HEAT TRANSFER IN FLUIDIZED BEDS

by

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AGENCY

Synopsis

This report contains:

(a) A summary of the state of the art for heat transfer to surfaces in contact with fluidized beds, including a compilation of published investigations; with a listing and discussion of the various theoretical, empirical, and semi-empirical expressions for predicting heat transfer coefficients in fluidized beds.

(b) A summary of the study carried out here, with a discussion relating these investigations to the general picture proposed above.

(c) A general discussion of the important factors to be considered in the design of fluidized-bed heat transfer units with recommendations for future research.

I. THE STATE OF THE ART

Appendix IIIA presents a compilation, in tabular form, of the major experimental investigations carried out in the field of heat transfer to surfaces in contact with fluidized beds. It has been found helpful to separate them into two groups; heat transfer in the dense phase region i.e., up to the point where there is a net movement of the solids bed relative to the walls of the confining vessel (low bed voidage of the order of magnitude 50-70 percent); heat transfer in the dilute phase region (bed voidage usually in excess of 90 percent). Although many investigations overlap into both regions, the various studies have been divided according to the region containing the majority of the data points. The reason for the division is that no one particular dense-phase heat-transfer correlation can be extended into the dilute zone, and vice versa. The first group has been further divided into heat transfer to the confining wall, and heat transfer to bodies immersed in the fluidized bed.

From the extensive amount of experimental data that have been collected by various investigators, a number of theoretical, empirical and semi-empiric correlations have been advanced. These are listed in Appendix IIIB. Each co relation is valid only within the limits of the experimental data used by the author.

Appendix IIIC lists the two generalized empirical correlations for heat transfer in fluidized beds, the Wen-Leva (38) and the Wender-Cooper (39) correlations. These two correlations are held to be the most useful in predictir heat transfer coefficients at surfaces in fluidized beds, and cover a wide range of conditions. Zenz (40), Kunii and Levenspiel (41), and Zabrodsky (42) all cite these generalized correlations in their books; Zabrodsky also considers the correlation of Vreedenburg (28) for horizontal tubes to be generall applicable.

Appendix IIID shows by means of a power function equation, the predicted effect of the many parameters which could affect the heat transfer in fluidize beds. The values in the columns under each parameter are the exponents in a standard power function equation of the type:

$$h \propto d_p^{a_1} \rho_g^{a_2} C_s^{a_3} C_g^{a_4} \text{ etc.}$$

where a_1, a_2, a_3, a_4 etc., are the exponents listed, and the power equation is derived from the empirical or theoretical correlation of the investigator. Th value of such an analysis is questionable, since many of the parameters are interrelated, but it has been suggested (Mickley (21)) that the effect of the thermal properties of the bed k_g, k_s, C_g, C_s on heat transfer can be separated from the effect of bed dynamics, which includes those properties related to th state of fluidization of the bed e.g., d_p, ρ_g, u etc. Bearing in mind the interrelationships of many of the parameters, it is possible to discuss the general effect of significant properties on the heat transfer process.

Effect of Variables (Empirical Models)

(a) Particle Diameter d_p

The majority of investigations show an inverse relationship between the heat transfer coefficient h and the particle diameter d_p, the most notable exception being the correlation of Leva (6). The latter investigator, however, included a fluidization efficiency factor of which makes allowance for the inverse effect of the particle diameter d_p on h. There is a wide variation in the precise dependence of h on d_p , as can be seen from Appendix IIID e.g.,

Dow (5) $h \alpha d_p^{-0.23}$

Miller (24) $h \alpha d_p^{-0.96}$

(b) Medium Thermal Conductivity k_g

All investigators showed a direct proportionality between h and k_g , ranging as follows:

Mickley (22) $h \alpha k_g^{0.33}$

Leva (6) $h \alpha k^{1.0}$

This variation in exponents is discussed later when the proposed mechanisms of heat transfer in fluidized beds are considered.

(c) Solids Heat Capacity C_s

Most investigators showed a direct proportionality between h and C_s

Dow (5) $h \alpha C_s^{0.25}$

Wender (39) $h \alpha C_s^{0.8}$

(d) Other Solid and Gas Properties

Because of the wide disparities between the various investigators as to the precise effects of properties such as bed-porosity ϵ_f; bed geometry D_t and H_t; superficial gas velocity u; gas density and viscosity ρ_g and μ; caused by the complex nature of the fluidized state of the bed, a more detailed discussion will have to await developments in the understanding of the physical nature of the complex behaviour of fluidized beds. However, for engineerir design purposes, Mickley (21) suggested the use of a stirring factor S whic accounts for bed motion and geometry to include the effects of those proper which modify bed dynamics. The present study attempted to relate these prc ties with a simplified model of gas bubble behaviour in fluidized beds.

(e) Gas Velocity u

A qualitative understanding of the overall picture of heat trans in fluidized systems can be obtained by considering the way in which the he transfer coefficient h varies with gas velocity. Up to the point of initia fluidization, the value of h is essentially the same as for heat transfer t packed bed. At the point of minimum fluidization, gas velocity u_{mf}, the he transfer coefficient increases abruptly and continues to increase with gas velocity until a maximum value h_{max} is reached. At higher gas velocities, heat transfer coefficient decreases slowly as the bed becomes "diluted" of solid particles. This general trend is borne out by all the investigations covered. Baerg (13), Kharchenko (20) and Varygin (27) have attempted to cc relate their data to predict the value of h_{max}, the maximum heat transfer c efficient to surfaces in contact with fluidized beds. Leva (43) suggested another reason for the variation in the predictions might be that individua investigators have studied different portions of the heat transfer - veloci curve. For example, Dow (5) gives $h \propto u^{0.8}$, whereas Van Heerden (11) gives $h \propto u^{0.45}$ which suggests that the latter author's data refer to a region cl to the maximum on the h vs u curve.

Theoretical Models

Of the various models which have been presented to suggest the physic mechanism of heat transfer between fluidized beds and contacting surfaces, general classes can be distinguished:

(a) The resistance to heat transfer lies within a relatively thin region at the wall [5,6,7,25].

(b) The resistance to heat transfer lies within a relatively thick emulsion layer which is being frequently replaced by fresh emulsion from the main core of the fluidized bed [11,17,19,22,30,35,37].

Botterill (44) suggests that the thermal conductivity of the fluidizing medium is the limiting factor in the heat transfer process. Mickley (21), proposing the contacting emulsion-packet mechanism of type (b) above, found that the heat transfer coefficient should vary as the square-root of the quiescent bed conductivity, i.e., the gas conductivity raised to the one-third power since the relationship between gas conductivity and quiescent bed conductivity has been found to be approximated by $k_e \propto k_g^{2/3}$. Alternatively, if a gas film were controlling the heat transfer process, (type (a)) as proposed by Leva (6), the heat transfer coefficient should vary as the first power of the gas conductivity. As shown previously, the exponent power on the gas conductivity varies from 0.33 to 1.0 in the various correlations proposed, giving an indication of the type of mechanism operating in each case.

II. THE PRESENT STUDY

The overall purpose of the study was to investigate the heat transfer characteristics between fluidized beds and contacting surfaces in order to develop heat-transfer correlations for use in large-scale fluidized bed design. The general approach was to consider the basic mechanics of fluidized bed heat transfer in order to provide a foundation for a better understanding of the many different correlations proposed in this field. To do this, generalized semi-empirical correlations were selected from the literature representing the two major theoretical models, and were tested using available data.

As previously mentioned, the physical picture of the heat transfer process has been debated by several groups of investigators, the outcome of which has been the division of opinion into two broad categories, the extremes of which are best represented by the theories of Leva (6) and Mickley (21). Leva (6) assumes that the chief resistance to heat exchange is in the laminar gas film at the boundary of the surface in contact with the fluidized bed. Heat flow through the film is by conduction. Further he suggests that the ver tical motion of the particles along the surface considerably lessens the thermal resistance of the laminar layer, causing the high heat-transfer coefficients observed in fluidized beds. The theory depends upon an understanding o the pattern of this particle motion and the velocities of the particles. Mickley (21) assumes that the controlling mechanism may be considered to be an unsteady-state diffusion of heat into mobile elements of quiescent bed materia "emulsion packets", in contact with the surface, which are constantly being renewed by fresh emulsion from the main core of the bed.

Neither of the theories can be tested directly since quantitative values for parameters such as Leva's interparticle friction factor β, or Mickley's emulsion packet contact frequency ϕ_d are not well known. However, it was decided to represent the two theories by the generalized correlations of Wen and Leva (38) for the "thin-film" model and Wender and Cooper (39) for the "emulsion market" model. The basis for this assumption came from the following observations:

(a) The Wen-Leva (38) correlation was developed directly from the Leva (6) 'thin-film' model.

(b) The Wender-Cooper (39) correlation for heat transfer to immersed surfaces was developed from the data of Mickley (21) and when tested independently gave a close alignment to the data of Pratt and Richards (45), Fairbanks (46), and Hawthorn (47) which were the source data of Mickley's (22) theoretical correlation.

(c) The Wen-Leva (38) correlation is based largely on data pertaining to heat-transfer between fluidized beds and the confining vessel wall [5,6,10,11]. The Wender-Cooper (39) correlation used was based entirely on data from investigations with surfaces immersed in the fluidized bed [10,13,21,23,35]. Toomey (10) reported simultaneous bed-exterior wall and bed-interior calrod heat-transfer coefficients and found large differences between the coefficients at the two surfaces although at high fluid mass velocities the coefficients approach each other. It was deduced that the differences might be due to different mechanisms operating and that the Wen-Leva correlation (bed-wall) represents the 'thin-film' mechanism and the Wender-Cooper correlation (bed-internal surface) represents the 'emulsion-contact' mechanism. Further it seems likely that in a bubbling bed, surfaces immersed in the bed will be contacted frequently with rising bubbles which allow fresh emulsion packets to sweep up to the surface, while the frequency of bubbles near the wall will be much lower with solids descending along the wall surface and the development of a thin gas film layer at the wall.

In order to test the Wen-Leva and the Wender-Cooper correlations, and the theoretical models that they were chosen to represent, the data of Van Heerden (11) and Dow and Jacob (5) for heat transfer from the bed to external surfaces and the data of Fairbanks (46), Hawthorn (47) and Baerg (13) for heat transfer to surfaces immersed in the bed were employed. The comprehensive data of Fairbanks and Hawthorn were available at M.I.T. The data of Van Heerden and Baerg are generally considered to be the most systematic and representative data on heat transfer between fluidized beds and contacting surfaces [cf Leva (43), Kunii and Levenspiel (41), Zenz (40)]. Of the other data sources listed in Appendix IIIA only Dow and Jacob gave information on bed voidage at different fluidization conditions which was necessary for application to the generalized correlations. An attempt to predict bed voidage from the 'M-plot' of

Leva (43) was made, but the results did not prove very satisfactory. Thus mc of the investigations in this study were based on the five data sources mentioned above which gave sufficient details to apply to the Wen-Leva and Wendc Cooper correlations.

Appendix IV, Graphs 1 and 2, shows the results of the data-testing. As might be expected, the bed-to-external surface heat transfer data of Van Heerd (11), and Dow and Jacob (5) (which had been employed by Wen and Leva, along wi other data, to develop their correlation) aligned closest to the Wen-Leva cor relation for bed-to-external surface transfer; also, the vertical immersed su face transfer data of Fairbanks (46) (used by Wender and Cooper to help develo their relationship) and Hawthorn (47) aligned closest to the Wender-Cooper co relation for transfer to vertical immersed tubes. The data of Baerg (13), ho ever, showed a tendency under certain conditions to follow the Wen-Leva correlation whereas it would be expected to be in line with Wender-Cooper for im mersed heat transfer surfaces.

In order to explain the anomalous results of Baerg, it was thought that not only the location of the heat transfer surface (external wall or internal immersed) but the geometry and size of that surface might be a governing fact as to which correlation, and which corresponding mechanism, might be valid in any circumstance. Baerg's internally heated tube was much larger than the internal cylindrical heaters of Fairbanks and Hawthorn, and the following hypothesis was therefore proposed:

1) Heat transfer coefficients to the containing walls of the fluidized bed can be predicted by the Wen-Leva (thin-film) correlation.

2) For surfaces immersed in the fluidized bed:

(a) If the dimensions of the surface (tube diameter) are small than some characteristic dimension for the fluidized system, then the Wender Cooper (emulsion-contact) correlation should predict the heat-exchange rates

(b) If the dimensions of the surface are large compared to this characteristic dimension then the Wen-Leva (thin-film) correlation will apply.

For the characteristic dimension, the bubble diameter was chosen based on the physical picture that only when the bubbles are of sufficient size, in comparison with the size of the immersed surface, to be able to sweep packets of emulsion to the immersed surface will the conditions for the 'contacting emulsion packet' mechanism be favorable. Estimations of the bubble diameter for different fluidizing gas rates were attempted in order to deduce values of Hawthorn's emulsion-packet contact frequency, ϕ_d. Assuming that the unsteady state behaviour of the emulsion packets is caused by the passage of the low density bubbles, it was proposed that the contact frequency of the emulsion packets can be equated to the frequency of bubbles past the surface. Values of bubble frequency were determined theoretically from Davidson's (48) simplified model. It was found that these values were from two to ten times higher than the emulsion contact frequencies measured by Hawthorn, which implies that the theoretically derived values of bubble diameter over-estimated the actual bubble size. Nonetheless, it was hoped that these estimates would serve as a comparative guide for assessing the characteristic dimension proposed above.

Graphs 3 and 4 in Appendix IV show the results of Baerg (13) and Vreedenburg (28) in the form of a plot of h_{exp}/h_{calc} versus the ratio of pseudo bubble diameter to internal tube diameter d_b/d_t. It was hoped to observe a transition from the 'thin-film' mechanism to the 'emulsion-contact' mechanism as d_b/d_t increased. The general scatter of the data points do not show a clear transition; although the data of Vreedenburg indicate an agreement with the above hypothesis i.e. at a certain d_b/d_t ratio (approx. 1.5) and greater the Wender-Cooper correlation gave a better fit to the experimental results, whilst below this value, the Wen-Leva correlation gave the closer fit. Vreedenburg himself found it necessary to propose two correlations of his own to represent his experimental results, noticing a transition at a Reynolds Group No. $(\frac{d_p G}{\mu \rho})$ of 2050, evidence of a change in nature of the fluidized bed and the mechanism controlling heat transfer.

For large-scale systems, the data of Highley (18) for transfer to immersed horizontal tubes in a 3-ft. diameter bed were applied to the correlations. Appendix V shows the coefficients predicted by the correlations compared with an average observed value of the heat-transfer coefficient from individual horizontal tubes in a multiple bank of tubes. The Wender-Cooper correlation gives the closer approximation, although neither of the correlations takes account of factors such as tube position in the bundle or the change in heat transfer coefficient around the circumference of the tube.

A large part of the investigation was centred on gaining physical insi as to the internal workings of a bubbling fluidized bed by investigating pro posed mechanisms for gas-particle motion and endeavouring to build a simplif model relating fluidized bed dynamics with easily measured physical properti of the system. Davidson's (48) bubbling-bed model has already been mentione in connection with bubble diameter and bubble frequency calculations. Howev the complex nature of the gas-particle interactions in most of the experimen systems studies make it unlikely that Davidson's simplifying assumptions appli in these cases. The Davidson model, therefore, is of limited quantitative u although it serves as an order-of-magnitude analytical tool for most fluidiz situations, and is readily amenable to design work since only well-known phy ical properties of the system under study are required for its application. other physically-based model has been found which combines simplicity in dat requirements with quantitative accuracy. This scarcity of viable theoretica models for the purely mechanical behaviour of fluidized beds is the largest obstacle to useful advance in understanding their heat transfer properties.

Overall Conclusions

The lack of theoretical understanding of mechanism and of general correlation work has left open the problem of predicting heat transfer coefficients between fluidized beds and the surfaces in contact with them. The reported empirical correlations of various individual investigators are valid only within the limits of their own experiment and do not appear to be able to encompass the data of other investigators.

Two generalized correlations, Wen-Leva (38) and Wender-Cooper (39) cover more extensive ranges of conditions and data and may be expected to be extended for use in large-scale design work after modification by the effect of such factors as:

(i) Tube spacing and arrangement in multiple tube banks

(ii) Position around tube circumference

(iii) Bed diameter (changing the flow properties of the fluidized solids)

(iv) Particle size distribution

(v) Gas entry configuration (distributor)

(vi) Baffles

However, it is important to note that even these correlations ordinarily require estimates of the void fraction; and in this regard, our present predictive abilities are very poor.

It is suggested, albeit very tentatively, that the choice of correlation is based on the hypothesis outlined previously which proposes that for surfaces immersed in a fluidized bed, the heat exchange rates will depend upon the physical state of the fluidized bed, either vigorously bubbling with the surfaces often swept clean of emulsion ("emulsion-packet model") or fairly smooth fluidization with particles descending near the surface "scouring" the gas film formed there ("thin-film model"). That there might be a transition from one

mechanism to the other with change in fluidization conditions is at least th
retically possible. Furthermore, transition would be expected to depend on relative sizes of the immersed element and a characteristic dimension of the bubbling bed (e.g., bubble diameter), and the choice of design correlation would depend on the physical mechanism operating i.e., Wen-Leva (38) for the 'thin-film' model and Wender-Cooper (39) for the 'emulsion-packet' model.

III. FUTURE RESEARCH REQUIRED

It is to be expected that future research in this field will be concer trated mainly on bridging the gap between laboratory and industrial fluidiz bed design with the emphasis on gaining information about the effects of pa meters such as those outlined in the conclusions.

Until the mechanics of gas-particle motion in a fluidized bed and its relation to the physical properties of the bed is better understood, no the retical equation describing the heat transfer properties of fluidized beds suitable for use as a design correlation can be formulated. For example, p diction of bed voidage at any fluidized conditions with any degree of accur is difficult with existing correlations, yet knowledge of this important pa meter is essential for specifying the state of fluidization of a bed.

The problem of scaling-up laboratory experimentation into full-scale units is to maintain the quality of fluidization. As more data are made av able from large-scale units, general design trends will become apparent. F example, Volk (49) used vertical surfaces in the form of tubes to modify th 'equivalent' diameter of his large-scale bed to conform with the diameter of his small-scale unit and Petrie (31) found that finned surfaces on the flu ized bed side of heat exchanger tubes increased heat transfer rates twofol Nonetheless, a sound interpretation of heat transfer in fluidized beds mus await future developments in our understanding of bed mechanics.

APPENDIX I

Nomenclature

Ar	Avogadro No. $\frac{d_p^3 \rho_s \rho_g g_c}{\mu^2}$
C_g	heat capacity of fluidizing gas
C_s	heat capacity of solid particles
C_R	correlation factor for non-axial location of internal tube
d_p	particle diameter
d_b	bubble diameter
d_t	external diameter of immersed object (tube, sphere)
d_t	diameter of containing vessel
G	superficial mass velocity of fluidizing gas
G_{mf}	superficial mass velocity of fluidizing gas at minimum fluidization
h	heat transfer coefficient
h_{exp}	heat transfer coefficient measured by investigator
h_{calc}	heat transfer coefficient calculated according to one of the generalized correlations
H_t	height of heat transfer surface exposed to fluidized bed
k_e	emulsion thermal conductivity
k_g	thermal conductivity of fluidizing gas
k_s	thermal conductivity of solid particles
ℓ_t	length of immersed tube
L_f	height of bubbling fluidized bed
L_{mf}	bed height at minimum fluidizing conditions
Nu	Nusselt No. $\frac{hd_p}{k_g}$
Nu_T	Nusselt No. $\frac{hd_t}{k_g}$
Pr	Prandtl No. $\frac{c_p \mu}{k_g}$

R expansion ratio, L_f/L_{mf}

Re Reynolds No. $\frac{Gd_p}{\mu}$

T_B bed temperature

u superficial velocity of fluidizing gas

u_{mf} superficial velocity of fluidizing gas at minimum fluidizing conditions

ε_f void fraction in fluidized bed

ε_{mf} void fraction in fluidized bed at minimum fluidizing conditions

η fluidization efficiency (defined by Leva (6))

μ viscosity of fluidizing gas

ρ_g density of fluidizing gas

ρ_s density of solid particles

ρ_{mf} bulk density of fluidized bed at minimum fluidization

ϕ_d emulsion-packet contact frequency (Mickley (21))

ν kinematic viscosity

APPENDIX II

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APPENDIX IIIA

SUMMARY OF DENSE - AND DILUTE - PHASE FLUIDIZED BED HEAT TRANSFER - SURFACE INVESTIGATIONS

(a) DENSE PHASE HEAT TRANSFER TO CONTAINING WALL

Investigator	System	Apparatus: Type	Apparatus: D_T (ins)	Apparatus: H_t (ins)	Flow Range G (lb/hr ft^2)	T_B (°F)	Particles: ε_f	Particles: ρ_s (lb/ft^3)	Particles: d_p (μ)	Reference
Agarwal	Air - Sand, Graphite	Steam Jacket	1.5	14.5	150-1200			24-26	8-100	1
Ciborowski	Air - Sand, Coke	Steam Jacket	1.93	39.4	24.6-1640	210		100-164	150-2000	3
	Alumina	Electric	1.06	39.4				121	44-210	2
Dow & Jakob	Air - Coke, Iron Powd., Aerocat	Steam Jacket	2 3	23-26.5	50-300	200-220	0.52-0.69	121-466	111-171	5
Drinkenburg	Air - Catalyst		3.94 11.3						20-150	4
Leva	Air, CO_2, He, N_2 - Sand, Iron, Silica-Gel	Steam Jacket	2 4	25-26	1.5-1100	258-413	0.35-0.75	80-500	39-454	6
Levenspiel	Air - Coal, Glass Beads, Catalyst	Electric Heating Jacket	4	3 Sect. 2,5,2	79-4350		0.42-0.86	64-180	76-433	7
Massimilla	Air - Glass Beads	Water Jacket	3.54	10 Sect 3.94	334-1290				700	9
Matsuyama	Air - Charcoal, Silica-Gel		3.5						80-1000	8
Toomey	Air - Glass Beads	Water Jacket	4.73	7 Sect. 5.0	24-1542	200	0.36	167-179	55-842	10
Van Heerden	Air, CH_4, CO_2, H_2-N_2 - Carborundum, Iron Oxide, Coke, Lead Alloy	Water Jacket	3.4	4	44-779	50-86		37.5-694	50-800	11

APPENDIX IIIA

(b) DENSE PHASE HEAT TRANSFER TO IMMERSED SURFACE

Investigator	System	Internals: Type & Position	Internals: d_t (ins)	Internals: ℓ_t (ins)	D_T (ins)	G (lb/hr ft^2)	T_B (°F)	Particles: ε_f	Particles: ρ_s (lb/ft^3)	Particles: d_p (μ)	Reference
Ainshtein	Air - Sand	Cyl Electric Heater	0.86 1.18	4.3 8.3	11.7 5.98	57.5-355	68-270		159	163-285	12
Baerg	Air - Iron Sand Glass Beads Catalyst	Cyl Electric Heater Vertical	1.25	10	1.25 5.5	1.85-605	24-65	0.39 0.75	119-434	60-878	13
Bondareva	Air - Sand	Cyl Electric Heater Vertical	0.39	7.9	3.22	47.2-1640	200		144	100-477	14
Campbell	Air - Sand Aluminium Graphite	Cooling Coil	Area 18-70 in^2		2.31	40-100	87-145		24.6-27.2	241-3840	15
Chechetkin	Air - Carbon Silica Catalyst	Cyl Cooler Vertical	0.5	39.4		595-2760	239-320			250-840	16
Ernst	Air - Quartz	Rect Heater Cyl Horizontal	0.394 0.236	Area= 34in^2	5.9	1.5-820	68-86		159	100-700	17
Highley	Air - Boiler Ash	Tube Bundle Heated Horizontal	1.37 2.34	36	Sq. 36x36	320-1280	77		193	450	18
Jacob	Air - Glass Beads CO_2 He H_2-N_2	Hot Wire Probe	(0.132 mm)	-	3.0	2.0-300	50-68	0.46-0.6	154	30-290	19
Kharchenko	Combustion Gas - Fire-clay Quartz	Sphere Cooled	2.34		8.65	0.1-2.4 m/sec.	1922		144-164	340-1660	20
Mickley	Air - Glass Beads CO_2 Microspheres He NH_3 CH_4	Cyl Electric Heater Vertical	0.25	6.0 18.0 -26.0	4	11-302	110	0.45-0.65	49-153	60-850 43-320	21 22

APPENDIX IIIA

(Continued)

Investigator	System	Internals: Type & Position	Internals: d_t (ins)	Internals: ℓ_t (ins)	D_T (ins)	G (lb/hr ft^2)	T_B (°F)	Particles: ε_f	Particles: ρ_s (lb/ft^3)	Particles: d_p (μ)	Reference
Miller	Air - Carborundum H_e Alumina CO_2 Silica-Gel	Cyl Cooler Vertical	0.37	22	2	6.4-200	120-414		70-243	88-249	24
Olin & Dean	Air - Glass Beads Sand	Cyl Vertical	0.375	22.5	4	11.1-166	0.45-0.6			40-450	23
Petrie	Air - Metal Oxides Sand	Cyl Heated Horizontal Tube Bundle	0.75 0.5	47	Sq. 24x24 Cyl 12	59-283	260-930		93-188	256-735	31
Sarkits	Air - Catalyst Silica-Gel Quartz Coke	Cooling Coil		Area 0.15 in^2	1.93 2.88 3.94	26.6-3480	250-300		31.1-83.4 (bulk)	127-4500	25
Shirai	Air - Alumina	Spherical Heater Cyl	0.76 1.81 0.394	0.394	11					100-240	26
Toomey	Air - Glass Beads	Cyl Electric Heater Vertical	0.5	7 Sect 5.0	4.73	24-1542	200	0.36-	167-179	55-848	10
Varygin	Air - Quartz Sand Ferrosilicon	Sphere Cooled	0.79		3.26 6.2	41-1150	68-86		156-425	82.5-1160	27
Vreedenburg	Air - Sand Iron Ore	Cyl Vertical Horizontal	1.36 0.67-2.0	39-42	22.2	65-300	100-400		165-330	234-600	28 29
Wicke	Air - Carborundum H_2 Sand CO_2 Aluminium Lead	Cyl Heated Vertical	0.52	1.06	3.44	u = 10-70 cm/sec			160-700	65-3000	37
Ziegler	Air - Cu, Ni, Pb Spheres	Cyl Heater	0.5	2.5	4	70-145		0.53-0.79	554-559	140	30

APPENDIX IIIA

(c) DILUTE PHASE

Investigator	System	Internals (If Any): Type & Position	Internals (If Any): d_t (ins)	Vessel: Type	Vessel: D_T (ins)	G (lb/hr ft^2)	T_B (°F)	Particles: ϵ_f	Particles: ρ_s (lb/ft^3)	Particles: d_p (μ)	Reference
Bartholomew	Air - Sand Aluminium Marble	-	-	Electric Heating Jacket	4 H_T=30	96.4-935	257-599	0.54-0.95	160-167	84-251	32
Brazelton	Air - Glass Beads	-	-	Electric Heating Jacket	2 H_T=12	95-3780				70-1100	33
Trense	Air - Glass Spheres	Cyl Heater		-	3.94	96.4-3900				150-1100	34
Trilling	Air - Glass Spheres			Electric Heating Jacket	4	135-2650	300-460	0.56-0.98	151-177	41-452	35
		Cyl Heater Vertical	0.49	-	2.88						
Urie	Air				2.88					41-450	36

APPENDIX IIIB

SUMMARY OF THEORETICAL AND EMPIRICAL CORRELATIONS FOR FLUIDIZED BED HEAT TRANSFER - SURFACE

Investigator	Correlation	Reported Accuracy	Special Comments
Ainshtein (12)	$\frac{h\delta}{k} = 0.96\left(\frac{Gd_p}{\mu\varepsilon}\right)^{0.34} Pr^{0.33}\left(\frac{\alpha}{D_T}\right)^{0.16}$	± 10%	Horizontal Tube $\delta = \frac{d_p}{6(1-\varepsilon)}$; α = vertical distance from grid
	$\frac{h\delta}{k} = 1.18\left(\frac{Gd_p}{\mu\varepsilon}\right)^{0.285}\left(\frac{u}{\varepsilon u_{mf}}\right)^{-0.2}\left(1 - \frac{\beta}{R_T}\right)^{0.36}$	± 15%	Vertical Tube β = horizontal distance from axis
Baerg (13)	$h_{max} = 49 \log\left(\frac{0.00037\rho_g}{d_p}\right)$	± 25%	
	$h = h_{max} - 55 \exp\left\{-0.012(G - 0.71\rho_g)\right\}$		
Bartholomew (32)	$\frac{h}{C_g G} = \frac{1.56 + \ln(ReC_m^{-2/3} - 0.0120)}{-0.227\, Pr^{2/3} C_m^{0.42}}$		$C_m = \frac{d_p^3 g_c \rho_g(\rho_s - \rho_g)}{\mu^2}$
Brazelton (33)	$\frac{h}{C_g G'} = 0.72\, Re'^{-0.87}(1 - \varepsilon_f)^{-1.3}$		$Re = \frac{D'G'}{\mu}$ D' - equivalent diameter of free area across bed; G' - based on void area of bed
Ciborowski (2)	$h = 55\, d_p^{0.468}(G - G_{mf})^{0.52}$		$163 < d_p < 590\mu$ } $1.05 < \frac{G}{G_{mf}} < 1.8$
	$h = 50\, \phi^{-0.54}(G - G_{mf})^{0.27}$		$1000 < d_p < 2160\mu$ }; ϕ - Leva shape factor
Dow (5)	$\frac{hD_t}{k} = 0.55\left(\frac{D_t}{L_f}\right)^{0.65}\left(\frac{D_t}{d_p}\right)^{0.17}\left(\frac{(1-\varepsilon_f)\rho_s C_s}{\varepsilon_f \rho_g C_g}\right)^{0.25}\left(\frac{D_t G}{\mu}\right)^{0.8}$	± 10%	For smooth, dense phase system. No channelling or slugging
Jacob (19)	$h = \alpha_o(1-\varepsilon_f)(1 - e^{-p k_g^r})$		α_o, P empirical constants; k_g^r - radial component of k_g
Kharchenko (20)	$h_{max} = 33.7\, \rho_s^{0.2} k_g^{0.6} d_p^{-0.36}$	± 8%	

APPENDIX IIIB
(Continued)

Investigator	Correlation	Reported Accuracy	Special Comments
Leva (6)	$h = 0.64\left(\frac{kG\eta}{\mu}\right)$	± 22%	$39<d_p<109\ \mu$
	$h = 3.0 \times 10^6\ k\ d_p\left(\frac{d_p G\eta}{\mu R}\right)^{0.6}$	± 27.5%	$161<d_p<452\ \mu$
Levenspiel (7)	$\frac{h}{C_g G} = 0.6\left(\frac{d_p G}{\mu}\right)^{-0.7}$		Empirical
	$\frac{hd_p}{k} = \frac{0.417(1-\epsilon_f)^{0.5}}{A_1} Re^{0.5}$ streamline flow at wall		Theoretical $A_1 = (1+B_1^2)^{1.5} - B_1^3$
			$B_1 = .0294(1-\epsilon_f)^{0.5} Re^{0.5}$
	$\frac{hd_p}{k} = \frac{8.02}{A_2}(1-\epsilon_f)^{0.8} Re^{0.2}$ turbulent flow near wall		$A_2 = (1+B_2^{1.25})^{1.8} - B_2^{2.25}$
			$B_2 = 0.478(1-\epsilon_f)^{0.8} Re^{0.2}$
Mickley (22)	$h = 1.13\,[(1-\epsilon_{mf})(1-f_o)\phi_d \rho_s C_s k_e]^{0.5}$		$1-f_o$ - fraction of surface bathed in emulsion ϕ_d - emulsion packet contact frequency k_e - emulsion thermal conductivity
Miller (24)	$h = 1.5\ \frac{G^{0.32}\ k_s^{0.072}\ k_g^{2.4}}{d_p^{0.96}\ C_g^{1.6}\ \mu^{0.8}}$		
Petrie (31)	$Nu_T = 14\left(\frac{G}{G_{mf}}\right)^{0.33} Pr^{0.33}\left(\frac{d_t}{d_p}\right)^{0.66}$		

APPENDIX IIIB

(Continued)

Investigator	Correlation	Reported Accuracy	Special Comments
Sarkits (25)	$Nu=0.0133Re^{0.41} Ar^{0.27} Pr^{0.33}\left(\frac{C_s}{C_g}\right)^{0.45}\left(\frac{D_T}{d_p}\right)^{0.16}\left(\frac{H_o}{d_p}\right)^{0.45}$	± 25%	0.34<Re<6.4 } laminar flow
	$Nu=0.00705Re^{-0.14} Ar^{0.49} Pr^{0.33}\left(\frac{C_s}{C_g}\right)^{0.45}\left(\frac{D_T}{d_p}\right)^{0.16}\left(\frac{H_o}{d_p}\right)^{0.45}$	± 15%	0.78<Re<16.5 } laminar flow
	$Nu=0.0528Re^{0.35} Ar^{0.25} Pr^{0.33}\left(\frac{C_s}{C_g}\right)^{0.1}\left(\frac{D_T}{d_p}\right)^{0.13}\left(\frac{H_o}{d_p}\right)^{0.16}$	± 15%	40<Re<600 } turbulent flow
	$Nu=0.018Re^{-0.12} Ar^{0.56} Pr^{0.33}\left(\frac{C_s}{C_g}\right)^{0.1}\left(\frac{D_T}{d_p}\right)^{0.13}\left(\frac{H_o}{d_p}\right)^{0.16}$	± 10%	100<Re<900 } turbulent flow
Toomey (10)	$\frac{hd_p}{k} = 3.75\left(\frac{d_p u_{mf}}{\mu} \rho_g \log \frac{u}{u_{mf}}\right)^{0.47}$	± 5%	
Trilling (35)	$h = 0.0118\left(\frac{\rho_f G}{d_p^3}\right)^{0.263}$		External
	$h = 0.0433\left(\frac{\rho_f^2}{d_p^3}\right)^{0.238}$		Internal
Van Heerden (11)	$\frac{Nu}{Pr^{0.5}} \frac{1}{\psi^{0.45}}\left(\frac{\rho_{mf}}{\rho_g}\right)^{0.18}\left(\frac{\rho_g C_g}{\rho_{mf} C_s}\right)^{0.36} = 0.028\left(\frac{G}{G_{mf}}\right)^{0.45}$		$\psi = \frac{\rho_g \rho_{mf} d_p^3 g_c}{\mu^2}$; $2<\frac{G}{G_{mf}}<20$ and Re<5
Varygin (27)	$Nu_{max} = 0.86\ Ar^{0.2}$		30<Ar<135,000
Vreedenburg (28)	$Nu = C_R\left\{\frac{G\nu}{(G\nu)_{mf}}\right\}^{0.35}$		Vertical C_R - correction factor for non-axial location of tube
	$Nu = 1.25\left\{\frac{G\nu}{(G\nu)_{mf}}\right\}^{0.35}$		Horizontal

APPENDIX IIIB

(Continued)

Investigator	Correlation	Reported Accuracy	Special Comments
Vreedenburg (29)	$Nu_T = 0.66 \left(\frac{G d_t \rho_s (1-\varepsilon_f)}{\rho_g \mu \varepsilon_f} \right)^{0.44} Pr^{0.3}$ $Nu_T = 420 \left(\frac{G d_t \rho_s}{\rho_g \mu} \frac{\mu^2}{d_p^3 \rho_s^2 g_c} \right)^{0.3} Pr^{0.3}$		$\frac{d_p G \rho_s}{\mu \rho_g} < 2050$ $\frac{d_p G \rho_s}{\mu \rho_g} > 2550$ } Horizontal tubes
Wicke (37)	$h = \frac{K_Z}{2H_T} \left\{ 1 - \exp - \left(\frac{2H_T}{\ell_g} \frac{k_g}{K_Z} \right) \right\}$		$K_Z = \rho_s (1-\varepsilon_{mf}) C_s u_s \ell_e$ ℓ_g - conduction layer thickness ℓ_e - emulsion layer thickness
Ziegler (30)	$Nu = \frac{7.2}{\left(1 + \frac{6 k_g \hat{t}}{\rho_s C_s d_p^2} \right)^2}$		$\hat{t}$ - mean contact time of particles at wall

APPENDIX IIIC

GENERALIZED CORRELATIONS FLUIDIZED BED HEAT TRANSFER TO SURFACE

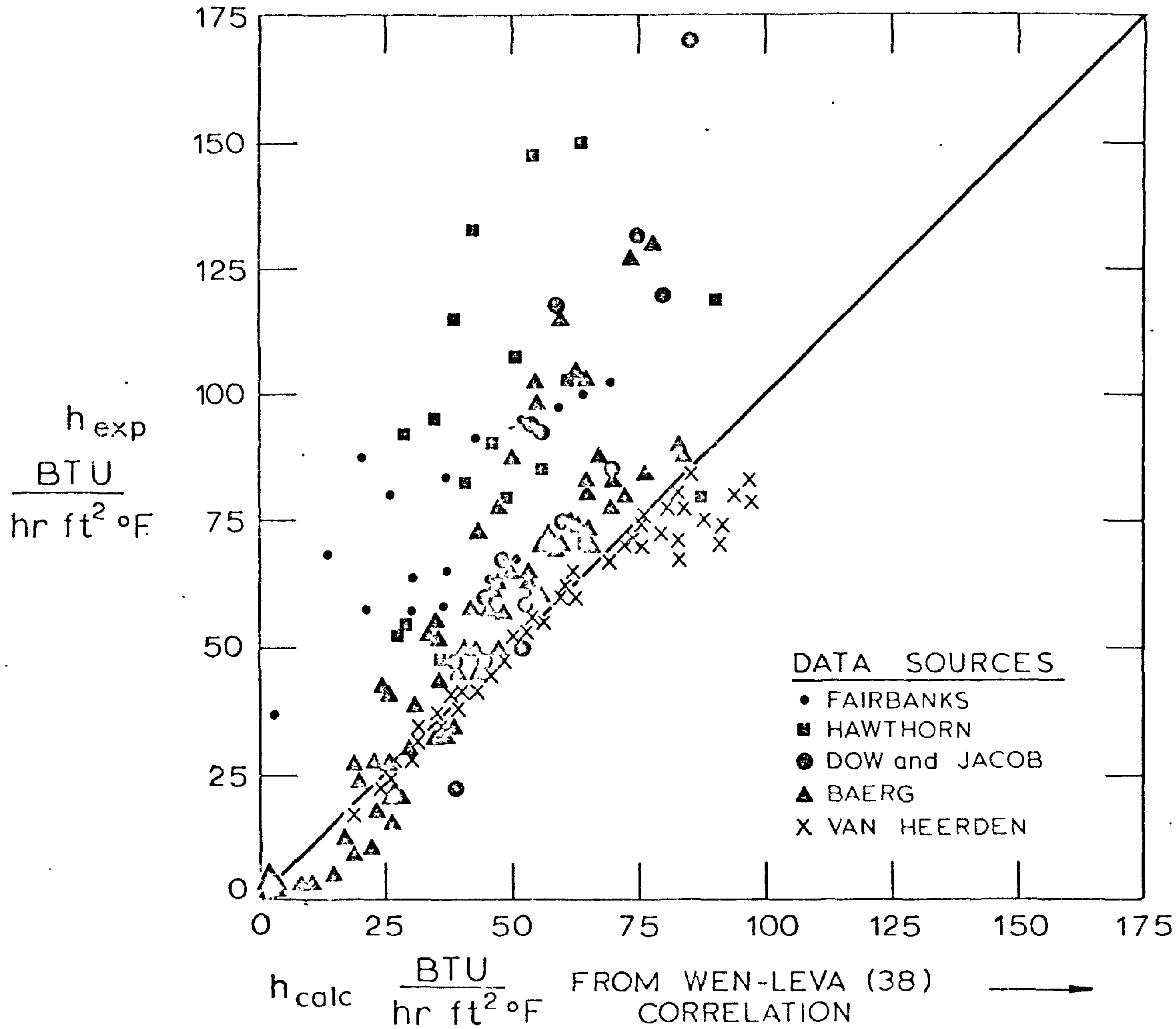
Investigator	Generalized Correlation	Range of Variables	Data Sources
Wen and Leva (38)	$\frac{hd_p}{k_g} = 0.16\, Pr^{0.4}\, Re^{0.76} \left(\frac{\rho_s C_s}{\rho_g C_g}\right)^{0.4} \left(\frac{u_o^2}{g d_p}\right)^{-0.2} \left(\eta \frac{L_{mf}}{L_f}\right)^{0.36}$	Accuracy $\pm$ 50% $39<d_p<848$ μ $37.5<\rho_s<694$ lb/ft^3 $0.35<\epsilon_f<0.75$ $0.009<k_g<0.102$ Btu/ft°F hr $2<D_t<4.73$ ins $50<T_B<400$°F	(5) (6) (10) (11) External Wall
Wender-Cooper (39)	Plot of y vs Re where $y = \frac{Nu/[(1-\epsilon_f)C_s\rho_s/C_g\rho_g]}{1+7.5\exp[-0.44(H_T/D_T)(C_g/C_s)]}$	Accuracy $\pm$ 22% $50<d_p<850$ μ $80<\rho_s<500$ lb/ft^3 $0.35<\epsilon_f<0.98$ $0.017<k_g<0.108$ Btu/ft°F hr $2<D_T<4.73$ ins $200<T_B<599$°F	(5) (6) (10) (32) (35) External Wall
	$\frac{hd_p}{k_g} = 0.01844\, C_R(1-\epsilon_f)\left(\frac{C_g\rho_g}{k_g}\right)^{0.43} Re^{0.23} \left(\frac{C_s}{C_g}\right)^{0.8} \left(\frac{\rho_s}{\rho_g}\right)^{0.66}$	Accuracy $\pm$ 19.5% $40<d_p<880$ μ $49<\rho_s<434$ lb/ft^3 $0.4<\epsilon_f<0.95$ $0.0061<k_g<0.089$ Btu/ft°F hr $2.88<D_T<6.4$ ft $24<T_B<460$°F	(10) (21) (23) (13) (35) + Commercial Data Internal Surface

APPENDIX IIID

EFFECT OF INDIVIDUAL BED PROPERTIES ON HEAT TRANSFER COEFFICIENT

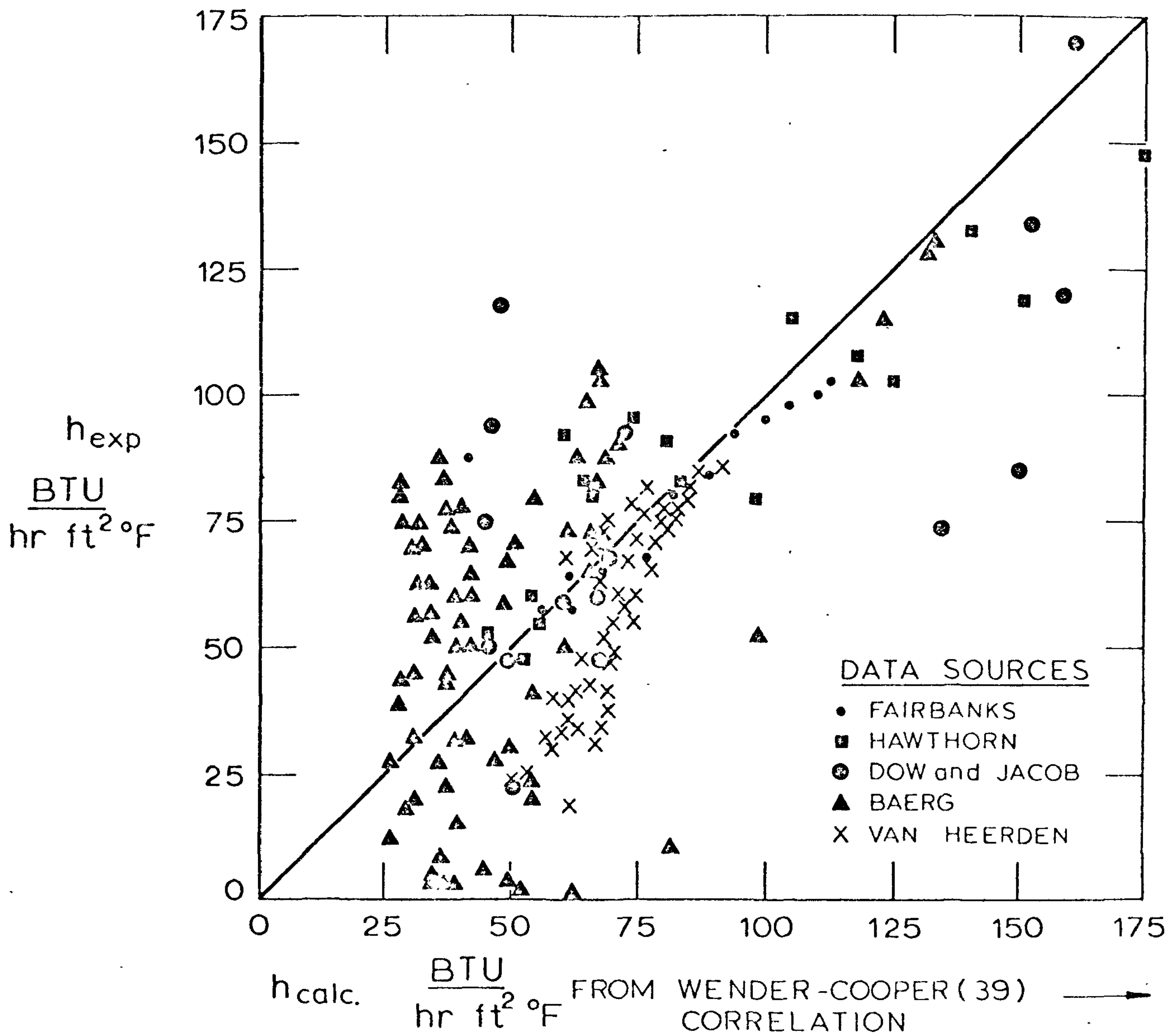
VALUES LISTED ARE EXPONENTS TO WHICH EACH PROPERTY IS RAISED

	Bed Thermal Propts.				Properties Affecting Bed Dynamics								
	k_g	C_g	k_s	C_s	ρ_g	μ	ρ_s	d_p	u	D_T	H_T	Voidage Function $f(\varepsilon_f)$	Other
Ainshtein (12)	+0.67	+0.3			+0.34	-0.01		-0.66	+0.34	-0.16		$\frac{1-\varepsilon_f}{\varepsilon_f^{1.34}}$	
	+1.0				+0.29	-0.29		-0.61	+0.085	-0.36		$\frac{1-\varepsilon_f}{\varepsilon_f^{1.1}}$	
Ciborowski (2)					0.52			0.47	+0.52				
Dow (5)	+1.0	-0.25		+0.25	+0.55	-0.8	0.9	-0.23	+0.8	+0.68	-0.65	$\left(\frac{1-\varepsilon_f}{\varepsilon_f}\right)^{0.25}$	
Leva (6)	+1.0				+1.0	-1.0			+1.0				$\eta^{1.0}$
	+1.0					-0.6		+1.6	+0.6				$\eta^{0.6}\ R^{-0.6}$
Levenspiel (7)		+1.0				-0.7		-0.7	+0.3				
Mickley (22)	+0.33			+0.5			+0.5					$(1-\varepsilon_{mf})^{0.5}$	$[(1-f_o)\phi_d]^{1/2}$
Miller (24)	+2.4	-1.6	+0.072		+0.32	-0.8		-0.96	+0.32				
Petrie (31)	+0.66	+0.33				+0.33		-0.66	+0.33				$d_t^{-0.33}u_{mf}^{-0.33}$
Sarkits (25)	+0.66	-0.12		+0.45	+0.35	-0.55		-0.25	-0.14	+0.16	+0.45		$\rho_{bulk}^{0.49}$
	+0.66	+0.23		+0.1	+0.6	-0.52		-0.39	+0.35	+0.13	+0.16		
Toomey (10)	1.0				+0.47	-0.47		-0.53					$\log\left(\frac{u}{u_{mf}}\right)^{0.47}$
Trilling (35)					+0.26			-0.78	+0.26				$\rho_{bulk}^{+0.26}$
Van Heerden (11)	+0.5	+0.14		+0.36	+0.63	-0.4		+0.35	+0.45				$u_{mf}^{-0.45}$
Vreedenburg (26)	+0.7	+0.3				-0.44	+0.44		+0.44			$\left(\frac{1-\varepsilon_f}{\varepsilon_f}\right)^{0.44}$	$d_t^{-0.54}$
	+0.7	+0.3				+0.6		-0.9	+0.3				$d_t^{-0.7}$
Wen-Leva (38)	±0.6			+0.4	+0.36	-0.76	+0.4	-0.04	+0.36			$\left(\frac{L_{mf}}{L_f}\right)^{0.36}$	$\eta^{0.36}$
Wender (39)	+0.57	-0.3		+0.8		-0.23	+0.66	-0.77	+0.23			$(1-\varepsilon_f)$	C_R



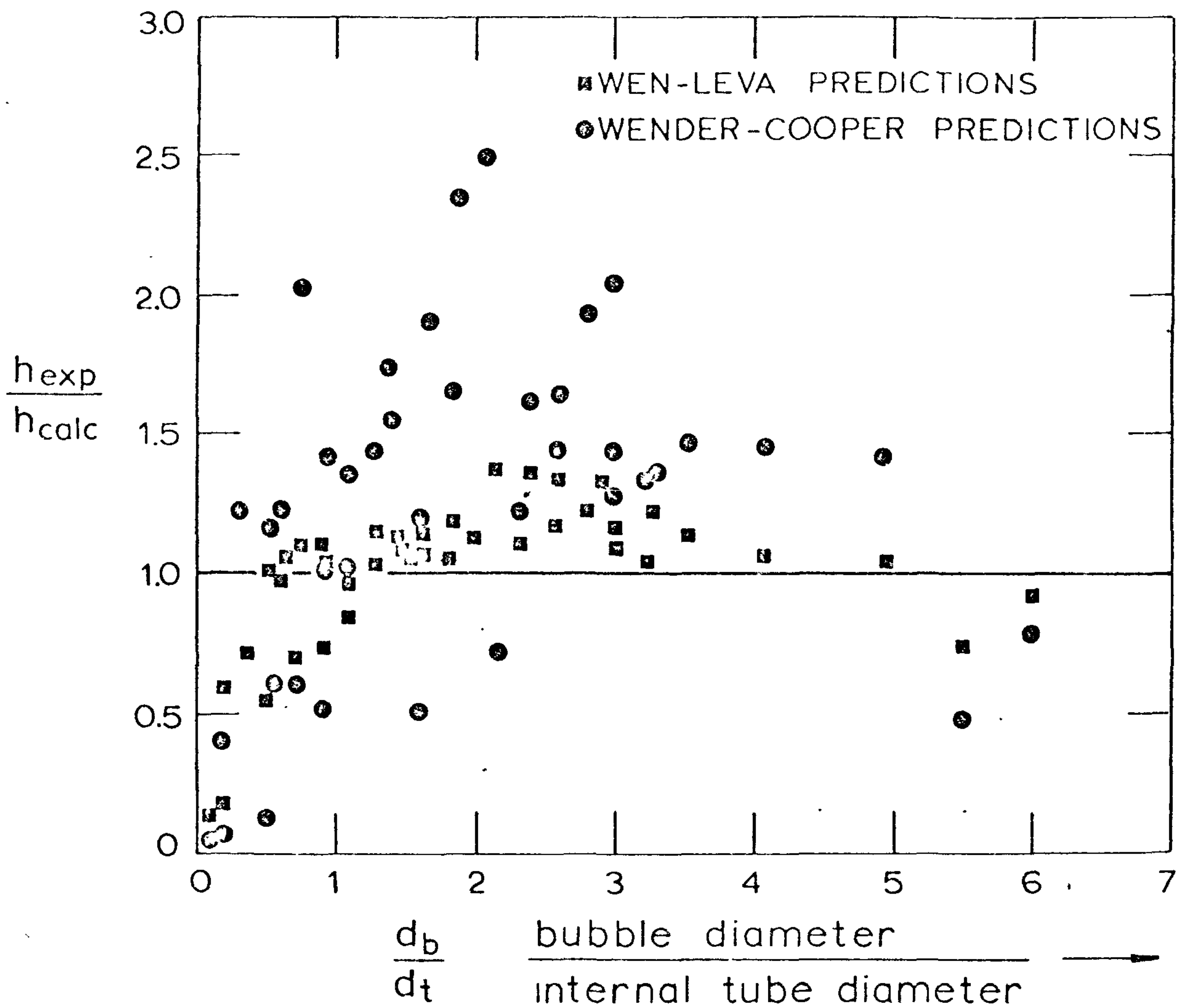
APPENDIX IV GRAPH 1

COMPARISON OF MEASURED AND CALCULATED HEAT-TRANSFER COEFFICIENTS.



APPENDIX IV GRAPH 2

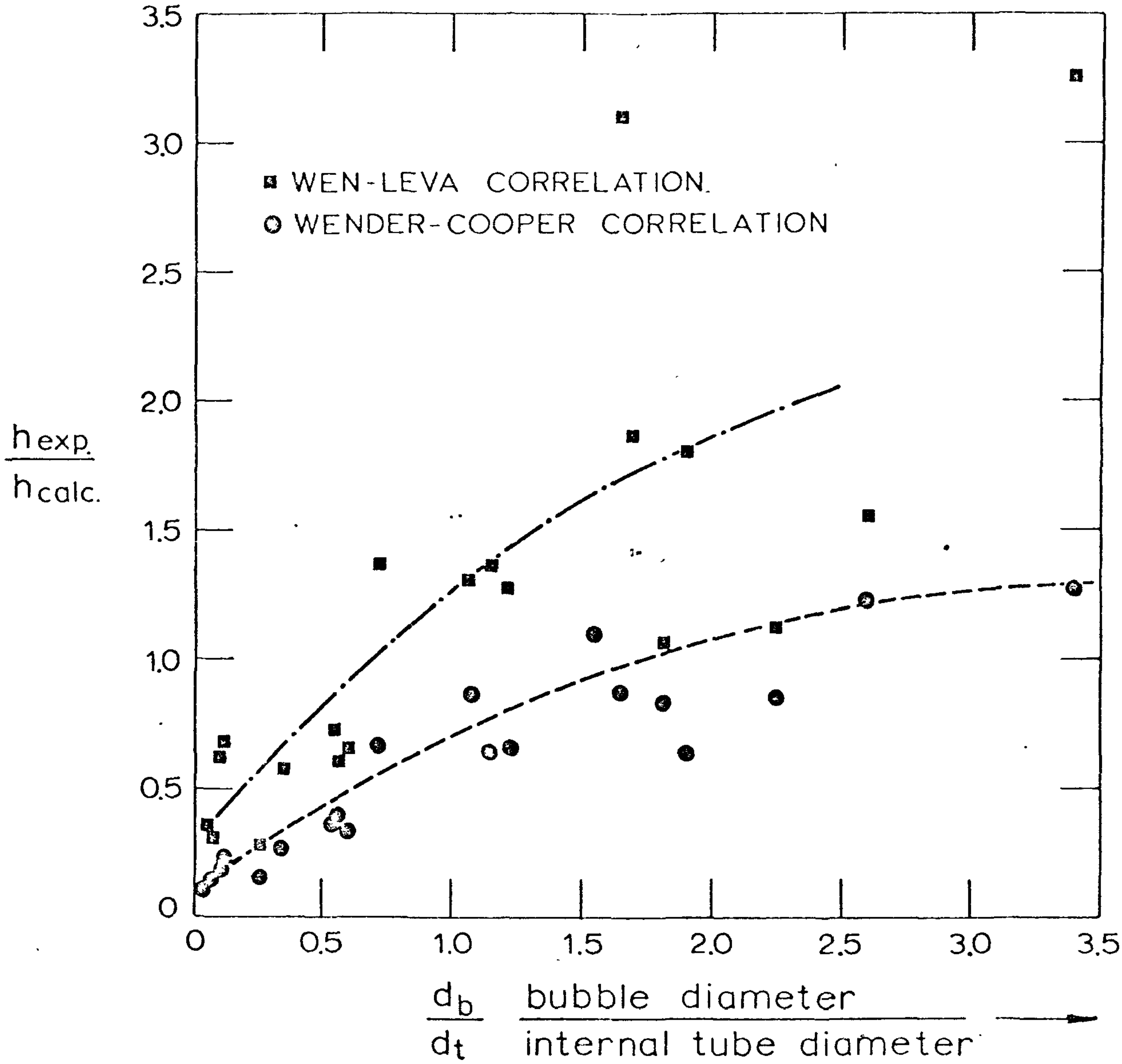
COMPARISON OF MEASURED AND CALCULATED HEAT TRANSFER COEFFICIENTS



APPENDIX IV GRAPH 3

DATA OF BAERG (13)

INTERNAL VERTICAL HEATED TUBE $d_t = 1.25''$



APPENDIX IV GRAPH 4
DATA OF VREEDENBURG (28)
SINGLE HORIZONTAL TUBES d_t = 0.66" and 1.35"

APPENDIX V

Calculation of Heat Transfer Coefficients in a Large Diameter Bed Compared with the Data of Highley (18)

Vessel Diameter - 3 ft.

h_{avg} - Average value of heat transfer coefficient from measurements to individual horizontal tubes in a multiple tube bank, Btu/hr-ft^2-°F.

h_{w-c} - Heat transfer coefficient calculated from the correlation of Wender and Cooper (39), Btu/hr-ft^2-°F.

h_{w-l} - Heat transfer coefficient calculated from the correlation of Wen and Leva (38), Btu/hr-ft^2-°F.

u - Superficial gas velocity, ft/sec.

u	h_{avg}	h_{w-c}	h_{w-l}
0.98	33.4	31.9	42.3
1.97	38.2	30.2	64.8
2.95	41.8	28.9	77.6
3.94	43.8	27.9	86.7