ABSORBER CALCULATIONS

USING ABSORPTION AND STRIPPING - FACTOR FUNCTIONS

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PREFACE

Calculation procedures for the design and evaluation of multicomponent hydrocarbon absorption systems have been a problem in the petroleum industry for many years. A recent system of calculations using absorption and stripping-factor functions is reviewed and tested, and applications are proposed.

A computer program has been developed using a medium capacity digital computer to test the reliability of this absorption and stripping calculation technique. The results are compared with results from another program of known accuracy to check the accuracy of the technique.

The calculation procedure, it is hoped, will assist in absorber design and evaluation work and provide a starting point for the design and evaluation of rich oil fractionators, stills and other complex systems.

The advice and guidance offered by Dr. Robert N. Maddox and the programming, check-out, and computer operational assistance of Messrs. W. A. Fling, Jr., and L. E. Williams are very warmly acknowledged.

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CHAPTER I

PROBLEM

Mass transfer phenomena in the form of gas absorption for the separation of desirable products have been employed for many years. Since its introduction in multi-component hydrocarbon systems in the early 1920's, there has been a continuing effort to develop calculation procedures that describe the complex systems involved. The desire for fast, accurate design and evaluation techniques is spurred by the continued gain in popularity of the absorption process. Today it is employed in essentially all natural gas processing plants where natural gas liquids are removed from a given stream.

During the intervening years, there have been improvements in absorption equipment to provide more consistent performance, in analytical methods to provide more reliable tests data, in equilibrium and enthalpy data to provide more accuracy in calculated results, in computation equipment to provide faster and more accurate solutions, and in calculation procedures to provide the best application of data and equipment available.

Improvements in design techniques have been hampered by the inherent difficulties in evaluating operating systems. Whether the process recovers chlorine in a packed tower or propane in a multitray system, the problems of process evaluation are well known. Unsteady state conditions, sampling and analytical problems, accuracy of test apparatus, and other difficulties always cast a shadow of doubt on evaluation results.

Absorption equipment has become more complex over the years with the introduction of intercoolers, reboilers, lean oil saturators, intermediate feed streams, and other means of bringing oil and gas together in the most effective manner. Changes in lean oil characterisitics, temperature level, pressure level, and other operating innovations have also been made, but these have not complicated the calculation procedure.

In recent years larger and faster computer systems have been established by many companies. These computers have led to an increased emphasis on accuracy from various calculation procedures which usually requires rigorous calculation techniques. At the same time, however, there is still need for rapid, accurate, short-cut calculation procedures.

Today, with the availability of high speed, large capacity computers, an ideal calculation procedure must be readily adaptable to computer solutions, applicable to complex processing systems, and capable of handling components from hydrogen to heavy crude oil fractions.

The primary purpose of this research is to review the Edmister absorber calculation method (5) which introduces absorption and stripping-factor functions, to test the reliability of its results, to check its accuracy by comparing it with another calculation procedure, and to propose applications for its use.

A secondary purpose of the research is to review the state of the art of equipment testing for operational evaluation to see if theoretical calculations with a small amount of testing might be a better approach.

CHAPTER II

SCOPE

The absorber calculation procedure using absorption and stripping-factor functions is an algebraic treatment of the absorption and stripping process. It is compared with other algebraic techniques that have been presented through the years. Although graphical solutions (12) have aided many students of absorption to visualize the different relations involved, this research is limited to algebraic calculation methods.

All calculations are limited to a simple absorber without intercoolers, reboilers, or other appurtenances. Also, all problems have vapor feed so that feed flash calculations are not required.

All systems in this study have a large number of components, especially in the lean oil. The computer program used for evaluating the systems is limited to twenty components which assumes that the lean oil characteristics can be adequately described within the twenty component limit.

Equilibrium and enthalpy data were obtained from commonly used sources (11, 9) and were applied as recommended. The calculated results indicate obvious discrepancies, but the results are adequate to analyze the calculation procedure which is the primary purpose.

The test information from the NGAA Absorption Data (16) is inconsistent and incomplete in some cases. The results are sufficiently consistent, and they correlate with adequate accuracy to be conclusive.

The computer program is not a sophisticated approach to this calculation procedure, but the program, as listed in the appendix, will provide fast, accurate answers to most problems.

CHAPTER III

LITERATURE SURVEY

Absorber calculations that describe multi-component multi-tray hydrocarbon systems were first proposed by Kremser (10) in 1930. The primary product of absorption at that time was gasoline, and the primary analytical method was the charcoal adsorption test. Component vapor pressure data were used for equilibrium relations assuming ideality. Pure component specific heats and heats of vaporization at atmospheric pressure were used for heat balances. The application of the procedure was not a complete description of the system, but it served the need of the day. The same basic calculation scheme with improvements is still in use.

Souders and Brown (13) modified the Kremser method in 1932 improving the ease of application. The absorption factor was defined and put in usable terms. The revised method calculated the equilibrium of a key component, usually normal butane, and related it to gasoline recovery. This so-called Kremser-Brown method was used extensively until the early 1950's.

In 1940 Horton and Franklin (6) improved the accuracy of the procedure by making a better approach to the effective point in an absorber. Up to that time, the "average" column conditions were used for the equilibrium constants, and this "average" was at the discretion of the user. The accuracy of the method was also improved by better equilibrium and analytical data. The assumption of ideality for equilibrium data was eliminated, and the low temperature fractional analysis improved test data. The method also recognized the effect of lean oil saturation.

The Horton and Franklin method was to be applied to each theoretical tray which made it slow and tedious. It was later modifed to improve speed of solution, but the faster method was only intended to be a starting point for the tray by tray approach. This method made a rigorous calculation possible, and at the same time, analytical and equilibrium data were of sufficient accuracy to justify the time required to make the calculations with a desk calculator.

Edmister (2) reviewed each of the previous methods and proposed his "short-cut" method in 1943. It combined the simplicity of the Kremser-Brown method with the accuracy of Horton-Franklin. It was arranged for calculation with a desk calculator, and it is in general use today for both design and evaluation work. It has also been adapted to computer calculations to some extent.

In 1947, the Edmister procedure was reviewed again (3). At that time graphical solutions were published for some of the intermediate equations. This procedure includes a correction for the effect of lean oil saturation, and it gave the industry a basis of evaluating the economics of lean oil stripping operations. This modified Edmister procedure was used throughout the industry by 1950, and it is still used today for quick estimates to absorption and stripping problems.

One of the best absorber calculation methods was presented in 1953 by Hull and Raymond (7). This method gives good accuracy for unusual absorber conditions. It is used exclusively with desk calculators because many curves and tables are required which are not adaptable to computers.

The Edmister procedure was modified in 1957 (5) introducing absorption and stripping-factor functions. The revised method is more adaptable to computer solutions, and it is more readily applicable to complex systems. This procedure has been reviewed, tested, and evaluated. Computer applications are proposed for its use in design, operations, and evaluation work.

Absorption data published by the Absorption Subcommittee of the Natural Gasoline Association of America in 1954 (16) were used for evaluating the new procedure.

Although other calculation procedures have been proposed recently (14), the modified Edmister procedure was felt to have greater promise for short-cut absorber design and evaluation calculations.

CHAPTER IV

DERIVATION OF ABSORBER EQUATIONS

In absorber calculations, the most accurate results are obtained when the system being studied operates at atmospheric pressure and temperature. These are the conditions where the ideal gas laws apply and where the best equilibrium and enthalpy data are available. As operating pressures are increased, the behavior of most components deviate from ideal, and the degree of deviation in multicomponent hydrocarbon systems is not accurately known.

It is appropriate then to begin the derivation of the equations utilized in this study with Raoult's Law which states that the partial pressure of an absorbed component is equal to its vapor pressure times its mol fraction in the liquid. It may be written:

$$P_p = P_v x$$

According to Dalton's Law, the partial pressure of a component is also equal to the mol fraction of the component in the vapor times the total system pressure. Thus Dalton's Law may be added:

$$P_p = P_v x = P y \qquad (1)$$

Rearranging equation (1) provides the relation which defines the equilibrium constant, K, for a given component.

$$K = \frac{y}{x} = \frac{P_v}{P} \qquad (2)$$

The vapor pressure-total pressure ratio defines the slope of the operating curve in graphical solutions. Souders and Brown (13) introduced the K-factor, and it is used universally today.

The mols of any component in the liquid, or vapor, are related to the total volume of liquid and vapor as mol fractions:

$$y = \frac{v}{V} \quad ; \quad x = \frac{l}{L}$$

Substituting these relations into equation (2), the equilibrium constant with this nomenclature becomes:

$$K = \frac{v \quad L}{V \quad l}$$

Rearranging and solving for l:

$$l = v \frac{L}{VK} \tag{3}$$

Or, solving for v:

$$v = l \frac{VK}{L} \tag{4}$$

Souders and Brown defined the absorption factor and stripping factor with relations similar to equations (3) and (4).

$$A = \frac{L}{VK} \quad ; \quad S = \frac{VK}{L}$$

Assuming that the theoretical tray concept holds for absorbers, equations (3) and (4) become the tray equilibrium relations for any theoretical tray (15).

$$l = v\,A \qquad (5)$$

$$v = l\,S \qquad (6)$$

Referring to Figure 1, a component material balance around the top tray of the absorber may be written:

$$v_2 + l_o = l_1 + v_1 \qquad (7)$$

Introducing the equilibrium relation for tray 1 from equation (5) and solving for v_2, equation (7) becomes:

$$v_2 = v_1\,(A_1 + 1) - l_o \qquad (8)$$

Equation (7) can be used to determine the liquid composition from tray 2 by substituting the tray 2 equilibrium and solving for l_2:

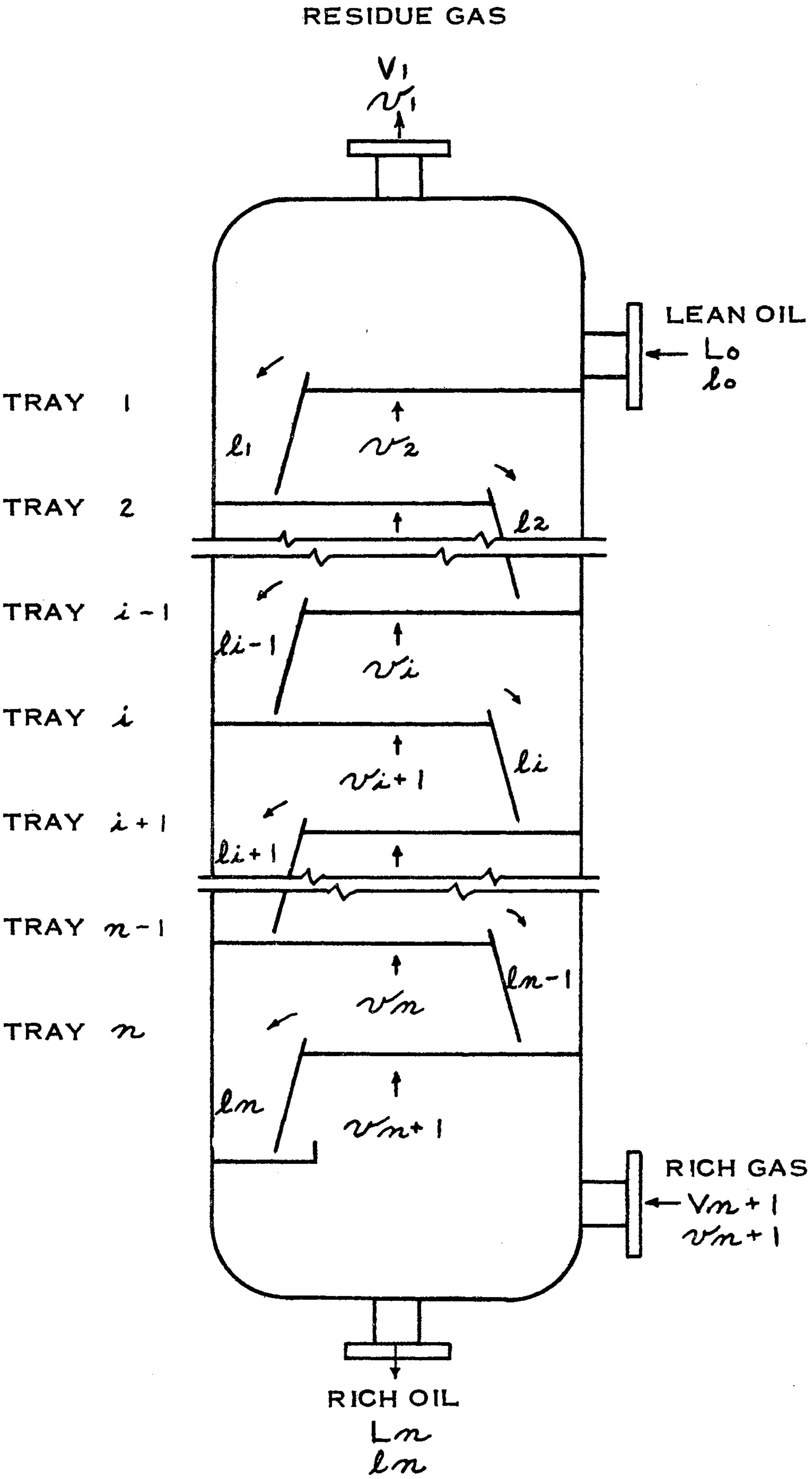
$$l_2 = v_1\,(A_1\,A_2 + A_2) - l_o\,A_2 \qquad (9)$$

A component material balance around the top two trays may be written:

$$v_3 + l_o = v_1 + l_2 \qquad (10)$$

By substituting equation (9) into equation (10) for l_2 and solving for v_3, an equation is obtained which is the same form as equation (8).

$$v_3 = v_1\,(A_1\,A_2 + A_2 + 1) - l_o\,(A_2 + 1) \qquad (11)$$



SIMPLE ABSORBER

FIGURE 1

Substituting tray 3 equilibrium into equation (11) and solving for l_3, the rich oil leaving tray 3 is obtained.

$$l_3 = v_1 \ (A_1 A_2 A_3 + A_2 A_3 + A_3) - l_o (A_2 A_3 + A_3) \qquad (12)$$

A component balance around the top of the absorber down to tray i may be written:

$$l_i + v_1 = l_o + v_{i+1} \qquad (13)$$

By substituting the equilibrium relation for tray i + 1 and solving for l_{i+1},

$$l_{i+1} = (l_i + v_1 - l_o) \ A_{i+1} \qquad (14)$$

Equation (14) is the same form as equation (9) for a two tray absorber and, continuing equations of this type for n trays, the rich oil stream can be determined as follows:

$$l_n = v_1 \ (A_1 A_2 A_3 \ldots A_n + A_2 A_3 \ldots A_n + \ldots + A_n)$$
$$- l_o (A_2 A_3 \ldots A_n + A_3 \ldots A_n + \ldots + A_n) \qquad (15)$$

To simplify equation (15) and the various relations for evaluating it, Edmister (5) defined two "recovery fraction" terms as follows:

$$\sum A = A_1 A_2 A_3 \ldots A_n + A_2 A_3 \ldots A_n + \ldots A_n$$

$$\Pi = A_1 A_2 A_3 \ldots A_n$$

Introducing these terms, equation (15) becomes:

$$l_n = v_1 \sum A - l_o \ (\sum A - \Pi) \qquad (16)$$

Equation (16) is the same basic relation presented by Kremser (10) with subsequent modifications by Souders and Brown (13), by Horton and Franklin (6), and by Edmister (2,5). By combining equation (16) with the overall material balance and solving for v_1, an equation is obtained that relates two normally known streams, rich gas and lean oil, with equilibrium conditions through the column to determine one of the terminal streams, residue gas.

$$v_1 = v_{n+1}\left(\frac{1}{\sum A+1}\right) + l_o\left[1 - \left(\frac{\Pi}{\sum A+1}\right)\right] \qquad (17)$$

A similar relation can be developed to determine the rich oil stream, but it is normally obtained by difference from the material balance.

Edmister (5) then defined absorption and stripping-factor functions which represent the fraction <u>not</u> absorbed and the fraction <u>not</u> stripped. The absorption-factor function is equal to one minus the fraction absorbed, and the stripping factor-function is equal to one minus the fraction stripped.

$$\phi_a = 1 - f_a \quad ; \quad \phi_s = 1 - f_s$$

The fractions absorbed and stripped are evaluated as follows:

$$f_a = \frac{\sum A}{\sum A+1} \quad ; \quad f_s = \frac{\sum S}{\sum S+1} \qquad (18)$$

Rearranging the expressions for the absorption and stripping-factor functions,

$$\phi_a = \frac{1}{\sum A+1} \quad ; \quad \phi_s = \frac{1}{\sum S+1} \qquad (19)$$

Similar relations can be used to develop equations for determining the rich oil stream instead of the residue gas.

$$l_n = v_{n+1} (1 - \emptyset_a) + l_o \emptyset_s \tag{22}$$

To help differentiate between absorption and stripping factors, effective absorption and stripping factors, and absorption and stripping-factor functions, the absorption and stripping-factor functions are sometimes referred to as "phi-factors" or "phi-factor functions." These terms are used here for clarification and easier understanding by the reader.

This is a rigorous calculation procedure as long as the absorption and stripping factors are evaluated on each try. This research, however, used the short-cut approach to the evaluation of the absorption and stripping-factor functions.

Edmister presented his short-cut method in 1943 at which time he defined effective absorption and stripping factors as follows:

$$\frac{A_e^{n+1} - A_e}{A_e^{n+1} - 1} = \frac{\sum A}{\sum A + 1}$$

$$\frac{S_e^{m+1} - S_e}{S_e^{m+1} - 1} = \frac{\sum S}{\sum S + 1}$$

The effective factors are mean values of the absorption and stripping factors that will give the same values for fraction absorbed or fraction stripped that would be obtained by using absorption and stripping factors on each tray. In relation to the absorption and stripping-factor functions, the definitions for effective factors apply as follows:

$$\phi_a = \frac{1}{A_e^{n} + A_e^{n-1} + \ldots + A_e^{2} + A_e + 1} = \frac{A_e - 1}{A_e^{n+1} - 1}$$

$$\phi_s = \frac{1}{S_e^{m} + S_e^{m-1} + \ldots + S_e^{2} + S_e + 1} = \frac{S_e - 1}{S_e^{m+1} - 1}$$

Figure 2 shows a plot of the effective absorption and stripping factors as functions of the absorption and stripping-factor functions and number of theoretical trays. The effective factors are shown on a logarithmic scale with the "phi-factor" functions on a probability scale. If rectangular coordinates were used for both factors, the plot would be similar to the well known Kremser-Brown plot of absorption factors as functions of fraction absorbed and number of theoretical trays.

Edmister derived equations for the evaluation of effective factors using a system with two theoretical trays. This approach assumes that the top and bottom theoretical trays can be related to describe any multi-tray section. The derivation results in the following relations.

$$A_e = \sqrt{A_n\ (A_1 + 1) + 0.25} - 0.5 \qquad (23)$$

$$S_e = \sqrt{S_m\ (S_1 + 1) + 0.25} - 0.5 \qquad (24)$$

These relations have been used extensively since the late 1940's, especially for desk calculations. In most absorbers, the number of actual

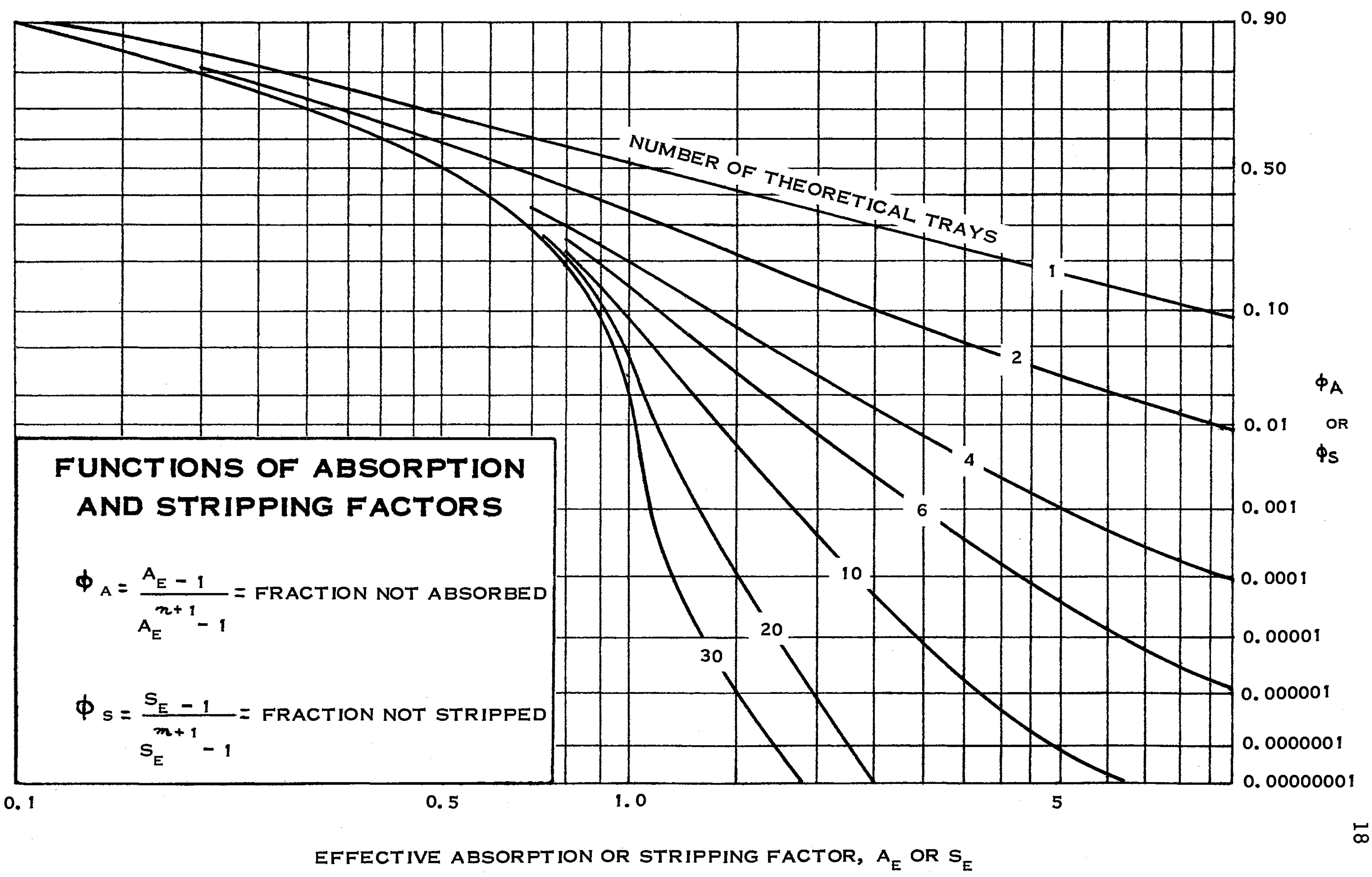

NUMBER OF THEORETICAL TRAYS
1
2
4
6
10
20
30
FUNCTIONS OF ABSORPTION AND STRIPPING FACTORS
$\phi_A = \frac{A_E - 1}{A_E^{m+1} - 1}$ = FRACTION NOT ABSORBED
$\phi_S = \frac{S_E - 1}{S_E^{m+1} - 1}$ = FRACTION NOT STRIPPED
0. 90
0. 50
0. 10
0. 01
0. 001
0. 0001
0. 00001
0. 000001
0. 0000001
0. 00000001
ϕ_A OR ϕ_S
0. 1
0. 5
1. 0
5
EFFECTIVE ABSORPTION OR STRIPPING FACTOR, A_E OR S_E

FIGURE 3

trays in each section is small, and the tray efficiencies are low. The resulting number of theoretical trays is almost always small which makes the "two-tray" approach quite accurate for most systems. This approach has been utilized with the "phi-factor" function procedure throughout this research.

CHAPTER V

COMPUTER PROGRAM

The computer program which was developed to evaluate the absorption and stripping-factor function method follows a calculation routine that is similar to the approach an engineer would make with the same problem at a desk calculator. Data requirements are essentially the same, also, including (1) number of theoretical trays, (2) lean oil and rich gas compositions, (3) lean oil and rich gas temperatures, (4) equilibrium and enthalpy data for each component, and (5) recovery of key component.

The program is set up as a design procedure which calculates the quantity of lean oil required to obtain the specified recovery of key component at the specified conditions of the system. This program is suitable for evaluation of test data also, since the recovery of key component is normally known, and the number of theoretical trays is the variable which controls the distribution of components other than key.

The computer program was developed by preparing a Fortran listing and compiling for the IBM 650 computer using a compilation program developed by IBM and modified by Janicek (8). The resulting program is not an efficient programming effort. The calculation procedure and results are of primary interest--not the evaluation method.

Equilibrium and enthalpy data are curve fitted outside of this program and only the equation constants are introduced. Equilibrium constants are calculated with a third degree polynomial, and the vapor and liquid enthalpies are calculated with a second degree polynomial. In each case, values at increments of 5°F over the operating range of the problem were used for the curve fit computations.

The computer program can be considered in four general parts--preliminary calculations, absorption-stripping relations, heat balance, and final relationships. Figure 3 shows the schematic sequence of the program. Results include lean oil requirements, component distributions, and terminal temperatures for a specified recovery of key component.

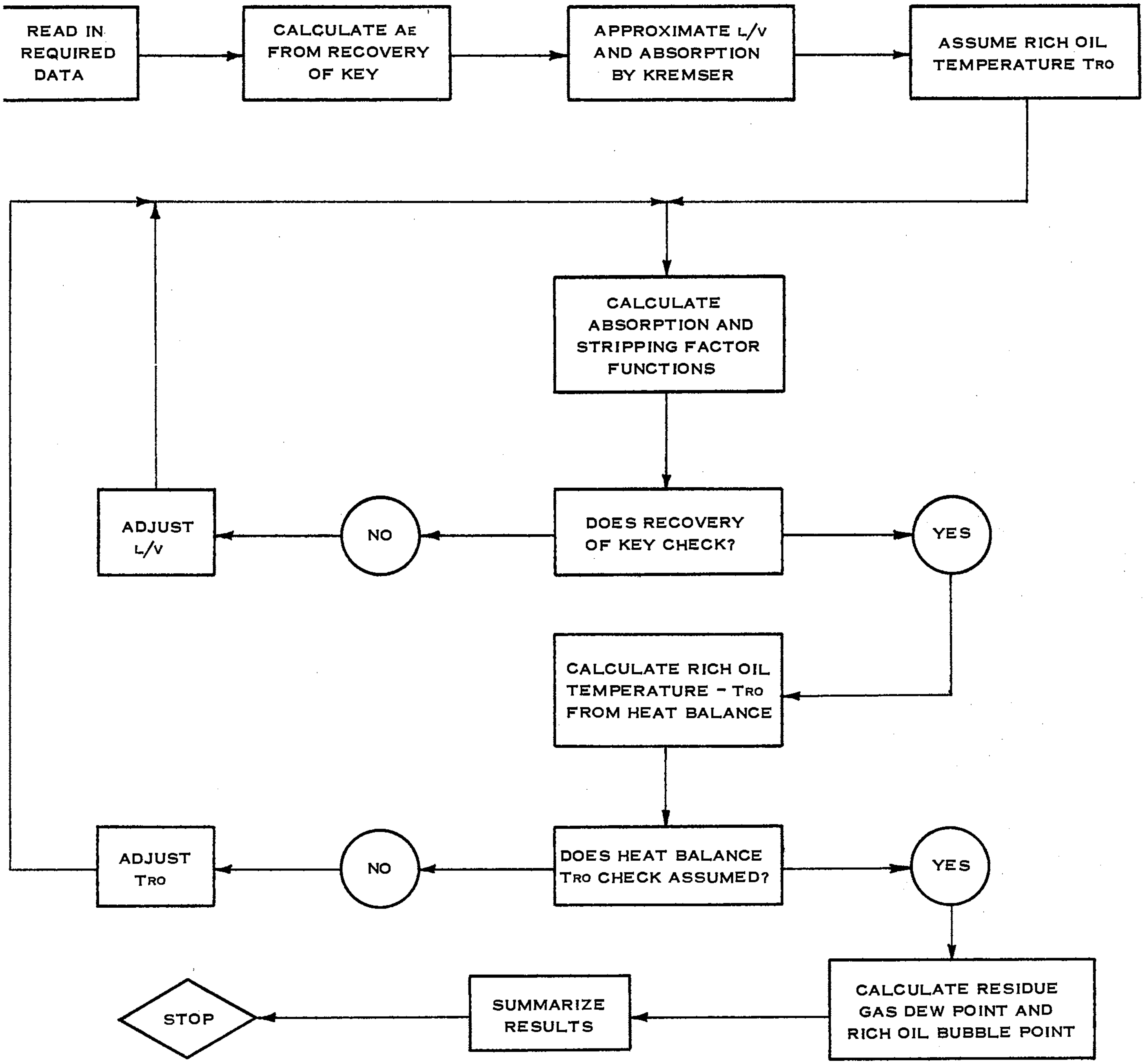
PRELIMINARY CALCULATIONS

The preliminary calculations section resolves the input data and specifications into a first estimate for the absorption-stripping relations.

The input data include lean oil and rich gas temperatures, system pressure, number of theoretical trays, specified recovery of key component, number of components, identity of key component, lean oil and rich gas compositions, equilibrium and enthalpy data, temperature difference between lean oil and residue gas, and various test tolerances. After the input data have been read, the effective absorption factor for the key component is calculated from the specified recovery and number of theoretical trays.

This calculation is a short iterative routine which assumes an

SCHEMATIC COMPUTER SEQUENCE SIMPLE ABSORBER PROGRAM



FIGURE

A_e and calculates ϕ_a. The specified ϕ_a is equal to one minus the specified fraction absorbed. The calculated ϕ_a converges to the specified value until the test tolerance is satisfied.

The residue gas temperature is then set relative to the lean oil temperature. This temperature difference is estimated based on experience.

The lean oil volume is then approximated from the effective absorption factor of the key component. Using the Kremser-Brown method (13), and assuming the residue gas temperature is the average effective temperature through the column, effective absorption factors are calculated for all components. The component distribution and total mols absorbed are then calculated. These results become the initial estimate for the absorption-stripping calculations.

For the first absorption-stripping iteration, the rich oil temperature, TB, is set equal to the residue gas temperature plus the mols absorbed. This approximation is based on the assumption that the difference in temperature between the top and bottom trays is equal to the total mols absorbed from a rich gas volume of 100 mols.

ABSORPTION-STRIPPING RELATIONS

The first step of the absorption-stripping relations is the calculation of the oil-gas ratios for the top and bottom theoretical trays. These ratios assume that straight line absorption occurs through the vessel. The relations are:

$$(L/V)_1 = \frac{L_0 + [(V_{n+1} - V_1) / n]}{V_1}$$

$$(L/V)_n = \frac{L_0 + (V_{n+1} - V_1)}{V_{n+1} - [(V_{n+1} - V_1) / n]}$$

The absorption and stripping factors for the top and bottom trays are then calculated for each component. These are then related to obtain effective absorption and stripping factors for each component. The effective factors are then related to the number of theoretical trays to obtain the absorption and stripping-factor functions. The residue gas composition and total mols absorbed are then calculated using the "phi-factor" functions, the rich gas composition, and the lean oil composition.

The recovery of key component is then checked against the quantity specified. If the recovery does not check, the lean oil rate is adjusted appropriately, and the absorption-stripping relations are repeated. When the calculated and specified recovery of key component agrees within the specified tolerance, the calculations proceed to the heat balance.

The convergence routine for adjusting the lean oil rate is quite arbitrary and works satisfactorily in most cases, but it is not foolproof. If the recovery of key is above 80 per cent, the routine converges faster if the next heavier component is selected as key.

HEAT BALANCE

For the heat balance, the total enthalpies are calculated for the rich gas, lean oil and residue gas streams. The rich oil enthalpy is then calculated by difference. The rich oil temperature, TRO, corresponding to the rich oil enthalpy is then calculated. The rich oil temperature, TRO, is obtained from a short iterative routine which assumes a temperature and calculates the corresponding rich oil enthalpy. The calculated enthalpy is compared with the heat balance enthalpy, and if it does not check, the rich oil temperature is adjusted in an appropriate direction, and the enthalpy is re-calculated. The calculated enthalpy and heat balance enthalpy converge to the proper rich oil temperature.

The calculated rich oil temperature is then compared with the assumed rich oil temperature from the preliminary calculations. If they do not agree, an adjustment is made, 0.1TB + 0.9TRO, and the new rich oil temperature is used to re-calculate the absorption and stripping relations.

When the rich oil temperature in the absorption-stripping relations agrees with the rich oil temperature from the heat balance within the specified tolerance, the final relationships are calculated.

FINAL RELATIONSHIPS

First, the residue gas dew point is calculated. Next, the rich oil bubble point is calculated. The residue dew point and rich oil bubble point are not compared with actual temperatures or used in any manner for adjusting terminal conditions. They are made available to indicate the consistency of the data, especially enthalpy and equilibrium.

The summarization of results includes (1) all input data, (2) residue gas and rich oil compositions, (3) total volumes of lean oil, residue gas, and rich oil, (4) calculated terminal temperatures, and (5) calculated residue dew point and rich oil bubble point. Stream compositions are given as mols per 100 mols of inlet gas.

GENERAL INFORMATION

The key tolerances in the program are the absorption-stripping check for recovery of key component, the heat balance check, and the rich oil temperature check. The dew point and bubble point tolerance could be important depending upon their purpose and use.

The absorption-stripping check converges to zero, and the normal tolerance is 0.001 or 0.1 per cent from specified. This deviation is well within the accuracy of the compositional, equilibrium, and enthalpy data available.

The heat balance check converges to zero, and the normal tolerance is 0.001 or 0.1 per cent. This deviation is usually equivalent to 0.5°F.

The rich oil temperature check converges to zero, and the normal tolerance is .005 or 0.5 per cent. This deviation is equivalent to 0.5°F with a temperature of 100°F.

In addition to the summary of results, intermediate information for the key component, oil-gas ratios, stream enthalpies, and other factors are punched out on each iteration. This intermediate information is helpful in checking convergence routines and effect of tolerances.

CHAPTER VI

RELIABILITY TESTS

To test the reliability of the computer program, eight NGAA Absorption Data (16) absorber tests were evaluated. Several runs were made for each set of test data with variations in the number of theoretical trays and in the recovery of key component until results were obtained which compared favorably with the actual test results.

These tests were selected based upon the system involved and the data presented. All tests cover a simple absorber without intercoolers, reboilers, or other complexities. Also, each test provided a lean oil distillation which was used to characterize the lean oil.

Calculated results check actual data for total volume absorbed, and a reasonable check is obtained on component distributions. The lean oil volume was calculated to provide the required absorption, and the primary mismatch between reported and calculated results is in the lean oil volumes.

An effort was made to approach the residue gas dew point and rich oil bubble point as close as practical; consequently the lean oil was divided into three or four fractions depending upon the distillation curve. The oil "components" were then handled as proposed by Edmister (4). The dew point and bubble point values could be improved by dividing the oil into more fractions.

Some lean oil distillations indicated lean oil saturation, but the information is insufficient to estimate the amount or the composition.

All calculations were made assuming a denuded lean oil.

PROBLEM ONE

NGAA Analysis No. 98, Volume I, page 39, was selected as Problem One. The absorber processed a stream containing 2.61 gallons per thousand cubic feet (GPM) of propane and heavier products at a pressure of 49 psig. The lean oil molecular weight was 195 with an initial boiling point (IBP) of 422°F which indicates zero saturation. The calculated results are compared with the experimental test data in Table I.

The calculated component distribution checks the actual test data throughout. The total amount absorbed checks within 0.5%.

The calculated lean oil volume is 0.564 mols or 0.6% less than the measured volume. The calculated rich oil temperature is 3.5°F less than the reported value.

Four theoretical trays are indicated by the calculations which represent a 22.2% tray efficiency for the 18 actual trays.

The following tabulation lists the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahreheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	72	75	3
Calculated Results	72	71.5	R 0.5
Dew Point and Bubble Point	72.5	73.7	1.2
Differences	0.5	2.2	

The enthalpy data is more accurate at this low pressure than at higher pressures encountered in the remaining problems. Consequently, the best overall check is obtained with Problem One.

TABLE I

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem One*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Carbon Dioxide	3.20	3.185	2.990
Methane	85.00	83.100	82.802
Ethane	3.74	2.875	3.184
Propane	3.65	1.692	1.833
Iso-Butane	0.76	.067	.075
N-Butane	1.66	.067	.064
Iso-Pentane	0.65	.000	.001
N-Pentane Plus	1.34	.000	.002
Totals	100.00	90.986	90.951

	Test Data	Calculated
Lean Oil Volume, Mols	89.898	89.334
Rich Oil Volume, Mols	98.909	98.382
Absorbed Mols	9.011	9.048
Theoretical Trays	–	4
Rich Oil Temperature, °F.	75	71.5

Operating Conditions

		Temperatures, °F.	
Pressure, psig	49		
Lean Oil Mol. Weight	195	Rich Gas	65
Actual Trays	18	Lean Oil	66
Oil-Gas Ratio	65.8	Residue Gas	72

* NGAA Analysis No. 98

PROBLEM TWO

NGAA Analysis No. 257, Volume I, page 149, was selected as Problem Two. The absorber processed a stream containing 4.16 GPM of propane and heavier at a pressure of 178 psig. The lean oil molecular weight was 178 with an initial boiling point of 362°F which indicates zero saturation. The calculated results are compared with the experimental test data in Table II.

The calculated component distribution and total mols absorbed provide an excellent check with the actual test data. The normal butane and pentanes plus content reported in the test data is apparently caused by some lean oil saturation although the lean oil IBP indicates none. The pentanes plus portion of the calculated residue gas is stripped lean oil.

The calculated lean oil volume is 4.638 moles or 12.9% lower than the measured volume which tends to confirm some lean oil saturation. The calculated rich oil temperature is 2.7°F less than the reported value. If the calculated lean oil volume were as high as the test, the calculated rich oil temperature would have been even lower.

Nine theoretical trays are indicated in the system from the calculations which represent a 45% tray efficiency for the 20 actual trays. This efficiency is probably a maximum, and the quantity and composition of any lean oil saturation would be required before a good evaluation could be made.

The following tabulation lists the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	64	90	26
Calculated Results	64	87.3	23.3
Dew Point and Bubble Point	68.2	90.0	21.8
Differences	4.2	2.7	

TABLE II

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Two*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Carbon Dioxide	4.90	4.577	4.535
Methane	75.69	73.409	73.484
Ethane	6.09	5.018	5.072
Propane	7.29	3.299	3.317
Iso-Butane	1.27	.030	.032
N-Butane	2.78	.013	.007
Iso-Pentane	0.60	-	.000
N-Pentane Plus	1.38	.017	.004
Totals	100.00	86.363	86.451
Lean Oil Volume, Mols		35.933	31.295
Rich Oil Volume, Mols		49.569	44.844
Absorbed Mols		13.636	13.549
Theoretical Trays		-	9
Rich Oil Temperature, °F.		90	87.3

Operating Conditions

		Temperatures, °F.	
Pressure, psig	178		
Lean Oil Mol. Weight	178	Rich Gas	80
Actual Trays	20	Lean Oil	57
Oil-Gas Ratio	24.2	Residue Gas	64

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* NGAA Analysis No. 257

PROBLEM THREE

NGAA Analysis No. 356, Volume I, page 152, was selected as Problem Three. The absorber processed a relatively rich gas containing 5.64 GPM of propane and heavier at 203 psig. The rich gas also contained 21.43 mol per cent air and nitrogen. The lean oil molecular weight was 244 with an initial boiling point of 238°F which indicates some lean oil saturation. The 1% distilled temperature was reported as 466°F which could mean that the saturation consists of a small amount of hexanes and heavier. The calculated results are compared with the experimental test data in Table III.

The calculated component distribution and the total mols absorbed provide an excellent check with the test data. The reported normal butane and heavier fraction shows the effect of some lean oil saturation, but it is not significant in the calculation.

The calculated lean oil volume is 3.123 mols or 7.4% higher than reported which is probably caused by measurement errors. The calculated rich oil temperature is 1.7°F higher than the reported value which is consistent with the higher than measured oil volume.

Six theoretical trays are indicated in the system from the calculations which represent a 25% tray efficiency for the 24 actual trays. This efficiency is probably a good representation for this system.

The following tabulation lists the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	90	108	18
Calculated Results	90	109.7	19.7
Dew Point and Bubble Point	93.3	113.1	19.8
Differences	3.3	3.4	

TABLE III

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Three*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Air & Nitrogen	21.43	21.774	21.171
Carbon Dioxide	.25	.200	.211
Methane	48.07	46.208	45.879
Ethane	12.04	8.076	9.197
Propane	10.18	3.249	3.233
Iso-Butane	1.07	.024	.027
N-Butane	4.59	.111	.026
Iso-Pentane	.68	-	.000
N-Pentane Plus	1.69	-	.000
Totals	100.00	79.642	79.744

	Test Data	Calculated
Lean Oil Volume, Mols	42.430	45.553
Rich Oil Volume, Mols	62.789	65.808
Absorbed Mols	20.359	20.255
Theoretical Trays	-	6
Rich Oil Temperature, °F.	108	109.7

Operating Conditions

		Temperatures, °F.	
Pressure, psig	203		
Lean Oil Mol. Weight	244	Rich Gas	87
Actual Trays	24	Lean Oil	84
Oil-Gas Ratio	39.1	Residue Gas	90

* NGAA Analysis No. 356

PROBLEM FOUR

NGAA Analysis No. 309, Volume I, page 193, was selected as Problem Four. The absorber processed a gas containing 2.94 GPM of propane and heavier at a pressure of 352 psig. The lean oil molecular weight was 161 with an initial boiling point of 280°F which indicates some saturation. Additional information was provided which indicates a lean oil saturation of nearly 10%. The calculated results are compared with the experimental test data in Table IV.

The calculated component distribution checks the actual test results very well down through iso-butane, but for the heavier components, the lean oil saturation caused lower absorption than expected. The calculated iso-pentanes plus fraction is stripped lean oil.

The calculated lean oil volume is 5.886 mols or 21.1% lower than the test volume for the same key component recovery which shows the effect of saturation. The calculated rich oil temperature is 3.5°F less than the reported value.

Eight theoretical trays are indicated for the system according to the calculations which represent a 40% tray efficiency for the 20 actual trays. This efficiency is probably a maximum value, and the quantity and composition of the lean oil saturation would be required before a more accurate figure could be developed.

The following tabulation shows the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	86	115	29
Calculated Results	86	111.5	25.5
Dew Point and Bubble Point	93.4	114.8	21.4
Differences	7.4	3.3	

The residue gas temperature was reported to be only 1°F higher than the lean oil temperature which is unusual to the point of being questionable. It could, however, reflect one of the many effects of high lean oil saturation.

TABLE IV

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Four*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Carbon Dioxide	.400	.267	.370
Methane	83.681	80.487	80.073
Ethane	6.672	5.522	5.477
Propane	4.883	2.487	2.448
Iso-Butane	.797	.060	.078
N-Butane	1.778	.087	.043
Iso-Pentane Plus	1.789	.052	.002
Totals	100.000	88.962	88.491
Lean Oil Volume, Mols		27.893	22.007
Rich Oil Volume, Mols		38.929	33.516
Absorbed Mols		11.036	11.509
Theoretical Trays		-	8
Rich Oil Temperature, °F.		115	111.5

Operating Conditions

Pressure, psig	352	Temperatures, °F.	
Lean Oil Molecular Weight	161	Rich Gas	96
Actual Trays	20	Lean Oil	85
Oil-Gas Ratio	17.3	Residue Gas	86

* NGAA Analysis 309

PROBLEM FIVE

NGAA Analysis No. 364, Volume II, page 23, was selected as Problem Five. The absorber processed an average stream containing 2.79 GPM of propane and heavier at a pressure of 560 psig. The lean oil molecular weight was 208 with an initial boiling point of 410°F which indicates zero saturation. Calculated results are compared with the experimental test data in Table V.

The calculated component distribution and total volume absorbed do not check too well with the actual test information. These differences are difficult to evaluate since the test data do not report a breakdown of the propane plus fraction in the residue gas.

The calculated lean oil volume is 12.551 mols or 32.4% lower than the test volume. The calculated rich oil temperature is 4.1°F less than the reported value. With the reported lean oil rate, the rich oil temperature would be considerably lower than either value which indicates a measuring discrepancy. The reported relative volumes of lean oil and rich gas probably did not exist.

Five theoretical trays are indicated for the system according to the calculations which represent a 20% tray efficiency for the 25 actual trays. This efficiency may be reasonably accurate, but with all of the discrepancies involved it should not be used as a primary case in any correlation.

The following tabulation shows the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	97	108	11
Calculated Results	97	103.9	6.9
Dew Point and Bubble Point	104.5	107.3	2.8
Differences	7.5	3.4	

TABLE V

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Five*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Nitrogen	1.24	1.245	1.213
Methane	80.78	73.979	72.179
Ethane	8.93	5.748	5.714
Propane	5.56	.909	1.100
Iso-Butane	.61	-	.024
N-Butane	1.63	-	.023
Iso-Pentane	.38	-	.000
N-Pentane Plus	.87	-	.002
Totals	100.00	81.881	80.256

	Test Data	Calculated
Lean Oil Volume, Mols	38.699	26.148
Rich Oil Volume, Mols	56.818	45.892
Absorbed Mols	18.119	19.744
Theoretical Trays	-	5
Rich Oil Temperature, °F	108	103.9

Operating Conditions

		Temperatures, °F	
Pressure, psig	560		
Lean Oil Mol. Weight	208	Rich Gas	83
Actual Trays	25	Lean Oil	93
Oil-Gas Ratio	29.9	Residue Gas	97

* NGAA Analysis No. 364

PROBLEM SIX

NGAA Analysis No. 91, Volume II, page 37, was selected as Problem Six. The absorber processed a relatively lean stream containing 2.34 GPM of propane and heavier at a pressure of 553 psig. The lean oil molecular weight was 178 with an initial boiling point of 202°F which indicates some saturation. Calculated results are compared with the experimental test data in Table VI.

The calculated component distribution and total volume absorbed check with the test data reasonably well. The reported iso-pentanes plus fraction was caused by lean oil saturation.

The calculated lean oil volume is 9.119 mols or 38.2% lower than the reported volume due at least in part to lean oil saturation. Some metering discrepancy could be involved, for it is difficult to conceive an operating system that would require over a third more oil because of saturation. The calculated rich oil temperature is 7.8°F less than the reported value.

Seven theoretical trays are indicated for the system from the calculations which represent a 35% tray efficiency for the 20 actual trays. This efficiency is probably a maximum value, and the quantity and composition of the lean oil saturation would be required before a more accurate evaluation could be made.

The following tabulation lists the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	87	97	10
Calculated Results	87	89.2	2.2
Dew Point and Bubble Point	97.3	95.5	1.8
Differences	10.3	6.3	

TABLE VI

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Six*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Carbon Dioxide	.60	.527	.506
Methane	82.81	78.328	77.473
Ethane	8.92	6.931	6.848
Propane	4.92	1.985	1.966
Iso-Butane	.68	.026	.046
N-Butane	1.20	.026	.023
Iso-Pentane	.26	.018	.000
N-Pentane Plus	.61	-	.001
Totals	100.00	87.841	86.863

	Test Data	Calculated
Lean Oil Volume, Mols	23.919	14.800
Rich Oil Volume, Mols	36.078	27.937
Absorbed Mols	12.159	13.137
Theoretical Trays	-	7
Rich Oil Temperature, °F	97	89.2

Operating Conditions

		Temperature, °F	
Pressure, psig	553		
Lean Oil Mol. Weight	178	Rich Gas	75
Actual Trays	20	Lean Oil	77
Oil-Gas Ratio	15.8	Residue Gas	87

* NGAA Analysis No. 91

PROBLEM SEVEN

NGAA Analysis No. 324, Volume II, page 70, was selected for Problem Seven. The absorber processed an average rich gas stream containing 2.79 GPM of propane and heavier at 385 psig. The lean oil molecular weight was 207 with an initial boiling point of 462°F which indicates zero saturation. Calculated results are compared with the experimental test data in Table VII.

The calculated component distribution and total volume absorbed check the test data. The residue gas analysis does not appear to be in the proper proportions when related to the rich gas composition, for excessive methane and ethane absorption is indicated from the calculations while insufficient propane and butanes are recovered. These differences are probably within the analytical accuracies obtained at that time.

The calculated lean oil volume is 3.522 mols or 18.0% lower than the actual test volume which indicates metering discrepancies. The calculated rich oil temperature is 2.2°F lower than the reported value.

Eight theoretical trays are indicated for the system from the calculations which represent a 33.3% tray efficiency for the 24 actual trays. According to the iso and normal butane distribution, the number of theoretical trays could be larger indicating a higher tray efficiency, but there are too many inconsistencies to rely upon this facet of the results.

The following tabulation lists the residue gas dew point and the rich oil bubble point with the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	78	92	14
Calculated Results	78	89.8	11.8
Dew Point and Bubble Point	87.6	94.3	6.7
Differences	9.6	4.5	

TABLE VII

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Seven*

Components	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Air	1.20	1.163	1.189
Carbon Dioxide	9.90	9.127	8.890
Methane	74.39	71.902	71.250
Ethane	5.57	4.811	4.572
Propane	4.99	2.420	2.455
Iso-Butane	.91	.040	.084
N-Butane	1.79	.008	.038
Iso-Pentane Plus	1.25	.009	.002
Totals	100.00	89.480	88.480

	Test Data	Calculated
Lean Oil Volume, Mols	19.515	15.993
Rich Oil Volume, Mols	30.034	27.515
Absorbed Mols	10.519	11.522
Theoretical Trays	-	8
Rich Oil Temperature, °F.	92	89.8

Operating Conditions

		Temperatures, °F.	
Pressure, psig	385		
Lean Oil Mol. Weight	207	Rich Gas	69
Actual Trays	24	Lean Oil	71
Oil-Gas Ratio	15.2	Residue Gas	78

* NGAA Analysis No. 324

PROBLEM EIGHT

NGAA Analysis No. 281, Volume II, page 62, was selected as Problem Eight. The absorber processed an average stream containing 2.90 GPM of propane and heavier at 390 psig. The lean oil molecular weight was 200 with an initial boiling point of 159°F which indicates some light end saturation. Calculated results are compared with the experimental test data in Table VIII.

The calculated component distribution and total volume absorbed check the actual test data very well through normal butane. The reported iso-pentanes plus content in the residue gas was caused by lean oil saturation.

The calculated lean oil rate is 3.835 mols or 19.2% less than the measured rate. The calculated rich oil temperature is 3.3°F less than the reported value. This inter-relation of results indicates the lean oil saturation is predominantly butanes and heavier. Propane saturation would have caused the calculated lean oil volume to be even lower than obtained.

Eight theoretical trays are indicated for the system which represent a 40% tray efficiency for the 20 actual trays. This is probably a maximum figure, and the lean oil saturation content and composition would have to be known to make a better evaluation.

The following tabulation shows the relation of the residue gas dew point and the rich oil bubble point to the reported and calculated terminal temperatures.

(Degrees Fahrenheit)	Residue Gas Temperature	Rich Oil Temperature	Differences
Actual Test Data	74	94	20
Calculation Results	74	90.7	16.7
Dew Point and Bubble Point	83.6	93.7	10.1
Differences	9.6	3.0	

TABLE VIII

COMPARISON OF CALCULATED RESULTS WITH EXPERIMENTAL TEST DATA

Problem Eight*

Component	Rich Gas Mols	Residue Gas Mols Test Data	Residue Gas Mols Calculated
Air	.50	.088	.495
Carbon Dioxide	1.00	.882	.896
Methane	79.81	76.886	76.359
Ethane	9.23	7.464	7.529
Propane	5.90	2.805	2.815
Iso-Butane	.88	.044	.070
N-Butane	1.70	.035	.030
Iso-Pentane Plus	.98	.018	.000
Totals	100.00	88.222	88.194
Lean Oil Volume, Mols		19.986	16.151
Rich Oil Volume, Mols		31.763	27.956
Absorbed Mols		11.777	11.805
Theoretical Trays		-	8
Rich Oil Temperature, °F.		94	90.7

Operating Conditions

Pressure, psig	390	Temperatures, °F.	
Lean Oil Mol. Weight	200	Rich Gas	78
Actual Trays	20	Lean Oil	64
Oil-Gas Ratio	15.02	Residue Gas	74

* NGAA Analysis No. 281

TERMINAL TEMPERATURE RELATIONS

The calculated terminal temperatures and the residue gas dew point and rich oil bubble point were studied in an effort to obtain a unique solution for a given absorber problem. It was concluded that this short-cut calculation procedure did not describe the systems with sufficient accuracy to approach a unique solution. It did develop, however, that the relations give a good indication of the consistency of the equilibrium and enthalpy data being employed and the temperatures assumed.

The following tabulation summarizes the differences between the calculated terminal temperatures and the appropriate equilibrium temperature for each of the absorber tests evaluated.

(Degrees Fahrenheit)	Residue Gas & Dew Point Difference	Rich Oil & Bubble Point Difference	Total Difference
Problem One	0.5	2.2	2.7
Problem Two	4.2	2.7	6.9
Problem Three	3.3	3.4	6.7
Problem Four	7.4	3.3	10.7
Problem Five	7.5	3.4	10.9
Problem Six	10.3	6.3	16.6
Problem Seven	9.6	4.5	14.1
Problem Eight	9.6	3.0	12.6

These total differences are plotted against absolute operating pressure in Figure 4. This relation indicates questionable enthalpy values at elevated pressures. Actual heats of vaporization are greater than reflected in the data. Effects of increased heat of absorption would include (1) higher rich oil temperature, (2) higher lean oil rate for a given absorption,

(3) higher rich oil bubble point, and (4) lower residue gas dew point. These changes would make all of the calculated and reported test results more agreeable.

Since the analysis of these terminal temperature relations is not a part of the evaluation of the "phi-factor" function method, they are presented for information only in relation to this research. The utilization of similar information for checking reliability of enthalpy data may hold some promise.

EFFECT OF PRESSURE
ON
TERMINAL TEMPERATURE MISMATCH

SUM OF TEMPERATURE DIFFERENCE EQUALS RESIDUE GAS TEMPERATURE MINUS RESIDUE GAS DEW POINT PLUS RICH OIL TEMPERATURE MINUS RICH OIL BUBBLE POINT.

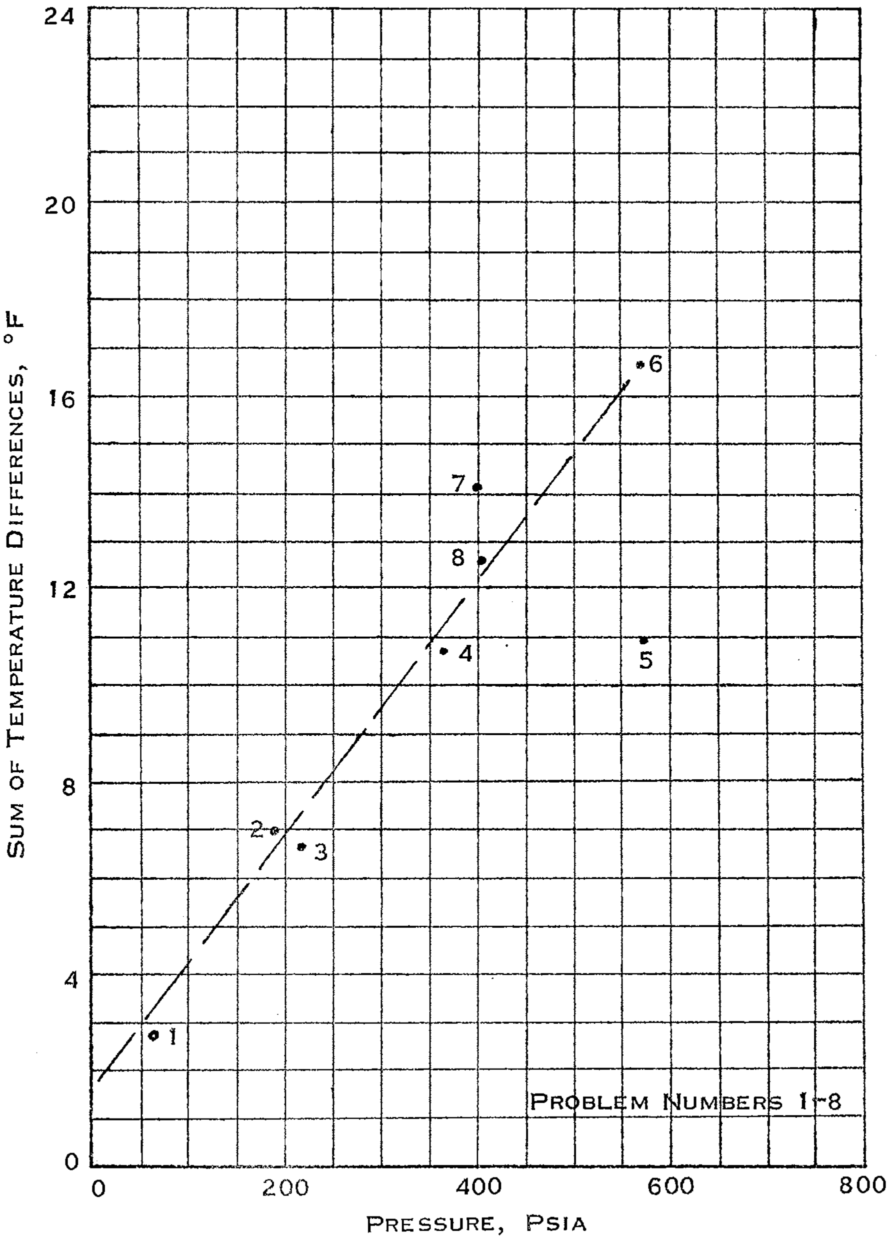

FIGURE 4

CHAPTER VII

ACCURACY TESTS

The accuracy of the "phi-factor" function calculation procedure has been determined by comparing calculated results with corresponding results from a different computer program using a different calculation procedure of known accuracy. The comparative results from two problems are presented.

The test computer program was developed by Cooper (1) in 1960, and it uses both the Kremser-Brown (13) and the earlier Edmister methods (2). Cooper reported that the Edmister procedure gave more accurate results; consequently the results from the Edmister portion of the program are utilized for comparison. This calculation technique has been used extensively, and accurate results have been obtained consistent with the "short-cut" approach involved.

PROBLEM ONE

Calculated results for Problem One are compared with the test program results in Table IX. The specified tolerances in each program were essentially the same, and the data required for calculation were identical. The results from the two calculation procedures are almost identical.

The total residue gas volume from the test program is 0.0025% higher than that from the new program. The lean oil volume from the test program is 0.25% lower than that from the new program. The calculated rich oil temperature from the test program is 0.9881°F less than that from the new program. In each case the rich oil temperature check tolerance was 0.1°F.

TABLE IX

COMPARISON OF CALCULATED RESULTS WITH PROCEDURE OF KNOWN ACCURACY

Problem One

Component	Residue Gas Mols New System	 Accurate System
Carbon Dioxide	2.99021	2.99022
Methane	82.80227	82.80375
Ethane	3.18449	3.18463
Propane	1.83305	1.83356
Iso-Butane	.07473	.07481
N-Butane	.06419	.06428
Iso-Pentane	.00110	.00111
N-Pentanes Plus	.00231	.00230
Total	90.95235	90.95466
Lean Oil Volume, Mols	89.33417	89.11202
Rich Oil Volume, Mols	98.38182	98.15737
Absorbed Mols	9.04765	9.04535
Theoretical Trays	4	4
Rich Oil Temperature, °F	71.64522	71.54641

PROBLEM FOUR

Calculated results for Problem Four are compared with the test program results in Table X. The specified tolerances in each program were more different than in Problem One, but the data required for calculation were identical. The results from the two calculation procedures are very similar, but they are not as nearly identical as in Problem One.

The total residue gas volume from the test program is 0.008% lower than that from the new program. The lean oil volume from the test program is 0.27% higher than that from the new program. The calculated rich oil temperature from the test program is 0.58700°F higher than that for the new program. The temperature test tolerance was 0.1°F for the new program and 0.5°F for the test program.

From the comparison of results it can be concluded that the earlier Edmister calculation procedure (2) and the "phi-factor" function method (5) are one and the same. This is not surprising when the derivation of equations is considered.

TABLE X

COMPARISON OF CALCULATED RESULTS WITH PROCEDURE OF KNOWN ACCURACY

Problem Four

	Residue Gas Mols	
Component	New System	Accurate System
Carbon Dioxide	.37005	.37012
Methane	80.07355	80.06515
Ethane	5.47662	5.47621
Propane	2.44835	2.44963
Iso-Butane	.07760	.07782
N-Butane	.04268	.04287
Iso-Pentane Plus	.00241	.00241
Total	88.49126	88.48422
Lean Oil Volume, Mols	22.00721	22.06709
Rich Oil Volume, Mols	33.51596	33.58287
Absorbed Mols	11.50875	11.51578
Theoretical Trays	8	8
Rich Oil Temperature, °F	111.46912	112.05612

CHAPTER VIII

PROPOSED APPLICATIONS

The short-cut calculation procedure using absorption and stripping-factor functions has many potential applications both in type of problem and in type of processing unit involved. Three areas of possible application are presented--design procedures, economic evaluation, and operational evaluation. Special emphasis is given to the problems and possible solutions for evaluation of existing absorption systems.

DESIGN PROCEDURES

The "phi-factor" function procedure can be used for many design problems, especially those of a preliminary nature. It is adaptable to many different processing units including absorption systems, rich oil fractionators, and stills.

When a new project is undertaken, the basis for design is rarely fixed. The gas volume, pressure and composition may not be known accurately. Also, the products and their value may be subject to further negotiation prior to firming the design basis. At the same time, the operating temperature level, pressure level, and lean oil characteristics are variables that require study. Under such conditions, the short-cut method is particularly appropriate, and its use will allow the study of a wider range of processing possibilities with lower computer cost.

For the complete design of a given system, a rigorous method may be required for final sizing and tray loading information, but the short-cut procedure can be employed to more thoroughly define the variables prior to making the more time consuming rigorous calculations. In fact in many cases, the short-cut results are used as initial estimates for the rigorous calculation routine.

In many cases, the short-cut procedure will give results of suitable accuracy with less computer cost per problem regardless of machine size and speed. The procedure is best for absorber problems where low tray efficiencies are normal, and it is especially suited to those problems where lean oil saturation is encountered.

ECONOMIC STUDIES

In the area of plant economics there are always a number of unanswered questions concerning optimum propane recovery, number of trays, temperature level, operating pressure and other variables. To make a true economic evaluation, the value of products and fuel, cost of operating materials and labor, and many other economic factors must be known. With a short, simple computer program, answers to optimizing problems would be approached more often.

Economic evaluations and operational evaluations can go hand in hand with some results being applicable to both. A complete mesh would take the form of developing an operating curve for the system during the design and construction period. In this way the operating curve is available for startup, acceptance tests, and operating uses. Unless radical changes are made, the curve should adequately describe the system throughout its operating life.

OPERATIONAL EVALUATION

The many difficulties in evaluating operating systems have hampered the progress of improvement in design techniques. Today, the ability to analyze operating systems in the petroleum industry is limited at best. Some of the difficulties include unsteady state conditions, accurate sampling and analytical procedures, accurate data measuring equipment, and lack of testing equipment on industrial absorption systems.

Today, these evaluations are becoming more and more difficult due to the changing economic picture. There is a trend toward minimum cost plants where permanent test apparatus is one of the first items cut. There is also a trend toward minimum personnel which removes from the plant scene those individuals who once made operational evaluations. Both trends are good and improve operational earnings, but plant tests become more and more difficult to arrange, and opportunities for good test data are few and far between.

The A.I.Ch.E. Standard Testing Procedure for Absorbers (17) is the only published procedure for the evaluation of absorption equipment, and it emphasizes the complexity of the testing process. The procedure stipulates that compositions, flow rates, temperatures, and pressures for all fluids entering and leaving the absorber are necessary for evaluation. It would be extremely difficult, if not impossible, to find an absorber in the natural gas processing industry that is equipped for such a test. The A.I.Ch.E. Standard Procedure is primarily for guidance where chemical absorption is involved. A multi-component supplement for use by the petroleum industry has been discussed.

The NGPA (Natural Gas Processors Association, formerly Natural Gasoline Association of America) is actively working on absorber evaluation in their Absorption Subcommittee of the Technical Committee. The Subcommittee has found it extremely difficult to draw conclusions from unreliable test data. On the side of member companies, it is obvious that most tests cannot be reliable since, in most cases, the absorbers were not equipped for testing. This problem has reduced the chance for success to the point that enthusiasm for the program, as originally defined, has dwindled.

It appears that a different approach is required to analyze the operation of existing absorption systems. This research, it is hoped, will help to initiate a new evaluation procedure. By working with typical rich gas and lean oil compositions, typical operating conditions and a range of theoretical trays, performance curves can be developed for individual absorption systems. Only minimum test data would then be required to establish the theoretical tray parameter.

Absorber test procedures have been improved over the years at about the same rate that the calculation procedure has been improved. Originally, charcoal adsorption tests were run on the inlet and outlet streams, and they were related directly for an efficiency check.

Until mid-1940, most tests utilized a spot check of operating temperatures and pressures, a spot sample and analysis of residue gas, daily average gas volumes, and daily production as measured in the storage tanks. Meters and other test apparatus were rarely calibrated for the test. This test procedure introduced many errors into every test.

Since the late 1940's, elaborate plant balances have made better absorber test data available. The test periods were usually 8-24 hours duration, and widely different conditions were averaged in the test data for evaluation purposes. Test equipment was calibrated, and sound sampling and analytical procedures were followed. This is the type of data that was made available to the NGPA for their evaluation work. However, only a small portion of the total data presented was complete and consistent.

During the past two years, a series of research quality absorber tests was planned for five non-reboiled absorbers at three plants. The planning progressed to analyzing the absorption systems, determining the required modifications for good tests, preparing data sheets for the individual systems, and tentative scheduling of the tests. All preparations progressed to the point of spending money to modify the systems for testing. At this point, the information to be obtained in light of the chances for success did not justify the estimated expenses, and the testing program was temporarily abandoned.

The specific problems are discussed below for the plants involved listed in the order of diminishing chance for success.

PLANT A

This plant has one simple absorber processing a relatively rich gas with refrigerated lean oil. The propane recovery is normally 95-98% and the lean oil always has some saturation. The column has 25 cross-flow bubble cap trays and five cross-flow perforated trays. All of the necessary flow measurements are available except rich oil. All pressure and temperature points are available. All sample points are satisfactory except rich oil.

One problem concerns the absorber bypass which may leak, and it is located on the absorber side of the metering equipment and part of the sample points. The only sure way to eliminate leakage would be to shut the plant down twice to install and to remove a blind in the bypass.

Another problem concerns an inlet scrubber in the base of the column which is not in use. It is always full of liquid which is presumably in equilibrium with the inlet gas, but the manner in which this liquid is lifted into the absorber is impossible to predict.

Still another problem, which is also true for most plants with severe proration, concerns tray loading. This column has been operating at 40-60% of design capacity, and results under design loading would have to be extrapolated, if obtained.

PLANT B

This plant has two large diameter simple absorbers which are mechanically identical and operate in parallel. They process a relatively lean gas with a light (170 molecular weight) absorption oil. Each column contains 20 split-flow perforated trays. The residue gas sampling point is not readily available for the individual absorbers, but it could be taken through the meter piping. The rich oil sampling point is questionable.

One problem involves a recycle stream that is introduced into the inlet of one of the absorbers which makes it mandatory that the columns be tested individually. The recycle stream consists of two individual streams that can be metered, sampled, and analyzed, but these data would be obtained upstream of compressors and dehydration facilities.

The propane recovery level is quite low (30%); consequently iso and normal butane are the key components. The lean oil is dehydrated by stripping with fuel gas, and it always contains significant quantities of methane and ethane. The quantities are small compared with the system, but the effect should be recognized.

This plant always presents an analytical problem due to the lean gas. Normal analytical accuracy may not be sufficient for good results. Modern chromatographic techniques have improved the chances, but great care would be required.

The column loading is in a good testing range most of the time. This absorption system offers considerable promise for testing, for both liquid and vapor loading can be varied individually or totally for short test periods.

PLANT C

Two parallel absorbers process a lean gas at Plant C. The columns contain 24 crossflow perforated trays that have been modified. The trays were some of the first large diameter perforated trays installed in absorber service. The towers operate at 0-20^{O}F and have two refrigerated intercoolers.

The chances for success are quite different here than at Plant B, for problems of poor sampling points, no rich oil temperatures, more complex recycle system, no positive distribution of recycle gas between vessels, and questionable absorber bypasses are encountered.

The residue gas and lean oil flow rates are the only measurements. The rich gas temperature is the same for both absorbers. The rich oil temperature is not available. The rich oil flow to the intercoolers is an unknown, and the intercooler refrigeration load would be difficult, if not impossible, to determine with accuracy.

There is a problem of liquid carryover from the inlet separators. This may not cause too much difficulty in itself, but the piping is arranged in such a way that all liquid flows into one vessel.

This absorption system is desirable for testing since the effects of temperature on tray efficiency could be evaluated along with vapor and liquid loading. The system design features, mechanical tray details after field modifications, and inherent analytical problems with lean gas make the chances for success quite remote.

ABSORBER TEST PROCEDURE

Prior to any testing, all meters, pressure gauges, and thermometers should be calibrated. Sample containers should be connected and filled with the displacement fluid. Data sheets should be prepared listing all pertinent information. The test period should be selected during the most stable portion of the operating day.

The test should be of three hours duration with data being recorded every 15 minutes throughout the test. Samples should be taken in four increments during the second hour. Samples should be taken by mercury or water displacement through well purged sample lines from carefully selected sample points. Duplicate samples are desirable.

All streams should be analyzed for inerts and all hydrocarbons through n-heptane with an octane plus fraction. A true boiling point distillation should be made on the lean oil with a fractional analysis of the saturation components. Flow rates, temperatures, and pressures should not vary significantly during the test period.

Several tests may be required for a given absorption system to make the information statistically usable. Since equipment tests are expensive, absorption systems should be selected carefully to insure good results.

The Absorbers Subcommittee of the A.I.Ch.E. Equipment Test Procedures Committee would like to formalize and publish a supplement for the testing of multi-component systems, but the present A.I.Ch.E.

approach would be used very rarely. Companies will be reluctant to spend extra money to meet all of the necessary stipulations.

The possibility of absorber evaluation by computer has been explored on a preliminary basis by developing performance curves for individual absorption systems. This approach to absorber evaluation appears feasible since the gas processed at each plant has a relatively constant amount of nitrogen, methane, and ethane, and these components comprise the bulk of the gas stream. The propane and heavier content varies from season to season and from day to night, but in most cases these components are absorbed, and the residue gas composition is essentially constant.

Absorption oil characteristics are the same month after month with only minor variations in saturation from day to day. Once the procedure is established, occasional simple tests should be sufficient to keep it current.

The proposed evaluation procedure would consist of estimating performance curves based on past plant tests and analyses. These curves would result from a series of computer runs with variations in temperatures, pressures, number of theoretical trays, etc. The accuracy of the curves would be confirmed by one or two simple tests which would also define the theoretical tray parameter.

To investigate this approach to operational evaluation, the computer was used to develop a performance curve for one particular absorption system. Sixteen computer runs were made with variations in lean oil temperature, rich gas temperature, propane recovery and theoretical trays. Additional variables might include pressure, lean oil saturation, and product content of rich gas.

The resulting absorber operating curve is shown in Figure 5. It is simple, easy to use, and probably accurate within 0.5% on the per cent of key absorbed scale.

This particular operating curve shows a sample problem which assumes (1) a molal oil-gas ratio by measurement of 15, (2) a lean oil temperature of 80°F, and (3) a rich gas temperature of 76°F. The lean oil temperature correction with an L/V of 15 is noted as .09 Δ L/V per °F. The difference between the base temperature and the lean oil temperature is - 10°F. The oil-gas ratio correction due to lean oil temperature becomes:

$$\Delta L/V = .09 \times -10 = -0.9$$

The difference between the base temperature and the rich gas temperature is - 6°F. The oil-gas ratio correction due to rich gas temperature is read directly as - 0.58. The corrected oil-gas ratio becomes:

Measured	15.0
Lean oil temperature correction	-0.9
Rich gas temperature correction	-0.58
Corrected oil-gas ratio	13.52

Entering the corrected oil-gas ratio on the main curve and assuming seven theoretical trays, a propane absorption of 59.3% is indicated.

The same relations are adaptable to solution by nomagraph,

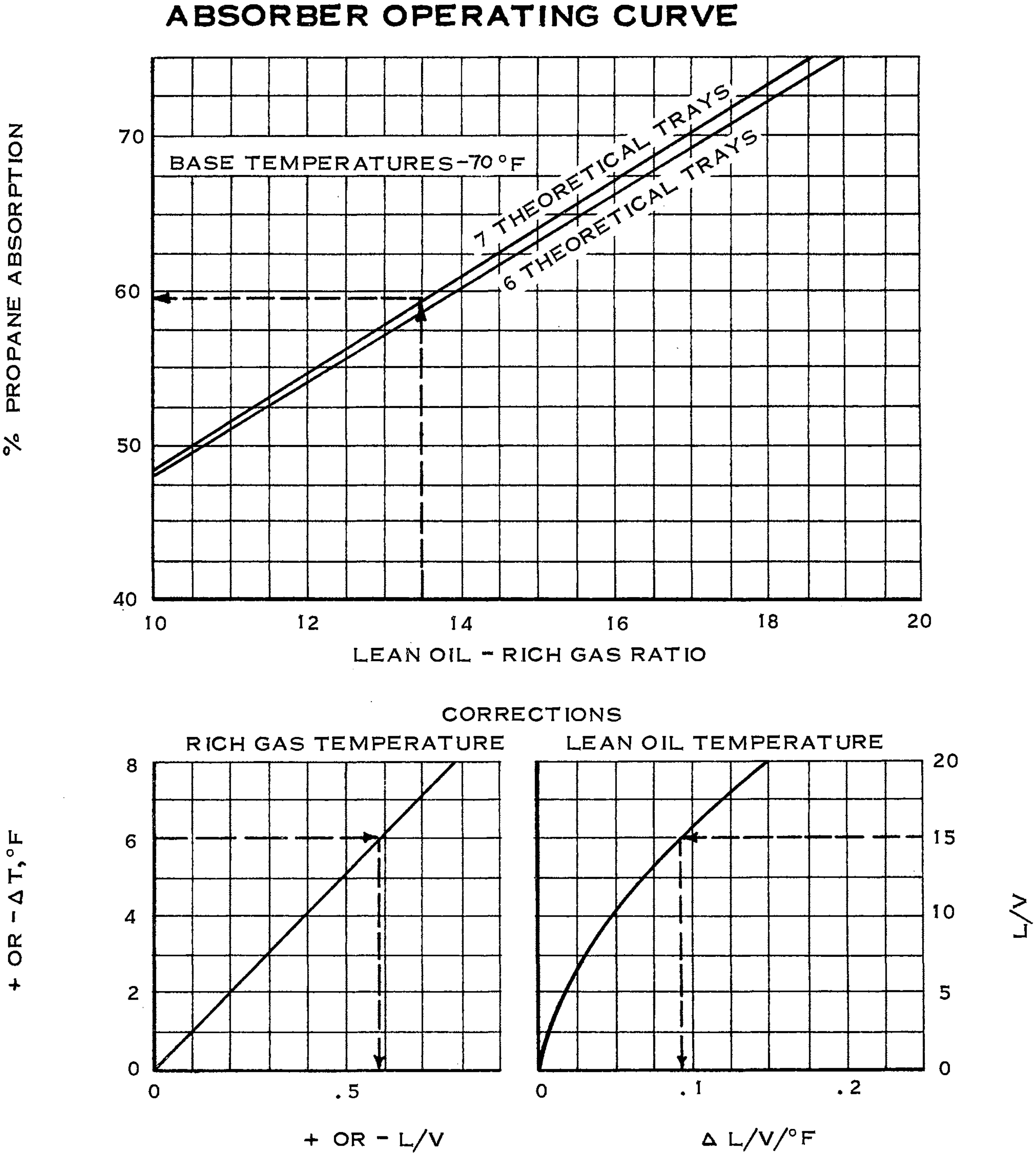
ABSORBER OPERATING CURVE
% PROPANE ABSORPTION
70
60
50
40
BASE TEMPERATURES-70 °F
7 THEORETICAL TRAYS
6 THEORETICAL TRAYS
10
12
14
16
18
20
LEAN OIL - RICH GAS RATIO
CORRECTIONS
RICH GAS TEMPERATURE
LEAN OIL TEMPERATURE
+ OR - ΔT, °F
8
6
4
2
0
0
.5
+ OR - L/V
0
.1
.2
Δ L/V/°F
20
15
10
5
0
L/V

FIGURE 5

and additional variables can be added. Depending upon the system involved, variables of rich gas product content, system pressure, and lean oil saturation may be included in the correction terms.

To justify this approach to operational evaluation, a short-cut calculation procedure is required to keep computer costs reasonable. The "phi-factor" function method is particularly adaptable for such evaluation programs. The approach may also be suitable for rich oil fractionators, stills and other process units in addition to absorbers where it would normally be applied.

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APPENDIX

COMPUTER PROGRAM DETAILS

FORTRAN LISTING
SIMPLE ABSORBER DESIGN PROGRAM
SPECIFY RECOVERY OF KEY
BASIS 100 MOLS RICH GAS

(# is =)

```
00000 0 DIMENSION TOIL(20),RG(20),
00000 1 AEE(20,4),APEL(20,3),APEV(20,
00000 2 3)
00000 0 DIMENSION EE(20,4),EFB(20,4),
00000 1 D(20),EFA(20),EFS(20),RO(20),
00000 2 TOL(20)
01999 0 READ,TOILT,RGT,PRS,TRY,REP,N,K
00003 0 READ,TTA,TTB,TTC,TTD,TTE,TTF,
00003 1 TTG
00000 0 DO 5 I#1,N
00005 0 READ,TOIL(I),RG(I)
00000 0 DO 6 I#1,N
00000 0 DO 6 J#1,4
00006 0 READ,AEE(I,J)
00000 0 DO 7 I#1,N
00000 0 DO 7 J#1,3
00007 0 READ,APEL(I,J),APEV(I,J)
00000 0 TT#TOILT+TTE
00000 0 DO 8 I#1,N
00000 0 EE(I,1)#(((((AEE(I,4)*TT)+AEE
00000 1 (I,3))*TT)+AEE(I,2))*TT)+AEE
00000 2 (I,1)
```

```
00000 0 EE(I,2)#APEL(I,1)+(APEL(I,2)*
00000 1 TOILT)+(APEL(I,3)*TOILT*TOILT)
00000 0 EE(I,3)#APEV(I,1)+(APEV(I,2)*
00000 1 RGT)+(APEV(I,3)*RGT*RGT)
00008 0 EE(I,4)#APEV(I,1)+(APEV(I,2)*
00008 1 TT)+(APEV(I,3)*TT*TT)
00000 0 HRG#0.0
00000 0 DO 9 I#1,N
00009 0 HRG#(EE(I,3)*RG(I))+HRG
00000 0 PHIP#(1.0-REP)
00000 0 EFD#REP
00011 0 PHIC#(EFD-1.0)/(EFD**(TRY+1.0)
00011 1 -1.0)
00012 0 IF(ABSF(PHIC-PHIP)-TTA)18,18,
00012 1 13
00013 0 EFD#EFD+PHIC-PHIP
00014 0 GO TO 11
00018 0 DT#0.0
00000 0 GOR#EFD*EE(K,1)
00000 0 DO 22 I#1,N
00022 0 TOL(I)#100.0*GOR*TOIL(I)
00000 0 DO 28 I#1,N
00000 0 EFA(I)#((GOR/EE(I,1))-1.0)/(((
00000 1 GOR/EE(I,1))**(TRY+1.0))-1.0)
00000 0 EFS(I)#((EE(I,1)/GOR)-1.0)/(((
00000 1 EE(I,1)/GOR)**(TRY+1.0))-1.0)
```

```
00000 0 D(I)#EFA(I)*RG(I)+(1.0-EFS(I))
00000 1 *TOL(I)
00028 0 DT#D(I)+DT
00000 0 SBA#100.0-DT
00000 0 IF(KONSF)29.30.30
00029 0 PUNCH.EFD.GOR.SBA.EFA(K).D(K).
00029 1 DT
00030 0 TB#TT+SBA
00031 0 GORT#((100.0*GOR)+(SBA/TRY))/
00031 1 DT
00000 0 GORB#((100.0*GOR)+SBA)/(100.0
00000 1 -(SBA/TRY))
00000 0 DT#0.0
00000 0 TOLT#0.0
00000 0 DO 33 I#1.N
00000 0 EFA(I)#(((((AEE(I.4)*TB)+AEE
00000 1 (I.3))*TB)+AEE(I.2))*TB)+AEE
00000 2 (I.1)
00000 0 EFB(I.1)#((((GORB/EFA(I))*((
00000 1 GORT/EE(I.1))+1.0))+0.25)**0.5
00000 2 )-0.5
00000 0 EFS(I)#((((EE(I.1)/GORT)*((EFA
00000 1 (I)/GORB)+1.0))+0.25)**0.5)-
00000 2 0.5
00000 0 TOL(I)#(100.0*GOR)*TOIL(I)
00033 0 TOLT#TOL(I)+TOLT
```

```
00000 0 IF(KONSF)34,37,37
00034 0 PUNCH,GORT,GORB,TB,TOLT,EFA(K)
00034 1 ,EFB(K,1),EFS(K)
00037 0 DO 40 I#1,N
00000 0 EFB(I,2)#(EFB(I,1)-1.0)/((EFB
00000 1 (I,1)**(TRY+1.0))-1.0)
00000 0 EFB(I,3)#(EFS(I)-1.0)/((EFS(I)
00000 1 **(TRY+1.0))-1.0)
00000 0 D(I)#EFB(I,2)*RG(I)+(1.0-EFB(
00000 1 I,3))*TOL(I)
00040 0 DT#D(I)+DT
00000 0 SBA#100.0-DT
00000 0 CONV#RG(K)*REP-RG(K)+D(K)
00000 0 IF(KONSF)41,44,44
00041 0 PUNCH,EFB(K,2),EFB(K,3),D(K),
00041 1 DT,SBA,CONV
00044 0 IF(ABSF(1.0-(RG(K)-D(K))/(RG(K
00044 1 )*REP))-TTB)57,57,52
00052 0 GOR#(CONV*((GOR*REP*REP)**TTG)
00052 1 )+GOR
00000 0 GO TO 31
00057 0 HLO#0.0
00000 0 HD#0.0
00000 0 DO 60 I#1,N
00000 0 HLO#(EE(I,2)*TOL(I))+HLO
00060 0 HD#(EE(I,4)*D(I))+HD
```

```
00000 0 HRO#HRG+HLO-HD
00000 0 IF(KONSF)62,65,65
00062 0 PUNCH,HRG,HLO,HD,HRO
00065 0 ROT#0.0
00000 0 TRO#TB
00000 0 DO 70 I#1,N
00000 0 RO(I)#TOL(I)+RG(I)-D(I)
00070 0 ROT#RO(I)+ROT
00071 0 HROC#0.0
00000 0 DO 72 I#1,N
00000 0 EFS(I)#APEL(I,1)+(APEL(I,2)*
00000 1 TRO)+(APEL(I,3)*TRO*TRO)
00072 0 HROC#EFS(I)*RO(I)+HROC
00000 0 IF(KONSF)73,76,76
00073 0 PUNCH,TRO,TB,EFS(K),RO(K),ROT,
00073 1 HROC
00076 0 IF(ABSF((HROC-HRO)/HROC)-TTC)
00076 1 79,79,77
00077 0 TRO#TRO-((HROC-HRO)/2000.0)
00000 0 GO TO 71
00079 0 IF(ABSF((TRO-TB)/TRO)-TTD)87,
00079 1 87,82
00082 0 TB#(TB+(9.0*TRO))*0.1
00000 0 GO TO 31
00087 0 TDP#TT
00090 0 DPA#0.0
```

```
00000 0 DO 91 I#1,N
00000 0 EFA(I)#(((((AEE(I,4)*TDP)+AEE
00000 1 (I,3))*TDP)+AEE(I,2))*TDP)+AEE
00000 2 (I,1)
00091 0 DPA#(D(I)/EFA(I))+DPA
00000 0 IF(KONSF)94,95,95
00094 0 PUNCH,EFA(K),DPA,DT
00095 0 IF(ABSF((DT-DPA)/DT)-TTF)99,
00095 1 99,96
00096 0 TDP#TDP-((DT-DPA)/2.0)
00000 0 GO TO 90
00099 0 TBP#TRO
00100 0 BPA#0.0
00000 0 DO 102 I#1,N
00000 0 EFA(I)#(((((AEE(I,4)*TBP)+AEE
00000 1 (I,3))*TBP)+AEE(I,2))*TBP)+AEE
00000 2 (I,1)
00102 0 BPA#(RO(I)*EFA(I))+BPA
00000 0 IF(KONSF)104,105,105
00105 0 IF(ABSF((ROT-BPA)/ROT)-TTF)120
00105 1 ,120,106
00106 0 TBP#TBP+((ROT-BPA)/0.3)
00000 0 GO TO 100
00120 0 PUNCH,TOILT,RGT,TRO,TT,PRS,N,K
00000 0 PUNCH,REP,TRY,TDP,TBP
```

```
00000 0 PUNCH,TTA,TTB,TTC,TTD,TTE,TTF,
00000 1 TTG
00000 0 DO 121 I#1,N
00121 0 PUNCH,TOL(I),RO(I),D(I),RG(I)
00000 0 PUNCH,TOLT,ROT,DT
00000 0 DO 122 I#1,N
00000 0 DO 122 J#1,4
00122 0 PUNCH,AEE(I,J)
00000 0 DO 123 I#1,N
00000 0 DO 123 J#1,3
00123 0 PUNCH,APEL(I,J),APEV(I,J)
00000 0 END
```

DEFINITION OF FORTRAN TERMS

GENERAL

All data are entered and all results are given in floating point notation.

The maximum number of individual components is 20, and terms relating to individual components are subscripted I. I represents all components in any consistent order, I through N.

Constant terms in equilibrium and enthalpy equations are subscripted J. J equals 1 for the constant A; J equals 2 for the constant B; J equals 3 for the constant C; and J equals 4 for the constant D.

The temperature, T, can be expressed in either degree Fahrenheit or Rankine, but consistent units must be used throughout.

Enthalpy terms are expressed as BTU per mol.

DIMENSIONED TERMS

Statement Number	Term	Definition
	TOIL (I)	Lean oil composition expressed as mol fraction.
	RG (I)	Rich gas composition expressed as 100 mols of inlet.
	AEE (I,J)	Constants for determining equilibrium factors using third degree polynomial, $K = A+BT+CT^2+DT^3$.

Statement Number	Term	Definition
	APEL (I,J)	Constants for determining partial enthalpies of liquids using second degree polynomial, $H_l = A+BT+CT^2$.
	APEV (I,J)	Constants for determining partial enthalpies of vapors using second degree polynomial, $H_v = A+BT+CT^2$.
	EE (I,J)	Computed factors that do not change during successive iterations. J equals 1 for equilibrium factors at the top tray (residue gas) temperature, TT. J equals 2 for partial enthalpies of liquids at the lean oil temperature, TOILT. J equals 3 for partial enthalpies of vapors at the rich gas temperature, RGT. J equals 4 for partial enthalpies of vapors at the residue gas temperature, TT.
	EFB (I,J)	Computed factors that may change during successive iterations.
	EFA (I,J)	Computed factors that may change during successive iterations.
	EFS (I)	Computed factors that may change during successive iterations.

Statement Number	Term	Definition
	D (I)	Computed quantity of individual components in residue gas, mols per 100 mols of rich gas.
	RO (I)	Computed quantity of individual components in rich oil, mols per 100 mols of rich gas.
	TOL (I)	Computed quantity of individual components in lean oil, mols per 100 mols of rich gas.
PROGRAM SECTION		
1999	TOILT	Lean oil temperature, consistent units.
	RGT	Rich gas temperature, consistent units.
	PRS	System pressure which is not used in the calculations, various units.
	TRY	Number of theoretical trays.
	REP	Recovery of key component expressed as a fraction.
	N	Number of individual components. Minimum is 1; maximum is 20.
	K	Designation for key components. In the order that individual components are listed, numerically for key I = K.
3	TTA	Tolerance for check of calculated absorption versus specified (see statement 12), normally specify .01.

Statement Number	Term	Definition
	TTB	Tolerance for check of quantity of key component absorbed versus specified (see statement 51). Expression converges to zero and normally specify 0.001.
	TTC	Tolerance in heat balance to determine rich oil temperature (see statement 76), normally specify .001.
	TTD	Tolerance in comparing calculated rich oil temperature with that assumed (see statement 79), normally specify .005.
	TTE	Temperature difference between lean oil and residue gas in consistent units. A positive number indicates that residue gas is at the higher temperature, (see statement 7).
	TTF	Tolerance in dew point and bubble point calculations (see statements 95 and 105), normally specify .001.
	TTG	Factor which varies the rate of adjustment in one convergence routine (see statement 52), normally specify 1.6.

Statement Number	Term	Definition
	TT	Temperature of residue gas and top theoretical tray, in consistent units.
9	HRG	Total enthalpy of rich gas.
	PHIP	Specified fraction of key component not absorbed, equal to one minus specified recovery of key.
	EFD	Effective absorption factor corresponding to fraction of key component not absorbed.
11	PHIC	Calculated $\varnothing_a$ from assumed A_e in routine to to determine specified A_e.
18	DT	Total residue gas volume, mols per 100 mols of rich gas.
	GOR	External oil-gas ratio, mols of lean oil per mol of rich gas.
	EFA (1)	Preliminary $\varnothing_a$ using assumed effective temperature.
	EFS (I)	Preliminary $\varnothing_s$ using assumed effective temperature.
	SBA	Total mols absorbed, equal to 100 minus DT.
30	TB	Temperature of rich oil and bottom theoretical tray, in consistent units.
31	GORT	Oil-Gas ratio (L/V) for the top theoretical tray which is external oil-gas ratio (GOR) adjusted for quantity absorbed.

Statement Number	Term	Definition
	GORB	Oil-gas ratio (L/V) for the bottom theoretical tray which is external oil-gas ratio (GOR) adjusted for quantity absorbed.
33	TOLT	Total lean oil volume, mols per 100 mols of rich gas.
	EFA (I)	Equilibrium factors for all components at rich oil temperature, TB.
	EFB (I,1)	Effective absorption factors using Edmister method.
	EFS (I)	Effective stripping factors using Edmister method.
	EFB (I,2)	Absorption-factor function.
	EFB (I,3)	Stripping-factor function.
57	HLO	Total enthalpy of lean oil.
	HD	Total enthalpy of residue gas.
	HRO	Total enthalpy of rich oil.
65	ROT	Total rich oil volume, mols per 100 mols of rich gas.
	TRO	Rich oil temperature calculated from heat balance, consistent units.
71	HROC	Total enthalpy of rich oil calculated at temperature TRO.

Statement Number	Term	Definition
	EFS (I)	Partial enthalpies of liquid at temperature TRO.
87	TDP	Residue gas dew point, consistent units.
90	DPA	Total dew point contribution, $\sum y/K$.
	EFA (I)	Equilibrium factors at dew point temperature TDP.
99	TBP	Rich oil bubble point, consistent units.
100	BPA	Total bubble point contribution, $\sum Kx$.
	EFA (I)	Equilibrium factors at bubble point temperature TBP.

DATA FORMAT

SIMPLE ABSORBER PROGRAM

INPUT DATA

Card 1

Word 1	TOILT	Lean oil temperature
Word 2	RGT	Rich gas temperature
Word 3	PRS	System pressure
Word 4	REP	Specified fraction of key component recovery
Word 5	TRY	Number of theoretical trays
Word 6	N	Number of components
Word 7	K	Number of key component

Card 2

Word 1	TTA	Test tolerance A
Word 2	TTB	Test tolerance B
Word 3	TTC	Test tolerance C
Word 4	TTD	Test tolerance D
Word 5	TTE	Test tolerance E
Word 6	TTF	Test tolerance F
Word 7	TTG	Test tolerance G

Card 3 through N + 2

Word 1	TOIL (I)	Lean oil analysis - mol fraction. I = 1 to N
Word 2	RG (I)	Rich gas analysis - 100 mols. I = 1 to N

Card N + 3 through 5N + 2

Word 1	AEE (I,J)	Equilibrium factor constants for third degree polynomial, one constant per card and four constants per component. I = 1 to N, J = 1 to 4

Card 5N + 3 through 8N + 2

Word 1	APEL (I,J)	Liquid enthalpy constants for second degree polynomial, one constant per card and three constants per component. I = 1 to N, J = 1 to 3
Word 2	APEV (I,J)	Vapor enthalpy constants for second degree polynomial, one constant per card and three constants per component. I = 1 to N, J = 1 to 3

OUTPUT DATA

Card 1

Word 1	TOILT	Lean oil temperature
Word 2	RGT	Rich gas temperature
Word 3	TB	Rich oil temperature
Word 4	TT	Residue gas temperature
Word 5	PRS	System pressure
Word 6	N	Number of components
Word 7	K	Number of key component

Card 2

Word 1	REP	Specified fraction of key component recovery
Word 2	TRT	Number of theoretical trays
Word 3	TBP	Rich oil bubble point
Word 4	TDP	Residue gas dew point

Card 3

Word 1	TTA	Test Tolerance A
Word 2	TTB	Test Tolerance B
Word 3	TTC	Test Tolerance C
Word 4	TTD	Test Tolerance D
Word 5	TTE	Test Tolerance E
Word 6	TTF	Test Tolerance F
Word 7	TTG	Test Tolerance G

Card 4 through N + 3

Word 1	TOL (I)	Lean oil composition, mols. I = 1 to N
Word 2	RO (I)	Rich oil composition, mols. I = 1 to N
Word 3	D (I)	Residue gas composition, mols. I = 1 to N
Word 4	RG (I)	Rich gas composition, mols. I = 1 to N

Card N + 4

Word 1	TOLT	Total lean oil, mols
Word 2	ROT	Total rich oil, mols
Word 3	DT	Total residue gas, mols

Card N + 5 through 5N + 4

Word 1	AEE (I,J)	Equilibrium factor constants, same as input

Card 5N + 5 through 8N + 4

Word 1	APEL (I,J)	Liquid enthalpy constants, same as input
Word 2	APEV (I,J)	Vapor enthalpy constants, same as input

Intermediate output on negative machine setting gives data on key component, temperatures, enthalpies, and other information to assist in evaluation of tolerances and convergence routines.

NOMENCLATURE

NOMENCLATURE

Tray	Theoretical contact zone where vapor and liquid leaving are in equilibrium.
P	Total system pressure
P_v	Vapor pressure of component
P_p	Partial pressure of component in liquid
x	Mol fraction of any component in liquid
y	Mol fraction of any component in vapor
K	Equilibrium constant, equals y/x
l	Mols of any component in liquid at designated point
v	Mols of any component in vapor at designated point
L	Mols of total liquid at designated point, equal $\sum l$
V	Mols of total vapor at designated point, equal $\sum v$
A	Absorption factor, equals L/KV
S	Stripping factor, equals KV/L
$\sum A$	$A_1A_2A_3\ldots A_n+A_2A_3\ldots A_n+\ldots+A_n$
$\sum S$	$S_1S_2S_3\ldots S_m+S_2S_3\ldots S_m+\ldots+S_m$
πa	$A_1A_2A_3\ldots A_n$
πs	$S_1S_2S_3\ldots S_m$
A_e	Effective absorption factor
S_e	Effective stripping factor

ϕ_a	Absorption - factor function
ϕ_s	Stripping - factor function
f_a	Fraction absorbed
f_s	Fraction stripped
T	Temperature

<u>Subscripts</u>

$1, 2, 3 \ldots n$	Tray numbers - top to bottom for absorbers
$1, 2, 3 \ldots m$	Tray numbers - bottom to top for strippers
o	Refers to lean oil in absorbers and stripping medium in strippers
$n + 1$	Refers to rich gas to tray n in absorbers
i	Any random tray

VITA

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Candidate for the Degree of

Master of Science

Thesis: ABSORBER CALCULATIONS USING ABSORPTION AND STRIPPING - FACTOR FUNCTIONS

Major Field: Chemical Engineering

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Publications: Duane F. Hogsett. "A Sponge Oil Unit May Do the Trick." Refining Engineer 32, No. 4, pp. C-9 (April, 1960).
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