the rate of physical adsorption is very difficult to obtain, as is pointed out by Brunauer (*8*), because of the difficulty of eliminating pore resistance and thermal effects which accompany adsorption, it is generally assumed that this process occurs with extreme rapidity and that the resistance involved is negligible compared with the other diffusional resistances involved. Such an assumption appears reasonable also from the quantity of heat involved in physical adsorption as compared with the much slower chemisorption process. The heat of physical adsorption is equal to or only slightly greater than that involved in condensation, which is assumed to occur instantaneously, while the heat evolved in chemisorption is about ten times that for physical adsorption. Eliminating gas-phase diffusion and physical adsorption as rate-controlling steps, we are left with only internal pore diffusion as the possible rate-determining step.

Other investigators have also reached the same conclusion. Foster and Daniels (*14*) concluded that for adsorption of nitrogen dioxide from air on silica gel the rate-determining step is the diffusion of the solute inside the particles. Eagleton and Bliss (*13*) also reported that diffusion inside the particles was controlling for the adsorption of water vapor on alumina.

That the resistance to mass transfer through the gas phase is negligibly small as compared with the adsorbed phase is also demonstrated by use of the present data by plotting the over-all transfer-unit height based on the gas phase, H_{OG}, vs. mG/L, the ratio of gas to carbon rates multiplied by the average slope of the equilibrium curve as shown in Figure 10. This plot expresses the usual relationship between the over-all and individual mass transfer unit heights, which is given by the equation

$$H_{OG} = H_G + \frac{mG}{L} H_A \qquad (7)$$

In the plot of H_{OG} vs. mG/L the slope of the line is equivalent to the value of H_A, and the y intercept is the value of H_G. Figure 10 shows H_G to be essentially zero, thus pointing to the negligible resistance to diffusion in the gas film. Allowing for scatter in the experimental data, the average value of the individual transfer-unit height based on the adsorbed phase, H_A, is observed to be about 11 in.

CONCLUSIONS

The performance of a pilot-scale moving-bed adsorber has been studied by means of a binary system of methane-acetylene, and a ternary system of methane–carbon dioxide–acetylene, with Columbia grade-HA activated carbon as the adsorbent. The data were correlated by plotting the over-all transfer unit height values against the ratio of the gas-to-carbon rate. A straight-line relationship was found to exist for both adsorption and enriching operations with over-all-transfer-unit-height values based either on the gas phase or the adsorbed phase ranging from 6.5 to 36.9 in., for ratios of gas-to-carbon rate of from 1.39 to 4.84 cu. ft. of gas (60 °F., 1 atm.)/lb. of carbon. The individual transfer-unit height for the gas phase, H_G, was observed to be essentially zero, an indication that the resistance to mass transfer occurred mostly in the adsorbed phase. The individual transfer-unit height based on the adsorbed phase, H_A, was estimated from the experimental data to be approximately 11 in.

II. EQUILIBRIUM CONSIDERATIONS

A method for estimating adsorption equilibria based on a modification of the Polanyi adsorption-potential theory was developed for use in the investigation described in Part I of this article. In addition, the recently published correlation method of Lewis, Gilliland, Chertow, and Cadogan, suitably modified for the present application, was successfully employed in the correlation and extrapolation of the ternary equilibrium data.

The ability of certain highly porous solids such as activated charcoal, silica gel, and alumina to adsorb large volumes of gases has been recognized and used industrially for over one hundred and fifty years. A large amount of experimental work has been carried out on the equilibrium relations of various individual gases on different adsorbents, and many theoretical and empirical equations have been developed to describe these relationships (*8*). Up to the past few years, however, very little work has been done on the equilibrium relations of binary or multicomponent gas mixtures that would be of interest to designers of a unit for the separation of gases by selective adsorption as described in Part I.

In an analysis of the performance of the moving-bed adsorber in terms of the height of a transfer unit it is necessary that both equilibrium data on the systems involved and operating data on the actual apparatus be available. The experimental determination of all the equilibrium data required, though highly desirable, was considered outside the scope of the present investigation. Since a relatively large body of equilibrium data on similar adsorbate-adsorbent systems is reported in the literature (*31*), it was decided to develop methods for application of these data to the problem. Accordingly, two methods for establishing adsorption equilibria for multicomponent systems were developed for use with the experimental-adsorber data to determine the transfer-unit height. The correlations were checked at random against certain experimental equilibrium measurements as shown in Tables 2 and 3.

METHOD I: BINARY-SYSTEM ADSORPTION EQUILIBRIA, MODIFIED POLANYI THEORY

The Polanyi approach (*29* and *30*) was first chosen because of its relative simplicity. A complete description of the theory and its modifications for use in the present investigation is outside the scope of this paper, and the reader is referred to standard works such as Brunauer (*8*). A few brief comments, however, are introduced here in order to furnish a background for this method of estimation (*20*).

With the combined effects of vertical and lateral interactions taken into account as well as those of other surface forces (*2*) the net effect of which is to create a potential field similar to that

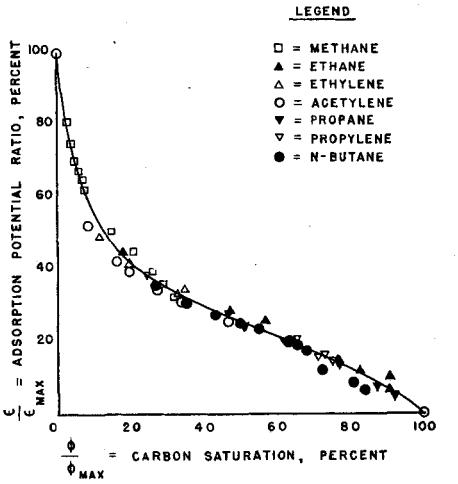

Fig. 11. Generalized adsorption potential curve for hydrocarbon gases. Adsorption on columbia grade L carbon.

of the atmosphere surrounding the earth, the adsorption potential, ϵ_i, is defined as

$$\epsilon_i = \int_{\rho_x}^{\rho_i} V\, dP \qquad (8)$$

where ρ_x and ρ_i are the densities of the adsorbate at point x in the external gas phase and at point i near the adsorbent surface, V is the molar volume, and dP the change in pressure between points x and i. Other investigators (*24* and *25*) have put Equation (8) into the form

$$\epsilon_i = RT_0 \ln \frac{f_0}{f_x} \qquad (9)$$

where f_0 is the fugacity of the adsorbate at temperature T_0, and f_x the fugacity in the external gas phase, R being the gas constant.

Associated with each equipotential surface, ϵ_i, is an enclosed volume, ϕ_i, which varies from zero at the surface of the solid to a maximum value at the boundary of the adsorbate and the external gas phase. The value of ϕ_i is given by

$$\phi_i = \frac{n}{\rho_T} \qquad (10)$$

where n is the quantity of adsorbate considered as a liquid and ρ_T is the density of the liquid at the adsorption temperature, T.

A plot of the relation of ϵ to ϕ is known as the characteristic curve, and it can be shown to be independent of temperature (*8*). On the basis of the adsorption data of Ray and Box (*31*), it was found possible (*20*) to generalize this relationship for the lower hydrocarbons, as shown in Figure 11 by plotting $\epsilon_i/\epsilon_{max}$ vs. ϕ/ϕ_{max}. The name *activity* has been applied (*22*) to ϕ_{max}, and the ratio $(\phi/\phi_{max}) \times (100)$ is also called (*4*) the *per cent saturation*.

To facilitate further the use of the adsorption data, Figures 12 to 15 were also constructed. Figures 12 and 13 are similar to the one published by Berg (*4*) for propane, and are analogous to vapor-pressure curves for liquids, except that for adsorption the additional variable of per cent saturation must be considered. In Figures 14 and 15 the adsorbed density, ρ_a, defined as the gram moles of adsorbate per cubic centimeters of total pore volume, is plotted against the per cent saturation.

On the basis of the foregoing structure it becomes possible to predict the amount and composition of the adsorbate for a given gas mixture from a knowledge of the external gas pressure, temperature, and composition. For gas mixtures it was assumed that the law of additive pressures applied, that the partial pressure of any component was directly proportional to its mole fraction in the gas mixture, and that to a first approximation the amount of one component adsorbed from the gas mixture was independent of the amounts of other components adsorbed. From the composition and total pressure of the gas phase, therefore, the partial pressures of each component were calculated as the product of the total pressure, π, and the mole fractions, y. With these partial pressures used as P_x values in the adsorbate vapor-pressure charts such as Figures 12 and 13, the corresponding values of per cent saturation were obtained at the temperature of the system. Knowing the per cent saturation, one could obtain the value of the average molal adsorbed density, ρ_a, from the appropriate chart (Figures 14 and 15), relating it to the per cent saturation at the same temperature. The amount of each gas adsorbed could then be calculated by multiplying the adsorbed density by the activity ϕ_{max}, from which the amount and composition of the adsorbate could be calculated.

The ability of the foregoing method to predict the adsorption equilibria for gas mixtures was tested against several different experimental determinations of binary mixtures of methane and acetylene and was found, as shown in Table 2, to give reasonably good agreement for acetylene, which was noted to be appreciably better than the agreement obtained for methane. This effect might result from the assumption that no interference occurred in the adsorption of one gas by the presence of a second more strongly adsorbed gas. In the present case the acetylene, being much more strongly adsorbed, would tend to be affected only slightly by the presence of methane, but the reverse would not be true. Similar effects have been reported (*33*) for the simultaneous adsorption of

TABLE 2. ADSORPTION EQUILIBRIA: COMPARISON OF OBSERVED WITH PREDICTED VALUES

Binary Systen: Methane–Acetylene: Pressure = 1 Atm.

Run	Temp., °F.	Component	Gas composition, %*	Adsorbate amount, Std. cu. ft./100 g. of carbon* Adsorption †	Desorption ‡	Predicted **	Predicted ¶
1	50	Methane	32.0	0.059	0.036	0.075	0.032
		Acetylene	68.0	0.241	0.244	0.280	0.294
		Total	100.0	0.300	0.280	0.355	0.326
2	45	Methane	31.4	0.048	0.043	0.082	0.032
		Acetylene	68.6	0.254	0.262	0.276	0.304
		Total	100.0	0.302	0.305	0.358	0.336
3	47	Methane	54.5	0.083	0.076	0.101	0.058
		Acetylene	45.5	0.196	0.204	0.244	0.213
		Total	100.0	0.279	0.280	0.345	0.271
4	55	Methane	54.5	0.045	0.050	0.097	0.056
		Acetylene	45.5	0.190	0.188	0.222	0.210
		Total	100.0	0.235	0.238	0.319	0.266
5	51	Methane	87.6	0.069	0.113	0.149	0.101
		Acetylene	12.4	0.051	0.071	0.154	0.063
		Total	100.0	0.120	0.184	0.303	0.164
6	51	Methane	87.7	0.082	0.113	0.149	0.102
		Acetylene	12.3	0.048	0.071	0.154	0.063
		Total	100.0	0.130	0.184	0.303	0.165

*Standard cubic feet of gas measured at 60°F., and 1 atm.
†Adsorbate amount and composition calculated from data on adsorption cycle. (*See* text.)
‡Adsorbate amount and composition calculated from data on desorption cycle. (*See* text.)
**Adsorbate amount and composition calculated by method I: Adsorption Equilibria, based on Polanyi theory. (*See* text.)
¶Adsorbate amount and composition calculated by methodI I: Adsorption Equilibria, based on method of Lewis et al.

TABLE 3. ADSORPTION EQUILIBRIA: COMPARISON OF OBSERVED WITH PREDICTED VALUES

Ternary System: Methane–Carbon Dioxide–Acetylene at 1 Atm. and 100°F.

Run	Component	Gas composition, %*	Adsorbate Amount, mmoles/g. of carbon Experimental	Method I†	Method II‡
1	Methane	55.1	0.636	0.645	0.610
	Carbon dioxide	33.8	0.530	1.080	0.544
	Acetylene	11.1	0.514	0.880	0.490
	Total	100.0	1.680	2.605	1.644
2	Methane	41.7	0.376	0.52	0.444
	Carbon dioxide	33.3	0.570	1.09	0.510
	Acetylene	25.0	1.234	1.50	1.050
	Total	100.0	2.180	3.10	2.000
3	Methane	32.0	0.339	0.430	0.376
	Carbon dioxide	56.4	1.027	1.480	1.000
	Acetylene	11.6	0.574	0.900	0.545
	Total	100.0	1.940	2.810	1.921

*Gas composition as mole per cent.
†Calculated from Method I: Adsorption Equilibrium, based on Polanyi theory (*29*). (*See* text.)
‡Calculated from Method II: Adsorption Equilibrium, based on empirical correlation of Lewis et al. (*25*). (*See* text.)

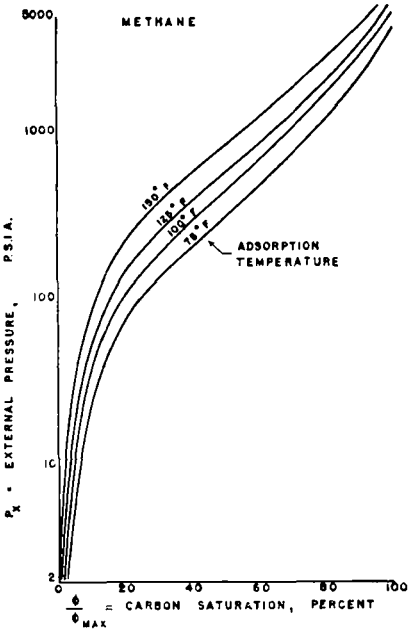

Fig. 12. Adsorption vapor pressure—methane.

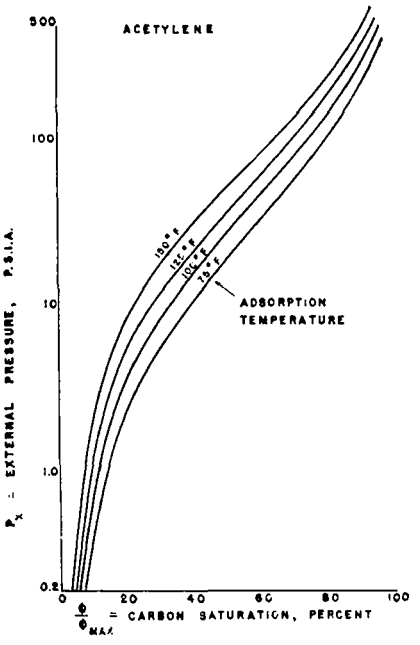

Fig. 13. Adsorption vapor pressure—acetylene.

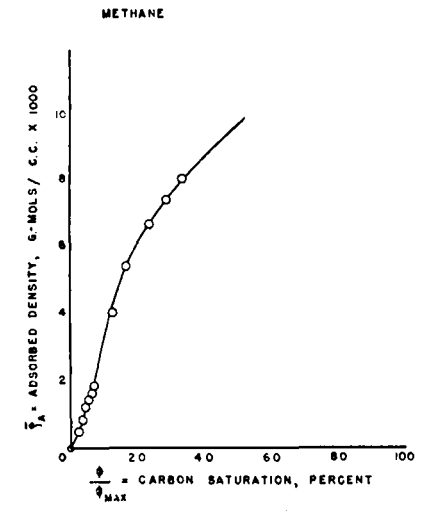

Fig. 14. Methane—adsorbed density at 100°F.

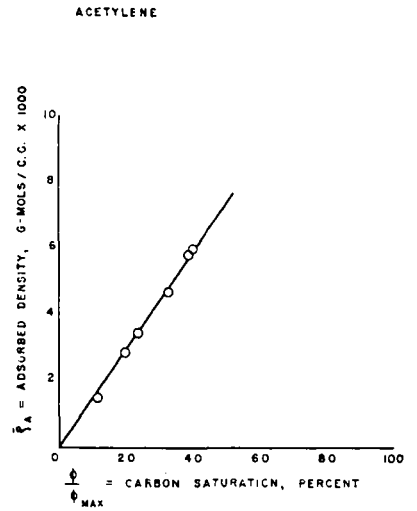

Fig. 15. Acetylene—adsorbed density at 100°F.

ethylene and propylene on carbon, where measurements showed that, even at high concentration of ethylene, the amount of propylene adsorbed was hardly affected, but small concentrations of propylene depressed the ethylene adsorption. Since only the acetylene adsorption formed the basis of the later analysis of the binary methane-acetylene separation in the moving-bed adsorber, the preceding method was considered to be acceptable in this case as a convenient means of obtaining consistent equilibrium data.

METHOD II: TERNARY-SYSTEM EQUILIBRIA

When method I, based on a modification of the Polanyi theory (*29* and *30*) and coupled with the assumption of independence of gas composition on the quantity of each component adsorbed, was tested on the ternary system methane–carbon dioxide–acetylene, it was found that in certain cases relatively large deviations between the predicted and the experimentally determined equilibria resulted. Accordingly, attention was turned to the correlations developed by Lewis, Gilliland, Chertow, and Cadogan (*24* and *25*). Based on these correlations as well as on certain experimental data on adsorption relative volatility (*26*), a method of predicting the adsorption equilibria satisfactory for the present system was developed and checked with a limited number of experimental equilibrium measurements. As shown in Table 3, the predicted results obtained by method II were in better agreement with the experimental equilibrium data than those predicted by method I. The reasoning behind method II is outlined briefly below.

Relative volatility, defined as the mole ratio of the more volatile to the less volatile components in the vapor divided by the mole ratio of the same components in the adsorbed phase, was noted to be fairly constant with respect to composition for a given system. In 80% of the cases studied by Lewis and coworkers (*24*), the vapor-adsorbate composition diagram obtained by using the relative volatility derived from a single experimental determination gave values of vapor composition for the various values of absorbate composition which were within 5% of the experimental values.

The relative volatilities of a multicomponent system appear to be equal to the values obtained from a binary mixture of the gases under question. Relative volatilities of 1.45 and 1.46 were found for the binary system: methane–carbon dioxide, as compared with an average value of 1.66 for methane–carbon dioxide in the ternary mixture. Likewise for the carbon dioxide–acetylene system, relative volatilities of 2.69 and 2.80 (average 2.75) were found for the binary combination, as compared with an average of 2.85 for the same components in a ternary mixture with methane. Based on ternary runs 1, 2, and 3 (Table 3) the average relative volatility for methane-acetylene is equal to 1.66 × 2.85, or 4.74, which compares favorably with the grand average value of 4.65 for these components calculated from both the adsorption and desorption data for the binary system.

The variation of the total quantity adsorbed was found (*24*) to be nonlinear with composition expressed as mole fraction in the adsorbate, but the data indicated a satisfactory correlation with the expression

$$\frac{n_1}{n_1'} + \frac{n_2}{n_2'} + \cdots + \frac{n_n}{n_n'} = 1 \quad (11)$$

where n_1, n_2, . . . n_n are the quantities of each component actually adsorbed from the mixture, and n_1', and n_2', . . . n_n' are the amounts of the pure components adsorbed at the same total pressure and temperature. The form of this equation may also be predicted from considerations based on Raoult's law (*26*).

According to this relationship the sum of the ratios of each component adsorbed to the quantity of that component which would be adsorbed if present alone at the total pressure and temperature should be unity. The values of n' were obtained by experimental determination of the quantity of each component adsorbed at 1 atm. and 100°F. For the ternary runs the value of n/n' for runs 1, 2, and 3 are 1.028, 0.988, and 1.024 respectively. For the binary runs 4, 5, 6, and 7 the values are 1.085, 1.031, 0.907, and 0.991 respectively. Thus the correlation can be con-

sidered satisfactory for the present mixture.

For the purpose of correlation and prediction of equilibrium data for the ternary system methane–carbon dioxide–acetylene, the Lewis et al. (*24*) method was employed in the following manner. The equation

$$\frac{n_1}{n_1'} + \frac{n_2}{n_2'} + \frac{n_3}{n_3'} = 1 \qquad (12)$$

was used. In this and subsequent equations for the ternary system the subscripts 1, 2, and 3 refer to methane, carbon dioxide, and acetylene respectively. n_1, n_2, and n_3 represent milligram moles of each component adsorbed from the mixture per gram of carbon at temperature T and total pressure π. The primed values refer to the milligram moles of each pure component which would be adsorbed at the same temperature and total pressure. In addition the adsorption equilibrium relative volatilities are defined as follows:

$$\alpha_{12} = \frac{y_1}{y_2}\cdot\frac{x_2}{x_1} \qquad (13)$$

$$\alpha_{23} = \frac{y_2}{y_3}\cdot\frac{x_3}{x_2} \qquad (14)$$

$$\alpha_{13} = \frac{y_1}{y_3}\cdot\frac{x_3}{x_1} = \frac{y_1}{y_2}\cdot\frac{x_2}{x_1}\cdot\frac{y_2}{y_3}\cdot\frac{x_3}{x_2} \qquad (15)$$

and

$$x_1 = \frac{n_1}{n_1 + n_2 + n_3} \qquad (16)$$

where x_1 is the mole fraction of methane in the adsorbate. Therefore, by substituting Equation (16) into Equation (13) one obtains

$$\alpha_{12} = \frac{y_1}{y_2}\cdot\frac{n_2}{n_1} \qquad (17)$$

and similar results for the other components. Also, solving Equation (17) for n_2, gives

$$n_2 = \alpha_{12}\frac{y_2}{y_1}\cdot n_1 \qquad (18)$$

Similarly,

$$n_3 = \alpha_{13}\frac{y_3}{y_1}\cdot n_1 \qquad (19)$$

Solving for n_2 and n_3, and substituting in (12) one obtains

$$\frac{n_1}{n_1'} + \alpha_{12}\frac{y_2}{y_1}\cdot\frac{n_1}{n_2'} + \alpha_{13}\frac{y_3}{y_1}\cdot\frac{n_1}{n_3'} = 1 \qquad (20)$$

For a given vapor mixture y_2/y_1 and y_3/y_1 are known, and n_1', n_2' and n_3' can be obtained from a plot of the quantity of each component adsorbed vs. temperature at 1-atm. pressure. Equation (20) can thereby be solved for n_1, the quantity of methane adsorbed from the gas mixture. The quantity of carbon dioxide, n_2, and acetylene, n_3, adsorbed can then be obtained from Equations (18) and (19) respectively. Thus the quantity of each component adsorbed can be determined from the gas composition when the relative volatilities and isotherms or isobars of the various components are known.

The applicability of this method for the prediction of adsorption data for the ternary system is shown under method II, Table 3. It is observed that the quantity and composition obtained by use of this method agrees well with the experimental data and is superior to the modified Polanyi method. The adsorbate quantities and compositions listed under method II were obtained by use of the binary-system relative volatilities. The good agreement of these results with the experimental data show that binary relative volatilities may be used to calculate ternary equilibria.

ADSORPTION EQUILIBRIA

Experimental measurements of adsorption equilibrium data were made with a flow or dynamic method with the apparatus shown in Figure 5. A sample of carbon was first treated with steam at 212°F. for at least 1 hr., and then air was dried in a steam-jacketed tube for approximately 2 hr. This procedure served to remove any residual adsorbed hydrocarbons. From this point on, two slightly different techniques were employed.

In the first case the carbon was further dried in an oven for 2 hr. at 400°F., and a known weight placed in the upper or adsorption chamber of the aforementioned apparatus. Water maintained at the desired temperature was passed through the surrounding jacket, and a metered quantity of the gas mixture to be adsorbed was allowed to flow up through the carbon bed. The saturation of the carbon by the gas mixture could be followed by observing the rise in bed temperature as indicated by the two iron-constantan thermocouples located in the carbon sample. After the heat wave accompanying the adsorption had passed the upper thermocouple, an additional 30 to 45 min. was allowed with full gas flow and, as a further check, chemical analysis of the exit-gas stream was made in order to see that it was identical with the feed-gas composition, thus ensuring that equilibrium had been attained. The total quantity of exit gas was metered and collected in a brine-sealed atmospheric-pressure gas holder. This mixture in the gas holder was then sampled and analyzed, whence from a knowledge of the total input and output quantities, a material balance for each component could be made. In this manner the amount and composition of the adsorbate in equilibrium with a given feed gas at a specific temperature and pressure could be calculated. When this analysis was complete, the saturated-carbon sample was transferred to the lower chamber of the apparatus (Figure 5) by opening the plug valve. Superheated-steam was then admitted into the bottom of the carbon bed, and the desorbed gas was metered and collected in the brine-sealed gas holder. Analysis of the desorbed gas was then made, to provide a check on the amount and composition of the adsorbate, which had been determined previously by a material balance in the adsorption step. Adsorption-equilibrium data obtained in this manner for the binary system methane-acetylene at atmospheric pressure and at temperatures ranging from 45° to 55°F. are summarized in Table 2, where they are compared with equilibria estimated by a modification of the Polanyi (*29* and *30*) equations as described previously under Method I: Adsorption Equilibria.

The second technique, which was employed for measuring the adsorption equilibria of the ternary system methane–carbon dioxide–acetylene as well as for the corresponding binary systems, was essentially the same as the one just described, except that the carbon sample, after air drying, was dried further in a vacuum oven at 400°F. for 3 to 4 hr. After cooling, the vacuum was broken by the admission of helium, which is not adsorbed under these conditions. The carbon sample was then weighed and charged to the adsorption section of the apparatus (Figure 5), which had previously been filled with helium or, in those runs where methane was a component, with methane. In this manner the amount of air contained on the carbon and in the apparatus was greatly reduced. In these cases a known feed gas was allowed to pass over the carbon until equilibrium had been reached in the same manner as described above. The saturated-carbon sample was then removed from the apparatus shown in Figure 5 and transferred to the apparatus shown in Figure 6, where it was stripped with superheated steam generated from freshly boiled distilled water. The desorbed gases were cooled and collected over a saturated-brine solution. The adsorbate amount and composition were then determined by chemical analysis of the known quantity of the desorbed gases. Adsorption-equilibrium data obtained by use of the second technique are summarized in Table 3, where they are compared with equilibrium estimated by a modification of the Lewis, Gilliland, Chertow, and Cadogan, correlation (*24* and *25*) described previously under Method II: Adsorption Equilibria. The systems methane–carbon dioxide, carbon dioxide–acetylene, and methane–carbon dioxide–acetylene were all investigated atmospheric pressure and at temperatures ranging from 77° to 100°F.

CONCLUSION

Two methods are described for estimating adsorption equilibria, one a modification of the Polanyi adsorption potential theory (*29* and *30*) and the other based on the recently published method of Lewis, Gilliland, Chertow, and Cadogan, (*24* and *25*). The latter method appears to be superior, the predicted equilibria being more consistent with limited experimental checks for both the binary and the ternary systems which were investigated.

ACKNOWLEDGMENT

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NOTATION

a = constant in van der Waal's equation of state

D = diffusion coefficient, sq. ft./hr.

D_p = particle diameter, ft.

f_0 = fugacity of a gas or vapor, corresponding to a vapor pressure P_0, lb./sq. in. abs.

f_x = fugacity of a gas or vapor corresponding to an external gas pressure, P_x, over an adsorbent, lb./sq. in. abs.

G = gas rate, cu. ft./hr. measured at 60°F., 1 atm.

G' = mass velocity, lb./(hr.)(sq. ft.)

G_M' = molar mass velocity, lb.-moles/(hr.)(sq. ft.)

h = height of a tower section, in.

H_A = individual transfer-unit height based on adsorbed phase, in.

H_{OA} = over-all transfer-unit height based on adsorbed phase, in.

H_G = individual transfer-unit height based on gas phase, in.

H_{OG} = over-all transfer-unit height based on gas phase, in.

j_M = mass transfer factor

k_G' = mass transfer coefficient, lb.-moles/(hr./sq. ft.)(mole fraction)

K = a constant

L = carbon rate, lb./hr.

m = average slope of equilibrium curve plotted on a phase-composition diagram

n = quantity of a gas component adsorbed from a gas mixture, g.-moles

n' = quantity of a pure gas component adsorbed at same total pressure and temperature as n, g.-moles

N_{OA} = number of transfer units, over-all, based on adsorbed phase

N_{OG} = number of transfer units, over-all, based on adsorbed gas phase

P = pressure, lb./sq. in. abs.

P_o = vapor pressure of a pure substance at temperature T, lb./sq. in. abs.

P_X = external equilibrium pressure of gas in contact with an adsorbent

R = gas constant, in consistent units, usually 1.987 cal./(g.-mole)(°K.)

s = external surface area of a particle, sq. ft./cu. ft.

T = absolute temperature, °K.

V = molar volume of a saturated vapor at temperature, T.

x = mole fraction of a component in an adsorbed phase consisting of more than one component

y = mole fraction of a component in a mixture of gases over an adsorbent.

Greek

α = adsorption relative volatility, defined by Equation (8)

ϵ = Polanyi adsorption potential, cal./g.-mole

ϵ_i = adsorption potential at point i in a potential field

π = total gas pressure, lb./sq. in. abs.

ρ = density, g./cc. or g.-moles/cc.

ρ_a = average adsorbate density based on maximum adsorbent pore volume, ϕ_{max}, if it is assumed that the adsorbate completely occupies the volume, ϕ_{max}. Units g.-moles/cc.

μ = viscosity, lb. (mass)/(ft.)(hr.)

ϕ_i = volume of adsorbent occupied by an adsorbate, considered as a liquid after Polyani (*24*), corresponding to an adsorption potential i; units, ml.

ϕ_{max} = maximum value of adsorbent pore volume, also called *activity*. units: cc./g. of adsorbent.

ϕ/ϕ_{max} = ratio of volume occupied by a given amount of adsorbate, considered as a highly compressed fluid, to the total available pore volume, ϕ_{max}. When converted to a percentage basis, this quantity is called adsorbent per cent saturation.

Subscripts

1, 2 . . . = components 1, 2, . . . in a mixture

A = adsorbed phase

G = gas phase

O = over-all

T = absolute temperature, T

g = gas-film conditions

i = point i in a field

max = maximum

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