PERFORMANCE OF THE KOCH KASKADE TRAY IN DISTILLATION

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SUMMARY

A 5 tray Koch KASKADE Column, 14" in diameter, was erected and all necessary pieces of auxiliary equipment were designed, constructed, and installed.

The tray spacing was 24" and the weir length 8".

The system was so designed as to make it possible to obtain data for rectifying, stripping, and total reflux. Vapor-liquid equilibrium data were obtained for the system cyclohexane-isooctane, analysis being made by refractive index. The performance of the Koch tower in the distillation of this system was investigated. Vapor loadings for the tower ranged from 3 to 7 feet per second, and liquid loadings ranged from 8 to 78 gallons per minute per square foot of tower cross section. The effects of liquid loading and of vapor loading on overall efficiency were about the same; in both cases increased loading resulted in decreased efficiency. An efficiency of about 45-50% was obtained for customary vapor loading of 4-4.5 feet per second at total reflux. The pressure drop per tray was found to be more responsive to vapor loading than to liquid loading. In determining this, the equation

$$\Delta P = cL^a V^b$$

was set up and a value of c of about 0.82 was found for a equal to 0.1 and b equal to 3.1; where ΔP is in inches of water, L is in pound mols of liquid per hour, and V is in pound mols of vapor per hour. Entrainment increased very rapidly with increasing vapor loading but was independent of liquid loading at low vapor loadings. However, at higher vapor loadings the liquid loading had a definite effect.

INTRODUCTION

The much used bubble-cap tray is admitted to have several weaknesses in distillation, among which are:

1. High pressure drop due to a flow reversal of the vapor in coming up the riser then back down the annulus between the cap and the riser.

2. Contact is limited to a short interval after the vapor leaves the slots, forming bubbles and rising to the surface of the liquid on the tray.

3. Quantity of liquid passing along each cap and the length of its travel is non-uniform.

4. Build-up of liquid gradient from inlet to outlet. As the loading increases this results in all of the vapor load being carried by the outer caps while liquid may flow down the inner ones.

The design of the Koch KASKADE tray is such as to remove these weaknesses[2], the principle advantages being:

1. Low pressure drop due to unrestricted flow paths.

2. Good contact as a result of violent mixing and large surface provided by the perforated baffles.

3. No liquid gradient build-up with increased loading because of the "stair-steps" arrangement. All liquid must travel the same path and receive the same treatment.

4. Path followed by the vapor-liquid mixture creates centrifugal action which separates the liquid from the vapor on the baffles. Any entrainment is knocked out by a backward curve on the end of the baffle.

This tray is now operating in commercial absorption, liquid-liquid extraction, and distillation units.[3,4] It was felt in this study, that the performance of the tray in the distillation of a hydrocarbon system should be quantitatively studied. The effects of various operating variables on efficiency and pressure drop in a laboratory plate column for conditions of total reflux and rectification have been investigated.[1] It was decided to conduct similar work on the 14" Koch tray and to include also stripping conditions as well as those of total reflux and rectification.

Isooctane (2,2,4-trimethylpentane)-cyclohexane was chosen as the system to be investigated for the following considerations:

1. Binary system for simplicity in calculation.

2. No azeotrope formation.

3. Sufficient difference in boiling points so that separation would not be difficult.

4. Boiling point of the heavier fraction is not so high as to necessitate the use of extremely high pressure steam for vaporization.

STATEMENT OF THE PROBLEM

Since there are no quantitative data readily available on the performance of the Koch KASKADE tray in distillation and since use of the tray in the Southwest area would be principally for hydrocarbon separations, the investigation concerned itself with the following:

1. Development of a convenient analytical method for the system cyclohexane-isooctane.

2. Determination of vapor-liquid equilibrium data for the above system.

3. Assembly of the Koch KASKADE column and design, construction, and installation of all auxiliaries.

4. Calibration of all temperature, pressure, and flow measuring devices.

5. Determination of overall distillation efficiencies in the Koch tower as affected by vapor and liquid loading.

6. Determination of pressure drop per tray as affected by vapor and liquid loading.

7. Determination of entrainment as affected by vapor and liquid loading.

MATERIALS USED

1. Cyclohexane and isooctane (2,2,4-trimethylpentane) were obtained from the Phillips Petroleum Company. Their properties are listed below:

	Cyclohexane	Isooctane
Grade	technical	pure
Assay	95 mol %	99 mol %
Boiling point	177.1-.3°F	210.6°F
Molecular weight	84.2	114.2
Specific heat of liquid	0.440 @ 79°F	0.489 @ 72°F
Specific gravity of liquid @ 60°F	0.783	0.696
Latent heat of vaporization @ normal b.p., BTU/#mol	12,940	13,300

2. An Abbe refractometer was used to determine refractive indices. The instrument was standardized with both distilled water and a standard prism. The temperature was controlled to 77°F ± 0.1°F by circulating water from a constant temperature bath.

3. The Othmer still[5] was used to obtain vapor-liquid equilibrium data.

4. The column used was fourteen inches in diameter and contained five Koch KASKADE trays. Refer to Figs. 2,3,4, and 5.

5. A throttling calorimeter to check the quality of the steam was made from std. 2" pipe fittings and a 3/4" needle valve.

6. Specifications for other auxiliary equipment such as pumps, thermometers, gages, flow meters, and heat exchangers can be found in the appendix.

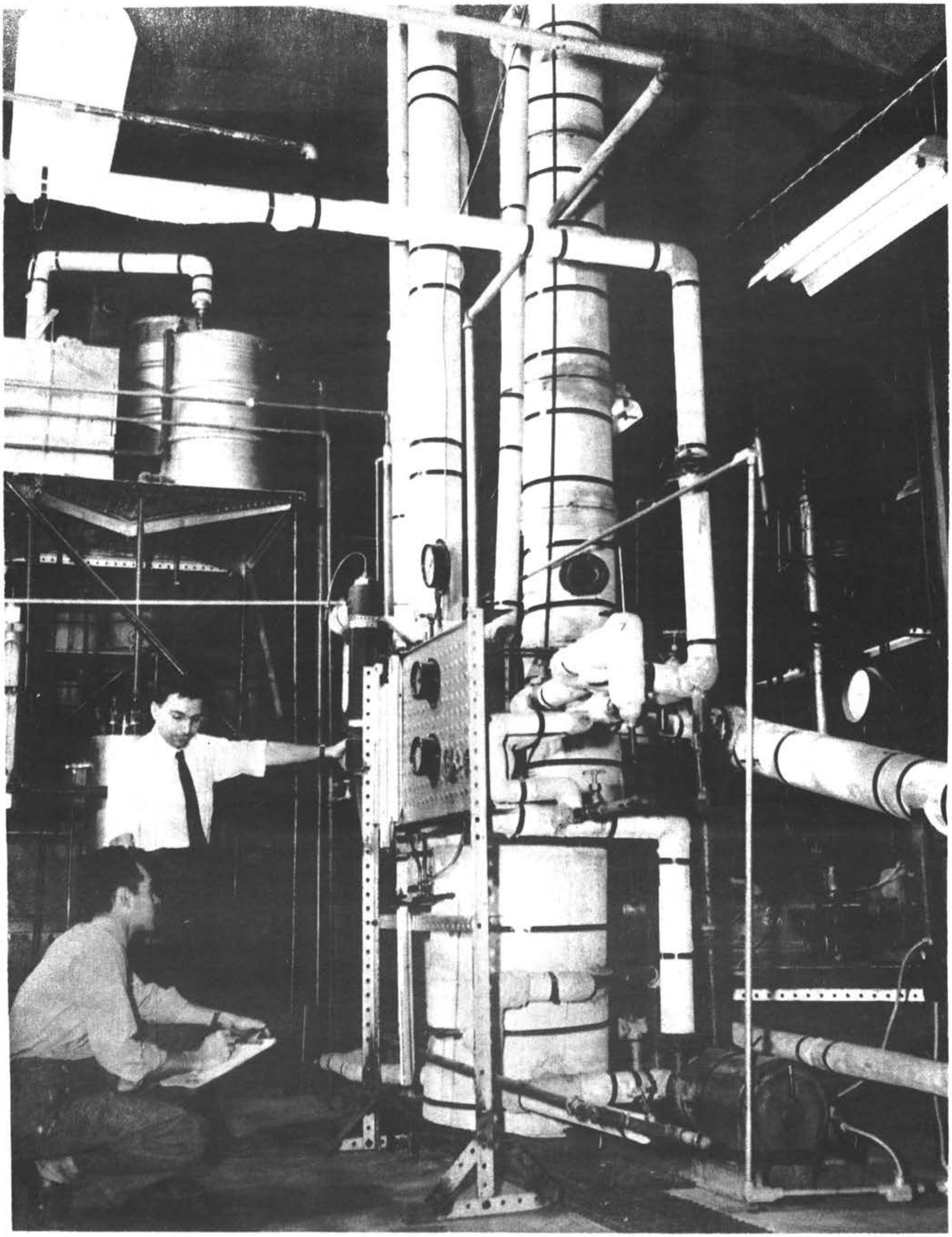

Figure 1. View of the Column and Auxiliaries

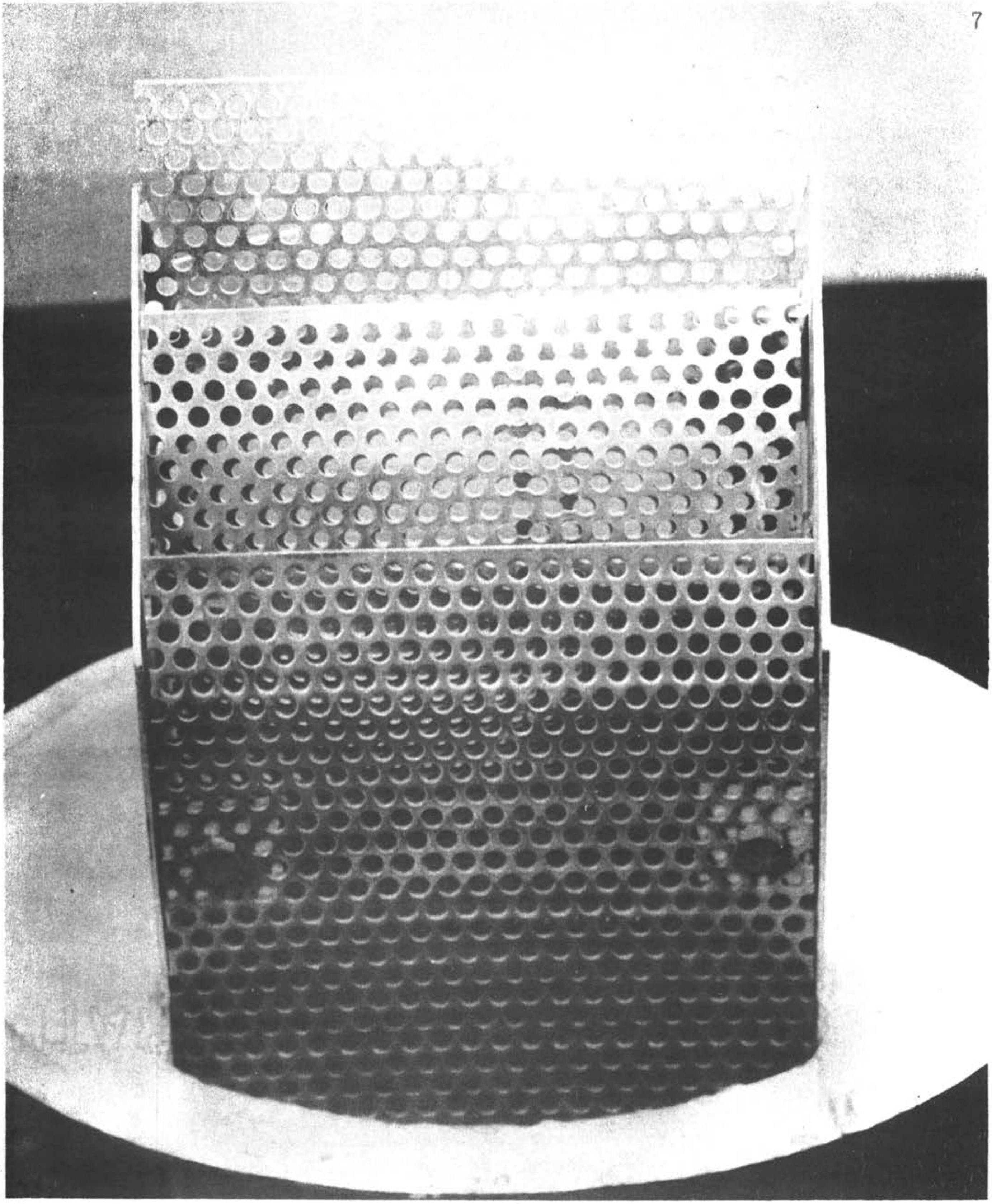

Figure 2. Top View of Tray

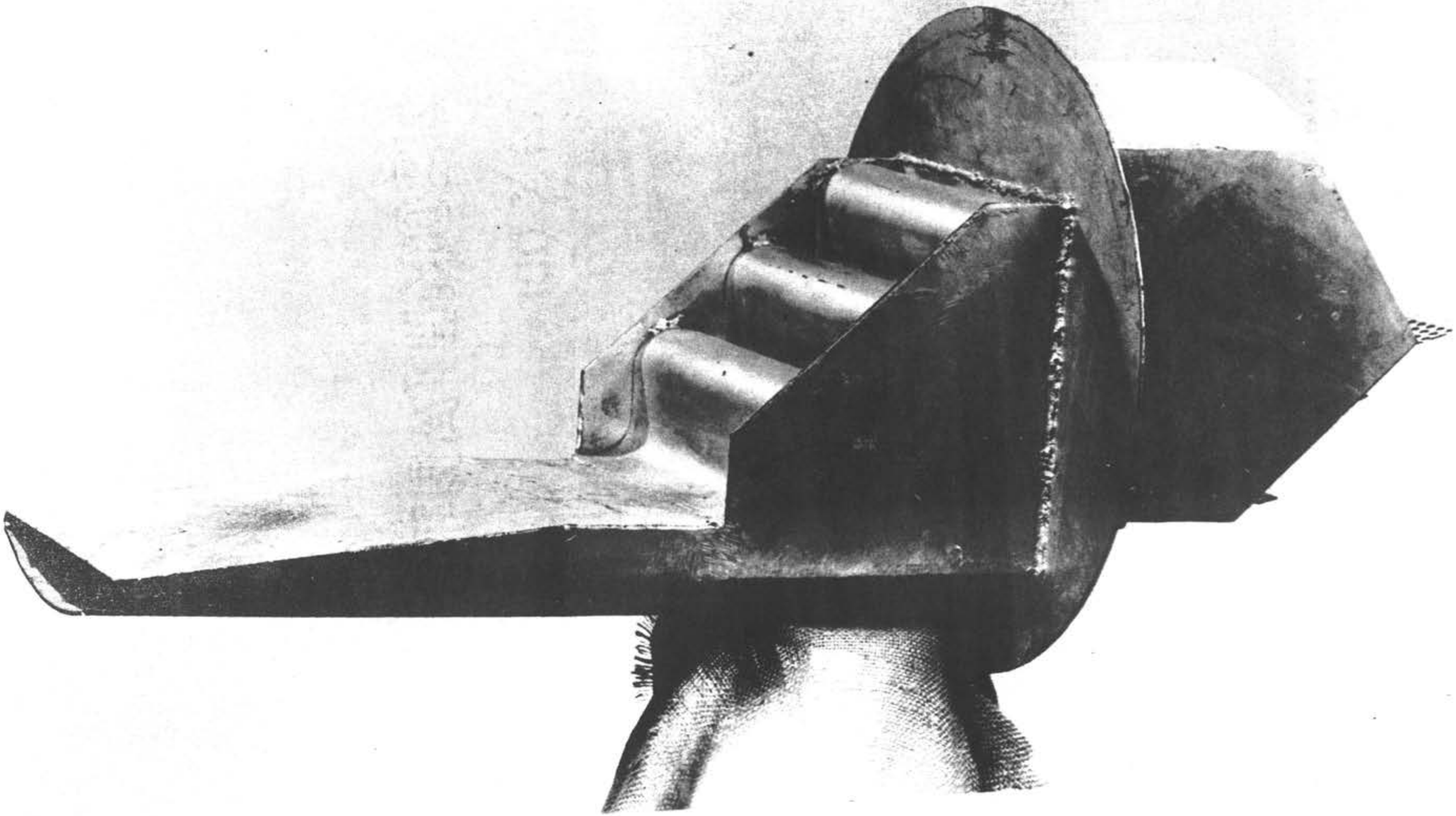

Figure 3. Side View of Tray

PROCEDURE

Analytical Technique:

The refractometer was first standardized against the standard prism and then refractive indices of the pure components were taken. During the course of analyzing the samples, periodic checks were made of the pure samples. Any difference between these check readings and the original ones was applied to the intervening readings as a correction. The refractive index-composition curve made up from known solutions of cyclohexane and isooctane is presented in Fig. 7.

The charge to the Othmer still was about 400 ml. It was determined that equilibrium was attained in one hour. Pressure in the still was maintained at 760 ± 1 mm. of mercury by means of a Model No. 5 Industrial Cartesian Manostat, available from the Emil Greiner Company of New York City. The monostat was operated from a compressed air source. The thermometer was checked in place by running samples of distilled water, cyclohexane, and isooctane; the average deviation from the respective true boiling points was less than 0.2°C. In the case of distilled water the deviation was 0.05°C. Table I contains the vapor-liquid equilibrium data and Figs. 8,9, and 10 present these data as a boiling-point diagram and as an equilibrium diagram with its accompanying smoothing curve. The smoothing curve was made by plotting (y-x) as a function of x for the experimental data and placing the best curve by "eye". This method greatly exaggerates the deviations so that the errors in judgment are insignificant when the smoothed data are reconverted for use in making the equilibrium curve.

Operational Technique:

A flow diagram of the Koch tower and auxiliaries is shown in Fig. 6.

The charge to the Koch column for the first twelve runs (1-4, 7-10, 12-15) was approximately 65 gallons. The initial charge was 75% cyclohexane and composition changes were made by withdrawing cyclohexane-rich liquid from the surge drum and replenishing with pure isooctane. For the last eighteen runs (1A through 17A) the higher vapor loadings with the resulting entrainment and holdup made it necessary to double the inventory to some 120-130 gallons in order to maintain visible liquid levels in the surge durm and the bottom of the column. As a result, there was not enough material to operate at other than one overall composition.

Since the liquid and vapor loadings were so large in comparison to the inventory, the reflux rate per minute often being half of it, equilibrium was quickly established. The limiting factor actually was how quickly the valves could be set to maintain a constant level in the surge drum. This usually took 15 to 20 minutes. Equilibrium was considered to be attained when constant liquid levels and constant temperature readings were maintained over a 5 to 10 minute period.

Three different operating conditions were set as follows:

Total Reflux: All product from the condenser was sent back as reflux to the top of the column.

Stripping: After operating at total reflux, valve V_2 (Fig. 6) was opened allowing pump No. 2 to take suction from the tower bottom (through pump No. 3). The line downstream of V_2 was "teed" into the suction of pump No. 1 and thus there was sent to the top tray some of the bottoms along with all the product from the condenser. Pump No. 3 boosted the pressure on pump No. 2 and hence the amount of the recycle stream during runs 1A through 17A. During the first twelve runs, pump No. 3 was not in the system and pump No. 2 took suction directly from the bottom of the tower.

Rectifying: After operating at total reflux, valve V_3 was opened allowing part of the product from the condenser to flow into the reboiler. It runs 1A trhough 17A, this was sent into the discharge line from the reboiler circulation pump (No. 3). The circulation pump was installed prior to run 1A to increase the boil-up rate.

To maintain atmospheric pressure in the system, the condenser and the surge drum were vented to the outside through approximately fifty feet of 1/2" pipe. Subsequent operation, carried out to give as little subcooling of the product in the condenser as possible, showed that quite a bit of material was being lost out the vent. It was thus necessary to put a valve in this line. This valve was left open while equilibrium was being reached but was closed during the remainder of the run. A slight change in conditions after this valve was closed sometimes resulted in the condenser's being under a slight vacuum. In some cases it was possible to "crack" the valve slightly but the product had to be subcooled considerably to prevent excessive loss in such cases.

After the first twelve runs it was noted that the reboiler design was such that partial vapor blanketing of the tubes on the hydrocarbon side was the controlling resistance. As a result, boil-up rate was completely insensitive to steam pressure and increased vapor loading of the column could not be obtained even with 120-pound steam in the reboiler. This was remedied with forced circulation through the reboiler. After the change, overall coefficients as high as 500 to 600 BTU/hr/ft^2/^{o}F were obtained in comparison with 90 to 110 BTU/hr/ft^2/^{o}F previously. At these higher vapor loadings it was necessary to start the run with the surge drum liquid level in the upper one-third and the tower liquid level between the look glasses and the second tray. After operation had begun the surge drum liquid level would drop to the lower one-third and tower liquid level to a point between the top of the sight glass and the first tray. It was necessary to maintain such a high level in the tower in order that the circulation pump would not lose suction by vapor locking.

Data for a run were taken by two persons working together over a period of 5 to 8 minutes. One operator, on the floor, began by turning the steam condensate into a tared barrel and then proceeded to take all temperature, pressure, and flow readings - coming back to catch the condensate weight at the end of the period. While this was taking place the other operator, on the tower, took vapor and liquid samples from each tray using a small hand condenser filled with ice water. The sampling sequence was begun and ended by the taking of an overhead vapor sample and a bottoms liquid sample to check constancy of conditions during the run.

RESULTS

The calibration data for the flow meters are presented in Figs. 11 and 12, and in Table II. The data for Fig. 12 were obtained by measuring the increase in level of the surge drum while drawing material from the bottom as in stripping; and by measuring the decrease in level of the surge drum while sending material from there to the bottom as in rectifying. It is to be noted that extrapolation of the experimental lines was necessary for low flows through both orifices. All thermometers and pressure gages were calibrated but the corrections to be applied were so much less than the experimental error that the calibrations were not used. The radiation correction for the reboiler was also small, ranging from 3 to 7 pounds of steam per hour, depending on the steam pressure.

In determining efficiencies for the first 12 runs a McCabe-Thiele plot was made for each run and the number of theoretical plates stepped off between x_1 and x_5, covering a range of 4 actual plates. The fifth or top tray was not used because of the marked subcooling effect from the cold reflux. It was likewise realized that x_b was a mixture of x_1 and the unvaporized liquid leaving the reboiler; it bore no operating line or equilibrium relationship to y_b and hence could not be used. Partial submergence of the first tray due to the high liquid level and high entrainment in the latter runs made it necessary to step off plates between x_2 and x_5, representing 3 actual plates. The operating line for the total reflux runs was taken as the 45° line and it is to be noted that the plotted points in Fig. 13 are in close agreement with this. For the other runs the "best" straight line was drawn through the operating points. The slope of such a line is L/V'. L/V' could be calculated by several methods based on either the reboiler duty, the condenser duty, or the y-intercept. The two listed in Table III showed the closest agreement and were based on the most reliable data (see sample calculations Nos. 8,9, and 10). Typical McCabe-Thiele diagrams for the three operating conditions are presented in Figs. 13, 14a, and 14b.

For the calculations, specific heat, specific gravity, and latent heat of vaporization data were taken from the Phillips Chemical Company's hydrocarbon handbook.[6] The effect of temperature on specific heat was taken from Perry[7] and an average correction factor of + 0.0003 (BTU/#/°F)/°F was used. The effect of composition on the above quantity was obtained by straight line interpolation between the extremes of 100% cyclohexane and 100% isooctane. The effect of temperature on specific gravity was obtained from Nelson;[8] a temperature compensation factor of 0.0005 units/°F was used for isooctane and 0.0004 units /°F for cyclohexane. Again composition effect was obtained by interpolation between the 100% extremes.

Overall efficiencies, liquid loadings, wet vapor loadings, dry vapor loadings, operating line slopes, and other pertinent data are summarized in Table III. It is to be noted that L, V, V', and L/V' are for that section of the column between trays 1 and 5, for the reason mentioned above.

Figures 15, 16, 17, and 18 were plotted from the data of Table III and show overall efficiency as a function of liquid loading, wet vapor loading, dry vapor loading and operating line slope, respectively.

Figure 19 was also plotted from Table III and shows the effect of vapor loading on pressure drop per tray. The "best" straight line (log-log plot) passing through all the points regardless of liquid loading could be drawn fairly easily, for the liquid loading effect is small. To determine the exponent to which L must be raised in an equation of the type

$$\Delta P = c'L^{a}V'^{b'} \text{ and } \Delta P = cL^{a}V^{b}$$

the exponent for V' or V was taken from the slope of Fig. 19 (0.8 and 3.1 respectively) and various values of a' and a were tried until the most constant values of c' and c were obtained. These were 1.80 ± 0.36 and 0.82 ± 0.18 respectively, the measure of precision being the average deviation of a single measurement. The a and a' were both found to be 0.1, emphasizing the small effect of L on pressure drop.

Entrainment as a function of dry vapor loading is presented in Fig. 20 and the "best" lines placed for total reflux and stripping conditions. A comparable line for rectifying was not drawn because of the small number of uncertain and widely scattered points for this condition. It appears that the entrainment drops to zero at values of V in the neighborhood of 30-35 pound mols per hour or 4-4.5 feet per second. Figure 21 is a plot of entrainment against liquid loading with V as a parameter. The rectifying points are not included because their values of V are subject to considerable error (as explained below). The noticeable effect of L on E at high vapor loadings is possibly due to the resulting decrease in disengaging space.

DISCUSSION

The data show that liquid and vapor loadings have about equal effect on overall efficiencies. This effect is a decrease in efficiency with increased loadings. At higher loadings, the effect is undoubtedly associated with the attending phenomenon of entrainment. Examination of Fig. 18 reveals that L/V' has no detectable effect on efficiency; this is further evidence of the approximate equality of the influence of L and V' on this quantity. The effect of L on efficiency is further shown in Fig. 17 where it is noted that the efficiency lines for stripping fall below that for total reflux. At values of V less than 35 pound mols per hour L/V' equals L/V so that in this normal operating area, the decrease in efficiency with increasing L at constant V is apparent.

In considering the pressure drop per tray, it is seen that vapor loadings exhibit the greater effect. Higher liquid loadings produce a small increase in pressure drop; this is essentially only that necessary to overcome the increase in liquid level in the cascades. The principle pressure drop is that necessary to increase "slot" velocity of the vapor. The effect of dry vapor loading is larger than that of wet vapor loading since the specific volume of the former is 2 to 3 times that of the latter.

The effect on entrainment of vapor loading, as shown by the data, is as might be expected, with an increase in vapor loading increasing the amount of liquid picked up. It is notable that there seems to be some range of vapor loadings below which the entrainment is negligible but above which the entrainment increases sharply. The entrainment seems somewhat higher for stripping than for total reflux. This is in the expected direction. There are four rectifying points, and three of these lie far to the left of the others. The V associated with these points is subject to considerable error (small difference between large numbers); these points then do not have sufficient weight in accuracy or number to justify any conclusion being drawn from them. The reliable portion of the data on entrainment indicate that the system should be operated at vapor loadings below about 30 pound mols per hour (4 feet per second) to insure negligible entrainment.

In considering the dimensions of the tray it will be noted that the downcomer is approximately 22 inches high. If this were filled with hot hydrocarbon (sp. gr. 0.7) it would be a pressure barrier equivalent only to some 15 inches of water. It is obvious from an examination of Table III or Fig. 19 that many of the later runs were made under conditions approximating or even exceeding this.

FUTURE WORK

The desirability of certain modifications in the equipment and system is evident. Of the more obvious might be included:

1. Redesign of the vapor sampling line so that it receives a composite sample all the way across the vapor space. This would tend to improve consistency and accuracy in the vapor compositions.

2. A proper reboiler capable of giving high boil-up rates without forced circulation. The high liquid level necessitated by the use of a circulation pump not only blanked out the bottom tray but may possibly have "induced" excessive entrainment in the bottom section, which excessive entrainment may have been partially continued all the way up the column.

3. Installation of a reflex preheater, and also a heat exchanger in the recycle line. These changes would remove the subcooling and flashing effects, which are considerable, thus allowing the use of all trays for calculation purposes.

4. Operate on a system having a higher relative volatility. This would produce appreciable composition changes per tray, allowing the calculation of individual tray efficiencies and much less error in the calculation of the overall efficiency.

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NOMENCLATURE

G_1	pressure in tower bottom, psia.
G_2	pressure in steam side of reboiler, psia.
G_3	pressure in steam line, psia.
G_4	pressure in discharge of circulation pump, psia.
G_5	pressure in tower top, psia.
M_1	flow of cooling water, lb./min.
M_4	flow of recycle stream, lb. mols/hr.
R_1	flow of reflux stream, lb. mols/hr.
T_1	temperature of reflux stream, °F.
T_2	temperature of vapor leaving tower, °F.
T_3	temperature of product leaving condenser, °F.
T_4	temperature of entering cooling water, °F.
T_5	temperature of leaving cooling water, °F.
T_6	temperature of tower bottoms, °F.
T_7	temperature in throttling calorimeter, °F.
V_1	reflux control valve.
V_2	recycle control valve for stripping.
V_3	recycle control valve for rectifying.
V_4	circulation control valve.
C_{pa}	average heat capacity, BTU/lb. mol/°F.
D	recycle stream during rectifying (M_4).
e	overall efficiency, per cent.
E	entrainment, lb. mols/hr.
L	liquid loading, lb. mols/hr.
λ	latent heat of vaporization, BTU/lb. mol.

ΔL_5	liquid loading increase below the 5th tray as a result of subcooled reflux, lb. mols/hr.
ΔV_b	vapor loading decrease above the reboiler as a result of subcooled recycle, lb. mols/hr.
n_a	number of actual trays.
n_t	number of theoretical trays.
ΔP	pressure drop per tray, inches of water.
S_n	net steam consumption, (total condensate - radiation correction) lb./hr.
V	dry vapor loading, lb. mols/hr.
V'	wet vapor loading $(V + E)$, lb. mols/hr.
W	recycle stream during stripping (M_4).
x	mol fraction cyclohexane in the liquid, subscripts refer to tray from which stream originates.
y	mol fraction cyclohexane in the vapor, subscripts refer to tray from which stream originates.
X_i	average overall composition of the inventory, mol per cent cyclohexane.

APPENDIX

Auxiliary Equipment Specifications

1. Thermometers:

   T_3, T_4, T_5, and T_6; Taylor Binoc Industrial Glass Stem thermometers reading in 2°F graduations.

   T_1 and T_2; Tagliabue Dial thermometers reading in 2°F graduations.

2. Pressure gages:

   G_1 and G_5; Foxboro, 0 to 15 psi, 0.2 psi graduations.

   G_2; Ashcroft, 0 to 160 psi, 2 psi graduations.

   G_3; Marshalltown ammonia gage, 0 to 300 psi, 10 psi graduations.

   G_4; Ashcroft, 0 to 60 psi, 1 psi graduations.

3. Flowmeters:

   M_1 and M_4; Meriam standard 20" cleanout manometers, orifice diameters 0.95 inches, modified vena contracta taps.

   R_1; Fischer & Porter Flowrator, 0 to 60 GPM at sp. gr. 1.0, 1 GPM graduations.

4. Pumps:

   No. 1; reflux pump, Allis Chalmers "Electrifugal", size 2" x 1½", type SS-DH, pump capacity 100 GPM at 120 ft. head, motor 3-phase 5 hp induction.

   No. 2; recycle pump, Magnaflux, size 1¼" x 1", Mod. H-25-SL, pump capacity 48 GPM at 4.6 ft. head and 5 GPM at 20.7 ft. head, motor 3-phase 1/4 hp induction.

   No. 3; circulation pump, Deming, size 2½" x 2", Fig. 4012, Type 2A, pump capacity 200 GPM at 60 ft. head with a bhp of 4.3, motor 3-phase 7½ hp induction.

5. Reboiler and condenser:

   Shell and tube heat exchanger; shell 6 5/8" I.D., 7" O.D., 28 tubes 19.0 ft. long with 0.50" I.D. and 0.84" O.D. outlets 3" flanged. Condenser installed vertically and the reboiler installed at an angle of approximately 15° with the horizontal.

6. Pressure regulator (steam):

   Davis counter-weighted piston, type no. 2, size 2", screwed.

7. Sampling lines:

1/4" copper tube brazed onto std. 1/8" pipe nipples fitted with plug type brass stop-cocks. Line x_1 was fitted with a special "bucket" that collected a sample for the tube to draw from. Three holes drilled in the bottom of this "bucket" insured constant liquid turnover and a true sample.

8. Surge drum:

Sixty gallon galvanized drum. This was connected to the condenser as shown in the sketch below. The volumetric constant for this system (drum + condenser shell) was 0.690 gallons/centimeter in the range of the sight glass.

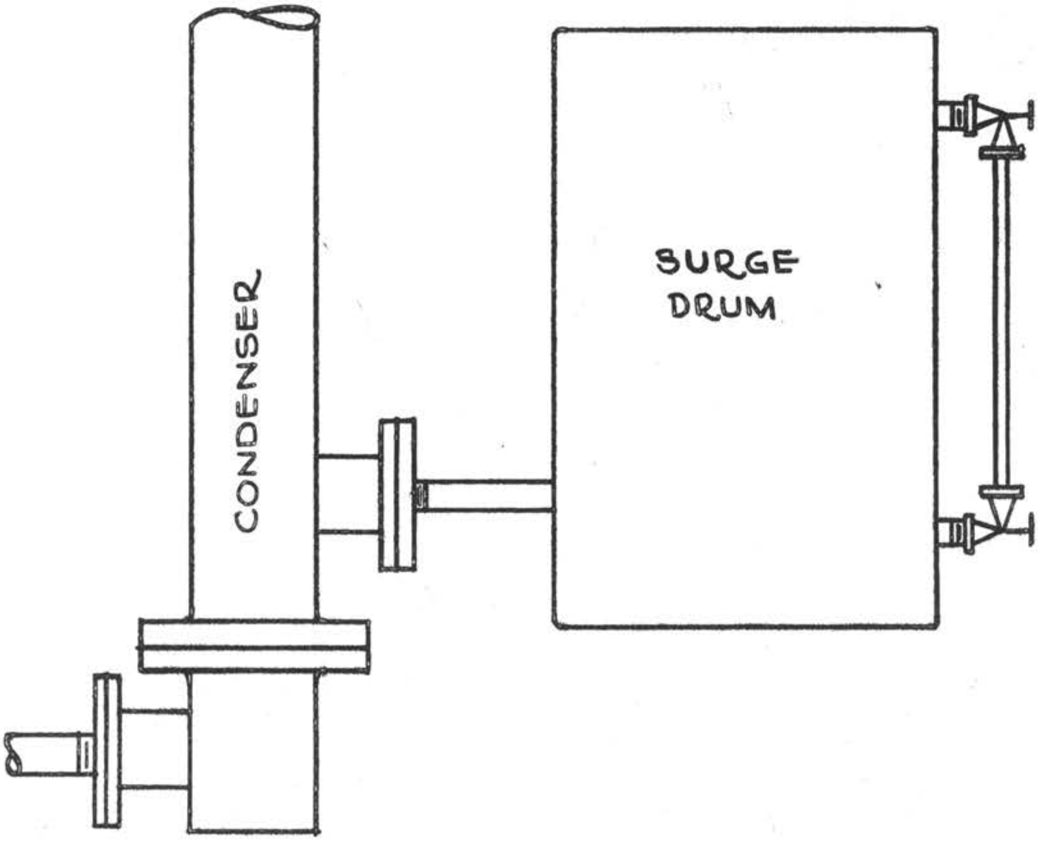


TABLE I

VAPOR-LIQUID EQUILIBRIUM IN THE SYSTEM CYCLOHEXANE-ISOOCTANE AT 760 mm.

Experimental Data

Temp. °C	Mol per cent cyclohexane in	
	Liquid	Vapor
97.8	6.0	10.5
96.8	13.2	18.6
94.7	20.6	29.2
94.0	25.7	34.8
92.1	31.2	43.3
91.0	37.3	49.6
89.8	44.5	56.8
88.8	49.5	61.8
87.6	56.8	68.0
85.7	68.1	77.4
---	73.1	82.6
82.9	85.5	90.1
81.4	92.5	94.9

TABLE II

CALIBRATION OF ROTAMETER AND DETERMINATION OF CONVERSION CONSTANT

Reading GPM	Actual Flow GPM	Conversion Constant
4.00	4.05	1.01
8.15	9.30	1.14
13.80	16.00	1.16
25.30	29.20	1.15
40.70	46.40	1.14
51.60	58.40	1.13
Average conversion constant (omitting first value)		1.15

TABLE III

RESULTS

Run No.	G_3	G_2	S_n	ΔV_b	V	T_1	T_2	R	ΔL_5
1	86	53.3	490	---	34.6	129	175	29.2	5.1
2	103	52.3	432	---	30.8	156	176	55.2	5.2
3	84	57.3	414	---	29.3	171	176	95.0	4.1
4	103	55.3	361	---	25.7	172	177	121	4.8
7	111	56.3	460	---	32.5	120	184	24.1	6.2
8	102	51.3	398	---	28.3	157	185	45.2	6.1
9	119	58.3	384	---	27.1	173	186	77.5	5.8
10	117	53.3	345	---	24.5	177	185	101	6.0
12	114	56.3	425	---	29.8	116	192	18.6	6.2
13	124	56.3	388	---	27.3	162	193	43.6	6.8
14	119	52.3	366	---	25.8	175	193	68.3	7.1
15	115	55.3	334	---	23.6	180	192	92.5	7.9
1A	109	43.3	580	---	41.6	88	187	82.1	31.9
2A	133	26.3	488	---	36.5	95	118	64.6	23.0
3A	132	39.3	574	---	41.9	100	118	86.2	28.5
4A	123	44.3	565	---	40.8	103	109	128	41.1
4B	133	40.0	599	---	43.7	108	188	128	38.2
4C	130	30.0	621	---	46.2	110	188	133	38.8
7A	132	53.3	627	---	45.0	116	118	165	44.6
8A	131	52.3	629	---	45.1	121	117	196	49.6
9A	130	54.0	697	---	49.9	129	116	224	50.9
10A	127	92.0	582	---	39.9	155	186	131	17.9
10B	132	26.3	460	---	34.4	159	193	131	16.2
12A	132	103.0	493	---	32.3	177	195	208	11.1
12B	132	26.3	459	---	34.4	176	194	216	12.3
13A	127	64.0	617	22.0	20.0	122	203	24	6.3
14A	130	112.0	613	---	39.8	151	185	208	31.6
15A	131	110.0	662	18.3	26.4	113	190	62.6	18.1
16A	126	104.0	706	31.1	17.0	119	191	66.6	17.7
17A	130	28.8	745	12.4	43.0	127	199	143	33.6

TABLE III (continued)

Run No.	L	T_6	M_4 [a]	V'	E	L/V' [b]	y	L/V (from y)
1	34.3	218	---	---	---	1	---	---
2	60.4	219	40	20	---	2.0	-0.913	2.3
3	99.1	224	82	17	---	3.4	-2.02	3.8
4	126	220	108	18	---	4.9	-3.10	5.2
7	30.3	234	---	---	---	1	---	---
8	51.3	231	34	17	---	1.8	-0.490	2.2
9	83.3	234	62	21	---	3.1	-0.976	3.0
10	107	250	84	23	---	4.3	-1.49	4.3
12	24.8	239	---	---	---	1	---	---
13	50.4	234	29	21	---	1.8	-0.223	1.5
14	75.4	235	53	22	---	2.9	-0.462	3.9
15	100	250	80	20	---	4.3	-0.855	4.3
1A	114	236	---	114	72	1	---	---
2A	88.0	224	---	88	51	1	---	---
3A	115	236	---	115	73	1	---	---
4A	169	230	---	169	128	1	---	---
4B	166	227	---	166	122	1	---	---
4C	172	219	---	172	125	1	---	---
7A	210	229	---	210	165	1	---	---
8A	246	227	---	246	201	1	---	---
9A	275	229	---	275	225	1	---	---
10A	149	230	65	84	44	1.8	-0.357	1.8
10B	147	230	69	79	45	1.8	-0.408	1.8
12A	219	230	139	80	48	2.7	---	---
12B	228	232	140	88	54	2.6	-0.794	2.6
13A	30.3	218	58	88	68	0.34	---	---
14A	240	220	88	151	111	1.6	---	---
15A	81.0	211	48	128	102	0.63	+0.192	0.63
16A	84.0	214	80	164	147	0.51	+0.255	0.53
17A	176	224	33	209	166	0.86	+0.085	0.86

[a] $M_4 \left.\begin{array}{l} \\ \\ \end{array}\right\} \begin{array}{l} = W \text{ (stripping)} \\ = D \text{ (rectifying)} \end{array}$

[b] Used in McCabe-Thiele diagrams

TABLE III (continued)

Run No.	e	ΔP	T_3	T_4	T_5	M_1	X_i
1	46	6.4	123	46	129	83	75
2	29	5.1	106	46	131	77	75
3	20	4.5	110	47	133	73	75
4	25	4.5	86	45	132	64	75
7	44	5.1	120	48	141	74	50
8	34	4.0	96	48	135	68	50
9	25	4.0	89	48	133	64	50
10	25	4.0	77	48	124	67	50
12	45	2.9	113	48	154	60	25
13	36	4.5	93	48	144	60	25
14	25	2.9	82	48	134	64	25
15	38	3.4	66	48	119	63	25
1A	29	13.6	188	54	86	267	40
2A	17	19.4	92	69	92	326	50
3A	18	23.3	98	69	97	300	50
4A	16	15.0	98	68	95	321	50
4B	18	15.5	103	72	101	333	50
4C	23	15.0	105	72	101	326	50
7A	25	13.9	112	69	99	326	50
8A	19	22.7	118	69	100	324	50
9A	17	30.0	124	68	100	326	50
10A	18	12.8	97	70	98	311	50
10B	9	13.3	93	71	93	333	50
12A	--	--	95	71	96	--	50
12B	14	11.1	93	71	93	329	50
13A	--	--	117	71	122	--	50
14A	--	--	104	71	101	--	50
15A	11	16.6	108	71	104	311	50
16A	8	15.5	117	71	108	302	50
17A	21	16.6	123	72	108	326	50

TABLE III (continued)

Run No.	x_b	x_1	x_2	x_3	x_4	x_5	x_R
1	0.628	0.638	0.683	0.733	0.776	0.813	0.865
2	0.704	0.706	0.733	0.746	0.751	0.758	0.753
3	0.728	0.725	0.746	0.753	0.756	0.758	0.748
4	0.735	0.733	0.748	0.751	0.753	0.753	0.756
7	0.303	0.322	0.367	0.414	0.487	0.535	0.632
8	0.402	0.399	0.433	0.452	0.471	0.481	0.481
9	0.426	0.423	0.449	0.455	0.462	0.465	0.459
10	0.437	0.440	---	0.462	0.471	0.465	0.459
12	0.118	0.132	0.154	0.175	0.225	0.281	0.395
13	0.221	0.213	0.233	0.250	0.246	0.270	0.265
14	0.227	0.230	0.246	0.254	0.270	9.273	0.250
15	0.252	0.246	0.254	0.258	0.265	0.270	0.258
1A	0.303	0.353	0.365	0.404	0.380	0.459	0.462
2A	0.417	0.433	0.446	0.468	0.481	0.500	0.513
3A	0.408	0.433	0.459	0.475	0.497	0.523	0.529
4A	0.435	0.465	0.491	0.504	0.535	0.541	0.545
4B	0.419	0.459	0.481	0.526	0.538	0.538	0.554
4C	0.399	0.442	0.475	0.516	0.538	0.560	0.554
7A	0.417	0.446	0.465	0.510	0.541	0.554	0.545
8A	0.423	0.459	0.484	0.523	0.535	0.551	0.541
9A	0.414	0.450	0.475	0.520	0.532	0.541	0.551
10A	0.459	0.472	0.478	0.500	0.504	0.510	0.504
10B	0.470	0.504	0.504	0.510	0.516	0.516	0.472
12A	0.489	0.497	0.504	0.510	0.510	0.510	0.497
12B	0.499	0.507	0.507	0.513	0.520	---	0.507
13A	0.522	0.475	0.481	0.500	0.497	0.484	0.516
14A	0.475	0.507	0.442	0.459	0.513	0.513	0.516
15A	0.509	0.481	0.484	0.494	0.500	0.513	0.516
16A	0.510	0.475	0.491	0.494	0.494	0.507	0.523
17A	0.458	0.472	0.472	0.507	0.500	0.541	0.545

TABLE III (continued)

Run No.	y_b	y_1	y_2	y_3	y_4	y_5
1	0.646	0.677	0.730	0.778	0.813	0.865
2	0.723	0.743	0.778	0.781	0.803	0.834
3	0.748	0.776	0.789	0.786	0.794	0.827
4	0.746	0.800	0.805	0.803	0.810	0.830
7	0.326	0.381	0.417	0.446	0.541	0.636
8	0.414	0.468	0.519	0.519	0.547	0.596
9	0.440	0.516	0.525	0.535	0.547	0.596
10	0.459	0.519	0.538	0.562	0.556	0.592
12	0.132	0.149	0.188	0.241	0.254	0.388
13	0.221	0.250	0.277	0.311	0.323	0.385
14	0.247	0.277	0.315	0.323	0.323	0.385
15	0.256	0.307	0.341	0.337	0.349	0.366
1A	0.305	0.350	0.394	0.430	0.475	0.459
2A	---	0.453	0.475	0.487	0.523	0.513
3A	---	0.453	0.478	0.453	0.529	0.526
4A	---	0.472	0.523	0.541	0.545	---
4B	---	0.475	0.513	0.535	0.563	0.551
4C	---	0.478	0.507	0.548	0.560	0.554
7A	---	0.478	0.507	0.541	0.569	0.536
8A	---	0.478	0.538	0.538	0.554	---
9A	---	0.481	0.520	0.538	0.538	---
10A	---	0.507	0.535	0.545	0.545	---
10B	---	0.510	0.532	0.532	0.545	0.537
12A	---	0.529	0.551	0.545	0.532	0.529
12B	---	0.516	0.545	0.541	0.535	0.531
13A	---	0.484	0.504	0.548	0.529	0.515
14A	---	0.535	0.563	0.560	0.548	0.531
15A	---	0.487	0.504	0.507	0.523	---
16A	---	0.504	0.523	0.510	0.523	0.510
17A	---	0.484	0.507	0.545	0.545	0.543

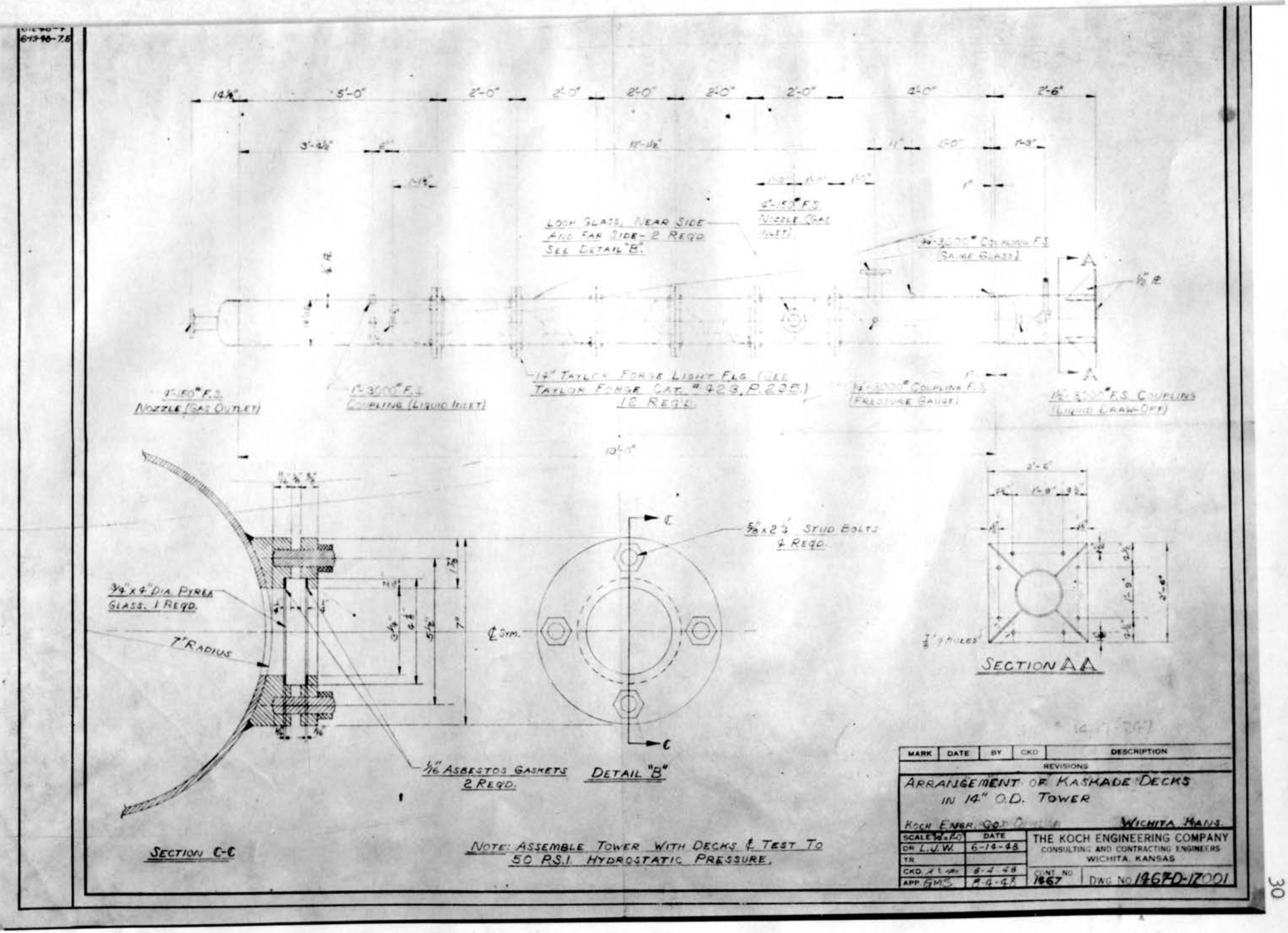


Figure 4. DETAIL OF THE COLUMN

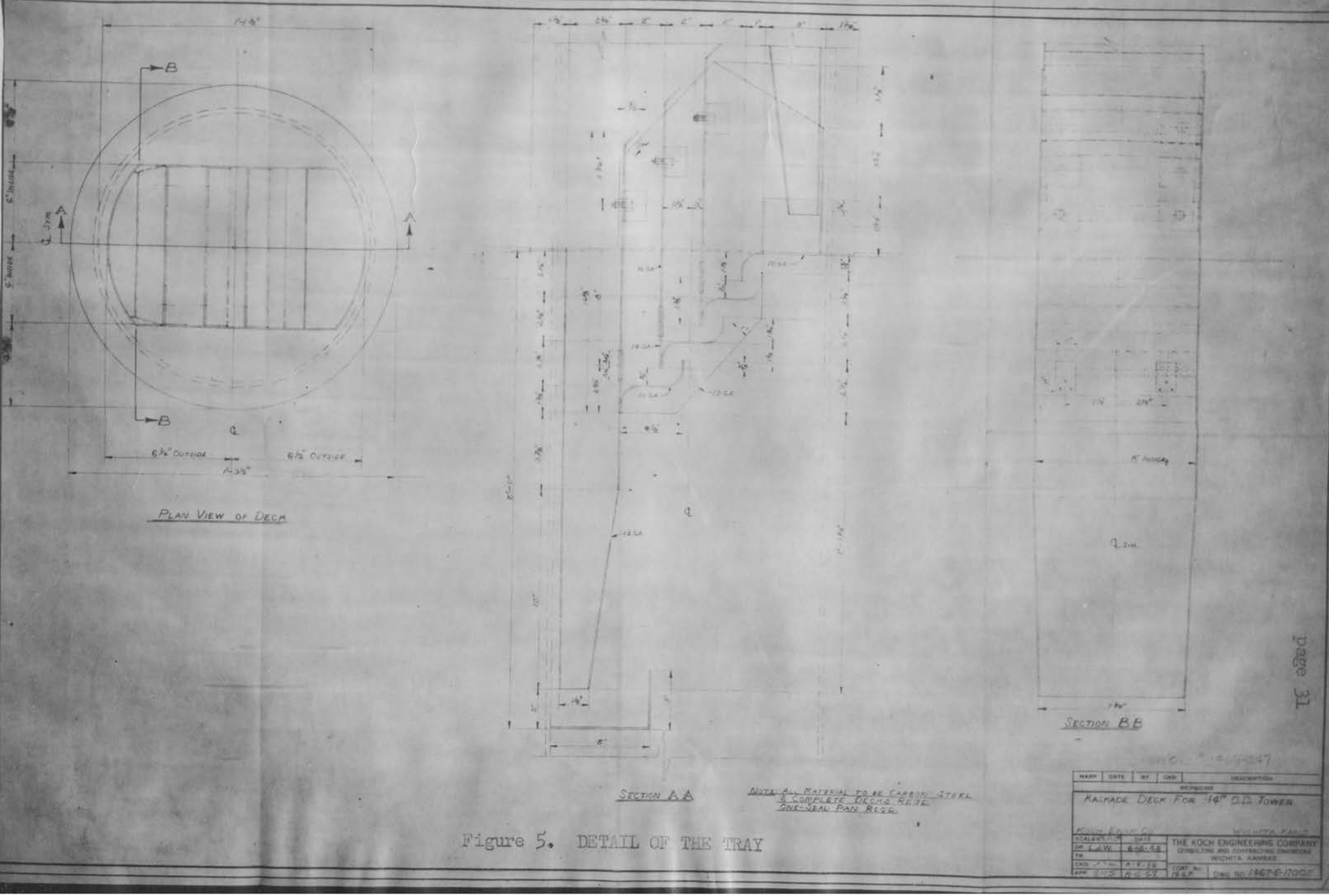


Figure 5. DETAIL OF THE TRAY

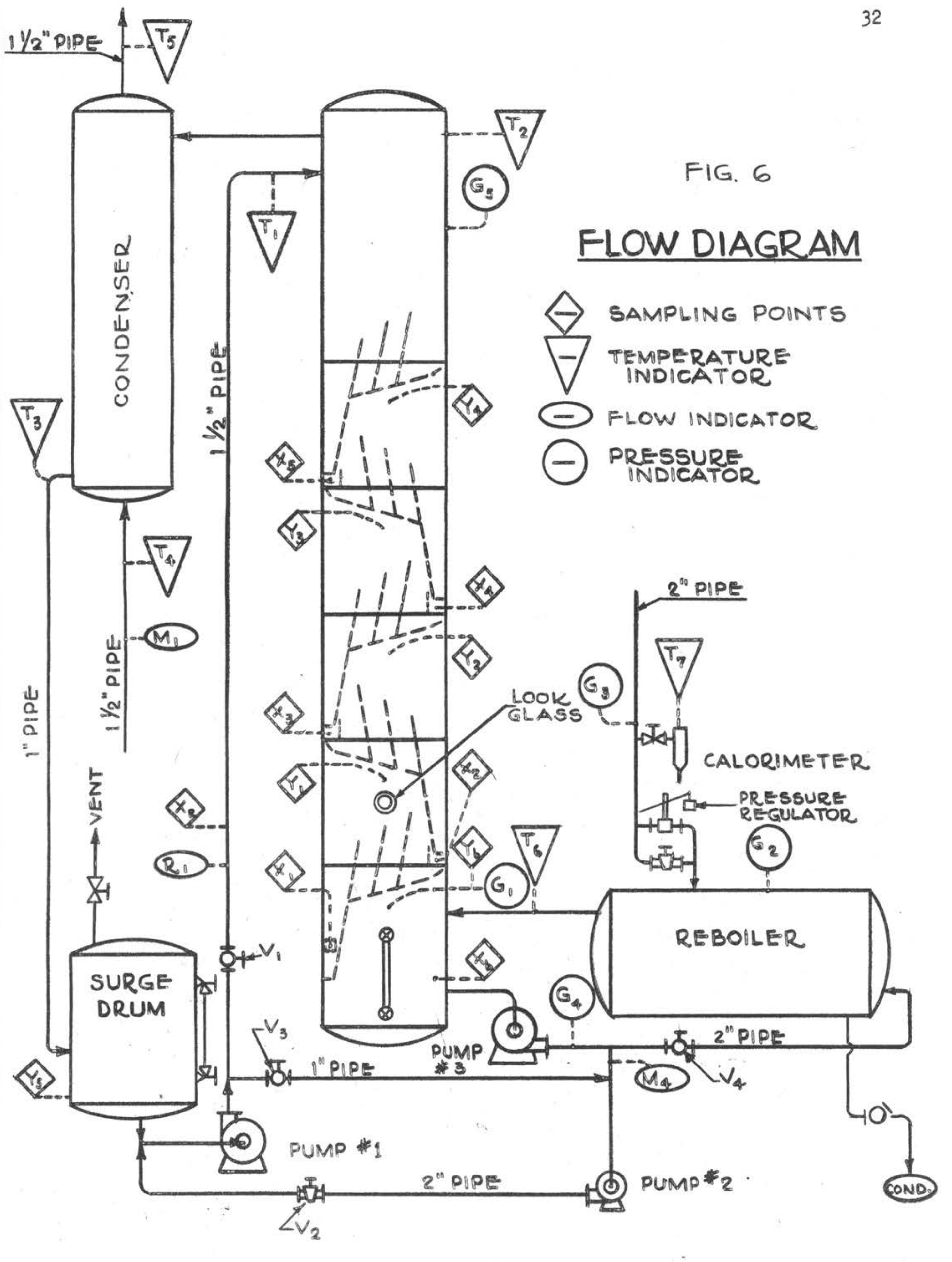
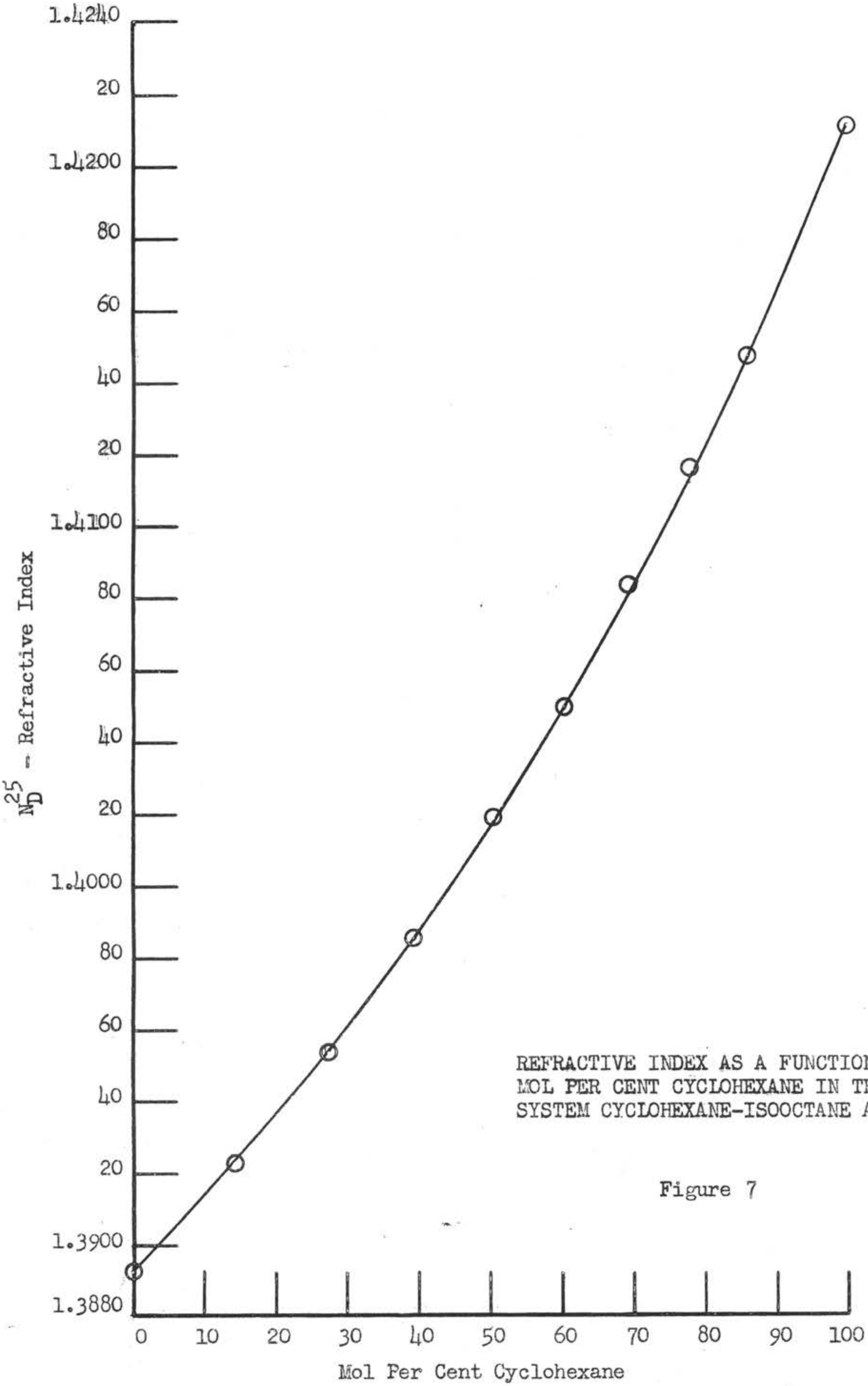


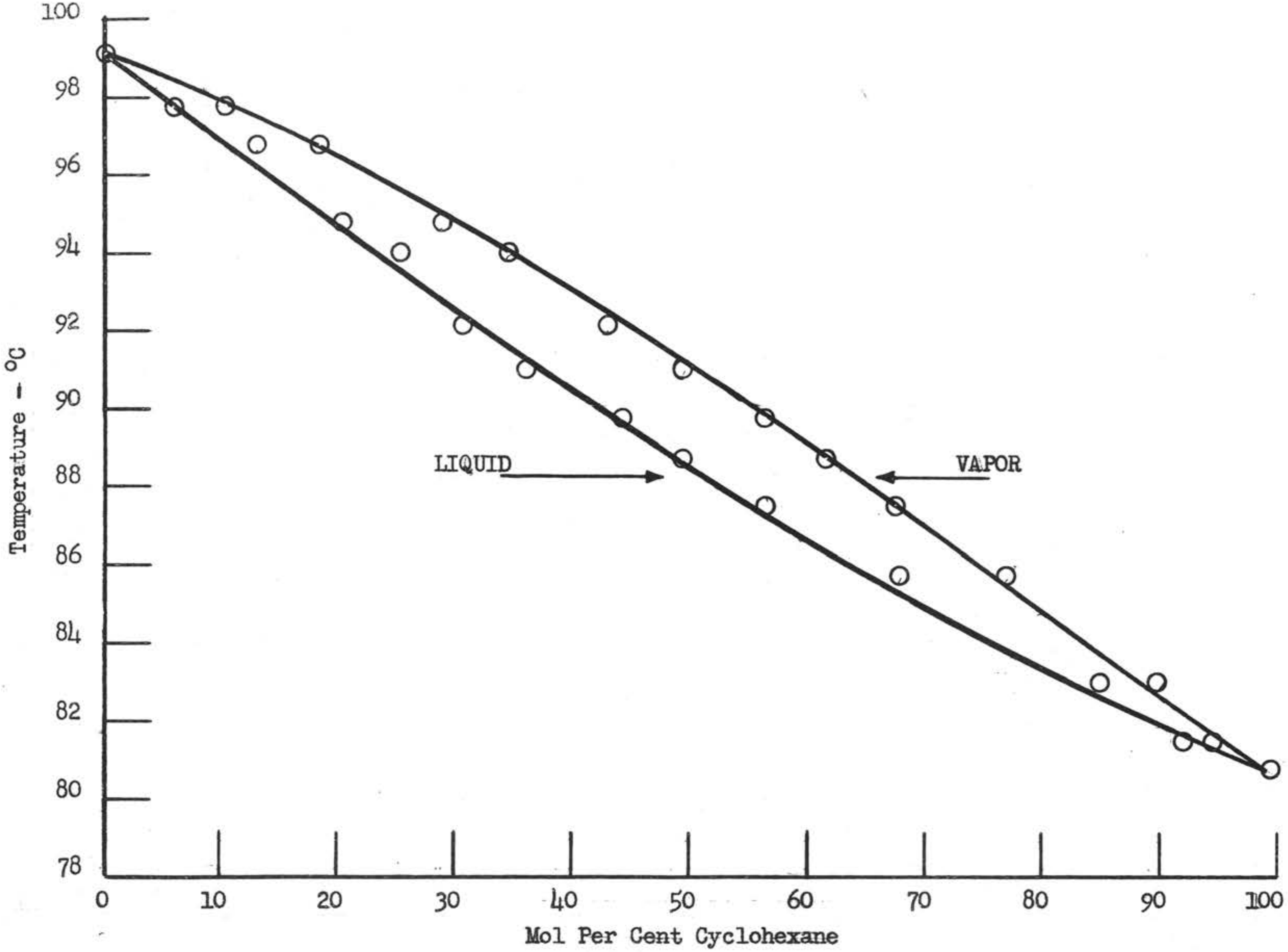
FIG. 6

FLOW DIAGRAM



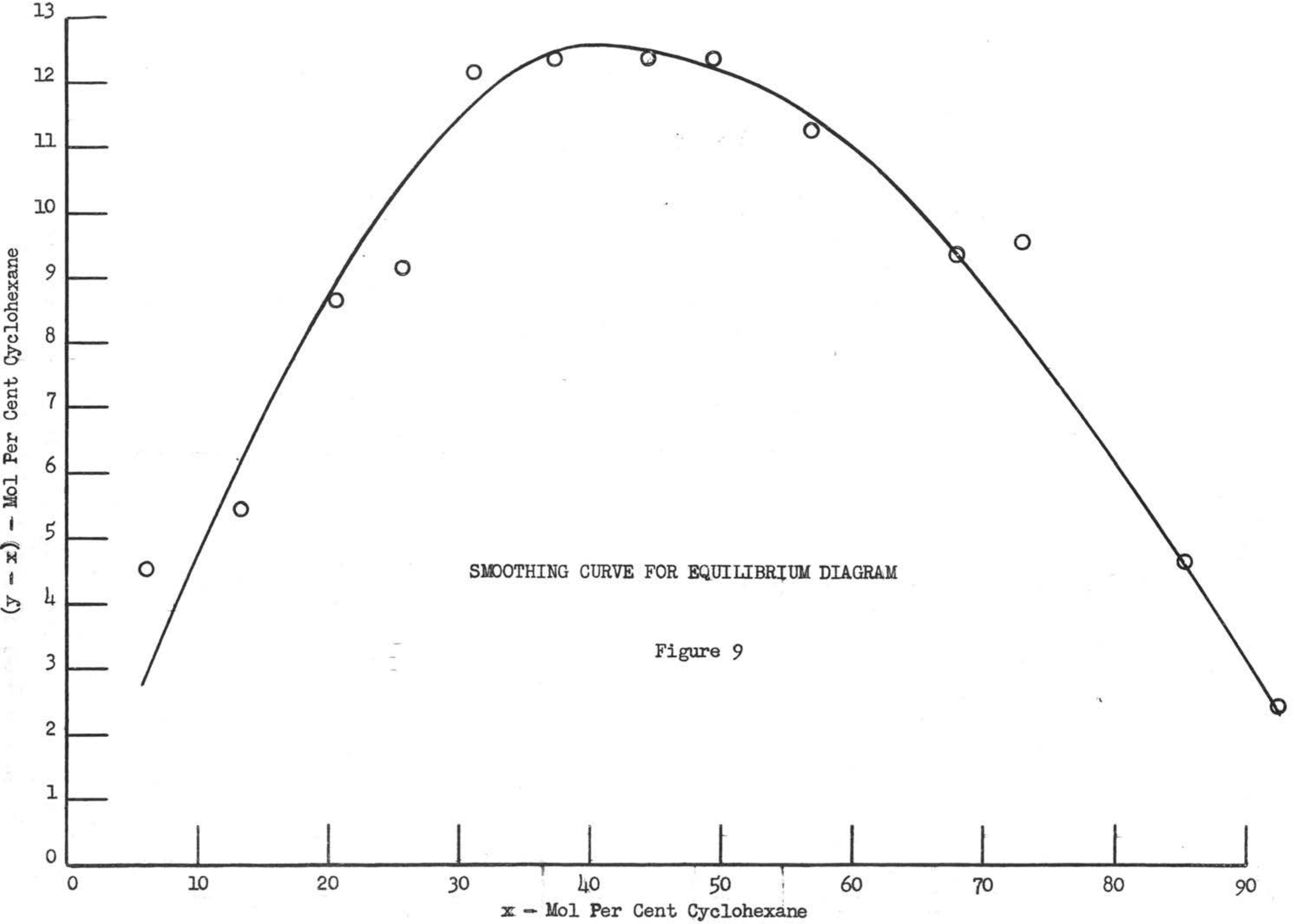
REFRACTIVE INDEX AS A FUNCTION OF
MOL PER CENT CYCLOHEXANE IN THE
SYSTEM CYCLOHEXANE-ISOOCTANE AT 25° C.

Figure 7



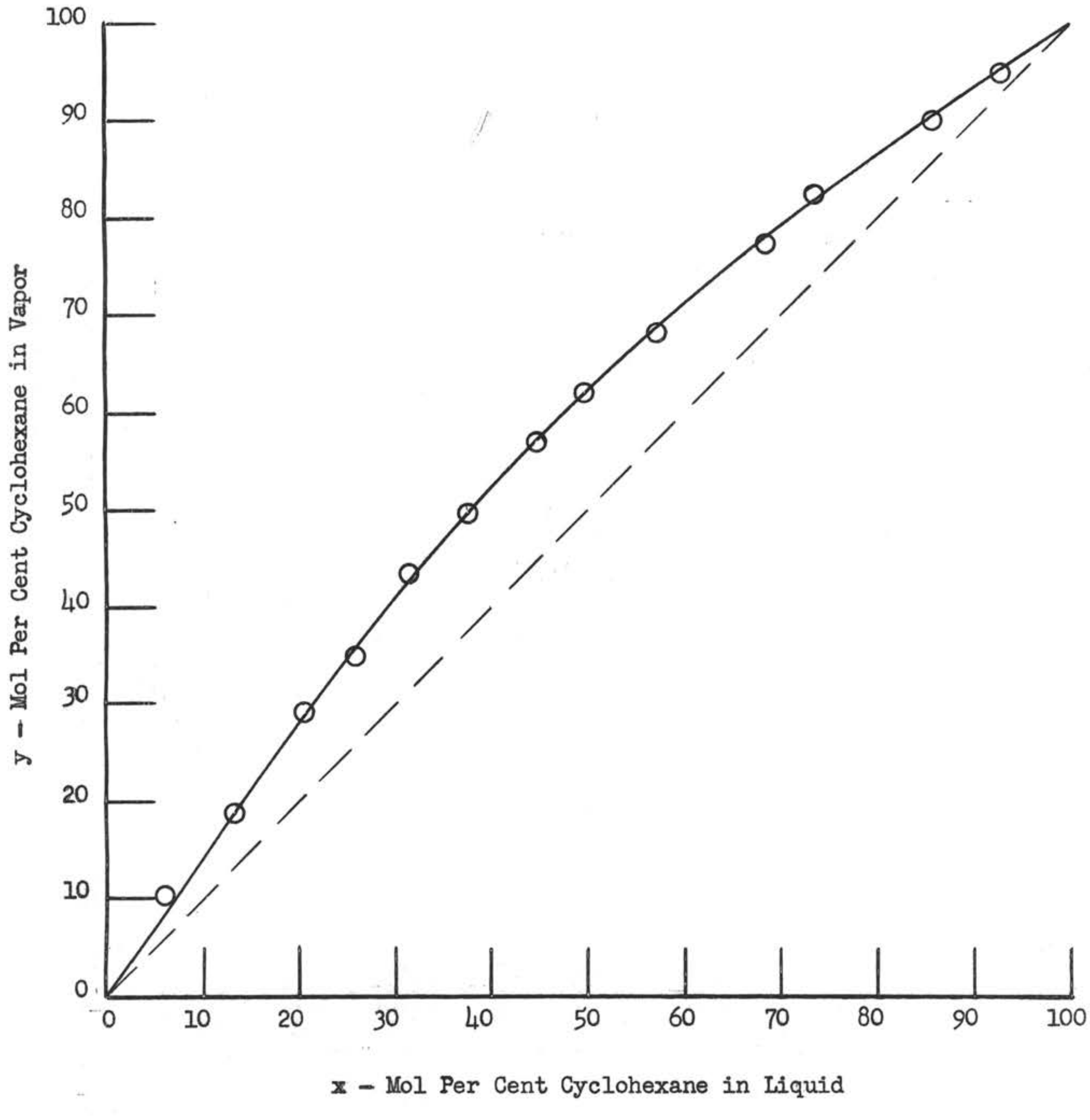
BOILING-POINT DIAGRAM FOR THE SYSTEM CYCLOHEXANE-ISOOCTANE AT 760 mm Hg

Figure 8



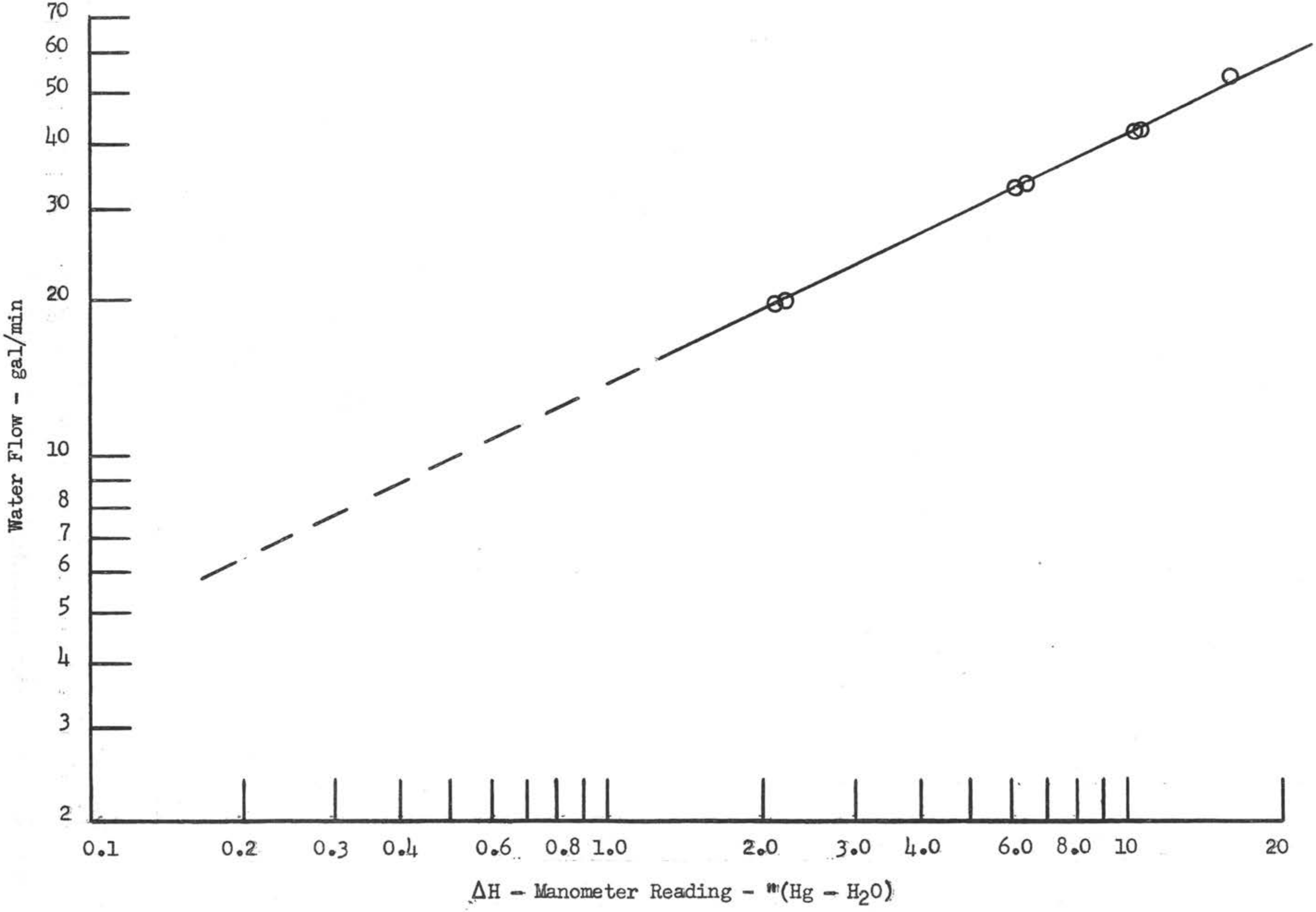
SMOOTHING CURVE FOR EQUILIBRIUM DIAGRAM

Figure 9



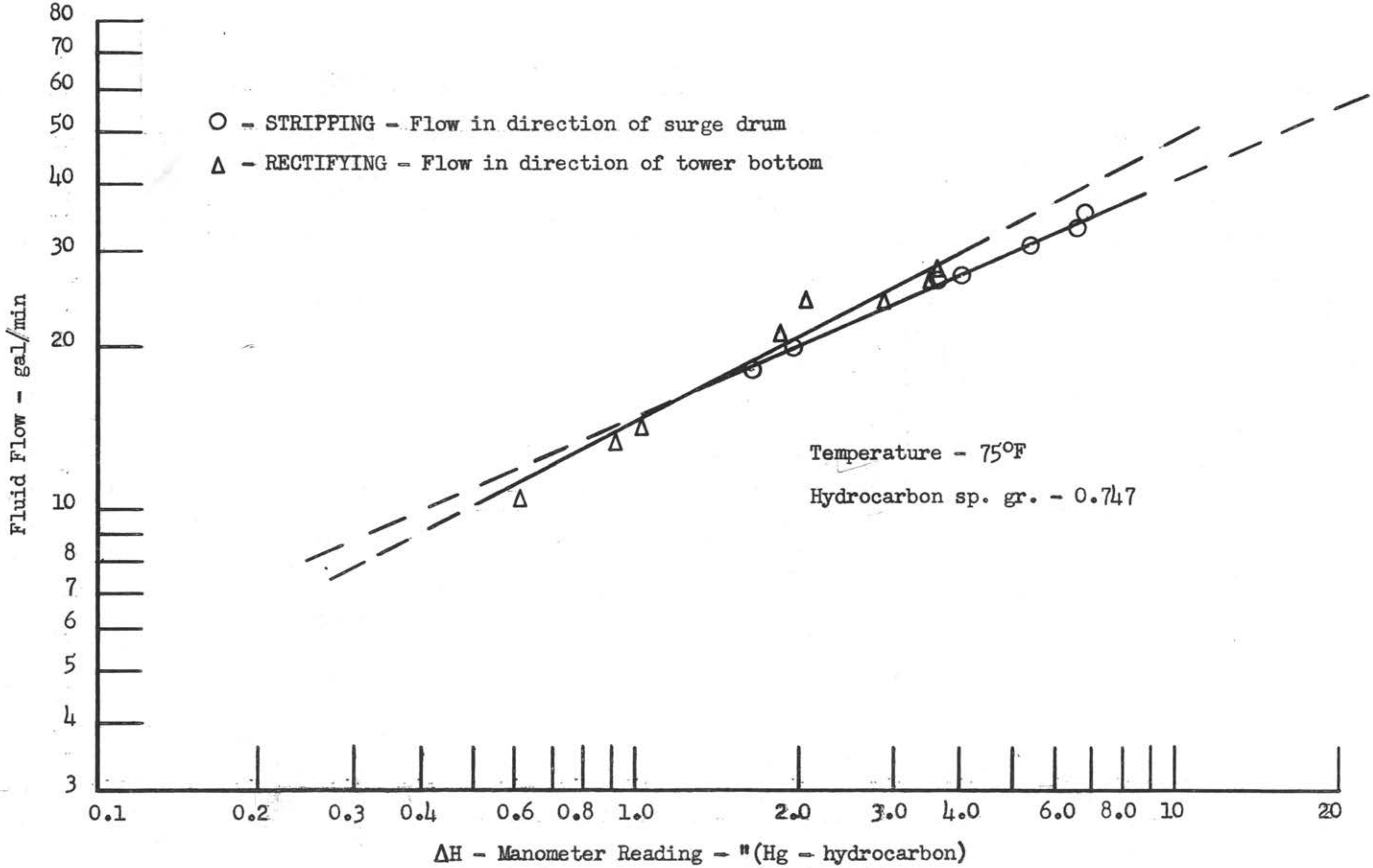
EQUILIBRIUM DIAGRAM FOR THE SYSTEM CYCLOHEXANE-ISOOCTANE
AT 760 mm Hg

Figure 10



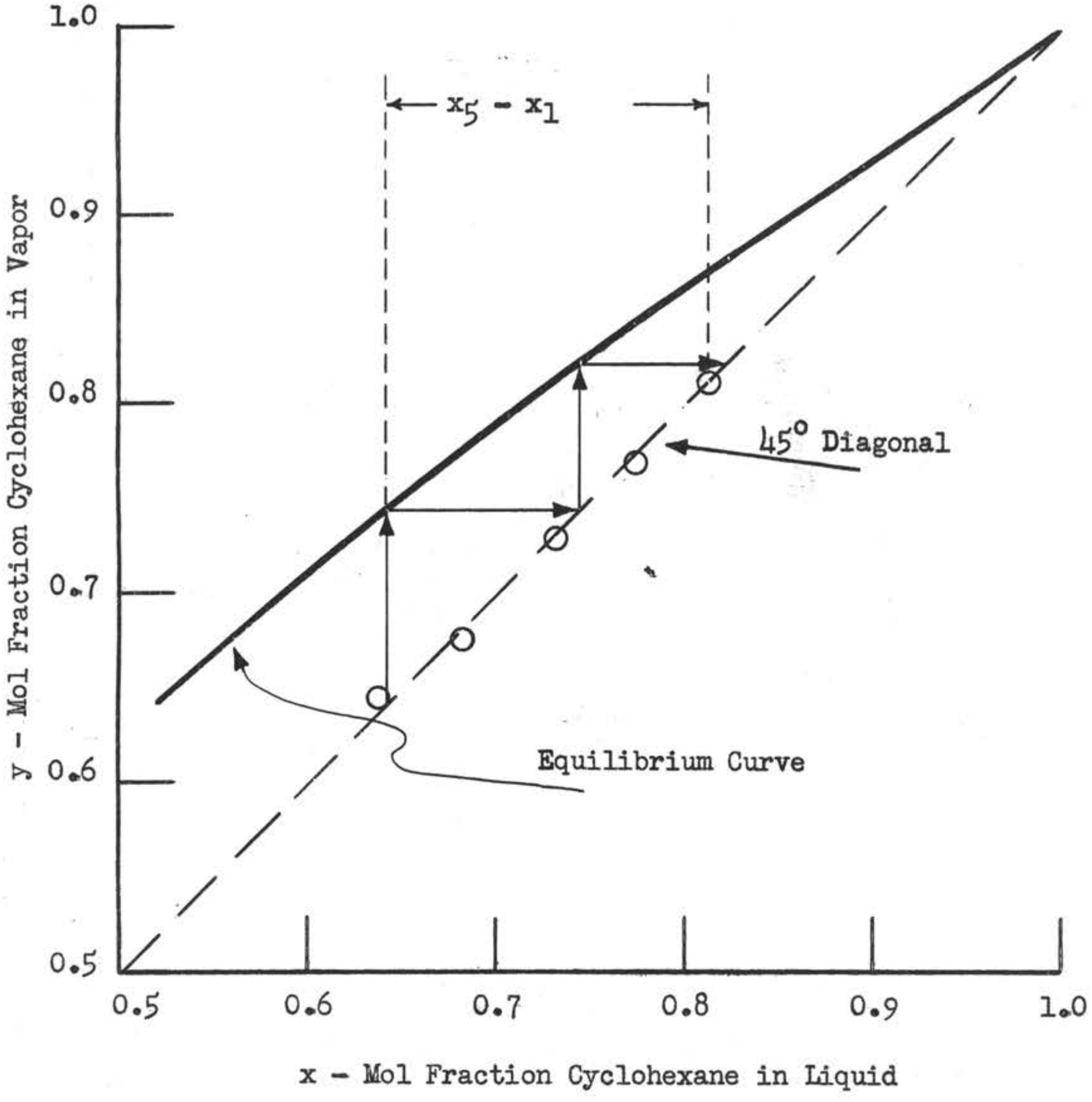
CALIBRATION PLOT OF WATER ORIFICE - M_1

Figure 11



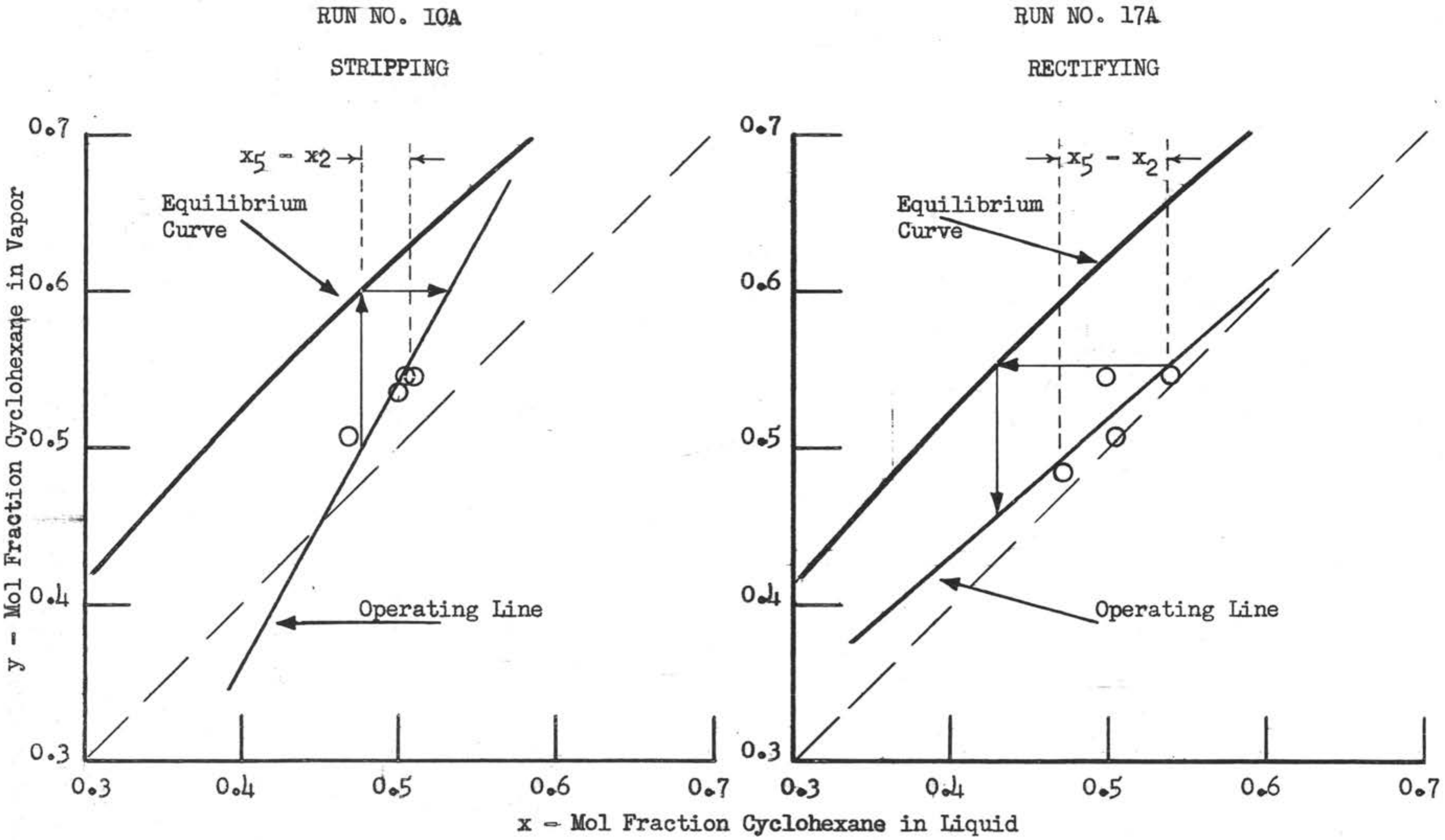
CALIBRATION PLOT OF HYDROCARBON RECYCLE ORIFICE – M_4

Figure 12



McCABE-THIELE DIAGRAM FOR RUN NO. 1
TOTAL REFLUX

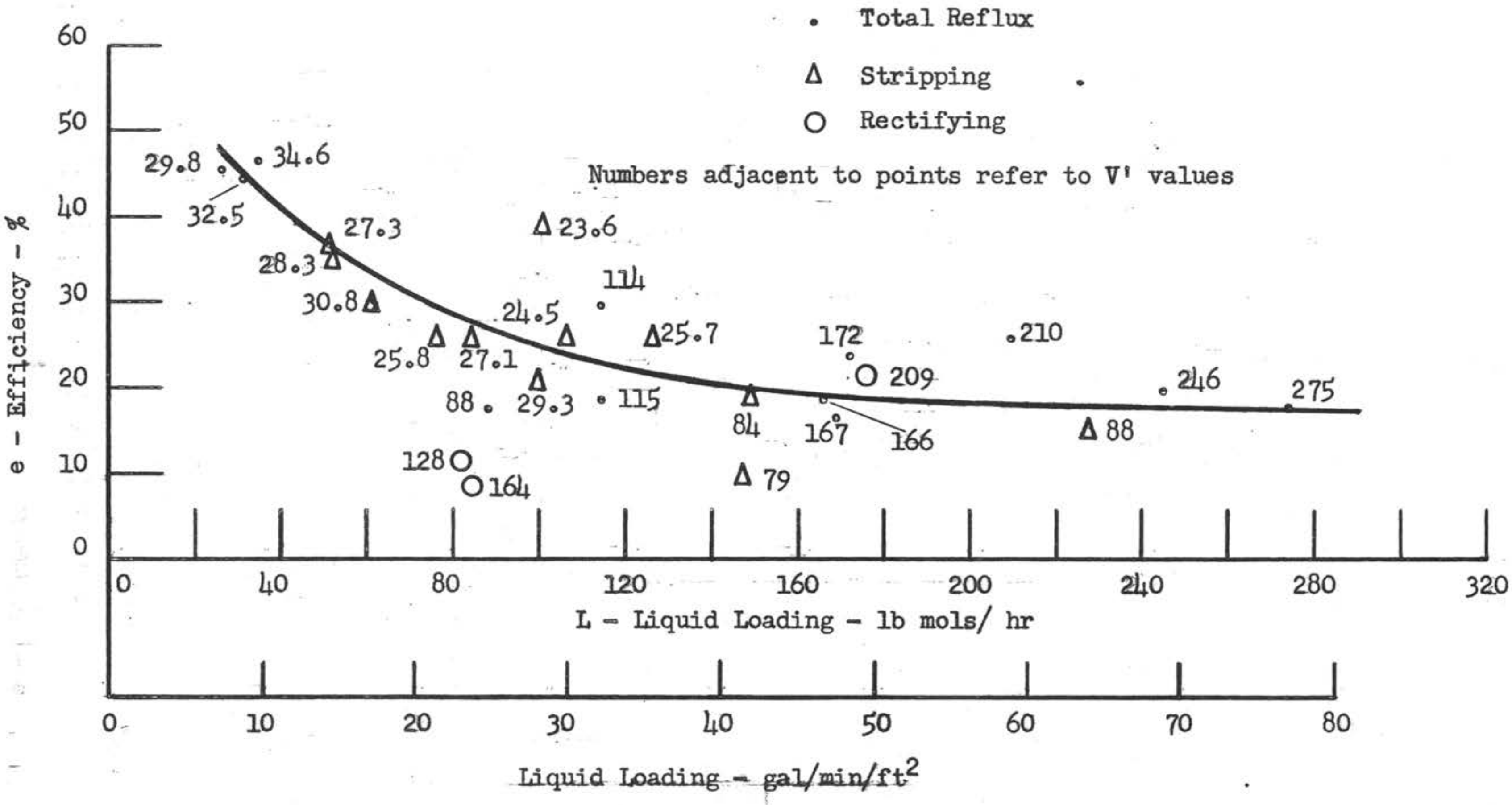
Figure 13



McCABE-THIELE DIAGRAMS

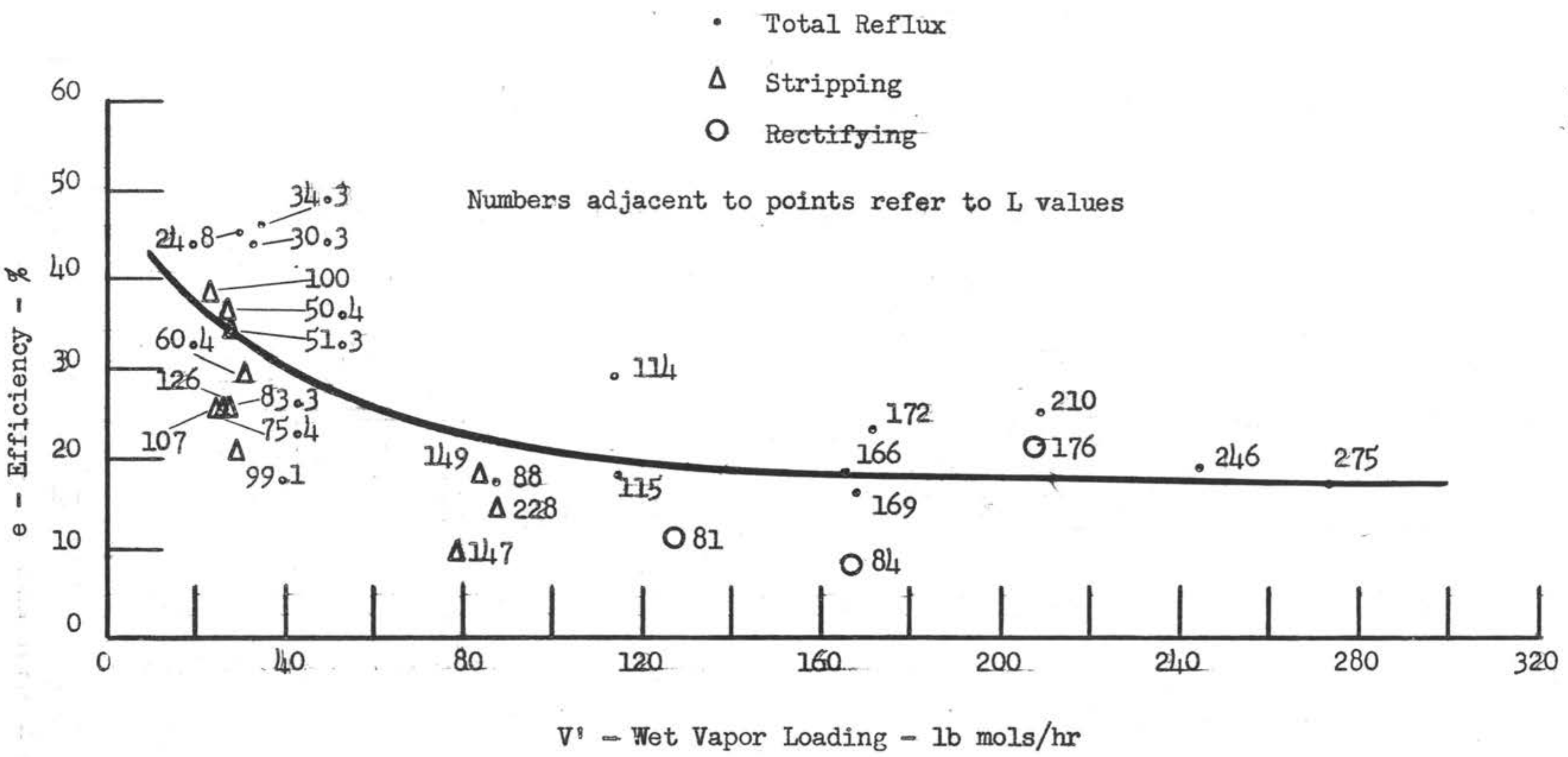
Figure 14a

Figure 14b



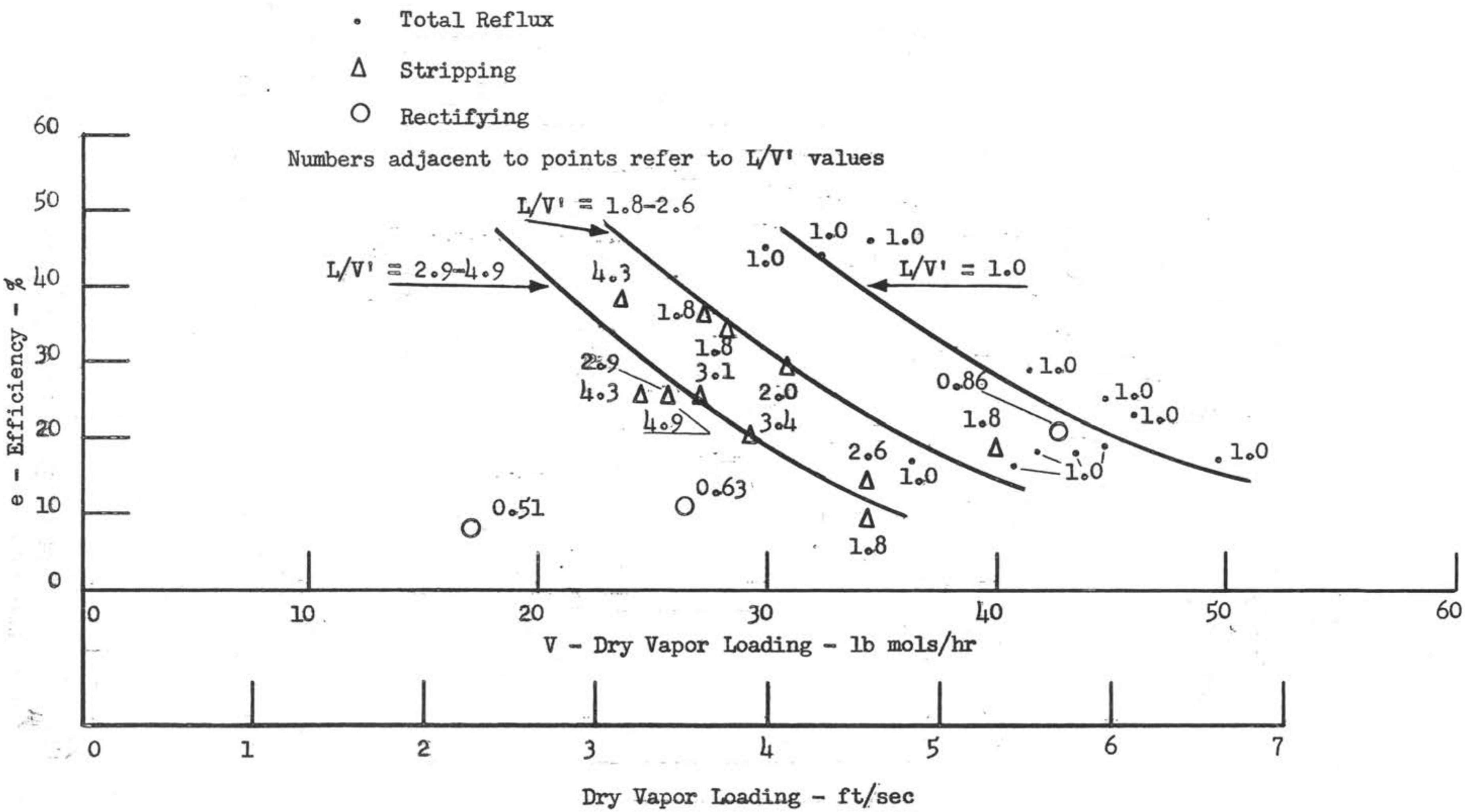
OVERALL EFFICIENCY AS A FUNCTION OF LIQUID LOADING

Figure 15



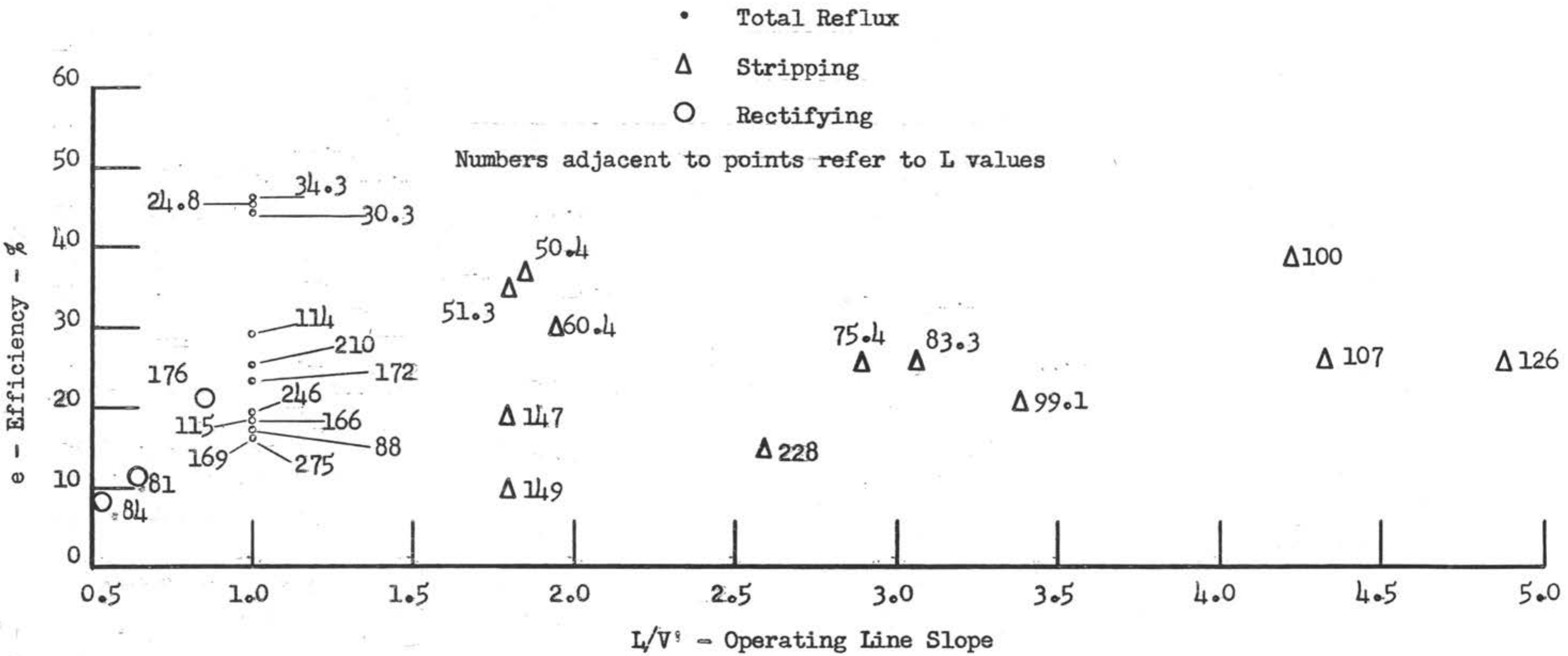
OVERALL EFFICIENCY AS A FUNCTION OF WET VAPOR LOADING

Figure 16



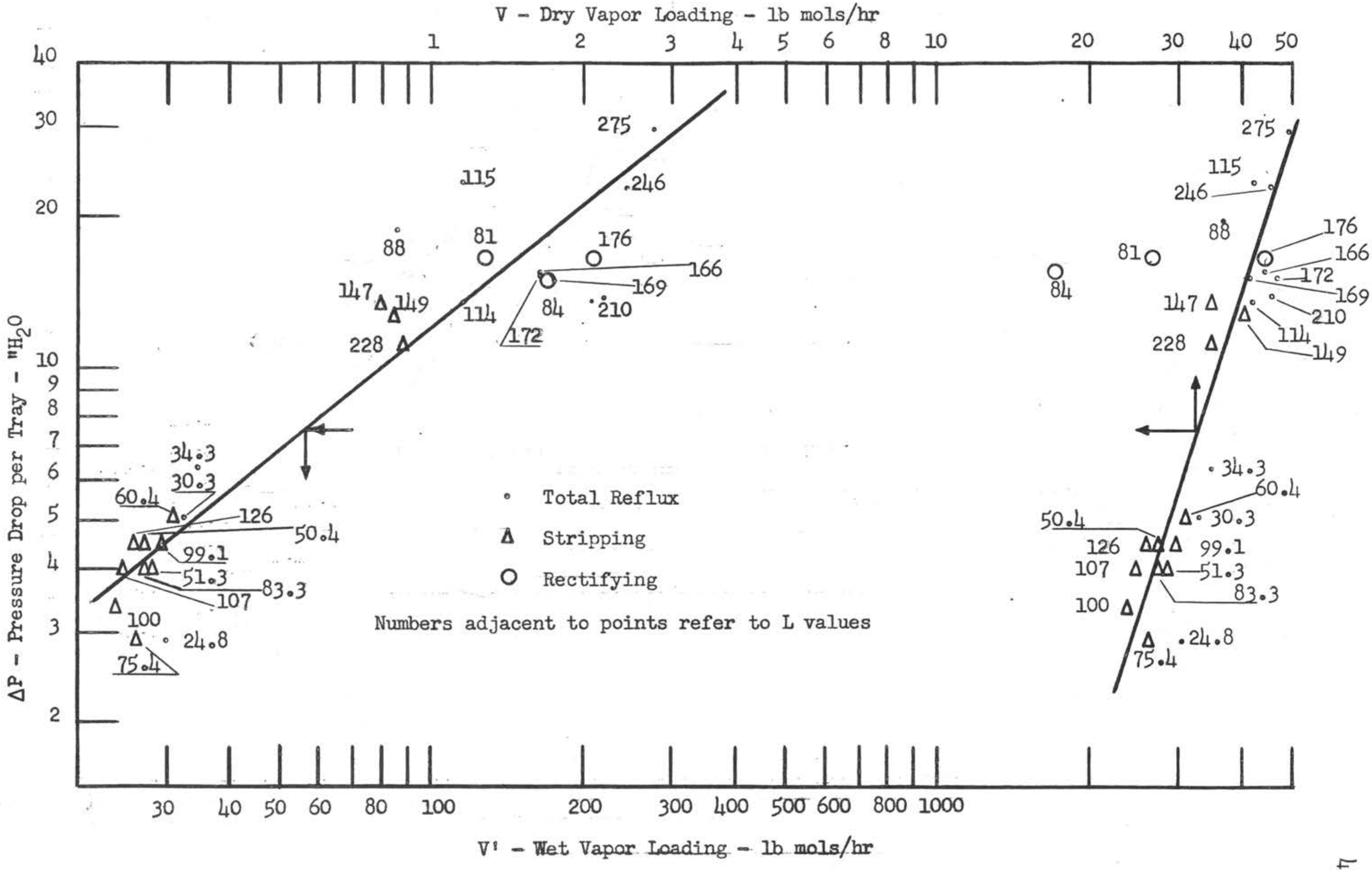
OVERALL EFFICIENCY AS A FUNCTION OF DRY VAPOR LOADING

Figure 17



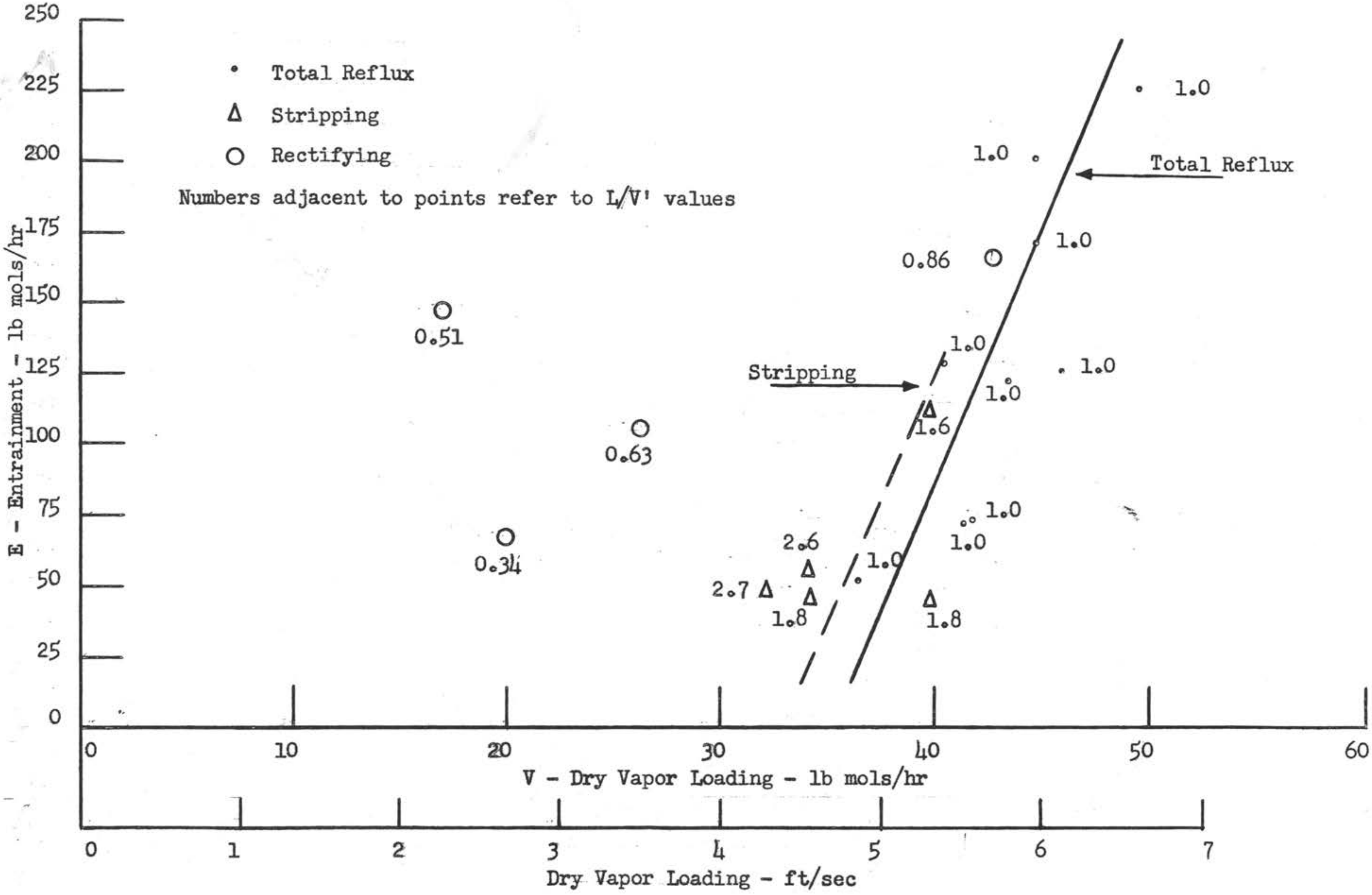
OVERALL EFFICIENCY AS A FUNCTION OF THE SLOPE OF THE OPERATING LINE

Figure 18



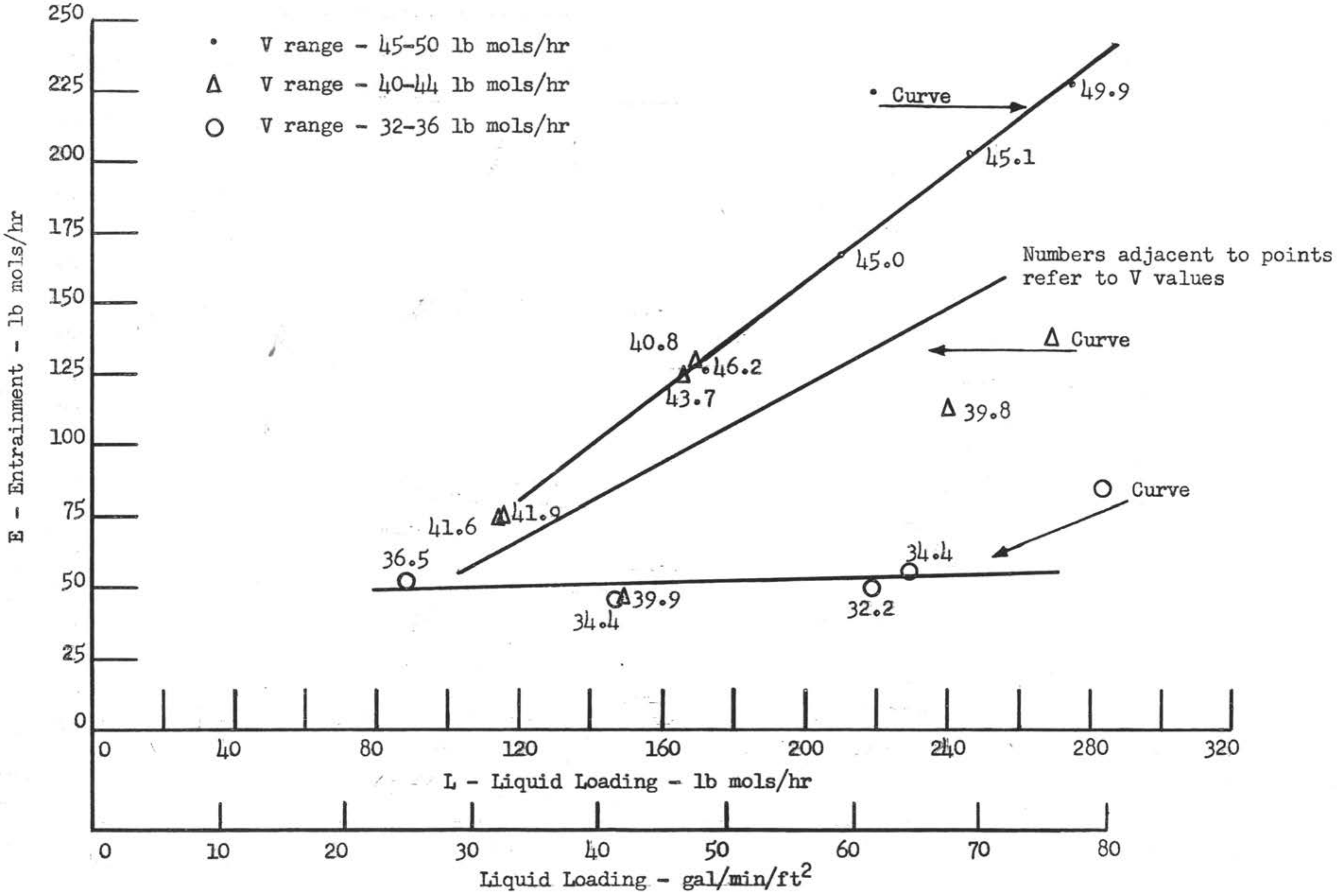
PRESSURE DROP PER TRAY AS A FUNCTION OF WET AND DRY VAPOR LOADINGS

Figure 19



ENTRAINMENT AS A FUNCTION OF DRY VAPOR LOADING

Figure 20



ENTRAINMENT AS A FUNCTION OF LIQUID LOADING

Figure 21

SAMPLE CALCULATION FORMULAS

(1) Dry vapor loading, V, lb. mols/hr.; non-rectifying runs.

$$V = S_n(h_g - h_f) \div \lambda_b$$

Where:

S_n = net steam consumption, lb./hr.

h_g = enthalpy of saturated steam at G_3, BTU/lb.

h_f = enthalpy of saturated water at G_2, BTU/lb.

λ_b = latent heat of vaporization of bottoms material, BTU/ lb. mol.

(2) Dry vapor loading, V, lb. mols/hr.; rectifying runs.

$$V = V_{(1)} - \underbrace{(D(T_6 - T_3) \times C_{pa} \div \lambda_b)}_{\Delta V_b}$$

Where:

$V_{(1)}$ = dry vapor loading as calculated in (1).

D = recycle stream sent to reboiler, lb. mols/hr.

T_6 = reboiler temperature, °F.

T_3 = recycle stream temperature, °F.

C_{pa} = average heat capacity between T_6 and T_3, BTU/lb. mol/°F.

λ_b = as in (1).

(3) Liquid loading, L, lb. mols/hr.

$$L = R + \underbrace{(R(T_2 - T_1) \times C_{pa} \div \lambda_5)}_{\Delta L_5}$$

Where:

R = reflux, lb. mols/hr.

T_2 = overhead temperature (bubble-point of x_5), °F.

T_1 = reflux temperature, °F.

C_{pa} = average heat capacity between T_2 and T_1, BTU/lb. mol/°F.

λ_5 = latent heat of vaporization of liquid on top tray, x_5, BTU/lb. mol.

(4) Wet vapor loading, V', lb. mols/hr.; total reflux.

$V' = L$ since L/V' must be unity.

(5) Wet vapor loading, V', lb. mols/hr.,; stripping.

$V' = L - W$

Where:
L = as in (3)

W = bottoms withdrawn and recycled as reflux, lb. mols/hr.

(6) Wet vapor loading, V', lb. mols/hr.; rectifying.

$V' = L + D$

Where:
L = as in (3)

D = as in (2)

(7) Entrainment, E, lb. mols/hr.

$E = V' - V$

Where:
V' = as in (4), (5), or (6).

V = as in (1) or (2).

(8) Operating line slope, L/V'.

Where:
L = as in (3)

V' = as in (5) or (6)

(9) Operating line y-intercept, y, mol fraction cyclohexane; stripping.

$7 = -(W \times x_w) \div V'$

Where:
W = as in (5)

$x_w = x_b$, mol fraction cyclohexane in bottoms.

V' = as in (5)

(10) Operating line y-intercept, y mol fraction cyclohexane; rectifying.

$$y = (D \times x_D) \div V'$$

Where:

D = as in (2).

$x_D = x_R$, mol fraction cyclohexane in reflux.

V' = as in (6).

(11) Overall efficiency, e, per cent.

$$e = n_t \div n_a$$

Where:

n_a = number of actual trays.

n_t = number of theoretical trays.

(12) Pressure drop per tray, ΔP, inches of water.

$$\Delta P = ((G_1 - G_5) \times 27.7) \div 5$$

Where:

$(G_1 - G_5)$ = total pressure drop, psi.

27.7 = inches of water ÷ by psi.

THESIS TITLE: PERFORMANCE OF THE KOCH KASKADE TRAY IN
DISTILLATION

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THESIS ADVISER: LEO GARWIN

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