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HEAT TRANSFER, PRESSURE DROP, VISUAL OBSERVATION, TEST DATA FOR AMMONIA EVAPORATING INSIDE PIPES

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INTRODUCTION

For rational design of evaporators and for developing correlations for the same, it is necessary to have detailed test data. The test data should give heat transfer and pressure drop rates as a function of pertinent parameters (such as mass flow rate, vapor quality, pressure, heat flux). Information on visual observations (for example flow patterns, extent of pipe surface wetted, behavior of oil) are also very useful.

Many such studies have been carried out for halocarbon refrigerants and a wide variety of experimental data are available in published papers, reports, doctorate theses, etc. The situation for ammonia evaporators is very different. Much of the experimental work was done long ago when the parameters affecting heat transfer and pressure drop were not known fully. Consequently, the published data from those studies do not contain important parameters such as mass flow rate and vapor quality, thus making it impossible to analyze and generalize them. Examples of such work are those of Cleis[1] and Schwind[2]. To the author's knowledge, the only analyzable data for ammonia evaporators containing oil, which are accesible to the general public, are those shown graphically in a paper by Shah[3]. Some analyzable data for oil-free ammonia are to be found in a paper by Noel[4] but their range is confined to subcooled boiling region.

The abundantly available test data for halocarbon refrigerants may not be applicable to ammonia evaporators because of the different behavior of oil. Oil dissolves in halocarbons while it is virtually insoluble in ammonia. As shown by Shah's experiments[3,5], insulating oil films form in ammonia evaporators which drastically reduce the heat transfer coefficients. Thus the need for test data on ammonia evaporators is quite acute. This has prompted the author to write this paper whose main objective is to make a large amount of test data available to the general public.

Data from two independent experimental studies are presented. The first study was done by the present author and the results were presented in two earlier papers[3,5]. However, these papers showed only a few data points and in graphical form which is not very convenient for accurate calculations. Here, data from 20 representative test runs are presented in tabular form and contain local heat transfer and pressure drop rates, and visual observations.

The other experimental study was carried out by Van Maale and Cosijn[6]. Ref 6 did not provide any analyzable data. Here the data taken from their original test records are presented in graphical form. These data give mean heat transfer coefficients and total pressure drops.

Extensive data analysis is not within the scope of this paper. However, the analyses presented in earlier papers[3,5] are critically reviewed. Suggestions for practical designs are made taking into account all available information. Some

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suggestions for further research are also given.

EXPERIMENTAL STUDY BY SHAH

Test data from 20 representative runs from Shah's experiments are given in Table 1. Information regarding experimental apparatus and data reduction techniques are given in the following.

Experimental Apparatus

The experimental apparatus has been described in Ref 3 and 5 and is shown here schematically in Fig. 1. The 140 m long evaporator was fabricated from commercial grade steel pipe of 26.2 mm ID, 33 mm OD. The length was divided into 12 sections , the first being 10.6 m long while the rest were 11.77 m long. Pressure and temperature measurements were done at 13 stations along the length of the evaporator. Sight glasses were provided at the last 12 stations. There was no sight glass at the first station(station 0). Sight glasses at stations 4, 10, 11, and 12 were later removed due to frequent breakages.

The ammonia compressors were of conventional reciprocating designs with standard oil separators in discharge lines. No devices for measuring or controlling the amount of oil in circulation were provided. The evaporator was frequently drained of oil. In no case were any extraordinary amounts of oil discovered. Liquid ammonia was pumped into the evaporator using one of the two pumps provided. Ammonia flow rate was measured with orifice plates constructed to DIN standards.

At each measuring station, a static pressure tap, a thermocouple inserted into the pipe, and seven thermocouples soldered to half the circumference of the pipe were provided. The pressure taps were connected to a set of U-tube manometers containing mercury and parrafin oil. The pressure drop in each section was read from the appropriate manometer depending on the magnitude of pressure drop. The pressure at the beginning, the middle, and the end of the coil was measured directly with a manometer open to the atmosphere. The total pressure drop across the whole evaporator was also measured directly.

Heating was done by three cables running along the length of the evaporator and located at 120 degrees from one another. Direct current was supplied to the cables. Sevety milimeter insulation covered the evaporator and was finished with a vapor barrier.

Data Reduction

Heat Transfer Arithmetic mean of the voltage shown by the seven thermocouples at each station was calculated. The mean voltages for the 13 stations were plotted on a graph along with the voltages shown by the thermocouples inserted into the ammonia stream. Smoothed out curves were drawn. External wall temperatures were calculated according to the voltages read from these smoothed out curves. Inner pipe wall temperature was then estimated by correcting for the temperature drop through the wall. Heat transfer coefficients were based on the saturation temperature in the boiling region and on the bulk temperature in the nonboiling region.

Pressure Drop The absolute pressures listed in Table 1 are the actual pressures calculated from measurements without curve smoothing. The total pressure drop in the evaporator was divided in the ratio of the measured sectional pressure drops in the sections. The absolute pressure at each station was calculated by successively subtracting the sectional pressure drops from the inlet pressure.

The pressure drop rates at the various stations were calculated from the absolute pressures at these stations by a computer subroutine as follows:

(a) For stations 1 to 11

$$\left(\frac{\Delta P}{\Delta L}\right)_i = \frac{P_{i-1} - P_{i+1}}{L_{i+1} - L_{i-1}} = \frac{P_{i-1} - P_{i+1}}{23.54} \qquad (1)$$

where

P_i = Static pressure at station i

L_i = Length of the test evaporator coil at station i measured from station 0

(b) For station 0

$$\left(\frac{\Delta P}{\Delta L}\right)_i = \frac{P_0 - P_1}{L_1 - L_0} = \frac{P_0 - P_1}{10.6} \qquad (2)$$

(c) For station 12

$$\left(\frac{\Delta P}{\Delta L}\right)_{12} = 2\left(\frac{P_{11} - P_{12}}{11.77}\right) - \left(\frac{\Delta P}{\Delta L}\right)_{11} \qquad (3)$$

Essentially, for stations 0 and 12 first order correct finite difference analog has been used while for the others, second order correct analog has been used.

Some limitations of the above calculation procedure must be pointed out. Eq 1 will not be accurate if boiling or dryout starts between stations (i-1) and (i+1). Eq 2 is inaccurate if boiling starts between station 0 and 1. Those using these data for testing of correlations, etc. are advised to correct the tabulated pressure drop rate values in the transition regions by graphical means.

Classification of Flow Patterns

The definitions of flow patterns used in Table 1 are as follows:

Wavy - Liquid separated on the bottom with a wavy interface. Occasional large waves or slugs may occur, but the top remains dry most of the time.

Slug - An extension of the wavy flow pattern. The frequency of slugs is greater so that the top remains wet most of the time.

Crescent - Basically annular but the liquid film is noticeably thicker at the bottom. The top 1/8 may occasionally be dry.

Semi-Crescent - Basically crescent but more than the top 1/8 remains dry at all times.

Annular - An apparently uniform liquid layer covers the entire pipe surface.

Semi-Annular - Basically annular but the top of the tube is not wet.

Accuracy and Utility of Data

As was pointed out in Ref 5, the wall temperature measurements in the boiling region showed considerable scatter and hence the reported values of boiling heat transfer coefficients should be regarded as approximate. However, the accuracy of these data has been confirmed by practical application. Peacock[7], to whom the author had made these data available, reported that these were used to rate a series of shell and tube coolers(ammonia inside the tubes) and that the ratings were later confirmed by full scale tests.

The pressure drop data presented are essentially without any curve smoothing and the author has no reason to doubt their accuracy. As both heat transfer and pressure drop data have been reported in the form of local values, they are well suited for developing and evaluating correlations.

The visual observations are of special value as no other visual study on ammonia evaporators has been reported in literature. The accuracy of the flow pattern data has been confirmed by the general agreement with the Baker correlation[12] as was reported in Ref 3.

A serious limitation of these data is that the quantity of oil in circulation was not measured or controlled. However, the test evaporator and conditions correspond closely to practical conditions and the quantity of oil in circulation would have been representative of the field conditions.

EXPERIMENTAL STUDY OF VAN MAALE AND COSIJN

Data from the tests by Van Maale and Cosijn[6] are shown in Fig. 2 to 9.

Test Apparatus

The test apparatus has been fully described in Ref 6. Only a brief description is given here.

The evaporator was composed of five coils connected in parallel, made of steel tubes 15 mm ID, 18 mm OD. Each coil had a total length of 18.9 m, bent into 12 horizontal passes, one above the other. The evaporator was enclosed in a well-insulated hermetically-closed vessel whose lower part was filled with R-11 liquid. Electric heaters were immersed in R-11 liquid. Electric heaters evaporated R-11 which then condensed on the surface of ammonia evaporator above, thus providing the heat flux for the evaporation of ammonia.

A conventional reciprocating compressor with standard oil separator was used. A pump was used to circulate ammonia through the evaporator. Total pressure drop through the evaporator was measured with a manometer. Evaporator wall temperature was not measured. Temperatures of ammonia and R-11 were measured with thermocouples

Data Reduction

All data reduction was done by the original authors and not by the present author. Their procedure is explained here.

Dividing the electric heat input to R-11 by the internal surface area of the coil, heat flux was known. This heat flux divided by the difference in temperature between ammonia and R-11 gave the mean overall heat transfer coefficient. The heat transfer coefficient of condensing R-11 was calculated by the Nusselt equation. Thus, knowing the overall heat transfer coefficient, outside heat transfer coefficient, and the thermal resistance of pipe wall, the mean heat transfer coefficient of ammonia was calculated in the usual way.

The recirculation number n is defined as the ratio of the actual ammonia flow rate to the flow required with saturated ammonia liquid for complete evaporation at that heat flux. Mathematically,

$$n = \frac{W\ h_{fg}}{3600\ Q} \tag{4}$$

where

W = Mass flow rate of ammonia, kg/hr
h_{fg} = Latent heat of ammonia, J/kg
Q = Total heat added to evaporator

In the case that ammonia liquid entering the evaporator is saturated, n is simply the reciprocal of exit vapor quality. With increasing pressure drop through the evaporator, entering liquid is more and more subcooled and then the exit vapor quality becomes progressively lower than the reciprocal of the recirculation number.

An important point to note is that the mass flow rate in Fig. 2 to 9 is that through all the 5 coils of the test evaporator. Hence the flow through each coil will be 1/5 of the total mass flow rate, assuming that the distribution is uniform.

Accuracy and Utility of Data

The heat transfer coefficients reported in Fig. 2 to 9 should be regarded as approximate because the pipe wall temperature was not measured. The heat transfer coefficients on ammonia and R-11 sides were comparable in magnitude. Hence errors in calculated values of condensing R-11 heat transfer coefficients would have caused comparable errors in the estimation of ammonia side coefficients. It should also be noted that the reported heat transfer coefficients include both boiling and non-boiling modes of heat transfer.

The pressure drop data are more accurate but very difficult to compare with any predictive technique. The total pressure drop includes losses due to friction, accelaration, change of elevation, as well as the effect of bends. The contribution of bends is particularly difficult to evaluate while it could be a significant part of the total pressure drop. The losses due to accelaration and elevation are also quite significant but difficult to estimate accurately, primarily because of lack of reliable information on void fractions. Hence these data by themselves are not adequate for testing the accuracy of any pressure drop correlation.

Finally, as in the tests by Shah, the quantity of oil in circulation is unknown and would have varied from test to test. As it appears that oil has a significant effect, at least on heat transfer, it makes it difficult to reach reliable conclusions regarding the accuracy of any correlation.

REVIEW OF ANALYSES IN REF 3 AND 5

The data presented here were analyzed in Ref 3 and 5 and on their basis, certain conclusions were drawn and certain recommendations for practical designs were made. Those are briefly reviewed here.

Single-Phase Pressure Drop

In Ref 5, single-phase liquid-flow friction factors for ammonia temperatures between 0 and -15 C were found to be consistently lower than the predictions of the Moody chart for commercial steel pipes of this size. Instead, they were found to approximate to 0.018. Furthermore, the two-phase drop data showed excellent agreement with the Lockhart-Martinelli correlation when a single-phase friction factor of 0.018 was used. As no other measurements on ammonia-oil flow were available, the author considered the possibility that oil may have a drag reducing effect similar to that of certain polymers on water.

While this hypothesis cannot be disproved on the basis of data in hand, other explanations are possible. As shown in Fig. 10, the range of data covered Reynolds numbers from $3x10^4$ to $1.6x10^5$. In this range, the data are fairly well satisfied with smooth pipe friction factors. Hence a possible explanation is that as long as oil remains fluid, its effect is to smooth out the pipe surface by filling out the cavities on the pipe surface. It must be stressed that this hypothesis is also purely speculative. Only through experiments with pure and oil-containing ammonia on the same test section can this question be satisfactorily resolved.

For ammonia temperatures between -22 to -37 C, friction factors were found to vary widely. Values ranged from those for smooth pipes to twice as high as for rough pipe. In the light of visual observations, it was concluded that at temperatures below the set point of oil, the possibility exists for oil forming thick semi-solid films around the pipe circumference which cause excessive drag. Such oil films are likely to form when oil content is more than usual and the pipe wall reaches low temperature without adequate liquid circulation. Table 1 contains several test runs in which thick oil films were seen.

Two-Phase Pressure Drop

As mentioned earlier, the predictions of the Lockhart-Martinelli correlation[8] were found to be too high if rough pipe friction factors were used. With a constant friction factor of 0.018, most of the data were correlated to within 30%. This was one of the reasons which had led the author to consider the hypothesis that oil mixed with ammonia has some drag reducing properties.

There is a strong possibility that the discrepency is due to flow pattern effects. At lower mass flow rates and in larger diameter pipes, considering horizontal pipes, over-prediction of pressure drop by the Lockhart-Martinelli method has been reported by several researchers including Baker[12] and Scheideman et al[13]. According to Baker[12], stratified flow occurs under such conditions. In stratified flow, the upper part of the pipe circumference is not in contact with liquid. Because of the lower viscosity of vapor compared to liquid, one would expect the pressure drop to be lower in stratified flow compared to a flow pattern in which the entire surface is wetted by liquid, other things being the same. Similarly, the heat transfer coefficient will be lower due to the lower thermal conductivity of vapor. Shah[9] found that at the same vapor quality and fluid properties, the boiling heat transfer coefficient increases with all liquid Froude number Fr_L until the latter reaches a value of 0.04. Apparently, at this value of Froude number, the entire pipe circumfernce is wetted, while at lower values, part of the circumference is dry.

Hence it appears possible that Fr_L may also be suitable for improving the Lockhart-Martinelli correlation for the effect of partly dry surfaces. To investigate this possibility, some data for adiabatic flow of ammonia vapor-liquid mixtures from tests by Chaddock et al.[10] were analyzed. The results are shown in Fig. 11 and appear to be encouraging. However, the suitability for Fr_L for this purpose can be fully established only after analyzing a large amount of varied data.

Considering all the factors discussed in the foregoing, one feels justified in concluding that this discrepency between Shah's data and the Lockhart-Martinelli correlation is probably due to flow pattern effect and not due to any drag reducing properties of oil. The fact that Shah's data were well correlated using a constant friction factor of 0.018 must be regarded as a mere coincidence. Either the Martinelli correlation should be improved by including flow pattern parameters or some other correlation must be used. Chaddock et al.[10] found best agreement with the Hughmark correlation[11].

Single-Phase Heat Transfer

Single-phase heat transfer coefficients were found to be invariably lower than the predictions of standard correlations such as the Dittus-Boelter equation. The conclusion was reached that oil films are formed on the pipe circumference which cause additional resistance to heat transfer. The thickness of these oil films was calculated by comparing the measured heat transfer coefficients with the predictions of the Dittus-Boelter equation. The calculated oil film thickness was found to decrease with increasing liquid ammonia Reynolds number and was correlated by the following dimensionless equation:

$$\delta/D = 0.028/Re_L^{0.23} \qquad (5)$$

where

δ = Thickness of oil film
D = Diameter of pipe

The measured heat transfer coefficients were correlated by the equation:

$$Nu = 0.1825\ Re_L^{0.509}\ Pr_l^{0.4} \qquad (6)$$

The data from which Eq 5 and 6 were derived covered a Reynolds number range of 5×10^3 to 1.6×10^5. Furthermore, these equations are based on only one series of tests on one pipe diameter. Hence their validity is not fully established until other data are available and in agreement.

Two-Phase Heat Transfer

In Ref 5, a new correlation for estimating the boiling heat transfer was presented. Two dimensionless parameters were defined:

$$\psi = h_{TP}/h_l \qquad (7)$$

$$Y = \left(\frac{1}{x} - 1\right)^{0.8} (\mu_g/\mu_l)^{0.4} (C_{pl}/C_{pg})^{0.4} (k_l/k_g)^{0.6} \qquad (7)$$

In Eq 7, h_l is the superficial heat transfer coefficient of liquid phase and is calculated with Eq 6(replacing Re_L with Re_l) for oil-containing ammonia and by the Dittus-Boelter equation for halocarbon refrigerants. When the data were plotted on Y-ψ graph, it was found that the mean curve through the ammonia data was lower than the mean curve through halocarbon data. Due to the large scatter in wall temperature measurements, the author concluded that the ammonia data were erroneous and the mean curve through halocarbon data represented the true correlation.

Subsequent investigations have shown that a general correlation is not obtainable in terms of Y and ψ(see Ref 15). Hence the use of this correlation should be confined only to oil containing ammonia and the mean through the ammonia data regarded as the correlating line.

In Ref 9, Shah presented a general correlation called the CHART for heat transfer during boiling of saturated fluids in pipes. It was verified by comparison with a wide variety of data on R-11, R-12, R-22, R-113, water and cyclohexane. At very low and negative qualities, it has also been verified for oil-free ammonia, various alcohols, and aqueous solutions of potassium carbonate, as described in Ref 14. Hence, one may feel confident that this correlation will predict the heat transfer coefficients for pure fluids fairly accurately. As shown in Ref 9, predictions of the CHART correlation were far higher than Shah's measurements as tabulated here. Furthermore, the discrepancy could be reasonably well corrected on the basis of oil film thickness calculated by Eq 5. This provides confirmation of the hypothesis that insulating oil films are formed in ammonia evaporators.

Flow pattern data showed good agreement with the Baker correlation[12] and provided more details of the regime called annular by Baker, which in the terminology used here includes annular, crescent, semi-crescent, and semi-annular regimes. The data were also well correlated in terms of Froude number and vapor volume fraction. As the Baker correlation has been tested over a wide range of varied data, it is suggested that this correlation as shown in Fig. 7 of Ref 3 be used for flow pattern predictions.

DESIGN RECOMMENDATIONS

From the foregoing discussions, it is apparent that the available information is not sufficient for designing ammonia evaporators with complete confidence. Much more remains to be learned through further experimentation and data analysis. In the meantime, designs have have to be done on the basis of available knowledge. Thus, the author's recommendations for oil containing ammonia evaporators are given in the following.

For single-phase pressure drop, use friction factors according to Moody chart. At temperatures below the set point of oil, friction factors may actually be higher, and above the set point of oil, friction factors may be lower. However, non-boiling length is generally a small part of the total evaporator length and the liquid pressure drop only a small fraction of total pressure drop. Therefore, errors due to deviations in single-phase friction factors are generally of little consequence.

For two-phase pressure drop, Hughmark correlation is recommended, based on the conclusion of Chaddock et al[10].

For single-phase heat transfer, use Eq 6 down to a Reynolds number of about 6,000.

For two-phase heat transfer, two alternative methods are available, neither fully satisfactory. First is the use of Y-ψ curve shown in Fig. 12, with h_l calculated with Eq 6. The other is to use the CHART correlation[9] to estimate the heat transfer coefficient of pure ammonia and correct it for the resistance of oil film thickness calculated with Eq 5. This will give low heat transfer coefficients at higher vapor qualities.

Most practical systems use compressors which add some oil to ammonia. In systems such as those using oil-free compressors, ammonia is completely free of oil. Heat transfer and pressure drop can then be calculated using generalized correlations for pure fluids.

SUGGESTIONS FOR FURTHER RESEARCH

As has hopefully been made clear in the foregoing, much remains to be leared about ammonia evaporators. To finally resolve some of the outstanding questions, several independent experimental studies are needed. This will take much time. In the meantime, much could be learned by analyzing the available data thoroughly. The available data are those presented here and the pressure drop data of Chaddock et al.[10]. Author's suggestions for further research are given in the following.

Data Analysis

Boiling heat transfer data would be analyzed to estimate the thickness of oil films in the boiling region. With increasing vapor velocity, the thickness of oil films should be substantially lower than that given by Eq 5 which is based on single-phase measurements. Heat transfer coefficients for pure ammonia would be calculated using some reliable general correlation(such as that by Shah[9]). The thickness of oil film to account for the difference between measured and predicted heat transfer coefficients would be calculated. An attempt should then be made to correlate the calculated thicknesses in terms of a suitably defined two-phase Reynolds number. One possible definition could be the sum of superficial liquid and vapor phase Reynolds numbers.

$$Re_{TP} = Re_l + Re_g \tag{9}$$

It is further suggested that Shah's data for two-phase pressure drop be compared with the Hughmark correlation. Attempts for improving the Lockhart-Martinelli correlation by introducing the Froude number are also suggested.

Experimental Studies

Heat transfer, pressure drop, and visual studies would be done in a loop in which quantity of oil can be controlled and measured. Ammonia circulating through the test section should be condensed back to liquid using an indirect refrigeration system so that tests with absolutely oil-free ammonia can be made. Use of a compressor to circulate the refrigerant directly would inevitably add some oil to ammonia, making oil-free tests impossible. Tests should be done on several pipe diameters, horizontal and vertical orientations, and a wide range of operating parameters. Tests on evaporators made of glass tubes could be of much help in studying the behavior of oil. Attempts to measure the thickness of oil films directly by optical methods are also desirable.

CONCLUDING REMARKS

The main purpose of this paper was to provide analyzable data on ammonia evaporators containing oil. It is realized that these data are not of the highest accuracy. But in the absence of virtually any other published data, it is believed that these will be valuable to researchers as well as practical engineers as a starting point for further research and for directly rating and developing evaporators.

Through the critical review of analyses and recommendations in the author's earlier papers, it is hoped that their deficiencies have been corrected. While a thoroughly satisfactory and reliable design approach can be developed only through further research, it is believed that the design recommendations provided here give the most reasonable interpretation of the available information and would be helpful in practical design.

NOMENCLATURE

A	Cross-sectional area of pipe
C_{pl}	Specific heat of liquid
C_{pg}	Specific heat of vapor at constant pressure
D	Internal diameter of pipe
Fr_L	All liquid Froude number = $W^2/(\rho_l^2 A^2 gD)$
g	Accelaration due to gravity
h	Heat transfer coefficient
h_{TP}	Two-phase heat transfer coefficient
h_l	Superficial heat transfer coefficient of liquid phase
h_g	Superficial heat transfer coefficient of vapor phase
h_{fg}	Latent heat of vaporization
k	Thermal conductivity
L	Length
Nu	Nusselt number = hD/k
n	Recirculation number, defined by Eq 4
P	Absolute pressure at a point
ΔP	Pressure drop
$\Delta P/\Delta L$	Pressure drop rate
Pr_l	Prandtl number of liquid ammonia = $C_{pl}\ \mu_l/k_l$
q	Heat flux
Re	Reynolds number
Re_l	Superficial Reynolds number of liquid phase
Re_g	Superficial Reynolds number of vapor phase
Re_L	Reynolds number for the total mass flowing as liquid
T_{OUT}	Temperature of ammonia at coil outlet
Y	Heat transfer correlating parameter defined by Eq 8
W	Mass Flow Rate
μ	Dynamic viscosity
ρ	Density
ψ	h_{TP}/h_l

Subscripts

l	For liquid
g	For vapor

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TABLE 1

Data of Shah for ammonia evaporating in a 26.2 mm ID pipe

TEST NO.	W kg/h	q W/m^2	STATION NO.	ABS. PRESS. 10^5 N/m^2	AMMONIA TEMP. C SAT.	AMMONIA TEMP. C BULK	h W/m^{2}deg C	(ΔP/ΔL) N/m^3	VAPOR QUALITY	FLOW PATTERN & OTHER VISUAL OBSERVATIONS
19.2	79.0	2312	0	2.3704	-14.93	-14.63	524	4.9	0.000	*
			1	2.3700	-14.92	-14.55	506	7.8	0.0725	Wavy. 1/3 wet
			2	2.3687	-14.94	-14.44	522	15.7	0.1520	" " "
			3	2.3663	-14.99	-14.49	542	45.1	0.2320	Wavy. 1/4 wet
			4	2.3581	-15.07	-14.60	571	101.0	0.3160	" " "
			5	2.3426	-15.18	-14.76	607	124.5	0.3900	Wavy. 1/2 wet
			6	2.3288	-15.33	-14.98	658	134.3	0.4700	" " "
			7	2.3109	-15.53	-15.25	728	171.6	0.5500	Wavy. 2/3 wet
			8	2.2883	-15.77	-15.57	834	214.8	0.6290	Semi-crescent. 2/3 wet
			9	2.2604	-16.06	-15.90	981	262.8	0.7100	" " " "
			10	2.2265	-16.39	-16.28		312.8	0.7750	" " 3/4 wet
			11	2.1869	-16.79	-16.71	956	337.3	0.8700	" " " "
			12	2.1560	-17.25	-17.17	767	338.3	0.9500	Some oil at bottom. Incomplete annular film at top and bottom.
55.0	112.5	2520	0	1.7248	-22.17	-27.07	687	22.5	0.0000	*
			1	1.7224	-22.23	-21.51	723	24.5	0.0370	Wavy
			2	1.7193	-22.28	-21.62	736	41.2	0.0963	"
			3	1.7126	-22.35	-21.84	783	79.4	0.1560	Slug
			4	1.7007	-22.47	-22.11	836	110.8	0.2160	Semi-crescent
			5	1.6866	-22.67	-22.50	924	161.8	0.2750	" "
			6	1.6626	-23.00	-23.00	1030	257.9	0.3360	Crescent
			7	1.6260	-23.47	-23.55	1167	356.0	0.3960	Annular
			8	1.5789	-24.15	-24.21	1326	482.5	0.4620	Annular
			9	1.5125	-25.07	-25.18	1510	649.2	0.5170	"
			10	1.4261	-26.31	-26.37	1721	829.6	0.5800	*
			11	1.3170	-27.93	-27.94	1960	988.5	0.6400	*
			12	1.1933	-30.07	-29.84	2169	1116.0	0.7050	*
56.2	362.0	2379	0	2.0328	-18.42	-3094	860	29.4	0.0000	*
			1	2.0297	-18.56	-26.04	970	23.0	0.0000	Liquid only
			2	2.0276	-18.64	-21.07	932	28.4	0.0000	" "
			3	2.0230	-18.68	-17.39	962	88.2	0.0093	Slug. 1/2 to 2/3 full
			4	2.0067	-18.76	-18.34	962	158.9	0.0270	*
			5	1.9856	-18.94	-18.61	988	211.8	0.0454	Slug
			6	1.9659	-19.27	-19.16	1013	326.5	0.0640	Slug turning to crescent
			7	1.9088	-19.87	-19.7	1031	492.3	0.0835	Crescent
			8	1.8411	-20.78	-20.52	1051	746.3	0.1030	"
			9	1.7332	-22.11	-21.89	1137	1031.6	0.1250	Crescent. Near to annular
			10	1.5983	-24.01	-24.1	1436	1432.7	0.1490	*
			11	1.3959	-26.68	-27.01	1666	1833.8	0.1740	*
			12	1.1665	-30.47	-30.23	1365	2063.3	0.2020	*
57.2	111.0	2346	0	1.6729	-22.85	-26.18	859	31.4	0.0000	*
			1	1.6696	-22.91	-20.93	570	31.4	0.0394	Slug. 1/3 full
			2	1.6658	-22.97	-21.07	579	47.1	0.0955	Wavy. 1/4 full
			3	1.6548	-23.07	-21.24	598	94.1	0.1520	Wavy. 1/8 full. Some film climbing.
			4	1.6436	-23.22	-21.57	626	147.1	0.2090	*
			5	1.6238	-23.47	-21.89	654	197.1	0.2660	Crescent
			6	1.5971	-23.84	-22.31	687	263.8	0.3220	Crescent but top not fully covered.
			7	1.5616	-24.39	-22.86	717	363.8	0.3800	Crescent
			8	1.5115	-25.14	-23.55	765	516.8	0.4360	"
			9	1.4400	-26.16	-24.52	784	722.7	0.4960	Annular.
			10	1.3414	-27.52	-25.76	769	890.4	0.5530	*
			11	1.2303	-29.31	-27.15	695	1002.2	0.6130	*
			12	1.1059	-31.67	-28.83	598	1117.9	0.6736	*

TABLE 1 (continued)

TEST NO.	W kg/h	q W/m²	STATION NO.	ABS. PRESS. 10^5 N/m²	AMMONIA TEMP. C SAT.	BULK	h W/m² °C	(ΔP/ΔL) N/m³	VAPOR QUALITY	FLOW PATTERN & OTHER VISUAL OBSERVATIONS
58.2	68.5	2368	0	1.5055	-25.2	-25.07	412	16.7	0.0000	*
			1	1.5037	-25.15	-24.38	437	20.6	0.0821	Wavy. 1/4 full. Top dry.
			2	1.5010	-25.17	-24.52	479	33.3	0.1730	" " " " "
			3	1.4960	-25.27	-24.65	535	71.6	0.2640	Wavy. 1/3 full.
			4	1.4841	-25.44	-24.93	596	117.7	0.3560	*
			5	1.4682	-25.69	-25.26	679	155.9	0.4490	Semi-crescent
			6	1.4474	-26.03	-25.62	770	201.0	0.5400	Semicrescent. 1/2 wet.
			7	1.4209	-26.45	-26.04	905	255.9	0.6300	" " "
			8	1.3871	-26.97	-26.6	1040	338.3	0.7210	Crescent. Top dry.
			9	1.3412	-27.58	-27.29	1232	400.1	0.8210	Crescent. Top layer thin.
			10	1.2903	-28.31	-28.13	1415	417.8	0.9100	*
			11	1.2412	-29.15	-28.97	319	418.7	0.9800	*
			12	1.1945	-30.12		44	405.0	1.0	*
59.2	580.0	2404	0	3.1473	-8.03	-16.06	1149	66.7	0.0000	*
			1	3.1411	-8.10	-13.2	1129	55.9	0.0000	Liquid only
			2	3.1357	-8.15	-9.99	1140	50.0	0.0000	Liquid only
			3	3.1291	-8.21	-6.81	855	110.8	0.0059	Wavy. Almost full.
			4	3.1105	-8.33	-7.60	904	207.9	0.0172	*
			5	3.0836	-8.52	-7.98	946	285.4	0.0298	Slug.
			6	3.0458	-8.84	-8.40	1008	412.8	0.0400	"
			7	2.9889	-9.33	-8.93	1062	598.2	0.0550	Crescent. Pulsating.
			8	2.9087	-10.03	-9.73	1151	810.0	0.0656	Crescent.
			9	2.7941	-10.99	-10.74	1239	1088.5	0.0800	"
			10	2.6525	-12.29	-12.00	1335	1420.0	0.0960	*
			11	2.4568	-14.01	-13.74	1506	1716.1	0.1120	*
			12	2.2411	-16.27	-15.84	1966	2071.1	0.1310	*
60.1	576.0	1635	0	2.8980	-10.06	-14.87	1039	55.9	0.0000	*
			1	2.8922	-10.14	-12.72	1142	46.1	0.0000	Liquid only.
			2	2.8876	-10.21	-10.42	1196	54.9	0.0000	Liquid only.
			3	2.8791	-10.30	-9.57	821	115.7	0.0064	Slug. 1/3 full.
			4	2.8605	-10.42	-9.70	814	187.3	0.0145	*
			5	2.8350	-10.6	-9.94	823	243.2	0.0228	Slug.
			6	2.8032	-10.86	-10.31	855	314.8	0.0314	"
			7	2.7608	-11.22	-10.74	884	398.1	0.0404	Crescent
			8	2.7095	-11.71	-11.33	960	500.1	0.0496	"
			9	2.6438	-12.37	-11.81	1015	648.2	0.0595	Crescent. Pulsating.
			10	2.5568	-13.21	-12.93	1133	885.5	0.0700	*
			11	2.4347	-14.30	-14.01	1592	1124.8	0.0812	*
			12	2.2921	-15.67	-15.19	1890	1298.4	0.0932	*
61.2	97.5	1635	0	2.4587	-14.06	-14.01	595	22.5	0.0000	*
			1	2.4563	-14.09	-13.74	620	23.5	0.0410	Wavy. 1/2 full.
			2	2.4534	-14.12	-13.77	650	33.3	0.0862	" " "
			3	2.4486	-14.17	-13.85	705	49.0	0.1310	" . 1/4 full. Some slugs.
			4	2.4419	-14.23	-13.95	766	66.7	0.1765	*
			5	2.4328	-14.3	-14.12	842	81.4	0.2220	Wavy
			6	2.4228	-14.4	-14.28	949	95.1	0.2680	Wavy. 1/4 full.
			7	2.4105	-14.53	-14.49	1070	117.7	0.3130	Semi-crescent. 1/2 tube wet.
			8	2.3951	-14.69	-14.71	1220	149.1	0.3580	" " . 2/3 tube wet.
			9	2.3753	-14.89	-14.98	1372	194.2	0.4050	" " " " "
			10	2.3495	-15.14	-15.25	1584	238.3	0.4500	*
			11	2.3193	-15.43	-15.52	1770	269.7	0.4960	*
			12	2.2859	-15.78	-15.82	1937	298.1	0.5410	*

TABLE 1 (continued)

TEST NO.	W kg/h	q W/m^2	STATION NO.	ABS. PRESS. 10^5 N/m^2	AMMONIA TEMP. °C SAT	BULK	h W/m^2 C	($\Delta P/\Delta L$) N/m^3	VAPOR QUALITY	FLOW PATTERN & OTHER VISUAL OBSERVATIONS
64.2	398.0	2298	0	3.9395	-02.27	-05.34	1145	38.2	0.0000	*
			1	3.9355	-02.33	-01.46	1082	52.9	0.0036	Wavy. Almost full.
			2	3.9276	-02.40	-01.77	1112	102.0	0.0198	Slug. 1/3 full.
			3	3.9115	-02.49	-02.06	1146	165.7	0.0362	" " "
			4	3.8885	-02.62	-02.35	1226	212.8	0.0525	*
			5	3.8614	-02.80	-02.64	1324	245.2	0.0696	Slug. 1/3 full.
			6	3.8284	-03.03	-02.92	1433	332.4	0.086	" " "
			7	3.7832	-03.33	-03.24	1539	562.9	0.1040	Crescent. Pulsating
			8	3.7303	-03.72	-03.68	1688	513.9	0.1210	" "
			9	3.6622	-04.22	-04.21	1833	659.0	0.1380	Crescent.
			10	3.5753	-04.83	-04.86	2031	820.8	0.1565	*
			11	3.4689	-05.57	-05.62	2399	965.0	0.1740	*
			12	3.3482	-06.48	-06.44	1785	1086.6	0.1930	*
65.2	121.0	2298	0	3.6490	-4.29	-3.81	612	23.5	0.0000	*
			1	3.6466	-4.30	-3.66	574	24.5	0.0596	Slug. 1/2 full
			2	3.6436	-4.33	-3.71	615	29.4	0.1020	" " "
			3	3.6396	-4.36	-3.81	660	49.0	0.1550	Wavy. 1/4 full.Some splashes.
			4	3.6321	-4.41	-3.94	719	78.4	0.2080	*
			5	3.6211	-4.48	-4.10	793	100.0	0.2600	Wavy. 1/3 full.
			6	3.6085	-4.57	-4.28	876	121.6	0.3140	Wavy. Film climbing to 1/2 the height.
			7	3.5925	-4.68	-4.47	991	149.1	0.3670	Semicrescent. Top 1/4 dry.
			8	3.5735	-4.83	-4.70	1122	182.4	0.4200	" 2/3 wet. Spray to top.
			9	3.5496	-5.01	-4.91	1270	232.4	0.4720	Semicrescent. Occasional splashes totop.
			10	3.5187	-5.23	-5.18	1475	293.2	0.5250	*
			11	3.4806	-5.49	-5.44	1709	332.4	0.5800	*
			12	3.4403	-5.80	-5.70	1986	352.0	0.6310	*
66.2	75.0	2298	0	3.5286		-4.7	519	14.7	0.0000	*
			1	3.5270	-5.16	-4.7	478	14.7	0.0781	Wavy. 1/3 full.
			2	3.5253	-5.17	-4.73	486	19.6	0.1670	Slug
			3	3.5223	-5.2	-4.81	499	38.2	0.2440	Wavy. 1/4 full.
			4	3.5162	-5.24	-4.86	508	62.8	0.3320	*
			5	3.5075	-5.3	-4.97	519	80.4	0.4160	Wavy. Film climbing to 1/3 height. Top dry.
			6	3.4973	-5.37	-5.10	532	93.2	0.5000	Semicrescent. Liquid film to 1/2 height.
			7	3.4855	-5.47	-5.23	541	111.8	0.5860	Semicrescent. 1/2 full.
			8	3.4710	-5.57	-5.39	553	134.3	0.6700	" 1/3 full.
			9	3.4539	-5.7	-5.55	561	163.8	0.7430	" " "
			10	3.4325	-5.84	-5.73	571	180.4	0.8410	*
			11	3.4114	-6.0	-5.91	396	173.6	0.9240	*
			12	3.3917	-6.17	18.2	65	161.8	superheated	*
66.1	77.5	2326	0	3.5450	-5.03	-4.44	577	13.7	0.0000	*
			1	3.5435	-5.04	-4.34	511	13.7	0.0755	Slug. 1/3 full.
			2	3.5419	-5.05	-4.39	522	17.6	0.1610	" " "
			3	3.5394	-5.07	-4.44	535	34.3	0.2440	Wavy. 1/8 full.
			4	3.5339	-5.11	-4.49	546	59.8	0.3280	*
			5	3.5254	-5.17	-4.60	562	74.5	0.4110	Wavy. 1/3 full.
			6	3.5163	-5.24	-4.73	577	86.3	0.4950	Semicrescent. 1/2 full.
			7	3.5051	-5.32	-4.86	598	103.0	0.5760	" " "
			8	3.4921	-5.41	-4.99	615	121.6	0.6600	" " "
			9	3.4763	-5.53	-5.15	634	153.0	0.7400	" " "
			10	3.4560	-5.68	-5.31	651	173.6	0.8260	*
			11	3.4354	-5.83	-5.49	647	169.6	0.9100	*
			12	3.4160	-5.97	+4.19	181	159.8	superheated	*

TABLE 1 (continued)

TEST NO.	W kg/h	q w/m^2	STATION NO.	ABS. PRESS. 10^5 N/m^2	AMMONIA TEMP. C SAT	AMMONIA TEMP. C BULK	h w/m^2 °C	($\Delta P/\Delta L$) N/m^3	VAPOR QUALITY	FLOW PATTERNS & OTHER VISUAL OBSERVATIONS
63.1	600.0	2354	0	4.1775	-00.72	-05.65	1372	62.8	0.0000	*
			1	4.1708	-00.78	-03.13	1321	55.9	0.0000	Liquid only
			2	4.1650	-00.85	-00.45	1115	84.3	0.0033	Wavy. Very little vapor.
			3	4.1510	-00.94	-00.11	1138	169.6	0.0146	Slug. 1/3 full.
			4	4.1249	-01.06	-00.81	1184	245.2	0.0254	*
			5	4.0933	-01.25	-01.18	1281	300.1	0.0378	Slug 1/2 full. Some elongated bubbles at the bottom.
			6	4.0542	-01.52	-01.46	1413	407.0	0.0496	Same as above.
			7	3.9975	-01.88	-01.88	1576	535.4	0.0618	Crescent
			8	3.9282	-02.38	-02.4	1833	665.9	0.0750	Crescent but pulsating.
			9	3.8406	-03.02	-03.03	2201	914.9	0.0878	" " "
			10	3.7128	-03.85	-03.87	2605	1201.3	0.1015	*
			11	3.5579	-04.89	-04.86	2964	1372.9	0.1160	*
			12	3.3895	-06.91	-06.02	3018	1487.7	0.1310	*
71.1	663.0	2807	0	3.1570	-07.97	-16.76	918	92.2	0.0000	*
			1	3.1476	-08.04	-14.33	999	75.5	0.0000	Dirty oil at bottom.
			2	3.1401	-08.10	-11.65	1055	65.7	0.0000	" " " "
			3	3.1321	-08.16	-09.04	1075	90.2	0.0000	" " " "
			4	3.1189	-08.27	-07.71	1143	117.5	0.0073	*
			5	3.0902	-08.49	-08.08	1171	270.7	0.0176	Slug. Rapidly pulsating.
			6	3.0551	-08.80	-08.56	1243	380.5	0.0280	" " "
			7	3.0006	-09.22	-09.14	1364	542.3	0.0392	Crescent. Pulsating.
			8	2.9276	-09.85	-09.83	1476	815.9	0.0509	Crescent.
			9	2.8085	-10.85	-10.79	1564	1179.7	0.0638	"
			10	2.6498	-12.27	-12.13	1554	1516.1	0.0780	*
			11	2.4515	-14.13	-13.90	1724	1909.3	0.0935	*
			12	2.2003	-16.67	-16.33	2137	2302.6	0.1111	*
72.0	627.1	2298	0	4.4107	00.71	-04.55	1106	73.5	0.0000	*
			1	4.4012	00.66	-02.14	1074	75.5	0 0000	Liquid only. Oil film moving slowly at bottom.
			2	4.3938	00.61	00.61	1084	85.3	0.0006	Same as above.
			3	4.3811	00.53	01.33	997	161.8	0.0112	Slug. Moving oil at bottom.
			4	4.3557	00.38	01.05	1022	251.0	0.0221	*
			5	4.3220	00.17	00.69	1068	323.6	0.0331	Slug. Oil at bottom.
			6	4.2796	-00.10	00.22	1143	432.5	0.0443	Slug. Tending to crescent. Oil at bottom.
			7	4.2202	-00.46	-00.27	1277	570.7	0.0559	Crescent. Oil film to 1/4 hgt.
			8	4.1452	-00.95	-00.84	1436	783.5	0.0678	Crescent. Oil at bottom.
			9	4.0358	-01.65	-01.54	1680	1046.6	0.0805	" " " "
			10	3.8989	-02.56	-02.45	1847	1250.3	0.0938	*
			11	3.7414	-03.64	-03.55	2268	1512.2	0.1076	*
			12	3.5429	-05.05	-04.86	2540	1774.0	0.1224	*
73.0	2675.6	2298	0	5.2261	5.37	-08.08	1910	1532.8	0.0000	*
			1	5.0696	4.52	-07.50	1914	1270.9	0.0000	Liquid only. Oil film.
			2	4.9417	3.82	-06.86	1917	1121.9	0.0000	" " " "
			3	4.8056	3.04	-06.23	1921	1144.4	0.0000	" " " "
			4	4.6722	2.29	-05.60	1924	1028.7	0.0000	*
			5	4.5634	1.62	-04.97	1886	908.1	0.0000	Liquid only. Oil film at bottom.
			6	4.4586	1.02	-04.36	1811	844.3	0.0000	Same as above.
			7	4.3647	0.42	-02.98	4533	901.2	0.0000	" " "
			8	4.2463	-0.30	-03.11	1818	1022.8	0.0000	" " "
			9	4.1239	-1.09	-02.48	1821	1000.3	0.0000	" " "
			10	4.0109	-1.82	-01.93	1961	1138.5	0.0010	*
			11	3.8557	-2.85	-02.22	1499	2740.0	0.0071	*
			12	3.3658	-6.35	-05.91	2421	4341.4	0.0220	*

* no sight glass

Heat transfer coefficients in nonboiling region are based on bulk ammonia temperature while those in boiling region are base on saturation temperatures.

TABLE 1 (continued)

TEST NO.	W kg/h	q w/m^2	STATION NO.	ABS. PRESS. 10^5 N/m^2	AMMONIA TEMP. °C SAT	AMMONIA TEMP. °C BULK	h w/m^2 °C	($\Delta P/\Delta L$) N/m^3	VAPOR QUALITY	FLOW PATTERN & OTHER VISUAL OBSERVATIONS
67.1	2827.6	2347	0	3.3342	-06.59	-33.57	924	3064.6	0.0000	*
			1	3.0702	-08.66	-32.69	1112	2496.8	0.0000	Liquid only. Oil film top 1/2
			2	2.7757	-11.13	-32.07	1160	2140.8	0.0000	Liquid only. Oil film all around. Moving faster than in station 1.
			3	2.5664	-13.07	-31.47	1496	1735.8	0.0000	Liquid only. No oil visible.
			4	2.3671	-14.92	-30.85	1749	1450.4	0.0000	*
			5	2.2250	-16.45	-30.23	1828	1116.0	0.0000	Liquid only. No oil visible.
			6	2.1045	-17.70	-29.64	1916	1024.8	0.0000	" " " " "
			7	1.9839	-19.00	-29.05	2011	992.4	0.0000	" " " " "
			8	1.8709	-20.43	-28.47	2063	1018.9	0.0000	" " " " "
			9	1.7440	-21.90	-27.88	2174	1056.2	0.0000	" " " " "
			10	1.6222	-23.55	-27.27	2360	1003.2	0.0000	*
			11	1.5080	-25.11	-26.71	2365	1008.1	0.0000	*
			12	1.3849	-26.93	-26.07	1462	1014.0	0.0039	*
68.1	1471.4	2312	0	2.1327	-17.39	-33.85	1048	439.3	0.0000	*
			1	2.0857	-17.91	-32.49	893	384.4	0.0000	Liquid only. Oil film.
			2	2.0467	-18.34	-31.30	1030	365.8	0.0000	" " " "
			3	1.9997	-18.87	-30.12	862	422.7	0.0000	Liquid only. Thick oil film all around, moving very slow.
			4	1.9473	-19.46	-28.97	1050	399.1	0.0000	*
			5	1.9056	-19.96	-27.80	1410	319.7	0.0000	Liquid only. No oil seen.
			6	1.8721	-20.37	-26.62	1369	296.1	0.0000	" " " " "
			7	1.8358	-20.78	-25.49	1330	297.1	0.0000	" " " " "
			8	1.8021	-21.23	-24.32	1314	296.1	0.0000	" " " " "
			9	1.7660	-21.65	-23.16	1279	313.8	0.0000	" " " " "
			10	1.7283	-22.14	-22.03	1880	588.4	0.0020	*
			11	1.6257	-23.46	-23.08	1884	1883.8	0.0105	*
			12	1.2849	-28.51	-27.99	1962	3178.3	0.0312	*
69.2	1324.7	2307	0	3.5180	-05.25	-15.57	1389	344.2	0.0000	*
			1	3.4834	-05.48	-14.28	1490	292.2	0.0000	Oil film all around. Top part moves very slow. Bottom part has jerky motion.
			2	3.4527	-05.70	-12.93	1498	265.7	0.0000	Liquid only. Oil at bottom.
			3	3.4207	-05.94	-11.65	1244	265.7	0.0000	" " " " "
			4	3.3901	-06.17	-10.37	1457	255.0	0.0000	*
			5	3.3608	-06.39	-09.06		238.3	0.0000	Liquid only. Oil at bottom.
			6	3.3341	-06.59	-07.79	1445	232.4	0.0000	" " " " "
			7	3.3062	-06.80	-06.49	1233	305.0	0.0018	Liquid only. Oil at bottom.
			8	3.2623	-07.15	-06.71	1181	582.5	0.0078	Slug. Very rapid pulsations.
			9	3.1690	-07.87	-07.71	1277	1074.8	0.0151	" " " "
			10	3.0092	-09.16	-08.93	1296	1697.5	0.0244	*
			11	3.7501	-11.20	-10.85	1358	2800.8	0.0361	*
			12	2.3499	-15.13	-14.82	1890	3904.0	0.0541	*
70.1	2815.9	2291	0	4.2164	-00.49	-17.12	1987	1401.4	0.0000	*
			1	4.0660	-01.46	-16.49	2135	1198.4	0.0000	Liquid only. No oil seen.
			2	3.9483	-02.23	-15.90	1818	1105.2	0.0000	" " . Oil at bottom.
			3	3.8057	-03.19	-15.33	1822	1200.3	0.0000	" " . Thick oil film all around.
			4	3.6658	-04.18	-14.76	1786	1177.8	0.0000	*
			5	3.5285	-05.13	-14.17	1828	1066.9	0.0000	Oil film all around.
			6	3.4147	-06.01	-13.63	1794	983.6	0.0000	Oil film at bottom upto 1/16 height of tube.
			7	3.2971	-06.84	-13.04	1797	966.9	0.0000	No oil visible
			8	3.1872	-07.75	-12.50	1762	1044.4	0.0000	Thin oil film on upper half and bottom.
			9	3.0512	-08.80	-11.92	1803	1088.5	0.0000	Liquid only. Thin lines and drops moving constantly. Probably oil.
			10	2.9309	-09.82	-11.38	1971	1011.1	0.0000	*
			11	2.8132	-10.82	-10.95	2647	1894.6	0.0020	*
			12	2.4850	-13.81	-13.47	2645	2777.2	0.0146	*

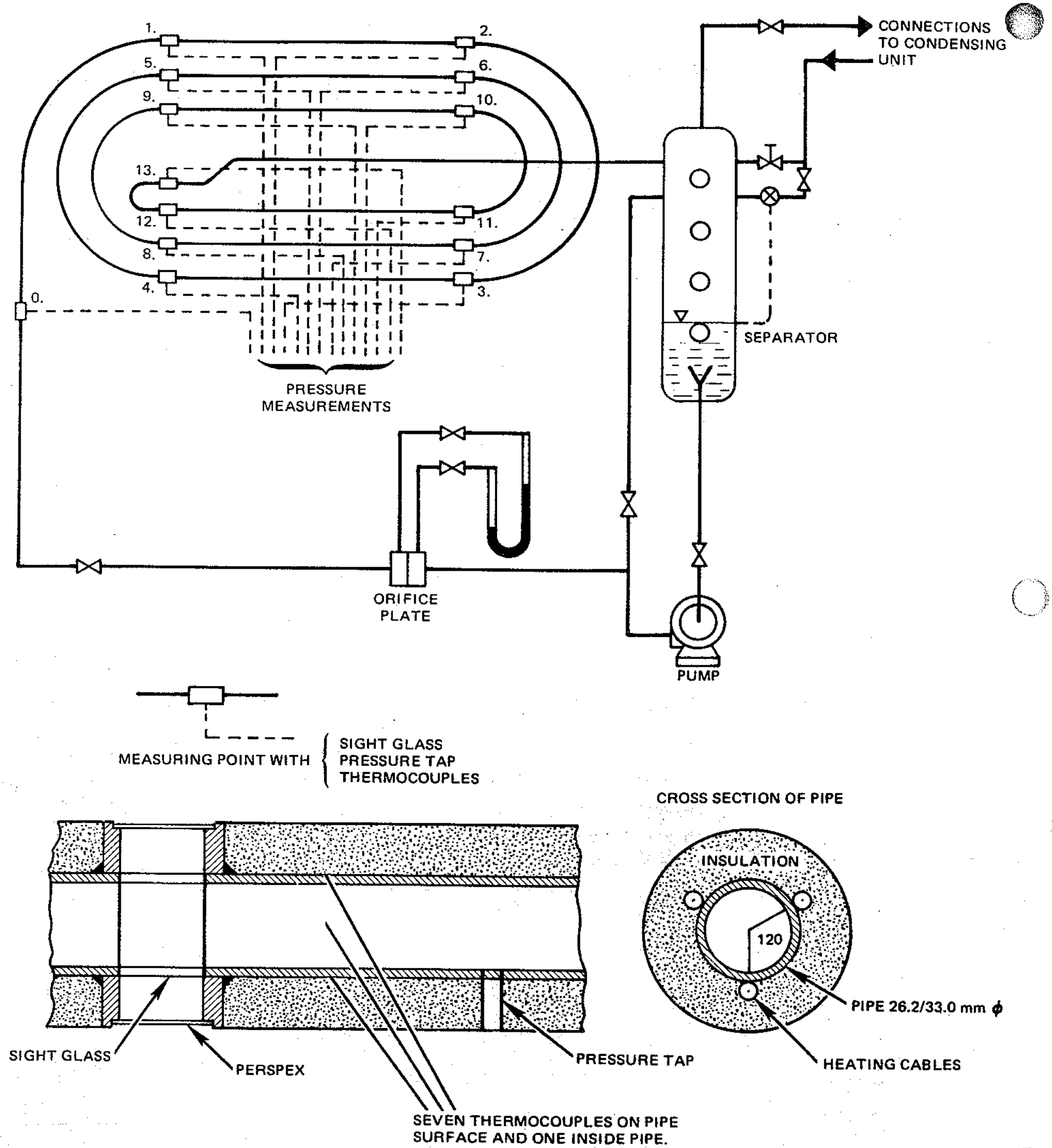

Fig. 1 Schematic arrangement and some design details of test evaporator used by Shah

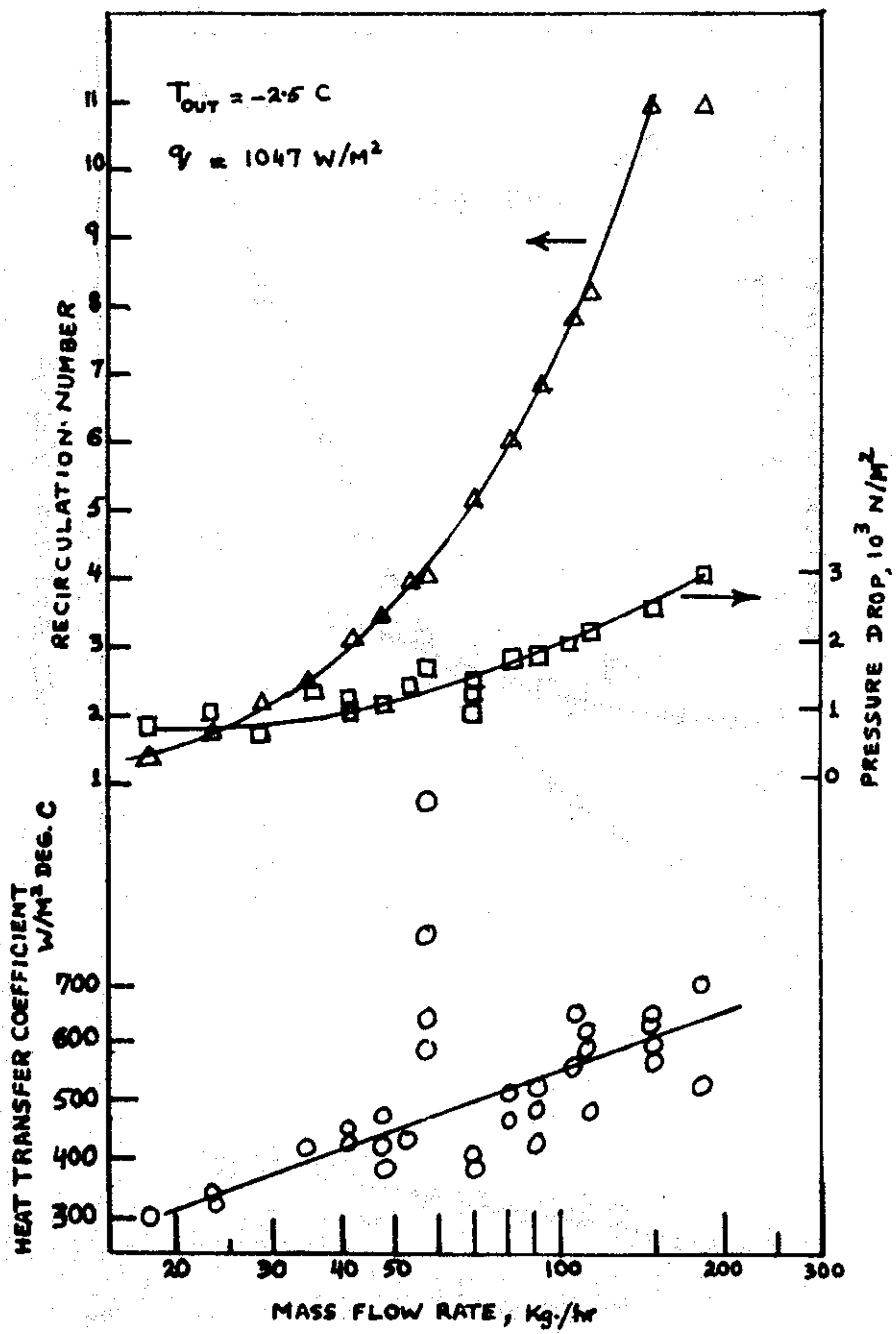

Fig. 2 Data of Van Maale and Cosijn

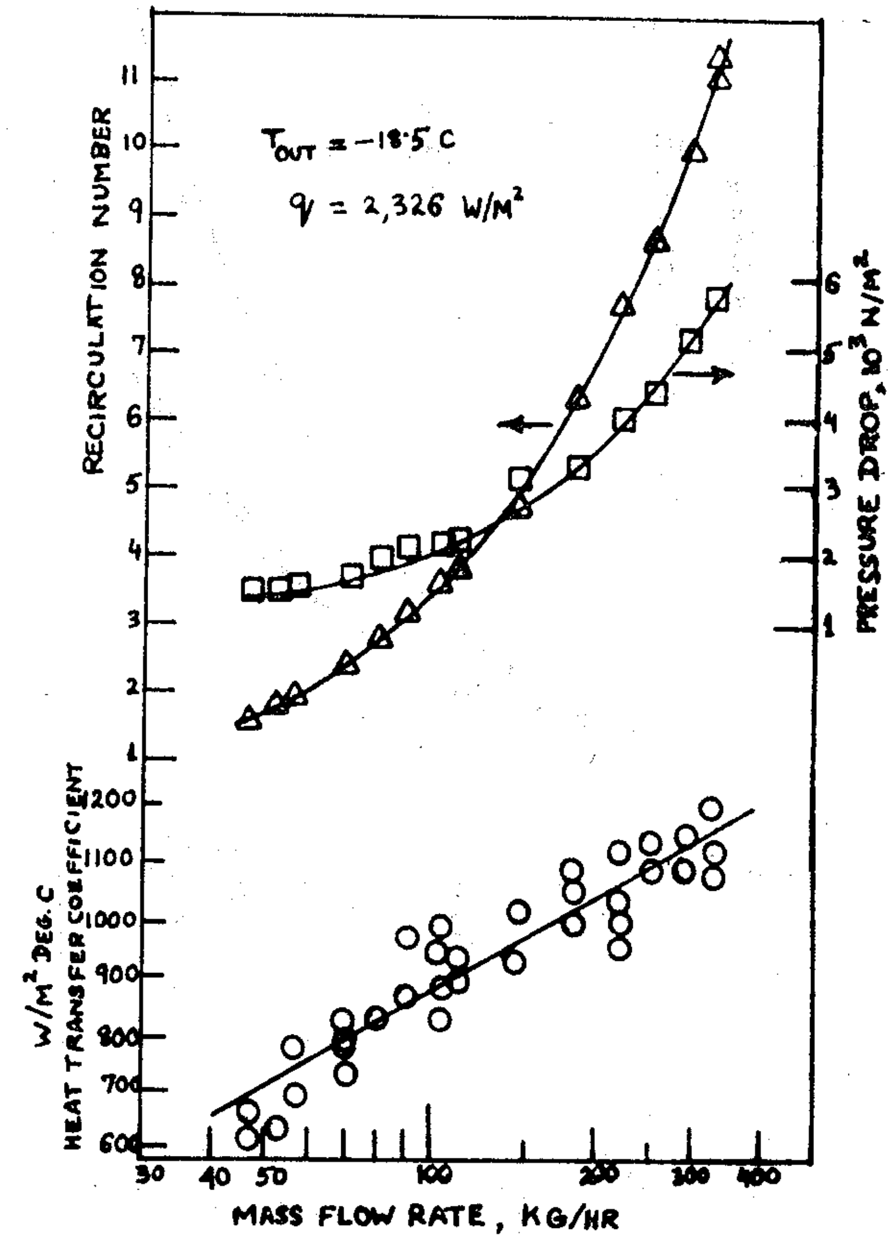

Fig. 3 Data of Van Maale and Cosijn

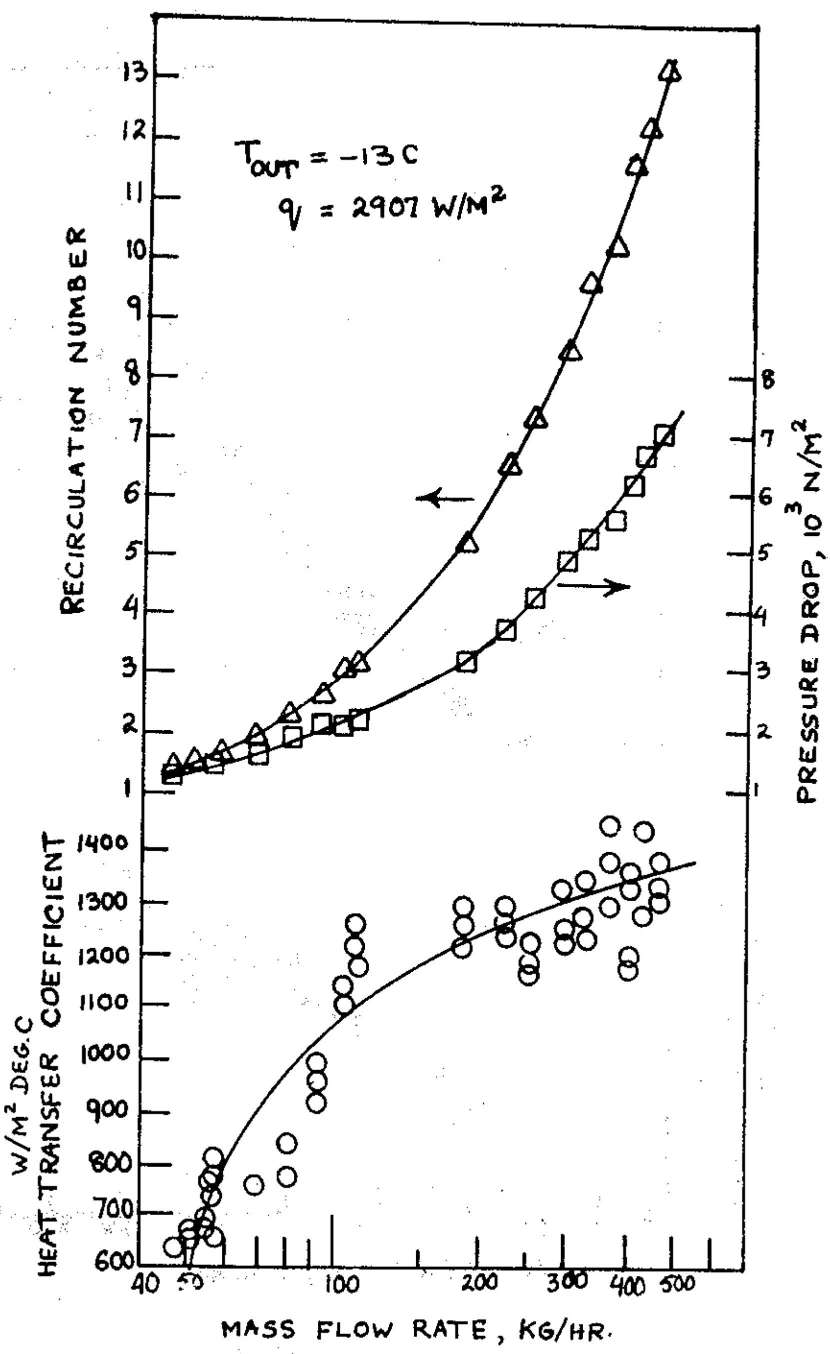

Fig. 4 Data of Van Maale and Cosijn

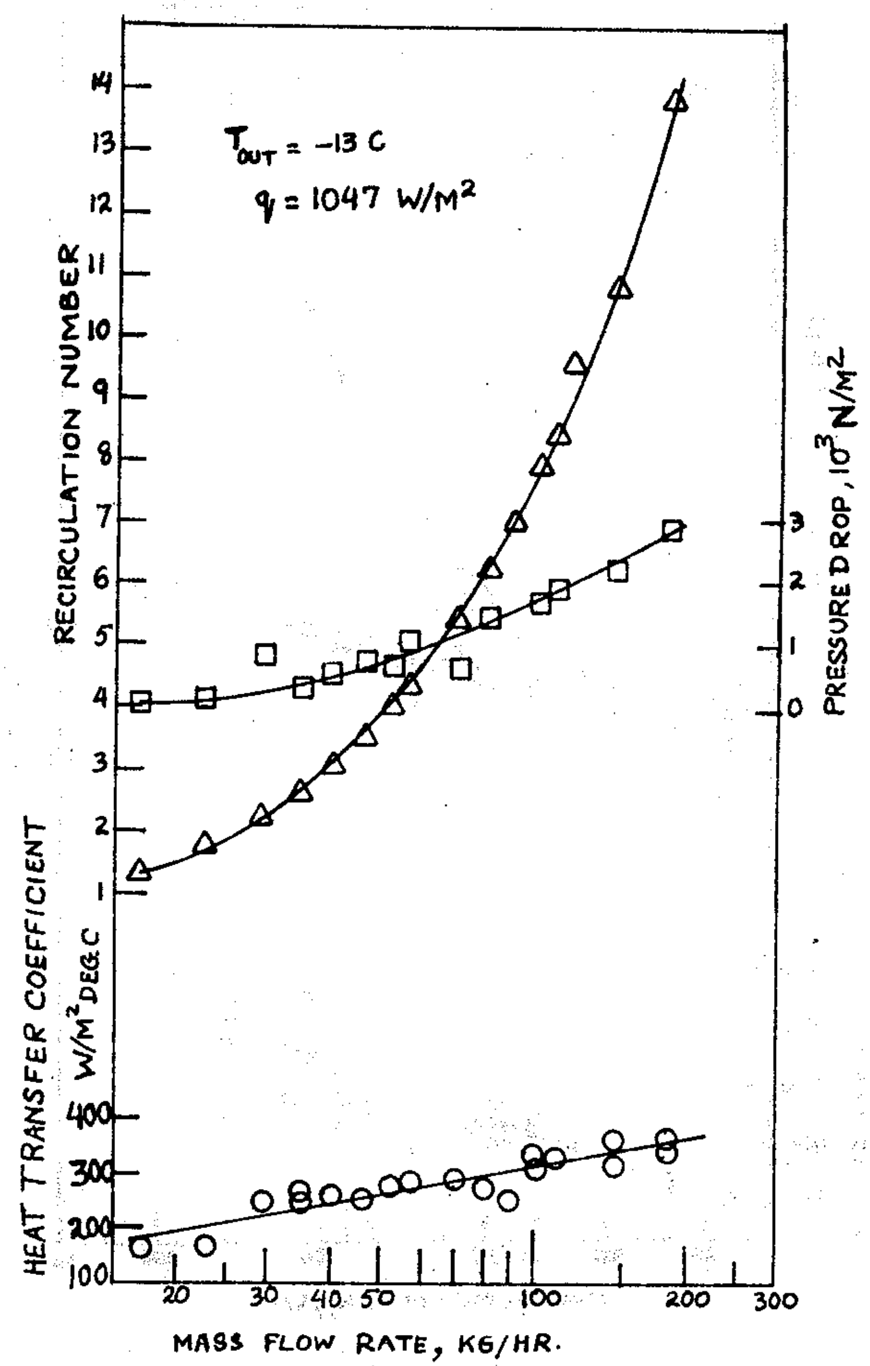

Fig. 5 Data of Van Maale and Cosijn

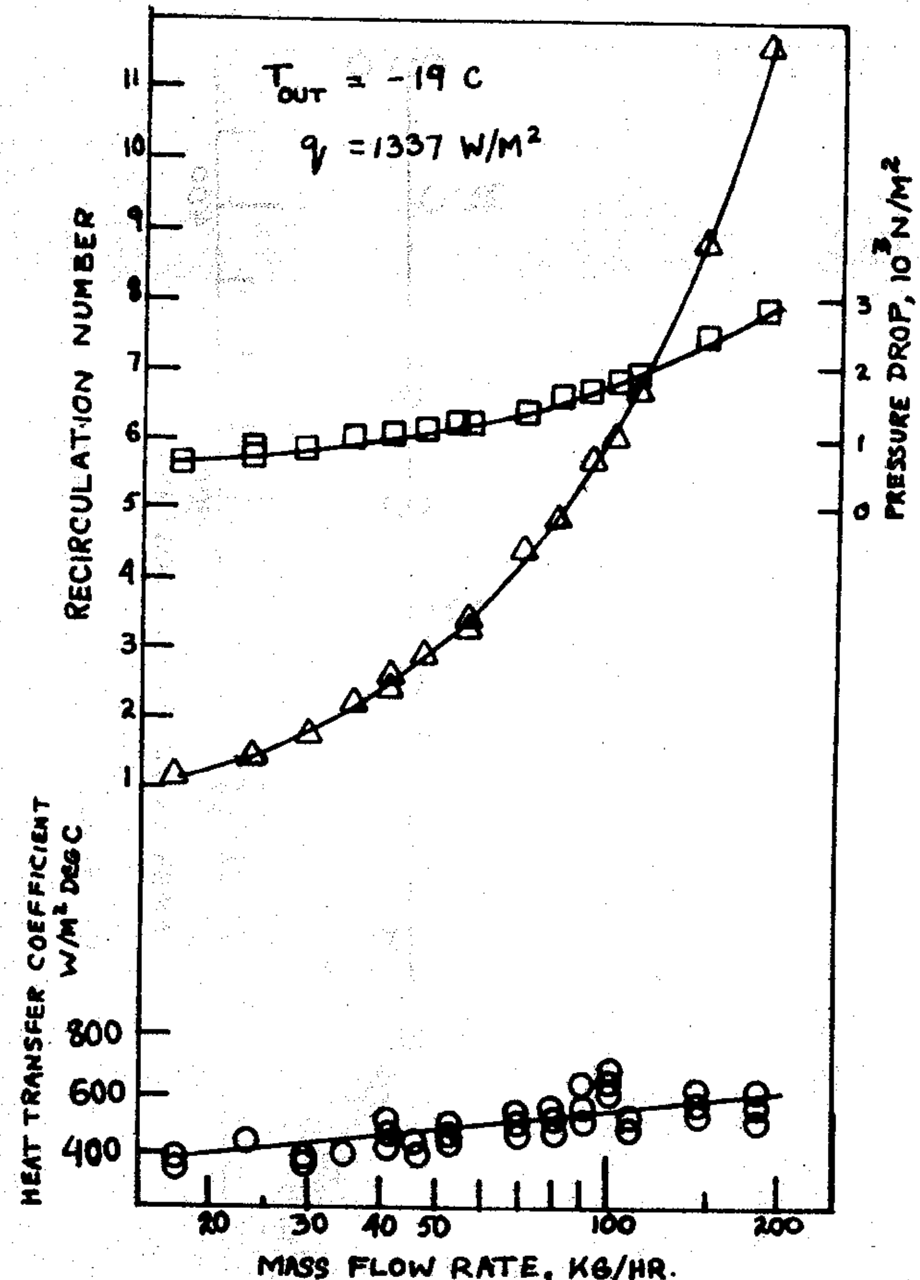

Fig. 6 Data of Van Maale and Cosijn

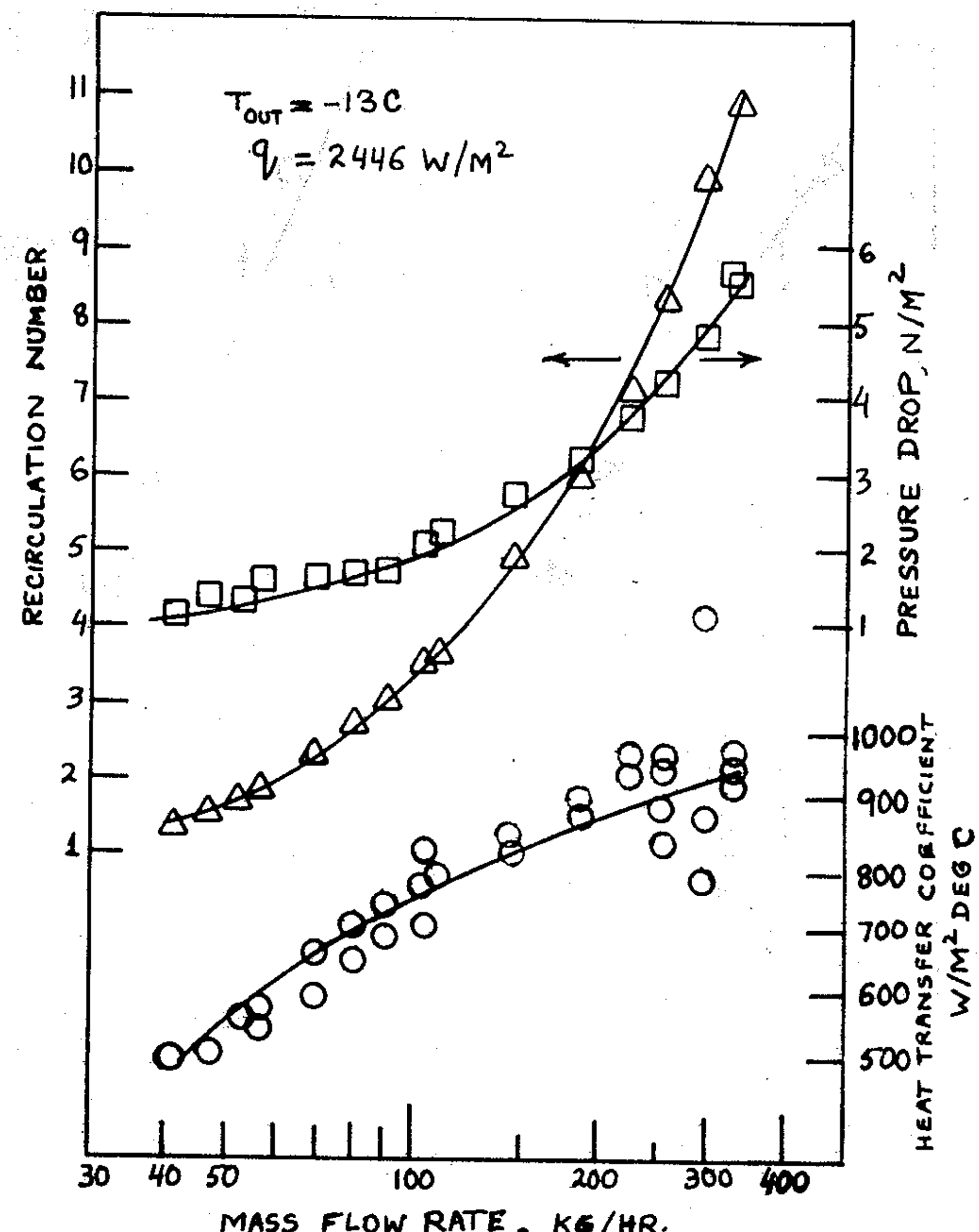

Fig. 7 Data of Van Maale and Cosijn

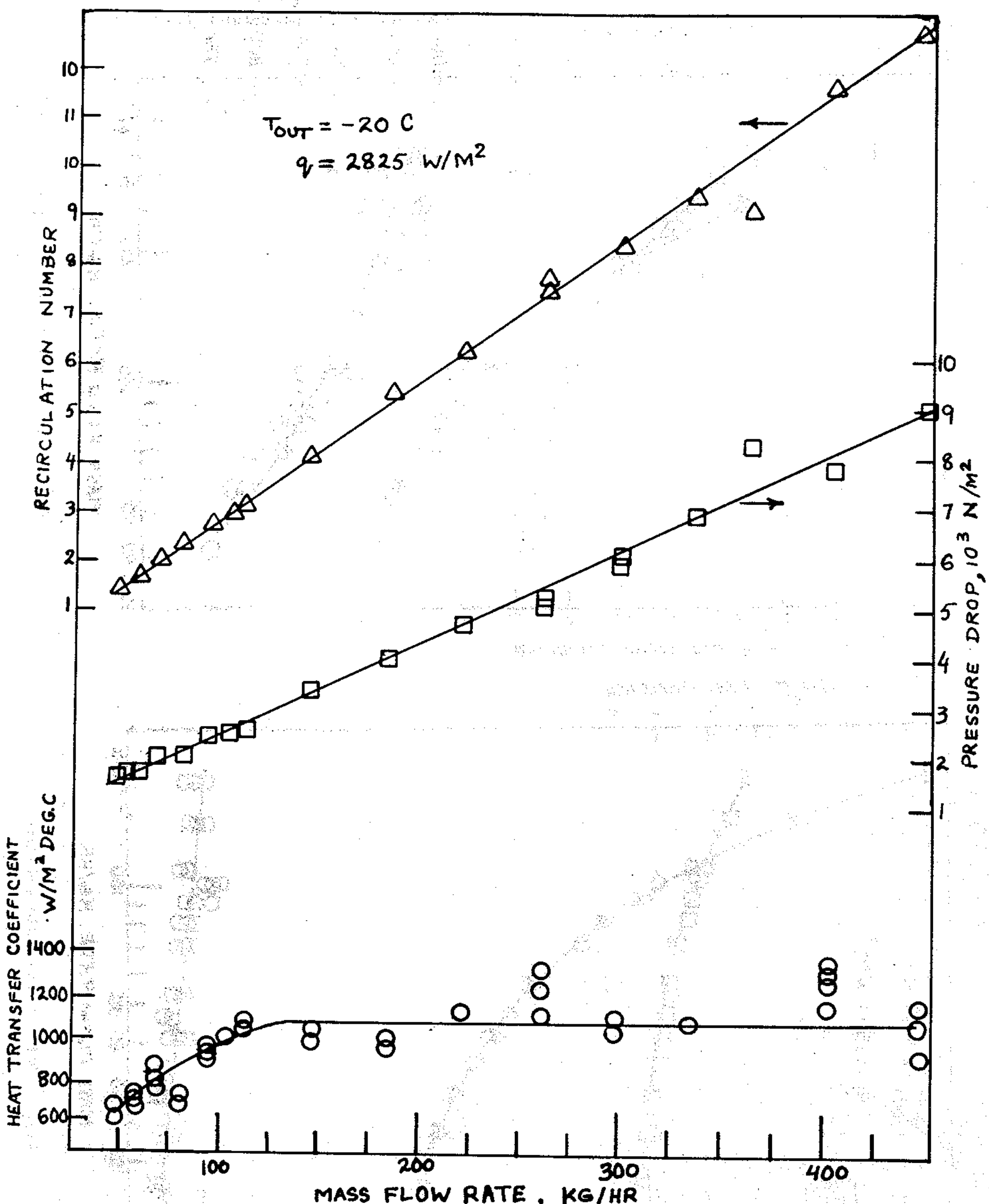

Fig. 8 Data of Van Maale and Cosijn

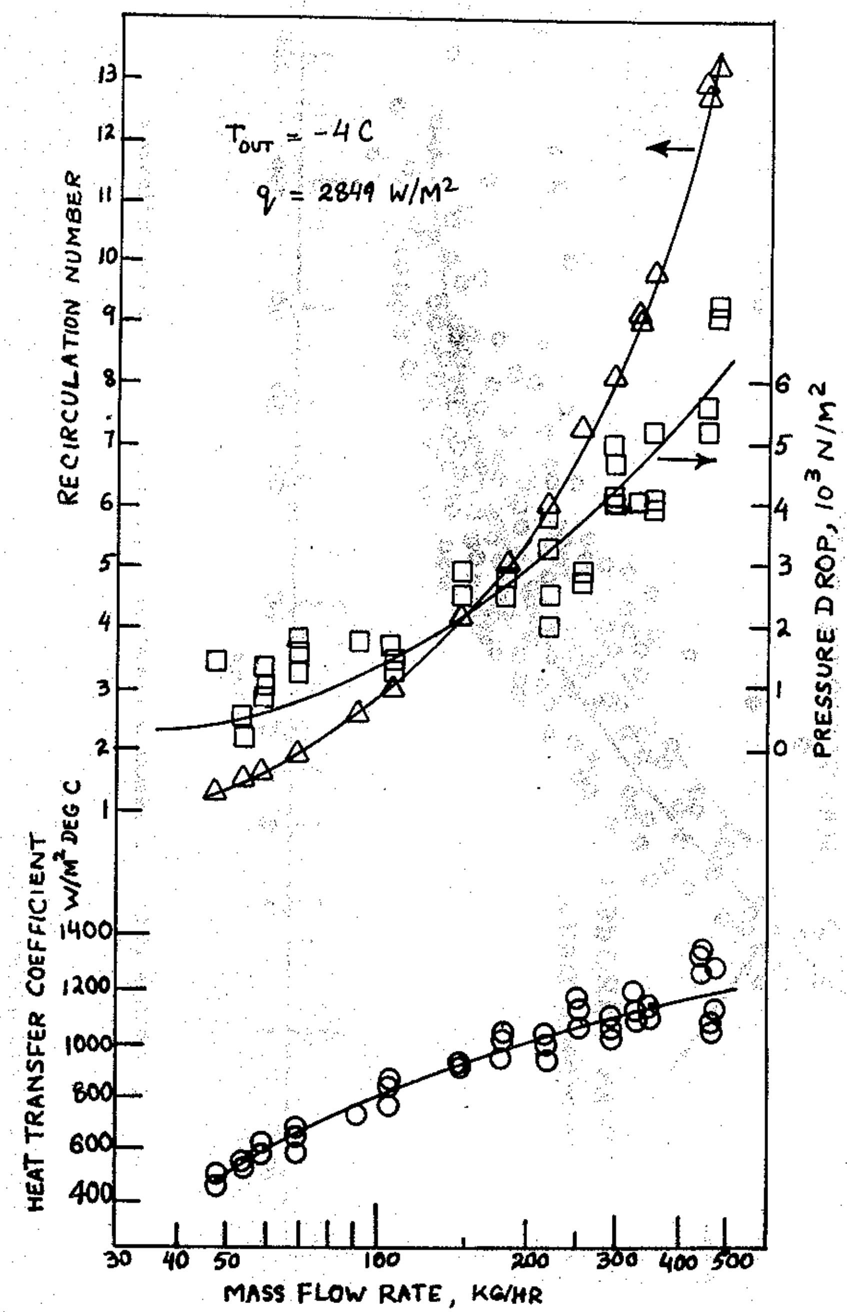

Fig. 9 Data of Van Maale and Cosijn

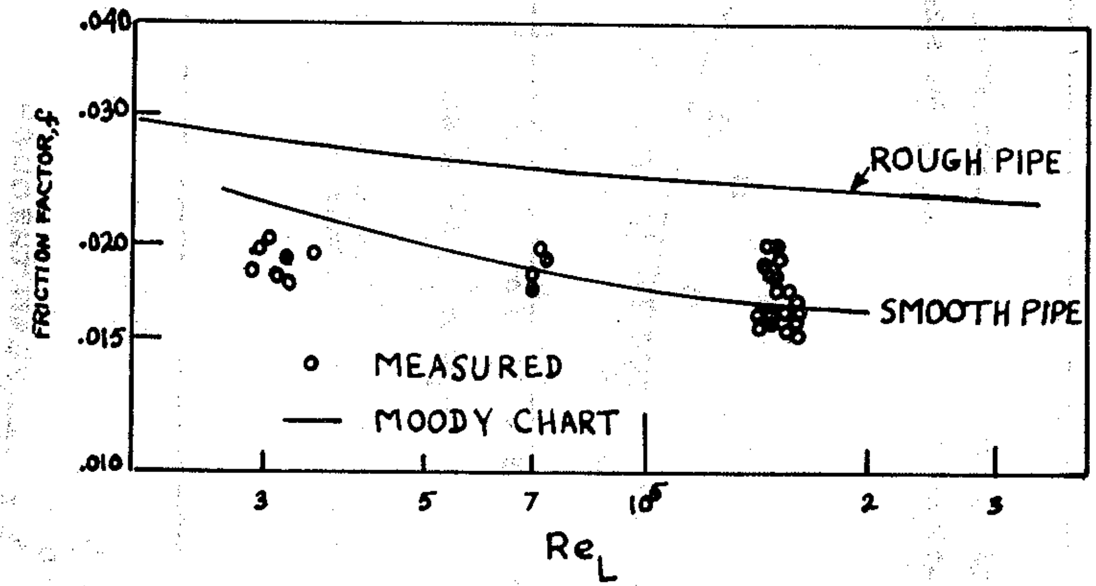

Fig. 10 Single-phase friction factors calculated from data of Shah; ammonia temperature from 0 to $-15^{o}C$

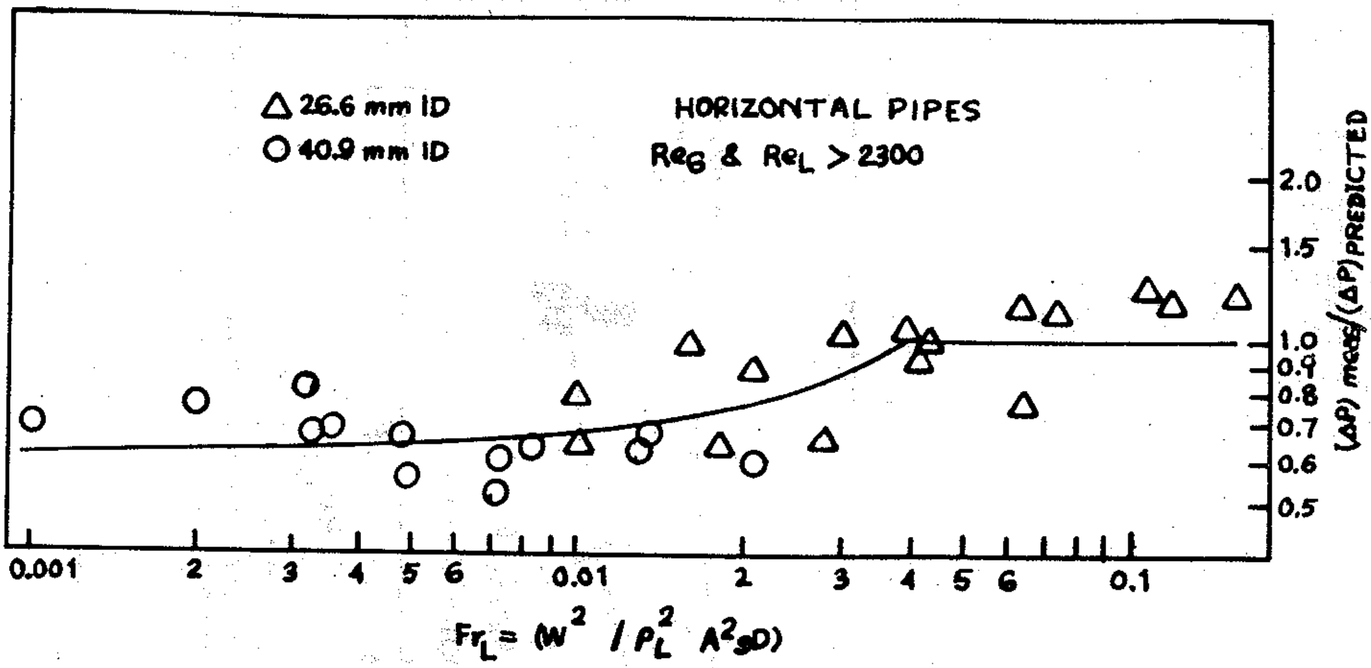

Fig. 11 Ratio of measured pressure drop to that predicted by Lockhart-Martinelli correlation as a function of an all liquid Froude number. Data of Chaddock et al[10].

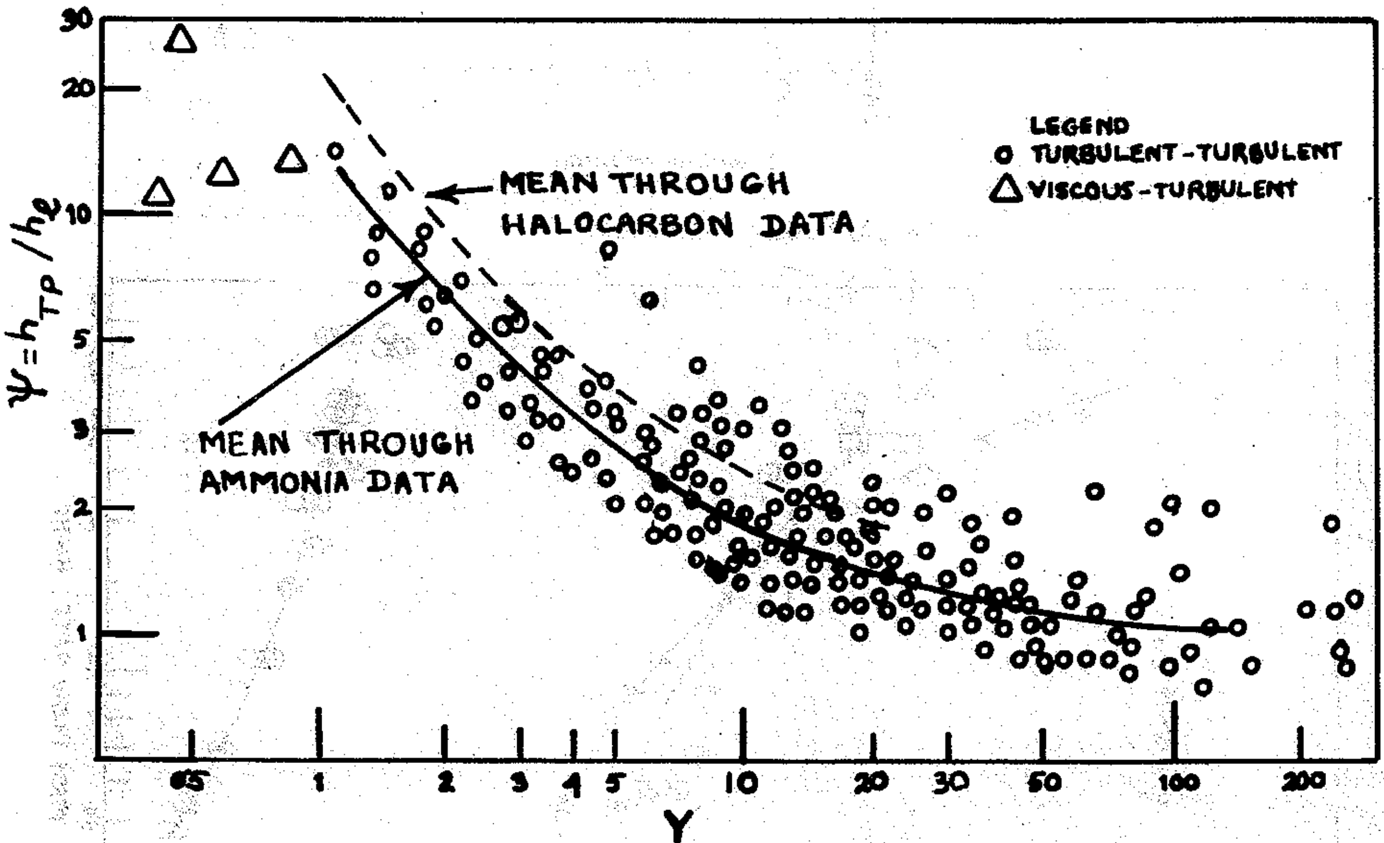

Fig. 12 Shah's data for ammonia evaporator in terms of Υ - ψ co-ordinates