MONTANA STATE UNIVERSITY

Heat transfer in the finned fluidized bed tubular heat exchanger
by Joon Taik Kim

A thesis submitted to the Graduate Faculty in partial fulfillment of the requirements for the degree of MASTER OF SCIENCE in Chemical Engineering
Montana State University
© Copyright by Joon Taik Kim (1970)

Abstract:
Local and average heat transfer coefficients for bed -to-wall heat transfer from a fluidized bed tubular heat exchanger with extended surfaces were investigated.

The fluidized bed tubular heat exchanger consisted of a 44-inch long, 5.5-inch inside diameter shell with 19, 3/4-inch diameter stainless steel tubes arranged in a 1-inch triangular pitch. Air was used as the fluidizing medium and glass spheres of 0.0185-inch average diameter were used in this study.

Tubes at the four possible tube locations tube were heated electrically. 3/16-inch stainless steel wire was used as an extended surface. The stainless steel wire was wrapped around the outside of each 19 tubes in a helical sprial. Variables studied include particle concentration, gas mass velocity and twist ratio.

The Results of This Study Are as Follows 1. Gradual increase in the weighted average heat transfer coefficients with respect to the number extended surface were observed. Local heat transfer coefficients decreased with the distance from the entrance of the fluidized bed. Tube locations had only a slight effect on local heat transfer coefficients.

2. The average sectional coefficients according to the particle mode heat transfer with the extended surface were higher than the coefficients calculated from the bare tube surface.

3. With local heat transfer coefficients and experimentally determined sectional particle fractions over the 11 different sections of the fluidized bed, Nusselt numbers were correlated with an equation based on a particle mode heat transfer mechanism.

4. It is concluded that particle Nusselt numbers are proportional to 0.48 of I-ε.

Statement of Permission to Copy

In presenting this thesis in partial fulfillment of the requirements for an advanced degree at Montana State University, I agree that the Library shall make it freely available for inspection. I further agree that permission for extensive copying of this thesis for scholarly purposes may be granted by my major professor, or, in his absence, by the Director of Libraries. It is understood thay any copying or publication of this thesis for financial gain shall not be allowed without my written permission.

Signature Joon Taik Kim

Date Feb. 27, 1970

HEAT TRANSFER IN THE FINNED FLUIDIZED BED
TUBULAR HEAT EXCHANGER

By

JOON T. KIM

A thesis submitted to the Graduate Faculty in partial fulfillment of the requirements for the degree

of

MASTER OF SCIENCE

in

Chemical Engineering

Approved:

Lloyd Berg
Head, Major Department

William E. Genetti
Chairman, Examining Committee

K Goering
Graduate Dean

MONTANA STATE UNIVERSITY
Bozeman, Montana

March, 1970

[redacted]
N378
K562
Cop. 2

VITA

Joon Taik Kim, the author, was born March 17, 1945, in Seoul, Korea. He is the son of Mr. Chang Hwa Kim and Mrs. Young Ja Chun. He received his elementary and secondary education at Seoul, Korea. He graduated from Pai Chai High School in February, 1963. In March, 1963, he enrolled at Yonsei University in Seoul, Korea, and graduated February, 1967 with a B. S. degree in Chemical Engineering. From January, 1967 to September, 1968, he worked with the Medium Industry Bank as a technical services engineer and conducted various feasibility studies on AID and OECF loan granted industries specializing in gas conversion.

He enrolled in Chemical Engineering Department at Montana State University as a research assistant in September 1968 working toward Master of Science degree.

ACKNOWLEDGEMENTS

I would like to thank Dr. William E. Genetti of Montana State University for his assistance and encouragement throughout the duration of the investigation. I wish to express my gratitude to Mr. Ronald D. Aafedt for his assistance which helped me to complete my study. I would like to thank Miss So Young Hahn for her heartful help to finish my study and research. Special thanks go to my parents for their encouragement throughout the period of research.

I wish to acknowledge the financial support I received from the Engineering Experiment Station at Montana State University.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

ABSTRACT

Local and average heat transfer coefficients for bed-to-wall heat transfer from a fluidized bed tubular heat exchanger with extended surfaces were investigated.

The fluidized bed tubular heat exchanger consisted of a 44-inch long, 5.5-inch inside diameter shell with 19, 3/4-inch diameter stainless steel tubes arranged in a 1-inch triangular pitch. Air was used as the fluidizing medium and glass spheres of 0.0185-inch average diameter were used in this study.

Tubes at the four possible tube locations tube were heated electrically. 3/16-inch stainless steel wire was used as an extended surface. The stainless steel wire was wrapped around the outside of each 19 tubes in a helical sprial. Variables studied include particle concentration, gas mass velocity and twist ratio.

The Results of This Study Are as Follows

1. Gradual increase in the weighted average heat transfer coefficients with respect to the number extended surface were observed. Local heat transfer coefficients decreased with the distance from the entrance of the fluidized bed. Tube locations had only a slight effect on local heat transfer coefficients.

2. The average sectional coefficients according to the particle mode heat transfer with the extended surface were higher than the coefficients calculated from the bare tube surface.

3. With local heat transfer coefficients and experimentally determined sectional particle fractions over the 11 different sections of the fluidized bed, Nusselt numbers were correlated with an equation based on a particle mode heat transfer mechanism.

4. It is concluded that particle Nusselt numbers are proportional to 0.48 of $1-\varepsilon$.

Introduction

The study of design methods for improved fluidized bed heat exchangers as industrial heat transfer equipment has become greatly important, because of its isothermal characteristic along with various applications in chemical and physical operations.

Fluidization is the phenomenon in which the gravitational force acting on a dense packet of particles is counteracted by an upward fluid stream, which causes these particles to be kept more or less in a floating state. The uniform temperature distribution of the fluidized bed is due to the turbulent fluid-solids suspended motion and the continuous, frequent contacts between the heating surface and a new swarm of particles by the rapid circulation of the fluidized stream.

Even if the fundamentals of fluidization were not fully and completely understood, many fluidized-solids processes have achieved commercial success. Fluid-bed catalytic cracking units with the conversion level of 50-60 percent is an example. The need for heat, mass, and momentum transfer correlations in fluidized beds was paramount for fluidized bed techniques to be a success.

One of the noticeable advantages of fluidized-beds to one-phase flow is high rates of heat transfer. It is well-known that higher heat transfer coefficients can be achieved with extended surfaces, however, this effect is sometimes dismissed as a means to improve performance of aheat exchanger because of the correspondingly higher power consumption.

Numerous studies have been made to find out the effect of various types of extended surfaces on the rate of heat transfer in turbulent flow for different industrial heat transfer equipment. These results are in no way general, but it is at least evident that extended surfaces like spiral wires or twisted strips have some effects on increasing the rate of heat transfer in industrial heat transfer equipment for single phase systems.

A number of studies on local and average heat transfer coefficients for heat transfer from an internal tube in a bundle of tubes to a fluidized-bed have been made. The present investigation is a study of local and average heat transfer rates in the fluidized bed tubular heat exchanger with the extended surfaces. Local heat transfer coeffcicents were measured with a moving thermocouple probe inside an electrically heated tube. This probe made it possible to measure local heat transfer coefficients at any point along a tube for various gas rates and particle concentrations. The rate of heat transfer was correlated with the area of extended surfaces.

Literature Survey and Theory

Fluid Mechanical Approach of Fluidized Beds

By assuming a fully developed circular bubble and solid-free, spherical in the three-dimensional case, cylindrical in the two-dimensional case, Davidson (3) adopted a continum approach, potential flow for both the fluid and particulate phases to postulate the motion of rising bubble. Even if Davidson model was not sufficiently accurate, it was the first useful approach to the motion of a rising bubble in the gas-solid fluidized bed.

Murray (17) adopted a continum approach for the two-phase motion of the fluid and solids in a fluidized bed. Momentum and mass conservation equations were applied to obtain solutions for the motion caused by rising bubbles in the bed. The stability analysis and the equations had been derived to provide a means of classifying fluidized systems, but he did not evaluate the roots of his secular equation, so that no quantitative results were obtained.

Anderson and Jackson (1) realized that for practical purposes it is necessary to seek some method of simplifying the actual problem so that it can be described by a small system of partial differential equations. Anderson and Jackson have used a formal mathematical definition of local mean variables to translate the point Navier-Stokes equations for the fluid and the Newtonian equations of motion for the particles somewhat devoid of number of terms whose forms are yet undetermined.

According to the renewal model proposed by Mickley and Fairbanks (18), owing to the bubbles, packets of particles are renewed continuously between the bulk of the bed and the vicinity of the wall. The rate of heat transfer between a non-homogeneous fluidized bed and a wall depends on the thermal conductivity of these packets and on their renewal frequency.

It appears a quantity, the renewal frequency ω, to be determined more specifically according to a more complete representation of hydrodynamic process. If the swarm of solid particles is assimilated to a quasi-fluid it is possible to write the equation of motion for the two-phase system. A volume which is encountered here is sufficiently small to be considered as infinitely small, but also large enough to contain a sufficiently large number of particles. Ruckenstein (24) suggested that a synthesis between the renewal model and the linearized instability theory might be able to give information with respect to the renewal frequency. The experimental results of Mickley and Fairbanks (18) show that ω is in the range 1 - 10 sec^{-1} and that it does not depend too much on the fluidization velocity. For glass spheres of 10^{-4} meter diameter Ruckenstein (24) computed that the renewal frequency is about 4 sec^{-1} and predicted a weak dependence of ω on velocity.

Newly Proposed Bubble-bed Model in Heat Transfer

Several theoretical mechanisms of fluidized bed models have been proposed and some of those have been briefly summarized by Genetti (8) for heat transfer in order to provide a basis for understanding correlations which have been presented in his recent work.

Kunii and Levenspiel (11, 12) have proposed a model for the flow of gas through a fluidized bed, the bubble-bed model, or a three-region model which have viewed as unifromly sized bubbles surrounded by clouds and followed by wakes. A sketch of this model is given in Fig. 1. They have shown that this model is good enough to fit the reported data for gas-solid heat transfer, gas-solid mass transfer and conversion in catalytic reactions.

Kunii and Levenspiel have only been concerned with the dense bubbling region. This region applies where a continuous exchange or floowing up of solid particles along the heat transfer surface is continuing. This concept has been sometimes explained on the basis of the "packet" theory (9).

For the bubble phase, Kunii and Levenspiel have started with several necessary assumptions to simplify the problem from the Davidson model (4). They have taken the bubble size to be uniform throughout the bed or section of bed and called it the effective bubble size. The velocity of rise of a crowd of bubbles has been related to the velocity of rise of

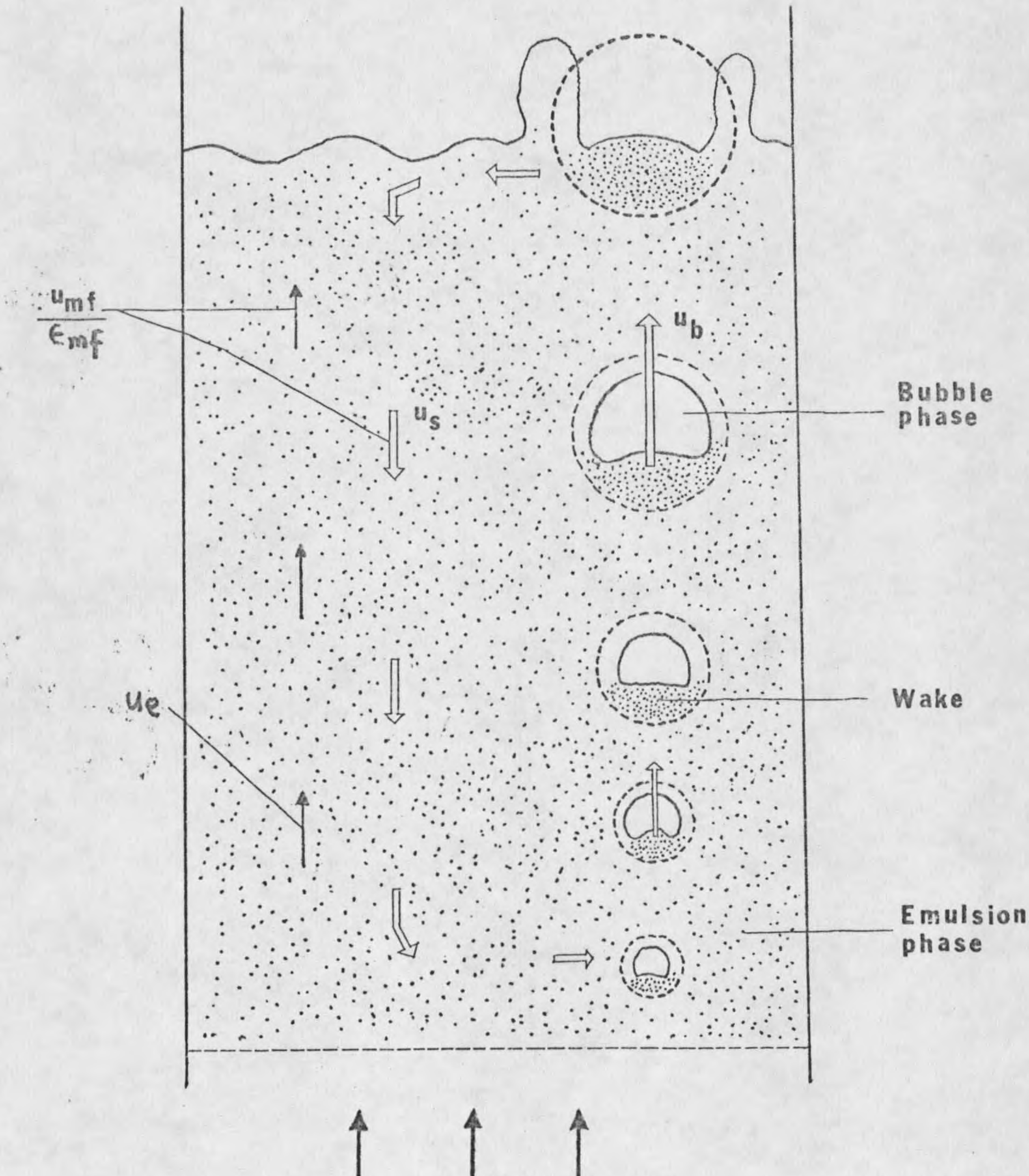

Figure 1. Solid movement and gas flow as visualized by bubbling bed model.

a single bubble by

$$u_b = u_o - u_{mf} + u_{br} = u_o - u_{mf} + 0.711(gd_b)^{\frac{1}{2}} \quad (1)$$

where $u_{br} = 0.711(d_b)^{\frac{1}{2}}$.

Ignoring small amounts of solids inside the rising bubble (measured to be from 0.2% to 1% by different investigators), they have taken the bubble void fraction $\varepsilon_b = 1$. The average bed voidage ε_f has been related to the voidage in bubbles and emulsion phase by

$$\varepsilon_f = \delta\varepsilon_b + (1-\delta)\varepsilon_e = \delta + (1-\delta)\varepsilon_e \quad (2)$$

where δ = the volume fraction of bubbles in the bed.

By assuming $\varepsilon_e = \varepsilon_{mf}$, voidages and bed heights have been correlated by

$$1 - \delta = \frac{L_{mf}}{L_f} = \frac{1 - \varepsilon_f}{1-\varepsilon_{mf}} \text{ and } \frac{L_{mf}}{L_m} = \frac{1 - \varepsilon_m}{1 - \varepsilon_{mf}} \quad (3)$$

The bubble phase variables d_b, u_b and δ have been correlated to give u_o and u_{mf}, i.e.

$$u_b = \frac{u_o - (1-\delta)u_{mf}}{\delta} \cong \frac{u_o - u_{mf}}{\delta}$$

For the emulsion phase, Kunii and Levenspiel (11,12) have assumed the void fraction of the wake to be that of the emulsion phase and have correlated the relative velocity of upward percolating gas, u_e, and of downward flowing solid,

u_s, to the superficial gas velocity of minimum fluidization, u_{mf};

$$u_e = u_f - u_s = \frac{u_{mf}}{\varepsilon_{mf}} - u_s . \qquad (5a)$$

The downward velocity of solid is given by

$$u_s = \frac{\alpha\delta\ u_b}{1-\delta-\alpha\delta}$$

$$\text{where } \alpha = \frac{\text{volume of wake dragged up to the bed behind a rising bubble}}{\text{(volume of bubble)}} .$$

Variables d_b, u_b, δ, u_e, and u_s have been correlated to give u_o and u_{mf}, i.e.

$$u_b = \frac{1}{\delta} [u_o - (1 - \delta - \alpha\delta)u_{mf}]. \qquad (6)$$

Equation 6 has been approximated to give the expression for the whole range of flows, and the result is identical to Equation 4;

$$u_b \cong \frac{u_o - (1-\delta)u_{mf}}{\delta} = \frac{u_o - u_{mf}}{\delta} . \qquad (7)$$

Bed-to-Wall Heat Transfer in Gas-Solid Fluidized Beds

Much work has been done on bed-to-wall heat transfer for numerous heating surface arrangements. Empirical equations have been developed to predict heat transfer coefficients for heat transfer from these surfaces.

Frantz (7) has summarized work done on surface-to-bed heat transfer in fluidized beds. Genetti (8) has discussed numerous proposed correlations and several heat transfer mechanisms for fluidized bed heat transfer.

Usually fluidized systems have low enough absolute temperatures so that thermal radiation may be safely discounted as a significant contributing factor. However, in many metallurgical applications of the fluidized bed techniques the processing temperatures are sufficiently high so that thermal radiation may be an important part of the over-all heat transfer process. Szekeley and Fisher (25) recently have studied bed-to-wall radiation heat transfer on the basis of the study of bed-to-wall heat transfer in the convective regime done by Botterill and et al (2).

While none of the proposed mechanisms has been sufficient to predict heat transfer coefficients for many different situations encountered, various mechanisms explaining the high rates of heat transfer between exchanger walls and fluidized beds have been suggested, and can be classified as follows:

(1) van Heerden et al (10) and Wicke and Fetting (32): the fluidizing medium (gas) acted as a stirring agent and the solids transported most of the heat, and the fluidized bed was visualized as well stirred liquid. Therefore, there is the steady state conduction through the emulsion phase.

(2) Leva et al (15), Dow and Jakob (5) and Levenspiel and Walton (16): the scouring action of the solids along the heat exchanger wall reduces the thickness of the laminar gas film and increases the rate of heat transfer.

(3) Mickley and Fairbanks (18): the fluidized bed was assumed to be composed of "packets" which renewed intermittently by the violent disturbance in the core portion of the fluidized bed, and where unsteady-state diffusion of heat to newly arrived mobile elements occurred.

(4) Ziegler et al (34) and Botterill and Williams (2): unsteady-state conduction by single particles in direct contact with heat exchanger walls. This model has been further modified by Genetti (8) to give more complete predictions with respect to experimental data.

This modified model is adopted in the present work. The results of this model predict two established facts. The Nusselt number is independent of thermal conductivity of the solids (19,34) and is proportional to about the square root of the particle fraction, $1 - \varepsilon$ (20). The results of the model also predict a maximum Nusselt number, which also is an established fact (15).

1. A Rather Different Explanation According to the Bubble Bed Model

Yoshida, Kunii, and Levenspiel (33) have tried to unify the mechanisms proposed by Mickley and Fairbanks (18), and van heerden et al (10) and Wicke and Fetting (32), and have applied the bubble model to postulate bed-to-wall heat transfer. However, the work is not complete, and some of the quantitative parameters are still not well-known.

They have started the prediction of the basis of the film-penetration theory for gas absorption into liquids proposed by Toor and Marchello (26). The equation which represents this film-penetration model for mass and heat transfer[1] is

$$\frac{\partial T}{\partial T} = \alpha \frac{\partial^2 T}{\partial^2 T} \qquad 0 \leq x \leq l_e$$

where $$\alpha = \frac{k_e}{\rho_e C_{ps}} = \frac{k_e}{\rho_s C_{ps}(1-\varepsilon_{mf})}$$

l_e = effective thickness of emulsion layer.

---

[1] A thin layer of emulsion of thickness l_e suddenly contacts the exchanger wall and after a short time it is suddenly moved away to replace a fresh element of emulsion from the core portion of the bed. Therefore it will include the steady state conduction of heat through an emulsion layer at the wall and the unsteady state absorption of heat by emulsion elements.

And boundary conditions are

$$T = T_b \text{ at } t = 0$$

$$T = T_w \text{ at } x = 0$$

$$T = T_b \text{ at } x = l_e$$

The solution of the above equation is

$$\frac{T_w - T}{T_w - T_b} = 2 \sum_{n=1}^{\infty} (-1)^{n+1} \frac{\sin n\pi(1 - x/l_e)}{n\pi} \exp\left(-n^2\pi^2 \frac{a\ t}{l_e^{\ 2}}\right) \tag{9}$$

From the above solution, the instantaneous local heat transfer coefficient is found to be

$$h_{ti} = \left[\frac{k_e \rho_e C_{ps}}{\pi t}\right]^{1/2} \left[1 + 2 \sum_{n=1}^{\infty} \exp\left(-\frac{n^2\ l_e^2}{a t}\right)\right]$$

$$= \frac{k_e}{l_e} \left[1 + 2 \sum_{n=1}^{\infty} \exp\left[n^2\pi^2 \frac{a\ t}{l_e}\right]\right] \tag{10}$$

The observed coefficient of heat transfer h_w is the time averaged value of the instantaneous coefficient, and that is given by

$$h_t = \int_o^{\infty} h_{ti}\ I(t)dt \tag{11}$$

where $I(t)$ = the age distribution function of emulsion elements on the surface.

Two types of age distribution functions have been considered: Case 1. random surface renewal, and Case 2. uniform surface renewal. A number of expressions for the observed heat transfer coefficients, h_w, have been derived according to the different criterions.

The bubbling bed model gives a simple representation of bubble flow, the emulsion flow and the interaction of these streams. However, two empirical expressions which have been reported by Wen and Leva (31) and Wender and Cooper (30). are usually recommended for the design of the bed-to-wall heat transfer.

2. <u>Some Generalized Design Equations for Wall-to-Bed Heat Transfer</u>

Bed-to-exterior wall heat transfer:

Wen and Leva (31) attempted to correlate some earlier observations of solids motion in the bed with heat transfer phenomena through a boundary layer. They considered that the chief resistance to heat exchange between a containing wall and a fluidized bed is in the laminar film at the vessel boundary. The thickness of the film is influenced by the velocity of particles along the wall parimarily due to the "scouring action" of the particles. By the usual method of plotting and cross plotting, the following relation was obtained:

$$\frac{hd_p}{k} = 0.16 \left(\frac{C_s \rho_s d_p^{1.5} g^{0.5}}{k}\right)^{0.4} \left(\frac{G_f k_p \eta}{\mu \; R}\right)^{0.36} \quad (12)$$

where η = the fluidization efficieny

R = the bed expansion ratio.

Nomographic solutions for evaluation of η and R have been given in Appendix II. Wender and Cooper (30) made a curve of the dimensionless group Y/F with the paricle Reynolds number as shown in Figure 2 with five data sets included. The term Y and F are defined as:

$$Y = \left(\frac{h \; D_p}{k[1-\varepsilon]}\right) \left(\frac{C_g \; \rho_g}{C_s \; \rho_s}\right) \quad (13)$$

$$F = 1 + 7.5e^{-0.44} \left(\frac{L_h}{D_t}\right) \left(\frac{C_g}{C_s}\right) \quad (14)$$

Bed-to-interior wall heat transfer:

Internal, vertical-tube heat transfer was found to be independent of tube length, particle shape and particle size distribution. Wender and Cooper (30) made a straight line plot of the dimensional Z/C_R group with the particle Reynolds number as shown in Figure 3 with six data sets included for the internal vertical heat transfer surface. The term Z is defined as:

$$Z = \frac{(hd_p/k_g)(k_g/C_g\rho_g)^{0.43}}{(1-\varepsilon)(C_s/C_g)^{0.80}\ (\rho_s/\rho_g)^{0.66}} \tag{15}$$

The term C_R allows for non-axial tube location and may be evaluated from Figure 4 which originally was given by Vreedenberg (28). The equation of this correlation is:

$$\left(\frac{hd_p}{k[1-\varepsilon]}\right)\left(\frac{k}{C_g\rho_g}\right)^{0.43} = 0.33C_R\left(\frac{d_pG}{\mu}\right)^{0.23}\left(\frac{C_s}{C_g}\right)^{0.80}\left(\frac{\rho_s}{\rho_g}\right) \tag{16}$$

Noë and Knudsen (22) have made a study of heat transfer from a vertical tube located in a bundle of vertical tubes in a fluidized bed. Their data was compared to the correlation of Wender and Cooper (30) using a value of $C_R = 2$. Almost all their experimental data was within $\pm$ 50 percent of the correlation. This is fair agreement considering the greatly different geometries.

Immersed, horizontal-tube heat transfer has been studied by Vreedenberg (29) along with the effect of tube diameter, particle size, shape, and density, and gas velocity on the heat transfer coefficient, including data on large-scale beds. The recommended correlations are:

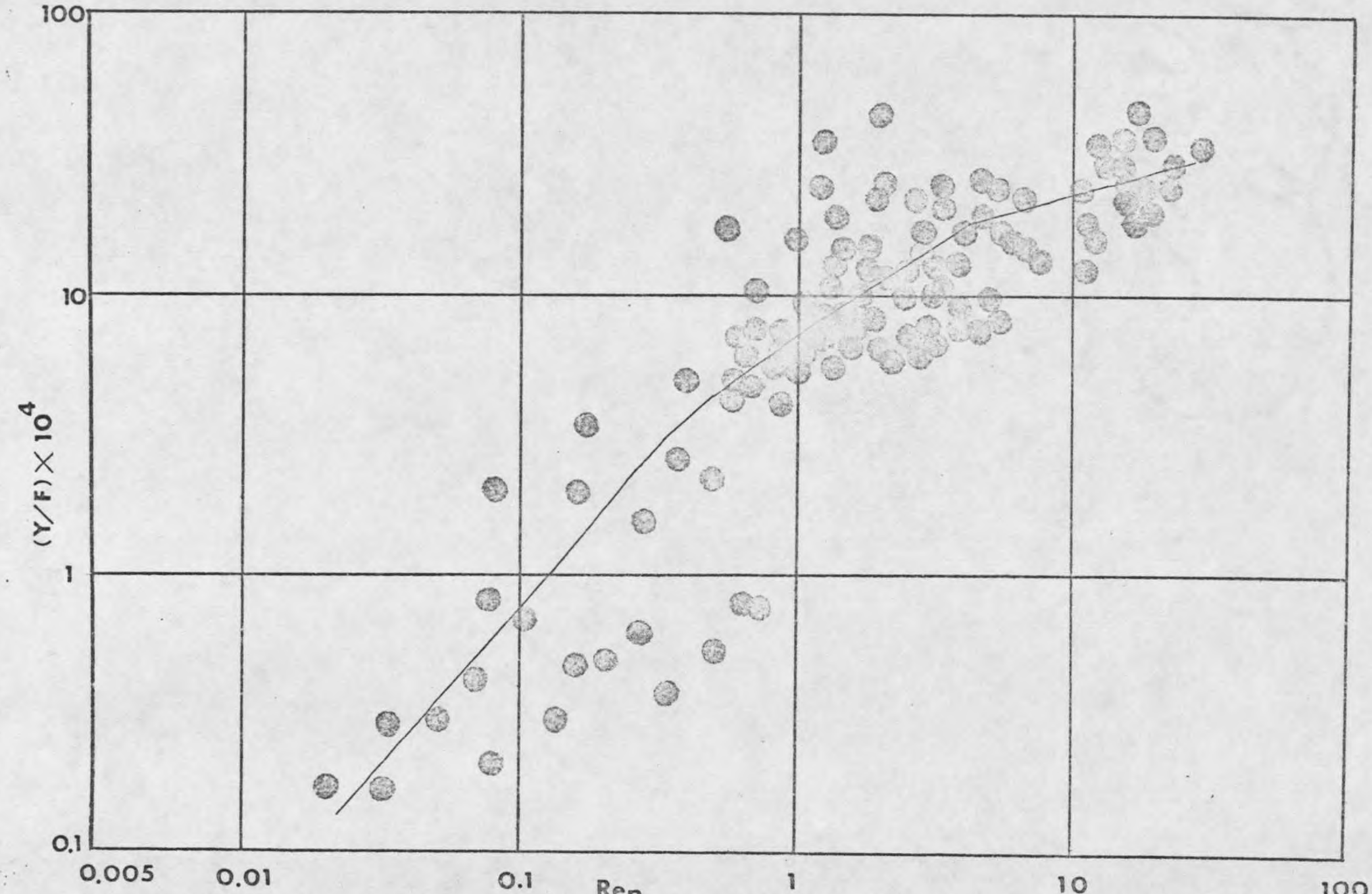

Figure 2. Correlation for heat transfer between containing wall and fluidized bed.

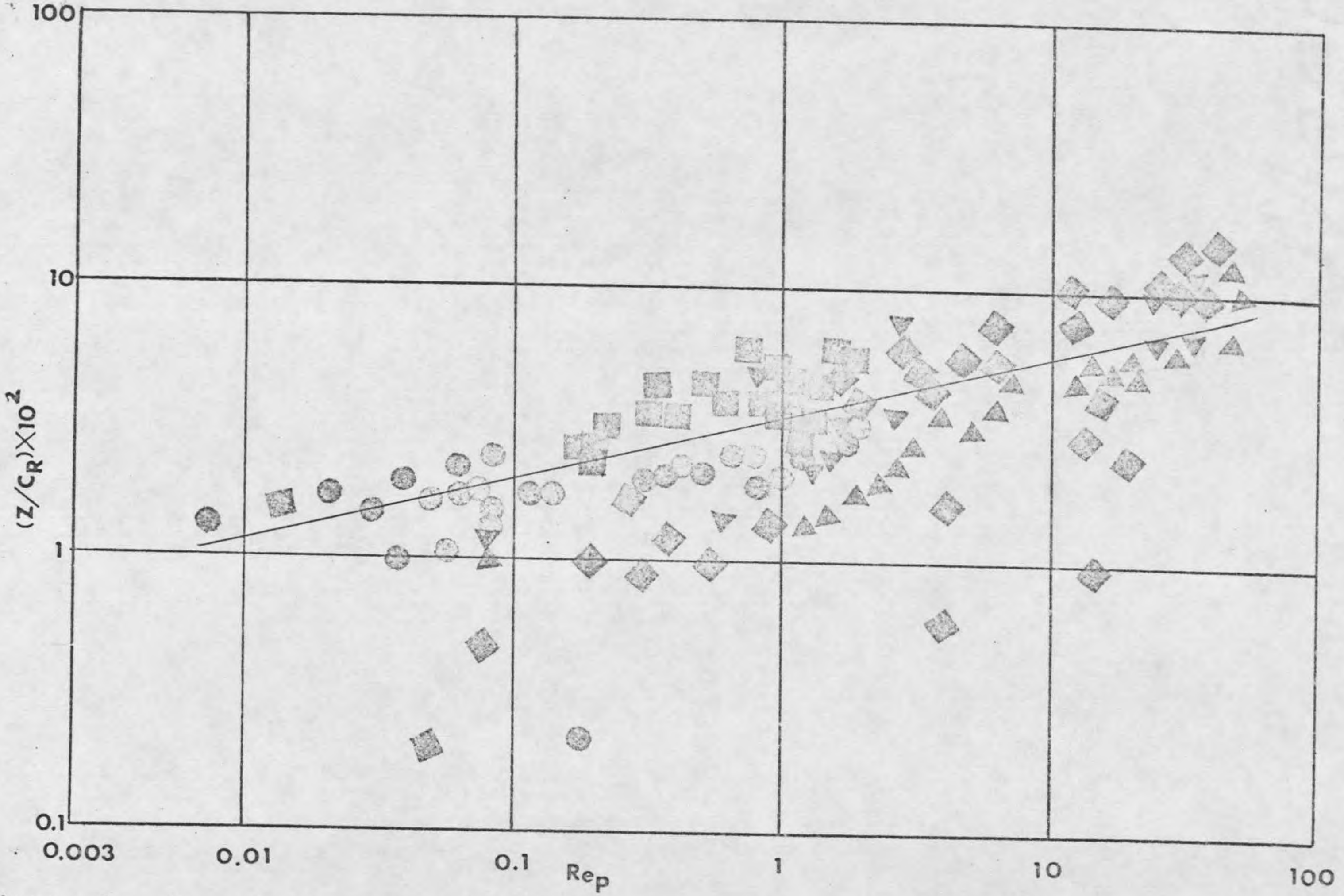

Figure 3. Correlation for heat transfer between internal tubes and bed.

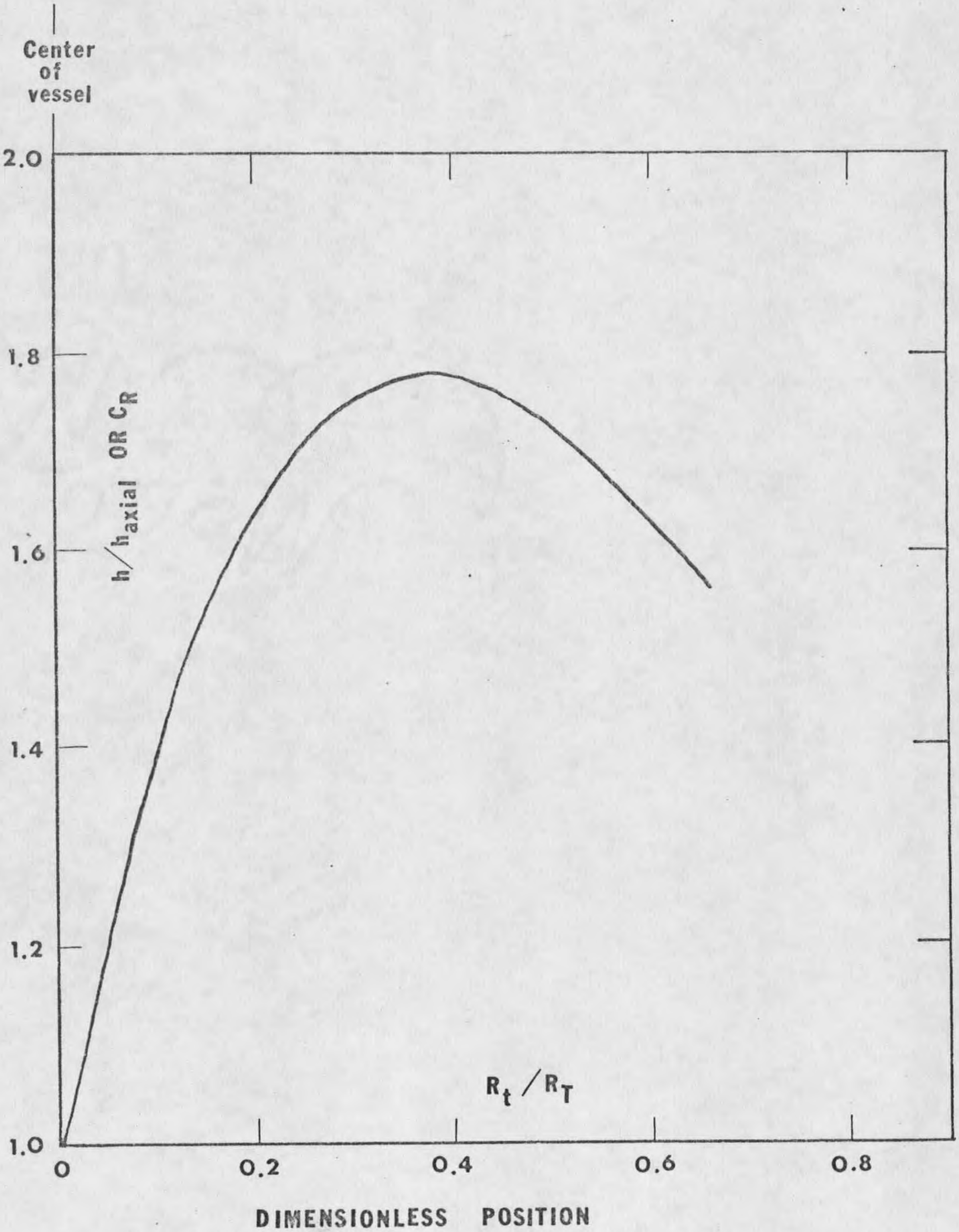

Figure 4

Correction factor C_R for nonaxial location of immersed tubes.