HEAT TRANSFER IN FLUIDIZED SYSTEMS

by

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A THESIS

submitted to

OREGON STATE COLLEGE

in partial fulfillment of
the requirements for the
degree of

MASTER OF SCIENCE

June 1949

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E. W. 7/6/49

ACKNOWLEDGMENT

Appreciation is gratefully expressed to Professor J. S. Walton, Head of the Chemical and Metallurgical Engineering Department for the assistance he has given in the preparation of this paper and under whose supervision this work was undertaken; to Bob Simon who proposed the project, constructed and calibrated the apparatus in its initial design and who gave many valuable suggestions.

Acknowledgments are also due Bob Mang for his assistance in the laboratory, Monta Montgomery for typing and correcting the rough drafts of the paper and Lillian Purdy for typing the manuscript.

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INTRODUCTION

Methods for producing fluidization of solids have been known since 1878 when a patent was granted to Frederic Luckenbach (1) for an improvement in apparatus for drying cereal grains. His invention consisted of a vessel with a conical false bottom through which hot or cold air or steam could be introduced into a bed of grain, the idea being similar to that used in fluidization today. He stated that "By this means...a powerful and rapid drying and heating action was produced rendering the process very rapid...". Subsequent patents (2,3,4) mentioned the uniformity of temperature in the bed, but none gave any indication of actual heat transfer rates in or through the bed. Since these earliest descriptions of such a process, most of the investigators in this field have focused their efforts on fluidization procedures and equipment design; (5,6,7,8,9,10,11,12,13) while in related fields work has been done on fixed and moving beds (14,15, 16,17,18,19,20) as well as on clouds of falling particles (21).

During World War II years, fluidization of solids attained industrial importance by the construction of a number of plants for catalytic cracking of petroleum, using a process developed by the Standard Oil Development Company. This process was first called "jiggling" and later "fluidization"--terms which were descriptive of the behavior of the solids in the reactor bed.

More recently, investigations into fluidization fundamentals have been concerned with the study of fluid flow (22,23,24,25) and, except for a paper from the Bureau of Mines Laboratories published late in 1948 by Leva and co-workers (26) another by Kettenring and Smith (27), and a Ph.D. thesis by Robert Simon (28), there has been no information available on heat transmission rates in fluidization process reactors. Leva et al examined the heat transfer rates between fluidized beds and the walls of the reactor. Sand and iron catalyst were employed ranging in size from 130 to 400 mesh with air mass velocities between 2 and 450 lb/hr.-sq.ft. An equation was derived for the calculation of h, the heat transfer coefficient, in terms of the mass velocity.

Reported in the present paper are data showing the effect of particle size and mass velocity on the coefficient h over a wide range of sizes and mass velocities. A number of narrowly sized cuts of pulverized Utah coal between 14 and 65 mesh were used in determining most of the data on the effect of particle size. To relate this information to the Bureau of Mines data, as reported by Leva and co-workers, a few subsequent runs were made with 100-150 mesh coal. Air mass velocities were varied from 50 to 1100, these being the minimum and maximum practical values which could be employed in the apparatus. It was believed this range was sufficiently wide to cover most of the cases where fluidized systems are now or will be employed; however, in a few instances with dense solids, still higher mass velocities may be necessary to produce fluidization.

METHOD OF ATTACK

A weighed quanitity of solid was introduced into a vertical four-inch I.D. galvanized iron tube after which a steady flow of hot air at constant temperature was passed upward through the tube at the desired rate. When the system came to thermal equilibrium, air temperature in the bed, tube-wall temperature, and temperatures at various points in the insulation were measured. The heat flow through the tube wall and insulation is calculated by the equation for the radial heat flow through cylinders (29, P.12).

$$q = \frac{2\pi kZ \ (t_2 - t_1)}{\ln \ r_2/r_1}$$

Where t_2 and t_1 are insulation temperatures measured at radial distances r_1 and r_2 from the center of the tube. Knowing the heat flux q permitted calculation of h, the local film coefficient at the inside wall of the tube by use of the equation.

$$h = \frac{q}{A \ (t_g - t_w)}$$

APPARATUS

A rotary blower driven by a 1 horsepower variable-speed motor supplied the air for all the experimental work. The air passed through an oil-mist filter consisting of a four-inch length of three-inch pipe filled with steel wool. A rupture disc, made by clamping four sheets of waxed paper between two halves of a flange union, was located just beyond this point. This part of the system was connected with the rest of the piping by a length of rubber hose in order to reduce vibrations. Just above the orifice section was an air bleed which, in combination with the variable-speed motor, made possible the control of the air flow at any rate between 50 and 1100 pounds per hour per square foot.

To measure the air-flow rate, two thin plate orifices were used. The smaller was 9/32 inches in diameter in a one-inch iron pipe and was used to measure mass velocities up to about 350 pounds per hour per square foot in the four-inch experimental tube. The larger one was 0.500 inches in diameter in a one-inch brass pipe. Both orifices were calibrated by means of a dry gas meter and were later rechecked after completion of the experimental work.

It was necessary to measure the gas temperature because the heat of compression was appreciable and was not all dissipated in the lines between the pump and the orifices. In operation, the air temperature was measured at the air bleed above the orifice section. A check valve was placed downstream of the orifices to prevent possible backflow of solids.

The air had three possible paths after it left the check valve.

1. It could pass through a 3/4 inch pipe to the bottom of the four-inch diameter (4.055 inches average inside diam.) glass tube, which was used to determine the bed height and fluidization characteristics of the solids used. Since only approximate determination of bed height was necessary, visual observation of the fluidized material in the glass tube gave ample accuracy. From the top of this tube, the air flowed via a 3/4 inch pipe to a cotton filter bag and then out into the room.

2. The second path for the air was through a one-inch line to the electric heater and then into the bottom of the four-inch diameter (3.98 inch average inside diameter) sheet-metal tube. From the

top of the tube, it flowed through a one-inch pipe to a filter bag and then out to the room. The heater was a 1400 watt, finned-space heater, and was enclosed in a two-inch by three-inch by eighteen-inch sheet-iron box with three baffles to force the air to flow parallel to the fins. Electrical contact was made by means of two spark plugs. The spark plugs were brazed to the box, and the central terminals were connected to the heater leads. The current to the heater was controlled by a 7.5 "Powerstat", variable transformer. A voltmeter and an ammeter measured the power input. Between the heater and the four-inch tube there was a one-inch three-way valve. The side opening was a bleed to allow preheat of the air without heating the tube. The connection between the one-inch entrance pipe and the four-inch metal tube was a sheet-metal cone 2 3/8 inches high. The 1/16-inch sheet-metal tube was connected by means of flanges 2½ inches above the top of the cone. The two flanges were thermally insulated from one another by means of asbestos paper gaskets. The whole system from the heater entrance to the top of the four-inch tube was insulated with one-inch of eighty-five per cent magnesia.

3. The third path for the air was through a one-inch pipe to the base of the tube bypassing the heater. A two-way, quick acting plug was located in this line.

The solid was fed into the sheet-metal tube from a hopper above the top of the tube and was connected to it by a one-inch pipe, which extended about one foot into the tube and was independent of the air exit line. In the glass tube the same pipe was used for the solid feed and exit of air.

The screen supporting the solid in the two tubes were different in design. In the metal tube, the brass-supporting frame for the 150-mesh screen was placed two inches above the flange and was made in the form of a damper which could be rotated manually from the outside. The maximum clearance between the frame and the tube wall was 0.002 inches which was small enough to prevent significant quantities of solid escaping around the edges of the screen. Solid was removed from the tube by rotating the damper to the vertical position and allowing the solid to run out through an arm of the tee below the neck of the cone. The frame of the sixty-five mesh glass-tube screen was a brass ring around which was wrapped enough rubber tape to give a tight seal when the screen was forced up into the glass tube. The bed of solid was removed by disconnecting the glass tube at the bottom flange, lifting the tube onto a special frame and then removing the screen.

Because a bed of fine solids packs quite tightly when air is not passing through it, air of several pounds pressure must be available to break it loose. High-pressure air from an eight-cubic-foot-per-minute compressor was connected into the line leading to the bottom of each tube to give a source of supply. Two additional lines were connected into the system above the orifices so that the compressor could be used either as an auxiliary supply, increasing the maximum air-flow rate slightly, or to replace the rotary blower when low-air flow rates were required. High pressure air was also used to blow out manometer connections and to clear the screens on the suction thermocouples.

A pressure tap about 2½ inches below the screen in the metal tube and another about five feet above the screen gave the combined pressure drop across the screen and the bed of solids. By subtracting the pressure drop at the same mass velocity with the tube empty, the 'net' pressure drop due to the bed was obtained. The pressure taps for the glass tube were located in the approach pipe and the air-exit pipe. A water manometer was used to read the pressure drop, and a second one gave the pressure drop across either orifice. A mercury manometer read the pressure at the upstream orifice tap.

Temperature in various points in the apparatus were determined by seventeen calibrated glass-fabric coated No. 20 gauge iron-constantan thermocouples. Lead wires were copper and extended from the cold junction, water at 70° F in a thermos bottle, to two multiple selector switches. The electromotive force was measured by a Leeds and Northrup Model 8662 Potentiometer which permitted reading temperatures to within ± 1/3°F.

Metal temperatures were measured by thermocouples brazed to the outside of the tube at the screen and 2,6,12 and 24 inches above it. Thermocouples of the following three types were used to measure gas and fluidized bed temperatures:-

(a) At the 2, 12 and 24 inch levels, screened suction thermocouples were used (Fig. 2, 3). These were made by inserting a glass-fabric covered two-wire couple into a length of 1/8" O.D. shelby steel tubing open at both ends with one end flared extending into the bed. This made it possible to draw hot air from the bed at a rapid rate past the thermocouple junction, thus giving true gas temperatures. Solid was prevented from entering the well by means of a fine screen soldered over the flared end of the well.

(b) Just below the screen where no solid was present, a bare suction thermocouple was installed similar to the ones just described, except that it had no screen over the end of the well.

(c) At the six inch level a bare thermocouple was installed with its junction bathed in solid.

To measure radial heat flow, six thermocouples were installed at various depths in the insulation2,6, and 12 inches above the screen. To prevent errors due to heat conduction along the thermocouple wire from the tip, the two inches of the thermocouple adjacent to the junction were installed in the insulation parallel to the walls of the metal tube and at the same depth as the junction. (Fig.3).

DISCUSSION AND RESULTS

Choice of Material:

The choice of the solid to be used in this study was governed by the following desired properties. The material should have a low density to facilitate fluidization in the particle-size range to be studied; and at the same time, a low attrition rate, thereby keeping the bed particle size constant. Based on a number of experimental runs with Ottawa sand, 3-A catalyst, charcoal, and a hard Utah coal, the latter was selected as the experimental material and proved to be entirely satisfactory.

Visual Observations:

When air is passed upward through a bed of granular material, one observes the following:

(1) At low rates of air flow there is no change in the bed appearance below a certain mass velocity which increases with particle size.

(2) As the rate of flow increases, the bed expands and some of the particles begin to vibrate. Air channels are formed. At these higher velocities the bed gives the appearance of bubbling and the particles assume a more violent motion and change relative positions.

(3) At still higher velocities, the fluidized, or pseudo-liquid state exists wherein the solid particles appear to be completely air-borne. In this state no stationary particles are observed. However, the particles' motion is not altogether random for they have a convective-like movement with ascending particles in the center of the tube with some small quantity carried up into the space above the dense or pseudo-liquid phase.

Channeling and slugging is observed at characteristic air flow rates depending on the weight and size of solid in the tube. The system is more susceptible to these effects with relatively large amounts of solids present or with narrow sized cuts.

Temperature measurements.

Since temperature measurements provide basic data for determining the values of h, it is pertinent to describe the variation in temperatures observed in the apparatus during the experimental runs.

In blank runs without solid present the bare thermocouple at the six inch level indicated lower temperatures than the vacuum couples, especially at low G values, apparently due to the low air velocity past the thermocouple junction.

In all runs with solid present in the tube the thermocouple readings at the two inch (vacuum) and six inch (bare) levles never differed by more than a fraction of one degree F, indicating that the bed temperature was uniform throughout. For G values greater than 400 lbs/hr.-ft.sq. this was also true for the vacuum thermocouple at the twelve inch level but at lower G values the latter thermocouple reading dropped considerably below the other two. From observations in the glass tube at comparable mass velocities it was determined that for these lower G values this thermocouple was above the dense phase of the fluidized bed, which when at rest was seven inches high. Horizontal temperature variation in the bed was found negligible at the six inch level by a traverse with the bare thermocouple, so all bed temperature measurements were subsequently taken at the center of the tube.

There may be some question as to the utility of the vacuum thermocouples which were designed so that air could be drawn past the junction at 5 ft./sec. giving a constant temperature reading within thirty seconds. In the bed they offer no advantage as they read the same temperature as the bare thermocouple; but when out of the bed or in blank runs their use is justified as shown in the previous paragraphs.

For all conditions when application of the vacuum was discontinued the readings dropped from 2 to 7°F. indicating that with enclosed thermocouples it is necessary to pass air rapidly past the junction. This will serve to warn other investigators that a protected thermocouple in a similar location may be expected to give unreliable readings unless this precaution is taken.

Calculation of h at the two inch level was not attempted because with fine solids present, a stagnant area adjacent to the tube wall was observed up to one inch above the screen. This dead space was caused by the ring which supported the screen blocking off air flow immediately above the ring support. Except at the high G values the thermocouple at the twelve inch level was above the dense phase of the bed as mentioned previously. A greater weight of solid in the bed which raised the dense phase above this thermocouples gave undesirable slugging effects. Hence while h values calculated at the two and twelve inch levels provided a check, only those values obtained for the six inch level were used in the final analysis of results. It may be noted that the bed temperature used in the calculations was the average of the readings of all the thermocouples within the bed.

Sequence of Experiments.

The first data were obtained in the air velocity range where true fluidization exists, because it was believed that this region was of primary interest from the engineering viewpoint. The coal fractions were sized with Tyler screens into cuts of 14-20, 20-28, 28-35, 35-48, and 48-65 mesh. By pressure drop measurements in the glass and metal tubes and by visual observations in the glass tube, it was found that the beginning of fluidization occurred at mass velocities of about 60 for the 48-65 mesh particles, but as particle size increased to 14-20 mesh the mass velocity to produce incipient fluidization progressively increased to 500.

After publication of the Bureau of Mines peper, in which values of h for 130 to 400 mesh particles at lower mass velocity values were reported, it was decided to extend the data into these lower mass velocity regions where partly expanded or partially fluidized beds exist, which corresponded to the conditions used by Leva et al in the afore mentioned Bureau of Mines paper. Hence, the data presented here cover the entire range of mass velocities values which it is possible to obtain in the apparatus employed.

Results:

Figures 4 and 5 and Tables I and II show the data obtained in these experiments. On Figure 4, the curve farthest to the right is for the largest mesh size fraction (14-20) coal employed, and for this fraction at mass velocities above about 200 the value of h increased very rapidly from its lowest value to near the maximum of about 15-16. At this point G was approximately 400. Thereafter, as G increased to 1100, the value of h decreased slightly. Whether this decrease was real may be argued because of the small magnitude of the change. The author holds no strong case for the decrease as shown on Figure 4, but believes that it may have been due to the fact that at the higher mass velocities the population of solid particles in the bed was relatively low (because the bed is expanded) and their efficacy in wiping off the film at the reactor wall was thus lessened. At the lowest mass velocities for the 14-20 mesh fraction, the two data points lying to the left of the curve may indicate leveling off of the values of h.

Again referring to Figure 4, it is seen that as particle size decreased the same trend of h variation with G existed but the differences between the static, expanded, and fluidized states were progressively less marked. This was not unexpected because pressure drop measurements and visual observations showed that fluidization commenced at lower G values with smaller particles. Values of h increased with decreasing particle size which showed that smaller particles were more efficient in decreasing bed to wall resistance. This may have been due to a better "wiping" effect and/or a greater surface contact with the smaller particles because they provided a greater number of point contacts with the wall surface for a given mass velocity.

It was not possible, when using the smallest particles, to determine whether h decreased at the highest velocities because solid carryover became excessive at these high velocities.

Figure 5 includes the same data shown on Figure 4, but plotted on a log log scale which allows inclusion of the h versus G curves (a a' and b b') obtained by Leva and co-workers. It will be seen that the Bureau of Mines data (for 130-400 mesh particles of sand and of iron catalyst) are approached by the data obtained in the present investigation. The authors attempted to obtain information on still smaller coal particles but encountered difficulty because the mesh size of the screen at the reactor bottom permitted the small coal particles to fall through the screen into the cone at the bottom of the reactor.

Sample Calculation:

Run 16M-17 has been selected in order to demonstrate the calculations. All the data on this run are given below:

Flow measurements

Orifice used - 9/32" diameter

Air temperature at orifice $t_o = 70^oF$

Pressure drop across orifice $\Delta p = 2.6'' H_2O$

For h calculations at $z = 6''$

$t_g = 270.33^oF$ (average of 4 readings on the t.c. at $z = 2''$ and $z = 6''$)

$t_w = 263.91^oF$ (average of 4 readings on t.c. at $z = 6''$)

$t_1 = 137^oF$ (average of 2 readings on the t.c. at $z = 6''$ and r_1)

$r_1 = 2^{25/32}''$

$t_2 = 218^oF$ (average of 2 readings on the t.c. at $z = 6''$ and r_2)

$r_2 = 2\ 7/32''$

<u>Calculations:</u>

Heat through insulation at $z = 6''$

$$q = \frac{2 \pi k Z (t_2 - t_1)}{\ln (r_2/r_1)}$$

where $k = 0.036$ Btu/hr-ft - ^oF

$Z = 1$ ft height

$q = 0.99 (t_2 - t_1)$

$= 79.3$ Btu/hr-ft ht.

Heat transfer coefficient

$$h = \frac{q}{A\,(t_g - t_w)}$$

where $q = 79.3$ Btu/hr-ft.ht.

$$A = \pi D_t Z = \pi (1/3)(1)$$

$$= 1/0.953$$

$$h = \frac{(0.953)(79.3)}{(7.42)}$$

$$= 10.2 \text{ Btu/hr - ft}^2 - {}^{o}F$$

Standard gas density

$$\rho_o = \frac{C_3\,(29.9 + p_1 - \Delta p/13.6)}{460 + t_o}$$

$$\text{where } C_3 = \frac{(492)(29.9)}{(28.97)(359)} = 1.33$$

$$\rho_o = \frac{1.33(29.9 + 0.7 - 26/13.6)}{530}$$

$$= 0.0764 \text{ \#/ft}^3$$

Air mass velocity

$$G = C_2\sqrt{(\rho_o)(\Delta p)}$$

where C_2 = 206, orifice Constant of the $^9/32"$ orifice

$$G = 206 \sqrt{(0.0764)(2.6)}$$

$$= 91.5 \text{ lbs/hr-ft}^2$$

CONCLUSIONS

The heat transfer coefficient between a bed of solid particles and its retaining wall increased sharply with mass velocity of entering air until the transition zone between the expanded and fluidized condition was reached. With still higher mass velocities the coefficient h increased less rapidly to a maximum and thereafter remained constant or decreased slightly.

Particle size of the solid exerted an effect upon the transfer coefficient h at constant mass velocity - the effect being in a direction inversely proportional to particle size.

For all particle sizes studied the coefficient h remained constant or decreased slightly above mass velocities of 400.

When fluidizing pulverized Utah coal in all sizes between 14 and 65 Tyler screen mesh, the coefficient h varied from a lower limit of 2 at low mass velocities to an upper limit of about 19 at a mass velocity of approximately 450.

Based on a comparison of Bureau of Mines data with the information reported here, it appears that the coefficient h continues to increase with decreasing particle size and it is also indicated that coefficients considerably higher than those reported here could be obtained when employing more dense solids than Utah coal.

RECOMMENDATIONS

It is recommended that further and more extensive work be carried out on the coefficient of heat transfer between the air stream and the tube wall. With the present available apparatus, the following factors which affect h may be examined: gas mass velocity, particle size and shape of materials of varying densities, and bed conditions.

The following apparatus changes are suggested:

1. A thermostat should be installed to regulate the temperature of the hot air entering the bed.

2. Should beds of particles smaller than 100 mesh be examined, provisions should be made for measuring air-mass velocities below 60 lbs/hr-sq ft., and the screens now in use should be replaced by finer ones.

Correlation of data by means of dimensionless equations should be attempted for the fluidized, partially expanded, and the static beds. The correlation if developed, would be of importance in showing whether the multifold decrease in the resistance to heat flow from the bed to the container walls is due to:

(1) The "scraping" effect of the particles in decreasing the gas film thickness.

(2) Actual contact between the solid particles with the wall thereby causing direct transfer of heat from one to the other.

(3) A combination of the above two effects where one or the other may be controlling depending on the experimental conditions.

This correlation, with the results of studies on other aspects of fluidized systems, would provide a better understanding of their mechanics.

BIBLIOGRAPHY

1. Luckenbach, Frederic, U. S. Patent 210,793. Dec. 10, 1878.

2. Canon, Frank A., U. S. Patent 1,355,105. Oct. 5, 1920.

3. Chappell, Fran. L., U. S. Patent 1,892,233. Dec. 27, 1932.

4. Odell, William W., U. S. Patent 1,984,380. Dec. 18, 1934.

5. Alexander, N. W. (Operating Manual for the Houston Research Catalytic Pilot Plant). Shell Oil Company, Houston Refinery Research Laboratory.

6. Carlsmith, L. E., Johnson, F. B., Ind. Eng. Chem., 37:451 (1945).

7. Kalbach, J. C., Chem. Eng., 54:1, 105-8 (1947).

8. Kalbach, J. C., ibid, 54:2, 136-8 (1947).

9. Lee, J. A., Chem. and Met. Eng., 52: 100-1, July (1945).

10. Murphree, E. V., Brown, C. I., Gohr, E. J., Jahnig, C. E., Martin, H. S., and Tyson, C. W., A.I.Ch.E. Trans. 41: 19, (1945).

11. Thomas, C. L., A. P. I. Proceedings 24: Sec. 3, 75-82 (1943).

12. Thomas, C. L., Hockstra, J. Ind. Eng. Chem., 37: 332-4 (1945).

13. Wickham, R. P., Petr. Refiner 24: 263-6, July (1945).

14. Brinn, M. S., Friedman S. J., Gluckert, F. A., Pigford, R. L., Ind. Chem. 40: 1050-61, June (1945).

15. Dietz, V. R., Robinson, H. E., Ind. Eng. Chem., 40: 1070-75, June (1948).

16. Leva, M., Ind. and Chem., 39: 857-62 (1947).

17. Leva, M., Grummer, M., Ind. Eng. Chem., 40: 415-419 (1948).

18. Leva, M., Weintraub, M., Grummer, M., Clark, E. L., Ind. Eng. Chem., 40: 747-52 (1948).

19. Furnas, J. J., U. S. Bureau of Mines, Bulletin 361, (1932).

20. Parent, J. D., Yagol, N., and Steiner, C. S., Chem. Eng. Progress, 43: 429-436, August (1947).

21. Johnstone, H. P., Pigford, R. L., Chapin, J. H., A.I.Ch.E. Trans., 37: 95-133 (1941).

22. Wilhelm, H. H., Kwauk, M., Chem. Eng. Progress 44: 3, 201 (1948).

23. Leva M., Grummer, M., Weintraub, M., Pollchik, M., Chem. Eng. Progress, 44: 511-520 (1948).

24. Leva, M., Grummer, M., Weintraub, M., Pollchik, M., Chem. Engr. Progress, 44: 619-626 (1948).

25. Frey, J. S., M. S. thesis in Chem. Eng., Oregon State College (1947).

26. Leva, M., Weintraub, M., and Grummer, M., "Heat Transmission Through Beds of Fine Particles", presented annual A.I.Ch.E. meeting, November, 1948 - New York to be published.

27. Kettenring, K. N., and Smith, J. M., "Heat and Mass Transfer in Fluidized Systems", presented annual A.I.Ch.E. meeting, November, 1948 - New York to be published.

28. Simon, R., Ph.D. thesis in Chem. Eng., Oregon State College (1948).

29. McAdams, "Heat Transmission". 2nd Edition, McGraw-Hill Book Co., (1940). 12p. Eqn. 10.

APPENDIX

FIGURE CAPTIONS

○ - Curve 1 - Run 15

◒ - Curve 2 - Run 14

◐ - Curve 3 - Run 13

⊙ - Curve 4 - Run 16

◑ - Curve 5 - Run 17

◓ - Curve 6 - Run 19

TABLE I

Run No.		Wt.(lbs.)	G	h	t_g
15 -	6	2	478	14.5	320
	7	2	625	16.2	298
	8	2	730	16.3	287
	9	2	545	14.8	309
	10	2	832	13.9	286
	11	2	1048	12.9	293
	12	2	943	13.9	305
	13	2	488	16.1	310
	14	2	678	15.9	312
	15	2	987	14.3	306
15-L-	1	2	292	8.94	273
	2	2	263	4.66	276
	3	2	219	1.28	292
	4	2	185	0.91	308
	5	2	159	1.26	292
	6	2	340	17.4	298
	7	2	387	15.1	289
14-a-	8	2	482	15.4	304
	9	2	562	16.8	303
	10	2	694	18.5	290
	11	2	830	16.9	287
	12	2	908	14.9	293
	13	2	1009	16.5	296
14-b-	14	2	485	16.0	303
	15	2	615	17.4	295
	16	2	523	17.2	306
	17	2	987	17.2	295
	18	2	863	16.5	315
	19	2	733	16.0	318
	20	2	817	16.6	315
	21	2	1010	18.1	303
	22	2	643	16.4	310
14-L-	1	2	430	15.5	278
	2	2	388	16.6	279
	3	2	340	17.2	280
	4	2	315	16.1	288
	5	2	274	15.9	291
	6	2	236	14.2	297
	7	2	185	11.5	297
	8	2.5	152	4.89	290

TABLE I (CONTINUED)

Run No.		Wt.(lbs.)	G	h	t_g
	9	2.5	208	11.5	307
	10	2.5	160	7.97	312
	11	2.5	128	2.69	303
13-a-	9	2	484	18.7	297
	10	2	562	19.6	300
	11	2	666	19.5	292
	12	2	746	18.5	285
	13	2	865	17.8	284
	14	2	986	17.1	285
	15	2	1014	17.4	290
	16	2	488	16.2	287
13-L-	1	2	436	15.3	303
	2	2	270	14.8	312
	3	2	197	14.2	333
	4	2	150	11.6	316
	5	2	124	8.7	304
	6	2	74	1.98	282
	7	2	314	15.2	315
	8	2	98.6	4.45	294
	9	2	59.6	1.34	298
	10	2	480	17.5	300
	11	2	388	17.0	303
16 -	6	2	472	18.4	304
	7	2	339	18.0	304
	8	2	641	17.7	304
	9	2	519	18.4	317
	10	2	767	17.0	314
16-M-	4	2	258	17.3	279
	5	2	179	14.2	278
	6	2	118.3	12.2	276
	7	2	91.5	10.2	270
	8	2	62.2	5.94	262
17 -	1	1.5	478	18.6	288
	2	1.5	390	19.5	318
	3	1.5	253	15.2	403
	5	2.5	182	14.7	258
	6	2.5	254	16.9	282
	7	2.5	306	17.5	289
	8	2.5	472	18.7	283
	9	2.5	540	18.4	285
	10	2.5	560	20.3	284

TABLE I (CONTINUED)

Run No.	Wt.(lbs.)	G	h	t_g
17-L- 1	2	307	15.9	260
2	2	250	16.2	267
3	2	182	16.8	276
4	2	106.6	13.2	271
5	2	59.5	7.53	261
6	2	145	13.8	299
8	2	72.7	9.37	290
9	2	298	16.6	280
10	2	132	14.0	304
11	2	207.5	16.5	281
12	2	380	17.9	287
19 - 1	2	244	16.9	272
2	2	185	16.9	298
3	2	119.5	15.2	302
4	2	75	13.2	299
5	2	48	11.0	282
6	2	304	19.2	293

TABLE II

PARTICLE SIZE DISTRIBUTION - WT.%

Tyler Mesh size	D		Run 15	Run 14	Run 13	Run 16	Run 17	Run 19
+ 14 - 10	+ 0.0464	- 0.0656	0.331					
+ 20 - 14	+ 0.0328	- 0,0464	95.322					
+ 28 - 20	+ 0.0232	- 0.0328	4.016	92.152	1.213			
+ 35 - 28	+ 0.0164	- 0.0232	0.221	7.517	88.457	0.679		
+ 48 - 35	+ 0.0116	- 0.0164	0.	0.132	9.497	83.593	0.285	
+ 65 - 48	+ 0.0082	- 0.0116	0.055	0.099	0.624	15.250	80.066	
+ 100 - 65	+ 0.0058	- 0.0082	0.055	0.066	0.069	0.358	18.510	39.982
+ 150 - 100	+ 0.0041	- 0.0058		0.	0.	0.040	0.854	34.974
+ 200 - 150	+ 0.0029	- 0.0041		0.033	0.138	0.040	0.047	18.014
+ 270 - 200	+ 0.00205	- 0.0029				0.040	0.	3.515
- 270		- 0.00205					0.237	3.515
	D_p =	(D X F)	0.0463	0.0269	0.0190	0.0131	0.009	0.00520

NOMENCLATURE

A = Mantle area of tube (ft)

D = Screen opening (inches)

D_p = Effective particle diameter (inches)

D_t = Experimental tube diameter (ft)

F = Wt. fraction

G = Mass velocity based on cross section of empty tube (lbs/hr-ft^2)

h = Heat transfer coefficient between the bed and the tube wall. (Btu/hr-ft^2 - ^oF)

K = Average thermal conductivity of insulation (Btu/hr-ft-^oF)

Δp = pressure drop across orifice (inches of H_2O)

p_1 = Pressure, upstream of orifice (inches of H_2O)

q = Heat flux (Btu/hr)

r_1 = $(i = 1,2)$ radial distance from the tube center (inches)

t_1 = $(i = 1,2)$ temperature of insulation at r_1 (^oF)

t_g = temperature, gas and bed (^oF)

t_w = temperature, tube wall (^oF)

t_o = temperature, air at orifice (^oF)

Z = Length of tube considered (ft)

z = Height above screen (inches)

ρ_o = Standard gas density (lbs/ft^3)

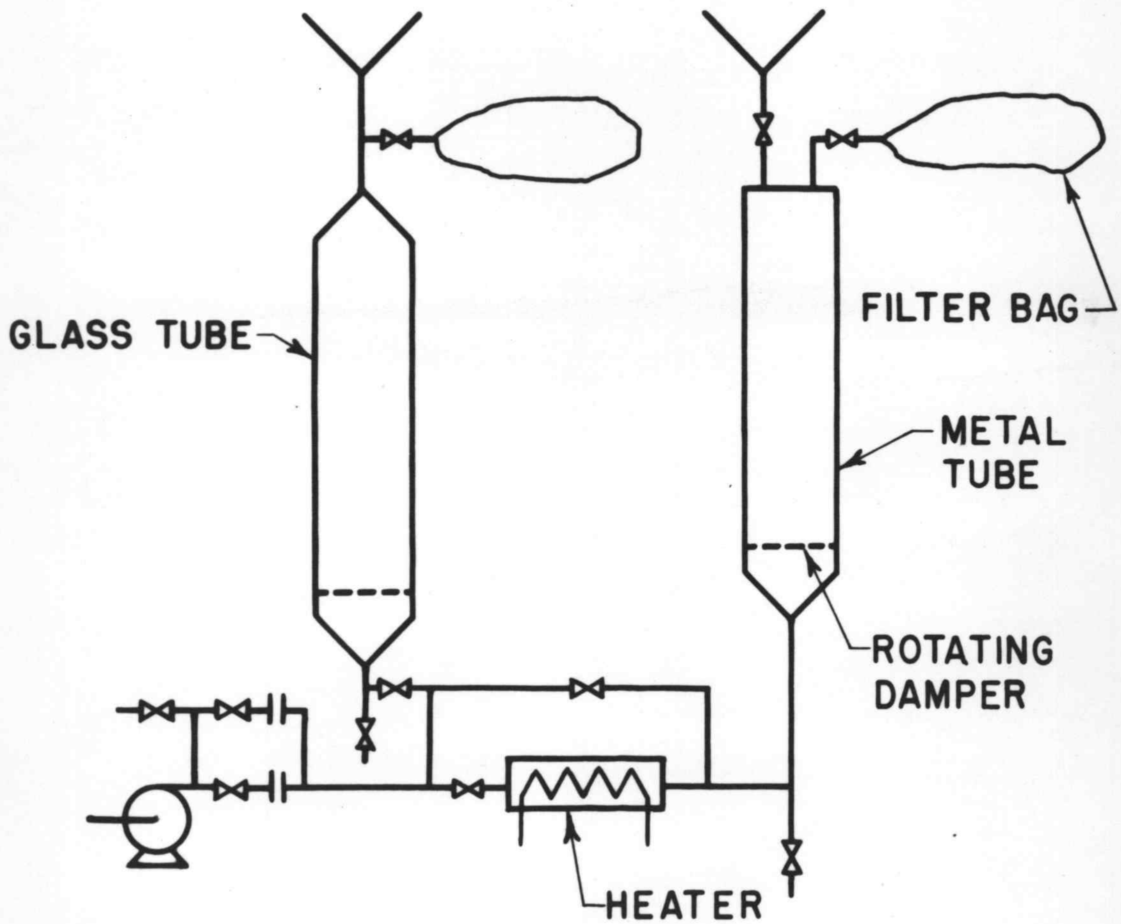

FIG. 1 - PROCESS FLOW DIAGRAM

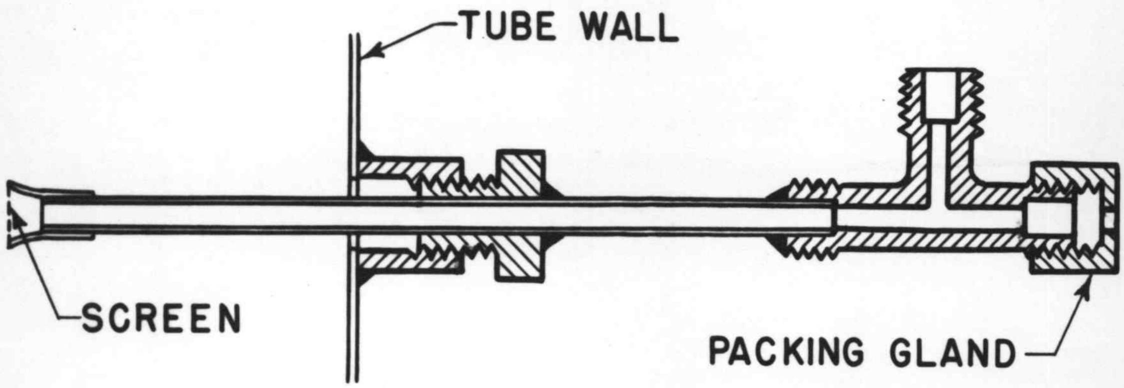

FIG. 2 - SUCTION THERMOCOUPLE WELL

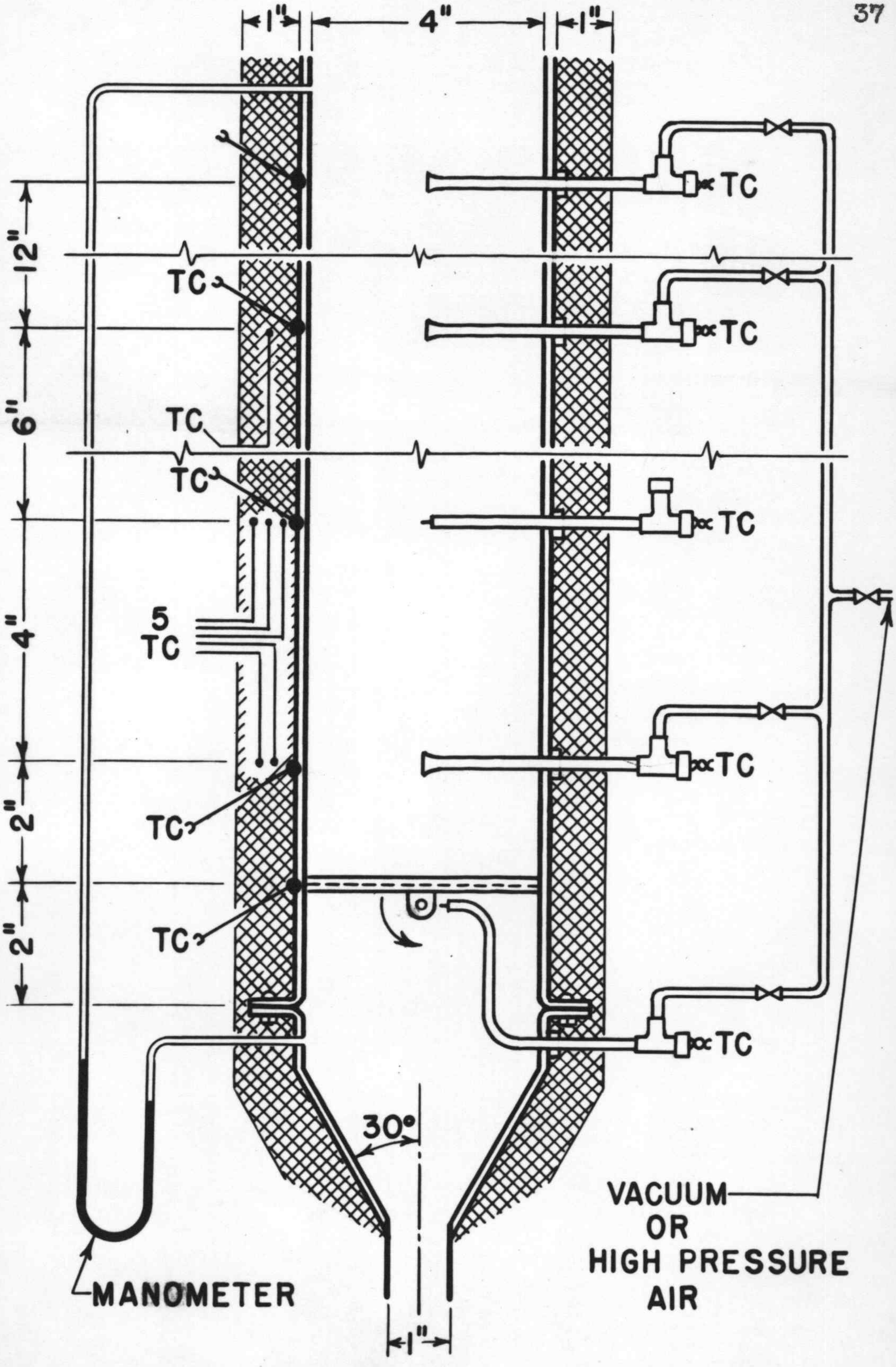

FIG. 3 — REACTOR TUBE AND FITTINGS

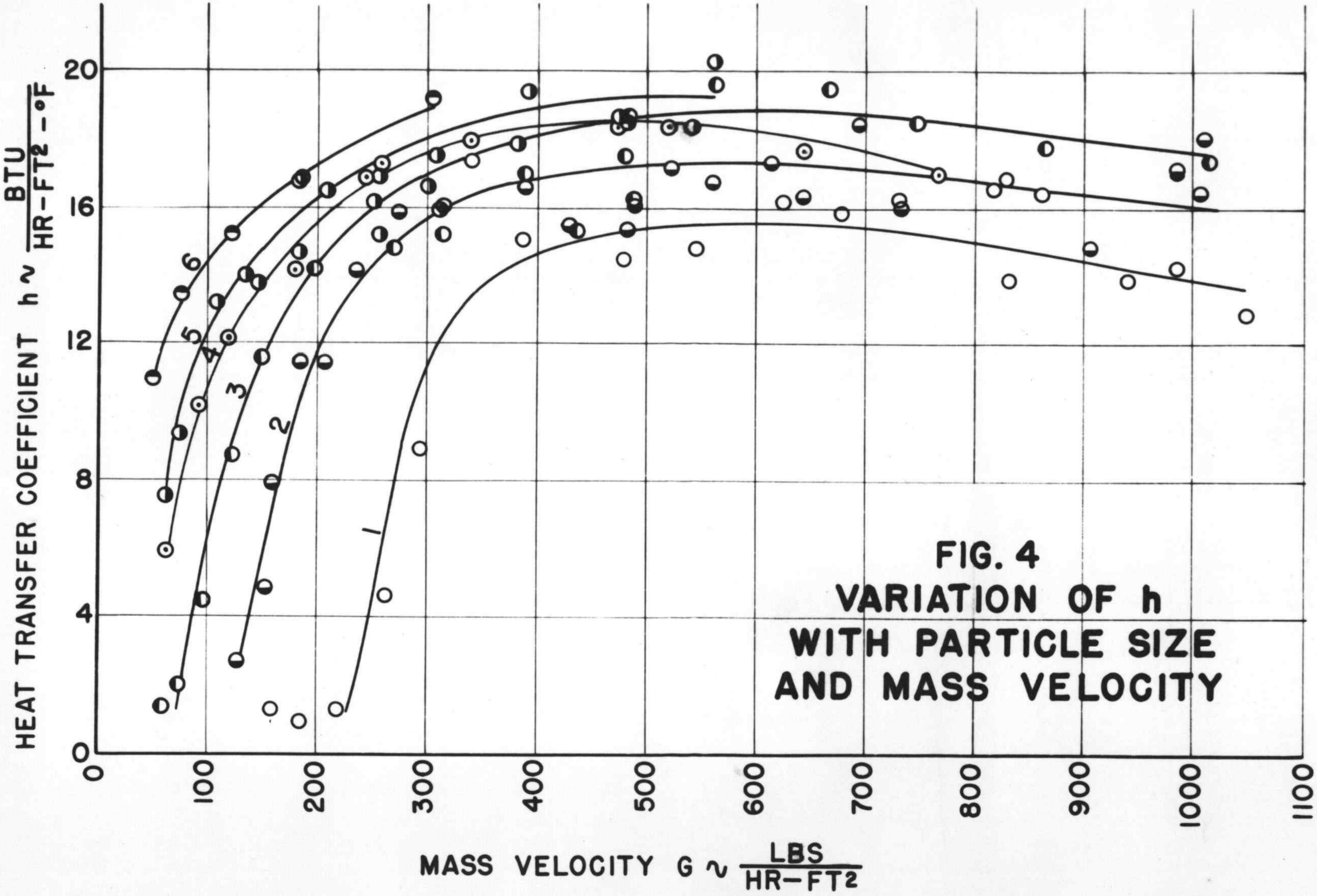

FIG. 4
VARIATION OF h
WITH PARTICLE SIZE
AND MASS VELOCITY

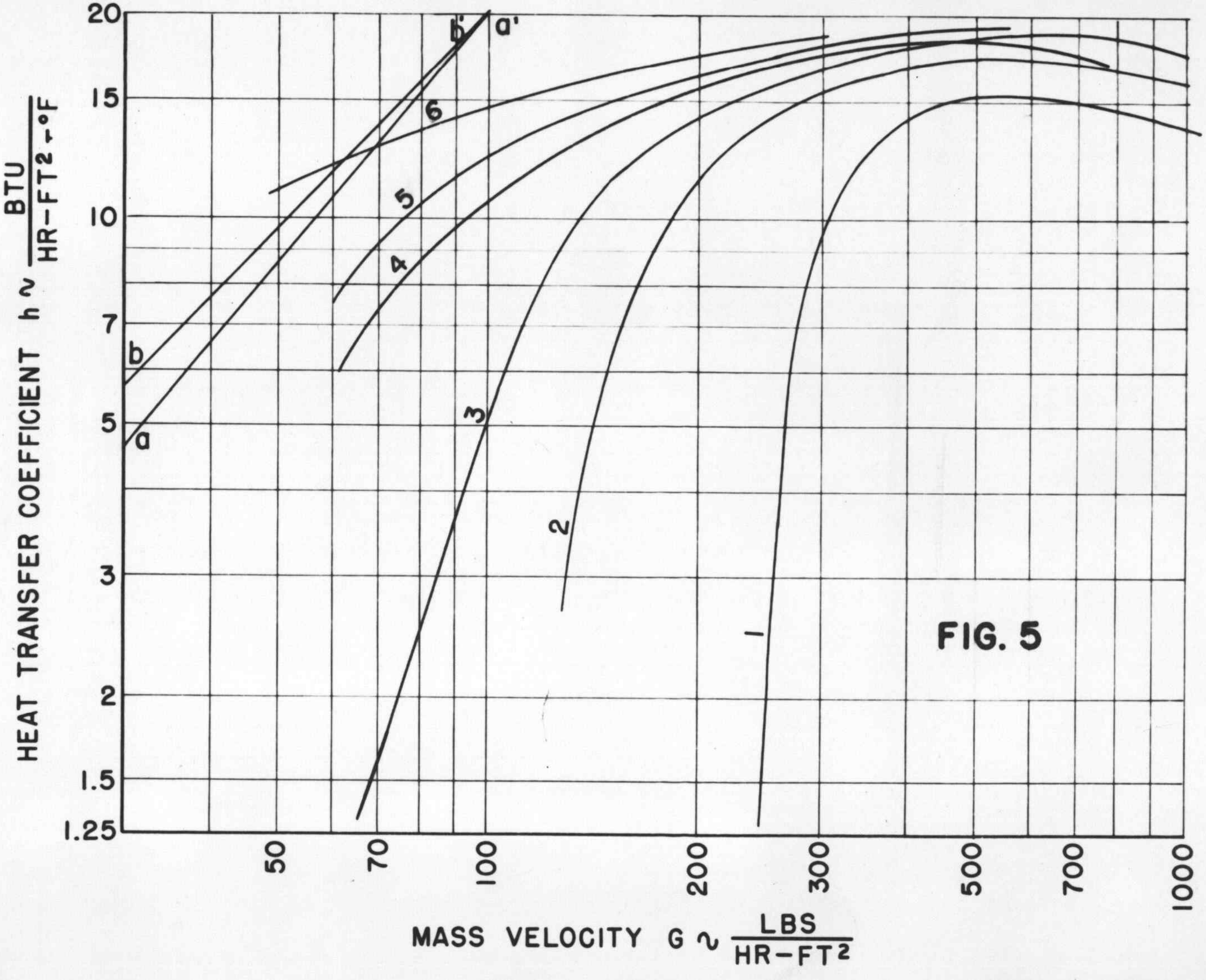
HEAT TRANSFER COEFFICIENT h ~ BTU / HR-FT2-°F
20
15
10
7
5
3
2
1.5
1.25
50
70
100
200
300
500
700
1000
MASS VELOCITY G ~ LBS / HR-FT2
a
b
a'
b'
1
2
3
4
5
6

FIG. 5