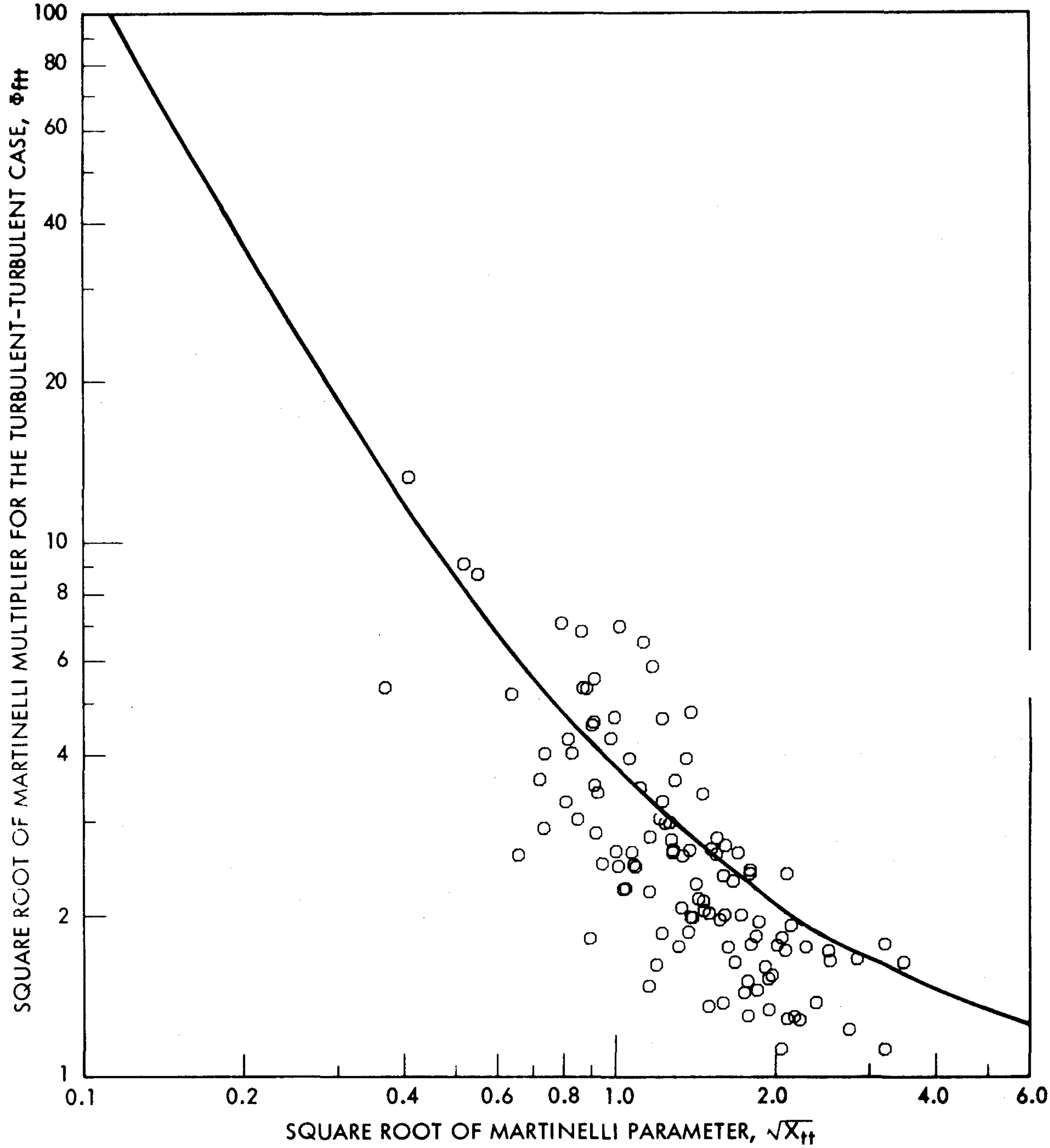

FIGURE IV-22. COMPARISON OF EXPERIMENTAL DATA WITH MARTINELLI-NELSON [1] MODEL, 150°F

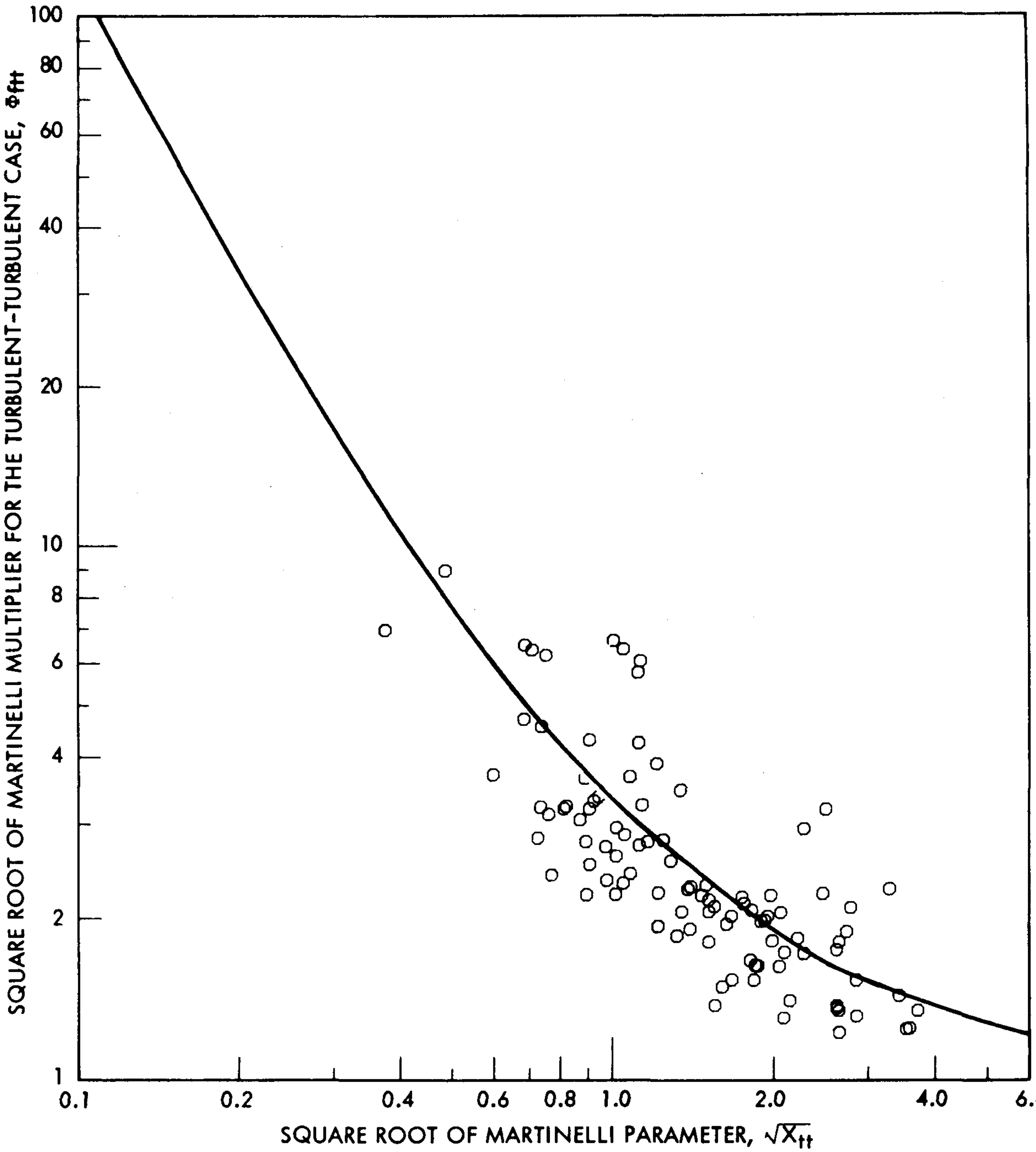

FIGURE IV-23. COMPARISON OF EXPERIMENTAL DATA WITH MARTINELLI-NELSON [1] MODEL, 200°F

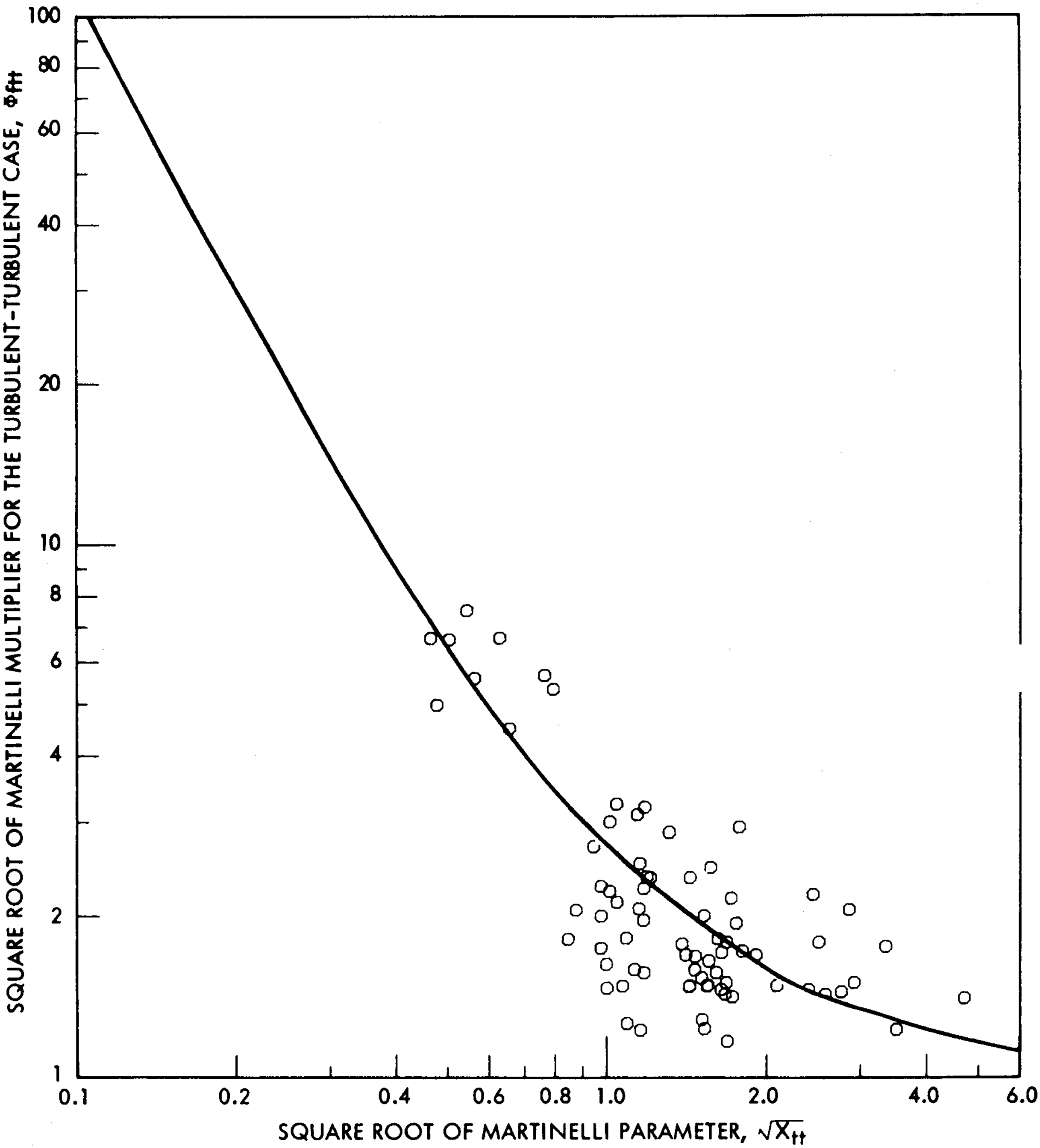

FIGURE IV-24. COMPARISON OF EXPERIMENTAL DATA WITH MARTINELLI-NELSON [1] MODEL, 250°F

here. Furthermore, even at the higher mass fluxes (1.70×10^6 to 3.40×10^6 lbm/hr-sq ft) where the Homogeneous [2] model should become accurate, the M-N model still gave reasonable results which tended to be on the conservative side.

The original M-N [1] model may be improved by including a different void fraction correlation which is more nearly representative of the particular fluid studied. Jallouk [42] employed this technique with some success when he modified Zuber, et al. [59] void fraction correlation and used it in the M-N [1] model to help analyze his R-114, two-phase pressure drop data. This same procedure was followed in the analysis of the data of this study; however, the results were slightly different. No significant improvement in agreement between model and experimental data resulted from inserting the Zuber, et al. [59] void fraction correlation, as modified by Jallouk [42], into the M-N [1] model. This can be explained by noting that Jallouk's [42] data reflected two-phase pressure drop in a vertical tube. Thus, he was forced to account for pressure losses due to gravitation, as well as acceleration and friction. Since the first two components are strongly dependent on void fraction, the particular void fraction correlation he employed had a greater impact on his results than on the results of this study where gravitational pressure losses did not exist.

Other Pressure Drop Models

Other investigators have also noted a mass flux effect on pressure drop data and have attempted to account for it by developing new correlations or modifying existing ones. The mass flux effect is believed to be caused by two factors: first, as flow or mass flux increases, frictional losses tend to become an increasingly larger portion of the total pressure loss. Thus, whatever inaccuracies that are built into the calculation of acceleration pressure losses, namely void fraction, become increasingly insignificant. Secondly, as mass flux increases, the stratified-smooth, stratified-wavy, and intermittent flow regimes, which are very inhomogeneous, are less likely to exist. This can usually be confirmed by flow regime studies. A reliable correlation should account for this effect. The text which follows is devoted to applying several of these correlations to the present set of experimental data. Although the data exhibited considerable scatter certain trends could be noted. An indication of the experimental uncertainty associated with the data is shown on some of the figures.

Baroczy [24] Correlation. The majority of the middle-to-high mass flux data of this study were predicted with fair accuracy by this correlation. Extrapolation was required to obtain curves in Figures IV-25 to IV-38 at mass fluxes of 0.12×10^6 lbm/hr-sq ft. It should also be noted that some of the curves generated exhibit inconsistent slopes in certain areas. This was caused by large variations in Baroczy's [24] mass flux correction factor. The Baroczy [24] correlation tended to underestimate the experimental data corresponding to stratified flow conditions.

Chisholm's [29, 30] Correlations. Two of Chisholm's [29, 30] correlations are shown in Figures IV-25 to IV-38, along with the experimental data. The most recent correlation (1973) is seen to plot slightly below the earlier work. Agreement with either correlation is fair-to-good for most of the low mass flux data. Chisholm's [29, 30] correlations have an advantage over Baroczy's [24] work in that they are more

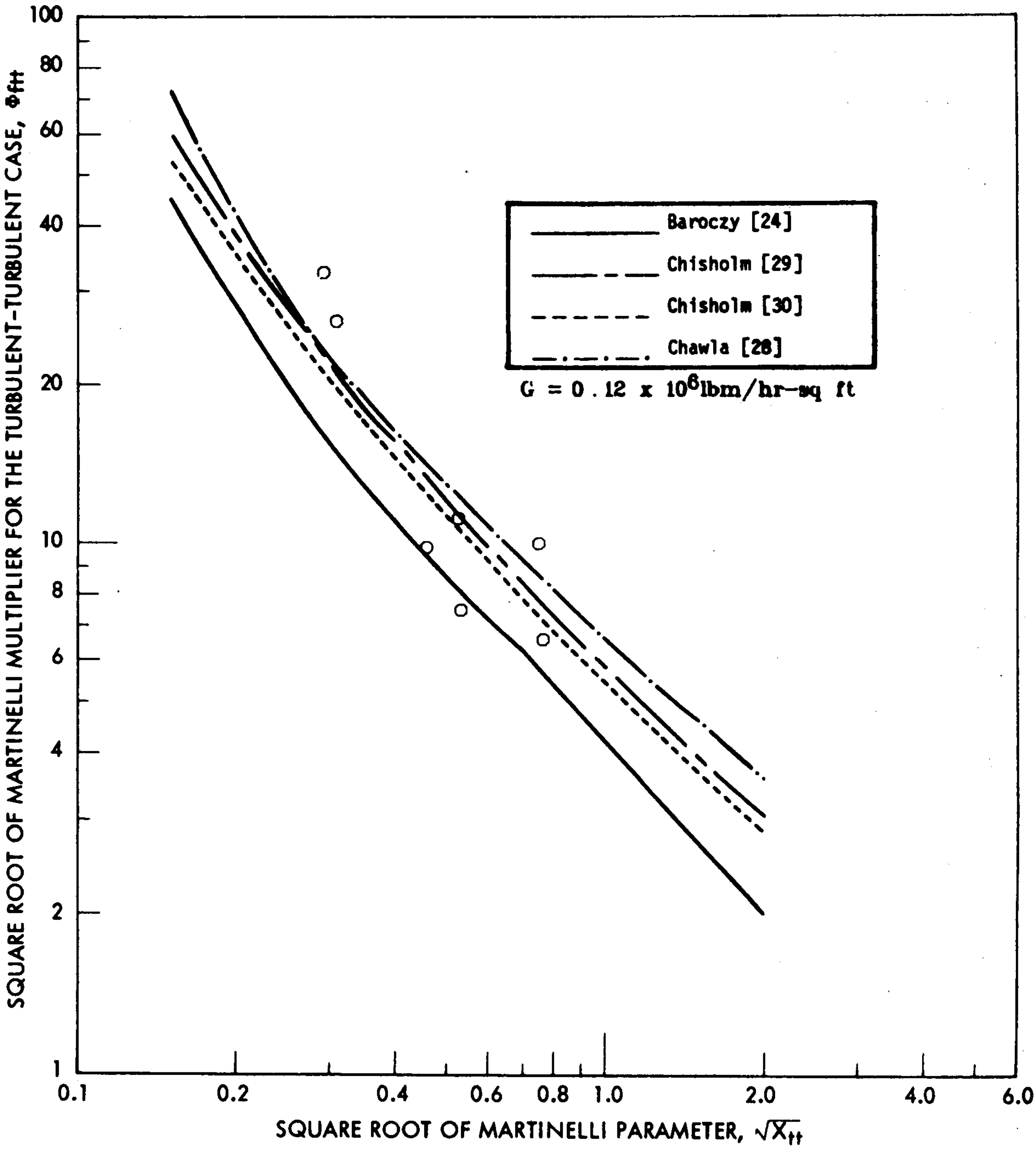

FIGURE IV-25. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 100°F, 0.12 x 10^6 lbm/hr-sq ft.

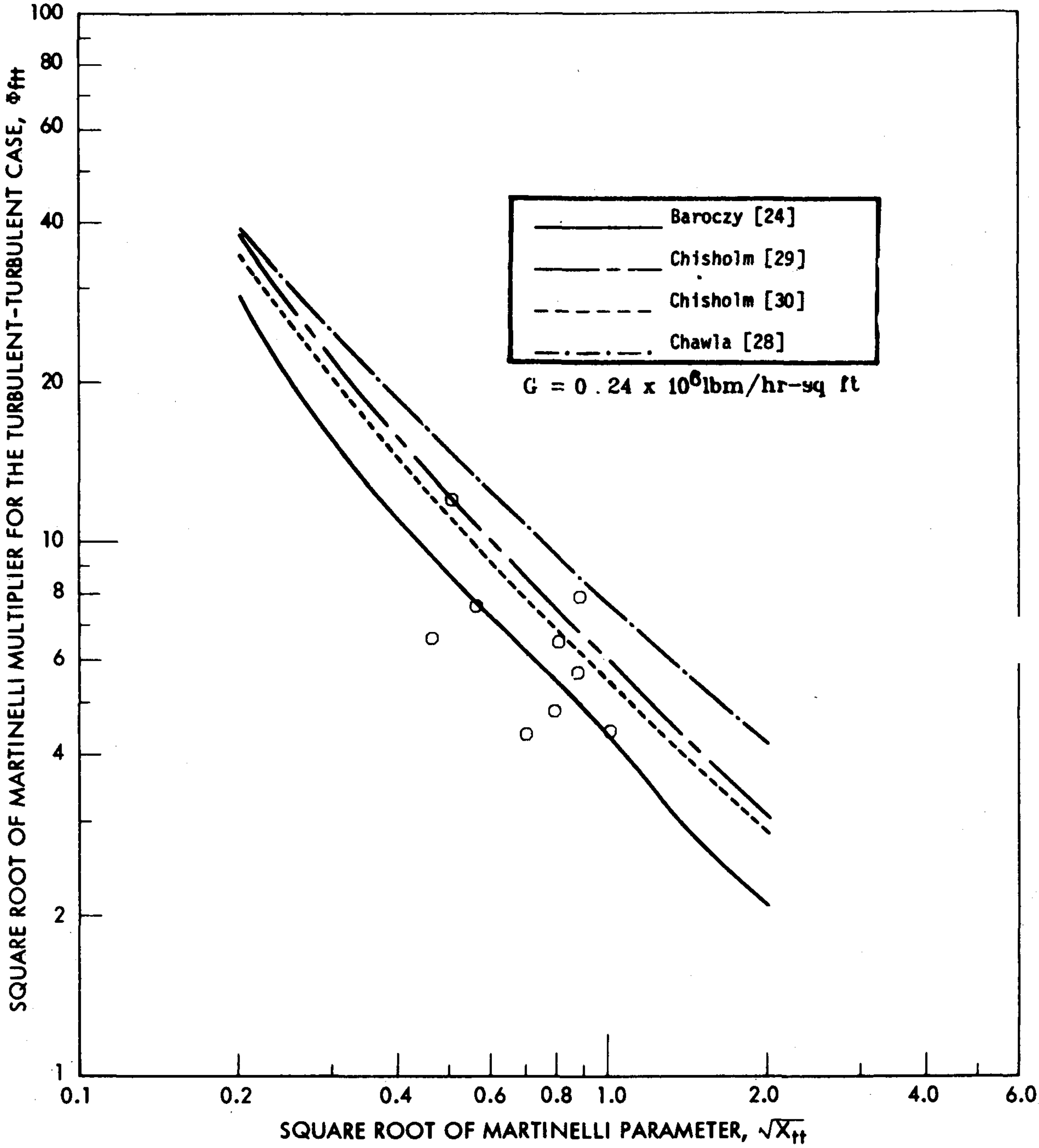

FIGURE IV-26. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 100°F , 0.24 x 10^6 lbm/hr-sq ft.

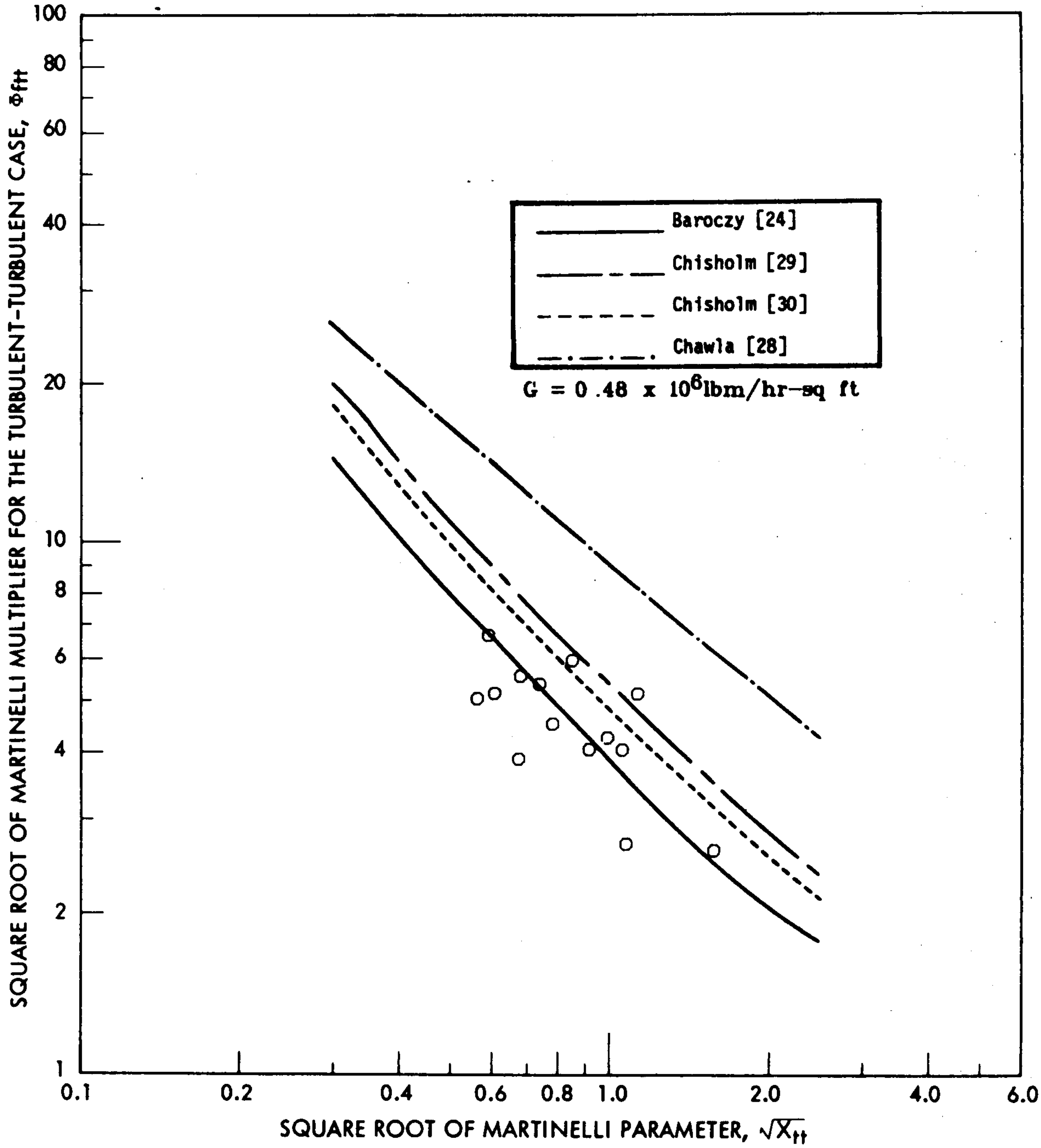

FIGURE IV-27. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 100°F, 0.48 x 10^6 lbm/hr-sq ft.

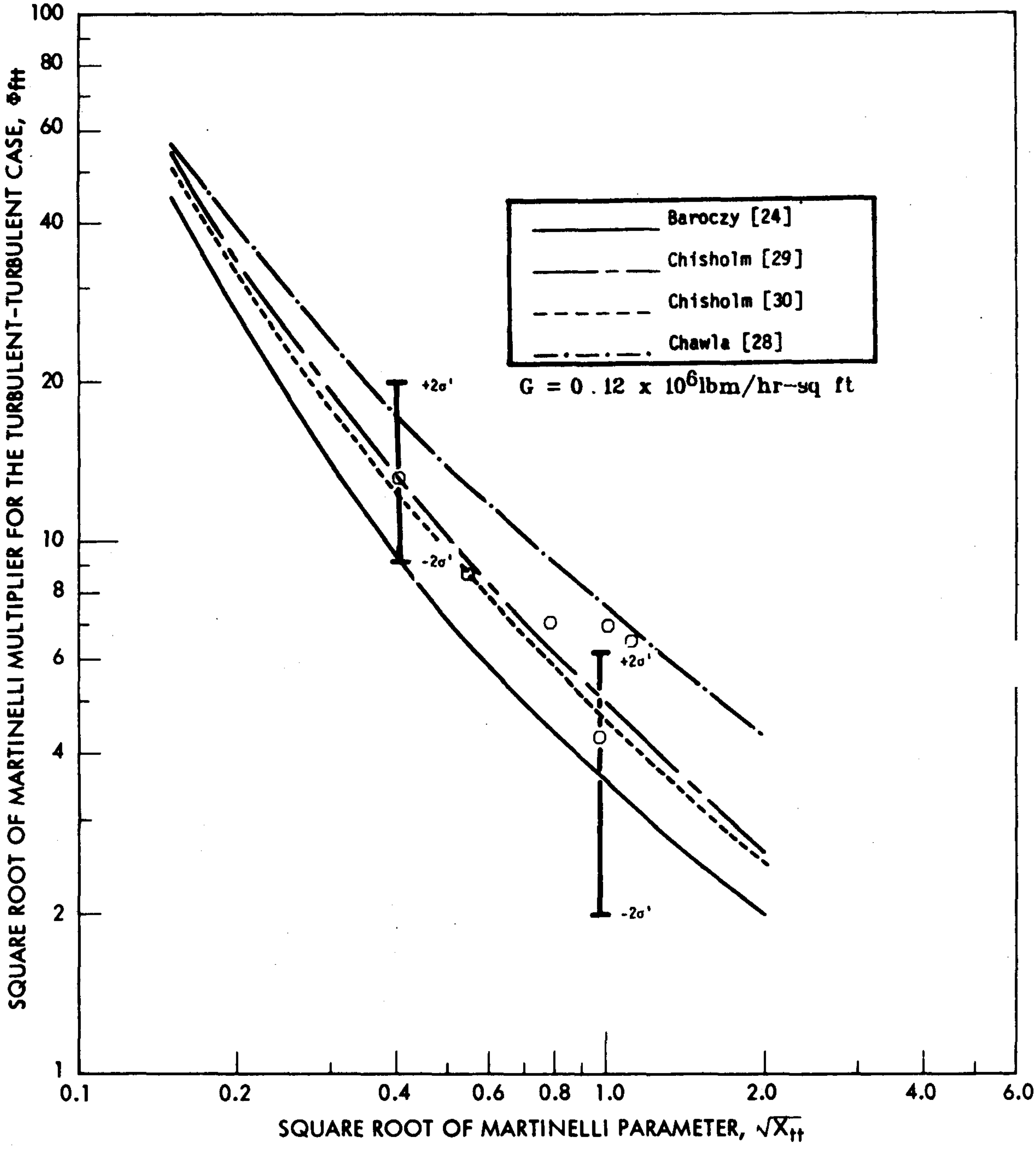

FIGURE IV-28. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F, 0.12 x 10^6 lbm/hr-sq ft.

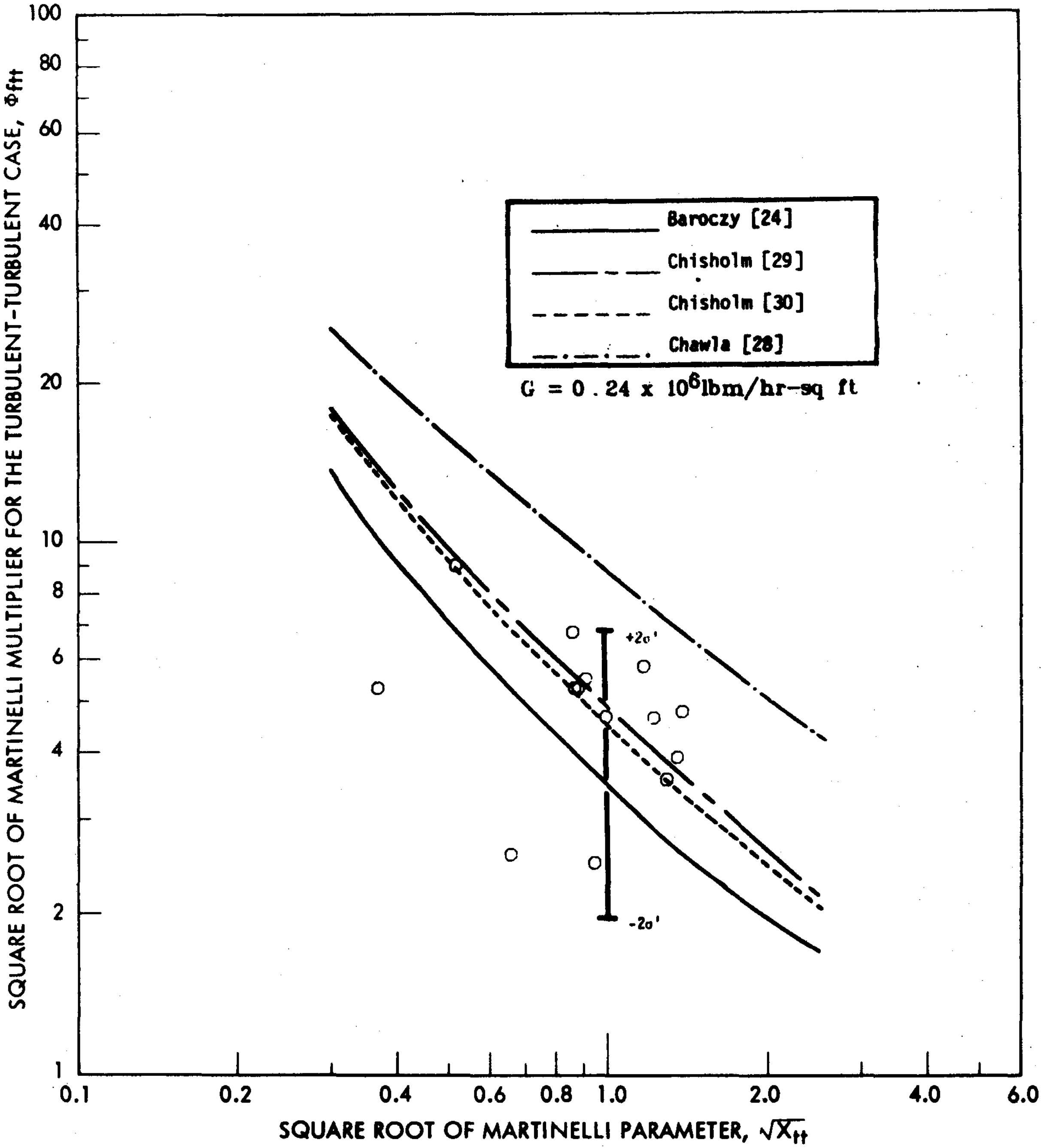

FIGURE IV-29. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F, 0.24 x 10^6 lbm/hr-sq ft.

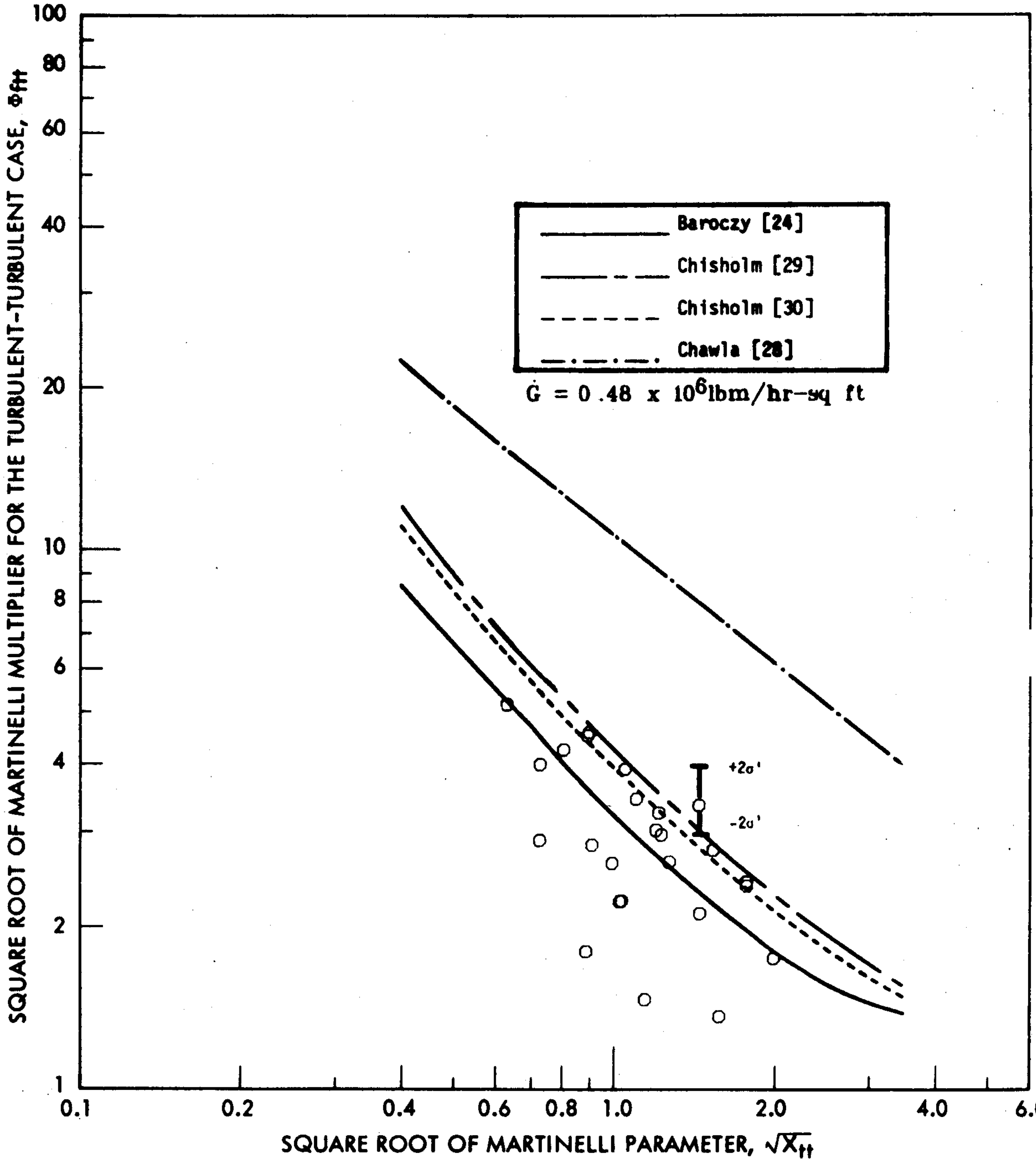

FIGURE IV-30. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F , 0.48 x 10^6 lbm/hr-sq ft.

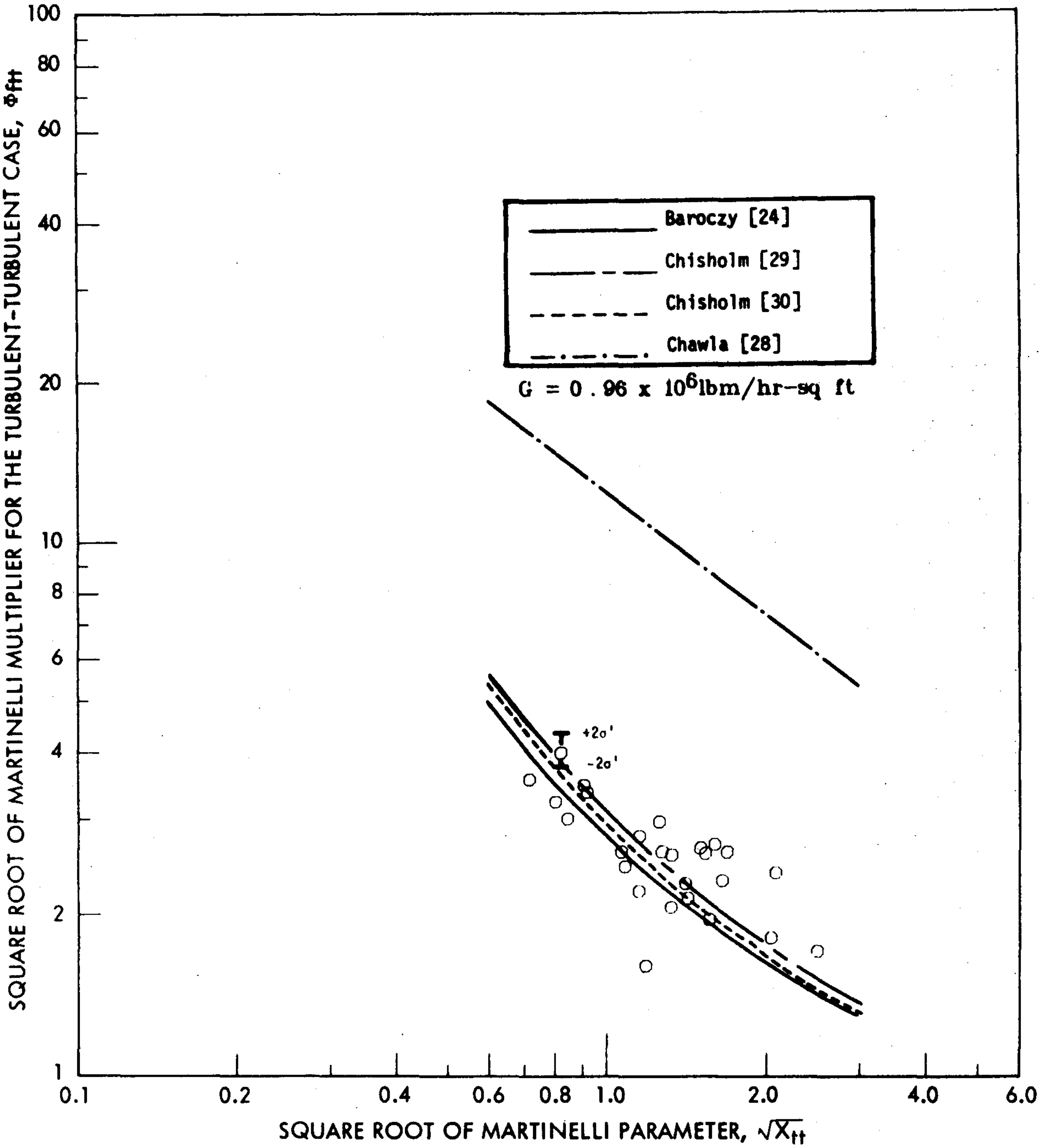

FIGURE IV-31. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F , 0.96 x 10^6 lbm/hr-sq ft.

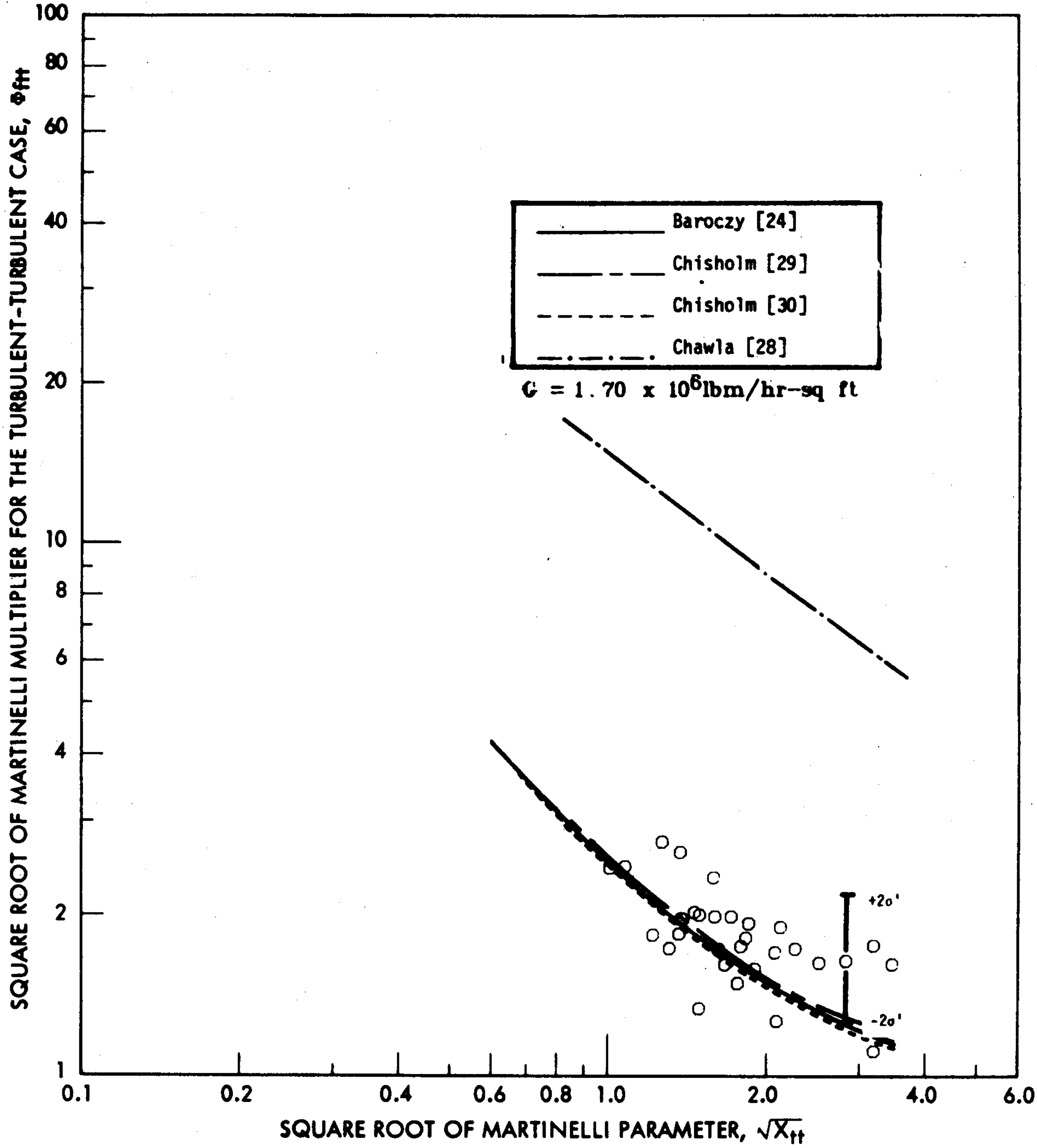

FIGURE IV-32. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F , 1.70 x 10^6 lbm/hr-sq ft.

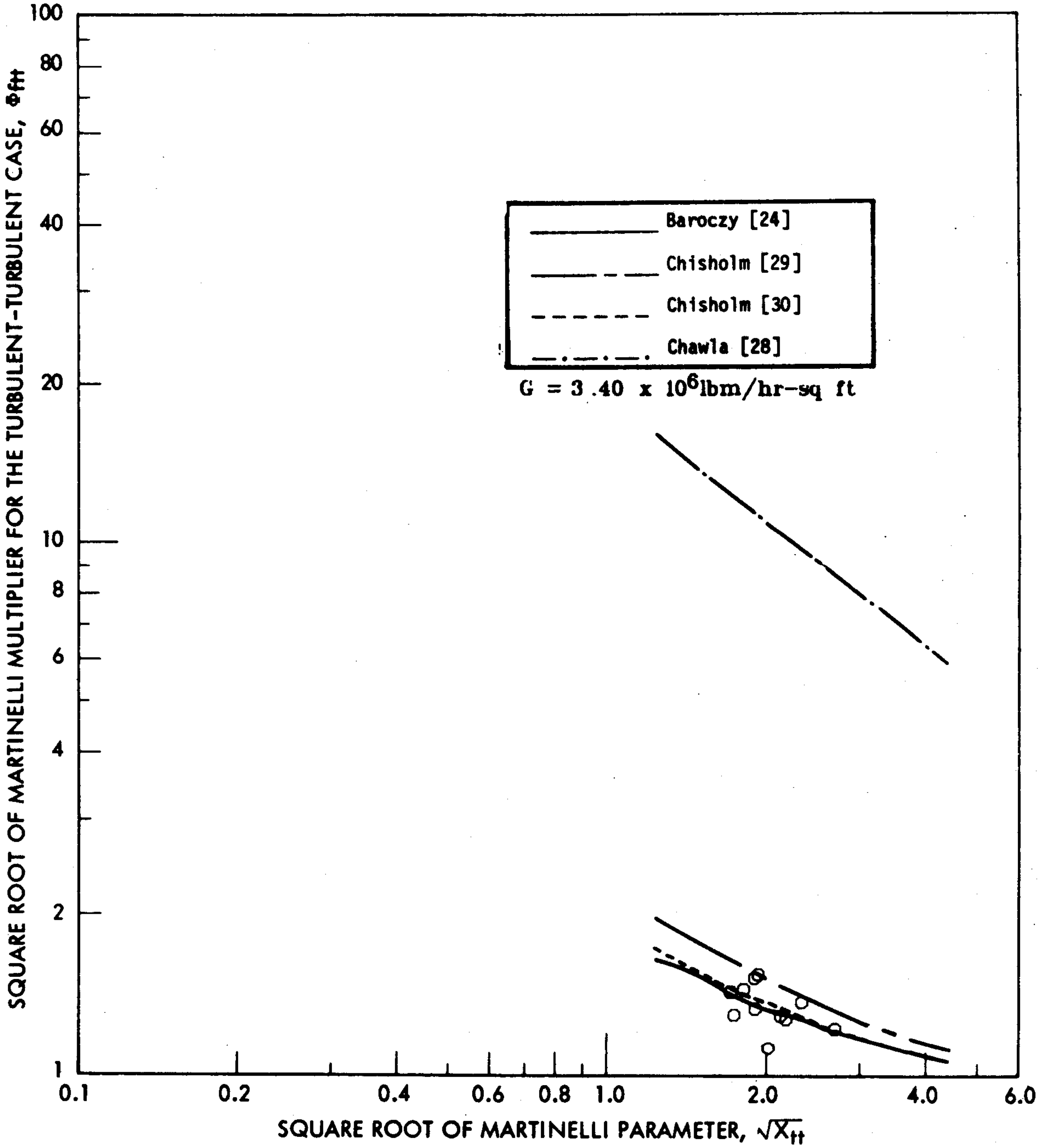

FIGURE IV-33. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 150°F, 3.40 x 10^6 lbm/hr-sq ft.

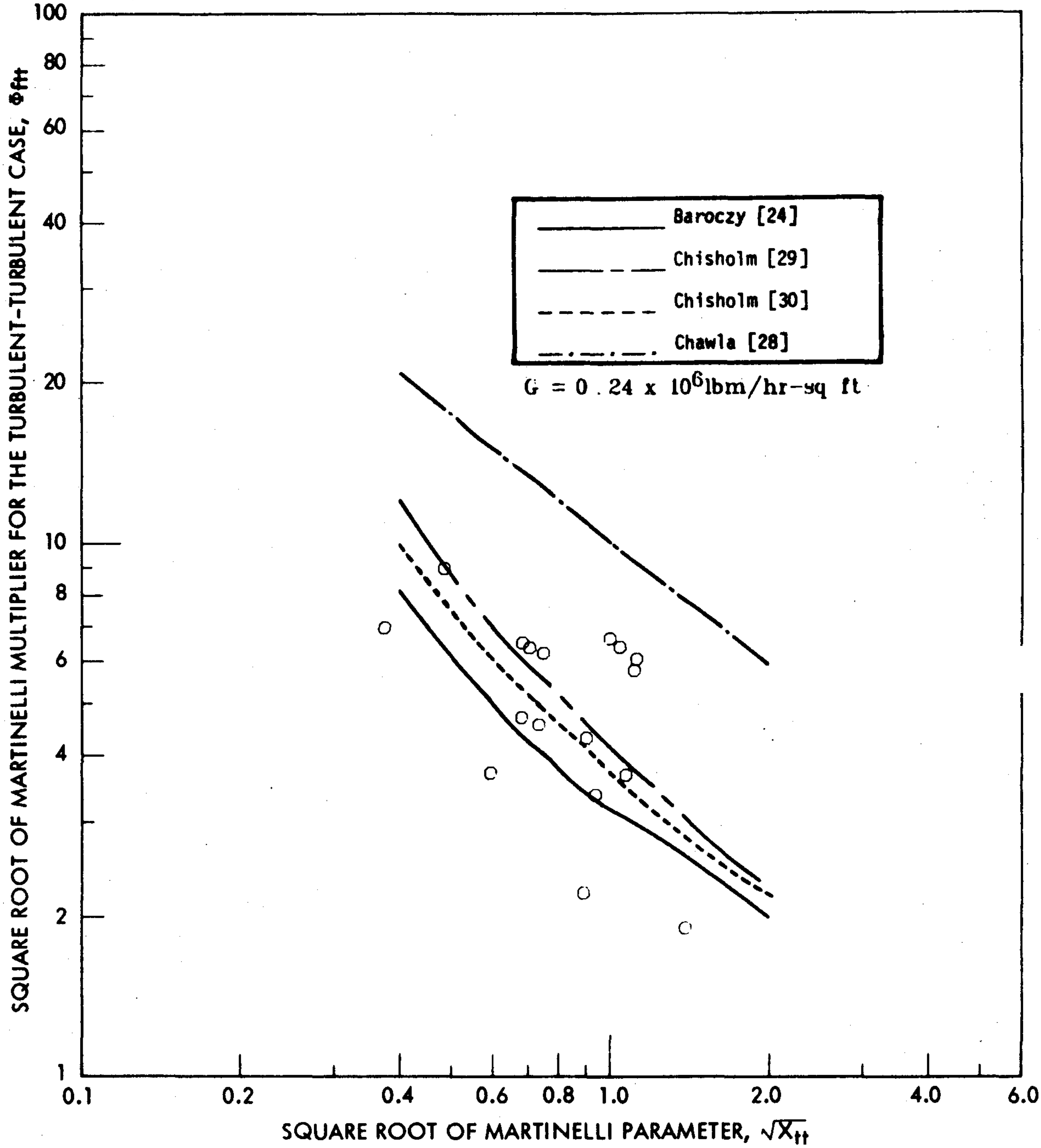

FIGURE IV-34. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 200°F , 0.24 x 10^6 lbm/hr-sq ft.

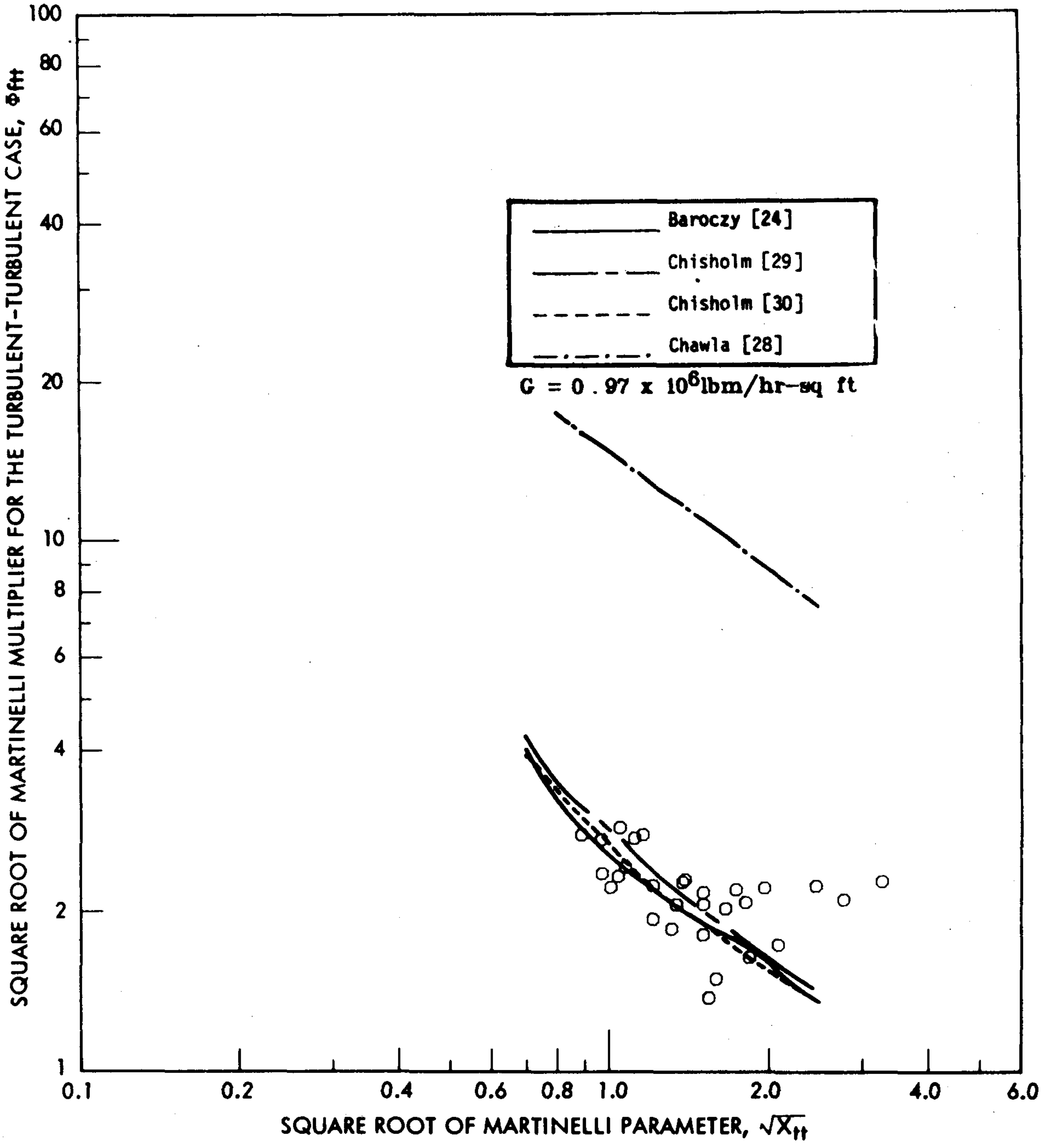

FIGURE IV-35. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 200°F , 0.97 x 10^6 lbm/hr-sq ft.

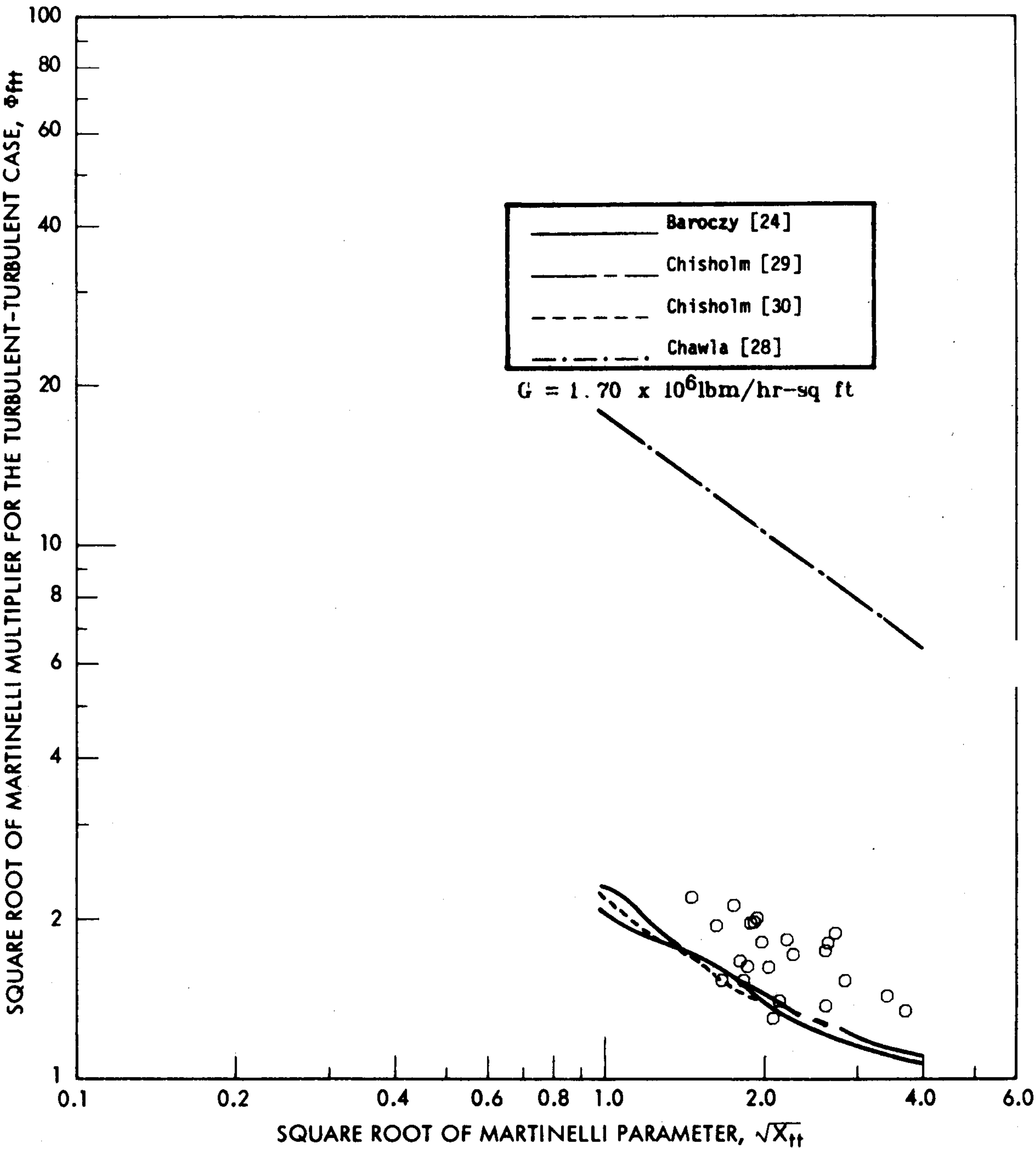

FIGURE IV-36. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 200°F , 1.70×10^6 lbm/hr-sq ft.

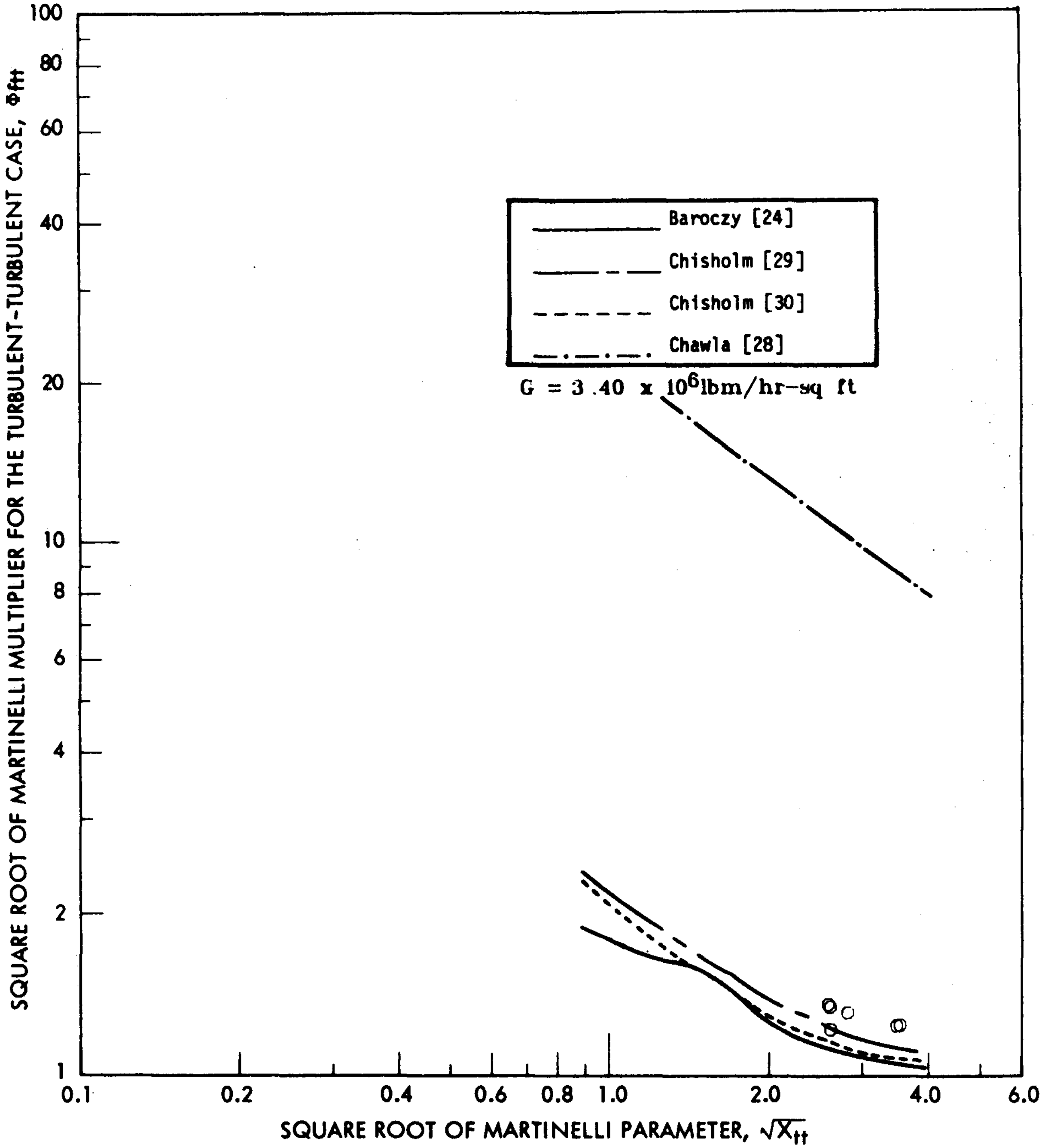

FIGURE IV-37. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 200°F , 3.40 x 10^6 lbm/hr-sq ft.

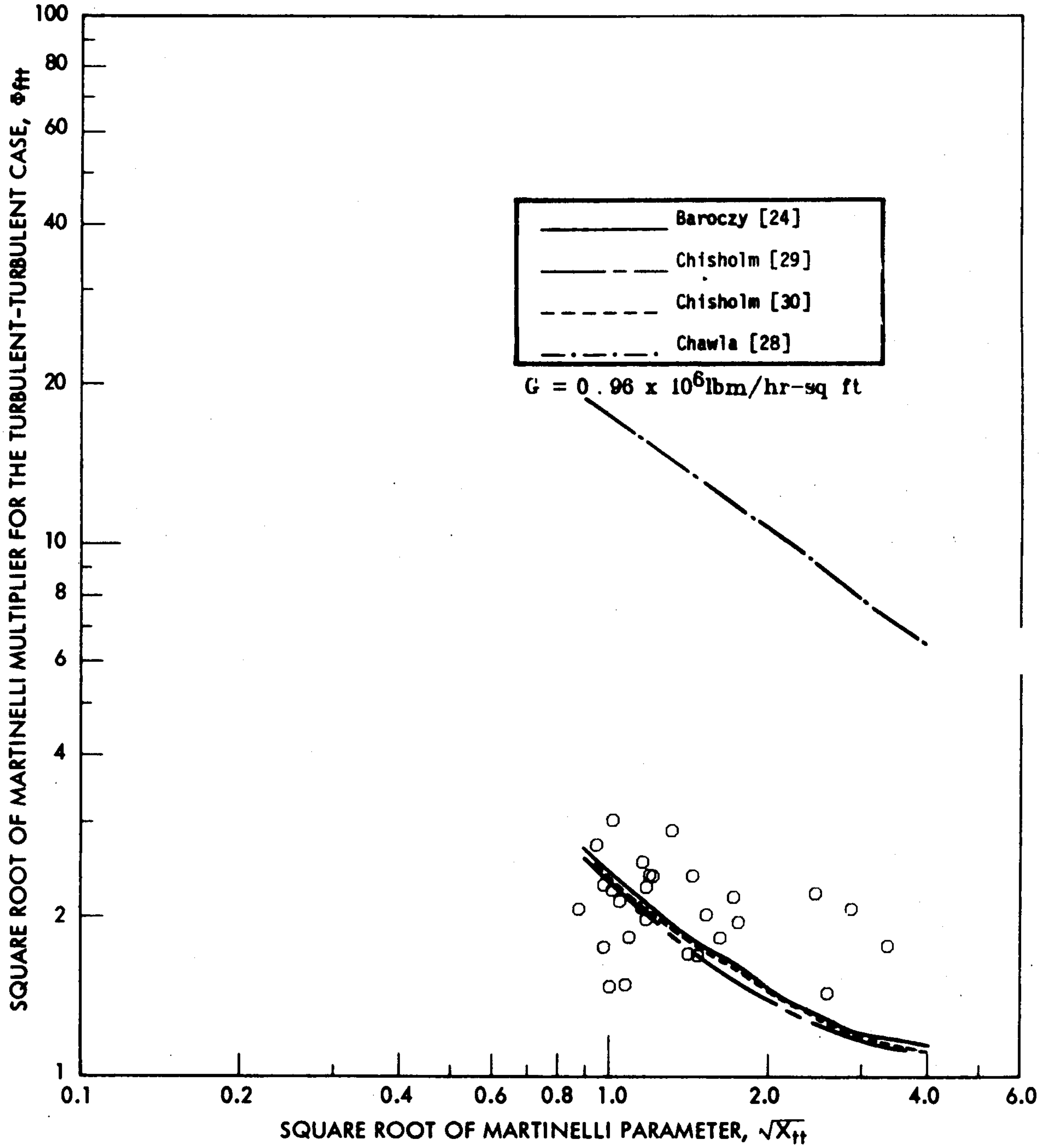

FIGURE IV-38. COMPARISON OF EXPERIMENTAL DATA WITH OTHER MODELS, 250°F , 0.96 x 10^6 lbm/hr-sq ft.

easily adapted to computerization. They also tended to provide fair accuracy when compared to the middle-to-high mass flux data.

Chawla [28] Correlation. The agreement between this correlation and the experimental data was not at all satisfactory, as the figures show. The nearest thing to acceptable agreement occurred at a boiling temperature of 100°F. This can most likely be explained by noting that Chawla [28] obtained his data at low temperatures ($\sim$11 to 20°F).

To summarize the results of the pressure drop studies, it can be stated that four of the correlations considered showed promise. First, the M-N [1] correlation gave fair-to-good results for all combinations of mass flux and saturation temperature. Even though the data did exhibit considerable scatter--partly due to experimental error, partly due to instrument inaccuracies and partly due to the nature of some of the flow regimes--the M-N [1] correlation faired remarkably well. One must also keep in mind that this correlation was developed from steam-water data. Thus, the fact that it even gave ballpark results when applied to R-114 should be somewhat gratifying. Second, since the experimental data exhibited a mass flux effect, it was possible to achieve slightly better results by utilizing three of the correlations which account for this phenomenon. Either of the two correlations proposed by Chisholm [29, 30] provided fair agreement over the range of mass flux considered. In addition, the correlation of Baroczy [24] provided fair results when compared with the medium-to-high mass flux data, however, Baroczy's [24] mass flux correction factor would be very difficult to computerize. Thus, either of the two correlations proposed by Chisholm [29, 30] should be considered superior to the other correlations. Some of the correlated results are listed in Appendix C.

C. HEAT TRANSFER COEFFICIENTS

Introduction

Over 1,200 experimental data points representative of two-phase heat transfer coefficients were obtained during this effort. A tabulation of these data can be found in Appendix C. Data were obtained during operation in both the natural and forced circulation modes. Since results were independent of the mode of operation, all heat transfer coefficient data were considered together. The discussion which follows will be limited to the subcritical heat flux region. Average heat transfer coefficient data for the entire 10-ft test section were included in many of the comparisons. Detailed study indicated that the results would not have been altered by omitting these data.

Results

The experimental heat transfer coefficients obtained during this investigation were compared to various published correlations. Such comparisions generally provide the most meaningful results, since trends are easier to identify and the effect of individual parameters is more readily determined. All graphs were constructed by computer.

Comparison of Experimental Results With Various Correlations

Riedle, Purcupile, and Schmidt's [58] Correlation. The results obtained from applying this correlation are shown in Figures IV-39 to IV-42. At a boiling temperature of 100°F, the great majority of the experimental data were from 0 to 50% above the calculated heat transfer coefficients. This deviation can be seen to get progressively worse as boiling temperature increases. By virtue of its simplicity, the correlation does not account for many parameters which are known to influence two-phase heat transfer. This is probably its most serious shortcoming.

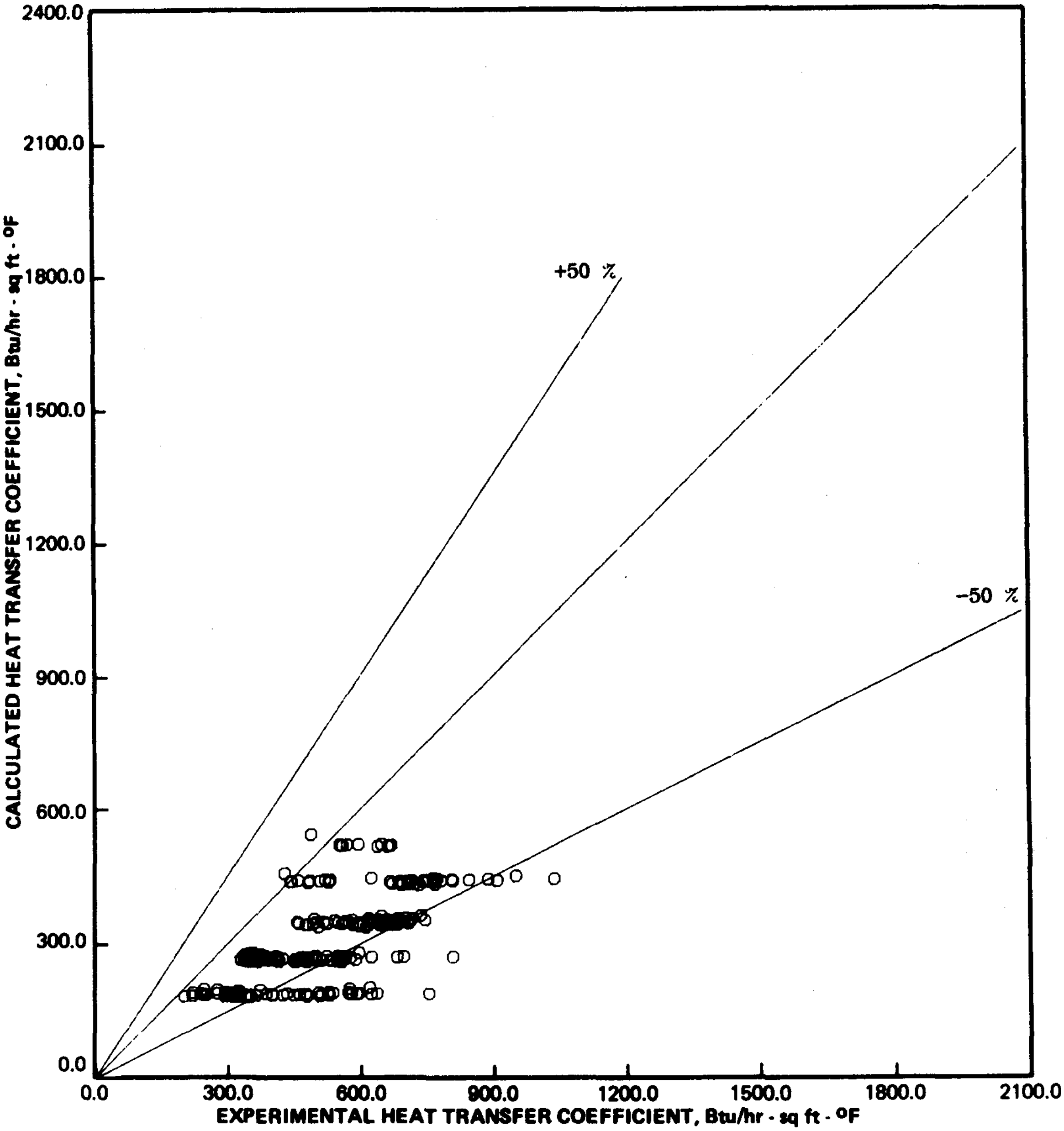

FIGURE IV-39. COMPARISON OF EXPERIMENTAL DATA WITH RIEDLE, PURCUPILE, AND SCHMIDT [58] CORRELATION, 100°F.

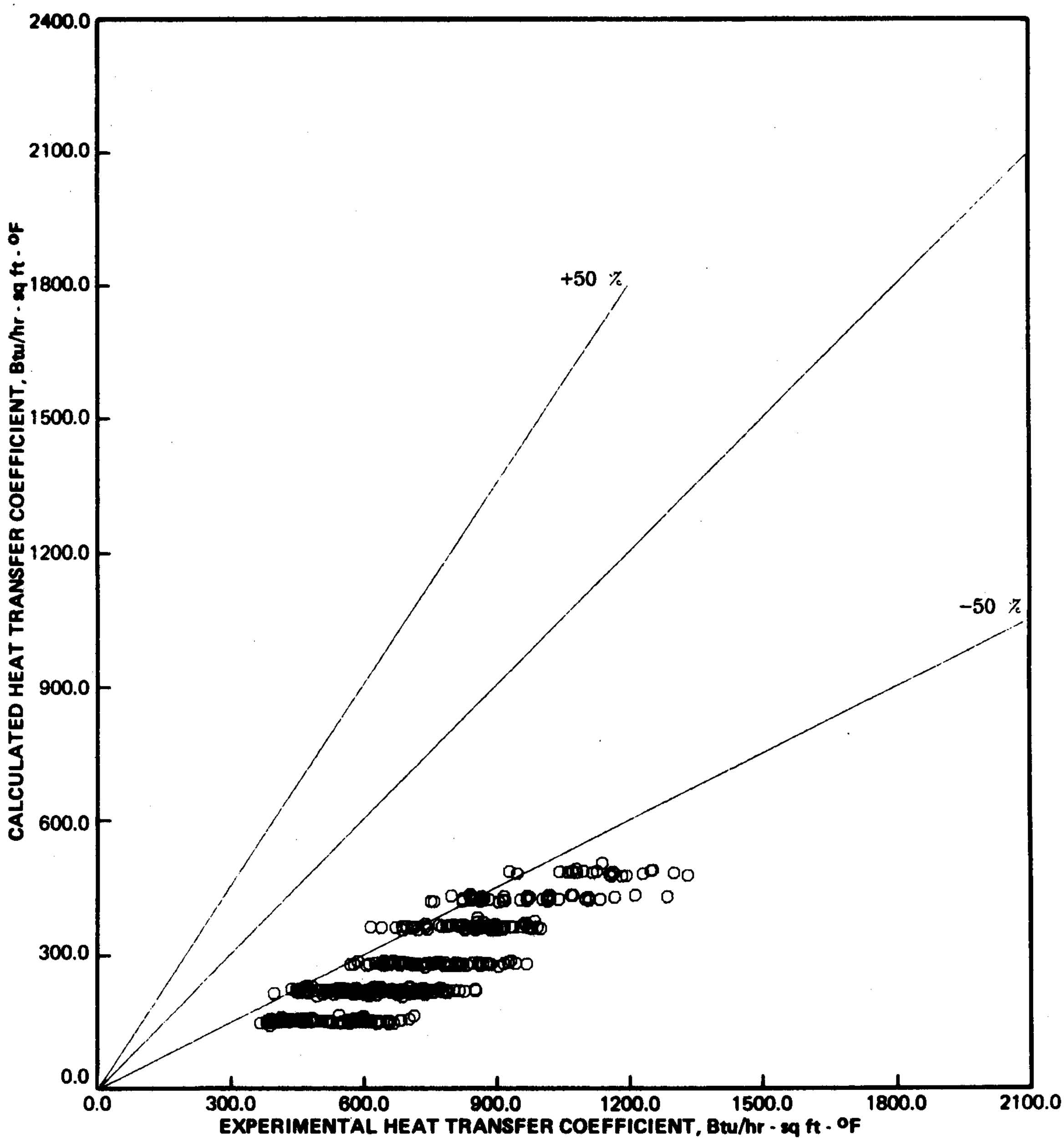

FIGURE IV-40. COMPARISON OF EXPERIMENTAL DATA WITH RIEDLE, PURCUPILE, AND SCHMIDT [58] CORRELATION, 150°F.

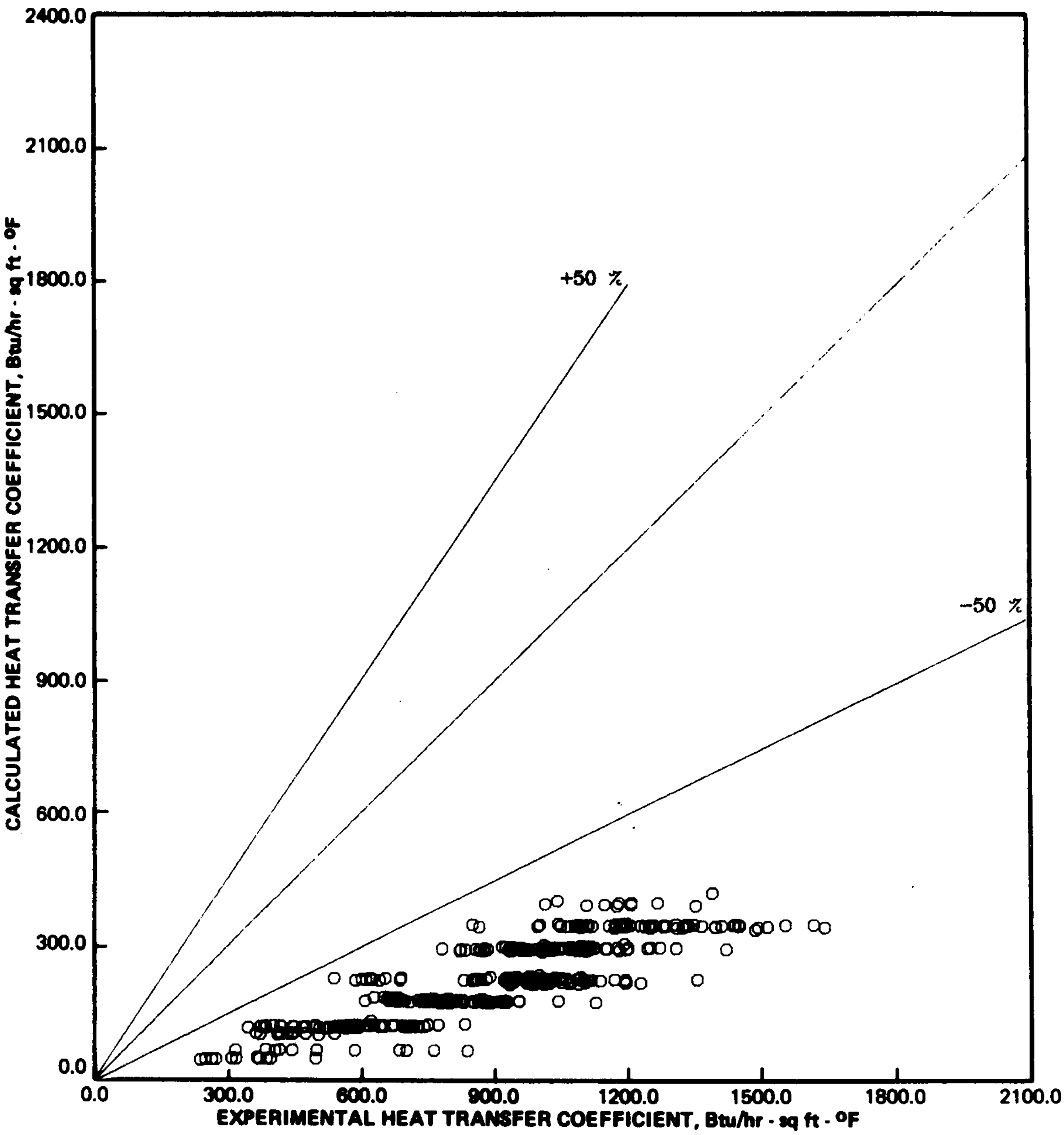

FIGURE IV-41. COMPARISON OF EXPERIMENTAL DATA WITH RIEDLE, PURCUPILE, AND SCHMIDT [58] CORRELATION, 200°F.

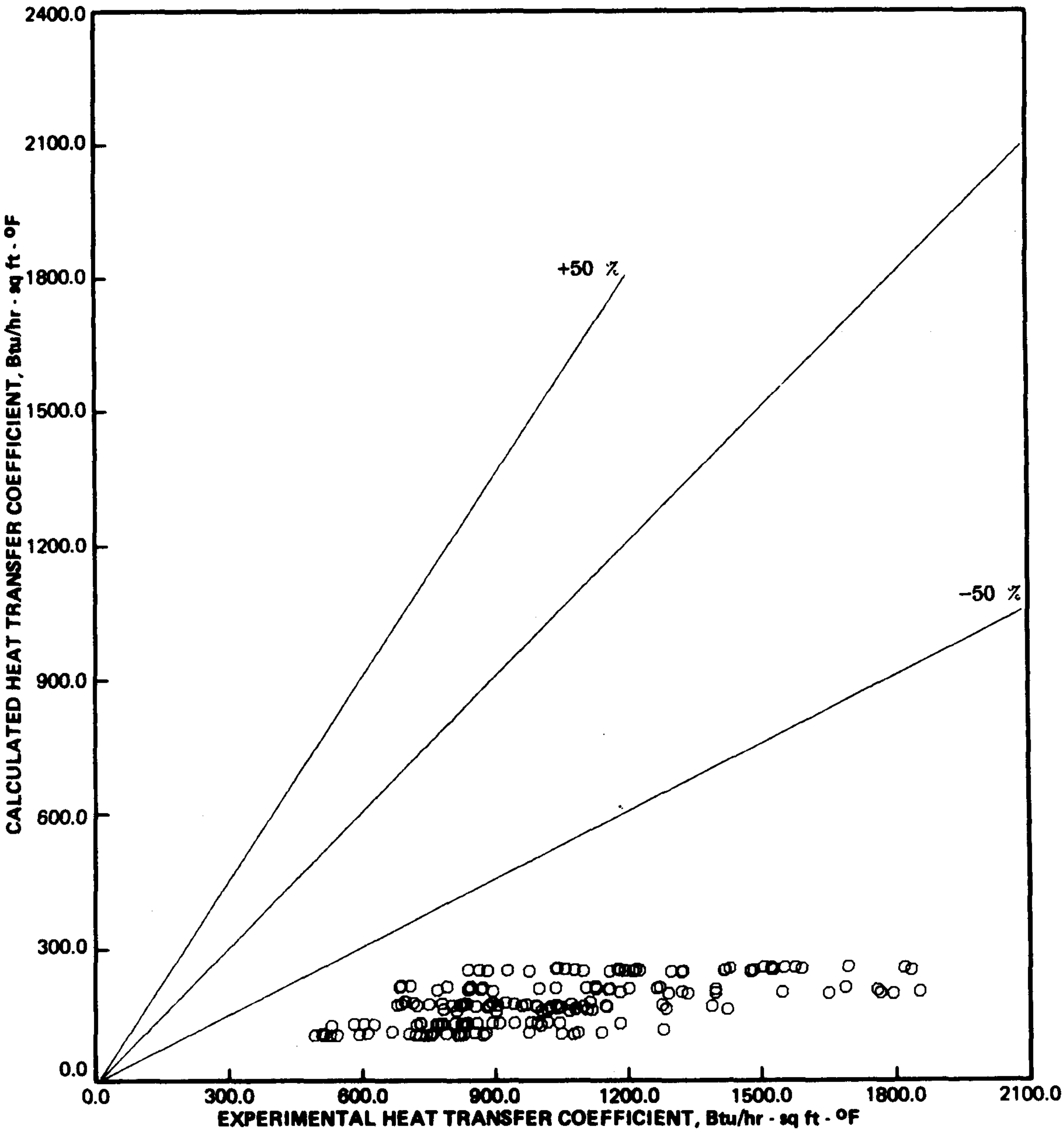

FIGURE IV-42. COMPARISON OF EXPERIMENTAL DATA WITH RIEDLE, PURCUPILE, AND SCHMIDT [58] CORRELATION, 250°F.

Davis and David [48] Correlations. These two correlations were not at all reliable, as Figures IV-43 to IV-50 indicate. The Homogeneous flow model had a tendency to overpredict the experimental data by as much as 300% in some cases. On the other hand, the great bulk of the experimental data were underpredicted by the separated-annular flow model. With both of these correlations, data scatter (an important consideration when judging the goodness of a correlation) was excessive.

Altman, Norris, and Staub [33] Correlation. The results obtained by utilizing this correlation are given in Figures IV-51 to IV-53. The bulk of the experimental data were 0 to 50% above the calculated coefficients when this correlation was applied. As the boiling temperature increased, the agreement between the measured and predicted quantities tended to worsen. A large portion of the experimental data (especially at 250°F) could not be compared with this correlation because the parameter, $\left(N_{Re_{fo}}^{2}\ K_F\right)^{0.5}$, was beyond the range of the correlation's applicability.

Shrock and Grossman [49] Correlation. The Shrock and Grossman [49] correlation tended to consistently underpredict the experimental data. The discrepancy was as much as 50% at a boiling temperature of 100°F and became progressively worse at the higher boiling temperatures. The results are shown in Figures IV-54 to IV-57.

Cheshire and Sterling [44] Correlation. When this correlation was applied to the data obtained during this investigation, poor agreement was obtained. This was not very suprising since other investigators, including Jallouk [42], have found that this correlation is applicable in only a very narrow range of operating conditions. For instance,

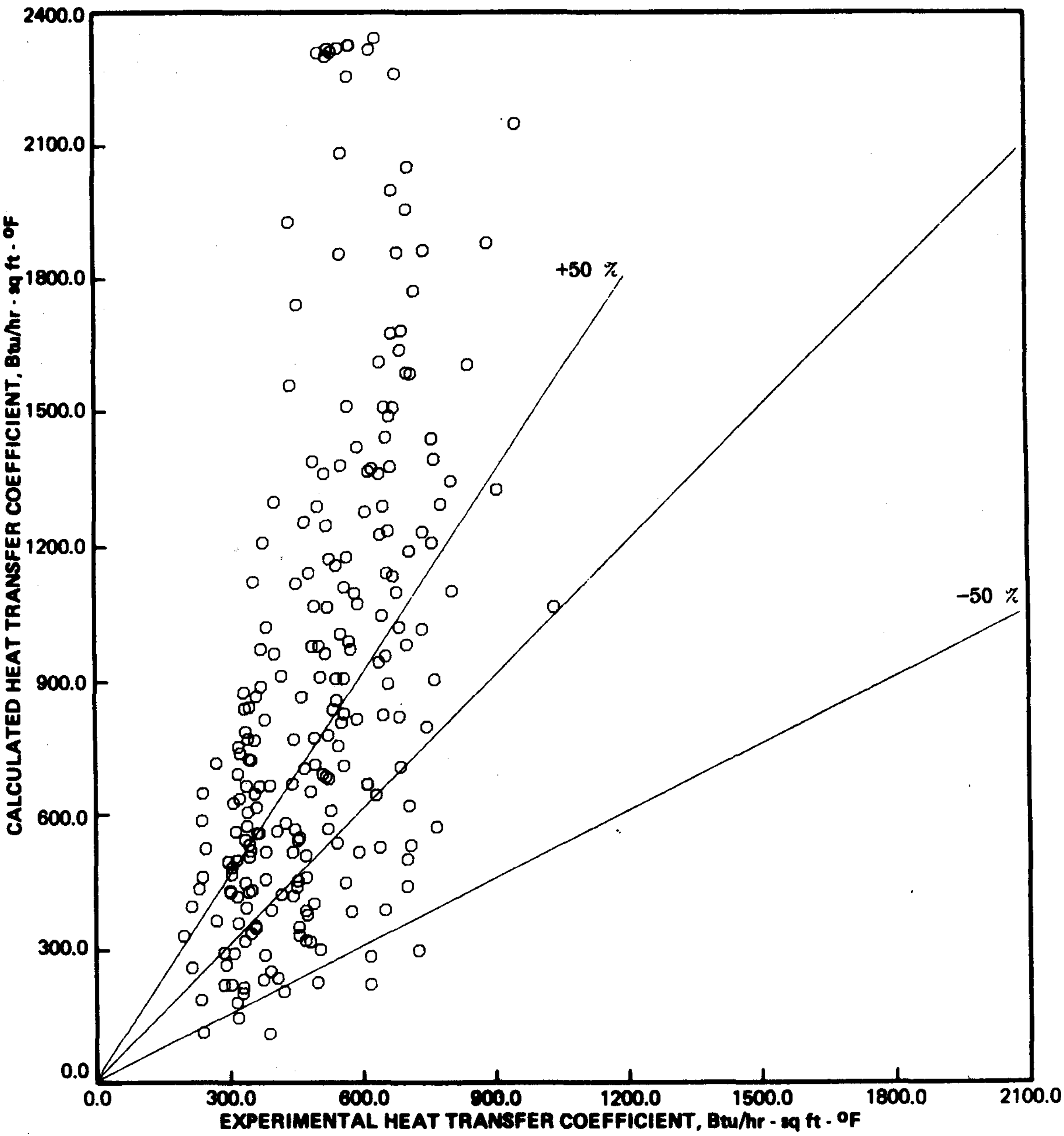

FIGURE IV-43. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] HOMOGENEOUS FLOW CORRELATION, 100°F.

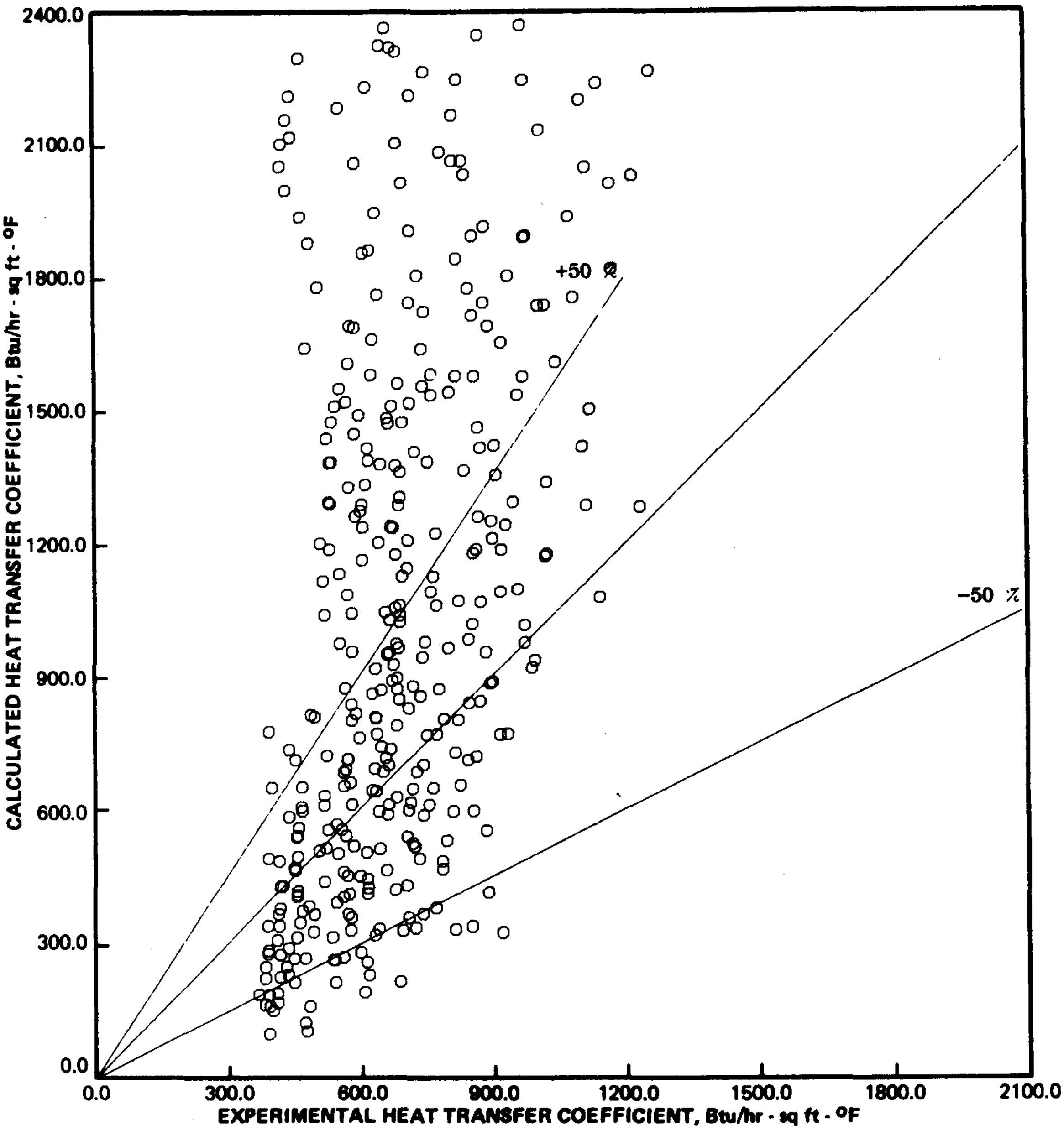

FIGURE IV-44. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] HOMOGENEOUS FLOW CORRELATION, 150°F.

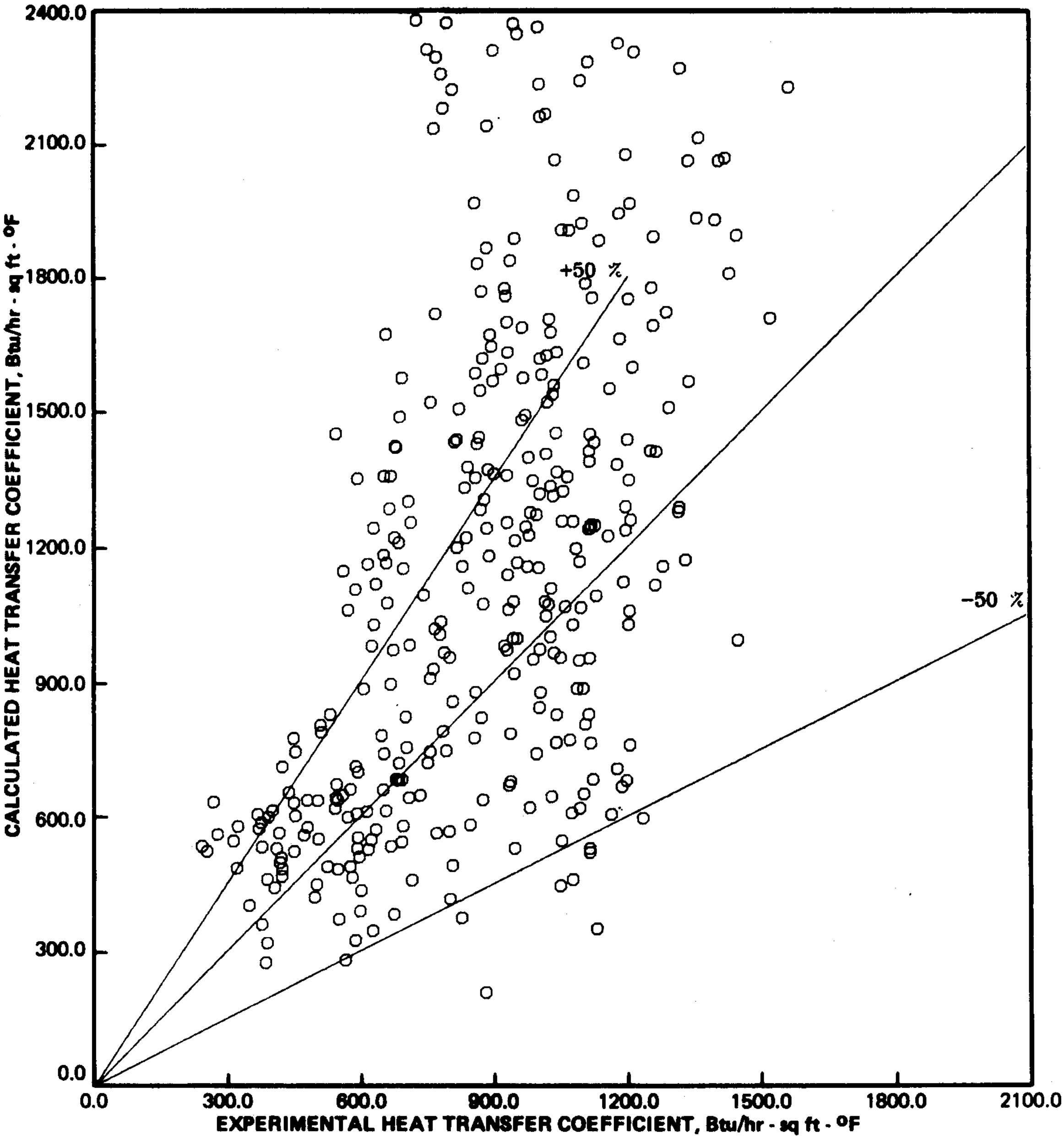

FIGURE IV-45. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] HOMOGENEOUS FLOW CORRELATION, 200°F.

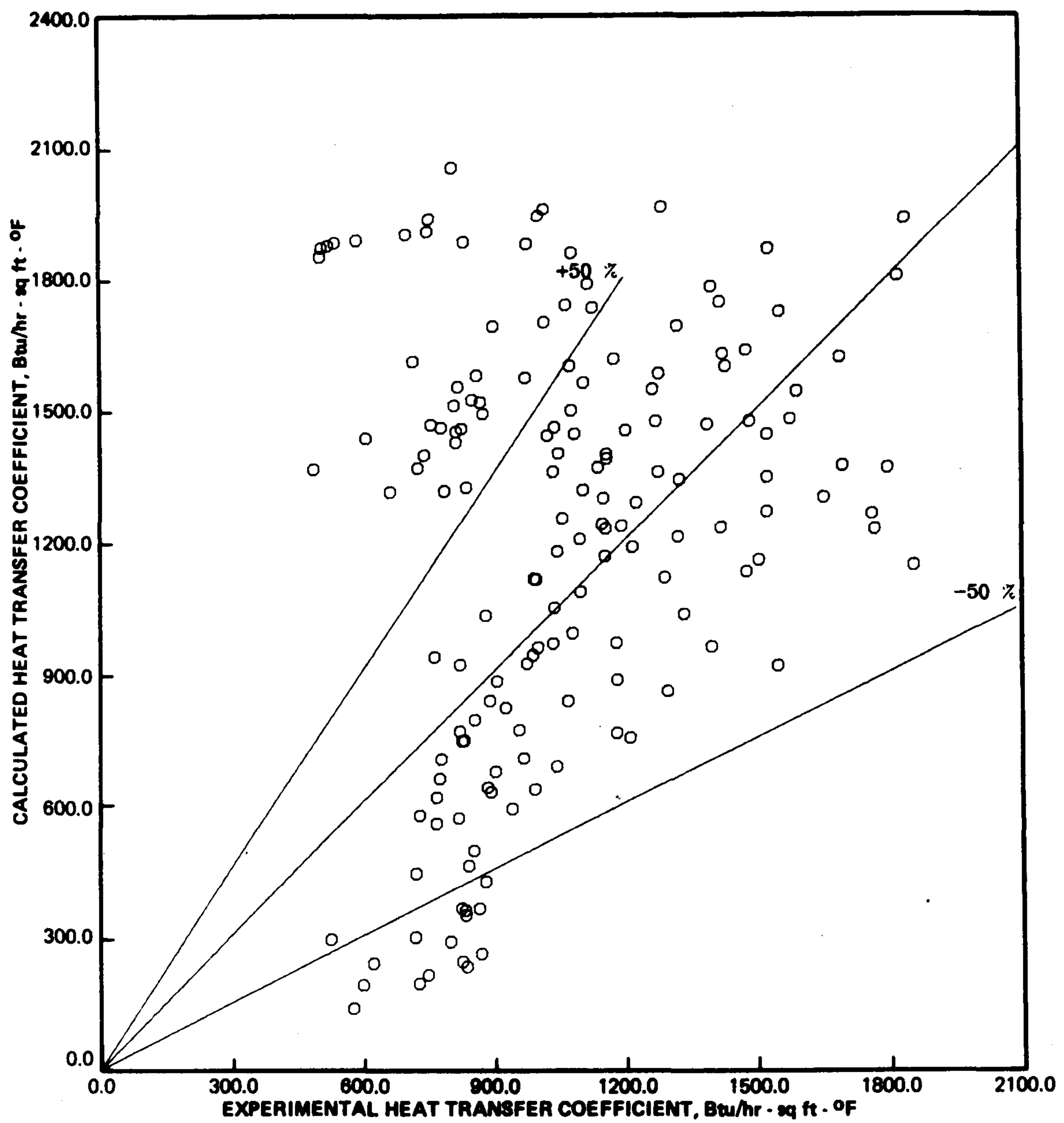

FIGURE IV-46. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] HOMOGENEOUS FLOW CORRELATION, 250°F.

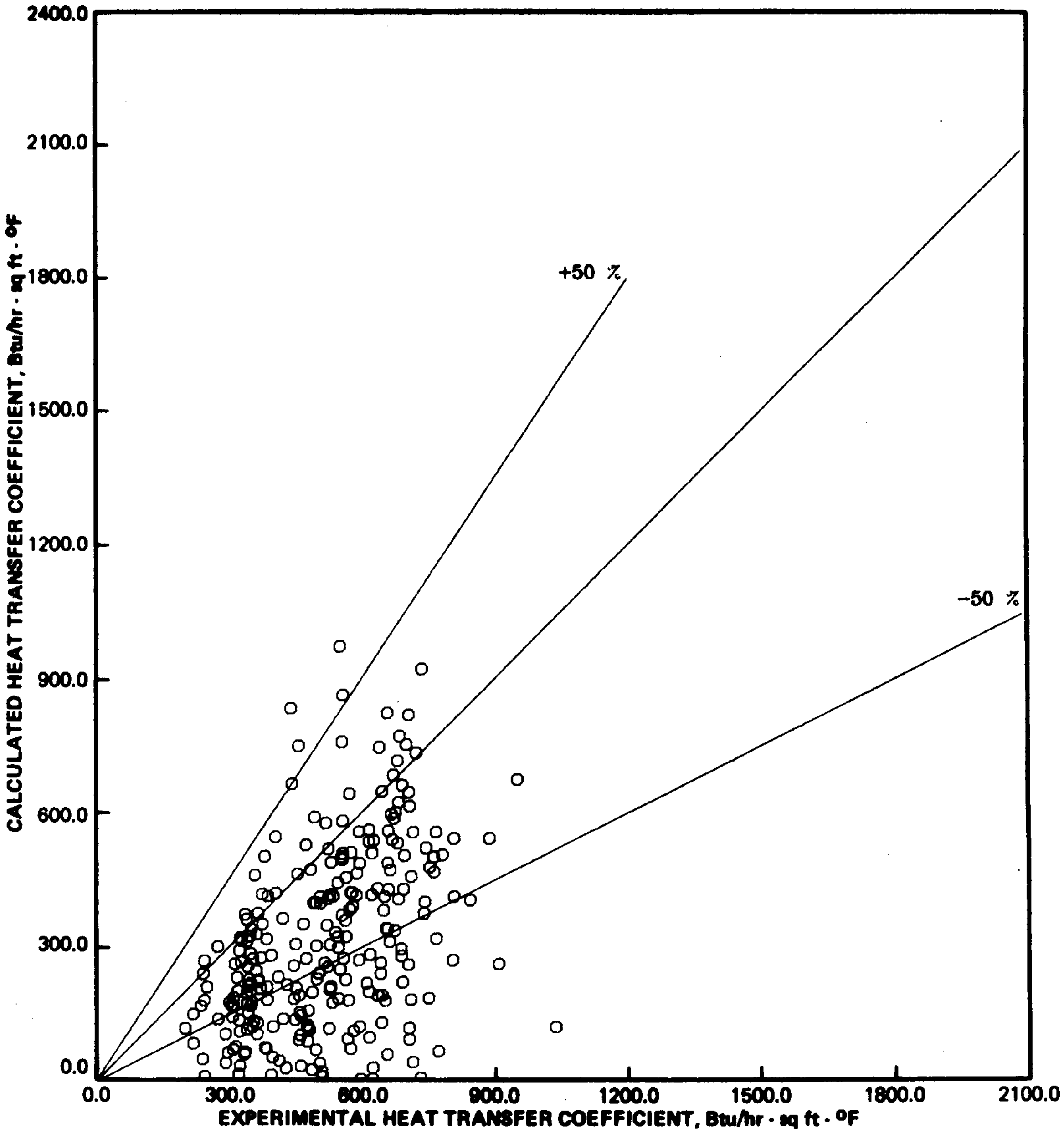

FIGURE IV-47. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] SEPARATED FLOW CORRELATION, 100°F.

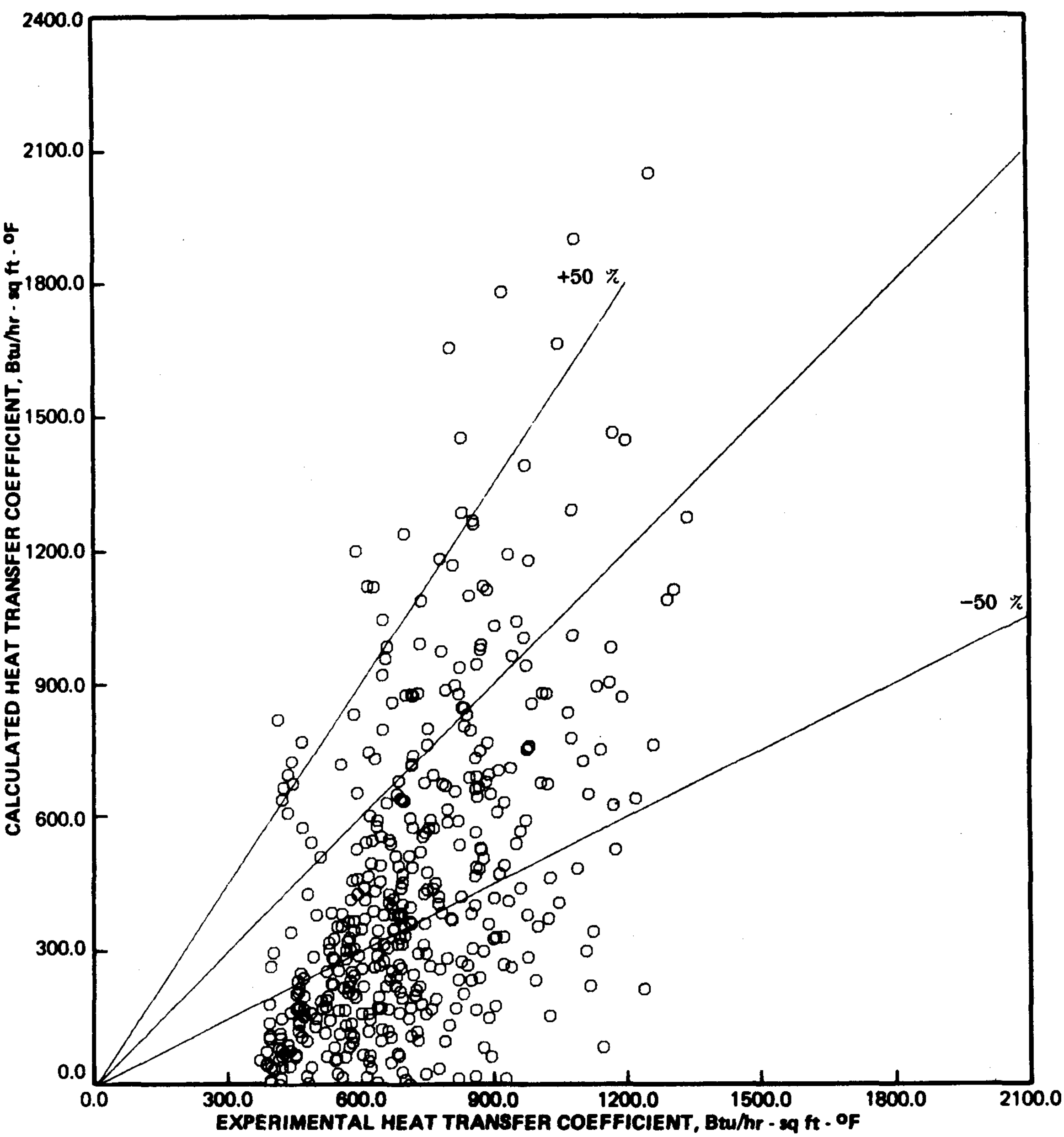

FIGURE IV-48. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] SEPARATED FLOW CORRELATION, 150°F.

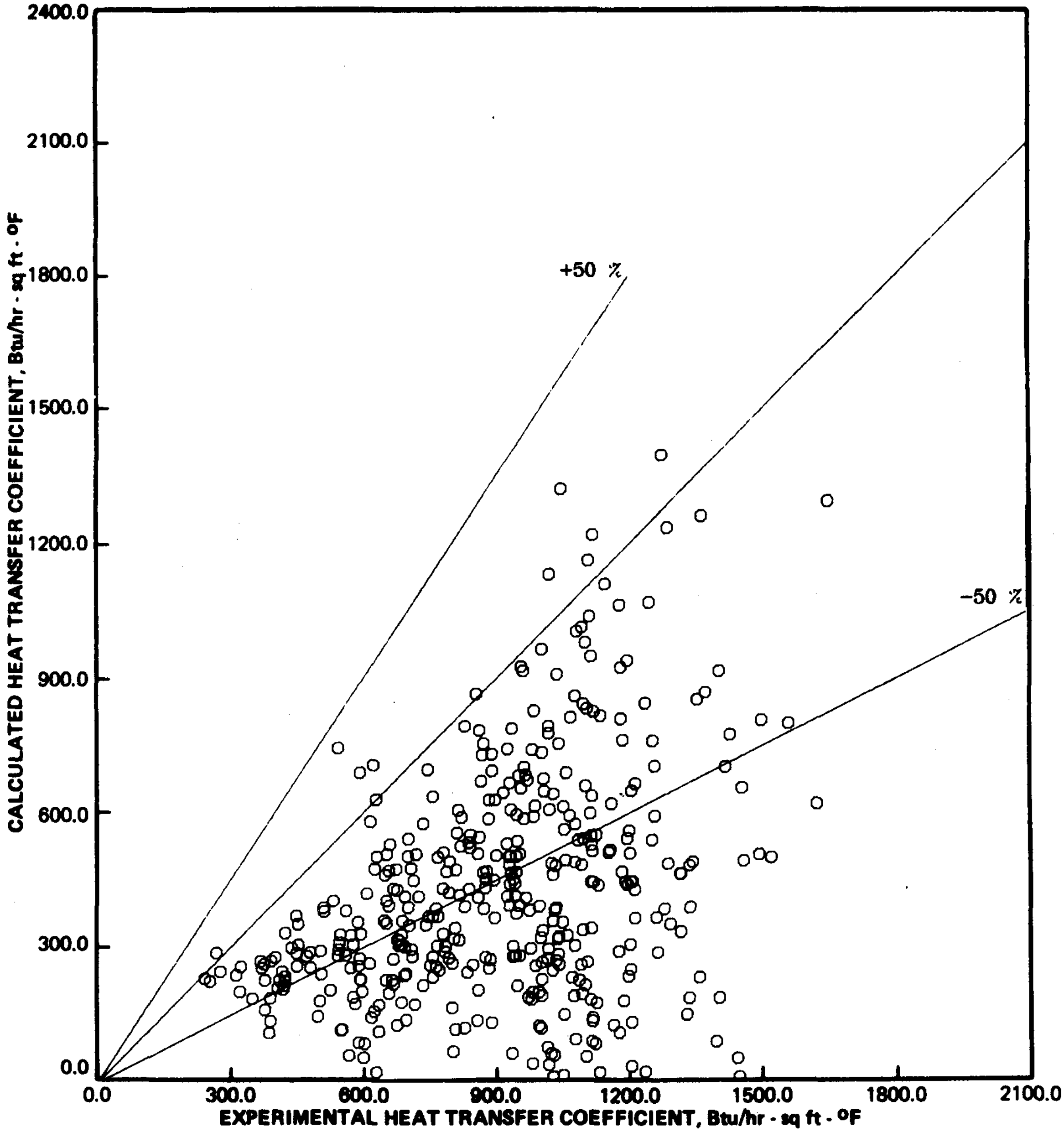

FIGURE IV-49. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] SEPARATED FLOW CORRELATION, 200°F.

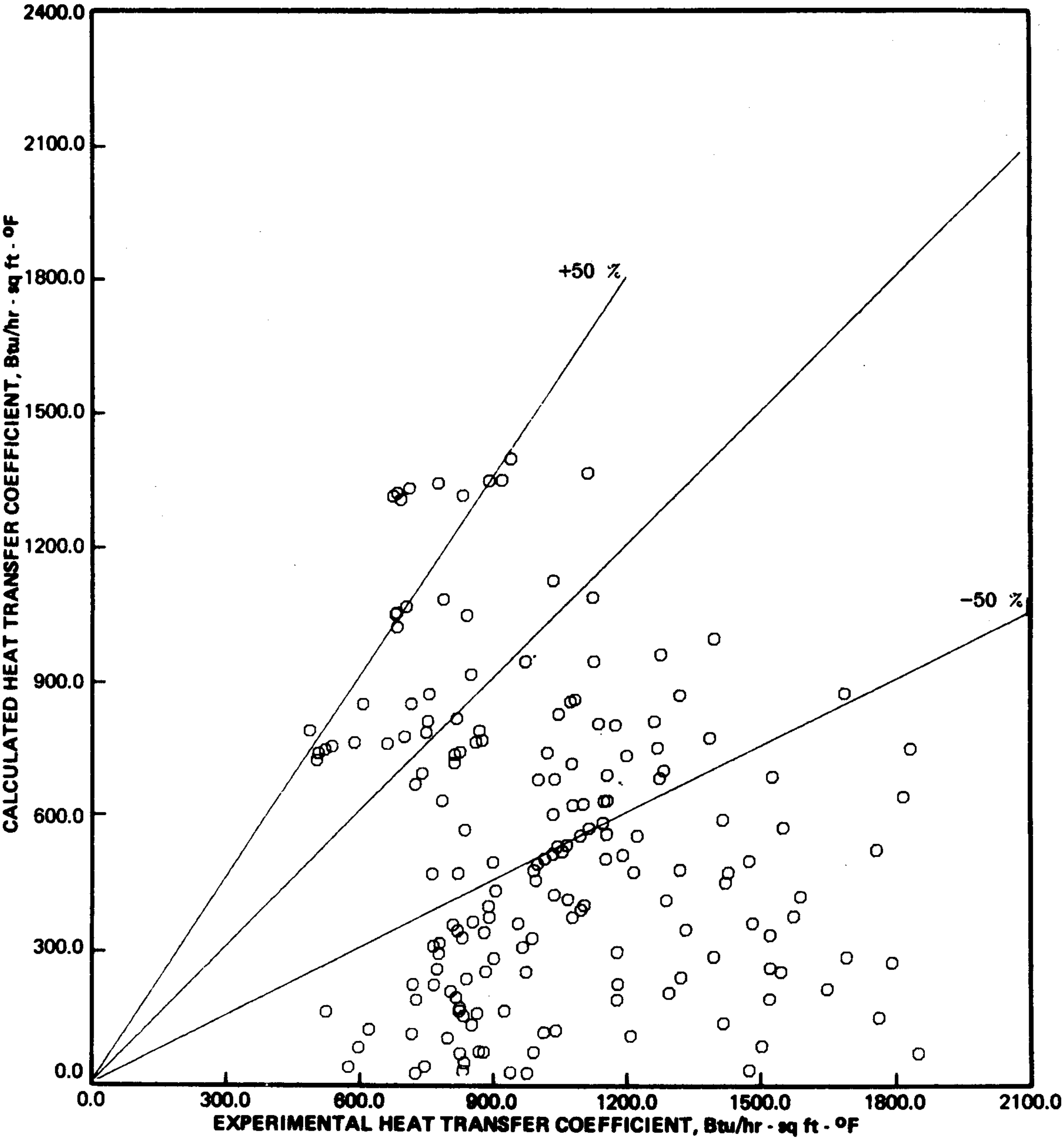

FIGURE IV-50. COMPARISON OF EXPERIMENTAL DATA WITH DAVIS AND DAVID [48] SEPARATED FLOW CORRELATION, 250°F.

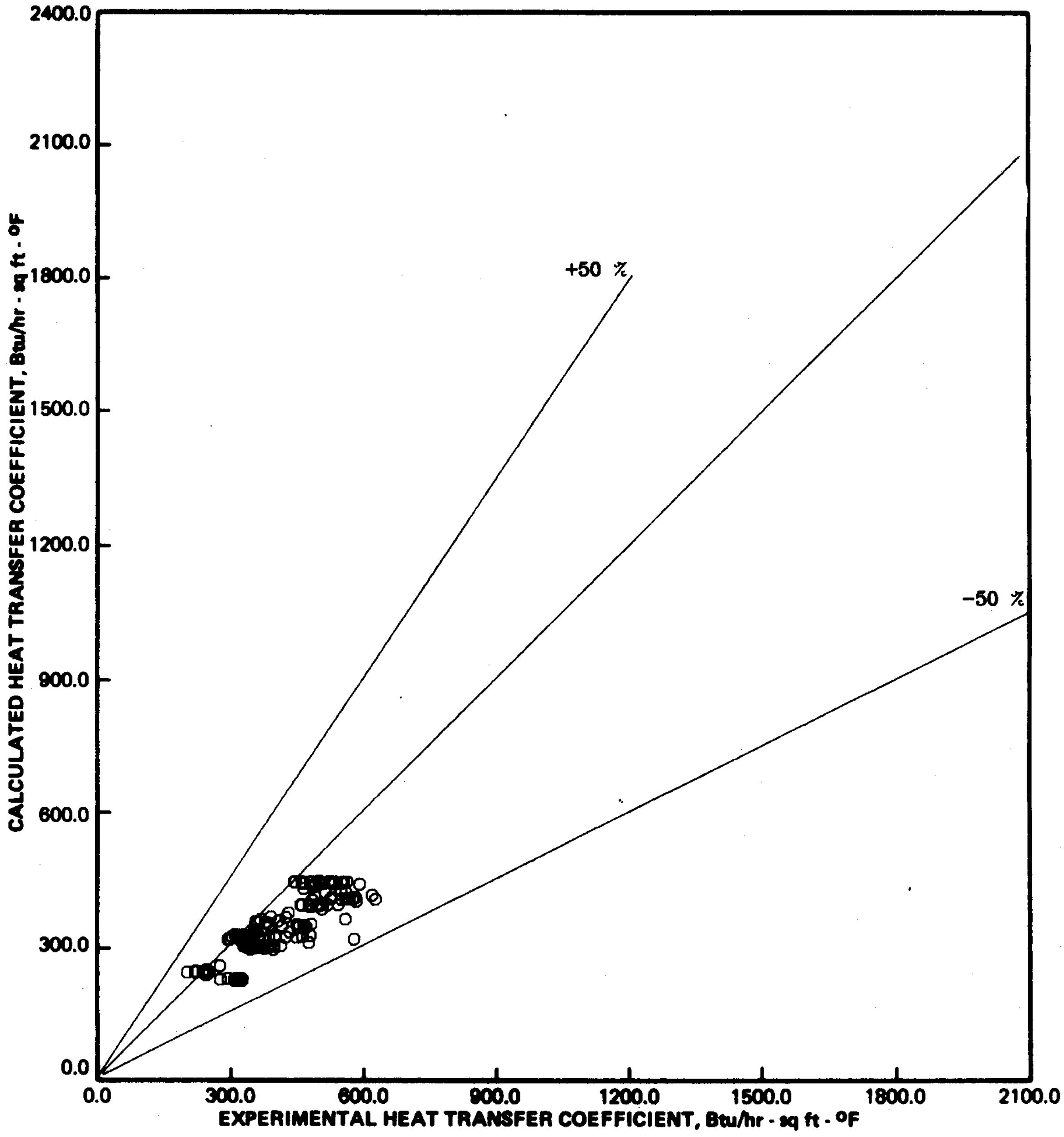

FIGURE IV-51. COMPARISON OF EXPERIMENTAL DATA WITH ALTMAN, NORRIS, AND STAUB [33] CORRELATION, 100°F.

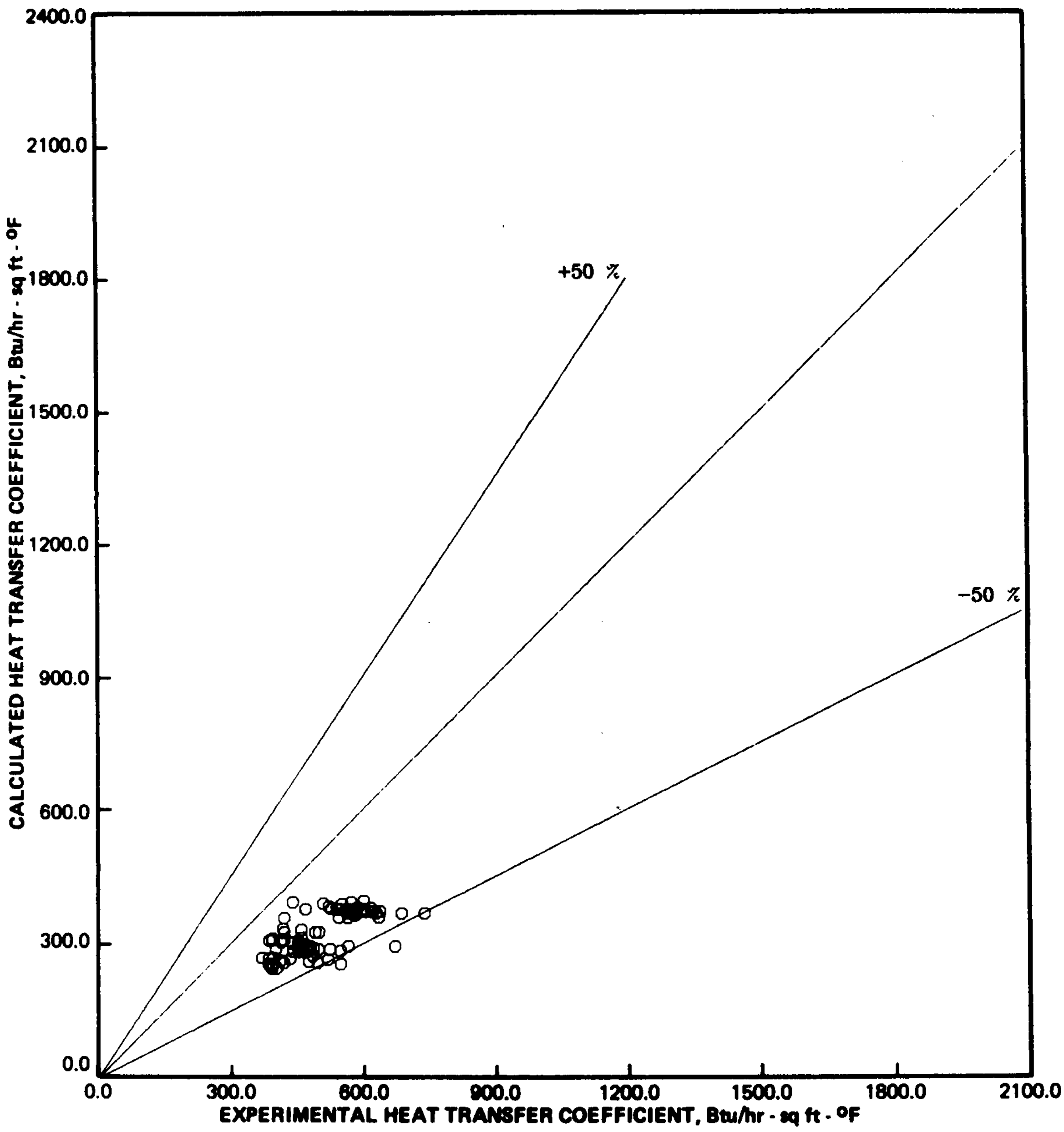

FIGURE IV-52. COMPARISON OF EXPERIMENTAL DATA WITH ALTMAN, NORRIS, AND STAUB [33] CORRELATION, 150°F.

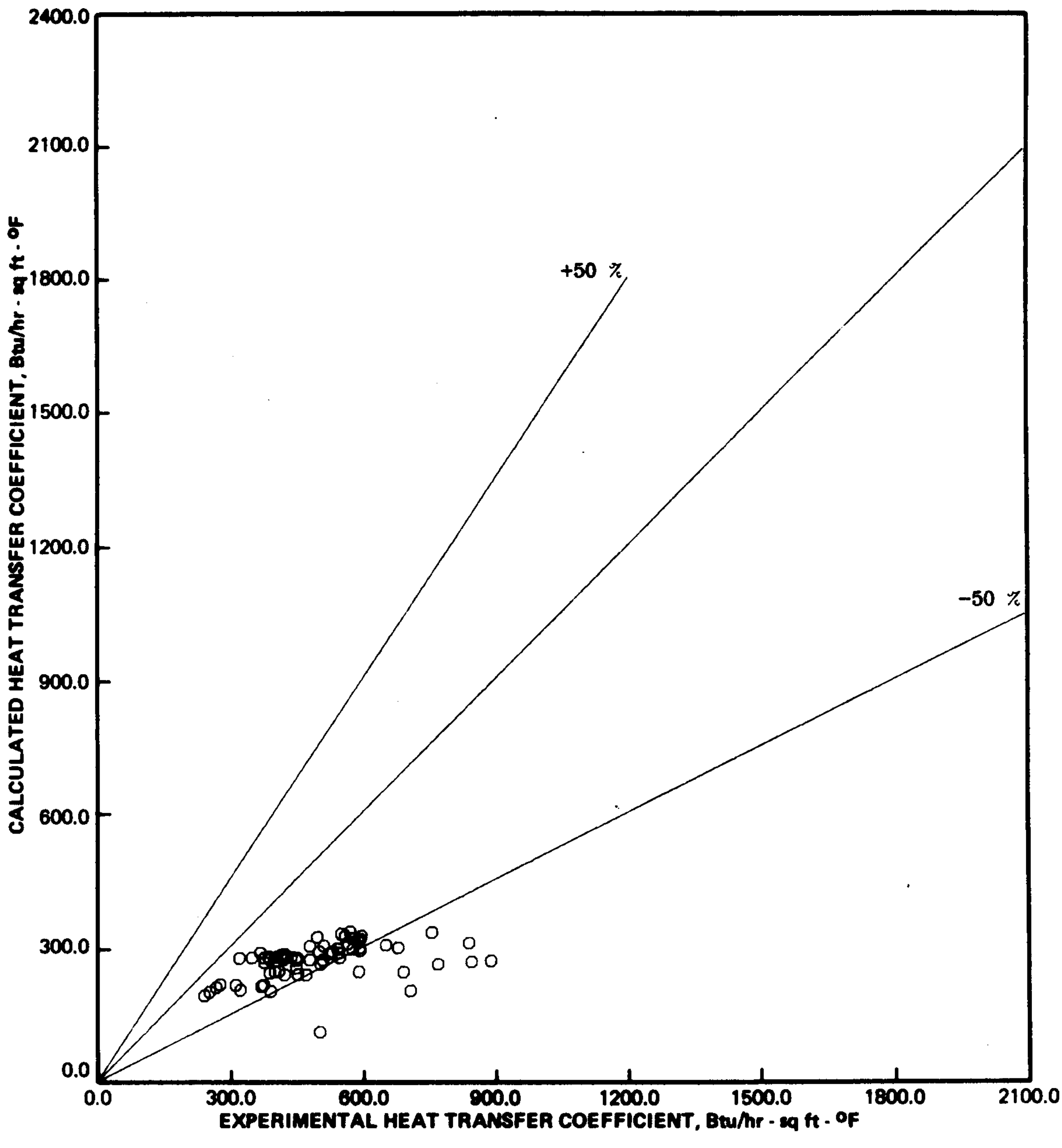

FIGURE IV-53. COMPARISON OF EXPERIMENTAL DATA WITH ALTMAN, NORRIS, AND STAUB [33] CORRELATION, 200°F.

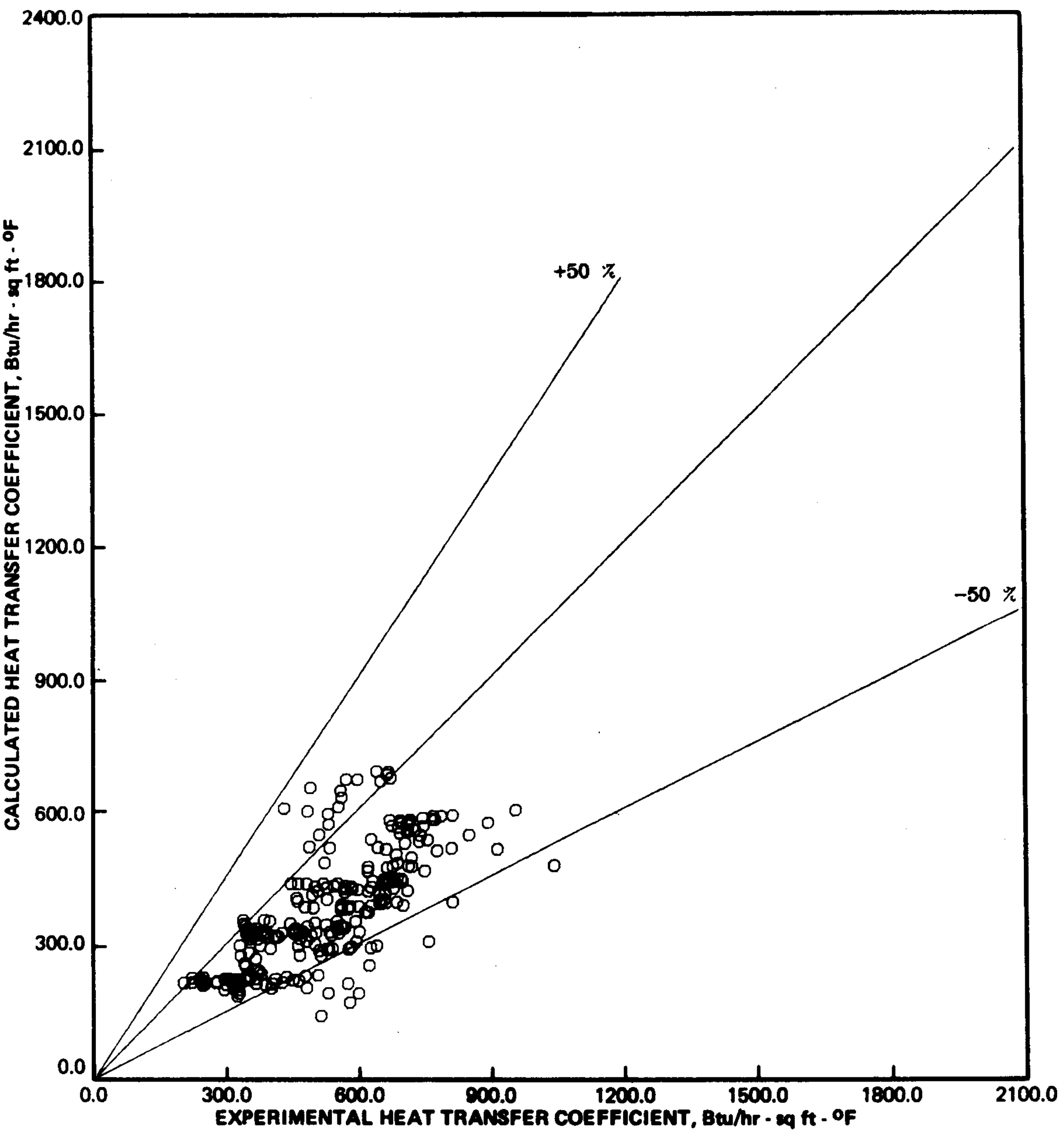

FIGURE IV-54. COMPARISON OF EXPERIMENTAL DATA WITH SCHROCK AND GROSSMAN [49] CORRELATION, 100°F.

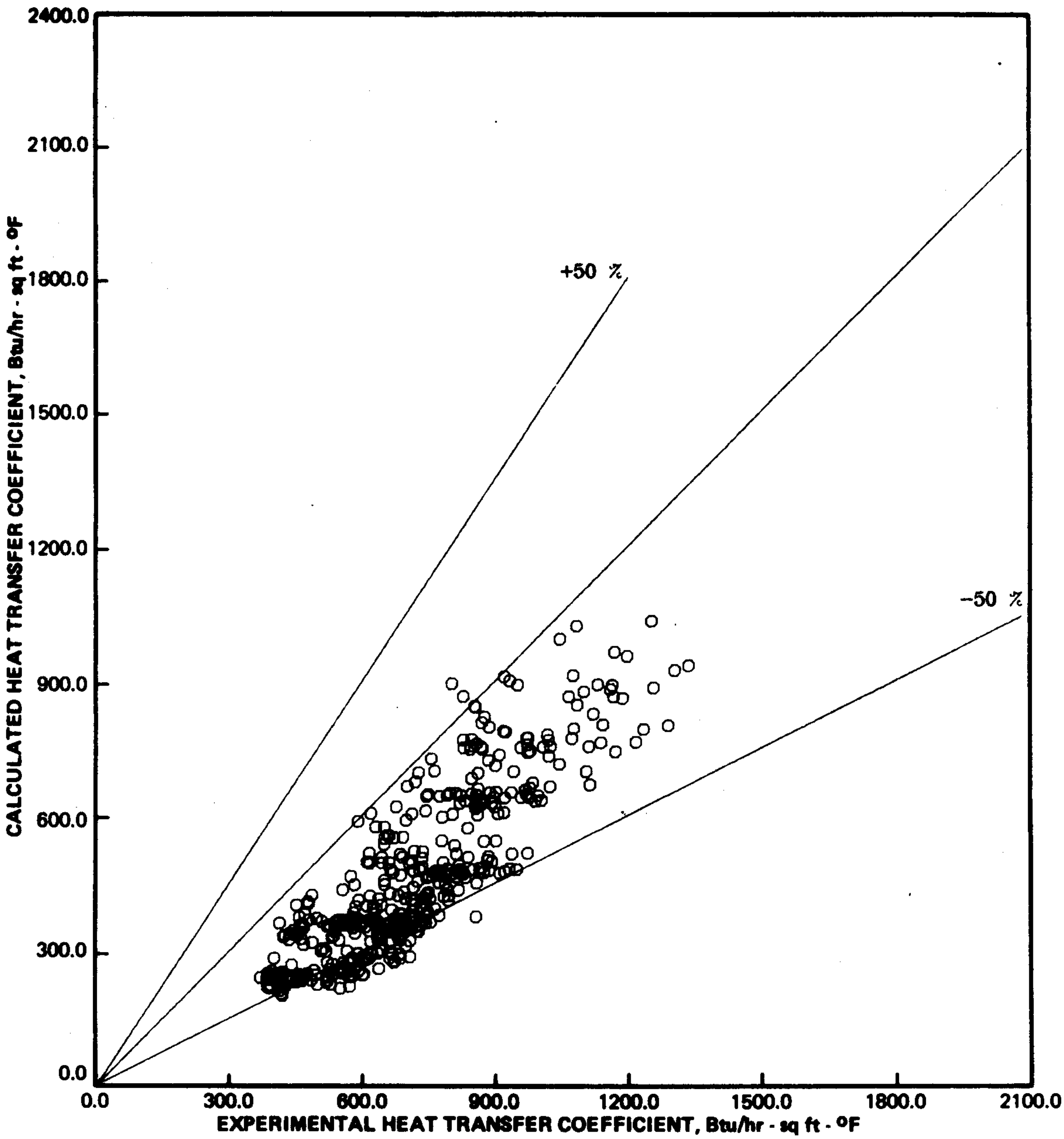

FIGURE IV-55. COMPARISON OF EXPERIMENTAL DATA WITH SCHROCK AND GROSSMAN [49] CORRELATION, 150°F.

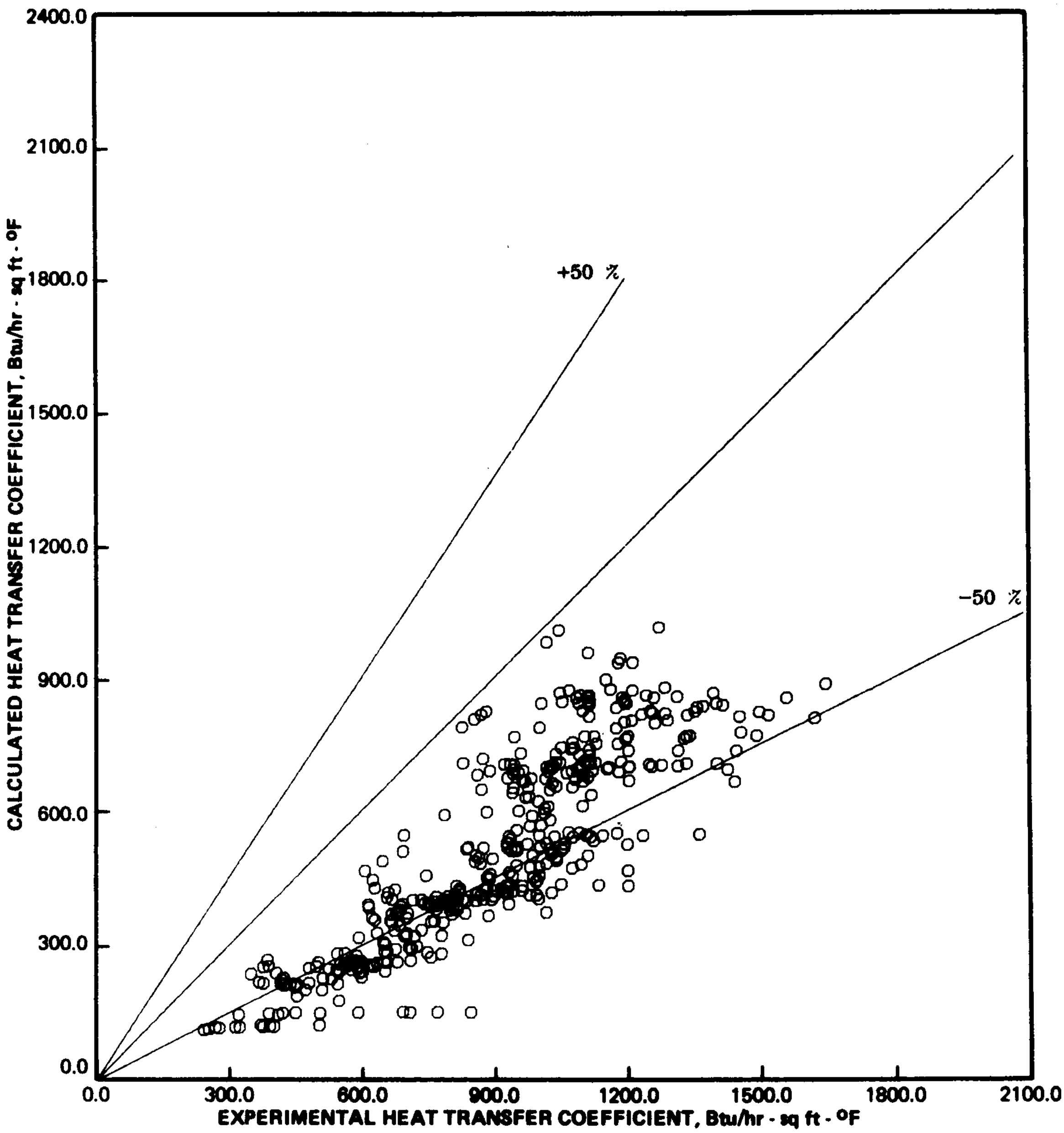

FIGURE IV-56. COMPARISON OF EXPERIMENTAL DATA WITH SCHROCK AND GROSSMAN [49] CORRELATION, 200°F.

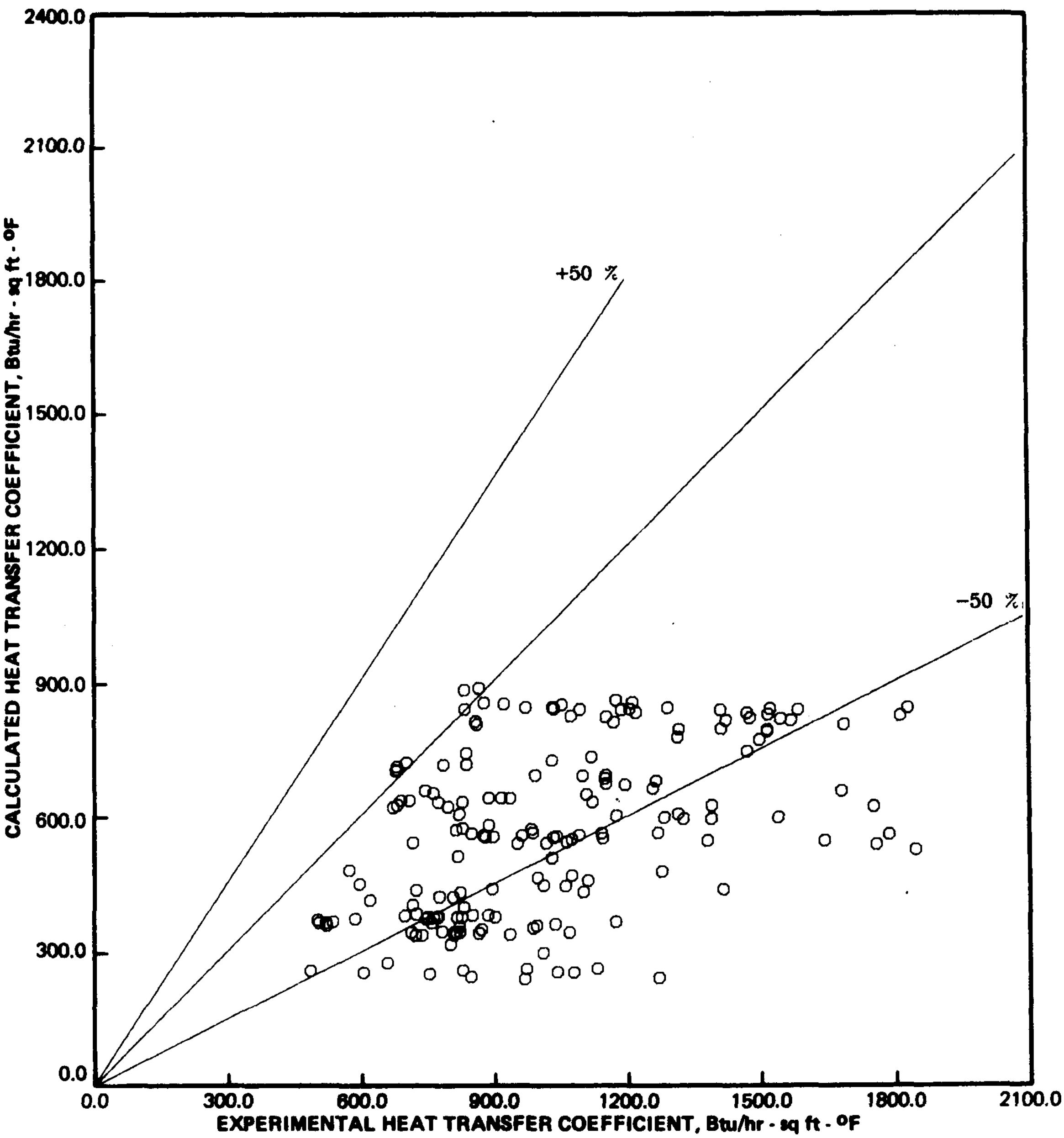

FIGURE IV-57. COMPARISON OF EXPERIMENTAL DATA WITH SCHROCK AND GROSSMAN [49] CORRELATION, 250°F.

Figures IV-58 to IV-61 illustrate the poor agreement between the correlation and the experimental data. The correlation appears to worsen as boiling temperature increases.

Dengler-Addoms [46] and Guerrieri-Talty [52] Correlations. These two correlations are similar in form. Dengler-Addoms [46] chose to base their correlation on the heat transfer coefficients of the *liquid only*, while Guerrieri and Talty [52] chose the heat transfer coefficient of the *liquid alone*. When compared to the data of this investigation, similar results were obtained for both correlations, as shown in Figures IV-62 to IV-69. The large amount of scatter in the correlated data should be apparent. Such poor agreement would imply that a correlation of this type is not well-suited for refrigerants.

Chawla [56] Correlation. The degree of agreement obtained with Chawla's [56] correlation was somewhat disappointing. As Figures IV-70 to IV-73 clearly indicate, this correlation severely underpredicts the experimental data at all saturation temperatures, especially the higher ones. As mentioned under the discussion of Chawla's [28] pressure-drop correlation, the discrepancy is probably due to the fact that Chawla's original data were taken at very low saturation temperatures. Based on this fact, it is not too surprising that the degree of agreement here is seen to be best at a saturation temperature of 100°F.

Chen [43] Correlation. The original Chen [43] correlation based on the Forster-Zuber [38] pool boiling correlation, predicted the experimental data of this study better than any of the previously discussed correlations. Agreement was best at boiling temperature of 100 and 150°F. At the higher boiling temperatures, 200 and 250°F, the Chen [43] correlation tended to underestimate the experimental data by more than 50% in some cases. These trends are shown in Figures IV-74 to IV-77.

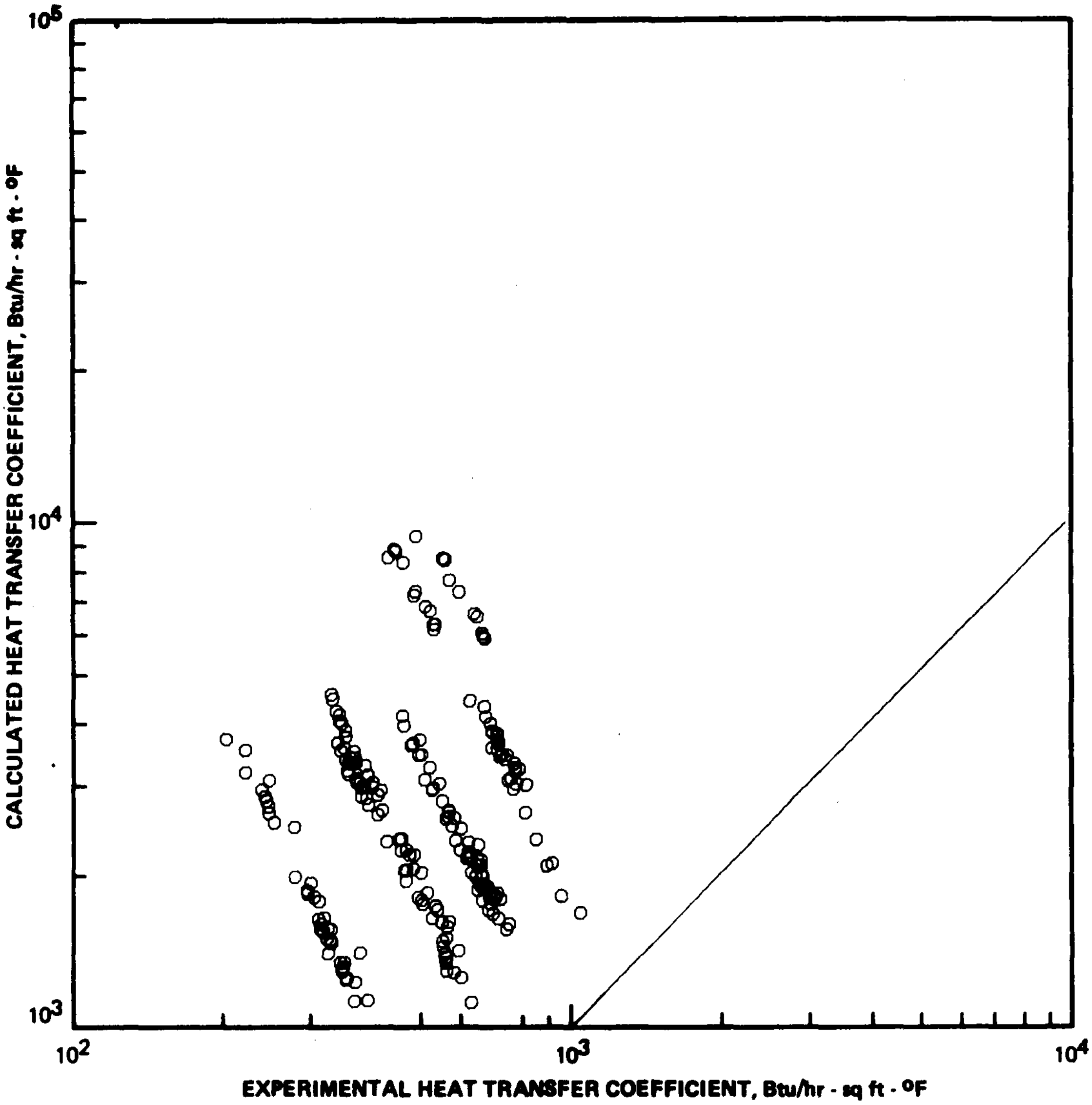

FIGURE IV-58. COMPARISON OF EXPERIMENTAL DATA WITH CHESHIRE AND STERLING [44] CORRELATION, 100°F.

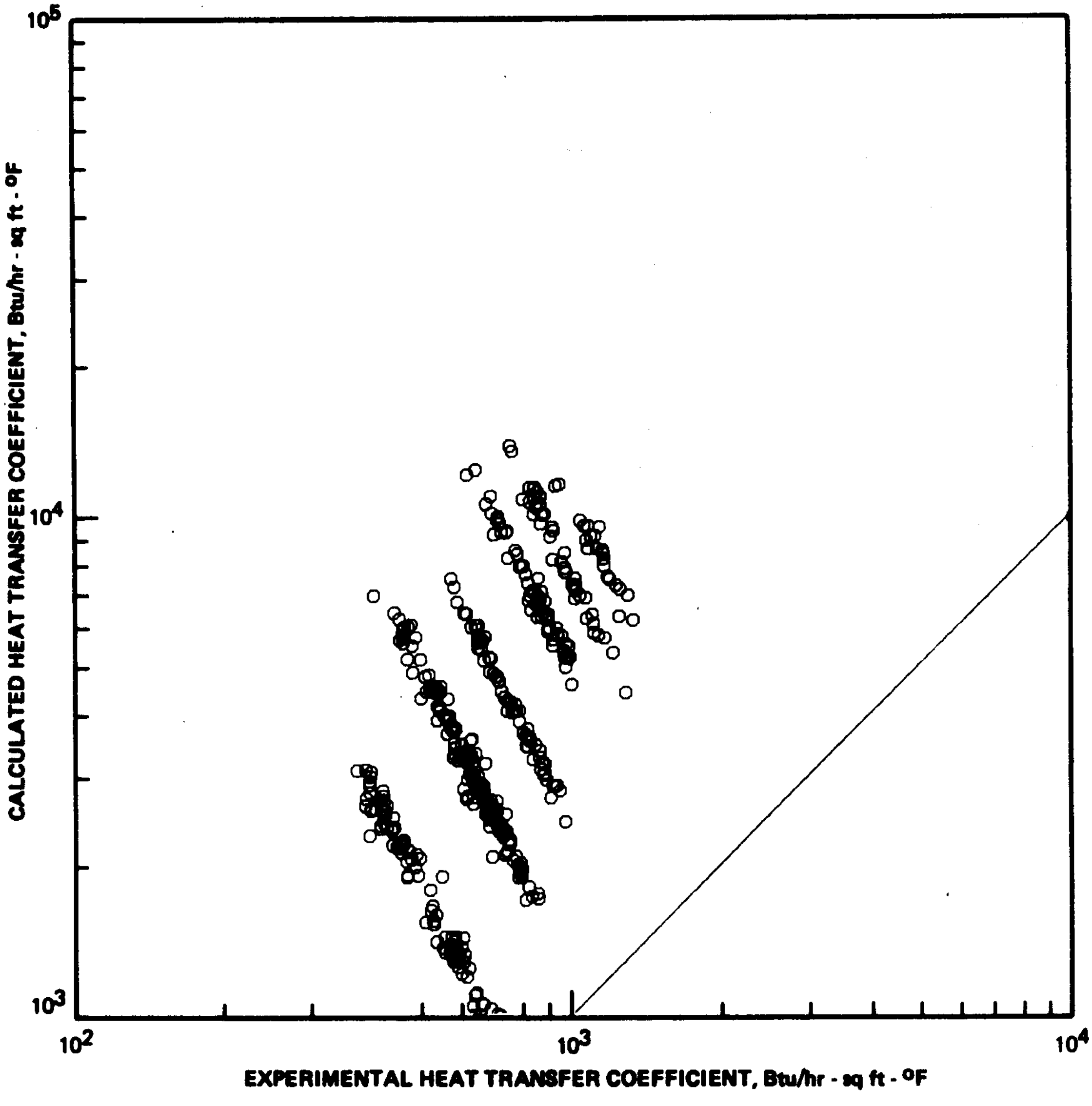

FIGURE IV-59. COMPARISON OF EXPERIMENTAL DATA WITH CHESHIRE AND STERLING [44] CORRELATION, 150°F.

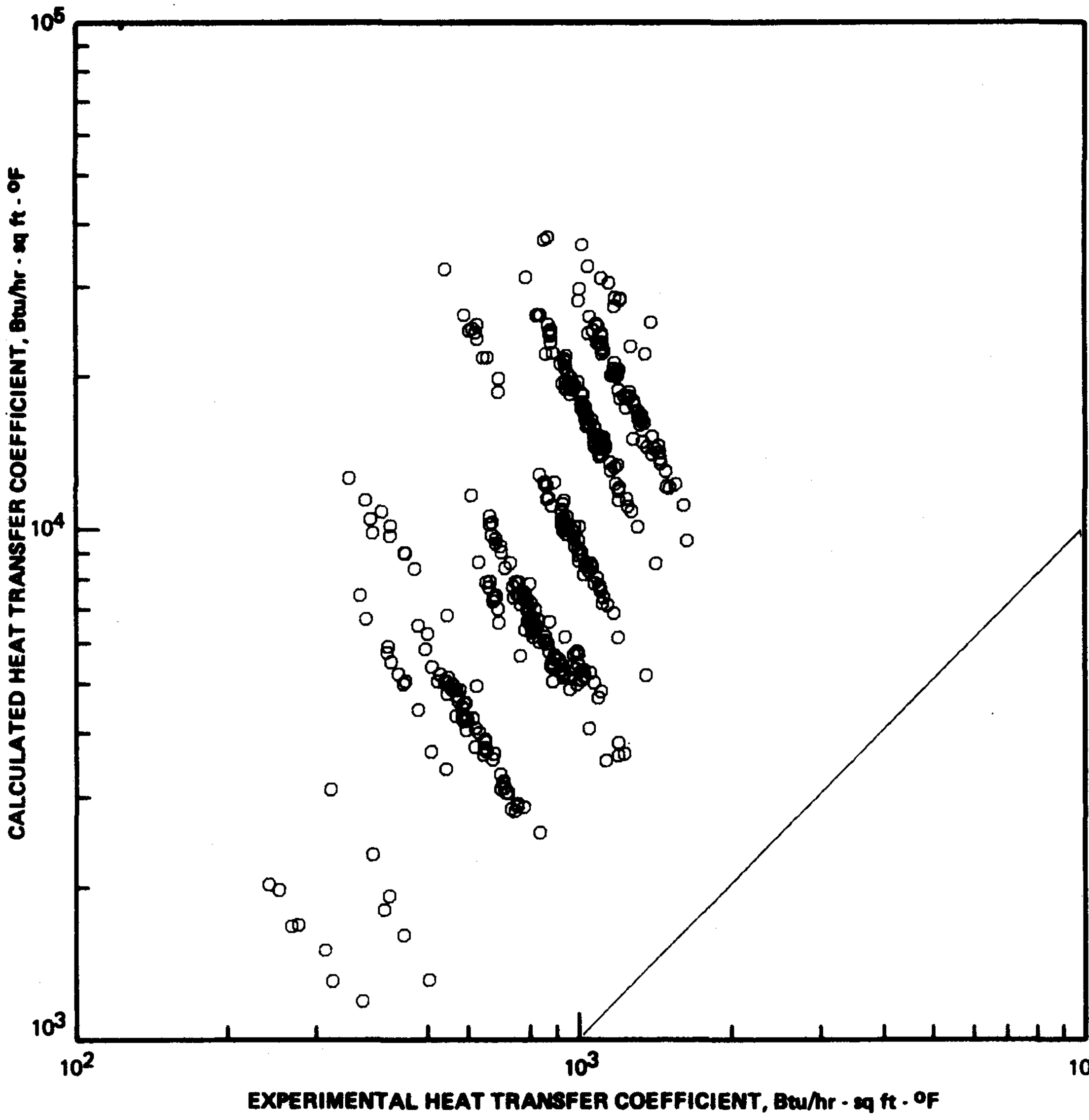

FIGURE IV-60. COMPARISON OF EXPERIMENTAL DATA WITH CHESHIRE AND STERLING [44] CORRELATION, 200°F.

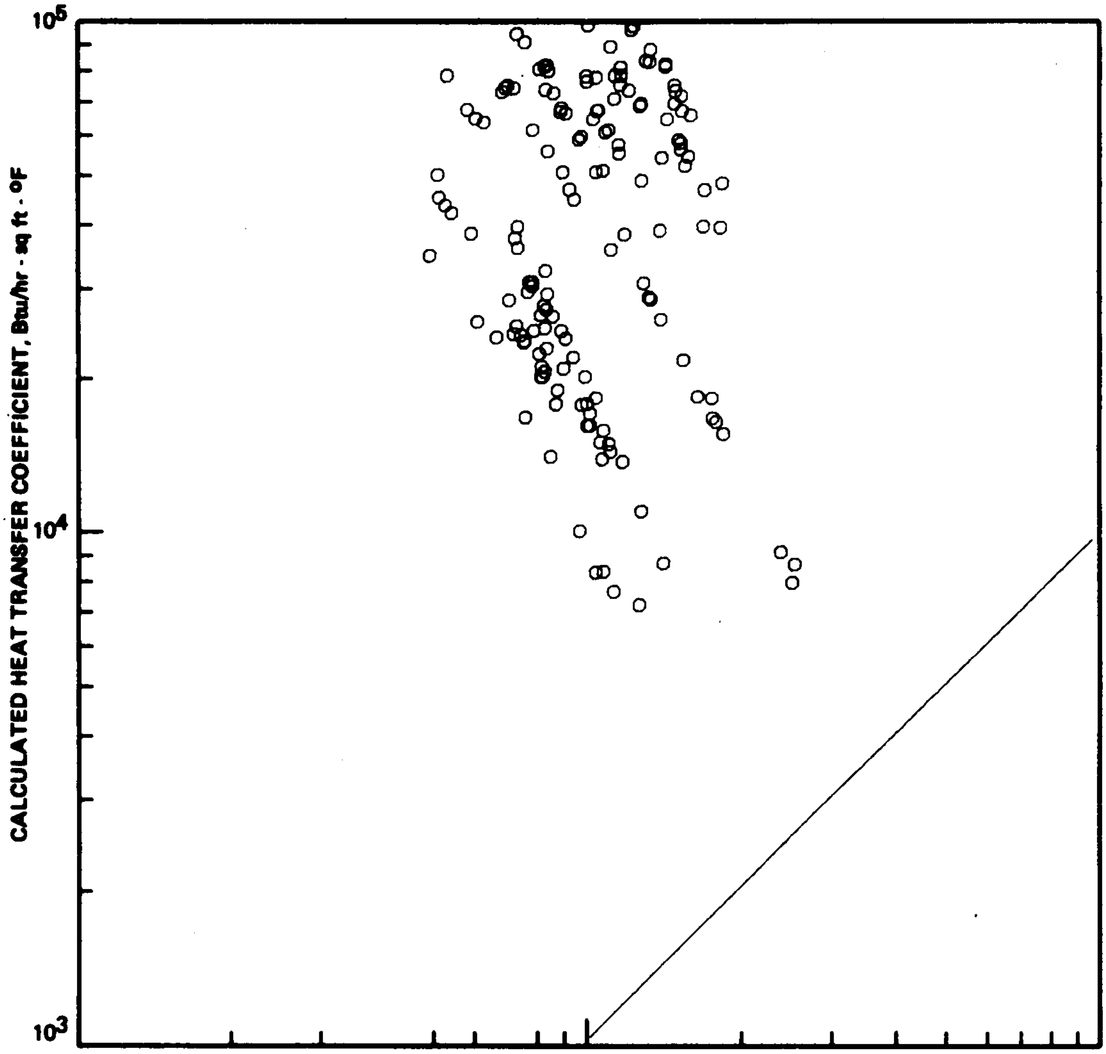

FIGURE IV-61. COMPARISON OF EXPERIMENTAL DATA WITH CHESHIRE AND STERLING [44] CORRELATION, 250°F.

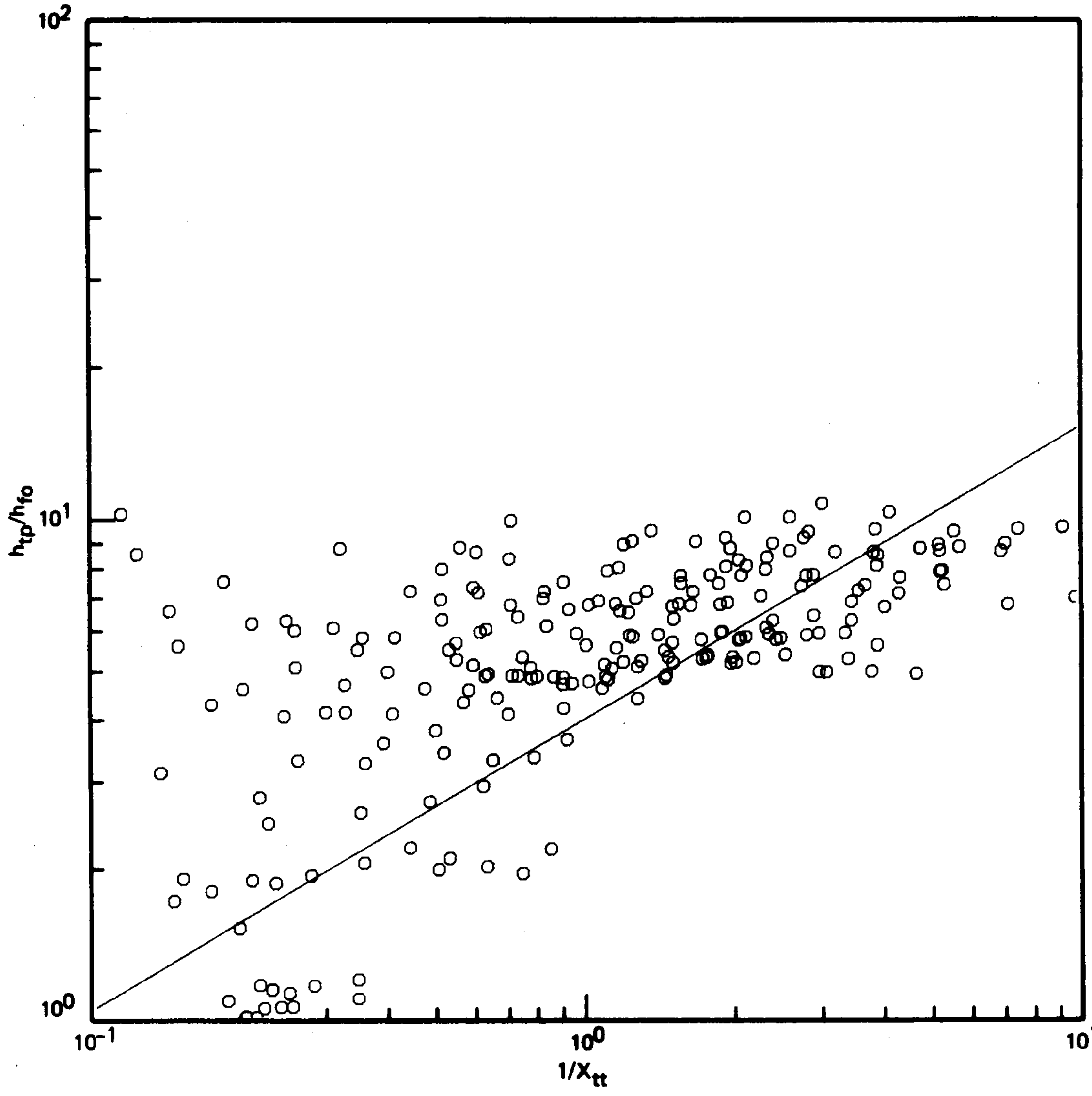

FIGURE IV-62. COMPARISON OF EXPERIMENTAL DATA WITH DENGLER AND ADDOMS [46] CORRELATION, 100°F.

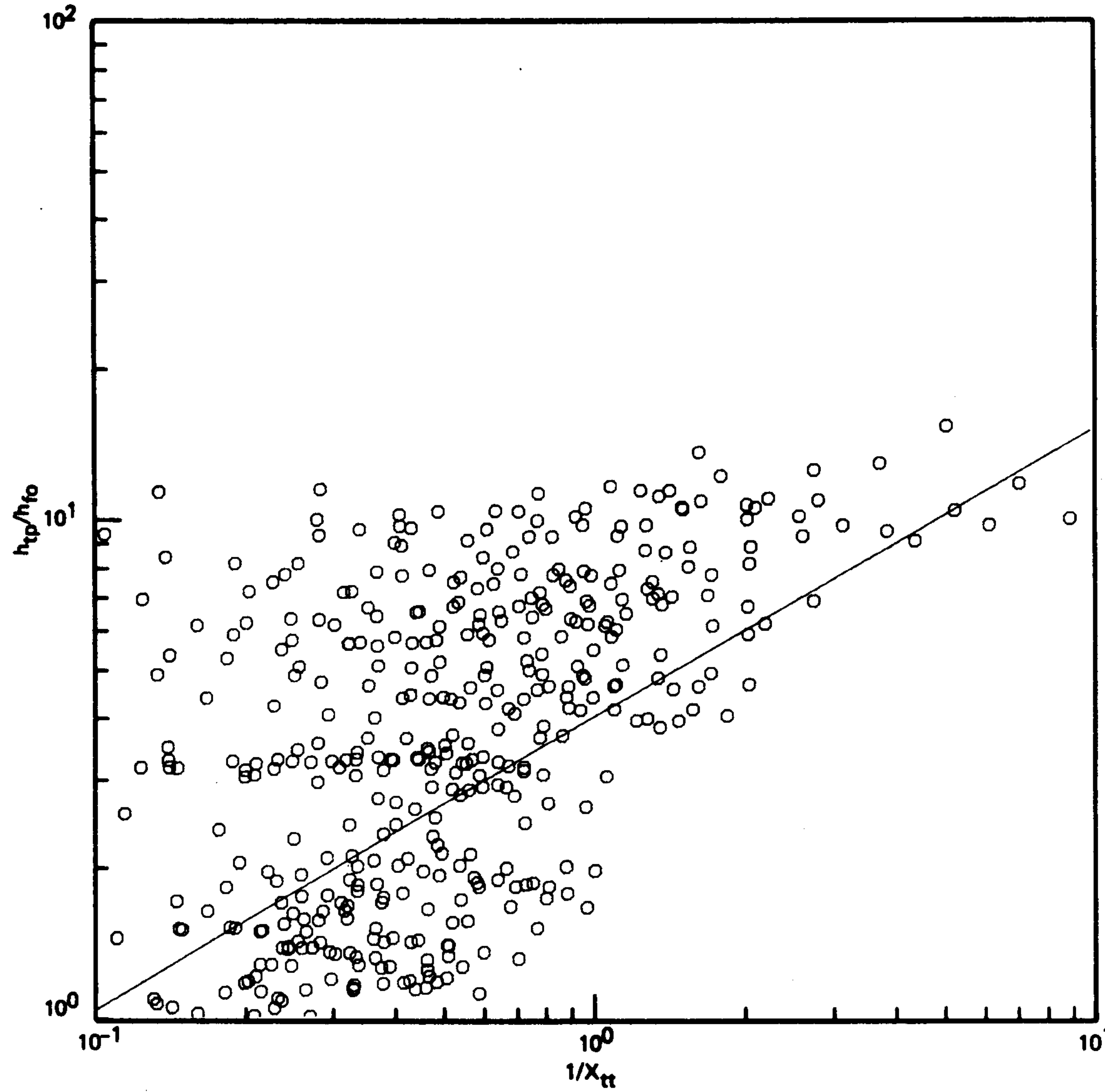

FIGURE IV-63. COMPARISON OF EXPERIMENTAL DATA WITH DENGLER AND ADDOMS [46] CORRELATION, 150°F.

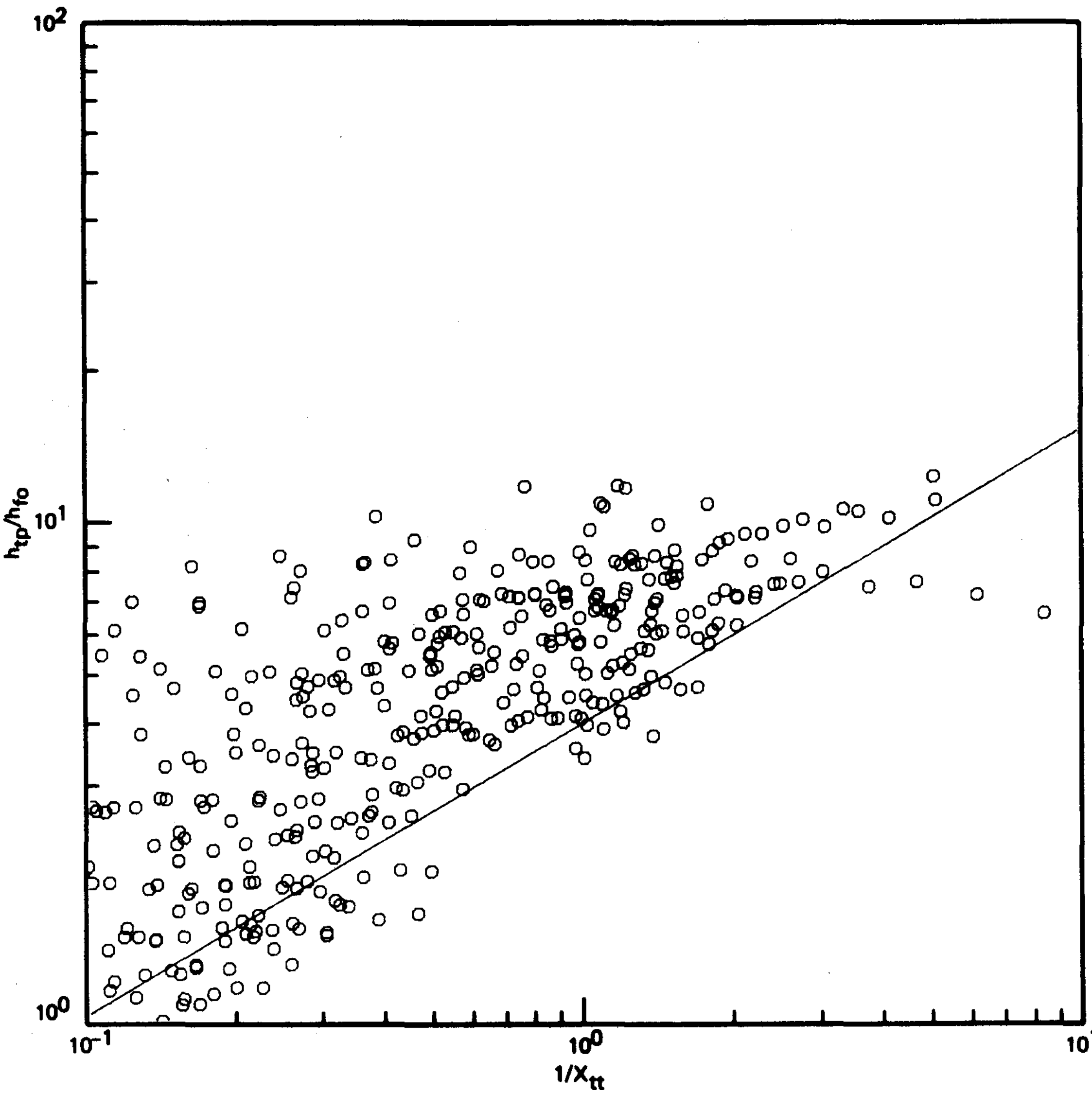

FIGURE IV-64. COMPARISON OF EXPERIMENTAL DATA WITH DENGLER AND ADDOMS [46] CORRELATION, 200°F.

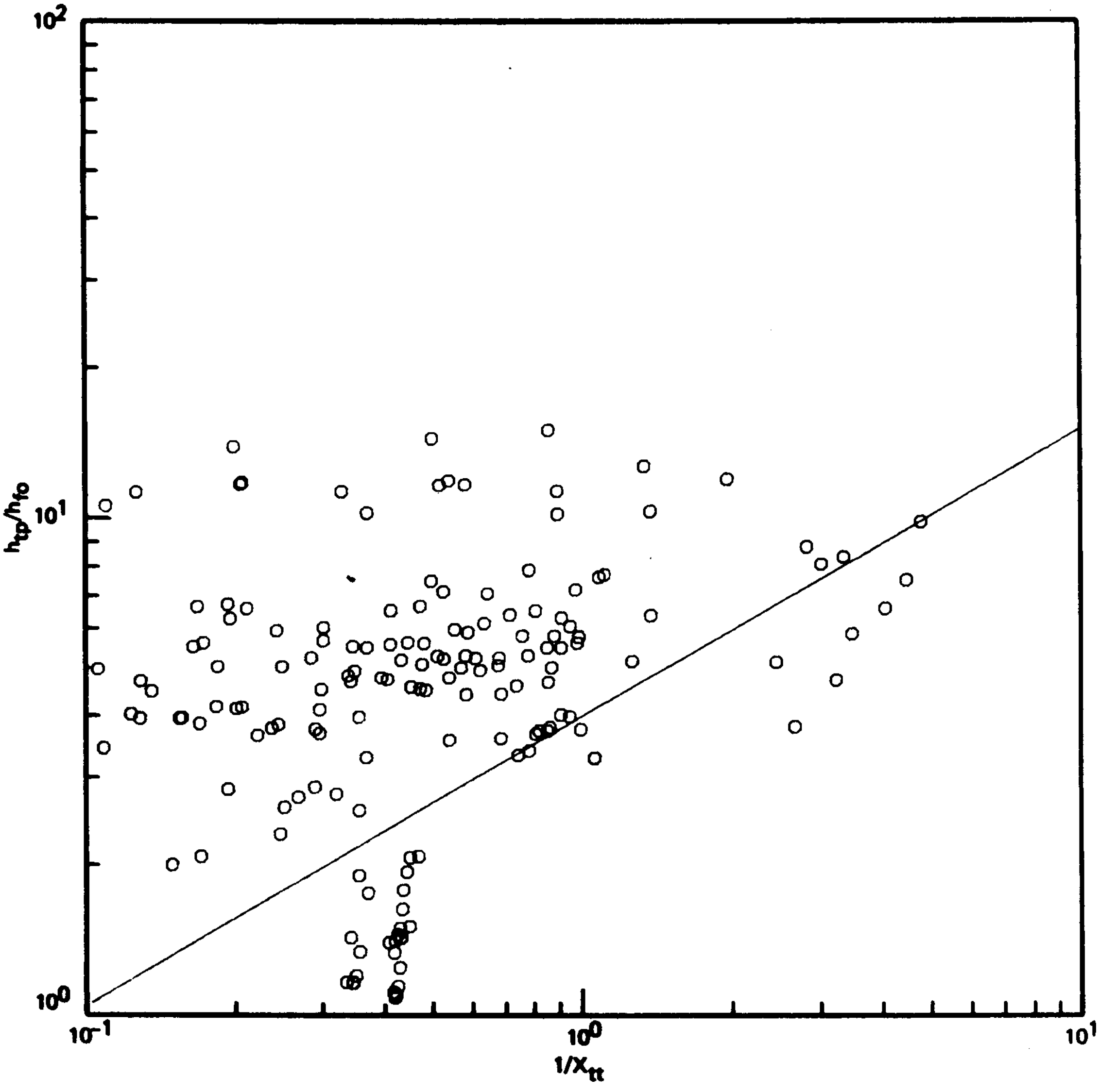

FIGURE IV-65. COMPARISON OF EXPERIMENTAL DATA WITH DENGLER AND ADDOMS [46] CORRELATION, 250°F.

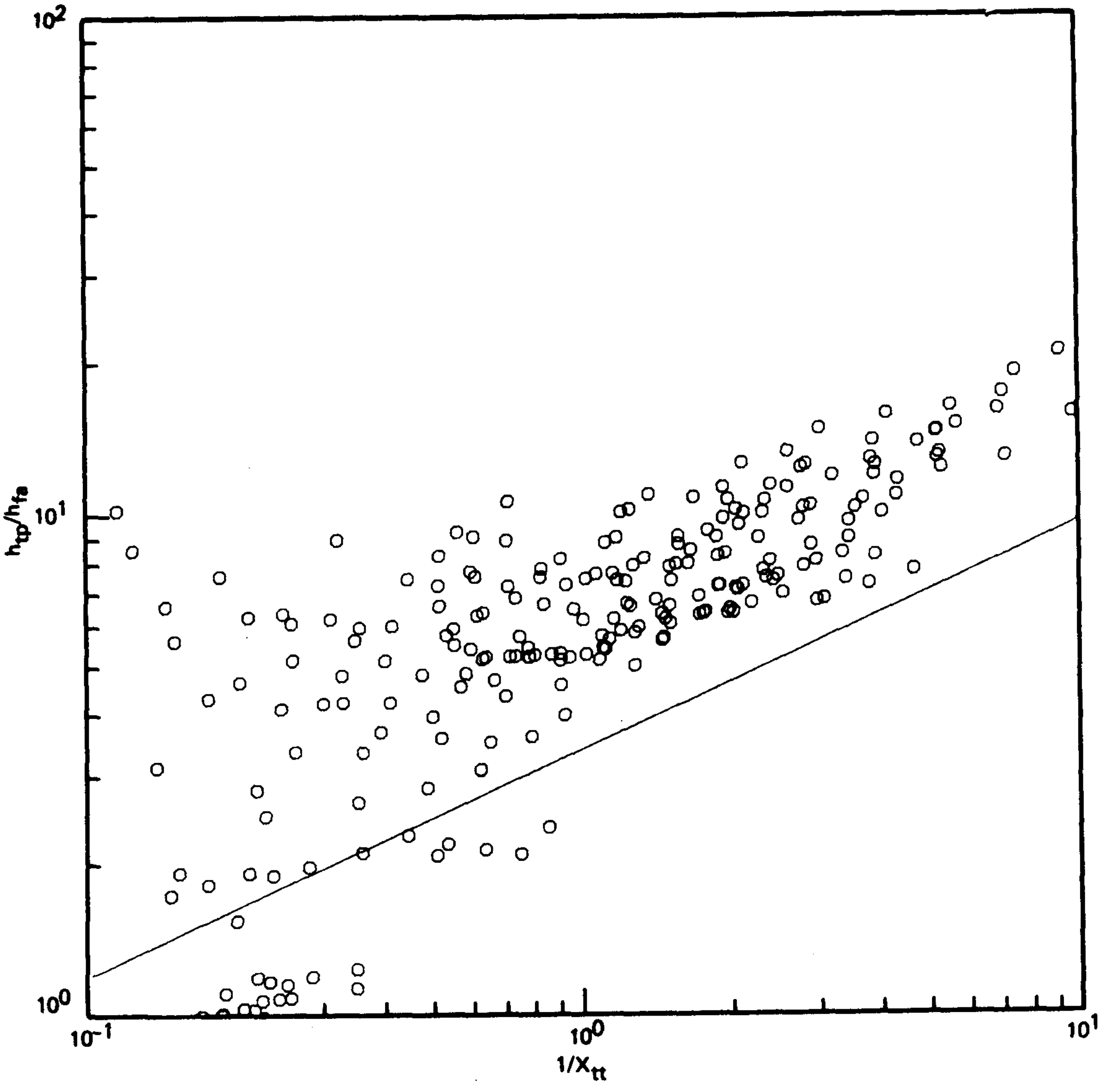

FIGURE IV-66. COMPARISON OF EXPERIMENTAL DATA WITH GUERRIERI AND TALTY [52] CORRELATION, 100°F.

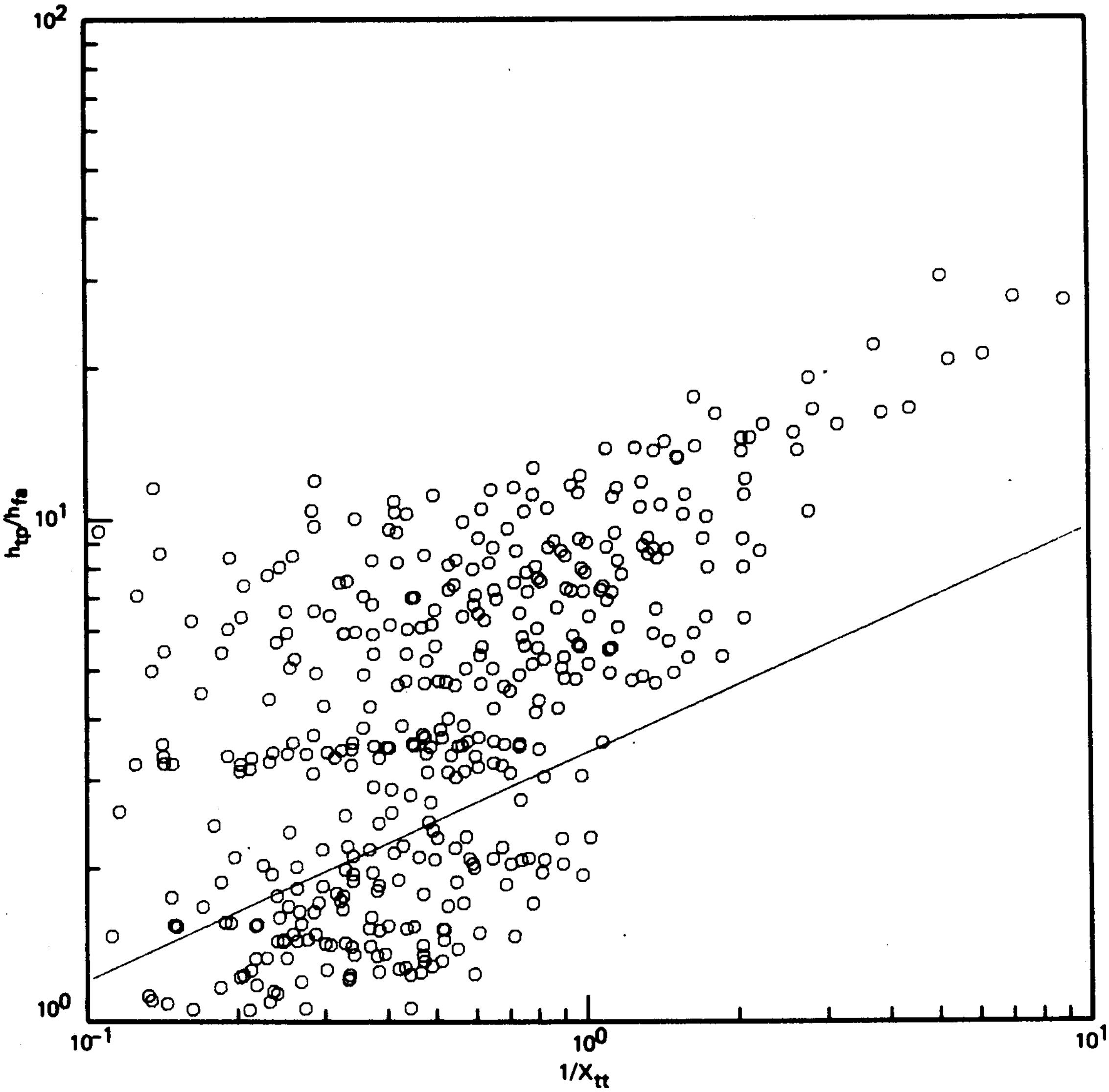

FIGURE IV-67. COMPARISON OF EXPERIMENTAL DATA WITH GUERRIERI AND TALTY [52] CORRELATION, 150°F.

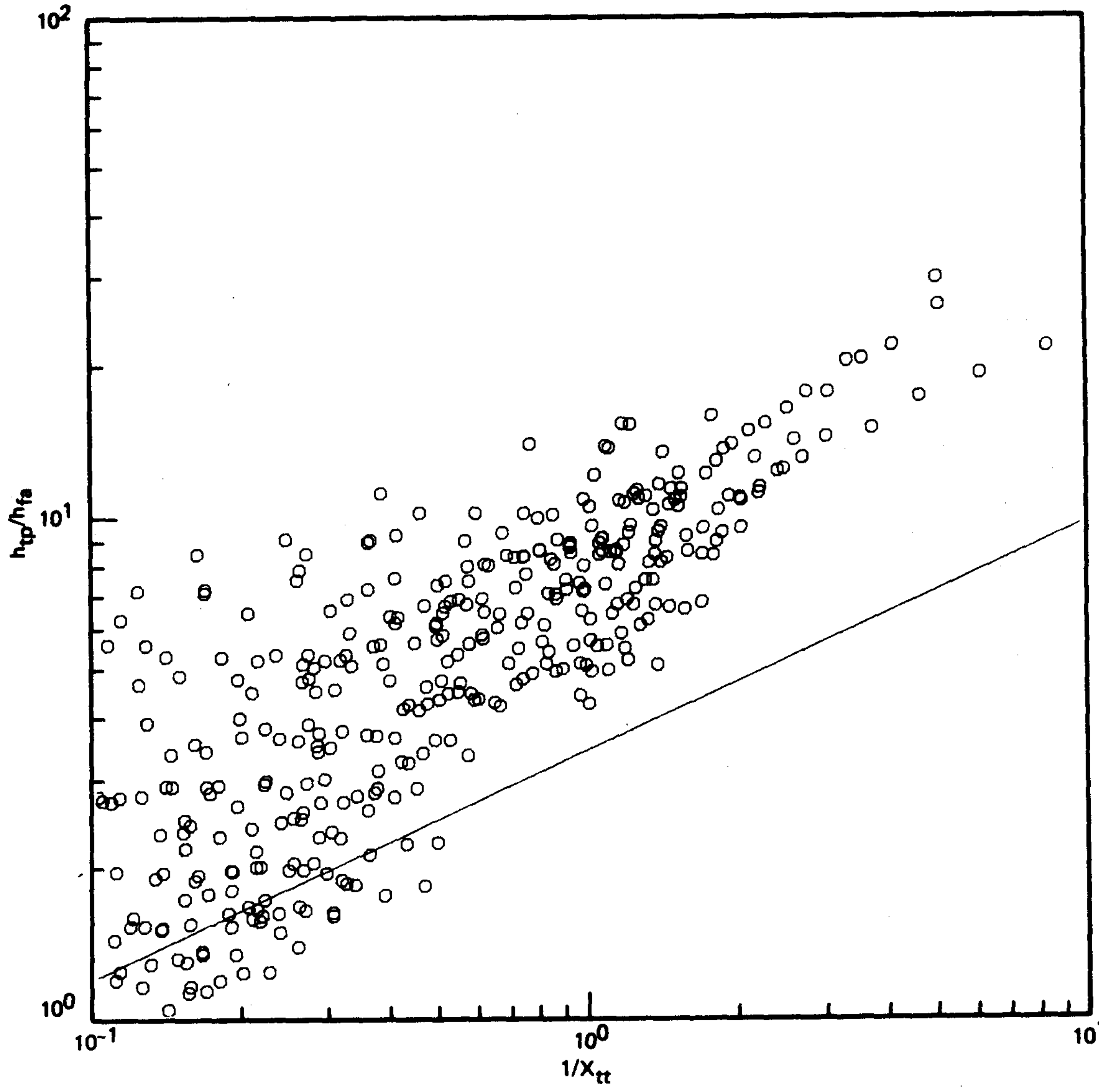

FIGURE IV-68. COMPARISON OF EXPERIMENTAL DATA WITH GUERRIERI AND TALTY [52] CORRELATION, 200°F.

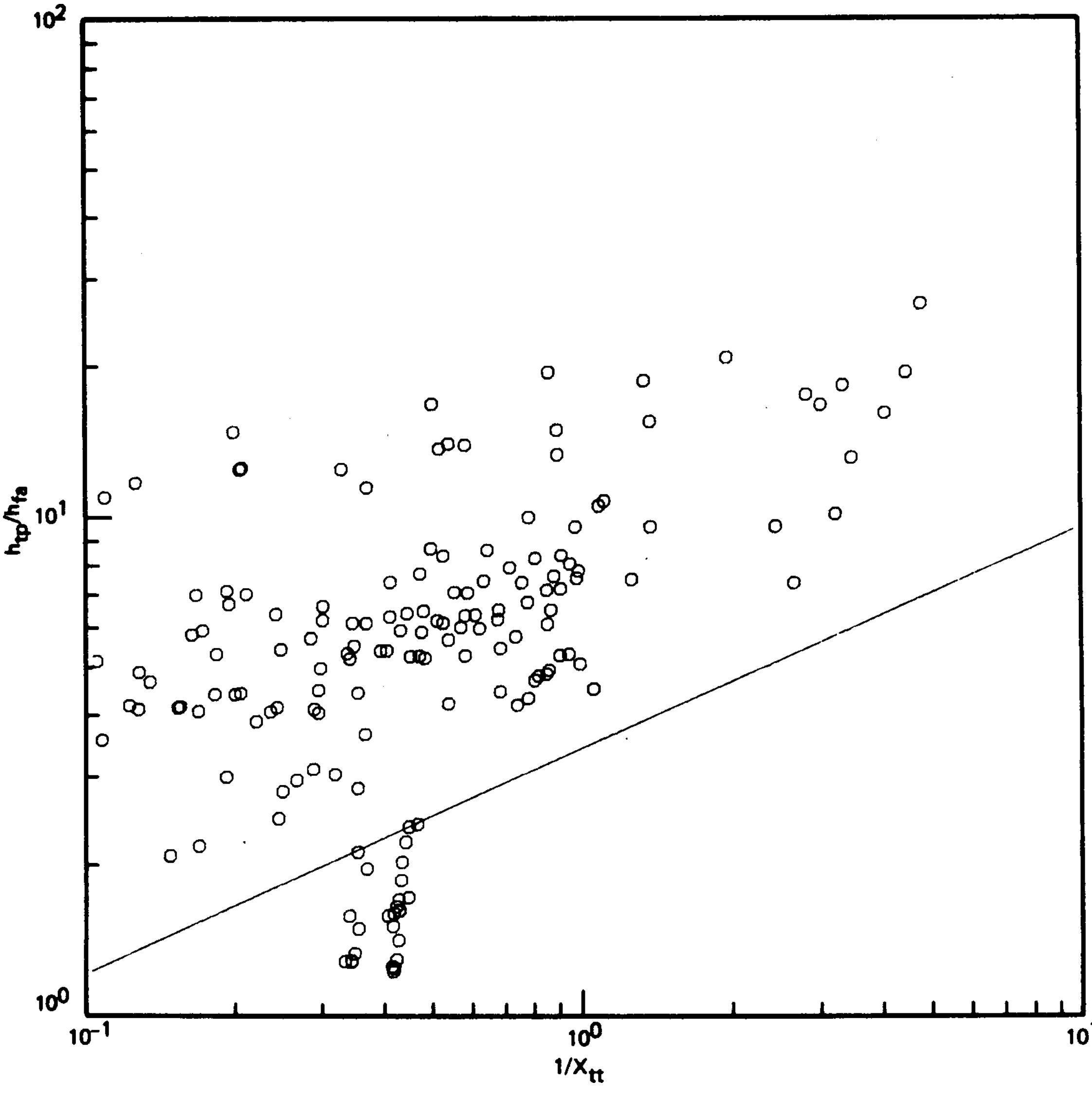

FIGURE IV-69. COMPARISON OF EXPERIMENTAL DATA WITH GUERRIERI AND TALTY [52] CORRELATION, 250°F.

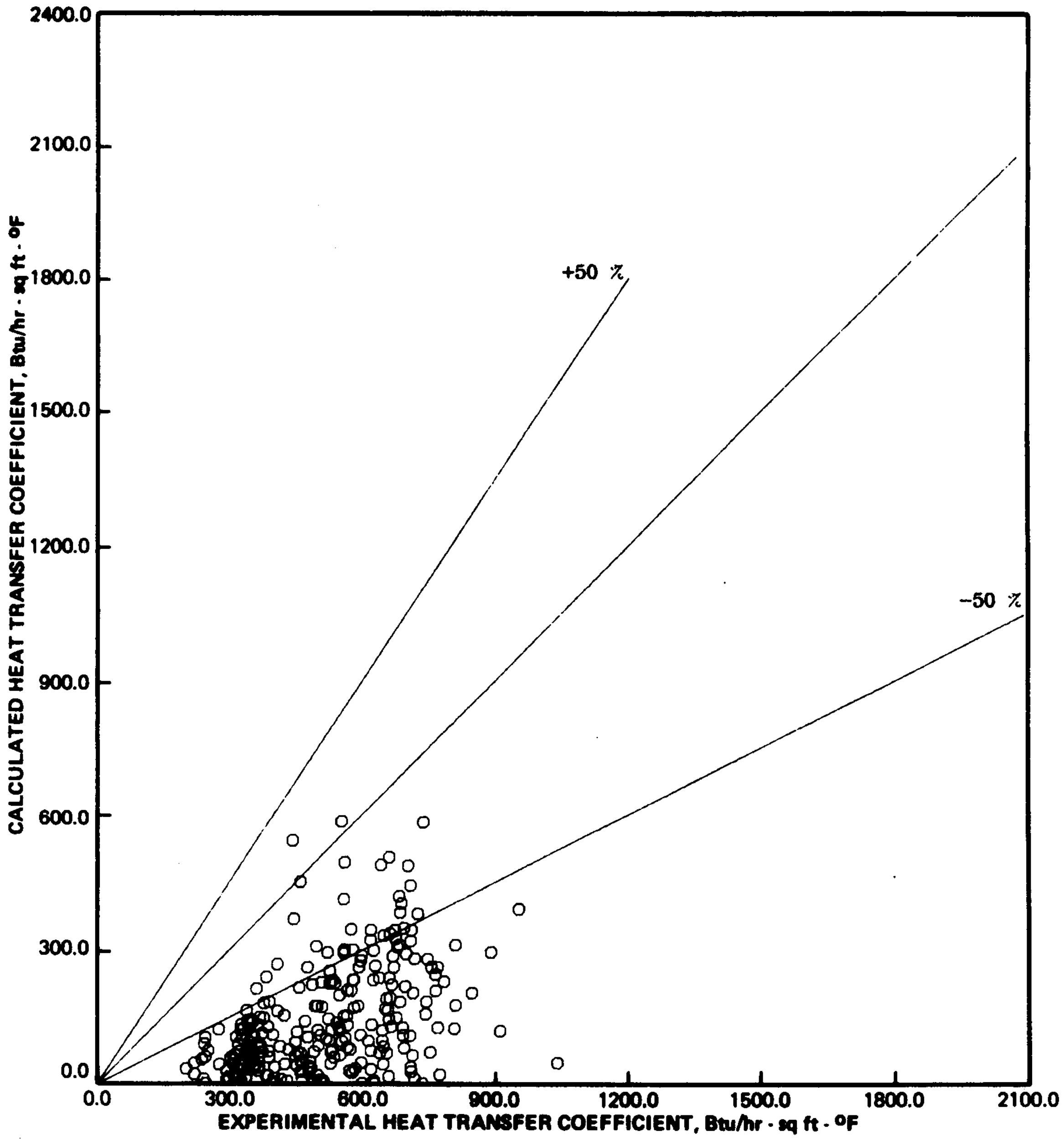

FIGURE IV-70. COMPARISON OF EXPERIMENTAL DATA WITH CHAWLA [56] CORRELATION, 100°F.

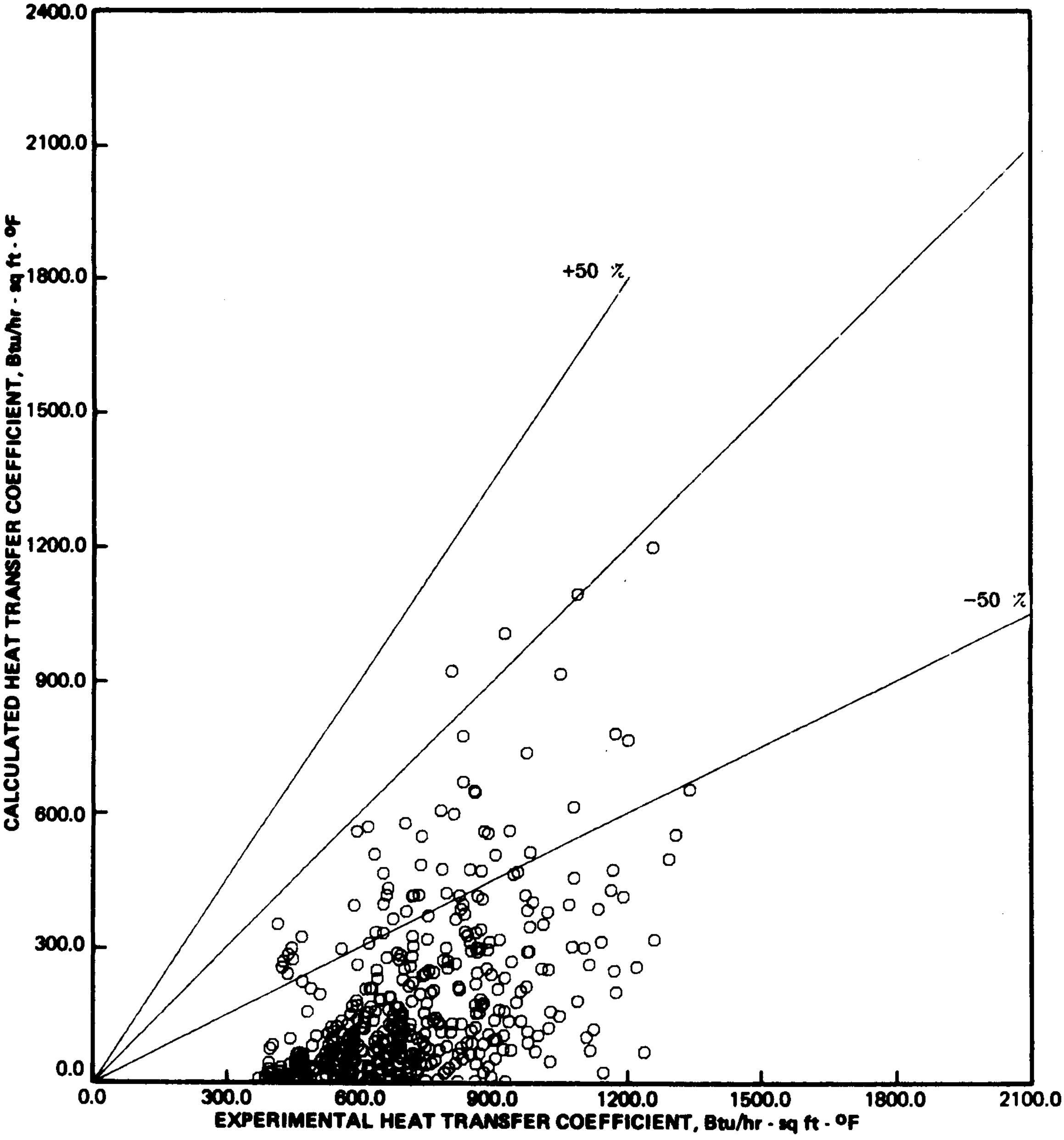

FIGURE IV-71. COMPARISON OF EXPERIMENTAL DATA WITH CHAWLA [56] CORRELATION, 150°F.

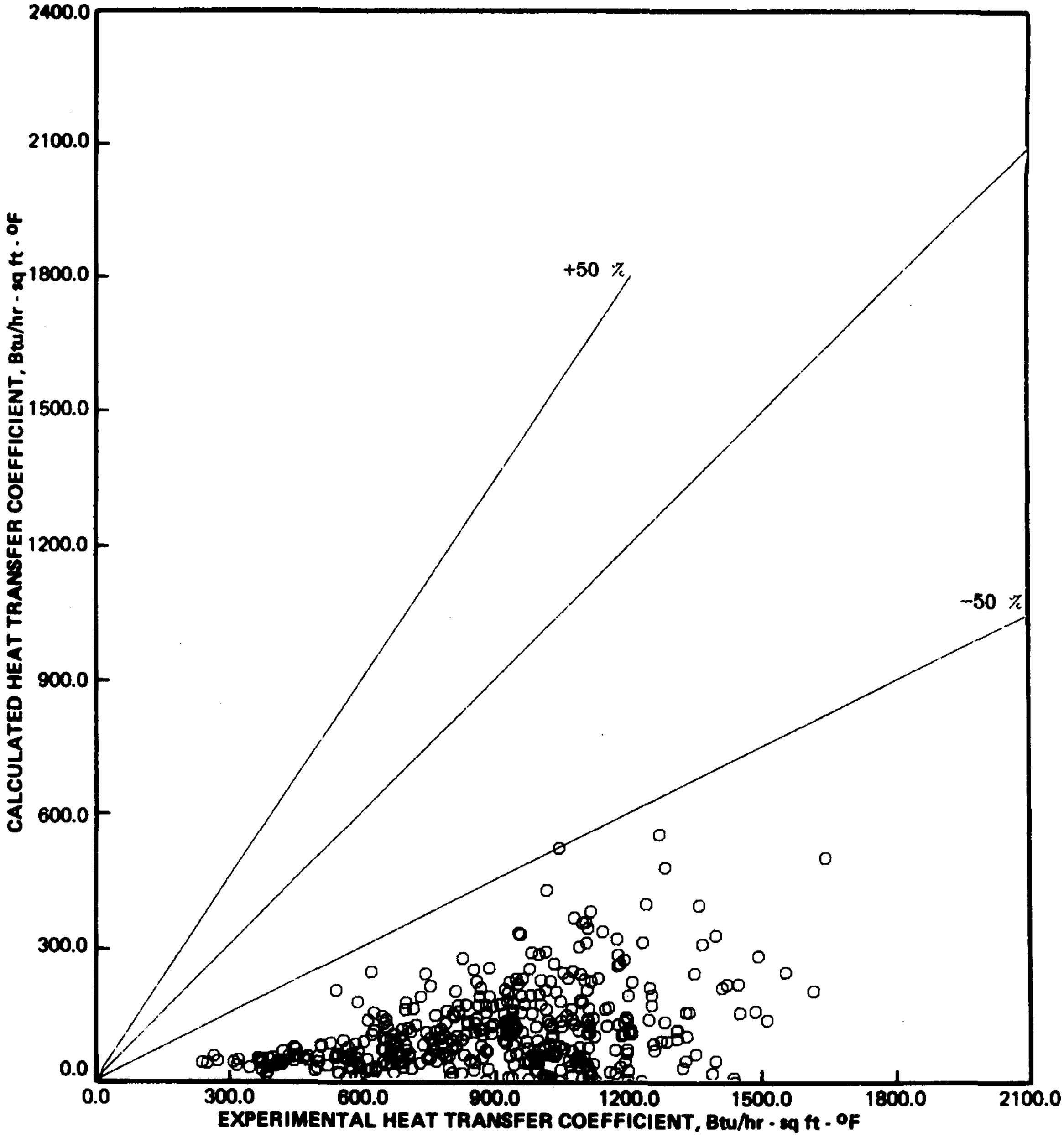

FIGURE IV-72. COMPARISON OF EXPERIMENTAL DATA WITH CHAWLA [56] CORRELATION, 200°F.

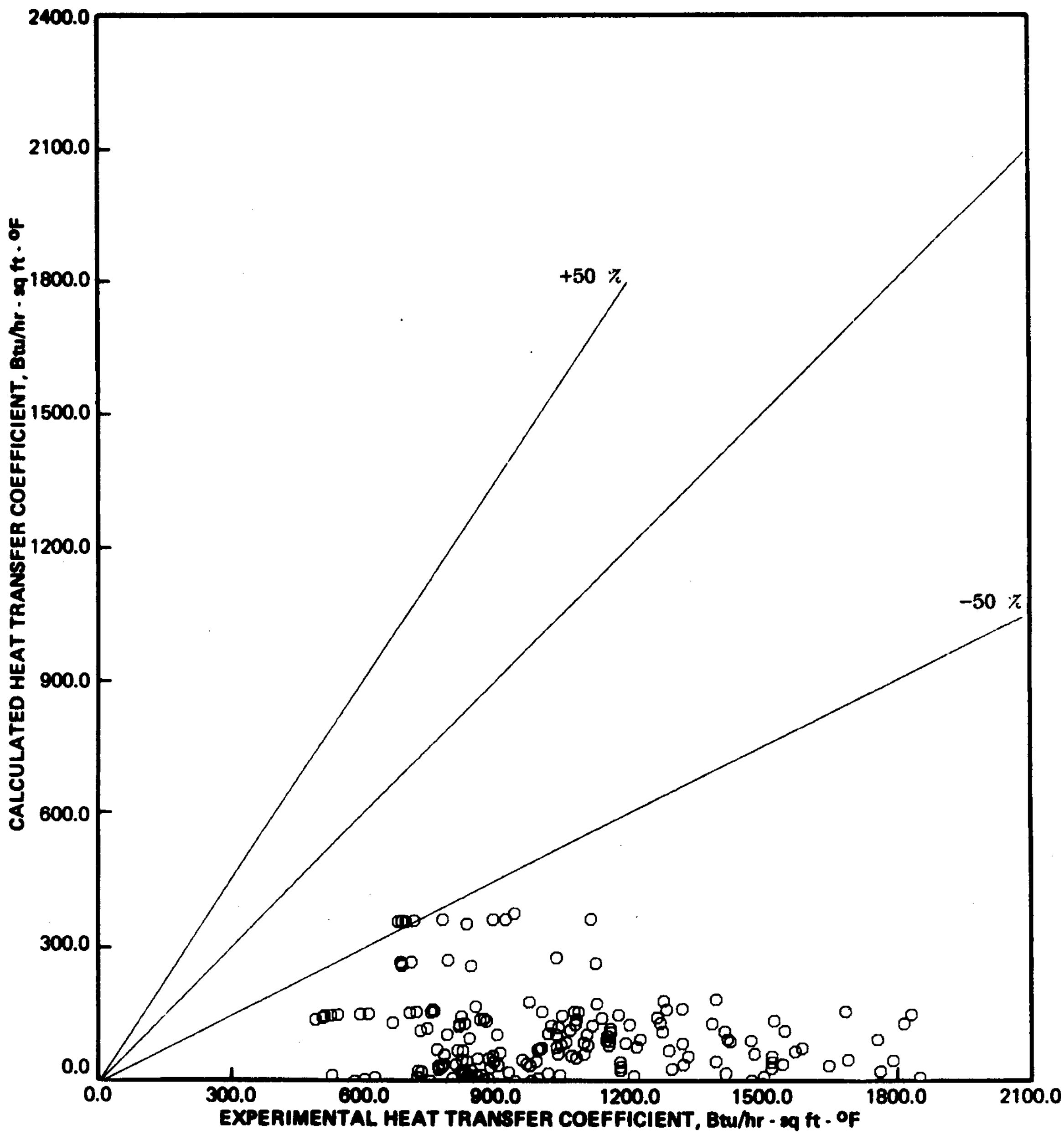

FIGURE IV-73. COMPARISON OF EXPERIMENTAL DATA WITH CHAWLA [56] CORRELATION, 250°F.

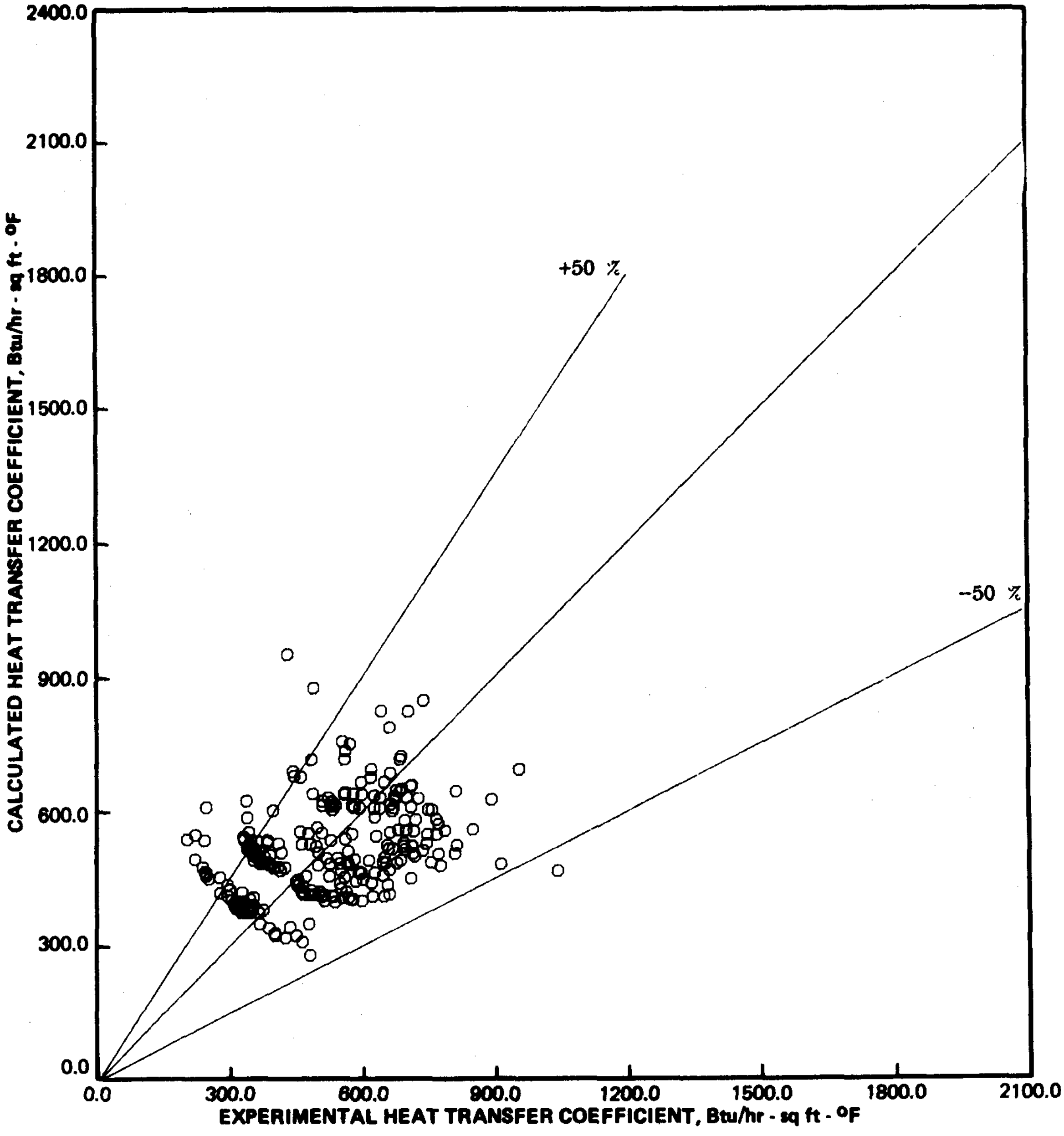

FIGURE IV-74. COMPARISON OF EXPERIMENTAL DATA WITH CHEN [43] CORRELATION, 100°F.

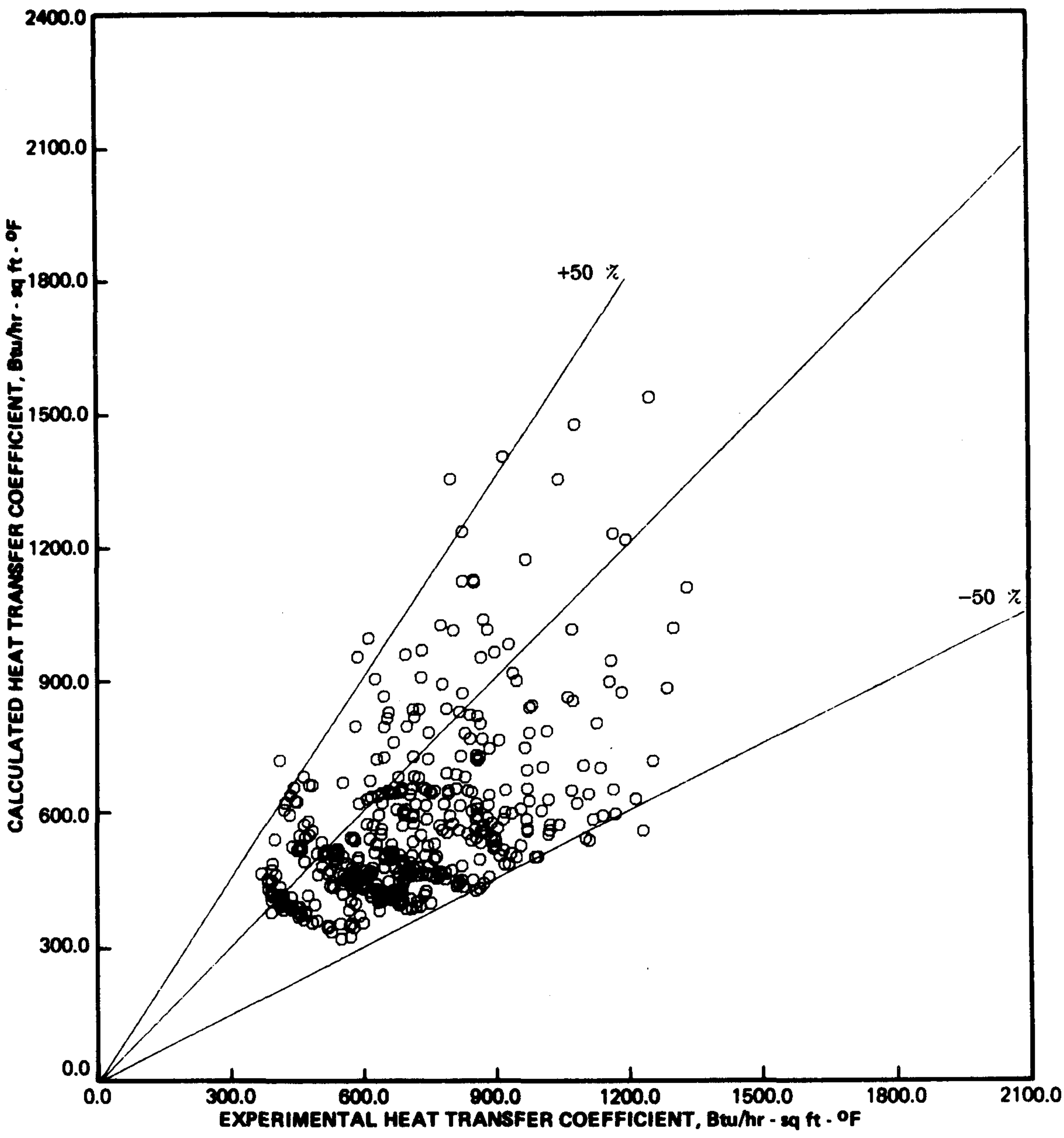

FIGURE IV-75. COMPARISON OF EXPERIMENTAL DATA WITH CHEN [43] CORRELATION, 150°F.

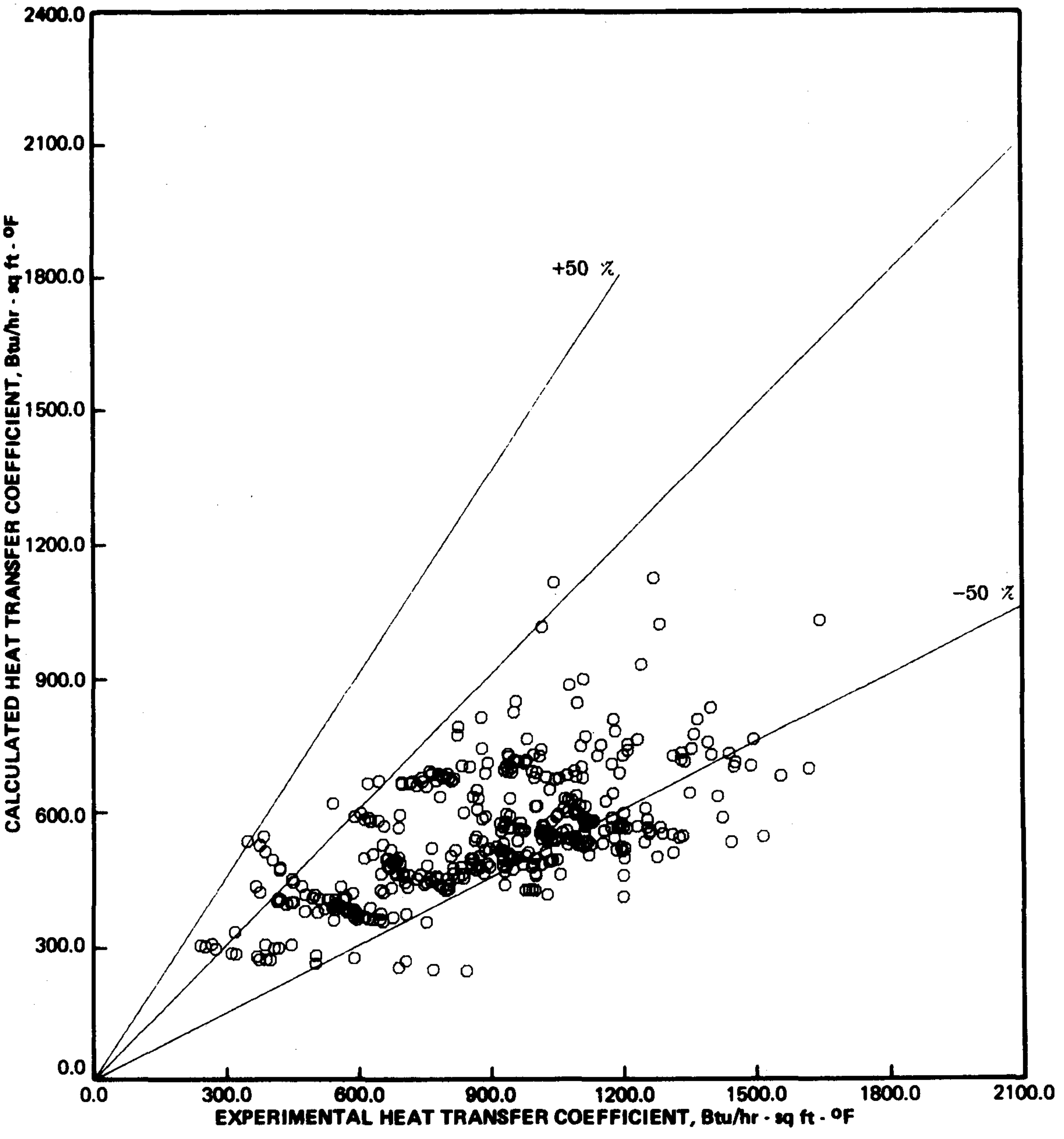

FIGURE IV-76. COMPARISON OF EXPERIMENTAL DATA WITH CHEN [43] CORRELATION, 200°F.

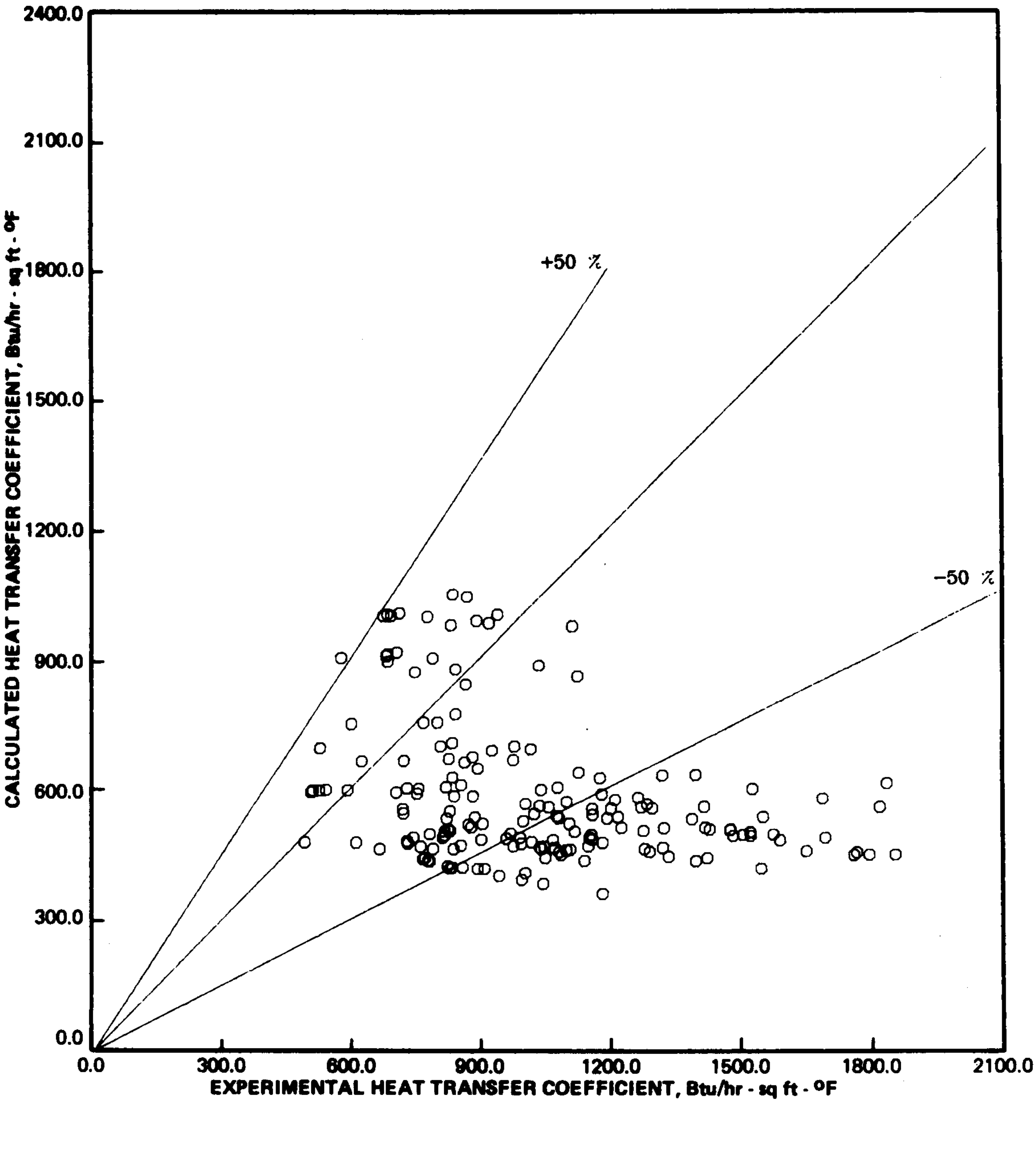

FIGURE IV-77. COMPARISON OF EXPERIMENTAL DATA WITH CHEN [43] CORRELATION, 250°F.

Jallouk's [42] Modification to Chen's [43] Correlation. Jallouk [42] has recently modified the correlation of Chen [43] in order to make it more applicable for refrigerants especially in vertical flow. A comparison of the present data with this new correlation is given by Figures IV-78 through IV-81. As the figures show, the degree of agreement tended to improve with increased boiling temperature. The agreement at boiling temperatures of 100 and 150°F was, however, unsatisfactory. At these boiling temperatures a significant portion of the data were vastly overestimated by the Jallouk [42] modification.

In summary, it can be stated that no existing correlation was completely adequate in predicting the experimental heat transfer coefficient data obtained during this effort. The original Chen [43] correlation provided the best results; however, it was rather poor at the higher boiling temperatures.

Proposed Correlation. Since there is a definite need for a correlation that is suitable for the ranges of operating conditions considered here, an attempt was made, beginning with the original Chen [43] correlation supplemented by Jallouk's [42] modification, at developing a more reliable correlation. A reduction in the degree of scatter exhibited by the experimental data was used as the criteria for measuring improvement.

As a first step the degree of agreement between all experimental data and these two correlations was closely studied. From this effort, it became evident that a mass flux effect was present. That is to say, the experimental data at mass fluxes of 0.12 to 0.24 x 10^6 lbm/hr-sq ft could be separated from the data obtained at the other mass fluxes by

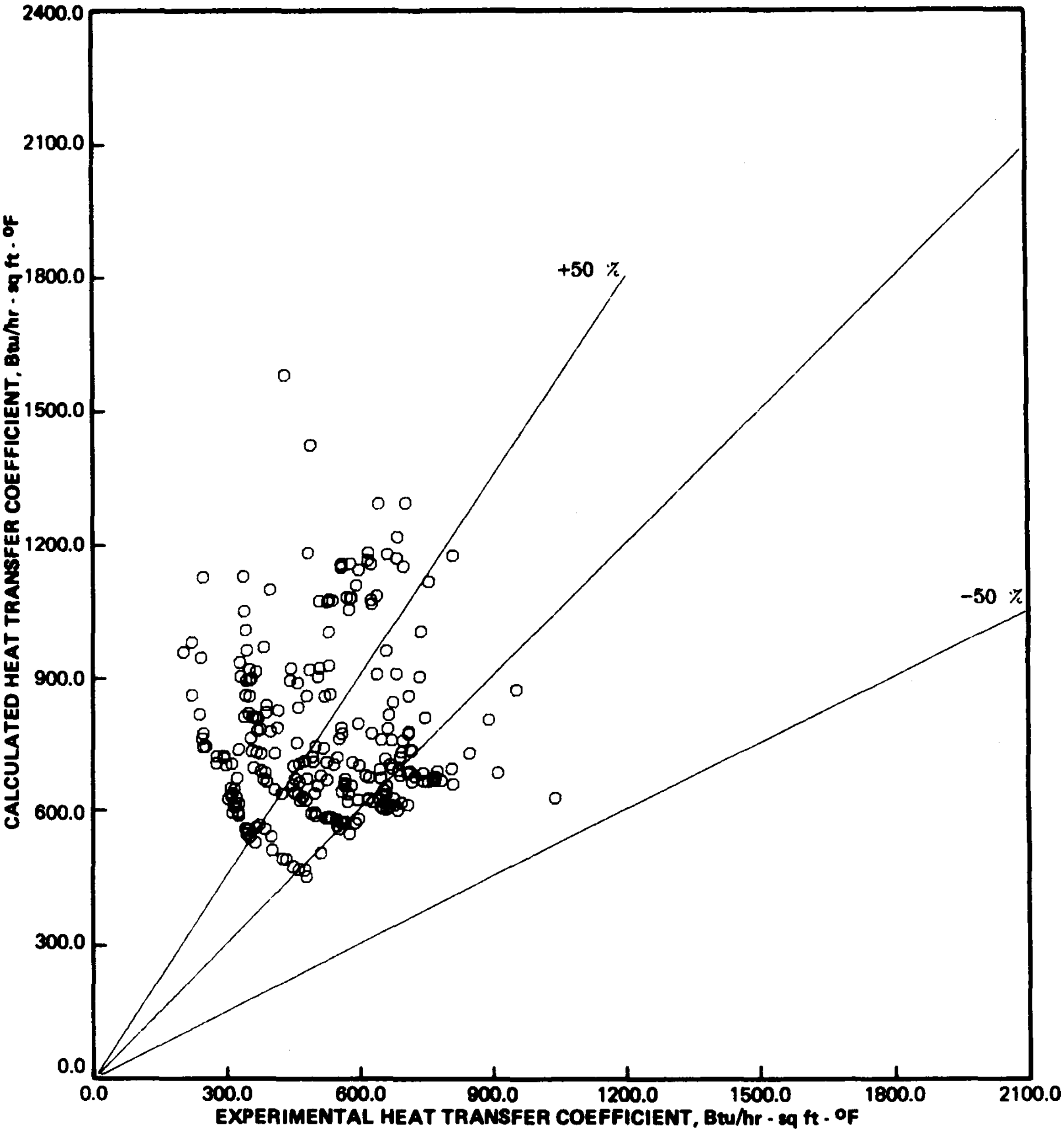

FIGURE IV-78. COMPARISON OF EXPERIMENTAL DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 100°F.

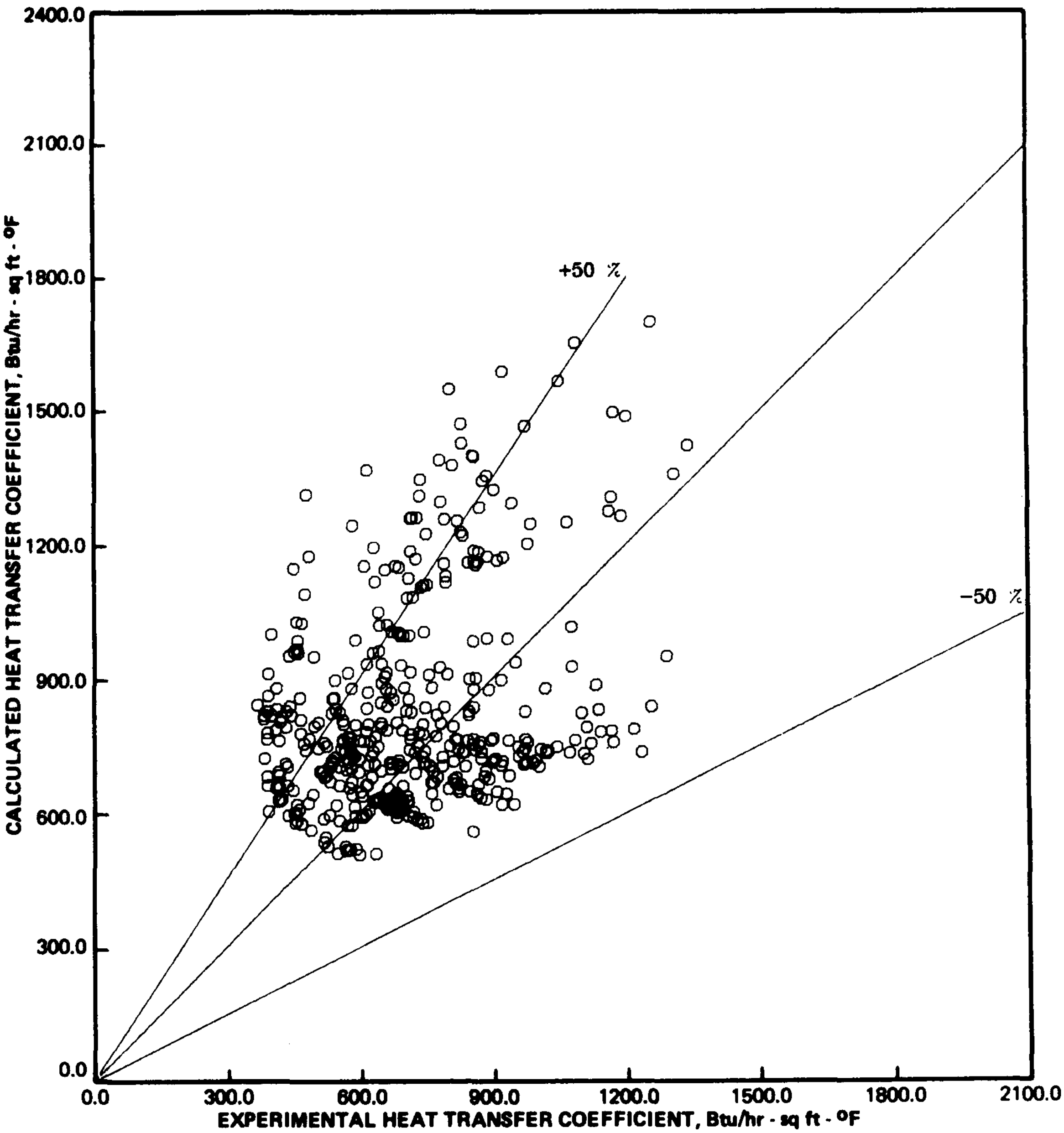

FIGURE IV-79. COMPARISON OF EXPERIMENTAL DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 150°F.

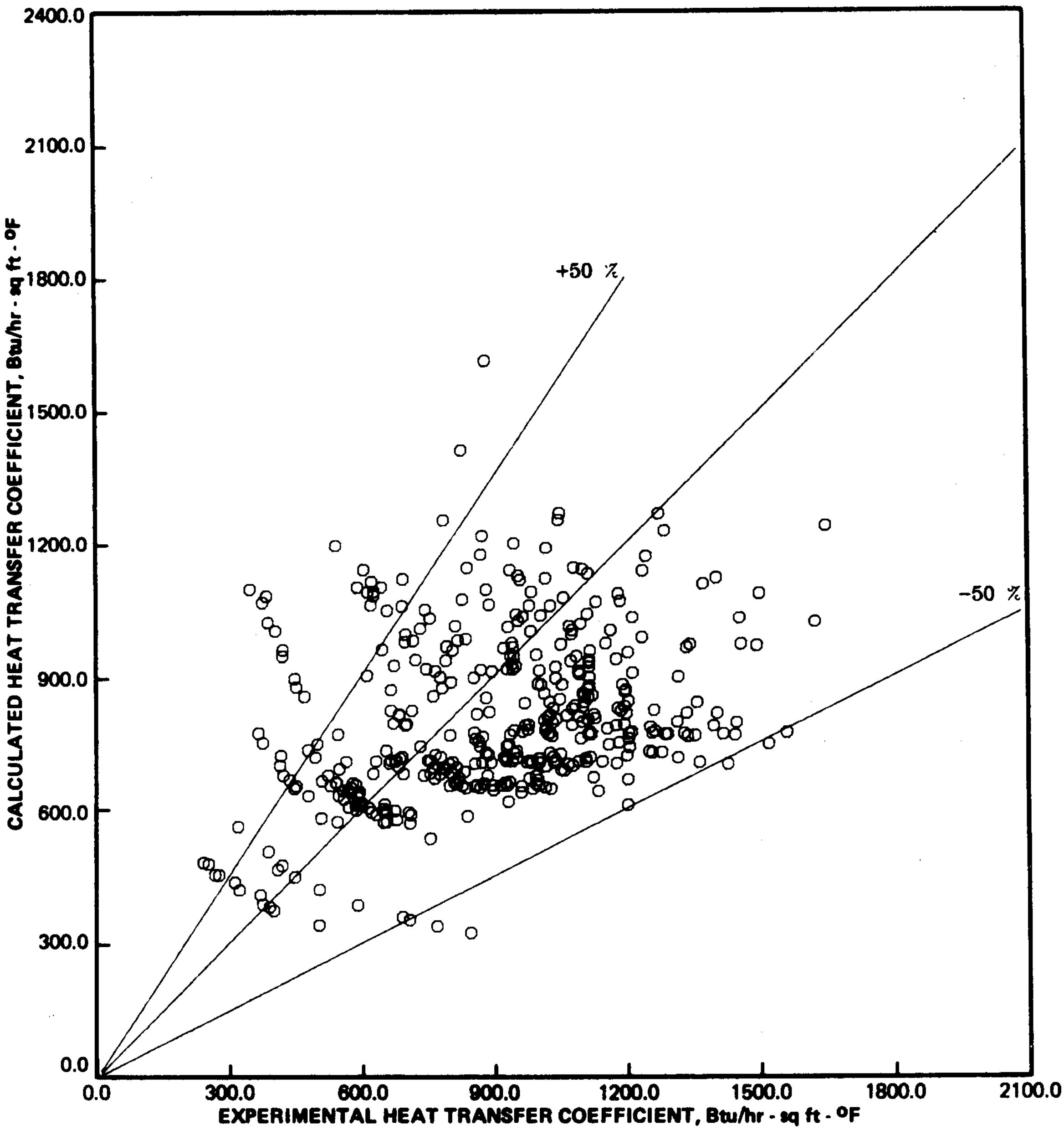

FIGURE IV-80. COMPARISON OF EXPERIMENTAL DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 200°F.

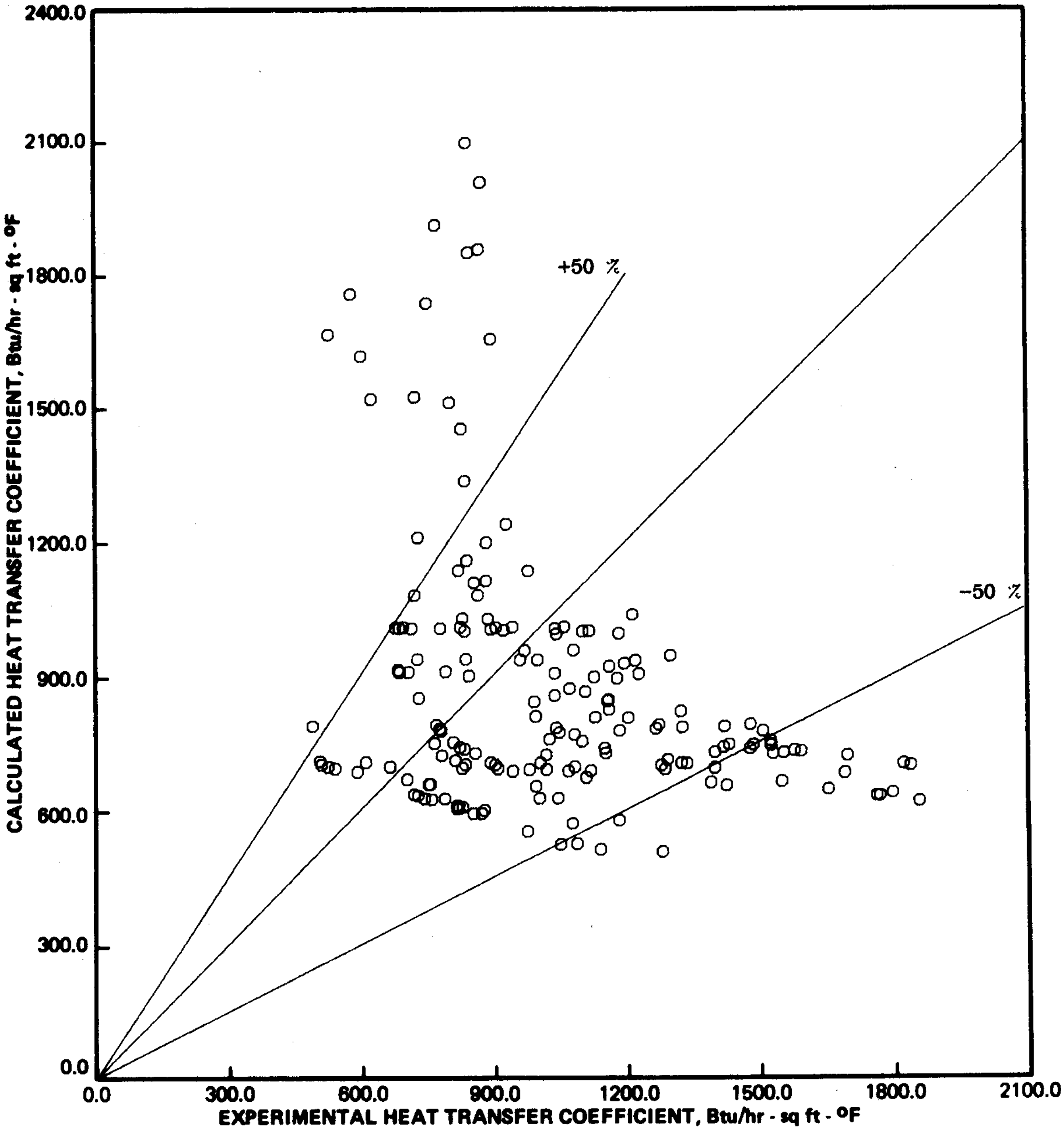

FIGURE IV-81. COMPARISON OF EXPERIMENTAL DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 250°F.

virtue of the smaller magnitude of the experimentally measured heat transfer coefficients. A correlation was then attempted which utilized this observation. The data obtained at mass fluxes between 0.12 and 0.24 x 10^6 lbm/hr-sq ft were compared to the original Chen [43] correlation and those obtained at the higher mass fluxes were compared to the Jallouk [42] modification. This was not an arbitrary decision but was based on a judicious examination of all data with particular emphasis on the results of the flow regime studies which were discussed earlier. At the low mass fluxes the stratified-smooth and stratified-wavy flow regimes were predominant. One would intuitively expect the heat transfer coefficients measured under these conditions to be smaller than those measured under the intermittent and annular flow regimes. This was noticed by Chaddock and Noerager [53] during their studies with R-12. They attributed the lower coefficients to the larger heat transfer resistance caused by the presence of the vapor phase at the top of their test section. The flow regime studies conducted as a part of the present investigation would not seem to verify Chaddock and Noerager's [53] conclusions. As stated earlier, during many instances when stratified flow was observed in the flow visualization section, a thin liquid film could also be observed at the top and sides. Thus, it is likely that the smaller heat transfer coefficient measurements at the low flow conditions were due to, perhaps, the smaller magnitude of the two-phase convective components of the total coefficients as well as other effects such as reduced vapor-liquid exchange and increased suppression.

The additive correlations of both Chen [43] and Jallouk [42] should be applicable to such situations so the Chen [43] correlation was compared to the low mass flux data since it typically gave lower calculated coefficients than Jallouk's [42] modification. By the same token Jallouk's [42] modification was compared to the medium-to-high mass flux data.

The result of the comparisons can be found in Figures IV-82 to IV-89. As the figures show, agreement between experimental data and correlation were vastly improved. In most cases, over 80% of the experimental data were predicted to within ±30%. Similar results may have been obtained by remodeling the F-factor from Jallouk's [42] work. Such an attempt was initiated but not completed.

At the 100°F saturation temperature, the original Chen [43] correlation did have a tendency to overestimate some of the experimental data. This was especially true of data obtained at low mass and heat flux conditions (0.12 x 10^6 lbm/hr-sq ft and 5,000 - 8,000 Btu/hr-sq ft). Moreover, Jallouk's [42] modification tended to break down at a saturation temperature of 250°F under high heat flux (11,000 - 16,000 Btu/hr-sq ft) conditions. The correlation tended to predict much lower coefficients than those measured, as shown by Figure IV-89. Potential experimental error could be significant in these latter cases since the measured temperature differentials were quite small (∿3°F). An indication of the experimental error associated with the data is shown on some of the figures.

In conclusion, no single existing correlation proved to be completely satisfactory in predicting the large quantity of experimental data obtained during this investigation. Good results were, however, obtained by applying the original Chen [43] correlation given by Equation (2-26), which is repeated below; to the low mass flux data.

$$h_{tp} = 0.023 \left(N_{Re}\right)_{fa}^{0.8} \left(N_{Pr}\right)_{f}^{0.4} \left(\frac{k_f}{D}\right)$$

$$+ 0.174 \left(\frac{k_f^{0.79}\, c_{pf}^{0.45}\, \rho_f^{0.49}}{h_{fg}^{0.24}\, \rho_g^{0.24}\, \sigma^{0.5}\, \mu_f^{0.29}}\right) \left(\Delta T_s\right)^{0.24} \left(\Delta p_s\right)^{0.75} S \quad \text{(2-26}$$

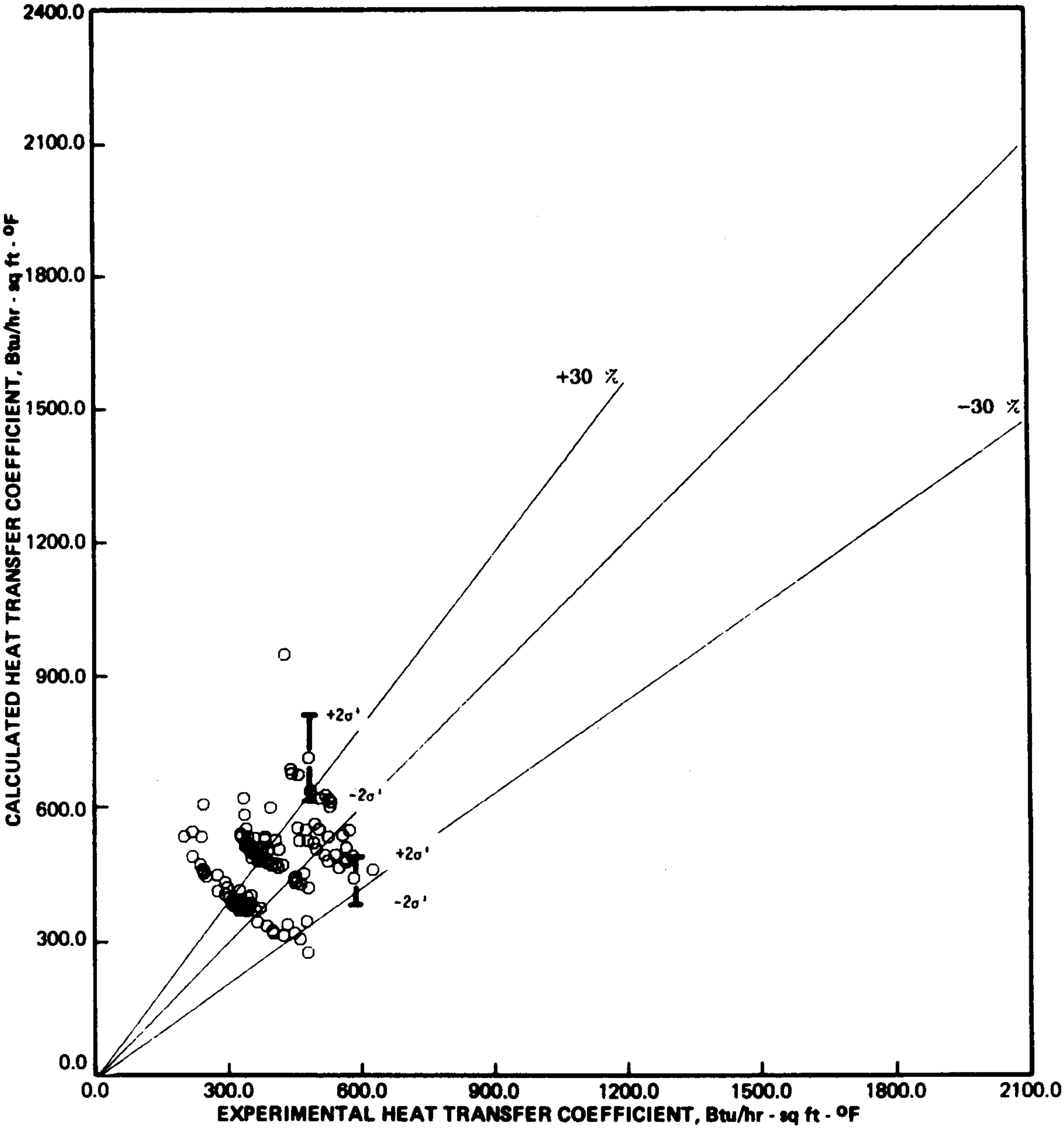

FIGURE IV-82. COMPARISON OF LOW MASS FLUX DATA WITH CHEN [43] CORRELATION, 100°F.

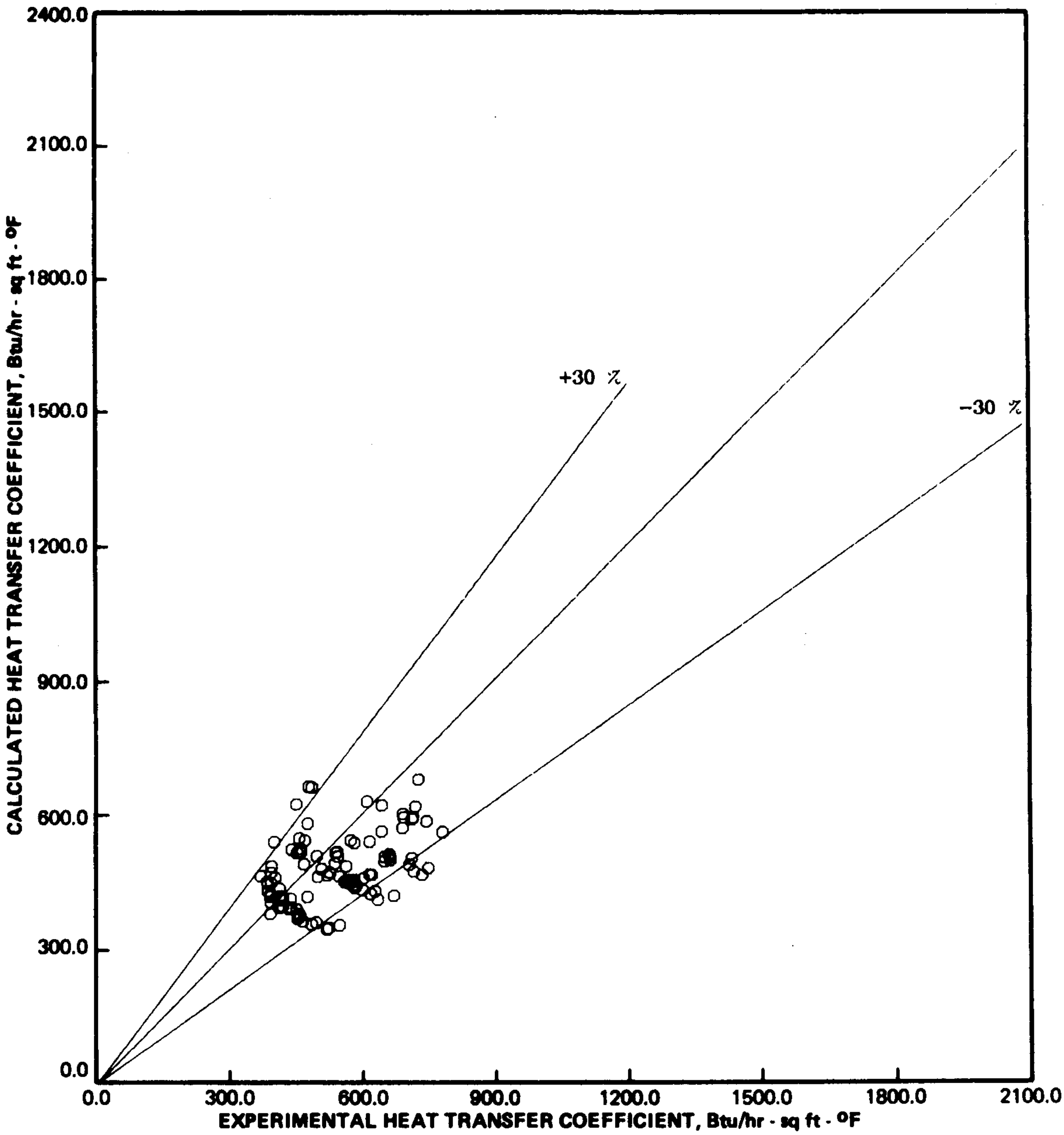

FIGURE IV-83. COMPARISON OF LOW MASS FLUX DATA WITH CHEN [43] CORRELATION, 150°F.

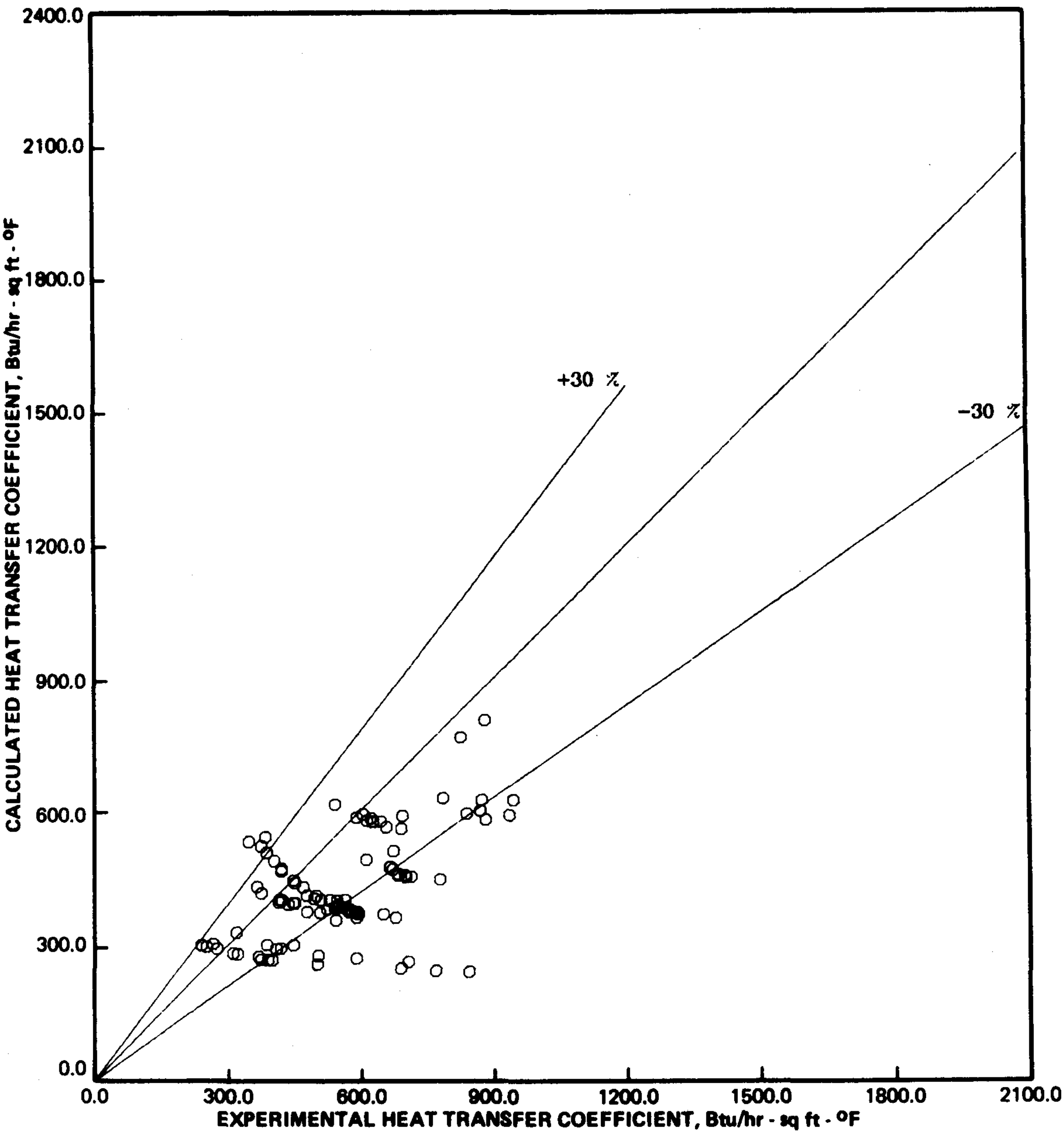

FIGURE IV-84. COMPARISON OF LOW MASS FLUX DATA WITH CHEN [43] CORRELATION, 200°F.

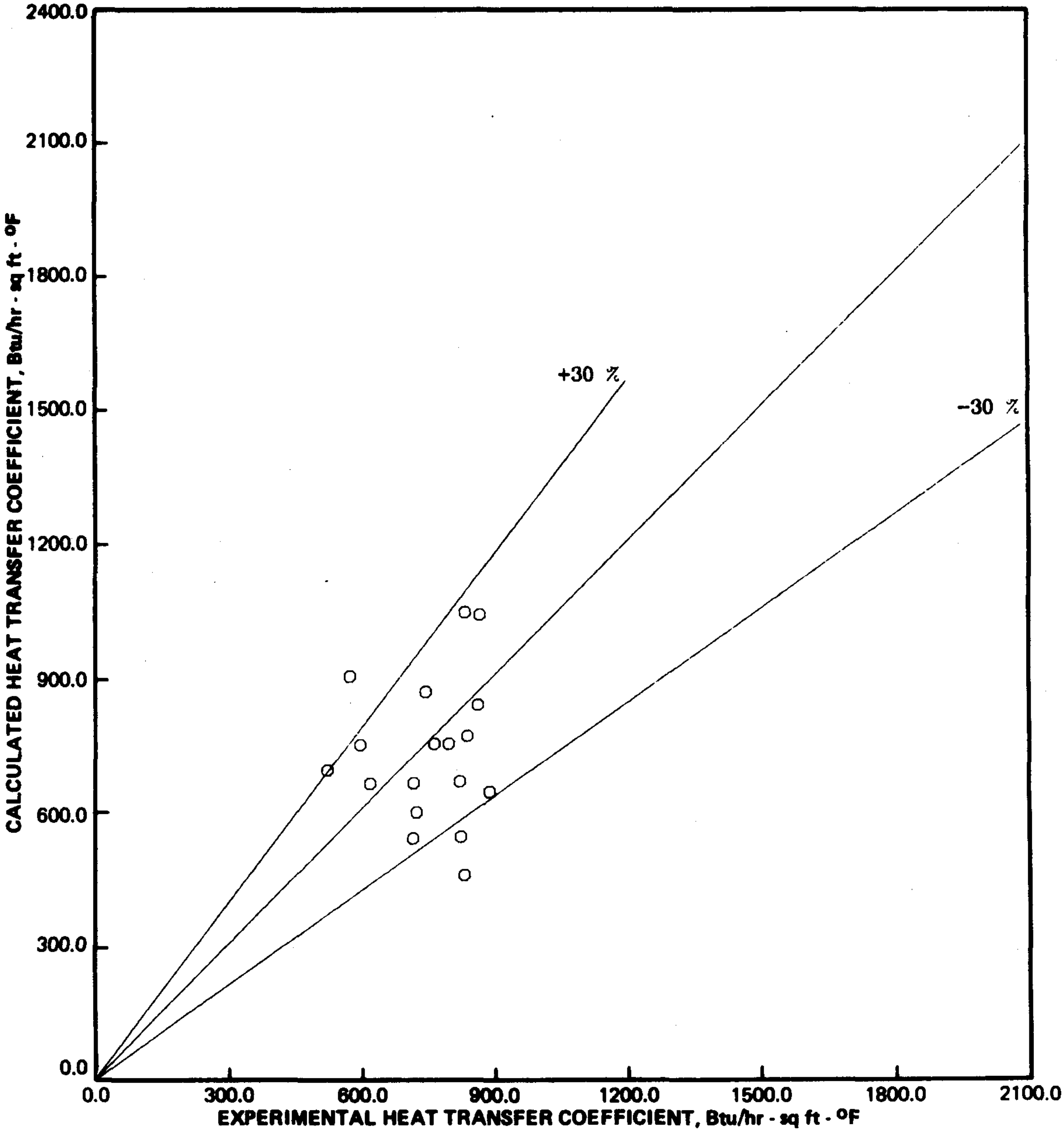

FIGURE IV-85. COMPARISON OF LOW MASS FLUX DATA WITH CHEN [43] CORRELATION, 250°F.

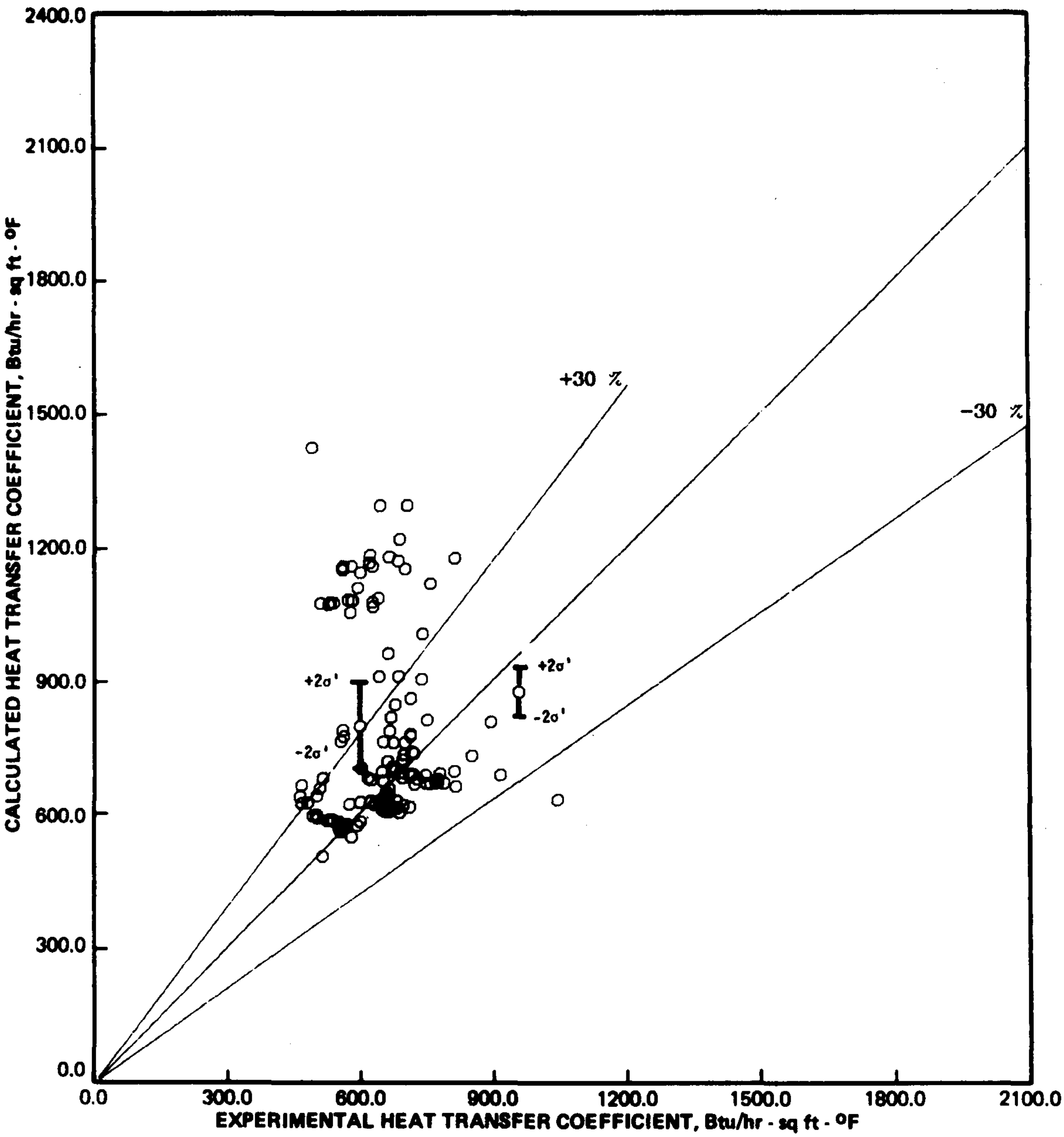

FIGURE IV-86. COMPARISON OF MEDIUM-TO-HIGH MASS FLUX DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 100°F.

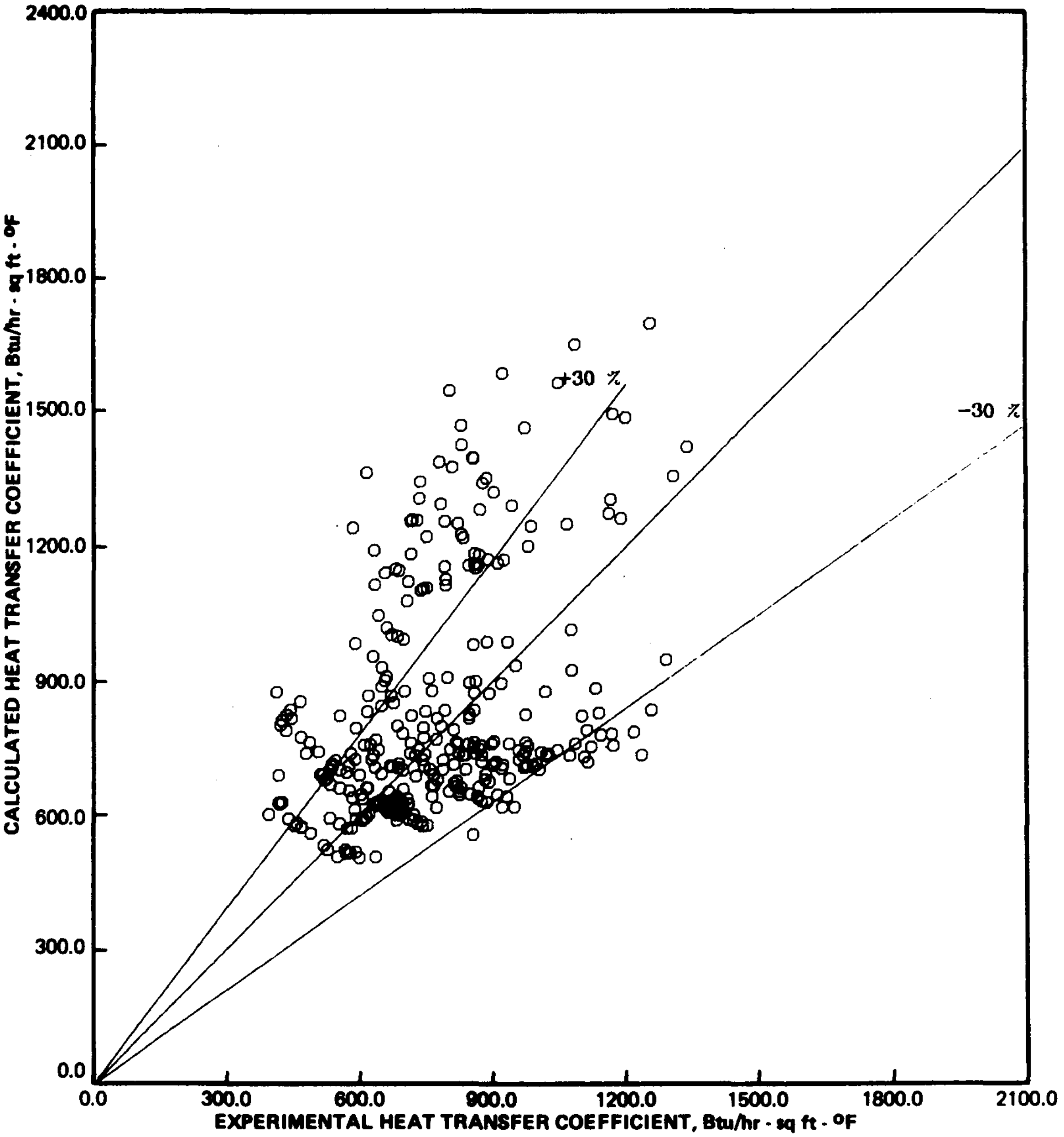

FIGURE IV-87. COMPARISON OF MEDIUM-TO-HIGH MASS FLUX DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 150°F.

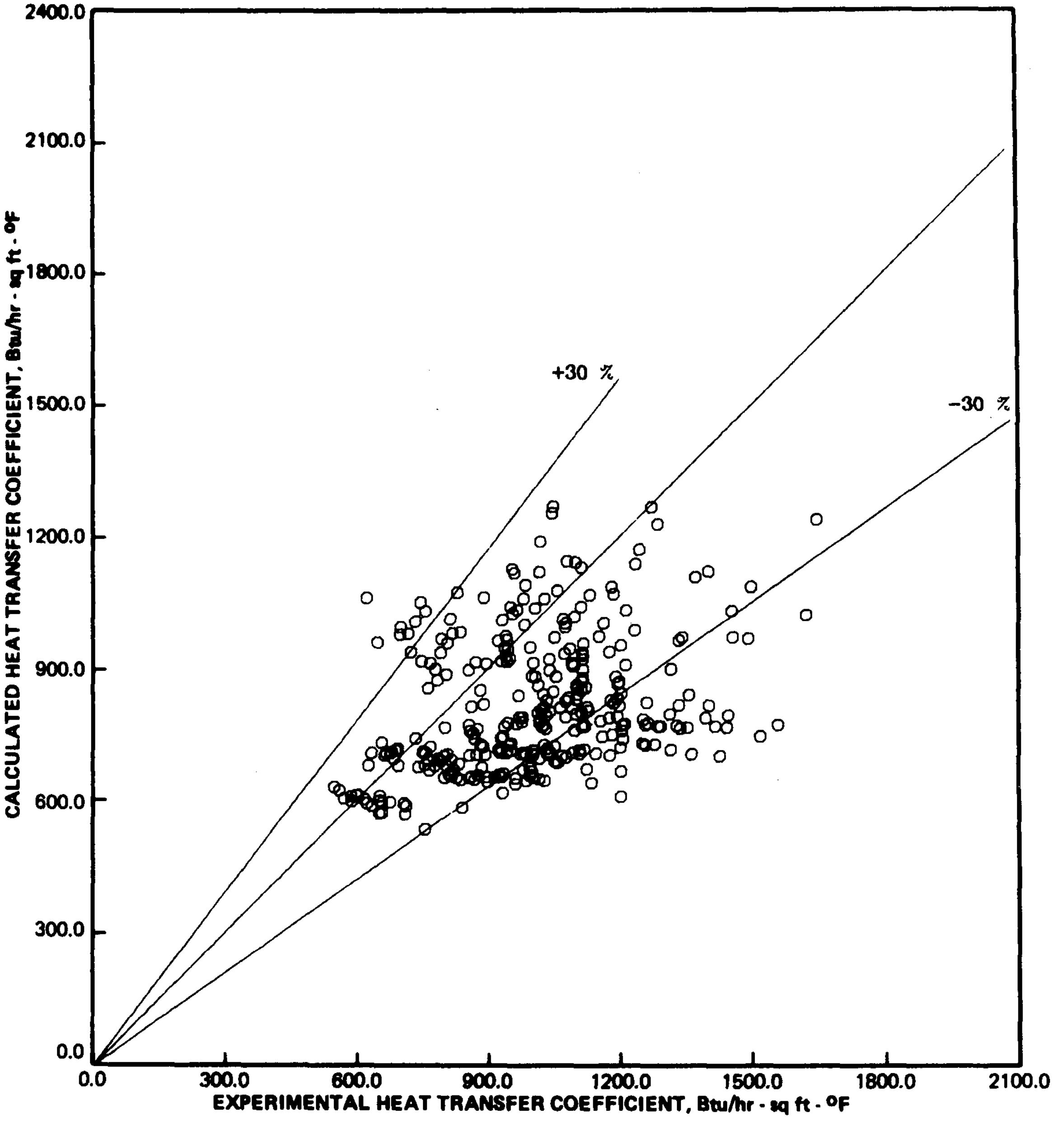

FIGURE IV-88. COMPARISON OF MEDIUM-TO-HIGH MASS FLUX DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 200°F.

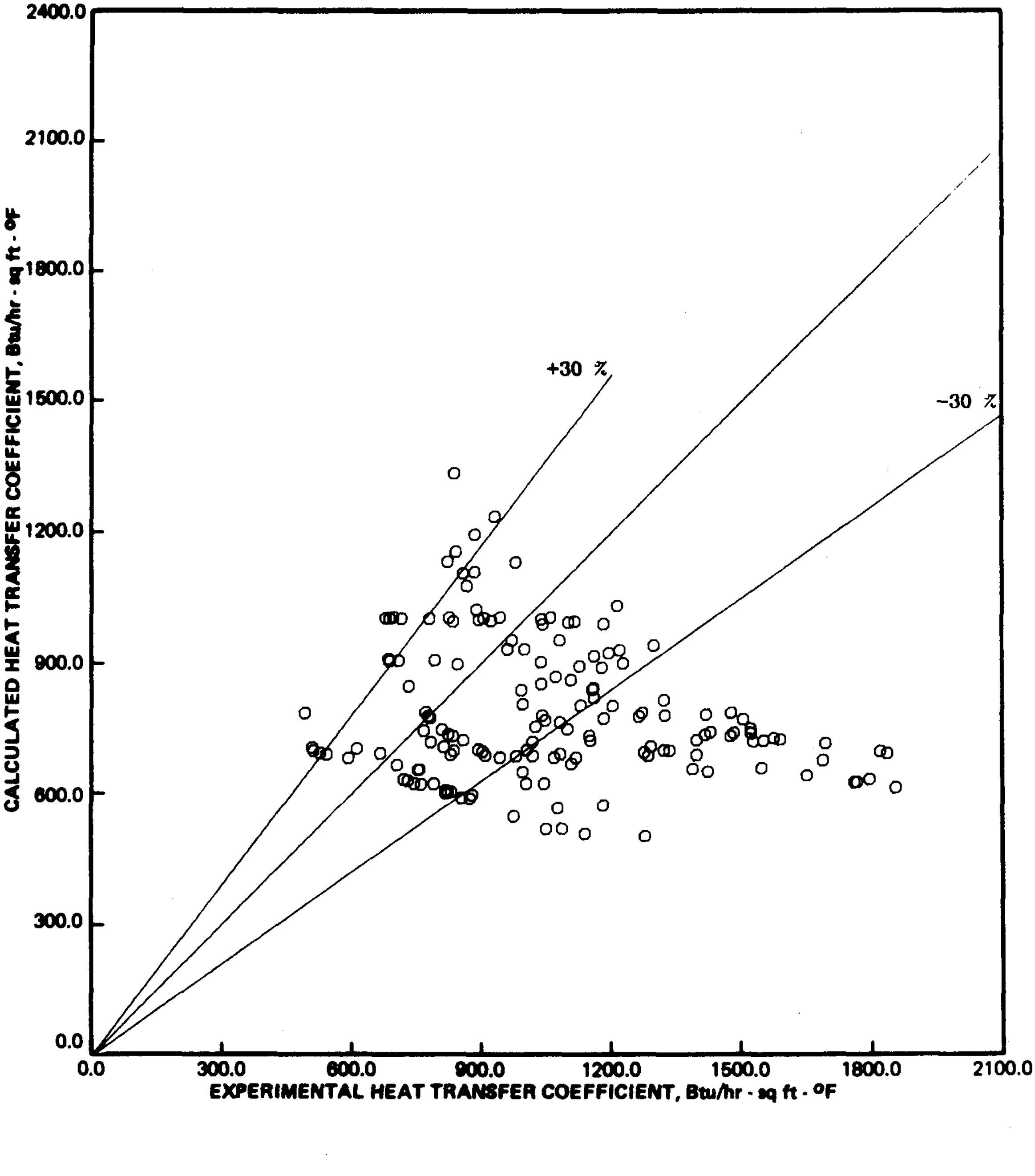

FIGURE IV-89. COMPARISON OF MEDIUM-TO-HIGH MASS FLUX DATA WITH JALLOUK'S [42] MODIFICATION TO CHEN'S [43] CORRELATION, 250°F.

Furthermore, the remaining data were well predicted by Jallouk's [42] modification to Chen's [43] correlation which is also repeated below.

$$h_{tp} = 0.023 \left(N_{Re}\right)_{fa}^{0.8} \left(N_{Pr}\right)_{f}^{0.4} \left(\frac{k_f}{D}\right) F$$

$$+ 0.3802 \left(\frac{k_f^{0.79} \, c_{pf}^{0.45} \, \rho_f^{0.49}}{h_{fg}^{0.24} \, \rho_g^{0.24} \, \sigma^{0.5} \, \mu_f^{0.29}}\right) \left(\Delta T_s\right)^{0.24} \left(\Delta p_s\right)^{0.75} S \quad (2\text{-}45)$$

where

$$S = -0.2074 \left(p_r\right)^{-0.0899} \ln \left(N_{Re_{fa}} F^{1.25}\right)$$

$$+ 2.85568 \left(p_r\right)^{-0.06203} \quad (2\text{-}47)$$

and

$$F = \left(\Phi_{ftt}\right)^{0.89}. \quad (2\text{-}48)$$

Generally speaking, approximately 80% of the experimental data were predicted within ±30% by the combination of these two correlations.

CHAPTER V

SUMMARY, CONCLUSIONS, AND RECOMMENDATIONS

Summary and Conclusions

The two-phase flow characteristics of R-114 in a horizontal tube have been studied extensively. Both analytic and experimental techniques have been employed in order to better understand the boiling characteristics of this fluid. By utilizing the information and equations presented herein, engineers should be able to determine the flow regime, pressure loss, and heat transfer coefficient for R-114 at conditions similar to those studied. The conclusions obtained from the three areas studied will be discussed below.

Flow Regimes. The flow regime data indicated that the intermittent flow pattern occurred most often while bubbly flow rarely occurred. Moreover, as mass flux increased at the same saturation temperature, the sizes of the intermittent and stratified flow regimes were reduced. In addition, as the saturation temperature was increased at the same mass flux, the stratified flow regime tended to diminish. A significant result of the flow regime studies was the detection of a thin liquid film at the top and sides of the flow visualization section on many occasions when stratified-smooth or stratified-wavy flow apparently existed.

Accurate flow regime prediction was obtained by combining Quandt's [9] analysis and Knowles' [10, 11] predictive procedures. The Quandt [9] analysis accurately predicted the existence of annular flow while Knowles' [10, 11] method was accurate for the stratified and intermittent regimes.

Pressure Drop. Four existing correlations provided reasonably accurate results when compared with the experimental data. One of these correlations, that proposed by Baroczy [24], unfortunately, would be rather difficult to computerize. The Martinelli-Nelson [1] correlation gave fair-to-good results for all combinations of mass flux and saturation temperature. Two correlations proposed by Chisholm [29, 30] also agreed fairly well with the experimental data. These two correlations go beyond the Martinelli-Nelson [1] model since they include a mass flux effect and account for momentum transfer between the two phases. Consequently, either of these two correlations should provide the most reliable prediction of two-phase frictional pressure loss.

Heat Transfer Coefficients. No existing correlation proved to be satisfactory in predicting the large quantity of heat transfer data obtained. Good results were obtained by applying the original Chen [43] correlation given by Equation (2-26), Page 33 to the low mass flux data (0.24×10^6 lbm/hr-sq ft and below). The remaining data were well predicted by Jallouk's [42] modification to Chen's [43] correlation, Equations (2-45), Page 71, (2-47), Page 72, (2-48), Page 72. Approximately 80% of the experimental data were predicted within ±30% by the combination of these two correlations.

Recommendations

Based on the results of this study, the following recommendations are made:

(1) Additional void fraction measurements leading to a correlation for R-114 are badly needed. From the standpoint of horizontal flow, a correlation similar to the one proposed by Eaton in Reference [10] could be a useful starting point.

(2) An extensive analysis of the refrigerant, two-phase heat transfer coefficient data available in the literature using the equations proposed in this study should be performed. An effort of this type could be extended by including the data of this investigation and attempting to remodel Jallouk's [42] F-factor to more closely represent the low mass flow conditions. This would require an improved void fraction relationship and could conceivably result in a more accurate suppression factor determination at these conditions.

(3) An extensive study of refrigerant pool boiling from a variety of surfaces with different finishes should be conducted. This would provide a better understanding of refrigerant pool boiling as well as allow verification of some of the more recent pool boiling correlations.

(4) A comprehensive study of flow boiling, critical heat flux in a horizontal tube with R-114 as the working fluid, is also needed. When combined with the studies of Jallouk [42], the CHF phenomenon of R-114 would be more completely understood.

(5) Additional pressure drop and heat transfer studies at a saturation temperature of 250°F are needed. A facility especially designed for these conditions may be required.

ACKNOWLEDGMENTS

This investigation was performed at the Oak Ridge Gaseous Diffusion Plant which is operated by Union Carbide Corporation for the Energy Research and Development Administration. Consequently, the author wishes to express appreciation to both Union Carbide Corporation and to the Energy Research and Development Administration for the use of the facilities and equipment and for the assistance of several of the staff members. The author is also grateful to Dr. Julian E. Mott who served as major professor. His assistance contributed to the success of this project. Furthermore, Dr. Philip A. Jallouk, a former co-worker, provided much needed advice and suggestions. Senior laboratory technician, Ulys C. Fulmer, served as chief operator of the Experimental Heat Transfer Facility and acquired the majority of the experimental data. Senior technicians, William D. Ayers and Richard P. Miller, Jr., assisted the author in tabulating and graphing the experimental data in addition to performing other important tasks. The support of Dr. George J. Kidd and Mr. John M. Vance of the Special Projects Department, Oak Ridge Gaseous Diffusion Plant is also appreciated. In addition, the author would like to express appreciation to his wife Terrye for her encouragement and understanding. Finally, a special word of gratitude is extended to the Technical Editing Group of the Oak Ridge Gaseous Diffusion Plant for preparing the manuscript.

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APPENDIX A

A DESCRIPTION OF THE EXPERIMENTAL HEAT TRANSFER FACILITY AND THE ASSOCIATED INSTRUMENTATION

APPENDIX A

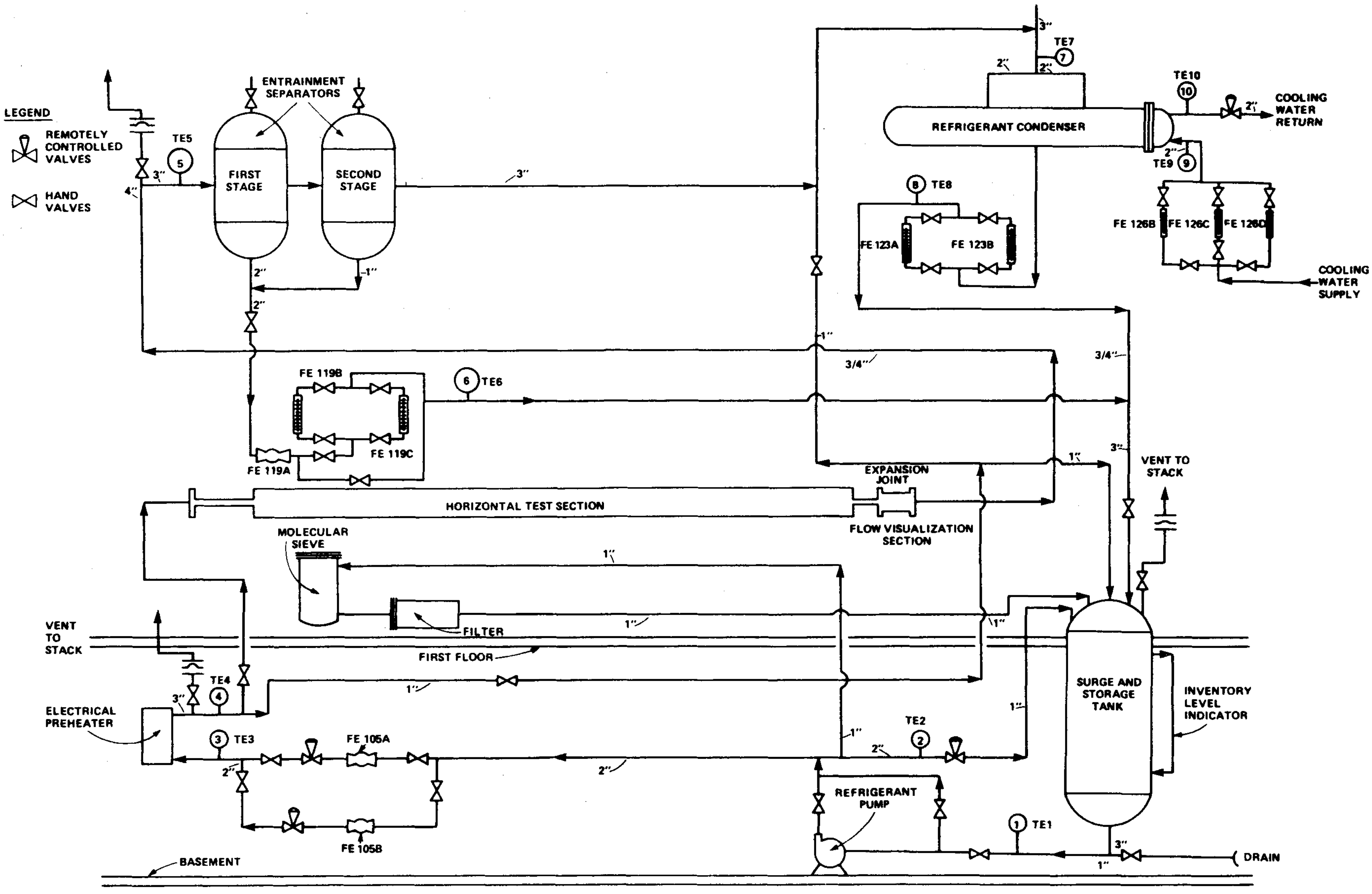
A DESCRIPTION OF THE EXPERIMENTAL HEAT TRANSFER FACILITY AND THE ASSOCIATED INSTRUMENTATION

A. THE EXPERIMENTAL HEAT TRANSFER FACILITY

The Experimental Heat Transfer Facility (EHTF), where the experimental work discussed herein was conducted, is located in Building K-1401 at the Oak Ridge Gaseous Diffusion Plant. A schematic drawing of this facility is shown in Figure A-1. All of the system piping and other pertinent equipment were designed to operate at pressures up to 300 psig and temperatures up to 250°F.

As Figure A-1 shows, during operation in the forced circulation mode, the working fluid, Refrigerant-114 (R-114), left the large storage tank and was forced, by means of an Eastern Industries canned rotor pump, to travel through one of two turbine flowmeters. One flowmeter, designated FE 105B, was used to measure small flow rates in the 0.5- to 6.0-gpm range. The other flowmeter, designated FE 105A, was used to measure flow rates in the 3.0- to 90.0-gpm range.

After leaving the appropriate flowmeter, the fluid entered a 30-kw preheater which was used to heat the fluid to either the desired saturation temperature or to a desired degree of subcooling. After exiting the preheater, the R-114 traveled through 3-in.-dia pipe until entering a reducer section which was followed by a 2-ft-long flow straightener which was 1 in. in diameter. At this point the flow entered the 10-ft horizontal test section where boiling was initiated. A flow visualization section was located at the outlet of the test



URE A-1. SCHEMATIC OF THE EXPERIMENTAL HEA RANSFER FACILITY.

section, and, after leaving this section, the fluid was sent to two entrainment separators. Since the internal diameters of the flow visualization section and the test section were very nearly equal, the flow was not appreciably disturbed before entering the flow visualization section. After the liquid and vapor were separated by the two entrainment separators in series, the flow rate of the liquid fraction was measured by either a turbine flowmeter (range 3.0 to 90.0 gpm) or one of two rotameters (ranges 0 to 1.12 gpm and 0 to 4.33 gpm). The vapor fraction was transported through the condenser, and the condensate flow rate was measured by either or both of two rotameters (ranges 0 to 1.12 gpm and 0 to 4.33 gpm). Both of these streams were then returned to the storage tank.

During forced circulation operation it was always necessary to reroute a portion of the flow produced by the single-speed, canned rotor pump. This was accomplished by a bypass line and the associated valves which sent the desired portion of the total flow back to the storage tank. In addition, a certain flow (a few gpm) was continuously passed through a molecular sieve and filters to remove the small amounts of moisture and iron oxide that might accumulate in the working fluid. Additional details pertaining to the facility can be obtained from Figure A-1. Moreover, Table A-1 should be consulted for specific information concerning the equipment and instrumentation used.

Operation in the natural circulation mode was not very different from the above description. The canned rotor pump was entirely bypassed, and no flow was sent through either the filters or the molecular sieve. Only low- to mid-range flows could be obtained because of limited head availability.

Table A-1

LISTING OF ALL MAJOR COMPONENTS IN THE EHTF

Instrument	Description
Pump	Eastern Industries, canned rotor, Mod. No. A1LG; 300-psig system pressure; 200°F operating temperature; 15 hp, 3-phase, 60 cycle; 85 gpm.
FE 105A	Cox Instruments, turbine Type 20 Mod. I-1 1/4-600; range 3.0 to 90.0 gpm ± 0.5%; internal body Type 300 SS, Type 416 SS rotor.
FE 105B	Same as FE 105A except range 0.5 to 6.0 gpm; Type 20 Mod. I-1/2(8-6)-600.
FE 119A	Same as FE 105A.
FE 119B	Fischer & Porter Co.; Rotameter Model 10A1735; Range 0-16.4 lpm; M.A.W.P. 235 psig at 100°F.
FE 119C	Fischer & Porter Co.; Rotameter Model 10A1735Z; Range 0-4.25 lpm; M.A.W.P. 290 psig at 100°F.
FE 123A	Same as FE 119B.
FE 123B	Same as FE 119C.
Condenser	Struthers Wells Corp., Type 12-6U14-6H; shell M.A.W.P. 300 psig, tube M.A.W.P. 150 psig; shell and tube maximum temperature 250°F.
Separators	Centrifix, Div. of Burgess Ind., Type V, M.A.W.P. 300 psig, maximum temperature 250°F.
Remote Control Valves	Skinner Uniflow Valve Div., Type 1020, for service at 200°F, 300 psig, $c_v = 12$ and $c_v = 80$.
Wattmeter	General Electric Watt-Hour Meter Standard, Type 13-10, Mod. AAA1; 120-240V, 1-5-12.5-50 amps.

Table A-1 (Continued)

Instrument	Description
Filter	Mueller Brass Co., Drymaster Micro-Guard Filter-Drier Mod. SD 14411.
Molecular Sieve	Linde Corp., Type 4A beads.
Pressure Transmitters	Foxboro Co., Mod. M-11GM, M.A.W.P. 750 psig; range 0-300 psi.
Preheater	E. L. Wiegand Co., Mod. GCH 83001; 30 kW, 480 V, 3-phase, M.A.W.P. 300 psig at 250°F.
FE 126B	Fischer & Porter Co., Rotameter Model 10A3665A; Range 0-17.6 gpm; M.A.W.P. 130 psig at 100°F.
FE 126C	Brooks Instrument Div.; Rotameter type 111009H3B1A; Range 0-3.2 gpm; M.A.W.P. 240 psig at 200°F.
FE 126D	Same as FE 119C.
SCR Unit	Robicon Corp.; SCR Type, 6KVA, 120 V, single-phase, zero voltage firing.
Flow Visualization Section	The Johnson Corp., regular type for 1 in. pipe 0-150 psig; modified for 0.785-in. ID and 300 psig.
L&N Controller	Leeds & Northrup Co.; Mod. 6432-6-40XX-200-0 20-206; 0-300°F range w/Type E thermocouples.
Loop Thermocouples	Conax Corp., Type E, 0.125-in.-OD SS sheath.
Piping	All piping Schedule 40 CRS 1/2 to 4 in.
Ball Valves	Contromatics, 300 series, 1000 psi. maximum, Teflon seats and gaskets.
Pipe Insulation	Hydrous Calcium Silicate, 70-1200°F, 2-in. thick.
Condenser Coolant Flow Controller	Foxboro Co., Model 58; 3-15 psig input; with proportional reset.

B. THE TEST SECTION AND THE ASSOCIATED INSTRUMENTATION

The 10-ft horizontal test section was fully instrumented so that the following quantities could be accurately measured and/or controlled:

1. Heat input per zone,
2. Instream temperature in each zone,
3. Circumferential wall temperature readings in each zone, and
4. Pressure drop at 2-ft increments.

Each of the systems used to detect these quantities is described and discussed in the text that follows. The test section was divided into 10 zones of 1 ft in length, and most of the discussion below applies to a single zone.

Heat Input System

The heat input to each of the 10 zones in the test section was accomplished through the use of Chromalox heaters. Each heater was capable of generating up to 1,120 watts. In each zone, 3 such heaters were connected in parallel and helically wrapped around the 1-ft length of copper tubing. Consequently, the heat flux capability of each zone was 3,360 watts, or about 55,800 Btu/hr-sq ft. The heaters were also able to withstand temperatures up to 1,200°F continuously. Thermon heat-conducting cement was used to cover the heaters, and approximately 2 in. of high-temperature pipe insulation was wrapped around the test section, including the Thermon.

A somewhat sophisticated control system was employed in conjunction with the Chromalox heaters, in order to achieve a uniform heat flux throughout the test section. A schematic diagram of the system is shown in Figure A-2. As the figure shows, adjustable potentiometers were used to send a 0- to 5-ma signal to each zero-firing Silicon Controlled Rectifier (SCR). The SCR, when activated by the 0- to 5-ma signal, fired on and off rapidly, allowing current to flow to the heaters. By adjusting the potentiometers and reading the watt-hour meter, any uniform heat flux up to about 55,800 Btu/hr-sq ft could be obtained.

Pressure Drop Measuring System

Although the pressure drop measuring system was discussed in a previous section, for completeness, it is also dealt with here. The system is shown schematically in Figure A-3. A total of 6 pressure taps were connected to the bottom of the test section, as shown in the figure. Thus, it was possible to experimentally measure the pressure drop in 5 different increments along the length of the test section.*

Each pressure tap was carefully constructed so as to minimize possible sources of error in pressure drop readings. Holes with a diameter of 1/16 in. were drilled completely through the test section wall, and 1/8-in.-dia holes were countersunk to a depth of 1/4 in. All holes were drilled perpendicular to the centerline of the tube, and each hole was inspected to make sure that no burrs existed on its inside surface. The holes were sized according to suggestions made by Benedict [60]. Copper tubing with an OD of 1/4 in. was silver-soldered into each of the countersunk holes, making sure that no solder or flux collected in the pressure tap holes. Each tube was bent in such a way as to extend from the test section at a slightly downward angle. This was done to keep noncondensable gases and refrigerant vapors from

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*Due to instrumentation problems, the third tap could not be used.

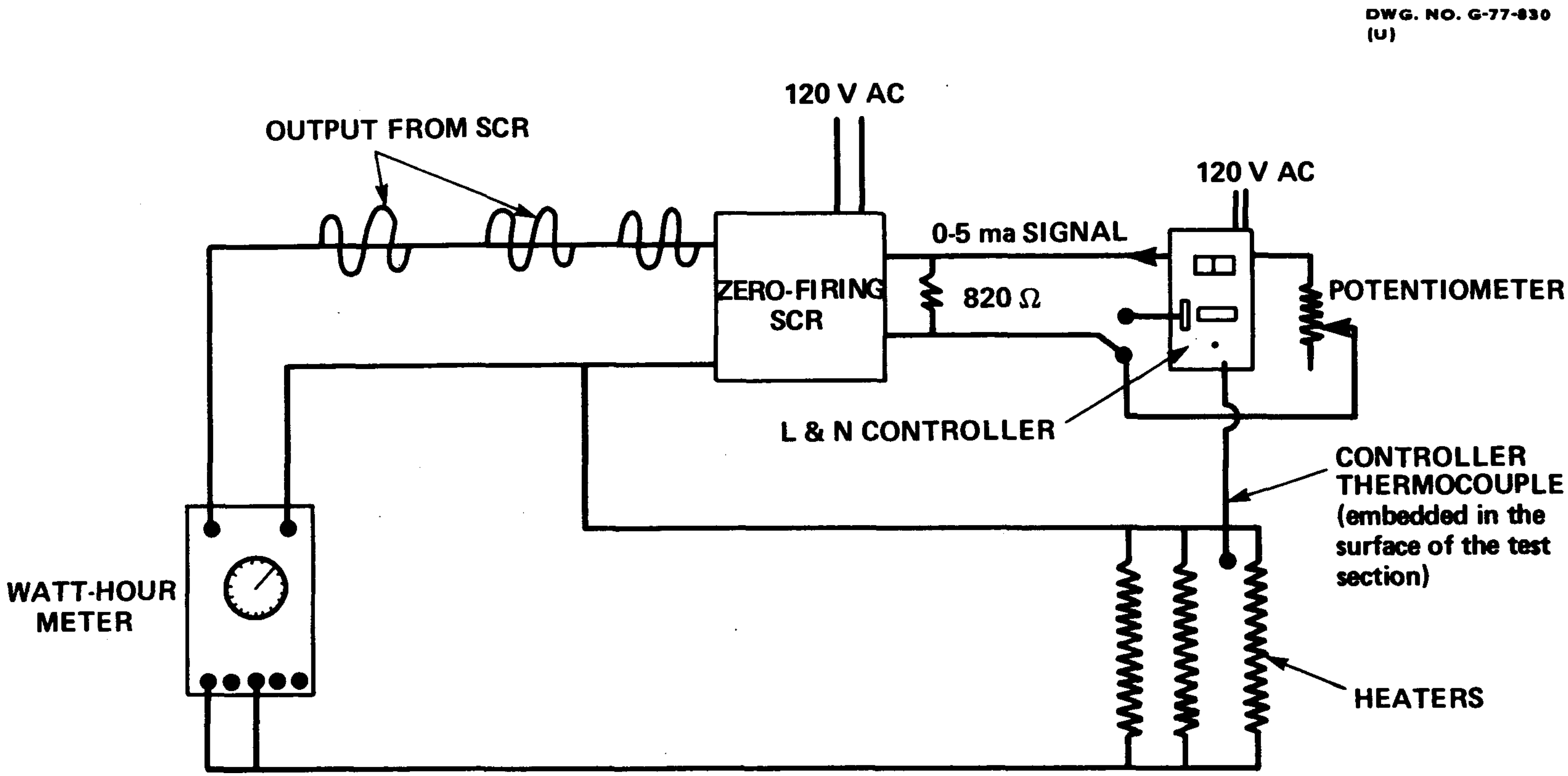

FIGURE A-2. CIRCUIT DIAGRAM FOR HEAT INPUT CONTROL SYSTEM

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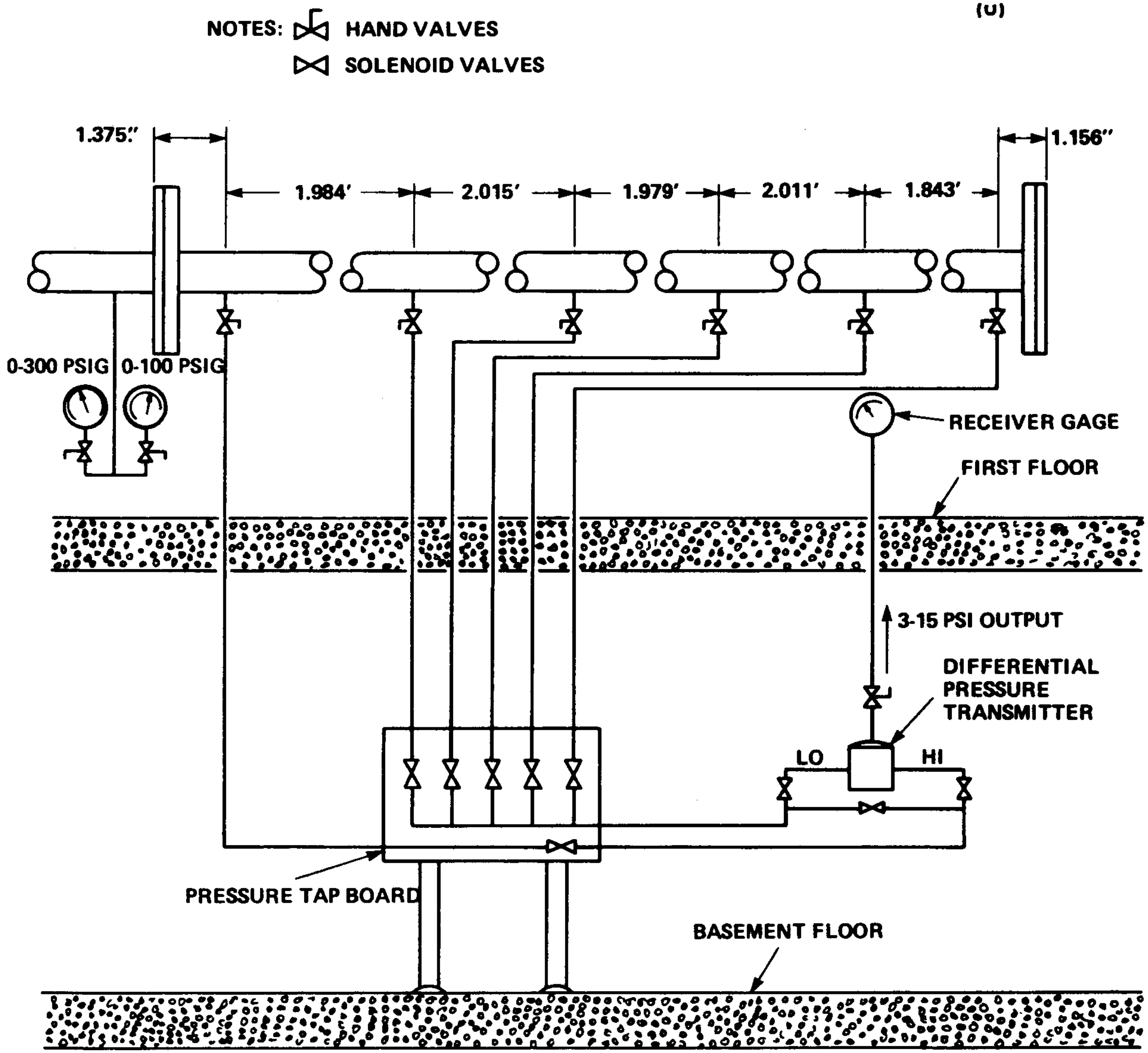

FIGURE A-3. THE PRESSURE DROP MEASURING SYSTEM.

collecting in the pressure tap lines. All pressure tap lines were connected to a board on the lower level of the test facility. This board contained the necessary solenoid valves and associated controls required to obtain the pressure drop between the desired set of taps. Two copper lines connected this board to a Foxboro D/P cell. One line represented the high-side pressure which always came from the first pressure tap in the test section, and the other line represented the low-side pressure which could come from any one of the 5 remaining taps. The calibrated D/P cell detected the difference between these two pressures and had a range of 0- to 100-in. water. The resulting pressure differential was converted to a 3- to 15-psi. signal which was transmitted to a receiver gage that had a range of 0 to 100 divisions.

The gage pressure of the refrigerant as it entered the test section was also measured. This was accomplished by reading one of two Robertshaw Acragages, both of which had been calibrated. One gage had a scale of 0 to 100 psig with a resolution of 0.5 psi., while the other gage had a scale of 0 to 300 psig with a resolution of 1.0 psi. This reading, coupled with the barometric pressure which was also taken, resulted in an accurate determination of the absolute inlet pressure.

Temperature Measuring System

The test section temperature measuring system was quite sophisticated. The reader is referred to Figure A-4 for important details. A total of 80, 24-gage, Chromel-Alumel thermocouples were used to measure tube wall temperatures. As Figure A-4 shows, 8 thermocouples were used in each zone. They were located in groups of 4 at two different cross

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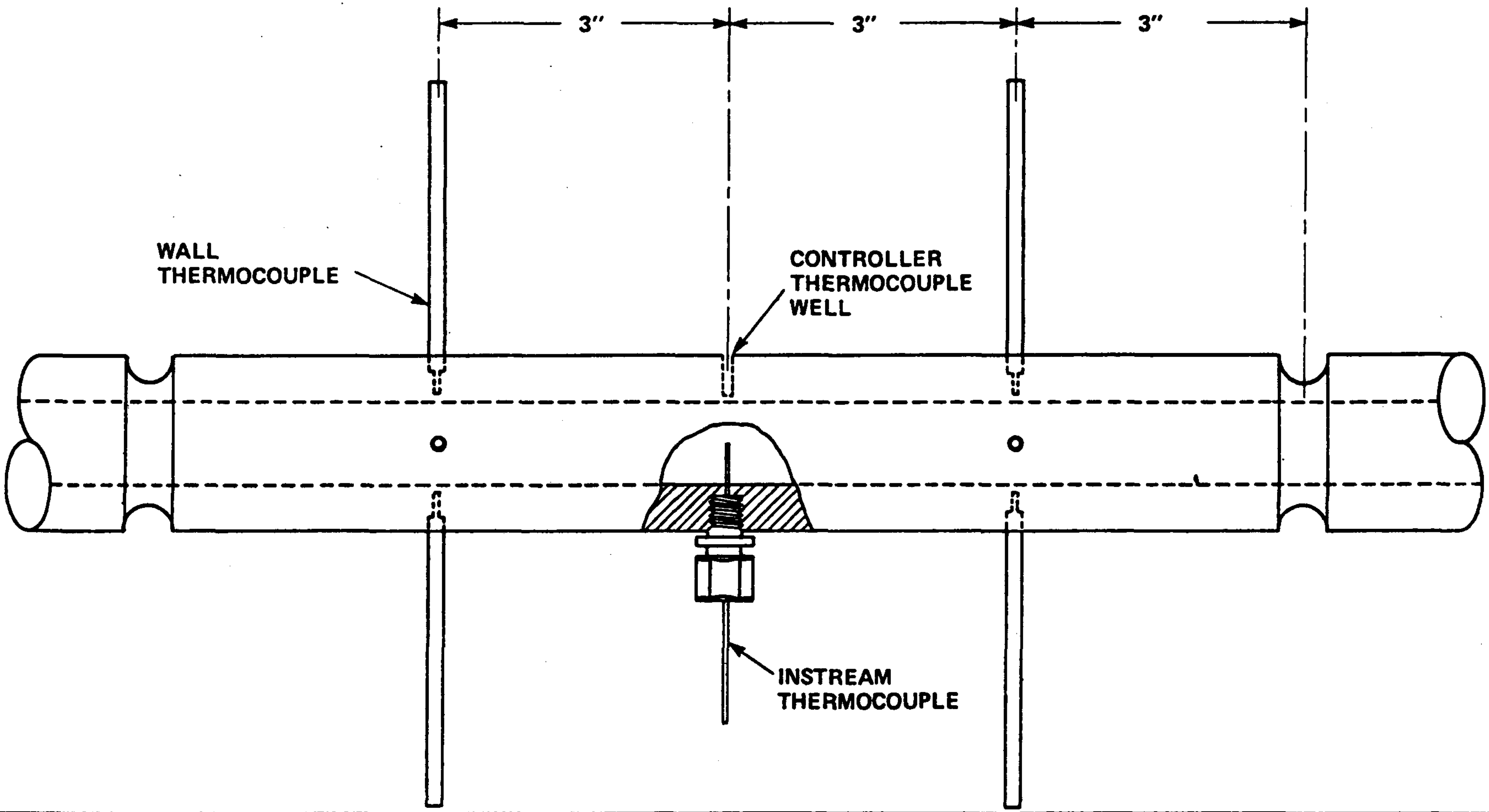

FIGURE A-4. A SCHEMATIC DRAWING SHOWING THERMOCOUPLE LOCATIONS.

sections within each zone. Thus, at each cross section, the wall temperatures at the top, bottom, and both sides were measured. Very little difference was detected between the corresponding wall temperatures in a zone. As expected, at low flows where pure stratification existed, the wall temperatures at the top of the tube were observed to be slightly higher than the temperatures at the bottom of the tube.

The wall thermocouples were attached to the tube wall with a low-temperature silver solder (96.5% Sn and 3.5% Ag) which melted at approximately 430°F. The author devised the following thermocouple attachment procedure which gave quite satisfactory results. After the holes had been drilled to the proper specifications (0.07-in.-dia, 3/8-in.-deep holes with 0.144-in.-dia, 1/8-in.-deep countersink) by an automatic drilling machine, the heaters were wrapped around the tube. The heaters had to be judiciously positioned so that no thermocouple holes were covered (Figure A-5). The test section was then positioned so that 4 of the holes in each zone were in the horizontal position. At this point, a piece of solder was placed at the bottom of each of the 4 holes in Zone 1, and a 10-amp Variac was connected to the heaters in the same zone. While monitoring the wall temperature, current was sent from the Variac to the heaters until the solder was observed to melt. The test section was never heated to temperatures in excess of 500°F, for fear of doing permanent damage to the copper. All 4 thermocouples were placed in the appropriate holes after the solder had melted. They were then taped in place so that their beads always touched the bottom of the thermocouple holes which were about 0.10 in. from the inside tube wall. The same procedure was repeated on Zone 2 and all remaining zones.

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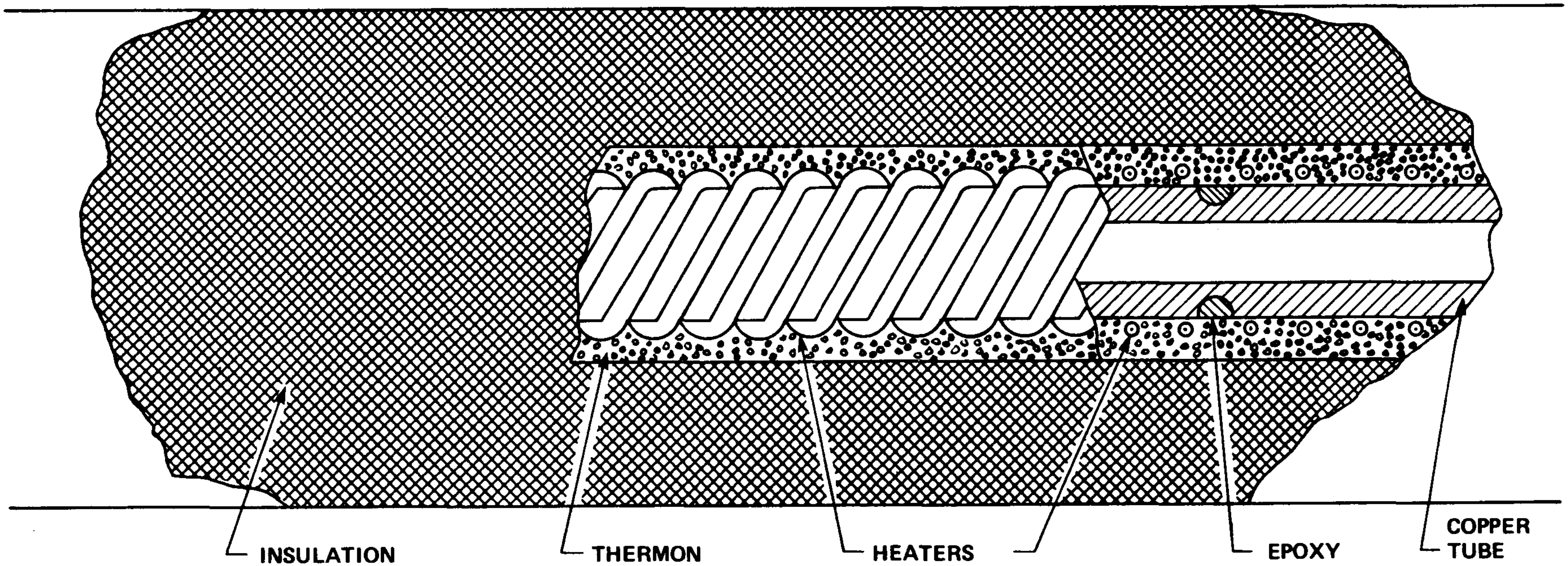

FIGURE A-5. SCHEMATIC SHOWING THE CHROMALOX HEATERS, INSULATION, ETC.

The test section was then rotated 90° and the above procedure was repeated.

The horizontal test section also contained 10 instream thermocouples which were used to measure the refrigerant temperature at the midpoint of each zone. These sheathed thermocouples of the Chromel-Alumel type were attached to the tube by stainless steel fittings with Teflon inserts, as shown in Figure A-4 on Page 181. They were inserted into the tube until the tip of each sheath was very near the tube centerline.

Ten control thermocouples were also installed on the test section wall. They were peened into the copper surface at a depth of 3/8 in. These thermocouples were continuously monitored through the L and N controllers to avoid possible excessive wall temperatures.

The 80 wall thermocouples and 10 instream thermocouples were connected to a Vidar 521 Integrating Digital Voltmeter (IDVM) using a zone box-type circuit [60]. A listing of all components and instrumentation associated with the test section is given in Table A-2.

Table A-2

LISTING OF ALL THE COMPONENTS AND INSTRUMENTATION ASSOCIATED WITH THE TEST SECTION

Instrument	Description
Tube	Wolverine Tube Division; CDA copper No. 122 Type DHP. fully annealed temperature; finish comparable to tubes produced under ASTM B-111; 1.750-in. OD, 0.785-in. ID.
TE, Wall	Claude S. Gordon Co.; Type K, Chromel-Alumel, B & S Gage No. 24, Duplex, Glass Braid Insulated each conductor.
Insulators	CGS/Thermodynamics; high-temperature, porcelain, one-hole, round, 0.140-in. OD, 0.093-in. ID, 3-in. long.
TE, In-stream	Conax Corp.; 304 SS sheathed, Type K, Chromel-Alumel, ungrounded, 0.040-in. OD, 12-in. long.
TE, Heater Control	Claude S. Gordon Co.; Type E, Chromel-Constantan, B & S Gage No. 24, duplex, glass braid insluated each conductor.
Heaters	E. L. Wiegand Co.; Chromalox Type TSSM-40, 115 V, 1120 W, 37-in. heated length.
Heat Transfer Cement	Thermon Mfg. Co.; high-temperature heat transfer cement, Grade T-63.
Differential Pressure Transmitter	Foxboro Co.; Type 13A, M.A.W.P. 1500-psi. range 0-100 in. water.
0-100 and 0-300 psi. Gages	Robertshaw Acragages; 0-1-0-psi. gage had a 0.5-psi. resolution. 0-300-psi. gage had a 1-psi. resolution.
Pressure Tap Lines	1/4-in.-OD copper tubing from taps to hand valves. 3/8-in.-OD copper tubing from hand valves to DP cell.
0-100 Division Gage	Ashcroft receiver gage. 3-15-psi. input, 0-100% output.
Solenoid Valves	1/4-in. pipe size, 3/64-in. orifice, 110 V normally closed, Automatic Switch Company, Florham Park, N. Y.

Table A-2 (Continued)

Instrument	Description
Hand Valves	Hoke 1/4-in.-OD tubing connections, primary rating 3000 psi. at 70°F, brass body with stainless steel stem.
Data Acquisition System	Vidar 521C Integrating digital voltmeter with Vidar 606 Master Scanner, Vidar 661 printer, Vidar 663 C/D Magnetic Tape Recorder, and Vidar 5403 System Controller.

APPENDIX B

CALIBRATION TECHNIQUES

APPENDIX B

CALIBRATION TECHNIQUES

Each piece of instrumentation used in the EHTF was carefully calibrated before being used. The calibration techniques employed were essentially the same as those used by Jallouk [42], and the reader is referred to this publication for additional information beyond that included here.

Pressure Drop

Pressure Taps. Even though the 6 pressure taps in the test section were carefully installed, a test was run to check their reliability, as well as the smoothness of the copper test section. Water at various flow rates was pumped through the test section after the pressure taps had been connected to an inclined mercury manometer. Pressure drops were measured at 2-, 4-, 6-, 8-, and 10-ft increments. The results agreed well with similar results obtained by Nikuradse [61] with smooth tubes, as shown in Figure B-1. Although this figure shows the results for the entire 10 ft, similar results were obtained for the other increments. A very good representation of the data is given by:

$$f = 0.06425\ (N_{Re})^{-0.2291} \qquad \text{(B-1)}$$

where f is the Fanning friction factor and N_{Re} is the liquid Reynolds number based on the tube equivalent diameter.

Pressure-Drop Measuring System. The Foxboro D/P cell was calibrated using a pneumatic deadweight tester which made it possible to expose the D/P cell to well-known pressure drops ranging from 0- to 100-in. water. The output from the D/P cell was then read on the 0- to

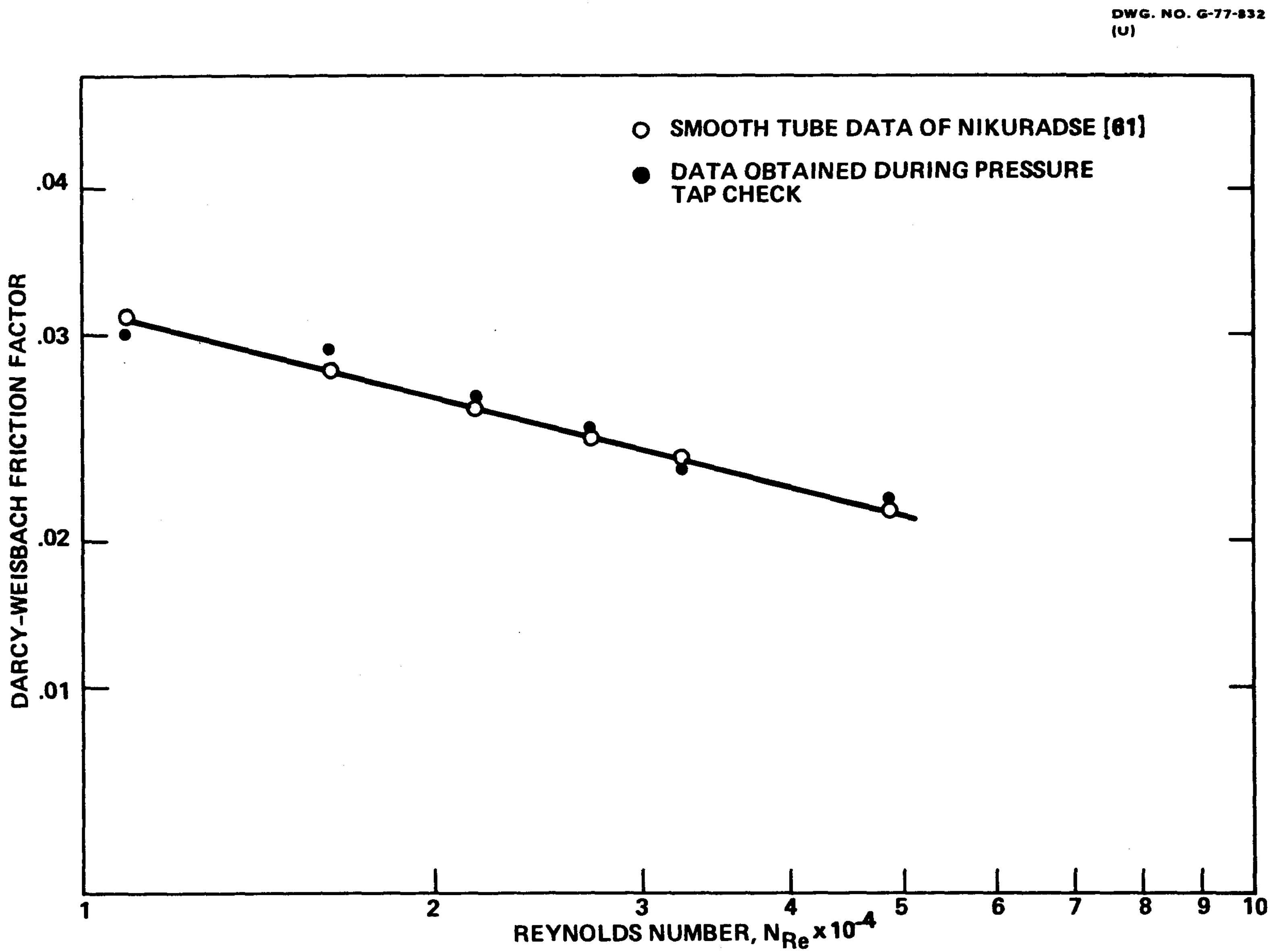

FIGURE B-1. FRICTION FACTOR DATA OBTAINED DURING PRESSURE TAP CHECK.

100-division receiver gage. The results of the calibration are shown in Figure B-2. This calibration was repeated several times throughout the course of the experiment with no detectable changes.

Flow Rates

Flows in the test facility were measured by turbine meters and rotameters. Each flowmeter was calibrated by pumping water at various flow rates through it and collecting the water in a barrel which was mounted on a set of scales. The mass of water collected per unit time was determined and converted into gpm. This result was compared to the reading obtained from the flowmeter. Generally speaking, good agreement was obtained between the experimental results and the manufacturer's recommendations.

Turbine Meters. All turbine meters used in the facility were manufactured by Cox Instruments Division, Detroit, Michigan. They were used to measure the inlet flow to the test section and the liquid flow from the entrainment separators. The calibration curves are shown in Figures B-3, B-4, and B-5. Table B-1 shows a list of all turbine meters used in the system, the flow measured, and the recommended range of accuracy.

Rotameters. Rotameters with special modifications to withstand high pressures were used to measure the remaining flows, as well as the entrained liquid flow, when it was less than 3.0 gpm. A list of these flowmeters is given in Table B-2.

DWG. NO. G-77-833
(U)

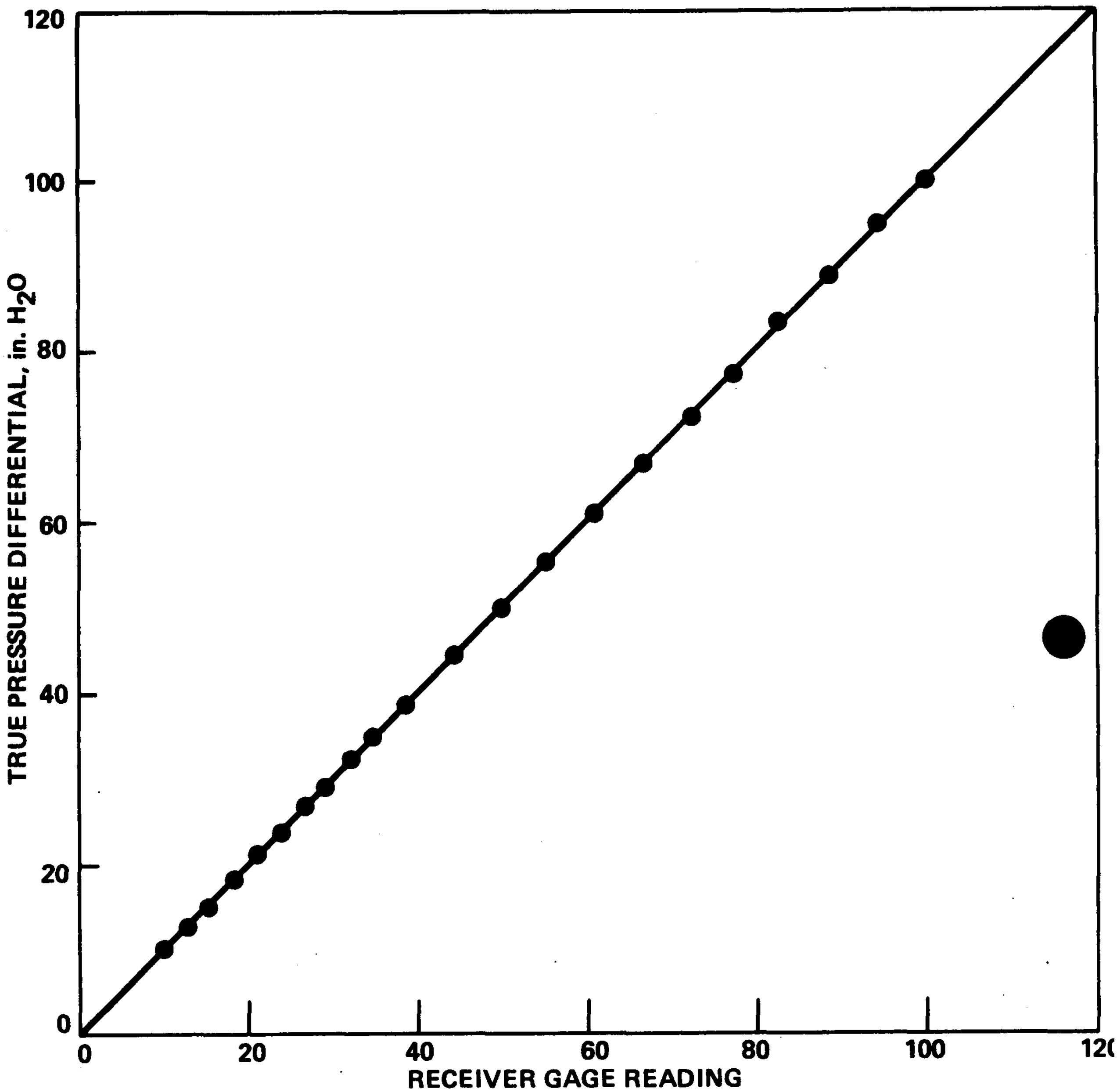

FIGURE B-2. RESULTS OF RECEIVER GAGE CALIBRATION

DWG. NO. G-77-834
(U)

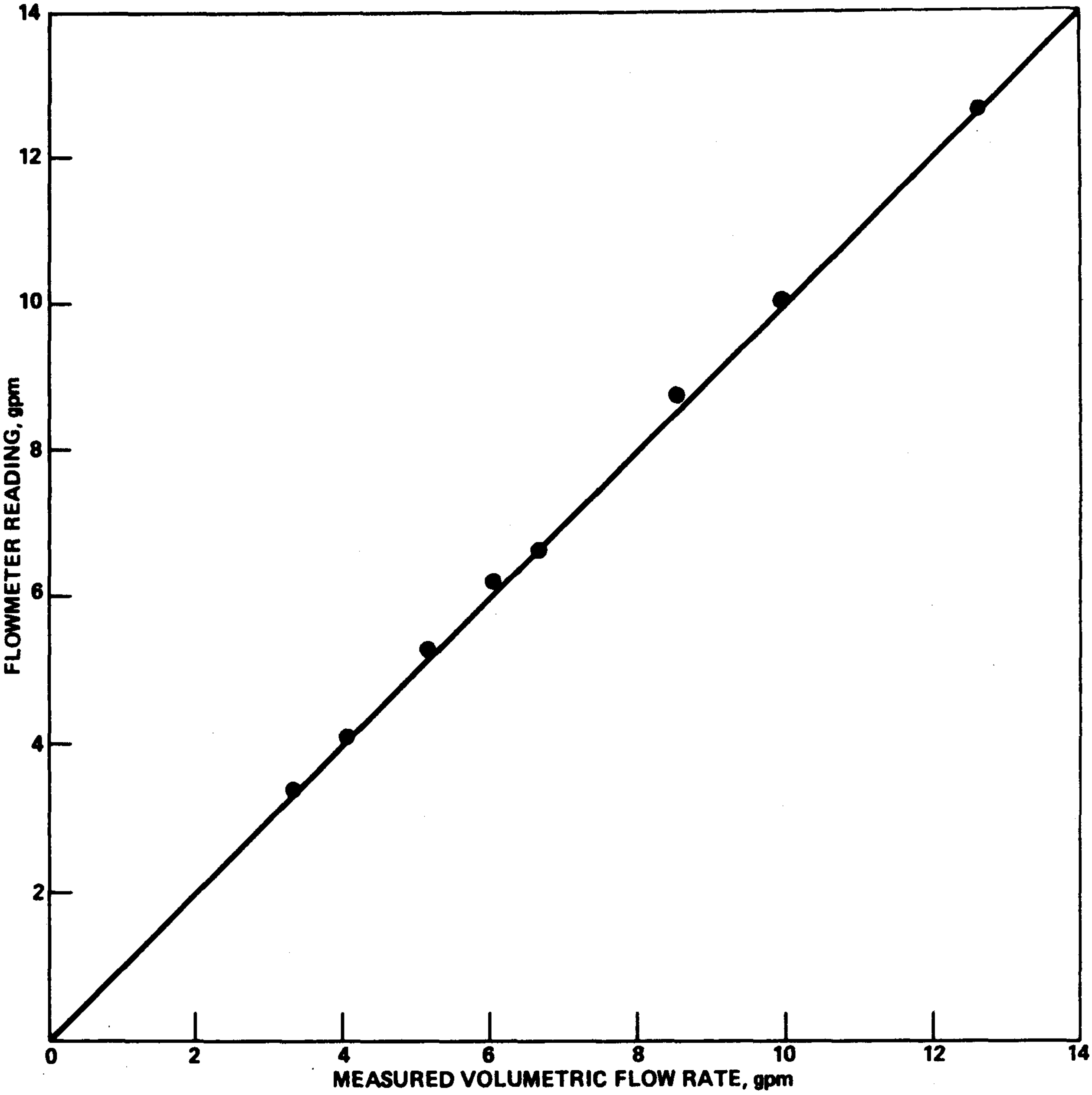

FIGURE B-3. CALIBRATION RESULTS FOR FE 105A.

DWG. NO. G-77-835
(U)

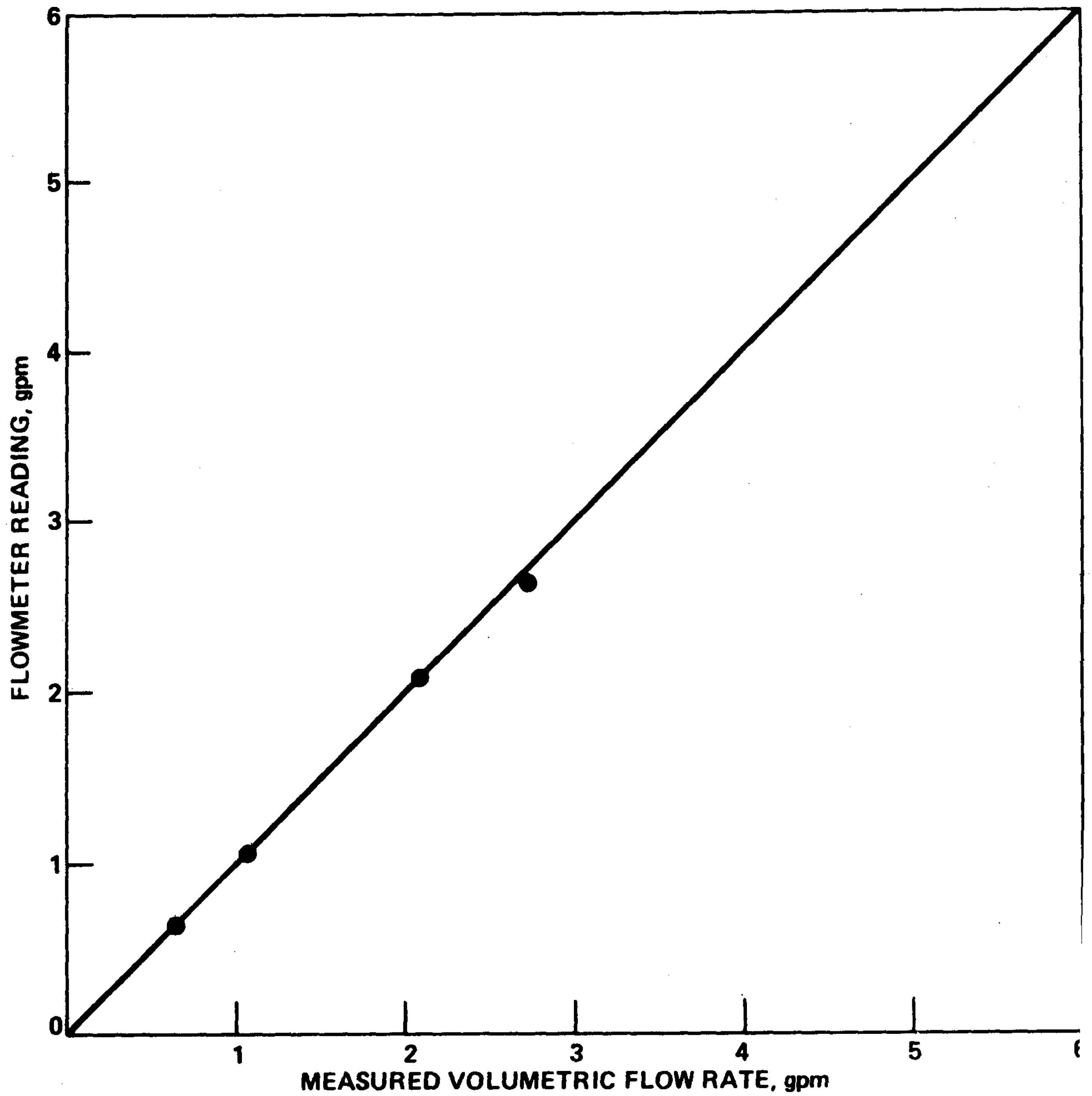

FIGURE B-4. CALIBRATION RESULTS FOR FE 105B

DWG. NO. G-77-836
(U)

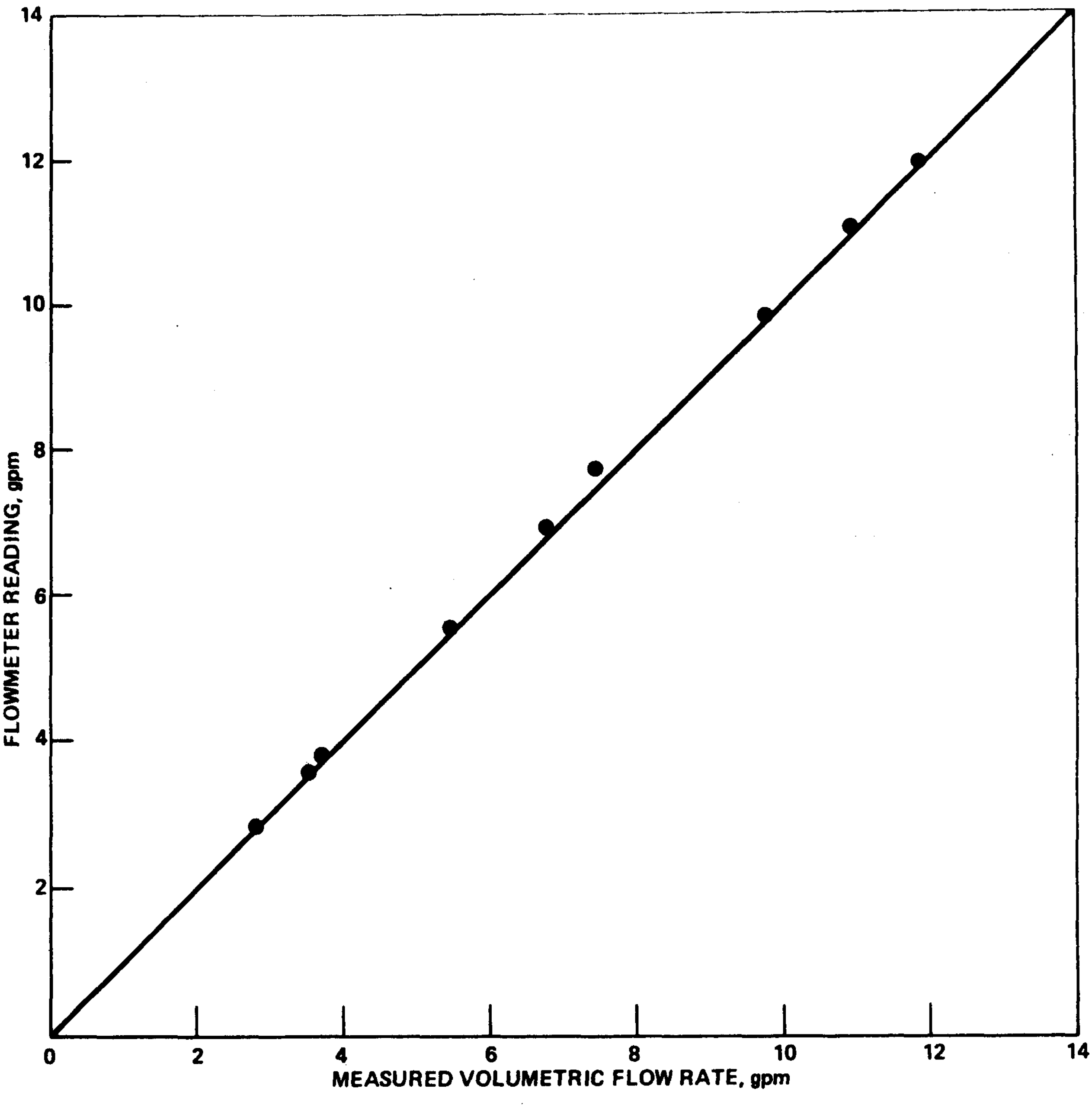

FIGURE B-5. CALIBRATION RESULTS FOR FE 119A.

Table B-1

TURBINE FLOWMETERS USED IN EHTF

Flowmeter Designation	Flow Measured	Range, gpm
FE 105A	Inlet	3.0 - 90.0
FE 105B	Inlet	0.5 - 6.0
FE 119A	Entrained Liquid	3.0 - 90.0

Table B-2

ROTAMETERS USED IN THE EHTF

Flowmeter Designation	Flow Measured	Range for Water, gpm
FE 119B	Entrained Liquid	0 - 4.330
FE 119C	Entrained Liquid	0 - 1.122
FE 123A	Condensed Vapor	0 - 4.330
FE 123A	Condensed Vapor	0 - 1.122
FE 126B	Cooling Water	0 - 17.60
FE 126C	Cooling Water	0 - 3.200
FE 126D	Cooling Water	0 - 1.122

As Jallouk [42] and others [60] have explained, even though some of the above flowmeters are used to measure R-114 flow, they may be calibrated with water and the resulting equation used. This equation can be written as:

$$Q = K\left(\frac{\rho_{float}}{\rho_{fluid}} - 1\right)^{1/2} R \qquad \text{(B-2)}$$

where Q is the actual flow in gpm, ρ_{float} is the density of the rotameter float, ρ_{fluid} is the density of the working fluid, R is the scale reading, and K is the constant determined from the calibration with water. The calibration curves for these rotameters are given in Figures B-6 through B-9.

Heat Input System

Heat Loss Through Insulation. The heat loss through the test section insulation was calibrated employing the same method used by Jallouk [42]. The system was evacuated and the test section was exposed to very low heat fluxes (0 to 100 watts/zone). The heat loss in each zone was correlated with the heater sheath temperature in that zone which was obtained from a thermocouple that was spot-welded to the heater. The linear relationships obtained are given by Equations (B-3) through (B-12), where Q_{loss} is the heat loss in watts/zone and T is the temperature, °F.

Zone 1: $$Q_{loss} = 0.1548T - 12.074 \qquad \text{(B-3)}$$

Zone 2: $$Q_{loss} = 0.1532T - 11.950 \qquad \text{(B-4)}$$

Zone 3: $$Q_{loss} = 0.1250T - 9.750 \qquad \text{(B-5)}$$

DWG. NO. G-77-837
(U)

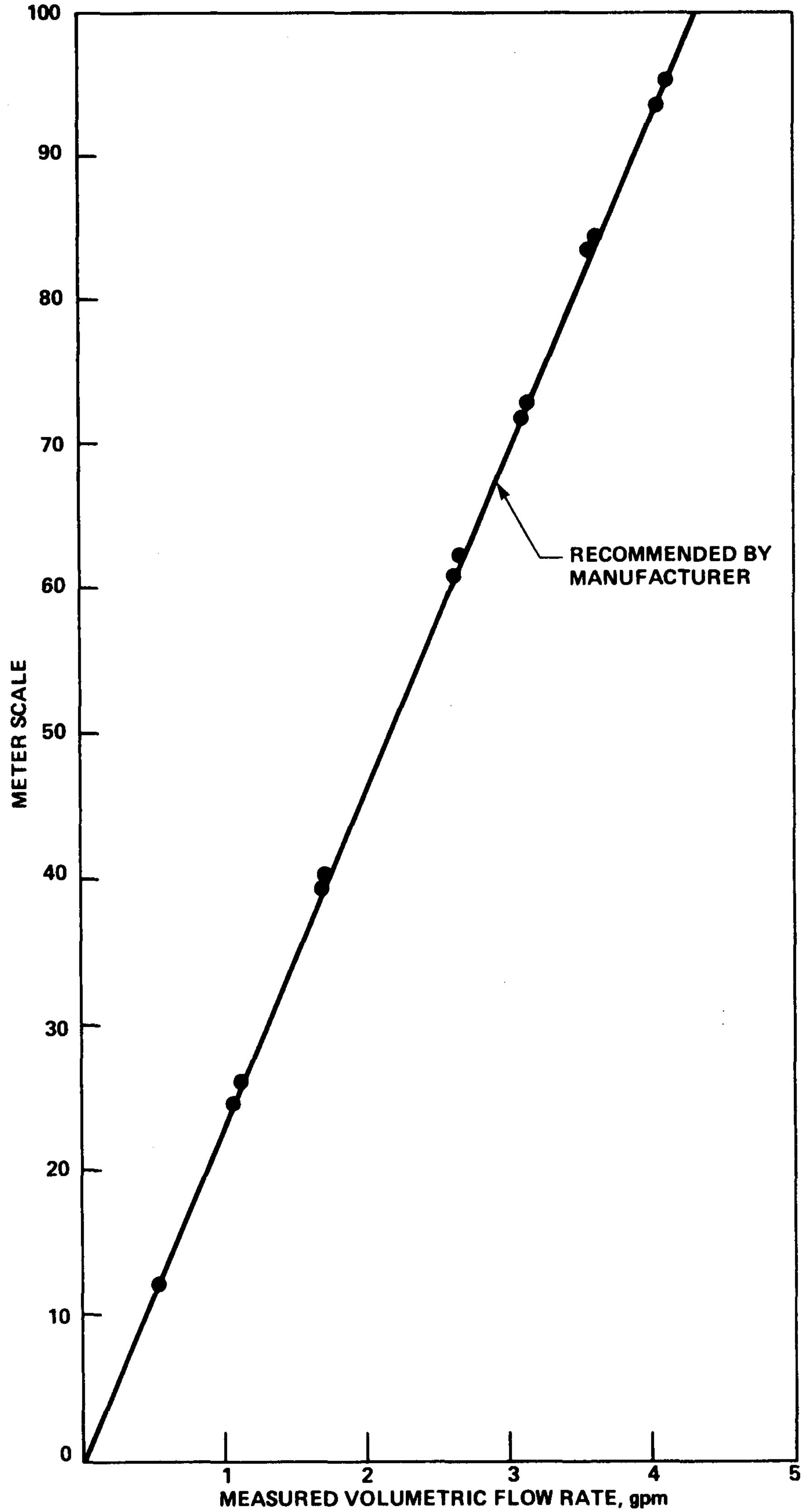

FIGURE B-6. CALIBRATION RESULTS FOR FE 119B AND FE 123A.

DWG. NO. G-77-838
(U)

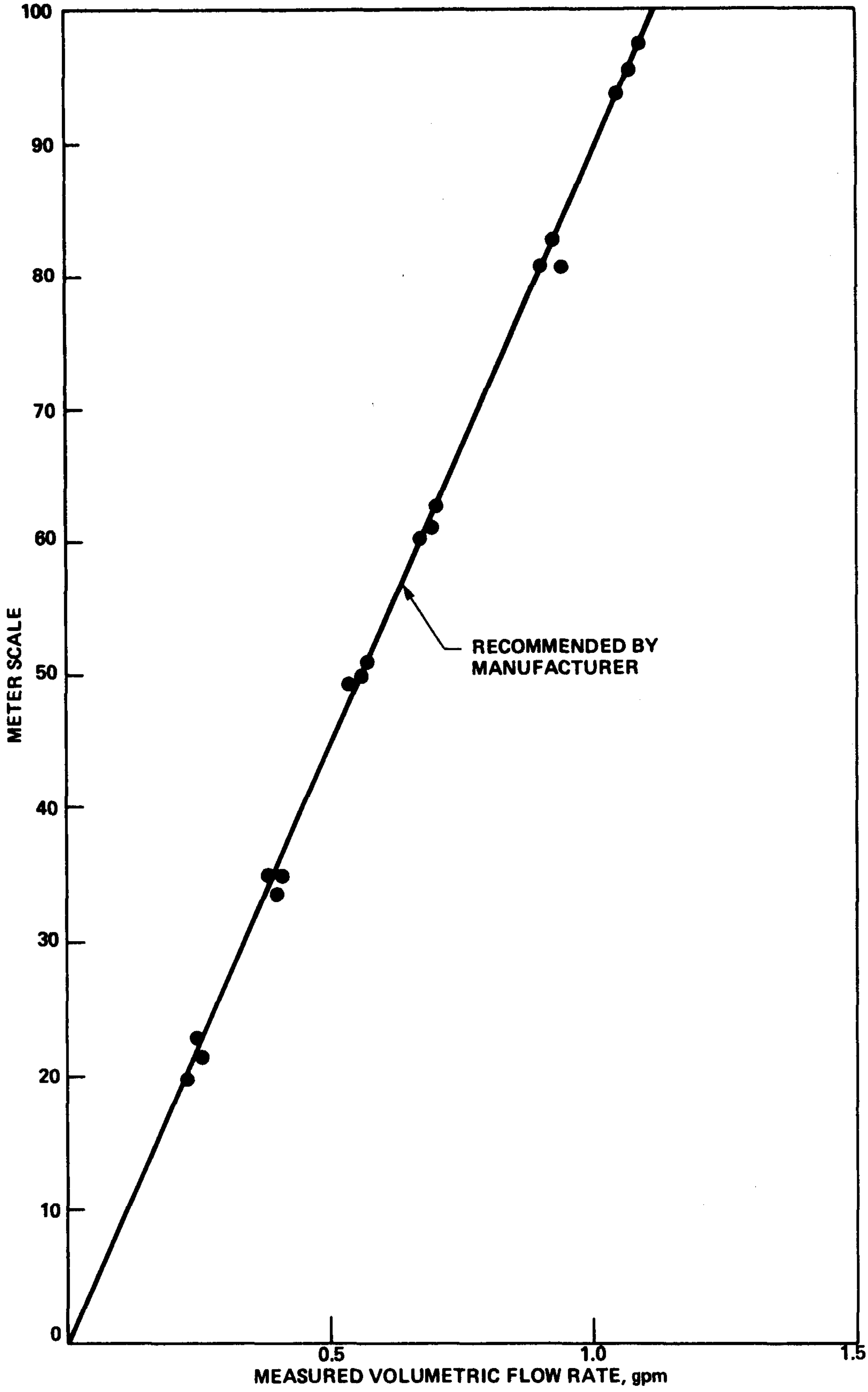

FIGURE B-7. CALIBRATION RESULTS FOR FE 119C, FE 123B, AND FE 126D.

DWG. NO. G-77-839
(U)

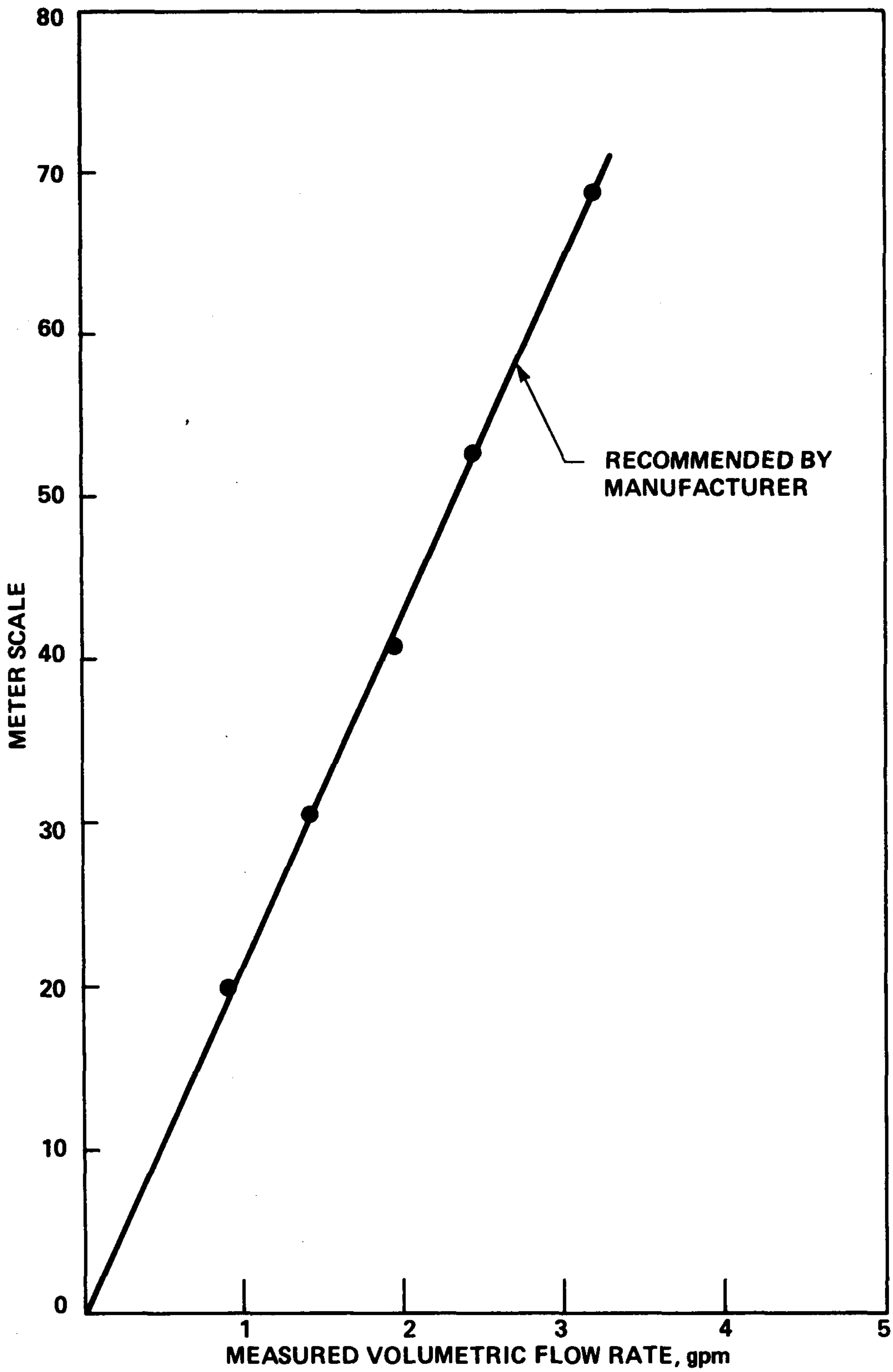

FIGURE B-8. CALIBRATION RESULTS FOR FE 126C.

DWG. NO. G-77-840
(U)

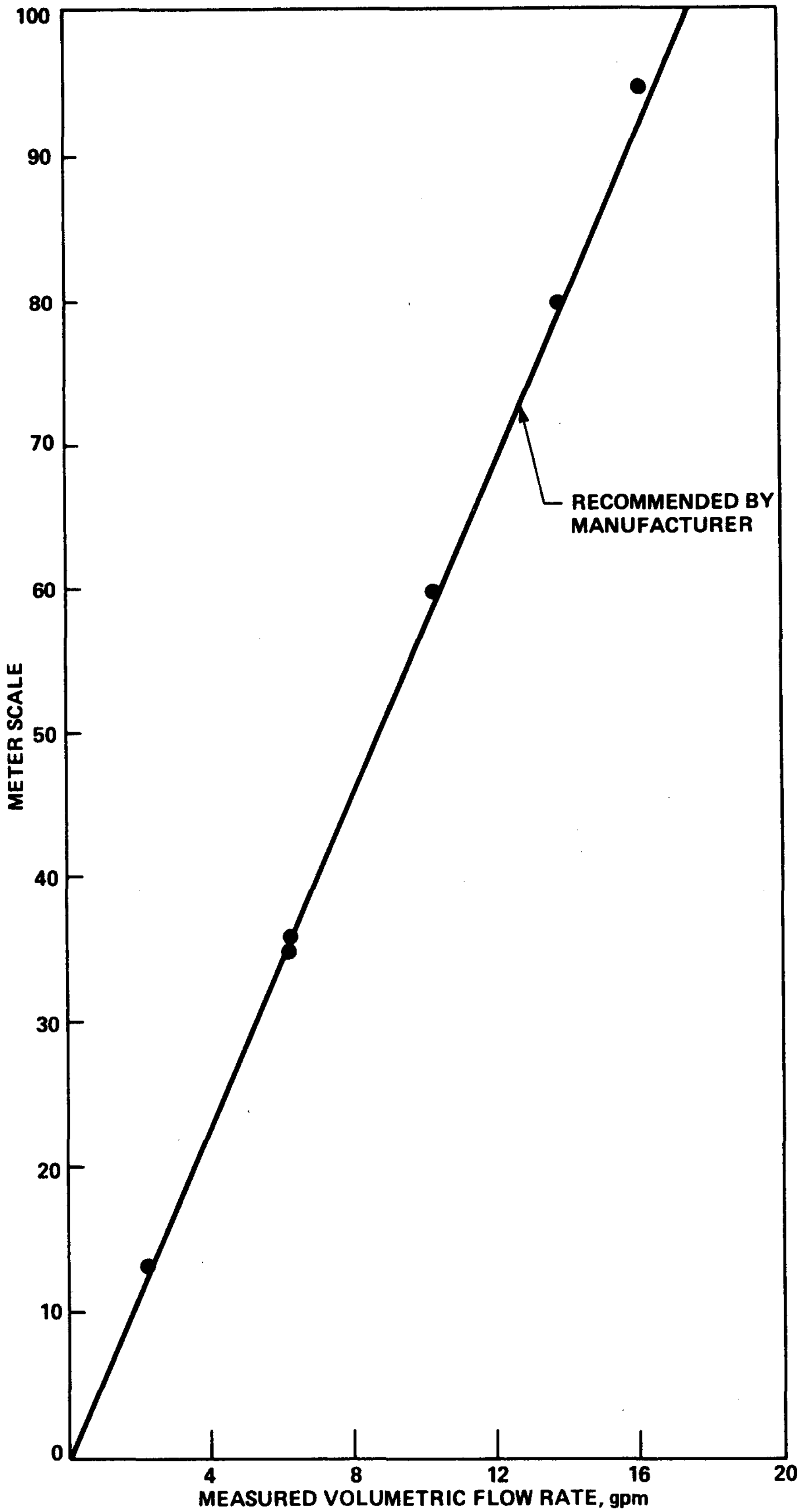

FIGURE B-9. CALIBRATION RESULTS FOR FE 126B.

Zone 4: $Q_{loss} = 0.1093T - 8.525$ (B-6)

Zone 5: $Q_{loss} = 0.1090T - 8.502$ (B-7)

Zone 6: $Q_{loss} = 0.1122T - 8.751$ (B-8)

Zone 7: $Q_{loss} = 0.1126T - 8.783$ (B-9)

Zone 8: $Q_{loss} = 0.1241T - 9.680$ (B-10)

Zone 9: $Q_{loss} = 0.1405T - 10.950$ (B-11)

Zone 10: $Q_{loss} = 0.1509T - 11.77$ (B-12)

Watt-Hour Meter Calibration. The General Electric watt-hour meter used to measure heat flux was calibrated only by the factory. This seemed reasonable, since information enclosed with the unit indicated that errors in excess of 0.5% should never be obtained.

Thermocouple Calibrations

Each Chromel-Alumel instream thermocouple was calibrated prior to installation. Plots of the deviation from the standard curve versus the actual temperature were constructed. These plots indicated that a reliable calibration was obtained for each instream thermocouple. Least squares fits were then obtained for each thermocouple which related millivolt reading to actual temperature. The best fits were obtained from relations of the form mv = mv(°F). The Newton-Rhapson Method was used to determine the true temperature from a given millivolt reading. The equations obtained are listed below where mv is the voltage difference in mv and T is the true temperature, °F.

Zone 1: $$mv = -0.68552 + 0.0210695T + 0.11745 \times 10^{-4}T^2 - 0.1994 \times 10^{-7}T^3 \quad \text{(B-13)}$$

$$\text{Zone 2:}\quad mv = -0.6865 + 0.02109T + 0.1185 \times 10^{-4}T^2 - 0.2089 \times 10^{-7}T^3 \qquad \text{(B-14)}$$

$$\text{Zone 3:}\quad mv = -0.6867 + 0.02111T + 0.11469 \times 10^{-4}T^2 - 0.19624 \times 10^{-7}T^3 \qquad \text{(B-15)}$$

$$\text{Zone 4:}\quad mv = -0.6849 + 0.02101T + 0.1287 \times 10^{-4}T^2 - 0.2335 \times 10^{-7}T^3 \qquad \text{(B-16)}$$

$$\text{Zone 5:}\quad mv = -0.6851 + 0.02104T + 0.1231 \times 10^{-4}T^2 - 0.2192 \times 10^{-7}T^3 \qquad \text{(B-17)}$$

$$\text{Zone 6:}\quad mv = -0.6917 + 0.02129T + 0.1066 \times 10^{-4}T^2 - 0.1859 \times 10^{-7}T^3 \qquad \text{(B-18)}$$

$$\text{Zone 7:}\quad mv = -0.6908 + 0.0212T + 0.1093 \times 10^{-4}T^2 - 0.1905 \times 10^{-7}T^3 \qquad \text{(B-19)}$$

$$\text{Zone 8:}\quad mv = -0.6890 + 0.02119T + 0.1129 \times 10^{-4}T^2 - 0.1954 \times 10^{-7}T^3 \qquad \text{(B-20)}$$

$$\text{Zone 9:}\quad mv = -0.6852 + 0.02133T + 0.1153 \times 10^{-4}T^2 - 0.1988 \times 10^{-7}T^3 \qquad \text{(B-21)}$$

$$\text{Zone 10:}\quad mv = -0.6873 + 0.02111T + 0.1209 \times 10^{-4}T^2 - 0.2171 \times 10^{-7}T^3 \qquad \text{(B-22)}$$

A calibration was also obtained for the Chromel-Alumel wall thermocouples. This was done by calibrating the first and last pieces of wire from both of the spools used. No significant difference could be detected from any of the four calibrations. The equation which best fits the data is:

$$mv = -0.6793 + 0.02086T + 0.1204 \times 10^{-4}T^2 - 0.2189 \times 10^{-7}T^3 \qquad \text{(B-23)}$$

Overall Calibration

For added assurance that all systems except the pressure-drop measuring system were adequately calibrated, a test was run using single-phase R-114. A heat flux of about 3,200 Btu/hr-sq ft-°F was imposed on the test section, and the refrigerant flow was set at about 14.0 gpm. At steady-state conditions, the wall and instream temperatures were recorded along with the refrigerant flow rate. Experimental heat transfer coefficients were calculated using the heat flux and measured temperature difference in each zone. The results were compared with heat transfer coefficients predicted by the well-known Seider-Tate [61] equation, as shown in Table B-3. As can be seen, the agreement was very good, except in Zones 9 and 10 where the experimental heat transfer coefficient exceeds that predicted by the Seider-Tate [61] relationship. This poorer agreement was probably caused by local boiling within these zones. In fact, small bubbles could be seen in the flow visualization section during this test. E. I. duPont de Nemours & Co. [62] recommends that the Seider-Tate equation be used to predict heat transfer coefficients for single-phase liquid refrigerants flowing in tubes.

Table B-3

COMPARISON OF CALCULATED AND EXPERIMENTAL SINGLE-PHASE HEAT TRANSFER COEFFICIENTS

Zone	Experimental Heat Transfer Coefficient, Btu/hr-sq ft-°F	Heat Transfer Coefficient Calculated From Seider-Tate Equation [61], Btu/hr-sq ft-°F
1	373.83	403.81
2	377.92	404.07
3	384.90	404.24
4	379.14	404.28
5	383.61	404.46
6	396.32	404.41
7	415.20	404.56
8	425.91	404.66
9	433.64	404.78
10	468.51	404.74

APPENDIX C

TABULATION OF DATA

APPENDIX C

TABULATION OF DATA

Tables C-1 through C-65 contain a complete listing of all acquired experimental data. Correlated results are also given for the pressure drop data.

Table C-1

A LISTING OF THE FLOW REGIME DATA

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-1	0.213	150	0.272	Intermittent
FR-2	0.450	150	0.200	Intermittent
FR-3	0.071	150	0.065	Intermittent
FR-4	1.681	150	0.065	Annular
FR-5	3.305	150	0.050	Annular
FR-6	0.114	150	0.247	Stratified-Wavy
FR-7	0.257	150	0.160	Intermittent
FR-8	0.518	150	0.095	Intermittent
FR-9	0.872	150	0.026	Intermittent
FR-10	1.643	150	0.009	Intermittent
FR-11	3.426	150	0.085	Annular
FR-12	0.223	150	0.458	Stratified-Wavy
FR-15	1.669	150	0.063	Annular
FR-16	3.352	150	0.039	Annular
FR-17	0.222	150	0.754	Stratified-Wavy
FR-18	0.428	150	0.483	Intermittent
FR-19	0.877	150	0.239	Intermittent
FR-20	1.670	150	0.126	Annular
FR-21	3.396	150	0.072	Annular
FR-23	0.483	200	0.470	Statified-Wavy
FR-24	0.910	200	0.326	Intermittent
FR-25	1.723	200	0.173	Annular

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-26	3.390	200	0.094	Annular
FR-27	3.381	200	0.078	Annular
FR-28	1.709	200	0.100	Annular
FR-29	0.890	200	0.156	Intermittent
FR-30	0.460	200	0.298	Stratified-Wavy
FR-32	0.207	200	0.524	Stratified-Wavy
FR-33	3.402	200	0.044	Annular
FR-34	1.699	200	0.046	Intermittent
FR-35	0.910	200	0.085	Intermittent
FR-37	0.209	200	0.524	Stratified-Wavy
FR-38	0.113	200	0.630	Stratified-Wavy
FR-39	3.411	200	0.015	Annular
FR-40	1.692	200	0.022	Intermittent
FR-41	0.906	200	0.034	Intermittent
FR-44	0.112	100	0.434	Stratified-Wavy
FR-45	0.212	100	0.413	Stratified-Wavy
FR-46	0.454	100	0.088	Intermittent
FR-47	0.854	100	0.048	Intermittent
FR-48	1.674	100	0.030	Intermittent
FR-49	3.369	100	0.026	Annular
FR-50	0.112	100	0.613	Stratified-Wavy
FR-51	0.252	100	0.413	Stratified-Wavy
FR-52	0.438	100	0.189	Intermittent

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-54	1.673	100	0.045	Annular
FR-55	3.455	100	0.029	Bubbly
FR-57	0.227	100	0.507	Stratified-Wavy
FR-58	0.450	100	0.270	Intermittent
FR-59	0.877	100	0.124	Intermittent
FR-60	1.719	100	0.110	Annular
FR-61	3.399	100	0.047	Annular
FR-62	0.224	100	0.651	Stratified-Wavy
FR-63	0.427	100	0.387	Stratified-Wavy
FR-64	0.794	100	0.126	Intermittent
FR-65	1.647	100	0.085	Annular
FR-66	3.338	100	0.020	Annular
FR-67	0.980	250	0.345	Intermittent
FR-68	0.407	250	0.428	Stratified-Wavy
FR-69	1.778	250	0.124	Annular
FR-70	0.407	250	0.348	Intermittent
FR-71	0.981	250	0.174	Intermittent
FR-72	1.804	250	0.069	Annular
FR-73	1.858	250	0.120	Annular
FR-74	0.852	250	0.344	Intermittent
FR-75	0.462	250	0.728	Intermittent
FR-76	0.465	250	0.725	Stratified-Wavy
FR-77	0.895	250	0.333	Intermittent

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-78	1.685	250	0.173	Annular
FR-79	0.116	100	0.510	Intermittent
FR-80	0.117	150	0.416	Intermittent
FR-81	0.227	150	0.218	Intermittent
FR-82	0.450	150	0.103	Intermittent
FR-83	0.880	150	0.128	Intermittent
FR-84	1.663	150	0.071	Intermittent
FR-85	3.442	150	0.039	Annular
FR-86	0.116	150	0.830	Stratified-Wavy
FR-87	0.223	150	0.357	Intermittent
FR-88	0.453	150	0.236	Intermittent
FR-89	0.886	150	0.128	Intermittent
FR-90	1.682	150	0.076	Intermittent
FR-91	3.386	150	0.048	Annular
FR-92	0.224	150	0.449	Intermittent
FR-93	0.451	150	0.239	Intermittent
FR-94	0.923	150	0.163	Intermittent
FR-95	1.672	150	0.081	Intermittent
FR-96	3.397	150	0.057	Annular
FR-98	0.226	150	0.256	Intermittent
FR-99	0.441	150	0.116	Intermittent
FR-100	0.922	150	0.123	Intermittent
FR-101	1.706	150	0.085	Intermittent

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-102	3.558	150	0.048	Annular
FR-103	0.118	200	0.734	Stratified-Wavy
FR-104	0.232	200	0.442	Intermittent
FR-105	0.478	200	0.303	Intermittent
FR-106	0.875	200	0.196	Intermittent
FR-107	1.673	200	0.102	Intermittent
FR-108	3.326	200	0.140	Annular
FR-109	0.228	200	0.558	Intermittent
FR-110	0.482	200	0.383	Intermittent
FR-111	0.897	200	0.218	Intermittent
FR-112	1.676	200	0.132	Annular
FR-113	3.286	200	0.067	Annular
FR-114	0.120	200	0.420	Stratified-Wavy
FR-115	0.244	200	0.194	Intermittent
FR-116	0.449	200	0.222	Intermittent
FR-117	0.892	200	0.118	Intermittent
FR-118	1.668	200	0.072	Intermittent
FR-119	3.339	200	0.045	Annular
FR-120	3.346	200	0.035	Annular
FR-123	0.112	100	0.139	Intermittent
FR-124	0.220	100	0.035	Intermittent
FR-126	0.888	100	0.088	Intermittent
FR-127	1.659	100	0.152	Intermittent

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-128	3.321	100	0.009	Stratified-Smooth
FR-129	0.108	100	0.559	Stratified-Smooth
FR-130	0.211	100	0.311	Intermittent
FR-131	0.432	100	0.076	Intermittent
FR-132	0.863	100	0.082	Intermittent
FR-133	1.668	100	0.009	Intermittent
FR-134	3.276	100	0.018	Annular
FR-135	3.271	100	0.002	Annular
FR-136	0.119	100	0.819	Stratified-Wavy
FR-137	0.215	100	0.385	Intermittent
FR-138	0.431	100	0.123	Intermittent
FR-139	0.875	100	0.169	Intermittent
FR-140	1.676	100	0.061	Intermittent
FR-141	3.326	100	0.036	Annular
FR-142	3.281	100	0.043	Annular
FR-143	0.228	100	0.541	Stratified-Wavy
FR-144	0.482	100	0.346	Intermittent
FR-145	0.859	100	0.233	Intermittent
FR-146	1.666	100	0.139	Intermittent
FR-147	0.231	250	0.553	Stratified-Wavy
FR-148	0.474	250	0.458	Stratified-Wavy
FR-149	0.893	250	0.411	Intermittent
FR-150	1.632	250	0.189	Annular

Table C-1 (Continued)

Run No.	Mass Flux x 10^{-6}, lbm/hr-sq ft	Temperature, °F	Quality	Flow Regime
FR-151	3.098	250	0.128	Annular
FR-152	0.122	250	0.719	Stratified-Wavy
FR-153	0.218	250	0.409	Stratified-Wavy
FR-154	0.492	250	0.350	Intermittent
FR-155	0.894	250	0.309	Intermittent
FR-157	3.337	250	0.146	Annular
FR-158	0.112	250	0.540	Stratified-Wavy
FR-159	0.216	250	0.212	Stratified-Wavy
FR-160	0.510	250	0.311	Intermittent
FR-161	0.851	250	0.060	Intermittent
FR-162	1.636	250	0.030	Intermittent
FR-163	3.354	250	0.014	Annular
FR-164	0.882	200	0.117	Intermittent
FR-165	0.227	200	0.219	Intermittent

Table C-2

PRESSURE DROP - EXPERIMENTAL DATA,
G = 0.12 x 10^6 lbm/hr-sq ft, T = 100°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
61-24	0.1129E 06	104.6	0.052	0.380	3.9948	0.0721
61-45	0.1129E 06	104.8	0.380	0.544	2.0104	0.0541
63-24	0.1090E 06	98.7	0.113	0.317	3.9948	0.1442
63-45	0.1090E 06	98.6	0.317	0.420	2.0104	0.0360
65-24	0.1227E 06	97.4	0.162	0.583	3.9948	0.1622
65-45	0.1227E 06	97.6	0.583	0.792	2.0104	0.2163
69-45	0.1169E 06	99.6	0.508	0.820	2.0104	0.1802

Table C-3

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 100^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
60-45	0.2247E 06	105.3	0.130	0.213	2.0104	0.1081
62-24	0.2241E 06	102.8	0.080	0.177	3.9948	0.1261
62-45	0.2241E 06	102.7	0.177	0.226	2.0104	0.0721
62-56	0.2241E 06	102.3	0.226	0.273	1.8437	0.0541
64-24	0.2540E 06	98.1	0.088	0.291	3.9948	0.3064
64-45	0.2540E 06	98.2	0.291	0.394	2.0104	0.1802
64-56	0.2540E 06	97.9	0.394	0.495	1.8437	0.1261
68-24	0.2292E 06	101.1	0.003	0.323	3.9948	0.3965
68-45	0.2292E 06	101.6	0.323	0.484	2.0104	0.3064

Table C-4

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 100^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
58-24	0.4341E 06	102.7	0.046	0.216	3.9948	0.4145
58-45	0.4341E 06	103.1	0.215	0.303	2.0104	0.3244
58-56	0.4341E 06	102.7	0.303	0.391	1.8437	0.2523
59-24	0.4468E 06	106.1	0.011	0.094	3.9948	0.1982
59-45	0.4468E 06	106.3	0.094	0.136	2.0104	0.1081
59-56	0.4468E 06	106.1	0.136	0.176	1.8437	0.1802
66-24	0.4613E 06	97.2	0.054	0.167	3.9948	0.4145
66-45	0.4613E 06	97.0	0.167	0.225	2.0104	0.2523
66-56	0.4613E 06	96.5	0.224	0.280	1.8437	0.1802
67-24	0.4500E 06	104.0	0.019	0.184	3.9948	0.6488
67-45	0.4500E 06	104.0	0.184	0.270	2.0104	0.3424
67-56	0.4500E 06	103.5	0.270	0.354	1.8437	0.2883
70-24	0.4607E 06	102.5	0.075	0.274	3.9948	0.7929
70-45	0.4607E 06	103.0	0.274	0.372	2.0104	0.4686

Table C-5

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 150°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
12-45	0.1110E 06	148.9	0.088	0.196	2.0104	0.0396
40-24	0.1182E 06	149.0	0.007	0.364	3.9948	0.0469
40-45	0.1182E 06	149.2	0.364	0.537	2.0104	0.0433
76-24	0.1215E 06	150.4	0.035	0.506	3.9948	0.0829
76-45	0.1215E 06	150.9	0.505	0.731	2.0104	0.0613
82-45	0.1184E 06	151.3	0.129	0.221	2.0104	0.0469

Table C-6

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 150^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
6-45	0.2426E 06	149.6	0.078	0.165	2.0104	0.0901
13-45	0.2348E 06	148.9	0.071	0.127	2.0104	0.0613
18-45	0.2370E 06	149.2	0.170	0.294	2.0104	0.1442
23-24	0.2378E 06	148.8	0.050	0.395	3.9948	0.2163
46-45	0.2367E 06	149.0	0.088	0.175	2.0104	0.1262
54-45	0.2426E 06	150.7	0.139	0.225	2.0104	0.0829
73-24	0.2257E 06	150.1	0.030	0.394	3.9948	0.1802
73-45	0.2257E 06	150.7	0.394	0.578	2.0104	0.1442
73-56	0.2257E 06	150.6	0.578	0.760	1.8437	0.0721
75-45	0.2274E 06	150.7	0.167	0.295	2.0104	0.0901
75-56	0.2274E 06	150.4	0.295	0.421	1.8437	0.0360
78-45	0.2315E 06	149.6	0.064	0.154	2.0104	0.0541
78-56	0.2315E 06	149.4	0.154	0.242	1.8437	0.0288
81-45	0.2263E 06	151.0	0.068	0.124	2.0104	0.0829

Table C-7

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 150^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
7-24	0.4648E 06	149.6	0.012	0.101	3.9948	0.1982
7-45	0.4648E 06	149.3	0.101	0.148	2.0104	0.1262
7-56	0.4648E 06	148.5	0.148	0.186	1.8437	0.0721
16-24	0.4790E 06	149.6	0.017	0.070	3.9948	0.1189
16-45	0.4790E 06	149.6	0.070	0.100	2.0104	0.0721
19-45	0.4656E 06	150.0	0.114	0.180	2.0104	0.1622
19-56	0.4656E 06	149.1	0.180	0.235	1.8437	0.1081
27-45	0.4644E 06	149.3	0.164	0.258	2.0104	0.2523
27-56	0.4644E 06	148.2	0.258	0.343	1.8437	0.1262
43-45	0.4902E 06	149.0	0.087	0.150	2.0104	0.1442
43-56	0.4902E 06	148.1	0.150	0.203	1.8437	0.1081
71-45	0.4505E 06	151.1	0.201	0.317	2.0104	0.2163
71-56	0.4505E 06	150.9	0.317	0.433	1.8437	0.2343
72-24	0.4744E 06	150.5	0.000	0.173	3.9948	0.3424
72-45	0.4744E 06	150.9	0.173	0.261	2.0104	0.2523
72-56	0.4744E 06	150.8	0.260	0.347	1.8437	0.1802
74-24	0.4977E 06	150.1	0.019	0.132	3.9948	0.2523
74-45	0.4977E 06	150.6	0.132	0.191	2.0104	0.2163
74-56	0.4977E 06	150.4	0.191	0.247	1.8437	0.0721
77-24	0.4687E 06	149.9	0.013	0.099	3.9948	0.1802
77-45	0.4687E 06	150.0	0.099	0.144	2.0104	0.1442
77-56	0.4687E 06	149.7	0.144	0.185	1.8437	0.0721
80-24	0.4822E 06	150.8	0.047	0.098	3.9948	0.0631
80-45	0.4822E 06	150.7	0.098	0.126	2.0104	0.0991
80-56	0.4822E 06	150.3	0.126	0.149	1.8437	0.0360

Table C-8

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.96 \times 10^6$ lbm/hr-sq ft, $T \approx 150^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
8-24	0.8761E 06	149.7	0.003	0.052	3.9948	0.3748
8-45	0.8761E 06	149.3	0.052	0.079	2.0104	0.2523
8-56	0.8761E 06	148.4	0.079	0.098	1.8437	0.1982
20-12	0.8940E 06	149.7	0.086	0.119	1.9844	0.3064
20-24	0.8940E 06	149.6	0.119	0.186	3.9948	0.5406
20-45	0.8940E 06	148.9	0.186	0.222	2.0104	0.4145
20-56	0.8940E 06	147.9	0.222	0.250	1.8437	0.3064
24-12	0.8760E 06	148.9	0.038	0.086	1.9844	0.3532
24-24	0.8760E 06	148.8	0.086	0.183	3.9948	0.7100
24-45	0.8760E 06	148.3	0.183	0.235	2.0104	0.4505
24-56	0.8760E 06	147.0	0.235	0.277	1.8437	0.3604
28-12	0.8790E 06	145.8	0.051	0.097	1.9844	0.3604
28-24	0.8790E 06	148.3	0.097	0.215	3.9948	0.6848
28-45	0.8790E 06	148.0	0.215	0.280	2.0104	0.5597
28-56	0.8790E 06	146.5	0.280	0.333	1.8437	0.4145
47-24	0.8653E 06	149.2	0.018	0.066	3.9948	0.3604
47-45	0.8653E 06	148.9	0.066	0.092	2.0104	0.3064
47-56	0.8653E 06	148.3	0.092	0.113	1.8437	0.1802
55-12	0.8761E 06	150.1	0.031	0.050	1.9844	0.2775
55-24	0.8761E 06	150.9	0.050	0.099	3.9948	0.3352
55-45	0.8761E 06	150.4	0.099	0.125	2.0104	0.2883
55-56	0.8761E 06	149.7	0.126	0.146	1.8437	0.1982
83-12	0.8858E 06	152.2	0.065	0.078	1.9844	0.3064
83-24	0.8858E 06	152.5	0.078	0.108	3.9948	0.4145
83-45	0.8858E 06	152.0	0.108	0.124	2.0104	0.3424
83-56	0.8858E 06	151.5	0.124	0.136	1.8437	0.1081

Table C-9

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 1.70 x 10^6 lbm/hr-sq ft, T ≈ 150°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
3-12	0.1755E 07	150.3	0.064	0.074	1.9844	0.4505
3-24	0.1755E 07	149.9	0.074	0.090	3.9948	0.5767
3-45	0.1755E 07	149.4	0.090	0.104	2.0104	0.4866
3-56	0.1755E 07	148.0	0.104	0.106	1.8437	0.3965
10-24	0.1673E 07	151.0	0.003	0.029	3.9948	0.6380
10-45	0.1673E 07	150.7	0.029	0.049	2.0104	0.5695
10-56	0.1673E 07	149.2	0.049	0.058	1.8437	0.4686
15-12	0.1667E 07	149.5	0.023	0.031	1.9844	0.4109
15-24	0.1667E 07	149.3	0.031	0.049	3.9948	0.5262
15-45	0.1667E 07	148.8	0.049	0.062	2.0104	0.4325
15-56	0.1667E 07	147.6	0.062	0.066	1.8437	0.3424
21-12	0.1685E 07	150.2	0.086	0.103	1.9844	0.5587
21-24	0.1685E 07	150.1	0.103	0.142	3.9948	0.9912
21-45	0.1685E 07	149.3	0.142	0.166	2.0104	0.7569
21-56	0.1685E 07	147.9	0.166	0.180	1.8437	0.6668
25-12	0.1675E 07	149.4	0.002	0.024	1.9844	0.6668
25-24	0.1675E 07	150.0	0.024	0.080	3.9948	1.3156
25-45	0.1675E 07	148.8	0.080	0.112	2.0104	0.9732
29-12	0.1676E 07	145.9	0.012	0.029	1.9844	0.6308
29-24	0.1676E 07	148.6	0.029	0.092	3.9948	1.4237
29-45	0.1676E 07	148.1	0.092	0.131	2.0104	1.0453
31-12	0.1690E 07	148.5	0.000	0.031	1.9844	0.8831
31-24	0.1690E 07	149.7	0.031	0.112	3.9948	1.9103
39-12	0.1702E 07	149.8	0.038	0.044	1.9844	0.4686
39-24	0.1702E 07	150.7	0.044	0.072	3.9948	0.7569
39-45	0.1702E 07	150.1	0.072	0.091	2.0104	0.5767
39-56	0.1702E 07	148.8	0.091	0.099	1.8437	0.5046
56-12	0.1683E 07	149.8	0.031	0.036	1.9844	0.4866
56-24	0.1683E 07	151.2	0.036	0.063	3.9948	0.7389
56-45	0.1683E 07	150.7	0.063	0.079	2.0104	0.5767
56-56	0.1683E 07	149.9	0.079	0.091	1.8437	0.5406

Table C-10

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 3.40×10^6 lbm/hr-sq ft, T ≈ 150°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
11-12	0.3398E 07	150.9	0.035	0.037	1.9844	0.9083
11-24	0.3398E 07	151.7	0.037	0.057	3.9948	2.4798
17-12	0.3401E 07	152.0	0.022	0.023	1.9844	0.8110
17-24	0.3401E 07	152.5	0.023	0.039	3.9948	1.9643
22-12	0.3415E 07	150.3	0.057	0.062	1.9844	1.0993
26-12	0.3393E 07	147.7	0.040	0.043	1.9844	0.8939
26-24	0.3393E 07	149.6	0.043	0.062	3.9948	2.5482
30-12	0.3317E 07	148.0	0.045	0.048	1.9844	1.1534
32-12	0.3432E 07	147.7	0.054	0.059	1.9844	1.2074
57-12	0.3457E 07	150.0	0.037	0.039	1.9844	0.9551
57-24	0.3457E 07	150.8	0.039	0.057	3.9948	2.3248

Table C-11

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
86-12	0.2682E 06	201.1	0.178	0.303	1.9844	0.1622
86-24	0.2682E 06	201.4	0.303	0.553	3.9948	0.1081
86-45	0.2682E 06	201.8	0.553	0.675	2.0104	0.1081
86-56	0.2682E 06	202.6	0.675	0.806	1.8437	0.0541
88-24	0.2662E 06	200.1	0.078	0.185	3.9948	0.0360
88-45	0.2662E 06	200.0	0.185	0.240	2.0104	0.0541
94-24	0.2557E 06	200.1	0.173	0.359	3.9948	0.0757
94-45	0.2557E 06	200.2	0.359	0.454	2.0104	0.0937
94-56	0.2557E 06	199.8	0.454	0.545	1.8437	0.0324
95-45	0.2574E 06	201.3	0.331	0.520	2.0104	0.1081
120-12	0.2187E 06	199.3	0.169	0.227	1.9844	0.0901
120-24	0.2187E 06	200.2	0.227	0.354	3.9948	0.0252
120-45	0.2187E 06	200.1	0.354	0.417	2.0104	0.0396
121-12	0.2245E 06	198.9	0.197	0.244	1.9844	0.1081
121-45	0.2245E 06	199.0	0.348	0.398	2.0104	0.0721
123-12	0.2236E 06	201.6	0.181	0.214	1.9844	0.1009
123-45	0.2236E 06	201.3	0.269	0.302	2.0104	0.0432

Table C-12

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 0.48 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
42-45	0.4670E 06	198.5	0.142	0.250	2.0104	0.2163
42-56	0.4670E 06	197.5	0.250	0.346	1.8437	0.1081
44-45	0.4735E 06	198.9	0.117	0.224	2.0104	0.1982
44-56	0.4735E 06	197.8	0.224	0.318	1.8437	0.1262
85-12	0.5270E 06	199.7	0.111	0.172	1.9844	0.1910
85-24	0.5270E 06	200.4	0.171	0.299	3.9948	0.2235
85-45	0.5270E 06	200.5	0.299	0.364	2.0104	0.1261
85-56	0.5270E 06	200.1	0.364	0.425	1.8437	0.0901
87-24	0.4600E 06	199.7	0.029	0.090	3.9948	0.1189
87-45	0.4600E 06	199.7	0.090	0.124	2.0104	0.0613
91-12	0.4712E 06	200.5	0.015	0.064	1.9844	0.1622
91-24	0.4712E 06	200.9	0.064	0.167	3.9948	0.1442
91-45	0.4712E 06	200.7	0.167	0.220	2.0104	0.1261
93-12	0.6052E 06	198.1	0.011	0.084	1.9844	0.2343
93-24	0.6052E 06	199.1	0.084	0.242	3.9948	0.3064
93-45	0.6052E 06	199.4	0.242	0.323	2.0104	0.1802
93-56	0.6052E 06	199.1	0.323	0.400	1.8437	0.1081
96-24	0.6412E 06	198.9	0.061	0.245	3.9948	0.3064
96-45	0.6412E 06	199.6	0.245	0.339	2.0104	0.2523
96-56	0.6412E 06	199.5	0.339	0.434	1.8437	0.1802

Table C-13

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
35-24	0.8759E 06	200.5	0.043	0.183	3.9948	0.4686
35-45	0.8759E 06	200.4	0.183	0.259	2.0104	0.3244
35-56	0.8759E 06	199.0	0.259	0.319	1.8437	0.2703
37-12	0.8984E 06	199.3	0.005	0.058	1.9844	0.2883
37-24	0.8984E 06	199.5	0.058	0.167	3.9948	0.3785
37-45	0.8984E 06	199.4	0.167	0.227	2.0104	0.3064
37-56	0.8984E 06	198.2	0.227	0.271	1.8437	0.2163
49-45	0.8974E 06	190.6	0.020	0.056	2.0104	0.2703
49-56	0.8974E 06	190.0	0.056	0.085	1.8437	0.1442
84-12	0.9999E 06	201.2	0.221	0.252	1.9844	0.2883
84-24	0.9999E 06	201.7	0.252	0.320	3.9948	0.4505
84-45	0.9999E 06	201.6	0.320	0.355	2.0104	0.3424
84-56	0.9999E 06	201.2	0.355	0.386	1.8437	0.2883
90-12	0.9429E 06	201.0	0.123	0.146	1.9844	0.2343
90-24	0.9429E 06	201.4	0.146	0.199	3.9948	0.3064
90-45	0.9429E 06	201.2	0.199	0.227	2.0104	0.2343
90-56	0.9429E 06	200.6	0.227	0.248	1.8437	0.1802
92-12	0.9115E 06	195.6	0.002	0.041	1.9844	0.3424
92-24	0.9115E 06	197.7	0.041	0.142	3.9948	0.3965
92-45	0.9115E 06	198.3	0.142	0.196	2.0104	0.2343
92-56	0.9115E 06	198.0	0.196	0.246	1.8437	0.2163
97-24	0.8887E 06	199.3	0.017	0.150	3.9948	0.4686
97-45	0.8887E 06	199.8	0.150	0.219	2.0104	0.3244
97-56	0.8887E 06	199.6	0.219	0.287	1.8437	0.2703
100-24	0.9317E 06	201.2	0.022	0.094	3.9948	0.3244
100-45	0.9317E 06	201.3	0.094	0.132	2.0104	0.2343
100-56	0.9317E 06	201.0	0.132	0.163	1.8437	0.1622
103-12	0.9027E 06	198.6	0.055	0.074	1.9844	0.2343
103-24	0.9027E 06	199.8	0.074	0.127	3.9948	0.2040
103-45	0.9027E 06	199.7	0.127	0.155	2.0104	0.1820
105-12	0.9155E 06	202.3	0.065	0.093	1.9844	0.1982
105-24	0.9155E 06	200.4	0.093	0.123	3.9948	0.1680
105-45	0.9155E 06	200.6	0.122	0.140	2.0104	0.2173

Table C-14

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 1.70 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
34-45	0.1658E 07	199.8	0.010	0.055	2.0104	0.6848
34-56	0.1658E 07	198.3	0.055	0.082	1.8437	0.5947
38-45	0.1670E 07	199.8	0.017	0.053	2.0104	0.5947
38-56	0.1670E 07	198.5	0.053	0.075	1.8437	0.5046
41-45	0.1692E 07	199.1	0.049	0.085	2.0104	0.6488
41-56	0.1692E 07	197.8	0.085	0.105	1.8437	0.5406
50-56	0.1715E 07	188.2	0.020	0.035	1.8437	0.4145
53-12	0.1656E 07	190.4	0.012	0.019	1.9844	0.3424
53-24	0.1656E 07	191.4	0.019	0.048	3.9948	0.5767
53-45	0.1656E 07	191.3	0.048	0.066	2.0104	0.3965
53-56	0.1656E 07	190.5	0.066	0.077	1.8437	0.3244
98-45	0.1730E 07	200.3	0.065	0.101	2.0104	0.7389
98-56	0.1730E 07	199.9	0.101	0.137	1.8437	0.6848
101-24	0.1730E 07	200.4	0.0	0.040	3.9948	0.8110
101-45	0.1730E 07	200.3	0.040	0.063	2.0104	0.5587
101-56	0.1730E 07	199.7	0.063	0.080	1.8437	0.5587
104-12	0.1710E 07	199.5	0.031	0.040	1.9844	0.4902
104-24	0.1710E 07	200.2	0.040	0.070	3.9948	0.5911
104-45	0.1710E 07	199.9	0.070	0.087	2.0104	0.4145
104-56	0.1710E 07	199.3	0.087	0.097	1.8437	0.3244
106-12	0.1711E 07	199.4	0.047	0.049	1.9844	0.4325
106-24	0.1711E 07	200.4	0.049	0.068	3.9948	0.4866
106-45	0.1711E 07	200.1	0.068	0.079	2.0104	0.3784
106-56	0.1711E 07	199.4	0.079	0.079	1.8437	0.2883

Table C-15

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
36-12	0.3387E 07	196.9	0.030	0.038	1.9844	1.0633
48-12	0.3371E 07	199.2	0.016	0.020	1.9844	1.0993
48-24	0.3371E 07	200.9	0.020	0.051	3.9948	2.3068
51-24	0.3438E 07	191.8	0.025	0.042	3.9948	1.9643
52-12	0.3370E 07	198.4	0.016	0.021	1.9844	0.9191
52-24	0.3370E 07	198.7	0.021	0.038	3.9948	1.7661

Table C-16

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 250^{\circ}F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P,TOTAL (PSI)
FR152-45	0.1223E 06	251.5	0.111	0.182	2.0104	0.0108
FR158-56	0.1116E 06	250.3	0.335	0.540	1.8437	0.0144

Table C-17

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 250^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
FR153-56	0.2176E 06	249.4	0.245	0.409	1.8437	0.0144

Table C-18

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
FR 75-12	0.4624E 06	244.1	0.421	0.469	1.9844	0.1766
FR 75-24	0.4624E 06	246.6	0.469	0.601	3.9948	0.1658
FR 75-45	0.4624E 06	246.6	0.601	0.668	2.0104	0.1514
FR 75-56	0.4624E 06	246.2	0.668	0.728	1.8437	0.0469
FR 76-12	0.4650E 06	245.2	0.533	0.576	1.9844	0.1622
FR 76-24	0.4650E 06	244.4	0.576	0.649	3.9948	0.1766
FR 76-45	0.4650E 06	244.8	0.649	0.689	2.0104	0.0937
FR 76-56	0.4650E 06	244.3	0.689	0.725	1.8437	0.0685
FR148-24	0.4742E 06	253.8	0.027	0.246	3.9948	0.0721
FR148-45	0.4742E 06	253.3	0.246	0.358	2.0104	0.1117
FR148-56	0.4742E 06	252.8	0.358	0.458	1.8437	0.0432
FR154-24	0.4923E 06	250.5	0.051	0.203	3.9948	0.1009
FR154-45	0.4923E 06	250.2	0.203	0.282	2.0104	0.1225
FR154-56	0.4923E 06	249.7	0.282	0.350	1.8437	0.0360
FR160-12	0.5097E 06	246.6	0.090	0.132	1.9844	0.1370
FR160-24	0.5097E 06	247.0	0.132	0.226	3.9948	0.0901
FR160-45	0.5097E 06	246.9	0.226	0.275	2.0104	0.1153

Table C-19

PRESSURE DROP - EXPERIMENTAL DATA,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 250°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
108-12	0.9775E 06	248.8	0.012	0.072	1.9844	0.2883
108-24	0.9775E 06	249.1	0.072	0.199	3.9948	0.3424
108-45	0.9775E 06	249.4	0.199	0.265	2.0104	0.2523
108-56	0.9775E 06	249.0	0.265	0.323	1.8437	0.1802
FR 67-12	0.9799E 06	248.2	0.030	0.087	1.9844	0.3064
FR 67-24	0.9799E 06	248.8	0.087	0.217	3.9948	0.3965
FR 67-45	0.9799E 06	248.9	0.216	0.283	2.0104	0.2703
FR 67-56	0.9799E 06	248.6	0.283	0.345	1.8437	0.1081
FR 71-24	0.9807E 06	243.8	0.008	0.094	3.9948	0.2703
FR 71-45	0.9807E 06	243.7	0.094	0.139	2.0104	0.2523
FR 71-56	0.9807E 06	243.1	0.139	0.174	1.8437	0.1442
FR 74-12	0.8518E 06	242.4	0.178	0.207	1.9844	0.2703
FR 74-24	0.8518E 06	243.2	0.207	0.278	3.9948	0.2523
FR 74-45	0.8518E 06	242.9	0.278	0.317	2.0104	0.2343
FR 74-56	0.8518E 06	242.3	0.317	0.344	1.8437	0.1622
FR 77-12	0.8953E 06	251.9	0.227	0.257	1.9844	0.1802
FR 77-24	0.8953E 06	250.9	0.257	0.299	3.9948	0.1910
FR 77-45	0.8953E 06	250.7	0.299	0.324	2.0104	0.1334
FR 77-56	0.8953E 06	250.0	0.324	0.333	1.8437	0.0721
FR149-12	0.8924E 06	249.9	0.151	0.192	1.9844	0.2307
FR149-24	0.8924E 06	251.6	0.192	0.303	3.9948	0.2667
FR149-45	0.8924E 06	251.8	0.303	0.362	2.0104	0.1694
FR149-56	0.8924E 06	251.4	0.362	0.411	1.8437	0.1261
FR155-12	0.8942E 06	254.6	0.102	0.140	1.9844	0.1730
FR155-24	0.8942E 06	255.2	0.140	0.229	3.9948	0.2235
FR155-45	0.8942E 06	254.8	0.229	0.275	2.0104	0.1802
FR155-56	0.8942E 06	254.3	0.275	0.308	1.8437	0.0829
FR161-45	0.8513E 06	246.9	0.013	0.046	2.0104	0.1586
FR161-56	0.8513E 06	246.3	0.046	0.060	1.8437	0.0505

Table C-20

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 1.70 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
109-24	0.1812E 07	250.3	0.011	0.081	3.9948	0.8038
109-45	0.1812E 07	250.5	0.080	0.118	2.0104	0.5046
109-56	0.1812E 07	250.1	0.118	0.147	1.8437	0.4505
FR 72-45	0.1804E 07	243.7	0.028	0.050	2.0104	0.4505
FR 72-56	0.1804E 07	243.8	0.050	0.069	1.8437	0.3784
FR 73-12	0.1858E 07	236.8	0.044	0.058	1.9844	0.5767
FR 73-24	0.1858E 07	236.9	0.058	0.091	3.9948	0.7389
FR 73-45	0.1858E 07	236.6	0.091	0.111	2.0104	0.4866
FR 73-56	0.1858E 07	235.9	0.111	0.120	1.8437	0.3424
FR 78-12	0.1685E 07	248.7	0.142	0.150	1.9844	0.3280
FR 78-24	0.1685E 07	249.1	0.150	0.159	3.9948	0.3857
FR 78-45	0.1685E 07	250.4	0.159	0.172	2.0104	0.2955
FR 78-56	0.1685E 07	250.0	0.172	0.173	1.8437	0.2343
FR150-12	0.1632E 07	245.0	0.129	0.139	1.9844	0.3280
FR150-24	0.1632E 07	247.4	0.139	0.153	3.9948	0.5370
FR150-56	0.1632E 07	254.1	0.167	0.189	1.8437	0.2703
FR156-12	0.1956E 07	249.2	0.240	0.245	1.9844	0.3244
FR156-24	0.1956E 07	250.9	0.245	0.258	3.9948	0.3965
FR156-45	0.1956E 07	253.9	0.258	0.278	2.0104	0.3532
FR156-56	0.1956E 07	253.8	0.278	0.287	1.8437	0.2235
FR162-45	0.1635E 07	249.6	0.004	0.025	2.0104	0.3604
FR162-56	0.1635E 07	248.8	0.025	0.030	1.8437	0.2343

Table C-21

PRESSURE DROP - EXPERIMENTAL DATA,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	LENGTH (FT)	DELTA-P, TOTAL (PSI)
FR151-12	0.3098E 07	243.1	0.112	0.119	1.9844	0.8290
FR151-24	0.3098E 07	244.1	0.119	0.123	3.9948	1.0830
FR151-45	0.3098E 07	247.2	0.123	0.129	2.0104	0.8470
FR157-12	0.3337E 07	242.2	0.140	0.143	1.9844	0.9155
FR157-24	0.3337E 07	243.2	0.143	0.148	3.9948	1.2471
FR157-45	0.3337E 07	245.0	0.148	0.153	2.0104	0.9912

Table C-22

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.12 x 10^6 lbm/hr-sq ft, T ≈ 100°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
61-24	0.7616	0.0213	0.0131	5.891	0.0088	0.0280	6.574	0.0136	0.0290	6.319
61-45	0.4571	0.0213	0.0161	9.354	0.0180	0.0383	9.824	0.0251	0.0159	8.795
63-24	0.7468	0.0138	0.0157	9.716	0.0060	0.0331	10.004	0.0080	0.0288	9.931
63-45	0.5313	0.0138	0.0142	6.886	0.0098	0.0327	7.479	0.0130	0.0147	7.005
65-24	0.5251	0.0359	0.0273	10.532	0.0207	0.0609	11.148	0.0343	0.0379	10.599
65-45	0.2912	0.0362	0.0278	33.220	0.0425	0.0638	32.635	0.0527	0.0285	31.662
69-45	0.3081	0.0502	0.0253	27.586	0.0613	0.0579	26.382	0.0406	0.0321	28.582

Table C-23

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 100^\circ F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
60-45	0.8696	0.0423	0.0265	4.886	0.0187	0.0577	5.696	0.0227	0.0628	5.567
62-24	1.0022	0.0259	0.0389	4.098	0.0096	0.0801	4.420	0.0106	0.1020	4.400
62-45	0.7864	0.0258	0.0311	4.250	0.0122	0.0680	4.832	0.0135	0.0651	4.777
62-56	0.6945	0.0260	0.0334	3.649	0.0138	0.0747	4.372	0.0159	0.0613	4.259
64-24	0.7999	0.0739	0.0618	5.963	0.0297	0.1288	6.505	0.0378	0.1704	6.409
64-45	0.5581	0.0748	0.0600	6.834	0.0496	0.1374	7.606	0.0668	0.0839	7.089
64-56	0.4595	0.0742	0.0668	5.828	0.0593	0.1564	6.614	0.0790	0.0803	5.554
68-24	0.8795	0.0932	0.0349	7.242	0.0360	0.0708	7.896	0.0489	0.1345	7.754
68-45	0.5028	0.0960	0.0555	11.541	0.0767	0.1295	12.058	0.1054	0.0714	11.279

Table C-24

PRESSURE DROP - CORRELATED RESULTS, $G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 100^{\circ}F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [illegible] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
58-24	0.9930	0.1692	0.1103	3.578	0.0618	0.2263	4.290	0.0723	0.4164	4.226
58-45	0.6801	0.1729	0.1184	4.578	0.0963	0.2669	5.618	0.1204	0.2402	5.312
58-56	0.5629	0.1714	0.1349	3.912	0.1157	0.3130	5.083	0.1503	0.2112	4.393
59-24	1.5792	0.0823	0.0603	2.218	0.0340	0.1181	2.640	0.0357	0.2180	2.627
59-45	1.0769	0.0827	0.0666	1.557	0.0309	0.1406	2.714	0.0342	0.2000	2.655
59-56	0.9169	0.0823	0.0758	3.331	0.0335	0.1642	4.078	0.0371	0.2088	4.027
66-24	1.0599	0.1330	0.1230	3.557	0.0459	0.2468	4.070	0.0525	0.4326	4.034
66-45	0.7824	0.1350	0.1141	3.543	0.0587	0.2458	4.552	0.0706	0.2843	4.410
66-56	0.6744	0.1342	0.1264	2.474	0.0654	0.2801	3.908	0.0841	0.2629	3.576
67-24	1.1344	0.1773	0.0923	4.664	0.0646	0.1845	5.192	0.0714	0.4008	5.162
67-45	0.7388	0.1831	0.1165	4.374	0.0975	0.2578	5.424	0.1148	0.2654	5.229
67-56	0.6067	0.1814	0.1355	4.156	0.1206	0.3098	5.205	0.1453	0.2301	4.807
70-24	0.8522	0.2303	0.1584	5.384	0.0914	0.3309	6.012	0.1126	0.5323	5.920
70-45	0.5916	0.2340	0.1603	5.858	0.1608	0.3676	6.711	0.2030	0.2620	6.234

Table C-25

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.12 x 10^6 lbm/hr-sq ft, T ≈ 150°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [[illegible]] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
12-45	1.1245	0.0077	0.0041	6.133	0.0036	0.0097	6.511	0.0046	0.0082	6.419
40-24	0.9758	0.0273	0.0104	3.381	0.0152	0.0253	4.293	0.0291	0.0219	3.220
40-45	0.5487	0.0130	0.0097	8.429	0.0110	0.0245	8.703	0.0150	0.0114	8.151
75-24	0.7879	0.0185	0.0095	6.593	0.0091	0.0227	7.054	0.0149	0.0215	6.772
76-45	0.4061	0.0187	0.0125	13.545	0.0206	0.0309	13.239	0.0263	0.0152	12.268
82-45	1.0146	0.0077	0.0051	6.644	0.0039	0.0123	6.959	0.0049	0.0095	6.872

Table C-26

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.24×10^6 lbm/hr-sq ft, T ≈ 150°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
6-45	1.2198	0.0275	0.0146	4.208	0.0125	0.0341	4.684	0.0143	0.0403	4.629
13-45	1.3524	0.0163	0.0122	3.588	0.0072	0.0280	3.933	0.0079	0.0304	3.907
18-45	0.8603	0.0367	0.0209	6.348	0.0204	0.0515	6.812	0.0260	0.0469	6.657
23-24	0.8795	0.1074	0.0406	4.471	0.0620	0.0999	5.320	0.1166	0.1019	4.276
46-45	1.1701	0.0269	0.0146	5.471	0.0124	0.0345	5.859	0.0144	0.0397	5.806
54-45	0.9919	0.0271	0.0188	4.232	0.0139	0.0454	4.705	0.0157	0.0463	4.642
73-24	0.9084	C.0495	0.0246	5.067	0.0234	0.0579	5.550	0.0319	0.0796	5.398
73-45	0.5168	0.0503	0.C316	8.891	0.0455	0.0800	9.116	0.0608	0.0475	8.378
73-56	0.3671	C.C495	0.0364	6.770	0.0580	0.0883	5.345	0.0703	0.0517	1.897
75-45	0.8660	0.0348	0.0190	4.718	0.0194	0.0471	5.335	0.0252	0.0423	5.112
75-56	0.6552	0.0349	0.0236	0.831	0.0249	0.0597	2.598	0.0329	0.0401	1.384
78-45	1.2935	0.0251	0.0122	2.944	0.0112	0.0284	3.578	0.0131	0.0341	3.501
78-56	0.9423	0.0256	0.0162	1.133	0.0129	0.0396	2.504	0.0158	0.0383	2.265
81-45	1.3829	0.0147	0.0109	4.547	0.0065	0.0251	4.812	0.0073	0.0267	4.789

Table C-27

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 150^{\circ}F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
7-24	1.7886	0.1032	0.0591	1.937	0.0524	0.1288	2.399	0.0524	0.2295	2.400
7-45	1.2083	0.0517	0.0465	2.586	0.0236	0.1091	3.036	0.0256	0.1313	3.006
7-56	1.0313	0.0518	0.0522	1.476	0.0252	0.1257	2.241	0.0268	0.1429	2.204
16-24	2.0085	0.0662	0.0563	1.389	0.0344	0.1202	1.759	0.0350	0.1757	1.752
16-45	1.4594	0.0327	0.0391	1.761	0.0145	0.0889	2.131	0.0166	0.0986	2.091
19-45	1.1103	0.0725	0.0512	2.900	0.0342	0.1219	3.466	0.0373	0.1514	3.423
19-56	0.9162	0.0728	0.0597	2.033	0.0385	0.1463	2.852	0.0419	0.1603	2.783
27-45	0.9072	0.1058	0.0652	3.989	0.0568	0.1596	4.608	0.0644	0.1804	4.518
27-56	0.7307	0.1062	0.0767	1.713	0.0686	0.1927	2.910	0.0813	0.1697	2.569
43-45	1.2368	0.0769	0.0495	2.332	0.0346	0.1157	2.976	0.0382	0.1512	2.926
43-56	0.9988	0.0774	0.0594	1.749	0.0382	0.1436	2.640	0.0416	0.1684	2.575
71-45	0.8104	0.1240	0.0689	3.437	0.0734	0.1718	4.277	0.0908	0.1698	4.009
71-56	0.6348	0.1243	0.0814	4.570	0.0923	0.2069	5.191	0.1169	0.1600	4.720
72-24	1.4566	0.1034	0.0539	3.106	0.0600	0.1133	3.376	0.0522	0.2150	3.423
72-45	0.9001	0.1052	0.0672	3.945	0.0569	0.1656	4.546	0.0650	0.1841	4.450
72-56	0.7324	0.1036	0.0775	3.306	0.0672	0.1951	4.015	0.0788	0.1729	3.804
74-24	1.5494	0.0754	0.0617	2.536	0.0375	0.1332	2.794	0.0389	0.1983	2.785
74-45	1.0573	0.0758	0.0604	3.476	0.0363	0.1455	3.935	0.0399	0.1751	3.895
74-56	0.8926	0.0772	0.0678	*******	0.0413	0.1671	1.811	0.0451	0.1808	1.696
77-24	1.7875	0.0504	0.0487	2.248	0.0277	0.1011	2.436	0.0271	0.1363	2.441
77-45	1.2228	0.0515	0.0457	2.857	0.0229	0.1075	3.268	0.0258	0.1287	3.229
77-56	1.0418	0.0511	0.0508	1.487	0.0240	0.1226	2.249	0.0265	0.1393	2.188
80-24	1.5881	0.0303	0.0638	1.119	0.0139	0.1409	1.371	0.0163	0.1506	1.337
80-45	1.2803	0.0309	0.0449	2.365	0.0131	0.1049	2.656	0.0156	0.1146	2.617
80-56	1.1508	0.0307	0.0469	0.708	0.0130	0.1117	1.475	0.0155	0.1214	1.392

Table C-28

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 150°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [illegible] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
8-24	2.5086	0.1951	0.1343	1.479	0.1333	0.2395	1.715	0.1130	0.3945	1.785
8-45	1.6610	C.0982	0.0986	2.001	0.0446	0.1969	2.324	0.0535	0.2317	2.273
8-56	1.4279	0.0962	0.1096	1.740	0.0420	0.2258	2.153	0.0499	0.2699	2.098
20-12	1.3342	C.1374	0.1295	2.149	0.0606	0.3000	2.592	0.0701	0.3806	2.541
20-24	1.0874	0.2772	0.3296	1.991	0.1353	0.7883	2.469	0.1508	1.0879	2.422
20-45	0.9241	C.1384	0.2010	3.042	0.0714	0.4917	3.391	0.0754	0.5751	3.371
20-56	0.8463	C.1393	0.2065	2.565	0.0741	0.5112	3.024	0.0812	0.5646	2.978
24-12	1.6983	C.1984	0.0959	2.014	0.0906	0.2125	2.622	0.1032	0.3106	2.559
24-24	1.1571	0.3918	0.2964	2.187	0.1856	0.7014	2.808	0.2153	1.0829	2.727
24-45	0.9108	0.1939	0.1979	3.001	0.1015	0.4843	3.500	0.1079	0.5777	3.468
24-56	0.8043	0.2009	0.2118	2.613	0.1128	0.5270	3.256	0.1255	0.5670	3.171
28-12	1.5454	0.2418	0.1096	1.779	0.1056	0.2490	2.606	0.1225	0.3772	2.519
28-24	1.0697	0.4935	0.3250	1.729	0.2424	0.7795	2.630	0.2960	1.1732	2.465
28-45	0.8239	0.2444	0.2207	3.465	0.1347	0.5476	4.025	0.1512	0.6129	3.946
28-56	0.7180	C.2479	0.2387	2.835	0.1475	0.6014	3.589	0.1796	0.5731	3.367
47-24	2.0512	0.1953	0.1581	1.453	0.1021	0.3013	1.818	0.1080	0.4653	1.797
47-45	1.5104	C.1002	0.1094	2.373	0.0454	0.2351	2.670	0.0527	0.2835	2.632
47-56	1.3270	0.0937	0.1181	1.642	0.0427	0.2455	2.071	0.0474	0.2960	2.035
55-12	2.0981	0.0971	C.0805	2.128	0.0483	0.1545	2.398	0.0559	0.1783	2.358
55-24	1.5686	C.0978	0.1930	1.777	0.0456	0.3842	1.963	0.0540	0.4521	1.934
55-45	1.2768	C.0974	0.1345	2.333	0.0457	0.2818	2.630	0.0495	0.3399	2.609
55-56	1.1550	0.0972	0.1406	1.816	0.0472	0.2996	2.220	0.0490	0.3606	2.207
83-12	1.6066	0.0563	0.1028	2.555	0.0250	0.2083	2.710	0.0314	0.2201	2.679
83-24	1.4148	0.0557	0.2251	2.203	0.0251	0.4629	2.295	0.0309	0.5006	2.277
83-45	1.2629	0.0572	0.1351	2.834	0.0256	0.2851	2.986	0.0300	0.3199	2.966
83-56	1.1894	C.0570	0.1342	1.271	0.0255	0.2862	1.616	0.0296	0.3240	1.575

Table C-29

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 1.70 x 10^6 lbm/hr-sq ft, T ≈ 150°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
3-12	1.6246	0.1188	0.3447	1.603	0.0575	0.5391	1.745	0.0724	0.5675	1.711
3-24	1.4906	0.2345	0.7637	1.162	0.1139	1.2125	1.352	0.1435	1.4317	1.308
3-45	1.3687	0.1163	0.4197	1.730	0.0579	0.6745	1.862	0.0692	0.7603	1.837
3-56	1.3100	0.1186	0.4171	1.578	0.0551	0.6774	1.749	0.0696	0.7828	1.712
10-24	3.1987	0.3707	0.3642	1.007	0.3062	0.5845	1.121	0.2527	0.7692	1.208
10-45	2.1351	0.1915	0.2385	1.723	0.1021	0.4484	1.916	0.1236	0.4602	1.871
10-56	1.8327	0.1861	0.2665	1.577	0.0888	0.4968	1.829	0.1133	0.5376	1.769
15-12	2.5240	0.1142	0.2789	1.525	0.0663	0.3149	1.643	0.0776	0.2834	1.616
15-24	2.0963	0.2204	0.4833	1.105	0.1161	0.7136	1.279	0.1445	0.7627	1.234
15-45	1.7941	0.1113	0.2802	1.619	0.0512	0.4279	1.764	0.0698	0.4265	1.721
15-56	1.6708	0.1142	0.2868	1.437	0.0479	0.4471	1.633	0.0691	0.4675	1.573
21-12	1.3918	0.2544	0.3728	1.632	0.1095	0.7735	1.983	0.1435	0.9306	1.907
21-24	1.2189	0.5199	0.8725	1.473	0.2454	1.8500	1.853	0.2999	2.5438	1.784
21-45	1.0796	0.2638	0.5058	2.196	0.1202	1.0904	2.495	0.1489	1.3541	2.438
21-56	1.0095	0.2635	0.5117	2.116	0.1153	1.1154	2.475	0.1520	1.3718	2.391
25-12	3.4742	0.3770	0.1791	1.482	0.3152	0.3263	1.633	0.2498	0.4337	1.778
25-24	1.8550	0.7562	0.5445	1.504	0.3797	1.1783	1.946	0.4367	1.9428	1.886
25-45	1.3747	0.3782	0.3822	2.283	0.1705	0.8769	2.651	0.2066	1.1237	2.591
29-12	2.8416	0.4490	0.2095	1.182	0.2729	0.4213	1.658	0.2808	0.5686	1.640
29-24	1.7198	0.9150	0.5827	1.446	0.4460	1.2880	2.005	0.5163	2.2459	1.931
29-45	1.2698	0.4574	0.4048	2.305	0.1940	0.9441	2.774	0.2475	1.2603	2.685
31-12	3.2026	0.5781	0.1938	1.513	0.4659	0.3647	1.769	0.3592	0.5899	1.983
31-24	1.5908	1.1450	0.6478	1.779	0.5439	1.4478	2.378	0.6301	2.7017	2.301
39-12	2.0834	0.1792	0.2599	1.498	0.0880	0.4447	1.718	0.1139	0.4547	1.659
39-24	1.7664	0.3864	0.5921	1.214	0.1868	1.1362	1.506	0.2355	1.4357	1.440
39-45	1.4967	0.1859	0.3535	1.798	0.0837	0.6798	2.020	0.1084	0.7700	1.969
39-56	1.3810	0.1835	0.3651	1.726	0.0788	0.7146	1.987	0.1042	0.8354	1.927
56-12	2.2747	0.1858	0.2422	1.533	0.0960	0.4319	1.746	0.1189	0.4407	1.694
56-24	1.9056	0.1877	0.5148	1.484	0.0969	0.9889	1.601	0.1220	1.0345	1.570
56-45	1.6019	0.1875	0.3292	1.794	0.0901	0.6379	2.006	0.1101	0.7013	1.965
56-56	1.4632	0.1877	0.3390	1.809	0.0900	0.6552	2.044	0.1069	0.7605	2.005

Table C-30

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T = 150^\circ F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
11-12	2.2136	0.3653	0.8345	1.107	0.1908	1.1426	1.273	0.2644	1.0726	1.206
11-24	1.9632	0.7233	1.7693	1.417	0.3921	2.3671	1.545	0.5401	2.5542	1.459
17-12	2.7401	0.2139	0.7050	1.146	0.1326	0.5593	1.221	0.1595	0.4601	1.197
17-24	2.3744	0.4293	1.4744	1.305	0.2656	1.1872	1.373	0.3225	1.0099	1.359
22-12	1.7399	0.4998	1.0261	1.185	0.2218	1.8242	1.433	0.3601	1.9476	1.316
26-12	2.0419	0.7389	0.8936	0.596	0.3458	1.6443	1.121	0.5125	1.7607	0.935
26-24	1.8409	1.4728	1.9638	1.117	0.7401	3.8939	1.449	1.0793	5.2025	1.306
30-12	1.9368	0.8990	0.9184	0.783	0.4160	1.8100	1.332	0.6090	2.0933	1.145
32-12	1.7692	1.0863	1.0451	0.529	0.4786	2.0940	1.297	0.7298	2.5362	1.059
57-12	2.1625	0.3621	0.8344	1.132	0.1824	1.1748	1.292	0.2610	1.0881	1.224
57-24	1.9337	0.3684	1.7251	1.463	0.2025	2.5862	1.523	0.2958	2.5217	1.490

Table C-31

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
86-12	1.0033	0.0241	0.0154	6.421	0.0147	0.0373	6.637	0.0181	0.0349	6.559
86-24	0.6819	0.0239	0.0397	4.563	0.0180	0.0976	4.721	0.0223	0.0743	4.605
86-45	0.4866	0.0241	0.0279	9.148	0.0267	0.0661	9.005	0.0299	0.0429	8.828
86-56	0.3752	0.0241	0.0289	8.141	0.0322	0.0645	6.952	0.0327	0.0450	6.879
88-24	1.3933	0.0099	0.0203	1.760	0.0052	0.0453	1.909	0.0058	0.0421	1.890
88-45	1.0750	0.0101	0.0144	3.509	0.0058	0.0345	3.677	0.0065	0.0304	3.650
94-24	0.9421	0.0162	0.0269	3.196	0.0093	0.0646	3.375	0.0112	0.0598	3.328
94-45	0.7074	0.0164	0.0194	6.218	0.0125	0.0478	6.373	0.0156	0.0337	6.249
94-56	0.5963	0.0163	0.0204	3.462	0.0139	0.0496	3.707	0.0175	0.0323	3.327
95-45	0.6850	0.0332	0.0203	6.267	0.0271	0.0498	6.515	0.0365	0.0362	6.128
120-12	1.1154	0.0080	0.0097	5.654	0.0044	0.0231	5.777	0.0052	0.0197	5.751
120-24	0.8920	0.0080	0.0224	2.037	0.0049	0.0544	2.215	0.0056	0.0437	2.173
120-45	0.7355	0.0082	0.0143	4.417	0.0060	0.0352	4.569	0.0072	0.0238	4.483
121-12	1.0479	0.0068	0.0106	6.298	0.0036	0.0255	6.396	0.0044	0.0210	6.370
121-45	0.7504	0.0068	0.0146	6.107	0.0042	0.0357	6.227	0.0058	0.0247	6.156
123-12	1.1268	0.0038	0.0098	6.017	0.0021	0.0233	6.069	0.0023	0.0184	6.062
123-45	0.9041	0.0038	0.0123	4.233	0.0023	0.0299	4.310	0.0026	0.0226	4.295

Table C-32

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.48 x 10^6 lbm/hr-sq ft, T ≈ 200°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
42-45	1.1190	0.0620	0.0378	3.926	0.0350	0.0904	4.256	0.0406	0.1030	4.190
42-56	0.8689	0.0620	0.0448	2.534	0.0410	0.1102	3.057	0.0483	0.1053	2.885
44-45	1.2082	0.0625	0.0359	3.537	0.0342	0.0848	3.889	0.0397	0.1001	3.823
44-56	0.9224	C.0629	0.0433	2.836	0.0397	0.1060	3.314	0.0466	0.1061	3.180
85-12	1.3402	C.0472	0.0388	3.229	0.0249	0.0896	3.470	0.0289	0.0992	3.429
85-24	1.0148	0.0467	0.0932	2.798	0.0264	0.2222	2.954	0.0298	0.2367	2.929
85-45	0.8188	C.0471	0.0619	2.976	0.0325	0.1520	3.240	0.0359	0.1403	3.181
85-56	0.7230	0.0473	0.0635	2.501	0.0355	0.1562	2.826	0.0411	0.1320	2.677
87-24	2.0593	0.0170	0.0398	1.991	0.0111	0.0784	2.049	0.0109	0.0724	2.050
87-45	1.5468	0.0175	0.0271	1.927	0.0092	0.0606	2.102	0.0107	0.0568	2.072
91-12	2.5061	C.0306	0.0203	3.082	0.0210	0.0393	3.193	0.0192	0.0439	3.213
91-24	1.4927	0.0302	0.0524	2.177	0.0163	0.1148	2.306	0.0184	0.1206	2.287
91-45	1.1357	C.0303	0.0377	3.058	0.0173	0.0898	3.258	0.0188	0.0915	3.236
93-12	2.2811	0.0772	0.0333	2.722	0.0513	0.0662	2.938	0.0481	0.0904	2.963
93-24	1.2409	0.0768	0.0926	2.603	0.0405	0.2098	2.802	0.0473	0.2545	2.766
93-45	0.9050	0.0776	0.0715	2.818	0.0474	0.1740	3.205	0.0557	0.1783	3.104
93-56	0.7675	0.0777	0.0762	1.781	0.0524	0.1871	2.409	0.0651	0.1703	2.117
96-24	1.2833	0.1028	0.0969	2.305	0.0551	0.2163	2.561	0.0632	0.2849	2.519
96-45	0.8863	0.1030	0.0814	3.270	0.0659	0.1988	3.653	0.0771	0.2041	3.542
96-56	0.7325	C.1045	0.0885	2.765	0.0770	0.2179	3.229	0.0941	0.1949	2.949

Table C-33

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 200°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
35-24	1.5112	0.2910	0.1735	1.565	0.1619	0.3907	2.057	0.1868	0.6146	1.971
35-45	1.0520	0.1447	0.1221	2.496	0.0871	0.2944	2.867	0.0929	0.3381	2.832
35-56	0.8904	0.1445	0.1333	2.369	0.0968	0.3275	2.782	0.1029	0.3441	2.733
37-12	2.7844	0.1188	0.0601	1.962	0.0953	0.1113	2.093	0.0763	0.1418	2.183
37-24	1.5070	0.2391	0.1820	1.356	0.1311	0.4106	1.807	0.1507	0.5929	1.734
37-45	1.1156	0.1178	0.1202	2.433	0.0677	0.2874	2.738	0.0732	0.3251	2.706
37-56	0.9734	0.1184	0.1280	1.946	0.0728	0.3122	2.356	0.0779	0.3390	2.313
49-45	2.4700	0.0834	0.0650	2.071	0.0552	0.1284	2.222	0.0548	0.1367	2.224
49-56	1.8412	0.0816	0.0746	1.291	0.0439	0.1634	1.635	0.0508	0.1715	1.577
84-12	1.0139	0.0886	0.1550	2.397	0.0508	0.3731	2.614	0.0571	0.3981	2.579
84-24	0.9044	0.0877	0.3348	2.418	0.0556	0.8130	2.523	0.0596	0.8560	2.510
84-45	0.8113	0.0884	0.1927	3.051	0.0622	0.4725	3.205	0.0648	0.4673	3.190
84-56	0.7592	0.0890	0.1881	2.957	0.0646	0.4622	3.102	0.0695	0.4370	3.097
90-12	1.3811	0.0607	0.1063	2.103	0.0326	0.2444	2.267	0.0378	0.2489	2.238
90-24	1.2126	0.0601	0.2307	1.838	0.0341	0.5392	1.932	0.0377	0.5569	1.919
90-45	1.0761	0.0603	0.1350	2.277	0.0368	0.3234	2.426	0.0378	0.3379	2.420
90-56	1.0100	0.0598	0.1326	2.036	0.0373	0.3207	2.218	0.0382	0.3336	2.212
92-12	3.2953	0.1166	0.0599	2.219	0.1065	0.1069	2.268	0.0780	0.1369	2.401
92-24	1.6630	0.1161	0.1506	1.861	0.0670	0.3153	2.017	0.0744	0.3675	1.994
92-45	1.2138	0.1174	0.1120	1.835	0.0610	0.2626	2.234	0.0728	0.2992	2.157
92-56	1.0441	0.1174	0.1189	1.867	0.0630	0.2855	2.326	0.0752	0.3173	2.231
97-24	1.7430	0.1416	0.1324	2.038	0.0916	0.2644	2.189	0.0909	0.3487	2.191
97-45	1.1620	0.1410	0.1114	2.389	0.0764	0.2628	2.778	0.0889	0.3088	2.708
97-56	0.9692	0.1433	0.1215	2.246	0.0838	0.2937	2.722	0.0967	0.3256	2.626
100-24	2.0910	0.0812	0.1346	1.643	0.0558	0.2586	1.726	0.0552	0.2701	1.729
100-45	1.5113	0.0824	0.0955	1.932	0.0433	0.2138	2.166	0.0518	0.2257	2.117
100-56	1.3159	0.0818	0.0999	1.522	0.0427	0.2307	1.854	0.0508	0.2467	1.791
103-12	1.9721	0.0564	0.0741	2.066	0.0317	0.1556	2.205	0.0371	0.1492	2.175
103-24	1.5969	0.0560	0.1625	1.375	0.0300	0.3531	1.491	0.0363	0.3438	1.463
103-45	1.3424	0.0569	0.1003	1.857	0.0287	0.2305	2.056	0.0355	0.2348	2.010
105-12	1.8172	0.0331	0.0776	1.988	0.0188	0.1481	2.073	0.0220	0.1302	2.054
105-24	1.5468	0.0339	0.1769	1.301	0.0179	0.3529	1.376	0.0220	0.3248	1.357
105-45	1.3967	0.0341	0.0998	2.196	0.0178	0.2053	2.292	0.0215	0.1963	2.271

Table C-34

PRESSURE DROP - CORRELATED RESULTS,
G = 1.70 x 10^6 lbm/hr-sq ft, T = 200°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
34-45	2.7365	0.2720	0.1761	1.769	0.2159	0.3220	1.886	0.1914	0.3515	1.934
34-56	1.9193	0.2741	0.2108	1.686	0.1525	0.4522	1.980	0.1901	0.4939	1.917
38-45	2.6446	0.2184	0.1820	1.682	0.1625	0.3393	1.803	0.1553	0.3477	1.818
38-56	1.9832	0.2217	0.2074	1.566	0.1255	0.4397	1.813	0.1486	0.4523	1.757
41-45	1.9446	0.2245	0.2261	1.821	0.1294	0.4750	2.014	0.1516	0.4657	1.971
41-56	1.6295	0.2249	0.2468	1.687	0.1200	0.5516	1.948	0.1455	0.5836	1.888
50-56	2.8476	0.1575	0.1769	1.417	0.1144	0.3029	1.532	0.1129	0.2777	1.535
53-12	3.7092	0.1146	0.1542	1.311	0.1029	0.2590	1.344	0.0949	0.2284	1.394
53-24	2.6176	0.1166	0.3515	1.334	0.0887	0.5895	1.374	0.0853	0.5153	1.379
53-45	2.0429	0.1166	0.2165	1.500	0.0677	0.4102	1.626	0.0801	0.3699	1.595
53-56	1.8303	0.1172	0.2203	1.367	0.0539	0.4342	1.533	0.0781	0.4061	1.490
98-45	1.7576	0.2768	0.2539	1.892	0.1518	0.5482	2.132	0.1829	0.5920	2.075
98-56	1.4637	0.2819	0.2717	1.914	0.1488	0.6152	2.207	0.1805	0.6828	2.141
101-24	3.4259	0.1151	0.3149	1.551	0.2173	0.3720	1.433	0.1178	0.4117	1.548
101-45	2.2119	0.1558	0.2162	1.713	0.0969	0.3911	1.834	0.1098	0.3516	1.809
101-56	1.8901	0.1549	0.2234	1.826	0.0867	0.4276	1.974	0.1055	0.4067	1.934
104-12	2.6177	0.1071	0.1942	1.675	0.0744	0.3348	1.745	0.0779	0.2876	1.738
104-24	2.1371	0.1064	0.4127	1.352	0.0583	0.7380	1.404	0.0769	0.6406	1.393
104-45	1.8025	0.1083	0.2431	1.550	0.0593	0.4584	1.669	0.0744	0.4284	1.633
104-56	1.6655	0.1076	0.2402	1.380	0.0559	0.4752	1.536	0.0725	0.4472	1.488
105-12	2.2715	0.0623	0.2112	1.665	0.0381	0.3195	1.719	0.0446	0.2622	1.764
106-24	2.0791	0.0638	0.4302	1.266	0.0394	0.7002	1.301	0.0463	0.5792	1.291
106-45	1.8613	0.0617	0.2388	1.567	0.0348	0.3554	1.632	0.0431	0.3053	1.612

Table C-35

PRESSURE DROP - CORRELATED RESUTLS,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
36-12	2.6430	0.5547	0.6610	1.054	0.3819	1.1543	1.220	0.4136	1.0663	1.191
48-12	3.5769	0.4252	0.5867	1.198	0.3723	0.9274	1.244	0.3277	0.8166	1.282
48-24	2.6381	0.8661	1.2728	1.254	0.6527	2.1584	1.343	0.6634	2.2574	1.339
51-24	2.6180	0.3000	1.3102	1.329	0.2257	1.9889	1.358	0.2351	1.6408	1.355
52-12	3.5277	0.2146	0.5721	1.226	0.1996	0.6869	1.239	0.1683	0.5476	1.266
52-24	2.8473	0.2191	1.1997	1.294	0.1816	1.5098	1.310	0.1725	1.2030	1.313

Table C-36

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 250^\circ F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
FR152-45	1.5755	0.0057	0.0026	2.141	0.0040	0.0050	2.473	0.0060	0.0041	2.076
FR153-56	0.7957	0.0033	0.0026	5.218	0.0027	0.0049	5.339	0.0036	0.0031	5.131

Table C-37

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.24 x 10^6 lbm/hr-sq ft, T ≈ 250°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
FR153-56	0.9799	0.0097	0.0073	1.596	0.0071	0.0135	1.999	0.0094	0.0120	1.649

Table C-38

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 250^{\circ}F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
FR 75-12	0.7665	0.0177	0.0346	5.485	0.0068	0.0650	5.671	0.0170	0.0594	5.498
FR 75-24	0.6578	0.0176	0.0732	4.370	0.0093	0.1374	4.491	0.0179	0.1224	4.365
FR 75-45	0.5470	0.0179	0.0417	7.254	0.0082	0.0764	7.512	0.0203	0.0664	7.187
FR 75-56	0.4801	0.0174	0.0404	4.227	0.0060	0.0726	4.978	0.0207	0.0640	3.987
FR 76-12	0.6313	0.0107	0.0395	6.491	0.0028	0.0751	6.660	0.0111	0.0641	6.483
FR 76-24	0.5652	0.0106	0.0829	5.435	0.0007	0.1563	5.594	0.0115	0.1320	5.420
FR 76-45	0.5066	0.0109	0.0445	6.235	0.0004	0.0823	6.618	0.0125	0.0695	6.175
FR 76-56	0.4673	0.0114	0.0423	6.031	-0.0010	0.0769	6.653	0.0134	0.0658	5.921
FR148-24	1.6517	0.0289	0.0404	1.314	0.0188	0.0648	1.459	0.0208	0.0797	1.432
FR148-45	1.0477	0.0291	0.0302	3.106	0.0214	0.0563	3.246	0.0237	0.0602	3.205
FR148-56	0.8482	0.0285	0.0319	1.588	0.0242	0.0602	1.808	0.0270	0.0579	1.665
FR154-24	1.6914	0.0220	0.0439	1.706	0.0144	0.0715	1.787	0.0159	0.0791	1.771
FR154-45	1.1854	0.0224	0.0295	3.079	0.0145	0.0538	3.199	0.0170	0.0582	3.162
FR154-56	1.0030	0.0221	0.0303	1.314	0.0148	0.0564	1.625	0.0178	0.0586	1.506
FR160-12	1.7876	0.0155	0.0249	2.873	0.0096	0.0433	2.942	0.0111	0.0431	2.925
FR160-24	1.3925	0.0154	0.0549	1.705	0.0093	0.0992	1.773	0.0110	0.0998	1.755
FR160-45	1.1482	0.0159	0.0327	3.012	0.0099	0.0620	3.101	0.0117	0.0634	3.074

Table C-39

PRESSURE DROP - CORRELATED RESULTS,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 250°F

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
108-12	2.8780	0.0751	0.0660	1.997	0.0620	0.0978	2.057	0.0563	0.1121	2.083
108-24	1.6274	0.0760	0.1544	1.725	0.0481	0.2601	1.813	0.0558	0.2910	1.789
108-45	1.2135	0.0766	0.0988	2.195	0.0490	0.1820	2.362	0.0570	0.2074	2.315
108-56	1.0490	0.0763	0.1003	1.902	0.0506	0.1884	2.124	0.0590	0.2124	2.055
FR 67-12	2.4595	0.0754	0.0695	2.107	0.0550	0.1079	2.198	0.0561	0.1215	2.193
FR 67-24	1.5317	0.0769	0.1597	1.918	0.0484	0.2720	2.002	0.0565	0.3063	1.978
FR 67-45	1.1594	0.0776	0.1016	2.344	0.0489	0.1872	2.513	0.0583	0.2140	2.459
FR 67-56	1.0036	0.0768	0.1026	1.070	0.0493	0.1921	1.467	0.0599	0.2156	1.327
FR 71-24	2.5938	0.0540	0.1285	1.430	0.0547	0.1835	1.428	0.0410	0.1938	1.472
FR 71-45	1.7245	0.0544	0.0811	2.057	0.0336	0.1424	2.163	0.0396	0.1472	2.133
FR 71-56	1.4747	0.0565	0.0822	1.491	0.0329	0.1504	1.680	0.0409	0.1600	1.619
FR 74-12	1.3167	0.0343	0.0738	2.779	0.0183	0.1383	2.872	0.0247	0.1430	2.835
FR 74-24	1.1551	0.0340	0.1565	1.995	0.0183	0.2972	2.066	0.0248	0.3124	2.037
FR 74-45	1.0174	0.0345	0.0864	2.879	0.0169	0.1715	3.004	0.0259	0.1813	2.941
FR 74-56	0.9480	0.0340	0.0857	2.515	0.0142	0.1673	2.702	0.0263	0.1759	2.590
FR 77-12	1.1953	0.0197	0.0820	2.306	0.0100	0.1316	2.374	0.0146	0.1364	2.342
FR 77-24	1.0934	0.0192	0.1726	1.759	0.0076	0.2799	1.817	0.0145	0.2934	1.782
FR 77-45	1.0163	0.0193	0.0922	2.108	0.0060	0.1509	2.228	0.0146	0.1595	2.149
FR 77-56	0.9785	0.0203	0.0874	1.519	0.0041	0.1437	1.740	0.0158	0.1521	1.582
FR149-12	1.4425	0.0545	0.0759	2.239	0.0339	0.1370	2.367	0.0397	0.1503	2.331
FR149-24	1.1783	0.0553	0.1645	1.883	0.0369	0.3032	1.963	0.0408	0.3363	1.946
FR149-45	0.9783	0.0555	0.0979	2.170	0.0439	0.1857	2.277	0.0434	0.2025	2.282
FR149-56	0.8784	0.0544	0.0968	1.941	0.0461	0.1847	2.051	0.0456	0.1915	2.057
FR155-12	1.7630	0.0394	0.0672	1.840	0.0244	0.1124	1.940	0.0291	0.1163	1.909
FR155-24	1.4142	0.0392	0.1452	1.627	0.0248	0.2504	1.690	0.0290	0.2641	1.672
FR155-45	1.1795	0.0400	0.0848	2.165	0.0278	0.1526	2.257	0.0300	0.1664	2.241
FR155-56	1.0754	0.0395	0.0831	1.321	0.0285	0.1515	1.480	0.0303	0.1654	1.454
FR161-45	3.3746	0.0270	0.0501	1.743	0.0253	0.0717	1.754	0.0205	0.0687	1.785
FR161-56	2.5596	0.0263	0.0505	0.798	0.0182	0.0815	0.922	0.0196	0.0757	0.901

Table C-40

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 1.70 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

		HOMOGENEOUS [2] MODEL			MARTINELLI-NELSON [1] MODEL			MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [99] GENERAL VOID FRACTION MODEL		
RUN #	$\sqrt{X_{tt}}$	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}	DELTA P (ACCELER) (PSI)	DELTA P (FRICTION) (PSI)	Φ_{ftt}
109-24	2.7801	0.1397	0.3806	1.441	0.1361	0.5291	1.445	0.1098	0.5400	1.473
109-45	1.9176	0.1403	0.2272	1.585	0.0904	0.3759	1.690	0.1063	0.3821	1.657
109-56	1.6539	0.1411	0.2258	1.578	0.0873	0.3898	1.710	0.1056	0.4084	1.666
FR 72-45	2.9406	0.0998	0.1965	1.480	0.0883	0.2550	1.504	0.0779	0.2481	1.525
FR 72-56	2.4096	0.1035	0.1918	1.394	0.0776	0.3086	1.459	0.0795	0.2895	1.454
FR 73-12	2.5214	0.0788	0.2147	1.757	0.0616	0.3188	1.787	0.0606	0.2789	1.789
FR 73-24	2.1003	0.0788	0.4524	1.459	0.0562	0.6991	1.483	0.0503	0.6187	1.479
FR 73-45	1.8077	0.0798	0.2503	1.657	0.0489	0.4109	1.719	0.0595	0.3773	1.698
FR 73-56	1.6857	0.0791	0.2418	1.413	0.0447	0.4073	1.502	0.0585	0.3829	1.467
FR 78-12	1.5645	0.0365	0.2142	1.598	0.0175	0.2882	1.650	0.0279	0.2790	1.622
FR 78-24	1.5189	0.0356	0.4374	1.246	0.0161	0.5933	1.230	0.0276	0.5802	1.260
FR 78-45	1.4704	0.0359	0.2242	1.529	0.0159	0.3061	1.587	0.0274	0.3023	1.554
FR 78-56	1.4369	0.0374	0.2103	1.401	0.0130	0.3079	1.482	0.0286	0.3083	1.432
FR150-12	1.6145	0.0978	0.2049	1.447	0.0568	0.3551	1.571	0.0731	0.3664	1.523
FR150-24	1.5563	0.0993	0.4208	1.422	0.0587	0.7363	1.486	0.0749	0.7579	1.461
FR150-56	1.4350	0.0978	0.2050	1.353	0.0650	0.3558	1.476	0.0727	0.3881	1.448
FR155-12	1.1803	0.0865	0.3351	1.410	0.0295	0.5443	1.569	0.0665	0.5897	1.469
FR156-24	1.1620	0.0863	0.6806	1.145	0.0423	1.1095	1.224	0.0690	1.2081	1.177
FR155-45	1.1322	0.0879	0.3495	1.519	0.0626	0.5720	1.590	0.0682	0.6253	1.575
FR156-56	1.0953	0.0869	0.3297	1.159	0.0525	0.5427	1.258	0.0679	0.5968	1.237
FR162-45	4.7438	0.0505	0.1514	1.477	0.0791	0.1599	1.407	0.0405	0.1512	1.501
FR162-56	3.5246	0.0493	0.1479	1.207	0.0433	0.1926	1.226	0.0389	0.1702	1.240

Table C-41

PRESSURE DROP - CORRELATED RESULTS,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 250^\circ F$

RUN #	$\sqrt{X_{tt}}$	HOMOGENEOUS [2] MODEL DELTA P (ACCELER) (PSI)	HOMOGENEOUS [2] MODEL DELTA P (FRICTION) (PSI)	HOMOGENEOUS [2] MODEL Φ_{ftt}	MARTINELLI-NELSON [1] MODEL DELTA P (ACCELER) (PSI)	MARTINELLI-NELSON [1] MODEL DELTA P (FRICTION) (PSI)	MARTINELLI-NELSON [1] MODEL Φ_{ftt}	MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL DELTA P (ACCELER) (PSI)	MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL DELTA P (FRICTION) (PSI)	MARTINELLI-NELSON [1] MODEL WITH ZUBER ET AL [59] GENERAL VOID FRACTION MODEL Φ_{ftt}
FR151-12	1.7299	0.1931	0.6179	1.339	0.1187	0.9752	1.415	0.1497	0.9382	1.384
FR151-24	1.6925	0.1965	1.2673	1.121	0.1235	2.0243	1.166	0.1574	1.9739	1.145
FR151-45	1.6769	0.1975	0.6444	1.356	0.1239	1.0345	1.431	0.1530	1.0097	1.402
FR157-12	1.5508	0.1572	0.7486	1.408	0.0755	1.0823	1.432	0.1241	1.0542	1.439
FR157-24	1.5343	0.1573	1.5178	1.194	0.0883	2.2083	1.231	0.1318	2.1657	1.208
FR157-45	1.5157	0.1602	0.7740	1.476	0.0931	1.1327	1.534	0.1268	1.1137	1.505

Table C-42

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 100^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
61- 2	0.1129E 06	104.4	-.026	0.052	19.87	7891.3	397.2
61- 3	0.1129E 06	104.8	0.052	0.134	20.81	7997.1	384.3
61- 4	0.1129E 06	104.5	0.134	0.215	21.27	7799.8	366.6
61- 5	0.1129E 06	104.9	0.215	0.297	20.31	7908.5	389.5
61- 6	0.1129E 06	104.5	0.297	0.380	19.70	8158.7	414.2
61- 7	0.1129E 06	104.9	0.380	0.462	19.98	7958.8	398.3
61- 8	0.1129E 06	104.8	0.462	0.544	21.52	7856.9	365.1
61- 9	0.1129E 06	104.5	0.544	0.625	21.00	7742.6	368.6
61-10	0.1129E 06	104.7	0.625	0.706	19.48	7956.5	408.5
61-AVER	0.1129E 06	104.5	-.102	0.706	20.42	7942.0	389.0
63- 1	0.1091E 06	98.2	0.014	0.063	14.89	4865.1	326.8
63- 2	0.1091E 06	98.5	0.063	0.113	15.30	4766.3	311.6
63- 3	0.1091E 06	98.8	0.113	0.164	16.51	4861.7	294.4
63- 4	0.1091E 06	98.3	0.164	0.215	17.22	4782.0	277.6
63- 5	0.1091E 06	98.8	0.215	0.266	15.58	4812.9	308.9
63- 6	0.1091E 06	98.5	0.266	0.317	15.30	4931.0	322.3
63- 7	0.1091E 06	98.7	0.317	0.368	14.92	4780.1	320.4
63- 8	0.1091E 06	98.7	0.368	0.420	15.17	4788.0	315.6
63- 9	0.1091E 06	98.2	0.420	0.471	14.95	4868.5	325.7
63-10	0.1091E 06	98.5	0.471	0.522	14.80	4848.4	327.7
63-AVER	0.1091E 06	98.5	0.014	0.522	15.46	4830.4	312.5
FR 44- 1	0.1243E 06	100.7	-.012	0.030	20.97	5165.5	246.3
FR 44- 2	0.1243E 06	101.8	0.030	0.074	20.00	4835.7	241.7
FR 44- 3	0.1243E 06	101.9	0.074	0.118	22.23	4915.5	221.1
FR 44- 4	0.1243E 06	101.9	0.118	0.162	22.83	4614.6	202.1
FR 44- 5	0.1243E 06	101.9	0.162	0.206	21.11	4663.2	220.9
FR 44- 6	0.1243E 06	101.7	0.206	0.251	19.29	4750.8	246.3
FR 44- 7	0.1243E 06	101.9	0.251	0.294	18.84	4755.5	252.3
FR 44- 8	0.1243E 06	102.0	0.294	0.338	19.76	4816.8	243.7
FR 44- 9	0.1243E 06	102.4	0.338	0.384	19.45	4785.4	246.0
FR 44-10	0.1243E 06	101.8	0.384	0.434	18.64	5163.6	277.0
FR 44-AVER	0.1243E 06	101.8	-.012	0.434	20.31	4846.7	238.6
FR 50- 4	0.1116E 06	105.4	0.032	0.115	23.49	7922.4	337.3
FR 50- 5	0.1116E 06	105.5	0.115	0.198	22.94	7845.8	342.0
FR 50- 6	0.1116E 06	105.3	0.198	0.279	22.14	7786.4	351.7
FR 50- 7	0.1116E 06	105.7	0.279	0.361	22.38	7869.2	351.7
FR 50- 8	0.1116E 06	105.7	0.361	0.444	24.33	8015.7	329.4
FR 50- 9	0.1116E 06	105.6	0.444	0.529	24.10	7993.8	331.6
FR 50-10	0.1116E 06	105.5	0.529	0.612	23.27	7966.7	342.4
FR 50-AVER	0.1116E 06	105.2	-.189	0.612	22.87	7908.1	345.8
61R- 3	0.1168E 06	97.6	0.036	0.114	23.55	7972.2	338.5
61R- 4	0.1168E 06	97.1	0.114	0.192	22.00	7825.7	355.7
61R- 5	0.1168E 06	97.5	0.192	0.270	23.16	7961.8	343.8
61R- 6	0.1168E 06	97.1	0.270	0.349	21.62	7997.6	370.0

Table C-42 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
61R- 7	0.1168E 06	97.5	0.349	0.428	21.47	8048.5	374.9
61R- 8	0.1168E 06	97.4	0.428	0.508	23.33	8145.5	349.2
61R- 9	0.1168E 06	97.1	0.508	0.587	23.65	8018.2	339.0
61R-10	0.1168E 06	97.4	0.587	0.664	20.84	7878.2	378.1
61R-AVER	0.1168E 06	97.3	-.121	0.664	22.70	7980.3	351.6

Table C-43

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 100^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
60- 4	0.2248E 06	105.1	0.006	0.047	18.74	7799.8	416.2
60- 5	0.2248E 06	105.6	0.047	0.088	17.03	7908.5	464.4
60- 6	0.2248E 06	105.1	0.088	0.130	16.92	8158.7	482.1
60- 7	0.2248E 06	105.4	0.130	0.171	17.50	7958.8	454.8
60- 8	0.2248E 06	105.3	0.171	0.213	17.48	7856.9	449.5
60- 9	0.2248E 06	105.1	0.213	0.253	17.11	7754.3	453.1
60-10	0.2248E 06	105.4	0.253	0.293	16.89	7956.5	471.1
60-AVER	0.2248E 06	105.1	-.107	0.293	17.57	7944.5	452.2
62- 1	0.2242E 06	102.5	0.033	0.056	12.47	4826.5	387.2
62- 2	0.2242E 06	102.7	0.056	0.080	12.98	4753.5	366.2
62- 3	0.2242E 06	102.9	0.080	0.105	14.27	4643.7	325.5
62- 4	0.2242E 06	102.5	0.105	0.129	14.71	4600.5	312.7
62- 5	0.2242E 06	102.9	0.129	0.153	13.56	4637.8	342.0
62- 6	0.2242E 06	102.5	0.153	0.177	13.61	4735.4	348.0
62- 7	0.2242E 06	102.8	0.177	0.201	13.31	4593.1	345.1
62- 8	0.2242E 06	102.7	0.201	0.226	13.05	4581.9	351.1
62- 9	0.2242E 06	102.1	0.226	0.250	13.47	4635.2	344.2
62-10	0.2242E 06	102.5	0.250	0.273	13.12	4621.8	352.2
62-AVER	0.2242E 06	102.6	0.033	0.273	13.46	4662.9	346.4
64- 1	0.2541E 06	97.7	-.012	0.039	18.12	11359.1	627.0
64- 2	0.2541E 06	97.8	0.039	0.088	18.76	10953.0	583.8
64- 3	0.2541E 06	98.3	0.088	0.138	20.41	11227.8	550.1
64- 4	0.2541E 06	97.9	0.138	0.189	21.04	11054.2	525.5
64- 5	0.2541E 06	98.4	0.189	0.240	19.91	11291.6	567.1
64- 6	0.2541E 06	98.0	0.240	0.291	19.71	11468.8	581.8
64- 7	0.2541E 06	98.3	0.291	0.342	19.98	11327.0	566.9
64- 8	0.2541E 06	98.1	0.342	0.394	21.05	11095.9	527.1
64- 9	0.2541E 06	97.8	0.394	0.445	19.72	11033.2	559.4
64-10	0.2541E 06	97.8	0.445	0.495	19.40	11155.9	575.2
64-AVER	0.2541E 06	98.0	-.012	0.495	19.78	11196.6	566.0
FR 45- 5	0.2121E 06	103.4	0.012	0.038	15.89	4672.3	294.1
FR 45- 6	0.2121E 06	103.0	0.038	0.064	15.00	4760.1	317.3
FR 45- 7	0.2121E 06	103.2	0.064	0.090	14.57	4764.8	326.9
FR 45- 8	0.2121E 06	103.1	0.090	0.117	15.56	4827.1	310.2
FR 45- 9	0.2121E 06	102.7	0.117	0.144	15.78	4785.4	303.3
FR 45-10	0.2121E 06	103.1	0.144	0.170	13.83	5176.1	374.2
FR 45-AVER	0.2121E 06	103.0	-.088	0.170	16.22	4849.3	299.0
FR 51- 1	0.2524E 06	97.4	0.064	0.094	22.82	8173.4	358.1
FR 51- 2	0.2524E 06	99.0	0.094	0.124	21.78	7712.7	354.1
FR 51- 3	0.2524E 06	99.9	0.124	0.161	21.19	7795.4	367.9
FR 51- 4	0.2524E 06	98.5	0.161	0.196	22.26	7931.5	356.4
FR 51- 5	0.2524E 06	100.0	0.196	0.230	20.24	7845.8	387.6
FR 51- 6	0.2524E 06	99.8	0.230	0.268	18.38	7805.1	424.6
FR 51- 7	0.2524E 06	98.9	0.268	0.307	20.17	7878.5	390.5

Table C-43 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 51- 8	0.2524E 06	98.6	0.307	0.342	22.21	8026.0	361.4
FR 51- 9	0.2524E 06	99.4	0.342	0.377	20.86	8005.4	383.8
FR 51-10	0.2524E 06	99.5	0.377	0.413	19.59	7991.8	407.9
FR 51-AVER	0.2524E 06	99.1	0.064	0.413	20.94	7916.5	378.1
FR 57- 2	0.2276E 06	99.0	-.005	0.049	21.32	10800.5	506.5
FR 57- 3	0.2276E 06	99.5	0.049	0.105	23.07	11085.9	480.5
FR 57- 4	0.2276E 06	98.9	0.105	0.162	24.18	11155.8	461.4
FR 57- 5	0.2276E 06	99.3	0.162	0.219	21.85	11330.4	518.5
FR 57- 6	0.2276E 06	98.9	0.219	0.277	21.12	11457.9	542.4
FR 57- 7	0.2276E 06	99.2	0.277	0.334	22.50	11109.1	493.8
FR 57- 8	0.2276E 06	99.1	0.334	0.391	24.70	11318.8	458.2
FR 57- 9	0.2276E 06	98.8	0.391	0.449	23.14	11027.3	476.5
FR 57-10	0.2276E 06	98.8	0.449	0.507	23.41	11615.1	496.2
FR 57-AVER	0.2276E 06	99.0	-.062	0.507	22.56	11279.7	499.9
62R- 2	0.2248E 06	100.7	-.003	0.019	14.16	4577.9	323.3
62R- 3	0.2248E 06	101.1	0.019	0.043	11.85	4745.6	400.4
62R- 4	0.2248E 06	100.7	0.043	0.068	9.53	4578.3	480.3
62R- 5	0.2248E 06	100.8	0.068	0.092	11.51	4669.0	402.1
62R- 6	0.2248E 06	100.5	0.092	0.117	10.24	4738.8	462.9
62R- 7	0.2248E 06	100.7	0.117	0.141	10.47	4705.0	449.5
62R- 8	0.2248E 06	100.6	0.141	0.167	11.12	4823.8	433.9
62R- 9	0.2248E 06	100.1	0.167	0.191	12.75	4641.9	364.1
62R-10	0.2248E 06	100.6	0.191	0.214	10.00	4776.1	477.4
62R-AVER	0.2248E 06	100.6	-.027	0.214	11.05	4697.1	425.1
68R- 1	0.2225E 06	93.3	-.011	0.043	37.47	16131.0	430.5
68R- 2	0.2225E 06	99.5	0.043	0.107	32.40	15666.9	483.6
68R- 3	0.2225E 06	100.9	0.107	0.188	29.88	15838.1	530.0
68R- 4	0.2225E 06	99.8	0.188	0.272	29.84	15845.2	531.1
68R- 5	0.2225E 06	100.5	0.272	0.354	31.27	15940.5	509.7
68R- 6	0.2225E 06	100.2	0.354	0.437	30.06	16041.8	533.7
68R- 7	0.2225E 06	100.6	0.437	0.521	30.91	16123.6	521.6
68R- 8	0.2225E 06	100.1	0.521	0.604	35.57	15847.8	445.5
68R- 9	0.2225E 06	100.2	0.604	0.687	34.58	15965.5	460.4
68R-10	0.2225E 06	99.9	0.687	0.771	35.81	15850.2	442.7
68R-AVER	0.2225E 06	99.5	-.011	0.771	32.59	15925.1	487.1

Table C-44

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 100°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
58- 1	0.4343E 06	98.6	-.012	0.015	25.66	16079.9	626.6
58- 2	0.4343E 06	101.8	0.015	0.046	22.21	15777.6	710.4
58- 3	0.4343E 06	103.3	0.046	0.087	22.30	15923.6	714.1
58- 4	0.4343E 06	102.7	0.087	0.129	22.96	15903.6	692.5
58- 5	0.4343E 06	103.2	0.129	0.172	21.62	16086.9	743.3
58- 6	0.4343E 06	102.8	0.172	0.215	21.28	16316.6	766.6
58- 7	0.4343E 06	103.3	0.215	0.259	20.97	16161.0	770.7
58- 8	0.4343E 06	102.8	0.259	0.303	22.74	16124.5	709.2
58- 9	0.4343E 06	102.6	0.303	0.347	21.78	15796.9	725.3
58-10	0.4343E 06	102.3	0.347	0.391	22.60	16021.1	709.0
58-AVER	0.4343E 06	102.3	-.012	0.391	22.42	16017.2	714.5
59- 2	0.4471E 06	105.9	-.006	0.011	15.41	7904.0	512.8
59- 3	0.4471E 06	106.3	0.011	0.032	16.06	8007.4	498.5
59- 4	0.4471E 06	105.9	0.032	0.052	16.26	7799.8	479.6
59- 5	0.4471E 06	106.4	0.052	0.073	14.74	7908.5	536.6
59- 6	0.4471E 06	106.0	0.073	0.094	14.40	8158.7	566.7
59- 7	0.4471E 06	106.5	0.094	0.114	14.17	7958.8	561.5
59- 8	0.4471E 06	106.3	0.114	0.136	14.36	7867.2	547.9
59- 9	0.4471E 06	105.9	0.136	0.157	13.91	7766.0	558.4
59-10	0.4471E 06	106.1	0.157	0.176	13.47	7956.5	590.7
59-AVER	0.4471E 06	106.0	-.022	0.176	14.94	7947.7	532.1
66- 1	0.4615E 06	96.2	0.004	0.029	17.14	11307.7	659.9
66- 2	0.4615E 06	96.9	0.029	0.054	16.92	10953.0	647.4
66- 3	0.4615E 06	97.5	0.054	0.082	18.12	11227.8	619.6
66- 4	0.4615E 06	96.8	0.082	0.110	18.53	11054.2	596.6
66- 5	0.4615E 06	97.2	0.110	0.138	17.09	11291.6	660.6
66- 6	0.4615E 06	96.8	0.138	0.167	16.75	11487.4	685.7
66- 7	0.4615E 06	97.1	0.167	0.195	16.99	11336.3	667.2
66- 8	0.4615E 06	96.7	0.195	0.224	17.65	11106.3	629.4
66- 9	0.4615E 06	96.4	0.224	0.253	16.25	11033.2	679.1
66-10	0.4615E 06	96.5	0.253	0.280	16.09	11168.4	694.1
66-AVER	0.4615E 06	96.8	0.004	0.280	17.17	11196.6	652.1
67- 2	0.4503E 06	103.4	-.016	0.019	21.38	15736.5	736.0
67- 3	0.4503E 06	104.4	0.019	0.059	22.47	15959.5	710.1
67- 4	0.4503E 06	103.8	0.059	0.101	22.75	15808.4	695.0
67- 5	0.4503E 06	104.3	0.101	0.142	20.61	15893.9	771.4
67- 6	0.4503E 06	103.9	0.142	0.184	20.05	16259.5	811.1
67- 7	0.4503E 06	104.0	0.184	0.227	20.75	16283.9	784.7
67- 8	0.4503E 06	103.6	0.227	0.270	24.13	16152.7	669.4
67- 9	0.4503E 06	103.5	0.270	0.312	23.59	15904.8	674.1
67-10	0.4503E 06	103.3	0.312	0.354	23.27	16020.5	688.6
FR 52- 2	0.4378E 06	100.9	-.000	0.019	16.63	7725.4	464.4
FR 52- 3	0.4378E 06	101.1	0.019	0.040	16.99	7826.5	460.7
FR 52- 4	0.4378E 06	100.7	0.040	0.062	17.07	7940.6	465.1

Table C-44 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 52- 5	0.4378E 06	101.0	0.062	0.083	16.00	7854.9	490.9
FR 52- 6	0.4378E 06	100.7	0.083	0.104	14.11	7814.4	554.0
FR 52- 7	0.4378E 06	100.6	0.104	0.125	14.35	7887.9	549.8
FR 52- 8	0.4378E 06	100.5	0.125	0.148	15.34	8036.3	523.9
FR 52- 9	0.4378E 06	99.8	0.148	0.170	16.08	8017.1	498.6
FR 52-10	0.4378E 06	100.3	0.170	0.189	14.59	8004.3	548.5
FR 52-AVER	0.4378E 06	100.6	-.021	0.189	15.82	7929.3	501.1
70R- 1	0.4357E 06	89.1	-.003	0.016	40.94	20058.7	489.9
70R- 2	0.4357E 06	96.4	0.016	0.045	34.54	19712.0	570.7
70R- 3	0.4357E 06	99.3	0.045	0.093	31.07	19873.1	639.6
70R- 4	0.4357E 06	98.2	0.093	0.147	29.82	19858.3	666.0
70R- 5	0.4357E 06	98.8	0.147	0.199	29.88	19813.7	663.1
70R- 6	0.4357E 06	98.5	0.199	0.251	29.61	19886.7	671.7
70R- 7	0.4357E 06	98.9	0.251	0.305	30.96	20086.3	648.7
70R- 8	0.4357E 06	98.2	0.305	0.359	35.55	19847.4	558.3
70R- 9	0.4357E 06	98.3	0.359	0.412	35.51	19878.8	559.8
70R-10	0.4357E 06	97.6	0.412	0.467	35.88	19850.2	553.3
70R-AVER	0.4357E 06	97.3	-.003	0.467	33.32	19886.5	596.8

Table C-45

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.96 \times 10^6$ lbm/hr-sq ft, $T \approx 100°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 59- 3	0.8778E 06	102.0	-.007	0.008	18.50	11096.3	599.7
FR 59- 4	0.8778E 06	101.5	0.008	0.024	18.02	11164.9	619.7
FR 59- 5	0.8778E 06	101.6	0.024	0.040	17.28	11339.4	656.3
FR 59- 6	0.8778E 06	100.8	0.040	0.057	16.16	11457.9	709.0
FR 59- 7	0.8778E 06	100.7	0.057	0.074	16.41	11118.5	677.5
FR 59- 8	0.8778E 06	100.2	0.074	0.091	17.33	11349.7	654.9
FR 59- 9	0.8778E 06	99.5	0.091	0.108	16.71	11050.6	661.5
FR 59-10	0.8778E 06	99.3	0.108	0.124	16.19	11640.2	719.2
FR 59-AVER	0.8778E 06	100.9	-.034	0.124	17.49	11292.7	645.8
FR 64- 6	0.7940E 06	100.4	-.001	0.024	20.80	16135.0	775.8
FR 64- 7	0.7940E 06	100.4	0.024	0.049	21.06	15895.3	754.6
FR 64- 8	0.7940E 06	99.5	0.049	0.075	22.82	15788.0	692.0
FR 64- 9	0.7940E 06	99.1	0.075	0.100	21.25	15832.8	745.2
FR 64-10	0.7940E 06	98.5	0.100	0.126	20.92	16026.5	766.0
FR 64-AVER	0.7940E 06	100.3	-.113	0.126	22.19	15928.9	717.9

Table C-46

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 1.70 \times 10^6$ lbm/hr-sq ft, $T \approx 100°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 48- 5	0.1675E 07	100.3	-.001	0.004	9.18	4699.4	511.8
FR 48- 6	0.1675E 07	99.9	0.004	0.008	8.30	4788.1	577.2
FR 48- 7	0.1675E 07	99.8	0.008	0.013	8.00	4783.5	597.8
FR 48- 8	0.1675E 07	99.2	0.013	0.020	8.46	4847.7	573.1
FR 48- 9	0.1675E 07	98.2	0.020	0.026	9.53	4820.4	506.0
FR 48-10	0.1675E 07	98.1	0.026	0.030	8.37	5201.2	621.6
FR 48-AVER	0.1675E 07	99.6	-.009	0.030	9.23	4880.5	528.8
FR 60- 1	0.1720E 07	98.9	0.016	0.019	18.37	11921.7	649.0
FR 60- 2	0.1720E 07	100.0	0.019	0.022	17.64	10826.0	613.8
FR 60- 3	0.1720E 07	101.0	0.022	0.029	17.18	11096.3	645.8
FR 60- 4	0.1720E 07	100.1	0.029	0.039	16.95	11164.9	658.6
FR 60- 5	0.1720E 07	99.9	0.039	0.049	16.25	11339.4	697.6
FR 60- 6	0.1720E 07	99.0	0.049	0.060	15.34	11467.2	747.5
FR 60- 7	0.1720E 07	98.3	0.060	0.072	15.64	11118.5	711.0
FR 60- 8	0.1720E 07	97.3	0.072	0.085	16.59	11339.4	683.5
FR 60- 9	0.1720E 07	95.7	0.085	0.098	16.71	11038.9	660.6
FR 60-10	0.1720E 07	94.9	0.098	0.110	15.78	11640.2	737.9
FR 60-AVER	0.1720E 07	98.5	0.016	0.110	16.72	11295.2	675.6
FR 65- 6	0.1648E 07	100.4	0.003	0.018	15.50	16135.0	1041.0
FR 65- 7	0.1648E 07	100.0	0.018	0.034	17.42	15895.3	912.3
FR 65- 8	0.1648E 07	98.4	0.034	0.051	18.58	15839.6	848.1
FR 65- 9	0.1648E 07	97.3	0.051	0.068	17.77	15832.8	891.1
FR 65-10	0.1648E 07	96.0	0.068	0.085	16.81	16039.0	954.2
FR 65-AVER	0.1648E 07	99.2	-.030	0.085	19.71	15939.1	808.6

Table C-47

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 3.40 x 10^6 lbm/hr-sq ft, T ≈ 100°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 49- 1	0.3370E 07	101.3	0.019	0.021	9.04	5204.1	575.5
FR 49- 2	0.3370E 07	101.2	0.021	0.022	9.23	4861.2	526.8
FR 49- 4	0.3370E 07	101.9	0.022	0.022	9.15	4650.9	508.1
FR 49- 5	0.3370E 07	102.4	0.022	0.022	8.84	4690.4	530.8
FR 49- 6	0.3370E 07	102.4	0.022	0.022	8.26	4778.8	578.3
FR 49- 8	0.3370E 07	103.3	0.022	0.024	7.59	4847.7	638.4
FR 49- 9	0.3370E 07	102.8	0.024	0.026	8.12	4808.7	592.3
FR 49-AVER	0.3370E 07	102.3	0.019	0.026	8.39	4876.3	581.2
FR 55- 1	0.3457E 07	97.9	0.025	0.026	13.72	8211.9	598.5
FR 55- 2	0.3457E 07	98.3	0.026	0.026	13.89	7763.6	559.0
FR 55- 3	0.3457E 07	99.0	0.026	0.026	14.06	7836.9	557.4
FR 55- 4	0.3457E 07	99.2	0.026	0.027	14.21	7949.7	559.3
FR 55- 5	0.3457E 07	100.0	0.027	0.027	13.61	7873.0	578.3
FR 55- 8	0.3457E 07	101.8	0.026	0.027	11.51	8046.6	699.1
FR 55- 9	0.3457E 07	101.6	0.027	0.029	11.72	8017.1	684.2
FR 55-AVER	0.3457E 07	100.2	0.025	0.029	12.68	7942.7	626.3
FR 61- 5	0.3401E 07	97.5	0.028	0.029	18.32	11339.4	618.9
FR 61- 6	0.3401E 07	96.5	0.029	0.031	18.50	11457.9	619.4
FR 61- 8	0.3401E 07	98.9	0.029	0.031	17.10	11329.1	662.4
FR 61- 9	0.3401E 07	99.0	0.031	0.037	16.07	11027.3	686.0
FR 61-10	0.3401E 07	97.8	0.037	0.047	16.51	11615.1	703.4
FR 61-AVER	0.3401E 07	96.0	0.036	0.047	17.58	11293.4	642.3

Table C-48

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.12 x 10^6 lbm/hr-sq ft, T ≈ 150°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
12- 5	0.1111E 06	149.1	-.020	0.034	11.53	4543.7	394.0
12- 6	0.1111E 06	148.8	0.034	0.088	11.70	4501.1	384.5
12- 7	0.1111E 06	148.9	0.088	0.143	11.57	4846.2	418.7
12- 8	0.1111E 06	149.0	0.143	0.196	12.35	4849.0	392.5
12- 9	0.1111E 06	149.5	0.196	0.247	10.83	4241.5	391.8
12-10	0.1111E 06	150.7	0.247	0.300	9.74	5338.4	548.0
12-AVER	0.1111E 06	148.7	-.201	0.300	11.62	4683.3	402.9
40- 3	0.1182E 06	149.2	0.007	0.095	17.19	8360.5	486.3
40- 4	0.1182E 06	149.5	0.095	0.187	17.64	8403.8	476.3
40- 5	0.1182E 06	149.0	0.187	0.278	17.66	8300.0	470.0
40- 6	0.1182E 06	148.8	0.278	0.364	17.17	7758.6	451.8
40- 7	0.1182E 06	149.4	0.364	0.450	17.57	8109.4	461.6
40- 8	0.1182E 06	149.2	0.450	0.537	18.95	7596.8	400.9
40- 9	0.1182E 06	148.8	0.537	0.620	18.29	8055.6	440.4
40-10	0.1182E 06	150.1	0.620	0.697	14.77	7368.3	499.0
40-AVER	0.1182E 06	149.0	-.152	0.697	17.25	7930.1	459.7
79- 3	0.1223E 06	150.3	-.037	0.047	16.65	7975.4	479.1
79- 4	0.1223E 06	149.7	0.047	0.132	17.85	8052.9	451.2
79- 5	0.1223E 06	150.2	0.132	0.214	17.08	7812.2	457.4
79- 6	0.1223E 06	150.0	0.214	0.296	16.81	7718.0	459.0
79- 7	0.1223E 06	150.5	0.296	0.371	17.26	7913.9	458.5
79- 8	0.1223E 06	153.9	0.371	0.449	14.54	7957.1	547.1
79- 9	0.1223E 06	153.5	0.449	0.535	14.20	8019.7	564.8
79-10	0.1223E 06	153.9	0.535	0.620	12.16	8160.0	671.2
79-AVER	0.1223E 06	150.9	-.184	0.620	16.02	7963.8	497.2
82- 5	0.1185E 06	151.3	0.025	0.077	12.28	4853.9	395.1
82- 6	0.1185E 06	151.0	0.077	0.129	12.39	4777.4	385.5
82- 7	0.1185E 06	151.5	0.129	0.173	11.90	4666.1	392.2
82- 8	0.1185E 06	154.2	0.173	0.221	9.75	4828.4	495.0
82- 9	0.1185E 06	153.2	0.221	0.275	9.67	4685.8	484.8
82-10	0.1185E 06	153.5	0.275	0.326	9.14	4741.7	518.9
82-AVER	0.1185E 06	151.9	-.175	0.326	11.45	4722.0	412.5
126- 8	0.1161E 06	152.5	-.002	0.049	9.93	4719.1	475.1
126- 9	0.1161E 06	151.7	0.049	0.104	12.28	4550.6	370.6
126-10	0.1161E 06	151.4	0.104	0.157	10.73	4650.8	433.6

Table C-49

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 150^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
6- 5	0.2428E 06	149.8	-.006	0.036	14.41	7844.2	544.3
6- 6	0.2428E 06	149.3	0.036	0.078	14.64	7869.6	537.6
6- 7	0.2428E 06	149.7	0.078	0.120	13.90	8011.1	576.2
6- 8	0.2428E 06	149.5	0.120	0.166	15.66	7953.8	507.8
6- 9	0.2428E 06	148.1	0.166	0.209	16.54	7740.7	468.0
6-10	0.2428E 06	149.3	0.209	0.245	14.25	8038.5	564.0
6-AVER	0.2428E 06	149.4	-.174	0.245	14.64	7912.8	540.3
13- 4	0.2350E 06	149.0	-.007	0.019	12.49	4913.4	393.4
13- 5	0.2350E 06	149.1	0.019	0.045	11.85	4575.2	386.1
13- 6	0.2350E 06	148.8	0.045	0.071	11.10	4580.0	412.5
13- 7	0.2350E 06	148.9	0.071	0.097	11.57	4808.9	415.7
13- 8	0.2350E 06	148.8	0.097	0.127	11.74	4922.4	419.3
13- 9	0.2350E 06	147.6	0.127	0.154	12.23	4805.4	392.8
13-10	0.2350E 06	148.7	0.154	0.175	10.70	4889.4	457.0
13-AVER	0.2350E 06	148.8	-.087	0.175	11.40	4781.5	419.3
18- 4	0.2371E 06	149.1	-.014	0.048	16.38	11314.4	690.9
18- 5	0.2371E 06	149.4	0.048	0.109	15.66	11130.3	710.9
18- 6	0.2371E 06	149.0	0.109	0.170	15.00	11014.9	734.3
18- 7	0.2371E 06	149.3	0.170	0.230	15.53	11104.3	715.2
18- 8	0.2371E 06	149.0	0.230	0.294	16.81	10921.6	649.5
18- 9	0.2371E 06	147.8	0.294	0.355	17.24	11203.5	650.0
18-10	0.2371E 06	148.9	0.355	0.411	14.88	11138.2	748.4
18-AVER	0.2371E 06	149.0	-.188	0.411	15.75	11110.8	705.6
46- 5	0.2368E 06	149.2	0.002	0.045	13.03	8084.8	620.5
46- 6	0.2368E 06	148.9	0.045	0.088	13.08	7571.4	578.9
46- 7	0.2368E 06	149.0	0.088	0.131	12.87	7962.5	618.6
46- 8	0.2368E 06	149.1	0.131	0.176	13.92	8157.1	585.9
46- 9	0.2368E 06	148.5	0.176	0.219	13.57	7899.3	582.1
46-10	0.2368E 06	149.0	0.219	0.258	11.72	7433.5	634.2
46-AVER	0.2368E 06	149.0	-.167	0.258	13.03	7844.8	602.0
54- 4	0.2428E 06	150.3	0.012	0.055	13.03	8019.7	615.5
54- 5	0.2428E 06	150.7	0.055	0.097	13.67	7943.9	581.0
54- 6	0.2428E 06	150.5	0.097	0.139	13.20	7921.2	600.1
54- 7	0.2428E 06	150.8	0.139	0.181	14.05	7989.8	568.5
54- 8	0.2428E 06	150.8	0.181	0.225	14.94	7776.7	520.4
54- 9	0.2428E 06	150.1	0.225	0.267	15.21	8000.9	526.1
54-10	0.2428E 06	150.7	0.267	0.306	13.07	7614.8	582.7
54-AVER	0.2428E 06	150.6	-.113	0.306	13.96	7866.8	563.7
73- 3	0.2259E 06	150.5	0.030	0.120	21.58	15680.8	726.5
73- 4	0.2259E 06	149.7	0.120	0.212	22.23	15995.5	719.6
73- 5	0.2259E 06	150.5	0.212	0.302	21.66	16132.2	744.7
73- 6	0.2259E 06	150.3	0.302	0.394	20.59	16081.2	780.9
73- 7	0.2259E 06	150.9	0.394	0.486	23.37	16161.3	691.4
73- 8	0.2259E 06	150.7	0.486	0.578	24.87	16020.8	644.1

Table C-49 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
73- 9	0.2259E 06	150.6	0.578	0.669	22.57	15643.2	693.0
73-10	0.2259E 06	150.7	0.669	0.760	22.35	15963.3	714.1
73-AVER	0.2259E 06	149.9	-.114	0.760	22.39	15936.8	711.7
75- 4	0.2275E 06	150.1	-.021	0.042	18.00	10987.3	610.3
75- 5	0.2275E 06	150.6	0.042	0.104	17.40	11192.2	643.2
75- 6	0.2275E 06	150.4	0.104	0.167	16.87	11140.3	660.3
75- 7	0.2275E 06	150.8	0.167	0.230	16.72	11090.8	663.4
75- 8	0.2275E 06	150.6	0.230	0.295	19.38	11121.1	574.0
75- 9	0.2275E 06	150.3	0.295	0.359	19.09	11119.3	582.4
75-10	0.2275E 06	150.5	0.359	0.421	16.81	11138.8	662.5
75-AVER	0.2275E 06	149.9	-.170	0.421	18.02	11102.6	616.2
78- 6	0.2316E 06	149.3	0.021	0.064	13.73	7727.3	562.7
78- 7	0.2316E 06	149.8	0.064	0.107	13.81	7923.3	573.7
78- 8	0.2316E 06	149.7	0.107	0.154	14.17	7967.4	562.1
78- 9	0.2316E 06	149.2	0.154	0.199	14.35	8019.7	559.0
78-10	0.2316E 06	149.6	0.199	0.242	13.00	8172.5	628.5
78-AVER	0.2316E 06	148.9	-.162	0.242	14.67	7970.2	543.2
81- 4	0.2265E 06	150.6	-.013	0.014	11.36	4700.8	413.7
81- 5	0.2265E 06	151.0	0.014	0.041	11.10	4853.9	437.4
81- 6	0.2265E 06	150.7	0.041	0.068	10.88	4768.0	438.3
81- 7	0.2265E 06	151.1	0.068	0.095	10.04	4666.1	464.6
81- 8	0.2265E 06	151.0	0.095	0.124	10.53	4828.4	458.5
81- 9	0.2265E 06	150.3	0.124	0.151	10.34	4685.8	453.4
81-10	0.2265E 06	151.0	0.151	0.175	9.06	4741.7	523.6
81-AVER	0.2265E 06	150.8	-.086	0.175	10.49	4726.2	450.5

Table C-50

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 150°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
7- 2	0.4650E 06	149.6	-.009	0.012	11.82	7496.0	634.3
7- 3	0.4650E 06	149.6	0.012	0.034	12.32	7612.9	617.9
7- 4	0.4650E 06	149.3	0.034	0.056	12.91	7943.5	615.4
7- 5	0.4650E 06	149.4	0.056	0.078	12.28	7899.8	643.5
7- 6	0.4650E 06	149.2	0.078	0.101	12.11	7926.7	654.3
7- 7	0.4650E 06	149.4	0.101	0.122	12.31	7871.0	639.6
7- 8	0.4650E 06	149.3	0.122	0.148	12.45	7842.9	630.0
7- 9	0.4650E 06	147.9	0.148	0.170	13.30	7755.8	583.2
7-10	0.4650E 06	149.1	0.170	0.186	11.48	7947.4	692.1
7-AVER	0.4650E 06	149.2	-.029	0.186	12.31	7845.5	637.3
16- 1	0.4793E 06	148.4	0.000	0.008	12.16	5067.2	416.9
16- 2	0.4793E 06	149.4	0.008	0.017	11.16	4680.6	419.3
16- 3	0.4793E 06	149.7	0.017	0.030	11.58	4921.8	424.9
16- 4	0.4793E 06	149.4	0.030	0.044	11.87	4955.2	417.4
16- 5	0.4793E 06	149.8	0.044	0.057	10.63	4884.3	459.4
16- 6	0.4793E 06	149.3	0.057	0.070	10.48	4925.6	470.1
16- 7	0.4793E 06	149.8	0.070	0.083	10.44	4900.0	469.5
16- 8	0.4793E 06	149.4	0.083	0.100	10.61	4825.1	454.8
16- 9	0.4793E 06	148.2	0.100	0.114	11.68	4616.4	395.2
16-10	0.4793E 06	149.3	0.114	0.121	9.92	4864.7	490.2
16-AVER	0.4793E 06	149.3	0.000	0.121	11.07	4864.1	439.3
19- 3	0.4659E 06	150.4	-.011	0.019	13.44	10976.7	816.8
19- 4	0.4659E 06	150.1	0.019	0.051	14.36	11302.4	787.1
19- 5	0.4659E 06	150.1	0.051	0.083	13.54	10993.0	811.8
19- 6	0.4659E 06	149.9	0.083	0.114	12.91	11164.8	864.6
19- 7	0.4659E 06	150.0	0.114	0.145	13.26	11254.5	849.1
19- 8	0.4659E 06	149.9	0.145	0.180	13.67	10953.7	801.2
19- 9	0.4659E 06	148.5	0.180	0.211	14.56	11084.7	761.4
19-10	0.4659E 06	149.8	0.211	0.235	12.11	11156.7	921.3
19-AVER	0.4659E 06	149.9	-.072	0.235	13.42	11120.5	828.4
27- 3	0.4647E 06	149.9	-.017	0.028	18.41	15741.1	855.2
27- 4	0.4647E 06	149.3	0.028	0.073	20.28	16164.2	797.0
27- 5	0.4647E 06	149.6	0.073	0.118	19.01	16079.5	845.9
27- 6	0.4647E 06	149.2	0.118	0.164	18.12	16344.0	902.0
27- 7	0.4647E 06	149.3	0.164	0.210	21.01	16264.5	774.2
27- 8	0.4647E 06	149.0	0.210	0.258	23.45	15854.7	676.1
27- 9	0.4647E 06	147.5	0.258	0.303	25.51	15796.2	619.2
27-10	0.4647E 06	148.7	0.303	0.343	22.04	16369.4	742.8
27-AVER	0.4647E 06	149.1	-.105	0.343	20.26	16045.5	791.8
43- 4	0.4905E 06	149.7	-.004	0.026	14.64	11320.6	773.1
43- 5	0.4905E 06	149.2	0.026	0.058	15.18	10993.0	724.3
43- 6	0.4905E 06	148.9	0.058	0.087	14.54	11142.4	766.2
43- 7	0.4905E 06	149.1	0.087	0.117	14.48	11226.4	775.4
43- 8	0.4905E 06	148.8	0.117	0.150	15.94	10933.1	685.9

Table C-50 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
43- 9	0.4905E 06	147.5	0.150	0.179	16.61	11101.0	668.4
43-10	0.4905E 06	148.6	0.179	0.203	14.56	11169.2	767.1
43-AVER	0.4905E 06	148.9	-.087	0.203	14.91	11104.2	744.8
71- 3	0.4508E 06	150.8	-.022	0.031	21.59	19945.8	923.8
71- 4	0.4508E 06	150.1	0.031	0.088	22.61	20038.9	886.3
71- 5	0.4508E 06	150.9	0.088	0.144	21.76	19987.8	918.5
71- 6	0.4508E 06	150.8	0.144	0.201	20.54	19991.8	973.1
71- 7	0.4508E 06	151.2	0.201	0.259	23.52	20165.8	857.5
71- 8	0.4508E 06	150.8	0.259	0.317	26.27	19856.3	755.8
71- 9	0.4508E 06	150.9	0.317	0.375	25.93	19784.0	763.0
71-10	0.4508E 06	150.7	0.375	0.433	23.89	20232.3	846.9
71-AVER	0.4508E 06	149.9	-.086	0.433	23.57	19952.6	846.7
72- 3	0.4747E 06	150.9	0.000	0.042	17.59	15691.1	892.0
72- 4	0.4747E 06	150.2	0.042	0.086	18.68	16013.7	857.3
72- 5	0.4747E 06	150.8	0.086	0.129	17.24	16141.2	936.2
72- 6	0.4747E 06	150.5	0.129	0.173	16.15	16090.5	996.6
72- 7	0.4747E 06	151.1	0.173	0.216	16.86	16161.3	958.5
72- 8	0.4747E 06	150.9	0.216	0.260	18.42	16000.2	868.8
72- 9	0.4747E 06	150.7	0.260	0.304	17.27	15654.9	906.2
72-10	0.4747E 06	150.8	0.304	0.347	18.57	15963.3	859.8
72-AVER	0.4747E 06	150.4	-.073	0.347	17.78	15940.6	896.8
74- 2	0.4980E 06	149.6	-.005	0.019	14.43	10736.1	743.8
74- 3	0.4980E 06	150.5	0.019	0.047	13.92	10953.4	786.8
74- 4	0.4980E 06	149.8	0.047	0.076	14.69	11121.8	757.1
74- 5	0.4980E 06	150.4	0.076	0.104	13.70	11192.2	817.1
74- 6	0.4980E 06	150.2	0.104	0.132	12.70	11098.5	873.6
74- 7	0.4980E 06	150.8	0.132	0.160	12.59	11153.9	885.6
74- 8	0.4980E 06	150.6	0.160	0.191	13.35	11010.8	824.7
74- 9	0.4980E 06	150.2	0.191	0.220	13.06	11292.9	865.0
74-10	0.4980E 06	150.6	0.220	0.247	11.85	11241.0	948.4
74-AVER	0.4980E 06	150.2	-.031	0.247	13.47	11102.9	824.1
77- 2	0.4690E 06	149.7	-.007	0.013	11.21	7813.3	697.1
77- 3	0.4690E 06	150.1	0.013	0.035	11.56	7872.2	680.8
77- 4	0.4690E 06	149.6	0.035	0.057	12.06	7778.6	645.3
77- 5	0.4690E 06	150.1	0.057	0.078	11.00	7821.2	710.8
77- 6	0.4690E 06	149.8	0.078	0.099	10.80	7879.4	729.9
77- 7	0.4690E 06	150.2	0.099	0.121	10.57	7959.1	753.3
77- 8	0.4690E 06	150.1	0.121	0.144	10.84	8009.3	738.6
77- 9	0.4690E 06	149.4	0.144	0.167	10.44	7749.8	742.3
77-10	0.4690E 06	150.0	0.167	0.186	9.44	8078.0	855.6
77-AVER	0.4690E 06	149.8	-.028	0.186	11.00	7918.8	720.2
80- 1	0.4824E 06	150.6	0.023	0.035	8.48	4849.3	571.5
80- 2	0.4824E 06	150.7	0.035	0.047	8.27	4555.4	550.5
80- 3	0.4824E 06	150.9	0.047	0.060	8.72	4605.2	528.0

Table C-50 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
80- 4	0.4824E 06	150.4	0.060	0.073	9.04	4700.8	520.0
80- 5	C.4824E 06	15C.9	0.073	0.085	8.38	4853.9	579.3
80- 6	0.4824E 06	150.5	0.085	0.098	8.34	4777.4	572.9
80- 7	C.4824E 06	150.8	0.098	0.110	7.78	4666.1	600.0
80- 8	C.4824E 06	150.7	0.110	0.126	8.15	4828.4	592.6
80- 9	C.4824E 06	150.0	0.126	0.139	8.25	4685.8	567.8
80-10	C.4824E 06	150.4	0.139	0.149	7.44	4741.7	637.0
80-AVER	C.4824E 06	150.6	0.023	0.149	8.30	4726.4	569.4

Table C-51

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 150°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
8- 3	0.8766E 06	149.7	0.003	0.015	11.65	7757.9	665.7
8- 4	0.8766E 06	149.6	0.015	0.027	12.12	8071.4	665.9
8- 5	0.8766E 06	149.6	0.027	0.039	11.36	7780.2	684.8
8- 6	0.8766E 06	149.3	0.039	0.052	11.12	8025.0	721.8
8- 7	0.8766E 06	149.4	0.052	0.063	11.43	7865.0	688.3
8- 8	0.8766E 06	149.3	0.063	0.079	11.59	8012.0	691.0
8- 9	0.8766E 06	147.8	0.079	0.092	12.29	7456.8	606.6
8-10	0.8766E 06	148.8	0.092	0.098	10.42	8050.0	772.8
8-AVER	0.8766E 06	149.2	-.016	0.098	11.47	7917.1	690.4
20- 1	0.8945E 06	149.6	0.086	0.103	12.13	11301.4	931.5
20- 2	0.8945E 06	149.4	0.103	0.119	11.83	10765.8	909.9
20- 3	0.8945E 06	149.7	0.119	0.135	12.59	10924.7	868.0
20- 4	0.8945E 06	149.2	0.135	0.152	13.35	10924.5	818.2
20- 5	0.8945E 06	149.2	0.152	0.169	12.35	11016.9	892.0
20- 6	0.8945E 06	148.9	0.169	0.186	12.24	11467.8	937.1
20- 7	0.8945E 06	148.9	0.186	0.202	12.59	11129.3	883.7
20- 8	0.8945E 06	148.8	0.202	0.222	13.02	10917.8	838.5
20- 9	0.8945E 06	147.2	0.222	0.240	13.66	11084.7	811.7
20-10	0.8945E 06	148.3	0.240	0.250	11.33	11025.2	972.7
20-AVER	0.8945E 06	148.9	0.086	0.250	12.54	11055.8	881.8
24- 1	0.8765E 06	148.8	0.038	0.063	16.90	16747.5	990.7
24- 2	0.8765E 06	148.6	0.063	0.086	16.96	15586.5	919.1
24- 3	0.8765E 06	149.0	0.086	0.110	17.49	15722.9	898.8
24- 4	0.8765E 06	148.4	0.110	0.134	18.21	15925.8	874.5
24- 5	0.8765E 06	148.6	0.134	0.159	16.74	16249.3	970.4
24- 6	0.8765E 06	148.2	0.159	0.183	15.59	15659.3	1004.5
24- 7	0.8765E 06	148.3	0.183	0.207	16.20	15823.2	977.0
24- 8	0.8765E 06	147.9	0.207	0.235	18.60	15475.7	832.2
24- 9	0.8765E 06	146.3	0.235	0.259	19.35	15918.8	822.9
24-10	0.8765E 06	147.4	0.259	0.277	16.91	16371.4	968.2
24-AVER	0.8765E 06	148.2	0.038	0.277	17.30	15948.0	921.8
28- 1	0.8795E 06	146.4	0.051	0.074	20.65	20113.6	974.1
28- 2	0.8795E 06	147.6	0.074	0.097	19.15	19539.0	1020.5
28- 3	0.8795E 06	148.8	0.097	0.124	19.47	19950.5	1024.6
28- 4	0.8795E 06	148.4	0.124	0.154	20.67	19800.0	957.9
28- 5	0.8795E 06	148.3	0.154	0.185	19.90	20304.3	1020.4
28- 6	0.8795E 06	148.1	0.185	0.215	19.08	20463.6	1072.8
28- 7	0.8795E 06	148.0	0.215	0.245	19.70	19859.1	1008.3
28- 8	0.8795E 06	147.7	0.245	0.280	22.76	19831.6	871.5
28- 9	0.8795E 06	149.8	0.280	0.310	23.77	20049.2	843.6
28-10	0.8795E 06	147.2	0.310	0.333	20.43	20002.8	979.0
28-AVER	0.8795E 06	147.6	0.051	0.333	20.55	19991.4	973.0
47- 1	0.8659E 06	149.1	-.006	0.006	11.78	8329.4	707.0
47- 2	0.8659E 06	149.1	0.006	0.018	11.11	7591.6	683.6

Table C-51 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
47- 3	0.8659E 06	149.3	0.018	0.030	12.01	7919.7	659.6
47- 4	0.8659E 06	148.7	0.030	0.043	12.55	7990.7	636.5
47- 5	0.8659E 06	149.1	0.043	0.054	11.51	7776.6	675.4
47- 6	0.8659E 06	148.8	0.054	0.066	11.34	7747.4	683.5
47- 7	0.8659E 06	148.9	0.066	0.078	11.61	7909.5	681.4
47- 8	0.8659E 06	148.9	0.078	0.092	11.55	8183.7	708.3
47- 9	0.8659E 06	147.9	0.092	0.105	11.59	7779.4	671.4
47-10	0.8659E 06	148.5	0.105	0.113	10.45	7225.8	691.7
47-AVER	0.8659E 06	148.8	-.006	0.113	11.58	7845.4	677.6
55- 1	0.8766E 06	150.3	0.031	0.040	12.53	7993.7	637.9
55- 2	0.8766E 06	150.8	0.040	0.050	11.28	7738.0	686.0
55- 3	0.8766E 06	151.0	0.050	0.063	11.74	7839.1	667.4
55- 4	0.8766E 06	150.3	0.063	0.076	12.11	7969.8	658.3
55- 5	0.8766E 06	150.6	0.076	0.087	11.40	7946.9	696.9
55- 6	0.8766E 06	150.3	0.087	0.099	11.05	7845.0	710.0
55- 7	0.8766E 06	150.5	0.099	0.111	11.47	7912.1	689.8
55- 8	0.8766E 06	150.3	0.111	0.126	11.40	7784.8	682.7
55- 9	0.8766E 06	149.3	0.126	0.138	11.79	7845.7	665.4
55-10	0.8766E 06	149.9	0.138	0.146	10.20	7779.6	763.0
55-AVER	0.8766E 06	150.3	0.031	0.146	11.54	7865.5	681.7
83- 1	0.8863E 06	152.3	0.065	0.072	8.32	4844.2	582.5
83- 2	0.8863E 06	152.4	0.072	0.078	7.98	4568.1	572.7
83- 3	0.8863E 06	152.5	0.078	0.086	8.30	4605.2	554.7
83- 4	0.8863E 06	152.0	0.086	0.093	8.78	4673.6	532.5
83- 5	0.8863E 06	152.2	0.093	0.101	8.00	4862.9	607.8
83- 6	0.8863E 06	151.9	0.101	0.108	7.90	4786.7	605.9
83- 7	0.8863E 06	152.1	0.108	0.115	7.61	4675.5	614.3
83- 8	0.8863E 06	152.0	0.115	0.124	7.77	4828.4	621.1
83- 9	0.8863E 06	151.2	0.124	0.132	7.98	4697.5	588.8
83-10	0.8863E 06	151.7	0.132	0.136	7.17	4746.7	662.5
83-AVER	0.8863E 06	152.0	0.065	0.136	8.00	4728.9	591.3

Table C-52

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 1.70 \times 10^6$ lbm/hr-sq ft, $T \approx 150^oF$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
3- 1	0.1756E 07	150.2	0.064	0.069	10.30	5013.4	486.8
3- 2	0.1756E 07	150.0	0.069	0.074	9.35	4601.9	467.3
3- 3	0.1756E 07	149.8	0.074	0.078	10.93	4753.0	434.8
3- 4	0.1756E 07	149.6	0.078	0.082	11.35	4791.4	422.1
3- 5	0.1756E 07	149.7	0.082	0.086	11.03	4687.1	424.9
3- 6	0.1756E 07	149.3	0.086	0.090	10.63	4632.6	435.8
3- 7	0.1756E 07	149.4	0.090	0.095	10.57	4696.8	444.2
3- 8	0.1756E 07	149.0	0.095	0.104	9.92	4625.9	466.3
3- 9	0.1756E 07	147.3	0.104	0.109	11.37	4691.2	412.7
3-AVER	0.1756E 07	149.3	0.064	0.106	10.58	4727.1	446.7
10- 2	0.1674E 07	150.7	-.001	0.003	14.38	7818.0	555.3
10- 3	0.1674E 07	151.2	0.003	0.009	14.98	7805.5	520.9
10- 4	0.1674E 07	150.9	0.009	0.015	15.45	7990.0	517.3
10- 5	0.1674E 07	151.0	0.015	0.022	14.87	7606.8	511.6
10- 6	0.1674E C7	150.7	0.022	0.029	15.33	8037.0	534.5
10- 7	0.1674E 07	150.7	0.029	0.037	14.88	7999.2	537.7
10- 8	0.1674E 07	150.2	0.037	0.049	14.98	8169.1	545.2
10- 9	0.1674E 07	148.5	0.049	0.057	15.98	7658.0	479.2
10-10	0.1674E 07	149.5	0.057	0.058	13.74	7969.3	580.0
10-AVER	0.1674E 07	150.4	-.006	0.058	14.94	7937.4	531.2
15- 1	0.1668E 07	149.4	0.023	0.027	8.58	5177.4	603.2
15- 2	0.1668E 07	149.4	0.027	0.031	8.15	4700.3	576.9
15- 3	0.1668E 07	149.3	0.031	0.036	8.52	4547.2	533.8
15- 4	0.1668E 07	149.0	0.036	0.040	8.97	4716.3	525.7
15- 5	0.1668E 07	149.2	0.040	0.044	8.41	5040.4	599.3
15- 6	0.1668E 07	148.7	0.044	0.049	8.38	4656.5	555.5
15- 7	0.1668E 07	148.8	0.049	0.053	8.26	4753.9	575.4
15- 8	0.1668E 07	148.5	0.053	0.062	8.11	4787.4	590.2
15- 9	0.1668E 07	147.0	0.062	0.067	9.15	4636.6	507.0
15-AVER	0.1668E 07	148.7	0.023	0.066	8.44	4812.8	570.1
21- 1	0.1686E 07	150.2	0.086	0.095	16.23	11117.8	684.9
21- 2	0.1686E 07	150.1	0.095	0.103	15.36	11011.8	716.7
21- 3	0.1686E 07	150.1	0.103	0.113	17.35	11256.9	648.8
21- 4	0.1686E 07	149.8	0.113	0.122	17.08	11465.6	671.2
21- 5	0.1686E 07	149.7	0.122	0.132	17.11	11108.2	649.3
21- 6	0.1686E 07	149.3	0.132	0.142	16.70	11024.2	660.2
21- 7	0.1686E 07	149.3	0.142	0.152	17.68	11488.8	649.6
21- 8	0.1686E 07	148.8	0.152	0.166	17.69	11132.5	629.5
21- 9	0.1686E 07	147.2	0.166	0.176	19.10	11269.3	589.9
21-10	0.1686E 07	148.1	0.176	0.180	16.01	11180.0	698.3
21-AVER	0.1686E 07	149.3	0.086	0.180	17.07	11205.5	656.6
25- 1	0.1676E 07	149.5	0.002	0.014	19.02	16633.2	874.5
25- 2	0.1676E 07	149.7	0.014	0.024	17.38	15671.0	901.5
25- 3	0.1676E 07	150.2	0.024	0.038	18.92	15854.7	837.8

Table C-52 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
25- 4	0.1676E 07	149.3	0.038	0.052	20.23	16251.0	803.3
25- 5	0.1676E 07	149.6	0.052	0.065	18.76	16055.6	855.9
25- 6	0.1676E 07	148.8	0.065	0.080	18.71	16035.4	856.9
25- 7	0.1676E 07	148.9	0.080	0.093	19.97	16198.7	811.2
25- 8	0.1676E 07	148.4	0.093	0.112	20.85	15599.6	748.2
25- 9	0.1676E 07	146.4	0.112	0.127	22.51	15784.7	701.2
25-10	0.1676E 07	147.4	0.127	0.135	19.19	16525.8	861.1
25-AVER	0.1676E 07	148.8	0.002	0.135	19.58	16061.0	820.4
29- 1	0.1677E 07	146.5	0.012	0.021	19.75	20221.0	1023.8
29- 2	0.1677E 07	147.7	0.021	0.029	17.67	19682.5	1114.0
29- 3	0.1677E 07	149.2	0.029	0.041	18.12	20030.1	1105.1
29- 4	0.1677E 07	148.7	0.041	0.058	19.07	19916.5	1044.5
29- 5	0.1677E 07	148.7	0.058	0.075	17.26	20220.7	1171.5
29- 6	0.1677E 07	148.1	0.075	0.092	16.78	20429.2	1217.7
29- 7	0.1677E 07	148.2	0.092	0.109	17.46	19871.0	1138.2
29- 8	0.1677E 07	147.5	0.109	0.131	19.67	20024.3	1017.9
29- 9	0.1677E 07	145.6	0.131	0.149	18.70	20138.6	1076.9
29-10	0.1677E 07	146.7	0.149	0.158	15.56	20072.6	1290.3
29-AVER	0.1677E 07	147.7	0.012	0.158	18.05	20060.6	1111.2
31- 1	0.1691E 07	148.8	0.000	0.016	22.30	25502.3	1143.4
31- 2	0.1691E 07	149.4	0.016	0.031	19.36	23914.3	1235.1
31- 3	0.1691E 07	149.9	0.031	0.050	21.55	24150.1	1120.9
31- 4	0.1691E 07	149.4	0.050	0.071	22.21	24098.9	1085.0
31- 5	0.1691E 07	149.0	0.071	0.091	20.56	24104.5	1166.9
31- 6	0.1691E 07	148.6	0.091	0.112	19.36	24335.7	1256.9
31- 7	0.1691E 07	148.3	0.112	0.133	21.33	24124.8	1131.3
31- 8	0.1691E 07	147.7	0.133	0.158	24.91	23678.8	950.8
31- 9	0.1691E 07	145.5	0.158	0.179	25.45	23739.1	932.8
31-10	0.1691E 07	146.5	0.179	0.192	22.19	23878.4	1075.9
31-AVER	0.1691E 07	148.3	0.000	0.192	21.96	24152.7	1099.7
39- 1	0.1703E 07	150.1	0.038	0.039	14.03	7529.8	536.9
39- 2	0.1703E 07	150.9	0.039	0.044	11.56	7785.0	673.5
39- 3	0.1703E 07	150.6	0.044	0.050	13.32	8356.4	627.3
39- 4	0.1703E 07	151.0	0.050	0.057	13.30	8389.3	630.9
39- 5	0.1703E 07	150.4	0.057	0.066	12.94	8309.1	642.3
39- 6	0.1703E 07	150.0	0.066	0.072	12.42	7749.3	623.9
39- 7	0.1703E 07	150.2	0.072	0.079	12.74	8103.8	636.3
39- 8	0.1703E 07	149.6	0.079	0.091	12.86	7607.1	591.4
39- 9	0.1703E 07	148.2	0.091	0.098	14.53	8067.2	555.3
39-10	0.1703E 07	149.0	0.098	0.099	11.92	7355.7	617.2
39-AVER	0.1703E 07	150.0	0.038	0.099	12.99	7925.3	610.0
56- 1	0.1684E 07	150.3	0.031	0.033	12.32	7993.7	648.6
56- 2	0.1684E 07	151.1	0.033	0.036	10.69	7763.5	726.0
56- 3	0.1684E 07	151.2	0.036	0.043	11.26	7849.5	697.2

Table C-52 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
56- 4	0.1684E 07	150.7	0.043	0.050	11.60	7978.9	687.6
56- 5	0.1684E 07	151.1	0.050	0.056	10.72	7946.9	741.4
56- 6	0.1684E 07	150.6	0.056	0.063	10.50	7845.0	747.2
56- 7	0.1684E 07	150.8	0.063	0.070	10.82	7921.4	732.4
56- 8	0.1684E 07	150.5	0.070	0.079	10.89	7795.1	715.5
56- 9	0.1684E 07	149.5	0.079	0.087	11.28	7857.4	696.6
56-10	0.1684E 07	149.8	0.087	0.091	9.94	7792.2	784.0
56-AVER	0.1684E 07	150.5	0.031	0.091	11.04	7874.4	713.4

Table C-53

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 150^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
11- 1	0.3400E 07	151.1	0.035	0.035	11.17	8243.2	737.9
11- 2	0.3400E 07	151.6	0.035	0.037	10.21	7589.1	743.5
11- 3	0.3400E 07	151.8	0.037	0.040	10.83	7690.1	709.8
11- 4	0.3400E 07	151.4	0.040	0.045	11.15	7589.1	680.4
11- 5	0.3400E 07	151.2	0.045	0.051	10.74	7672.0	714.6
11- 6	0.3400E 07	150.4	0.051	0.057	10.63	7973.1	750.2
11- 7	0.3400E 07	150.0	0.057	0.063	10.76	7846.7	728.9
11- 8	0.3400E 07	149.2	0.063	0.074	10.91	8006.0	733.8
11- 9	0.3400E 07	146.9	0.074	0.083	12.14	7468.5	615.0
11-10	0.3400E 07	147.2	0.083	0.084	10.41	8109.6	778.9
11-AVER	0.3400E 07	150.1	0.035	0.084	10.95	7818.7	713.9
17- 1	0.3403E 07	152.2	0.022	0.023	7.07	4851.9	686.0
17- 2	0.3403E 07	152.5	0.023	0.023	6.78	4556.8	672.1
17- 3	0.3403E 07	152.6	0.023	0.026	6.85	4531.3	661.1
17- 4	0.3403E 07	152.0	0.026	0.030	7.34	4710.3	642.1
17- 5	0.3403E 07	151.9	0.030	0.035	7.00	4952.6	707.0
17- 6	0.3403E 07	151.0	0.035	0.039	7.06	4477.2	634.4
17- 7	0.3403E 07	150.9	0.039	0.044	6.95	4783.8	688.1
17- 8	0.3403E 07	150.0	0.044	0.053	7.30	4616.0	632.1
17- 9	0.3403E 07	148.3	0.053	0.059	8.51	4966.8	583.9
17-10	0.3403E 07	148.5	0.059	0.059	7.25	5208.0	717.9
17-AVER	0.3403E 07	151.0	0.022	0.059	7.25	4765.5	657.3
26- 1	0.3395E 07	148.1	0.040	0.043	19.96	17180.6	860.9
26- 2	0.3395E 07	148.8	0.043	0.043	18.12	15539.4	857.8
26- 3	0.3395E 07	150.2	0.043	0.044	18.37	15894.2	865.2
26- 4	0.3395E 07	150.7	0.044	0.048	18.37	16301.2	887.6
26- 5	0.3395E 07	151.2	0.048	0.054	16.53	16169.2	978.3
26- 6	0.3395E 07	150.5	0.054	0.062	16.12	15867.6	984.5
26- 7	0.3395E 07	150.4	0.062	0.071	17.08	16075.0	941.4
26- 8	0.3395E 07	149.6	0.071	0.086	18.14	16041.2	884.4
26- 9	0.3395E 07	147.2	0.086	0.098	19.40	16074.3	828.5
26-10	0.3395E 07	147.3	0.098	0.103	16.76	16273.4	971.2
26-AVER	0.3395E 07	149.4	0.040	0.103	17.92	16141.6	900.6
30- 1	0.3319E 07	148.6	0.045	0.047	24.40	20593.2	843.9
30- 2	0.3319E 07	149.8	0.047	0.048	21.55	19609.0	909.8
30- 3	0.3319E 07	150.8	0.048	0.054	22.96	19916.5	867.6
30- 4	0.3319E 07	150.5	0.054	0.064	23.95	19822.1	827.6
30- 5	0.3319E 07	150.2	0.064	0.075	23.52	20454.9	869.7
30- 6	0.3319E 07	149.2	0.075	0.087	22.85	20003.2	875.3
30- 7	0.3319E 07	148.7	0.087	0.099	23.50	20050.4	853.3
30- 8	0.3319E 07	147.5	0.099	0.118	23.99	19850.3	827.5
30- 9	0.3319E 07	144.7	0.118	0.132	24.88	19945.9	801.6
30-10	0.3319E 07	144.8	0.132	0.139	21.62	19902.3	920.3
30-AVER	0.3319E 07	148.5	0.045	0.139	23.40	20014.8	855.3

Table C-53 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
32- 1	0.3434E 07	148.2	0.054	0.057	22.50	23979.5	1066.0
32- 2	0.3434E 07	149.2	0.057	0.059	19.78	23489.8	1187.8
32- 3	0.3434E 07	150.7	0.059	0.063	20.67	23969.1	1159.7
32- 4	0.3434E 07	151.1	0.063	0.072	20.42	23770.1	1164.1
32- 5	0.3434E 07	150.9	0.072	0.084	18.59	24256.9	1305.1
32- 6	0.3434E 07	149.9	0.084	0.098	17.74	23708.6	1336.1
32- 7	0.3434E 07	149.3	0.098	0.112	19.67	23562.8	1197.6
32- 8	0.3434E 07	147.7	0.112	0.133	22.93	23983.1	1046.0
32- 9	0.3434E 07	144.8	0.133	0.150	22.19	24055.6	1083.9
32-10	0.3434E 07	144.7	0.150	0.159	18.97	23766.5	1253.1
32-AVER	0.3434E 07	148.6	0.054	0.159	20.41	23854.2	1168.9
57- 1	0.3365E 07	150.2	0.037	0.038	10.63	7993.7	752.0
57- 2	0.3365E 07	150.6	0.038	0.039	9.76	7750.8	793.9
57- 3	0.3365E 07	150.9	0.039	0.042	9.88	7849.5	794.2
57- 4	0.3365E 07	150.4	0.042	0.047	10.08	7969.8	790.4
57- 5	0.3365E 07	150.4	0.047	0.052	9.28	7956.0	857.8
57- 6	0.3365E 07	149.6	0.052	0.057	9.42	7835.7	832.2
57- 7	0.3365E 07	149.4	0.057	0.063	9.65	7912.1	820.2
57- 8	0.3365E 07	148.7	0.063	0.072	9.95	7784.8	782.1
57- 9	0.3365E 07	147.0	0.072	0.081	10.68	7857.4	735.5
57-10	0.3365E 07	146.7	0.081	0.085	9.65	7792.2	807.8
57-AVER	0.3365E 07	149.4	0.037	0.085	9.95	7870.2	791.1

Table C-54

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
89- 1	0.1237E 06	200.0	0.159	0.212	11.96	4627.7	386.9
89- 2	0.1237E 06	200.6	0.212	0.263	11.49	4483.5	390.2
89- 3	0.1237E 06	201.7	0.263	0.315	12.17	4603.1	378.3
89- 4	0.1237E 06	201.9	0.315	0.368	12.75	4463.2	350.1
89- 5	0.1237E 06	203.1	0.368	0.421	11.58	4712.6	407.0
89- 6	0.1237E 06	203.5	0.421	0.477	11.14	4716.7	423.5
89- 7	0.1237E 06	204.5	0.477	0.529	10.32	4656.3	451.0
89- 8	0.1237E 06	205.8	0.529	0.585	9.75	4611.8	473.1
89- 9	0.1237E 06	205.3	0.585	0.642	10.16	4610.4	453.8
89-10	0.1237E 06	206.8	0.642	0.695	8.64	4741.3	548.8
89-AVER	0.1237E 06	203.3	0.159	0.695	10.93	4622.7	422.9

Table C-55

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.24 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
86- 1	0.2684E 06	201.1	0.178	0.242	16.16	11243.7	695.9
86- 2	0.2684E 06	201.0	0.242	0.303	15.68	10884.7	694.0
86- 3	0.2684E 06	201.6	0.303	0.366	16.81	10885.5	647.5
86- 4	0.2684E 06	200.9	0.366	0.429	18.09	10989.1	607.5
86- 5	0.2684E 06	201.4	0.429	0.491	17.85	11158.6	625.2
86- 6	0.2684E 06	201.3	0.491	0.553	16.88	11142.8	659.9
86- 7	0.2684E 06	202.1	0.553	0.611	17.82	10980.1	616.1
86- 8	0.2684E 06	203.5	0.611	0.675	17.57	11061.6	629.4
86- 9	0.2684E 06	201.9	0.675	0.742	18.42	10924.3	593.1
86-10	0.2684E 06	201.7	0.742	0.806	20.51	11162.9	544.3
86-AVER	0.2684E 06	201.7	0.178	0.806	17.53	11043.3	630.0
88- 1	0.2663E 06	199.7	0.030	0.053	8.17	4627.7	566.6
88- 2	0.2663E 06	200.2	0.053	0.078	7.61	4483.5	588.8
88- 3	0.2663E 06	200.1	0.078	0.106	8.35	4603.1	551.4
88- 4	0.2663E 06	199.5	0.106	0.132	8.98	4463.2	497.1
88- 5	0.2663E 06	200.1	0.132	0.158	8.12	4721.7	581.4
88- 6	0.2663E 06	199.8	0.158	0.185	7.92	4726.0	597.0
88- 7	0.2663E 06	200.1	0.185	0.210	7.87	4675.0	594.2
88- 8	0.2663E 06	200.2	0.210	0.240	8.11	4632.4	571.5
88- 9	0.2663E 06	199.2	0.240	0.266	8.25	4622.1	560.2
88-10	0.2663E 06	200.2	0.266	0.288	7.02	4778.9	680.5
88-AVER	0.2663E 06	199.9	0.030	0.288	8.03	4633.4	576.7
94- 1	0.2559E 06	199.7	0.082	0.128	11.92	8056.5	676.0
94- 2	0.2559E 06	199.9	0.128	0.173	10.72	7677.4	716.0
94- 3	0.2559E 06	200.3	0.173	0.220	11.77	7866.9	668.3
94- 4	0.2559E 06	199.6	0.220	0.267	12.70	7806.9	614.7
94- 5	0.2559E 06	200.2	0.267	0.313	11.28	7745.9	686.8
94- 6	0.2559E 06	199.9	0.313	0.359	11.12	7828.1	703.9
94- 7	0.2559E 06	200.4	0.359	0.404	11.18	7848.5	701.8
94- 8	0.2559E 06	200.3	0.404	0.454	11.99	8008.3	668.0
94- 9	0.2559E 06	199.5	0.454	0.501	11.63	7842.4	674.1
94-10	0.2559E 06	200.2	0.501	0.545	10.11	7887.5	780.5
94-AVER	0.2559E 06	200.0	0.082	0.545	11.44	7856.8	686.5
95- 3	0.2576E 06	201.1	-.039	0.051	17.80	15714.1	882.8
95- 4	0.2576E 06	200.3	0.051	0.145	18.87	15621.1	827.9
95- 5	0.2576E 06	201.1	0.145	0.238	17.00	16096.0	946.8
95- 6	0.2576E 06	201.0	0.238	0.331	16.92	15877.1	938.2
95- 7	0.2576E 06	201.5	0.331	0.424	18.14	15814.9	871.8
95- 8	0.2576E 06	201.4	0.424	0.520	20.24	15953.6	788.2
95- 9	0.2576E 06	200.9	0.520	0.615	18.72	15758.4	841.6
95-10	0.2576E 06	201.2	0.615	0.708	18.00	15910.8	883.9
95-AVER	0.2576E 06	200.5	-.192	0.708	18.07	15820.2	875.3
120- 1	0.2188E 06	199.5	0.169	0.199	9.31	4673.8	501.8
120- 2	0.2188E 06	199.9	0.199	0.227	8.30	4360.8	525.6

Table C-55 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
120- 3	0.2188E 06	200.4	0.227	0.259	7.53	4463.1	592.8
120- 4	0.2188E 06	199.7	0.259	0.291	9.45	4543.2	480.7
120- 5	0.2188E 06	200.0	0.291	0.323	8.28	4497.1	543.0
120- 6	0.2188E 06	199.8	0.323	0.354	7.96	4593.7	577.1
120- 7	0.2188E 06	200.2	0.354	0.384	7.58	4504.5	594.6
120- 8	0.2188E 06	200.5	0.384	0.417	7.21	4706.8	653.0
120- 9	0.2188E 06	199.6	0.417	0.450	8.62	4410.0	511.9
120-10	0.2188E 06	199.9	0.450	0.480	8.44	4484.3	531.6
120-AVER	0.2188E 06	199.9	0.169	0.480	8.30	4523.7	545.2
121- 1	0.2247E 06	198.9	0.197	0.222	9.13	3825.7	418.9
121- 2	0.2247E 06	198.9	0.222	0.244	9.71	3669.6	378.1
121- 3	0.2247E 06	200.0	0.244	0.270	8.85	3691.5	417.0
121- 4	0.2247E 06	198.5	0.270	0.297	10.31	3799.4	368.7
121- 5	0.2247E 06	199.5	0.297	0.321	7.84	3768.0	480.5
121- 6	0.2247E 06	199.3	0.321	0.348	6.88	3755.4	545.9
121- 7	0.2247E 06	198.8	0.348	0.374	8.82	3746.0	424.6
121- 8	0.2247E 06	199.0	0.374	0.398	8.42	3820.3	453.5
121- 9	0.2247E 06	199.5	0.398	0.421	8.31	3739.3	449.7
121-10	0.2247E 06	199.8	0.421	0.444	7.10	3619.1	509.9
121-AVER	0.2247E 06	199.2	0.197	0.444	8.53	3743.4	438.9
123- 1	0.2237E 06	201.4	0.181	0.198	5.49	2144.2	390.2
123- 2	0.2237E 06	200.9	0.198	0.214	6.41	2061.5	321.7
123- 3	0.2237E 06	200.9	0.214	0.230	5.04	2122.8	421.3
123- 4	0.2237E 06	200.4	0.230	0.242	4.92	2024.8	411.4
123- 5	0.2237E 06	201.4	0.242	0.255	3.11	2157.7	692.9
123- 6	0.2237E 06	201.1	0.255	0.269	2.76	2129.4	770.9
123- 7	0.2237E 06	201.5	0.269	0.284	2.49	2110.9	847.0
123- 8	0.2237E 06	200.9	0.284	0.302	3.55	2102.1	591.6
123- 9	0.2237E 06	200.2	0.302	0.316	4.66	2098.5	450.0
123-10	0.2237E 06	201.2	0.316	0.324	2.95	2092.5	709.3
123-AVER	0.2237E 06	201.0	0.181	0.324	4.16	2104.4	505.4
124- 1	0.2228E 06	201.0	0.230	0.239	5.10	1297.0	254.1
124- 2	0.2228E 06	201.1	0.239	0.246	5.16	1250.7	242.5
124- 3	0.2228E 06	201.4	0.246	0.256	4.42	1387.1	313.8
124- 4	0.2228E 06	200.9	0.256	0.266	4.72	1311.1	277.9
124- 5	0.2228E 06	201.2	0.266	0.276	3.95	1467.4	371.8
124- 6	0.2228E 06	200.9	0.276	0.285	3.59	1353.5	377.0
124- 7	0.2228E 06	201.1	0.285	0.294	3.49	1371.1	392.7
124- 8	0.2228E 06	201.2	0.294	0.307	3.34	1345.5	402.4
124- 9	0.2228E 06	199.9	0.307	0.316	4.78	1284.2	268.9
124-10	0.2228E 06	201.1	0.316	0.319	2.75	1388.2	504.2
124-AVER	0.2228E 06	201.0	0.230	0.319	4.15	1345.6	324.1

Table C-56

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.48 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
42- 4	0.4673E 06	199.5	-.019	0.033	14.21	16086.8	1131.8
42- 5	0.4673E 06	198.5	0.033	0.090	14.15	15809.4	1116.4
42- 6	0.4673E 06	198.3	0.090	0.142	13.87	16622.2	1198.5
42- 7	0.4673E 06	198.6	0.142	0.193	15.08	15693.5	1040.4
42- 8	0.4673E 06	198.4	0.193	0.250	17.41	16079.8	923.6
42- 9	0.4673E 06	196.8	0.250	0.302	17.08	15891.0	930.5
42-10	0.4673E 06	198.2	0.302	0.346	14.67	15808.1	1077.5
42-AVER	0.4673E 06	198.3	-.164	0.346	14.84	15944.0	1074.6
44- 5	0.4738E 06	198.8	0.013	0.067	14.96	16129.8	1078.0
44- 6	0.4738E 06	198.6	0.067	0.117	14.31	15652.7	1093.5
44- 7	0.4738E 06	199.0	0.117	0.168	15.36	15974.5	1040.2
44- 8	0.4738E 06	198.8	0.168	0.224	16.65	15751.0	945.8
44- 9	0.4738E 06	197.0	0.224	0.276	16.85	15903.9	943.8
44-10	0.4738E 06	198.5	0.276	0.318	14.73	15975.5	1084.5
44-AVER	0.4738E 06	198.6	-.182	0.318	15.15	15946.9	1052.6
85- 1	0.5273E 06	199.9	0.111	0.142	11.99	11243.7	937.6
85- 2	0.5273E 06	200.2	0.142	0.171	10.85	10897.4	1004.7
85- 3	0.5273E 06	200.5	0.171	0.204	11.72	10895.9	929.6
85- 4	0.5273E 06	200.0	0.204	0.236	12.55	10989.1	875.4
85- 5	0.5273E 06	200.4	0.236	0.267	11.71	11158.6	952.6
85- 6	0.5273E 06	200.3	0.267	0.299	10.59	11152.1	1053.2
85- 7	0.5273E 06	200.7	0.299	0.330	11.13	10989.4	987.1
85- 8	0.5273E 06	200.6	0.330	0.364	12.77	11061.6	866.2
85- 9	0.5273E 06	199.7	0.364	0.397	12.58	10936.0	869.2
85-10	0.5273E 06	200.4	0.397	0.425	10.99	11188.0	1018.1
85-AVER	0.5273E 06	200.3	0.111	0.425	11.67	11051.2	946.9
87- 1	0.4603E 06	199.4	0.000	0.015	8.28	5199.9	627.9
87- 2	0.4603E 06	199.7	0.015	0.029	7.46	4470.8	599.4
87- 3	0.4603E 06	199.8	0.029	0.045	7.64	4603.1	602.2
87- 4	0.4603E 06	199.4	0.045	0.060	8.14	4472.3	549.2
87- 5	0.4603E 06	199.8	0.060	0.075	7.65	4721.7	617.6
87- 6	0.4603E 06	199.5	0.075	0.090	7.44	4726.0	635.1
87- 7	0.4603E 06	199.8	0.090	0.105	7.11	4675.0	657.6
87- 8	0.4603E 06	199.8	0.105	0.124	7.13	4642.7	651.1
87- 9	0.4603E 06	198.7	0.124	0.140	7.84	4622.1	589.6
87-10	0.4603E 06	199.7	0.140	0.149	6.32	4778.9	756.7
87-AVER	0.4603E 06	199.5	0.000	0.149	7.51	4691.2	624.3
91- 1	0.4714E 06	200.6	0.015	0.040	10.23	8196.7	801.5
91- 2	0.4714E 06	200.8	0.040	0.064	9.50	7665.4	807.1
91- 3	0.4714E 06	201.0	0.064	0.091	9.65	7723.4	800.1
91- 4	0.4714E 06	200.2	0.091	0.118	10.80	7932.5	734.4
91- 5	0.4714E 06	200.6	0.118	0.143	10.29	7730.9	751.4
91- 6	0.4714E 06	200.4	0.143	0.167	10.04	7887.9	785.5
91- 7	0.4714E 06	200.8	0.167	0.191	9.76	7878.4	807.1

Table C-56 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
91- 8	0.4714E 06	200.8	0.191	0.220	10.23	7813.3	763.9
91- 9	0.4714E 06	199.9	0.220	0.245	10.01	7799.6	778.8
91-10	0.4714E 06	200.8	0.245	0.265	8.31	7750.6	933.1
91-AVER	0.4714E 06	200.6	0.015	0.265	9.89	7837.9	792.8
93- 1	0.6056E 06	198.3	0.011	0.048	14.44	16105.6	1115.5
93- 2	0.6056E 06	198.8	0.048	0.084	13.20	15677.8	1187.4
93- 3	0.6056E 06	199.4	0.084	0.124	14.24	15742.2	1105.4
93- 4	0.6056E 06	198.6	0.124	0.164	15.09	15842.3	1049.7
93- 5	0.6056E 06	199.2	0.164	0.203	14.20	16036.2	1129.2
93- 6	0.6056E 06	199.0	0.203	0.242	13.80	15958.1	1156.6
93- 7	0.6056E 06	199.6	0.242	0.281	15.03	16012.2	1065.1
93- 8	0.6056E 06	199.5	0.281	0.323	16.34	15858.0	970.6
93- 9	0.6056E 06	198.9	0.323	0.363	15.38	16015.6	1041.4
93-10	0.6056E 06	199.3	0.363	0.400	14.16	15858.2	1120.3
93-AVER	0.6056E 06	199.0	0.011	0.400	14.57	15910.6	1092.2
96- 1	0.6416E 06	196.9	-.016	0.022	19.08	20022.8	1049.5
96- 2	0.6416E 06	198.2	0.022	0.061	17.01	19800.8	1164.0
96- 3	0.6416E 06	199.4	0.061	0.106	18.48	19780.3	1070.6
96- 4	0.6416E 06	198.5	0.106	0.154	17.78	19828.8	1115.2
96- 5	0.6416E 06	199.3	0.154	0.199	16.82	20023.5	1190.3
96- 6	0.6416E 06	199.2	0.199	0.245	16.93	20240.1	1195.5
96- 7	0.6416E 06	199.8	0.245	0.291	17.80	19865.4	1115.8
96- 8	0.6416E 06	199.8	0.291	0.339	19.69	19755.9	1003.5
96- 9	0.6416E 06	199.3	0.339	0.387	18.03	19928.6	1105.5
96-10	0.6416E 06	199.5	0.387	0.434	17.13	20226.5	1180.6
96-AVER	0.6416E 06	199.0	-.016	0.434	17.87	19947.3	1116.0

Table C-57

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
35- 2	0.8764E 06	200.3	0.009	0.043	16.70	19653.0	1176.6
35- 3	0.8764E 06	200.7	0.043	0.077	18.39	19985.7	1087.0
35- 4	0.8764E 06	200.4	0.077	0.113	18.39	20147.2	1095.4
35- 5	0.8764E 06	200.5	0.113	0.148	18.05	20090.1	1113.2
35- 6	0.8764E 06	200.4	0.148	0.183	17.98	20024.3	1113.8
35- 7	0.8764E 06	200.5	0.183	0.219	19.98	20125.1	1007.3
35- 8	0.8764E 06	200.2	0.218	0.259	22.60	19697.8	871.5
35- 9	0.8764E 06	198.1	0.259	0.295	23.19	19852.3	855.9
35-10	0.8764E 06	199.9	0.295	0.319	16.62	19868.4	1195.4
35-AVER	0.8764E 06	200.1	-.026	0.319	18.88	19911.3	1054.7
37- 1	0.8989E 06	199.3	0.005	0.033	14.53	16314.6	1122.8
37- 2	0.8989E 06	199.2	0.033	0.058	14.02	15610.0	1113.4
37- 3	0.8989E 06	199.8	0.058	0.084	15.44	15973.4	1034.5
37- 4	0.8989E 06	199.4	0.084	0.112	15.49	15921.8	1027.8
37- 5	0.8989E 06	199.4	0.112	0.140	14.37	16070.7	1118.1
37- 6	0.8989E 06	199.1	0.140	0.167	14.50	16160.9	1114.6
37- 7	0.8989E 06	199.5	0.167	0.194	15.54	15842.9	1019.3
37- 8	0.8989E 06	199.2	0.194	0.227	17.63	15704.2	890.7
37- 9	0.8989E 06	197.4	0.227	0.254	18.08	15598.0	862.5
37-10	0.8989E 06	199.0	0.254	0.271	14.65	16071.7	1097.0
37-AVER	0.8989E 06	199.1	0.005	0.271	15.43	15926.8	1032.5
49- 6	0.8979E 06	190.5	0.003	0.020	9.64	9929.3	1029.9
49- 7	0.8979E 06	190.6	0.020	0.036	10.20	10157.3	995.4
49- 8	0.8979E 06	190.7	0.036	0.056	10.13	10153.2	1001.9
49- 9	0.8979E 06	189.6	0.056	0.073	9.94	9827.9	988.5
49-10	0.8979E 06	190.3	0.073	0.085	8.33	10020.2	1202.4
49-AVER	0.8979E 06	190.4	-.069	0.085	10.15	9954.1	980.5
90- 1	0.9435E 06	201.1	0.123	0.134	9.32	8183.8	877.8
90- 2	0.9435E 06	201.4	0.134	0.146	8.52	7665.4	899.6
90- 3	0.9435E 06	201.5	0.146	0.161	8.96	7723.4	861.5
90- 4	0.9435E 06	200.7	0.161	0.174	9.65	7932.5	821.8
90- 5	0.9435E 06	201.2	0.174	0.187	8.62	7739.9	898.0
90- 6	0.9435E 06	200.9	0.187	0.199	8.47	7887.9	930.9
90- 7	0.9435E 06	201.3	0.199	0.211	8.18	7878.4	963.4
90- 8	0.9435E 06	201.2	0.211	0.227	8.44	7813.3	926.0
90- 9	0.9435E 06	200.3	0.227	0.240	8.33	7799.6	936.1
90-10	0.9435E 06	201.0	0.240	0.248	6.82	7750.6	1136.2
90-AVER	0.9435E 06	201.1	0.123	0.248	8.56	7837.5	915.5
92- 1	0.9121E 06	196.1	0.002	0.021	14.62	16118.5	1102.4
92- 2	0.9121E 06	197.3	0.021	0.041	13.01	15685.5	1205.6
92- 3	0.9121E 06	197.9	0.041	0.066	14.31	15748.4	1100.8
92- 4	0.9121E 06	197.3	0.066	0.092	14.71	15842.3	1077.0
92- 5	0.9121E 06	198.1	0.092	0.117	12.53	16036.2	1279.3
92- 6	0.9121E 06	197.9	0.117	0.142	12.13	15958.1	1316.0

Table C-57 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
92- 7	0.9121E 06	198.5	0.142	0.167	12.81	16021.5	1251.0
92- 8	0.9121E 06	198.3	0.167	0.196	13.69	15878.6	1160.2
92- 9	0.9121E 06	197.8	0.196	0.222	12.73	16015.6	1258.0
92-10	0.9121E 06	198.3	0.222	0.246	11.10	15858.2	1428.1
92-AVER	0.9121E 06	197.7	0.002	0.246	13.19	15916.3	1206.6
97- 2	0.8893E 06	198.7	-.009	0.017	16.11	19903.2	1235.2
97- 3	0.8893E 06	199.6	0.017	0.051	17.89	19991.1	1117.4
97- 4	0.8893E 06	198.7	0.051	0.086	18.19	19852.7	1091.6
97- 5	0.8893E 06	199.5	0.086	0.118	15.90	20059.3	1261.9
97- 6	0.8893E 06	199.5	0.118	0.150	15.38	20213.6	1314.3
97- 7	0.8893E 06	200.0	0.150	0.183	16.60	19915.8	1199.8
97- 8	0.8893E 06	199.8	0.183	0.219	17.86	19670.6	1101.3
97- 9	0.8893E 06	199.4	0.219	0.254	15.90	19940.3	1253.8
97-10	0.8893E 06	199.7	0.254	0.287	14.93	20215.7	1354.3
97-AVER	0.8893E 06	199.3	-.035	0.287	16.72	19994.6	1195.6
100- 2	0.9323E 06	201.0	0.003	0.022	11.47	10718.1	934.5
100- 3	0.9323E 06	201.3	0.022	0.041	12.74	10915.4	856.6
100- 4	0.9323E 06	200.8	0.041	0.059	12.80	10986.5	858.6
100- 5	0.9323E 06	201.2	0.059	0.077	11.17	11206.6	1003.1
100- 6	0.9323E 06	201.0	0.077	0.094	10.64	11274.3	1060.0
100- 7	0.9323E 06	201.5	0.094	0.111	11.39	11112.4	975.3
100- 8	0.9323E 06	201.5	0.111	0.132	11.92	11075.2	929.5
100- 9	0.9323E 06	200.7	0.132	0.150	11.81	10965.6	928.8
100-10	0.9323E 06	201.4	0.150	0.163	9.80	11021.2	1124.9
100-AVER	0.9323E 06	201.2	-.019	0.163	11.58	11030.4	952.6
103- 1	0.9033E 06	198.9	0.055	0.064	10.52	7962.6	756.7
103- 2	0.9033E 06	199.6	0.064	0.074	9.51	7617.6	800.8
103- 3	0.9033E 06	199.9	0.074	0.088	10.08	7735.3	767.2
103- 4	0.9033E 06	199.2	0.088	0.102	10.40	7711.5	741.7
103- 5	0.9033E 06	199.7	0.102	0.114	9.41	7799.7	829.0
103- 6	0.9033E 06	199.5	0.114	0.127	9.13	7654.1	838.0
103- 7	0.9033E 06	199.8	0.127	0.139	9.05	7872.4	869.6
103- 8	0.9033E 06	199.8	0.139	0.155	9.11	7817.6	858.1
103- 9	0.9033E 06	198.9	0.155	0.169	9.47	7686.0	811.3
103-10	0.9033E 06	199.6	0.169	0.177	8.18	7876.1	962.7
103-AVER	0.9033E 06	199.5	0.055	0.177	9.52	7773.3	816.8
105- 1	0.9161E 06	201.6	0.065	0.082	6.45	4579.9	710.3
105- 2	0.9161E 06	200.2	0.082	0.093	7.65	4375.9	572.1
105- 3	0.9161E 06	200.6	0.093	0.100	7.73	4549.3	588.4
105- 4	0.9161E 06	200.2	0.100	0.108	8.21	4606.9	561.1
105- 5	0.9161E 06	200.6	0.108	0.115	7.16	4665.0	651.8
105- 6	0.9161E 06	200.3	0.115	0.122	6.92	4678.2	676.1
105- 7	0.9161E 06	200.8	0.122	0.129	6.44	4591.3	713.2
105- 8	0.9161E 06	200.7	0.129	0.140	6.52	4614.4	707.7

Table C-57 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
105- 9	0.9161E 06	199.6	0.140	0.149	7.18	4687.8	652.7
105-10	0.9161E 06	200.5	0.149	0.151	5.84	4909.9	841.0
105-AVER	0.9161E 06	200.5	0.065	0.151	7.03	4625.9	658.0

Table C-58

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 1.70 x 10^6 lbm/hr-sq ft, T ≈ 200°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
34- 6	0.1659E 07	199.6	-.008	0.010	13.97	20219.0	1447.4
34- 7	0.1659E 07	199.8	0.010	0.029	14.83	19709.9	1329.5
34- 8	0.1659E 07	199.5	0.029	0.055	16.73	19676.3	1176.0
34- 9	0.1659E C7	197.4	0.055	0.074	16.32	19768.3	1211.4
34-10	0.1659E C7	199.1	0.074	0.082	13.01	19773.3	1519.7
38- 6	0.1671E 07	199.8	0.001	0.017	15.62	15841.4	1014.4
38- 7	0.1671E 07	199.9	0.017	0.032	16.13	15783.1	978.5
38- 8	0.1671E 07	199.5	0.032	0.053	16.14	15771.6	977.3
38- 9	0.1671E 07	197.8	0.053	0.069	16.40	15831.2	965.3
38-10	0.1671E 07	199.2	0.069	0.075	13.61	16094.5	1182.6
38-AVER	0.1671E 07	199.8	-.073	0.075	15.47	15884.7	1026.9
41- 4	0.1693E 07	200.4	-.001	0.015	15.73	16083.2	1022.3
41- 5	0.1693E 07	199.2	0.015	0.035	16.27	15804.0	971.3
41- 6	0.1693E 07	198.9	0.035	0.049	16.15	16426.6	1016.9
41- 7	0.1693E 07	199.3	0.049	0.063	15.20	15693.5	1032.3
41- 8	0.1693E 07	198.8	0.063	0.085	15.70	16079.8	1023.9
41- 9	0.1693E 07	197.0	0.085	0.101	16.81	15902.6	946.0
41-10	0.1693E 07	198.4	0.101	0.105	13.14	15833.1	1205.2
41-AVER	0.1693E 07	199.0	-.037	0.105	15.51	15930.0	1026.8
50- 7	0.1716E 07	189.0	-.000	0.008	9.99	10157.3	1017.0
50- 8	0.1716E 07	188.8	0.008	0.020	10.16	10163.5	1000.6
50- 9	0.1716E 07	187.8	0.020	0.030	9.85	9804.5	995.3
50-10	0.1716E 07	188.3	0.030	0.035	8.33	10020.2	1203.1
53- 1	0.1657E 07	190.7	0.012	0.015	12.47	7913.9	634.4
53- 2	0.1657E 07	191.3	0.015	0.019	10.78	7502.6	695.9
53- 3	C.1657E 07	191.5	0.019	0.027	11.44	7851.8	686.1
53- 4	0.1657E 07	191.1	0.027	0.034	11.71	7789.8	665.2
53- 5	0.1657E 07	191.3	0.034	0.041	11.83	7901.4	668.1
53- 6	0.1657E 07	191.1	0.041	0.048	11.35	7689.4	677.4
53- 7	0.1657E 07	191.3	0.048	0.055	11.43	7867.2	688.4
53- 8	0.1657E 07	191.1	0.055	0.066	11.15	7726.3	693.1
53- 9	0.1657E 07	190.0	0.066	0.074	12.03	7901.6	656.9
53-10	0.1657E 07	190.7	0.074	0.077	10.08	7746.3	768.1
53-AVER	0.1657E 07	191.0	0.012	0.077	11.45	7789.0	680.0
84- 1	0.1000E 07	201.3	0.221	0.237	10.52	11243.7	1068.8
84- 2	0.1000E 07	201.4	0.237	0.252	10.11	10897.4	1078.2
84- 3	0.1000E 07	201.8	0.252	0.270	10.50	10885.5	1036.6
84- 4	0.1000E 07	201.2	0.270	0.287	10.97	10989.1	1002.1
84- 5	0.1000E 07	201.6	0.287	0.303	10.21	11158.6	1092.9
84- 6	0.1000E 07	201.4	0.303	0.320	9.46	11142.8	1178.2
84- 7	C.1000E 07	201.7	0.320	0.336	9.58	10970.8	1145.4
84- 8	0.1000E 07	201.7	0.336	0.355	9.97	11051.3	1107.9
84- 9	0.1000E 07	200.8	0.355	0.373	9.77	10936.0	1118.8
84-10	0.1000E 07	201.5	0.373	0.386	8.20	11188.0	1364.1

Table C-58 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
84-AVER	0.1000E 07	201.4	0.221	0.386	9.95	11046.3	1109.6
98- 3	0.1731E 07	200.6	-.006	0.013	16.60	19991.1	1204.6
98- 4	0.1731E 07	199.4	0.013	0.032	17.52	19843.7	1126.4
98- 5	0.1731E 07	200.3	0.032	0.048	15.87	20059.3	1264.1
98- 6	0.1731E 07	200.0	0.048	0.065	15.11	20213.6	1337.8
98- 7	0.1731E 07	200.5	0.065	0.081	15.48	19925.2	1287.0
98- 8	0.1731E 07	200.2	0.081	0.101	15.64	19670.6	1257.8
98- 9	0.1731E 07	199.7	0.101	0.120	14.08	19951.9	1417.2
98-10	0.1731E 07	199.8	0.120	0.137	12.97	20228.3	1559.9
98-AVER	0.1731E 07	200.0	-.037	0.137	15.47	19997.0	1292.3
101- 3	0.1731E 07	200.6	0.0	0.011	11.59	10925.8	942.6
101- 4	0.1731E 07	199.9	0.011	0.022	12.35	10977.4	888.7
101- 5	0.1731E 07	200.3	0.022	0.031	11.41	11206.6	981.8
101- 6	0.1731E 07	200.1	0.031	0.040	10.81	11265.0	1041.7
101- 7	0.1731E 07	200.4	0.040	0.049	10.68	11103.0	1039.3
101- 8	0.1731E 07	200.3	0.049	0.063	10.71	11075.2	1034.4
101- 9	0.1731E 07	199.3	0.063	0.074	10.67	10965.6	1027.3
101-10	0.1731E 07	199.8	0.074	0.080	9.17	11021.2	1201.6
101-AVER	0.1731E 07	200.1	-.019	0.080	11.01	11031.3	1002.3
104- 1	0.1711E 07	199.7	0.031	0.035	9.56	7975.5	834.1
104- 2	0.1711E 07	200.2	0.035	0.040	8.61	7630.3	886.5
104- 3	0.1711E 07	200.3	0.040	0.049	9.49	7745.7	816.1
104- 4	0.1711E 07	199.5	0.049	0.057	10.18	7711.5	757.4
104- 5	0.1711E 07	200.0	0.057	0.064	9.09	7799.7	858.4
104- 6	0.1711E 07	199.6	0.064	0.070	8.59	7682.0	894.0
104- 7	0.1711E 07	200.1	0.070	0.076	8.49	7881.8	928.8
104- 8	0.1711E 07	200.0	0.076	0.087	8.47	7817.6	923.5
104- 9	0.1711E 07	198.9	0.087	0.095	8.72	7697.7	882.9
104-10	0.1711E 07	199.6	0.095	0.097	7.49	7876.1	1051.2
104-AVER	0.1711E 07	199.8	0.031	0.097	8.90	7781.8	874.3

Table C-59

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 200°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
33- 1	0.3377E 07	199.5	-.000	0.011	18.79	26235.0	1395.9
33- 2	0.3377E 07	199.6	0.011	0.021	17.48	23741.5	1358.6
33- 3	0.3377E 07	199.9	0.021	0.033	19.73	23941.1	1213.3
33- 4	0.3377E 07	199.3	0.033	0.046	20.56	23823.4	1153.1
33- 5	0.3377E 07	199.0	0.046	0.060	19.95	24220.7	1214.1
33- 6	0.3377E 07	198.3	0.060	0.073	19.87	23477.9	1181.3
33- 7	0.3377E 07	198.3	0.073	0.087	21.18	23572.5	1113.0
33- 8	0.3377E 07	197.4	0.087	0.108	23.19	23646.3	1019.6
33- 9	0.3377E 07	195.0	0.108	0.122	22.87	23949.8	1047.0
33-10	0.3377E 07	196.3	0.122	0.124	18.70	23810.9	1273.3
33-AVER	0.3377E 07	198.3	-.000	0.124	20.28	24041.9	1185.7
36- 1	0.3389E 07	197.1	0.030	0.036	16.73	20129.3	1203.3
36- 2	0.3389E 07	197.7	0.036	0.038	14.75	19808.5	1343.0
36- 4	0.3389E 07	201.3	0.035	0.043	13.05	19492.6	1493.2
36- 5	0.3389E 07	199.2	0.043	0.061	13.85	20149.8	1454.4
36- 6	0.3389E 07	198.7	0.061	0.071	13.13	19680.9	1498.8
36- 7	0.3389E 07	198.8	0.071	0.082	14.10	19773.1	1402.5
36- 8	0.3389E 07	198.1	0.082	0.100	15.83	19702.1	1244.4
36- 9	0.3389E 07	195.8	0.100	0.113	15.27	19648.2	1286.8
36-10	0.3389E 07	197.1	0.113	0.113	11.90	19606.5	1648.1
36-AVER	0.3389E 07	198.3	0.030	0.113	14.44	19822.6	1372.5
48- 1	0.3373E 07	199.6	0.016	0.019	17.95	15850.0	883.1
48- 2	0.3373E 07	200.3	0.019	0.020	15.30	15529.3	1014.8
48- 3	0.3373E 07	201.3	0.020	0.027	15.97	15975.9	1000.4
48- 4	0.3373E 07	200.5	0.027	0.035	16.94	15956.1	942.1
48- 5	0.3373E 07	200.8	0.035	0.043	16.50	15671.2	943.8
48- 6	0.3373E 07	200.4	0.043	0.051	16.05	15798.0	984.3
48- 7	0.3373E 07	200.4	0.051	0.060	16.36	15800.3	966.1
48- 8	0.3373E 07	200.0	0.060	0.072	18.99	15756.3	829.9
48- 9	0.3373E 07	198.9	0.072	0.082	16.54	15890.4	960.7
48-10	0.3373E 07	199.0	0.082	0.089	14.64	15834.3	1081.6
48-AVER	0.3373E 07	200.1	0.016	0.089	16.54	15806.2	955.7
51- 3	0.3440E 07	192.1	0.025	0.029	9.93	9895.5	996.7
51- 4	0.3440E 07	191.6	0.029	0.034	10.40	9791.4	941.4
51- 5	0.3440E 07	192.1	0.034	0.038	9.30	10011.2	1076.8
51- 6	0.3440E 07	191.7	0.038	0.042	9.04	9920.0	1097.0
51- 7	0.3440E 07	191.9	0.042	0.047	9.14	10157.3	1111.7
51- 8	0.3440E 07	191.4	0.047	0.058	9.50	10153.2	1057.8
51- 9	0.3440E 07	190.1	0.058	0.065	9.64	9816.2	1018.0
51-10	0.3440E 07	190.4	0.065	0.067	8.11	10032.7	1236.7
51-AVER	0.3440E 07	191.3	0.030	0.067	9.57	9947.9	1028.5
52- 1	0.3372E 07	198.4	0.016	0.020	10.34	7901.1	764.1
52- 2	0.3372E 07	198.4	0.020	0.021	9.55	7489.9	784.1
52- 3	0.3372E 07	199.0	0.021	0.025	9.74	7841.4	804.7

Table C-59 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
52- 4	0.3372E 07	198.5	0.025	0.029	10.11	7771.7	768.9
52- 5	0.3372E 07	198.6	0.029	0.034	9.96	7892.3	792.6
52- 6	0.3372E 07	198.3	0.034	0.038	9.50	7680.1	808.6
52- 7	0.3372E 07	198.3	0.038	0.043	9.58	7857.8	819.8
52- 8	0.3372E 07	197.9	0.043	0.052	9.47	7715.9	814.4
52- 9	0.3372E 07	196.5	0.052	0.059	10.58	7889.9	746.0
52-10	0.3372E 07	197.0	0.059	0.059	8.58	7733.8	890.8
52-AVER	0.3372E 07	198.1	0.016	0.059	9.79	7777.4	794.5
99- 6	0.3411E 07	200.5	-.002	0.008	14.00	20213.6	1443.6
99- 7	0.3411E 07	200.8	0.008	0.017	14.20	19934.5	1403.4
99- 8	0.3411E 07	200.3	0.017	0.031	14.94	19670.6	1316.6
99- 9	0.3411E 07	199.4	0.031	0.044	13.69	19951.9	1457.3
99-10	0.3411E 07	199.2	0.044	0.054	12.48	20253.3	1623.5
99-AVER	0.3411E 07	200.1	-.029	0.054	14.98	20006.0	1335.3
102- 1	0.3411E 07	201.2	0.026	0.028	12.75	11429.1	896.4
102- 2	0.3411E 07	201.6	0.028	0.030	11.26	10707.6	950.8
102- 3	0.3411E 07	201.8	0.030	0.038	12.12	11381.8	938.7
102- 4	0.3411E 07	200.9	0.038	0.045	13.03	10911.4	837.7
102- 5	0.3411E 07	201.2	0.045	0.051	12.14	11329.2	932.9
102- 6	0.3411E 07	200.7	0.051	0.057	11.53	11704.8	1006.8
102- 7	0.3411E 07	200.8	0.057	0.063	11.46	11243.9	981.1
102- 8	0.3411E 07	200.4	0.063	0.073	11.67	11505.0	985.9
102- 9	0.3411E 07	199.3	0.073	0.081	11.85	11336.0	956.2
102-10	0.3411E 07	199.6	0.081	0.084	10.35	11393.0	1100.5
102-AVER	0.3411E 07	200.8	0.026	0.084	11.88	11294.2	951.0
107- 1	0.3429E 07	200.6	0.022	0.025	6.18	4825.0	780.3
107- 2	0.3429E 07	200.5	0.025	0.028	6.14	4597.1	748.5
107- 3	0.3429E 07	200.3	0.028	0.033	6.42	4653.7	724.4
107- 4	0.3429E 07	199.8	0.033	0.036	7.05	4571.1	648.5
107- 5	0.3429E 07	199.8	0.036	0.040	6.74	4718.8	700.2
107- 6	0.3429E 07	199.4	0.040	0.043	6.57	4606.5	701.6
107- 7	0.3429E 07	199.4	0.043	0.046	6.27	4612.6	735.3
107- 8	0.3429E 07	199.1	0.046	0.054	6.34	4798.1	757.3
107- 9	0.3429E 07	197.7	0.054	0.058	7.41	4621.8	623.4
107-AVER	0.3429E 07	199.5	0.022	0.055	6.50	4674.3	719.1

Table C-60

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.12 \times 10^6$ lbm/hr-sq ft, $T \approx 250^{\circ}F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR158- 6	0.1117E 06	249.3	0.017	0.124	10.54	6101.4	578.7
FR158- 7	0.1117E 06	249.8	0.124	0.227	10.17	6115.3	601.2
FR158- 8	0.1117E 06	250.9	0.227	0.335	9.74	6086.0	624.8
FR158- 9	0.1117E 06	249.9	0.335	0.441	11.16	5897.1	528.4

Table C-61

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.24 \times 10^6$ lbm/hr-sq ft, $T \approx 250^\circ F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR147- 6	0.2308E 06	252.7	0.019	0.130	14.53	12659.2	871.3
FR147- 7	0.2308E 06	252.9	0.130	0.239	14.30	12398.0	867.2
FR147- 8	0.2308E 06	253.2	0.239	0.350	14.83	12506.0	843.1
FR147- 9	0.2308E 06	252.9	0.350	0.454	15.93	12252.6	769.0
FR147-10	0.2308E 06	254.8	0.454	0.553	13.62	12174.2	894.1
FR147-AVER	0.2308E 06	252.6	-.463	0.553	14.76	12372.1	838.3
FR153- 6	0.2177E 06	248.8	-.003	0.078	12.46	9341.7	749.6
FR153- 7	0.2177E 06	249.6	0.078	0.157	11.43	9154.6	800.9
FR153- 8	0.2177E 06	249.9	0.157	0.245	11.41	9437.0	827.2
FR153- 9	0.2177E 06	249.0	0.245	0.330	12.59	9094.4	722.5
FR159- 7	0.2158E 06	248.0	-.004	0.046	8.41	6134.0	729.1
FR159- 8	0.2158E 06	249.2	0.046	0.101	7.35	6096.4	828.9
FR159- 9	0.2158E 06	248.1	0.101	0.160	8.18	5897.1	720.5
FR159-10	0.2158E 06	248.3	0.160	0.212	7.16	5988.8	836.4

Table C-62

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 0.48 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR 68- 4	0.4068E 06	247.0	-.026	0.052	24.78	16128.9	650.8
FR 68- 5	0.4068E 06	247.4	0.052	0.128	25.03	16011.6	639.7
FR 68- 6	0.4068E 06	247.3	0.128	0.204	24.79	16118.6	650.3
FR 68- 7	0.4068E 06	247.9	0.204	0.280	25.17	16260.2	646.1
FR 76- 1	0.4653E 06	244.7	0.533	0.558	7.22	4802.3	665.1
FR 76- 2	0.4653E 06	243.7	0.558	0.576	8.92	4379.4	490.9
FR 76- 3	0.4653E 06	244.9	0.576	0.592	4.04	4609.7	1139.6
FR 76- 4	0.4653E 06	244.7	0.592	0.611	4.25	4459.7	1048.9
FR 76- 5	0.4653E 06	244.8	0.611	0.630	7.44	4541.5	610.1
FR 76- 6	0.4653E 06	244.6	0.630	0.649	6.03	4577.0	759.5
FR 76- 7	0.4653E 06	245.0	0.668	0.689	5.47	4668.5	853.3
FR 76- 9	0.4653E 06	243.7	0.689	0.709	4.79	4667.1	974.9
FR 76-10	0.4653E 06	244.9	0.709	0.725	3.93	5022.6	1279.0
FR 76-AVER	0.4653E 06	244.6	0.533	0.725	4.27	4634.8	1086.4
FR148- 2	0.4744E 06	253.4	-.021	0.027	24.56	12167.7	495.5
FR148- 3	0.4744E 06	254.1	0.027	0.082	24.85	12340.0	496.6
FR148- 7	0.4744E 06	253.1	0.246	0.301	27.03	12388.6	458.3
FR148- 8	0.4744E 06	253.4	0.301	0.358	24.41	12495.7	512.0
FR148- 9	0.4744E 06	252.3	0.358	0.412	26.60	12240.9	460.2
FR148-10	0.4744E 06	253.3	0.412	0.458	22.68	12174.2	536.8
FR154- 1	0.4926E 06	249.9	-.016	0.017	11.27	9414.9	835.0
FR154- 2	0.4926E 06	250.5	0.017	0.051	10.12	8931.8	882.3
FR154- 3	0.4926E 06	250.5	0.051	0.089	10.55	9022.4	855.3
FR154- 4	0.4926E 06	250.2	0.089	0.128	11.29	9260.0	819.8
FR154- 5	0.4926E 06	250.2	0.128	0.166	10.30	9125.2	886.3
FR154- 6	0.4926E 06	250.2	0.166	0.203	9.64	9341.7	968.9
FR154- 7	0.4926E 06	250.3	0.203	0.240	9.55	9164.0	959.2
FR154- 8	0.4926E 06	250.5	0.240	0.282	8.82	9447.3	1070.7
FR154- 9	0.4926E 06	249.1	0.282	0.321	11.09	9141.0	824.2
FR154-10	0.4926E 06	250.2	0.321	0.350	8.89	9224.1	1037.2
FR154-AVER	0.4926E 06	250.2	-.016	0.350	10.18	9207.2	904.3
FR160- 1	0.5100E 06	246.8	0.090	0.111	8.32	6074.0	730.2
FR160- 2	0.5100E 06	247.0	0.111	0.132	7.65	5891.9	770.1
FR160- 3	0.5100E 06	247.1	0.132	0.155	7.63	5928.9	777.4
FR160- 4	0.5100E 06	247.0	0.155	0.178	7.66	5985.9	781.0
FR160- 5	0.5100E 06	247.1	0.178	0.201	7.17	5975.6	833.6
FR160- 6	0.5100E 06	246.8	0.201	0.226	7.12	6110.7	857.8
FR160- 7	0.5100E 06	246.9	0.226	0.248	6.88	6143.3	892.3
FR160- 8	0.5100E 06	247.1	0.248	0.275	6.72	6106.7	908.8
FR160- 9	0.5100E 06	245.9	0.275	0.298	7.73	5920.4	765.6
FR160-10	0.5100E 06	247.2	0.298	0.310	5.78	5800.9	1002.9
FR160-AVER	0.5100E 06	246.9	0.090	0.310	7.28	5993.8	822.8

Table C-63

HEAT TRANSFER - EXPERIMENTAL DATA,
G ≈ 0.96 x 10^6 lbm/hr-sq ft, T ≈ 250°F

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
108- 1	0.9781E 06	248.8	0.012	0.043	12.81	15524.5	1211.5
108- 2	0.9781E 06	248.9	0.043	0.072	11.94	15488.8	1297.4
108- 3	0.9781E 06	249.3	0.072	0.105	13.37	15793.8	1180.9
108- 4	0.9781E 06	248.6	0.105	0.138	14.16	15566.4	1099.5
108- 5	0.9781E 06	249.2	0.138	0.168	13.11	15970.1	1218.4
108- 6	0.9781E 06	249.1	0.168	0.199	12.80	15691.6	1225.8
108- 7	0.9781E 06	249.5	0.199	0.229	13.49	15629.8	1159.0
108- 8	0.9781E 06	249.5	0.229	0.265	14.79	15950.6	1078.8
108- 9	0.9781E 06	248.7	0.265	0.297	13.63	16041.3	1176.6
108-10	0.9781E 06	249.5	0.297	0.323	11.66	15399.2	1320.8
108-AVER	0.9781E 06	249.1	0.012	0.323	13.16	15705.6	1193.2
FR 67- 1	0.9805E 06	248.3	0.030	0.059	16.97	15749.9	927.9
FR 67- 2	0.9805E 06	248.6	0.059	0.087	15.91	15547.1	977.1
FR 67- 3	0.9805E 06	249.0	0.087	0.120	17.84	15751.9	883.0
FR 67- 5	0.9805E 06	248.8	0.154	0.185	15.15	16020.7	1057.8
FR 67- 6	0.9805E 06	248.7	0.185	0.216	15.54	16118.6	1037.2
FR 67- 7	0.9805E 06	249.0	0.216	0.248	15.63	16260.2	1040.4
FR 67- 8	0.9805E 06	249.0	0.248	0.283	18.39	15889.1	863.8
FR 67-AVER	0.9805E 06	248.7	0.030	0.345	18.98	15922.9	838.9
FR 71- 5	0.9813E 06	243.6	0.053	0.073	9.39	11092.3	1181.9
FR 71- 6	0.9813E 06	243.5	0.073	0.094	7.79	10885.4	1397.5
FR 71- 7	0.9813E 06	243.7	0.094	0.113	8.10	10809.6	1334.5
FR 71- 8	0.9813E 06	243.8	0.113	0.139	8.39	10835.0	1291.0
FR 71- 9	0.9813E 06	242.6	0.139	0.161	8.39	11085.1	1321.4
FR 71-10	0.9813E 06	243.7	0.161	0.174	6.48	11409.8	1761.7
FR 71-AVER	0.9813E 06	243.5	-.030	0.174	7.11	11000.0	1547.4
FR 77- 1	0.8959E 06	251.5	0.227	0.244	5.96	4700.7	788.6
FR 77- 2	0.8959E 06	250.7	0.244	0.257	6.18	4499.0	728.5
FR 77- 3	0.8959E 06	250.9	0.257	0.267	6.01	4466.2	743.3
FR 77- 4	0.8959E 06	250.8	0.267	0.277	5.49	4483.6	816.1
FR 77- 5	0.8959E 06	250.9	0.277	0.288	5.55	4601.3	829.0
FR 77- 6	0.8959E 06	250.6	0.288	0.299	5.37	4717.1	878.2
FR 77- 7	0.8959E 06	250.8	0.299	0.309	5.17	4509.8	872.2
FR 77- 8	0.8959E 06	250.9	0.309	0.324	5.49	4511.4	821.3
FR 77- 9	0.8959E 06	249.4	0.324	0.334	6.33	4546.9	718.9
FR 77-AVER	0.8959E 06	250.7	0.227	0.333	5.63	4593.0	816.4
FR149- 1	0.8930E 06	250.3	0.151	0.171	12.34	12325.3	998.6
FR149- 2	0.8930E 06	251.1	0.171	0.192	10.53	12167.7	1155.7
FR149- 3	0.8930E 06	251.9	0.192	0.218	10.66	12340.0	1157.3
FR149- 4	0.8930E 06	251.5	0.218	0.248	11.32	12508.2	1105.4
FR149- 5	0.8930E 06	251.6	0.248	0.276	10.60	12286.7	1158.8
FR149- 6	0.8930E 06	251.7	0.276	0.303	9.92	12612.6	1271.5
FR149- 7	0.8930E 06	251.9	0.303	0.329	9.80	12388.6	1264.5
FR149- 8	0.8930E 06	252.1	0.329	0.362	7.40	12506.0	1689.7

Table C-63 (Continued)

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR149- 9	0.8930E 06	250.8	0.362	0.391	10.85	12240.9	1127.7
FR149-10	0.8930E 06	251.9	0.391	0.411	8.71	12174.2	1397.3
FR149-AVER	0.8930E 06	251.5	0.151	0.411	10.28	12355.0	1202.3
FR155- 1	0.8948E 06	254.8	0.102	0.121	9.49	9402.0	990.8
FR155- 2	0.8948E 06	255.1	0.121	0.140	8.26	8931.8	1081.2
FR155- 3	0.8948E 06	255.2	0.140	0.162	8.65	9001.6	1040.2
FR155- 4	0.8948E 06	254.9	0.162	0.184	9.32	9260.0	993.7
FR155- 5	0.8948E 06	255.0	0.184	0.207	8.72	9125.2	1047.0
FR155- 6	0.8948E 06	254.8	0.207	0.229	8.12	9332.4	1149.0
FR155- 7	0.8948E 06	254.8	0.229	0.249	7.95	9154.6	1151.7
FR155- 8	0.8948E 06	255.1	0.249	0.275	7.40	9447.3	1276.1
FR155- 9	0.8948E 06	253.8	0.275	0.297	8.92	9141.0	1024.5
FR155-10	0.8948E 06	255.0	0.297	0.308	6.64	9224.1	1388.4
FR155-AVER	0.8948E 06	254.9	0.102	0.308	8.38	9202.0	1098.4
FR161- 6	0.8518E 06	246.8	-.001	0.013	6.50	6120.0	942.0
FR161- 7	0.8518E 06	247.0	0.013	0.025	6.18	6143.3	994.3
FR161- 8	0.8518E 06	247.2	0.025	0.046	5.86	6117.0	1044.4
FR161- 9	0.8518E 06	245.7	0.046	0.060	7.17	5932.1	827.9
FR161-10	0.8518E 06	247.1	0.060	0.060	5.09	6013.9	1181.3

Table C-64

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 1.70 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
109- 2	0.1813E 07	249.9	-.004	0.011	10.51	15525.4	1477.1
109- 3	0.1813E 07	250.6	0.011	0.028	11.21	15911.8	1419.0
109- 4	0.1813E 07	250.0	0.028	0.047	11.30	15623.3	1323.7
109- 5	0.1813E 07	250.4	0.047	0.064	10.53	16047.8	1523.8
109- 6	0.1813E 07	250.2	0.064	0.080	10.14	16134.7	1591.5
109- 7	0.1813E 07	250.6	0.080	0.096	10.69	15785.3	1476.4
109- 8	0.1813E 07	250.6	0.096	0.118	11.15	15805.4	1416.9
109- 9	0.1813E 07	249.7	0.118	0.136	10.41	15909.2	1528.1
109-10	0.1813E 07	250.5	0.136	0.147	8.61	15817.7	1838.0
109-AVER	0.1813E 07	250.3	-.026	0.147	10.68	15842.2	1483.4
FR 69- 4	0.1779E 07	246.5	0.002	0.021	10.71	16119.8	1505.6
FR 69- 5	0.1779E 07	246.9	0.021	0.038	10.53	16029.7	1522.3
FR 69- 6	0.1779E 07	246.7	0.038	0.055	9.51	16127.9	1695.1
FR 69- 7	0.1779E 07	246.9	0.055	0.072	10.21	16082.6	1575.7
FR 69- 8	0.1779E 07	246.9	0.072	0.094	11.12	15899.4	1429.2
FR 69- 9	0.1779E 07	246.0	0.094	0.113	10.27	15951.6	1552.6
FR 69-10	0.1779E 07	246.6	0.113	0.124	8.77	15966.6	1821.5
FR 69-AVER	0.1779E 07	246.6	-.041	0.124	10.45	15922.6	1523.0
FR 72- 5	0.1805E 07	243.8	0.003	0.015	5.97	11092.3	1857.1
FR 72- 6	0.1805E 07	244.0	0.015	0.028	6.15	10876.1	1768.1
FR 72- 7	0.1805E 07	243.5	0.028	0.037	6.54	10809.6	1651.9
FR 72- 8	0.1805E 07	244.5	0.037	0.050	6.01	10804.1	1797.2
FR 72- 9	0.1805E 07	243.3	0.050	0.063	4.33	11015.1	2543.8
FR 72-10	0.1805E 07	244.0	0.063	0.068	4.42	11372.2	2570.7
FR 72-AVER	0.1805E 07	243.9	-.035	0.068	4.56	10986.3	2411.6
FR 73- 1	0.1859E 07	236.8	0.044	0.051	9.99	7815.3	782.2
FR 73- 2	0.1859E 07	236.8	0.051	0.058	9.36	7610.7	812.8
FR 73- 3	0.1859E 07	237.0	0.058	0.067	6.96	7710.5	1107.9
FR 73- 4	0.1859E 07	236.4	0.067	0.077	5.40	7683.8	1423.3
FR 73- 5	0.1859E 07	236.6	0.077	0.084	8.34	7518.1	901.2
FR 73- 6	0.1859E 07	236.4	0.084	0.091	7.08	7553.5	1067.5
FR 73- 7	0.1859E 07	236.6	0.091	0.098	6.90	7713.9	1117.7
FR 73- 8	0.1859E 07	236.5	0.098	0.111	7.25	7838.4	1081.4
FR 73- 9	0.1859E 07	235.4	0.111	0.119	7.54	7568.8	1003.9
FR 73-10	0.1859E 07	236.3	0.119	0.120	6.09	7829.7	1285.6
FR 73-AVER	0.1859E 07	236.5	0.044	0.120	7.55	7684.3	1018.0
FR 78- 1	0.1686E 07	248.7	0.142	0.147	9.24	4687.8	507.3
FR 78- 2	0.1686E 07	248.7	0.147	0.150	8.78	4486.3	511.0
FR 78- 3	0.1686E 07	249.3	0.150	0.151	8.47	4455.9	525.8
FR 78- 4	0.1686E 07	249.5	0.151	0.153	8.27	4474.5	541.3
FR 78- 5	0.1686E 07	250.1	0.153	0.156	7.76	4592.2	591.7
FR 78- 6	0.1686E 07	250.1	0.156	0.159	6.69	4707.8	703.9
FR 78- 7	0.1686E 07	250.6	0.159	0.161	5.98	4500.4	752.4
FR 78- 8	0.1686E 07	250.9	0.161	0.172	5.95	4501.1	756.3

Table C-65

HEAT TRANSFER - EXPERIMENTAL DATA,
$G \approx 3.40 \times 10^6$ lbm/hr-sq ft, $T \approx 250°F$

RUN #	G (LBM/HR-SQ FT)	T BOIL (DEG F)	X IN	X OUT	DELTA T (DEG F)	HEAT FLUX (BTU/HR-SQ FT)	H EXP (BTU/HR-SQ FT-DEG F)
FR151- 1	C.3100E 07	243.2	0.112	0.118	17.98	12338.1	686.4
FR151- 2	0.3100E 07	243.4	0.118	0.119	17.91	12167.7	683.3
FR151- 3	0.3100E 07	244.6	0.119	0.120	17.99	12340.0	685.9
FR151- 4	C.3100E 07	245.1	0.120	0.123	17.66	12499.2	707.7
FR151- 5	0.3100E 07	245.5	0.123	0.124	15.55	12295.8	790.7
FR151- 8	0.3100E 07	248.4	0.123	0.129	11.12	12516.3	1125.5
FR151- 9	0.3100E 07	247.9	0.129	0.133	11.82	12252.6	1036.6
FR151-AVER	0.3100E 07	246.2	0.112	0.127	14.66	12360.4	843.0
FR157- 1	0.3339E 07	242.4	0.140	0.142	13.59	9440.6	694.7
FR157- 2	0.3339E 07	242.9	0.142	0.143	13.22	8957.3	677.5
FR157- 3	0.3339E 07	243.4	0.143	0.144	13.18	9043.2	686.4
FR157- 4	0.3339E 07	243.8	0.144	0.146	13.00	9287.2	714.3
FR157- 5	0.3339E 07	244.2	0.146	0.147	11.71	9125.2	779.0
FR157- 6	0.3339E 07	244.7	0.147	0.148	10.47	9351.0	893.0
FR157- 8	0.3339E 07	246.0	0.148	0.153	8.49	9457.6	1114.3
FR157- 9	0.3339E 07	245.2	0.153	0.155	9.71	9141.0	941.5
FR157-AVER	0.3339E 07	244.5	0.140	0.146	11.06	9221.6	833.8
FR163- 7	0.3356E 07	244.3	-.000	0.003	6.27	6143.3	979.6
FR163- 8	0.3356E 07	244.1	0.003	0.014	6.01	6117.0	1017.0
FR163- 9	0.3356E 07	242.7	0.014	0.019	7.35	5932.1	807.5
FR163-AVER	0.3356E 07	243.9	-.010	0.014	7.21	6024.1	835.3

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