THE EFFECT OF LIQUID CONCENTRATION ON THE MASS TRANSFER COEFFICIENT IN THE LIQUID FILM

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ABSTRACT

The effect of liquid concentration on the resistance to mass transfer across the liquid film is studied for an acetone-methyl isobutyl ketone-air system in a packed tower. The liquid film resistance is computed by measuring the overall resistance for various concentrations and subtracting the gas film resistance. The gas film resistance is determined at the same liquid and gas rates by measurements on the pure components.

In the theory developed to handle the calculations, it is assumed that the interface temperatures are the wet-bulb temperatures based on a dynamic equilibrium. This supposition is supported experimentally at a liquid rate of 1200 lbs/hr.-ft^2 and a gas rate of 313 lbs/hr-ft^2 within experimental error. Humidity charts for air-acetone, air-MIK, and air-n butanol, at one atmosphere pressure, are included for reference.

Finally, the liquid film resistances are correlated by the equations below. The first is based on the liquid diffusivity data of Wilke[28] and the second on that of Scheibel[19].

(1) $1/k_L a = 0.376 - 0.0000102 (C_L^1)_f M_m (\mu/\rho D_L)$

(2) $1/k_L a = 0.376 - 0.0000152 (C_L^1)_f M_m (\mu/\rho D_L)$

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PREFACE

This thesis is based on original experimental data compiled to investigate the effect of liquid concentration on the mass transfer coefficient across the liquid film. The data are obtained for an acetone-methyl isobutyl ketone-air system in a packed tower. The problem was proposed and supported by Dr. E. G. Scheibel for the purpose of investigating his contention that such correlations should be based on the assumption of a dynamic equilibrium at the interface. Sincere appreciation is expressed by the authors for the guidance so generously rendered. Acknowledgement is also made to Professor G. C. Keeffe for his assistance in procuring and assembling the equipment.

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INTRODUCTION

One of the more fundamental unit operations in the chemical industries is that of mass transfer between fluid phases. A mass transfer operation can be defined as the migration of the molecules of a particular component from one homogeneous phase to another phase. Common examples of this type of system are partially immiscible liquid pairs and liquid-gas combinations. Numerous types of apparatus have been devised to effect the phenomena such as distillation, extraction, absorption, and desorption columns as ordinarily encountered in industry. A device common to most is some sort of mechanism to bring the phases into intimate contact. This usually takes one of two forms: discrete plates or stages (with or without mechanical mixing) or continuous packing.

A mathematical, idealized model, first proposed by Whitman[27] and known as "the two-film theory" has been devised to handle the design of such units. The theory postulates that there exist, adjacent to the interface of the phases, films which in effect provide the major resistance to the transfer of molecules. In the case of liquid-gas systems, a film would exist in the gas phase and a film would exist in the liquid phase, each providing its own resistance to transfer. At the interface itself, zero resistance is assumed. The rate of mass transfer is then the product of a driving force (concentration or pressure potential) and a conductance (reciprocal of the film resistance) the latter being known as a mass transfer coefficient. Since all such idealized models assume a condition of "steady state" (no net build-up or decrease of material or energy at any point in the system), the mathematics can be applied to either or both of the conductances in combination. When the latter technique is used, the conductance is known as an "overall coefficient"; when either of the two film driving forces are considered individually, the conductances so defined are known as "gas-film and liquid film coefficients", respectively. In the present work, only liquid-gas combinations in packed columns will be considered. Hence, no further reference will be made to liquid-liquid systems or to plate columns.

Numerous investigators have attempted generalized correlations of the individual coefficients. The most successful of these has been the Chilton-Colburn modification[2] of the Reynolds analogy between heat transfer and fluid friction. These workers provided the basis for a relation between heat transfer and diffusion in gas films that has found general utility in predicting k_G, the gas-film coefficient. Numerous other workers have studied the effect of flow conditions and the physical properties of the phases on the respective coefficients. It is the primary purpose of this thesis to study the effect of one of these variables, liquid concentration, on the liquid film coefficient. A secondary purpose is to critically examine the methods that have previously been used to compute a mass transfer coefficient (overall or individual film) from data. The fundamental approach used in the study is analogous to that commonly used in water-air systems where a dynamic equilibrium rather than a static equilibrium is assumed at the interface.

THEORY AND PREVIOUS WORK

By virtue of the two film assumption discussed in the introduction, and by analogy to heat transfer and electrical energy transfer rate equations (where potential is proportional to rate times resistance), the basic rate equations for mass transfer can be written as follows:

(1) $N_A = K_G\,(p_G - p_e) = K_L\,(C_e - C_L) = k_G(p_G - p_i) = k_L(C_i - C_L)$

For a wide class of systems where the quantity $(p_G - p_e)$ remains proportional to $(C_e - C_L)$ as pressure and concentration are varied, it can be stated that Henry's law applies; i.e.;

(2) $C = \mathcal{H} p$

Eliminating C_i, C_L and p_i from (1):

(3a) $K_G(p_G - p_e) = k_G\,(p_G - p_i) = \mathcal{H} k_L\,(p_i - p_e)$

(3b) $p_i = p_e + \dfrac{K_G(p_G - p_e)}{\mathcal{H}\,k_L}$

(3c) $k_G(p_G - p_e)\,\left(1 - \dfrac{K_G}{\mathcal{H}\,k_L}\right) = K_G\,(p_G - p_e)$

(3d) $\dfrac{1}{K_G} = \dfrac{1}{k_G} + \dfrac{1}{\mathcal{H}\,k_L}$ $\qquad \dfrac{1}{K_G a} = \dfrac{1}{k_G a} + \dfrac{1}{\mathcal{H}\,k_L a}$

Similarly, it can be shown that:

(4) $\dfrac{1}{K_L a} = \dfrac{1}{k_L a} + \dfrac{\mathcal{H}}{k_G a}$

Thus, it is evident that the overall resistances are the sums of the individual resistances consistent with the two-film theory; the constant, $\mathcal{K}$, merely effects a change in units. This is again analogous to series resistances in electrical energy and heat transfer problems.

The theoretical justification of (1), for turbulent flow, encounters the same difficulties as the analogous situation for heat transfer and fluid friction; the eddy motion under these conditions renders the flow so complex, that the equations of motion have yet to be written in rigorous fashion. A quasi-theoretical justification can be derived by assuming that the liquid head in the system lost due to friction, divided by the momentum of the stream, is equal to the ratio of the actual material transferred between phases to the material which would be transferred were the stream to come to equilibrium with the other phase. This is similar to the Reynolds analogy[16] between heat transfer and fluid friction which is well substantiated by data[14]. The assumption can be expressed analytically by:

(5) $$\frac{\pi r^2 dP}{\frac{W\mu}{g}} = \frac{dC_G}{C_G - C_i}$$

Combining (5) with a material balance:

(6) $$N_A = \frac{W\,dC_G}{\rho\,2\pi r\,dX}$$

and the Fanning equation for turbulent friction:

$$(7)\quad dP = \frac{f \rho u^2 dX}{gr}$$

yields

$$(8)\quad \frac{N_A}{C_G - C_1} = \frac{fu}{2}$$

For ideal gases where:

$$(9)\quad C = \frac{P}{RT}$$

there results:

$$(10)\quad \frac{N_A}{p_G - p_1} = k_G = \frac{fu}{2RT} = \frac{f \rho \mu}{2pM_m}$$

Equation (10) can be expected to apply under the same conditions as does the corresponding equation for heat transfer, i.e., for gas flow, under conditions where the friction factor represents the true skin friction and not a combination of skin friction and turbulent losses[21].

Colburn[7] modified the basic Reynolds assumption by treating the transfer through the laminar layer only as a process of true diffusion. This followed from work done by Prandtl[15] for the analogous heat flow situation. The resulting relation is:

$$(11)\quad k_G = \frac{f \rho \mu}{2p_f M_m} \phi_D$$

where:

(12) $\phi_D = 1 - 5.9\sqrt{f} + 5.9\sqrt{f}\left(\frac{\mu_f}{\rho D}\right)$

For most gases, the variation in ϕ_D with u is small, and since f is approximately proportional to $u^{-0.2}$ (for turbulent flow), the net result is:

(13) $k_G \cong Ku^{0.8}$

Later empirical work by Colburn[8] and by Chilton and Colburn[2] provided the basis for predicting k_G from values of h (the heat transfer coefficient) obtained in wetted wall columns. The relation which has found general utility is:

(14) $J_H = \frac{h}{C_p \mu \rho}\left(\frac{C_p \mu_f}{k_H}\right)^{2/3} = \frac{1}{2}f = J_D = \frac{k_G p_f M_m}{\mu \rho}\left(\frac{\mu_f}{\rho D}\right)^{2/3}$

(14a) $k_G = \frac{h}{C_p p_f M_m}\left(\frac{C_p \rho D}{k_H}\right)^{2/3}$ and since $C_p \mu / k_H \cong 0.74$

for all gases.

(14b) $k_G = \frac{h}{C_p P_f M_m}\left(\frac{0.74}{\mu/\rho D}\right)^{2/3}$

(14c) $\frac{h}{k^1} = C_p\left(\frac{\mu/\rho D}{0.74}\right)^{2/3}$ with $k^1 = k_G P_f M_m$

It is noted that a simple heat balance on a wet-bulb psychrometer assumes the form:

(15) $ha(T_G - T_{WB}) = k^1 a\,(H_{WB} - H_G)\lambda_{WB}$

Thus, by combination with (14c):

(16) $\left(\frac{H_{WB} - H_G}{T_G - T_{WB}}\right)\lambda_{WB} = C_p\left(\frac{\mu/\rho D}{0.74}\right)^{2/3}$

Predictions of T_{WB} by means of (16) is generally within engineering accuracy for a large number of air-liquid combinations[22].

The preceding has summarized the state of knowledge on k_G, the gas film coefficient. The liquid film coefficient, k_L, has also been studied extensively. Sherwood and Holloway[23] proposed a generalized empirical correlation of the form:

$$(17) \quad \frac{k_L a}{D} = \alpha \left(\frac{L}{\mu}\right)^{1-n} \left(\frac{\mu}{\rho D}\right)^{1-s}$$

with the constant, s, being approximately 0.5 and α and n being dependent on the packing used. For the systems studied (desorption of oxygen from water for various packings), n was found to vary between 0.16 and 0.46 making:

$$(18) \quad k_L a = K D^{0.5} L^{0.54-0.84}$$

Scheibel and Othmer[18], working with several ketone-water systems, however, report the following:

$$(19) \quad k_L a = KDL^{0.8}$$

Equation (19) was found to fit the data on the ketone-water systems studied as well as the oxygen-water system of Sherwood and Holloway and others.

In general, most workers have ignored the effect of liquid concentration on $k_L a$ primarily due to the fact that virtually all of the experiments have been performed in dilute solutions. Scheibel[18] noticed an effect of liquid concentration on k_La and suggested that it was due to the effect of concentration on liquid diffusivity. However, lacking reliable liquid diffusivity data, no substantiation was offered for the dilute solutions used (up to 2 mole %). Hobson and Thodos[10] used a log mean inert composition across the liquid film to correlate their data, but, again, very dilute solutions were used.

In a pioneering paper, Stutzman et.at.[25] undertook the study of the effect of liquid concentration on k_La for butanol-hexanol-air systems in a packed tower. Their technique was to first measure a quantity which they called k_Ga for the pure butanol-air system. Then, without disturbing the column in any way, (same liquid and air rate), they varied the concentration of the liquid and measured K_Ga. Use of Equation (4) then permitted calculation of k_La which was consequently correlated with liquid concentration. Their final equation assumed the form:

$$(20)\quad \frac{1}{k_La} = K\,(C_L^1)_f\; M_m\,\left(\frac{\mu}{\rho D_L}\right)$$

This can be rewritten (dimensionally) as:

$$(20a)\quad \frac{1}{k_La} = K\,(1-x)_f\;\mu/D_L$$

In computing the quantity which was called $k_G a$, Stutzman used the following relation:

(21) $$k_G a = \frac{G_o S}{M_1 P V} \left[\frac{1}{1 - \frac{p_{i_2} - p_{i_1}}{p_2}} \right] \ln \frac{p_{i_1}}{p_{i_2} - p_2}$$

This can be derived as follows:

From (1):

(22a) $$N_A \, a dV = k_G a \, (p_G - p_i) \, dV$$

From a material balance:

(22b) $$\frac{G_o S}{M_s} dH = N_A \, a dV = k_G a \, (p_G - p_i) dV$$

Assuming that the liquid temperature varies linearly with the amount evaporated:

(22c) $$dt \propto dH$$

and that over small ranges of temperature:

(22d) $$dt \propto dp$$

there results:

(22e) $$dH \propto dp$$

However, by definition:

(22f) $$H = \frac{\text{Wt. Solvent}}{\text{Wt. Inert}} = \frac{p}{P - p} \; \frac{M_s}{M_1}$$

(22g) $$dH = \left[\frac{-pdp-(P-p)dp}{(P-p)^2} \right] \left[\frac{M_s}{M_1} \right] = \frac{-pdp}{(P-p)^2} \; \frac{M_s}{M_1}$$

For relatively nonvolatile liquids, p can be neglected in comparison to P and hence:

(22h) $dH \cong \frac{-dp}{P}\left[\frac{M_a}{M_1}\right]$

Combining with (22b) and integrating:

(22i) $$k_G a = \frac{G_o S}{M_1 PV}\int\frac{dp}{p_i - p_G} = \frac{G_o S}{M_1 PV}\left\{\frac{(p_2 - p_1)}{([p_i - p_G]_{LM})}\right\}$$

$$= \frac{G_o S}{M_i PV}\left[\frac{p_2 - p_1}{\frac{(p_{i_2} - p_2) - (p_{i_1} - p_1)}{\ln \left(\frac{p_{i_2} - p_2}{p_{i_1} - p_1}\right.}}\right]$$

and since solvent-free air was run into the bottom of the column ($p_1 = 0$):

(22j) $$k_G a = \frac{G_o S}{M_1 PV}\left[\frac{p_2}{(p_{i_2} - p_2 - p_{i_1})}\right] \ln \frac{p_{i_2} - p_2}{p_{i_1}}$$

which upon rearranging yields (21).

In order to evaluate (21), it is evident that information on the temperature (and hence the partial pressure) at the interface is necessary. Stutzman assumed that, since a pure component was evaporating, the partial pressure at the interface was the same as the equilibrium pressure in the main liquid stream. This is tantamount to assuming that there is no resistance to diffusion or heat transfer across the liquid film when evaporating a pure solvent, or, stated differently, that the liquid film is non-existant and that:

(23) $K_G a = k_G a$

There is considerable theoretical evidence that the above assumption is not compatible with the facts. First there is the knowledge that the process of evaporation of a pure component can be expressed in terms of high energy molecules diffusing through low energy molecules and finally evaporating; this clearly is consistent with the two film theory if resistance to self diffusion is considered. Second, there is the analogy between heat transfer and mass transfer. It is hypothesized that if there is a heat transfer across the liquid film, (as evidenced by the fact that the liquid has changed in temperature) there should be an equivalent mass transfer. The only conditions under which there is zero heat transfer across the liquid film is when the column is operated as an adiabatic humidifier. Under these conditions, the liquid leaving the column is fed to the top of the column at the wet-bulb temperature of the gas and does not change in temperature in passing through the column. Many wetted-wall columns, operating in this manner, have been described in the literature. If the liquid entering the top of the column is at a higher temperature than the wet-bulb temperature, and if the column is long enough, then the liquid leaving the column will still be at the wet-bulb temperature. However, since the liquid in this case has given up heat for evaporation, there must be a temperature gradient across a liquid-film at the top of the column even though a pure liquid is used. Stated differently, the interface temperature at the top of the column must be lower than the main liquid stream temperature; however, this interface temperature cannot be lower than the wet-bulb temperature. In shorter columns, the liquid leaving the column would tend toward the wet-bulb temperature, but never reach it. The interface temperature at the bottom of the column would then be somewhere between the main liquid stream temperature and the wet-bulb temperature of the gas at the bottom of the column.

From the above discussion, it is clear that the transfer coefficients measured by Stutzman were "over-all coefficients" since the driving forces used were between liquid and gas main-body streams. Mathematically, this is stated by:

$$(24) \quad K_G a = \frac{G_0 S}{M_1 P V}\left[\frac{1}{1 - \dfrac{p_{e_2} - p_{e_1}}{p_2}}\right] \ln \frac{p_{e_1}}{p_{e_2} - p_2}$$

In the present work, the basic technique used by Stutzman is repeated for acetone-air and methyl isobutyl ketone (MIK)-air systems. However, in computing $k_G a$, an attempt is made to be more realistic about the interface temperatures (and partial pressures). The interface temperatures are assumed to be the wet-bulb temperatures and the partial pressures at the interface to be those existing at this temperature.

Although it is realized that the above assumption is not entirely rigorous, it is believed to give a better estimate of the true conductances than if the interface temperatures were taken to be those of the main body liquid. It is, in fact, shown experimentally that for one particular set of flow rates, the wet-bulb assumption is justified within the accuracy of the experiment.

In order to facilitate the calculations of the present work, a different form of Equation (21) was used:

$$(25) \quad k_G a = \frac{G_o}{Z\,M_s} \quad \frac{\Delta H}{(p_i - p_G)_{LM}}$$

with the p_i's being computed from the wet-bulb assumption. All other parameters were obtained experimentally.

Wet bulb temperatures were interpolated from humidity charts which were constructed for acetone-air; MIK-air and butanol-air by means of equation (16). The technique for this is given in the sample calculations. Having computed $k_G a$ by means of (25), $K_G a$ was next computed (from the same data) by a modified version of (24):

$$(26) \quad K_G a = \frac{G_o}{Z\,M_s} \quad \frac{\Delta H}{(p_e - p_G)_{LM}}$$

Liquid film coefficients, $k_L a$ were next computed, for the pure component, by means of (3d). The value of $\mathcal{H}$ (Henry's Constant) was computed for the mean liquid film conditions by:

$$(27)\quad \mathcal{H} = \frac{1}{P^{o} v_{m}}$$

Having established the liquid and gas film coefficients for the pure components, further data were taken, at a fixed liquid and gas rate, to determine the effect of liquid composition on $k_L a$. Since $k_G a$ is essentially a function of the state of motion of the gas relative to the liquid (gas rate, liquid rate, and packing) and of the properties of the gas film to a lesser extent (viscosity, density, and diffusivity), the gas film coefficient would not be expected to change with liquid composition providing these conditions were not changed. Since the liquid and gas rates were not changed, the state of motion was kept constant; some error was introduced because the properties of the gas film changed due to the presence of the second component. However, this effect was expected to be small.

Finally, $K_G a$ for the various liquid concentrations, was computed by (26) making the assumption of an ideal mixture in order to compute p_e. Hence:

$$(28)\quad p_e = p^{*} = x\, P^{o}$$

The liquid film coefficients were again computed from (3d) using a mean liquid film condition to obtain P^o and v_m. The wet-bulb temperatures for the mixtures were obtained by assuming a linear variation with liquid mole fraction; a further discussion of this is given in the section on interpretation of data. Liquid film coefficients were then correlated by the method proposed by Stutzman[25] (equation 20).

It should be noted that the technique described above permits calculation of $k_L a$ for the pure component by two different methods. (The first is direct measurement on the pure component and the second is extrapolation of the $k_L a$ - concentration plot to the pure component intercept.) A comparison of these supposedly identical values gives an indication of the reliability of the assumption made for the interface condition. Hence, for any given set of data, it is possible to calculate $k_L a$ by several assumed interface temperatures (e.g., one for wet-bulb, one for main-body and one half way between). If the deviation obtained for each of the several points are plotted against the assumed interface temperature, the point at which they minimize should indicate the correct assumption. This, then, was the technique used in checking the wet-bulb assumption used throughout this work. As pointed out previously, at the liquid and gas rate at which the runs were made, this deviation was zero (within the limits of accuracy of the experiment) for a wet-bulb interface.

SCOPE OF WORK-DESCRIPTION OF EQUIPMENT

Measurement of overall mass transfer coefficients requires that inlet and outlet liquid compositions and temperatures; inlet and outlet gas compositions and temperatures; and liquid and gas rates be known. All temperatures were measured directly by thermometers in the fluid streams. The liquid inlet composition was measured by analysis with a refractometer; the gas inlet composition was zero. By mounting the tower on a scale and insuring that all liquid leaving the column was added again at the top, the amount of liquid in the system was maintained constant except for that lost by evaporation. Keeping the scale reading constant by the measured addition of fresh liquid from a burette, the amount of evaporation and, hence, the exit gas composition was known. Liquid and gas rates were measured by rotameters. All other streams being known, the exit liquid composition could be calculated from a material balance.

If the liquid consists of a mixture of two different compounds the above procedure requires that the ratio of components in the exit gas be the same as that in the liquid added to the system. (This will only be true for one particular circulating liquid composition, the steady state composition.) If one component is present in greater proportion in the liquid added to the system than in the vapor, then the circulating liquid composition will change. As the composition of the circulating liquid changes, the composition of the exit gas changes. Thus, the criterion for steady state is the constancy of the circulating liquid composition with time; when this is attained, the exit gas composition can be computed.

Figure 1 is a diagram of the apparatus used. In summary, the apparatus consisted of the following pieces of equipment:

1. A <u>packed tower</u> mounted on
2. <u>Scale Balance</u> and fed by an
3. <u>Air Supply</u> which passed into the bottom of the tower.
4. The air supply was pre-treated by passing it through a <u>Drying Column</u>
5. and an <u>Air Heat Exchanger</u>. It was measured by an
6. <u>Air Rotameter</u> and
7. <u>Manometer</u> installed just prior to the column.
8. Air stream temperatures were measured by <u>Thermometers</u> installed before and after the column.
9. The <u>Liquid Supply</u> was stored in a separatory funnel and fed into a
10. <u>Burette</u> installed just prior to the
11. <u>Liquid Circulating Pump</u>. The stream next passed into a
12. <u>Liquid Rotameter</u> and a

13. Liquid Drying Column before passing through a
14. Liquid Heat Exchanger and into the top of the column.
15. Liquid stream temperatures were again measured by Thermometers. Liquid from the bottom of the column passed through
16. Flexible Tubing merging with the feed stream from the burette.
17. A Sample Valve for the circulating liquid was installed just prior to the stream entrance into the top of the tower in order to permit
18. composition analysis by Refractometer.

For ease of discussion, the system will be broken down into three sections:

A. Tower and Scale
B. Air System
C. Liquid System

A. Tower and Scale - The tower was a 4-inch ID glass column, 18-inches long, and packed to a depth of 4.75-inches with 9/16 inch diameter glass marbles. There were 77 ft^2 of packing area per ft^3 of packing volume; the fractional void space was 48%. The packing was supported by a wire screen resting on a cruciform foundation suspended 6-inches above the bottom of the column.

The lower section of the column served as a liquid reservoir for the suction side of the pump. A stainless steel flange was bolted to the bottom of the column with the seal being maintained by gasket. A liquid down-comer and a gas entrance line were inserted into this flange. The gas line extended up to the bottom of the packing and was fitted with a protective cap; this served to prevent liquid from draining into the gas line and to distribute the air over the entire packed surface.

Fitted over the top of the tower, but in no way connected to it, was a 6-inch length of 5-inch standard pipe. The liquid line was inserted into the side of this pipe and terminated in a spray head which extended down into the column. The spray head consisted of a ½-inch steel cap which had numerous 1/8-inch holes drilled in it. Fitted over the top of the 5-inch pipe was a cast iron reducing cap which contained the gas exit line. Also inserted into the side of this cap was a thermometer which measured the gas exit temperature.

Three legs were welded to the bottom flange of the column to support the column on the scale balance. The balance was an Eimer and Amend platform type with a 6-pound range. The scale was graduated in 2 gram divisions and covered a 500 gram range; absolute readings could be obtained to the nearest ½ gram but deflections from a fixed point could be noted with even greater accuracy.

B. Air System - The air supply was delivered at 80 psig from a Worthington 2-stage reciprocating compressor. It was then reduced to from 10-25 psig by a Leslie diaphragm pressure reducing valve. Operating in this manner minimized pulsations in the air flow.

The air stream then passed into a Silica Gel drying tower designed to reduce the humidity of the entering air to a maximum of 0.00084 pounds of water per pound of air. This figure was calculated to be the maximum water content permissible so as not to cause a change of more than 0.3°C in the acetone liquid temperature under minimum flow conditions (the temperature change to be caused by desorption of the water into the acetone). The drying column was designed in accordance with the procedure outlined in Perry[3]. The unit consisted of a 6-foot length of 3-inch standard iron pipe packed to a depth of 5.1 feet with 6-8 mesh Tel-Tale Silica Gel. This is sufficient material to permit 6 hours of continuous operation under the conditions specified. A color change of the drying agent from deep blue to pale pink specifies the necessity for regeneration which is accomplished by heating the material for 24 hours in an oven at 300-350°F. Concentrically fitted around the outside of the drying column was a four foot length of 8-inch standard pipe; cold water was circulated through this to prevent the Silica Gel from overheating during operation. Three inch steel plugs were fitted into the top and bottom of the drying column to facilitate quick removal of the Silica Gel for regeneration.

The air stream was next directed into a shell and tube heat exchanger which permitted heating to a maximum of 55°C.

The air flow was controlled by a needle valve located downstream from the heat exchanger. The stream was then directed through the air rotameter and into the bottom of the column. The rotameter was a Fischer and Porter, Tube No. CA-25-B, Dwg. G17170-0, Serial J2-3236, capable of 30 SCFM but reading in arbitrary units. Figure 2 is a calibration curve for the meter. The final connection from the rotameter to the column was made by flexible rubber tubing. A U-Tube Manometer, containing CCl_4, was installed just prior to the column to measure air pressure. Air temperature was measured by thermometer also installed at this point.

C. <u>Liquid System</u> - Flow energy for the liquid system was supplied by an Eastern Industries centrifugal pump (Model D-11) powered by an 1/8 HP explosion-proof motor (3450 RPM). The capacity of the pump was 60 gal/hr at 5 ft H_2O head. The liquid flow stream was directed through a needle valve into a solvent rotameter fitted with special acetone gaskets. The rotameter read directly in percentages of the maximum reading which was 0.94 gpm for a liquid S.G. of 0.79. The rotameter was a Fischer-Porter, Tube B4-27-10/70G, Serial #12-4142/2.

The liquid stream passed from the top of the rotameter into the top of a liquid drying column. This consisted of a 2-foot length of standard 2-inch pipe which was packed with Molecular Sieve-5A (A product of the Linde Air Products Company). Although the exact performance of this unit was not checked, it was claimed by the drying agent manufacturer to be capable of removal of the last traces of moisture from the liquid. Following this, the liquid passed through an 8-foot length of 3/8-inch coiled copper tubing placed in a 5-gal. can. This served as a constant temperature bath for the liquid stream since either hot or cold water, as desired, could be circulated through the can. The stream was then directed into the spray head located over the column. Just prior to entrance to the column, the liquid inlet temperature was measured by thermometer. In addition, a sample valve was here installed to permit measurement of the liquid circulating composition.

Analysis of the circulating liquid composition was accomplished with a Spencer, No. 491 Refractometer. This instrument does not read the refractive index directly but measures an angle that is related to refractive index. The device was pre-calibrated (Fig.3) for the acetone-MIK systems used; the difference in refractive index for the pure compounds was 22.8°. The instrument was capable of readings reproducible to $\frac{1}{10}$ degree when properly controlled for temperature variation.

A one liter separatory funnel served as a liquid reservoir. It was located at a sufficient height above the system so that liquid could drain by gravity either into the burette or into a connection in the bottom of the coil located in the constant temperature bath. (The latter connection was installed to permit complete purging of air from the system when filling - a purging valve was located at the top of the drying column to accomplish this.)

Liquid from the bottom of the packed column passed through a plasticized Kel-F flexible connection to a point where its temperature could be measured. Following this, it merged with the feed stream from the burette and passed to the suction side of the pump.

The entire apparatus (except air supply, air drying and air heating units) was located in a laboratory hood as a safety precaution. All thermometers had a range of -2 to 100°C and could be read to 0.05°C. he thermometers were checked against each other over the range of temperatures involved in the work and were found to agree within 0.1°C. The specifications of the solvents used in the work are listed in Table I.

OPERATING TECHNIQUE

The liquid system was first filled so as to expel all air through the purging valve. Liquid circulation was then established at the desired rate and temperature; weights were added to or removed from the scale to set the pointer at an arbitrarily chosen reading. The liquid was allowed to circulate for at least 15 minutes, and if no change in weight was observed, the system was assumed to have no leaks and to be operating satisfactorily. The air valve was then opened, and the air rate was adjusted to the desired rate and temperature. Liquid was then fed in through the burette so as to maintain the scale reading constant.

The rate of evaporation was determined as follows:

a. The burette was filled from the overhead liquid reservoir. This was accomplished while the liquid was feeding into the system at the fixed rate described above. Upon filling the burette, the valve from the reservoir was closed and the level of liquid in the burette was allowed to fall.

b. When the level in the burette passed a fixed point, a stop-watch was started, and the reading of the scale balance was noted.

c. The burette stop-cock was then opened wide and a measured volume of liquid was dumped into the system.

d. When the scale balance returned to its original point, the stopwatch was stopped, and a record of the rate of evaporation was available.

e. Liquid and gas temperatures were then recorded and the run repeated and averaged.

The above procedure was utilized to determine the evaporation rates of pure acetone, pure MIK and the liquid mixtures.

The steady state liquid compositions were determined prior to the runs by removing samples at fifteen minute intervals for analysis in the refractometer. If there was no change in concentration, the exit gas composition (on an air-free basis) was the same as the particular feed composition being used for the run. If the composition changed, an adjustment in composition in the proper direction was made and the system kept running until steady state was attained.

OBSERVED DATA

Runs with pure acetone were made at four liquid rates (400,800,1200, and 1600 lbs/hr-ft^2) and over a gas rate range of 113-555 lb/hr-ft^2. These data, along with pertinent calculations, are tabulated in tables II - V. Runs with pure MIK were made at three liquid rates (800, 1200 and 1600 lbs/hr-ft^2) and over a gas rate range of 202-646 lbs/hr-ft^2; these data are tabulated in tables VI - VIII. Runs with acetone-MIK mixtures were made at a liquid rate of 1200 lbs/hr-ft^2 and a gas rate of 313 lb/hr-ft^2; the liquid feed compositions were 40, 60, 80, 90, 92.5 and 95% acetone by volume. These data, along with pertinent calculations, are given in table IX. Additional runs were made with pure acetone at elevated gas inlet temperatures and decreased liquid inlet temperatures; the gas temperatures covered the range of 25.3 - 55.4°C and the liquid temperatures covered the range of 14.1 - 20.7°C. All the latter runs were made at a liquid rate of 1200 lb/hr-ft^2 and a gas rate of 313 lb/hr-ft^2; the data are listed in table X.

CALCULATIONS

The calculated values of k_Ga and K_Ga for the pure components are shown in Figures 4 - 5. These mass transfer coefficients are plotted against G_o at constant L on a log - log scale. The values of k_Ga were calculated assuming the interface temperature to be the wet-bulb temperature. Hence, they are pseudo values. They may not necessarily represent gas film coefficients based on actual conditions but rather maximum obtainable values. If the k_Ga represents an actual gas film coefficient the liquid film coefficient could be calculated from equation 3d. These values of k_La are shown in Figure 6 as a function of mol % M.I.K. in the liquid film. Also included are the so-called liquid film coefficients k_L^*a calculated by the method of Stutzman[25] and k_L^xa calculated by assuming the interface temperature half-way between the wet-bulb temperature and the main body liquid temperature. Figure 7 shows the variation of resistance to mass transfer through the liquid film with the correlating function $(C_L^1)_f \, M_m \left(\frac{\mu}{(\rho D_L)}\right)$. This is done only for the resistances computed by the wet-bulb assumption which is shown to be the correlation in Figure 8.

A summary of the effect of elevated gas and decreased liquid temperatures is given in Figure 9. This graph shows

k_Ga computed on the basis of a wet-bulb interface temperature plotted against the log-mean temperature difference across the liquid film. Extrapolation of this curve to zero mean temperature difference indicates the value of K_Ga for the pure liquid as used by Stutzman in his correlation.

The method of calculation is divided into four basic parts:

A. Calculation of Humidity Charts

B. Calculation of Mass Transfer Coefficients for Pure Components

C. Calculation of Mass Transfer Coefficients for Acetone-MIK Mixtures

D. Correlation of Liquid Film Coefficients as a Function of Concentration

A. Calculation of Humidity Charts - Construction of a humidity chart is accomplished by means of Equation (16):

$$(16) \quad \frac{(H_{WB}-H_G)\ \lambda_{WB}}{(T_G - T_{WB})} = c_p \left(\frac{\mu/\rho D}{0.74}\right)^{2/3}$$

The technique is illustrated by the following sample calculation for acetone-air:

a. Choose a set of conditions; i.e. $H_G=0$, $T_G= 36^\circ C$.

b. As a first trial, assume $T_{WB}= -7.6^\circ C$.

c. From a vapor pressure chart for acetone, p_{WB}=44mm Hg.

Figure 10 is an Othmer plot for acetone, n-butanol and MIK. The data for acetone and n-butanol were obtained from Perry[5]; the data for MIK were obtained from a Shell Chemical Co. publication[20]. It was found that the vapor pressure data in Perry for MIK are substantially in error at the lower temperatures.

d. $H_{AB} = \frac{p_{AB}}{P-p_{AB}} \frac{M_{air}}{M_{acetone}} = \frac{44}{760-44} \left(\frac{29}{58.1}\right) = 0.124$ lb/lb.

e. $\lambda_{AB} = 138$ Cal/gm

(Figure 11 is a Watson plot of $\frac{\lambda}{T}$ (q) vs. T_r and was used for acetone, MIK, and n-butanol. The acetone data shown were obtained from Perry[4]. The MIK and butanol critical data were extrapolated from boiling point data[20] by methods outlined in Hougen and Watson[12].)

f. C_p, and μ are determined for pure air neglecting the small amount of acetone in the gas film. The values[6] are found to be $C_p = 0.238$ cal/g - °C and $\mu = 0.0407$ lb/ft-hr. ρ is computed from the perfect gas law using a mean molecular weight for the gas film (29.85#/mole) and found to be 0.0789 lb/ft^3.

g. The diffusivity, D, is computed from the Gilliland equation[9]: $D = \frac{0.0043\, T^{3/2}}{p(V_A{}^{1/3} + V_B{}^{1/3})^2} \sqrt{\frac{1}{M_1} + \frac{1}{M_2}}$. From the molecular structures, $V_{air} = 29.9$ and $V_{acetone} = 74.0$. This yields the dimensional equation: $D = 7.07 \times 10^{-5} T^{1.5}$ (ft^2/hr) and for a mean film temperature of 287.5°K, D=0.343.

h. Substituting in (16) and solving for T_{WB} yields T_{WB}=-7.6°C. This checks the assumed value and thus for a T_G of 36.0°C and H_G=0 the wet-bulb temperature, T_{WB}=-7.6°C and H_{WB}= 0.124 lb/lb.

i. Selection of other combinations of T_G and H_G yields a family of wet-bulb lines. Figures 12-14 are humidity charts constructed for acetone, MIK and butanol by this method.

B. <u>Calculation of Mass Transfer Coefficients for Pure Components</u> - The sample calculation is based on the data from Run 51 (Table III). The basic equations used for the calculations are:

$$(25)\quad k_G a = \frac{G_o}{\pi M_s}\ \frac{\Delta H}{(p_i - p_G)_{LM}}$$

$$(26)\quad K_G a = \frac{G_o}{\pi M_s}\ \frac{\Delta H}{(p_e - p_G)_{LM}}$$

$$(3d)\quad \frac{1}{K_G a} = \frac{1}{k_G a} + \frac{1}{\mathcal{H}\, k_L^o a}$$

a. The pounds of acetone evaporated per hour per square foot of superficial column area, E, is first computed.

$$E = \text{Evaporation Rate (ml/sec)} \times 3600\ \frac{\text{sec}}{\text{hr}} \times .0000353\ \frac{\text{ft}^3}{\text{ml}} \times .79 \times 62.4\ \frac{\text{lb}}{\text{ft}^3} \times \frac{1}{0.093\ \text{ft}^2}$$

$$E = 66.7 \times \text{ml/sec} = 66.7(0.859) = 57.4\ \text{lb/hr-ft}^2.$$

b. $H_2 = \quad H = \frac{E}{G_o} = \frac{57.4}{313} = 0.183 \; \frac{\text{lb Acetone}}{\text{lb Air}}$

c. $p_2 = \frac{\frac{H_2}{M_s}}{\frac{1}{M_1} + \frac{H_2}{M_s}} = \frac{H}{2+H}$ (since $M_1 = 29$ and $M_s = 58$) $= 0.0838$ Atm.

d. Since $T_2 = 8.9°C$ and $H_2 = 0.183$, $T_{2-WB} = 0.9°C$ (From Fig. 12)

e. From Fig. 10, $p_{2_{WB}} = 0.0932$ Atm.

f. In a similar fashion, $p_{1_{WB}} = \Delta p_1 = 0.0473$ Atm.

g. $$(p_{WB}-p_G)_{LM} = \frac{(p_{2_{WB}}-p_2) - (p_{1_{WB}}-p_1)}{\text{Ln}\frac{(p_{2_{WB}}-p_2)}{(p_{1_{WB}}-p_1)}} =$$

$$\frac{(.0932-0.0838)-(0.0473-0)}{\text{Ln}\frac{(0.0932-0.0838)}{(0.0473-0)}} = 0.0235 \text{ Atm.}$$

h. $k_G a = \frac{(313)}{(.396)(58)} \; \frac{(.183)}{(.0235)} = 106.2 \; \frac{\text{lb.Mol}}{\text{Hr-Ft}^3\text{-At}}$

i. Since $t_2 = 18.8°C$ and $t_1 = 5.7°C$, $p_{2_e} = 0.223$ and $p_{1_e} = 0.120$ Atm. (From Fig. 10)

j. $(p_e-p_G)_{LM} = \frac{(0.223-0.0838)-(0.120-0)}{\text{Ln}\frac{(0.223-0.0838)}{0.120-0}} = 0.130$ At.

k. $K_G a = \frac{(313)}{(.396)(58)} \; \frac{(.183)}{(0.130)} = 19.2 \; \frac{\text{lb.Mol}}{\text{Hr-Ft}^3\text{-At}}$

Assuming the interface temperature half-way between the wet-bulb and the main body liquid temperature, then

l. $t_{2_i} = \frac{T_{2_{WB}} + t_2}{2} = \frac{0.9 + 18.8}{2} = 9.8°C$

Similarly $t_{i_1} = \frac{-11.0 + 5.7}{2} = -2.7°C$

m. From Figure 10, $p_{2_i} = 0.1470$ atm. and $p_{1_i} = 0.0755$atm.

n. $$(p_i - p_G)_{LM} = \frac{(p_{2_i} - p_2) - (p_{1_i} - p_1)}{\mathrm{Ln}\,\dfrac{(p_{2_i} - p_2)}{p_{1_i} - p_1}} =$$

$$\frac{(0.147-0.0838)-(0.0755-0)}{\mathrm{Ln}\dfrac{(0.147-0.0838)}{(0.0755-0)}} = 0.0694 \text{ atm.}$$

o. $$k_G^x a = \frac{(313)\ (0.183)}{(0.396)(58)(0.0694)} = 36.0\ \frac{\text{lb.Mol}}{\text{Hr-Ft}^3\text{-At.}}$$

p. The average liquid film temperature corresponding to $k_G^x a$ is obtained by averaging t_1, t_2, t_{1_i} and t_{2_i} giving a value of 7.9°C.

p_f^o-av. is obtained from Fig. 10 and equals 0.134 Atm. Since the molal volume, v_m, for acetone equals 1.174 $\frac{ft^3}{mol}$ and $\mathcal{H} = \frac{1}{p^o v_m}$, Henry's constant is found to be 6.36 $\frac{\text{lb.mols}}{\text{ft}^3\text{-Atm.}}$

q. Substituting in (3d) and solving for $k_L^x a$, the interfacial area times the mass transfer coefficient for self-diffusion based on this assumption, gives a value of 6.48 $\frac{\text{ft}^3 \text{ acetone}}{\text{ft}^3\text{-hr.}}$

r. Similarly, taking the interface temperature equal to the wet-bulb temperature, $t_{f\text{-av}} = 3.6°C$, $p_f^o\text{av} = 0.108$ atm., and $k_L a = 2.95$ ft³/ft³-hr

C. <u>Calculation of Mass Transfer Coefficients for Acetone-MIK Mixtures</u> -

The data on which the calculation is based are those of Run 142 (Table IX). The basic equation for $K_G a$ is:

(24) $$K_Ga = \frac{G_o}{M_1 pz}\left[\frac{1}{1 - \frac{p_{e_2} - p_{e_1}}{p_2}}\right] \ln \frac{p_{e_1}}{p_{e_2} - p_2}$$

a. The weight % of acetone in the feed = $\frac{\text{Sp.Gr.Ac.}}{\text{Sp.Gr.Mix.}}$ x VOL%

$W\%_F = \frac{0.792}{0.798} \times 40.0 = 39.7\%$

b. The pounds of acetone evaporated per hour per ft^2 is:

E = 66.7 x ml/sec x Vol.% = (66.7)(0.324)(0.4) = 8.65 $\frac{\text{lb Acetone}}{\text{hr-ft}^2}$ (this is the same as the feed rate.)

c. $$p_2 = \frac{\frac{E}{58}}{\frac{G_o}{29} + \frac{E}{58} + \frac{(1-wt\%)}{wt\%}\frac{E}{100.2}} =$$

$$\frac{8.65}{2(313) + 8.65 + (0.58)\left(\frac{0.603}{0.397}\right)(8.65)} = 0.138 \text{ Atm.}$$

d. $p_{e_2} = p^*_2 = x_2 p^\circ_2$. Since $t_2 = 20.7°C$, $p_2^\circ = 0.246$ Atm. x_2 is determined from the sample taken of the entering liquid. Hence, Vol.$\%_2$ = 0.075;

$$\text{Mole }\%_2 = \frac{\text{Vol.}\%_2}{0.406 \text{ Vol.}\%_2 + 0.594} = x_2 =$$

$$\frac{0.0705}{0.406(0.0705) + 0.594} = 0.122$$

assuming an ideal mixture, $p^*_2 = (0.122)(0.246) =$ 0.0298 At.

e. Wt. % Acetone in liquid at bottom of column is: $W\%_1 = \frac{W\%_2 L - E}{L - E/W\%_F}$ where $W\%_2 = \frac{0.58x_2}{1 - 0.42x_2} = 0.0745$

Hence: $$Wt\%_1 = \frac{0.0745\ (1200) - (8.65)}{1200 - \frac{8.65}{0.397}} = 0.0685$$

f. $$x_1 = \frac{wt\%_1}{0.42\ wt\%_1 + 0.58} = 0.112$$

g. $p_1^* = x_1 p_1^o$. Since $t_1 = 17.6^o$, $p_1^o = 0.2105$ and $p_1^* = 0.0236$ At.

h. Substitution in (24), with $G_o = 313$, $M_1 = 29$, $Z = 0.396$ and $P = 1$ yields: $K_Ga = 19.3 \frac{lb.Mol}{Hr\text{-}Ft^3\text{-}at.}$

i. The value of $\mathcal{H}_{f-av}$ depends upon the temperature across the liquid film. The interface temperature at the top of the column is obtained as follows:

H_2 (acetone) $= \frac{E}{G_o} = \frac{8.65}{313} = 0.0277$. Since $T_2 = 18.6^oC$, T_{2-WB} (acetone) $= -10.0^oC$.

Similarly, $H_2(MIK) = \frac{(1-wt\%_F)}{wt\%_F} \frac{E}{G_o} =$

$$\frac{0.603}{0.397} \left(\frac{8.65}{313}\right) = 0.042$$

Hence, T_{2-WB} (MIK) $= 15.6^oC$. The combined wet-bulb temperature for the mixture is obtained by assuming a linear variation of T_{WB} with liquid mole fraction. Hence, for an $x_2 = 0.122$, $t_{2_i} = T_{WB-2} = 12.5^oC$.

j. Similarly, $t_{1_i} = 12.4^oC$ and $t_{f-av} =$

$$\frac{t_1 + t_2 + t_{1_i} + t_{2_i}}{4} = 15.8^oC.$$

k. $p^{o}_{f\text{-}av} = 0.194$ Atm. (from Fig. 9)

l. Making use of the relation: $\frac{K_G a}{k_G a} = \frac{(p_i - p)}{(p^* - p)}$ av.,

it is further calculated that $p^*_{f\text{-}av} = 0.0186$ Atm.

m. Since $x_{f\text{-}av} = \frac{p^*_{f\text{-}av}}{p^{o}_{f\text{-}av}}$, it is also seen that

$$x_{f\text{-}av} = \frac{0.0186}{0.194} = 0.0958$$

n. Finally, since $v_m = 1.93 \ \frac{ft^3}{mol}$

o. $\mathcal{H}_f = \frac{1}{p^{o}_f \ v_{m_f}} = 2.67 \ \frac{mol}{ft^3\text{-Atm.}}$

p. $k_L a = \frac{1}{(2.67)\left(\frac{1}{19.3} - \frac{1}{106.1}\right)} = 8.84 \ \frac{Ft^3 \ Ac.}{Ft^3\text{-hr.}}$

(106.1 is the $k_G a$ from Fig. 4 at these flow rates.)

q. It can also be shown by assuming the interface temperature is half-way between the wet-bulb and themain body liquid temperature and following the steps of Section C (a through p) that

$t_{f av} = 17.5°C$
$p^{o}_{f av} = 0.209$ Atm.
$K_G a / k_G a = 0.529$ (based on $k^x_G a = 36.5$)
$p^*_{f\text{-}av} = 0.0236$ atm.
$x_{f\text{-}av} = 0.113$
$v_{m\text{-}f \ av} = 1.92$ ft³/mol
$\mathcal{H}_f = 2.51$ mol/ft³-atm.
$k^x_L a = 13.6$

r. Calculation of $k^*_L a$ by Stutzman's method is accomplished by the following steps:

$p^o_{f\text{-}av} = \frac{p^o_1 \neq p^o_2}{2} = \frac{0.2105 \neq 0.2455}{2} =$ 0.228 Atm.

$\frac{K_G a}{k^*_G a} = \frac{19.3}{19.6} = 0.985$ (based on $K_G a = k_G a$ for pure component = 19.6)

By following a procedure similar to that just illustrated, $x_f = 0.1173$, $v_m = 1.91$ ft^3/mol and $\mathcal{H}_f = 2.29$ mol/ft^3- atm.

$$k^*_L a = \frac{1}{229 \left(\frac{1}{19.3} - \frac{1}{19.6}\right)} = 624$$

D. Calculation of Correlating Function For Liquid Film Coefficients

Correlation of the $k_L a$'s computed for the liquid mixtures was accomplished by equation (20):

$\frac{1}{k_L a} = K \, (C^1{}_L)_f \, M_m \left\{\frac{\mu}{\rho D_L}\right\}$. This was done based on the wet-bulb temperatures with the results being shown in Table XI.

It has been established theoretically[17] and experimentally[1] that the product $(D_L \mu)$ is linear with mole fraction for ideal mixtures at constant temperature. Since this quantity (in the form of $F = \frac{T}{D_L \mu}$) has been expressed graphically by Wilke[28] for dilute solutions and generalized by use of the parameter $\phi = 0.9$, a convenient method was available for expressing D_L as a function of mole fraction. Thus, since V_A-acetone = 74 and V_A-MIK = 140.6, $F = 2.45 \times 10^7$ and 3.65×10^7 $\left(\frac{°K - sec}{Cm^2 - Cp}\right)$ for acetone and MIK, respectively. (A second empirical correlation of Wilke's data has been proposed by Scheibel[19]. This takes the form $F = 1.22 \times 10^7 \left[\frac{V_A^{1/3}}{1 + \left(\frac{3V_B}{V_A}\right)^{2/3}}\right]$ Use of this equation yields values of 1.23×10^7 and 2.70×10^7 for acetone and MIK, respectively.)

a. Using the data from Runs 142-144 (Table XI)

$$t_{f-av} = \frac{t_{1_{WB}} + t_{2_{WB}} + t_1 + t_2}{4} = 15.8°C$$

b. $D_L \mu = \frac{T}{F}$

$$D_L \mu \text{ ac} = \frac{289.0}{2.45 \times 10^7} = 118 \times 10^{-7} \frac{Cm^2\text{-}Cp}{sec.}$$

$$D_L \mu \text{ MIK} = \frac{289.0}{3.65 \times 10^7} = 79.2 \times 10^{-7} \frac{Cm^2\text{-}Cp}{sec.}$$

c. $$(1-w_f) = 1 - \frac{x_{fac} M_{ac}}{x_{fac} M_{ac} + (1-x_{fac}) M_{MIK}}$$

$$= 1 - \frac{0.0958(58.1)}{0.0958(58.1) + (1-0.0958)(100.2)} =$$

$$0.943 \frac{\text{lbs MIK}}{\text{lbs Mixture}}$$

x_{fac} is obtained from Table IX

d. μ then equals 0.604 cp. This is based on a linear variation of μ vs. x_f with the values for the pure components being obtained from the literature[26]. Thus μ-acetone = 0.335 and μ-MIK = 0.62 cp at 15.8°C.

e. $D_L\mu$ equals 82.9×10^{-7}. This is obtained from a linear variation of $D_L\mu$ vs. x_f (between the limits of 118×10^{-7} and 79.2×10^{-7} for this temperature).

f. M_m (molecular weight of liquid in the film) is obtained from the linear variation with x_f. For this case $x_{fac} = 0.0958$ and $M_m = 96.3$.

g. ρ (lb/ft^3) is obtained from a linear variation with x_f. For this case $\rho = 50.45$ lb/ft^3.

h. $$\mu/\rho D_L = \frac{\mu^2}{\rho D_L\mu} = \frac{(0.604)^2\,(0.624)}{(50.45)(82.9 \times 10^{-7})} = 545$$

0.624 is a conversion constant to change cp to lb/ft-hr and cm^2 to ft^2. The Schmidt number is then rendered dimensionless.

i. $(C^1{}_L)_f$ is the mols MIK/ft^3 mixture.

This is obtained from $\dfrac{(1-x_{fac})}{(v_{mf\text{-}av.})}$

$v_{mf\text{-}av} = 1.93$ (from Table IX)

therefore $(C^1{}_L)_f = \left(\dfrac{1-0.0958}{193}\right) = 0.470 \dfrac{\text{mols MIK}}{\text{ft}^3 \text{ Mix.}}$

j. $(C^1{}_L)_f M_m\,(\mu/\rho D_L) = (0.470)(96.3)(545) = 24{,}700$

(Note that $(C^1_L)_f M_m = (1-x)_f \rho_f$)

$$\frac{\text{moles MIK}}{\text{ft}^3\ \text{Mixture}} \times \frac{\text{lb Mixture}}{\text{moles Mixture}}$$

k. By a similar procedure, but using the equation of Scheibel[19] to predict D_L, $(C^1_L)_f M_m (\mu / \rho D_L) =$ 17,080.

INTERPRETATION OF DATA - DISCUSSION

Examination of Figures 4-5 reveals that the usual methods of correlating the conductances for mass transfer can be applied to this system. Both K_Ga, and k_Ga when plotted on a log-log scale vs. G_o at constant L, form straight lines with slopes varying from 0.6 to 1.0. This was true for all runs (below the flooding point) except for the 400 lb/hr-ft^2 run with acetone. Apparently, at this low liquid rate, the packing was not completely wetted and the resulting slopes were closer to 0.5. The dashed curves shown on the plots are flooding velocities calculated from the data of Lobo[13]. This correlation was found to agree completely with the present data. All acetone runs were below the flooding point and the following runs with MIA were observed to be above the flooding point, as indicated on Fig. 5:

Runs	Liquid Rate (lb/hr-ft^2)	Gas Rate (lb/hr-ft^2)
139-141	800	646
133-135	800	591
109-111	1200	556

All points shown are the averages of from two to six runs at the specified conditions. This technique serves to reduce the scattering of data. The average correlation coefficient obtained by the method of least squares[11], was 85%; i.e., $(0.85)^2 \times 100 = 72\%$ of the slope of the lines is explained by the data. 28% can be attributed to experimental error and unknown sources and remain unexplained.

As discussed in the section on theory, the values of K_ga computed are maximum values since it was assumed that the interface temperatures were wet-bulb temperatures. It is shown below that this assumption is valid within experimental error for at least one set of conditions; i.e., at a liquid rate of 1200 lb/hr-ft^2 and a gas rate of 313 lb/hr-ft^2 for acetone. Theoretically, for all other runs the true K_ga could be anywhere between the indicated K_ga and the lower values of K_ga. It is probable, however, that the K_ga's plotted are good engineering estimations of the true values.

Figure 6 is a plot of the liquid film resistance ($1/k_La$) plotted vs. mole % MIK in the liquid film. The calculation was made by each of three methods:

a. Assuming the interface temperature is a wet-bulb temperature

b. Assuming the interface temperature is the main body liquid temperature (this is the method of Stutzman[25])

c. Assuming the interface temperatures is midway between the wet-bulb and the main body liquid temperature.

Examination of the curves reveals two interesting effects. First, the shape of all the curves indicates that acetone has more difficulty diffusing through itself than through MIK. This is the case since the liquid film resistance to acetone is greater for pure acetone than for pure MIK. Although this result might be intuitively unappealing, it is not without some empirical justification.

Wilke[28], in studying liquid diffusivities for a large number of materials in dilute solution, proposed a generalized correlation of the form:

(29) $\frac{T}{D_L \mu} = f(V_A)$ (V_A is the molecular volume of the solute).

The function was expressed graphically in terms of a parameter ϕ which was recommended at 0.9 for general use.

However, Scheibel[19], recorrelated the data published by Wilke and expressed his results by the empirical equation:

$$(30)\quad D_L = K \frac{T}{\mu} \left[\frac{1 + \left(\frac{3V_B}{V_A}\right)^{2/3}}{V_A^{1/3}} \right]$$

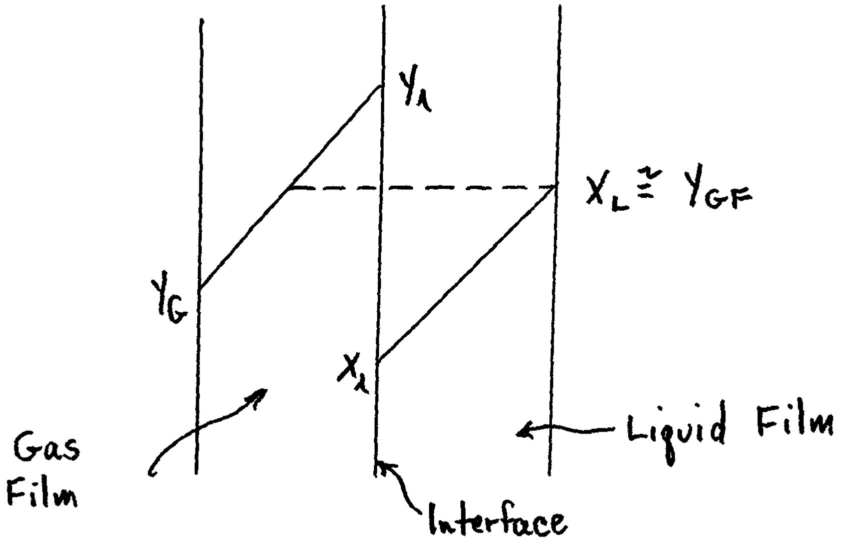
It was found that (30) provided a better fit to the data than the $\phi = 0.9$ recommendation of Wilke for general use. It is noted, however, that (30) also provides the explanation for the shape of the curves on Figure 6. The molecular volume of MIK, being greater than that for acetone, yields larger diffusivities when acetone diffuses through it than when it diffuses through itself. Thus, in general, the diffusivity increases with the molecular volume of the solvent. (See also Sherwood and Pigford[24].)

The second observation of importance about Fig. 6 concerns the intercepts corresponding to 100% acetone. If these intercepts are compared to the resistances obtained from the pure component data, some indication of the reliability of the assumption made in computing the data is obtained. Thus:

		$1/k_L a$		
Interface Temperature	Pure Component Computation	Liquid Conc. Intercept @ 100% Acetone	Discrepancy	Pure Component Conc. Intercept
t_L	0	0.0901	0.0901	0
$\frac{t_{w.B.}+t_L}{2}$	0.152	0.213	0.061	0.715
$t_{w.B.}$	0.338	0.376	0.038	0.900

This table is also shown graphically in Fig. 8. It is thus evident that at this particular gas rate and liquid rate, (313 and 1200 lb/hr-ft^2, respectively) the wet-bulb assumption gives agreement well within experimental error. Although, it is noted that there is no guarantee that the same is true for other liquid rates and gas rates, the wet-bulb method appears to offer promise of predicting liquid film coefficients within engineering accuracy for all rates.

Figure 7 contains plots of $1/k_L a$ (computed by the wet-bulb assumption) vs. the correlating variable suggested by Stutzman $Z = (J^1{}_L)_f \, M_m \, (\frac{\mu}{\rho D})$. Two curves are shown, the higher one being computed using Wilke's diffusivity curves, and the lower one is equation (30). It is evident that all the points (except one) fall close to a single straight line, thus, supporting the correlation. The single point that deviates is that generated by Runs 174-176 for 95% Acetone (by volume) feed. The most probable reason for the deviation is that it was exceedingly difficult to bring the system to equilibrium in this range of feed concentrations; small, momentary fluctuations in feed rate (possibly due to the method of obtaining the evaporation rates) tended to unbalance the system and hence the data in this range are viewed with suspicion. It is also possible that the method of computing wet-bulb temperatures for the mixtures contributed to the error. It is recalled that this method consisted of weighting the wet-bulb temperatures computed for the pure components by the mole fraction in the liquid at that point. The choice of liquid mole fraction (rather than gas mole fraction) was dictated by the fact that it gave answers consistant with the heat balance in the system. Also, at several liquid compositions, choice of the gas mole fraction to weight the wet-bulb temperatures would have resulted in impossible situations (higher gas partial pressures than interface partial pressures). It is noted that another possible weighting factor for wet-bulb temperature is mole fraction in the gas film. If this is the case, an explanation for the apparently satisfactory results obtained by assuming proportionality to main body liquid mole fraction is given by the diagram below. It is probable that main body liquid mole fraction and gas film mole fraction (on an air free basis) are approximately the same.



Thus, from Fig. 7, the following equations are recommended to describe the effect of liquid concentration on $k_L a$ for the present system at the rates specified:

(31) $1/k_L a = 0.376-0.0000102(C_L^{1})_f M_m\ (\mu/\rho D)$ (used with Wilke's[28] Diffusivity Data)

(31a) $1/k_L a = 0.376-0.0000152(C_L^{1})_f M_m\ (\mu/\rho D)$ (used with Scheibel's [19] Diffusivity Data)

The attempt to investigate the effect of increased gas-decreased liquid inlet temperatures is summarized in Figure 9. This plots $k_G a$ (computed on the wet-bulb assumption) vs. the mean temperature difference across the liquid film. As discussed in the theory, when this temperature difference is zero, then $k_G a = K_G a$; this is shown by the dashed extrapolation (the intercept being 19.6 computed from main body liquid temperatures). The extent of error in assuming $K_G a = k_G a$ at other Δ t's is evident for the present system. Note that the same is true for the calculations performed by Stutzman[25], although to a lesser extent since involatile liquids, n-butanol and n-hexanol were used.

Attempts to analyse the data presented by Stutzman by the present methods were generally unsuccessful. This was due, in large measure, to the fact that gas outlet temperatures were not reported. Attempts to calculate these, by and large, resulted in impossible heat balances (outlet gas temperatures higher than inlet gas or inlet liquid temperatures). These discrepancies could be due to simple inaccuracy of measurement, but they also might possibly be traced to the use of "wet" air. In the present experiments, where the air was dried by passing it through a silica gel column, the heat balance agreements were excellent. The only useful information remaining from the attempts to correlate Stutzman's data is the humidity chart constructed for n-butanol. This is presented as Fig. 14 due to the scarcity of these plots in the literature.

CONCLUSIONS - RECOMMENDATIONS

A. The effect of liquid concentration on the resistance to mass transfer across the liquid film was found to follow the correlating variables proposed by Stutzman[25]. For the specific system at hand, and at a liquid rate of 1200 lb/hr-ft^2 and a gas rate of 313 lb/hr-ft^2, the effect can be described by:

(31) $1/k_L a = 0.376-0.0000102\ (C_L^1)_f\ M_m\ (\mu/\rho D)$ when diffusivity data of Wilke[28] are used, and

(31a) $1/k_L a = 0.376-0.0000152\ (C_L^1)_f\ M_m\ (\mu/\rho D)$ when diffusivity data of Scheibel[19] are used.

B. The transfer of acetone through a liquid film of methyl isobutyl ketone is accomplished with less apparent difficulty than the transfer of acetone through a liquid film of acetone. This was established experimentally and supported by the empirical equation of Scheibel[19].

C. The temperature at the interface between liquid and gas films should be somewhere between the wet-bulb and main body liquid temperatures. For the system at hand, and at a liquid rate of 1200 lb/hr-ft^2 and a gas rate of 313 lb/hr-ft^2, this temperature was found to be equal to the wet-bulb temperature (within the accuracy of the experiment). This would serve as a more fundamental basis for the design of absorption equipment than the use of main-body liquid temperatures to approximate the interface.

The following recommendations are proposed for future work in this field:

A. A study should be undertaken to determine the relationship between liquid and/or gas concentration and the wet-bulb temperature of a gas passing over a liquid mixture. It has been estimated throughout the calculations herein that this temperature might be arrived at by computing the wet-bulb temperatures for each of the pure components and the assumption of a linear relationship of T_{wb} with the main body liquid mole fraction. This technique apparently was justified by the heat balances in the system. Any carefully planned experiment, however, could easily confirm or establish some other relationship.

B. It is recommended that further studies be made of the acetone-MIK system to determine how the interfacial temperature varies with liquid rate, gas rate, height, size and type of packing. The technique to be used for this study should follow that illustrated in Figure 8. Specifically, mass transfer coefficients should be determined for both pure acetone and mixtures of this with MIK all at constant liquid and gas rates. $k_L a$ should be computed by each of three methods for the mixtures (wet-bulb, intermediate, and main body liquid assumption for interface temperatures) and extrapolated to 100% acetone composition. These intercepts should then be plotted along with the k_La's determined from the pure acetone run (again by each of the three methods) and the point where the curves intersect should be taken as the true interface condition. The experiment will require carefully determined K_Ga's at low MIK feed compositions.

C. A similar study should be initiated for non-ideal mixtures and an attempt made to correlate using activity coefficients.

SUMMARY

A study of the effect of liquid concentration on the resistance to mass transfer across the liquid film was undertaken for the acetone-MIK-air system in a packed tower. The technique used was to first calculate the gas film resistance from data taken on the pure components, and then, without changing the flow conditions in the column, to determine the overall coefficients for mixtures. The liquid film resistance is subsequently calculated by subtracting gas film resistance from overall resistance. The experimental data taken can be correlated by the equations:

$\frac{1}{k_L a} = 0.376-0.0000102\ (C_L^1)_f\ M_m(\mu/\rho D)$ using diffusivity data of Wilke[28]

$\frac{1}{k_L a} = 0.376-0.0000152\ (C_L^1)_f\ M_m\ (\mu/\rho D)$ using diffusivity data of Scheibel [19].

The techniques for calculating gas film coefficients from data are critically examined. While theoretically, the interface temperature should be somewhere between the wet-bulb temperature and the main body liquid temperature, it was found experimentally that for the set of flow conditions thoroughly investigated in this work the correct interface temperature is very close to the wet-bulb temperature. Humidity charts for acetone-air, methyl isobutyl ketone-air, and n-butanol-air, all at one atmosphere pressure, were calculated and constructed.

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NOMENCLATURE

a = Interfacial contact area per unit volume, $\frac{Ft^2}{Ft^3}$

C_p = Specific Heat, cal/G_m°C

C_e = concentration of solute in liquid in equilibrium with p_G, $\frac{lb.mols}{Ft^3}$

C_G = Partial molal density of the diffusing component in the fluid stream $\frac{lb.Mols}{Ft^3}$

C_i = Concentration at phase boundary, $\frac{lb.mols}{Ft^3}$

C_1 = Concentration, $\frac{lb.mols}{Ft^3}$

D_G = Molecular Diffusivity in gas, $\frac{Ft^2}{Hr.}$

D_L = Molecular Diffusivity in liquid C_m^2/sec.

E = Evaporation rate, lbs/hr-ft^2

f = Friction factor in Fanning equation

g = Acceleration due to gravity, Ft/sec^2

G_o = Gas rate, lbs/hr-ft^2

H = Humidity, $\frac{lbs\ solute}{lbs.\ gas}$

$\mathcal{H}$ = Henry's Law constant, moles/ft^3-Atm

k_G = Gas film coefficient for wet-bulb method, lb.moles/hr-ft^2-At.

K_G = Over-all mass transfer coefficient, lb.moles/Hr-Ft^2-Atm.

K_L = Over-all mass transfer coefficient, lb.moles/hr-ft^2-Atm

k_L = Liquid film coefficient for wet-bulb method, lb.moles/hr-ft^2-Atm.

L_o = Liquid Rate, lbs/hr-ft^2

M = Molecular weight.

N_A = Rate of mass transfer, lb.moles/hr-ft^2

p_G = Partial pressure of diffusing gas in main stream, Atm.

p_e^* = Partial pressure of solute gas in equilibrium with C_e, Atm.

p_i = Partial pressure of diffusing gas at phase boundary, Atm.

P = Total pressure

q = constant in Watson latent heat correlation

r = Radius, Ft.

R = Gas constant, Atm-ft^3/lb-mol-°R

S = X-sectional area - Ft^2

T = Gas temperatures

t = Liquid temperatures

u = Average velocity, ft/sec.

v = Molal volume, ft^3/mol

V = Molecular volume

W = Weight rate of flow, lb.-mols/sec.

w = Weight fraction

x = Mol fraction

X = Length, ft.

Z = Height of packing.

Greek Symbols

ϕ: Parameter in Wilke's Correlation[28]

ρ: Density, lbs/ft^3

μ: Viscosity, C_p

λ: Latent heat of vaporization

Conditions

1 = Bottom of column

2 = Top of column

i = Interface

Subscripts

A = Solute

B =Solvent

F = Feed

f = Film

G = Gas

L = Liquid

LM= Log Mean Value

M = Mean Value

O = Pure Component

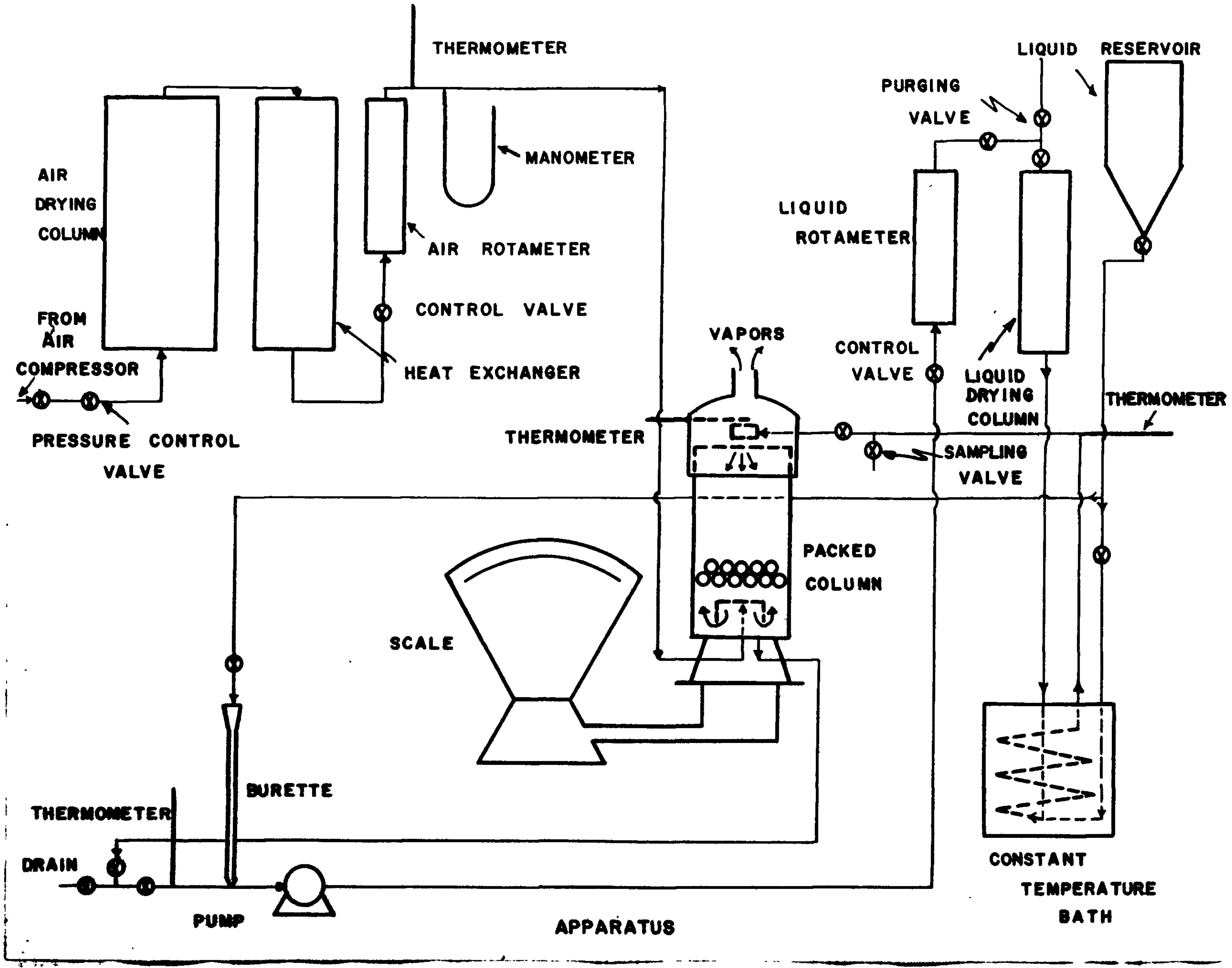
WB= Wet-Bulb

Superscripts

o = Pure Component

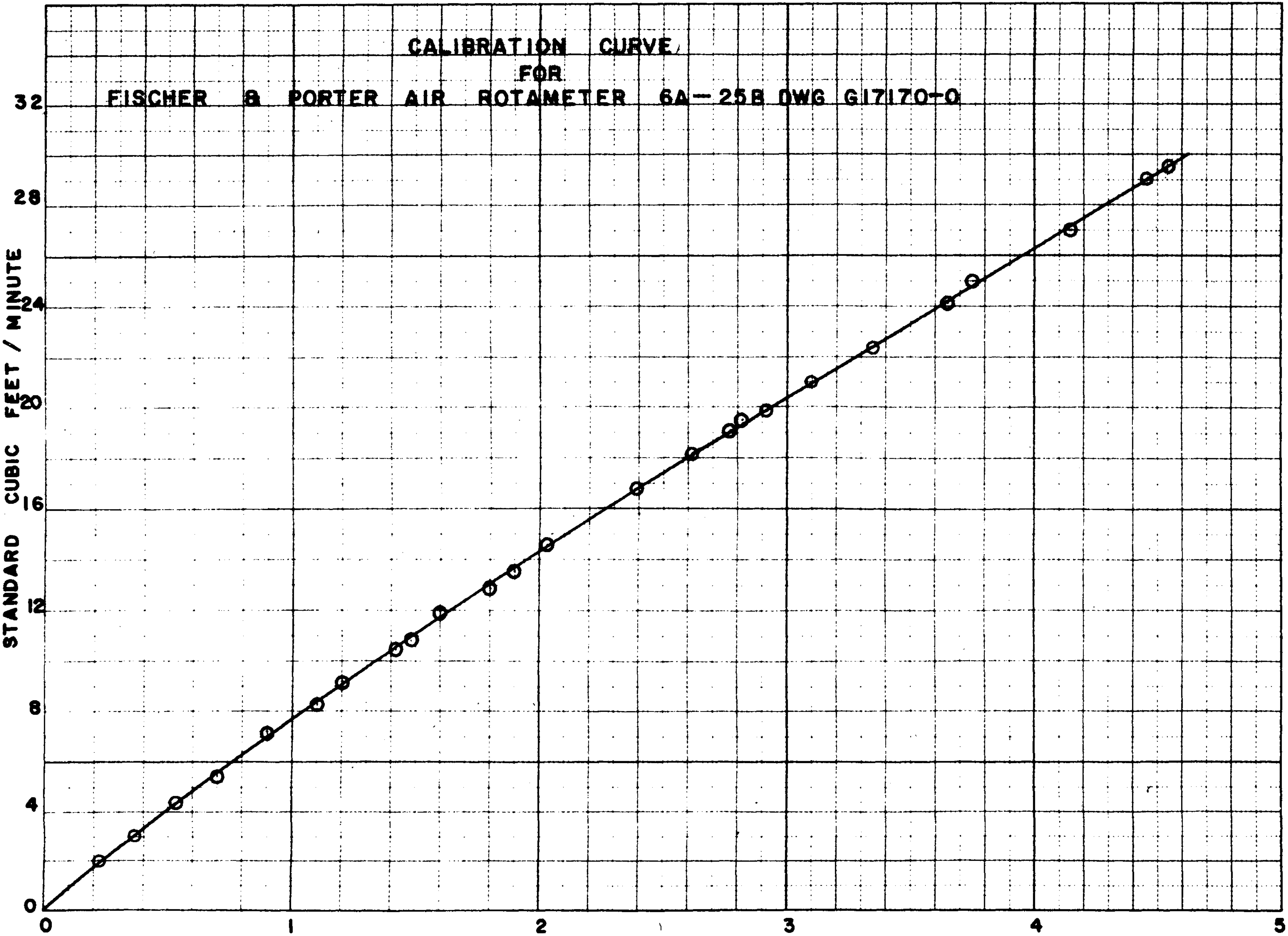
x = based on interface midway between wet-bulb and liquid temperatures.

* = Equilibrium liquid.



APPARATUS

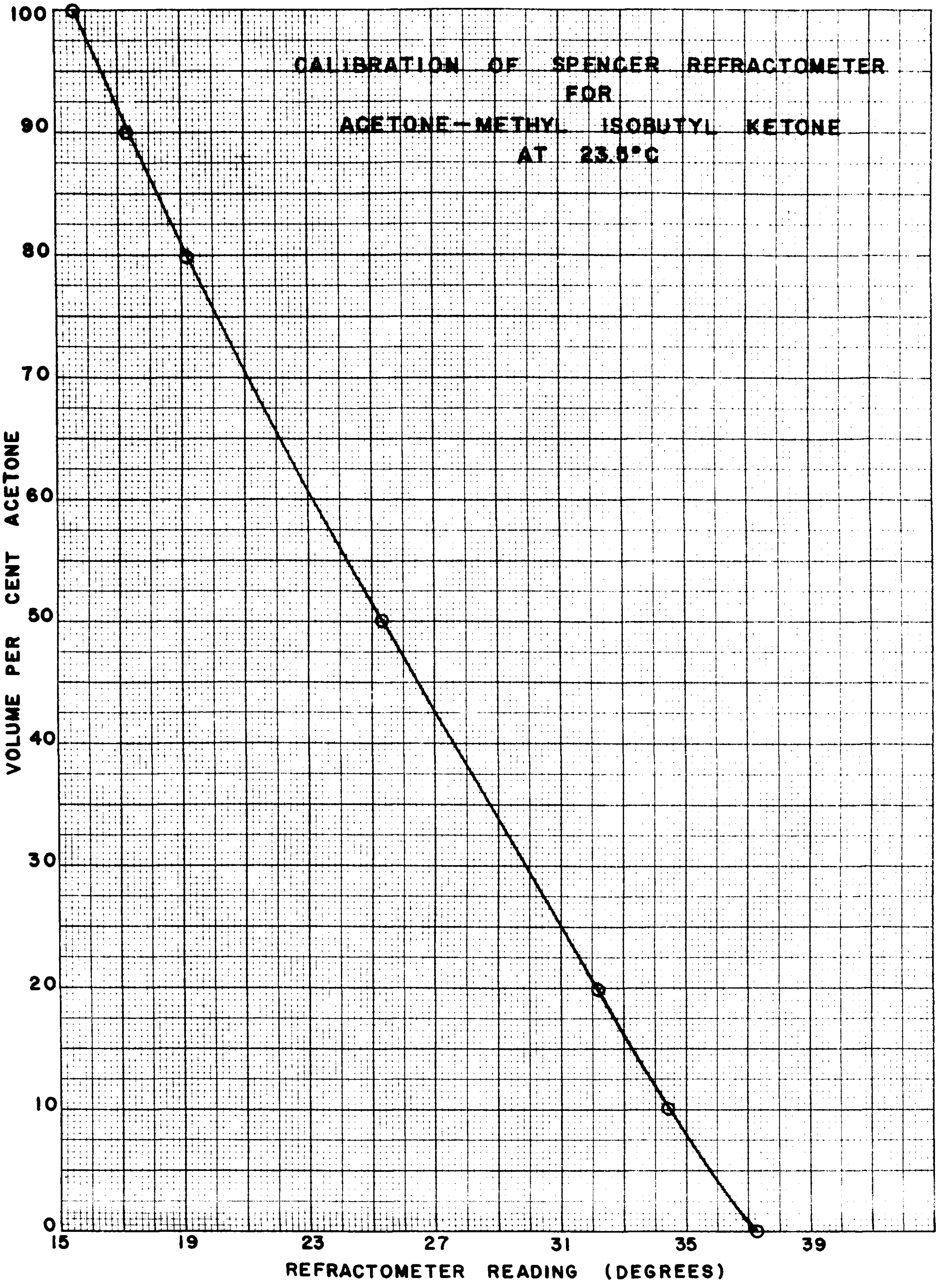
Fig. 1



CALIBRATION CURVE
FOR
FISCHER & PORTER AIR ROTAMETER 6A-25B DWG G17170-0

Fig. 2

Fig. 3



CALIBRATION OF SPENCER REFRACTOMETER
FOR
ACETONE—METHYL ISOBUTYL KETONE
AT 23.5°C

Fig. 4

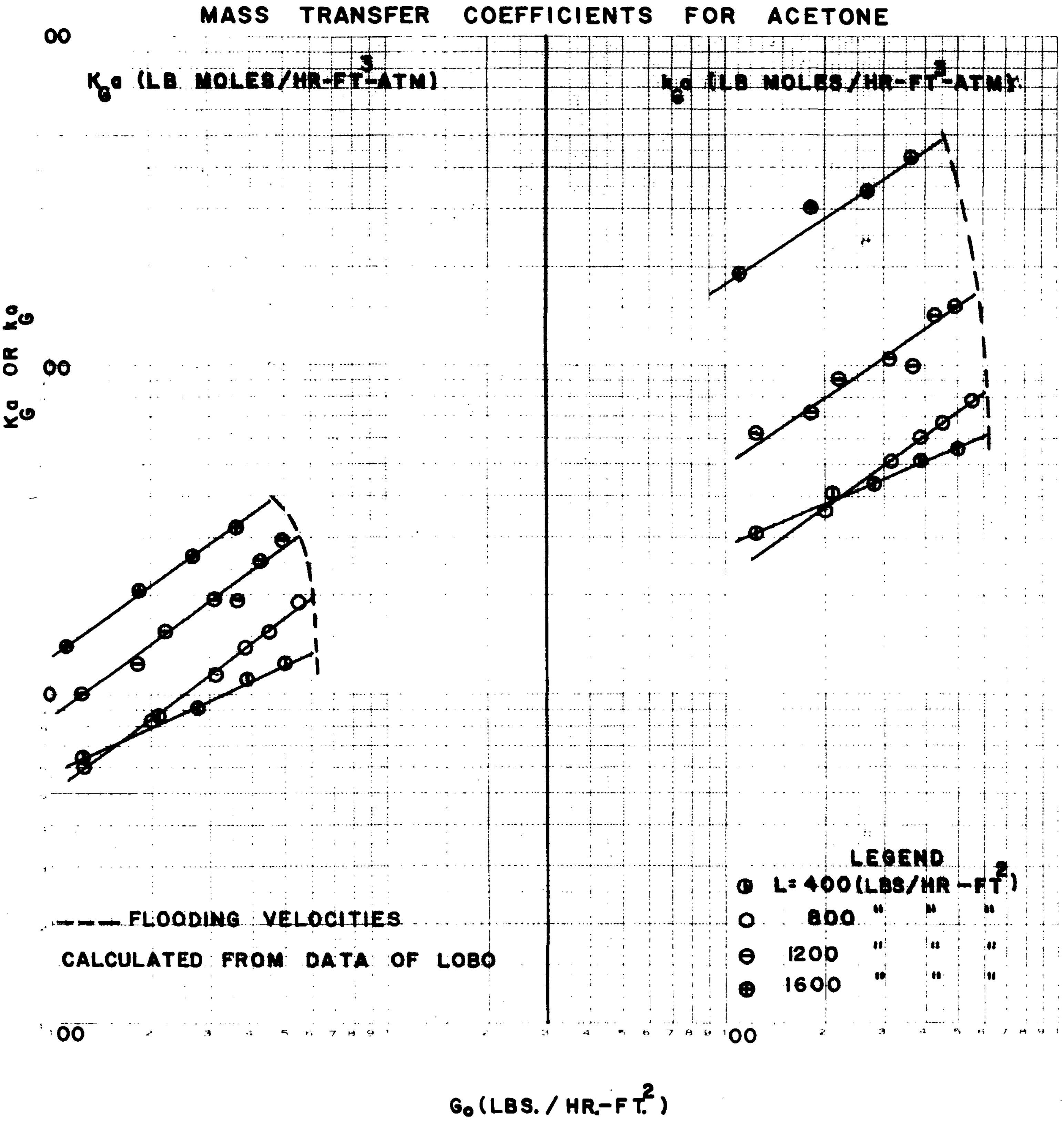

MASS TRANSFER COEFFICIENTS FOR ACETONE
K_Ga (LB MOLES/HR-FT^3-ATM)
k_Ga (LB MOLES/HR-FT^3-ATM)
K_Ga OR k_Ga
00
00
00
00
00
G_0 (LBS./HR-FT^2)
LEGEND
⦶ L= 400 (LBS/HR-FT^2)
○ 800 " " "
⊖ 1200 " " "
⊕ 1600 " " "
--- FLOODING VELOCITIES
CALCULATED FROM DATA OF LOBO

Fig. 5

MASS TRANSFER COEFFICIENTS FOR METHYL ISOBUTYL KETONE

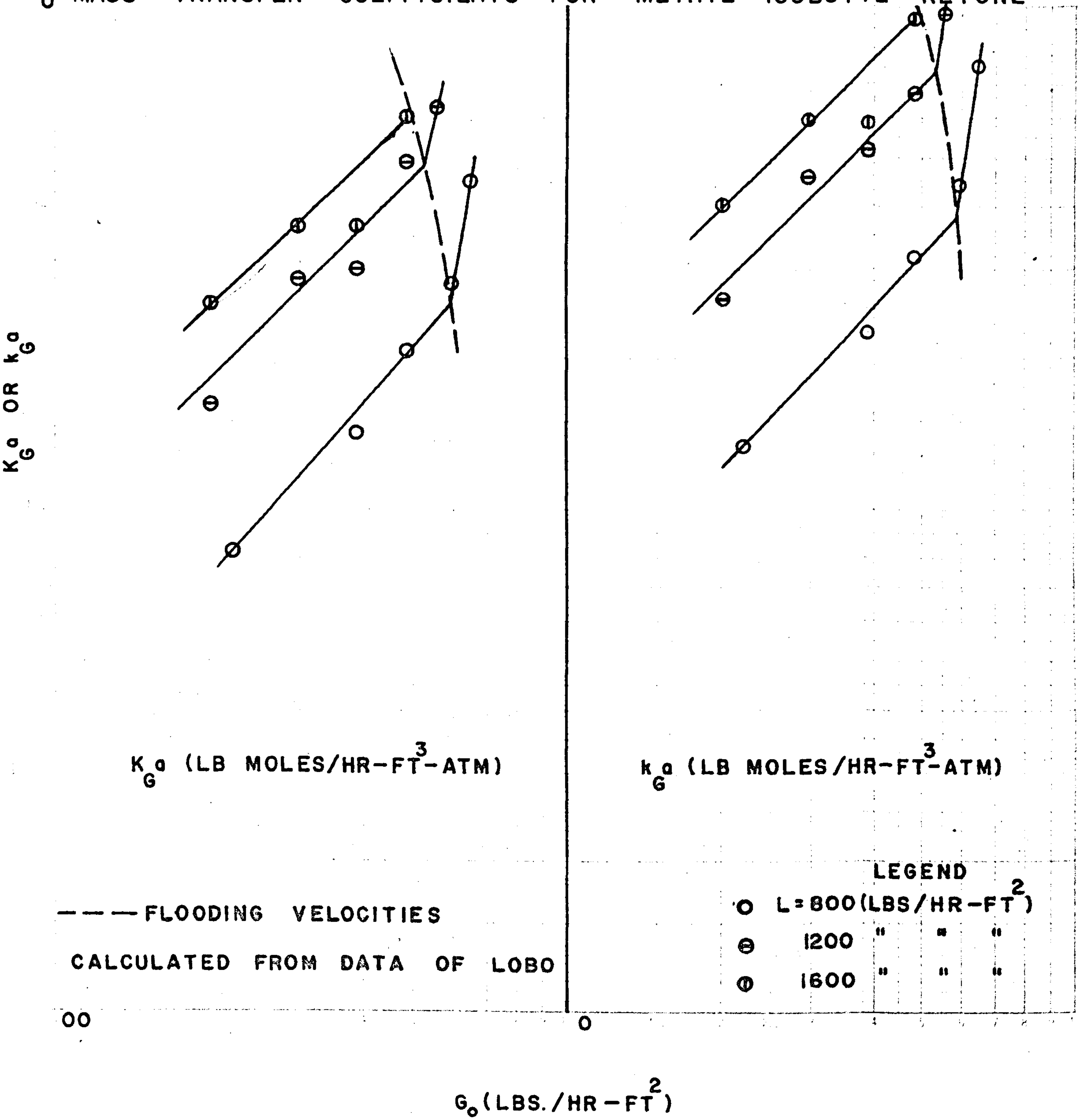


Fig. 6

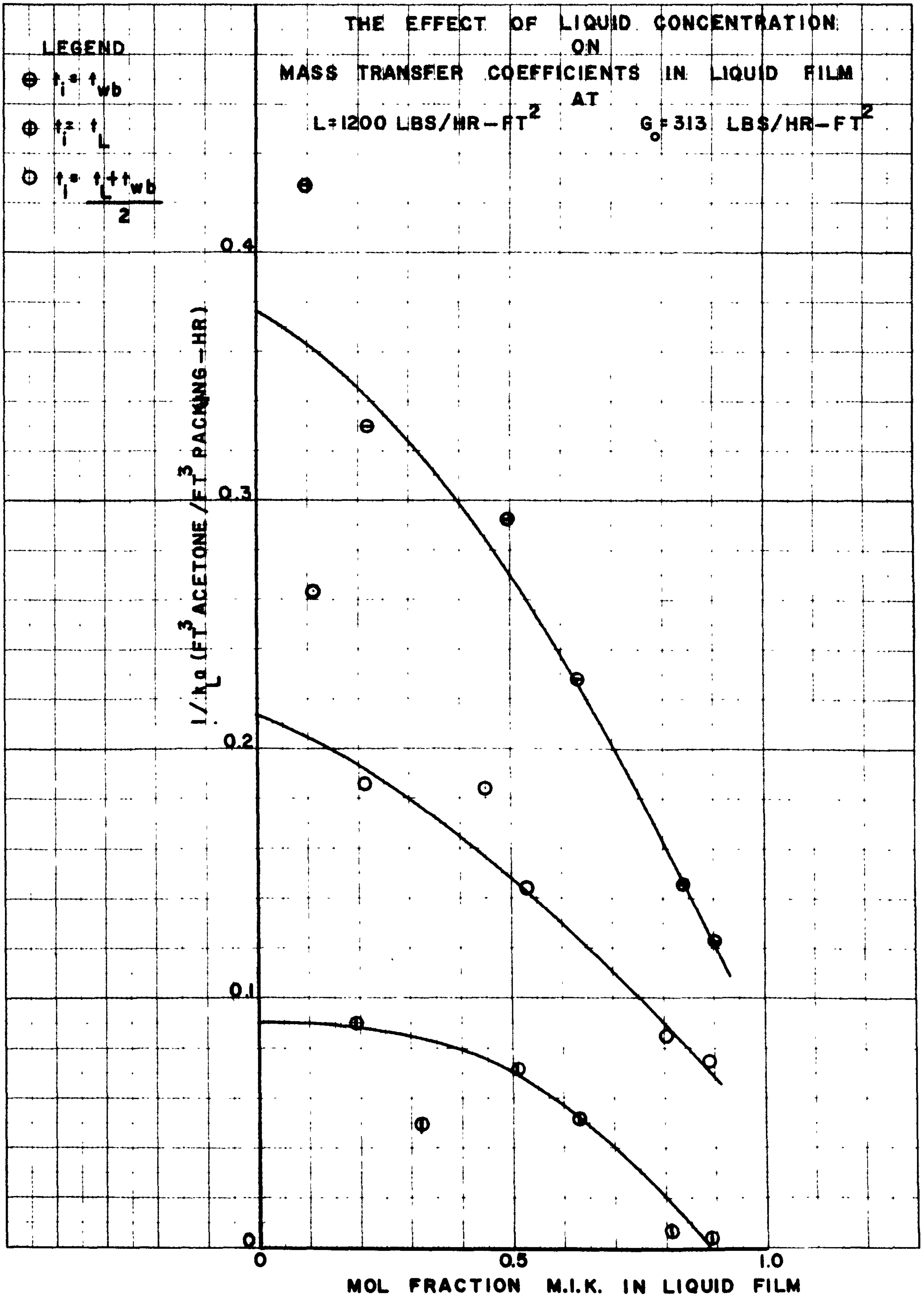

THE EFFECT OF LIQUID CONCENTRATION
ON
MASS TRANSFER COEFFICIENTS IN LIQUID FILM
AT
L=1200 LBS/HR−FT^2
G_o=313 LBS/HR−FT^2
LEGEND
⊖ $t_i = t_{wb}$
⦶ $t_i = t_L$
⊙ $t_i = \frac{t_L + t_{wb}}{2}$
$1/k_L a$ (FT^3 ACETONE / FT^3 PACKING−HR)
0.4
0.3
0.2
0.1
0
0
0.5
1.0
MOL FRACTION M.I.K. IN LIQUID FILM

Fig. 7

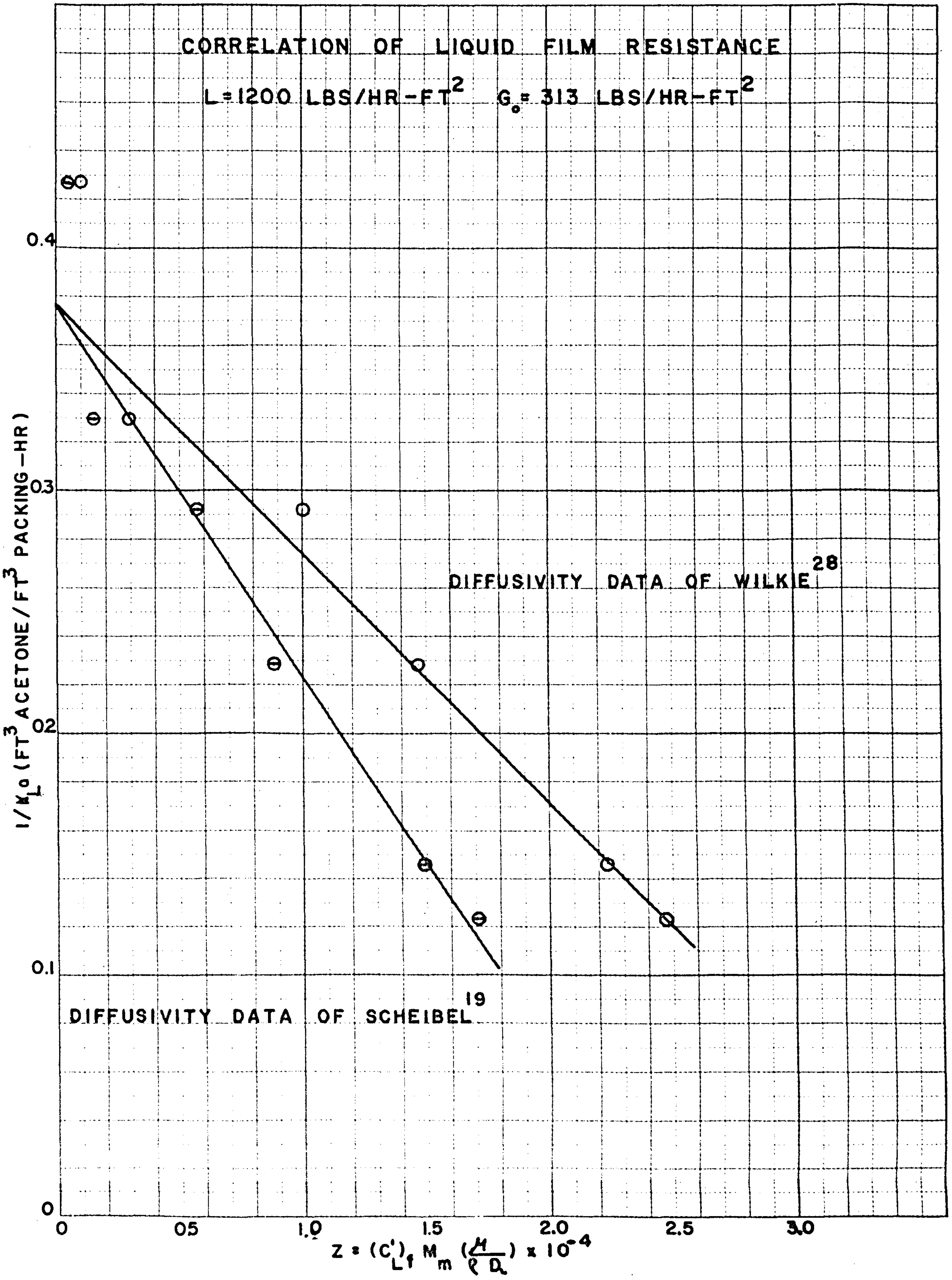
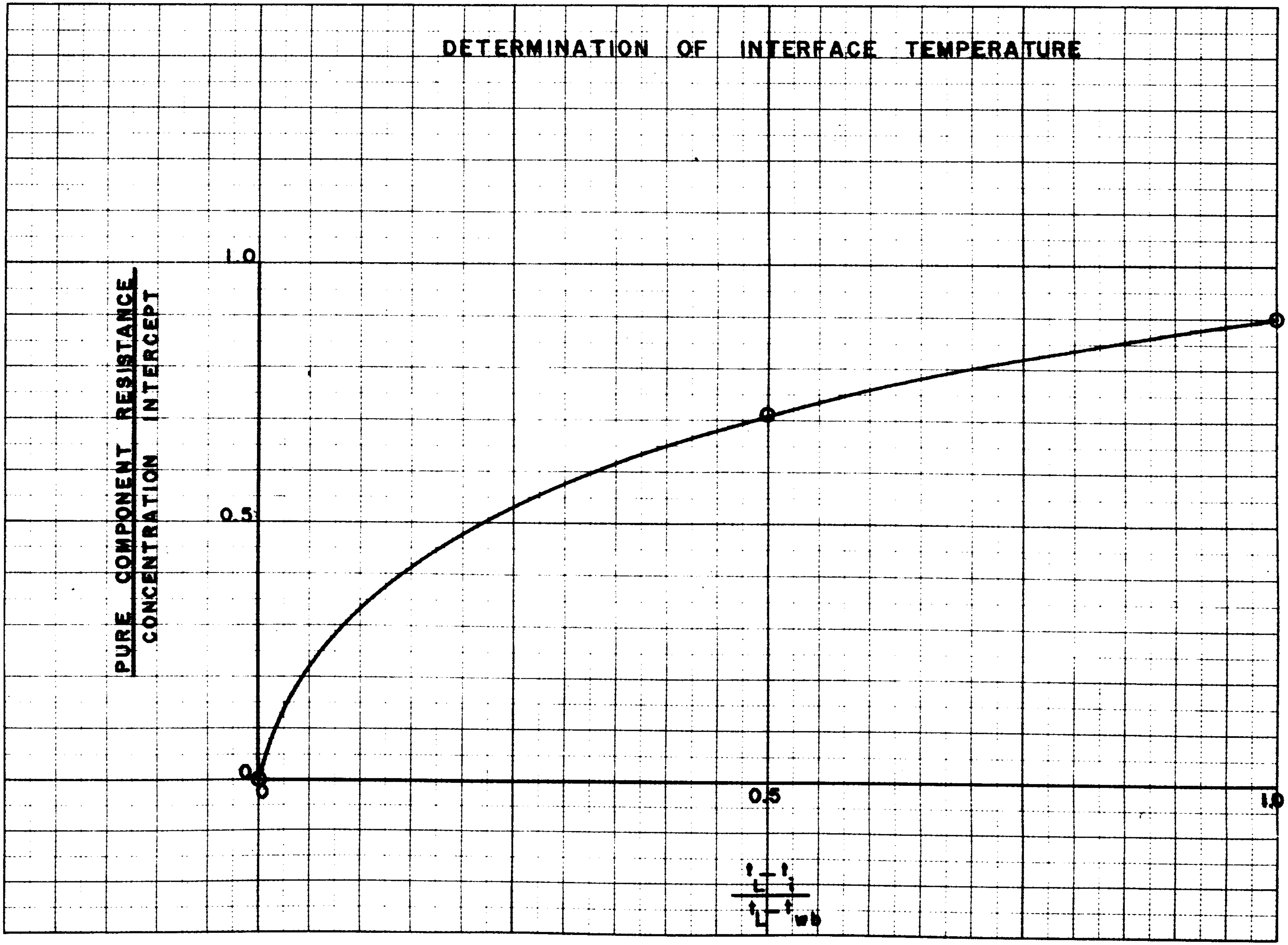
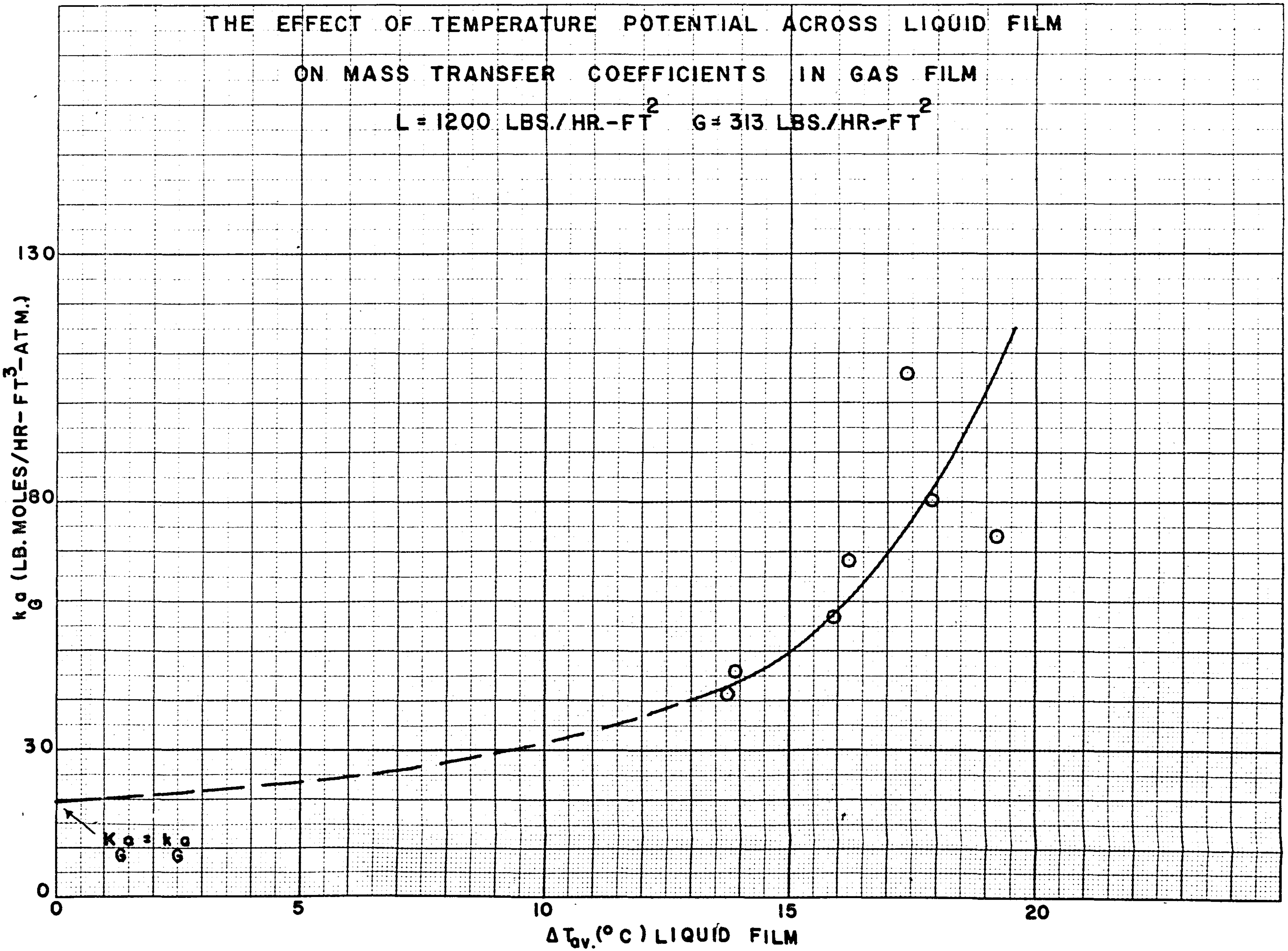

CORRELATION OF LIQUID FILM RESISTANCE
L=1200 LBS/HR-FT^2 G_o=313 LBS/HR-FT^2
DIFFUSIVITY DATA OF WILKIE[28]
DIFFUSIVITY DATA OF SCHEIBEL[19]
$1/K_La$ (FT^3 ACETONE/FT^3 PACKING-HR)
0.4
0.3
0.2
0.1
0
0
0.5
1.0
1.5
2.0
2.5
3.0
$Z = (C'_L)_f M_m \left(\frac{\mu}{\rho D_L}\right) \times 10^{-4}$

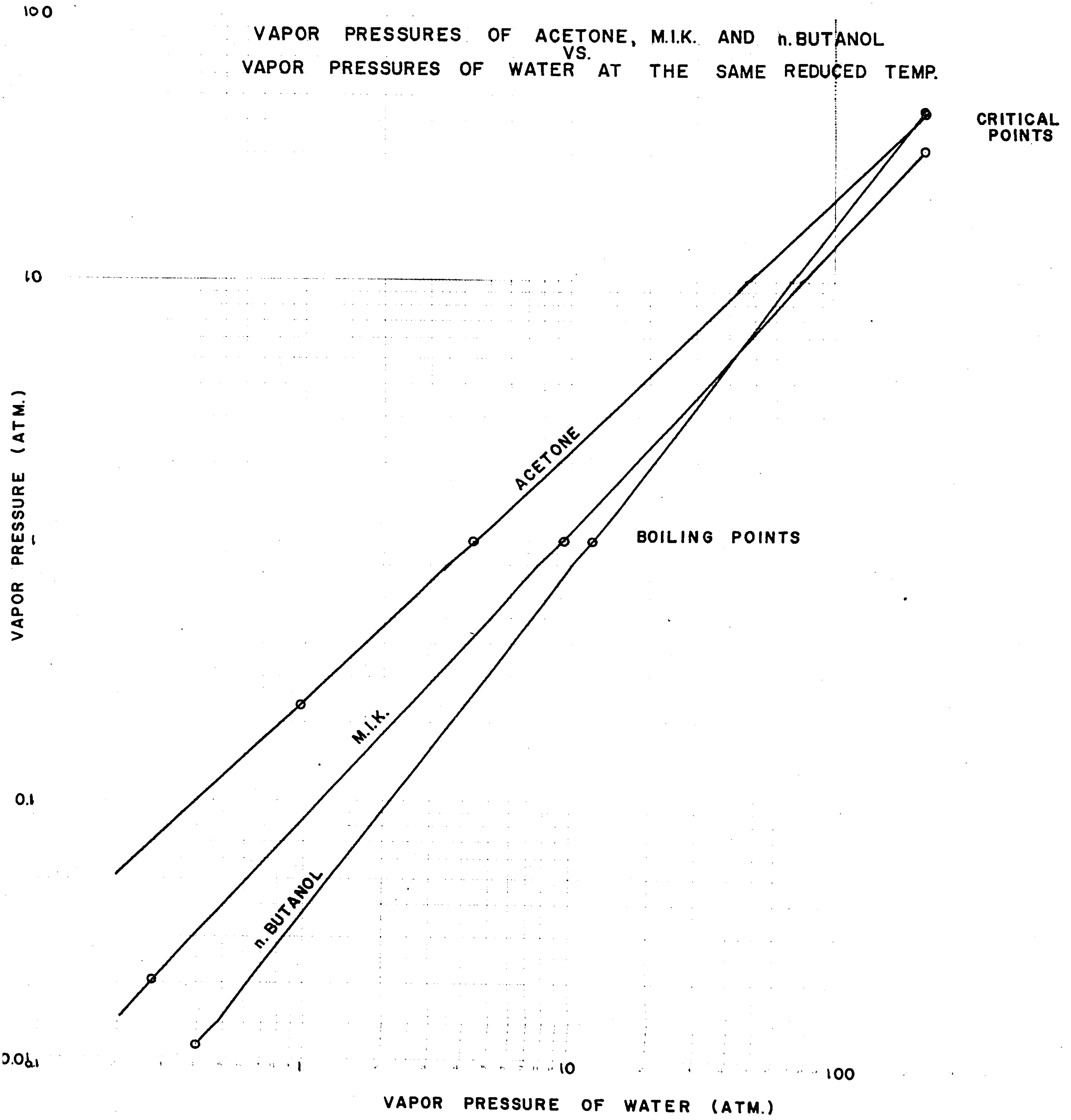
Fig. 8


DETERMINATION OF INTERFACE TEMPERATURE
PURE COMPONENT RESISTANCE / CONCENTRATION INTERCEPT
1.0
0.5
0
0.5
1.0
$\frac{t_L - t_i}{t_L - t_{wb}}$



THE EFFECT OF TEMPERATURE POTENTIAL ACROSS LIQUID FILM ON MASS TRANSFER COEFFICIENTS IN GAS FILM

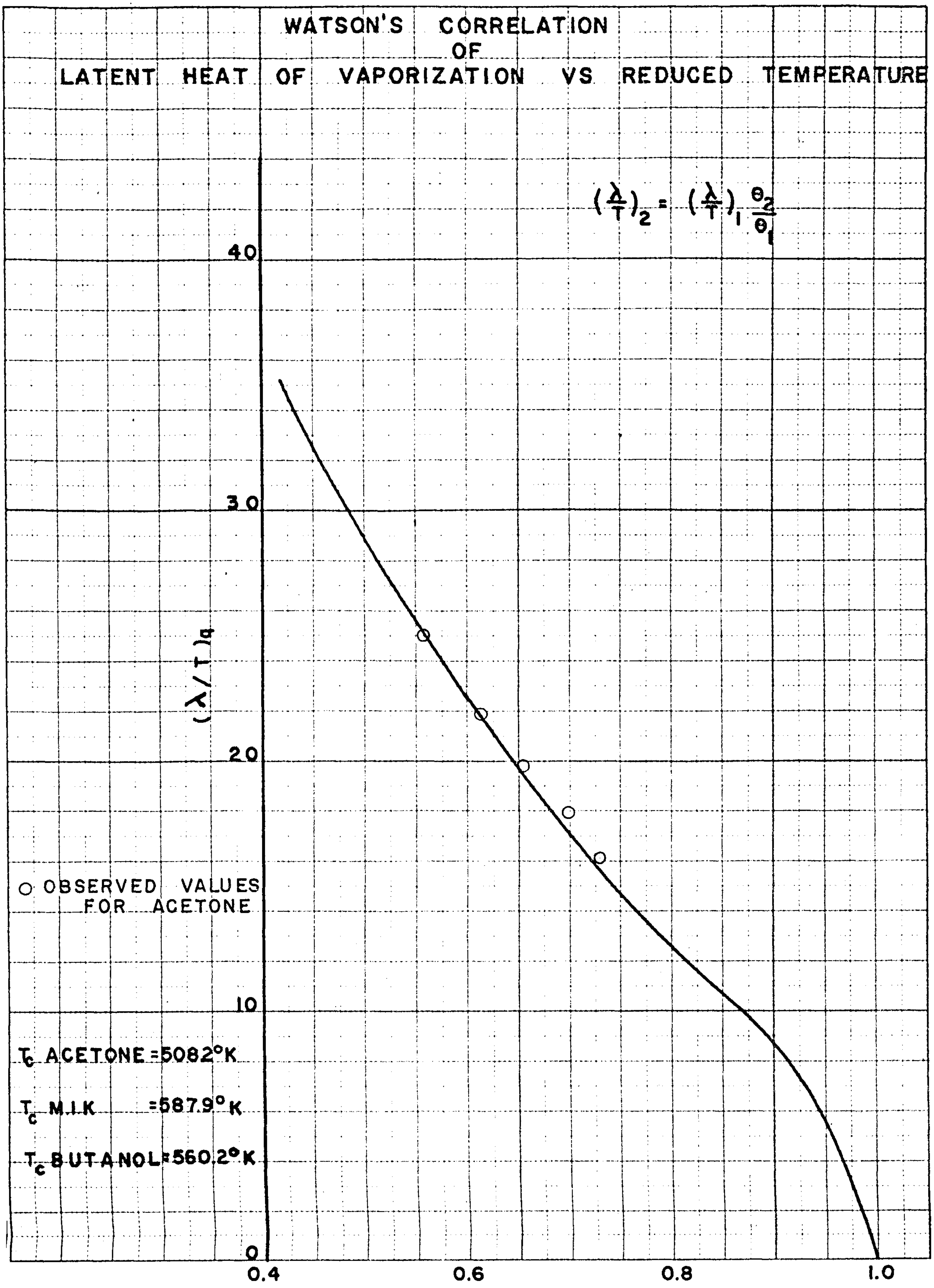
L = 1200 LBS./HR.-FT^2 G = 313 LBS./HR.-FT^2



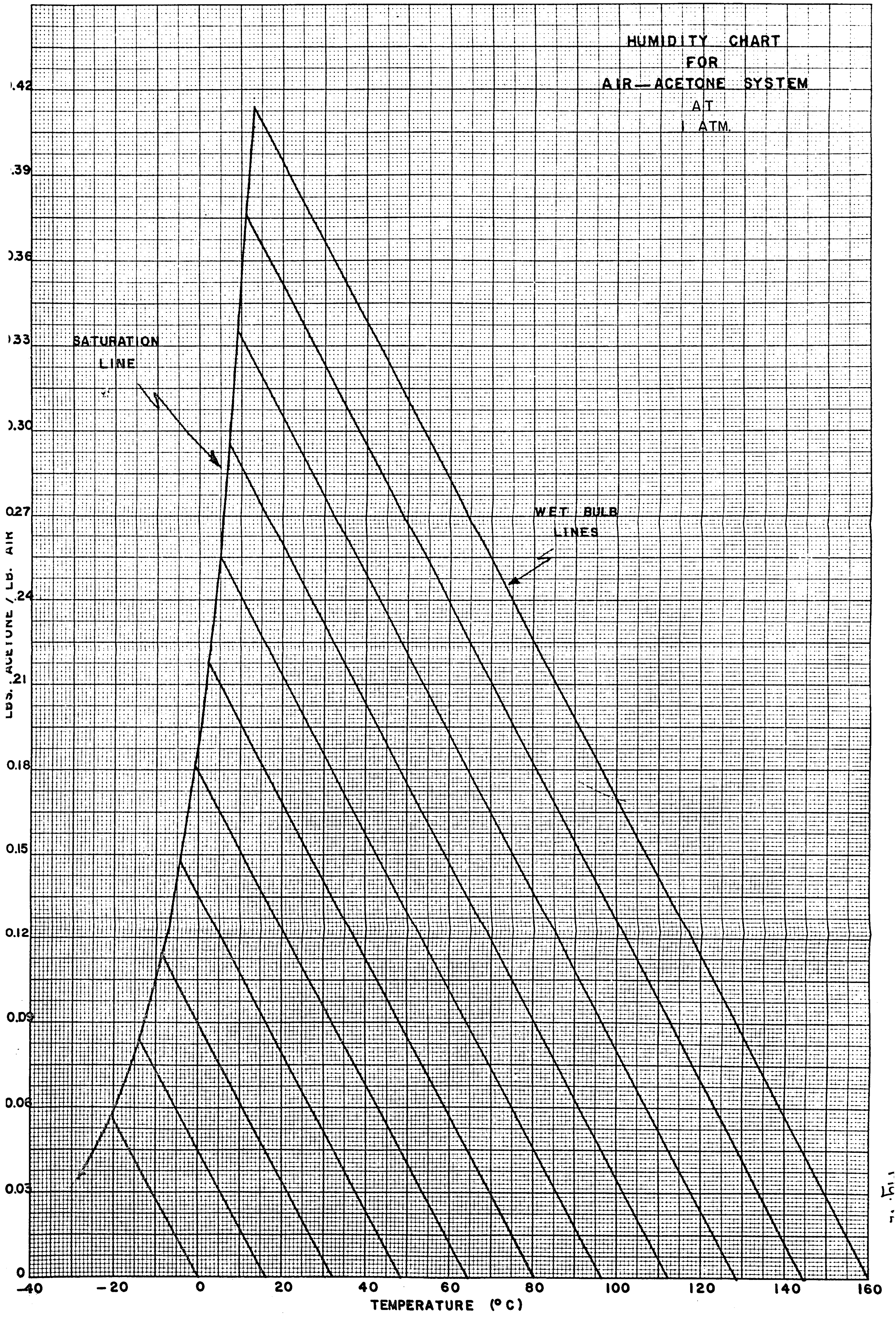
VAPOR PRESSURES OF ACETONE, M.I.K. AND n. BUTANOL VS. VAPOR PRESSURES OF WATER AT THE SAME REDUCED TEMP.

Fig. 10

Fig. 11



WATSON'S CORRELATION OF LATENT HEAT OF VAPORIZATION VS REDUCED TEMPERATURE


HUMIDITY CHART
FOR
AIR—ACETONE SYSTEM
AT
1 ATM.
SATURATION
LINE
WET BULB
LINES
LBS. ACETONE / LB. AIR
.42
.39
.36
.33
.30
0.27
.24
.21
0.18
0.15
0.12
0.09
0.06
0.03
0
-40
-20
0
20
40
60
80
100
120
140
160
TEMPERATURE (°C)

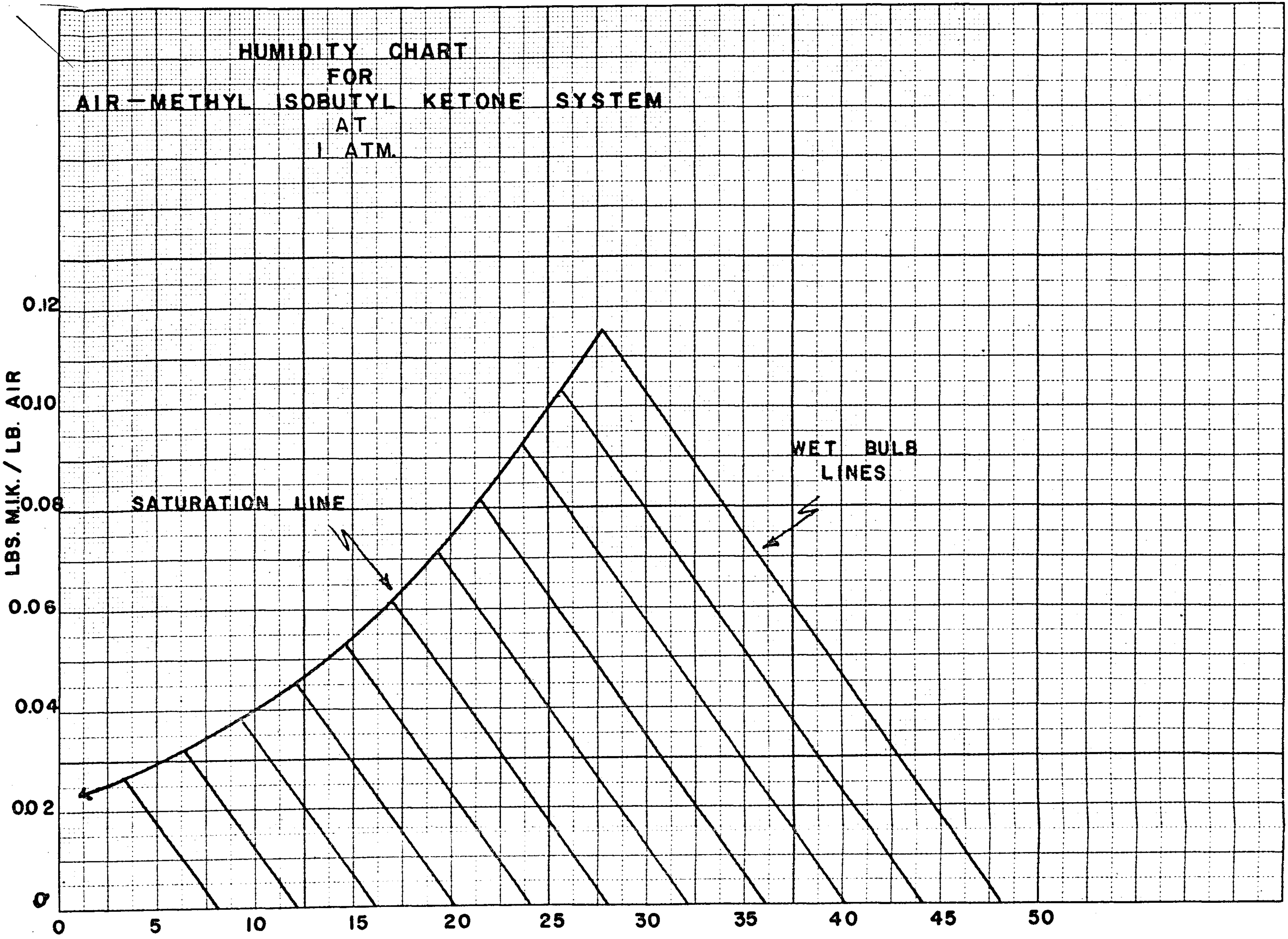

HUMIDITY CHART
FOR
AIR-METHYL ISOBUTYL KETONE SYSTEM
AT
1 ATM.
LBS. M.I.K. / LB. AIR
0.12
0.10
0.08
0.06
0.04
0.02
0
SATURATION LINE
WET BULB
LINES
0
5
10
15
20
25
30
35
40
45
50

Fig. 13

Fig. 14

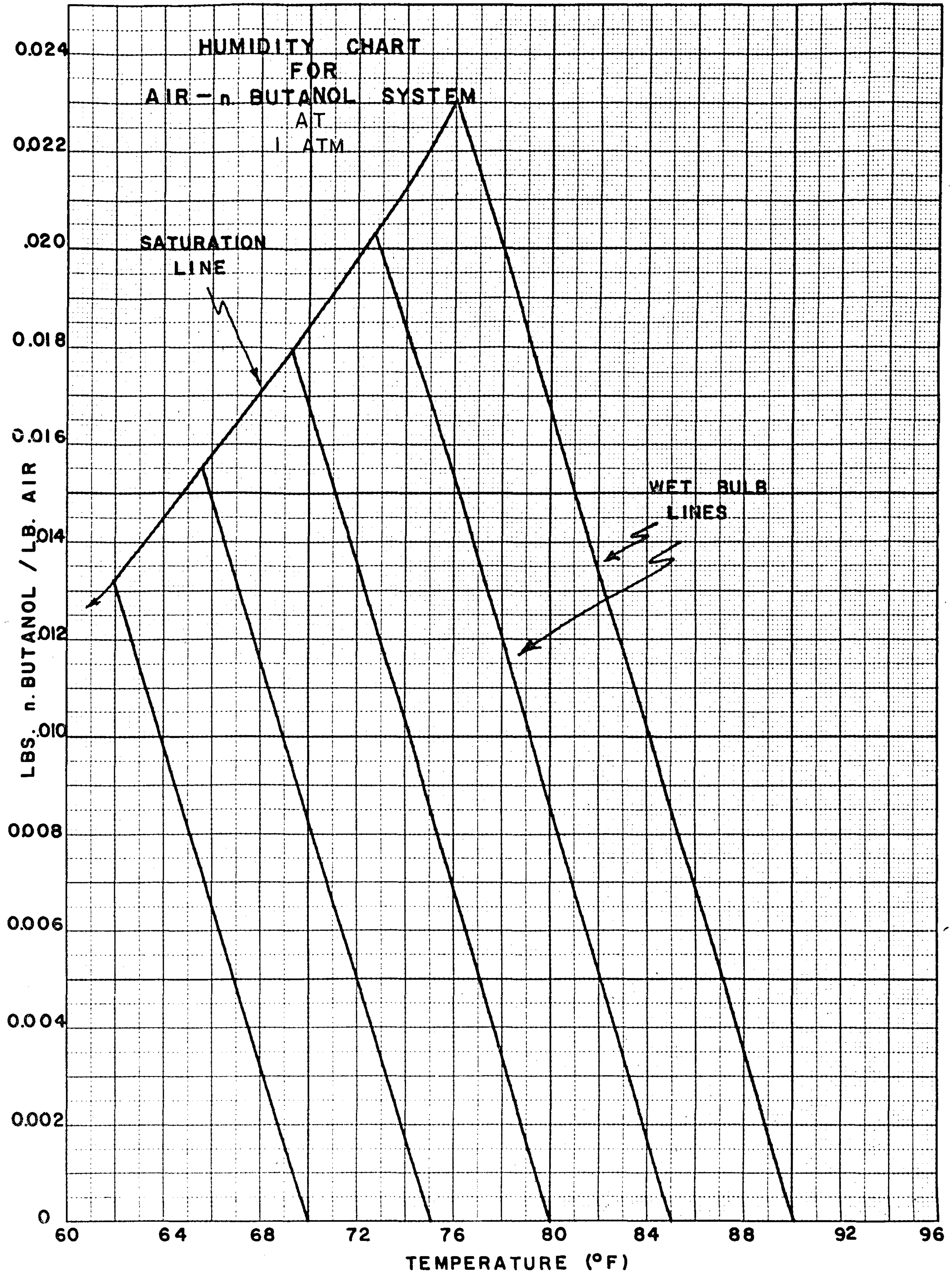

HUMIDITY CHART
FOR
AIR – n. BUTANOL SYSTEM
AT
1 ATM
SATURATION LINE
WET BULB LINES
LBS. n. BUTANOL / LB. AIR
0.024
0.022
.020
0.018
0.016
.014
.012
.010
0.008
0.006
0.004
0.002
0
60
64
68
72
76
80
84
88
92
96
TEMPERATURE (°F)

TABLE I

SPECIFICATIONS OF SOLVENTS USED

Methyl Isobutyl Ketone (Pract. P5617 209531)

Manufacturer: Matheson, Coleman & Bell

Purity: 99% by weight.

Specific Gravity 20°/20°C: 0.800 - 0.804

Color: Maximum 15 platinum cobalt (Hazen) standard.

Water: Miscible w/o turbidity w/19 volumes 60° Bé gasoline @ 20° C.

Acidity: Less than 0.01% HA_c

Distillation Range: 114.0 - 117.0°C.

Acetone (two lots)	(1)	(2)
Manufacturer:	Matheson, Coleman & Bell	Merck
Grade:	ACS Reagent	ACS Reagent
Specific Gravity 25°/25°C:	0.788	0.788
Water:	--	0.40%
Acidity:	0.003%	0.002%
Distillation Range:	0.5°C	95% within °.5°C 55.5 - 57.0°C

TABLE II

EVAPORATION OF PURE ACETONE; L = 800 Lb/Hr-Ft^2

3 March 1956

Run	23	24	25	26	27	28	29	30	31	32	33	34	35	36	37	38	39
G_0(Lbs/Hr-Ft^2)	202	202	318	318	318	555	555	555	555	454	454	388	388	126	126	126	126
Evaporation Rate (Ml./Sec)	0.407	0.395	0.612	0.504	0.567	0.829	0.772	0.847	0.888	0.729	0.699	0.660	0.632	0.298	0.300	0.316	0.323
$\overset{\Delta}{H_2} = H$(Lbs Acetone/Lb.Air)	0.134	0.130	0.129	0.106	0.119	0.0995	0.0927	0.102	0.107	0.107	0.103	0.113	0.109	0.158	0.159	0.167	0.171
P_2 (Atm.)	0.0630	0.0613	0.0604	0.0503	0.0563	0.0475	0.0443	0.0484	0.0507	0.0508	0.0488	0.0537	0.0515	0.0732	0.0737	0.0772	0.0789
T_2(Gas Out - °C)	15.3	15.2	13.2	13.1	12.8	9.6	9.2	9.1	9.1	9.7	9.6	10.2	10.4	12.8	13.2	14.0	14.7
T_2-WB (°C)	-1.1	-1.4	-2.0	-4.0	-2.8	-5.3	-6.0	-5.3	-4.8	-4.6	-5.0	-4.0	-4.4	0.1	0.3	0.8	1.1
t_2(Liquid In - °C)	17.3	17.3	18.6	18.5	18.3	17.9	17.7	17.2	17.0	18.0	17.7	17.8	17.9	17.8	17.9	18.7	19.4
P_2-WB (Atm.)	0.0828	0.0815	0.0791	0.0717	0.0760	0.0663	0.0638	0.0663	0.0678	0.0684	0.0673	0.0712	0.0696	0.0885	0.0893	0.0927	0.0949
T_1 (Gas In - °C)	24.2	24.2	24.2	24.2	24.2	24.4	24.4	24.5	24.6	24.8	24.9	25.0	25.0	24.8	24.8	24.7	24.7
T_1-WB. (°C)	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.9	-10.8	-10.7	-10.6	-10.6	-10.4	-10.4	-10.5	-10.5	-10.7	-10.7
t_1 (Liquid Out - °C)	8.6	8.5	7.0	6.9	7.0	2.5	2.3	2.1	2.1	4.3	4.6	5.3	5.4	10.3	10.4	11.3	11.9
P_1-WB. (Atm.)	0.0477	0.0477	0.0477	0.0477	0.0477	0.0479	0.0479	0.0480	0.0482	0.0485	0.0486	0.0487	0.0487	0.0485	0.0485	0.0483	0.0483
k_Ga (Lb.Moles/Hr-Ft^3-Atm.)	37.2	35.7	57.7	44.9	52.2	77.3	77.1	80.3	86.2	69.4	65.2	62.8	59.4	30.1	30.1	31.9	32.1
Av. k_Ga	36.5		51.6			78.7				67.3		61.1		31.1			
P_{2e} (Atm.)	0.2065	0.2065	0.2200	0.2190	0.2170	0.2125	0.2100	0.2060	0.2040	0.2.35	0.2100	0.2110	0.2120	0.2110	0.2120	0.2210	0.2285
P_{1e} (Atm.)	0.1365	0.1355	0.1255	0.1245	0.1255	0.1000	0.0990	0.0975	0.0975	0.1093	0.1110	0.1150	0.1155	0.1478	0.1483	0.1550	0.1600
K_Ga (Lb-Moles/Hr-Ft^3 -Atm.)	8.5	8.3	12.6	10.2	11.6	18.7	17.4	19.7	21.1	15.9	15.3	14.3	13.5	6.1	6.1	6.1	6.1
Av. K_Ga	8.4		11.5			19.2				15.6		13.9		6.1			

TABLE III

EVAPORATION OF PURE ACETONE; L=1200 Lb/hr-Ft^2

17 March 1956

Run	40	41	42	43	44	45	46	47
G_o	368	368	368	368	494	494	494	494
Evaporation Rate	0.817	0.843	0.938	0.907	1.24	1.14	1.22	1.20
$H_2 - \Delta H$	0.148	0.153	0.170	0.164	0.167	0.154	0.165	0.162
P_2	0.0689	0.0709	0.0784	0.0760	0.0772	0.0706	0.0761	0.0752
T_2	10.2	9.9	10.0	10.0	8.5	7.3	6.9	6.6
$T_{2\ WB}$	-1.0	-1.1	0.2	0.0	-0.1	-1.3	-0.5	-0.8
t_2	18.5	18.4	18.2	19.2	18.6	18.8	18.5	18.1
P_{2WB}	0.0832	0.0826	0.0889	0.0879	0.0872	0.0818	0.0851	0.0838
T_1	23.2	23.2	23.2	23.2	23.2	23.4	23.5	23.6
$T_{1\ WB}$	-11.3	-11.3	-11.3	-11.3	-11.3	-11.2	-11.2	-11.0
t_1	6.3	5.8	5.4	6.1	1.6	2.6	1.4	1.2
$P_{1\ WB}$	0.0466	0.0466	0.0464	0.0464	0.0465	0.0467	0.0470	0.0472
k_Ga	87.0	96.8	113.0	103.7	151.3	133.1	154.2	156.1
Av.	100.1				151.7			
P_{2e}	0.2200	0.2180	0.2155	0.2265	0.2220	0.2230	0.2205	0.2160
P_{1e}	0.1210	0.1185	0.1155	0.1200	0.0960	0.1020	0.0955	0.0945
K_Ga	17.4	18.4	21.6	19.5	28.9	26.1	30.0	29.9
Av.	19.2				29.4			

Run	48	49	50	51	52	53	54	55
G_o	494	429	429	313	313	313	222	222
Evaporation Rate	1.25	1.12	1.12	0.859	0.848	0.895	0.726	0.711
$H_2 - \Delta H$	0.169	0.173	0.174	0.183	0.181	0.191	0.218	0.214
P_2	0.0778	0.0798	0.0802	0.0838	0.0829	0.0870	0.0983	0.0966
T_2	6.7	8.0	8.0	8.9	9.3	9.3	10.6	10.8
$T_{2\ WB}$	-0.3	0.3	0.3	0.9	0.8	1.4	3.4	3.2
t_2	17.8	19.4	19.1	18.8	18.9	18.8	18.6	18.7
P_{2WB}	0.0859	0.0890	0.0893	0.0932	0.0930	0.0965	0.1065	0.1049
T_1	23.6	23.8	23.8	23.8	23.8	23.8	23.7	23.7
$T_{1\ WB}$	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0
t_1	1.1	3.4	3.3	5.7	6.0	6.1	7.9	8.0
$P_{1\ WB}$	0.0472	0.0473	0.0473	0.0473	0.0473	0.0473	0.0472	0.0472
k_Ga	163.7	138.5	147	106.2	102	110.2	94.5	92.0
Av.		143.2		106.1			92.4	
P_{2e}	0.2130	0.2310	0.2265	0.2230	0.2250	0.2230	0.2220	0.2230
P_{1e}	0.0940	0.1065	0.1060	0.1200	0.1220	0.1225	0.1340	0.1345
K_Ga	31.9	25.4	25.9	19.2	18.7	20.0	16.0	15.3
Av.		25.6		19.6			15.6	

Run	56	57	58	59	60	61	62	63
G_o	222	182	182	182	124	124	124	124
Evaporation Rate	0.712	0.624	0.594	0.588	0.480	0.503	0.477	0.477
$H_2 - \Delta H$	0.214	0.230	0.219	0.217	0.259	0.271	0.257	0.257
P_2	0.0967	0.103	0.0986	0.0978	0.115	0.119	0.114	0.114
T_2	10.8	12.3	12.4	12.4	13.9	14.2	14.3	14.4
$T_{2\ WB}$	3.2	4.5	3.8	3.5	6.3	7.0	6.3	6.3
t_2	18.8	19.0	19.1	19.1	19.2	19.4	19.4	19.4
P_{2WB}	0.1054	0.1122	0.1082	0.1072	0.1229	0.1270	0.1229	0.1229
T_1	23.7	23.5	23.4	23.4	23.3	23.3	23.3	23.3
$T_{1\ WB}$	-11.0	-11.0	-11.0	-11.0	-11.3	-11.3	-11.3	-11.3
t_1	8.1	10.1	10.1	10.2	11.7	11.8	12.0	12.0
$P_{1\ WB}$	0.0472	0.0471	0.0471	0.0471	0.0466	0.0466	0.0466	0.0466
k_Ga	90.7	70.8	73.2	72.9	62.8	67.7	60.5	60.5
Av.		72.3			62.9			
P_{2e}	0.2230	0.2270	0.2275	0.2265	0.2280	0.2310	0.2320	0.2320
P_{1e}	0.1350	0.1485	0.1490	0.1495	0.1600	0.1610	0.1620	0.1620
K_Ga	15.3	13.1	12.3	12.1	10.3	10.7	10.0	10.0
Av.		12.5			10.2			

TABLE III (Cont'd)

Run	51	52	53
Calculation of k_La by Wet-bulb Assumption			
t_f-Av(^oC)	3.6	3.7	3.8
$P^o{}_f$-Av (AT.)	0.1075	0.1080	0.1085
$\mathcal{H}_f$-Av	7.92	7.88	7.85
k_La	2.95	2.93	3.00
Av.	2.96		
Calculation of k_G^xa and k_L^xa at $t_i=\frac{1}{2}(t_L\ T_{WB})$			
t_{2i}	9.8	9.8	10.1
p_{2i}	0.1470	0.1470	0.1490
t_{1i}	-2.7	-2.5	-2.5
p_{1i}	0.0755	0.0765	0.1490
k_G^xa	36.0	35.2	37.5
Av.		36.2	
T_f av.	7.9	8.1	8.1
$P^o{}_f$ av.	0.134	0.136	0.136
$\mathcal{H}_f$ av.	6.36	6.27	6.27
k_L^xa	6.48	6.36	6.84
Av.		6.56	

TABLE IV

EVAPORATION OF PURE ACETONE; L=400 Lb/Hr-Ft^2

7 April 1956

Run	64	65	66	67	68	69	70	71	72	73	74	75	76	77	78	79	80	81	82	83	84	85
G_o	212	212	212	212	212	212	212	126	126	126	126	126	126	388	388	388	280	280	280	504	504	504
Evaporation Rate	.409	.414	.408	.416	.403	.416	.395	.295	.299	.301	.293	.295	.294	.549	.555	.547	.466	.453	.454	.606	.617	.587
$H_2=\Delta H$	.129	.130	.128	.131	.126	.131	.124	.155	.158	.159	.155	.156	.156	.0944	.0955	.0940	.111	.108	.108	.0802	.0816	.0777
P_2	0.0605	0.0612	0.0604	0.0616	0.0593	0.0616	0.0586	0.0721	.0734	.0738	.0721	.0726	.0724	.0451	.0456	.0449	.0528	.0512	.0513	.0386	.0392	.0374
T_2	11.5	11.6	10.9	10.6	10.4	10.2	10.2	11.9	12.1	12.1	12.3	12.4	12.3	7.9	7.9	7.8	8.7	8.8	8.8	6.8	6.8	6.7
T_{2-WB}	-2.4	-2.4	-2.7	-2.5	-2.9	-2.5	-3.1	-.3	0	.1	-.2	-.1	-1.	-6.2	-6.1	-6.2	-4.3	-4.8	-4.7	-7.8	-7.7	-8.1
t_2	20.0	20.1	20.1	20.0	19.9	19.8	19.8	19.4	19.4	19.3	19.3	19.3	19.4	22.0	22.2	22.2	22.2	22.2	22.2	21.8	21.8	21.7
P_{2WB}	0.0771	0.0774	0.0762	0.0768	0.0755	0.0767	0.0747	0.0859	0.0877	0.0882	0.0868	0.0871	0.0869	0.0632	0.0635	0.0631	0.0699	0.0682	.0683	.0577	.0580	.0565
T_1	24.0	24.1	24.0	24.0	24.0	24.0	24.0	24.0	24.0	24.0	24.0	24.0	24.1	24.0	24.0	24.1	24.1	24.1	24.1	24.5	24.7	24.7
T_{1-WB}	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.8	-10.7	-10.7
t_1	4.2	4.0	3.8	3.7	3.5	3.3	3.2	4.7	4.9	5.1	5.1	5.2	5.2	0.3	0.4	0.4	1.0	1.2	1.2	-0.6	-.8	-.7
P_{1WB}	0.0474	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0476	0.0477	0.0477	.0477	.0478	.0479	.0479
k_Ga	40.6	41.3	42.4	43.9	40.0	42.7	39.7	31.2	31.4	31.5	30.5	30.8	30.7	50.8	53.1	52.0	45.4	44.2	44.3	56.5	57.8	54.7
Av.	41.5							31.1						52.0			44.6			56.3		
P_{2e}	.238	.237	.237	.236	.234	.233	.233	.230	.229	.229	.229	.229	.229	.259	.262	.263	.263	.262	.262	.257	.257	.256
P_{1e}	.111	.107	.107	.106	.105	.104	.103	.112	.113	.114	.114	.114	.114	.0900	.0900	.0900	.0923	.0932	.0937	.0854	.0847	.0850
K_Ga	8.37	8.69	8.56	8.84	8.53	9.02	8.52	6.42	6.56	6.60	6.38	6.42	6.39	11.2	11.2	11.0	9.46	9.13	9.10	12.5	12.7	12.1
Av.	8.65							6.46						11.2			9.23			12.5		

TABLE V

EVAPORATION OF PURE ACETONE; L=1600 Lb/Hr-Ft^2

7 April 1956

Run	86	87	88	89	90	91	92	93	94	95	96	97	98	99
G_o	268	268	268	113	113	113	113	184	184	184	184	363	363	363
Evaporation Rate	1.16	1.14	1.17	0.649	0.652	0.655	0.617	0.924	1.10	1.05	0.963	1.53	1.29	1.45
$H_2=\Delta H$	0.290	0.283	0.290	0.382	0.390	0.385	0.363	0.337	0.401	0.382	0.351	0.281	0.237	0.267
P_2	0.127	0.124	0.127	0.160	0.163	0.161	0.154	0.144	0.167	0.160	0.149	0.123	0.106	0.118
T_2	14.1	14.3	14.3	20.0	21.8	21.8	21.9	17.6	18.1	18.0	17.3	13.4	12.9	13.3
T_{2WB}	8.0	7.6	8.0	12.6	13.1	12.9	12.1	10.6	13.1	12.4	11.1	7.4	5.1	6.6
t_2	20.2	20.7	20.8	21.9	21.7	21.3	20.9	25.0	24.0	22.8	21.4	20.8	20.9	22.5
P_{2WB}	0.1328	0.1308	0.1331	0.1676	0.1720	0.1702	0.1643	0.1515	0.1722	0.1662	0.1556	0.1292	0.1148	0.1250
T_1	24.3	24.3	24.3	24.3	24.4	24.4	24.4	24.4	24.5	24.5	24.5	24.6	24.6	24.7
T_{1WB}	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7	-10.7
t_1	7.9	7.9	7.9	14.2	14.4	14.1	13.9	11.7	11.8	11.5	10.8	5.3	5.2	5.7
P_{1WB}	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479	0.0479
k_Ga	352	339	349	194	197	194	183	278	341	321	293	474	384	435
Av.	347			192				308				431		
P_{2e}	0.241	0.246	0.249	0.261	0.258	0.255	0.250	0.300	0.289	0.274	0.256	0.248	0.251	0.268
P_{1e}	0.134	0.134	0.134	0.179	0.181	0.179	0.177	0.161	0.162	0.159	0.155	0.118	0.117	0.120
K_Ga	27.2	25.7	26.4	13.9	14.4	14.4	13.5	17.0	22.7	22.5	21.5	37.4	28.7	31.2
Av.	26.4			14.1				20.9				32.4		

TABLE VI

EVAPORATION OF PURE METHYL ISOBUTYL KETONE; L=1200 Lb/Hr-Ft^2

20 April 1956

Run	100	101	102	103	104	105	106	107	108	109	110	111	112	113	114
G_o (Lb/Hr-Ft^2)	297	297	297	388	388	388	484	484	484	556	556	556	202	202	202
Evaporation Rate (Ml/Sec)	0.230	0.217	0.212	0.253	0.261	0.248	0.367	0.339	0.375	0.400	0.438	0.428	0.125	0.142	0.134
$H_2 = \Delta H$(Lb.MIK/Lb.Air)	0.0525	0.0497	0.0484	0.0444	0.0458	0.0435	0.0516	0.0476	0.0527	0.0489	0.0536	0.0524	0.0421	0.0476	0.0450
P_2 (At.)	0.0150	0.0142	0.0138	0.0127	0.0131	0.0125	0.0148	0.0136	0.0150	0.0140	0.0153	0.0149	0.0121	0.0136	0.0129
T_2 (Gas Out- °C)	19.9	20.1	20.1	19.5	19.4	19.5	19.9	19.5	19.4	18.8	18.9	18.6	19.0	19.3	19.5
T_2 - WB (°C)	17.8	17.6	17.5	16.6	16.7	16.5	17.7	17.1	17.5	16.7	17.4	16.9	16.1	16.8	16.6
t_2 (Liquid In-°C)	21.4	21.4	21.4	21.6	21.7	21.7	21.3	21.1	21.0	20.7	20.5	20.4	20.6	21.1	21.3
$P_{2\,WB}$ (At.)	0.0184	0.0182	0.0181	0.0173	0.0174	0.0172	0.0183	0.0177	0.0182	0.0174	0.0180	0.0176	0.0168	0.0175	0.0173
T_1 (Gas In-°C)	22.9	22.9	22.8	22.1	22.1	22.1	23.3	23.5	23.7	24.1	24.2	24.4	24.3	24.3	24.3
T_1 - WB (°C)	13.9	13.9	13.8	13.5	13.4	13.4	14.2	14.2	14.4	14.6	14.6	14.9	14.8	14.8	14.8
t_1 (Liquid Out-°C)	19.1	19.1	19.2	19.1	19.1	19.1	18.7	18.5	18.4	17.4	17.3	17.0	19.4	19.8	20.0
$P_{1\,WB}$ (At.)	0.0149	0.0149	0.0148	0.0146	0.0145	0.0145	0.0151	0.0152	0.0153	0.0155	0.0155	0.0157	0.0157	0.0157	0.0157
k_Ga (Lb.Mol/Hr-Ft^2 At)	50.4	45.1	42.9	58.2	53.7	49.0	79.3	68.9	84.3	86.2	103.6	101.2	21.5	29.1	28.1
k_Ga Av.	46.1			53.6			77.5			97.0			26.2		
P_{2_e} (At.)	0.0224	0.0224	0.0224	0.0227	0.0228	0.0228	0.0223	0.0221	0.0220	0.0217	0.0215	0.0214	0.0216	0.0221	0.0224
P_{1_e} (At.)	0.0199	0.0200	0.0201	0.0200	0.0200	0.0200	0.0196	0.0194	0.0192	0.0183	0.0182	0.0179	0.0202	0.0207	0.0209
K_Ga (Lb.Mol/Hr-Ft^3-At.)	31.1	28.2	27.1	30.3	31.8	29.4	50.6	44.2	53.6	56.4	67.0	65.9	15.2	17.8	16.0
K_Ga Av.	28.8			30.5			49.5			63.1			16.3		

TABLE VII

EVAPORATION OF PURE METHYL ISOBUTYL KETONE; $L=1600$ Lb/Hr-Ft^2

20 April 1956

Run	115	116	117	118	119	120	121	122	123	124	125	126
G_o	202	202	202	388	388	388	297	297	297	484	484	484
Evaporation Rate	0.171	0.185	0.180	0.303	0.288	0.284	0.238	0.239	0.247	0.407	0.427	0.391
$H_2=\Delta H$	0.0574	0.0621	0.0607	0.0531	0.0505	0.0497	0.0581	0.0583	0.0602	0.0572	0.0600	0.0550
p_2	0.0164	0.0177	0.0173	0.0152	0.0144	0.0142	0.0166	0.0166	0.0172	0.0163	0.0171	0.0157
T_2	20.6	20.9	21.0	20.6	20.6	20.5	20.6	20.7	20.7	20.7	20.6	20.5
T_2-WB	18.6	19.3	19.1	18.1	17.9	17.8	18.7	18.8	19.0	18.6	18.9	17.4
t_2	21.7	21.8	21.9	21.9	21.8	21.8	21.7	21.7	21.7	21.6	21.4	21.4
p_{2WB}	0.0192	0.0199	0.0197	0.0187	0.0185	0.0184	0.0193	0.0193	0.0195	0.0192	0.0195	0.0180
T_1	24.1	24.5	24.5	25.0	25.0	25.1	25.1	25.1	25.2	25.5	25.6	25.7
T_1-WB	14.9	14.9	14.9	15.2	15.2	15.3	15.4	15.4	15.4	15.6	15.7	15.7
t_1	20.1	20.5	20.7	20.0	19.9	19.9	20.0	20.1	20.2	19.1	19.0	18.9
p_{1WB}	0.0157	0.0158	0.0158	0.0160	0.0160	0.0161	0.0162	0.0162	0.0162	0.0164	0.0165	0.0165
k_Ga	39.0	39.3	43.4	63.1	56.8	55.2	58.1	58.2	63.1	90.0	100.8	93.5
Av.	40.6			58.4			59.8			94.8		
p_{2e}	0.0228	0.0229	0.0230	0.0230	0.0229	0.0228	0.0228	0.0228	0.0228	0.0226	0.0224	0.0224
p_{1e}	0.0214	0.0215	0.0216	0.0209	0.0208	0.0208	0.0209	0.0211	0.0211	0.0199	0.0198	0.0198
K_Ga	23.7	27.7	25.9	39.2	36.1	35.2	36.1	36.1	38.5	59.4	66.4	55.8
Av.	25.8			36.8			36.9			60.5		

TABLE VIII

EVAPORATION OF PURE METHYL ISOBUTYL KETONE; L=800 Lb/Hr-Ft^2

20 April 1956

Run	127	128	129	130	131	132	133	134	135	136	137	138	139	140	141
G_o	484	484	484	388	388	388	591	591	591	222	222	222	646	646	646
Evaporation Rate	0.209	0.207	0.223	0.164	0.149	0.159	0.309	0.281	0.289	0.0947	0.0987	0.0893	0.434	0.420	0.410
$H_2=\Delta H$	0.0294	0.0291	0.0313	0.0288	0.0261	0.0278	0.0355	0.0323	0.0332	0.0290	0.0302	0.0273	0.0457	0.0442	0.0431
P_2	0.00846	0.00838	0.00900	0.00828	0.00751	0.00801	0.0102	0.00928	0.00959	0.00835	0.00868	0.00786	0.0131	0.0127	0.0124
T_2	19.7	20.0	20.2	20.6	20.7	20.4	20.7	20.6	20.5	21.0	21.1	21.3	19.6	19.7	19.7
T_2- WB	15.2	15.4	15.8	15.8	15.5	15.4	16.4	16.1	16.2	16.0	16.0	15.9	16.8	16.8	16.7
t_2	22.1	22.4	22.6	22.9	22.9	23.1	23.4	23.4	23.3	23.4	23.5	23.7	24.0	24.0	24.0
P_{2WB}	0.0160	0.0162	0.0165	0.0165	0.0163	0.0162	0.0171	0.0168	0.0169	0.0167	0.0167	0.0166	0.0175	0.0174	0.0173
T_1	26.0	26.0	26.0	26.0	26.0	26.0	26.1	26.2	26.2	25.8	25.7	25.7	26.1	26.1	26.2
T_1- WB	15.9	15.9	15.9	15.9	15.9	15.9	16.0	16.1	16.1	15.8	15.7	15.7	15.9	16.0	16.1
t_1	19.7	19.9	20.0	20.7	20.9	21.0	20.0	20.1	20.0	21.6	21.7	22.0	19.2	19.1	19.1
P_{1WB}	0.0166	0.0166	0.0166	0.0166	0.0166	0.0166	0.0168	0.0168	0.0168	0.0166	0.0165	0.0165	0.0167	0.0168	0.0168
k_Ga	31.4	30.6	33.4	23.6	20.9	22.9	47.7	41.8	43.5	13.6	14.4	12.6	81.0	75.8	72.7
Av.	31.8			22.5			44.3			13.5			76.5		
P_{2e}	0.0233	0.0236	0.0238	0.0242	0.0242	0.0244	0.0248	0.0248	0.0247	0.0248	0.0250	0.0252	0.0255	0.0256	0.0255
P_{1e}	0.0206	0.0208	0.0209	0.0217	0.0219	0.0220	0.0209	0.0211	0.0209	0.0227	0.0228	0.0231	0.0201	0.0199	0.0200
K_Ga	20.6	20.0	21.8	15.2	13.4	14.4	30.3	26.6	27.9	8.42	8.78	7.69	47.6	44.7	43.2
Av.	20.8			14.3			28.3			8.30			45.2		

TABLE IX

EVAPORATION OF ACETONE - MIK MIXTURES ; L=1200 Lb/Hr-Ft^2; G_o=313 Lb/Hr-Ft^2

Runs 142-155, 21 April 1956; Runs 174-179, 5 May 1956

Run	142	143	144	145	146	147	148	149	150	151	152	153	154	155	174	175	176	177	178	179
Vol. % Acetone Feed	40	40	40	60	60	60	80	80	80	90	90	90	90	90	95	95	95	92.5	92.5	92.5
Evaporation Rate (Ml./Sec.)	0.324	0.300	0.298	0.343	0.318	0.304	0.414	0.441	0.427	0.459	0.435	0.445	0.457	0.447	0.739	0.716	0.696	0.664	0.649	0.663
p_2- Acetone (Atm.)	0.01384	0.01286	0.01278	0.02190	0.0203	0.0193	0.03485	0.0369	0.0358	0.0432	0.0411	0.0419	0.0430	0.0421	0.0713	0.0692	0.0676	0.0629	0.0616	0.0627
T_2 (Air Out °C)	18.6	18.6	18.7	17.8	17.8	17.8	16.1	16.1	16.1	14.8	14.7	14.7	14.6	14.6	10.9	10.9	10.7	13.2	13.1	13.0
T_1 (Air In - °C)	25.2	25.2	25.2	25.4	25.4	25.5	25.7	25.7	25.7	25.5	25.5	25.5	25.5	25.5	27.2	27.1	27.0	25.5	25.5	25.5
Vol. % Acetone @ 2 (Liquid)	0.075	0.075	0.075	0.127	0.127	0.127	0.260	0.260	0.260	0.420	0.420	0.420	0.420	0.420	0.860	0.860	0.860	0.705	0.705	0.705
x_2 (Mol. Fraction)	0.1216	0.1216	0.1216	0.196	0.196	0.196	0.375	0.375	0.375	0.552	0.552	0.552	0.552	0.552	0.912	0.912	0.912	0.804	0.804	0.804
t_2 (Liquid In °C)	20.7	20.7	20.8	20.4	20.3	20.3	20.1	20.1	20.2	19.1	19.1	19.1	19.1	19.2	19.2	19.2	19.1	20.1	19.9	19.8
P_2 (Atm.)	0.2455	0.2460	0.2470	0.2430	0.2400	0.2400	0.2380	0.2390	0.2390	0.2265	0.2265	0.2265	0.2265	0.2280	0.2290	0.2280	0.2265	0.2390	0.2360	0.2350
p_2^* (Atm)	0.0298	0.0299	0.0300	0.0476	0.0470	0.0470	0.0892	0.0896	0.0896	0.1250	0.1250	0.1250	0.1250	0.1258	0.2078	0.2078	0.2065	0.1685	0.1662	0.1657
x_1 (Mol Fraction)	0.1118	0.1127	0.1129	0.1853	0.1867	0.1868	0.3600	0.3595	0.3595	0.5400	0.5400	0.5400	0.5400	0.5400	0.911	0.909	0.909	0.798	0.798	0.798
t_1 (Liquid Out °C)	17.6	17.8	17.8	16.6	16.6	16.5	14.6	14.7	14.7	11.8	12.0	12.1	12.2	12.2	8.5	8.8	8.7	9.9	9.7	9.7
p_1^o (Atm)	0.2105	0.2110	0.212	0.201	0.201	0.204	0.183	0.183	0.1835	0.161	0.163	0.163	0.164	0.1645	0.138	0.1400	0.1395	0.1480	0.1470	0.1460
p_1^* (Atm)	0.02355	0.02380	0.02390	0.0372	0.03755	0.03810	0.08410	0.08400	0.0843	0.0869	0.0881	0.0881	0.0886	0.0888	0.1257	0.1273	0.1268	0.1182	0.1173	0.1165
K_Ga (Mol/Hr-Ft^3-Atm)	19.3	17.4	17.1	19.2	15.5	16.3	14.0	15.0	14.3	13.9	12.9	13.8	13.9	13.6	14.6	14.0	13.8	15.4	15.3	15.7
Av.																				

TABLE IX (Cont'd.)

RECOMPUTATION ASSUMING INTERFACE IS AT WET-BULB TEMPERATURE ($k_G a$=106.1)

Run	142	143	144	145	146	147	148	149	150	151	152	153	154	155	174	175	176	177	178	179
T_{2-WB} (°C)	12.5	12.2	12.2	9.6	9.4	9.1	4.7	5.0	4.8	1.2	0.9	1.1	1.2	1.1	-0.3	-0.56	-0.77	-0.9	-1.1	-0.9
T_{1-WB} (°C)	12.4	12.4	12.4	10.8	10.8	10.7	3.6	3.7	3.7	1.6	1.6	1.7	1.6	1.6	-7.8	-7.8	-7.8	-5.3	-5.3	-5.3
$t_{f\text{-av.}}$ (°C)	15.8	15.8	15.8	14.4	14.3	14.2	10.8	10.9	10.9	8.4	8.4	8.5	8.5	8.5	4.90	4.91	4.81	6.0	5.8	5.8
$p_{f\text{-av.}}$ (Atm)	0.194	0.194	0.194	0.181	0.181	0.180	0.154	0.155	0.155	0.137	0.137	0.138	0.138	0.138	0.115	0.115	0.1145	0.122	0.121	0.121
$\frac{K_G a}{k_G a}$	0.182	0.164	0.161	0.181	0.146	0.154	0.132	0.141	0.135	0.131	0.122	0.130	0.131	0.128	0.1376	0.1319	0.1300	0.145	0.144	0.148
$p^*_{f\text{-av.}}$ (Atm)	0.0188	0.0183	0.0183	0.0295	0.0286	0.0287	0.0566	0.0574	0.0571	0.0694	0.0678	0.0694	0.0697	0.0697	0.1041	0.1039	0.1035	0.0956	0.0943	0.0944
$x_{f\text{-av}}$ (Mol Fraction)	0.0958	0.0943	0.0943	0.163	0.158	0.159	0.367	0.371	0.368	0.506	0.495	0.502	0.505	0.505	0.905	0.904	0.903	0.784	0.780	0.781
$v_{m\text{-}f}$ Av. (Ft /Mol)	1.93	1.93	1.93	1.87	1.87	1.87	1.70	1.70	1.70	1.59	1.59	1.59	1.59	1.59	1.25	1.25	1.25	1.36	1.36	1.36
$\mathcal{H}_f$ av. (Mol/Ft^3- Atm)	2.67	2.67	2.67	2.95	2.95	2.97	3.82	3.80	3.80	4.59	4.59	4.56	4.56	4.56	6.95	6.95	6.99	6.02	6.07	6.07
$k_L a$ (Ft^3 Acetone/Ft^3 Hr)	8.84	7.80	7.63	7.93	6.14	6.47	4.22	4.59	4.34	3.49	3.20	3.48	3.51	3.42	2.44	2.32	2.29	3.05	2.95	3.09
Av	8.09			6.85			4.37			3.42					2.34			3.03		

TABLE IX (Cont'd.)

CALCULATION OF k^x_La BY ASSUMING INTERFACE HALF WAY BETWEEN LIQUID AND WET-BULB

Run	142	143	144	145	146	147	148	149	150	151	152	153	154	155	174	175	176	177	178	179
t_{1_i} (°C)	15.0	15.1	15.1	13.7	13.7	13.6	9.1	9.2	9.2	6.7	6.8	6.9	6.9	6.9	0.5	0.7	0.6	2.3	2.2	2.2
t_{2_i} (°C)	16.6	16.5	16.5	15.0	14.9	14.7	12.4	12.6	12.5	10.2	10.0	10.1	10.2	10.2	8.5	8.4	8.2	9.6	9.4	9.4
$t_{f\text{-av.}}$(°C)	17.5	17.5	17.6	16.4	16.4	16.3	14.1	14.2	14.5	12.0	12.0	12.1	12.1	12.1	9.2	9.3	9.2	10.5	10.3	10.3
p^o_f-av.(Atm)	0.209	0.209	0.210	0.197	0.197	0.196	0.178	0.179	0.180	0.160	0.160	0.162	0.162	0.162	0.140	0.140	0.140	0.149	0.147	0.147
$\frac{K_Ga}{K_Ga}=\frac{(p_i-P)}{(p^*-P)}$ av.	0.529	0.477	0.469	0.526	0.425	0.447	0.384	[illegible]	0.392	0.381	0.354	0.378	0.381	0.373	0.384	0.370	0.359	0.422	0.419	0.431
$p^*_{f\text{-av.}}$(Atm)	0.02355	0.02380	0.02390	0.0372	0.03755	0.0381	0.0841	0.0840	0.0843	0.0869	0.0881	0.0881	0.0886	0.0888	0.1242	0.1259	0.1255	0.1182	0.1173	0.1165
$x_{f\text{-av.}}$(Mol Fraction)	0.113	0.114	0.114	0.189	0.191	0.194	0.473	0.469	0.468	0.543	0.551	0.544	0.547	0.549	0.889	0.894	0.898	0.793	0.798	0.793
$v_{m\text{-}f}$ av(Ft^3/Mol)	1.915	1.910	1.910	1.85	1.845	1.841	1.612	1.616	1.616	1.555	1.547	1.552	1.550	1.549	1.264	1.260	1.258	1.345	1.340	1.345
$\mathcal{H}_{f\text{-av.}}$(Mol/Ft^3-Atm)	2.51	2.51	2.49	2.75	2.75	2.77	3.49	3.46	3.44	4.02	4.04	3.98	3.99	3.99	5.65	5.63	5.68	4.99	5.08	5.06
k^x_La (Ft^3 Acetone/Ft^3-Hr)	13.6	13.3	12.9	14.7	9.8	10.6	6.52	7.37	6.85	5.60	4.94	5.57	5.64	5.43	4.07	3.72	3.60	5.35	5.19	5.60
Av.	13.3			11.7			6.91			5.44					3.80			5.38		

TABLE IX (Cont'd.)

RECOMPUTATION BY METHOD OF STUTZMAN ($K_Ga=k_Ga$ FOR PURE COMPONENT=19.6)

Run	142	143	144	145	146	147	148	149	150	151	152	153	154	155	174	175	176	177	178	179
$p^o_{av.}$ (Atm)	0.2280	0.2285	0.2295	0.2220	0.2205	0.2220	0.2105	0.2110	0.2113	0.1938	0.1948	0.1948	0.1953	0.1963	0.1835	0.1840	0.1830	0.1935	0.1915	0.1905
$\frac{K_Ga}{k_Ga}$	0.985	0.887	0.872	0.979	0.791	0.831	0.714	0.766	0.729	0.709	0.658	0.704	0.709	0.694	0.719	0.688	0.668	0.786	0.781	0.801
$x_{f-av.}$ (Mol Fraction)	0.1173	0.1124	0.1108	0.1886	0.1775	0.1790	0.3645	0.373	0.367	0.483	0.471	0.482	0.483	0.481	0.707	0.828	0.819	0.679	0.676	0.683
v_{m-f} av. (Ft^3/Mol)	1.91	1.91	1.915	1.85	1.855	1.855	1.70	1.69	1.70	1.61	1.62	1.61	1.61	1.61	1.42	1.32	1.33	1.44	1.44	1.44
$\mathcal{H}_{f-av.}$ (Mol/Ft^3-Atm)	2.29	2.29	2.28	2.43	2.44	2.43	2.79	2.80	2.79	3.21	3.17	3.19	3.18	3.17	3.83	4.12	4.11	3.59	3.63	3.65
k^x_La	62$_x$4	67.2	58.6	375	30.4	39.6	17.5	21.4	19.0	14.9	11.9	14.7	15.0	14.0	13.1	10.5	9.6	20.0	19.1	21.5
Av.	250			148			19.3			14.1					11.1			20.2		

TABLE X

EVAPORATION OF PURE ACETONE AT ELEVATED GAS TEMPERATURES; $L=1200$ Lb/Hr-Ft^2; $G_o=313$ Lb/Hr-Ft^2

5 May 1956

Run	156	157	158	159	160	161	162	163	164	165	166	167	168
Evaporation Rate (Ml./Sec)	0.672	0.716	0.752	0.790	0.848	0.832	0.768	0.783	0.791	0.748	0.709	0.691	0.639
$H_2=\Delta H$	0.143	0.153	0.160	0.168	0.181	0.177	0.164	0.167	0.169	0.160	0.151	0.147	0.136
P_2	0.0668	0.0709	0.0742	0.0776	0.0828	0.0816	0.0756	0.0770	0.0777	0.0739	0.0703	0.0686	0.0638
T_2	11.9	11.8	11.6	12.6	12.0	11.8	11.3	11.3	11.3	11.6	11.7	11.9	10.4
T_{2-WB}	-1.1	-0.5	0	0.7	1.2	1.1	0.2	0.3	0.4	0	-0.6	-0.8	-2.1
p_{2-WB}	0.0826	0.0846	0.0881	0.0919	0.0955	0.0947	0.0889	0.0899	0.0905	0.0879	0.0848	0.0838	0.0788
T_1	25.3	25.4	25.5	34.7	33.8	33.4	44.2	44.7	45.0	52.7	52.7	52.7	53.9
T_{1-WB}	-10.6	-10.5	-10.4	-7.8	-8.1	-8.3	-5.4	-5.2	-5.1	-3.5	-3.5	-3.5	-3.3
$p_{1-WB}=\Delta p_1$	0.0483	0.0484	0.0485	0.0575	0.0559	0.0558	0.0660	0.0666	0.0668	0.0735	0.0735	0.0735	0.0741
$\Delta p_{(LM-Gf)}$	0.0292	0.0276	0.0278	0.0311	0.0292	0.0295	0.0336	0.0328	0.0328	0.0360	0.0364	0.0371	0.0371
$k_G a$	66.8	75.5	78.6	73.8	84.3	82.0	66.4	69.4	70.1	60.5	56.7	54.1	50.0
Av. $k_G a$		73.6			80.3			68.6			57.1		
t_2	20.1	19.9	19.9	20.7	20.5	20.3	19.3	19.3	19.3	18.9	18.9	18.8	15.2
t_1	7.8	7.8	7.8	8.7	8.1	7.7	8.2	8.1	8.1	8.8	8.9	9.1	6.8
$\Delta t_{(LM-Lf)}$	19.2	19.4	19.0	18.3	17.7	17.6	16.4	16.1	16.0	15.6	15.9	16.1	13.7
Av. Δt		19.2			17.9			16.2			15.9		

TABLE XI

CORRELATION OF k_La USING DIFFUSIVITY DATA OF WILKE

Runs	142-144	145-147	148-150	151-155	174-176	177-179
t_f-av. (^{o}K)	289.0	287.5	284.1	281.7	280.0	279.1
$D_L\mu$(Acetone) x10^7 (cm^2-cp./sec)	118	117.2	116	115.0	113.2	114
$D_L\mu$(M.I.K.) x10^7(cm^2-cp/sec)	79.2	78.8	77.9	77.2	76.0	76.5
μ Acetone (cp.)	0.335	0.342	0.357	0.363	0.380	0.375
μ M.I.K.(cp.)	0.62	0.63	0.665	0.675	0.72	0.705
(1-x_f-ac) (Mol Fraction)	0.9042	0.840	0.631	0.497	0.096	0.218
(1-W_f-ac) (Wt. Fraction)	0.943	0.900	0.747	0.630	0.155	0.325
μ (cp.)	0.604	0.601	0.587	0.560	0.433	0.481
$D_L\mu$ x10^7 (cm^2-cp/sec)	82.9	84.8	91.9	96.1	109.6	105.8
M_m (Lbs/Mol)	96.3	93.7	84.7	79.0	62.03	67.2
ρ (Lbs/Ft^3)	50.45	50.4	50.23	50.1	49.5	49.68
$\mu/\rho D_L$	545	528	465	407	215	275
$(C_L^1)_f$ (Mols M.I.K./Ft^3)	0.470	0.450	0.372	0.313	0.0768	0.1602
$(C_L^1)_f$ M_m ($\mu/\rho D_L$)	24,700	22,300	14,650	10,080	1,024	2,985
$1/k_La$	0.1235	0.146	0.2285	0.2923	0.427	0.330

CORRELATION OF k_La USING DIFFUSIVITY DATA OF SCHEIBEL

Runs	142-144	145-147	148-150	151-155	174-176	177-179
t_f-av. (^{o}K)	289.0	287.5	284.1	281.7	280.0	279.1
$D_L\mu$(Acetone) x10^7 (cm^2-cp./sec)	236	235	232	230	228.5	228
$D_L\mu$(M.I.K.) x10^7(cm^2-cp/sec)	107.2	106.7	105.4	104.3	103.9	103.5
μ Acetone (cp.)						
μ M.I.K.(cp.)						
(1-x_f-ac) (Mol Fraction)						
(1-W_f-ac) (Wt. Fraction)						
μ (cp.)						
$D_L\mu$ x10^7 (cm^2-cp/sec)	119.7	127.2	152.2	167.6	216.6	200.9
M_m (Lbs/Mol)						
ρ (Lbs/Ft^3)						
$\mu/\rho D_L$	377	352	280	234	108.7	144.6
$(C_L^1)_f$ (Mols M.I.K./Ft^3)						
$(C_L^1)_f$ M_m ($\mu/\rho D_L$)	17,080	14,880	8,840	5,790	518	1,570
$1/k_La$	0.1235	0.146	0.2285	0.2923	0.427	0.330