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Effect of concentration on the mass transfer coefficient in the liquid film

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

BY

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

APPROVAL OF THESIS

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APPROVED: _____

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

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) $\quad N_A = K_G (p_G - p_e) = K_L (C_e - C_L) = k_G(p_G - p_1) = k_L(C_1 - C_L)$

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) $\quad C = \mathcal{H} p$

Eliminating C_1, C_L and p_1 from (1):

(3a) $\quad K_G(p_G - p_e) = k_G (p_G - p_1) = \mathcal{H} k_L (p_1 - p_e)$

(3b) $\quad p_1 = p_e + \dfrac{K_G(p_G - p_e)}{\mathcal{H} k_L}$

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

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

Similarly, it can be shown that:

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

Thus, it is evident that the overall resistances are the sums of the individual resistances consistent with the two-film theory; the constant, $\mathcal{H}$, 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)
$$\frac{\pi r^2 \, dP}{\frac{W \mu}{\gamma}} = \frac{d C_G}{C_G - C_i}$$

Combining (5) with a material balance:

(6) $\quad N_A = \dfrac{W \, d C_G}{P \, 2\pi r \, dX}$

and the Fanning equation for turbulent friction:

$$(7) \quad dP = \frac{f \rho u^2 \, dX}{8 r}$$

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:

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

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:

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

where:

(12) $\quad \phi_D = 1-5.9\sqrt{f} \nleq 5.9\sqrt{f} \quad \left(\dfrac{\mu_f}{\rho D}\right)$

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) $\quad k_G \cong Ku^{0.8}$

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) $\quad J_H = \dfrac{h}{c_p \mu \rho}\left(\dfrac{c_p \mu_f}{k_H}\right)^{2/3} = \tfrac{1}{2}f = J_D = \dfrac{k_G p_f M_m}{\mu \rho}\left(\dfrac{\mu_f}{\rho D}\right)^{2/3}$

(14a) $\quad k_G = \dfrac{h}{c_p \rho_f M_m}\left(\dfrac{c_p \rho D}{k_H}\right)^{2/3}$ and since $c_p \mu /k_H \cong 0.74$

for all gases.

(14b) $\quad k_G = \dfrac{h}{c_p p_f M_m}\left(\dfrac{0.74}{\mu/\rho D}\right)^{2/3}$

(14c) $\quad \dfrac{h}{k^1} = c_p\left(\dfrac{\mu/\rho D}{0.74}\right)^{2/3}$ 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) $\quad ha(T_G - T_{WB}) = k^1 a (H_{WB} - H_G)\lambda_{WB}$

Thus, by combination with (14c):

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

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:

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

with the constant, s, being approximately 0.5 and $\propto$ 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 $k_L a$ 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 $k_L a$ for butanol-hexanol-air systems in a packed tower. Their technique was to first measure a quantity which they called $k_G a$ 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 $K_G a$. Use of Equation (4) then permitted calculation of $k_L a$ which was consequently correlated with liquid concentration. Their final equation assumed the form:

$$(20) \quad \frac{1}{k_L a} = K \, (c_L^1)_f \, M_m \, \left(\frac{\mu}{\rho \, D_L} \right)$$

This can be rewritten (dimensionally) as:

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

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

$$(21) \quad k_G a = \frac{G_o S}{M_1 P V} \left[\frac{1}{\frac{1-p_{1_2}-p_{1_1}}{p_2}} \right] \ln \frac{p_{1_1}}{p_{1_2}-p_2}$$

This can be derived as follows:

From (1):

$$(22a) \quad N_A \, adV = k_G a \, (p_G - p_1) \, dV$$

From a material balance:

$$(22b) \quad \frac{G_o S}{M_s} \, dH = N_A \, adV = k_G a \, (p_G - p_1) dV$$

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

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

and that over small ranges of temperature:

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

there results:

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

However, by definition:

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

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

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

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

Combining with (22b) and integrating:

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

$$= \frac{G_o S}{M_i P V}\left[\frac{p_2-p_1}{\dfrac{(p_{1_2}-p_2) - (p_{1_1}-p_1)}{\ln\dfrac{(p_{1_2}-p_2}{p_{1_1}-p_1}}}\right]$$

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

(22j) $\quad k_G a = \frac{G_o S}{M_1 P V}\left[\frac{p_2}{(p_{1_2}-p_2-p_{1_1})}\right]\ Ln\ \frac{p_{1_2}-p_2}{p_{1_1}}$

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) $\quad K_G a = k_G a$

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 con-
ditions, 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:

$$(24) \quad K_Ga = \frac{G_oS}{M_1PV} \left[\frac{1}{1 - \frac{p_{e_2}-p_{e_1}}{p_2}} \right] \ln \frac{p_{e_1}}{p_{e_2}-p_2}$$

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_Ga, 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^0 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 tempera-
tures; 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 packed tower mounted on

2. Scale Balance and fed by an

3. Air Supply which passed into the bottom of the tower.

4. The air supply was pre-treated by passing it through a Drying Column

5. and an Air Heat Exchanger. It was measured by an

6. Air Rotameter and

7. Manometer installed just prior to the column.

8. Air stream temperatures were measured by Thermometers installed before and after the column.

9. The Liquid Supply was stored in a separatory funnel and fed into a

10. Burette installed just prior to the

11. Liquid Circulating Pump. The stream next passed into a

12. Liquid Rotameter 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 $\frac{1}{2}$-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 $\frac{1}{2}$ 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 $0.3^{\circ}C$ 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 $300-350^{\circ}F$. 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 $55^\circ C$.

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. Liquid System - 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 #s 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 tempera-

tures involved in the work and were found to agree within 0.1°C. .he 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 stoppea, 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 $k_G a$ and $K_G a$ 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 $k_G a$ 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 $k_G a$ represents an actual gas film coefficient
the liquid film coefficient could be calculated from
equation 3d. These values of $k_L a$ 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 $k_L^x a$ calculated
by assuming the interface temperature half-way between
the wet-bulb temperature and the main body liquid tempera-
ture. Figure 7 shows the variation of resistance to mass
transfer through the liquid film with the correlating
function $(C_L^1)_f M_m \left(\dfrac{\mu}{\rho D_L}\right)$. This is done only for the resis-
tances 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

$k_Q a$ 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 $K_G a$ 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):

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

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

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

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

c. From a vapor pressure chart for acetone, $p_{WB} = 44$mm 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. $H_{AB} = \dfrac{P_{AB}}{P-P_{AB}} \dfrac{M_{air}}{M_{acetone}} = \dfrac{44}{760-44} \left(\dfrac{29}{58.1}\right) = 0.124$ lb/lb.

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

(Figure 11 is a Watson plot of $\dfrac{\lambda}{T}$ (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 - ^{o}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]: $D = \dfrac{0.0043\ T^{3/2}}{P(V_A^{1/3} + V_B^{1/3})^2}\sqrt{\dfrac{1}{M_1} + \dfrac{1}{M_2}}$. 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, b=0.343.

    h. Substituting in (16) and solving for T_{AB} yields T_{AB}=-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_{AB}=-7.6°C and H_{AB}= 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) $k_G a = \dfrac{G_o}{\overline{z M_s}} \left(\dfrac{\Delta H}{p_i - p_G}\right)_{LM}$

(26) $K_G a = \dfrac{G_o}{\overline{z M_s}} \left(\dfrac{\Delta H}{p_e - p_G}\right)_{LM}$

(3d) $\dfrac{1}{K_G a} = \dfrac{1}{k_G a}_{LM} + \dfrac{1}{\mathcal{H} k_L^o a}$

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

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

E = 66.7 x ml/sec = 66.7(0.859) = 57.4 lb/hr-ft².

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

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

d. Since $T_2 = 8.9^\circ C$ and $H_2 = 0.183$, $T_{2-WB} = 0.9^\circ C$ (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. $(p_{WB}-p_G)_{LM} = \dfrac{(p_{2_{WB}}-p_2) - (p_{1_{WB}}-p_1)}{Ln\dfrac{(p_{2_{WB}}-p_2)}{(p_{1_{WB}}-p_1)}} =$

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

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

i. Since $t_2 = 18.8^\circ C$ and $t_1 = 5.7^\circ C$, $p_{2_e} = 0.223$ and

$p_{1_e} = 0.120$ Atm. (From Fig. 10)

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

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

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

1. $t_{2_i} = \dfrac{T_{2_{WB}} + t_2}{2} = \dfrac{0.9 + 18.8}{2} = 9.8^\circ C$

Similarly $t_{1_i} = \dfrac{-11.0 + 5.7}{2} = -2.7^\circ C$

m. From Figure 10, $p_{2_1} = 0.1470$ atm. and $p_{1_1} = 0.0755$ atm.

n. $(p_1 - p_G)_{LM} = \dfrac{(p_{2_1}-p_2) - (p_{1_1}-p_1)}{Ln \dfrac{(p_{2_1}-p_2)}{p_{1_1}-p_1}} =$

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

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

p. The average liquid film temperature corresponding to $k_G^x a$ is obtained by averaging t_1, t_2, t_{1_1} and t_{2_1} giving a value of 7.9^oC.

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 $\dfrac{ft^3}{mol}$

and $\dfrac{1}{p^o V_m}$, Henry's constant is found to be 6.36 $\dfrac{lb.mols}{ft^3-Atm.}$

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 $\dfrac{ft^3 \text{ acetone}}{ft^3-hr.}$

r. Similarly, taking the interface temperature equal to the wet-bulb temperature, $t_{f-av} = 3.6^oC$, $p_{f-av}^o = 0.108$ atm., and $k_L a = 2.95$ ft^3/ft^3-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) $K_G a = \dfrac{G_o}{M_1 p z}\left[\dfrac{1}{1-\dfrac{p_{e_2}-p_{e_1}}{p_2}}\right]$ $\ln \dfrac{p_{e_1}}{p_{e_2}-p_2}$

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

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

b. The pounds of acetone evaporated per hour

   per ft^2 is:

   $E = 66.7 \times ml/sec \times Vol.\% = (66.7)(0.324)(0.4) =$

   $8.65 \dfrac{lb\ Acetone}{hr-ft^2}$ (this is the same as the feed

   rate.)

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

   $\dfrac{8.65}{2(313) + 8.65 + (0.58)\dfrac{(0.603)}{0.397}(8.65)} = 0.138\ Atm.$

d. $p_{e_2} = p_2^* = x_2 p^o_2$. Since $t_2 = 20.7^\circ C$, $p_2^o = 0.246\ Atm.$

   x_2 is determined from the sample taken of

   the entering liquid. Hence, Vol.$\%_2 = 0.075$;

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

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

   assuming an ideal mixture, $p_2^* = (0.122)(0.246) =$

   $0.0298\ At.$

e. Wt. % Acetone in liquid at bottom of column

   is: $W\%_1 = \dfrac{W\%_2 L - E}{L - E/W\%_F}$ where $W\%_2 = \dfrac{0.58 x_2}{1-0.42 x_2} = 0.0745$

Hence: $M_1 = \dfrac{0.0745 \ (1200) - (8.65)}{1200 - \dfrac{8.65}{0.397}} = 0.0685$

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

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 $G_0 = 313$, $M_1 = 29$, $z = 0.396$ and $P = 1$ yields: $K_G a = 19.3 \ \dfrac{lb.Mol}{Hr-Ft^3-at.}$

i. The value of π_{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) $= \dfrac{E}{G_0} = \dfrac{8.65}{313} = 0.0277$. Since

$T_2 = 18.6^oC$, T_{2-wB} (acetone) $= -10.0^oC$.

Similarly, H_2(MIK) $= \dfrac{(1-wt\%_F)}{wt\%_F} \dfrac{E}{G_0} =$

$\dfrac{0.603}{0.397} \dfrac{(8.65)}{313} = 0.042$

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 $x_2 = 0.122$, $t_{2_1} = T_{wB-2} = 12.5^oC$.

j. Similarly, $t_{1_1} = 12.4^oC$ and $t_{f-av} =$

$\dfrac{t_1 + t_2 + t_{1_1} + t_{2_1}}{4} = 15.8^oC$.

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

l. Making use of the relation: $\dfrac{K_Ga}{k_Ga} = \dfrac{(p_i-p)}{(p^*-p)}\text{av.}$,

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

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

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

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

o. $\cancel{\rtimes}_f = \dfrac{1}{p^o_f \, v_{m_f}} = 2.67 \dfrac{mol}{ft^3\text{-Atm.}}$

p. $k_La = \dfrac{1}{(2.67)\left(\dfrac{1}{19.3} - \dfrac{1}{106.1}\right)} = 8.84 \dfrac{Ft^3 \, Ac.}{Ft^3\text{-hr.}}$

(106.1 is the k_Ga from Fig. 4 at these flow rates.)

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

$t_{f\,av} = 17.5^oC$
$p^o_{f\,av} = 0.209$ Atm.
$K_Ga/k_Ga = 0.529$ (based on $k^x_Ga = 36.5$)
$p^*_{f\text{-av}} = 0.0236$ atm.
$x_{f\text{-av}} = 0.113$
$v_{m\text{-}f\,av} = 1.92$ ft^3/mol
$\cancel{\rtimes}_f = 2.51$ mol/ft^3-atm.
$k^x_La = 13.6$

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

$$p^o_{f-av} = \frac{p^o_1 + p^o_2}{2} = \frac{0.2105 + 0.2455}{2} = $$

0.228 Atm.

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

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 $\aleph_f = 2.29$ mol/ft^3- atm.

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

D. <u>Calculation of Correlating Function For Liquid Film Coefficients</u>

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 $\emptyset = 0.9$, 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 $\left(\frac{{}^{\circ}K - sec}{Cm^2 - Cp}\right)$ for acetone and MIK, respectively. (A second empirical correlation of Wilke's data has been proposed by Scheibel[19]. This takes the form $F = 1.22 \times 10^7 \left[\frac{V_A^{1/3}}{1 + \frac{(3V_B)^{2/3}}{(V_A)}}\right]$ 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)

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

b. $D_L \mathcal{H} = \frac{T}{F}$

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

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

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

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

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

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 x 10⁻⁷. This is obtained
from a linear variation of $D_L\mu$ vs. x_f
(between the limits of 118 x 10⁻⁷ and 79.2 x 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/ft³) is obtained from a linear variation
with x_f. For this case ρ = 50.45 lb/ft³.

h. $\mu / \rho D_L = \dfrac{\mu^2}{\rho D_L \mu} = \dfrac{(0.604)^2 (0.624)}{(50.45)(82.9 \times 10^{-7})} = 545$
0.624 is a conversion constant to change cp to
lb/ft-hr and cm² to ft². The Schmidt number
is then rendered dimensionless.

i. $(C^1_L)_f$ is the mols MIK/ft³ mixture.
This is obtained from $\dfrac{(1-x_{fac})}{(v_{mf-av.})}$
v_{mf-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 $(c^1{}_L)_f$ x_m = $(1-x)_f$ ρ_f)

$$\frac{\text{moles mix}}{\text{ft}^3 \text{ mixture}} \times \frac{\text{lb mixture}}{\text{moles mixture}}$$

k. By a similar procedure, but using the equation of Scheibel[19] to predict ∂_L, $(c^1{}_L)_f$ x_m $(\mu/\rho \vartheta_L)$ = 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 MIK 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., $(0.85)^2$ x 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 $k_g a$ 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 $k_g a$ could be anywhere between the indicated $k_g a$ and the lower values of $k_g a$. It is probable, however, that the $k_g a$'s plotted are good engineering estimations of the true values.

Figure 6 is a plot of the liquid film resistance ($1/k_L a$) plotted vs. mole % MA 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 Surman[25])

   c.   Assuming the interface temperatures is alway 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) $\dfrac{T}{D_L \mu} = f(V_A)$ (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) $D_L = K \dfrac{T}{\mu} \left[\dfrac{1 + \dfrac{(3V_B)^{2/3}}{(V_A)}}{V_A^{1/3}} \right]$

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 re-
liability of the assumption made in computing the data is
obtained. Thus:

Interface Temperature	Pure Component Computation	$1/k_L a$ Liquid Conc. Intercept @ 100% Acetone	Discrepancy	Pure Component Conc. Intercept
t_L	0	0.0901	0.0901	0
$\dfrac{t_{w.s.}+t_L}{2}$	0.152	0.213	0.061	0.715
$t_{w.s.}$	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 $1/k_La$ (computed by the
wet-bulb assumption) vs. the correlating variable suggested
by Stutsman $z = (\rho^1_L)_f M_B (\frac{\mu}{\rho D})$. Two curves are shown,

the higher one being computed using Wilke's diffusivity
curves, and the lower one is equation (30). It is evi-
dent 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 prob-
able 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.
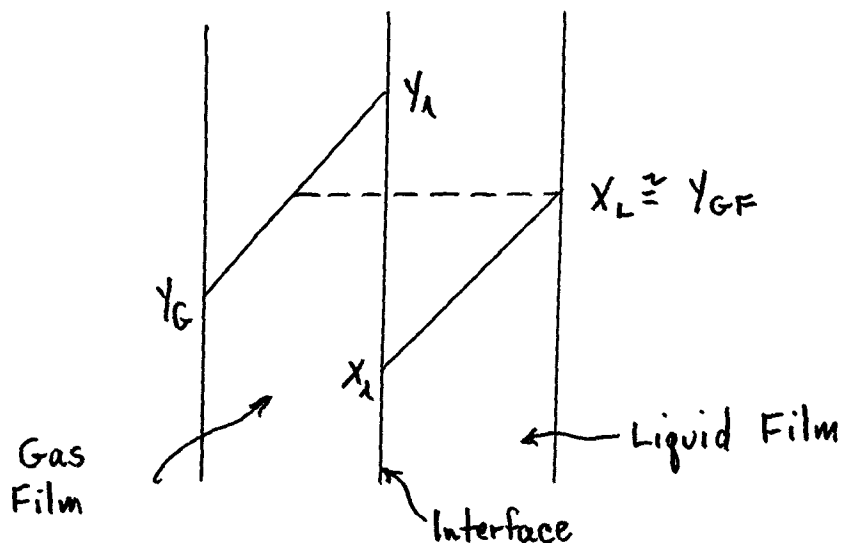
Thus, from Fig. 7, the following equations are recommended to describe the effect of liquid concentration on $k_L a$ for the present system at the rates specified:

(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 $k_G a = k_L a$; this is shown by the dashed extrapolation (the intercept being 19.6 computed from main body liquid temperatures). The extent of error in assuming $k_G a = k_L a$ at other Δt'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 analyze 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 Stutaman'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) $1/k_L a = 0.376 - 0.0000102 (c_L^{\frac{1}{2}})_f M_m (\mu / \rho \omega)$ when diffusivity data of Wilke[28] are used, and

(31a) $1/k_L a = 0.376 - 0.0000152 (c_L^{\frac{1}{2}})_f M_m (\mu / \rho \omega)$ 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 comput-
ing the wet-bulb temperatures for each of the pure components
and the assumption of a linear relationship of T_{w} 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 con-
firm or establish some other relationship.

B. It is recommended that further studies be made of
the acetone-AIR 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 deter-
mined for both pure acetone and mixtures of this with AIR
all at constant liquid and gas rates. $k_{L}a$ should be com-

puted 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 $k_L a$'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 $k_G a$'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:

$$\frac{1}{k_L a} = 0.376 - 0.0000102 \ (c_L^1)_f \ M_m (\mu / \rho D) \ \text{using diffu-}$$

sivity data of Wilke[28]

$$\frac{1}{k_L a} = 0.376 - 0.0000152 \ (c_L^1)_f \ M_m \ (\mu / \rho D) \ \text{using diffu-}$$

sivity 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, $\dfrac{Ft^2}{Ft^3}$

C_p = Specific Heat, $cal/G_m{}^\circ C$

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

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

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

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

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

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, $\dfrac{lbs\ solute}{lbs.\ gas}$

$\cancel{H}$ = 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_b = Partial pressure of diffusing gas in main stream, atm.

p_e^* = Partial pressure of solute gas in equilibrium with J_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-^{o}R

J = 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.-moles/sec.

w = Weight fraction

x = Mol fraction

X = Length, ft.

s = 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

C = Pure Component

WB = Wet-Bulb

Superscripts

o = Pure Component

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

* = Equilibrium liquid.
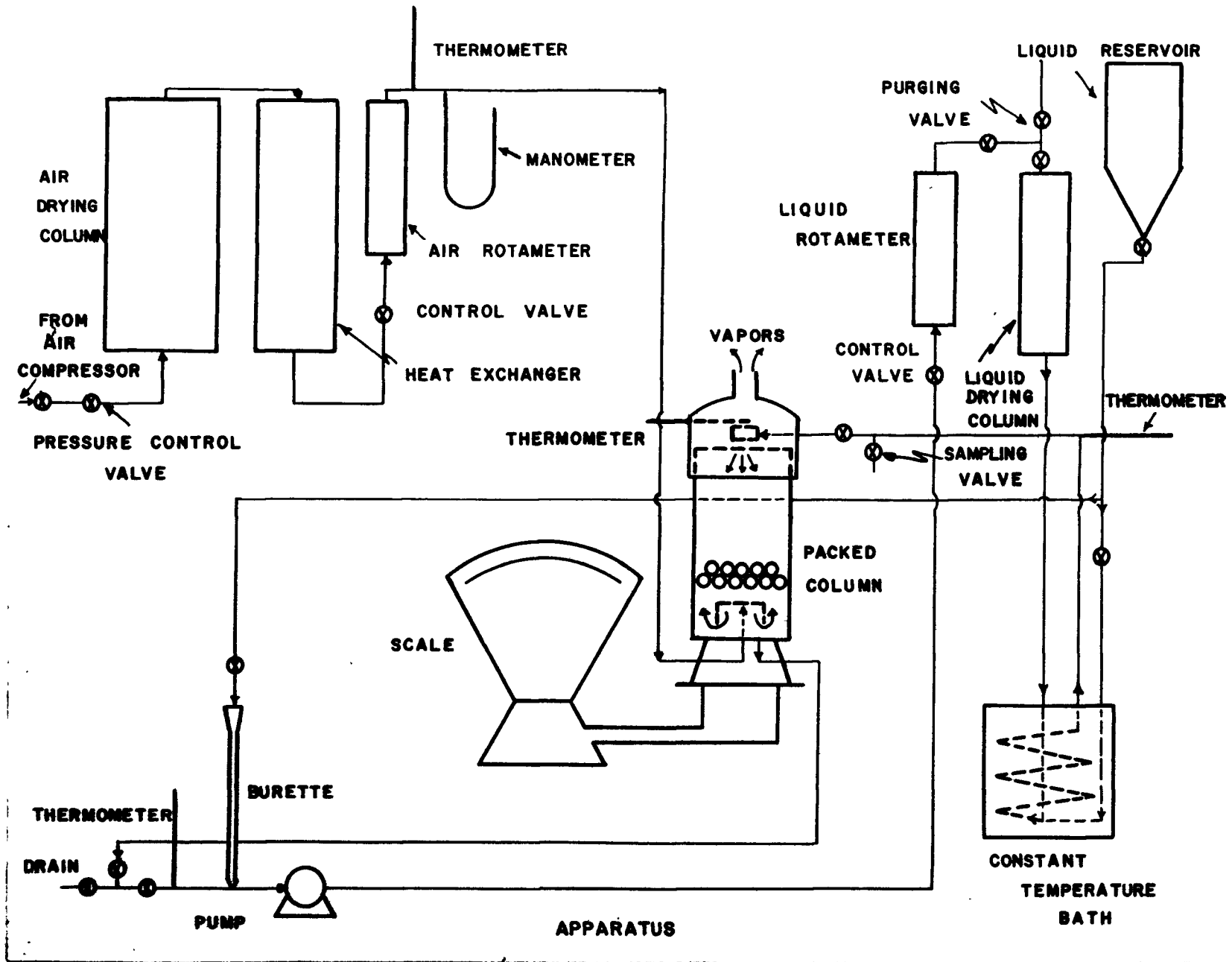
THERMOMETER

LIQUID RESERVOIR

PURGING VALVE

MANOMETER

AIR DRYING COLUMN

AIR ROTAMETER

LIQUID ROTAMETER

CONTROL VALVE

FROM AIR COMPRESSOR

HEAT EXCHANGER

VAPORS

CONTROL VALVE

LIQUID DRYING COLUMN

THERMOMETER

PRESSURE CONTROL VALVE

THERMOMETER

SAMPLING VALVE

PACKED COLUMN

SCALE

BURETTE

THERMOMETER

DRAIN

CONSTANT TEMPERATURE BATH

PUMP

APPARATUS

Fig. 1

Fig. 2

Fig. 5
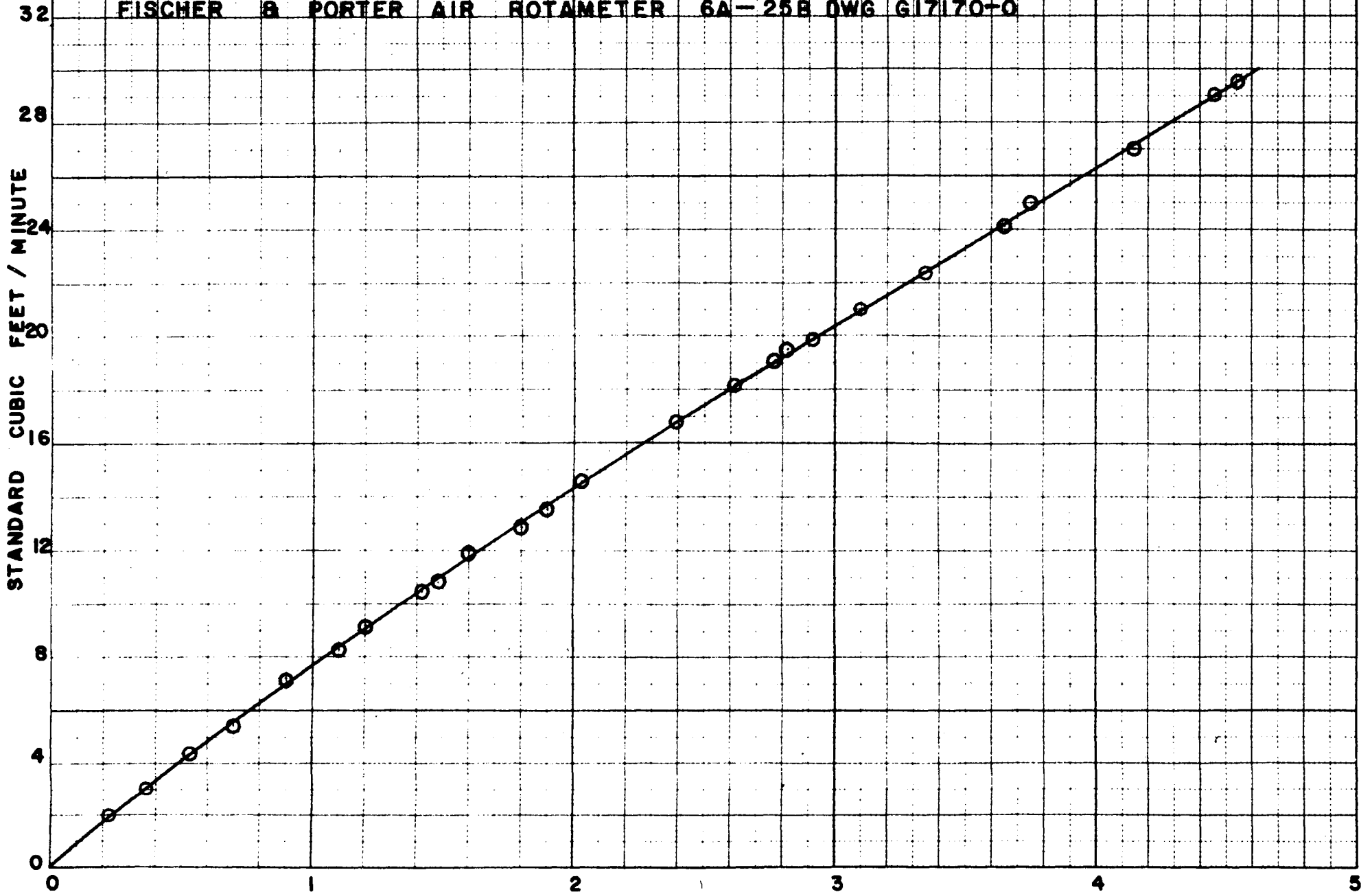
CALIBRATION OF SPENCER REFRACTOMETER
FOR
ACETONE—METHYL ISOBUTYL KETONE
AT 23.5°C

VOLUME PER CENT ACETONE

REFRACTOMETER READING (DEGREES)

Fig. 4

MASS TRANSFER COEFFICIENTS FOR ACETONE

Fig. 5
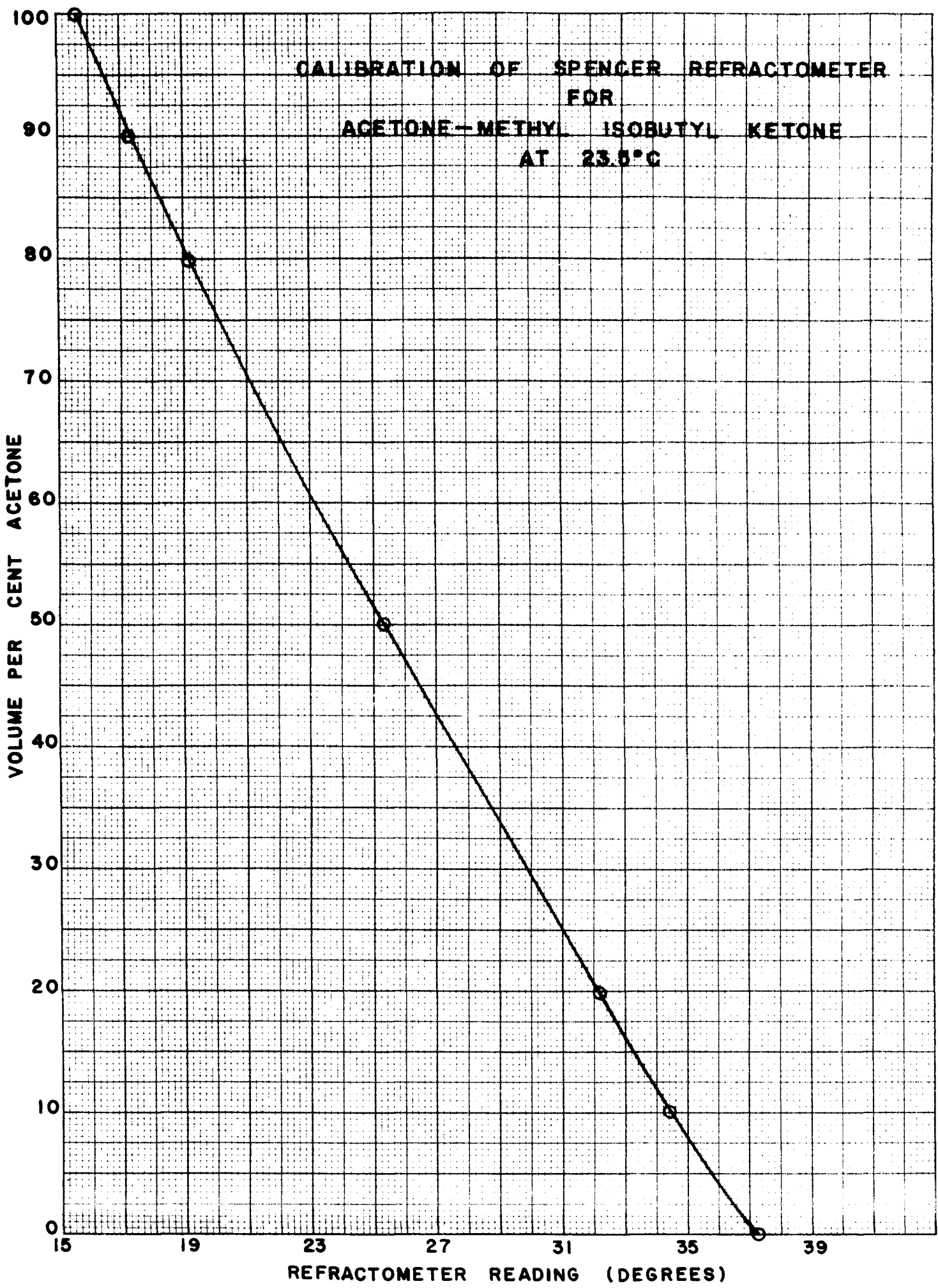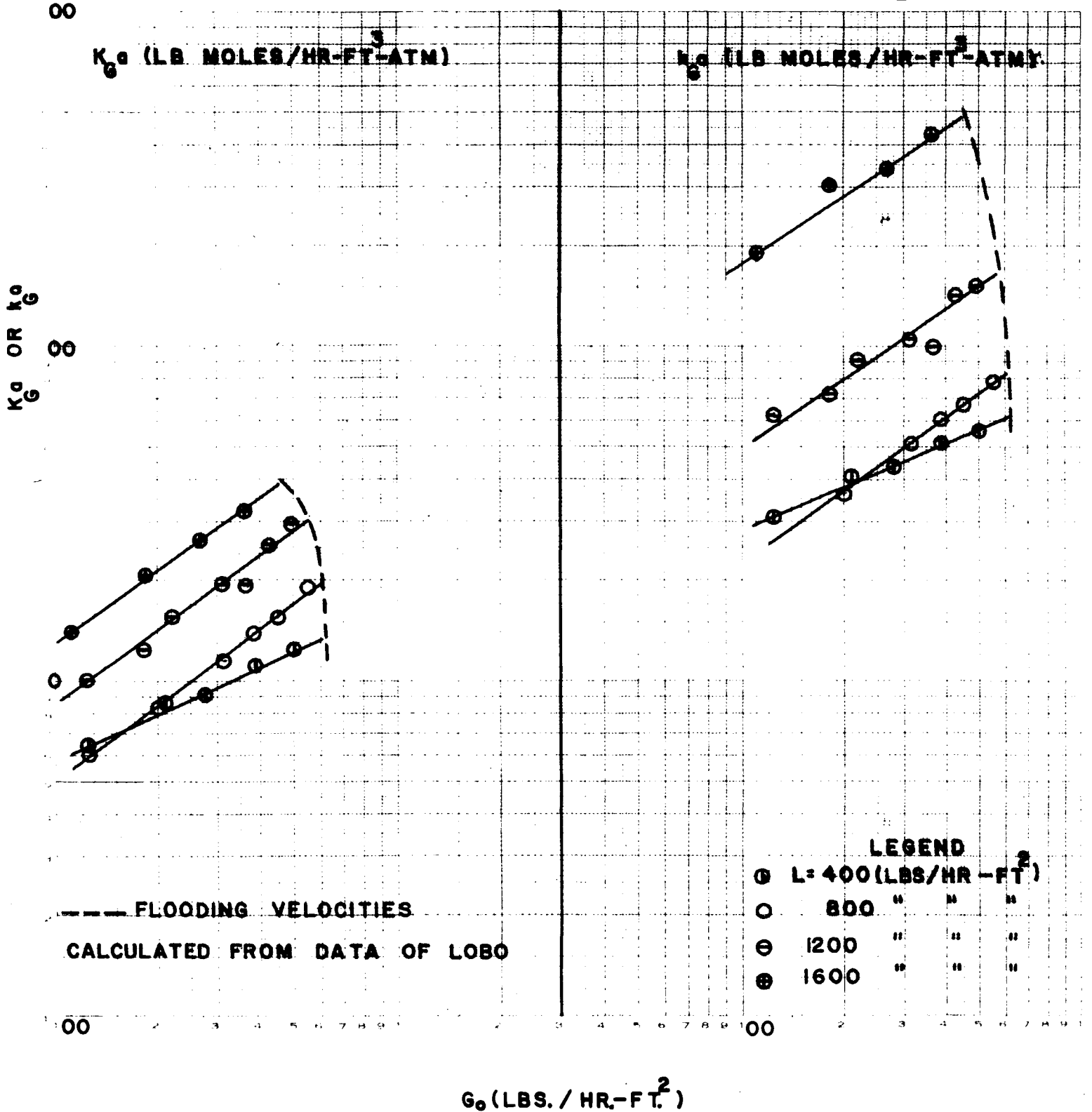
MASS TRANSFER COEFFICIENTS FOR METHYL ISOBUTYL KETONE

K_Ga OR k_Ga

K_Ga (LB MOLES/HR-FT^3-ATM)

k_Ga (LB MOLES/HR-FT^3-ATM)

— — —FLOODING VELOCITIES

CALCULATED FROM DATA OF LOBO

LEGEND

O L=800 (LBS/HR-FT^2)

⊖ 1200 " " "

⊘ 1600 " " "

G_o (LBS./HR-FT^2)

Fig.6

THE EFFECT OF LIQUID CONCENTRATION
ON
MASS TRANSFER COEFFICIENTS IN LIQUID FILM
AT
L = 1200 LBS/HR−FT^2 G = 313 LBS/HR−FT^2

LEGEND

⊕ $t_i = t_{wb}$

⊕ $t_i = t_L$

⊕ $t_i = \dfrac{t_L + t_{wb}}{2}$

$1/k_L a$ (FT^3 ACETONE /FT^3 PACKING −HR)

MOL FRACTION M.I.K. IN LIQUID FILM

Fig. 7
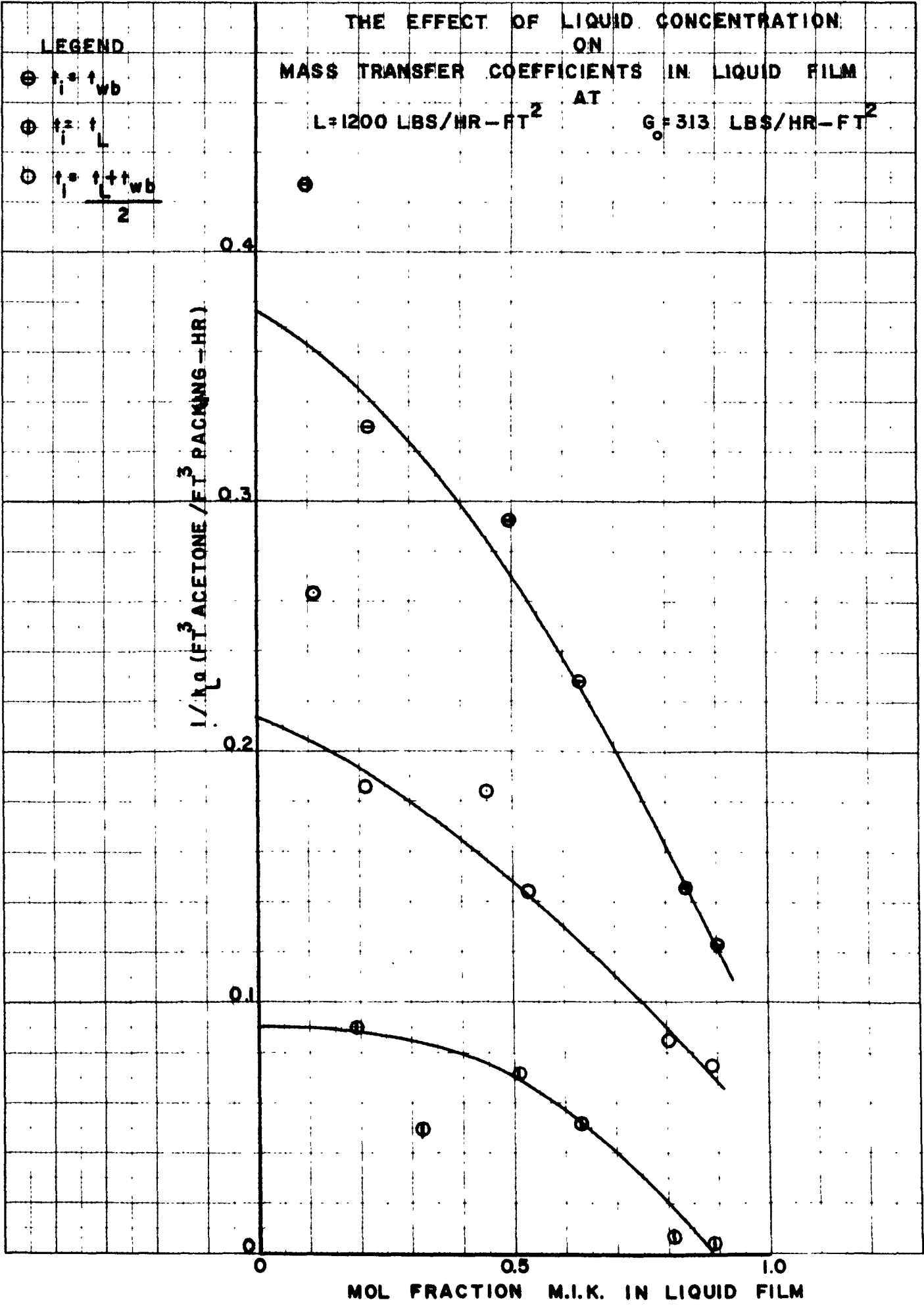
CORRELATION OF LIQUID FILM RESISTANCE

$L = 1200$ LBS/HR-FT^2 $G_o = 313$ LBS/HR-FT^2

DIFFUSIVITY DATA OF WILKIE[28]

DIFFUSIVITY DATA OF SCHEIBEL[19]

$1/K_L^o$ (FT^3 ACETONE/FT^3 PACKING-HR)

$Z = (C'_L)_f M_m (\frac{\mu}{\rho D_L}) \times 10^{-4}$

Fig. 8

DETERMINATION OF INTERFACE TEMPERATURE

PURE COMPONENT RESISTANCE / CONCENTRATION INTERCEPT
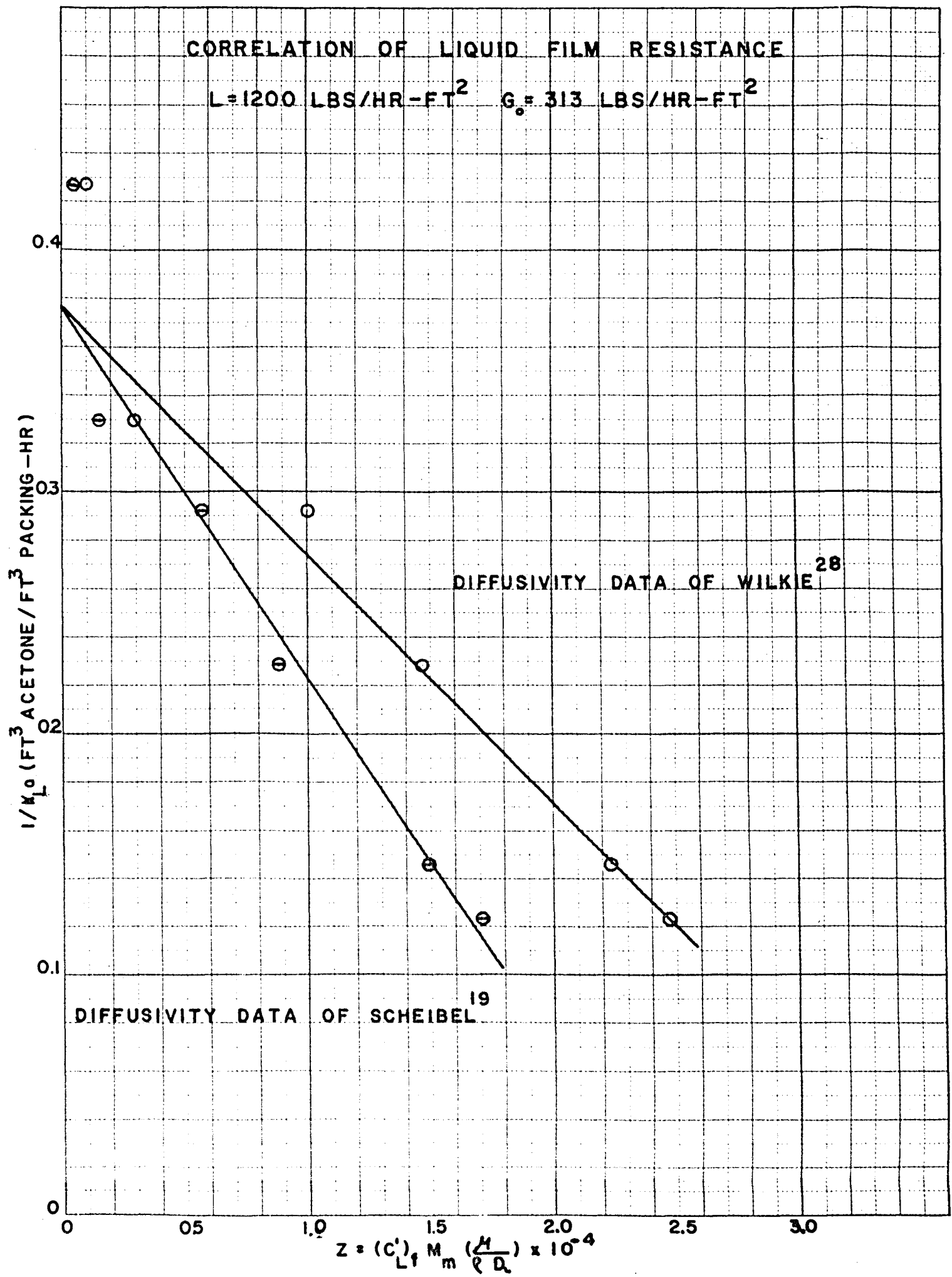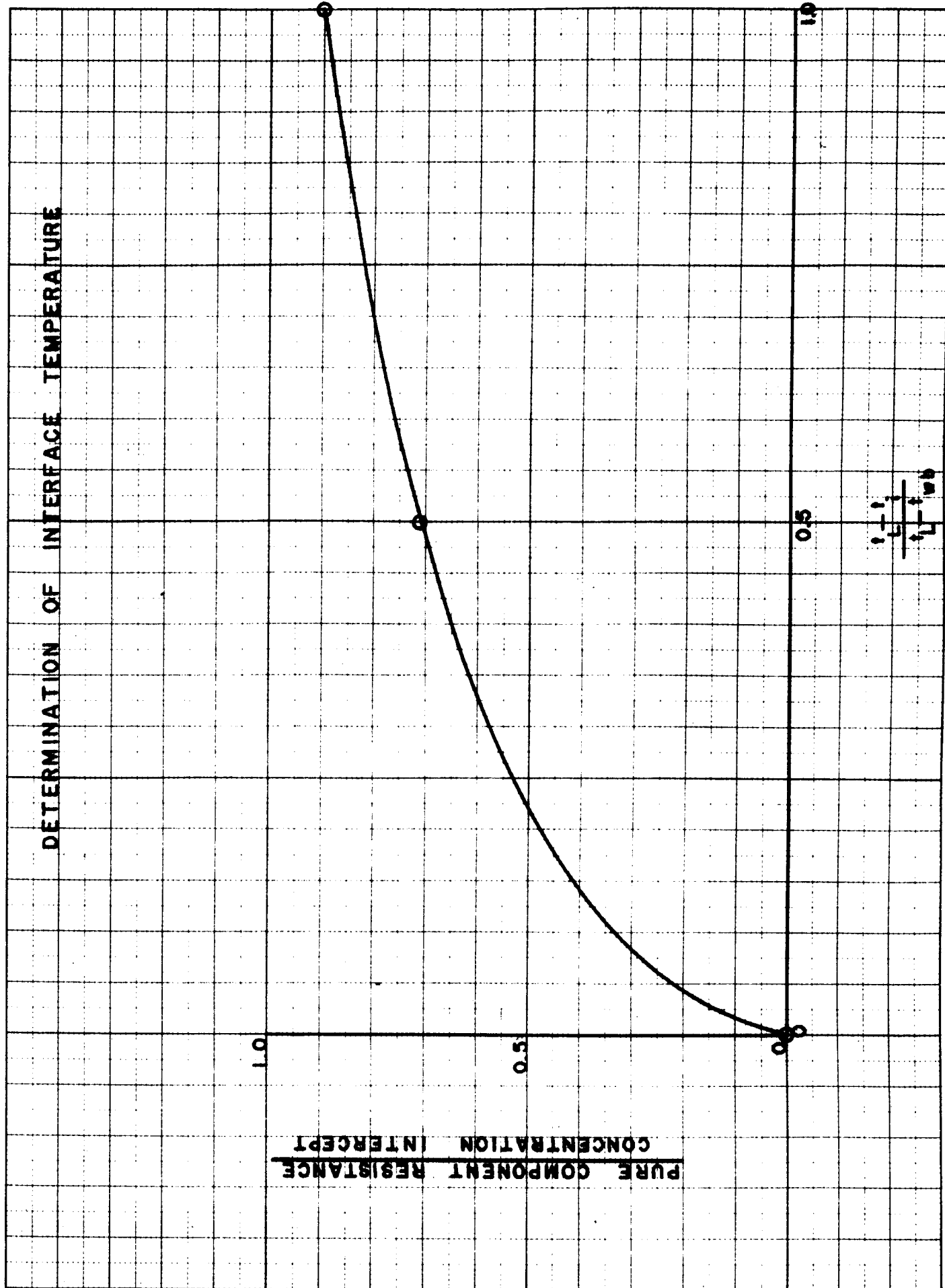
THE EFFECT OF TEMPERATURE POTENTIAL ACROSS LIQUID FILM

ON MASS TRANSFER COEFFICIENTS IN GAS FILM

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

130

$k_G a$ (LB. MOLES/HR-FT^3-ATM.)

80

30

0

$K_G a = k_G a$

0   5   10   15   20

$\Delta T_{av.}$ (°C) LIQUID FILM
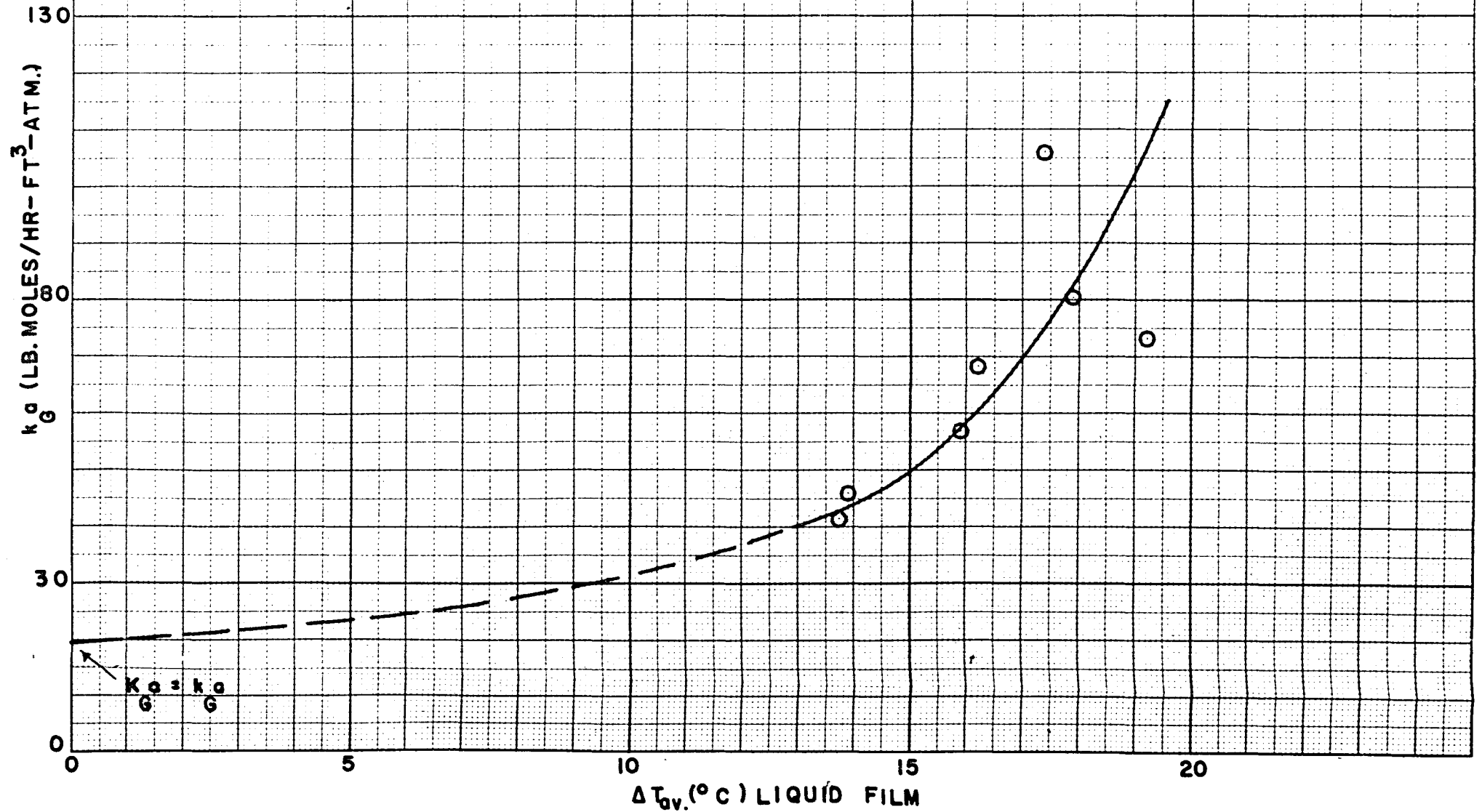
VAPOR PRESSURES OF ACETONE, M.I.K. AND n.BUTANOL
VS.
VAPOR PRESSURES OF WATER AT THE SAME REDUCED TEMP.
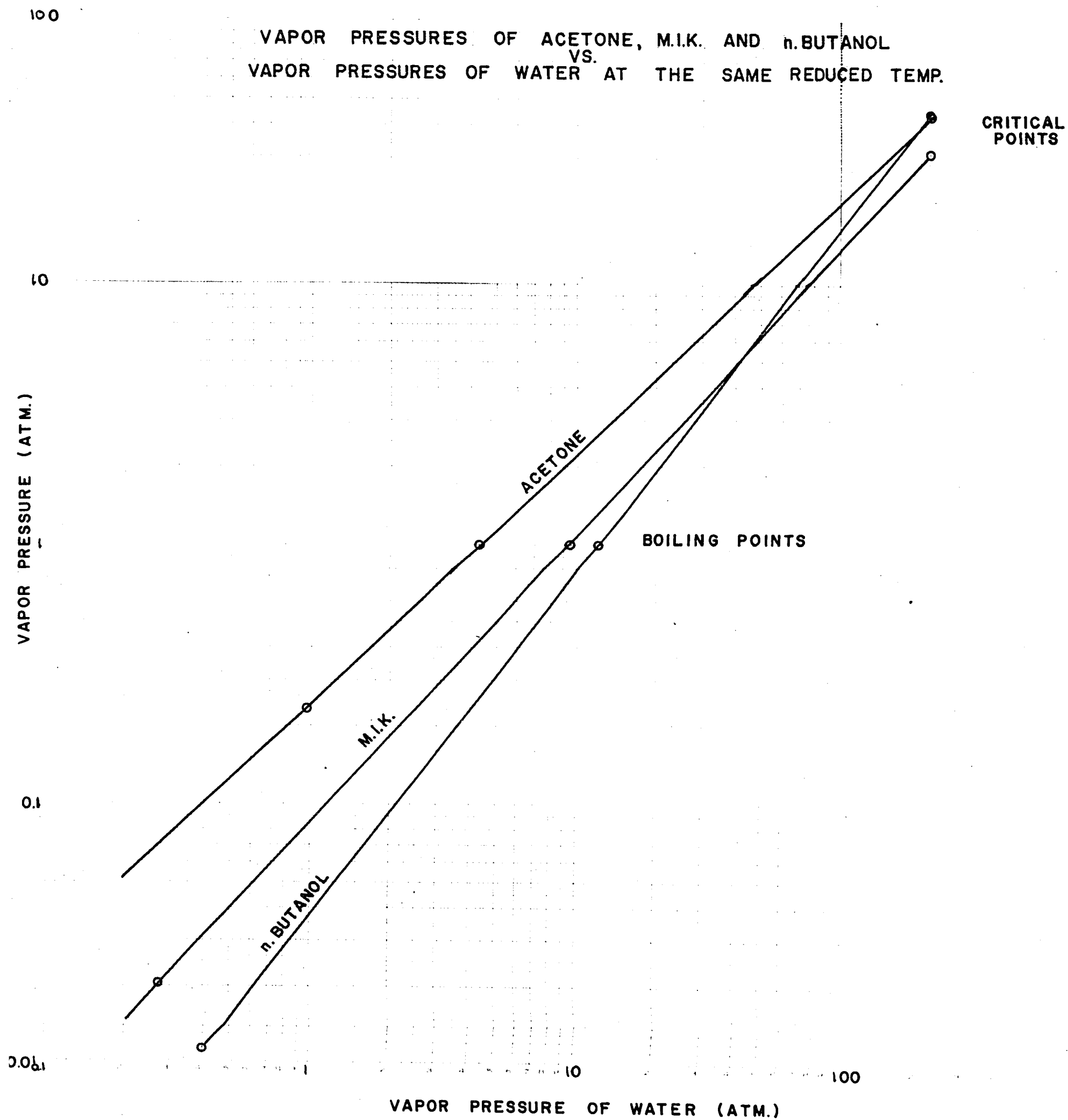
Fig. 11

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

$$\left(\frac{\lambda}{T}\right)_2 = \left(\frac{\lambda}{T}\right)_1 \frac{\theta_2}{\theta_1}$$

$(\lambda/T)_q$

○ OBSERVED VALUES
FOR ACETONE

T_c ACETONE $= 508.2°K$

T_c M.I.K $= 587.9°K$

T_c BUTANOL $= 560.2°K$

$T_r = T/T_c$
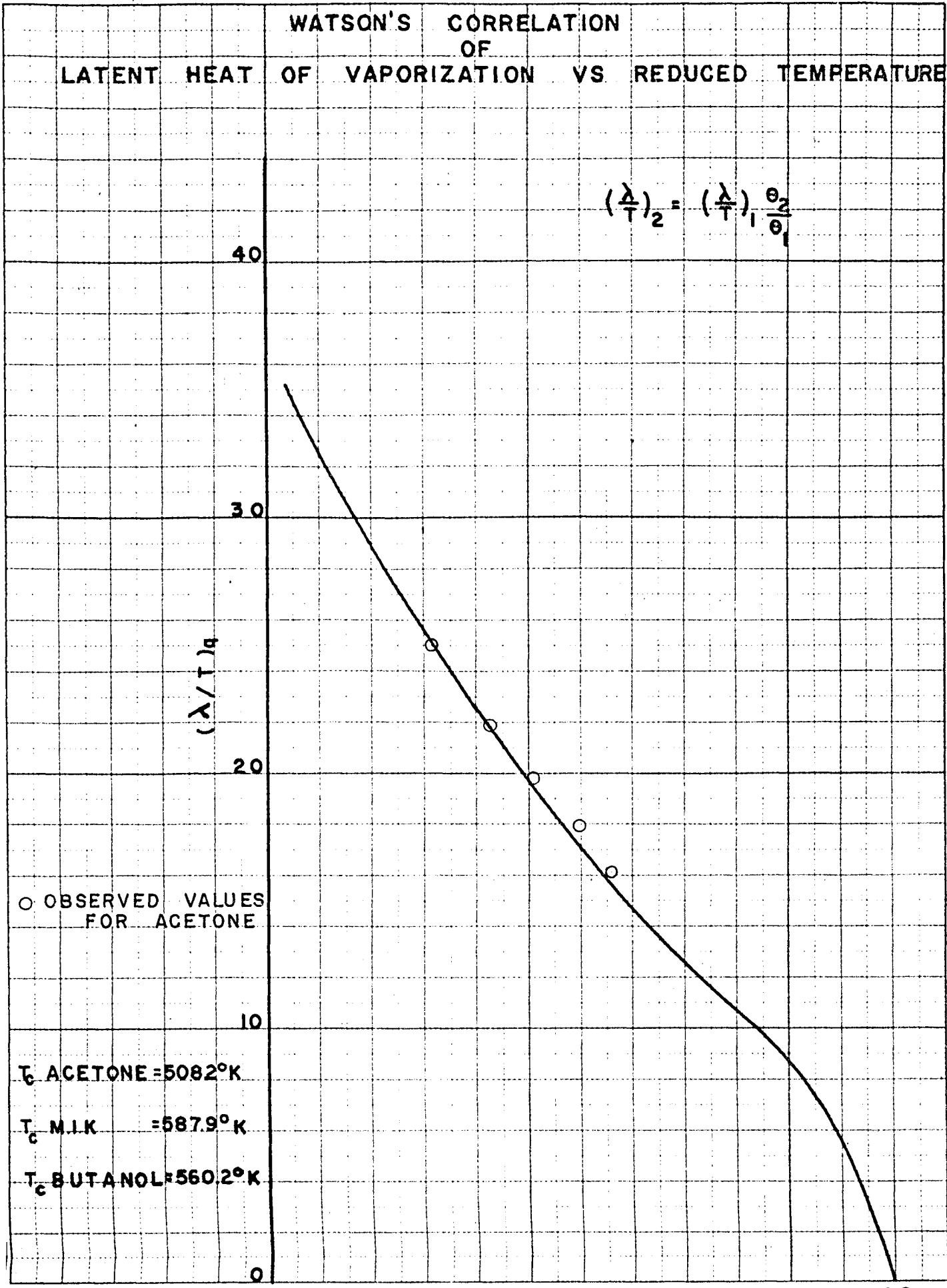
HUMIDITY CHART
FOR
AIR — ACETONE SYSTEM
AT
1 ATM.

SATURATION
LINE

WET BULB
LINES

LBS. ACETONE / LB. AIR

TEMPERATURE (°C)
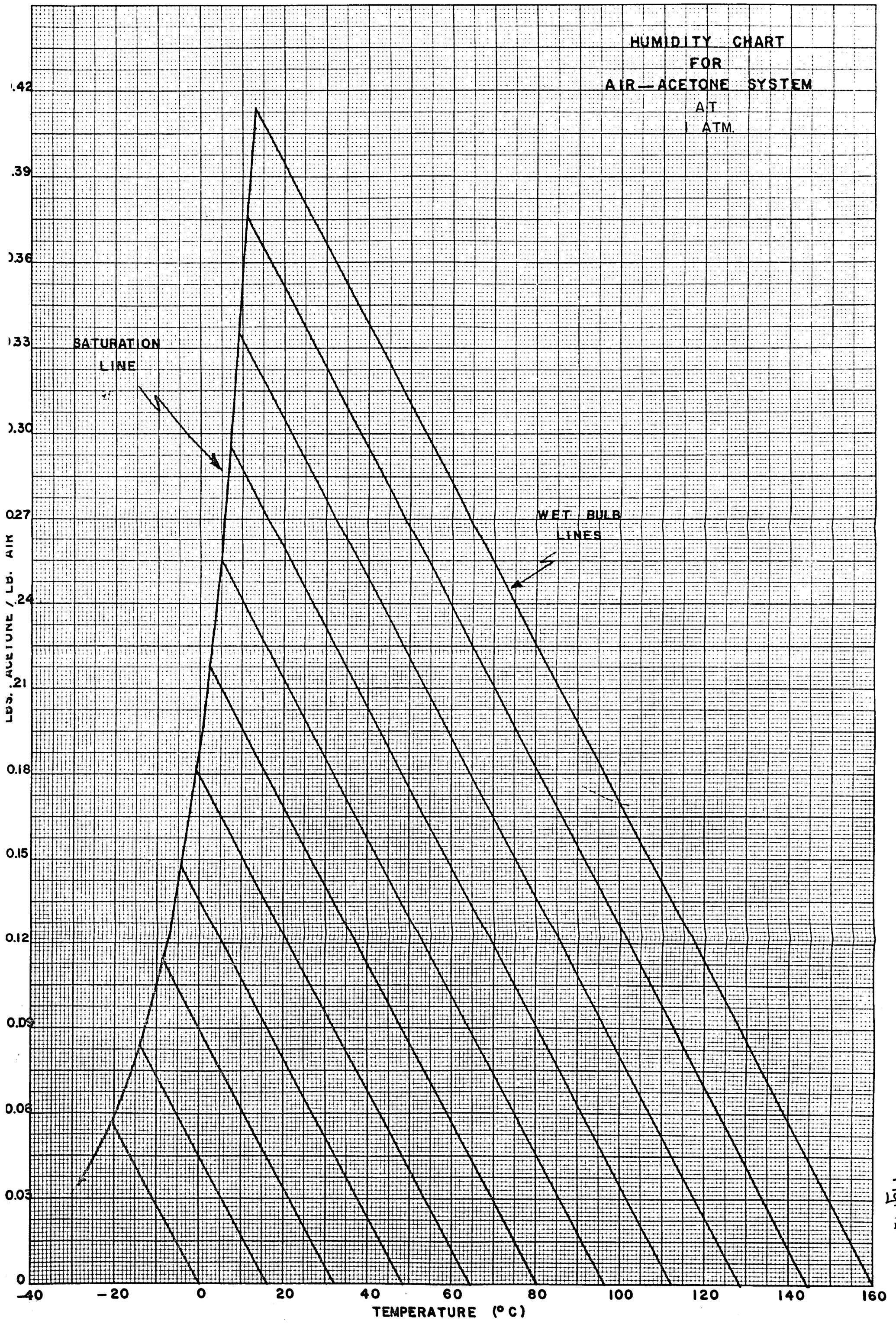
HUMIDITY CHART
FOR
AIR—METHYL ISOBUTYL KETONE SYSTEM
AT
1 ATM.

LBS. M.I.K. / LB. AIR

SATURATION LINE

WET BULB LINES

Fig. 13

Fig. 14

HUMIDITY CHART
FOR
AIR – n. BUTANOL SYSTEM
AT
1 ATM

SATURATION LINE

WET BULB LINES

LBS. n. BUTANOL / LB. AIR

TEMPERATURE (°F)
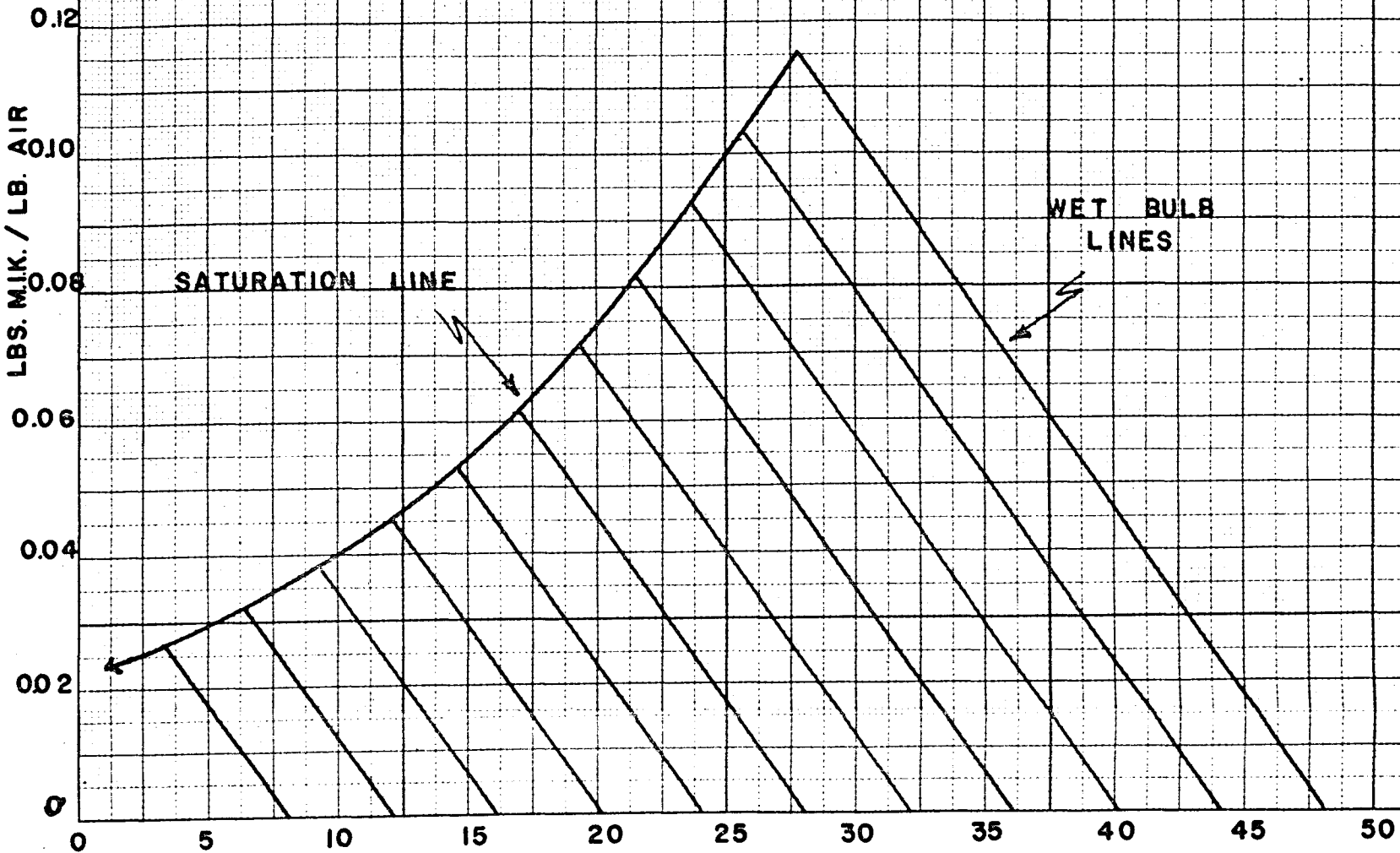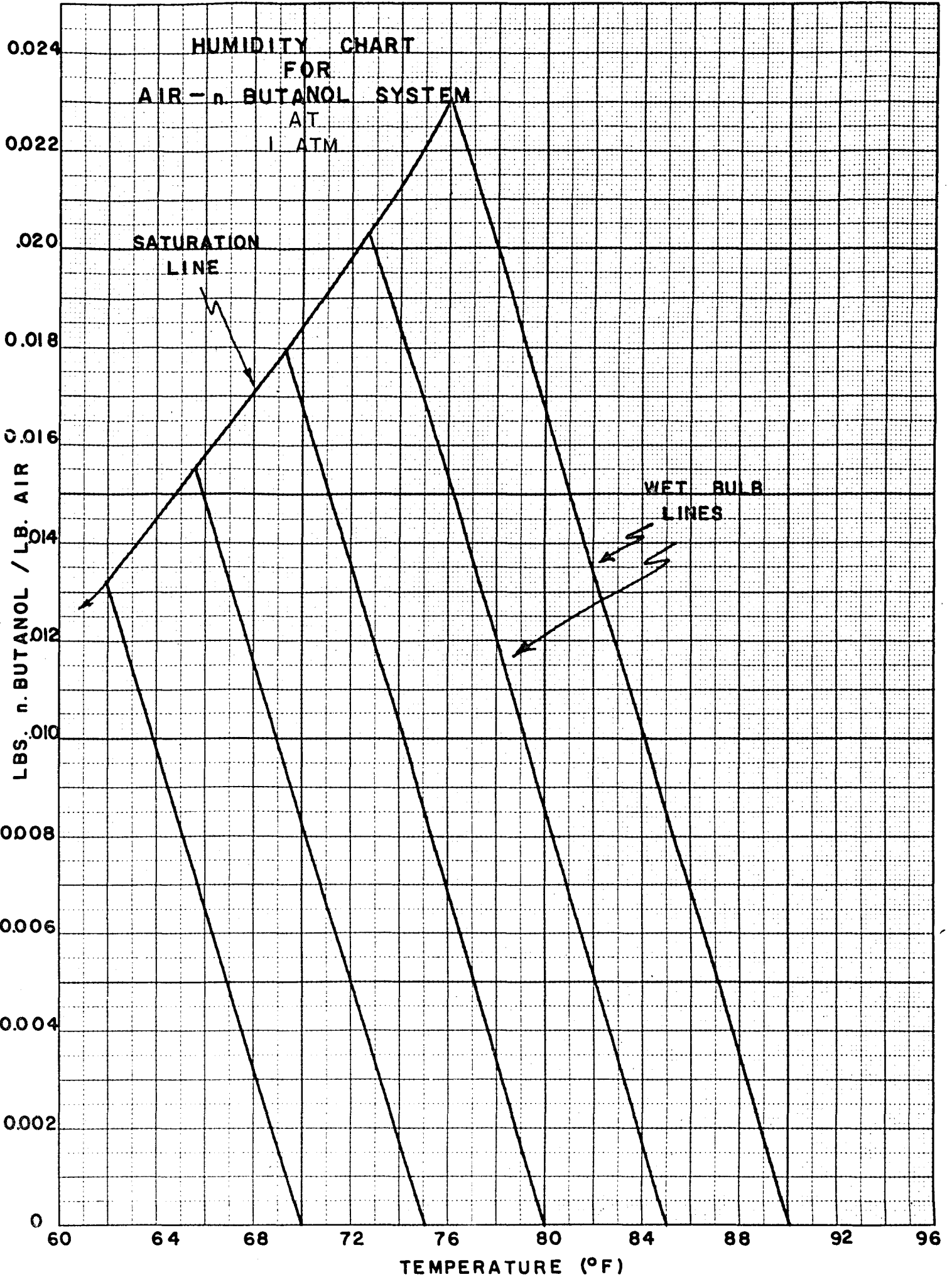
TABLE I

SPECIFICATIONS OF SOLVENTS USED

<u>Methyl Isobutyl Ketone</u> (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: $114.0 - 117.0°C$.

<u>Acetone</u> (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
$H_2 = H$ (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	0.235	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	48	49	50	51	52	53	54	55	56	57	58	59	60	61	62	63
G_o	368	368	368	368	494	494	494	494	494	429	429	313	313	313	222	222	222	182	182	182	124	124	124	124
Evaporation Rate	0.817	0.843	0.938	0.907	1.24	1.14	1.22	1.20	1.25	1.12	1.12	0.859	0.848	0.895	0.726	0.711	0.712	0.624	0.594	0.588	0.480	0.503	0.477	0.477
$H_2 - \Delta H$	0.148	0.153	0.170	0.164	0.167	0.154	0.165	0.162	0.169	0.173	0.174	0.183	0.181	0.191	0.218	0.214	0.214	0.230	0.219	0.217	0.259	0.271	0.257	0.257
P_2	0.0689	0.0709	0.0784	0.0760	0.0772	0.0706	0.0761	0.0752	0.0778	0.0798	0.0802	0.0838	0.0829	0.0870	0.0983	0.0966	0.0967	0.103	0.0986	0.0978	0.115	0.119	0.114	0.114
T_2	10.2	9.9	10.0	10.0	8.5	7.3	6.9	6.6	6.7	8.0	8.0	8.9	9.3	9.3	10.6	10.8	10.8	12.3	12.4	12.4	13.9	14.2	14.3	14.4
$T_{2\,WB}$	-1.0	-1.1	0.2	0.0	-0.1	-1.3	-0.5	-0.8	-0.3	0.3	0.3	0.9	0.8	1.4	3.4	3.2	3.2	4.5	3.8	3.5	6.3	7.0	6.3	6.3
t_2	18.5	18.4	18.2	19.2	18.6	18.8	18.5	18.1	17.8	19.4	19.1	18.8	18.9	18.8	18.6	18.7	18.8	19.0	19.1	19.1	19.2	19.4	19.4	19.4
$P_{2\,WB}$	0.0832	0.0826	0.0889	0.0879	0.0872	0.0818	0.0851	0.0838	0.0859	0.0890	0.0893	0.0932	0.0930	0.0965	0.1065	0.1049	0.1054	0.1122	0.1082	0.1072	0.1229	0.1270	0.1229	0.1229
T_1	23.2	23.2	23.2	23.2	23.2	23.4	23.5	23.6	23.6	23.8	23.8	23.8	23.8	23.8	23.7	23.7	23.7	23.5	23.4	23.4	23.3	23.3	23.3	23.3
$T_{1\,WB}$	-11.3	-11.3	-11.3	-11.3	-11.3	-11.2	-11.2	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.0	-11.3	-11.3	-11.3	-11.3
t_1	6.3	5.8	5.4	6.1	1.6	2.6	1.4	1.2	1.1	3.4	3.3	5.7	6.0	6.1	7.9	8.0	8.1	10.1	10.1	10.2	11.7	11.8	12.0	12.0
$P_{1\,WB}$	0.0466	0.0466	0.0464	0.0464	0.0465	0.0467	0.0470	0.0472	0.0472	0.0473	0.0473	0.0473	0.0473	0.0473	0.0472	0.0472	0.0472	0.0471	0.0471	0.0471	0.0466	0.0466	0.0466	0.0466
$k_G a$	87.0	96.8	113.0	103.7	151.3	133.1	154.2	156.1	163.7	138.5	147	106.2	102	110.2	94.5	92.0	90.7	70.8	73.2	72.9	62.8	67.7	60.5	60.5
Av.	100.1				151.7					143.2		106.1			92.4			72.3			62.9			
P_{2e}	0.2200	0.2180	0.2155	0.2265	0.2220	0.2230	0.2205	0.2160	0.2130	0.2310	0.2265	0.2230	0.2250	0.2230	0.2220	0.2230	0.2230	0.2270	0.2275	0.2265	0.2280	0.2310	0.2320	0.2320
P_{1e}	0.1210	0.1185	0.1155	0.1200	0.0960	0.1020	0.0955	0.0945	0.0940	0.1065	0.1060	0.1200	0.1220	0.1225	0.1340	0.1345	0.1350	0.1485	0.1430	0.1495	0.1600	0.1610	0.1620	0.1620
$K_G a$	17.4	18.4	21.6	19.5	28.9	26.1	30.0	29.9	31.9	25.4	25.9	19.2	18.7	20.0	16.0	15.3	15.3	13.1	12.3	12.1	10.3	10.7	10.0	10.0
Av.	19.2				29.4					25.6		19.6			15.6			12.5			10.2			

TABLE III (Cont'd)

Run		51	52	53
Calculation of k_La by Wet-bulb Assumption				
t_f-Av($^\circ$C)		3.6	3.7	3.8
P^0_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 $k_G^X a$ and $k_L^X r$ 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
$k_G^X a$		36.0	35.2	37.5
Av.			36.2	
T_f av.		7.9	8.1	8.1
P^0_f av.		0.134	0.136	0.136
$\mathcal{H}_f$ av.		6.36	6.27	6.27
$k_L^X a$		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_0	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 7

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_0	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_ua	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_ua	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_0 (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
$k_{G}a$ (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
$k_G a$ Av.	46.1			53.6			77.5			97.0			26.2		
p_{2e} (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_{1e} (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
$K_{G}a$ (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
$K_G a$ 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	0.0531	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_G^1	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_G^1	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_0	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_{2s}	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_{1s}	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
$k_G a$	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
$K_G a$	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$_G$a (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 ($k_Ga=106.1$)

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
$t_{f-av.}$ (°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
$p_{f-av.}$ (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
$\dfrac{K_Ga}{k_Ga}$	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
$p^*_{f-av.}$ (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-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-f} Av. (Ft^3/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
η_{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_La (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	3.49	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 $k_L^x a$ 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_{l_1} (°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_1} (°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
$t_{f-av.}$ (°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
$\dfrac{K_Ga=(p_1-P)}{K_Ga=(\overline{p^*-P})}$ av.	0.529	0.477	0.469	0.526	0.425	0.447	0.334	0.411	0.392	0.381	0.354	0.378	0.381	0.373	0.384	0.370	0.359	0.422	0.419	0.431
$p^*_{f-av.}$ (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
$x_{f-av.}$ (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-f} av (Ft³/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
$\rho_{f-av.}$ (Mol/Ft³-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
$k_L^x a$ (Ft³ Acetone/Ft³-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
$\dfrac{K_Ga}{k_Ga}$	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_L^xa	62.4	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	0.0636	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_Ga	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_Ga		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 $k_L a$ USING DIFFUSIVITY DATA OF WILKE						CORRELATION OF $k_L a$ USING DIFFUSIVITY DATA OF SCHEIBEL					
Runs	142-144	145-147	148-150	151-155	174-176	177-179	142-144	145-147	148-150	151-155	174-176	177-179
t_f-av. (°K)	289.0	287.5	284.1	281.7	280.0	279.1	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	236	235	232	230	228.5	228
$D_L \mu$(M.I.K.) x10^7(cm^2-cp/sec)	79.2	78.8	77.9	77.2	76.0	76.5	107.2	106.7	105.4	104.3	103.9	103.5
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	119.7	127.2	152.2	167.6	216.6	200.9
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	377	352	280	234	108.7	144.6
$(c_L^1)_f$ (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	17,080	14,880	8,840	5,790	518	1,570
$1/k_L a$	0.1235	0.146	0.2285	0.2923	0.427	0.330	0.1235	0.146	0.2285	0.2923	0.427	0.330