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I hereby recommend that the thesis prepared under my supervision by Dinshaw Naoroji Daruwalla

entitled Experimental Investigations Into Steam Stripping Operations.

be accepted as fulfilling this part of the requirements for the degree of Doctor of Philosophy.

Approved by:

H. J. Garber

Wm. Licht Jr.

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EXPERIMENTAL INVESTIGATIONS INTO STEAM STRIPPING OPERATIONS

A dissertation submitted to the
Graduate School
of the University of Cincinnati

in partial fulfillment of the
requirements for the degree of

DOCTOR OF PHILOSOPHY

1943

by

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Table of Contents

List of Illustrations

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To professor Reuben Tour goes the credit for having suggested the simple and efficient method of external heating of the batch unit.

Introduction

It must be stated at the very beginning that, although the process of steam distillation has diverse applications in industry, there is very scant information on this unit operation either in the present literature or in the text books. To overcome these shortcomings a comprehensive theory on the principles of steam stripping was worked out by Garber and Lerman (5) and presented at the A.I.Ch.E. convention on November I7, I942 at Cincinnati. The task of obtaining experimental verification of the theoretical postulates put forth was entrusted to the author by professor Garber and the experimental work was begun in September I94I. The purpose of these investigations was:

I) To obtain experimental confirmation as to the degree of accuracy of the quantitative relationships postulated about the principles of steam stripping.

2) To show the vast superiority of the continuous counter current method of operation over the batch process, both as to saving in steam consumption and equipment, and as to the ease and smoothness of operation.

3) To set forth a method based on these investigations for designing steam distillation equipment for either batch or continuous counter current operation.

In steam stripping, the steam which is insoluble in the liquid mixture is blown into the distillation unit, and serves as a carrier to remove the volatile constituents. Boiling occurs when the additive pressures of the steam and the vapor pressure of the solution equals the total pressure imposed upon the system.

The steam which is intimately mixed with the solution counteracts part of the total pressure imposed upon the system and has the net effect of causing volatilization at a lower temperature than would be obtained by straight distillation. This could also be looked upon as the apparent reduction in pressure over the liquid or the imposition of a vacuum on the system. This is the key idea behind all so called "partial pressure distillations".

Any vapor or gas insoluble in the liquid may be used for stripping operations. Steam however, is preferred for the following reasons: The stripping steam can be readily condensed after leaving the system. Hence the volatile component can be easily recovered by mere decantation. Steam also serves as a h heat carrier to the system and simpifies the maintenance of a vacuum on the system because of its ease of condensation and low back pressure exerted by it at normal temperatures attainable with usual industrial cooling water.

Steam distillation finds varied and extensive uses in industry. A comprehensive discussion of these is given by Weis(IO) in his thesis. Typical among these uses are: the concentration and purification of the essential oils; the purification of high molecular weight fatty acida; the deodorization of fats and oils; the stripping of benzene, toluene and xylene and other aromatic compounds from absorption oils (straw or cycle oils) in by-product coke production; the recovery of propane, butane and pentane in the debutanizing processes of the petroleum industry; the removal of volatile

components from reduced crude prior to further processing; and the concentration of beers to light wine in the fermentation industry.

Theory Of Steam Stripping Operation

Batch Process:

Carey (4) has presented an expression for batch steam distillation for systems that follow Raoult's law and has credited these to McAdams. Bailey (I) in discussing the deodorization of edible fats and oils, developed a modified and approximated form of Carey's equation. Brash (2,3) also considered steam distillation in connection with the treatment of fats and their deodoriztion using the adiabatic expansion of steam as the basis of exposition. Garber and Lerman (5) have worked out a general equation for batch stripping using the method outlined by Carey. For an infinitesimal time interval $d\theta$, assuming that steam is sparged into the still at a constant rate and that enough additional heat is added, either from an outside source or from the condensing steam itself, so that the vapor phase is composed of only the steam and the volatile component, a differential material balance for the volatile component gives

$$-N_b\, dX = S_b\, Y\, d\theta \qquad (1)$$

Here N_b is the moles of inert component in the still charge, X is the instantaneous composition of the liquid in the still in moles of volatile component per mole of inert component, S_b is the moles of steam admitted per unit of time, and Y is

the instantaneous composition of the vapor leaving the still in moles of volatile component per mole of stripping steam. Equation (I) assumes that no entrainment occurs in the vapor. The variables in equation (I) may be reduced to two by obtaining Y in terms of other variables.

Carey's definition of vaporization efficiency (4) is $E = p/p^*$ where p is the partial pressure of the volatile component in the vapor phase, p* is the equilibrium partial pressure of the volatile component as dictated by the phase properties of the system and the temperature, E is the vaporization efficiency or the ratio of actual to full attainment of the equilibrium pressure in the vapor phase by the volatile component.

Applying Daltons law to the component in the vapor phase Y is the ratio of the partial pressure p of the volatile component to the partial pressure ($\pi - p$) of the stripping medium where π is the total pressure maintained in the vapor space of the still.

Hence $Y = p/\pi - p = Ep^*/\pi - Ep^*$ (3)

In accord with the phase rule for a three component system with two phases present- liquid and vapor - and with pressure specified, p* is a function of two variables, such as temperature T, and X the composition.

Hence $p^* = p^*(T,X)$ (4)

Inserting this implicit function for p* into equation (3) for Y and substituting the resulting value of Y into equation (I) the general equation for batch stripping is obtained.

$$-N_b\left[\frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}\right]dX = S_b\theta_b$$

For specific initial and final conditions, if X_f is the composition of the feed and X_r is the composition of the residue and θ_b the total time of steaming, the above expression can be integrated over the set limits and the general equation for batch stripping is

$$N_b\int_{X_r}^{X_f}\frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}\,dX = \int_0^{\theta_b} S_b\,d\theta = S_b\theta_b \qquad (5)$$

Implicit equation (5) involves assumptions that are usually valid. If experimentally determined data for the influence of T and X on p*,and E for the process, assumed constant, be inserted into equation (5) an explicit expression for the total steam consumed results. When two liquid layers are present in the still, the variation of the system is reduced to two, so that either with pressure fixed only one degree of freedom remains. Considering X to be the remaining independent variable, T is no longer subject to choice. Thus, in this case after π has been specified p* is a function of X only.

If the exact nature of the p*(T,X) function is not known estimates of stripping medium requirements for a batch process may be obtained by introducing some simple or approximate function for p* into equation (5). Three such functions may be considered, depending on whether Raoult's, Henry's or Lewis and Luke's (7) laws apply. Since the three laws have

the same analytical form all three case can be dealt with simultaneously.

Raoult's Law $p^* = P \frac{X}{I+X}$

Henry's Law $p^* = H \frac{X}{I+X}$

Lewis & Luke's Law $p^* = \pi K \frac{X}{I+X}$

This leads to the general expression for Y in terms of X:

$$Y = \frac{X}{(k - I)X + k} \qquad (6)$$

where $k = \frac{\pi}{EP}, \frac{\pi}{EH}, \frac{I}{EK}$ when Raoult's, Henry's or Lewis and Luke's laws apply respectively.

Inserting this value of Y in equation (I) and integrating

$$N_b \int_{x_r}^{x_f} \frac{1}{Y}\, dx = S_b \theta_b$$

$$N_b \int_{x_r}^{x_f} \frac{(k-1)X + k}{X}\, dX = S_b \theta_b$$

$$N_b \int_{x_r}^{x_f} \left[(k-1) + \frac{k}{X}\right] dX = S_b \theta_b$$

$$N_b \left[k \ln \frac{x_f}{x_r} + (k-1)(x_f - x_r)\right] = S_b \theta_b \qquad (7)$$

The moles of volatile component vaporized is $N_b (X_f - X_r)$ hence the ratio of the moles of steam per mole of volatile component distilled over is

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} = k \frac{\ln \left(\frac{X_f}{X_r} \right)}{(X_f - X_r)} + (k-1)$$

$$= \left[\frac{k}{X_{lm}} + (k-1) \right] \qquad (8)$$

where X_{lm} is the logarithmic mean value of X_f and X_r. When equations (7) and (8) are written for a system obeying Raoult's law the expression presented by Carey (4) is obtained thus:

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} = \frac{\pi}{EP} \left[\frac{1}{X_{lm}} + \frac{\pi - EP}{\pi} \right]$$

When Henry's law or Lewis and Luke's law applies, the corresponding answers are respectively:

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} = \frac{\pi}{EH} \left[\frac{1}{X_{lm}} + \frac{\pi - EH}{\pi} \right]$$

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} = \frac{1}{EK} \left[\frac{1}{X_{lm}} + 1 - EK \right]$$

It must be remembered that equations (7) amd (8) have limited application and serve for estimating purposes only. For more exact calculations the precise functional relationship for p* demanded by equation (4) must be determined experimentally over the required operating range and used in equation (5). The degree of accuracy of the approximations and the exact nature

of the function for the system light turbine oil- carbon tetrachloride studied will be discussed in detail in the later sections.

Equation (7) also shows that the amount of steam per ponnd of volatile component distilled over will be low when the ratio ofthe molecular weight of the volatile component to the inert component is low, and when the escaping tendency of the volatile component is high i.e. when k is small. It also indicates quantitatively how the total staem consumption is reduced by operating at high temperatures and low pressures.

At this stage it might also be shown that the approximate steam distillation equation found in most text books can be derived by making certain assumptions. These assumptions are as follows:

I) The effect of the inert component is negligible i.e. $N \longrightarrow 0$ If this is the case then $p^* \rightarrow P$ provided Raoult's law is followed thus:

$p^* = P \frac{X}{I+X} = P \frac{V}{V+N}$ where V is the moles of volatile component.

If $N \rightarrow 0$ $\quad p^* \rightarrow P$

2) Also assume $E = I$

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} = \frac{\pi - p^*}{p^*} = \frac{p_{steam}}{p_{volatile}}$$

$$\frac{S_b \theta_b}{N_b (X_f - X_r)} \frac{M_s}{M_v} = \frac{P_s}{P_v} \frac{M_s}{M_v}$$

$$\frac{W_s}{W_v} = \frac{M_s}{M_v} \frac{P_s}{P_v}$$

Since the assumptions made to arrive at the above expression are seldom valid, the limitations of this equation are self evident.

Continuous Steam Distillation:

In contrast to batch steam stripping two other methods can be employed, continuous counter current and continuous parallel flow stripping processes. In the former case the vapor last comes in contact with the fattest liquid in the system, hence will tend to be richer as regards the volatile component than in the parallel flow process where the exit vapors last contact the leanest liquid. Parallel flow stripping finds application only in a few cases. Thus in the refining of fatty acids the Wecker still employs this rather inefficient principle of distillation.

Let N_c be the moles of inert component fed to the still per unit of time in the counter current process. Assuming steady flow conditions, N_c is also the moles of inert component removed from the still per unit time.

The volatile component mole feed rate is $N_c X_f$ and the moles of volatile component leaving the system per unit time are $N_c X_r$. Then $N_c(X_f - X_r)$ is the mole rate of distillation of the volatile component. In the counter current process the vapor last engages the feed liquid with the composition X_f. If this vapor were in equilibrium with the incoming feed its composition would be predicted by the equilibrium properties of the system, as $p^* = p^*(T,X)$.

However the partial pressure of the volatile component in the vapor does not reach its full equilibrium value, but only some fraction of it. Calling this fraction E as in equation (2) the vapor leaving the still has the composition

$$Y_f = \frac{E p^*(T, X_f)}{\pi - E p^*(T, X_f)} \qquad (9)$$

In the parallel flow the vapor last encounters liquid of composition X_r and hence has composition

$$Y_r = \frac{E p^*(T, X_r)}{\pi - E p^*(T, X_r)} \qquad (10)$$

The staem consumption rates for the two processes is then determined by writing a material balance for the volatile component.

$$S_c = N_c(X_f - X_r)\,\frac{\pi - E p^*(T, X_f)}{E p^*(T, X_f)} \qquad (11)$$

$$S_p = N_p (X_f - X_r) \frac{\pi - E p^*(T, X_r)}{E p^*(T, X_r)} \qquad (12)$$

The moles of steam required per mole of volatile distilled is

$$\frac{S_c}{N_c (X_f - X_r)} = \frac{\pi - E p^*(T, X_f)}{E p^*(T, X_f)} \qquad (13)$$

$$\frac{S_p}{N_p (X_f - X_r)} = \frac{\pi - E p^*(T, X_f)}{E p^*(T, X_r)} \qquad (14)$$

Equations (I2) and (I3) are implicit exact relationships. If an explicit approximation is desired the relationships of equation (6) can be used, whence

$$S_c = N_c \left[h \frac{X_f - X_r}{X_f} + (h-1)(X_f - X_r) \right] \qquad (15)$$

$$S_p = N_p \left[h \frac{X_f - X_r}{X_f} + (h-1)(X_f - X_r) \right] \qquad (16)$$

$$\frac{S_c}{N_c (X_f - X_r)} = \left[\frac{h}{X_f} + (h-1) \right] \qquad (17)$$

$$\frac{S_p}{N_p(X_f - X_r)} = \left[\frac{k}{X_r} + (k-1)\right] \qquad (18)$$

Depending upon which solution law fits the case under consideration best the appropriate value of k may be used in equations (I5), (I6), (17), (18).

Comparison of the Batch and Counter Current Processes:

It can readily be seen that equation (8) for the batch process is very similar to equations (17) and (18) for the continuous processes. The difference is in the first term of the bracketed quantities where $\frac{1}{X_{lm}}$ replaces $\frac{1}{X_f}$ or $\frac{1}{X_r}$.

Hence a comparison of the steam consumption of the three types of processes for equal amounts of stripping can be made in a direct manner. Since $X_f > X_{lm} > X_r$ then $\frac{1}{X_f} < \frac{1}{X_{lm}} < \frac{1}{X_r}$

Hence when applied to equations (8), (17) and (18) it can be readily seen that for the same operating comditions and for same values of E, $\frac{S_p}{N_p(X_f - X_r)} > \frac{S_b\theta_b}{N_b(X_f - X_r)} > \frac{S_c}{N_c(X_f - X_r)}$

This is merely a qualitative deduction based on the approximated explicit results. To ascertain the exact differences in the steam consumption for the batch and counter current processes for the same operating conditions, let $\Delta S = \frac{S_b\theta_b - S_c}{N_b(X_f - X_r)}$

$$\Delta S = \frac{1}{(X_f - X_r)} \int_{X_r}^{X_f} \frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}\, dx - \frac{\pi - Ep^*(T,X_f)}{Ep^*(T,X_f)} \qquad (19)$$

According to the law of the mean for integrals

$$\frac{1}{(X_f - X_r)} \int_{X_r}^{X_f} \frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}\, dx = \frac{\pi - Ep^*(T,X_\xi)}{Ep^*(T,X_\xi)}$$

where $X_f > X_\xi > X_r$

Now from the physical nature of the mixtures, irrespective of what solution law applies to the system provided the carrier is really inert and that the system does not exhibit azeotropism, since $X_f > X_\xi$ then $p^*(T,X_f) > p^*(T,X_\xi)$. Thus ΔS in equation (19) is always positive and the stripping medium consumption for the batch process for the same operating conditions is always greater than it is for the counter current process.

Considerable knowledge could be gained by calculating an estimate of the superiority of the counter current process over the batch process assuming one of the solution laws to apply.

Let $a = X_r / X_f \leq 1$ (21)

Then

$$\Delta S = \frac{k}{X_f}\left[\frac{\ln \frac{1}{a}}{(1-a)} - 1\right] \qquad (22)$$

$$\% \Delta S = 100\, \frac{\frac{\ln \frac{1}{a}}{1-a} - 1}{X_f\left(1 - \frac{1}{k}\right) + 1} \qquad (23)$$

A quantitative picture of the effect of a and X_f alone on ΔS is given in fig. (6), the other variables are held constant. This plot clearly shows the pronounced relative superiority of the counter current flow process over the batch process, particularly when the concentration of the volatile component in the residual liquid is reduced to a very low figure as is done in deodorization. Qualitative estimates of relative superiority for several systems are cited by Weis (10), Kure (6) and Nelson (8).

Besides the considerable saving in steam the continuous counter current process also has other pronounced advantages These are, steady smooth running operation; constancy of liquid levels; constancy of liquid composition distribution and of temperature throughout the process; short exposure time of the liquid to the high temperature zone, thereby decreasing the amount of undesirable side reactions; higher product throughput per unit volume of apparatus; better dispersion of steam throughout the solution.

Selection of Materials and Determination of their Physical Properties

To test the validity of the theory stated it was decided to prepare a mixture of an inert medium and a volatile solvent mutually soluble in all proportions, and to subject this mixture to steam stripping under carefully regulated conditions.

The inert solvent should be non-corrosive and have a high boiling point or negligible vapor pressure. It should also be immiscible and non-reactive with water and have a low emulsification point. After due consideration it was decided to use a light turbine oil because it seemed to satisfy approximately all the properties desired.

Carbon tetrachloride was selected as the volatile solute for the following reasons:

1) It is immiscible with water
2) It is cheap and was available even in war-time.
3) It is non-inflammable and non-explosive and very slightly toxic.
4) It is relatively non-corrosive and only slightly reactive with water at the conditions encountered in the experimental work.
5) It is completely miscible with the solvent in all proportions.
6) It has a high vapor pressure and a low boiling point.
7) Its density is decidedly different from that of the turbine oil.

Hence the mixture selected for study consisted of varying amounts of carbon tetrachloride in a fixed quantity of light turbine oil.

Method of Analysis of Mixture:

It was realized right at the start that only such mixtures should be used that could be subjected to some method of rapid routine analysis during the course of distillation that may enable the operator to determine the composition. After prolonged investigation the most satisfying method found was that of determining the density of the mixture at a fixed temperature. The densities of different mixtures of varying proportions of carbon tetrachloride in turbine oil were determined carefully both by means of a pycnometer and a Westphal chainomatic balance. The results of these determinations are plotted (Refer figure 1). It must be pointed out that the pycnometer method though more accurate is time consuming and rather messy. It was found extremely difficult to get the last traces of the oil washed away from the outside surface of the bottle before weighing.

For purposes of comparison, the pycnometer was used to determine the density and the composition of all the samples obtained from the batch process of distillation, and the Westphal chainomatic balance was used for all samples from the continuous counter current method of operation. This was necessitated due to the fact that the batch process theory demands a greater precision in measuring the initial and final composition of the liquid mixture, since the steam consumption depends upon the reciprocal of the logarithmic mean value of the concentration. A slight error in either X_f or X_r would

produce a appreciable error in the value of $S_b \theta_b$.

Assuming that the system follows Raoult's, Henry's or Lewis and Lukes laws

$$S_b \theta_b = N_b \left[k \ln \frac{X_f}{X_r} + (k-1)(X_f - X_r) \right]$$

Now let an error in density measurement cause a small error ΔX in the values of X_f and X_r. The apparent steam consumption $S_b' \theta_b$ will then be:

$$S_b' \theta_b = N_b \left[k \ln \frac{X_f + \Delta X}{X_r + \Delta X} + (k-1)(X_f - X_r) \right]$$

The percent error in steam consumption would therefore be

$$\frac{S_b \theta_b - S_b' \theta_b}{S_b \theta_b} \times 100$$

$$= \left[N_b k \ln \frac{X_f}{X_r} - N_b k \ln \frac{(X_f + \Delta X)}{(X_r + \Delta X)} \right] 100$$

$$= 100 N_b k \ln \frac{\frac{X_f}{X_r}}{\frac{X_f + \Delta X}{X_r + \Delta X}}$$

This theoretical deduction is now applied to the actual results of experiment of run No. 1 in table 11. In this case ΔX is taken to be one percent of the real value of X_f.

Hence $\Delta X = 0.008181$

$X_f + \Delta X = 0.8263$

$X_r + \Delta X = 0.0900$

$$\% \text{ error} = 100 \times 0.0436 \times \frac{742}{1040} \times 2.303 \times \log \frac{0.8181 / 0.08181}{0.8263 / 0.0900}$$

$$= 0.27 \%$$

Determination of Molecular Weight of Solvent:

The average molecular weight of the light turbine oil was determined by the standard cryoscopic method. A known weight of the oil was introduced into a known weight of pure recrystallized benzene and the depression in the freezing point determined on a Beckman thermometer. Knowing the F.P. constant for benzene as 5.12, the mean molecular weight of the oil was found to be 450 from the following formula.

$$M = \frac{1000 K_f g}{G \Delta T_f}$$

Here M is the molecular weight of the oil, T_f is the depression in freezing point observed, K_f a constant and g the grams of oil and G the grams of benzene.

The value of 450 for the turbine oil was used in all the other calculations.

Determination of the Vapor Pressure of the Oil CCl4 Mixture:

The vapor pressures of mixtures of varying compositions were determined at different temperatures by the static method. The apparatus (see fig. 7) consisted of a two litre round bottom flask "A" fitted with a thermometer and placed inside a thermostat "B". Flask "A" was connected to a 250 c.c. flask "C" placed inside an ice bath and acting as an auxiliary trap to catch any carbon tetrachloride that may happen to leave the mixture and condense out. Flask "C" was connected to a manometer "D" and a two litre bottle "E" acting as a vapor container and a pressure equalizer. The whole system was connected to a Cenco Hyvac vacuum pump.

The method of procedure was as follows: A large volume of mixture of known composition was placed in flask "A". The system was then evacuated at a constant temperature until the liquid in "A" just begins to boil. An interval of 15 minutes was allowed for the vapor to come in equilibrium with the liquid and the vapor pressure at the fixed temperature was read. The temperature of the thermostat was progressively raised and several vapor pressure readings taken. Due care was taken during measurements to see that no liquid carbon tetrachloride collected in flask "C", thus ensuring a reasonably constant liquid composition. It was found that there was no appreciable change in the density of the liquid mixture before and after the run. Hence constant composition was assumed.

The vapor pressure of the oil at the relatively low temperatures of measurement was assumed to be negligible.

The results of the experiment were plotted (see fig. 3) using vapor pressure as the ordinate and temperature as the abscissa. The percent composition by volume i.e. the c.c. of carbon tetrachloride per 100 c.c. of oil at 20 degrees C. was the parameter. From these family of curves, another set of curves was constructed (see fig. 4) using percentage volume composition as abscissa and vapor pressures as ordinate and the temperature as parameter.

It must be noted that the percent volume composition of the mixture can readily be changed to the percent weight composition using the known values of the density of the pure oil and carbontetrachloride. The weight percent composition can then be expressed in terms of moles of carbontetrachloride per mole of oil from a knowledge of the molecular weights of the components. Detailed sample calculations on these transformations will be given in later sections.

BATCH STEAM DISTILLATION

For investigation into batch staem distillation operations the unit constructed by Weis (10) was employed with a few modifications. The unit constructed was of pilot plant size and originally intended to be used for both batch and counter current methods of operation. It consisted of a column still 6" inside diameter and 5' high (see fig. 9) supplied with a Spanish prison baffle, a steam sparger, a steam superheater, a steam and product combination condenser cooler, a feed preheater, a residue cooler, a residue pump, and feed, product and residue tanks. Arrangements were made for working the column both under atmospheric pressure and under vacuum.

The following control instruments were fitted:

1) Manometers measuring pressure both at the top of the still and over the product tank.

2) Orifice flometers to measure the rate of flow of steam, feed, and residue.

3) Steam pressure gauge

4) Thermometers for measuring the temperature of the steam, preheated feed and also the still temperature.

For further details of the unit refer to Weis's original thesis (10).

A few preliminary runs made with the existing equipment proved that certain modifications were necessary for efficient and accurate performance and control. Hence the

following changes were introduced:

1) The steam superheater was found dangerously close to the product tank containing the highly volatile CCl_4 which would induce losses by vaporiztion to the vacuum system during low pressure operation. The superheater was therefore relocated to avoid such losses.

2) No provision was made for heating the column externally in order to avoid condensation of the steam in the column. Electrical resistance heating, electrical induction heating, and heating with steam coils inside and outside the column were all considered. Due to the scarcity of materials imposed by war conditions and because of adequate machine shop facilities it was finally decided as suggested by professor Tour to wind a quarter inch diameter copper tube round the body of the column and use it as an external steam coil. To provide for adequate metal to metal contact the copper coil was soldered to the still body. Low pressure steam was used for heating in these coils.

3) The original steam sparger had exceedingly fine holes which easily got plugged up. It was also found that over 50 psi steam pressure was necessary to force any appreciable quantity of steam through these fine openings. To use 15 psi steam and avoid constant plugging the sparger holes were enlarged.

4) To have more flexible control of the steam rate, the original globe valve was replaced by a precision "Lunkenheimer" needle valve with a graduated indicator wheel to help control the opening.

5) The feed tank did not create a sufficient head to force the viscous feed into the still at the desired rate. Hence a system was rigged up whereby compressed air at a fixed pressure could be used to drive the feed into the column.

Method of Operation:

The method of operation was as follows: An approximately 25 lb. batch of turbine oil carbon tetrachloride mixture was made up and analysed to determine the initial composition. It was then introduced into the feed tank. Steam was blown into the empty column for over two hours to ensure a uniform temperature of about 200 degrees F inside the column. The feed was then slowly led into the column through the feed heater. The feed usually entered the column around a temperature of 150-180 degrees F. When about 25 pounds of the batch was charged into the column, the feed valve was shut, the condenser cooling water was turned on and the product tank valve opened. Superheated steam at around 15 lb pressure and around 250 degrees F (corresponding to about 30 deg. F superheat) was gently led in through the sparger at a steady rate of flow as indicated by the flometer. At regular intervals the still pressure and temperature; the steam pressure temperature and flow rate were noted. Periodically samples were taken of the batch liquid and analysed. From a few preliminary test runs the best allowable steam rate of flow was determined, so that the stripping operation could complete itself within an hour

or an hour and a half. Since density methods of analysis for the composition were not reliable below a concentration of 1% by weight, no attempt was made to strip the oil of CCl_4 below 1% concentration. When the CCL_4-oil mixture was stripped to about 1-2%, the steam was shut off. The residue was passed through the residue cooler and discharged in a bucket at approximately room temperature and weighed. The steam and CCl_4 were collected into a two litre bottle and separated by means of a separatory funnel. The individual portions were then weighed. The two main variables studied were changing initial composition of the feed, other conditions being kept the same; and changing pressure in the column using a constant feed composition.

Difficulties of Operation:

The operation of the unit was by no means a simple matter. The major obstacle was the exact control of all the variable factors liable to affect the final results. As in all chemical engineering development work, more variables had to be looked after than was efficiently feasible for one man. Improvements had to be constantly effected to see that all the valves, gauges dials, thermometers and flometers were easily accessible and easily read at once by a single operator. The most difficult task was to keep the steam rate down to a very low rate of about 1 lb per hour and maintain it constant. Any oversight or carelessness in controlling the steam rate caused considerable

entrainment losses, the oil passing over into the condenser and the product tank, in spite of the elaborate baffle system. Because of these intrinsic difficulties of operation due to the lack of sufficient control devices, numerous runs had to be discarded time and again.

Discussion of Data:

All the data for batch steam stripping runs are given in tables I & V; Table II gives a summary of results for comparison and correlation purposes. These tables present the result of 10 typical runs. Run numbers 1-5 are at atmospheric pressure, initial feed composition being varied from approximately 15% by volume CCl_4 in light turbine oil to 3%.
Run numbers 6-10 were all carried out at approximately constant feed composition of 14-15%. The stripping was performed under a vacuum varying from an absolute pressure of 564.5 mm to 195.6 mm mercury. Every attempt was made to see to it that all other variables like the steam pressure, the steam temperature, the steam flow and the batch temperature were held constant as much as possible. It was comparatively easy to maintain a steady rate of steam flow,but despite all precautions, both the batch temperature and the vacuum would fluctuate within narrow limits during the entire run. A careful scrutiny of the various runs in table I will show the exact fluctuations mentioned above.

It must be realized that, since the unit was of a pilot

plant type, and since no expensive control devices were available, the slight fluctuations in the other variables are to be expected and due allowances should be made for them. For calculation purposes therefore the following devices were employed.

1) For runs at atmospheric pressure the best values for the temperature of the batch liquid was calculated by Gauss's method(9). This mathematical device is highly effective and corrects for the fluctuations actually observed during the process.

2) For runs under vacuum, since the fluctuations were of the order of 0.1 inch of mercury on 5 inches total pressure, an arithmetic average value was used as the pressure prevailing over the batch during distillation.

Table V gives the values of the function $\frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}$ for specified values of the percentage composition by volume of the CCl_4 - oil mixture. Here π represents the total pressure in mm of mercury over the system, p* the equilibrium partial pressure of CCl_4 exerted at the temperature prevailing in the column during stripping (see fig. 4) and E the vaporization efficiency, which was found to be 80% for this particular unit (using the same sparger and fluid depth in all runs). The tabulated data in table V is plotted and gives the curves of fig. 5 . These curves are then employed for graphical integration in order to evaluate the exact value of the quantity of steam necessary to bring about a definite amount of stripping.

A few sample calculations of the amount of steam required will make this discussion clear.

Sample Calculations:

Refer tables I, II, V and figs. 1 to 5

Run No. 3

Weight of batch	25.25 lbs
Moles of inert oil, N	0.04301
Feed composition Vol. %, V_f	12.84
Moles CCl_4/ mole oil, X_f	0.7013
Residue comp. Vol. %, V_r	1.55
Moles CCl_4/ mole oil X_r	0.0847
Still press. mm Hg.	744
Still temperature as per Gauss method	96 deg. C
Vaporization efficiency	80 %

From fig. 5, using Simpson's rule

$$\int_{1.55}^{12.835} \frac{\pi - Ep^*(T, V)}{Ep^*(T, V)}\, dV = 55.57 \text{ units}$$

Hence $$\int_{X_r}^{X_f} \frac{\pi - Ep^*(T, X)}{\pi - Ep^*(T, X)}\, dX = \frac{66.47 \times (0.70129 - 0.08474)}{12.835 - 1.55}$$

Hence $S_b\theta_b = 0.0431 \times 18 \times 5.887 \times 0.61655$

$= 2.35$ lbs

Steam observed $= 2.31$ "

CCl_4 present in feed $= \frac{25.25 \times 24}{124} = 4.887$ lbs

CCl_4 present in residue $= \frac{25.25 \times 2.9}{102.9} = 0.7116$ lbs

Hence lb steam/ lb CCl_4 $= \frac{2.31}{(4.887 - 0.7116)} = 0.5531$

Note that the percent composition of feed and residue are determined from the density composition chart (see fig. 1).

Calculations assuming Raoults law to apply:

Full vapor press. of CCl_4 at 96° C, P = 1310 mm Hg.

EP = 1040 mm Hg., N = 0.0430, E = 80 %, X_f = 0.7013

X_r = 0. 0847, π - 744 mm Hg.

$$S_b \theta_b = \frac{744 \times 0.0430 \times 18}{1040} \left[2.303 \log \frac{0.7012}{0.0847} - \frac{(744 - 1040)}{744} \right]$$

= 1.602 lbs.

Run No. 6:

Weight of batch	25 lbs
Moles of inert, N	0.0435
Feed Comp. Vol. %, V_f	14.81 %
X_f	0.8074
Residue comp. V_r	1.6 %
X_r	0.0435
Barometric press.	29.65" Hg.

Manometer reading over batch	7.425" Hg.
Still press. = (29.65 - 7.425)25.4 =	564.5 mm Hg.
Still temp. per Gauss method	76.6 deg. C
Vaporization efficiency	80 %

$$\int_{V_1}^{V_f} \frac{\pi - Ep^*(T,V)}{Ep^*(T,V)}\, dV = 52.117$$

Hence

$$\int_{X_r}^{X_f} \frac{\pi - Ep^*(T,X)}{Ep^*(T,X)}\, dX = \frac{52.117 \times 0.7217^5}{13.2}$$

$$\text{Steam consumed} = \frac{0.0435 \times 18 \times 52.117 \times 0.7217^5}{13.2}$$

$$= 2.231 \text{ lbs}$$

Steam collected = 2.34 lbs

Assuming Raoult's law to obey:

Full vapor press. CCl_4 at 76.6° C, P (see fig. 2) = 760 mm Hg.

EP = 608, = 564.5 mm, N = 0.0435, $X_f = X_r = 0.7217$

$$S_b\theta_b = \frac{564.5 \times 0.435^5 \times 18}{608}\left[2.303 \log \frac{0.8094}{0.0435^5} - \frac{43.5^5 \times 0.7217}{564.5^5}\right]$$

$$= 1.575^5 \text{ lbs}$$

The rest of the calculations for the other runs are made in a similar manner.

It would be very instructive to compare the results of the ten runs given in table II and see what important deductions could be made. The first thing to notice is that the

nominal working capacity of the unit is 25 lbs of oil CCl_4 mixture, of a composition not exceeding 15 % by volume . This upper limit of composition is fixed due to the fact that if higher concentrations of CCl_4 of in oil are used, appreciable losses of CCl_4 vapor from the mixture results during the initial stage of charging the feed tank because of the volatility of the CCl_4. It is also to be noted that the lowest limit of stripping is 1.28 %. No stripping below 1.28 % was carried out because of the unreliability of the density measurements as an indication of the compositon below this range.

The first five runs were made at atmospheric pressure and the initial feed composition changed from 14.97 % to 3.21 %. Also the pounds of steam required to strip one ton of CCl_4 from each batch progressively increases for decreasing feed composition. The lbs of steam/ ton of CCl_4 ranges from 1038 lbs to 2440 lbs, an approximate 100 % increase. This is in accord with the theory of steam stripping as postulated before (5). The increase in steam consumption here is due to the decrease in the driving potential or concentration difference, $(X_f - X_r)$.

It is to be noticed that there is excellent agreement between the calculated values of the steam consumed and the actual values of the steam collected. This goes to prove that the theory of steam stripping as postulated by Garber and Lerman (5) is fundamentally sound.

Runs number 6 to 10 are made at still pressures varying from 564.5 mm Hg. to 125.6 mm. The feed composition here was

kept approximately constant at 14 to 15 %. The most remarkable fact to notice in this case is the considerable saving in steam affected per ton of CCl4, other conditions being approximately the same. Thus compare the results of run No. 1 with run No. 10. In both cases 25 lbs of batch was charged at approximately the same initial composition and stripped to a concentration of 1.49 % and 1.97 %. At atmospheric pressure of 742 mm the steam required to strip one ton of CCl_4 is 1038 lbs as compared to a mere 182 lbs at a pressure of 195.6 mm. Hence a decrease of pressure of 74 % has produced a steam economy of about 83 %. The values of steam actually collected and the values calculated from theory are also in close agreement whilst operating under vacuum . This agreement and the great saving in steam economy lend further support to the theory of Garber and Lerman (5).

The final point to observe is that steam consumption values calculated on the basis that Raoult's law may apply to the system are in error by as much as 30 to 60 %. Hence it is concluded that for this particular system Raoult's law does not apply, which fact is also borne out by the phase data of the system. All the same, for design purposes it could be assumed that Raoult's law applies and an estimate of the steam consumption necessary to bring about a specified amount of stripping could be obtained. The error in calculation would eventually be compensated for by using a safety factor 100 to 200 %. This procedure should only be applied when exact vapor pressure temperature relationships for a mixture are not

available. The value of vaporization efficiency of 80 % was arrived at by a process of adjustments. The method of calculation was much the same as in multiple effect evaporator calculations. An efficiency of 70 % was assumed and the steam calculated for all the runs. The efficiency values are then progressively adjusted until the best agreement was obtained between the calculated and the observed values for steam consumption. In accord with Carey's (4) theory for the vaporization efficiency, E should be a function only of the steam sparger hole diameter and the fluid depth through which the bubbles of steam rise. The constancy of E at 80% substantiates this theory. Weis (10) in his thesis mentions that in industrial steam distillation units, the vaporization efficiencies vary between 50 % in the distillation of high molecular weight oils to 100 % in the distillation of organic compounds with fairly low molecular weights. Bailey (1) states that his plant tests indicated efficiencies of 70 to 90 % with palm oil. Hence the value of 80 % realized with the exixsting pilot plant seems to be quite reasonable.

COUNTER CURRENT CONTINUOUS STEAM STRIPPING —

The close agreement between theory and experimental results acted as an incentive towards further experimental work in continuous counter current mode of operation. The existing pilot plant unit was not well suited for this method of operation. To make the necessary alterations would have consumed quite a lot of valuable time since it involved major machining operations on the column proper which would have to be dismantled. To save time therefore, an experimental unit of glass was built. This unit (see fig. 10) consisted of the packed column "D" made of 2" diameter pyrex glass tube 4' 8" long. Column "D" was well insulated and fitted with an internal electric heating coil "E" connected to a variable rheostat. This coil was placed inside a central 8 mm pyrex glass tube running through the entire glass column. Arrangements were made for measuring the temperatures at the top and bottom of the column as well as the pressure within the column by means of the manometer "F". The other connections to the column were the feed and steam inlet, the steam and CCl_4 vapor outlet and the residue outlet. Two five gallon capacity bottles "B" and "M" were used as the feed and residue receivers. Capillary flometers "J" and "C" were installed to measure the steam and feed rate. The product after passing through the condenser "G" could be collected in either of the two product receivers "H" . The residue passed through the cooler "K" into the residue bottle "M". Arrangement was also made to withdraw samples

into test tube "L" at intervals, even under a vacuum. The whole system could be operated under atmospheric pressure and under vacuum. Steam was generated in a small metal boiler fitted with a manometer and a copper coil superheater. The rate of flow of feed was measured by directly noting on the balance "A" the loss in weight of the feed bottle at regular intervals The flometer therefore acted only as an indicating device. The residue was continuously withdrawn at such a rate as to ensure a constant residue level of about 3" at the bottom of the column. This liquid level also acted as a seal and prevented the steam from short circuiting directly into the residue cooler "K". Broken porcelain crushed and graded to mesh was used as a packing. Over 8" of vapor space was left over the packing inside the column to prevent entrainment losses.

Method of Operation:

After a considerable amount of experimenting, it was found that the following method of procedure gave the best results. The entire column with the packing was brought to a uniform temperature of around 90 to 100 degrees C before the feed was led in. This was done by blowing steam into the column for over two hours until the top and bottom thermometers registered approximately the same temperature and the steam just began to come over into the condenser. At this stage the electric heat was turned on, the column quickly drained of all the condensed steam

and the feed led in at a constant rate. The steam rate was held constant and the feed rate was so adjusted as to avoid "tower hold up" effects and flooding. Since feed rate cannot be arbitrarily varied over a wide range, once the steam rate was fixed, the composition of the residue was also fixed. During the stripping process the electric heat was so adjusted as to keep the temperature inside the column constant. As soon as steady conditions were established, the rate of feed, the still temperature, the still pressure, the residue composition and the steam flow were noted at regular intervals. The entire run lasted about one hour. At the end of the run, the residue was analysed the steam and CCl_4 separated and weighed.

Difficulties of Opeartion:

Since two additional variable factors, the feed rate and the residue withdrawal rate were introduced, considerably greater difficulties were experienced in operating this unit than the previous batch still. As usual it required an acute alertness and constant watchingto see that steady condition was reached and maintained. A slight change in any one variable factor would have serious repurcussions throughout and would ruin an entire run. After some practice it was however possible to read the loss of feed weight; the steam temperature, pressure and flometer; the feed flometer; the still temperature and pressure within an interval of 30 seconds. The greatest difficulty

however was to judge the steam rate so as to avoid over steaming. The amount of steam required to perform a given stripping was so small that it taxed all ones patience and ingenuity to realize and maintain this small rate of flow.

Discussion of Experimental Results:

The data for 14 representative runs are given in table III. Table IV is made up for comparison and correlation purpose and gives a quick and easy reference to all pertinent and relevant figures.

Table III includes 14 different runs made under atmospheric pressure and constant feed rates. The variables employed are the initial feed composition and the still temperatures. Values for the feed rate; the steam temperature, pressure and flow; the still temperature and pressure and the residue density and composition are also given. Considering the nature of the unit and the inherent difficulties of operation, from a chemical engineering point of view the results given are fairly constant. Referring to table IV it is noticed that the feed rate varies from 3.7 lbs per hour to 9.87 lbs per hour, the highest rate allowable if flooding of the column is to be avoided. The rate of feed is determined by the allowable steam rate which is fixed arbitrarily. For the sake of convenience each run was over a period of one hour. The feed composition has been varied progressively from 3.62 % by volume to 15.94 %

by volume and the temperature of stripping from 90 degrees C to 110 degrees C. In each case consistent agreement was obtained between the amount of steam actually collected and the amount calculated using the theoretical method of Garber and Lerman(5). The average deviation is on the order of about 2 %. These deviations are calculated on the basis of a vaporization efficiency of 90 %, selected as the best value. The manner in which this value was arrived at was to first assume a perfect agreement between theoretical and calculated steam quantities and solve for the efficiency. A suitable average value is then selected and the steam required calculated on this basis. A few sample calculations will make the procedure clear.

Sample Calculations:

(see table IV and figs. 1, 3, 4)

Run number 1

Feed weight 8.813 lbs

Feed comp. vol. %, V_f 3.62

Feed comp. Wt. %, $= \frac{3.62 \times 1.595}{0.853} =$ 6.768

Moles of inert, $N = \frac{8.8125 \times 1.595}{106.77 \times 450} =$ 0.01834

Residue density 0.8580

Residue comp. vol. % 0.94 %

Moles CCl_4/ mole oil, $X_f = \frac{3.62 \times 1.595 \times 450}{154 \times 0.853 \times 100} = 0.7758$

Moles CCl_4/mole oil $X_r = \frac{0.1492}{154 \times 0.01834} = 0.0528$

CCl_4 in the feed $= \frac{8.8125 \times 6.768}{106.76} = 0.5586$ lbs

CCl_4 recovered $= 186.1 \times 0.0022 = 0.4094$ lbs

CCl_4 in residue $= 0.1492$ lbs

$X_r = \frac{0.1492}{154 \times 0.01834} = 0.05282$

$\pi = 729$ mm, $T = 110$ C, $p^*(T, X_f) = 156$ mm (see fig. 4)

Steam collected $= 84.8$ grams $= \frac{84.8 \times 0.0022}{18} = 0.01036$ lb moles

$$E = \frac{\pi}{\left[\frac{S}{N(X_f - X_r)} + 1\right] p^*(T, X_f)} = \frac{729}{156\left[\frac{0.010354}{0.01834 \times 0.14494} + 1\right]}$$

$$= 95\ \%$$

Now recalculate back assuming E = 90 %

$$S_c = \frac{N(X_f - X_f)(\pi - E p^*(T, X_f))}{E p^*(T, X_f)}$$

Steam $= \frac{0.01834 \times 0.14494 \times 588.6 \times 18}{140.4 \times 0.0022} = 91.1$ grams

Steam collected $= 88$ grams

% difference $= \frac{91.1 - 88}{91.1} = 3.2\ \%$

Lbs steam/ton $CCl_4 = \frac{88 \times 2000}{186} = 437$ lbs

SUMMARY

An experimental investigation into the phenomena of steam stripping was carried out to verify the theoretical postulates set forth by Garber and Lerman (5).

The system chosen for investigation was a mixture of inert light turbine oil and carbon tetrachloride. The vapor pressure data and other physical data for the system were investigated first. The mixture was then subjected to steam stripping in a batch still and also in a continuous counter current still. The variables employed for the batch process were initial feed composition and the pressure in the still over the batch. Those for the counter current process were initial feed composition and the temperature within the column. In both cases excellent agreement was obtained between the amount of steam actually collected and the amount calculated from theory to perform a specified stripping operation. The consistent agreement between calculated and observed values in steam consumption over a wide range of variation in still temperature and pressure and in initial feed composition lend ample support to the validity of the theory of steam stripping put forth by Garber and Lerman.

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Nomenclature

a - ratio of final to initial conc. of volatile component in liquid.

E - Vaporization efficiency

H - Henry's coefficient

k - Constant in liquid vapor expression when one of the three simple solution laws apply

K - Equilibrium onstant in Lewis and Luke's law, $y = Kx$

N - Number of moles of inert

P - Full vapor pressure of volatile component

p - Partial pressure of the volatile component in the vapor

p*- Equilibrium partial pressure of the volatile component in the vapor

S - Moles of steam

T - Temperature

V - Volume % composition of the liquid mixture

X - Composition of volatile component in liquid, moles per mole of inert

x - Mole fraction of volatile component in liquid

Y - Concentration of volatile component in vapor, moles per mole of ~~inert~~ steam

y - Mole fraction of volatile component in vapor

θ - Time

π - Pressure in the vapor

Subscripts:

b - batch process

c - counter current process

f - feed

p - parallel flow

r - residue

TABLE I

Run No. 1

Time	Feed		Still		Steam	
Mins.	Density	Wt. % comp.	Press ins.Hg	Temp. °F	Temp. °F	Flometer
0	0.9495	28.0	29.2	170	215	2
5	0.9191	18.8	29.2	188	215	2
10	0.9097	15.7	29.2	192	220	2
15	----	----	29.2	196	220	2
20	----	----	29.2	200	220	2
30	0.8875	8.7	29.2	206	220	2
40	0.8722	4.7	29.2	208	223	2
50	0.8667	3.9	29.2	210	223	2
60	0.8644	2.8	29.2	210	221	2
70	----	----	29.2	210	220	2
80	----	----	29.2	210	220	2
90	0.8642	2.8	29.2	211	220	2
Run No. 2						
0	0.9467	27.0	29.2	180	215	4
10	0.9040	13.7	29.2	190	220	4
20	0.8906	9.4	29.2	198	220	4
30	0.8795	6.6	29.2	204	232	4
40	0.8712	4.5	29.2	208	240	4
50	0.8673	3.6	29.2	210	280	4
65	0.8638	2.7	29.2	211	280	4
80	0.8630	2.5	29.2	211	310	4
95	0.8625	2.4	29.2	211	310	4
Run No. 3						
0	0.9369	24.0	29.3	192	215	4
10	0.8907	9.5	29.3	195	215	4
17	0.8802	6.7	29.3	205	215	4
27	0.8681	3.8	29.3	208	215	4
37	0.8656	3.15	29.3	210	215	4
47	0.8649	2.9	29.3	210	215	4
Run No. 4						
0	0.9280	19.7	29.2	188	215	4
5	0.9217	19.3	29.2	193	215	4
10	0.8935	10.4	29.2	197	220	4
15	----	----	29.2	200	220	4
20	0.8861	8.3	29.2	202	220	4
25	----	----	29.2	204	220	4
30	0.8763	5.9	29.2	206	220	4
35	0.8722	4.7	29.2	208	220	4
45	----	----	29.2	210	220	4
60	0.8689	4.0	29.2	210	220	4
70	0.8649	3.0	29.2	210	220	4
80	0.8641	2.8	29.2	210	220	4
90	0.8641	2.8	29.2	210	220	4

TABLE I

Run No. 5

Time	Feed		Still		Steam	
Mins.	Density	Wt. % comp.	Press ins.Hg.	Temp. °F	Temp. °F	Flometer
0.	0.8856	6.0	29.2	180	215	4
10	0.8884	?	29.2	199	215	4
20	0.8751	5.5	29.2	204	225	4
30	----	---	29.2	208	245	4
35	0.8674	4.5	29.2	208	250	4
40	----	---	29.2	209	250	4
45	0.8648	3.6	29.2	210	250	4
50	----	---	29.2	210	235	4
55	----	---	29.2	210	225	4
60	0.8644	2.9	29.2	210	220	4
Run No. 6						
0	0.9470	27.1	6.8	184	150	2
5	0.8997	12.4	8.0	174	175	2
10	0.8861	8.3	8.3	168	175	2
15	0.8784	6.4	8.3	166	175	2
20	----	---	8.3	192	175	2
25	0.8688	3.7	8.3	192	175	2
Run No.7						
0	0.9498	28.0	14.9	160	180	2
5	----	----	14.5	162	185	2
10	0.9080	15.0	13.4	162	185	2
15	----	----	13.5	160	185	2
20	0.8961	11.1	13.5	160	190	2
25	----	----	13.5	163	195	2
30	0.8809	7.0	13.4	168	200	2
55	0.8690	4.0	13.4	180	205	2
Run No. 8						
0	.92711	21	23.7	160	200	2
5	---	--	22.5	172	205	2
10	---	--	22.0	176	205	2
15	---	--	20.1	177	205	2
20	---	--	20.7	182	205	2
25	---	--	21.0	186	205	2
30	---	--	21.3	190	205	2
35	---	--	21.3	190	205	2
40	---	--	21.3	192	205	2
45	---	--	21.5	194	205	2
50	---	--	21.5	194	205	2
55	---	--	21.5	195	205	2
60	.8750	5.6	21.5	195	205	2

TABLE I

Run No. 9

Time	Feed		Still		Steam	
Mins.	Density	Wt. % comp.	Press ins. Hg	Temp. °F	Temp. °F	Flometer
0	0.9376	24.2	21.75	170	200	4
5	----	----	22.05	168	205	4
10	----	----	22.05	166	208	4
20	----	----	22.15	168	205	4
35	----	----	22.15	176	208	4
40	----	----	22.15	180	212	4
45	0.8677	3.7	22.05	180	215	4
50	----	----	21.65	180	215	4
55	----	----	22.15	182	215	4
60	0.8639	2.8	22.15	182	215	4

Run No. 10

0	0.9489	27.7	22.3	160	200	2
5	----	----	22.3	165	200	2
10	0.9026	13.3	22.3	165	205	2
15	0.8853	8.0	22.3	167	205	2
20	----	----	22.3	169	200	2
25	0.8797	6.7	22.3	170	205	2
30	----	----	22.3	171	205	2
35	0.8735	5.1	22.3	172	205	2
40	----	----	22.0	173	205	2
45	0.8685	3.9	22.0	174	205	2
50	----	----	22.0	174	205	2
55	0.8648	3.0	22.5	175	200	2

TABLE II - BATCH STEAM STRIPPING DATA

Run No.	1	2	3	4	5	6	
Feed wt. lbs.	25.125	25	25.25	19.62	24	25	2
Feed lb. mols.	0.04361	0.04374	0.04301	0.03642	0.04929	0.04350	0
Feed comp. vol. %	14.97	14.49	12.835	10.53	3.21	14.81	1
Residue comp. vol. %	1.49	1.28	1.55	1.49	1.49	1.6	1
X_f	0.8181	0.79188	0.70129	0.57568	0.2396	0.80941	0
X_r	0.08181	0.07012	0.08474	0.08181	0.08181	0.04350	0
Still press. mm. Hg	742	742	744	742	742	564.5	5
Still temp. °C	96	96	96	94	96	77	7
Efficiency %	80	80	80	80	80	80	8
$\int_{X_r}^{X_f} \frac{\pi - E_p^*(T,X)}{E_p^*(T,X)} dx$	3.453	4.220	3.034	2.624	1.638	2.849	2
Steam calculated lbs.	2.70	2.75	2.35	1.73	1.45	2.231	2
Steam observed lbs.	2.62	2.75	2.31	1.70	1.50	2.34	2
Difference %	2.9	0	1.7	1.73	3.4	4.9	2
Steam calculated assuming Raoult's law	1.124	1.345	1.60	0.903	0.431	1.575	1
% difference	57	60	30	47	71	32	2
Lbs. steam /ton CCl_4 observed	1038	1114	1106	1212	2440	974	9

NG DATA

4	5	6	7	8	9	10
19.62	24	25	25.93	19.687	25	25.125
0.03642	0.04929	0.04350	0.0464	0.036155	0.04340	0.04393
10.53	3.21	14.81	12.9	11.2	14.97	14.492
1.49	1.49	1.6	1.49	3	2.13	1.978
0.57568	0.2396	0.80941	0.70714	0.61365	6.81814	0.79188
0.08181	0.08181	0.04350	0.08181	0.02108	0.11688	0.10811
742	742	564.5	558	548.6	274	195.6
94	96	77	79	85	73	80
80	80	80	80	80	80	80
2.624	1.638	2.849	2.499	1.690	1.131	0.5065
1.73	1.45	2.231	2.08	1.09	0.884	0.400
1.70	1.50	2.34	2.02	1	0.812	0.437
1.73	3.4	4.9	2.8	8.2	8.1	9.2
0.903	0.431	1.575	1.543	1.206	0.492	0.393
47	71	32	23	20.6	65	10
1212	2440	974	920	761	236	182

TABLE III

Run No. 1

Time Mins.	Feed Weight Lb.	Oz.	Flo-meter	Rate lb./hr.	Still Temp. °C Top	Bot.	Press. Mm.Hg	Steam Press. Cm.	Flometer Cm.	Resi Density
0	32	0	3.1	0	112	100	729	1.4	0.7	0.8783
10	30	9	3.1	8.62	108	100	729	1.5	0.7	----
15	--	--	3.2	---	110	104	729	1.5	0.7	0.8590
20	29	3	3.2	8.25	110	101	729	1.5	0.7	----
30	27	13	3.2	8.25	110	101	729	1.5	0.7	0.8590
40	26	7	3.2	8.25	110	103	729	1.5	0.7	----
45	--	--	3.2	---	110	103	729	1.5	0.7	0.8590
50	25	1	3.4	8.25	110	103	729	1.5	0.7	----
64	23	3	3.6	8.03	109	103	729	1.5	0.7	0.8590

Run No. 2

0	32	0	7.2	0	97	98	737	7.1	3.0	
10	31	4	7.6	4.5	97	98	737	7.6	3.0	
15	--	--	6.2	---	97	98	737	7.6	3.0	
20	30	5	6.2	5.5	97	98	737	7.6	3.3	
30	29	6	6.2	5.4	97	98	737	7.6	3.3	
40	28	8	6.3	5.25	97	98	737	7.6	3.3	
45	--	--	6.2	---	97	98	737	7.6	3.3	
50	27	9	6.2	5.4	97	98	737	7.6	3.3	
60	26	10	6.3	5.4	97	98	737	7.6	3.3	
70	25	11	6.3	5.4	97	98	737	7.6	3.3	

Run No. 3

0	32	3	3.2	6	112	101	735	1.4	0.7	0.8846
10	30	10	3.2	9.37	109	105	735	1.4	0.7	----
15	--	--	3.2	---	109	102	735	1.4	0.7	0.8637
20	29	1	3.2	9.37	110	102	735	1.4	0.7	----
25	28	6	3.2	9.20	110	102	735	1.4	0.7	0.8637
40	26	1	3.2	9.25	110	102	735	1.4	0.7	----
45	--	--	3.2	---	110	102	735	1.4	0.7	0.8648
50	24	7	3.2	9.75	110	102	735	1.4	0.7	----
60	22	15	3.2	9.0	110	98	735	1.4	0.7	0.8632

Run No. 4

0	30	0	3.1	0	110	100	747	0.8	0.5	
10	28	8	3.2	9.0	102	100	747	0.8	0.5	
15	--	--	3.2	--	103	104	747	0.8	0.5	0.8790
20	27	2	3.2	8.2	104	104	747	0.8	0.5	
30	25	11	3.3	8.6	104	104	747	0.8	0.5	0.8762
40	24	4	3.4	8.6	104	104	747	0.8	0.5	
45	--	--	3.4	---	104	104	747	0.8	0.5	0.8672
50	22	14	3.4	8.2	105	104	747	0.8	0.5	
60	--	--	3.4	---	--	104	747	0.8	0.5	

TABLE III

o-ter	Rate lb./hr.	Still Temp.°C Top	Still Temp.°C Bot.	Still Press. Mm.Hg	Steam Press. Cm.	Steam Flometer Cm.	Residue Density	Residue Vol. % comp.
1	0	112	100	729	1.4	0.7	0.8783	3.6
1	8.62	108	100	729	1.5	0.7	0.8590	1
2	---	110	104	729	1.5	0.7	0.8590	1
2	8.25	110	101	729	1.5	0.7	0.8590	1
2	8.25	110	103	729	1.5	0.7	0.8590	1
2	---	110	103	729	1.5	0.7	0.8590	1
4	8.25	110	103	729	1.5	0.7		
6	8.03	109	103	729	1.5	0.7		
2	0	97	98	737	7.1	3.0		
6	4.5	97	98	737	7.6	3.0		
2	---	97	98	737	7.6	3.0		
2	5.5	97	98	737	7.6	3.3		
2	5.4	97	98	737	7.6	3.3		
3	5.25	97	98	737	7.6	3.3		
2	---	97	98	737	7.6	3.3		
2	5.4	97	98	737	7.6	3.3		
3	5.4	97	98	737	7.6	3.3		
3	5.4	97	98	737	7.6	3.3		
2	0	112	101	735	1.4	0.7	0.8846	4.4
2	9.37	109	105	735	1.4	0.7	0.8637	1.5
2	---	109	102	735	1.4	0.7	0.8637	1.5
2	9.37	110	102	735	1.4	0.7	0.8648	1.9
2	9.20	110	102	735	1.4	0.7	0.8632	1.5
2	9.25	110	102	735	1.4	0.7		
2	---	110	102	735	1.4	0.7		
2	9.75	110	102	735	1.4	0.7		
2	9.0	110	98	735	1.4	0.7		
1	0	110	100	747	0.8	0.5		
2	9.0	102	100	747	0.8	0.5		
2	---	103	104	747	0.8	0.5	0.8790	
2	8.2	104	104	747	0.8	0.5		
3	8.6	104	104	747	0.8	0.5	0.8762	
4	8.6	104	104	747	0.8	0.5		
4	---	104	104	747	0.8	0.5	0.8672	
4	8.2	105	104	747	0.8	0.5		
4	---	--	104	747	0.8	0.5		

TABLE III

Run No. 5

Feed					Still			Steam	
Time Mins.	Weight Lb.	Oz.	Flo-meter	Rate lb./hr.	Temp. °C Top	Bot.	Press. Mm.Hg	Press. Cm.	Flometer Cm.
0	30	9	4.3		113	100	752	1.1	0.7
10	29	5	4.4		111	102	752	1.3	0.7
15	--	--	4.4		111	102	752	1.3	0.7
20	28	2	4.4		110	102	752	1.2	0.7
30	27	0	4.4		109	102	752	1.2	0.7
40	25	13	4.4		110	103	752	1.2	0.7
45	--	--	4.4		110	103	752	1.2	0.7
50	24	10	4.4		110	103	752	1.2	0.7
60	23	6	4.4		111	103	752	1.2	0.7

Run No. 6

0	28	14	3.1		103	100	747	1.5	0.3
10	27	3	3.1		103	102	747	2	1.1
15	--	--	3.1		103	101	747	2	1.1
20	25	7	3.1		102	101	747	2.4	1.3
30	23	13	3.1		102	104	747	1.9	1.0
40	22	3	3.1		103	105	747	2.0	1.0
45	--	--	3.3		102	104	747	2.1	1.0
50	20	9	3.4		101	105	747	2.0	1.1

Run No. 7

0	19	6	1.6		110	105	737	2.1	0.6
10	18	2	1.5		100	106	737	2.1	0.6
15	--	--	1.45		100	105	737	2.1	0.6
20	16	14	1.4		100	105	737	2.1	0.6
30	15	10	1.4		100	105	737	2.1	0.6
40	14	7	1.4		101	105	737	2.1	0.6
45	13	13	1.4		101	105	737	2.1	0.6

Run No. 8

0	25	2	2.2	0	98	98	740	7.7	4.3
10	23	10	2.2	9.0	99	99	740	8.8	4.3
15	--	--	2.2	---	94	98	740	8.8	4.3
20	22	3	2.2	8.63	95	98	740	8.8	4.3
30	--	--	2.2	---	94	98	740	8.8	4.3
33	20	6	2.2	8.36	94	98	740	8.8	4.3
40	19	6	2.2	8.57	95	98	740	8.8	4.3
45	--	--	2.3	---	95	98	740	8.8	4.3
50	17	15	2.3	8.63	95	98	740	8.8	4.3
58.5	16	13	2.3	7.94	95	98	740	8.8	4.3

TABLE III

Flo-meter	Rate lb./hr.	Still Temp. °C Top	Still Temp. °C Bot.	Still Press. Mm.Hg	Steam Press. Cm.	Steam Flometer Cm.	Residue Density	Residue Vol. % comp.
4.3		113	100	752	1.1	0.7	0.8919	
4.4		111	102	752	1.3	0.7	----	
4.4		111	102	752	1.3	0.7	0.8611	1.3
4.4		110	102	752	1.2	0.7	----	
4.4		109	102	752	1.2	0.7	0.8611	1.3
4.4		110	103	752	1.2	0.7	----	
4.4		110	103	752	1.2	0.7	0.8611	1.3
4.4		110	103	752	1.2	0.7	----	
4.4		111	103	752	1.2	0.7	0.8611	1.3
5.1		103	100	747	1.5	0.3	0.9000	6.6
5.1		103	102	747	2	1.1	----	
5.1		103	101	747	2	1.1	0.8637	1.7
5.1		102	101	747	2.4	1.3	----	
5.1		102	104	747	1.9	1.0	0.8650	1.8
5.1		103	105	747	2.0	1.0	----	
5.3		102	104	747	2.1	1.0	0.8600	1.2
5.4		101	105	747	2.0	1.1	0.8687	2.2
.6		110	105	737	2.1	0.6	0.8818	7.30
.5		100	106	737	2.1	0.6	----	
.45		100	105	737	2.1	0.6	0.8580	0.9
.4		100	105	737	2.1	0.6	----	
.4		100	105	737	2.1	0.6	0.8615	1.3
.4		101	105	737	2.1	0.6	----	
.4		101	105	737	2.1	0.6	0.8605	1.15
.2	0	98	98	740	7.7	4.3	0.9094	8
.2	9.0	99	99	740	8.8	4.3	----	
.2	---	94	98	740	8.8	4.3	0.8516	0
.2	8.63	95	98	740	8.8	4.3	----	
.2	---	94	98	740	8.8	4.3	0.8530	0.2
.2	8.36	94	98	740	8.8	4.3	----	
.2	8.57	95	98	740	8.8	4.3	----	
.3	---	95	98	740	8.8	4.3	0.8507	0
.3	8.63	95	98	740	8.8	4.3	----	
.3	7.94	95	98	740	8.8	4.3	0.8526	0.25

TABLE III

Run No. 9

Time Mins.	Feed Weight Lb.	Feed Weight Oz.	Feed Flo-meter	Feed Rate lb./hr.	Still Temp. °C Top	Still Temp. °C Bot.	Still Press. Mm.Hg	Steam Press. Cm.	Steam Flometer Cm.
0	34	13	3.6		113	96	731	3.2	1.5
10	33	0	3.6		112	96	731	3.4	1.5
15	--	--	3.6		108	98	731	3.4	1.5
20	31	3	3.6		110	98	731	3.4	1.5
30	29	12	3.6		111	101	731	3.4	1.5
40	27	14	3.6		110	101	731	4.0	1.8
45	--	--	3.6		109	101	731	3.4	1.7
50	26	0	3.6		109	100	731	3.4	1.8
60									

Run No. 10

0	26	12	19.6		93	99	744.5	7.1	3.0
11	24	12	19.6	10.90	93	98	744.5	7.1	3.0
15	--	--	19.6	---	93	99	744.5	7.1	3.0
20	23	6	19.6	9.16	93	99	744.5	7.1	3.0
32	21	8	19.6	9.37	97	99	744.5	7.1	3.0
40	20	5	19.6	8.90	97	98	744.5	7.1	3.0
45	--	--	19.6	---	92	99	744.5	7.1	3.0
50	18	10	19.6	10.12	97	98	744.5	7.1	3.0
60	17	3	19.6	8.62	92	99	744.5	7.1	3.0

Run No. 11

0	25	8	3.3	0	100	98	737	4.7	2.4
11	23	10	3.3	10.22	100	98	737	4.7	2.4
15	--	--	3.4	---	100	98	737	4.7	2.4
20	22	3	3.4	8.62	100	98	737	4.7	2.4
30	20	10	3.3	9.37	100	98	737	4.7	2.4
40	18	5	3.3	10.12	100	98	737	4.7	2.4
45	--	--	3.3	---	100	100	737	4.7	2.4
50	17	4	3.3	10.12	100	101	737	4.7	2.4
60	15	10	3.3	9.72	100	100	737	4.7	2.4

Run No. 12

0	34	5			88	100	754	6.8	3.2
5	--	--			88	100	754	6.8	3.2
12	32	8			86	100	754	6.8	3.2
15	--	--			87	100	754	6.8	3.2
22	32	0			97	101	754	6.8	3.2
33	31	8			105	101	754	6.8	3.2
39	--	--			100	100	754	6.8	3.2
45	31	2			85	100	754	6.8	3.2
56	--	--			89	100	754	6.8	3.2
63	30	2			88	100	754	6.8	3.2
77	29	9			88	100	754	6.8	3.2

TABLE III

lo-eter	Rate lb./hr.	Still Temp. °C Top	Still Temp. °C Bot.	Still Press. Mm.Hg	Steam Press. Cm.	Steam Flometer Cm.	Residue Density	Residue Vol. % comp.
.6		113	96	731	3.2	1.5	0.9140	
.6		112	96	731	3.4	1.5	----	
.6		108	98	731	3.4	1.5	0.8637	1.6
.6		110	98	731	3.4	1.5	----	
.6		111	101	731	3.4	1.5	0.8633	1.5
.6		110	101	731	4.0	1.8	----	
.6		109	101	731	3.4	1.7	0.8635	1.5
.6		109	100	731	3.4	1.8	----	
9.6		93	99	744.5	7.1	3.0	0.9238	9.8
9.6	10.90	93	98	744.5	7.1	3.0	0.8578	0.84
9.6	---	93	99	744.5	7.1	3.0		
9.6	9.16	93	99	744.5	7.1	3.0	0.8560	0.83
9.6	9.37	97	99	744.5	7.1	3.0	0.8526	0.1
9.6	8.90	97	98	744.5	7.1	3.0		
9.6	---	92	99	744.5	7.1	3.0	0.8526	0.1
9.6	10.12	97	98	744.5	7.1	3.0		
9.6	8.62	92	99	744.5	7.1	3.0	0.8519	0
.3	0	100	98	737	4.7	2.4	0.9490	13.25
.3	10.22	100	98	737	4.7	2.4	----	
.4	---	100	98	737	4.7	2.4	0.8610	1.3
.4	8.62	100	98	737	4.7	2.4	----	
.3	9.37	100	98	737	4.7	2.4	0.8610	1.3
.3	10.12	100	98	737	4.7	2.4	----	
.3	---	100	100	737	4.7	2.4	0.8610	1.3
.3	10.12	100	101	737	4.7	2.4	----	
.3	9.72	100	100	737	4.7	2.4	0.8610	1.3
		88	100	754	6.8	3.2	0.9553	14.20
		88	100	754	6.8	3.2		
		86	100	754	6.8	3.2		
		87	100	754	6.8	3.2	0.8936	5.7
		97	101	754	6.8	3.2		
		105	101	754	6.8	3.2		
		100	100	754	6.8	3.2	0.8526	0.2
		85	100	754	6.8	3.2		
		89	100	754	6.8	3.2	0.9264	10.1
		88	100	754	6.8	3.2		
		88	100	754	6.8	3.2	0.9125	8.3

TABLE III

Run No. 13

Feed					Still			Steam		R
Time	Weight		Flo-	Rate	Temp. °C		Press.	Press.	Flometer	Dens
Mins.	Lb.	Oz.	meter	lb./hr.	Top	Bot.	Mm.Hg	Cm.	Cm.	
0	35	10		--	130	92	741	12.7	11.2	
17	33	1		15.0	--	--	741	12.7	11.2	0.88
32	31	4		7.2	126	94	741	13.5	11.2	---
53	28	8		7.8	121	93	741	13.5	11.2	0.89
80	25	0		7.7	123	96	741	13.5	11.2	0.86
105	21	13		7.6	125	96	741	13.5	11.2	0.89
Run No. 14										
0	27	15	7.05		94	98	737	8.6	3.5	0.90
10	26	11	7.05		94	98	737	8.6	3.5	---
15	--	--	7.05		94	98	737	8.6	3.5	0.8
20	25	7	7.05		94	98	737	8.6	3.5	---
25	--	--	7.05		94	98	737	8.6	3.5	---
30	24	3	7.05		94	98	737	8.6	3.5	0.8
40	22	15	7.05		94	98	737	8.2	3.4	--
45	--	--	7.05		94	98	737	8.2	3.4	0.8
50	21	11	7.05		94	98	737	8.2	3.4	---
59.5	20	8.5	7.05		94	98	737	8.2	3.4	0.8

TABLE III

i Flo- meter	Rate lb./hr.	Still Temp. °C Top	Still Temp. °C Bot.	Still Press. Mm.Hg	Steam Press. Cm.	Steam Flometer Cm.	Residue Density	Residue Vol. % comp.
	--	130	92	741	12.7	11.2		
	15.0	--	--	741	12.7	11.2	0.8884	5.03
	7.2	126	94	741	13.5	11.2	----	--
	7.8	121	93	741	13.5	11.2	0.8900	5.02
	7.7	123	96	741	13.5	11.2	0.8682	2.5
	7.6	125	96	741	13.5	11.2	0.8900	5.02
7.05		94	98	737	8.6	3.5	0.9684	15.94
7.05		94	98	737	8.6	3.5	----	
7.05		94	98	737	8.6	3.5	0.8530	0.2
7.05		94	98	737	8.6	3.5	----	
7.05		94	98	737	8.6	3.5	----	
7.05		94	98	737	8.6	3.5	0.8537	0.3
7.05		94	98	737	8.2	3.4	----	
7.05		94	98	737	8.2	3.4	0.8536	0.3
7.05		94	98	737	8.2	3.4	----	
7.05		94	98	737	8.2	3.4	0.8534	0.2

COUNTER-CURRENT STRIPPING TABLE IV

Run No.	1	2	3	4	5	6	7
Feed rate lbs./hr.	8.262	5.41	9.25	8.55	7.18	9.97	7.4
Feed lb. mols.	0.01834	0.01306	0.01897	0.0144	0.01448	0.01644	0.0
Feed comp. vol. %	3.62	3.95	4.45	5.3	5.49	6.6	7.3
Residue comp. vol. %	0.94	0	2.69	3.22	1.17	1.5	4
X_f	0.19776	0.21582	0.2431	0.2895	0.29991	0.3605	0.3
X_r	0.0528	0	0.1474	0.1761	0.06433	0.08456	0.2
Still press mm. Hg	729	740	735	747	752	747	737
Still temp. °C	110	98	110	104	110	102	100
Efficiency %	90	90	90	90	90	90	90
$p^*(T,X_f)$	156	150	165	178	186	184	187
Steam observed grams	88	100	58	52	95	122	52
Steam calculated grams	91.1	103	58	49	97.4	130.3	54
% difference	3.2	2.9	0	6.1	2.4	6	3.7
Lbs. steam / ton CCl_4	946	943	515	882	795	766	769

7	8	9	10	11	12	13	14
7.41	8.525	8.81	9.56	9.87	3.701	7.89	7.492
0.01087	0.01606	0.01689	0.01795	0.01758	0.00834	0.02395	0.01267
7.3	8	8.5	9.8	13.25	14.2	15.05	15.94
4	0	0	0	0	6.9	5.02	3.5
0.39886	0.43711	0.46435	0.53819	0.72384	0.7758	0.08223	0.87094
0.21875	0	0	0	0	0.3775	0.27427	0.19232
737	740	731	744	737	754	741	737
100	96	110	96	100	90	96	96
90	90	90	90	90	90	90	90
187	185	204	207	220	205	245	230
52	200	200	215	293	82.5	249	200
54	198	191	237	283	81.0	254	180
3.7	1	4.7	9	3.3	1.8	1.9	11
769	788	648	612	644	993		602

Table V for Graphical Integration

Runs No. 1-4

Vol. %	p*	Ep*	π - Ep*	$\frac{\pi - Ep^*}{Ep^*}$
1.0	68.2	56.5	687.1	12.594
1.5	76.0	60.8	680.9	11.199
2.0	85.0	68.0	673.7	9.907
2.5	96.0	76.8	664.9	8.657
3.0	113.5	90.8	650.9	7.168
3.5	131.5	105.2	636.5	6.050
4.0	141.5	113.2	628.5	5.552
5.0	167.4	133.9	607.8	4.539
6.0	173.0	138.4	603.3	4.359
7.0	178.8	143.0	598.7	4.186
8.0	184.3	147.4	594.3	4.032
9.0	190.0	152.0	589.7	3.879
10.0	195.8	156.6	585.1	3.736
12.0	207.0	165.6	576.1	3.478
14.0	218.4	174.7	567.0	3.245
15.0	224.1	179.3	562.4	3.136

Run No. 5

0.4	60.3	45.225	695.17	15.372
0.6	63.0	47.25	693.15	14.669
0.8	65.6	49.2	691.2	14.048
1.0	68.4	51.3	689.1	13.433
1.5	76.0	57.0	683.4	11.989
2.0	85.0	63.73	676.6	10.620
2.5	96.0	72.0	668.4	99.283
3.0	113.5	85.13	655.2	7.697
3.5	131.5	98.63	641.7	6.507

Table V cont'd.

Run No. 6

Vol. %	p*	Ep*	π - Ep*	$\frac{\pi - E\mu^*}{E\mu^*}$
1.0	61.0	48.8	146.8	3.008
1.5	69.0	55.2	140.4	2.543
2.0	78.0	62.4	133.2	2.134
2.5	81.0	64.8	130.8	2.018
3.0	98.4	78.7	116.8	1.484
3.5	110.0	88.0	107.6	1.223
4.0	121.0	96.8	98.8	1.020
5.0	146.2	116.9	78.7	0.673
6.0	154.0	123.2	72.4	0.587
7.0	159.2	127.3	68.3	0.536
8.0	16433	131.4	6422	0.488
9.0	169.5	135.6	60.0	0.442
10.0	174.8	139.8	55.8	0.399
11.0	180.0	144.0	51.6	0.358
13.0	190.5	152.4	43.2	0.283
15.0	201.0	160.8	34.8	0.216

Run No. 7

2.0	74.0	59.2	214.8	3.62
2.5	83.0	66.4	207.6	3.12
3.0	93.0	74.4	199.6	2.68
3.5	103.0	82.4	191.6	2.32
4.0	113.0	90.4	183.6	2.03
4.5	123.0	98.4	175.6	1.78
5.0	134.0	107.2	166.8	1.55
6.0	141.0	112.8	161.2	1.42
8.0	151.5	121.2	152.8	1.26
10.0	161.5	129.2	144.8	1.12
12.0	171.5	137.2	136.8	0.99
14.0	181.5	145.2	128.8	0.88
16.0	191.5	153.2	120.8	0.78

FIG. 1 ANALYSIS CHART FOR OIL CCl4 MIXTURE AT 20°C.

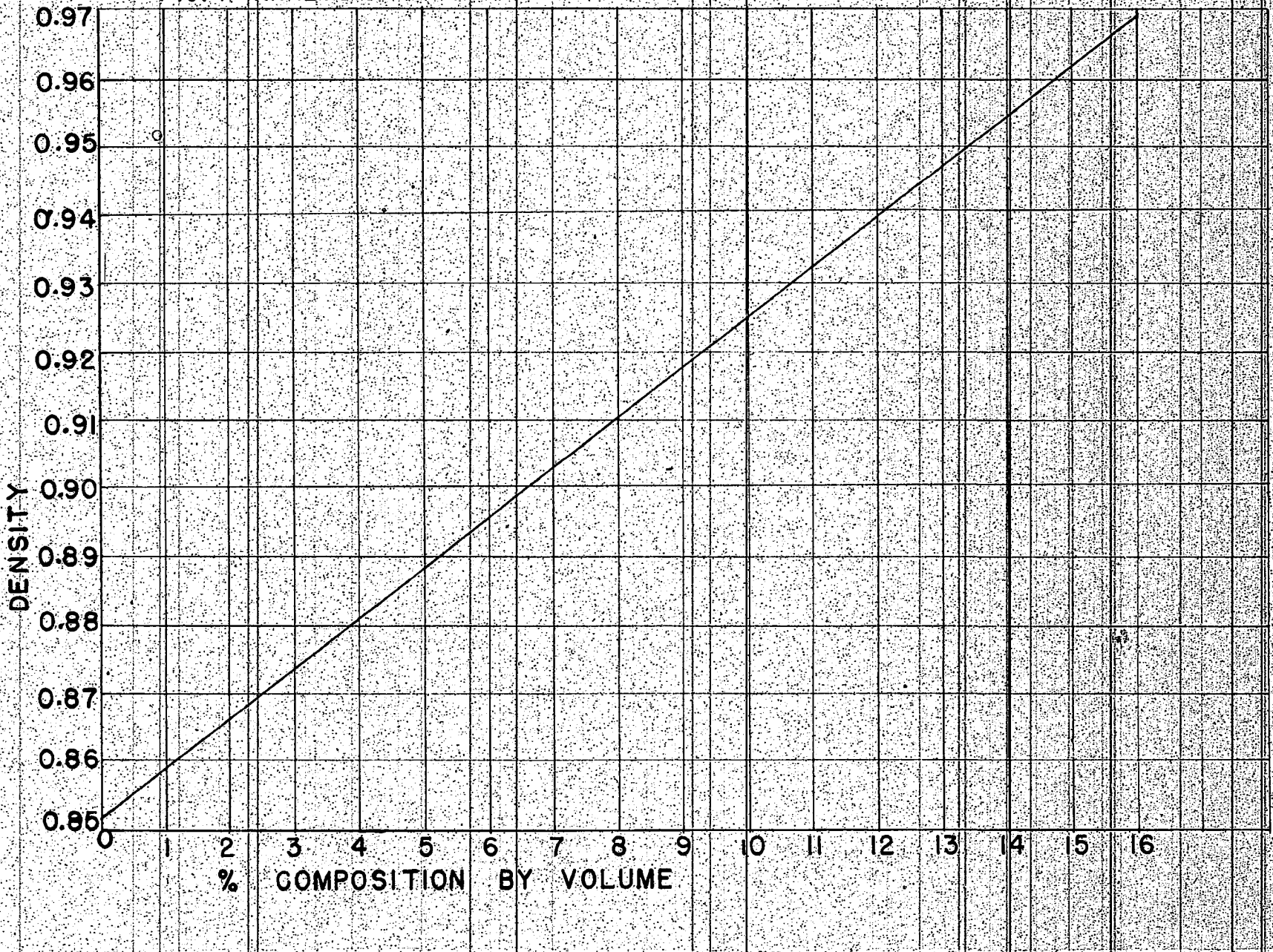

FIG. 2 VAPOR PRESSURE OF CCl4

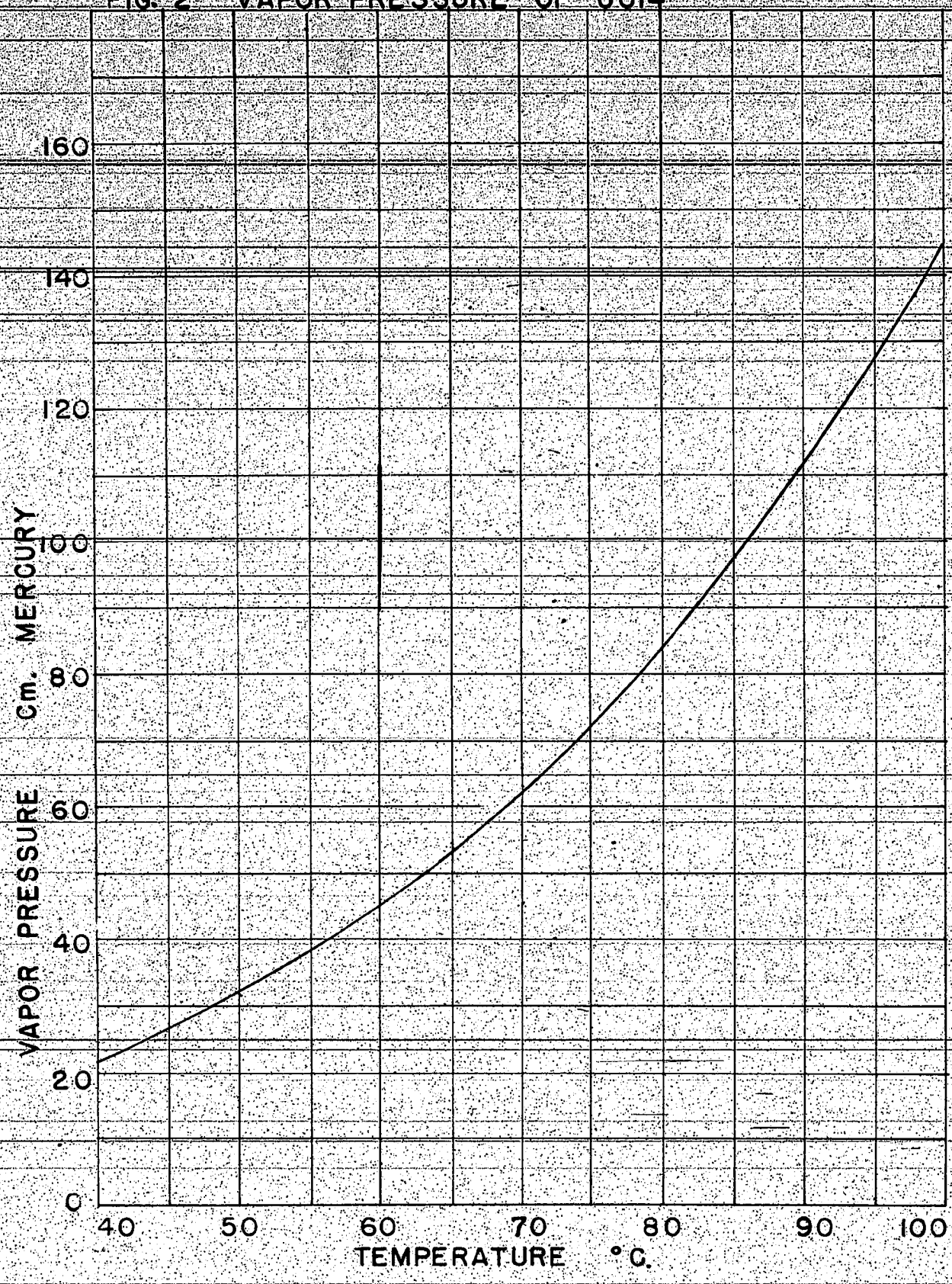

FIG. 3

VAPOR PRESSURE Vs. TEMPERATURE

OIL - CCl_4 MIXTURE

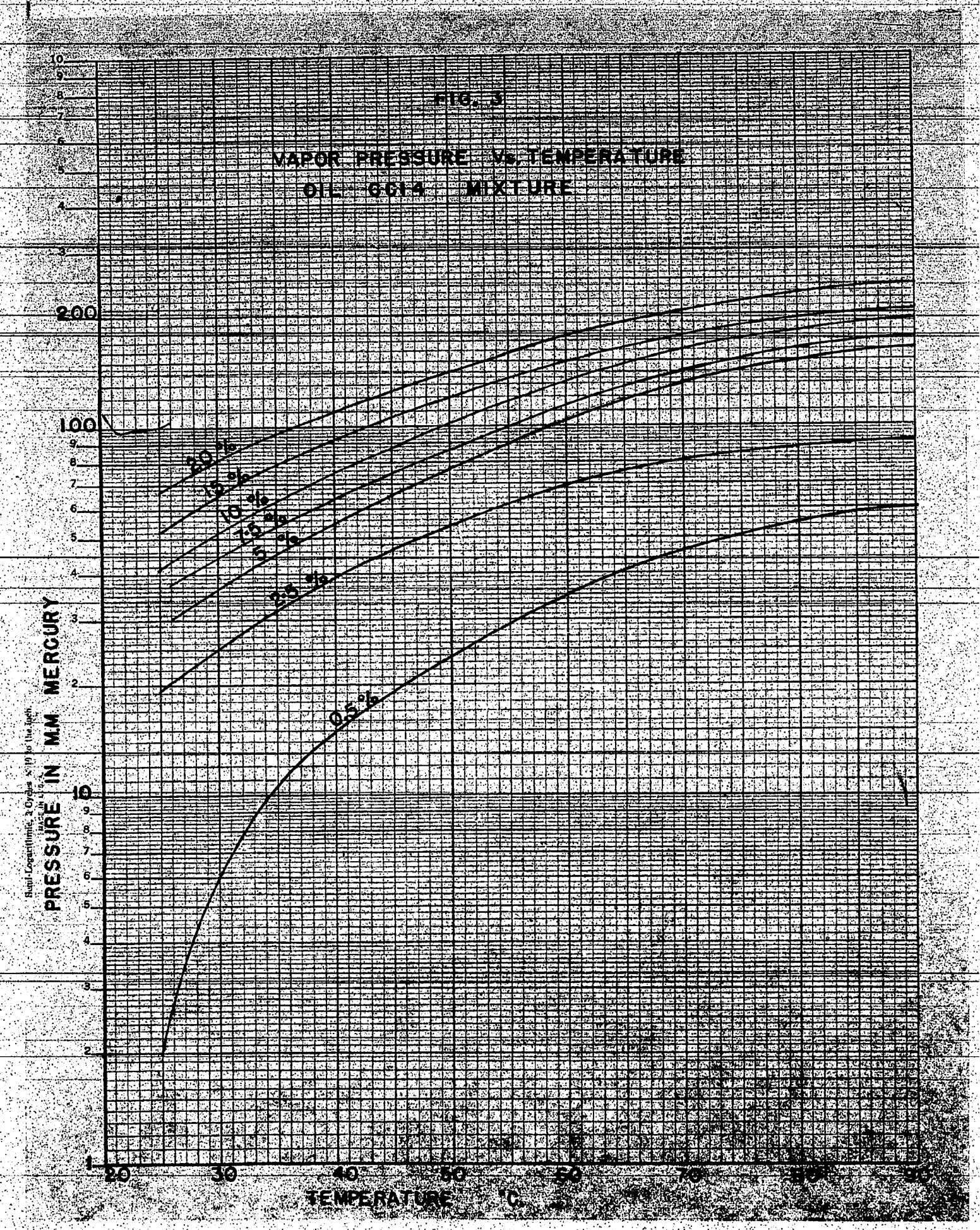

FIG. 4

VAPOR PRESSURE Vs. COMPOSITION

OIL CCl4 MIXTURE

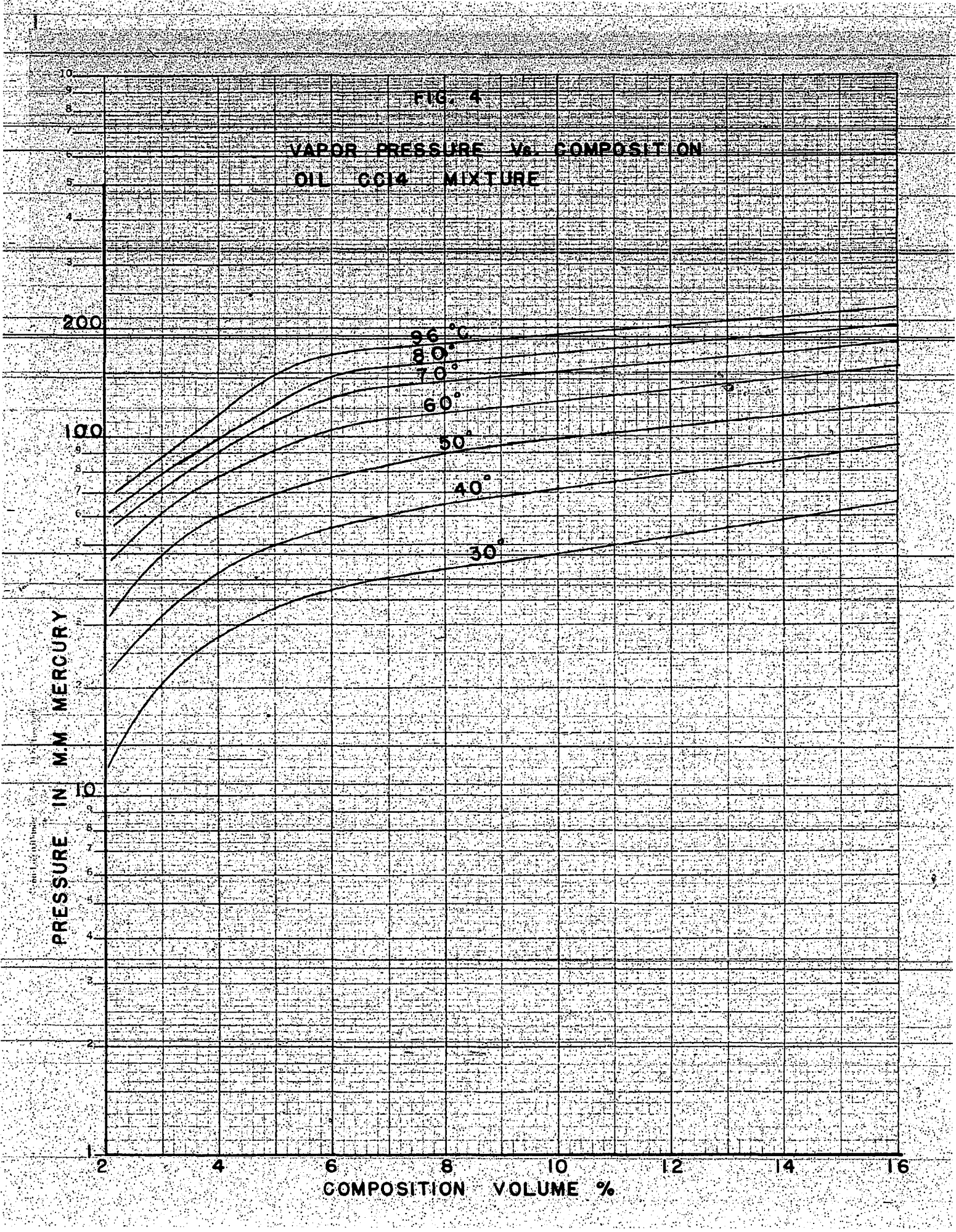

FIG. 5

GRAPHICAL INTEGRATION CURVES

REFER TABLE RUN NUMBERS 1, 6, 7, 8, 9

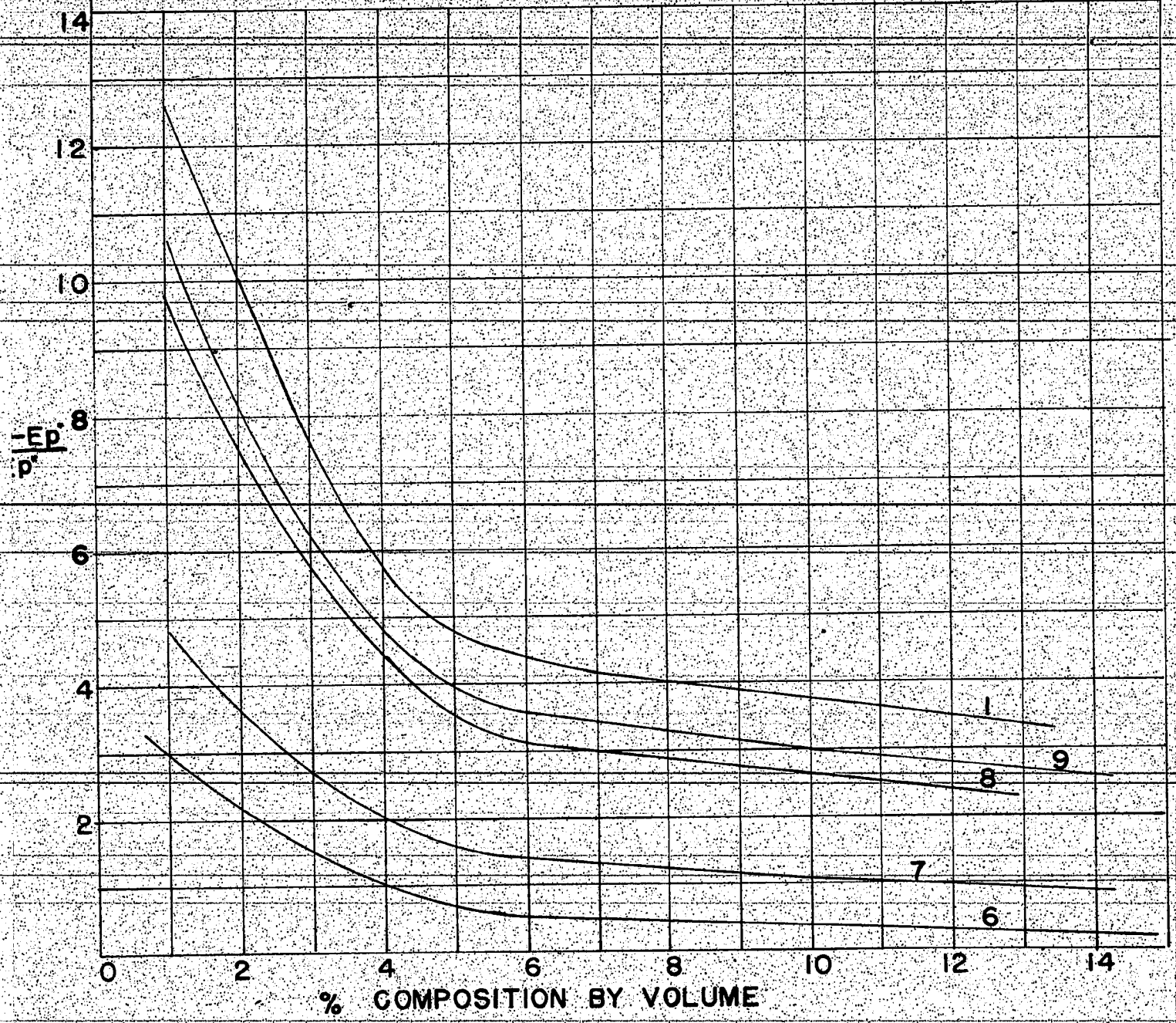

Figure 6

Relative Superiority Curves for Batch & Counter Current Processes

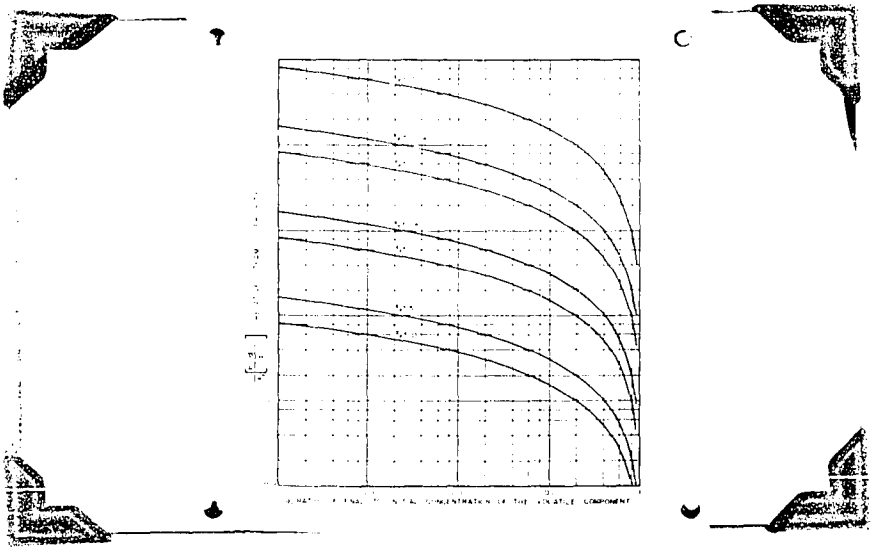

FIG. 7

APPARATUS FOR VAPOR PRESSURE DETERMINATION

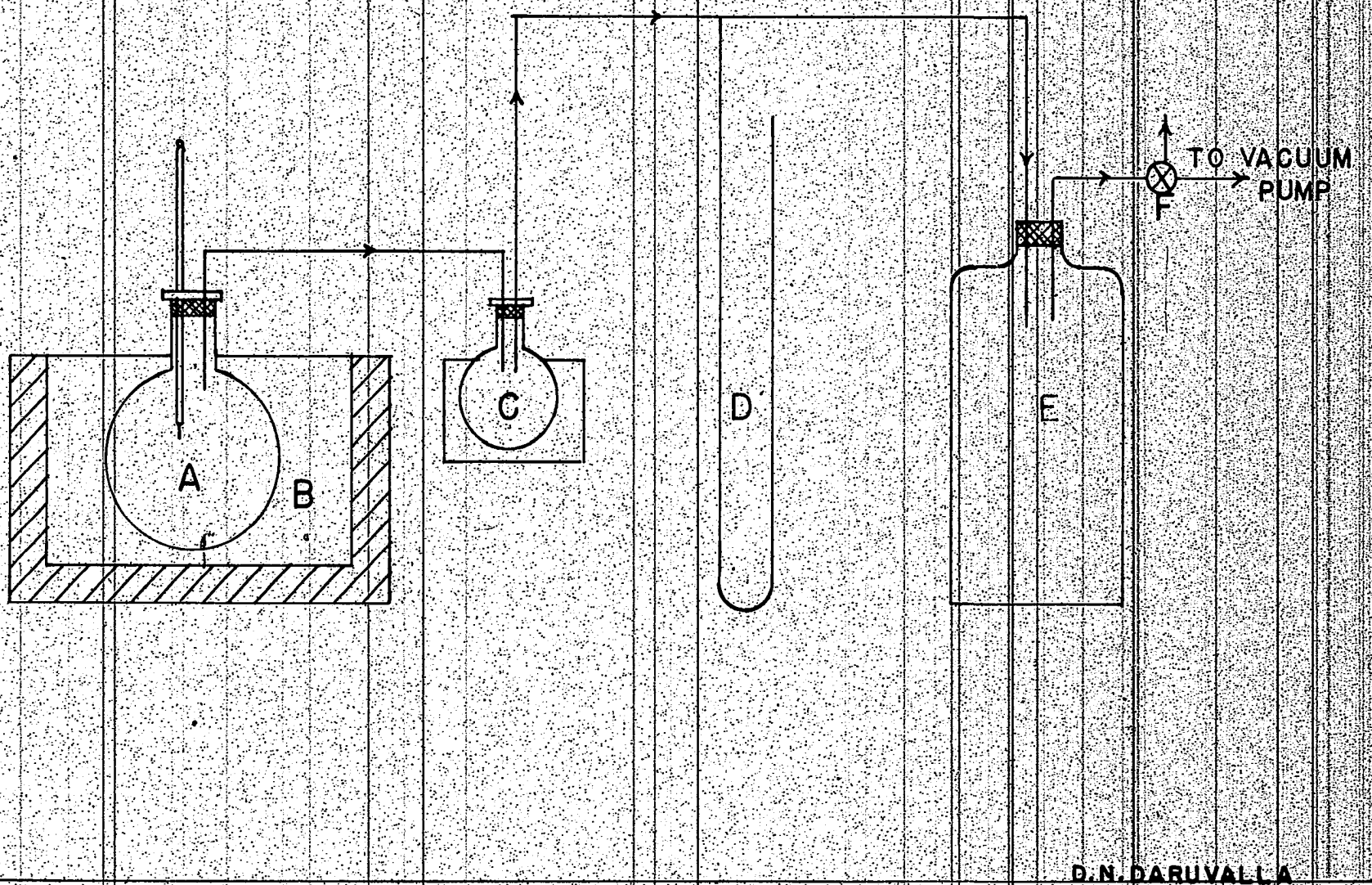

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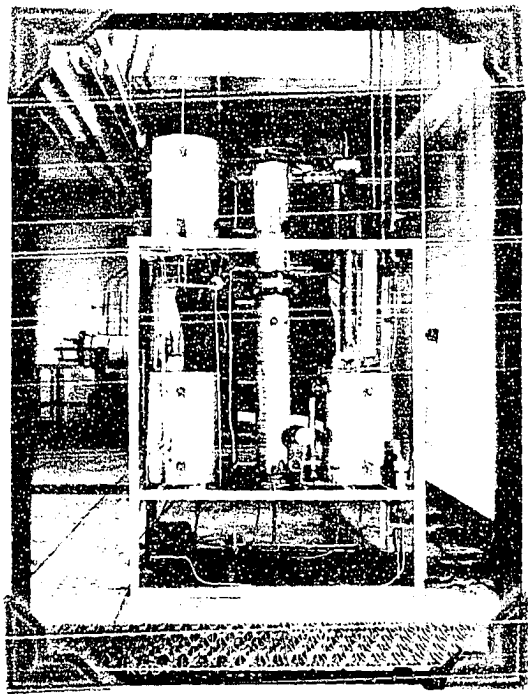
Figure 9

Pilot plant unit for batch steam stripping operation



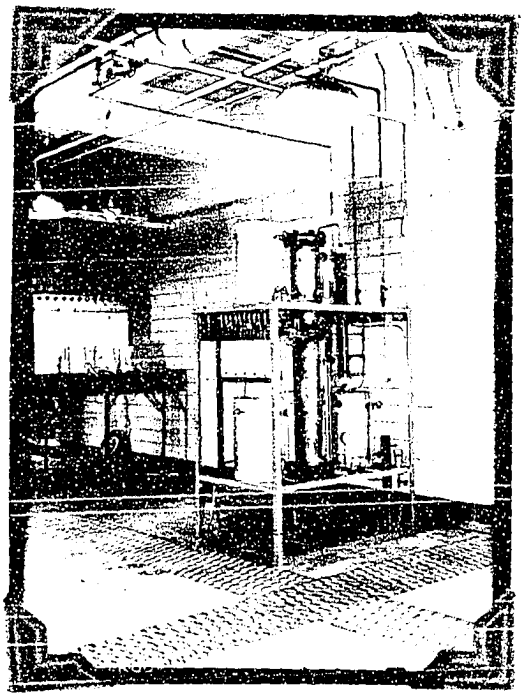
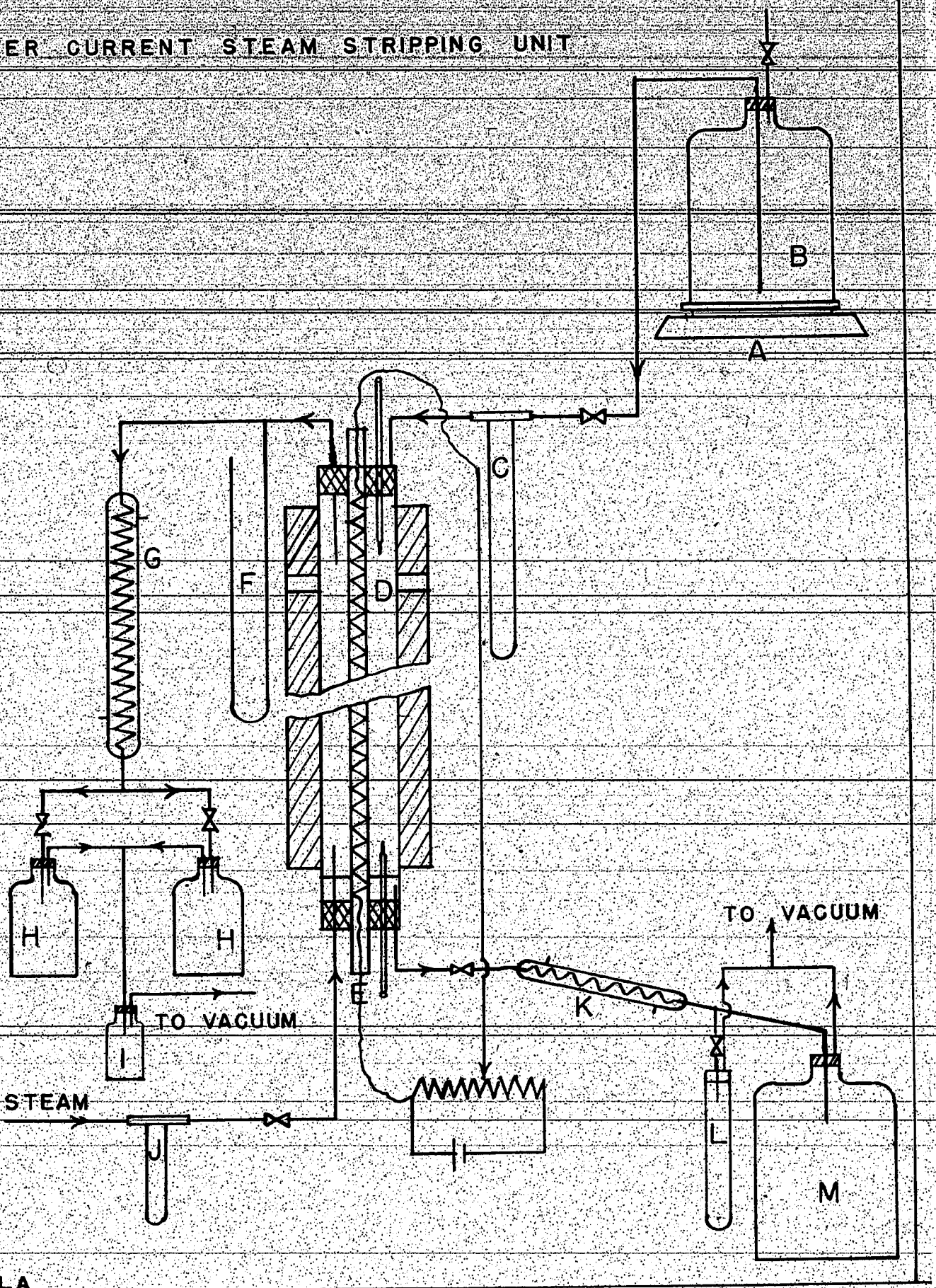


FIG. 10

COUNTER CURRENT STEAM STRIPPING UNIT



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