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Characteristics of a falling film distillation column.

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CHARACTERISTICS OF A FALLING FILM DISTILLATION COLUMN

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ABSTRACT

The characteristics of a Falling Film Distillation Column are presented. The work undertaken by the writer has been limited to the study of a column of fixed length and diameter under the conditions of atmospheric pressure. Heat transfer studies have been conducted on the systems of water and on varying compositions of ethanol-water in an attempt to evaluate the quantitative factors on a pilot plant sized film still so that the design of a commercial alcohol unit might be attempted.

The author wishes to acknowledge
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who directed this research.

INTRODUCTION

Up to the present time a number of types of distillation columns have been developed. The most prominent of these are sieve plate and bubble cap units. The sieve plate is the simpler of the two and consists of a vertical tower of punched plates where liquid is contacted by vapor to effect a material interchange. The bubble cap tower is a type in which an attempt is made to improve fractionation by a more intimate and longer contact between liquid and vapor. A further development for still greater contact area is the packed tower, in which greater intimacy of contact is obtained by passing the vapor over even greater surface areas for the same tower volume. The packed tower differs from the first two in that the concentration gradient is gradual from top to bottom instead of stepwise.

More modern still designs are based on theories formulated from packed tower studies. The liquid film covering the tower packing is seen as an ideal medium for vapor contact, and the spray still, the cascade column, and the wetted walled tower follow this concept.

Each design had its own characteristics and it soon channeled to industrial fields favoring the special attributes of the type. For example, the fermentation industries utilized sieve plate columns; the petroleum industries used batch stills and bubble tray towers. One reason for the former was the suitability of the sieve plate for handling a liquid of relatively high solids content, whereas the bubble cap unit was more readily adaptable for usages entailing negligible solids materials where high fractionation efficiency was demanded. In fields where low absolute pressures were necessary and little entrainment loss could be tolerated, the spray still was introduced. It was only partially successful because of its inability to reduce entrainment. Maintenance costs also ran higher than expected, and the simplified design emerged as a cascade or baffle plate still.

The baffle or cascade type completely eliminated the disadvantage of the mist formation in a spray still, but with the advent of very high vacuum processes, it was found that the cascade still had too high a pressure drop even though it was considerably below that of previous plate column designs. The next and probably latest design of commercial importance was the wetted walled tower. This column is similar to a tubular film type evaporator, and although there is little information on this apparatus in the field of distillation, it is conceivable that data may be utilized from that of evaporation for certain design factors. However, far too little information is available in the literature covering this important piece of equipment, and it is the purpose of this investigation to study the characteristics of a film type distillation column from the standpoint of heat and material interchange.

The falling film type of distillation unit is considered since it may be an ideal unit for the removal of alcohol from the beer formed in the fermentation of cereal grains. Recent advances in the alcohol industry have shown the importance of low temperature beer stripping, necessitating the use of high vacuums. The conventional sieve plate, grid packed, and ring packed columns, have proved comparatively poor in extremely high vacuum work because of excessively high pressure drops. The falling film type of still because of the free vapor space obviously achieves the minimum in pressure drop through the column length.

Certain column packing materials, particularly wood grids, have created a cleaning problem in beer distillation that has remained unsolved after many years of study. It seems that the solids in the beer adhere to the grid surfaces where they decompose thermally during operation, thereby imparting disagreeable odors and/or taste to the distillate. This factor may be eliminated by the film type unit with the elimination of a packing to become fouled and irregular surfaces to permit build-up of solids.

A very desirable feature of film still design is the elimination of the dilution effect caused by sparged steam which is normally present in the conventional beer still. This feature permits a maximum ratio of alcohol to water during processing and thereby increases the steam economy.

An important item in distillation equipment design is simplicity of construction and maintenance. The falling film still is of relatively simple design in comparison with other beer columns.

A final feature of the film type column must be emphasized. There have been heat transfer coefficients reported in the literature (1) as high as 3,000 Btu./hr.-sq. ft.-°F. for certain sizes of vertical film heaters. Such a coefficient would effect an economy that could be realized as in no other beer stripping unit. This feature already has been realized on short tube film boilers.

HISTORICAL

A film still may be defined as a unit in which fractionation occurs during film flow. This film type fractionation may vary in a number of ways, depending on the manner of introducing feed, the method of heating the film, the manner of removing the vapor formed in the process, and the subsequent removal or reflux of the condensed overhead vapor.

The major portion of the data in the literature at present has been obtained on a film type column in which the feed has been introduced as a vapor from a reboiler to the base section, while the overhead vapor has been condensed and returned as total reflux (2, 3, 4, 5, 6). With such a system, one of the usual methods of column evaluation may be employed, and the efficiency has been reported as a height equivalent to a theoretical plate (H.E.T.P.), a height of a transfer unit (H.T.U.), a number of theoretical plates at total reflux, or some type of correlating equation involving film resistances, heat, or material transfers.

The problem undertaken by the author in this thesis was the evaluation of column performance under possible plant operating conditions of liquid feed to the top of the column, vaporisation by jacketed steam, absence of reflux, and single passage of material-in-process through the column. Comparable operating conditions are not represented among data available in the literature.

THEORETICAL

The exact mechanism of heat transfer for film type distillation equipment is not to be found in the literature. Data have been presented on individual coefficients but are lacking for overall performance. It is possible, however, for one to postulate a mechanism by the use of quantitative relations that have evolved from studies on film heaters. The Nusselt equation (7) has approximated conditions in the film heater and is

$$\frac{h\,L}{C_p\Gamma} = 2.62\left(\frac{k\,L}{3^{1.33}}\,\frac{\rho^{0.66}}{C_p\,\mu^{0.33}}\,\frac{g^{0.33}}{\Gamma^{1.33}}\right)^{0.66}$$

where h = liquid film coefficient Btu./hr.-sq. ft.-°F.

L = height of pipe, ft.

C_p = specific heat of wall material, Btu./lb.-°F.

Γ = fluid rate, lbs./hr.-ft.

k = thermal conductivity, Btu./hr.-sq.ft.-°F/ft. of tube wall.

ρ = density of fluid, lbs./cu. ft.

g = acceleration of gravity, ft./sq. hr.

μ = fluid viscosity, lbs./hr.-ft.

It is seen from this equation that the individual heat transfer coefficient varies directly as the feed rate, and inversely as the column length and steam pressure (or temperature drop between heating medium and liquid film). This can be expected to hold approximately for the overall transfer of heat, as the length of the column used in these studies might be considered abnormally high in comparison with that of the film heaters reported.

For the distillation theory involved in the study, possible mechanisms must be reviewed. A batch distillation quantitative relationship may be considered initially. By definition, batch distillation is a process in which a portion of material is charged to a pot, subjected to heat, and, upon vaporization, is removed immediately and condensed. The mathematical analysis, which is a differential one, was first proposed by Rayleigh (8). The relationship he presented was that the logarithm of the ratio of liquid fed to the pot to residual liquid equals the integral value of $dx/(y-x)$, where x = the composition of the liquid and y = the vapor composition in equilibrium with the liquid. The limits are taken between the composition of the residual charge and the original charge with reference to the more volatile component. This relation is easily evaluated by means of graphical integration.

Modern exponents of continuous processing techniques state that a continuous process is a series of batch processes occurring in rapid succession. Therefore, it is possible that a batch relationship might approximate the results obtained by continuous processing, particularly in a film type column.

If equilibrium is assumed to be obtained in each area of a film still, the relationship would be that for equilibrium distillation. The quantitative study of such a process would demand the use of a graphical procedure with the equilibrium diagram.

It is reasonable to expect that the mechanism of film type distillation is intermediate between the two extremes of batch distillation and true equilibrium distillation. There has been some evidence to show that the mechanism of film or wetted wall distillation approaches more closely the batch type evaluation. This similarity is indicated in the work of Woodfield and Copeland (9), who dealt with materials that closely adhered to Raoult's law. One object of the present investigation is to determine the validity of the theory or the degree of divergence from it.

EXPERIMENTAL

The apparatus for this investigation was a steam-jacketed copper tube 2.938 inches inside diameter by 15.20 feet in length (see Fig. 1). The upper portion of the heating surface had an 8 inch calming section, the upper edge of which was notched as a weir, and over which feed to the column could be distributed. The base of the column was fitted with a liquid-seal vapor trap through which stripped liquid passed to a "bottoms" receiver. It was very important that the wall surface be perfectly vertical; therefore a plumb line was arranged alongside the column. The external surface of the column was not lagged and therefore necessitated heat corrections on all determinations involving steam-side heat transfer coefficients.

One of the most critical factors involved in film still operation is liquid distribution to the column. A weir arrangement was used in this work. This weir was so designed that 19 V notches, 2" deep, were evenly spaced around the upper edge of a 3 inch copper tube. The points jutting upward from this section were bent outward so that they lay on the circumference of a circle of 3½ inch diameter. This spreading of points accomplished a rather smooth flow of liquid into the column with no perceptible channeling effect.

After the liquid was spread in a uniform film, boiling occurred. The vapor moved up the column, left the unit through a vapor line, and passed through an entrainment separator into the top of the condenser. The distillate left the condenser at the bottom through a non-condensables vent bottle and flowed into a "tops" product receiver.

By preliminary inspection, it was observed that a complete and

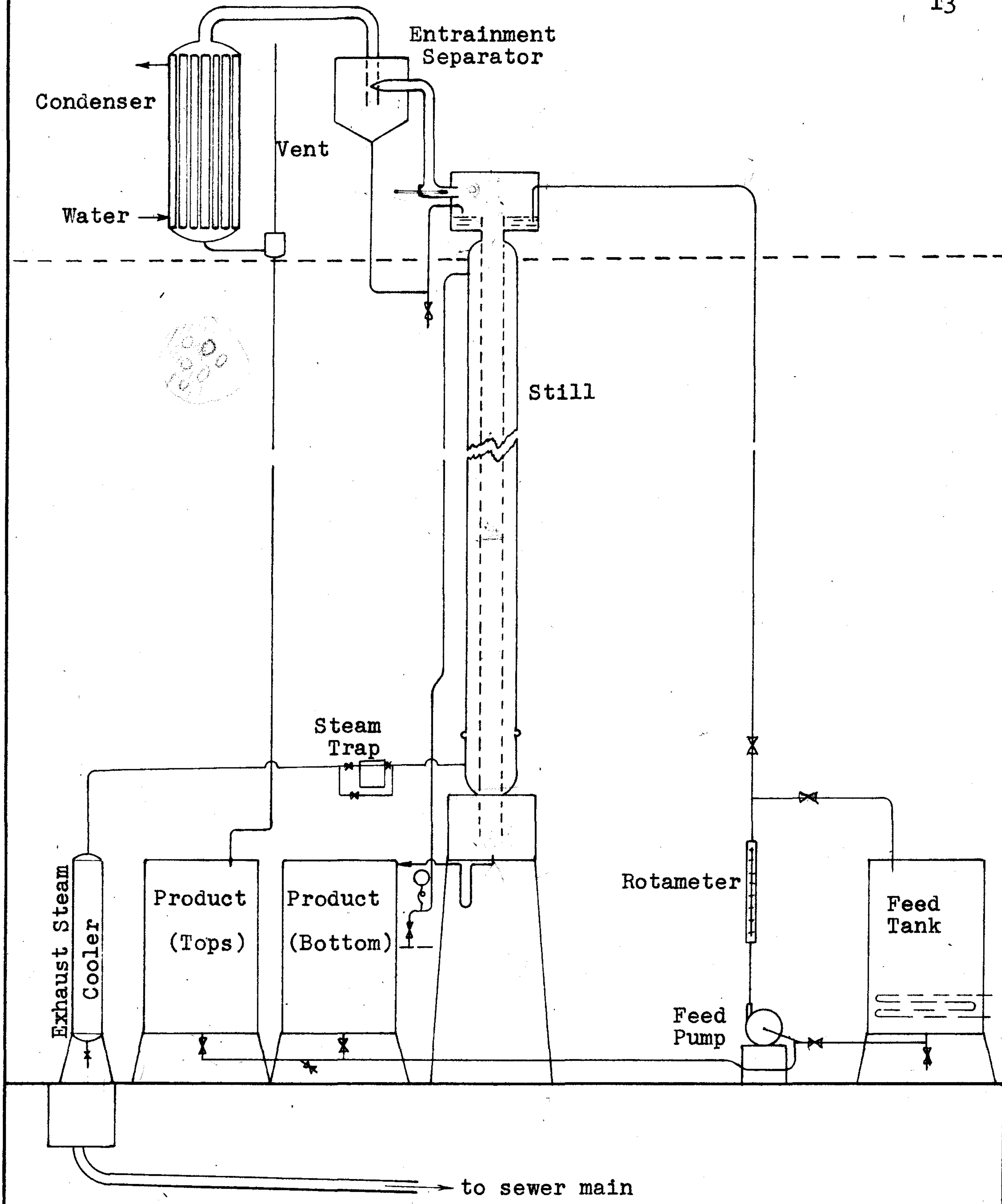

Fig. 1. Flow Diagram of Falling Film Still.

uniform film could be formed, as long as no boiling occurred with a feed of less than 5 gallons per hour; however, when boiling occurred, a feed rate slightly over 5 gallons per hour was required.

In the study of the heat transfer characteristics of the film column, water was charged to a feed pot and heated by a steam coil to the desired temperature. This feed pot was fitted with a recirculation line to distribute the heat throughout the batch of liquid. The feed was transferred from the pot by a centrifugal pump to the upper section of the film still, thence into the head of the column, overflowing the weir, and passing in a uniform film down the column. The liquid was raised to the boiling point and partially vaporized with the vapor passing up the column. The residual liquid flowed down the tower in gradually lessening amounts and out the bottom through a liquid leg seal and into the "bottoms" product drum. The vapor from the operation was condensed and fed into the "tops" receiver. The rates of these streams were measured volumetrically, and a run was not begun until the products flowed at a constant rate as determined by incremental measurements.

In addition to the rate of flow measurements, the temperatures of the feed liquor, the vapor, and the residual liquid to bottoms were obtained.

The steam to the jacket of the column was maintained constant by a Brown "Air-o-line" pressure recorder and controller. The steam, after giving up its latent heat, was removed through a trap at the base of the jacket. The condensed steam leaving the trap was passed through a small condenser in order to liquefy any flash vapor that might form. The rate of the steam condensate was measured in order to ascertain the steam side coefficients.

After the conclusion of the heat transfer experiments using water alone, mixtures of ethanol and water were studied. These runs were similar in nature to the water runs with the exception that the feed was adjusted to the approximate desired composition. After the feed was distilled, the "tops" and "bottoms" fractions were analyzed for alcohol content by the standard procedure of laboratory distillation and refractive index measurement. This supplied all the necessary information for the fractionation studies.

1. Calibration of Rotameter

The rotameter was calibrated in place by utilizing the following method. Water was pumped at a constant rotameter setting, and the actual flow was caught and measured volumetrically until a constant flow was measured per unit of time. An average of these measured volumes is recorded in Table I as "metered flow - G.P.H", and is graphically presented in Fig. 2.

Calculations indicated that for feeds used in this investigation, deviations due to density were negligible.

Table I. Rotameter Calibration

Run No.	Rotameter Reading G.P.H.	Metered Flow G.P.H.
1	4.9	4.1
2	11.3	11.8
3	12.0	12.6
4	14.2	15.3
5	18.7	20.2
6	19.1	20.6
7	20.1	21.7
8	20.2	21.8
9	21.0	22.7
10	24.8	26.8
11	30.0	32.4
12	39.0	42.1
13	39.9	43.1
14	43.3	46.8
15	59.8	64.6
16	65.5	70.7
17	95.5	103.1
18	95.9	103.6
19	99.5	107.5
20	103.0	111.2

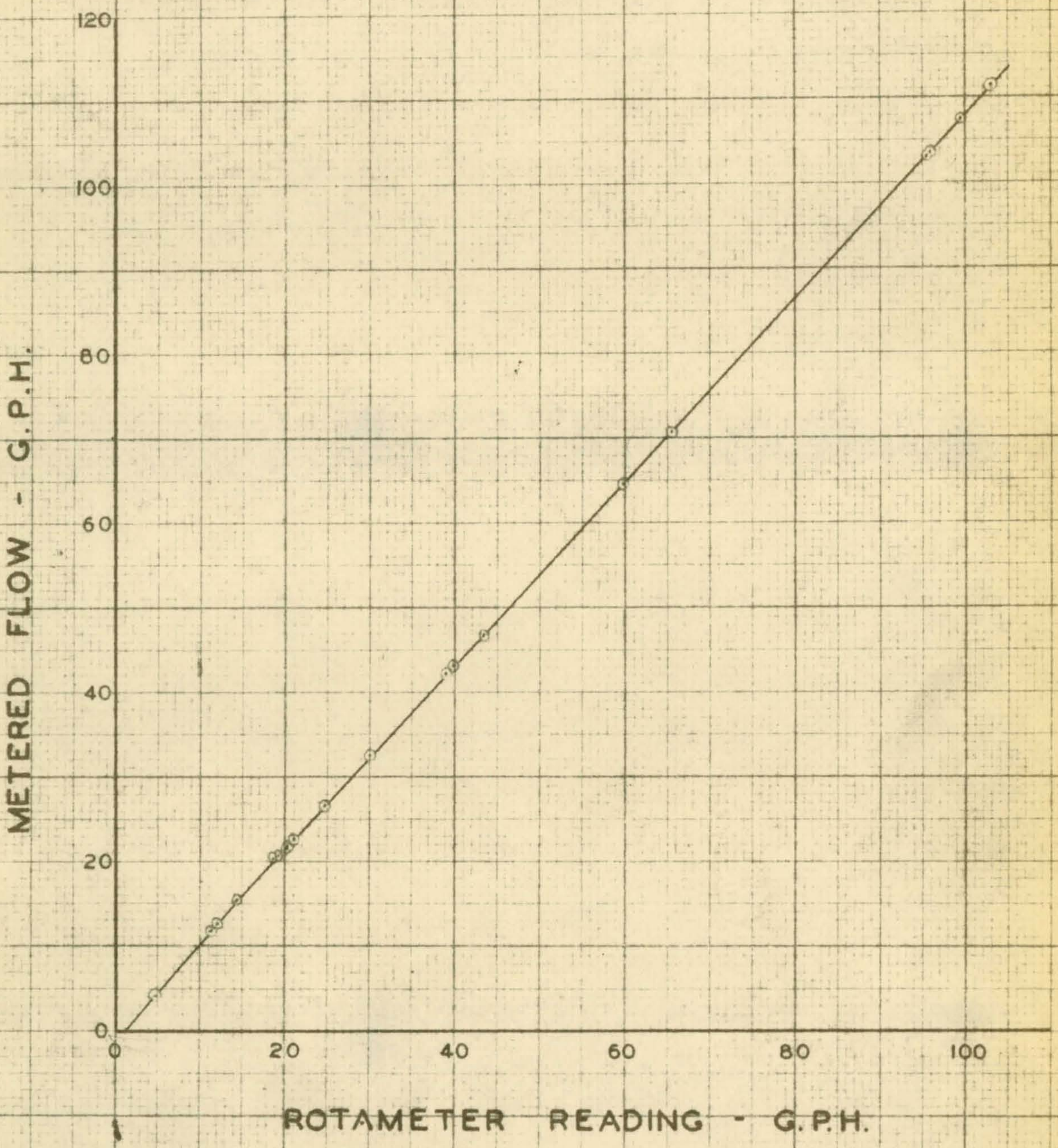

FIG. 2 ROTAMETER CALIBRATION CURVE

2. External heat losses from the column

The heat losses of the column were determined by sealing off the top and bottom sections of the column to prevent undue convection on the inner surfaces, maintaining a constant steam jacket pressure and measuring the rate of condensate from the steam jacket.

The condensate rate was calculated back to total heat necessary to maintain column temperature without feed. These values corresponded to heat losses by radiation and convection. Data are presented in Table II and are correlated in Fig. 3.

Table No. II Heat Losses from Column

Run No.	Pressure Hg.	Pressure p.s.i. abs.	Corresponding Temp., °F.	Loss in B.t.u./hr.
SB-1	1.85	15.605	215.02	8,159.3
2	3.85	16.587	218.15	11,987.6
3	6.25	17.766	221.72	13,999.9
4	8.35	18.797	224.67	18,431.9
5	10.55	19.879	227.63	19,484.7
6	0.80	15.089	213.34	8,799.4
7	3.00	16.170	216.89	10,698.9
8	6.00	17.643	221.33	13,896.6
9	9.00	19.116	225.58	19,890.9
10	12.00	20.588	229.51	20,090.9

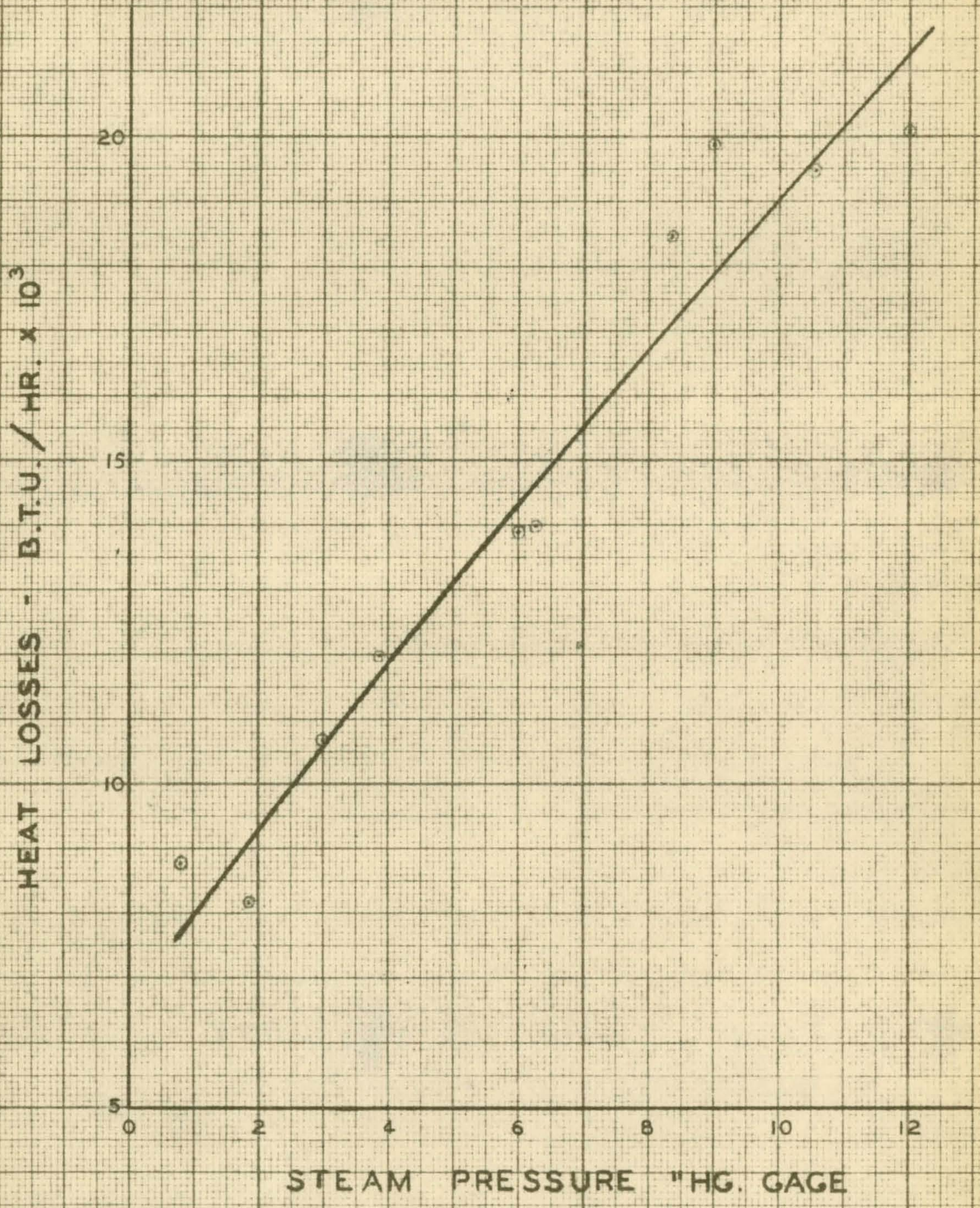

FIG. 3 HEAT LOSSES ON FILM TYPE STILL

Experimental Results - Water Runs

Complete data were obtained for heat transfer calculations on the falling film still as described in the Experimental Procedure and illustrated by a sample run in Fig. 4. The calculations from these data are explained in the Sample Calculations (Table IX), and are tabulated in Table X. A summary of these results is presented in Table No. III.

The values of the overall heat transfer coefficients are presented for both steam side and liquid side. These values were obtained by considering Δt as the log mean difference between steam jacket temperature and liquid film temperature, area as a constant value obtained by measurement of the column and total heat transferred as the product of pounds of steam exhausted and its true latent heat, i. e. the difference between the specific enthalpy of saturated vapor at the inlet pressure and that of the saturated boiling liquid at jacket pressure assuming adiabatic conditions across the inlet steam valve less radiation losses.

The value of the overall liquid side coefficient has the same Δt value and area constant for the corresponding steam side coefficient, but the total heat transferred in this case is the summation of sensible heat to the vapor, latent heat to the vapor, and sensible heat to the bottoms. This coefficient probably represents more nearly the true heat transfer coefficient, as there is always a possibility of varying radiation losses due to atmospheric disturbances about the column, and the effect produced by intermittent trap action on the steam condensate discharge.

A correlation of data for heat transfer coefficients relating

Operating Data:

Run No.	W-43
Date	4-17-45
Time	8:52 P. M.
Jacket Steam Pressure	6.37" Hg.
Feed Temperature	58.0°C.
Feed Rate	42.9 G.P.H.
Tops Product Rate	178.0 cc/min.
Bottoms Product Temperature	98.0°C.
Bottoms Product Rate	2536.0 cc/min.
Vapor Temperature	100.0°C.
Steam Condensate Rate	565.0 cc/min.
Room Temperature	23.0°C.
Barometric Pressure	756 mm.

Remarks: Operations steady; good jacket pressure control

Fig. No. 4 - Sample Data Sheet

Table III - Summary Data for Water Runs

Run No.	Feed Rate (G.P.H.)	Steam Rate (#/hr.)	Jacket Pressure ("Hg.)	U. Liq. Side Btu./hr. ft.^{2}-°F.	U. Steam Side Btu./hr. ft.^{2}-°F.
W-1	11.20	45.00	4.02	72.99	103.90
W-2	11.80	83.80	9.2	90.40	104.15
W-3	21.44	82.10	9.2	105.60	120.13
W-4	13.50	59.75	9.2	75.59	59.35
W-5	12.60	58.56	9.2	74.27	59.32
W-6	12.60	57.42	9.26	72.16	56.69
W-7	4.10	57.51	9.1	43.95	42.45
W-8	3.20	57.77	9.1	66.09	77.08
W-9	22.25	56.85	9.0	74.32	66.22
W-10	21.82	56.85	9.2	71.93	62.56
W-11	21.71	56.58	9.0	70.08	62.48
W-12	33.05	57.37	9.0	73.08	68.71
W-13	32.40	57.64	9.1	69.83	68.40
W-17	46.87	89.90	9.2	125.80	142.46
W-18	46.87	89.40	9.2	120.09	140.91
W-19	46.76	86.59	9.2	121.38	134.42
W-20	15.70	85.93	9.2	70.57	81.51
W-21	15.30	86.72	9.15	77.79	94.14
W-22	27.43	83.29	9.4	102.71	114.85
W-23	26.78	84.61	9.05	105.02	121.45
W-24	26.24	86.99	9.4	105.05	123.37

Table III - Summary Data for Water Runs (Cont.)

Run No.	Feed Rate (G.P.H.)	Steam Rate (#/hr.)	Jacket Pressure ("Hg.)	U. Liq. Side Btu./hr. ft.^{2}-°F.	U. Steam Side Btu./hr. ft.^{2}-°F.
W-25	25.92	86.19	9.1	102.44	123.73
W-26	25.60	91.61	9.3	108.59	132.22
W-27	111.24	92.54	8.8	138.29	155.86
W-28	71.60	104.17	9.2	129.82	170.66
W-29	70.74	89.90	9.2	130.07	142.11
W-30	22.68	36.62	2.2	20.24	47.80
W-31	22.68	39.66	2.2	21.61	54.88
W-32	22.68	42.83	2.3	33.80	78.42
W-33	22.68	42.57	2.75	38.72	75.48
W-34	22.68	40.32	2.6	37.74	71.48
W-35	22.68	40.32	2.5	35.41	72.19
W-36	22.68	69.14	4.5	61.47	119.66
W-37	42.12	44.95	3.25	26.68	50.86
W-38	42.66	55.13	4.1	45.00	80.64
W-39	40.50	-	3.7	62.16	-
W-40	41.90	54.86	3.0	67.45	114.05
W-41	41.90	63.98	4.7	86.16	117.68
W-42	42.44	68.61	5.5	99.54	120.42
W-43	42.98	74.69	6.4	106.72	125.37
W-44	64.04	62.80	3.7	106.73	136.32
W-45	63.94	64.78	4.9	97.89	126.61

Table III - Summary Data for Water Runs (Cont.)

Run No.	Feed Rate (G.P.H.)	Steam Rate (#/hr.)	Jacket Pressure ("Hg.)	U. Liq. Side Btu./hr. ft.^{2}-°F.	U. Steam Side Btu./hr. ft.^{2}-°F.
W-46	63.61	74.50	6.8	99.17	124.25
W-47	63.94	73.11	5.58	109.70	136.16
W-48	61.56	84.61	7.54	100.91	142.37
W-49	107.46	84.87	6.3	173.99	168.84
W-50	107.03	86.99	6.3	163.75	164.55
W-51	105.30	86.19	6.4	143.03	161.78
W-52	104.76	86.19	6.4	151.31	161.78
W-53	104.76	84.21	8.5	124.42	131.27
W-54	104.76	85.93	8.4	124.93	135.23
W-55	103.57	113.43	10.6	168.39	173.37
W-56	103.14	112.50	10.8	160.98	170.02
W-57	22.14	68.22	6.5	66.48	86.71
W-58	20.92	72.05	6.6	77.76	99.45
W-59	20.63	71.65	6.5	80.98	100.67
W-60	20.20	69.80	6.5	78.66	97.47
W-61	21.71	81.03	8.58	98.95	111.90
W-64	21.38	61.06	5.8	68.21	91.48
W-65	42.93	88.84	10.17	80.62	128.10
W-66	43.09	65.44	4.9	77.99	123.16
W-67	43.42	65.44	5.15	76.89	120.61
W-68	43.09	79.58	7.7	102.03	128.02
W-69	42.66	105.76	12.3	134.16	142.21

the overall coefficients on the liquid side with those on the steam side is presented in Fig. 5. It is noted that the curve is displaced from the 45° line slightly for the lower values of the overall heat transfer coefficient and to a much less extent as the higher values of heat transfer were obtained. Moreover, the overall heat transfer coefficients from the liquid side enthalpies are presented in functional relationship to liquid feed rate in Fig. 6, where average values of jacket pressures at different feed rates are connected by tie lines.

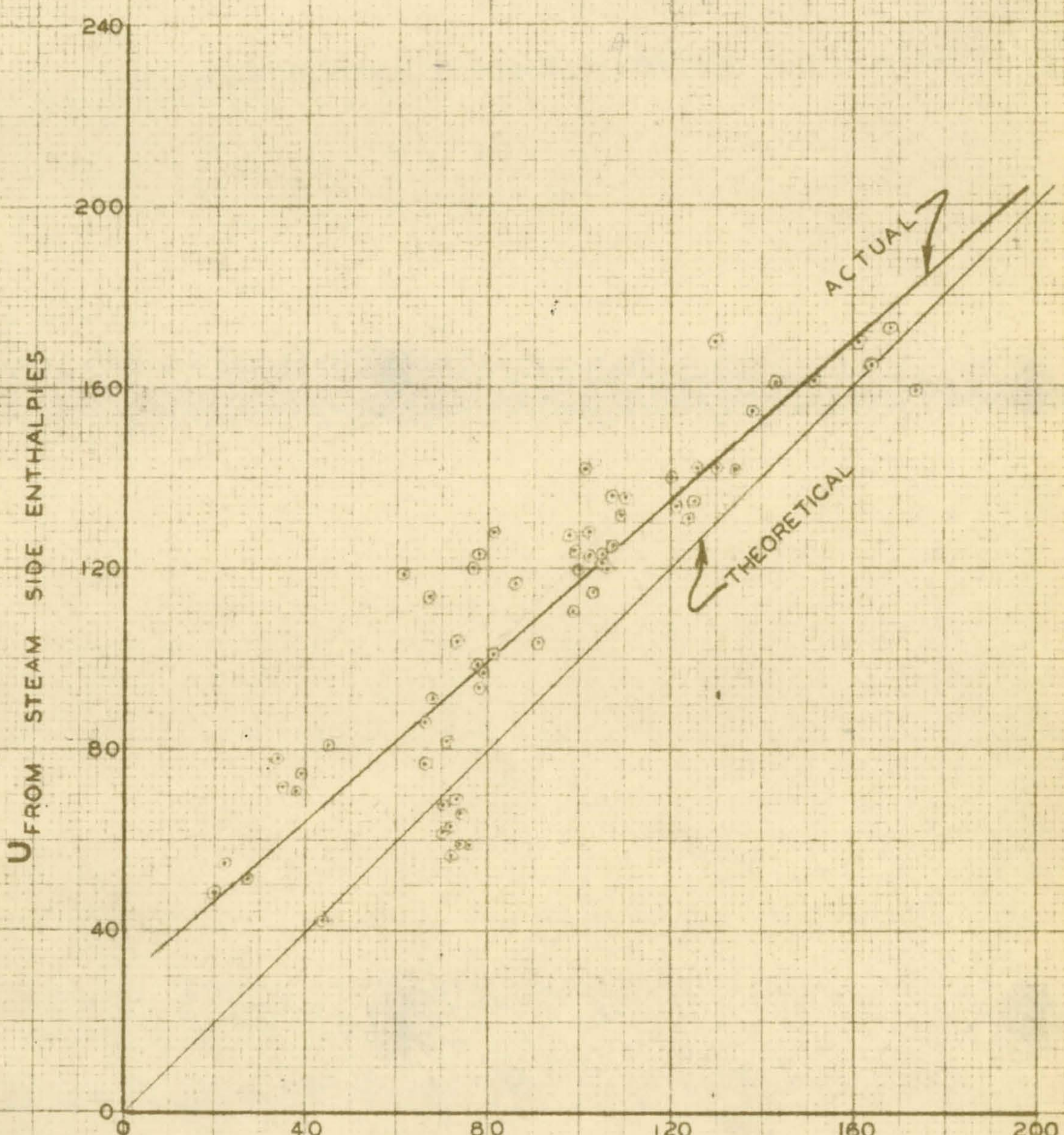

FIG. 5 COMPARISON OF OVERALL HEAT TRANSFER COEFFICIENTS

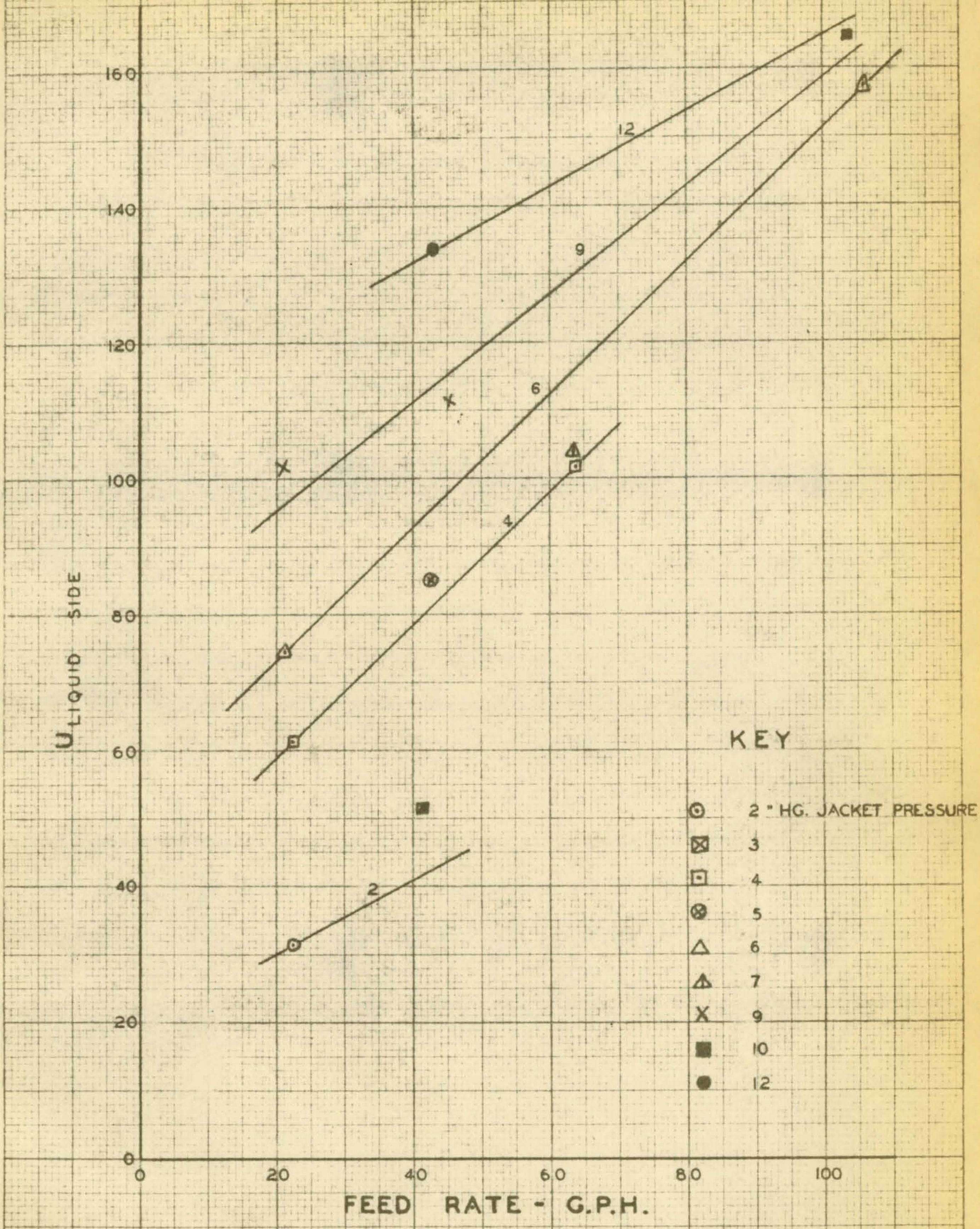

FIG. 6 $U_{LIQUID\ SIDE}$ VERSUS FEED RATE

Experimental Results - Alcohol Runs

The data of the alcohol runs, besides including that for heat transfer relations, contained necessary data for fractionation studies. The procedure was to mix ethanol and water to approximate a desired composition and to make any temperature adjustment as required. Feed was then introduced to the column and maintained constant. Product samples were taken until volumetric rates were constant. Spot samples were taken after this point and later mixed for average analyses. Alcohol compositions were obtained by a standard laboratory procedure of laboratory fractionation and refractometric measurement. The latest edition of the Wagner Tables was consulted for conversion of refractometer readings to percent alcohol by volume. The data for heat transfer on the alcohol mixtures are presented in Fig. 7. This figure shows the relation of overall heat transfer coefficients to the alcohol feed rate of the column.

The results of the fractionation are presented in Table IV. This summary data sheet is supplemented by additional information concerning these runs and is included in the Appendix (Table XI).

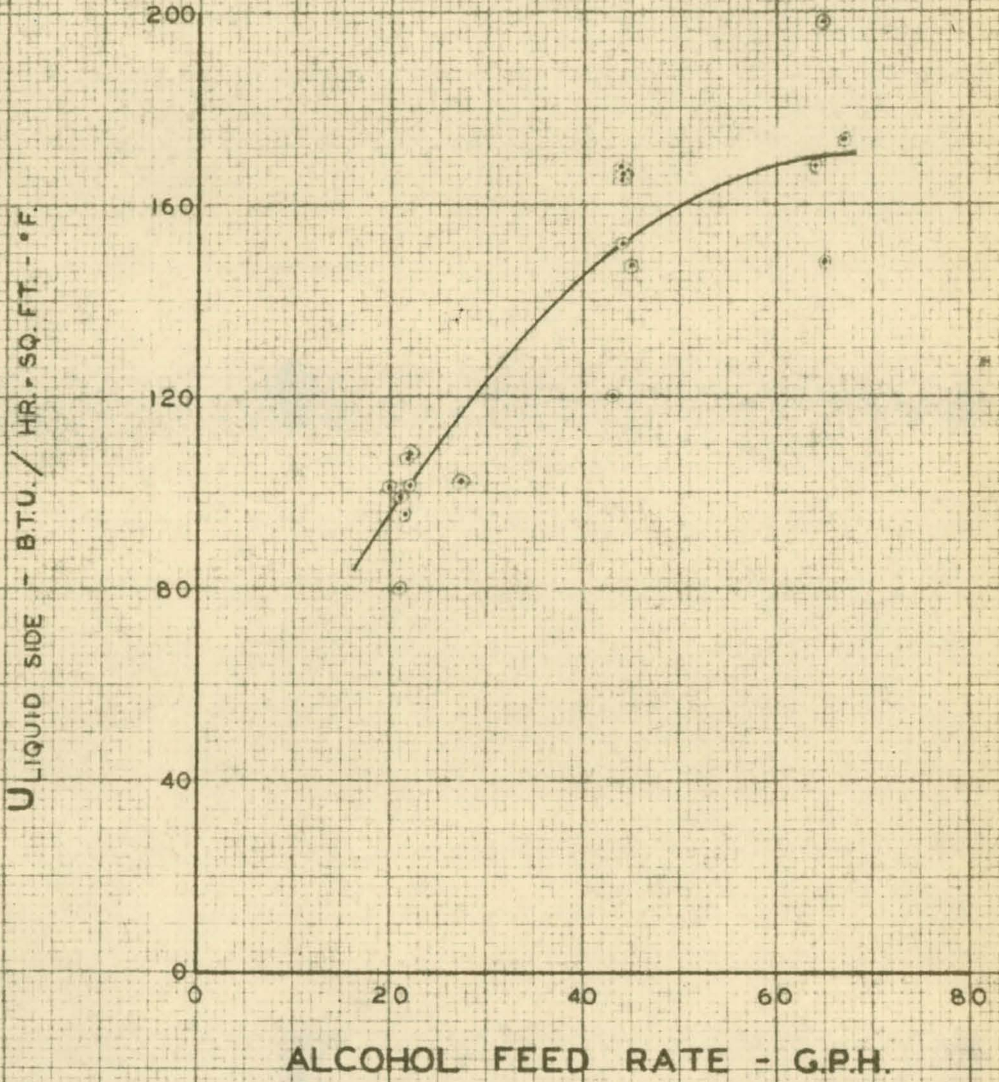

FIG. 7 $U_{LIQUID\ SIDE}$ VERSUS ALCOHOL FEED RATE

Table IV - Summary Data for Ethanol Runs

Liquid Side Coefficients

Run No.	Abs. Pressure (p.s.i.)	Flow Rate (G.P.H.)	Total Heat (Btu./hr.)	LMTD °F.	U liq.
AA-1	18.14	27.30	52510.68	44.0	102.09
AA-2	18.04	21.92	56369.39	48.0	101.30
AA-3	17.92	46.09	70562.57	41.1	146.86
AA-4	17.99	66.91	81437.67	40.1	173.73
AA-5	18.97	21.22	64845.12	56.0	99.05
AA-6	19.05	43.85	79577.68	44.7	152.29
AA-7	19.24	64.42	88749.52	45.3	167.59
AA-8	17.89	19.82	54804.11	46.3	101.25
AA-9	17.84	43.85	77308.78	40.1	164.92
AA-10	18.16	21.44	55463.63	59.0	80.42
AA-11	18.14	42.77	70698.36	50.6	119.52
AA-12	18.19	64.58	85278.15	49.2	148.27
AA-13	17.70	21.38	67432.85	60.4	95.50
AA-14	17.75	44.28	97998.16	50.6	165.67
AA-15	19.12	21.65	75873.42	60.6	107.10
AA-16	18.24	21.92	63337.75	50.2	107.93
AA-17	18.11	43.69	84587.99	43.0	168.28
AA-18	18.06	64.75	99644.20	43.0	198.23

Table IV - Summary Data for Ethanol Runs

Composition Data

Run No.	Steam Pressure "Hg ga.	Steam Temperature °F.	Compositions Feed wt. %	Tops wt. %	Bottoms wt. %
AA-1	7.00	223	3.39	17.470	0.192
AA-2	6.80	223	4.39	16.220	0.064
AA-3	6.55	222	4.39	28.350	0.584
AA-4	6.70	222	4.39	38.600	1.150
AA-5	8.70	225	4.024	12.266	0.000
AA-6	8.85	225	4.024	23.960	0.256
AA-7	9.25	226	4.024	30.718	0.703
AA-8	6.50	222	3.912	13.548	0.032
AA-9	6.40	222	3.912	22.888	0.392
AA-10	7.05	223	3.848	15.930	0.096
AA-11	7.00	223	3.848	28.117	0.584
AA-12	7.10	223	3.848	39.930	1.214
AA-13	6.10	222	11.422	29.776	0.128
AA-14	6.20	222	11.422	41.354	1.158
AA-15	9.00	226	11.422	25.295	0.096
AA-16	7.20	223	8.692	25.818	0.128
AA-17	6.95	223	8.692	39.589	0.894
AA-18	6.85	223	8.692	45.066	1.942

CORRELATION I

A development of theoretical relations for film type distillation by a direct approach appeared to be extremely complicated; therefore, an indirect method of attack was utilized. Previous investigators (9) have shown a similiarity between batch design calculations for falling film boilers and experimentally determined data. Their work assumed a limitation to substances whose vapor-liquid equilibrium followed Raoult's Law closely. This limitation can be circumvented by the use of some algebraic equation to relate the calculated values to the actual values. In the development of this correlation, Rayleigh's Equation for differential distillation should apply since by definition of this process, the composition of the liquid changes continually during the distillation. From the Rayleigh Equation

$$\ln \frac{B}{F} = \int_{x_F}^{x_B} \frac{dx}{y - x}$$

where

B = Mols (lbs.) of stripped material,
F = Mols (lbs.) of feed material,
X_B = Composition of stripped material,
X_F = Composition of feed material,

there are as knowns; B, F, X_F; therefore it is possible to find X_B by graphical integration. This procedure yields a calculated composition of the stripped material from which the function of the actual experimental composition may be evaluated. The summary data from this procedure are presented in Table V.

Table V - Calculated Compositions by Rayleigh Procedure

Run No.	Feed Rate (G.P.H.)	Bottoms Composition (wt. %)	Calculated Bottoms Composition (wt. %)
AA-1	27.30	0.192	0.452
AA-2	21.92	0.064	0.271
AA-3	45.09	0.584	1.447
AA-4	66.91	1.150	2.154
AA-5	21.22	0.000	0.000
AA-6	43.85	0.256	0.850
AA-7	64.42	0.703	1.575
AA-8	19.82	0.032	0.280
AA-9	43.85	0.392	1.047
AA-10	21.44	0.096	0.189
AA-11	42.77	0.584	1.368
AA-12	64.58	1.214	2.285
AA-13	21.38	0.128	0.000
AA-14	44.28	1.158	2.102
AA-15	21.65	0.096	0.000
AA-16	21.92	0.128	0.226
AA-17	43.69	0.894	1.633
AA-18	64.75	1.942	3.124

The relation between the actual and calculated Bottoms Composition may be expressed by the approximate equation:

$$X_{calculated} = (1.96)\,(X_{actual})^{0.7405}$$

This equation was derived from the curve shown in Fig. 8 which becomes a straight line when plotted on log-log coordinates as presented in Fig. 9.

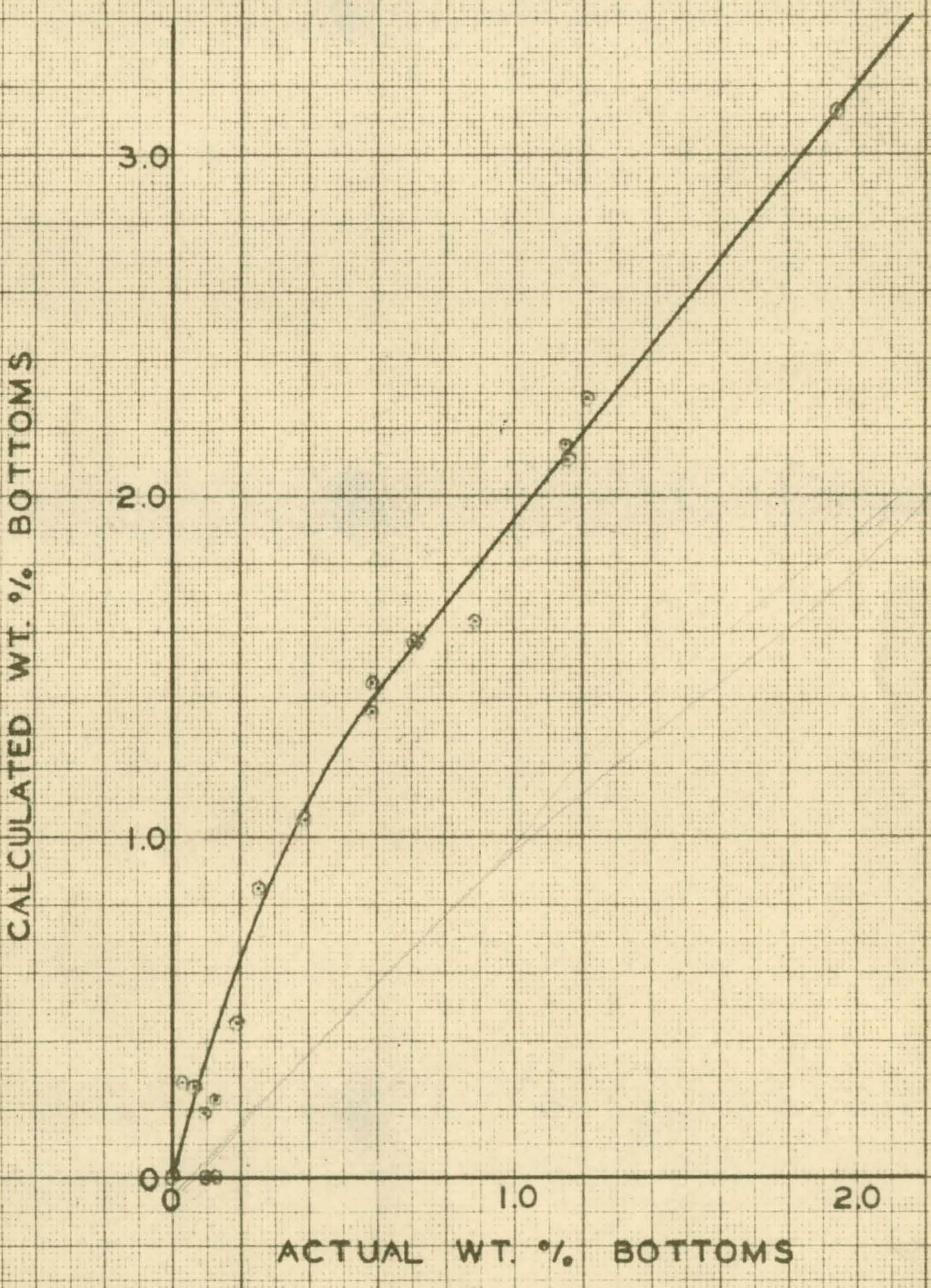

FIG. 8 CALCULATED versus ACTUAL BOTTOMS COMPOSITION

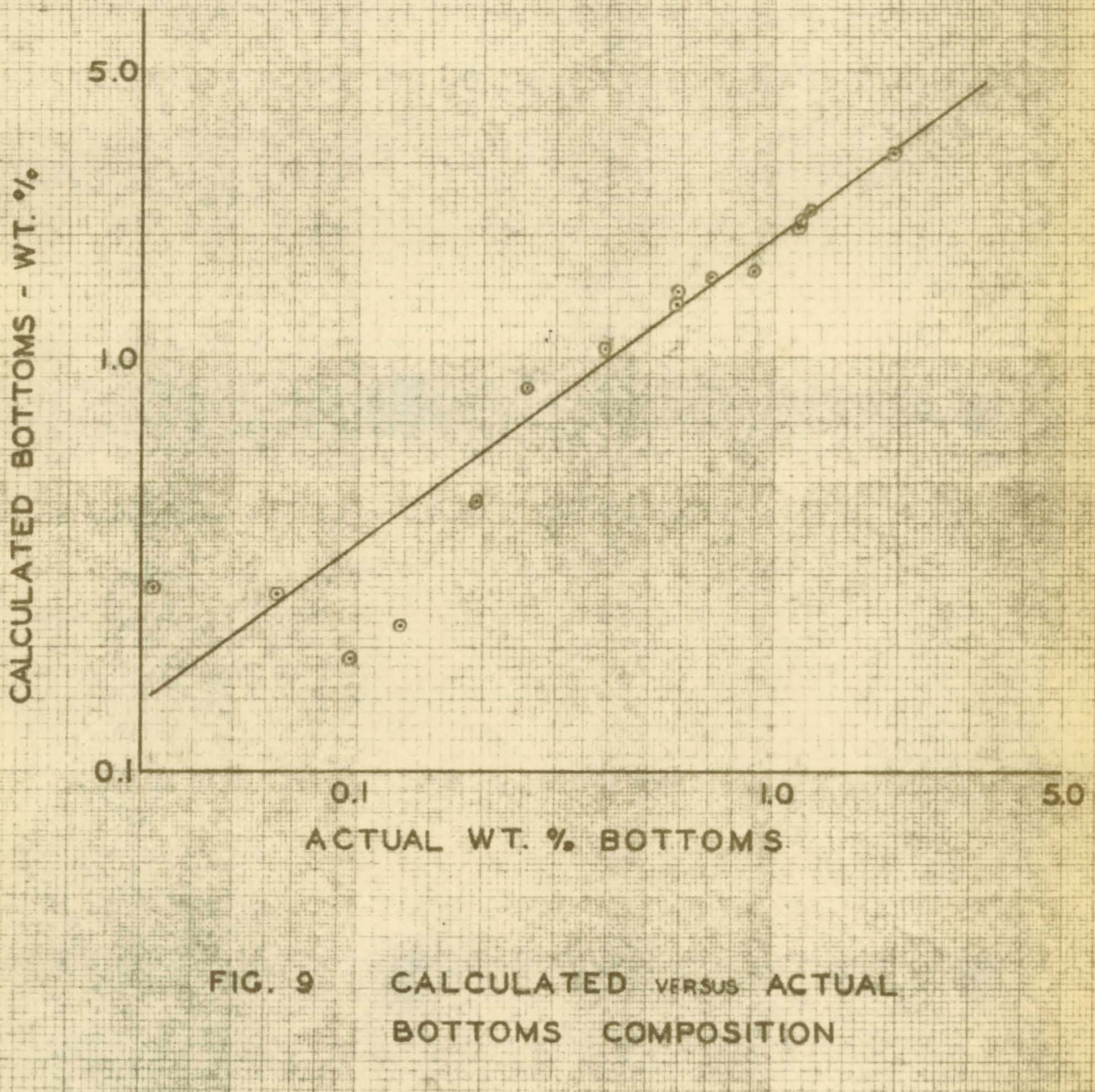

FIG. 9 CALCULATED VERSUS ACTUAL BOTTOMS COMPOSITION

CORRELATION II

A second correlation was attempted on the assumption of a constant rate of heat transfer from steam to falling liquid, and with the resultant vapor passing off without subsequent material interchange between vapor and liquid.

This assumption leads to a calculation of vapor and liquid rates throughout the column and therefore a value for the slope of the equilibrium operating line.

This procedure failed, as the slope of the operating line was not a continuously increasing function from feed entrance to liquid exhausting position as postulated. It appeared that a point of inflection of the operating line was necessary, and the concept was untenable for this reason.

CORRELATION III

Number of Theoretical Plates at Total Reflux

Although this method is theoretically unsound for reporting data for distillation operations in which there is no external reflux, it does present characteristics of distillation equipment which have some functional value.

The data from this procedure are:

Table VI - Number of Theoretical Plates at Total Reflux

Run No.	Feed Rate G.P.H.	No. of Theoretical Plates at Total Reflux
AA-1	27.10	2.37
AA-2	21.90	2.87
AA-3	45.15	1.96
AA-4	66.70	1.87
AA-5	21.20	4.10
AA-6	44.40	2.36
AA-7	63.60	1.94
AA-8	19.80	3.00
AA-9	43.90	1.98
AA-10	21.40	2.69
AA-11	41.90	1.88
AA-12	63.60	1.87
AA-13	20.20	1.13
AA-14	41.90	1.05
AA-15	19.20	0.78
AA-16	21.83	2.78
AA-17	43.70	1.96
AA-18	64.10	1.83

An examination of the data indicates the critical importance of feed rate which has been previously pointed out by Surowiec and Furnas (3).

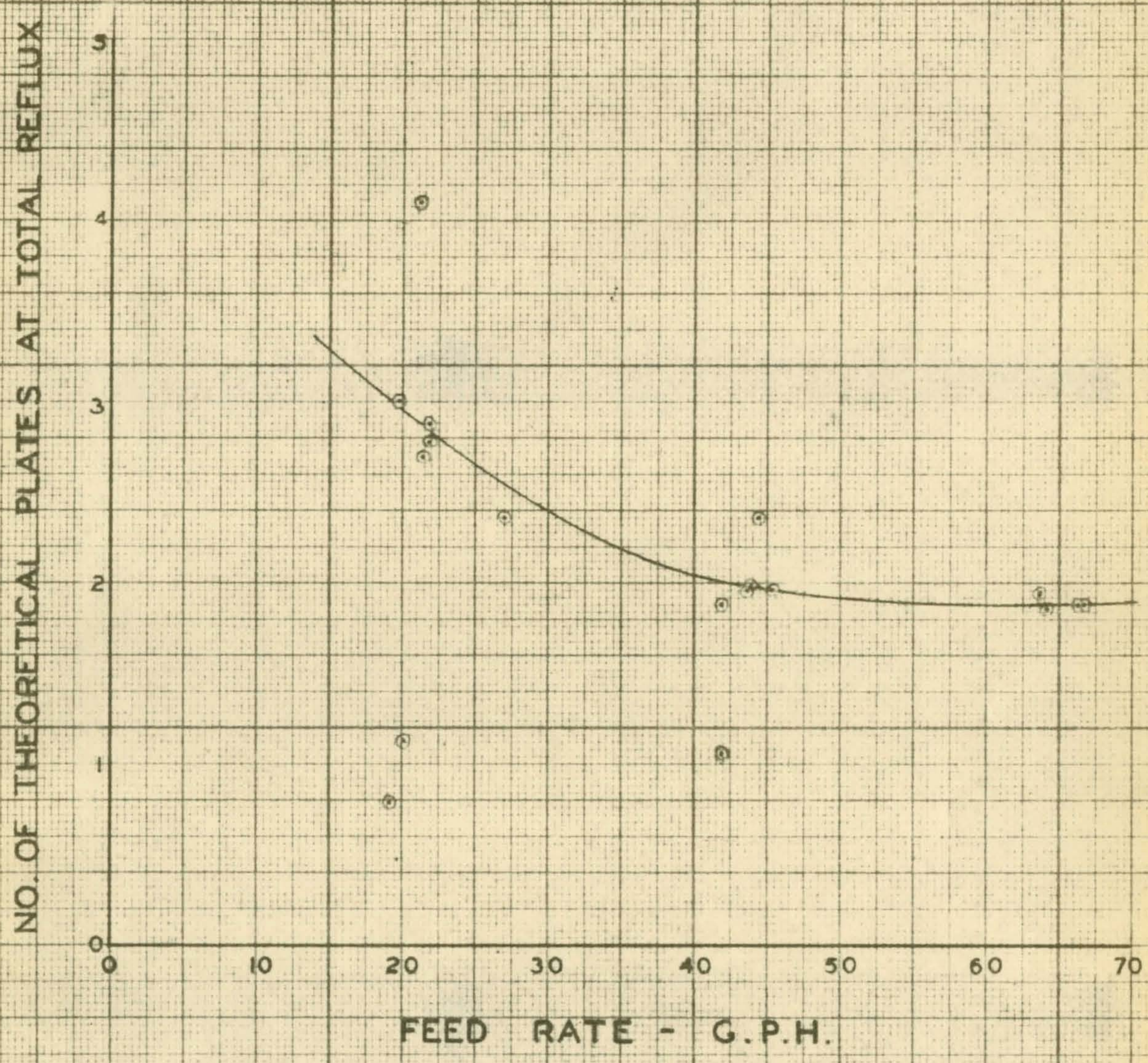

FIG. 10 NO. OF THEORETICAL PLATES AT TOTAL REFLUX versus FEED RATE

SUMMARY AND CONCLUSIONS

(1) Heat transfer data on the film type still indicate fairly normal overall heat transfer coefficients. The only data available on a unit of comparable size is that presented by Bays in the Mc Adams (10) text. The heat transfer data presented by Mc Adams (10) and shown in Fig. 11, were obtained on a column of 2.5 inches in diameter and 2.0 feet high. Their data are for film flow in a vertical tube heater, without vaporization. The Nüsselt Equation would indicate that the transfer coefficient is inversely proportional to the tube length. The 15.56 feet long tube used in this investigation is considerably longer than that of Bays, and therefore the lower coefficients are qualitatively predictable. Data shown in the author's curve are only for runs in which little or no vaporization occurred.

(2) Higher values of the overall heat transfer coefficient were obtained in the alcohol runs when compared with the data of the water runs. However, this difference becomes smaller as higher feed rates are utilized.

(3) An alcohol-water mixture of approximately 7% by volume can be stripped completely by proper balance of feed rate and percent vaporization.

(4) The higher the concentration of alcohol in the feed mixture, the greater is the entrainment and the more difficult it is to dealcoholize the feed in a film type column.

(5) Change in flow rate has a very noticeable effect on fractionation and stripping efficiency, whereas changes in steam jacket pressure

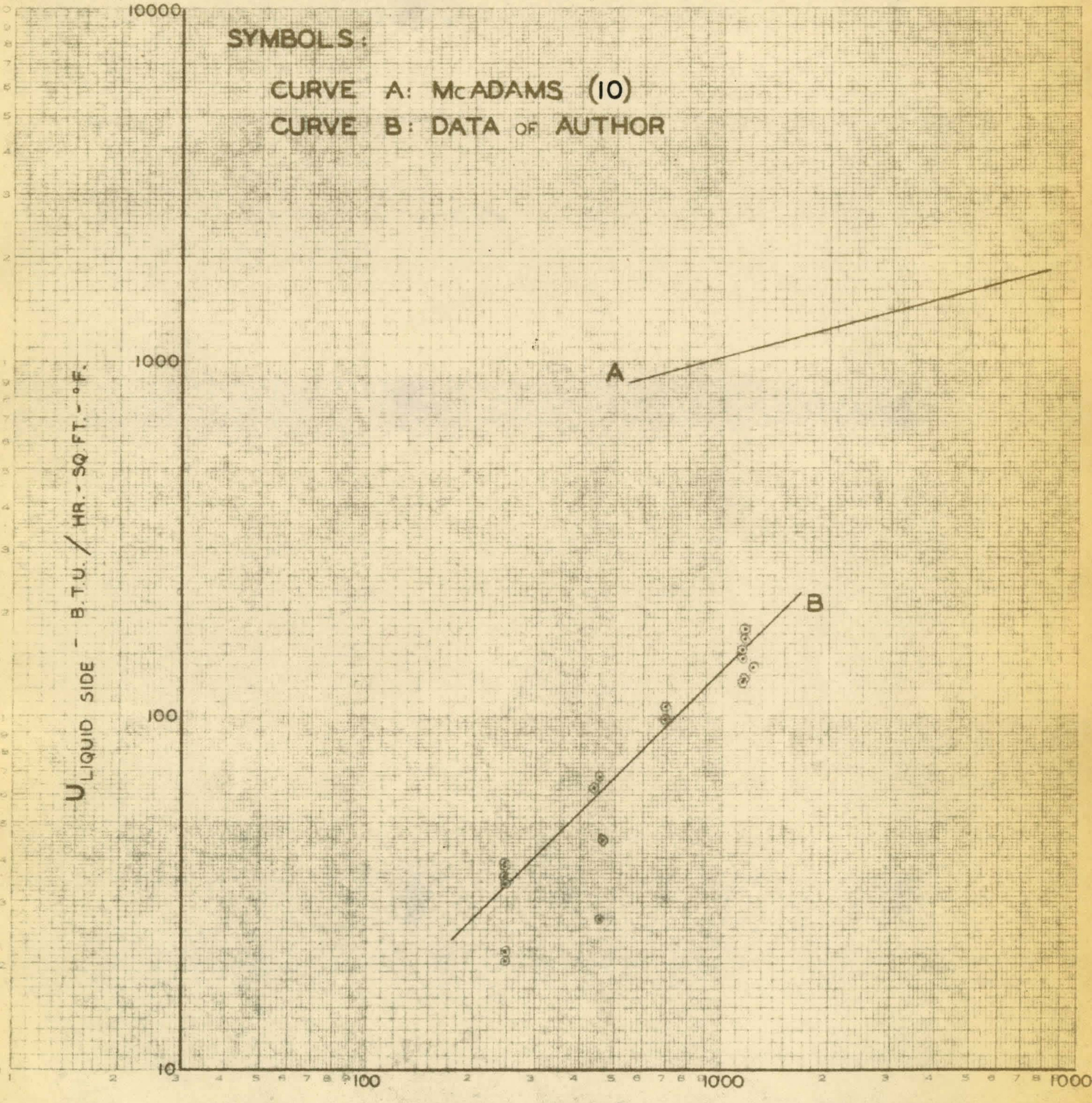

FIG. 11 COMPARISON OF FALLING FILM HEATERS

or feed temperature have apparently little effect.

(6) The higher the alcohol concentration, the lower the number of theoretical plates.

(7) The number of theoretical plates is an inverse function of the feed rate.

(8) A modification of the Rayleigh Equation probably affords the best method of evaluating film still operation.

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APPENDIX

TABLE VII - PHYSICAL MEASUREMENTS ON FILM TYPE DISTILLATION UNIT

FILM STILL		
	Inside diameter	2 15/16"
	Height	15' 6 3/4"
	Wall thickness	3/32"
	Metallic surface	Copper
	Heat transfer area	11.69 sq. ft.
VAPOR HEAD		
	Inside diameter	11 3/4"
	Volume	1,190 cu. in.
ENTRAINMENT SEPARATOR		
	Volume	1,155 cu. in.
CONDENSER		
	Tube diameter	7/8"
	Tube length	69 1/2"
	Number of tubes	14.0
	Metal composition	Copper
	Heat transfer surface	18.47 sq. ft.
	Type system	Water in shell
FEED TANK		
	Capacity	50.0 gallons
TOPS RECEIVER		
	Capacity	55.0 gallons

TABLE VII - PHYSICAL MEASUREMENTS ON FILM TYPE DISTILLATION UNIT (Cont.)

BOTTOMS RECEIVER	
Capacity	55.0 gallons
ROTAMETER	
Range	3-106 g. p. h.
Sp. gr. specification	1.05

TABLE VIII - SUMMARY COMPARISON OF INDUSTRIAL BEER STRIPPING EQUIPMENT

SIMPLICITY OF DESIGN

UNIT	CONSTRUCTION	MAINTENANCE	PRESSURE DROP	EFFICIENCY	CAPACITY	LIQUOR HOLD-UP	ENTRAINMENT	FOULING TENDENCY	EASE OF CLEANING	COST
Sieve Plate	Poor	Poor	Fairly High	High	High	High	Medium	Low	Fair	High
Bubble Cap[1]	Poor	Poor	Very High	High	Moderate	Very High	High	High	Poor	High
Packed Tower[1]	Fair	Fair	Very High	Good	Moderate	High	Very High	High	Poor	Medium
Batch Still	Fair	Fair	Low	Low	Low	Low	High	Moderate	Fair	Low
Spray Still	Poor	Poor	Low	Low	Very Low	High	High	Low	Good	Very High
Cascade Still	Good	Good	Fairly Low	High	Moderate	High	Low	Low	Good	Low
Film Still	Good	Good	Extremely Low	Fair	Low	Very Low	Low	Low	Excellent	Low

1 Commercially unsatisfactory as beer columns.

Table IX - Sample Calculations

(1) Heat Losses from Column

Run SB-7

Data: Jacket Pressure: 3.0" Hg. = 16.169 psi abs.

Condensate Collected: 82 cc./min. average.

The true Latent Heat is the difference between the Enthalpy of the Saturated Vapor (at the line pressure) and the specific enthalpy of the Saturated Boiling Liquid, therefore:

$$\Delta H_{\text{vapor}} - \Delta H_{\text{liquid}} = \text{True Latent Heat}$$

1171.9 Btu./lb. – 184.95 Btu./lb. = 986.95 Btu./lb.

$$\frac{(82 \text{ cc./min.})\ (60 \text{ min./hr.})\ (8.337\ \#/\text{gal.})}{(3{,}785 \text{ cc./gal.})} = 10.84\ \#/\text{hr.}$$

(10.84 #/hr.) (986.95 Btu./lb.) = 10,698.9 Btu./hr.

(2) Overall Heat Transfer Coefficients

(a) Based on Steam Side Enthalpies: Run 7:

Data: Condensate collected : 435 cc./min.

Feed Temperature: 138.2 ° F.

Bottoms Temperature: 162.5° F.

Steam Temperature: 225.7 °F.

$$\frac{(435)(60)(8.337)}{(3{,}785)} = 57.6\ \#/\text{hr.}$$

True Latent Heat: 961.6 Btu./lb.

(57.6)(961.6) = 55,388.0 Btu./hr.

Heat Loss from curve of Fig. 3 for this run amounts to 17,920 Btu./hr.

Net Heat: 55,388.0 – 17,920 = 37,468 Btu./hr.

$$LMTD = \frac{(225.7 - 138.2) - (225.7 - 162.5)}{\ln \frac{87.5}{63.2}} = 75.5\ ^\circ F.$$

$$U = \frac{(37,468\ Btu/hr.)}{(11.69\ ft.^2)(75.5^\circ\ F.)} = 42.45\ Btu./hr. - ft.^2 - ^\circ F.$$

Overall Heat Transfer Coefficients

(b) Based on Liquid Side Enthalpies: Run 7

Data: Tops Product Rate: 281 cc./min. (37.15 #/hr.)

Bottoms Product Rate: 3.0 cc./min. (0.40 #/hr.)

Feed Temperature: 138.2 °F.

Bottoms Temperature: 162.5 °F.

Steam Temperature: 225.7 ° F.

Sensible heat to Vapor: (37.15) (212 − 138.2) = 2749.1 $\frac{BTU.}{Hr.}$

Latent heat to Vapor: (37.15) (970) = 36035.5 Btu./hr.

Sensible heat to Bottoms: (0.4) (162.5 − 138.2) = 9.72 Btu./hr.

Net Heat: 2749.1 + 36035.5 + 9.72 = 38,794.3 Btu./hr.

$$U: \frac{(38,794.3)}{(11.69)\ (75.5)} = 43.95\ Btu./hr. - ft.^2 - ^\circ F.$$

(3) Correlation I

Data: Feed Rate: 530.0 #/hr.

Bottoms Rate: 471.95 #/hr.

(Bottoms/Feed) = 0.8905

ln (Bottoms/Feed) = −0.116

The area on a plot of "x" vs "(1/y−x)" counting from "x_{feed}" of 4.024 wt. % equal to the value −0.116 has a calculated composition of 1.575 wt. %.

(4) Evaluating the equation constants as shown on page 34.

Two points are chosen from the curve of Fig. 9, as,

x_{actual}.	$x_{calculated}$.
0.076	0.29
2.300	4.20

Substituting these values in the general equation of the curve shown gives:

$$0.29 = a(0.076)^m \text{ and } 4.2 = a(2.8)^m$$

Solution by simultaneous equations gives:

$$\log 0.29 = \log a + m \log (0.076)$$

$$-1.238 = \log a + m (-2.579)$$

$$\log 4.2 = \log a + m \log (2.8)$$

$$1.43508 = \log a + m (1.02962)$$

$$2.67308 = m (3.60862)$$

$$m = 0.7405$$

$$1.4351 = \log a + 0.762$$

$$a = 1.96$$

Table X (a) - Data on Water Runs

Run No.	Feed Rate (G.P.H.)	Steam Rate (#/hr.)	Product Rates Tops (#/hr.)	Product Rates Bottoms (#/hr.)	Temperatures Feed °F.	Temperatures Steam °F.	Temperatures Bottoms °F.	Temperatures LMTD °F.
W-1	11.20	45.00	22.00	68.00	195.8	218.4	188.6	26.0
W-2	11.80	83.80	49.64	48.91	137.1	225.8	199.4	51.4
W-3	21.44	82.10	42.50	127.24	136.9	225.8	208.9	43.4
W-4	13.50	59.75	44.29	71.39	137.3	225.8	192.2	56.8
W-5	12.60	58.56	42.83	66.76	140.9	225.8	192.2	55.3
W-6	12.60	57.42	42.04	65.31	139.6	225.9	192.2	56.0
W-7	4.10	57.51	37.15	0.40	138.2	225.7	162.5	75.5
W-8	3.20	57.77	30.80	0.00	131.0	225.7	212.0	41.9
W-9	22.25	56.85	31.60	154.81	139.1	225.6	201.2	49.0
W-10	21.82	56.85	30.93	152.29	136.4	225.8	201.2	50.2
W-11	21.71	56.58	29.88	151.24	135.5	225.6	201.2	50.3
W-12	33.05	57.37	20.49	260.70	135.0	225.6	205.7	46.7
W-13	32.40	57.64	19.30	253.03	133.7	225.7	205.7	47.1
W-17	46.87	89.90	33.71	362.89	140.4	225.8	210.2	41.1
W-18	46.87	89.40	31.46	356.94	140.0	225.8	210.2	41.2
W-19	46.76	86.59	31.46	365.93	138.6	225.8	210.2	41.6
W-20	15.70	85.93	50.90	80.11	138.2	225.8	174.2	67.9
W-21	15.30	86.72	51.03	75.35	138.2	225.8	183.2	62.3
W-22	27.43	83.29	40.98	188.12	138.2	226.1	205.7	46.2
W-23	26.78	84.61	40.98	178.73	138.2	225.6	206.6	44.8
W-24	26.24	86.99	42.30	169.61	138.2	226.1	206.6	45.4
W-25	25.92	86.19	40.45	170.54	138.2	225.7	206.6	45.0

Table X (a) - Data on Water Runs
(Cont.)

	Feed Rate (G.P.H.)	Steam Rate (#/hr.)	Product Tops (#/hr.)	Rates Bottoms (#/hr.)	Temperatures Feed °F.	Steam °F.	Bottoms °F.	LMTD °F.
W-26	25.60	91.61	44.15	166.97	138.2	228.0	206.6	45.3
W-27	111.24	92.54	1.98	832.86	138.2	225.3	212.0	30.3
W-28	71.60	104.17	18.51	576.39	136.4	225.8	211.3	41.2
W-29	70.74	89.90	19.04	570.58	136.4	225.8	211.3	41.2
W-30	22.68	36.62	0.00	190.63	134.6	215.6	192.2	46.4
W-31	22.68	39.66	0.00	191.16	134.6	215.6	194.0	44.9
W-32	22.68	42.83	0.00	189.05	132.8	215.8	205.5	34.8
W-33	22.68	42.57	1.72	185.87	131.0	216.5	206.6	35.0
W-34	22.68	40.32	0.93	189.57	131.0	216.3	206.6	34.7
W-35	22.68	40.32	0.00	188.78	131.0	216.1	206.6	34.5
W-36	22.68	69.14	14.01	174.50	131.0	219.2	206.6	38.9
W-37	42.12	44.95	0.00	351.65	134.6	217.2	183.2	54.8
W-38	42.66	55.13	0.00	355.88	134.6	218.5	199.4	43.8
W-39	40.50	-	0.00	338.17	136.4	217.9	208.4	33.5
W-40	41.90	54.86	0.00	349.54	136.4	216.9	208.4	31.9
W-41	41.90	63.98	11.10	337.77	136.4	219.4	208.4	35.7
W-42	42.44	68.61	18.51	334.99	136.4	220.7	208.4	37.4
W-43	42.98	74.69	23.53	335.26	136.4	222.0	208.4	39.1
W-44	64.04	62.80	0.00	534.09	138.2	217.9	210.2	30.8
W-45	63.94	64.78	0.00	533.56	138.2	219.8	210.2	33.6
W-46	63.61	74.30	6.61	523.51	136.4	222.5	209.3	38.9

Table X (a) - Data on Water Runs
(Cont.)

Run No.	Feed Rate (G.P.H)	Steam Rate (#/hr.)	Product Tops (#/hr.)	Rates Bottoms (#/hr.)	Temperatures Feed °F.	Steam °F.	Bottoms °F.	LMTD °F.
W-47	63.94	73.11	6.61	526.68	136.4	220.9	210.2	35.7
W-48	61.56	84.61	8.86	503.95	136.4	223.6	210.2	39.4
W-49	107.46	84.87	0.00	1011.33	138.2	221.8	210.6	36.0
W-50	107.03	86.99	0.00	951.84	138.2	221.8	210.6	36.0
W-51	105.30	86.19	0.00	836.03	138.2	221.9	210.6	36.2
W-52	104.76	86.19	0.00	884.42	138.2	221.9	210.6	36.2
W-53	104.76	84.21	0.00	856.66	138.2	224.9	209.0	41.7
W-54	104.76	85.93	0.00	856.66	138.2	224.7	209.0	41.5
W-55	103.57	113.43	23.53	864.59	138.2	227.7	210.2	44.1
W-56	103.14	112.50	23.53	819.64	138.2	228.0	210.2	44.4
W-57	22.14	68.22	29.08	155.47	138.2	222.1	194.9	50.3
W-58	20.92	72.05	32.52	141.98	138.2	222.2	199.4	46.9
W-59	20.63	71.65	33.58	138.28	138.2	222.1	200.3	46.1
W-60	20.20	69.80	32.52	136.03	138.2	222.1	200.3	46.1
W-61	21.71	81.03	42.70	138.55	138.2	225.0	203.9	46.4
W-64	21.38	61.08	21.95	158.38	138.2	221.1	205.0	42.0
W-65	42.93	88.84	18.24	318.07	138.2	227.1	209.3	44.2
W-66	43.09	65.44	7.01	350.33	138.2	219.3	208.4	35.0
W-67	43.42	65.44	7.01	350.33	138.2	219.7	208.4	35.5
W-68	43.09	79.58	23.53	336.58	138.2	223.4	208.4	40.4
W-69	42.66	105.76	51.82	304.06	138.2	229.6	208.4	48.1

Table X (b) — Data on Water Runs

Run No.	HEAT QUANTITIES ON STEAM SIDE: In Steam Btu./hr.	Heat Loss Btu./hr.	Net Steam Btu./hr.	OVERALL COEFFICIENT OF HEAT TRANSFER Btu./hr. sq. ft.-°F.
W-1	43479.00	11900	31579.00	103.90
W-2	80669.85	18090	62579.85	104.15
W-3	79035.30	18090	60945.30	120.13
W-4	57497.70	18090	39407.70	59.35
W-5	56440.05	18090	38350.05	59.32
W-6	55209.30	18100	37109.30	56.69
W-7	55388.16	17920	37468.20	42.45
W-8	55676.64	17920	37756.64	77.08
W-9	57720.73	17790	39930.73	66.22
W-10	54805.50	18090	36715.50	62.56
W-11	54528.39	17790	36738.39	62.48
W-12	55297.75	17790	37507.75	68.71
W-13	55580.48	17092	37660.48	68.40
W-17	86535.00	18090	68445.00	142.46
W-18	85958.10	18090	67868.10	140.91
W-19	83458.20	18090	65368.20	134.42
W-20	82785.15	18090	64695.15	81.51
W-21	86563.04	18000	68563.04	94.14
W-22	80180.76	18260	61920.76	114.65
W-23	81447.52	17840	63607.52	121.45
W-24	83737.94	18260	65477.94	123.37

Table X (b) - Data on Water Runs
(Cont.)

Run No.	HEAT QUANTITIES ON STEAM SIDE			OVERALL COEFFICIENT OF HEAT TRANSFER
	In Steam Btu./hr.	Heat Loss Btu./hr.	Net Steam Btu./hr.	Btu./hr. sq. ft.-°F.
W-25	82986.08	17900	65086.08	123.73
W-26	88160.38	18140	70020.38	132.22
W-27	89168.13	17560	71608.13	155.86
W-28	100284.45	18090	82194.45	170.66
W-29	86535.00	18090	68445.00	142.11
W-30	35529.27	9600	25929.27	47.60
W-31	38433.57	9600	28833.57	54.88
W-32	41585.28	9700	31885.28	78.42
W-33	41219.76	10300	30919.76	75.48
W-34	39095.08	10100	28995.08	71.48
W-35	39099.12	10000	29099.12	72.19
W-36	66826.44	12480	54346.44	119.66
W-37	43515.00	10930	32585.00	50.86
W-38	53328.72	12000	41328.72	80.64
W-39	-	11500	-	-
W-40	53196.00	10640	42556.00	114.05
W-41	61792.00	12720	49072.00	117.68
W-42	66274.89	13680	52594.89	120.42
W-43	72003.33	14740	57263.33	125.37
W-44	60792.85	11500	49292.85	136.82

Table X (b) - Data on Water Runs - (Cont.)

Run No.	HEAT QUANTITIES ON STEAM SIDE: In Steam Btu./hr.	Heat Loss Btu./hr.	Net Steam Btu./hr.	OVERALL COEFFICIENT OF HEAT TRANSFER Btu./hr. sq. ft.-°F.
W-45	62647.97	12960	49687.97	126.61
W-46	71684.40	15210	56474.40	124.25
W-47	70608.72	13780	56828.72	136.16
W-48	81557.63	16020	65537.63	142.37
W-49	81458.00	14610	66848.00	158.84
W-50	83858.40	14610	69248.40	164.55
W-51	83184.57	14720	68464.57	161.78
W-52	83184.57	14720	68464.57	161.78
W-53	81192.80	17200	63992.80	131.27
W-54	82740.60	17090	65650.60	135.23
W-55	109005.40	19630	89375.40	173.37
W-56	108107.26	19840	88267.26	170.02
W-57	65827.54	14840	50987.54	88.71
W-58	69482.77	14960	54522.77	99.45
W-59	69097.29	14840	54257.29	100.67
W-60	67369.62	14840	52529.62	97.47
W-61	78018.20	17320	60698.20	111.90
W-64	58924.84	14010	44914.84	91.48
W-65	85348.60	19160	66188.60	128.10
W-66	63349.92	12960	50389.92	123.16
W-67	63323.68	13270	50053.68	120.61
W-68	76712.58	16250	60462.58	128.02
W-69	101462.20	21500	79962.20	142.21

Table X (e) - Heat Transfer Based on Liquid Side Data

Run No.	HEAT QUANTITIES ON LIQUID SIDE Vapor (sensible) Btu./hr.	Vapor (latent) Btu./hr.	Bottoms (sensible) Btu./hr.	Net Heat Btu./hr.	Coefficient of Heat Transfer Btu./hr.-ft.^{2}-°F.
W-1	356.40	21340.00	489.60	22186.00	72.99
W-2	3723.00	48150.80	3047.09	54920.89	90.40
W-3	3191.75	41225.00	9161.28	53578.03	105.60
W-4	3308.46	42961.30	3919.31	50189.07	75.59
W-5	3040.93	41545.10	3424.79	48010.82	74.27
W-6	3026.88	40778.80	3435.31	47240.99	72.16
W-7	2749.10	36035.50	9.72	38794.32	43.95
W-8	2494.80	29876.00	-	32370.80	66.09
W-9	2306.80	30652.00	9613.70	42572.50	74.32
W-10	2338.31	30002.10	9868.39	42208.80	71.93
W-11	2285.82	28983.60	9936.47	41205.89	70.06
W-12	1577.73	19875.30	18431.49	39884.52	73.06
W-13	1511.19	18721.00	18218.16	38450.35	69.83
W-17	2413.64	32698.70	25329.72	60442.06	125.80
W-18	2265.12	30516.20	25057.19	57838.51	120.09
W-19	2309.16	30516.20	26200.59	59025.95	121.38
W-20	3756.42	49373.00	2883.96	56013.38	70.57
W-21	3766.01	49499.10	3390.75	56655.86	77.79
W-22	3024.32	39750.60	12698.10	55473.02	102.71
W-23	3024.32	39750.60	12225.13	55000.05	105.02
W-24	3121.74	41031.00	11601.32	55754.06	105.05

Table X (c) - Heat Transfer Based on Liquid Side Data (Cont).

Run No.	HEAT QUANTITIES ON LIQUID SIDE Vapor (sensible) Btu./hr.	Vapor (latent) Btu./hr.	Bottoms (sensible) Btu./hr.	Net Heat Btu./hr.	Coefficient of Heat Transfer Btu./hr. -ft.^{2}-°F.
W-25	2985.21	39236.50	11664.94	53886.65	102.44
W-26	3258.27	42825.50	11420.75	57504.52	108.59
W-27	146.12	1920.60	61465.07	63531.79	138.29
W-28	1399.36	17954.70	43171.61	62525.67	129.82
W-29	1439.42	18468.80	42736.44	62644.66	130.07
W-30	-	-	10980.29	10980.29	20.24
W-31	-	-	11354.90	11354.90	21.61
W-32	-	-	13743.94	13743.94	33.80
W-33	139.32	1668.40	14051.77	15859.49	38.72
W-34	75.33	902.10	14331.49	15308.92	37.74
W-35	-	-	14271.77	14271.77	35.41
W-36	1134.81	13589.70	13192.20	27916.71	61.47
W-37	-	-	17090.19	17090.19	26.68
W-38	-	-	23061.02	23061.02	45.00
W-39	-	-	24348.24	24348.24	62.16
W-40	-	-	25166.88	25166.88	67.45
W-41	839.16	10767.00	24319.44	35925.60	86.16
W-42	1399.36	17954.70	24119.28	43473.34	99.54
W-43	1778.87	22824.10	24138.72	48741.69	106.72
W-44	-	-	38454.48	38454.48	106.73
W-45	-	-	38416.32	38416.32	97.89

Table X (c) - Heat Transfer Based on Liquid Side Data (Cont.)

Run No.	HEAT QUANTITIES ON LIQUID SIDE Vapor (sensible) Btu./hr.	Vapor (latent) Btu./hr.	Bottoms (sensible) Btu./hr.	Net Heat Btu./hr.	Coefficient of Heat Transfer Btu./hr.-ft.^{2}-°F.
W-46	499.72	6411.70	38163.88	45075.30	99.17
W-47	499.72	6411.70	38868.98	45780.40	109.70
W-48	669.82	8594.20	37191.51	46455.53	100.91
W-49	-	-	73220.29	73220.29	173.99
W-50	-	-	68913.22	68913.22	163.75
W-51	-	-	60528.57	60528.57	143.03
W-52	-	-	64032.01	64032.01	151.31
W-53	-	-	60651.53	60651.53	124.42
W-54	-	-	60651.53	60651.53	124.93
W-55	1736.51	22824.10	62250.48	86811.09	168.39
W-56	1736.51	22824.10	59014.08	83574.69	160.98
W-57	2146.10	28207.60	8737.41	39091.11	66.48
W-58	2399.98	31544.40	8689.18	42633.56	77.76
W-59	2478.20	32572.60	8587.19	43637.99	80.98
W-60	2399.98	31544.40	8447.46	42391.84	78.66
W-61	3151.26	41419.00	9102.74	53673.00	98.95
W-64	1619.91	21291.50	10579.78	33491.19	68.21
W-65	1346.11	17692.80	22614.78	41653.69	80.62
W-66	517.34	6799.70	24593.17	31910.21	77.99
W-67	517.34	6799.70	24593.17	31910.21	76.89
W-68	1736.51	22824.10	23627.92	48188.53	102.03
W-69	3824.32	50265.40	21345.01	75434.73	134.16

Table XI - Data on Alcohol Runs

Run No.	Feed Rate #/hr.	Tops Product #/hr.	Bottoms Product #/hr.	Temperatures Feed °F.	Bottoms °F.	Vapor °F.	LMTD °F.
AA-1	226.2	42.30	179.79	140.0	203.0	208.6	44.0
AA-2	182.4	48.91	133.26	134.6	201.2	208.6	47.6
AA-3	376.0	50.50	327.33	134.6	207.5	205.4	41.1
AA-4	556.0	48.12	506.33	134.6	208.4	203.2	40.1
AA-5	176.7	58.70	116.34	134.6	194.0	209.6	56.0
AA-6	370.0	60.55	308.03	134.6	207.3	207.2	44.7
AA-7	530.0	57.64	471.95	134.6	208.4	205.0	45.3
AA-8	165.0	47.86	122.95	134.6	201.2	209.6	46.3
AA-9	366.0	57.37	313.31	134.6	208.4	207.5	40.1
AA-10	178.2	43.10	129.66	105.8	198.5	209.0	59.0
AA-11	349.0	40.59	308.03	105.8	207.3	205.5	50.6
AA-12	530.0	35.96	495.75	105.8	208.4	202.1	49.2
AA-13	168.3	63.46	93.86	111.2	194.0	205.4	60.4
AA-14	349.0	83.55	264.40	111.2	204.8	199.4	50.6
AA-15	159.9	71.92	80.64	111.2	199.4	206.8	60.6
AA-16	182.0	60.94	116.60	136.4	197.6	206.8	50.2
AA-17	364.0	74.83	285.55	136.4	205.7	202.1	43.0
AA-18	534.0	80.91	453.45	136.4	205.8	199.0	43.0

Table XI - Data on Alcohol Runs (Cont.)

Run No.	Latent Heat Of Vapor	HEAT TO VAPOR Sensible	Latent	Heat To Bottoms	Total Heat Transferred	U
AA-1	905	2901.78	38281.50	11327.40	52510.68	102.09
AA-2	897	3619.34	43872.27	8877.78	56369.39	101.30
AA-3	854	3575.40	43127.00	23860.17	70562.57	146.86
AA-4	847	3301.03	40757.64	37379.00	81437.67	173.73
AA-5	912	4402.50	53534.40	6908.22	64845.12	99.05
AA-6	872	4395.90	52790.88	22391.60	79577.68	152.29
AA-7	865	4057.86	49858.60	34833.60	88749.52	167.59
AA-8	899	3589.50	43026.14	8188.47	54804.11	101.25
AA-9	871	4184.46	49995.40	23128.92	77308.78	164.92
AA-10	905	4447.92	39005.50	12010.21	55463.63	80.42
AA-11	875	4033.86	35402.50	31262.00	70698.36	119.52
AA-12	860	3462.95	30925.60	50889.60	85278.15	148.27
AA-13	846	5978.87	53695.62	7758.50	67432.85	95.50
AA-14	788	7373.52	65876.80	24747.80	97998.16	165.67
AA-15	862	6864.08	61891.60	7117.70	75873.42	107.10
AA-16	852	4289.47	51912.36	7135.92	63337.75	107.93
AA-17	800	4916.99	59872.00	19799.01	84587.99	168.28
AA-18	780	5064.97	63109.80	31469.43	99644.20	198.23

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VITA

Edward Joseph Kimmel was born of Albert Joseph and Louise Bromm Kimmel on March 2, 1920, at Louisville, Kentucky. He received his primary education at St. Joseph Parochial School, his secondary training at St. Xavier High School, and his college training at the Speed Scientific School, receiving his Bachelor's Degree in Chemical Engineering in May, 1942. He entered the night division of Adult Education in July, 1943, and received his Master's Degree in Chemical Engineering from the University of Louisville in March, 1947.

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