THE MEASUREMENT AND PREDICTION OF THE ENTHALPY OF FLUID MIXTURES UNDER PRESSURE

by
Victor Francis Yesavage

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Doctoral Committee:

Professor John E. Powers, Chairman
Associate Professor Robert H. Kadlec
Professor Donald L. Katz
Professor Joseph J. Martin
Professor Richard E. Sonntag

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NOMEMCLATURE

a,b,c,d,e	Constants used in Equation (18)
a,b,c	Constants in BWR Equation
A	Helmholtz free energy
A,B,C,D	Constants in flowmeter calibration equation
A_0, B_0, C_0	Constants in BWR Equation
A',B',C',D',E'	Empirical constants for mixing rules
A",B",C",D",E"	Empirical constants for mixing rules
B	Second virial coefficient
B.P.	Normal boiling point
C_p	Isobaric heat capacity
D	Correction term of Lydersen, Greenkorn, Hougen correlation
e	Base of natural logarithm
E	Total energy
f	Symbol for a functional relationship
F	Mass flow rate
F	Root mean square deviation of enthalpy departure
$g^{(2)}$	Pair distribution function
$\underline{H}$	Specific enthalpy
$\overline{H}$	Partial molal enthalpy
J	Expression in mixing rule of Joffe-Stewart, Burkhardt, Voo
k	Boltzmann constant
K	Expression in mixing rule of Joffe-Stewart, Burkhardt, Voo
M.A.B.P.	Molal average boiling point

N	Number of molecules
P	Pressure
$\dot{q}$	Rate of transfer of heat
Q	Configurational integral
r	Correction factor in mixing rule of Prausnitz, Gunn
r	Distance
R	Gas constant
s	Correction factor in mixing rule of Prausnitz-Gunn
T	Temperature
u	Configurational energy
$\underline{U}$	Specific internal energy
V	Volume
$\underline{V}$	Specific volume
$\dot{w}$	Rate of transfer of work
W	Third parameter used in this investigation
x	Mole fraction
y	Mole fraction
z	Compressibility factor
Z	Canonical ensemble partition function
α	Constant in mixing rule of Leland-Mueller
α, γ	Constants in BWR Equation
β, γ	Expression in mixing rule of Prausnitz, Gunn
Δ	Difference
ϵ	Molecular energy parameter
μ	Joule-Thomson coefficient
μ'	Viscosity

ρ	Density
σ	Molecular distance parameter
$\sum$	Summation
φ	Isothermal throttling coefficient
Φ	Intermolecular potential energy
ω	Accentric factor

<u>Subscripts</u>

c	Critical point property
i,j	Components in a mixture
j	Energy state
m	Mean Value
mix	Mixture property
r	Rotational contribution
r	Reduced property
t	Translational portion
v	Vibrational portion
x	Mixture property
0	Reference substance property
1	Component 1
1	Inlet condition
11	Component 1
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2	Component 2
2	Outlet condition
22	Component 2

<u>Superscript</u>

0	Zero pressure value

Conversion Factors for Units Used in This Study

1 Btu	=	1054.6 J(kg m^2s^{-2})
1 lb	=	0.45359 kg
1 psi	=	1.48907×10^6 kg $m^{-1}s^{-2}$
1°R	=	°K/1.8
0°F	=	459.6°R

ABSTRACT

THE MEASUREMENT AND PREDICTION OF THE ENTHALPY
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Victor Francis Yesavage

Chairman: Professor John E. Powers

The objectives of this research are (1) to modify the existing recycle flow calorimetry system in order to extend its capabilities to mixtures with higher critical temperature, (2) to complete the experimental investigation of the effects of pressure and temperature on the enthalpy of the methane-propane system at elavated pressures, (3) to use the data obtained as a basis for comparison of methods of prediction of enthalpies of mixtures, and (4) to extend methods of prediction to represent the available data.

Before obtaining measurements of enthalpy for propane and propane rich methane-propane mixtures with the recycle flow system, it was necessary to eliminate the possibility of severe composition and flow upsets. These upsets were caused by large parts of the system being at room temperature, and the presence of two-phase flow in parts of the system where liquid holdup was present. The modified flow system uses a steam heated/cooled Corblin diaphragm compressor for recycling the fluid. The valve panels, buffer tanks, and connecting lines are maintained at temperatures of 250°F. The calorimeter bath section consists of a series of baths connected such that the fluid, which may be a two-phase mixture in this part of the system, always flows in a downward direction. Two calorimeters, an isobaric and a throttling calorimeter, are used interchangeably in the system. The throttling calorimeter is

used isothermally at higher temperatures and as a Joule-Thomson device for liquids.

The effects of pressure and temperature on the enthalpy of propane, and 76.6 and 50.6 mole percent propane in methane mixtures were measured. Data were obtained at temperatures from -250 to +300°F at pressures from 100 to 2000 psia in the liquid, two-phase, critical, and gaseous regions. The data were internally self consistent for each mixture to about ±0.2 percent. Enthalpy-pressure-temperature tables and diagrams were prepared at pressures up to 2000 psia with the aid of data from the literature. For propane the table extended from -280 to +500°F and for the mixtures from -280 to +300°F. These tables are believed to be accurate to 1 Btu/lb. These results together with the previous experimental results for methane and a 5.1, 11.7, and 28.0 mole percent propane in methane mixture should adequately represent the methane-propane binary system. Direct experimental data of this type and accuracy are rare in the literature.

Experimental data were used to compare results from several published methods of prediction. This study indicated that the corresponding states principle is a fruitful approach for extending methods of prediction to represent enthalpy behavior.

A three parameter corresponding states correlation was developed which uses reference reduced enthalpy tables developed from data for methane and propane. The correlation is valid between reduced temperatures of 0.5 and 1.5 at values of reduced pressure up to 3.0. Values of enthalpy departures for nitrogen calculated from the correlation agreed with experimental results to within the experimental uncertainty of the data. This supports the validity of the three parameter corresponding states principle for pure component enthalpy departures.

The correlation was extended to methane-propane mixtures by determining an optimum set of mixing rules containing six empirical constants. This correlation predicted enthalpy departures for the mixtures which agreed with experimental results almost to within experimental uncertainty (generally 1 Btu/lb). This established the validity of the principle of corresponding states for mixtures of nonpolar, mutually nonconformal components. The mixing rules obtained were compared with other rules available in the literature, and the correlation was successfully applied to a methane-nitrogen mixture.

INTRODUCTION

A knowledge of enthalpies of fluid mixtures over a wide range of pressure and temperature is necessary for accurate engineering designs of thermal processes. The goal of this research is to increase this knowledge by obtaining accurate experimental data and extending the available methods of prediction of enthalpies of fluids making maximum use of the experimental data.

In the past most enthalpies of fluids at elevated pressures have been calculated by differentiation of volumetric data. However, this method limits the accuracy of the enthalpy data obtained. In addition, accurate volumetric data for mixtures are not in great abundance and the determination of enthalpy changes across the two-phase region involves the use of not only volumetric data and derivatives but also vapor-liquid equilibrium data and derivatives. For these reasons it is desirable to have direct experimental determinations of the enthalpy behavior of fluid mixtures under pressure. These data are quite scarce in the literature.

Due to the unlimited number of mixtures which may exist and the limited amount of reliable data, an experimental approach must be used which allows considerable generalization. The Thermal Properties of Fluids Laboratory at the University of Michigan has in the past accurately measured enthalpies of fixed gases and their mixtures over a wide range of conditions by direct flow calorimetry. Systems investigated have included methane, nitrogen, one methane-nitrogen mixture, and several mixtures of methane-rich methane-propane mixtures. A specific purpose of this research has been to expand the capabilities of the calorimetric facility so that data could be obtained for propane and propane-rich methane-propane mixtures. Determinations for these mixtures and propane would result in an accurate knowledge of the enthalpy behavior of a binary system with two components of considerably different molecular type. In addition the modified system could be used in the future to make additional measurements of mixtures or pure components with critical temperatures less than that of propane. Hopefully, such an approach would permit evaluation and extension of methods of prediction so that eventually further experimental determinations of light hydrocarbon and fixed gas mixtures could be calculated _a priori_.

There are many different methods of prediction which have been proposed for enthalpies of fluid mixtures. A comparison study[140] has been made by the American Petroleum Institute of these numerous enthalpy correlations. The data used in the investigation, however, were not plentiful and consisted mainly of data derived from volumetric properties. Data were especially lacking in the critical region. Therefore, current methods of prediction are still limited to questionable uncertainty at least in the critical region. The availability of more accurate enthalpy data based on calorimetric determinations provides an improved basis for comparison of available methods of prediction. Such comparisons serve to focus attention on methods that have potential both for accurate prediction of enthalpy data and extension to systems for which data are not available.

Therefore, the specific goals of the present research were (1) to extend the capabilities of the recycle flow system, (2) to make experimental enthalpy determinations on propane and two propane-rich methane-propane mixtures, (3) to evaluate the available methods of prediction, and (4) to select a potentially fruitful method and extend it to accurately represent the available data over a wide range of conditions.

SECTION I - PRELIMINARY CONSIDERATIONS

This introductory section presents a discussion of flow calorimetry and the necessary equations which are applied, a discussion of the data in the literature, and a summary of the techniques used in interpreting the experimental data obtained in this investigation.

Flow Calorimeters

The measurement of enthalpies of fluids at elevated pressures has been made by a number of methods. A review of experimental methods published recently by Barieau[3] supplements earlier work by Masi[84] and Faulkner.[40] Flow calorimeters have many advantages and have been used widely. The calorimeter may be designed to operate in a number of differing modes depending upon the type of enthalpy data desired. In general they are used to measure the isobaric effect of temperature on enthalpy, the isenthalpic effect of pressure on temperature, the isothermal effect of pressure on enthalpy, and the effect of composition on enthalpy at constant temperature and pressure. In the present investigation the first three of the above effects have been studied.

The first law of thermodynamics, applied to a flow calorimeter with negligible potential and kinetic energy effects, is

$$\left(\underline{H}_{T_2,P_2} - \underline{H}_{T_1,P_1} \right)_x = \frac{\mathring{q} - \mathring{w}}{F} \qquad (1)$$

where $\mathring{q}$ is the rate of heat transfer, $\mathring{w}$ the rate of work done, and F is the mass flow rate.

Isobaric Effect of Temperature on Enthalpy

To measure the isobaric effect of temperature on enthalpy the pressure difference $P_2 - P_1$ is made as small as possible, and energy is added to the fluid to change its temperature. A small correction is made for the fact that the pressure is not constant. For a calorimeter with electrical energy input, $-\dot{w}$, and negligible heat leak, $\dot{q}$, Equation (1) becomes

$$\left(\underline{H}_{T_2} - \underline{H}_{T_1} \right)_{P_2, x} = - \frac{\dot{w}}{F} - \int_{P_1}^{P_2} \left(\frac{\partial \underline{H}}{\partial P} \right)_T dP \Bigg|_{T_1} \tag{2}$$

Several types of flow calorimeters have been developed to determine the isobaric effect of temperature on enthalpy. Partington and Shilling[98] present a review of the early designs. The isobaric calorimeter of Faulkner,[40] which is used in this investigation, utilizes electrical heating in the internals of the calorimeter followed by passage of the fluid through a series of concentric shells to ensure uniformity of temperature. Heat leakage is reduced to negligible proportions by heating a radiation shield located in the vacuum jacket to the temperature of the exiting gas.

Isothermal Effect of Pressure on Enthalpy

The measurement of the effect of pressure on enthalpy can be accomplished by causing a pressure drop in the flowing fluid. Energy is added so that the inlet temperature is equal to the outlet temperature. From a knowledge of the flow rate, the energy input, and the pressure drop, the isothermal effect of pressure on enthalpy can be determined by

$$\left(\underline{H}_{P_2} - \underline{H}_{P_1} \right)_{T_1, x} = - \frac{\dot{w}}{F} - \int_{T_1}^{T_2} C_p \, dT \Bigg|_{P_2} \tag{3}$$

where $\dot{q}$ is assumed to be negligible and the integral term is a correction for any slight difference between inlet and outlet temperatures. The calorimeter recently developed by Mather[85] and used in this study uses a capillary coil to induce the pressure drop with an insulated heating wire coaxially inserted into the capillary. The gas exiting from the capillary is passed through a series of concentric baffles to ensure uniformity of temperature, and the input of electrical energy to the internal heating wire is adjusted to make the outlet temperature of the fluid equal to the inlet temperature, therefore making the entire process very nearly isothermal.

Isenthalpic Effect of Pressure on Temperature

The isenthalpic mode of flow calorimetry was first studied by Joule and Thomson.[64] In this type of experiment the fluid is throttled from a high pressure to a low pressure in an isenthalpic expansion. Equation (1) becomes

$$\underline{H}_{T_2,P_2} - \underline{H}_{T_1,P_1} = 0 \quad (4)$$

Hoxton[56] gives a good summary of the experimental work in this area prior to 1920. Johnston and White[59] review the Joule-Thomson determinations from 1920 to 1948. These summaries have been brought up to date (1967) by Yesavage *et al.*[152]

Previous Experimental Data

All of the above methods have been used to obtain thermal data of fluids at elevated pressures. Most of the results have been for pure components. Reviews of data have been prepared by Masi,[84] Barieau,[3] Johnston and White,[59] Potter,[105] and Mage.[77] A recent review of the available mixture data at elevated pressures has been presented by Yesavage et al.[152] and Mather.[85]

Methane

For methane a compilation of thermodynamic properties including a comprehensive literature search is presented by Tester et al.[139] Since this compilation was published additional determinations were made by Jones,[62] Sahgal et al.,[129] Colwell, Gill, and Morrison,[18] Vennix,[142] Douslin et al.,[34] and Huang.[57]

Propane

A similar compilation of the thermodynamic properties of propane is presented by Kuloor, Newitt, and Bateman.[69] Since that time additional determinations have been made by Dittmar, Schulz, and Strese,[33] Helgeson and Sage,[51] Yarborough and Edmister,[148] Ernst,[39] Huang,[57] and Finn.[42]

Several of these previous determinations are similar to the present investigation in that they contain experimental measurements of the thermal properties of propane. Gaseous heat capacities at atmospheric pressure have been reported by Kistiakowski and Rice,[67] Dailey and Felsing,[25] and Sage, Webster, and Lacey.[128] Saturated liquid heat capacities are given by Kemp and Egan,[66] and Dana et al.[26] Joule-Thomson coefficients have been measured by Sage, Kennedy, and Lacey[125] in the gaseous region. The isothermal effect of pressure on enthalpy has been measured by Yarborough and Edmister[148] at pressures up to 1000 psia at temperatures between 200 and 400°F. Latent heats of vaporization have been reported by Dana et al.,[26] Kemp and Egan,[66] Sage, Evans, and Lacey,[124] and Helgeson and Sage.[51] The specific heat at constant pressure at 700 psia, about 80 psi above the critical point, has been reported by Finn.[42] Gaseous heat capacities have been measured by Ernst[39] between temperatures of 68 and 176°F at pressures up to 118 psia.

Methane-Propane Mixtures

A review of the physical properties data available for the methane-propane system, including vapor-liquid equilibrium, compressibilities, Joule-Thomson coefficient, enthalpy changes on vaporization, heat capacities, viscosities, thermal conductivities, and surface tension has been presented by Mather.[85]

For these mixtures a few experimental thermal properties investigations have been made. Joule-Thomson coefficients have been reported for three mixtures of methane-propane by Budenholzer et al.[11] at pressures up to 1500 psia in the temperature range between 70 and 310°F. Head[50] measured Joule-Thomson coefficients for a mixture of 51.1 mole percent propane in methane at pressures up to 40 atm between 260 and 360°K in the single phase region.

Cutler and Morrison[23] have measured the vapor pressures and heat capacities of saturated liquid mixtures of methane-propane in the temperature range of 90 to 110°K. Dillard[32] presents values of the isothermal effect of pressure on enthalpy for two methane-propane mixtures in the region between 90 and 200°F at pressures up to 2000 psia.

Manker[80] obtained isobaric data on a nominal 5 percent propane in methane mixture at temperatures from -245 to 87°F at pressures from 250 to 2000 psia. Mather[85] obtained data on the isothermal effect of pressure on enthalpy for this mixture at pressures up to 2000 psia in the temperature range of -147 to 201°F and extended the isobaric data to +257°F. In addition, Mather[85] reports isobaric data for a 12 percent and a 28 percent propane in methane mixture at pressures between 250 and 2000 psia and temperatures between -230 and 150°F.

Interpretation of Data

The basic data are recorded in terms of quantities that can be readily measured, such as microvolts, height of a fluid and weight. These quantities are converted to temperatures, pressures, power input and flow rate in the manner discussed by previous authors.[40,62,80,85] At this point Equations (2), (3), and (4) can be used to determine integral changes of enthalpy. In the single phase region these data may be interpreted to yield the derivative properties, heat capacity, C_p, Joule-Thomson coefficient, μ, and isothermal throttling coefficient, φ:

$$C_p = \left(\frac{\partial \underline{H}}{\partial T} \right)_{P,x} = \lim_{\Delta T \to 0} \left[\frac{\underline{H}_{T_2} - \underline{H}_{T_1}}{T_2 - T_1} \right]_{P,x} \tag{5}$$

$$\mu = \left(\frac{\partial T}{\partial P} \right)_{\underline{H},x} = \lim_{\Delta P \to 0} \left[\frac{T_2 - T_1}{P_2 - P_1} \right]_{\underline{H},x} \tag{6}$$

$$\varphi = \left(\frac{\partial \underline{H}}{\partial P} \right)_{T,x} = \lim_{\Delta P \to 0} \left[\frac{\underline{H}_{P_2} - \underline{H}_{P_1}}{P_2 - P_1} \right]_{T,x} \tag{7}$$

These three derivatives are related by the mathematical identity

$$\varphi = -\mu C_p \tag{8}$$

These derivitive properties are obtained from the integral data by several techniques. Both graphical and computer methods are used with the choice depending upon the type of data and the region where the data are taken.

Graphical Interpretation

Graphical interpretation is used to obtain differential properties for isothermal and isenthalpic data and for isobaric data in the regions of rapid change of heat capacity with respect to temperature and in regions where extrapolation is required.

As an example one may consider determining C_p at constant pressure as a function of T. The data available from the experimental determinations are reported as sets of P, T_1, T_2, and $\left(\underline{H}_{T_2,P} - \underline{H}_{T_1,P}\right)$ over a wide range of temperature. For any one data point

$$\underline{H}_{T_2,P} - \underline{H}_{T_1,P} = \int_{T_1}^{T_2} C_p \, dT \tag{9}$$

or

$$\frac{\underline{H}_{T_2,P} - \underline{H}_{T_1,P}}{T_2 - T_1} = C_{p_m} = \frac{\int_{T_1}^{T_2} C_p \, dT}{T_2 - T_1} \tag{10}$$

where C_{p_m} is the mean heat capacity between temperatures T_1 and T_2. Values of C_{p_m} are plotted as horizontal lines as shown in Figure 12. The smooth C_p curve is obtained by satisfying Equation (10): the area under the horizontal line segments, $C_{p_m}*(T_2 - T_1)$, should equal the area under the smooth heat capacity curve. This same technique is used to determine the isothermal throttling coefficient as a function of pressure. For determining μ as a function of P, however, this technique can be applied only if all of the data points obtained are on the same line of constant enthalpy. This type of an experimental approach is not practical with the present facility and some other procedure must be used.

In this investigation for isenthalpic determinations data at different pressures were obtained with the same inlet temperature, T_1. The inlet pressure was continually reduced as data points were taken, One approximate technique can be used if the temperature difference, $T_2 - T_1$, is small and if μ is not a strong function of temperature. This is to assume that

$$\left(\frac{T_2 - T_1}{P_2 - P_1} \right)_{\underline{H}} = \mu_m \tag{11}$$

where μ_m is the mean Joule-Thomson coefficient between pressures P_2 and P_1 at an arithmetic average temperature, $T_m = (T_1 + T_2)/2$. From values of μ_m, P_2, and P_1, the function μ versus P at an average temperature, T_m, can be approximated by the equal area graphical technique described above.

A second technique relates the isenthalpic temperature and pressures to the isothermal throttling coefficient. For an isenthalpic determination Equation (4) applies, and if $\underline{H}_{P_2,T_1}$ is added to both sides

$$\left(\underline{H}_{P_2} - \underline{H}_{P_1} \right)_{T_1} = - \left(\underline{H}_{T_2} - \underline{H}_{T_1} \right)_{P_2} \tag{12}$$

By dividing both sides by $P_2 - P_1$

$$\frac{\left(\underline{H}_{P_2} - \underline{H}_{P_1} \right)_{T_1}}{P_2 - P_1} = - \frac{\left(\underline{H}_{T_2} - \underline{H}_{T_1} \right)_{P_2}}{T_2 - T_1} \quad \frac{T_2 - T_1}{P_2 - P_1} \tag{13}$$

but

$$\varphi_m \Big|_{T_1} = \frac{\left(\underline{H}_{P_2} - \underline{H}_{P_1} \right)_{T_1}}{P_2 - P_1} \tag{14}$$

$$C_{p_m}\Big|_{P_2} = \frac{\left(\underline{H}_{T_2} - \underline{H}_{T_1}\right)_{P_2}}{T_2 - T_1} \tag{15}$$

and

$$\mu_m = \frac{T_2 - T_1}{P_2 - P_1} \tag{16}$$

where μ_m is the average Joule-Thomson coefficient measured in the experiment, $C_{p_m}\Big|_{P_2}$ is the average heat capacity which is determined from isobaric data, and $\varphi_m\Big|_{T_1}$ is the average isothermal throttling coefficient at temperature T_1, thus

$$\varphi_m\Big|_{T_1} = -C_{p_m}\Big|_{P_2}\mu_m \tag{17}$$

The smooth isothermal throttling coefficient can be obtained as a function of pressure at temperature T_1 by the same graphical technique described above.

Computer Reduction

The above graphical procedures are extremely time consuming, and in addition the equal area construction can easily lead to errors. Therefore, a computer program was developed for the interpretation of integral data. The equations used in the program are obtained as follows. Let us assume that the enthalpy at any temperature and a given pressure can be well represented over a limited temperature range by a truncated power series.

$$\underline{H} = a + bT + cT^2 + dT^3 + eT^4 \tag{18}$$

For any two temperatures

$$\underline{H}_1 = a + bT_1 + cT_1^2 + dT_1^3 + eT_1^4 \tag{19}$$

$$\underline{H}_2 = a + bT_2 + cT_2^2 + dT_2^3 + eT_2^4 \tag{20}$$

Subtracting Equation (19) from Equation (20) yields

$$\underline{H}_2 - \underline{H}_1 = b(T_2 - T_1) + c(T_2^2 - T_1^2) + d(T_2^3 - T_1^3) + e(T_2^4 - T_1^4) \quad (21)$$

Factoring $(T_2 - T_1)$ from the left hand side and dividing, results in

$$\frac{\underline{H}_2 - \underline{H}_1}{T_2 - T_1} = C_{p_m} = b + c(T_1 + T_2) + d(T_1^2 + T_1T_2 + T_2^2) + e(T_1^3 + T_1^2T_2 + T_1T_2^2 + T_2^3) \quad (22)$$

From data sets of $\left(\underline{H}_{T_2} - \underline{H}_{T_1}\right)$, T_1, and T_2, the constants b, c, d, and e can be determined by a multivariable least squares regression. $\underline{H}$ can then be determined from Equation (18) relative to a base enthalpy, a. Differentiating Equation (18) results in

$$C_p = b + 2cT + 3dT^2 + 4eT^3 \quad (23)$$

Thus in this way C_p can be determined as a function of temperature over a given temperature range.

Since data are obtained over a wide range of temperatures and it would not be expected of Equations (18) and (23) to accurately represent the data over such a range, the computer program was written to use the following approach. All experimental data for a given isobar are read in, and the program sorts the data in ascending order of the average temperature. Values of the enthalpy (relative to an arbitrary base) and the heat capacity are calculated at equal temperature intervals between a set initial and final temperature. The constants are determined at a temperature by fitting only eight data points. The data points used are the four nearest points below and above the temperature of the calculation. At the ends of the interval the first or last eight data points are used to determine the constants. The constant a is selected to make the enthalpy a continuous function of temperature.

The constants in the fitting equation are continually changing from calculation to calculation. This procedure ensures that the equation will more closely fit the experimental data. However, this will cause the heat capacity to be discontinuous. This is not a serious drawback in regions where there are abundant data and the heat capacity is not a strong function of temperature. Nonetheless, in obtaining the final heat capacity values the results from the computer output are plotted and a smooth curve drawn through these points.

Where the heat capacity is a strong function of temperature the fit is generally so poor that the graphical technique must still be used. In addition, it has been found that the equations do a poor job of extrapolating and when this is necessary again graphical methods are used.

Although the program does not eliminate graphical methods, it does reduce the amount of equal area curve fitting which can lead to error. It also eliminates much of the graphical integration needed to determine enthalpy; and even when it cannot be applied it still can be used as a check of the graphical results. A listing of the MAD computer program which determines $\underline{H}$ and C_p is given in Table LVIII of Apendix C.

Extension to Low Pressures

Since the lower limit in pressure of the recycle flow system is 100 psia, data from the literature are used in extending the enthalpy results to zero pressure. Isobaric enthalpy differences at zero pressure can be determined from ideal gas heat capacities which are related to the rotational, vibrational, and translational motion of a single molecule. In addition, experimental values of heat capacities at elevated pressures when plotted versus pressure should extrapolate to the ideal gas heat capacity at zero pressure.

The effect of pressure on enthalpy is determined at low pressures using $P\underline{V}T$ data and second virial coefficients. The isothermal throttling coefficient can be related to volumetric properties by

$$\varphi = \underline{V} - T\left(\frac{\partial \underline{V}}{\partial T}\right)_P \tag{24}$$

At zero pressure this equation can be used to relate φ_0 to the second virial coefficient

$$\varphi^0 = B - T\left(\frac{dB}{dT}\right) \tag{25}$$

The second virial coefficient is related to the interaction of two molecules. For binary mixtures B is of the form

$$B_{mix} = x_1^2 B_{11} + 2x_1 x_2 B_{12} + x_2^2 B_{22} \tag{26}$$

where B_{11} and B_{22} are the pure component second virial coefficients and B_{12} is the interaction virial coefficient. Also

$$\frac{dB_{mix}}{dT} = x_1^2\left(\frac{dB_{11}}{dT}\right) + 2x_1 x_2\left(\frac{dB_{12}}{dT}\right) + x_2^2\left(\frac{dB_{22}}{dT}\right) \tag{27}$$

The experimental isothermal throttling coefficients obtained at elevated pressure should extrapolate to the zero pressure values derived from the second virial equation using Equation (25). The resulting curve is integrated to determine the effect of pressure on enthalpy at low pressure.

<u>Enthalpy Change on Vaporization</u>

For pure component data in the two-phase region Equation (2) and

(3) can be used directly to determine enthalpies of vaporization. For pure components vaporization occurs at constant pressure and temperature so that the resulting enthalpy change of either an isobaric or isothermal experiment should be identical. For mixtures, however, vaporization must cause change of temperature, pressure, or both and thus the enthalpy of vaporization will depend on both the initial and final conditions. Equation (2) can be used to interpret isobaric data through the two-phase region and thus one obtains an isobaric enthalpy of vaporization for a mixture. If the inlet temperature, T_1, is such that a condition in the two-phase region results, an appropriate value of φ must be used in the pressure correction term. Equation (3) is used to reduce isothermal data, and, therefore, one obtains an isothermal enthalpy of vaporization for a mixture. An appropriate value of C_p must be used if the outlet fluid is in the two-phase region.

SECTION II - METHODS OF PREDICTION

This section presents a review of the methods of prediction of thermodynamic properties and particularly enthalpy of mixtures. As indicated a limited number of accurate enthalpy data are available for mixtures. However, the number of mixtures which have been investigated is infinitesimal relative to the number of systems of interest so that it is fruitless to contemplate the possibility of obtaining data for all such systems. Therefore, it is essential that reliable methods of predicting enthalpy behavior be developed.

Statistical Mechanics Background

With the advent of advanced calculational techniques including electronic computers, one would hope to be able to calculate enthalpy data for mixtures from a detailed knowledge of the behavior of individual molecules and of the interaction between molecules. Statistical mechanics has been applied with some success in this endeavor especially with respect to the behavior of gases containing relatively simple molecules. However, for mixtures in the liquid or dense fluid region this approach has not been extremely successful in obtaining quantitative representation of the macroscopic behavior. A brief discussion of the advances and problems in this endeavor will now be presented.

Fundamental Equation of State

The statistical mechanical theory relevant to the determination of thermodynamic properties of fluid mixtures at elevated pressure has been derived and discussed in numerous textbooks and reviews, e.g., Hill,[52] Prigogine,[107] Rowlinson,[120] Guggenheim.[47] The main goals of statistical mechanics in this endeavor are the determination of the potential energy of interaction of molecules, and the derivation of an equation of state which relates the thermodynamic properties of mixtures to the intermolecular potential.

Intermolecular Potential Functions

There are numerous discussions of the methods used in the determination of intermolecular potential functions, such as Rowlinson,[120] Hirschfelder, Curtiss, and Bird,[55] and more recently Klein.[68] These functions can in principle be calculated from quantum mechanics. However, except for the most simple molecules, such an approach is not mathematically feasible since it involves a detailed knowledge of the behavior of every electron for each molecule. Thus, the potential energy function must be obtained from macroscopic data by use of a simplified equation of state in a region where the equation applies accuarately. A functional form is assumed for the potential energy and the parameters fit to represent the macroscopic data.[155] Such an approach is often quite arbitrary, especially for non-spherical molecules where the potential is a function of orientation as well as distance. In addition, the potential energy function between two molecules is effected by the presence of additional molecules. Most studies of intermolecular potentials, however, are limited to systems which assume pairwise additivity and ignore this effect.

If a knowledge of this potential energy function were available it would then be necessary to substitute it into an equation of state in order to determine macroscopic properties. The basic equations of statistical mechanics are the expressions for the partition functions.

Partition Functions

An example of such an equation which can be used for the determination of thermodynamic properties is the expression for the canonical ensemble partition function, $Z(N,V,T)$

$$Z(N,V,T) = \sum_j e^{-E_j(N,V)/kT} \tag{28}$$

where E_j is the total energy of a system in an energy state j, and k is the Boltzmann constant. This partition function can be related to macroscopic thermodynamic functions by

$$A = -kT\ell nZ(N,V,T) \tag{29}$$

where A is the Helmholtz free energy

If it is assumed that the different contributions to the partition function are independent then

$$Z(N,V,T) = Z_t Z_r Z_v Q \tag{30}$$

where Z_t, Z_r, and Z_v are the contributions of the partition function for translational rotational and vibration motion, and Q is the configurational integral.

If the configurational portion is assumed to obey the laws of classical mechanics then

$$Q = \frac{1}{N!} \int \cdots \int e^{-u/kT} \, d\vec{r}_1 \ldots d\vec{r}_N \tag{31}$$

where u is the configurational energy of a molecular arrangement. This can be readily generalized to mixtures by

$$Q = \frac{1}{\Pi_i (N_i!)} \int \cdots \int e^{-u/kT} \, d\vec{r}_1 \ldots d\vec{r}_N \tag{32}$$

where

$$\sum_i N_i = N \tag{33}$$

Π_i designates the Π product and there are i species.

Physical Models

Since the evaluation of the above expression even for the simplest potential is extremely complex a simplified physical model is often assumed to represent the behavior of a fluid. An example of such a model is the cell theory of the liquid where it is assumed that each liquid molecule occupies a single cell or lattice site. Next an assumption is made regarding the potential energy at positions inside the cell. The partition function and then the thermodynamic properties can thus be evaluated. This procedure can be similarly applied to mixtures often by making a random mixing assumption and can be applied to elongated molecules by assuming that a molecule can occupy more than one site. Additional refinements can be made to improve the physical model itself.

Several attempts have been made to apply these methods to the determination of heats of mixing of simple fluids at low pressures with reasonable success.[36,49,23] This is, however, partly due to the fact that heats of mixing represent only a very small part of the configurational enthalpy and may in fact be easier to represent.[107] In general, however, due to the number of approximations involved such methods give only a qualitative description of the properties of fluids. In addition, it is felt that due to the basic error in the physical assumption itself such methods can never be improved to the point of quantitatively predicting thermodynamic properties of fluids.[107] Finally, such methods which incorporate physical assumptions must be limited to narrow regions of applications.

Direct Calculations

A more direct approach is the solution for the macroscopic properties from the individual properties of a system of several hundred particles on a digital computer. There are two basic approaches. The method of molecular dynamics uses the classical equations of motions for each particle and averages over time.[2] The Monte Carlo method uses a statistical sampling process to select configurations in the canonical ensemble and averages over these configurations with equal weight.[119] Due to the complex calculations involved, these methods have, however, only been applied to the simplest type of molecular arrangements and potential functions which cannot be expected to represent the behavior of actual systems. Their main value has been in testing physical and mathematical models.[4,5]

Pair Distribution Functions

A third approach uses the pair distribution function, $g^{(2)}(r)$, defined as the probability of finding a molecule in a volume element dr_1 and a second molecule in a second volume element dr_2. For molecules with a force field depending only on r (spherical pairwise additive molecules) this expression can be related directly to the configurational energy

$$\underline{U} - \underline{U}_0 = \frac{N\rho}{2} \int_0^{\infty} g^{(2)}(r)\Phi(r)4\pi r^2 dr \qquad (34)$$

where ρ is the density and $\Phi(r)$ the potential function. Knowledge of the distribution function can be obtained from diffraction experiments[38,90] or calculated by various approximate techniques.[121] The above expression can be readily generalized to mixtures. Again, however, the method has severe limitations in accurately predicting the thermodynamic behavior of fluids. If the molecule is not spherical the distribution function and potential function must be made functions of orientation. Little or no progress has been made in the solution of such a problem. In addition calculation of thermodynamic functions are very sensitive to small errors either in the distribution function or potential function.[114]

Recently an interesting approach to the calculation of thermodynamic properties of simple fluids, limited to moderate densities, has been developed by Orentlicher and Prausnitz.[94,95] Simplifying models for the distribution function and potential function in terms of three parameters are made. These models are substituted into an equation for enthalpy departure analogous to Equation 34. The three parameters are then empirically fit to best represent the thermodynamic data. Although the parameters have no physical significance, the flexibility of the expressions which are developed allow the authors to claim a minimum accuracy of 10 percent. Such an approach is similar to both equation of state methods and corresponding states correlations which are discussed in later sections.

Although advances have been made in the area of predicting macroscopic enthalpy behavior from a study of microscopic properties it appears that for some time to come less sophisticated methods of prediction must play an important role. In the first place, the mathematical difficulties encountered not only make exceedingly difficult the solution of an equation of state, based on statistical mechanics, but also prevent the determination of accurate intermolecular potential functions. The simplifying assumptions which have been made are quite often themselves extremely complex mathematical problems, usually give only qualitative results, and are in general limited to narrow regions of application. To obtain quantitative representation of thermodynamic properties assumptions are required which make use of experimental data and contain parameters or functional forms having no direct fundamental significance. These approaches and other methods developed from correlating thermodynamic data will now be discussed.

Methods of Prediction Based on Thermodynamic Data

A review of the available methods of prediction of enthalpies of fluid mixtures at elevated pressures has been published recently by Nathan.[92]

The predictive methods in common use can be classified in five categories.

1) Estimation of partial enthalpies
2) Equivalent pure component method
3) Application of PVT data
4) Application of equations of state
5) Generalized corresponding states correlations

This classification is somewhat arbitrary since first of all, the application of PVT data can be used to determine enthalpies for all of the other methods. In addition, corresponding states and equations of state methods can be difficult to distinguish and fundamentally are the same. However, for the sake of classification the five categories and examples of each will now be summarized.

Estimation of Partial Enthalpies

The enthalpy per mole of a mixture, $\underline{H}_m$, can be determined exactly from a knowledge of the partial molal enthalpies of the individual components, $\overline{H}_i$, by application of the expression

$$\underline{H}_m = \sum_i x_i \overline{H}_i \qquad (35)$$

The partial molal enthalpy of a component is generally a function of composition and is, therefore, truly a mixture property.

Accurate enthalpy data are available at low pressure for a large number of pure compounds, e.g., Rossini et al.[118] For gases at zero pressure the assumption that the enthalpy of a component in a mixture is the same as the enthalpy of the pure component is an accurate one. Therefore, Equation (35) can be applied rigorously to establish the enthalpy of gaseous mixtures at zero pressure. In general, however, partial molal data are not abundant and, therefore, several methods of estimation have been suggested.

For non-polar fluids either as gases or as liquids it is often adequate to assume that the partial molal enthalpy $\overline{H}_i$ is independent of composition, i.e., that the enthalpy of the component in the mixture is the same as the enthalpy of the pure component at the same temperature and pressure. However, extreme care must be taken when applying this procedure since as Mather[85] has shown extreme values of the heat of mixing do exist, especially in the critical region. For this and other reasons, this method has been discouraged by the prediction evaluation study conducted by the American Petroleum Institute.[140] Peters[99] and Maxwell[89] suggest a modification of this procedure in applying data from their published enthalpy diagrams to predict the enthalpy of gaseous mixtures. In applying this general procedure to liquid mixtures, one is faced with the problem of estimating the partial molal enthalpy of a component as a liquid above the critical temperature of the component. Peters relates the partial molal enthalpy of hydrocarbons above their critical temperatures to the molal average boiling point of the mixture and plots values of liquid partial molal enthalpies on enthalpy diagrams for the pure components. Unfortunately the method cannot be applied for pressures above 600 psia. Maxwell includes a single line on the enthalpy diagrams of pure components to represent the partial molal enthalpy of the component in the liquid phase above its critical. In general, the plots presented by Peters are limited to a temperature range between -260 and +420°F at pressures up to 600 psia. Similarly, Maxwell's plots extend between -200 and +1200°F at pressures below 150 atm.

A major drawback of this method is the fact that an enthalpy diagram must be available for every component present in a mixture. This diagram is constructed by relying primarily on empiricism and would be difficult to extend.

Equivalent Pure Component Method

Application of this method is the simplest, generally only involving one parameter and is usually restricted to mixtures of components of homologous series. Scheibel and Jenny[130] present nomographs based on the average molecular weight of a hydrocarbon mixture.

Papadopoulos et al.[97] showed that the molal average boiling point of a mixture defined as

$$\text{M.A.B.P.} = \sum_i x_i (\text{B.P.})_i \qquad (36)$$

where the $(\text{B.P.})_i$'s are the pure component boiling points, for lighter hydrocarbons served to correlate values of a partial molal enthalpies calculated from an equation of state. Similar results were obtained at about the same time by Canjar and Edmister.[13] Canjar and Peterka[15] prepared plots of the isothermal enthalpy departure as a function of temperature and pressure for mixtures with different molal average boiling points. The plots of Canjar and Peterka are restricted to the temperature range -200 to + 500°F at pressures below 1500 psia for mixtures with molal average boiling points from -270 to +190°F.

Application of PVT Data

As mentioned earlier accurate P$\underline{V}$T data can be used to calculate the effect of pressure on the enthalpy of fluids and fluid mixtures. In general, the enthalpy of a mixture at a specified temperature and zero pressure, $\underline{H}(T,0)$, can be determined for most simple fluids. This is evaluated from ideal gas enthalpies of pure fluids either from statistical mechanics or from measured data. The mixture enthalpy is determined by applying Equation (35) as already described.

Thus, if accurate volumetric data are available for the mixtures, the enthalpy can be calculated by use of the relation

$$\underline{H}(T,P) = \underline{H}(T,0) + \int_0^P \left[\underline{V} - T \left(\frac{\partial \underline{V}}{\partial T} \right)_P \right] dP_T \qquad (37)$$

In principle, the use of this relation in the single-phase region is fairly straightforward. In contrast to pure components for which $\left(\frac{\partial \underline{V}}{\partial T} \right)_P$ is infinite in the two-phase region, this derivative is finite in the two-phase region for mixtures (with the exception of mixtures of azeotropic composition). Thus Equation (37) applies throughout the two-phase region for mixtures in which case $\underline{V}$ is the total volume per mole of the mixture.

In application, the use of Equation (37) is not quite so straightforward. The term in brackets under the integral sign involves the difference between two terms, one of which includes a derivative. As a result, extremely accurate volumetric data are required to yield reliable estimates of the effect of pressure on enthalpy. A reduction of accuracy of one order of magnitude is to be expected. Volumetric data for mixtures of the required accuracy are available but somewhat rare.

Direct experimental determination of the volumetric behavior of two-phase mixtures is even more rate. The total volume, $\underline{V}$, can be related to the properties of the individual equilibrium phases. Strickland-Constable[138] has presented the resulting equations which demonstrate that in order to make use of the properties of the individual phases, extremely accurate vapor-liquid equilibrium data are required in addition to the volumetric data.

Another practical consideration is the method of obtaining accurate values of the derivative in Equation (37) from experimental data. In general, some method of curve fitting is applied. Quite often an equation of state is used for this purpose.

The next secti on describes the application of equations of state in the determination of enthalpies of mixtures.

Application of Equations of State

Reviews dealing with equations of state have been presented by Van Ness,[141] Martin,[83] and Rowlinson.[122] In general most equations of state serve to relate pressure, P, as the dependent variable to temperature, T, and specific volume, $\underline{V}$, as independent variables. As a result it is convenient to transform Equation (37) to

$$\underline{H}(T,\infty) - \underline{H}(T,\underline{V}) = RT - P\underline{V} + \int_{\infty}^{\underline{V}} \left[P - T\left(\frac{\partial P}{\partial T}\right)_{\underline{V}} \right] d\underline{V}_{\ T} \tag{38}$$

The Benedict-Webb-Rubin (BWR) equation of state[7] is commonly used to fit volumetric data and thereby estimate the enthalpy of fluids at elevated pressures. In terms of the eight constants which are used in this equation of state Equation (38) becomes

$$\underline{H}(T,\underline{V}) = \underline{H}(T,\infty) + \frac{1}{\underline{V}}\left[B_0RT - 2A_0 - \frac{4C_0}{T^2}\right] + \frac{1}{2\underline{V}^2}\left[2bRT - 3a\right] + \frac{6}{5}\,\frac{a\alpha}{\underline{V}^5} + \frac{c}{\underline{V}^2T^2}\left[\frac{3\underline{V}^2}{\gamma}\left(1 - e^{-\frac{\gamma}{\underline{V}^2}}\right) - \frac{1}{2}e^{-\frac{\gamma}{\underline{V}^2}} + \frac{\gamma}{\underline{V}^2}e^{-\frac{\gamma}{\underline{V}^2}}\right] \tag{39}$$

The eight constants required for application of this relation are usually determined from volumetric data. It is risky to use the equation to extrapolate beyond the range of the original data. Experimental data are quite extensive in some cases as is illustrated by the fact that Crain and Sonntag[22] recently published BWR constants for nitrogen which apply fairly well to reduced densities of approximately 2.

The BWR constants are only extremely rarely determined directly from volumetric data for mixtures. Instead, empirical rules have been developed for estimating values to be applied to a mixture from a knowledge of the constants for the pure components and the composition of the mixture. Combining rules for the eight constants have been suggested by Benedict, Webb, and Rubin.[8]

Constants for the BWR equation of state are available for a considerable number of pure components, and therefore Equation (39) together with the appropriate mixing rules provides a convenient means of estimating the effect of pressure on enthalpy for mixtures.

Recently constants for the BWR equation of state[75] or a modification of it[133,21,115] have been determined by fitting volumetric and enthalpy relations simultaneously. This ensures thermodynamic consistency of both types of data. Other equations of state have been used to calculate enthalpies of fluid mixtures. The virial equation can be derived from statistical mechanics considerations.[55] The nth virial coefficient is related to an n particle interaction system.

Unfortunately neither accurate thermodynamic data nor constants for equations of state such as the BWR are available for many components. Several equations of state such as the Redlich-Kwong equation[112] and the equations due to Hirschfelder et al.[53,54] use constants which are directly related to macroscopic constants for the individual components. These equations analytically represent a generalized correlation. Such correlations were developed to permit one to approximate the enthalpy of materials at elevated pressures using a very limited amount of available data. These corresponding states correlations will be described in the following paragraphs.

Generalized Corresponding States Correlations

Recently reviews of the application of the corresponding states principle have been presented by Stiel[137] and Leland and Chappelear.[74] The principle of corresponding states was first applied in its simplest form to develop generalized correlations of PVT behavior by Cope, Lewis, and Weber[20] and Brown, Souders, and Smith.[10] The law was stated as

$$z = f(P_r, T_r) \tag{40}$$

where P_r is the reduced pressure, $\frac{P}{P_c}$, and T_r is the reduced temperature, $\frac{T}{T_c}$. Later Pitzer[100] derived the expression from statistical mechanics limiting its validity to spherical molecules having a potention function depending on only two parameters. In addition the translational and configuration portions of the partition function are assumed to be independent of quantum effects. This eliminates light molecules such as H_2, He, and Ne. In addition the canonical ensemble partition function is assumed to be separable into independent internal and external factors. As expressed by Pitzer

$$z = f\left(\frac{\epsilon}{kT}, \frac{V}{N\sigma^3}\right) \tag{41}$$

where f is a universal function. The ϵ and σ can be related directly to the critical constants[107] by applying Equation (41) at the critical point. To account for the nonconformity of compounds, which do not satisfy the above restrictions later contributors have suggested use of a third correlating parameter in addition to the critical temperature and pressure. Lydersen, Greenkorn, and Hougen[76] made use of the critical compressibility factor, z_c. Pitzer et al.[102] employed the accentric factor, ω, which is related to the shape of the reduced vapor pressure curve. Substances with the same value of the third parameter are thus mutually conformal. The validity of the insertion of a third parameter from statistical mechanical considerations has been established by Pople[104] and Cook and Rowlinson.[19]

The correlations of Lydersen et al.[76] and Pitzer et al.[102] of $P\underline{V}T$ data have been used as the basis for generalized correlations of the isothermal effect of pressure on enthalpy. The total effect is considered to be the sum of two factors. The first factor gives the reduced enthalpy departure for a fluid with a standard value of the third correlating parameter and the second factor represents the influence of the deviation of this correlating parameter from the standard value. The correlation of Lydersen, Greenkorn, and Hougen has the form

$$\frac{\underline{\underline{H}}(T,0) - \underline{\underline{H}}(T,P)}{T_c} = \left[\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{T_c}\right] + (z_c - 0.27)\,[D] \tag{42}$$

where the bracketed terms are presented as generalized functions in tabular form by the authors. The conditions covered include $P_r \leq 30$ for $0.5 \leq T_r \leq 15$. Yen and Alexander[150] have published revisions of these functions in equation and graphical form. The data used to develop these revised correlations include enthalpy data at elevated pressures and extend the upper limit on T_r to 30. A recent further modification by Yen[149] applies for $P_r \leq 100$ for $0.4 \leq T_r \leq 60$.

The correlation of Curl and Pitzer[24] is given in the form

$$\frac{\underline{\underline{H}}(T,0) - \underline{\underline{H}}(T,P)}{RT_c} = \left[\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{RT_c}\right]_0 + \omega\left[\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{RT_c}\right]_1 \tag{43}$$

where the bracketed terms are somewhat different generalized functions presented in tabular form by the authors. These tables cover the range of pressures for $P_r \leq 9$ for $0.8 \leq T_r \leq 4$. Revisions of the original correlations which incorporate enthalpy data at elevated pressures in addition to $P\underline{V}T$ data have been presented recently (Yarborough[147]). The range of conditions include $P_4 \leq 10$ for $0.5 \leq T_r \leq 4$.

An approach for incorporation of the third parameter which is equivalent to the Curl-Pitzer method is used later in this investigation. This involves the use of two reference substances instead of the correction term. In this case the departure is represented as

$$\frac{\underline{\underline{H}}(T,0) - \underline{\underline{H}}(T,P)}{RT_c} = \left[\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{RT_c}\right]_1 W_1 + \left[\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{RT_c}\right]_2 W_2 \tag{44}$$

where

$$W_2 = 1 - W_1 \qquad (45)$$

and the bracketed terms are actual reduced enthalpy functions of reduced temperature and reduced pressure for two substances which are not conformal. The normalized third parameter, W_1, has a value of 1 for component one and 2 for component two. These factors weigh the reduced enthalpy departure for a substance according to its relative conformity to component one or component two. These weighing factors W_1 and W_2 can be related to another third parameter such as ω by

$$W_1 = \frac{\omega_2 - \omega}{\omega_2 - \omega_1} \qquad (46)$$

$$W_2 = \frac{\omega - \omega_1}{\omega_2 - \omega_1} \qquad (47)$$

where ω_1 and ω_2 are the values of ω for the two reference components. This latter approach was most convenient in this investigation, since accurate enthalpy data were availalbe for two reference components, methane and propane. This will be discussed further in the section on extensions of methods of prediction.

Another approach used in the application of the corresponding states principle to non-conformal substances has been developed by Leach and Leland.[70,71] Shape factors for a given species are introduced which adjust the reduced temperature and pressure to force the component to conform to a reference substance. Its main disadvantage is that it requires the addition of two more parameters even for components which differ only slightly from a reference substance.

In applying these correlations to mixtures it is assumed that the reduced functions for a mixture behave in the same manner as those for a pure component. It is necessary to establish values of the three parameters for a mixture. The actual critical properties of a mixture are in general not used, since, as is shown in several texts such as Prigogine and Defay,[108] the critical properties of a mixture do not have the same physical and thermodynamic significance as those for a pure component. Therefore, some kind of mixing rule must be developed which relates the mixture parameters to the parameters of the pure components. The study conducted by the American Petroleum Institute[140] however, does recommend use of the true critical properties in the vicinity of the mixture critical. Since heat capacity is infinite for a pure component but not for a mixture at the critical, as is pointed out in later sections, it seems doubtful that a reduced enthalpy function developed from pure component data would, at the critical, represent the mixture at its critical.

Under certain conditions the generalization of the corresponding states principle to mixtures can be justified from theoretical considerations and mixing rules suggested. For molecules of equal size with a Lennard-Jones type potential function molecular pseudoparameters can be determined from pure component parameters. This development is given in numerous texts, e.g., Rowlinson.[120] This approach uses a random mixing assumption which is not valid for molecules of different size.

Leland, Chappalear, and Gamson[73] use an approximate technique based on expansion of the distribution function to determine equations for pseudomolecular parameters. This technique is limited to mixtures of mutually conformal molecular species. However, little is known about the molecular properties of non-conformal mixtures and since the statistical mechanics equations for such systems are extremely complex there is no fundamental basis for the application of mixing rules to such complex systems. Thus, empirical methods must be used for substantiation and application of the corresponding states principle for such mixtures.

A number of mixing rules have been developed for use in applying the corresponding states principle to mixtures. The simplest mixing rules used to define mixture pseudoparameters are those suggested by Kay[65] which combine the pure component properties in a linear fashion with regard to mole fraction

$$T_{cx} = \sum_i T_{ci} x_i \tag{48}$$

$$P_{cx} = \sum_i P_{ci} x_i \tag{49}$$

$$W_x = \sum_i W_i x_i \tag{50}$$

Curl and Pitzer[24] recommend the rules of Pitzer and Hultgren[101] for their correlation. These rules add a third empirical interaction term for each parameter which is a function of the components in the mixture

$$T_{cx} = x_1 T_{c1} + x_2 T_{c2} + 2x_1 x_2 (2T_{c12} - T_{c1} - T_{c2}) \tag{51}$$

$$P_{cx} = x_1 P_{c1} + x_2 P_{c2} + 2x_1 x_2 (2P_{c12} - P_{c1} - P_{c2}) \tag{52}$$

$$\omega_x = x_1 \omega_1 + x_2 \omega_2 + 2x_1 x_2 (2\omega_{12} - \omega_1 - \omega_2) \tag{53}$$

These rules produce pseudoparameters having deviations from Kay's rule which are symmetrical with respect to composition. A generalization of these rules can be made by increasing the number of constants in the equation in order to better represent mixture behavior. Such an approach was used in this investigation and will be discussed more fully in a later section. The equations which were used are

$$T_{cx} = \sum x_i T_{ci} + x_i x_j [A' + (1 - 2x_i) B' + (1 - 2x_i)^2 C'] \qquad (54)$$

$$P_{cx} = \sum x_i P_{ci} + x_i x_j [D' + (1 - 2x_i) E'] \qquad (55)$$

and again

$$W_x = \sum x_i W_i + F' x_i x_j \qquad (56)$$

In addition there are several other mixing rules which have been proposed in the literature. The Joffe-Stewart, Burkhardt, Voo[60,61,136] equations are given as

$$T_{cx} = \frac{K^2}{J} \qquad (57)$$

$$P_{cx} = \frac{T_{cx}}{J} \qquad (58)$$

$$K^2 = \frac{T_{cx}^{\ 2}}{P_{cx}} = \left[\frac{\sum_i y_i T_{ci}}{P_{ci}^{\frac{1}{2}}} \right]^2 \qquad (59)$$

$$J = \frac{T_{cx}}{P_{cx}} = \frac{1}{8} \sum_i \sum_j y_i y_j \left[\left(\frac{T_{ci}}{P_{ci}} \right)^{\frac{1}{3}} + \left(\frac{T_{cj}}{P_{cj}} \right)^{\frac{1}{3}} \right]^3 \qquad (60)$$

These rules were originally suggested by the mixing rules for the constants in the Van der Waals equation. Since the Van der Waals equation is a two parameter equation, the strict development is limited to mutually conformal substances. However, the rules can be applied to other mixtures by assuming a linear variation in the third parameter. The rules due to Leland and Meuller[74] are given as

$$T_{cx} = \left[\frac{\sum_i \sum_j x_i x_j \left(\frac{z_c T_c^{\alpha+1}}{P_c} \right)_i^{\frac{1}{2}} \left(\frac{z_c T_c^{\alpha+1}}{P_c} \right)_j^{\frac{1}{2}}}{\sum_i \sum_j x_i x_j \left[\frac{1}{2} \left(\frac{z_c T_c}{P_c} \right)_i^{\frac{1}{3}} + \frac{1}{2} \left(\frac{z_c T_c}{P_c} \right)_j^{\frac{1}{3}} \right]^3} \right]^{\frac{1}{\alpha}} \tag{61}$$

where α is a function of

$$\frac{T \sum_i x_i P_{ci}}{P \sum_i x_i T_{ci}}$$

and is tabulated by the authors. The equation for the pseudocritical pressure is given as

$$P_{cx} = \frac{T_{cx} \sum_i x_i z_{ci}}{\sum_i \sum_j x_i x_j \left[\frac{1}{2} \left(\frac{z_c T_c}{P_c} \right)_i^{\frac{1}{3}} + \frac{1}{2} \left(\frac{z_c T_c}{P_c} \right)_j^{\frac{1}{3}} \right]^3} \tag{62}$$

These rules are obtained by equating terms of a second virial coefficient expression for mixtures with those of a pure component and making appropriate simplifying assumptions. Again, since the development was limited to mutually conformal substances the third parameter z_{cx} is given as

$$z_{cx} = \sum_i x_i z_{ci} \tag{63}$$

The mixing rules of Prausnitz and Gunn[106] are also suggested from relations for the second virial coefficient of conformal substances. These rules are given by

$$T_{cx} = \frac{\beta + \sqrt{\beta^2 + r\underline{V}_{cx}\gamma}}{2s\underline{V}_{cx}} \tag{64}$$

$$\underline{V}_{cx} = \sum_i \sum_j y_i y_j \underline{V}_{cij} \tag{65}$$

$$P_{cx} = \frac{RT_{cx}}{\underline{V}_{cx}} \sum_i y_i z_{ci} \tag{66}$$

$$\omega_x = \sum_i y_i \omega_i \tag{67}$$

The z_{ci} can be obtained from

$$z_c = 0.291 - 0.08\,\omega \tag{68}$$

The quantities β and γ are computed

$$\beta = \sum_{i,j} y_i y_j (T_c \underline{V}_c)_{ij} \tag{69}$$

$$\gamma = \sum_{i,j} y_i y_j (\underline{V}_c T_c^2)_{ij} \tag{70}$$

where

$$T_{cij} = (T_{ci} T_{cj})^{\frac{1}{2}} - \Delta T_{cij} \tag{71}$$

$$\underline{V}_{cij} = \frac{1}{2}(\underline{V}_{ci} + \underline{V}_{cj}) - \Delta\underline{V}_{cij} \tag{72}$$

where the Δ terms are small correction terms depending upon individual mixtures. Finally the quantities r and s are functions of T_r and ω tabulated by the authors. Prausnitz and Gunn also recommend a simplified rule

$$\omega_x = \sum_i y_i \omega_i \tag{73}$$

$$T_{cx} = \sum_i y_i T_{ci} \tag{74}$$

$$P_{cx} = \frac{RT_{cx}}{\sum_i y_i V_{ci}} \sum_i y_i z_{ci} \tag{75}$$

where again z_{ci} is given by Equation (68).

Finally, Reid and Leland[113] have obtained molecular mixing rules from the general expressions developed by Leland, Chappalear, and Gamson[73] for mutually conformal mixtures. Molecular rules can be obtained which are equivalent to the equations of Joffe-Stewart, Buikhardt, and Voo,[136] Leland and Mueller,[74] and Prausnitz and Gunn,[106] depending on the assumptions used for approximating higher order terms.

SECTION III - THE MODIFIED FLOW SYSTEM

The flow calorimetry facility as described by Mather,[85] Manker,[80] and Jones[62] was capable of making accurate determinations of the effect of pressure and temperature on the enthalpy of pure gases and mixtures behaving as fixed gases in the liquid two-phase, critical, and gaseous states. As originally developed by Faulkner[40] and Jones[62] it was capable of measuring the isobaric effect of temperature on the enthalpy of pure fluids under pressure. It could operate between temperatures of -250 and +50°F and at pressures between 250 and 2000 psia. Such a system as described by Jones[62] required in addition to the calorimeter itself, a series of baths to obtain constant temperature in the calorimeter, a compressor to enable recycle at a steady flow, a flow metering section, numerous buffer tanks to maintain stable pressures, storage tanks to enable adjustment of operating pressure, metering valves, and a host of accurate measuring devices. Thus, the calorimeter was only a small part of a large and complex system. When modifying the system to enable measurements on mixtures behaving as fixed gases Manker[80] and Mage[77] found that a major problem was preventing fractionation and other sources of composition upsets throughout the system. In addition in order to complete a pressure-temperature-enthalpy network for any pure or mixed system it was necessary to determine the isothermal effect of temperature on enthalpy. Thus, Mather[85] incorporated a throttling calorimeter into the system. In addition Mather increased the maximum operating temperature to +300°F.

The next step in the process was to obtain measurements on more complex systems other than pure or mixed fixed gases. Since data were available for mixtures of methane-rich methane and propane and, in addition, this system is of great practical interest, the completion of the methane-propane system was a logical direction in which to proceed. This required making measurements on pure propane as well as several mixtures of methane-propane.

The phase behavior of the methane-propane system is shown in Figure 1. It can be noted that at room temperature mixtures which are greater than 30 mole percent propane are in the two-phase region over quite a range of pressures. The desired measurement pressures are between 250 and 2000 psia and the operating pressures would thus be between 95 and 2500 psia. Much of the original system (tanks, compressor, throttling valves) was at room temperature. Therefore, it would be impossible to maintain a constant composition for mixtures which were heavy in propane in this facility. In addition for mixtures with large propane content the temperature range covered by the two-phase region at pressures below 1000 psia is as large or larger than 200°F. That is complete vaporization of a liquid at constant pressure requires at least a 200°F temperature rise. It would be very desirable to obtain isobaric enthalpy data over the complete temperature range between -250 and +300°F including the two-phase region at specified pressures. The system as decribed by Mather could not be used to obtain data all of the way across the two-phase region for mixtures with large propane content. The available energy from the power supply was not great enough at lowest flow rates to vaporize as well as heat a mixture 200°F. Also, flow and composition instabilities would not allow the fluid to enter the calorimeter in the two-phase region. Finally even with pure propane where composition fluctuations are no problem there would still

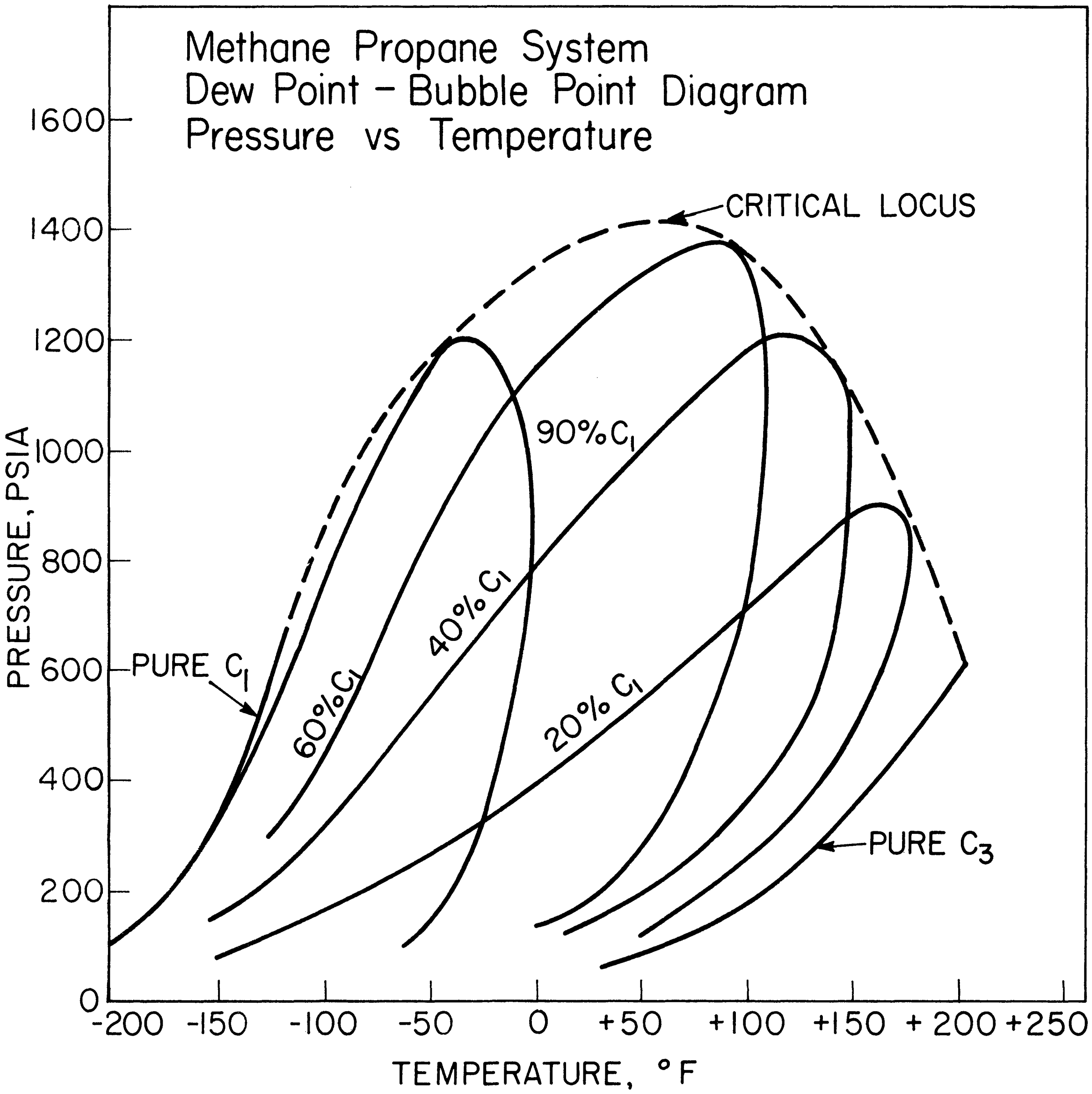


Figure 1. Vapor-Liquid Equilibrium Envelope of the Methane-Propane System

be problems of flow and pressure instabilities caused by operating the system with the fluid in the dense gas or liquid state. In addition the gas compressor could not be used to pump propane without modification.

It was therefore necessary to modify the original system in light of the above considerations. Basically the main modifications consisted of placing all parts of the system which could be in the two-phase region in a high temperature environment, with the exception of the calorimeter bath section itself since it was desirable to obtain data at temperatures as low as -250°F. The calorimeter bath section consisted of a series of cooling coils and baths arranged so that the flow was always downward since in this section two-phase flow would occur. The baths bring the fluid to the desired temperature where it enters the calorimeter. Thus the high temperature environment would eliminate composition and pressure instability. The downward flow in the baths would in addition to removing instabilities allow for measurements with the fluid entering the calorimeter in the two-phase region.

The final result would be the creation of a facility capable of accurately measuring enthalpies of mixtures as heavy as propane and as light as any fixed gas over a wide range of temperature and pressure. Accurate data of this type for mixtures over a wide range of conditions, where one component is a fixed gas and the other a heavier component, are at present almost entirely nonexistant in the literature.

Description of Equipment

The modified recirculating system capable of operating with propane as the test fluid and incorporating both isobaric and throttling calorimeters for direct experimental determinations of the effects of both temperature and pressure on enthalpy is illustrated in Figure 2. Important features of the facility are described below.

A Corblin A2CCV50-250 two stage-diaphragm compressor is used to provide recycle capabilities. Either water or steam can be used to cool or heat the heads ensuring that only a gas phase is compressed. Additional heating can be applied to the gas leaving the high pressure stages as well as between stages. The compressor is located in a transite enclosed area on the floor above the laboratory. The compressed fluid flows through two large bombs situated in an insulated metal box heated to 250°F. The first bomb contains glass wool with a layer of copper filings in the center. The glass wool is used to trap any oil which may have leaked into the system from the compressor. The copper filings are used to remove any oxygen in the system as copper oxide. Before installation of the copper filings, particles of copper oxide would form in the calorimeter bath coils and eventually plug up the calorimeters. The second bomb is filled with dehydrite and is used to remove any water from the system. When operating the calorimeter at low temperatures water in the bath coils would freeze and plug up the coils.

The compressor is operated at a constant volume rate of 4 SCFM and a bypass line is provided to permit variation in flow rate through the calorimeter. The fluid passes in a heated line from the bomb box to the valve panel which is also located in an insulated box. The fluid is heated by a heating tape and throttled by a metering valve to approximately the calorimeter inlet pressure. Although the fluid is located in a heated box the auxiliary heater is needed to prevent condensation

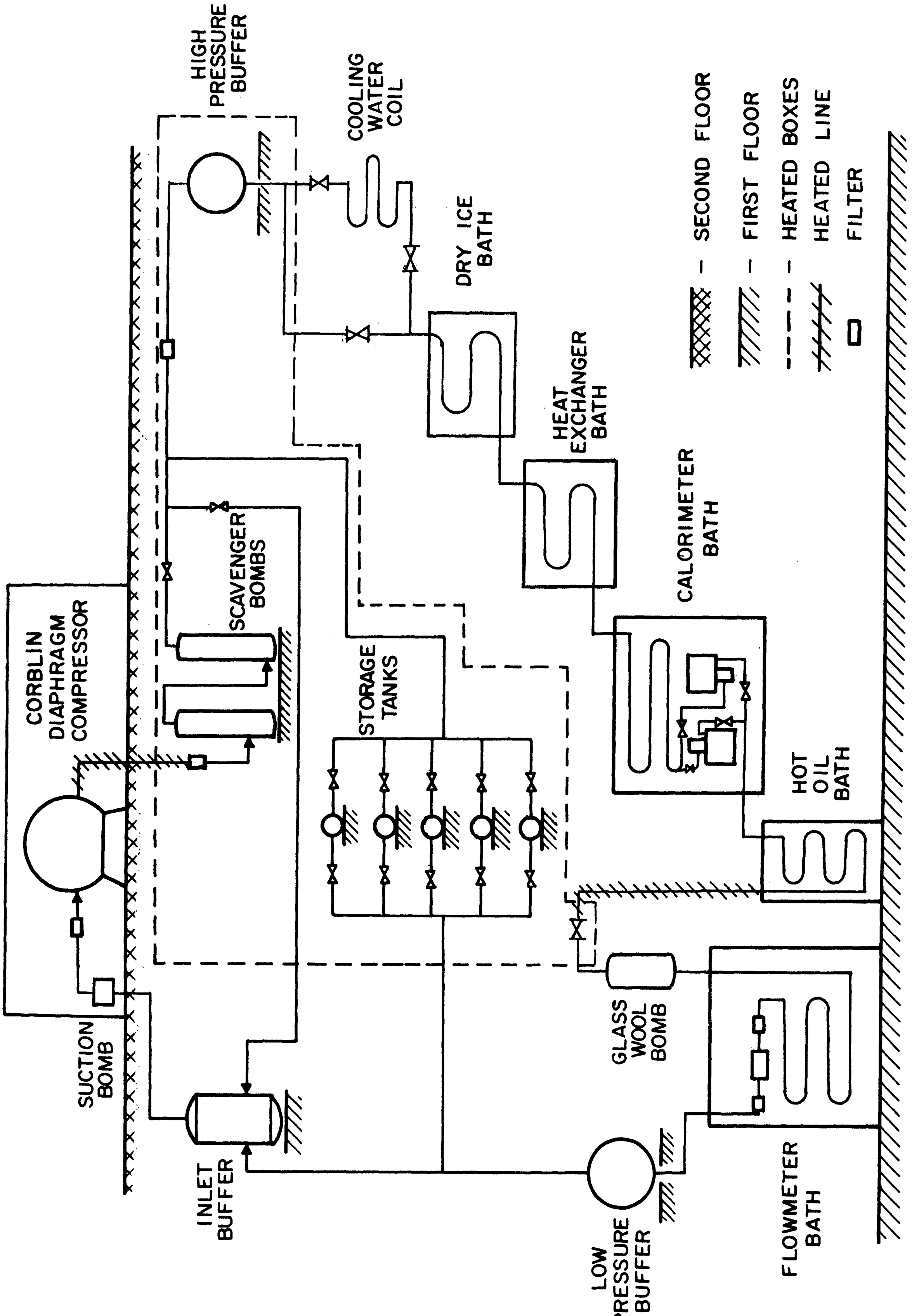


Figure 2. Flow Diagram of Modified Recycle System

caused by the cooling effect on throttling. The fluid then enters a high pressure manifold in the valve panel. The fluid which is bypassed is first heated with a heating tape and then throttled by a metering valve to the compressor suction pressure. This gas is then recycled. A bank of five storage tanks located in another heated box is provided to allow operation of the calorimeter at various pressures up to 2000 psia. Figure 3 is a photograph of this box with the cover removed.

The calorimeter high pressure buffer tank is located in the controlled temperature box with the storage tanks. The fluid entering the calorimeter section first passes through this buffer tank and then goes through a micron filter at the entrance to the calorimeter section. Should temperatures get too high in the heated section and decomposition occurs this filter acts to prevent any decomposition products from entering the calorimeter section.

The fluid at elevated temperature enters the calorimeter section at a height of about eleven feet above the floor level. As condensation and/or boiling may occur within the calorimeter section every effort has been made to reduce holdup in this section and to arrange piping such that flow of fluid is downward throughout this section.

A water cooling coil is used when operating a calorimeter below room temperature and may be bypassed.

A dry ice bath provides a convenient means of reducing the temperature of the recycle gas to about -100°F when operating a calorimeter at low temperatures. This bath contains 175 feet of 3/16" O.D. copper tubing. The bath is constructed of stainless steel and is contained in a wooden box with a layer of Styrofoam insulation between the bath and the wood.

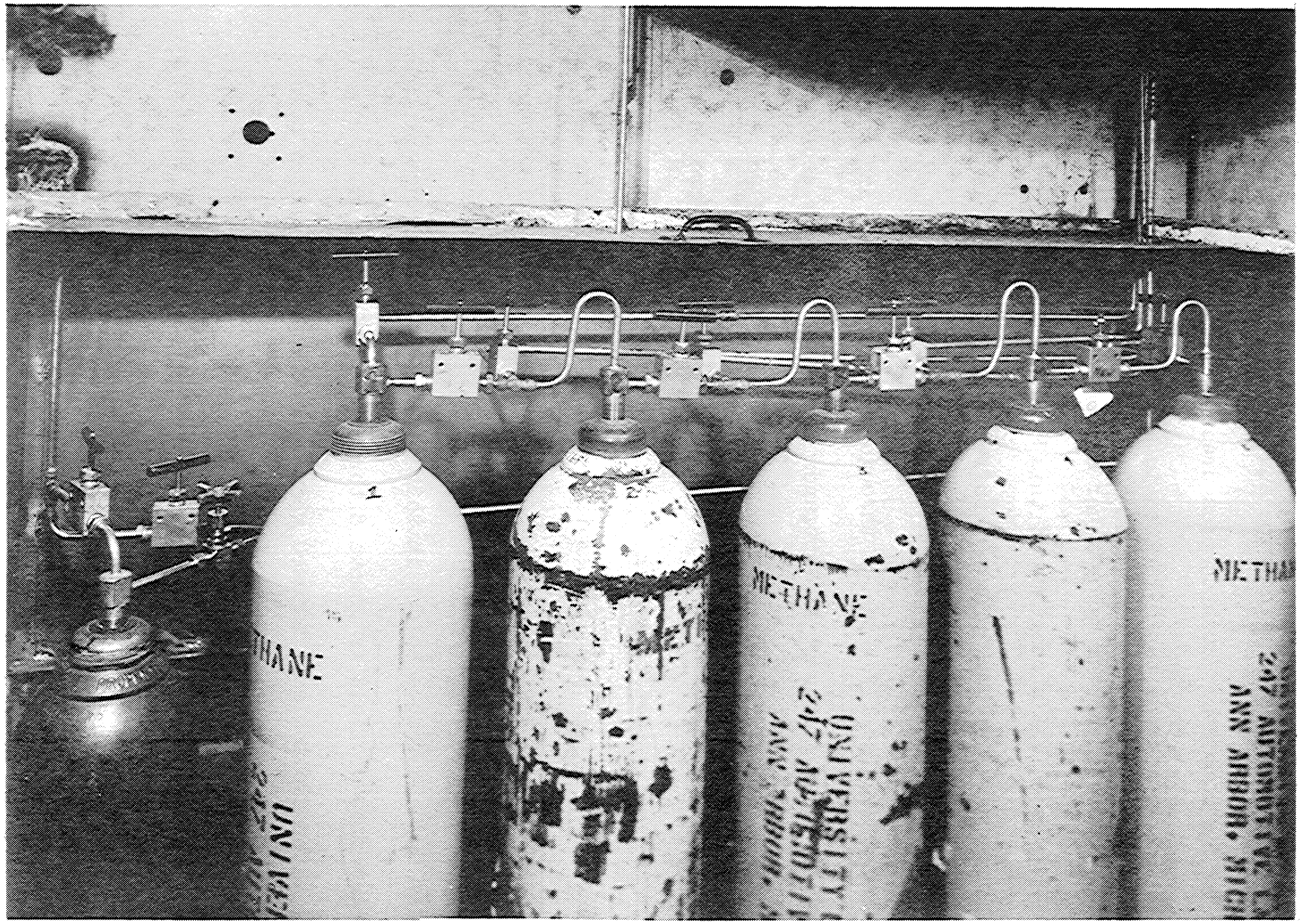


Figure 3. View of Storage Tanks with the Front of the Insulated Box Removed

The heat exchanger bath is designed to bring the temperature of the fluid close to that of the calorimeter bath. In this bath the fluid passes through 325 feet of 3/16" O.D. copper tubing. Cooling is provided by use of liquid nitrogen at low temperature and compressed air at elevated temperatures. The bath is well stirred and the temperature is controlled to within ±0.5°F by a Bailey electronic controller driving a 50 watt immersion knife heater.

The fluid from the heat exchanger bath passes through 100 feet of 3/16" O.D. copper tubing in the calorimeter bath before entering either the isobaric or throttling calorimeter. Four packless valves with stainless steel bellows (Hoke TY-445) located within the bath make it possible to operate either calorimeter with only minor adjustments. The calorimeter bath is well stirred and a Honeywell proportional-reset controller is used to obtain stable temperatures within ±0.1°F. A photograph of the outside of this bath is shown in Figure 4. The hot oil bath at the exit of the calorimeter bath serves the purpose of ensuring that the fluid leaves the calorimeter section as a single phase. The temperature of this bath is controlled to within ±2°F by a Fenwal controller driving a coil immersion heater. When operating at low temperatures additional immersion heaters are used to supply a sufficient heat input in order to vaporize all of the flowing fluid.

After leaving the calorimeter section the fluid returns to the valve panel where it is heated by a heating tape and throttled to approximately 80 psig, the flow meter pressure. The fluid next passes through a second glass wool bomb just outside of the flowmeter bath and then is brought to 27°C, ±0.02°C in the water filled flowmeter bath. The temperature is controlled by a mercury contact switch and a Fisher relay. The

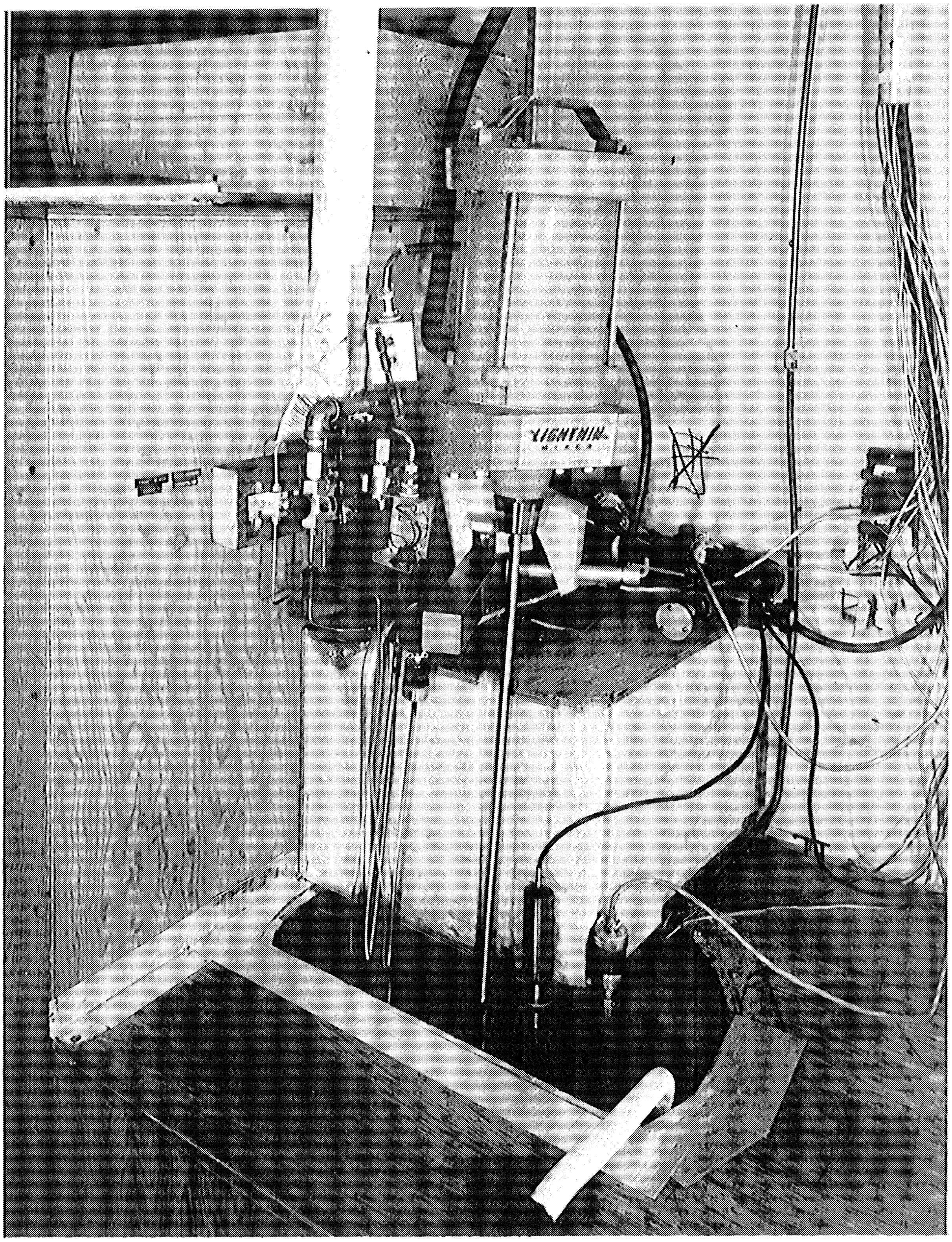


Figure 4. View of Calorimeter Bath Area

flowmeter is further isolated from the rest of the system by two micron filers. After leaving the flowmeter the fluid passes through several buffer tanks and is recycled back to the compressor to maintain a steady flow.

The heat exchanger bath and calorimeter bath are both contained in the same wooden structure. The baths were designed to operate at temperatures up to 300°F. Both are of double wall stainless steel construction in order to provide a vacuum jacket. They are each seated on 16 inch diameter transite pipe and are covered with a 1 inch layer of glass wool insulation. The wooden framework is filled with styrofoam insulation providing at least 5 inches around each bath. In addition when assembling several inches of urethane foam was formed between each bath and the styrofoam. This provides adequate insulation at bath temperatures in the range from -250 to about +275°F. Three bath fluids are used depending upon the bath temperature. Between -250 and -50°F isopentane is used, between -50 and +125°F kerosene and above +125°F industrial oil. To ensure adequate mixing at any condition a variable speed Lighnin' mixer is used for stirring.

With operation of much of the system at temperatures as high as 300°F it was found that propane or one of the trace components was decomposing and leaving a brownish black residue in the system. Since the flowmeter is the element in the system most affected by fouling, it was necessary to isolate that part from the rest of the system. In addition oil soluble in compressed propane could condense at the low flowmeter pressure. Thus the glass wool bomb was required right before the flowmeter bath. In case of backflow during startup and shutdown the low pressure buffer could act as a knockout for oil. In addition a sight glass was incorporated right before entry into the flowmeter bath so that oil could be visually detected if present. Finally the two filters, one before and one after the flowmeter, were required to eliminate solid matter which passed through the bombs.

After leaving the flowmeter buffer the fluid passes through the compressor inlet buffer located in the laboratory. When the revised system was first tested with the compressor located on the floor above the laboratory, the compressor was connected to this inlet buffer by a long 1/2" O.D. stainless steel piece of tubing. Operating under these conditions the compressor suction pressure would drop to a very low level during the intake stage, and the compressor discharge pressure would not build up. That is the compressor did not have an ample enough supply of fluid to accept during the intake stage for it to operate in a satisfactory fashion. The 1/2 inch line was, therefore, replaced with a 3/4" O.D. line and a small suction bomb was placed upstairs, located physically near the compressor. After these modifications the compressor discharge would build up significantly and the gas could be compressed at a reasonable rate, however, a regulating valve was located between this second buffer tank and the compressor in order to eliminate large oscillations of pressure in the flowmeter section.

A low vacuum in the calorimeters is obtained by use of a Hyvac 7 vacuum pump in series with an oil diffusion pump and a liquid nitrogen cold trap. The vacuum jacket of either or both calorimeters could be connected to the vacuum system by opening or closing two Veeco vacuum valves. Duplicate vacuum measurements can be made with both an ionization vacuum gauge and a McLeod gauge.

Pressure must be transmitted from the calorimeter to the pressure measuring devices. The 1/8 inch taps for each calorimeter come out of the calorimeter bath containing the test fluid. All four lines are connected to shutoff valves after which the two high pressure taps and the two low pressure taps are combined. The two lines then enter a heated conduit and above the baths the conduit and lines are teed. One heated conduit and one set of lines go to a Meriam high pressure mercury manometer. This manometer is located in a heated box and is used to measure the small pressure drops occuring in the isobaric calorimeter. The second conduit and the second set of lines lead to a heated valve manifold. The high pressure tap passes through the manifold and into the gas leg of a Ruska diaphragm pressure null detector. The low pressure line passes through a mercury knockout bomb located in the heated box and into one leg of a mercury U-leg. The leg extends below the box and is insulated along the part which normally contains the system fluid. The U-leg is used to transmit pressure from fluid to oil. The mercury in both legs is kept at the same level by addition or removal of oil. The mercury levels are detected by a series of iron electrical contact wires on both sides of the leg.

In addition, since the temperature of many parts of the system can be a critical factor in obtaining accurate data, it was necessary to install many thermocouples into the system. For example, when a fluid is throttled additional heat from a heating tape is generally required. There is normally one thermocouple located before the fluid is heated, one after the fluid is heated to ensure that the fluid temperature has not become too hot, and one after the fluid is throttled to ensure that the fluid is not too cold and therefore fractionating.

In such a system operating over a wide range of temperature and pressure leaks become a severe source of trouble. In order to minimize this problem heliarc welds were used to connect joints wherever possible. Valves, however, were connected into the system with standard high pressure connectors so that they could be readily removed in case of failure. As an example the valve panel was heliarced together to eliminate 38 trouble spots. A photograph of the valve panel is given in Figure 5. The photograph illustrates the high and low pressure manifolds which are heliarced together, numerous thermocouples and the neat valve arrangement. The various U-bends which are seen are wrapped with heating tapes and are located in the system right before a throttling operation. Figure 6 shows the valve panel with cover on and again illustrates the neat valve arrangement. Figure 7 shows an overall view of the control panel area of the laboratory.

The flowmeter calibration system is similar to that described by Jones[62] with several modifications necessary for running with propane and propane rich mixtures. The fluid to be collected is taken from one of several of the storage tanks and is throttled to flowmeter pressure in a special Tescom pressure regulator inside the heated valve panel box. The Tescom is a precision single stage regulator with a special viton diaphragm for withstanding the elevated temperature. Due to the temperature coefficient of pressure on the regulator and the severe effect of throttling propane near the critical point it is necessary to reach a steady state condition in the regulator before calibrating. This is accomplished by flowing through the regulator and flowmeter at approximately the desired flow rate for a short period before actually beginning the calibration. This produces a much more stable pressure and flow

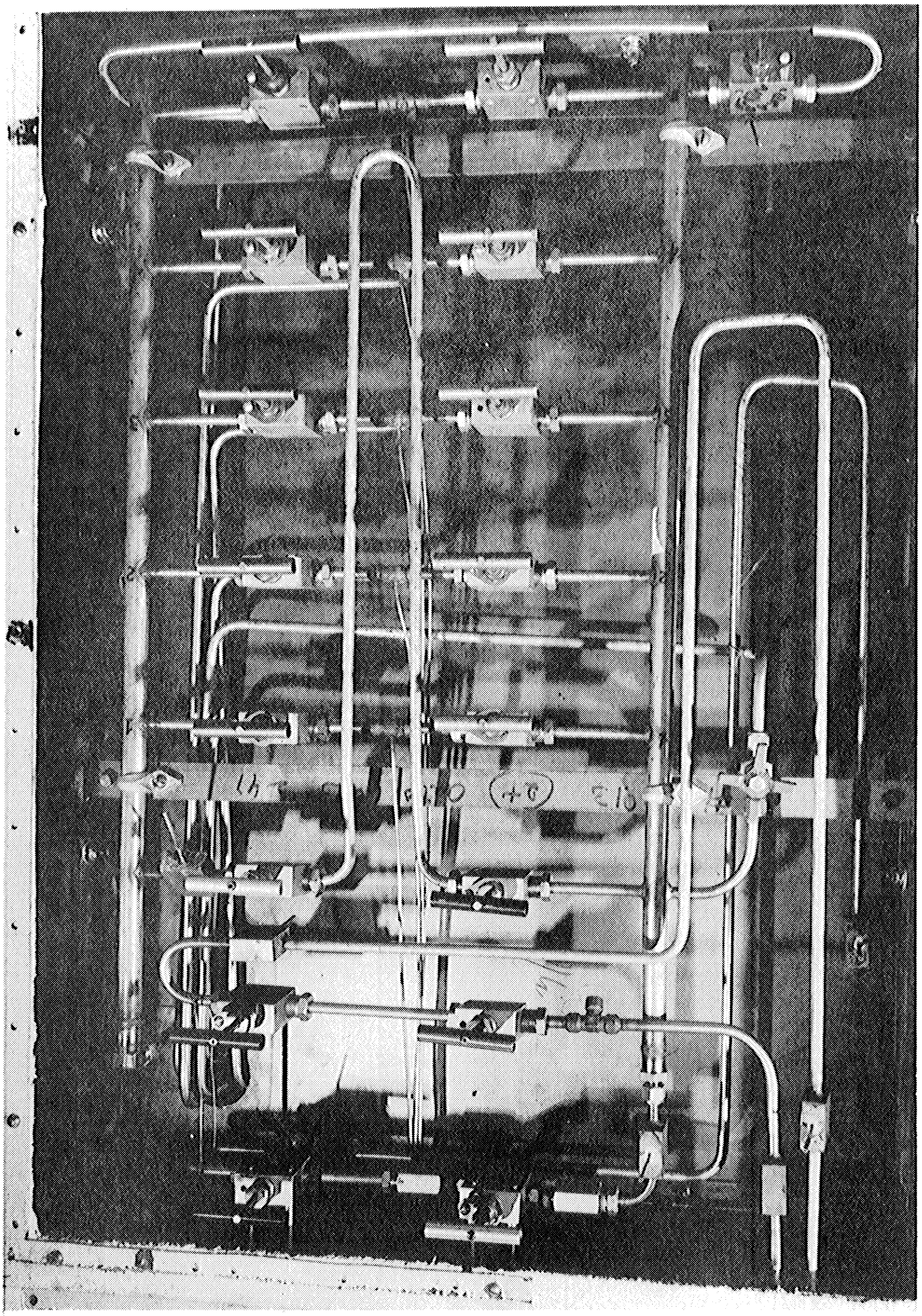


Figure 5. Valve Panel with the Front of Insulated Box Removed

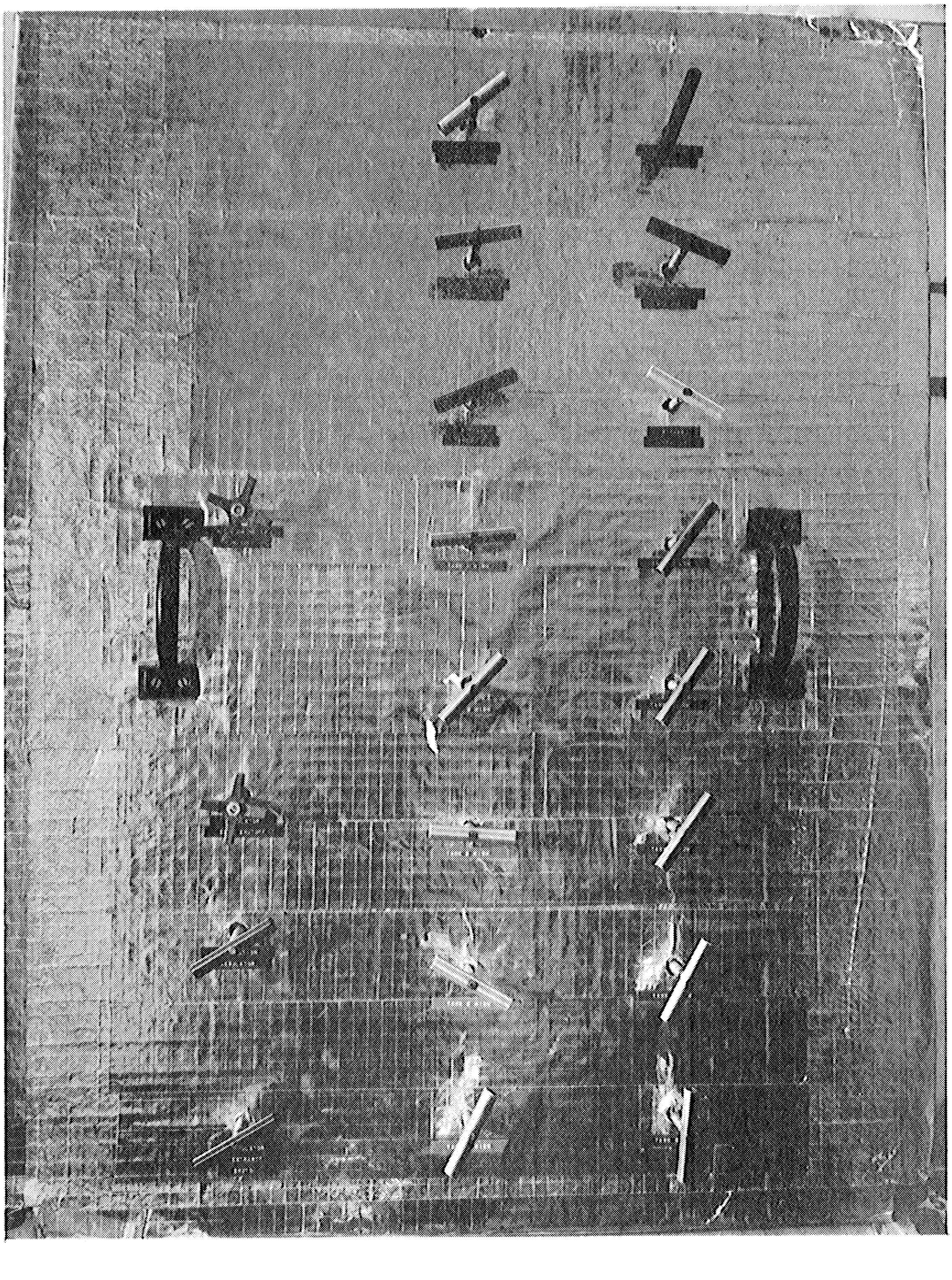


Figure 6. View of Insulated Valve Panel

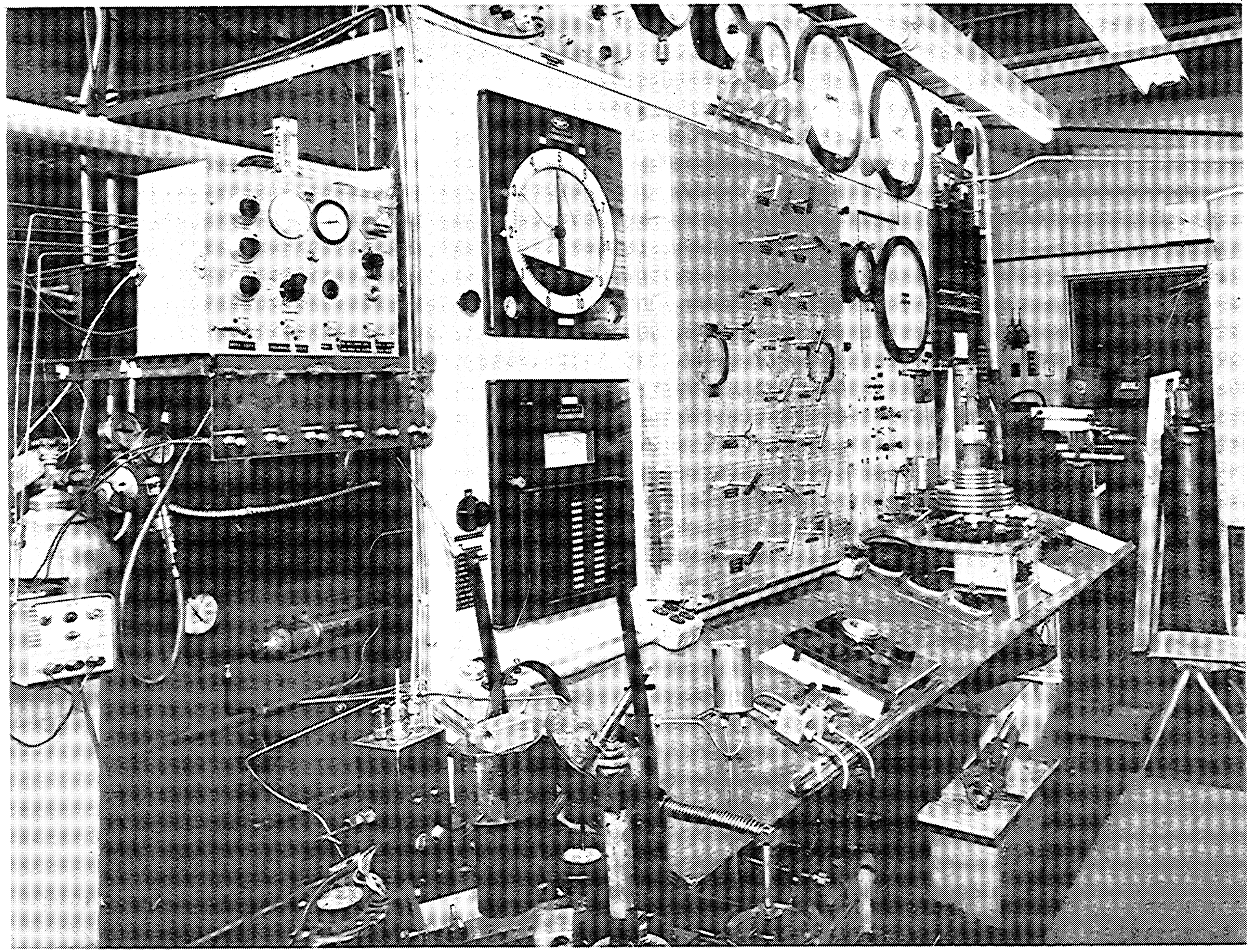


Figure 7. Control Area of Modified Flow System

rate of fluid through the flowmeter. The flow rate and pressure in the flowmeter are regulated by this pressure regulator, and the fluid passes through the meter and is collected in aluminum bombs as described by Jones.[62]

Calorimeters

The isobaric calorimeter used in this investigation is the one used by Faulkner[40] and Jones[62] with several minor modifications. The throttling calorimeter is that used by Mather[85] modified to allow for Joule-Thomson measurements in the compressed liquid region.

In the isobaric calorimeter new six-junction copper-constantan thermocouples replaced the ones used by Manker[80] and Mather.[85] The width of the inlet thermowell was increased to allow for removing and inserting the duplicate thermocouples more readily without damage. The lead wires to the nicrome heating wire in the calorimeter capsule was insulated with teflon to permit higher temperature operation without shorting. The previous vacuum electrical seals were replaced with Cervac vacuum electrical seals to facilitate removing the calorimeter and to prevent leakage. The vacuum line from the calorimeter was replaced with a one piece construction line to eliminate the possibility of leaks.

For the isothermal calorimeter the stainless steel O-ring and mylar sheets used for a high pressure seal were replaced with a teflon coated stainless steel O-ring. Several additional capillary coils were made for the calorimeter to enable measurement over a larger range of mixture compositions and temperatures at reasonable flow rates.

To operate the throttling calorimeter in the liquid region where throttling causes a temperature rise the radiation shield was wound with a nicrome heating wire in order to act as a guard heater. In addition a thermocouple was added between the guard heater and the outlet line to measure the temperature difference between the calorimeter body and the radiation shield.

Measuring Instruments

A detailed description of the measuring instruments has been given by Jones[62] as modified by Manker[80] and Mather.[85] Changes made from the above works and important features are listed below.

(1) The inlet temperature to the calorimeter is assumed to be equal to the temperature of the calorimeter bath which is measured using a platinum resistance thermometer. The calibration constants are given by Jones.[62] The thermometer was checked at the ice point and found to agree with the original calibration to ±0.1°C.

(2) The temperature rise in each calorimeter is measured by duplicate six-junction copper constantan thermocouples. These were calibrated at the oxygen and nitrogen points and compared with a platinum resistance thermometer at 20°C intervals by the National Bureau of Standards. These thermocouples are calibrated from -196 to +150°C and the calibration data are given in Tables XLII through XLV of Appendix A. The accuracy of the temperature rise measurement is about ±0.05 percent.

(3) The electrical energy input to the calorimeter is supplied by a regulated DC power supply. The energy input is measured by a K-3 potentiometer using standard resistors to scale the voltage to the range of the Potentiometer. This measurement circuit is described below. The accuracy of the electrical energy determination is ±0.05 percent.

(4) The mass flow rate of gas is determined by the measurement of pressure drop by a 10 inch precision water manometer across a Meriam

laminar flow element together with the temperature and pressure of the element. These data are used to obtain the mass flow rate, F, from the equation

$$\frac{\rho \Delta P}{F\mu'} = B + A\left(\frac{F}{\mu'}\right) + C\left(\frac{F}{\mu'}\right)^2 + D\left(\frac{F}{\mu'}\right)^3 \tag{76}$$

The calibration constants A, B, C, D are obtained by a least squares fit of the calibration data. As indicated by Equation (76) small density, ρ, and viscosity, μ', corrections must be applied for minor variations in flowmeter conditions. In this investigation the results of Giddings[44,45] were used to correct for viscosity and the density data of Reamer, Sage, and Lacey[111] and the BWR equation used for the density correction. The accuracy of the mass flow rate determination is about ±0.2 percent.

(5) The pressure at the inlet of the calorimeter is measured with a calibrated Mansfield and Green dead weight gauge. The calibration data for this gauge are given by Mather,[85] and it is accurate to 0.03 percent. The pressure is checked with a calibrated Heise gauge during measurements. The accuracy of the gauge is 1 percent of full scale. In addition the dead weight gauge is occasionally checked with the differential dead weight gauge of Roebuck.[116] The pressure is transmitted from the calorimeter fluid to the oil in the pressure measurement system by means of a Ruska differential pressure indicator.

(6) The pressure drop across the isobaric calorimeter is measured with a 40 inch high pressure mercury manometer. The manometer is located in a special heated box to prevent condensation and fractionation within it. The pressure measurement is accurate to ± 0.1 inch of mercury pressure drop.

(7) The pressure drop across the throttling calorimeter is measured

over the differential dead weight balance due to Roebuck[116] and modified by Mather.[85] The inlet pressure, as already mentioned, is transmitted to the oil in the pressure measurement system by means of a Ruska differential pressure indicator. The outlet pressure is transmitted by a 36 inch mercury U-leg. The level of mercury in the U-leg is sensed by iron electrical probes sealed into each leg.

It was found that the Viton O-ring dynamic pressure seal used by Mather[85] for the Roebuck differential pressure balance would become defective after a short period of time. The Viton was extruded by the motion of the piston and eventually the seal would blow and oil would burst out into the laboratory. The seal was replaced with a teflon O-ring and backup ring set-up which worked satisfactorily. The sensitivity of the Roebuck pressure balance at higher pressures with the new seal is about 1 psia. This limits the lower pressure drop measurements to an accuracy of 1 percent.

Electrical Measurement

The electrical measurement system for the isobaric calorimeters has been described by Jones.[62] The electrical system for the isothermal calorimeter is essentially the same as that used for the isobaric calorimeter. In fact both calorimeters have been connected to the same measuring system where a series of switches can connect either one or the other. The fact that the isothermal calorimeter heater is grounded forced some modifications to prevent ground loops.

Measurements of power input, temperature, and temperature difference are obtained from voltages recorded with a K-3 potentiometer. The power measurement circuit used to eliminate the possibility of more than one ground is shown in Figure 8.

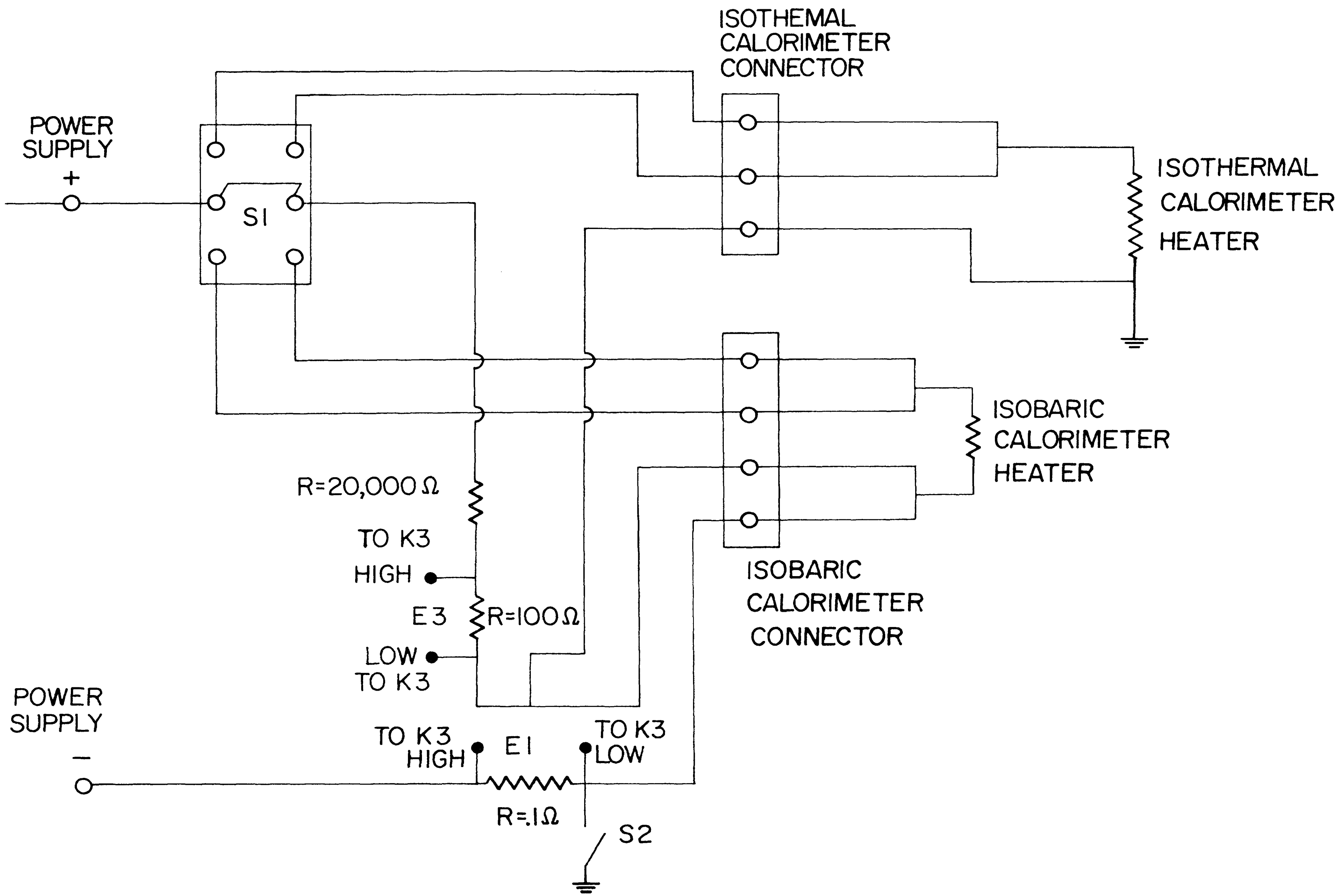


Figure 8. Wiring Diagram of Power Measurement Circuit

When operating the isobaric calorimeter (the one used most frequently) double pole double throw switch S1 is connected to the positive side of the measurement and power leads of the isobaric calorimeter. Each calorimeter heater can be connected from the electrical system to the heater leads in the vacuum jacket by an electrical vacuum connector. During operation of the isobaric calorimeter the plug for the isobaric calorimeter is connected and the plug for the throttling calorimeter must be disconnected. Switch S2, which can be grounded, is open. Readings are then made for current (related to E1) and voltage (related to E3). At the same time readings are made of the voltages across the thermocouples and the current through and voltage across the platinum resistance thermometer (see Jones[62]). The low side of the K-3 potentiometer is the only point of the circuit that is grounded.

When the isothermal calorimeter is in operation S1 is connected to the positive side of the isothermal calorimeter heater. The isothermal calorimeter connector is connected and the isobaric vacuum connector may or may not be connected. Switch S2 is closed and the current through the heating wire passes through ground. When reading E1 and E3, the ground on the low side of the K-3 potentiometer must be disconnected in order to prevent a ground loop. Such a loop creates an error of as large as 20 percent in the power measurement. When making the other readings with the potentiometer the ground is again connected.

Procedure

In obtaining data, inlet conditions of temperature and pressure are established and flow rate and power input are adjusted to desired values. Readings are taken and adjustments made as necessary until the condition of steady state operation is obtained and maintained for at least 15 minutes. A single determination generally lasts from one to two hours depending on the magnitude of the changes made between determinations.

SECTION IV - THE ENTHALPY OF PROPANE UNDER PRESSURE

After modifying the recycle flow facility the first system investigated was propane for several reasons. First of all it would permit testing the facility with the heaviest system that could possible be investigated. Thus, if propane could be investigated any system with a lower critical temperature could also. Secondly, it would allow checks of the results obtained by the new system with data in the literature. Finally, data could be obtained for propane in regions where at present they are nonexistant. Thus, the final result would be to obtain an accurate knowledge of the enthalpy of propane over a wide region of temperature and pressure. The data obtained would hopefully be used as a standard for investigations in the future.

Regions of Measurements

The range of experimental determinations is indicated on a PT diagram in Figure 9. In the single phase region isobaric measurements were usually made in groups of four runs each having the same inlet temperature rises of approximately 10, 20, 40, and 80°F. In regions where C_p varied significantly with temperature (such as near the critical point) much smaller temperature rises were investigated (as small as 1°F). Isobaric determinations across the two-phase region included determinations of the heat capacities of both liquid and vapor at temperatures respectively below and above the saturation temperatures. These runs are indicated by asterisks on Figure 9. The number of isobars was reduced at low temperatures because it was found that C_p did not vary appreciably with pressure in this region and values of C_p are available below

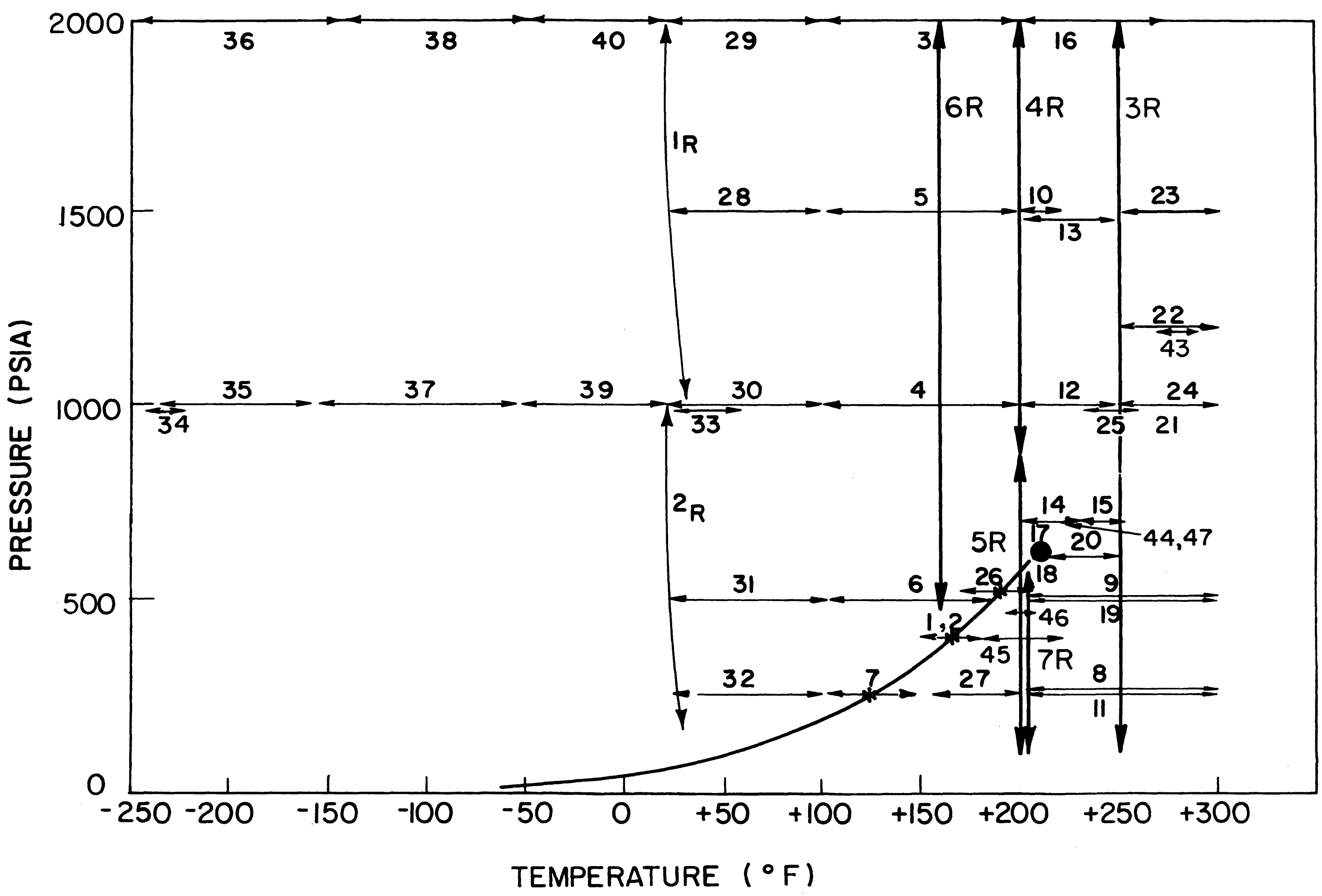


Figure 9. Temperatures and Pressures of Measurement for Propane

atmospheric pressure for the saturated liquid.[66]

Isothermal determinations were made mainly in the single-phase region. One isothermal enthalpy change on vaporization was obtained. Pressure drops between 100 and 500 psia were used.

Isenthalpic determinations were made at conditions of constant inlet temperature for different inlet pressures. In general, pressure drops of approximately 300, 600, and 900-1000 psia were used at two different inlet pressures.

The isobaric results for individual runs in the single- and two-phase region are given in Table XLVI of Appendix B. The basic isothermal and Joule-Thomson data are given in Tables XLVII and XLVIII of Appendix B, respectively.

Composition of Gas

Phillips Instrument Grade propane was used. This material contained approximately 1/4 percent impurities as determined by mass spectrometer analysis and reported in Table I.

TABLE I

IMPURITY CONTENT OF PROPANE

	Mass Spectrometer
Nitrogen	0.15%
Oxygen	0.04%
Methane	0.02%
Ethane	0.03%
Propylene	0.01%
Propane (by difference)	99.75%
	100.00%

Flowmeter Calibrations

Seven sets of flowmeter calibration runs (usually ten determinations to a run) were made during the course of this investigation of propane. In contrast to previous experience with other pure components and with mixtures[5,9,11,13] the calibration function was decidedly nonlinear with marked curvature in the middle of the operating range as shown in Figure 10. Therefore, one set of constants in Equation (76) was used to represent the calibration data at low-flow rates (0.1 to 0.25 lb/min) and another for higher flows (0.25 to 0.4 lb/min).

Data from the calibration runs were correlated in three groups. The results of the first two runs were in excellent agreement. These were fit with the lower curve in Figure 10. A third set showed deviations of as much as 1 percent from this pair at high flow rates (the upper curve on Figure 10). After oil was found in the flowmeter, the glass wool bomb described in the previous section was installed. The flowmeter was cleaned ultrasonically, and subsequently, four sets of calibration runs were made which yielded results in excellent agreement. These sets are fit by the middle line in Figure 10.

A single correlating equation represents the data from these four runs in the high flow region with an average deviation of ±0.17 percent. Another equation serves to correlate the data at low flows with an average deviation of ±0.22 percent. In processing the data some improvement in precision was obtained by using correlating equations based on the two sets of calibration data obtained preceding and following a run when the calibration did not change. When the calibration changed, repeat runs were made to determine which calibration equation would most adequately represent the flow rate at the time of an experimental run.

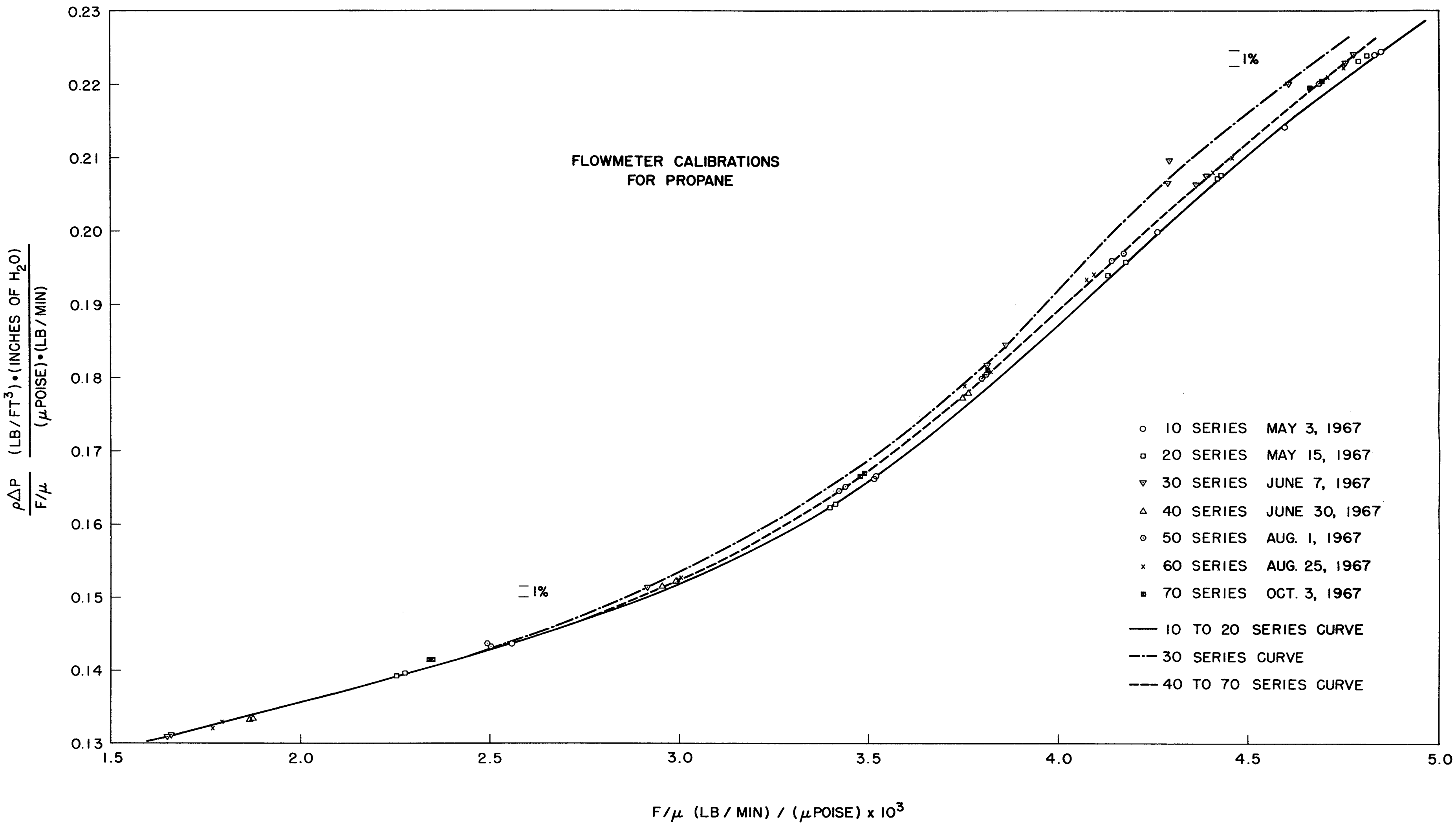


Figure 10. Results of Flowmeter Calibrations for Propane

TABLE II

Calibration Data Used in
Interpreting Experimental Results

Experimental Runs	Calibration Runs	Number of Calibration Points	Average Deviation
1-7	10,20	High: 14 Low: 7	0.14% 0.01%
8-10	30	High: 6 Low: 6	0.10% 0.06%
11-20	40,50	High: 15 Low: 12	0.14% 0.14%
21-31	50,60	High: 15 Low: 12	0.07% 0.11%
32-47 1R-7R	60,70	High: 16 Low: 12	0.17% 0.17%

TABLE III

Illustration of Consistency of Calibration
Equations for High and Low Flow Rates
at Intermediate Flow Rates

Run	Flow Rate lb/min		Mean Heat Capacity Btu/lb °F	
	Calibration Equation Used			
	Low	High	Low	High
20.010	0.29304	0.29276	1.4773	1.4787
20.020	0.29367	0.29336	1.2323	1.2335
20.030	0.29278	0.29251	0.9889	0.9898

A summary of the calibration equations which were used to interpret specefic data is presented in Table II, together with values of the total number of points included in the calibration sets and the standard deviation of each set from the calibration equation. The high and low flow rate correlating equations appear to give reasonable results in the region of overlap. For example, Run 20 was obtained in the middle of the flow rate region and both correlating equations were used to calculate the flow rate. Table III shows the small effect (0.1 percent) of calculating flow rates by either equation for this run. All of the sets of calibration constants of Equation (76) used to interpret experimental data are presented in Table XLIX of Appendix B.

Results

Enthalpy Change on Vaporization

Isobaric determinations were made across the two-phase region at 250, 400, and 500 psia as indicated by asterisks on the vapor pressure curve of Figure 9. Results from the run at 500 psia are presented on Figure 11. Note that the transition is not isothermal because of the presence of impurities. (See Table I.) This factor was taken into account by extending the horizontal portion of the curve to intersect with an extension of the liquid phase curve to determine the equivalent initiation of vaporization. Estimation of the point of complete vaporization was complicated by the fact that some scattering of points in the vapor region resulted bacause relatively high rates of electrical energy input were required to span the two-phase region and as a result very small changes in the flow rate caused very large changes in temperature. The data in the vapor region (difference values) were used together

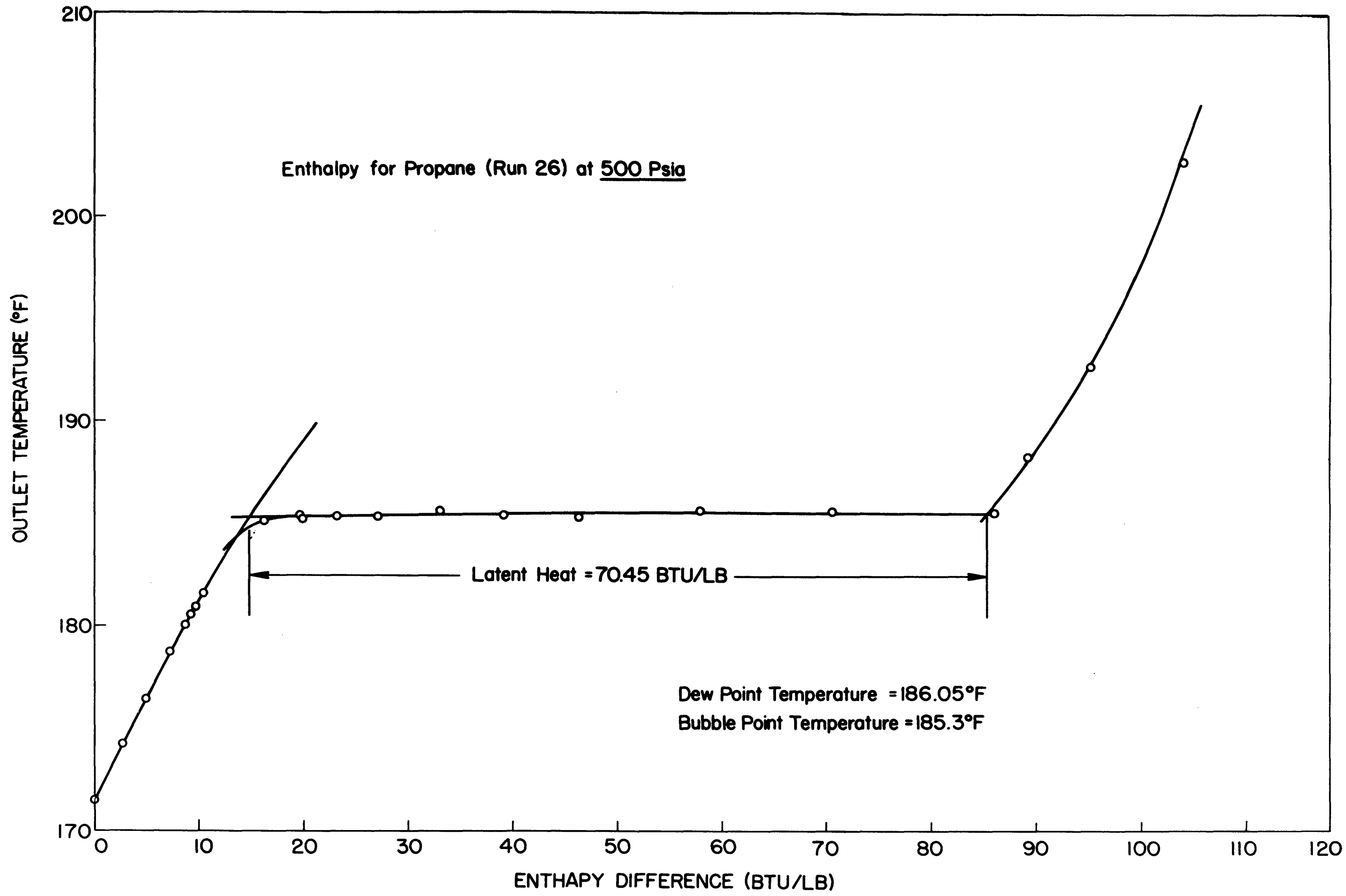


Figure 11. Enthalpy Differences for Propane in the Two-Phase Region

with other runs made entirely in the vapor region to estimate $C_p = f(T)$ for the vapor near the saturation line. A linear relation served to represent the data and upon integration yielded the curve drawn on Figure 11. The resulting value for the latent heat of vaporization is listed in Table IV together with experimental values at 250 and 400 psia.

TABLE IV

EXPERIMENTAL VALUES OF LATENT HEATS OF VAPORIZATION OF PROPANE

	Btu/lb				
		Other Investigators			
Pressure psia	This Work	Helgeson & Sage (51)	Kuloor et al. (69)	Sage, Evans, & Lacey (124)	Dana et al. (26)
250	122.60	122.35[b]	123.82[a]	124.25[b]	127.87[c]
400	93.07	94.5[b]	93.20[a]	98.0[b]	---
500	70.45	71.7[c]	67.56[a]	80.5[c]	---
588	44.9^{+1}_{-2}	42.3[d]	35.63[a]	---	---

a Calculated from equation fit to experimental data published prior to 1965.
b Interpolated based on plot of experimental values.
c Extrapolation based on vapor pressure data.
d Calculated from equation given by authors.

Isobaric Data

Typical isobaric data in the single phase region are presented in Figure 12. Mean values of C_p determined both from the direct measurements (solid horizontal lines) and by differencing experimental data (dashed horizontal lines) are plotted versus temperature.

A curve is constructed to determine point values of $C_p = f(T)$ as indicated by the solid curved line. In regions where the heat capacity does not vary greatly with temperature Equations (18) and (23) are used to obtain the point values of heat capacity and enthalpy with the aid

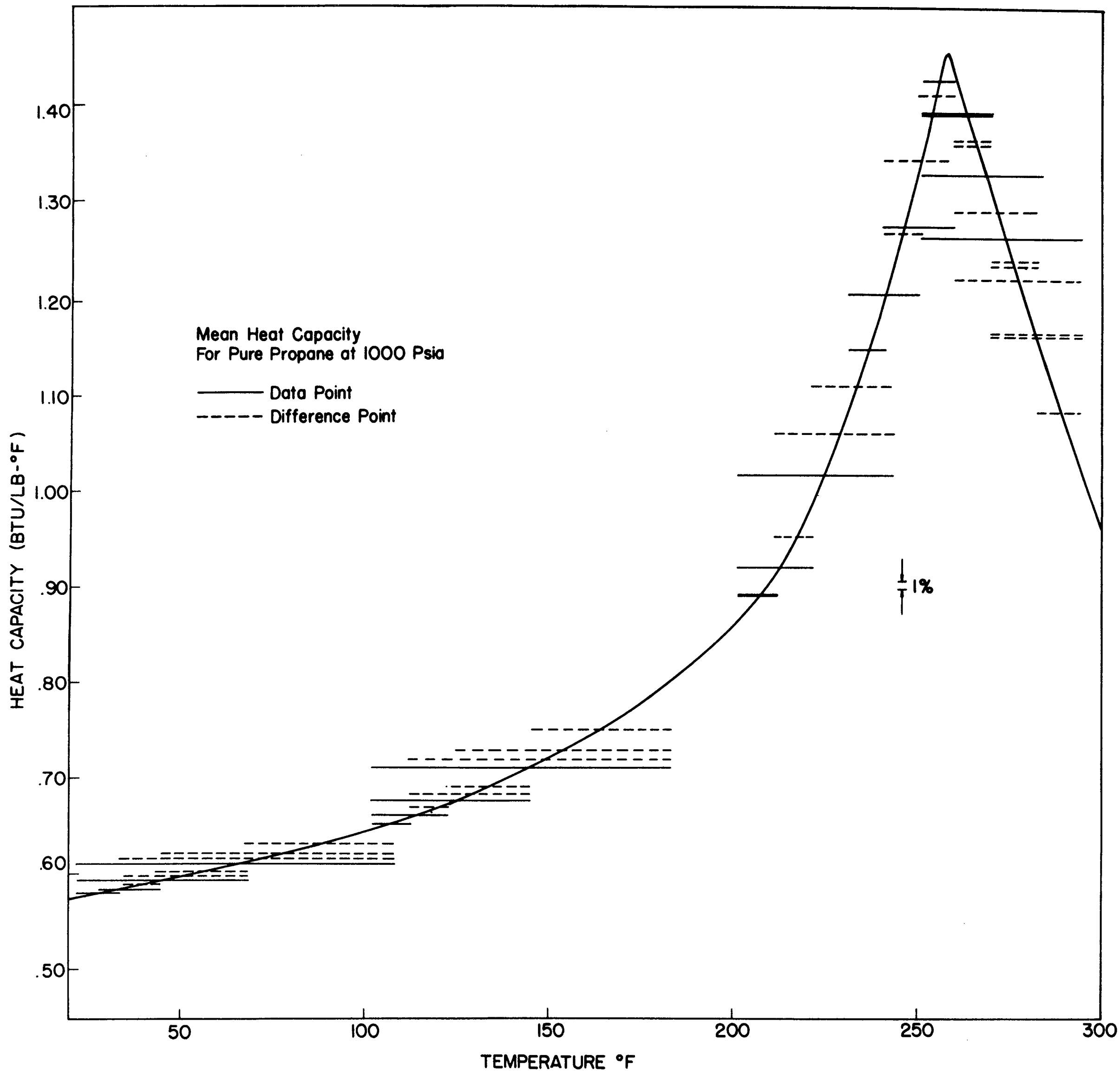


Figure 12. Isobaric Heat Capacity for Propane at 1000 psia in the Upper Temperature Range

of a digital computer. The smooth curve is obtained by fitting the calculated point values of the heat capacity. Where C_p is a strong function of temperature the fit of Equation (18) and (23) of eight data points becomes quite poor. In this region the graphical technique using Equation (10) is used. Also extrapolation to temperatures, above and below the region of data, is done graphically.

A distinct maximum in C_p of 1.47 Btu/lb-°F was located at 257°F for this isobar. Data at lower temperatures showed less dependence of temperature on heat capacity (see Table VI). Maximum values of C_p were determined at several pressures above the critical. The results are summarized in Table V.

TABLE V

MAXIMUM VALUES OF C_p FOR PROPANE

Pressure psia	Temperature at Maximum, °F	Maximum Value of C_p Btu/lb-°F
617	206	$>$ 60
700	219	5.4
1000	257	1.47
1200	284	1.18

An isobaric run was made at the critical pressure, 617 psia.[69] The results are presented in Figure 13. A very large value of C_p ($>$ 60 Btu/lb-°F) was determined. From these data, it was determined that the maximum value of C_p occured at a temperature of 206.3 ±0.3°F. This is in good agreement with the accepted value of 206.3°F as the critical temperature.[69]

As can be seen by Figure 14, there is a strong variation in the heat capacity with respect to temperature at pressures below the critical

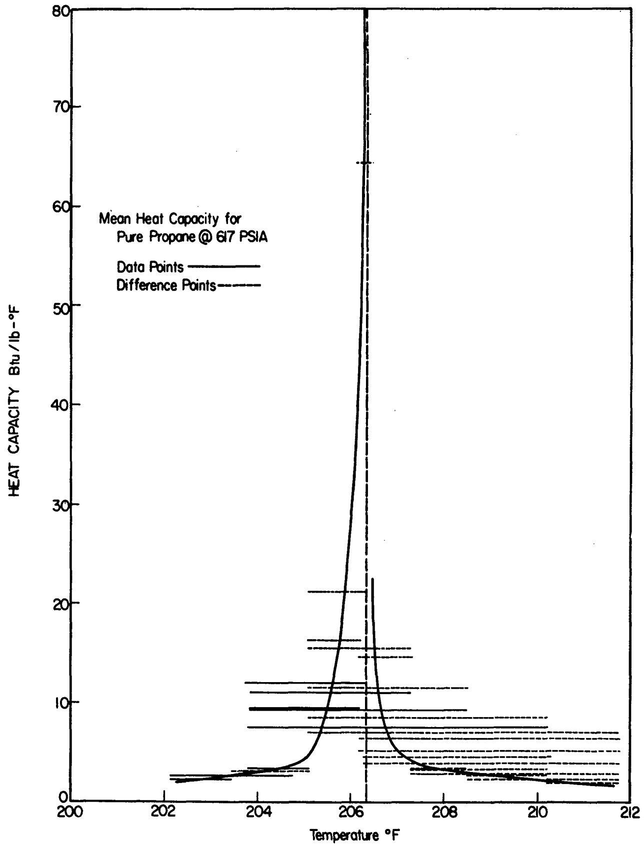


Figure 13. Isobaric Heat Capacity in the Critical Region at the Critical Pressure for Propane

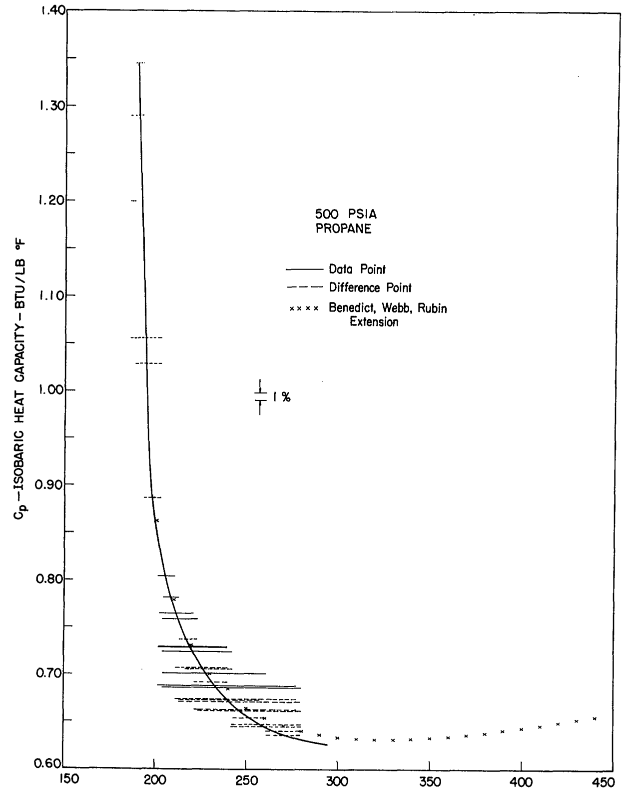


Figure 14. Isobaric Heat Capacity at 500 psia in the Gaseous Region for Propane.

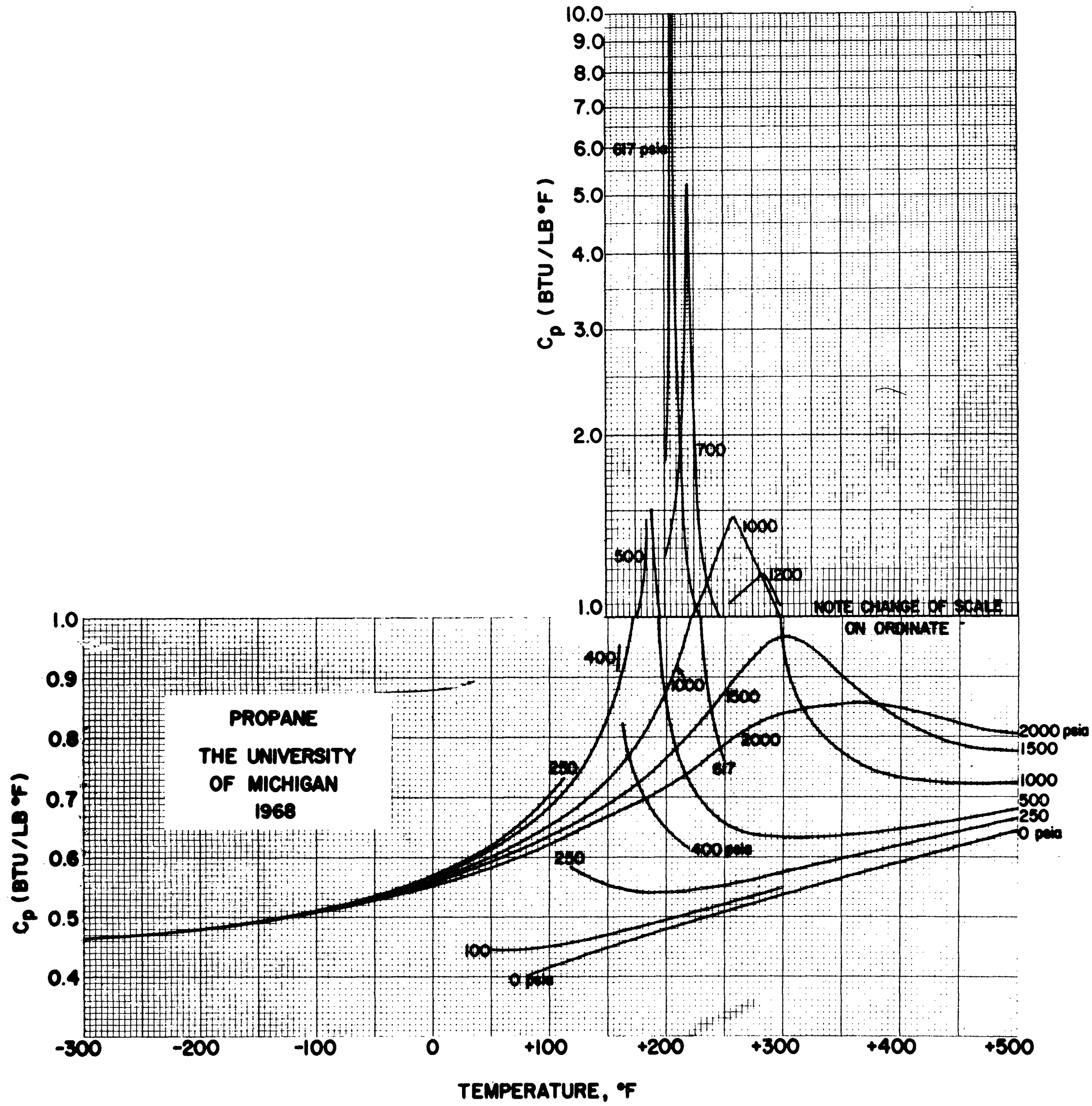


Figure 15. Isobaric Heat Capacity for Propane

also. At 500 psia the heat capacity of the liquid as well as the gas changes by a factor of 2 in the region near the saturation point.

A table of C_p values is presented in Table VI. Interpolated values are indicated underlined. Additional experimental values of the isobaric heat capacity, C_p, determined from data obtained in regions of rapid change with respect to temperature are presented in Table VII. Figure 15 summarizes all of the experimental values of isobaric heat capacity for propane.

Isothermal Data

Typical isothermal data are presented in Figure 16. Average values of $\varphi_m = (\Delta H/\Delta P)_T$ are plotted as horizontal lines and the graphical equal area method was used to determine point values of $\varphi = f(P)$ as illustrated by the solid curve. This was necessary because of the large variations in φ with pressure. In extrapolating the results of the experimental investigation to zero pressure it was necessary to use Equation (32) for φ^0 in terms of the second virial coefficient. Values of φ^0 were calculated from the virial coefficients given by Diaz-Pena and Cervena,[30,31] and Huff and Reed.[58] Both results are plotted on Figure 16. In addition the value of φ^0 obtained from the BWR equation of state[7] using the original constants for propane is also plotted. There are significant differences in the results (±7 percent) and heavy reliance was placed on the value of Diaz-Pena since it was the intermediate value.

A table of φ values for all of the experimental isotherms is presented as Table VIII. These results are summarized in Figure 17. One isothermal run was made through the two-phase region at 201°F (see Figure 22). The estimate of the heat of vaporization at this temperature (588 psia) is listed in Table IV.

Table VI

Experimental Values of Isobaric Heat Capacity, C_p, for Propane (Btu/lb - °F)

(Pressure psia)

Temperature °F	p[a]	250	500	1000	1500	2000
-250	0.4651	0.465	0.466	0.466	0.467	0.467
-225	0.4697	0.470	0.470	0.470	0.471	0.472
-200	0.4751	0.475	0.475	0.475	0.476	0.477
-175	0.4812	0.480	0.480	0.480	0.481	0.483
-150	0.4878	0.487	0.486	0.485	0.487	0.489
-125	0.4962	0.495	0.493	0.492	0.493	0.494
-100	0.5060	0.505	0.503	0.502	0.501	0.501
-75	0.5173	0.517	0.516	0.514	0.513	0.511
-50	0.5307	0.530	0.529	0.526	0.524	0.522
-25		0.546	0.543	0.540	0.537	0.535
0		0.564	0.561	0.557	0.553	0.549
+25		0.585	0.583	0.576	0.570	0.568
+50		0.611	0.608	0.597	0.588	0.581
+75		0.645	0.637	0.619	0.607	0.597
+100		0.696	0.673	0.644	0.631	0.617
+125	(p°)[b]	0.755(l)	0.754	0.677	0.657	0.640
+150	0.4434	0.552(g)	0.815	0.722	0.684	0.664
+175	0.4590	0.542	0.987(l)	0.780	0.715	0.687
+200	0.4744	0.536	0.854(g)	0.863	0.755	0.713
+225	0.4898	0.540	0.711	1.026	0.807	0.743
+250	0.5050	0.553	0.662	1.45	0.867	0.792
+275	0.5199	0.569	0.634	1.250	0.928	---
+300	0.5343	---	0.618	0.956	0.963	---

a Values for saturated liquid at $p < 15$ psia from Kemp and Egan (66).

b Values for ideal gas at zero pressure (118).

Table VII

Experimental Values of Isobaric Heat Capacity, C_P,
Near the Saturation Curve and in the Vicinity of $C_P(T)$ Maxima

Pressure (psia)										
	Temperature (°F)									
	100	105	110	115	119.9 (l)	122.4 (g)	125	130	135	140
250	C_P (Btu/lb°F)									
	0.696	.7047	.7159	.7290	.7424	.5794	.5756	.5684	.5629	.5586

Pressure (psia)										
	Temperature (°F)									
	160	161.1 (l)	163.5 (g)	165	170	175	180	190	200	220
400	C_P (Btu/lb°F)									
	.912	.943	.8214	.8078	.7693	.7399	.7157	.6769	.6478	.6130

Pressure (psia)										
	Temperature (°F)									
	165	170	175	180	185.3 (l)	186.0 (g)	190	195	200	205
500	C_P (Btu/lb°F)									
	.8970	.9347	.9889	1.11		(1.54)	1.32	.985	.8543	.8069

Pressure (psia)										
	Temperature (°F)									
	202	203	204	205	206	206.3 (c.p.)	207	208	209	210
617	C_P (Btu/lb°F)									
	1.84	2.37	3.09	4.3	29	-	6.5	4.0	2.95	2.35

Pressure (psia)										
	Temperature (°F)									
	212	214	217	220	225	230	235	240	245	250
617	C_P (Btu/lb°F)									
	1.88	1.68	1.30	1.17	1.06	0.97	0.90	0.84	0.79	0.76

Table VII (continued)

Pressure (psia)										
	Temperature (°F)									
	200	207	213	217	219	220	225	230	240	250
700	C_P (Btu/lb°F)									
	1.265	1.410	2.150	3.795	5.280	3.910	2.018	1.351	1.061	.986
	Temperature (°F)									
	220	235	245	250	255	257	260	265	270	280
1000	C_P (Btu/lb°F)									
	.980	1.134	1.272	1.345	1.436	1.466	1.439	1.377	1.314	1.189
	Temperature (°F)									
	255	260	270	275	280	284	285	290	295	300
1200	C_P (Btu/lb°F)									
	1.050	1.077	1.121	1.139	1.161	1.177	1.172	1.118	1.070	
	Temperature (°F)									
	220	230	240	250	260	270	280	290	295	300
1500	C_P (Btu/lb°F)									
	.796	.819	.843	.867	.891	.914	.938	.956	.961	.963

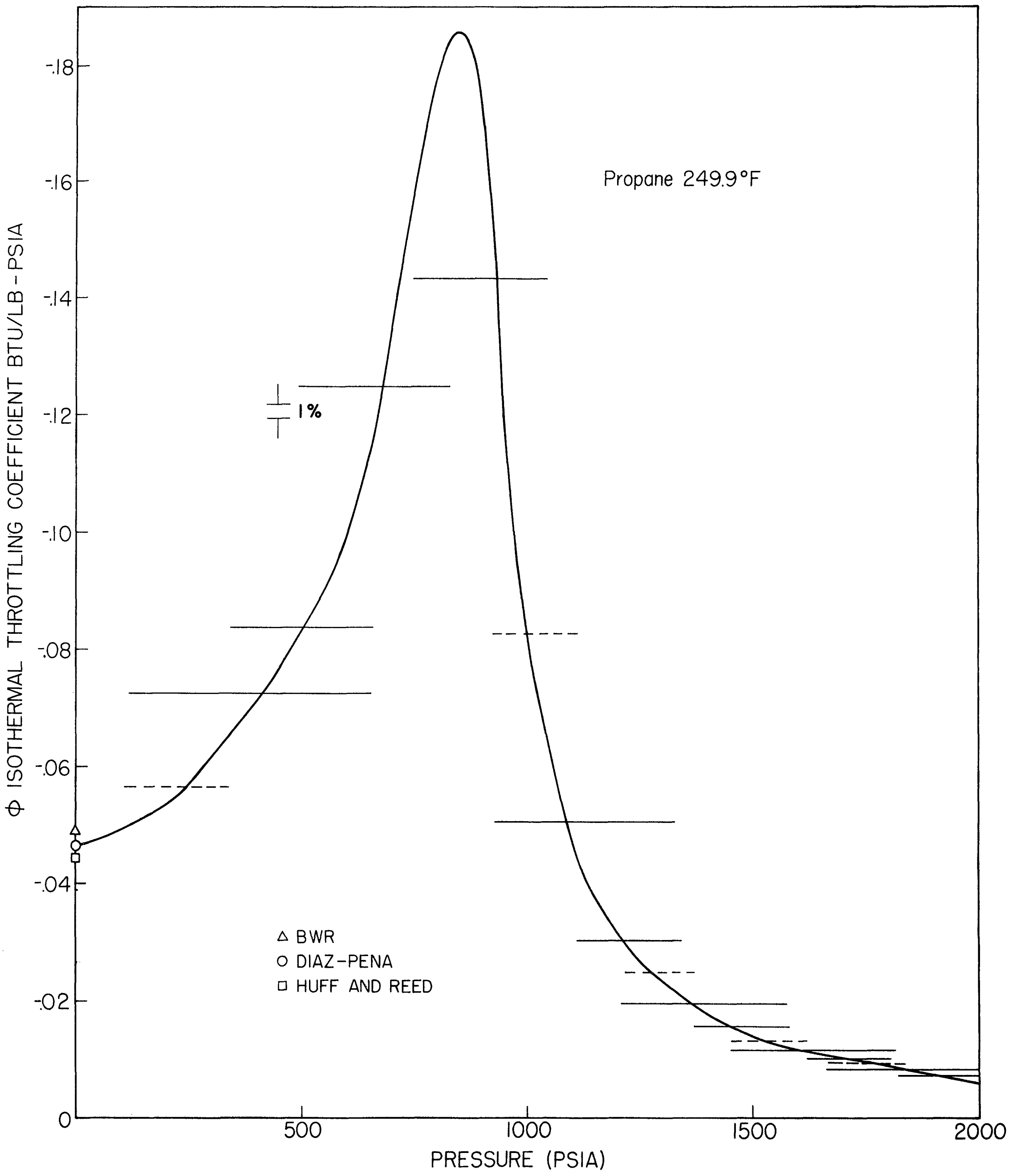


Figure 16. Isothermal Throttling Coefficient for Propane Above the Critical Temperature.

Table VIII

Experimental Values of the
Isothermal Throttling Coefficient for Propane
$\phi \times 10^2$ (Btu/lb-psia)

Pressure psia	Temperature - °F: 21.2[a]	160.5	201.0	249.9
2000	+0.253	-0.009	-0.207	-0.57
1800	+0.248	-0.059	-0.304	-0.87
1600	+0.244	-0.110	-0.410	-1.16
1400	+0.239	-0.167	-0.567	-1.79
1200	+0.235	-0.239	-0.843	-3.18
1000	+0.230	-0.350	-1.292	-8.28
800	+0.226	-0.518	-2.198 (l)	-17.69
600	+0.221	-0.802		-9.90
400	+0.217		-10.41 (g)	-7.09
200			-6.58	-5.70
0[b]			-4.7	-3.84

a Estimated from Figure 18

b Based on $\phi^\circ = B - T(dB/dT)$

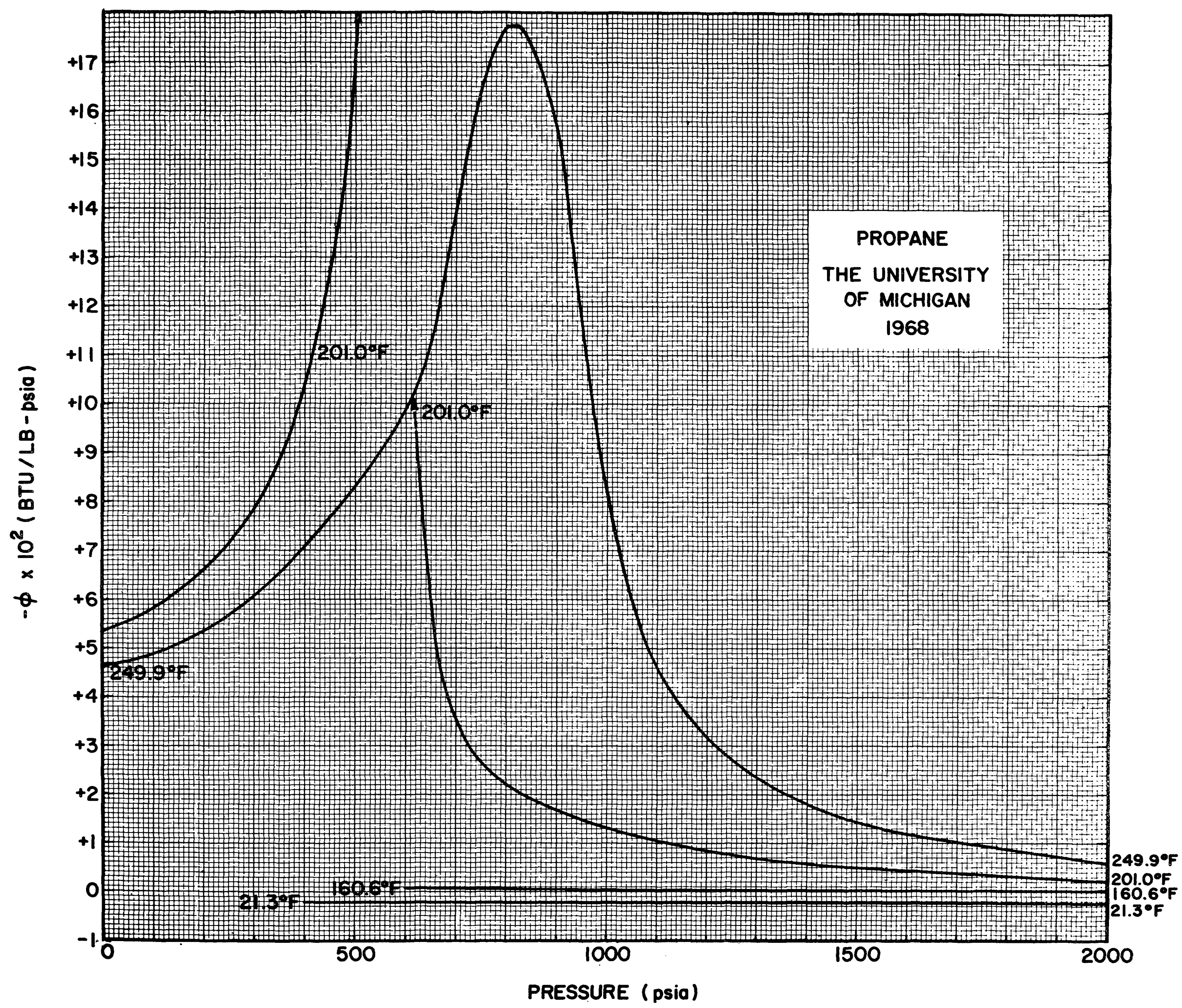


Figure 17. Isothermal Throttling Coefficients for Propane

Isenthalpic Data

Isenthalpic determinations were made at an inlet temperature of 21.2°F and inlet pressures of 2000 psia and 1100 psia. The basic results are presented on Figure 18. Values of $\mu \equiv (\partial T/\partial P)_{\underline{H}}$ estimated from a plot of $(\Delta T/\Delta P)_{\underline{H}}$ versus P are presented in Table IX. The isenthalpic data were used together with C_{p_m} data in Equation (17) to generate an isotherm at 21.2°F. The results of these calculations are summarized on Figure 19. Values of μ estimated from these data are included in Table IX.

TABLE IX

EXPERIMENTAL VALUES OF THE JOULE-THOMSON COEFFICIENT OF PROPANE

Pressure psia	Temperature °F	$-\mu \times 10^3$ °F/psi
2000	21.20	4.61
1700	22.50	4.49
1300	24.23	4.12
1000	25.50	3.99
1000	21.58	3.99
700	22.74	3.71
300	24.15	3.27
150	24.61	3.08

Analysis and Comparison of Results

Isobaric Data

Values of $\Delta \underline{H}_V$ at 250, 400, and 500 psia and at 201°F (588 psia) which have been estimated from other published data are also presented in Table IV. Results from the present investigation are in reasonable agreement with values reported in 1967 by Helgeson and Sage.[51] The discrepancies between the experimental values obtained using the isobaric

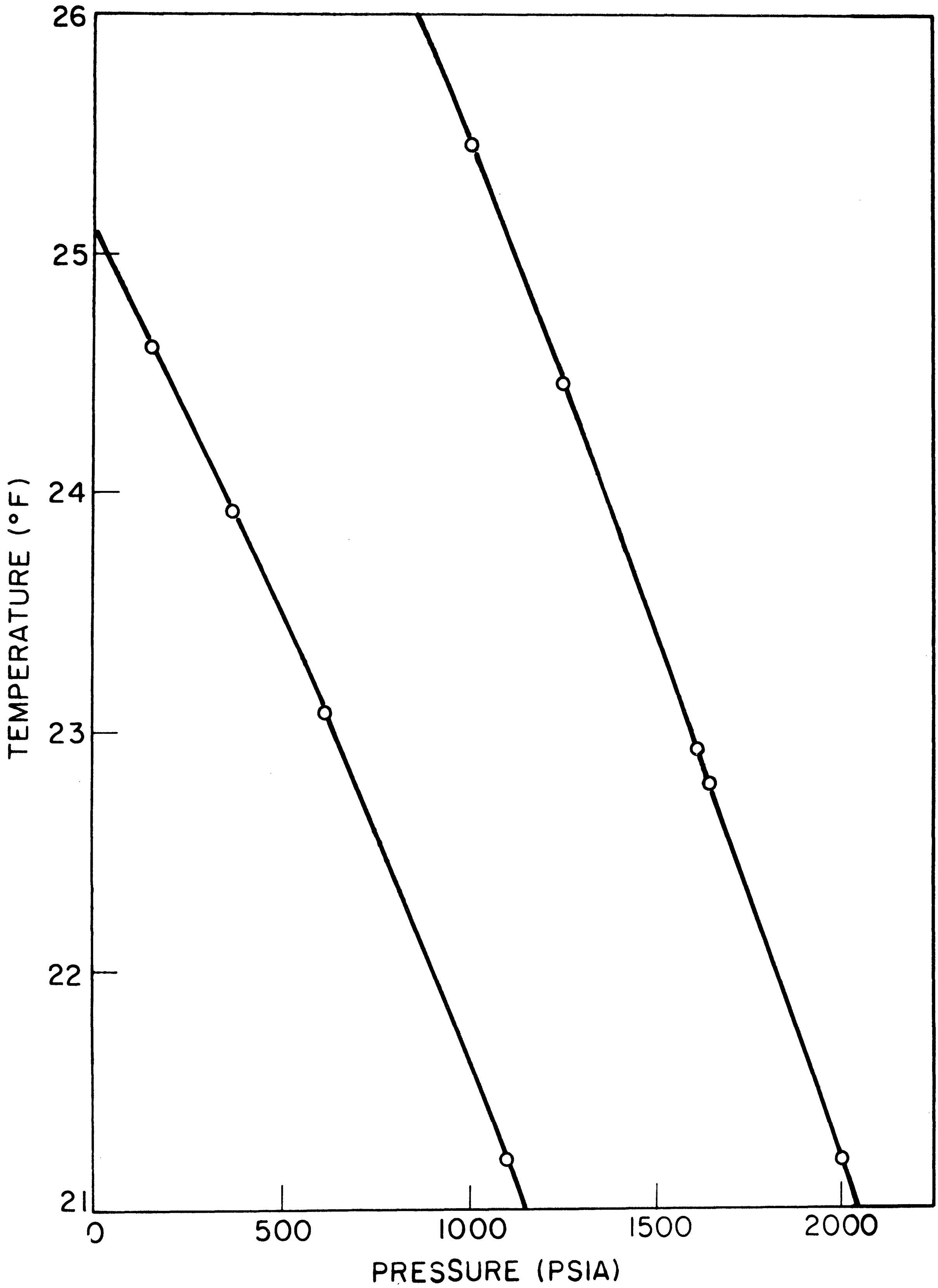


Figure 18. Isenthalpic Curves for Propane with a 21.2°F Initial Temperature

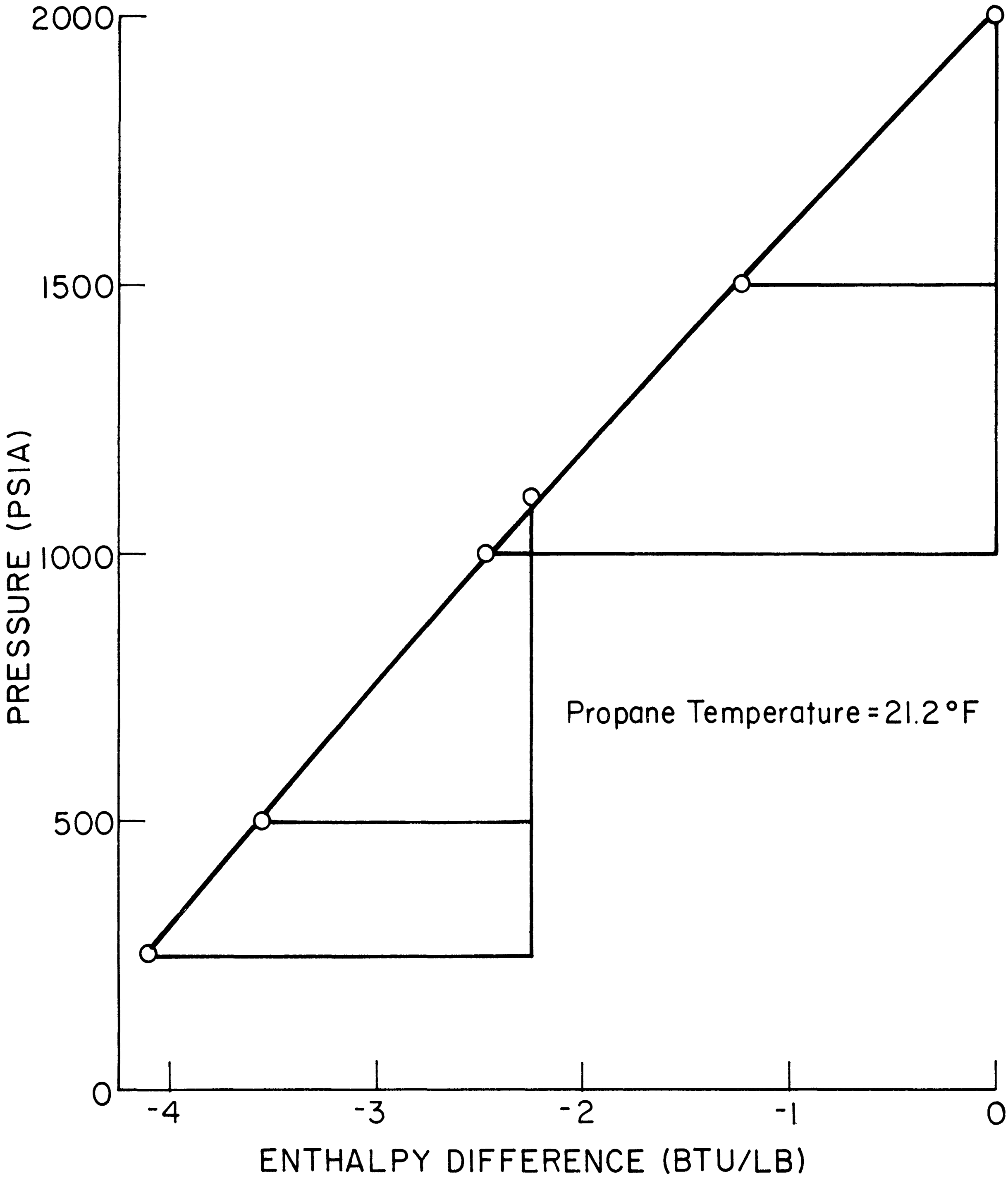


Figure 19. Pressure-Enthalpy Isotherm Generated from Isenthalpic and Isobaric Data

calorimeter in this investigation and experimental values recently reported by Helgeson and Sage[51] are within the 1.5 Btu/lb which is the uncertainty the latter claimed. At 588 psia (201°F) the value of this investigation is about 2 Btu/lb (4 percent) higher than the calculated value which appears in their publication. Values from the other sources vary from these results by as much as 10 Btu/lb (11 percent).

Isobaric heat capacity data at pressures in excess of 1 atm have only recently become available.[39,42] Figure 20 summarizes the results of Finn[42] and of this investigation at 700 psia, only 80 psi or so above the critical point. There is excellent agreement between the two investigations with respect to the temperature of the maximum in C_p (219°F) but the values reported by Finn are between 7 and 26 percent higher than those reported in this contribution with the maximum deviation occurring at the peak.

Ernst[39] reports values of C_p in the temperature range from 68 to 176°F at pressures up to 118 psia. This is below the lowest pressure used for isobaric determinations in the present investigation (250 psia) and also at temperatures lower than any used in the determination of φ for propane as a gas (201°F) so that direct comparisons cannot be made. However, when values of C_p from the two investigations were plotted versus either T or P, smooth isobaric and isothermal curves could be passed through all the data.

Values of C_p presented in Table VI were compared with values estimated by Kuloor, Newitt, and Bateman[69] using low pressure C_p values and PVT data. The results of this comparison are summarized in Figure 21.

Sciance <u>et al.</u>[131] recently published calculated values of the properties of the saturated phases of propane. Comparison of his reported

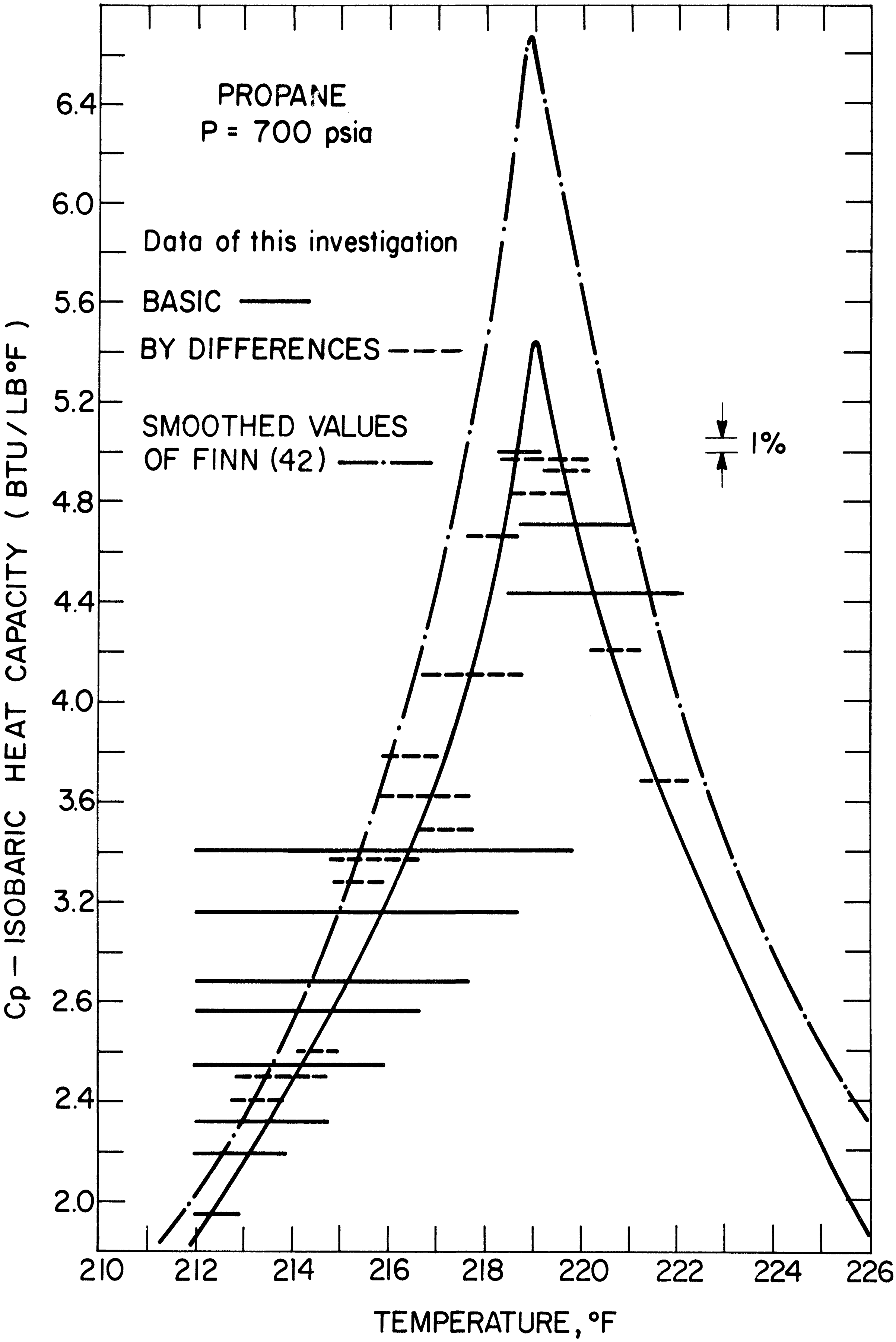


Figure 20. Experimental Data at 700 psia and Comparison with Results of Finn

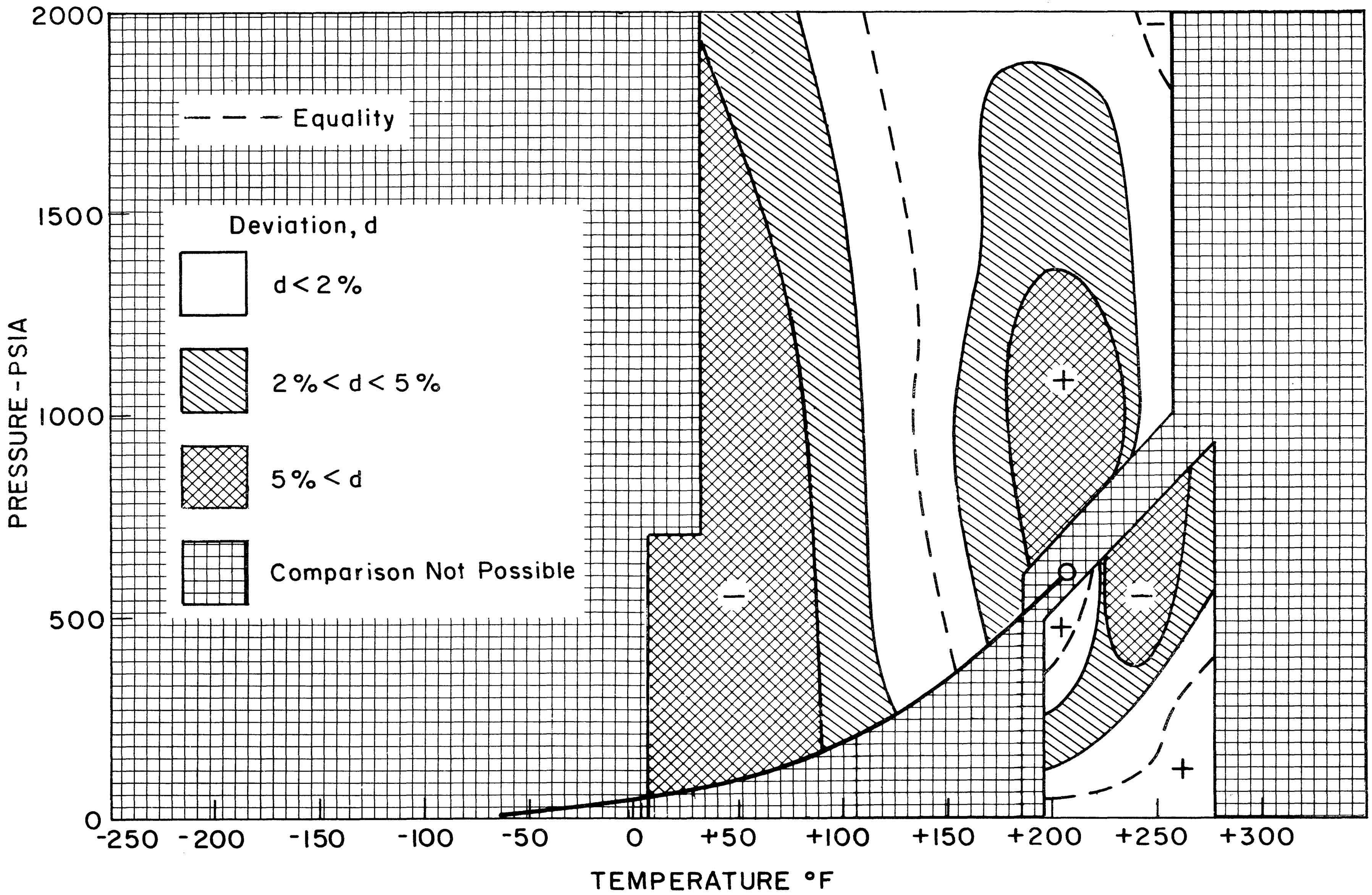


Figure 21. Comparison of Experimental Heat Capacities with Tabulated Values of Kuloor et al. (69)

values with values of C_p for the saturated vapor determined in the course of this study and reported in Table VII together with a value from the investigation of Ernst[39] indicates that the values reported by Sciance *et al.* are uniformly high by about 0.02 Btu/lb-°F ($\sim$4 percent).

Isothermal Data

Yarborough and Edmister[148] report results of isothermal throttling experiments for propane at 200, 300, and 400°F. Their results at 200 and 300°F are plotted as points on Figure 22.

Data from Runs 4R, 5R at 201°F, and 7R at 200.5°F were interpreted to yield values of the isothermal enthalpy departure and the results are represented by a solid line in Figure 22. When corrected to 200°F the results of the two investigations differ by about 5 Btu/lb above the boiling point (about 4 percent). This agreement seems reasonable when it is considered that the determinations are made through the two-phase region within 5°F of the critical temperature.

Isothermal measurements were not made at 300°F but isobaric determinations extended to this temperature. Therefore the isothermal data at 249.9°F (see Table VIII) were used in conjunction with isobaric results at elevated temperatures (see Table VI) to calculate values of enthalpy departure at 300°F. The results are plotted as a dashed line on Figure 22. The agreement at this temperature is good with a deviation of about 3 Btu/lb (5 percent at 1000 psia).

Isenthalpic Data

Unfortunately, it was not possible to make a direct comparison with other published experimental Joule-Thomson data[125] because the isenthalpic data of the investigation were obtained in the liquid region. However, it is possible to apply Equation (8) to check the consistency

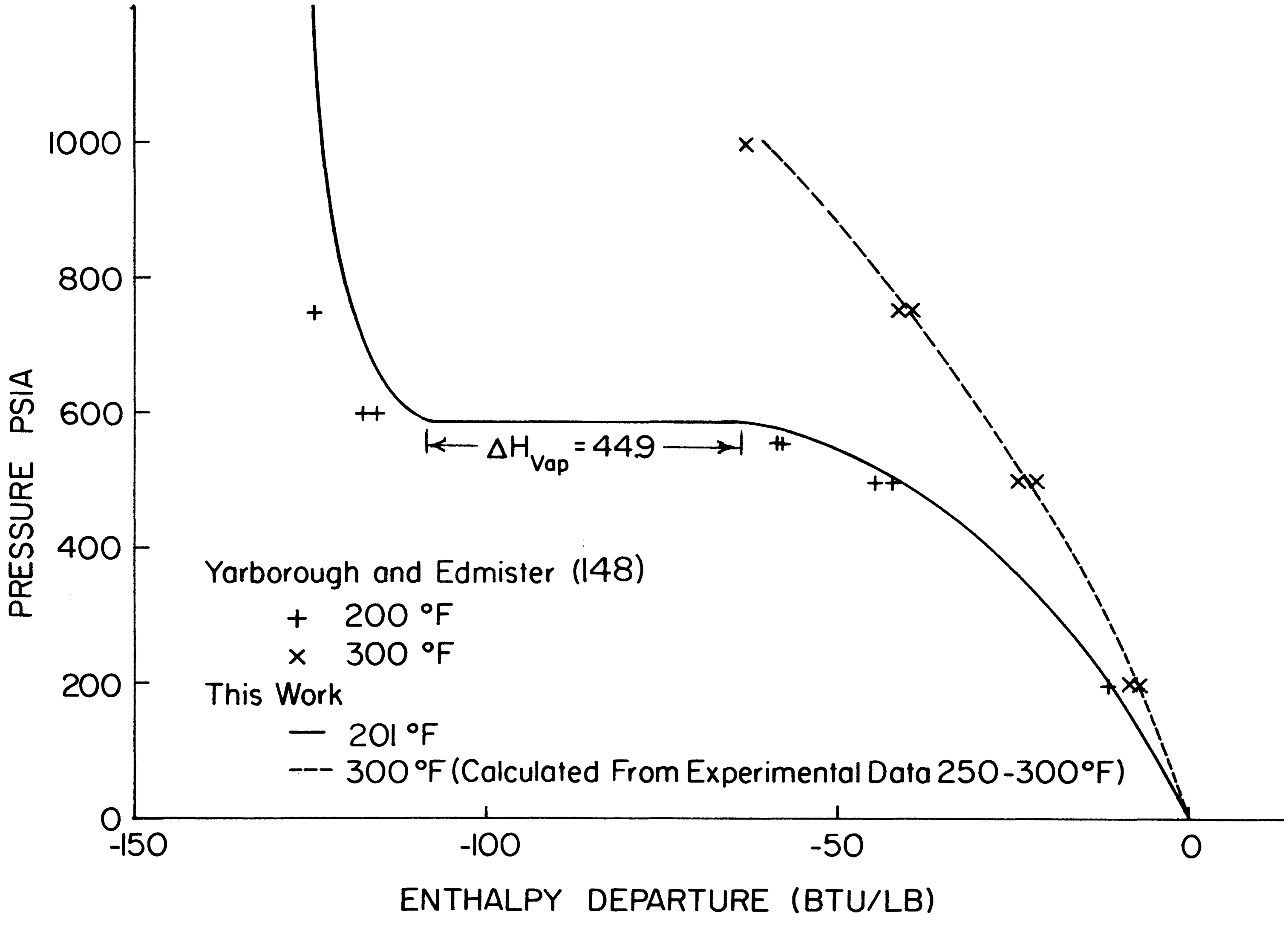


Figure 22. Comparison of Isothermal Enthalpy Departure with Data for Yarborough and Edmister

of φ and C_p data from this investigation with experimental values of μ from Reference 125. Smoothed experimental values of these properties are listed in Table X together with the ratio $(-\mu C_p/\varphi)$ which, according to Equation (8) should have a value of 1.0. The agreement is on the order of ±10 percent.

Enthalpy Diagram and Table

A number of tables and charts of thermodynamic properties of propane have been published.[12,14,17,26,69,103,127,135,153] The ranges in temperature and pressure covered by these published tables are indicated in Figure 23. The values of the thermal properties, enthalpy, and entropy, presented in these tables were calculated using heat capacity data at low pressure[25,26,66,67] and volumetric (P$\underline{V}$T) data. This general procedure has been followed because there have been very few other experimental data at elevated pressures with the exception of data on the latent heat of vaporization[26,51,124] and results of Joule-Thomson experiments.[125] The pressure-temperature-enthalpy diagram developed in this investigation, although developed mainly from the experimental portion of this research, summarizes all published experimental data on the thermal properties of propane over the range -250 to +500°F and 0 to 2000 psia. Figure 24 presents a list of the data used and the regions covered.

The reference for enthalpy was taken as H = 0 at T = -280°F for liquid propane at its saturation pressure. This is consistent with the reference previously used in reporting enthalpy values for mixtures of methane and propane.[81,82,85,86,87,88] Values of the enthalpy of propane as a gas at zero pressure were calculated using data on the liquid phase

Table X

Test of Consistency of Data

Based on Equation (8)

200°F

Pressure psia	C_p Btu/lb°F	ϕ Btu/lb psi	μ[a] °F/psi	$\frac{-\mu C_p}{\phi}$
500	0.854	0.172	0.1846	0.919
450	0.730	0.127	0.1752	1.005
400	0.652	0.104	0.1677	1.051
350	0.599	0.0903	0.1613	1.070
300	0.563	0.0802	0.1537	1.079
250	0.536	0.0723	0.1470	1.090
200	0.515	0.0662	0.1407	1.095
150	0.502	0.0614	0.1370	1.120
100	0.491	0.0580	0.1259	1.066
50	0.482	0.0555	0.1165	1.012
0	0.4744[b]	0.0536[c]		

a Interpolated from data in Table I in Reference (125)

b Value for ideal gas at zero pressure (118)

c Based on $\phi° = B - T\ (dB/dT)$

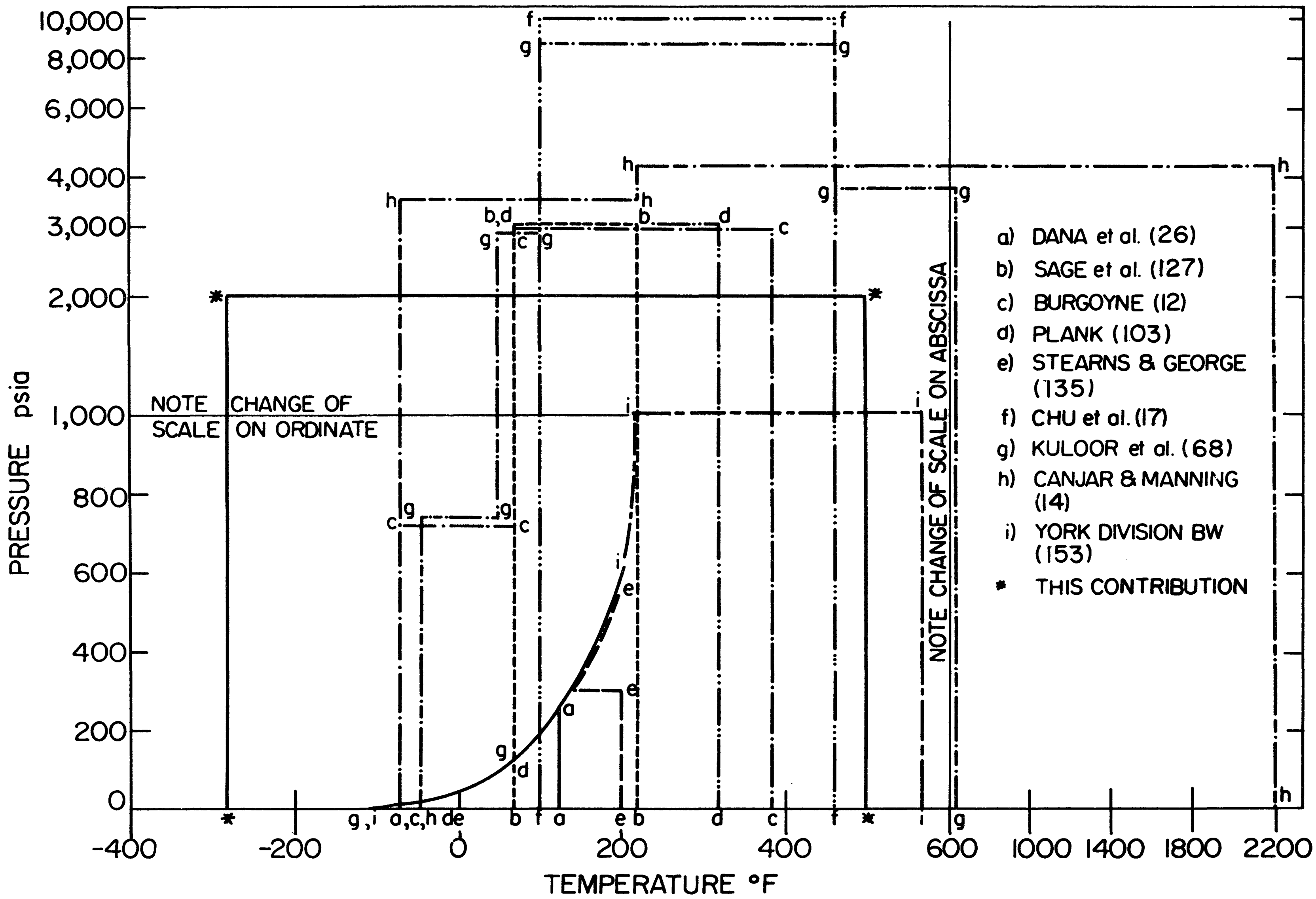


Figure 23. Range of Tables and Charts of Thermodynamic Properties of Propane

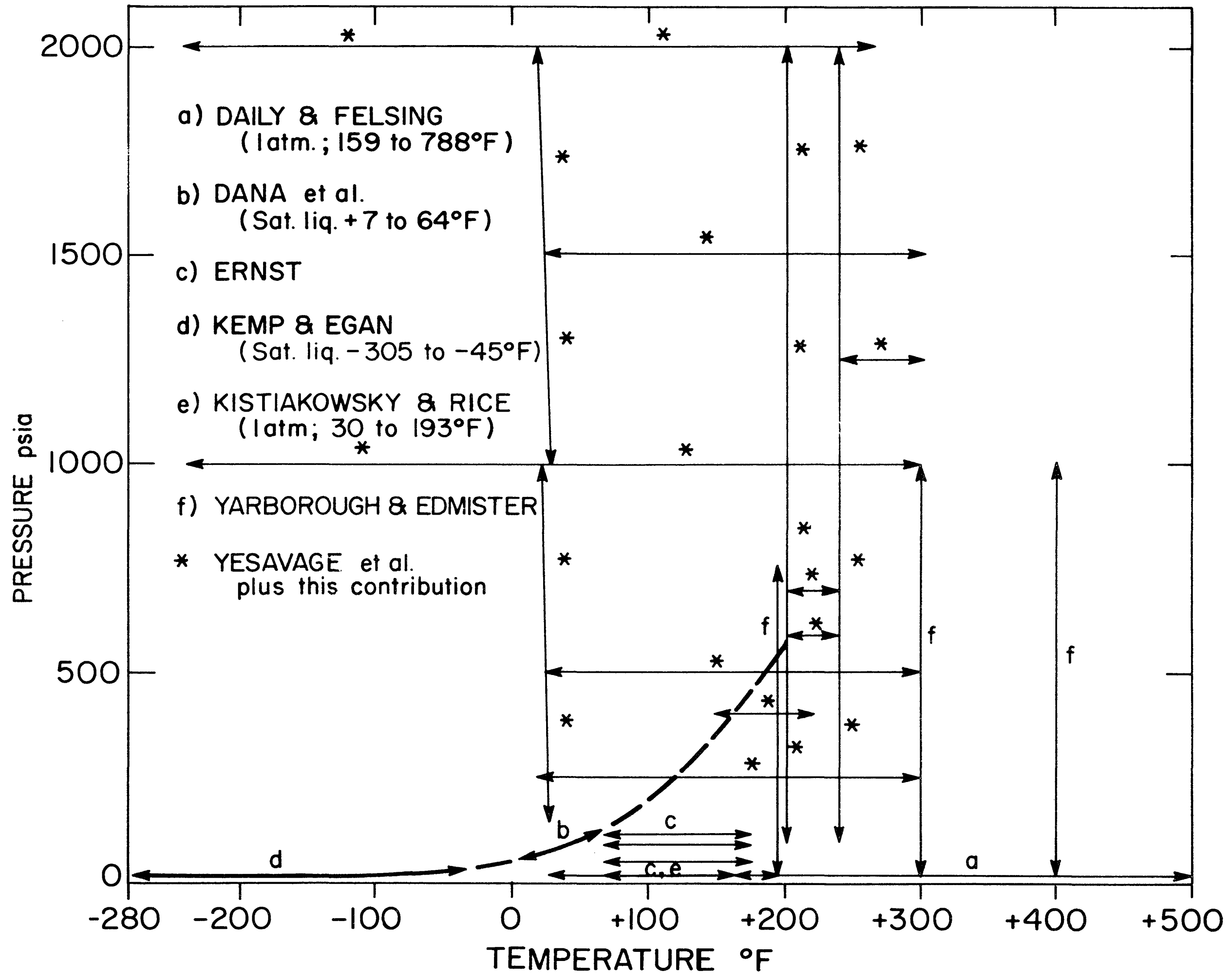


Figure 24. Range of Calorimetric Data Used in Preparation of Pressure-Enthalpy-Temperature Table for Propane

heat capacity,[66] the latent heat of vaporization at 1 atmosphere,[66] the BWR equqtion of state[7] to correct from 1 atm to zero pressure at the normal boiling point, and values of the ideal gas heat capacity[118] to account for a change in temperature from the normal boiling point to other temperatures at zero pressure. The calculations of the enthalpy at 250°F and zero pressure are summarized below.

	$\underline{H}$(Btu/lb)
Saturated liquid (at -280°F)	0
Saturated liquid (-280 to -43.7°F)	115.30
Enthalpy change on vaporization (at -43.7°F)	183.17
Effect of pressure on enthalpy (14.7 to 0 psia)	2.70
Effect of temperature on zero pressure enthalpy (-43.7 to +250°F)	121.93
$\underline{H}^0$ (Propane at zero pressure and +250°F)	423.1

Isobaric enthalpy differences at elevated pressures were calculated from the data presented in Table VI plus other published values of C_p.[39] The pressures at which such data are available are indicated by horizontal lines on Figure 24. The values of C_p recently reported by Ernst[39] are in excellent agreement with those at 1 atm previously published by Kistiakowsky and Rice.[62]

Above 100 psia, values of φ from this investigation were used to determine the effect of pressure on enthalpy. These basic data were supplemented by published experimental values of enthalpy departures, $(\underline{H}^0 - \underline{H})$.[146,148] The temperatures as which such data are available are indicated by vertical lines on Figure 24. As already mentioned the effect of pressure on enthalpy at low pressures (100 psia) was estimated using published experimental values of the pressure dependence of enthalpy[146,148] together with values of φ^0 estimated from published correlations of the second virial coefficient, B, and its temperature dependence[30,31,58] (see Equation (32)) and estimates made using the BWR equation of state[7] with primary reliance placed on the correlation of experimental values of B as made available by Diaz-Pena.[30,31]

The interpreted isenthalpic data yielding isothermal differences in enthalpy as shown in Figure 19 was used in the compressed liquid region.

In establishing the enthalpy change on vaporization, experimental data from Dana et al.[26] and this investigation were used together with "critically chosen values" reported by Helgeson and Sage.[51] The values from Helgeson and Sage are in excellent agreement (better than 1 percent) with the results of this investigation in the region of overlap as already shown in Table IV.

No attempt was made to correlate the extensive vapor pressure data reported in the literature. Instead values were taken from recent tabulations.[14,69]

As indicated by Figure 9 redundant thermal data are available between 21 and 300°F at pressures up to 2000 psia. Consistency checks were made as permitted by these redundancies and the results are summarized in Figure 25. Consider the loop between the pressures of 500 and 1000 psia and the temperatures of 21.2 and 160.5°F. As indicated on the figure there is some error in the constituent experimental determinations because the algebraic sum of the enthalpy differences around the complete loop, $\Delta \underline{H}_1$, is not zero but instead is -0.64 Btu/lb. The percentage deviation, defined as

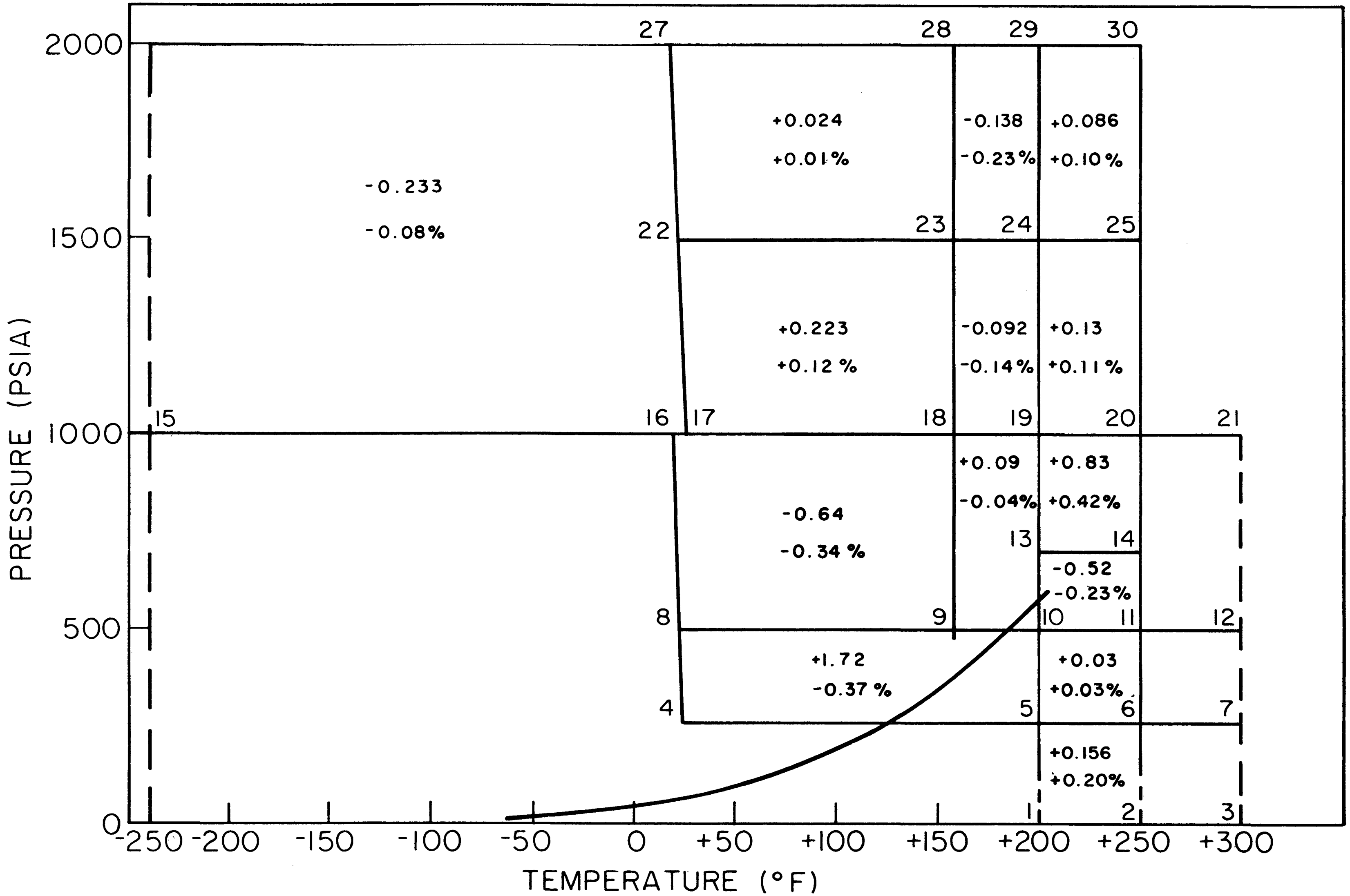


Figure 25. Checks of Thermodynamic Consistency of Thermal Data for Propane

$$\text{percentage deviation} = \frac{\sum_i \Delta\underline{H}_i}{\sum_i \left|\Delta\underline{H}_i\right|} \times 100 \qquad (77)$$

is determined to be -0.34. The maximum percentage deviation for any loop is +0.42 percent and the average absolute deviation of all such checks is 0.18 percent. It is felt that this is indicative of the accuracy of the enthalpy differences in this region.

The differences in enthalpy at a point calculated from different experimental data are small but nevertheless it was necessary to make minor adjustments in preparing the final compilation. This was done by adjusting individual values of isobaric and isothermal enthalpy differences to make each loop thermodynamically consistent. These adjustments were made within the limits of experimental uncertainty of the data. In general, the uncertainty in $\Delta\underline{H}_P$ is ±0.3 percent except near the critical region. The uncertainty in $\Delta\underline{H}_T$ is ±1 percent except at pressures below 200 psia.

In extending the calculations of enthalpy down to -280°F use was made of experimental values of C_p for the saturated liquid[66] and values at 1000 and 2000 psia at temperatures of -240°F and above were extrapolated graphically to the lower temperatures. The variation in value of C_p over the range of extrapolation was about ±1 percent.

Above 300°F primary reliance was placed on the C_p data of Daily and Felsing[25] at 1 atm. Between 300 and 400°F and at elevated pressures smoothed curves were drawn to blend experimental values at 300°F with values calculated using the BWR equation of state[7] with the low pressure data. The blending was carried out so that the results were consistent with the experimental values of Yarborough and Edmister at 400°F.[146,148] Between 400 and 500°F enthalpy departures were calculated using the BWR equation of state.[7]

After all adjustments and extrapolations had been made as described above, a skeleton table of values of enthalpy was prepared. These values were then plotted on a diagram and smooth curves drawn to connent all points and to represent interpolated values. The results are presented as Figures 26a and 26b. Values were then read from the master plot and are reported in Tables XI and XII. From -250 to +300°F at the temperatures and pressures of measurement as indicated in Figure 9, little or no interpolation is involved and therefore the numbers listed for these temperatures and pressures can be considered to be smoothed experimental values.

Since the values in Table XII are based on extensive data on the thermal properties of propane between -240 and +300°F at pressures up to 2000 psia, the values in this region are believed to be accurate enough to permit determinations of enthalpy to within ±1 Btu/lb or less. The graphical representation is slightly less accurate. Extrapolation of C_p data to -280°F was carried out by graphical means. Extension of the values of +500°F was based on a combination of limited experimental data on the thermal properties and the BWR equation of state, and the uncertainty in these regions may be somewhat greater.

A comparison of the results of this table has been made with the results of a most recent compilation by Canjar and Manning[14] and is given in Figure 27. The deviations are quite significant in the region right above the critical temperature (as large as 7 Btu/lb).

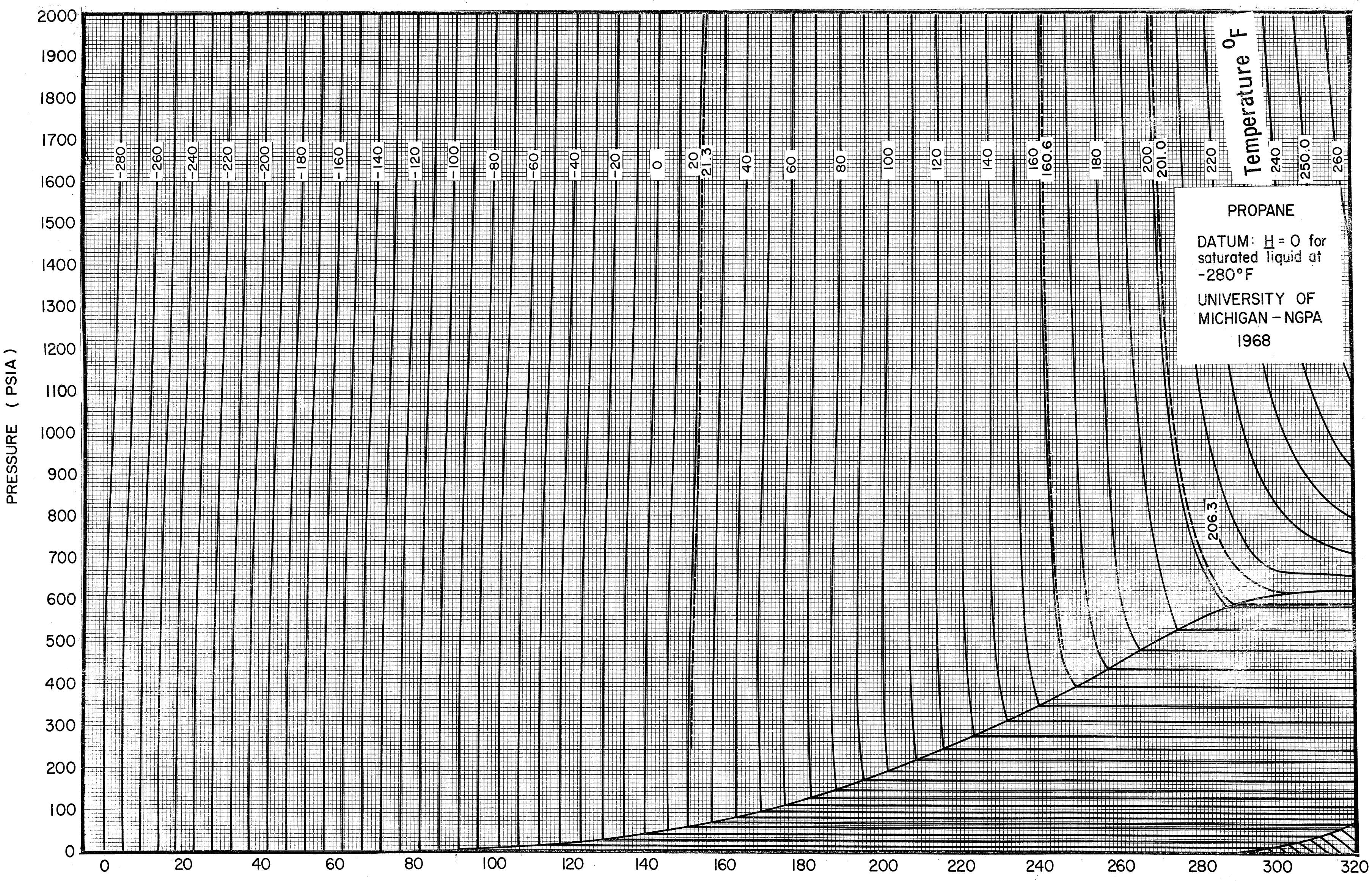


Figure 26a. Pressure-Temperature-Enthalpy Diagram for Propane

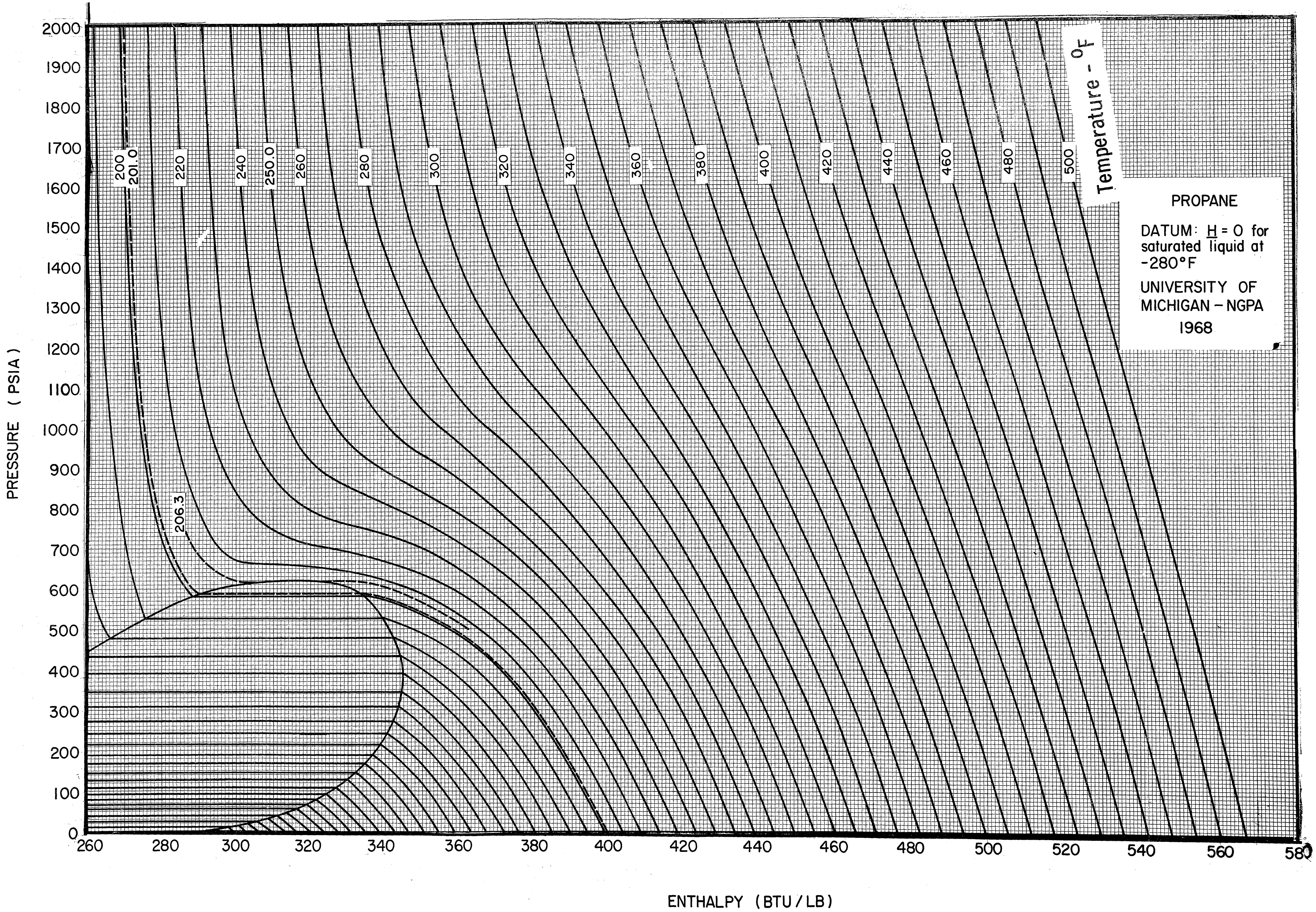


Figure 26b. Pressure-Temperature-Enthalpy Diagram for Propane

TABLE XI

TABULATED VALUES OF ENTHALPY
FOR PROPANE AT SATURATED CONDITIONS

Pressure (psia)	Temperature (°F)	Saturated Liquid Enthalpy (Btu/lb)	Saturated Vapor Enthalpy (Btu/lb)	Latent Heat of Vaporization (Btu/lb)
14.7	-43.7	116.2	299.4	183.2
50	14.1	147.2	314.5	167.3
100	55.0	172.2	325.7	153.5
150	82.7	190.1	332.7	142.6
200	104.5	204.6	337.6	133.0
250	122.5	217.4	340.8	123.4
300	137.6	230.0	343.4	113.4
350	151.4	241.4	344.6	103.2
400	163.5	252.0	344.8	92.8
450	174.7	261.4	343.4	82.0
500	184.8	270.4	340.8	70.4
550	194.4	279.8	337.1	57.3
588	201.0	289.7	333.1	43.4
600	203.5	296.4	330.7	34.3
617	206.3	315.0	315.0	0.0

TABLE XII

TABULATED VALUES OF ENTHALPY
FOR PROPANE

$\underline{H}$ (Btu/lb)

Temperature (°F)	Pressure, psia: 0	100	200	250	300	350	400	450	500	550
-280		.1	.2	.3	.4	.5	.6	.7	.8	.9
-270		4.8	4.9	5.0	5.0	5.0	5.1	5.1	5.2	5.4
-260		9.3	9.5	9.6	9.7	9.6	9.7	9.8	10.0	10.1
-250	242.9	14.0	14.1	14.2	14.3	14.3	14.4	14.5	14.5	14.7
-240	245.4	18.5	18.6	18.8	18.9	19.0	19.1	19.2	19.4	19.5
-230	248.0	23.2	23.3	23.4	23.5	23.5	23.6	23.8	23.9	24.0
-220	250.5	28.0	28.1	28.2	28.3	28.4	28.5	28.6	28.7	28.9
-210	253.1	32.9	33.0	33.0	33.1	33.0	33.1	33.2	33.4	33.5
-200	255.7	37.5	37.6	37.6	37.7	37.8	37.9	38.0	38.0	38.2
-190	258.3	42.2	42.3	42.4	42.5	42.5	42.7	42.8	43.0	43.0
-180	261.0	46.9	47.0	47.1	47.2	47.3	47.5	47.6	47.7	47.9
-170	263.7	51.7	51.8	51.9	52.0	52.0	52.1	52.2	52.4	52.5
-160	266.4	56.7	56.8	56.9	57.0	57.0	57.1	57.2	57.4	57.5
-150	269.2	61.5	61.6	61.6	61.7	61.7	61.8	62.0	62.1	62.4
-140	272.0	66.5	66.6	66.6	66.7	66.8	66.9	67.0	67.1	67.2
-130	274.8	71.4	71.5	71.6	71.7	71.8	71.9	72.0	72.1	72.3
-120	277.7	76.4	76.5	76.6	76.7	76.8	76.9	76.9	77.0	77.0
-110	280.6	81.4	81.5	81.5	81.6	81.7	81.8	81.8	81.9	82.0
-100	283.6	86.5	86.7	86.7	86.8	86.8	86.9	87.0	87.0	87.0
-90	286.6	91.5	91.7	91.7	91.8	91.8	91.9	91.9	92.0	92.0
-80	289.7	96.5	96.6	96.6	96.7	96.7	96.8	96.9	97.0	97.1
-70	292.8	101.8	101.9	102.0	102.0	102.0	102.0	102.0	102.1	102.3
-60	296.0	106.8	106.9	107.0	107.0	107.0	107.1	107.2	107.4	107.5
-50	299.2	112.2	112.2	112.3	112.4	112.3	112.4	112.5	112.6	112.8
-40	302.5	117.5	117.5	117.5	117.5	117.7	117.8	117.8	118.0	118.1
-30	305.8	123.0	123.0	123.0	123.0	123.1	123.1	123.2	123.4	123.5
-20	309.2	128.5	128.5	128.5	128.5	128.6	128.7	128.7	128.7	128.9
-10	312.6	133.9	134.0	134.0	134.0	134.1	134.1	134.1	134.2	134.3
0	316.1	139.2	139.4	139.5	139.5	139.6	139.6	139.7	139.8	139.9
10	319.7	145.1	145.2	145.3	145.3	145.3	145.3	145.3	145.4	145.5
20	323.3	150.4	150.5	150.6	150.7	150.8	150.8	150.9	151.1	151.2
21.3				151.2	151.3	151.4	151.5	151.6	151.7	151.9
30	327.0	156.5	156.5	156.5	156.6	156.7	156.8	156.9	157.0	157.1
40	330.7	162.5	162.5	162.5	162.6	162.7	162.7	162.7	162.8	162.9
50	334.5	169.0	168.8	168.7	168.5	168.4	168.4	168.3	168.4	168.4
60	338.3	327.6	174.9	174.8	174.8	174.8	174.8	174.8	174.9	175.0
70	342.2	332.9	181.5	181.3	181.2	181.1	181.1	181.1	181.1	181.1
80	346.2	337.5	187.8	187.6	187.5	187.3	187.2	187.2	187.2	187.3
90	350.2	342.1	195.0	194.4	194.1	193.9	193.7	193.7	193.7	193.8
100	354.3	346.6	201.3	201.0	200.7	200.5	200.4	200.3	200.2	200.2
110	358.5	351.3	340.5	208.2	207.9	207.6	207.3	207.1	206.9	206.8
120	362.7	355.5	346.0	215.5	215.2	214.9	214.5	214.3	214.0	214.0
130	367.0	360.2	351.2	344.8	223.1	222.5	222.0	221.6	221.3	221.1
140	371.3	364.7	356.1	350.6	344.7	230.8	229.9	229.2	228.8	228.6
150	375.7	369.1	361.0	356.0	350.4	240.0	238.5	237.4	236.8	236.5
160	380.2	373.7	366.3	361.7	356.6	350.5	248.9	246.4	245.2	244.8
160.6			368.2	363.8	358.7	352.5	249.8	247.0	245.9	245.3
170	384.7	378.6	371.7	367.5	362.3	356.5	349.6	256.4	254.2	253.2
180	389.3	383.3	376.4	372.5	368.1	363.0	356.7	348.2	264.3	262.3
190	394.0	388.1	381.6	378.1	374.1	369.5	363.9	356.8	347.1	274.5
200	398.7	392.9	386.6	383.4	379.7	375.6	370.8	364.8	356.9	345.9
201.0	399.2	393.5	387.5	384.0	380.3	376.4	371.4	365.4	357.8	347.0
210	403.5	398.0	392.1	388.8	385.2	381.2	376.7	371.5	365.2	357.1
220	408.3	403.0	397.3	394.2	390.8	387.1	383.0	378.4	372.7	366.2
230	413.2	408.2	402.7	399.7	396.3	392.8	388.8	384.6	379.6	374.0
240	418.1	413.1	407.9	405.0	401.8	398.6	395.0	391.2	386.8	381.9
250	423.1	418.3	413.3	410.6	407.5	404.3	401.0	397.3	393.3	389.1
260	428.2	423.5	418.7	416.0	413.3	410.3	407.0	403.6	399.8	395.5
270	433.3	428.7	424.0	421.5	419.0	416.2	413.1	409.9	406.3	402.4
280	438.5	434.1	429.6	427.2	424.6	421.8	418.9	415.8	412.6	409.0
290	443.7	439.6	435.1	432.8	430.3	427.7	425.0	421.9	418.8	415.5
300	449.0	445.2	440.8	438.5	436.1	433.5	430.9	428.1	425.2	422.2
310	454.4	450.7	446.4	444.1	441.7	439.2	436.6	434.0	431.2	428.4
320	459.8	455.9	451.8	449.9	447.5	445.2	442.8	440.3	437.7	435.1
330	465.3	461.7	457.7	455.6	453.4	451.1	448.8	446.3	443.9	441.4
340	470.9	467.2	463.4	461.5	459.4	457.2	455.0	452.7	450.2	447.8
350	476.5	473.0	469.3	467.3	465.3	463.1	460.9	458.7	456.4	454.1
360	482.1	478.8	475.2	473.3	471.3	469.2	467.1	464.9	462.8	460.5
370	487.9	484.5	481.1	479.2	477.3	475.4	473.3	471.3	469.1	467.1
380	493.6	490.5	487.0	485.2	483.3	481.4	479.4	477.4	475.3	473.3
390	499.5	496.3	492.9	491.2	489.4	487.5	485.7	483.7	481.7	479.8
400	505.4	502.2	498.9	497.3	495.4	493.6	491.7	489.8	487.8	486.1
410	511.3	508.3	505.1	503.5	501.8	500.0	498.3	496.5	494.6	492.8
420	517.3	514.4	511.3	509.7	508.0	506.3	504.6	502.8	501.1	499.3
430	523.4	520.4	517.4	515.9	514.3	512.7	511.0	509.4	507.6	506.0
440	529.5	526.7	523.7	522.3	520.7	519.1	517.5	515.7	514.0	512.5
450	535.6	532.9	530.0	527.5	527.0	525.5	523.9	522.4	520.7	519.1
460	541.8	539.3	536.4	535.0	533.5	532.0	530.5	528.9	527.3	525.8
470	548.0	545.4	542.7	541.3	539.8	538.4	536.9	535.5	534.0	532.5
480	554.3	551.7	549.2	547.8	546.5	545.1	543.7	542.3	540.8	539.4
490	560.7	558.2	555.6	554.3	553.0	551.6	550.2	548.8	547.4	546.0
500	567.1	564.7	562.2	561.0	559.7	558.3	556.9	555.5	554.1	552.8

TABLE XII (continued)

TABULATED VALUES OF ENTHALPY
FOR PROPANE

$\underline{H}$ (Btu/lb)

Temperature (°F)	Pressure, psia 600	617	700	800	900	1000	1250	1500	1750	2000
-280	1.0	1.1	1.4	1.7	1.9	2.5	3.4	4.0	4.4	4.6
-270	5.5	5.5	6.0	6.3	6.7	7.1	8.0	8.8	9.2	9.3
-260	10.2	10.3	10.5	10.9	11.4	11.7	12.6	13.3	13.7	13.9
-250	14.9	15.0	15.1	15.5	15.9	16.3	17.2	18.0	18.5	18.6
-240	19.7	19.8	20.0	20.5	20.8	21.1	22.0	22.6	23.1	23.3
-230	24.3	24.4	24.7	25.1	25.5	25.8	26.8	27.5	27.7	27.9
-220	29.0	29.0	29.4	29.7	30.0	30.4	31.4	32.0	32.5	32.6
-210	33.7	33.7	34.0	34.5	34.9	35.3	36.1	36.9	37.2	37.4
-200	38.5	38.5	38.7	39.1	39.5	39.8	40.8	41.4	42.0	42.2
-190	43.3	43.4	43.6	44.0	44.3	44.7	45.7	46.3	46.8	47.0
-180	48.0	48.1	48.4	48.6	49.0	49.4	50.4	51.0	51.5	51.8
-170	52.7	52.8	53.1	53.5	53.8	54.1	55.2	55.9	56.3	56.6
-160	57.7	57.8	58.0	58.4	58.7	59.0	60.2	60.8	61.3	61.5
-150	62.5	62.5	62.9	63.1	63.5	63.9	65.0	65.5	66.1	66.5
-140	67.4	67.5	67.8	68.1	68.5	68.9	69.8	70.4	71.0	71.3
-130	72.4	72.5	72.8	73.1	73.5	73.8	74.6	75.3	75.8	76.2
-120	77.2	77.2	77.6	78.0	78.3	78.6	79.6	80.3	80.9	81.2
-110	82.0	82.1	82.5	82.8	83.2	83.5	84.5	85.4	85.9	86.0
-100	87.1	87.2	87.5	87.8	88.2	88.5	89.5	90.3	90.8	91.0
-90	92.1	92.2	92.5	92.9	93.2	93.5	94.5	95.3	95.8	96.2
-80	97.3	97.4	97.6	98.0	98.3	98.6	99.7	100.5	100.9	101.3
-70	102.5	102.5	102.8	103.1	103.4	103.7	104.7	105.5	106.1	106.5
-60	107.6	107.7	108.0	108.4	108.7	109.0	110.0	110.7	111.3	111.7
-50	113.0	113.0	113.3	113.6	113.9	114.3	115.4	115.9	116.4	116.9
-40	118.2	118.3	108.5	108.9	109.2	119.5	120.6	121.2	121.7	122.0
-30	123.5	123.6	123.9	124.2	124.5	124.9	125.8	126.5	127.0	127.5
-20	129.0	129.0	129.3	129.6	129.9	130.3	131.1	131.7	132.3	132.7
-10	134.4	134.5	134.8	135.0	135.4	135.8	136.5	137.1	137.7	138.1
0	140.0	140.0	140.3	140.7	141.0	141.2	142.0	142.6	143.0	143.5
10	145.6	145.7	146.0	146.3	146.6	146.9	147.6	148.1	148.5	149.0
20	151.3	151.4	151.7	152.0	152.4	152.5	153.1	153.5	154.0	154.6
21.3	152.1	152.1	152.5	152.7	153.0	153.2	153.8	154.0	154.6	155.2
30	157.3	157.3	157.6	157.9	158.1	158.3	158.9	159.4	160.0	160.4
40	163.0	163.0	163.4	163.6	163.9	164.2	164.9	165.3	165.9	166.2
50	168.5	168.6	168.9	169.3	169.6	170.0	170.8	171.3	171.7	171.9
60	175.1	175.2	175.4	175.6	175.9	176.1	176.9	177.2	177.4	177.7
70	181.2	181.2	181.4	181.6	181.9	182.2	182.9	183.2	183.2	183.5
80	187.3	187.4	187.6	187.9	188.1	188.4	188.9	189.3	189.5	189.5
90	193.9	193.9	194.1	194.2	194.5	194.6	194.9	195.4	195.8	195.8
100	200.2	200.2	200.4	200.6	200.9	201.0	201.3	201.6	201.8	201.9
110	206.8	206.8	207.0	207.1	207.4	207.5	207.5	208.0	208.0	208.0
120	213.9	213.9	214.0	214.1	214.1	214.2	214.3	214.5	214.6	214.5
130	221.0	220.9	220.8	220.8	220.9	220.9	221.0	221.0	221.0	220.8
140	228.4	228.4	228.2	228.0	227.9	228.0	228.0	227.8	227.6	227.4
150	236.2	236.1	235.8	235.4	235.1	235.0	234.7	235.5	234.2	233.9
160	244.3	244.2	243.7	243.2	242.6	242.4	241.5	241.2	240.8	240.4
160.6	244.9	244.8	244.3	243.7	243.1	242.9	242.0	241.7	241.5	241.1
170	252.5	252.3	251.4	250.7	250.3	250.0	249.2	248.4	247.9	247.4
180	261.1	260.8	259.8	258.9	258.1	257.5	256.5	255.6	255.0	254.4
190	272.7	272.1	269.9	268.1	267.0	266.0	264.0	262.7	261.9	261.4
200	286.3	285.2	280.8	277.6	275.6	274.1	271.5	270.1	269.2	268.4
201.0	287.8	286.6	282.4	278.5	276.6	275.1	272.6	271.1	270.0	269.3
210	345.3	340.3	295.1	289.1	285.6	283.1	279.6	277.9	276.7	275.6
220	357.8	354.7	326.2	303.1	296.4	292.6	287.9	285.7	284.3	283.0
230	367.6	365.3	350.5	318.8	308.0	302.8	296.2	293.8	291.9	290.3
240	376.3	374.5	362.4	342.5	321.9	314.2	305.7	302.1	299.6	298.1
250	384.2	382.5	372.9	357.4	339.2	326.8	315.3	310.5	307.8	305.8
260	391.0	389.3	381.2	369.3	355.2	341.0	326.0	319.3	315.8	313.7
270	398.2	396.7	389.1	378.6	367.1	354.7	337.0	328.4	324.1	321.8
280	405.2	403.8	397.1	388.5	378.9	367.2	348.2	337.8	332.9	330.0
290	412.2	411.1	405.2	397.4	388.5	378.4	358.6	347.1	341.4	338.3
300	419.0	418.0	412.5	405.3	397.4	388.6	368.1	356.8	350.8	346.5
310	425.5	424.5	419.5	413.0	406.0	398.0	378.6	366.4	359.7	355.0
320	432.4	431.5	426.8	420.8	414.1	406.6	388.3	375.8	368.5	363.4
330	438.8	437.8	433.3	427.5	421.5	415.0	398.1	385.3	377.5	371.8
340	445.3	444.4	440.1	434.7	429.0	423.0	407.1	394.6	386.5	380.3
350	451.6	450.8	446.6	441.6	436.3	430.8	416.3	403.9	395.2	388.7
360	458.3	457.5	453.5	448.6	443.6	438.4	424.7	412.7	403.9	397.4
370	464.9	464.2	460.5	455.8	451.0	445.8	432.7	421.4	412.8	405.8
380	471.2	470.5	467.0	462.6	458.1	453.3	441.0	430.1	421.5	414.2
390	477.7	477.0	473.6	469.4	465.1	460.7	449.1	438.5	430.0	422.7
400	484.0	483.5	480.1	476.1	472.0	467.8	456.9	446.7	438.4	431.1
410	491.0	490.3	487.1	483.2	479.3	475.4	465.0	455.0	447.2	439.7
420	497.4	496.8	493.7	489.9	486.2	482.5	472.8	463.2	455.3	448.1
430	504.3	503.6	500.7	497.2	493.6	490.0	480.5	471.2	463.5	456.2
440	510.8	510.2	507.4	504.0	500.6	497.1	488.1	479.2	471.5	464.8
450	517.6	517.0	514.2	511.0	507.6	504.2	495.6	487.0	479.5	473.0
460	524.2	523.7	521.2	518.0	514.8	511.5	503.1	494.9	487.5	481.2
470	531.1	530.6	528.0	525.0	521.8	518.6	510.6	502.6	495.7	489.3
480	538.0	537.5	535.0	532.0	531.5	525.9	518.2	510.5	503.8	497.5
490	544.6	544.1	541.7	538.8	536.0	533.0	525.7	518.2	511.6	505.5
500	551.3	550.9	548.6	545.8	543.0	540.1	533.0	525.9	519.5	513.5

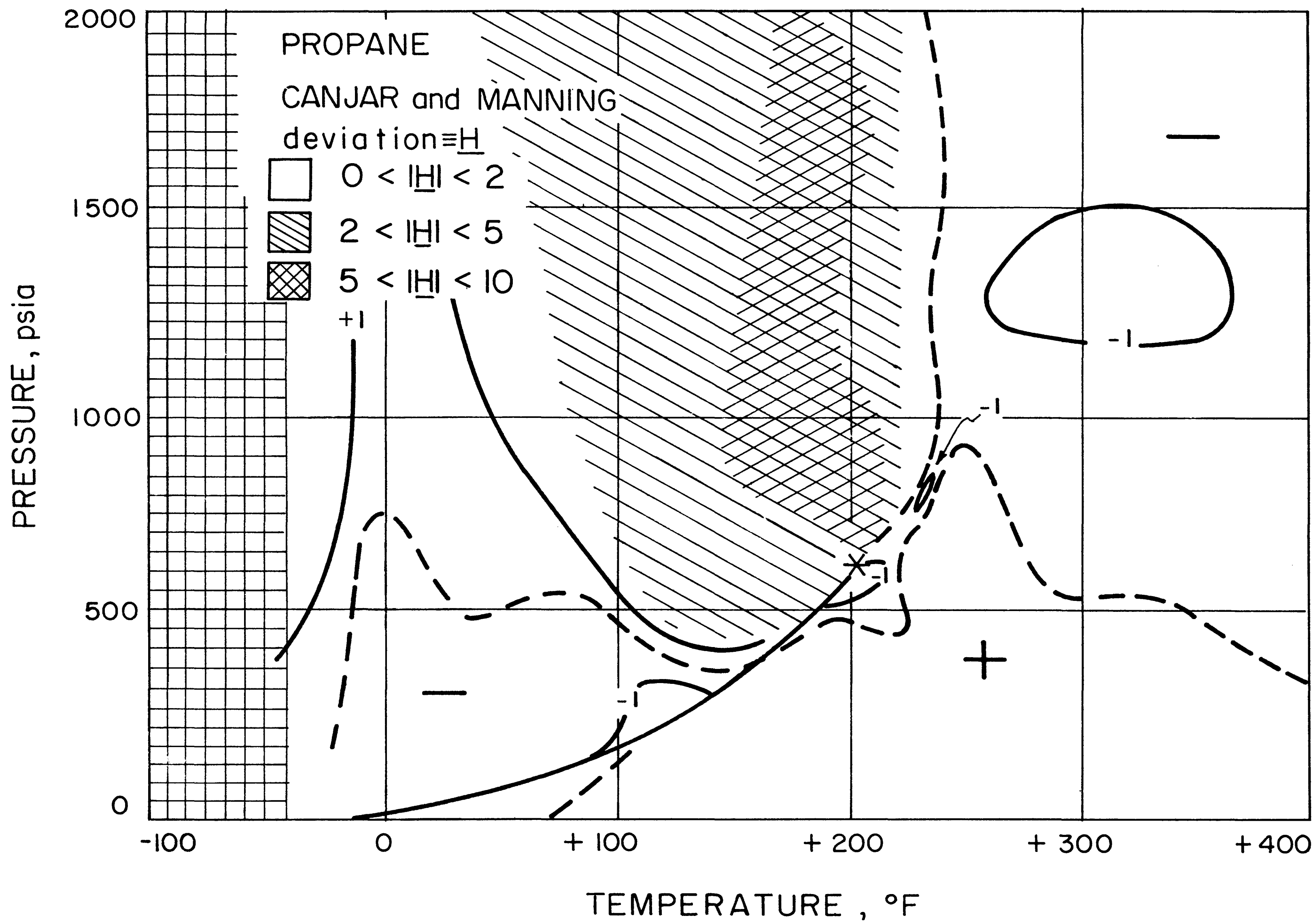


Figure 27. Comparison of Tabulated Enthalpies of This Investigation with Those of Canjar and Manning

SECTION V - THE ENTHALPY OF METHANE-PROPANE MIXTURES UNDER PRESSURE

As already mentioned the enthalpy of a 5.1 mole percent,[80,85] a 12 and a 28 mole percent[85] mixture of propane in methane had already been obtained over a wide range of temperature and pressure before the start of this investigation. Thus, to complete the methane-propane system data for two additional mixtures with propane as the major component were obtained.

The 76.6 Mole Percent Propane in Methane Mixture

After completing the experimental investigation of propane, methane was added to the fluid in the system to obtain a nominal 77 mole percent propane in methane mixture.

Composition of Gas

The methane used in the investigation of the mixture was obtained from the Southern California Gas Company. The same instrument grade propane as obtained from the Phillips Petroleum Company was used for the mixtures as was used for the determinations of pure propane. The composition of this mixture as determined by mass spectrometer and chromatographic analyses are reported in Table XIII.

A chromatograph which is incorporated as part of the recycle flow facility was used for frequent checks on the composition of the gas in the system. The chromatograph was calibrated using samples of known composition prepared by direct weighing.

From time to time the fluid composition did change. This was most frequent during periods of excessive leakage of fluid from the system. The composition was re-established within reasonable limits by the

TABLE XIII

COMPOSITION OF NOMINAL 77 MOLE PERCENT PROPANE IN METHANE MIXTURE

	Chromatograph[a] (mole percent)	Mass Spectrometer[b] (mole percent)
Nitrogen	0.20	$<$ 0.05
Methane	23.16	24.23
Ethane	$<$ 0.05	$<$ 0.05
Carbon Dioxide	$<$ 0.05	$<$ 0.05
Propane	76.53	75.77
Butane	0.11	$<$ 0.05
	$<$ 100.10	$<$ 100.20

a Sample taken from system in March 1968.
b Sample taken from system in October 1967.

addition of one component of the mixture and remixing the entire system.

The chromatographic analyses on a day to day basis are summarized in Figure 28. There is no indication of a change in average composition with time.

Regions of Measurement

Experimental measurements of isobaric, isothermal, and isenthalpic changes in enthalpy for the mixture containing 76.6 mole percent propane in methane were made in the liquid, two-phase, critical, and gaseous regions at temperatures between -240 and +300°F and pressures between 100 and 2000 psia. The ranges of pressures and temperatures covered by these experiments are indicated by lines drawn on a pressure versus temperature diagram in Figure 29. In addition to the mixture data normally obtained with the recycle facility the first successful attempt to obtain an isothermal enthalpy of vaporization was made. Experimental results for each run are presented in Appendix B. The isobaric single phase and

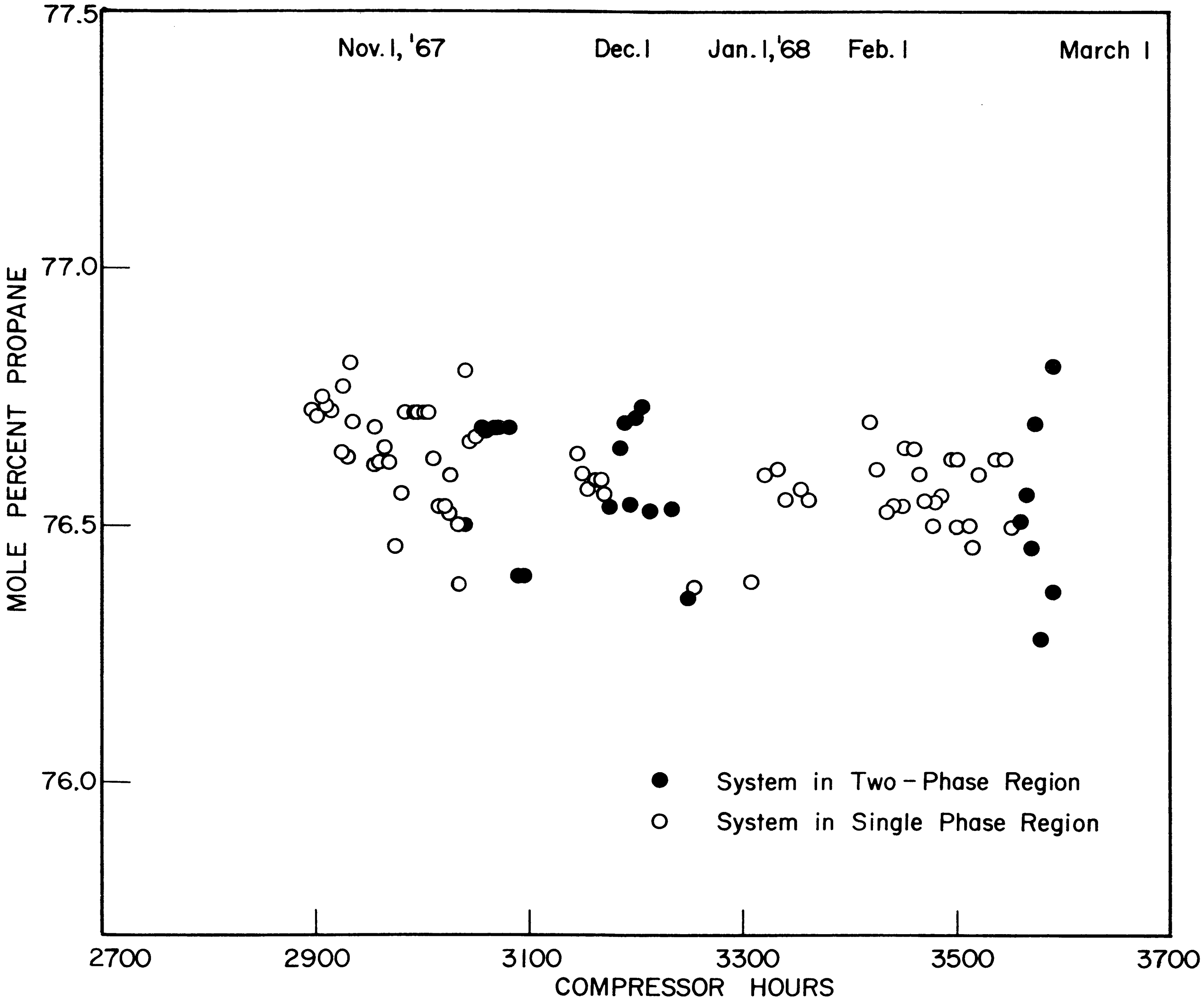


Figure 28. Composition of the Nominal 77 Percent Mixture as a Function of Time

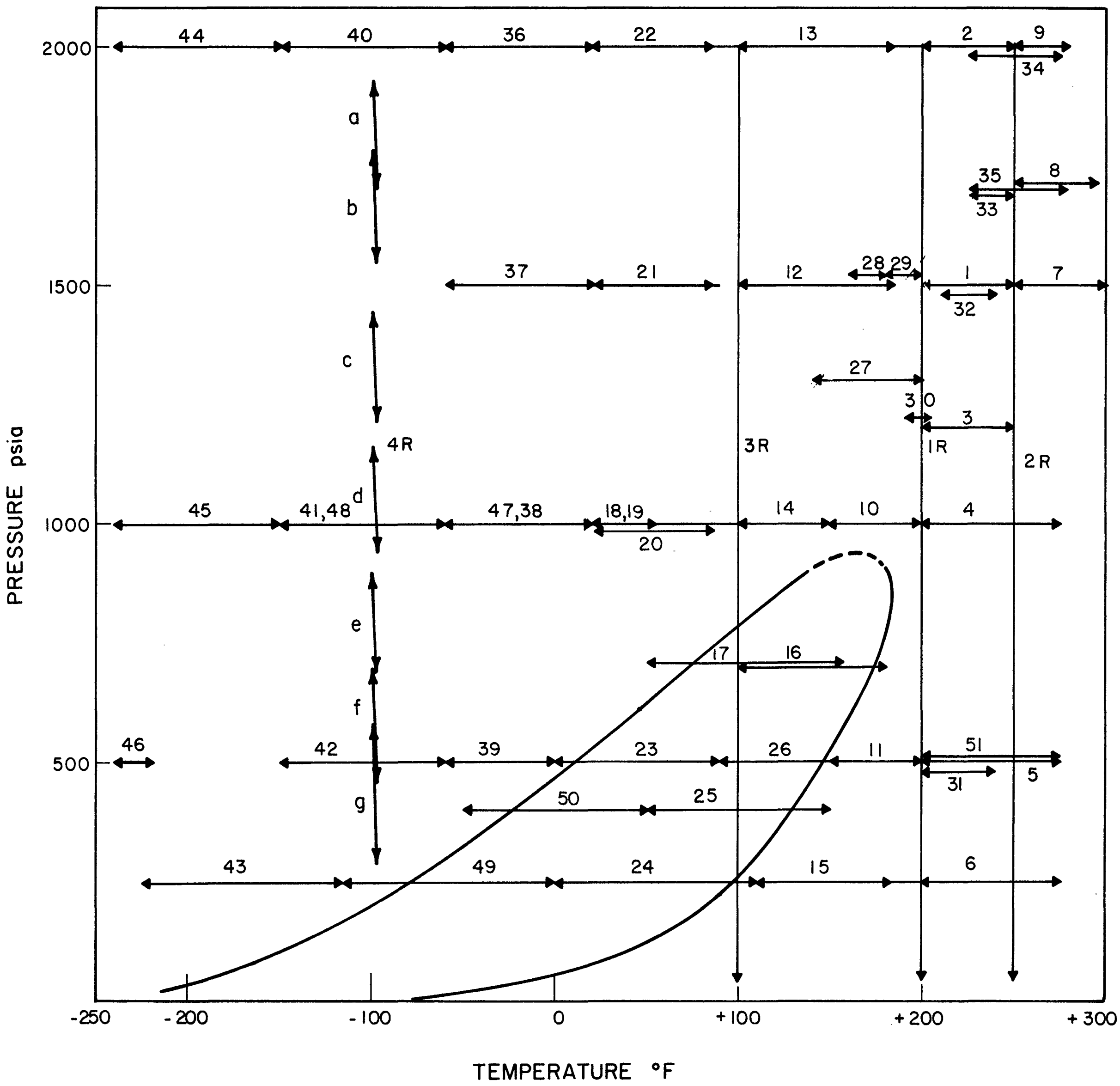


Figure 29. Temperatures and Pressures of Measurement for the Nominal 77 Percent Mixture

two-phase data are listed in Table L. The isothermal and Joule-Thomson data are presented in Table LI and Table LII, respectively.

Flowmeter Calibrations

The first three calibration runs made for the 77 percent propane in methane mixture were very successful and yielded reproducible results which are presented as the lowest curve in Figure 30. All curves illustrate that as with propane the flow is not strictly laminar in the flowmeter; if it were, the results would lie on a horizontal line on this type of plot. The calibration data for the first three calibration runs lie essentially on a single curve. Two sets of constants for Equation (76) were used to fit the calibration curve. The average deviations of the experimental calibration points from the correlating equation was ±0.16 for the 19 points in the low flow rate range and ±0.18 for the 19 points at high flow rates.

The success of these early calibrations led to a deviation from standard practice and no recalibration was made between Runs 26 to 46. During this interval of nine weeks, the calibration changed. Calibrations were then made after Run 46 and before Run 49. These calibrations are represented by the upper curve on Figure 30 and are fairly consistent (average deviation ±0.17 for the 12 low flow rate points and ±0.25 for the 13 points at high flow rate points). Runs 47 and 48 are repeats of previous runs and were made in an attempt to establish when the flowmeter calibration changed.

Following the calibration after Run 48, the flowmeter was removed from the system, cleaned ultrasonically, and recalibrated. Additional check runs were made to aid in the interpretation of the previous data and new data were obtained. Data from the last calibration agreed well

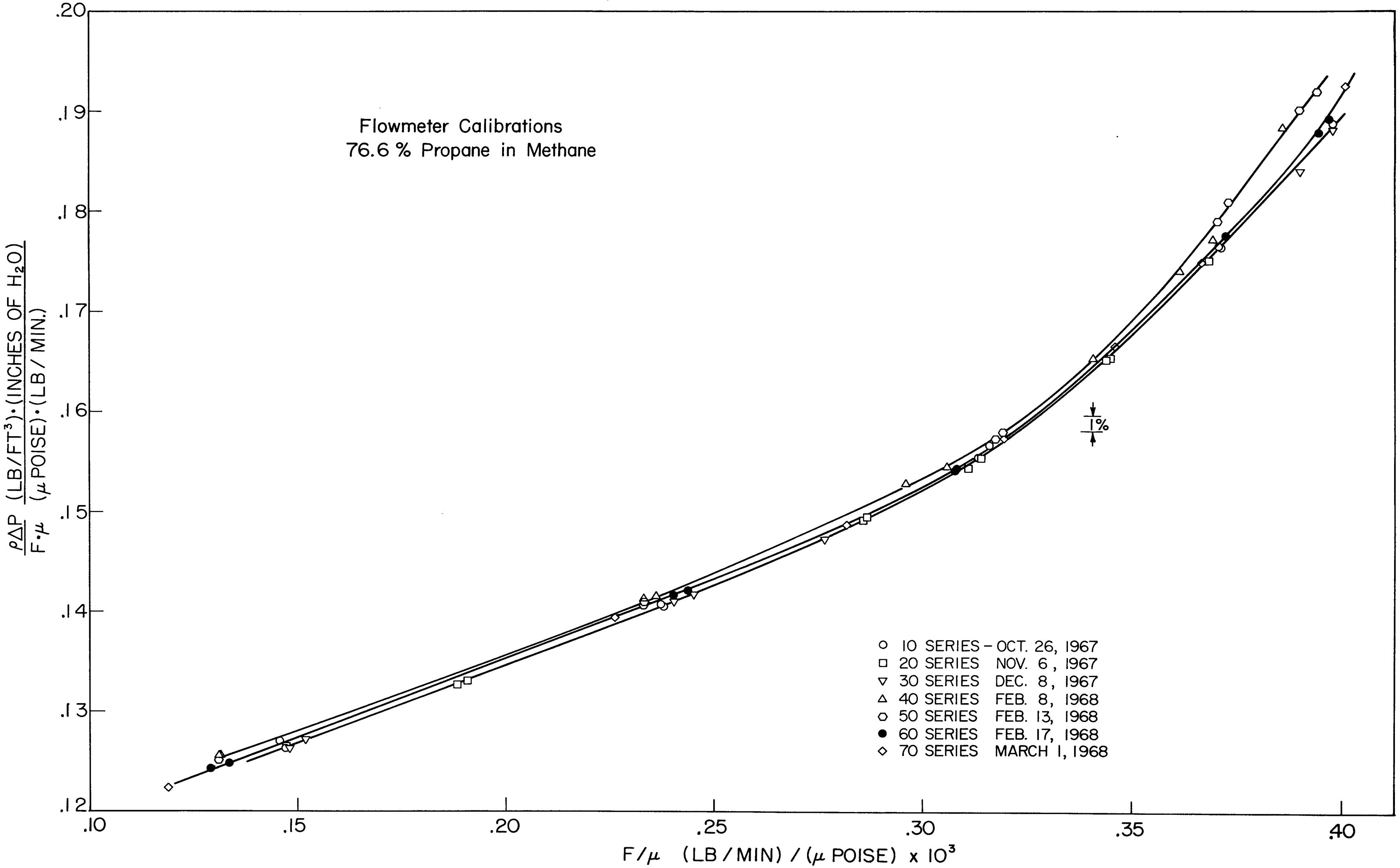


Figure 30. Results of Flowmeter Calibrations for the Nominal 77 Percent Mixture

with the one made subsequent to ultrasonic cleaning as illustrated by the middle curve on Figure 30 (average deviation ±0.08 for 10 low flow rate points and ±0.12 for 12 high flow rate points).

In using the results of flowmeter calibrations to establish the experimental values of the flow rate, results from the pair of calibration runs which bracket the experimental runs are usually used. The calibration runs used to interpret specific data for this mixture are indicated in Table XIV together with values of the total number of points included in the calibration sets and the average deviation of each set from the calibration equation. The calibration constants used in Equation (76) obtained for each set are presented in Table LII of Appendix B.

TABLE XIV

CALIBRATION DATA USED IN INTERPRETING EXPERIMENTAL RESULTS

Experimental Runs	Calibration Runs		Number of Calibration Points	Average Deviation (percent)
1 - 26, 1R - 3R	10,20,30	high	19	0.18
		low	19	0.16
27 - 47	40,50	high	13	0.25
		low	12	0.17
48,51 4R	60,70	high	12	0.16
		low	10	0.08

Check on Assumption of Adiabaticity

In applying Equation (2) to interpret experimental data, it is assumed that the calorimeter is adiabatic. It has been established for the isobaric mode[91] that this condition is satisfied if the heat capacity determined using the calorimeter is independent of the flow rate. Therefore, a series of isobaric determinations (Runs 18 and 20) was made at four different flow rates to test the assumption of zero heat leakage. As illustrated in Figure 31, the heat capacity obtained is essentially independent of flow rate within the limits of precision of the measurements (±0.3 percent).

Interpretation of Results

Isobaric

The isobaric data in the single phase region were again interpreted using Equation (2). Typical results are shown as Figure 32 on which average values of heat capacity calculated from experimental results are plotted versus temperature. Again solid lines indicate basic results obtained in accordance with the procedure described previously and dashed lines are values obtained by difference from the basic results. Point values of heat capacity were obtained in the same manner as the method used for propane, implementing both a graphical and computer technique. Figure 32 illustrates the broad maximum in the heat capacity which occurs in the region just above the critical point for the mixture, which closely resembles the curve for propane at 1000 psia (Figure 12).

A majority of the values of heat capacity, C_p, obtained from interpretation of isobaric data in the single phase region are summarized in Table XV for equal intervals of temperature and pressure.

Significant changes in the value of the heat capacity as with pure propane occur not only in the region above the critical point for the mixture but also near the two-phase locus. Figure 33 illustrates typical values of experimental data with the smoothed curve through the data representing values of the heat capacity near the two-phase region. Table XVI lists values of heat capacity in the regions of significant

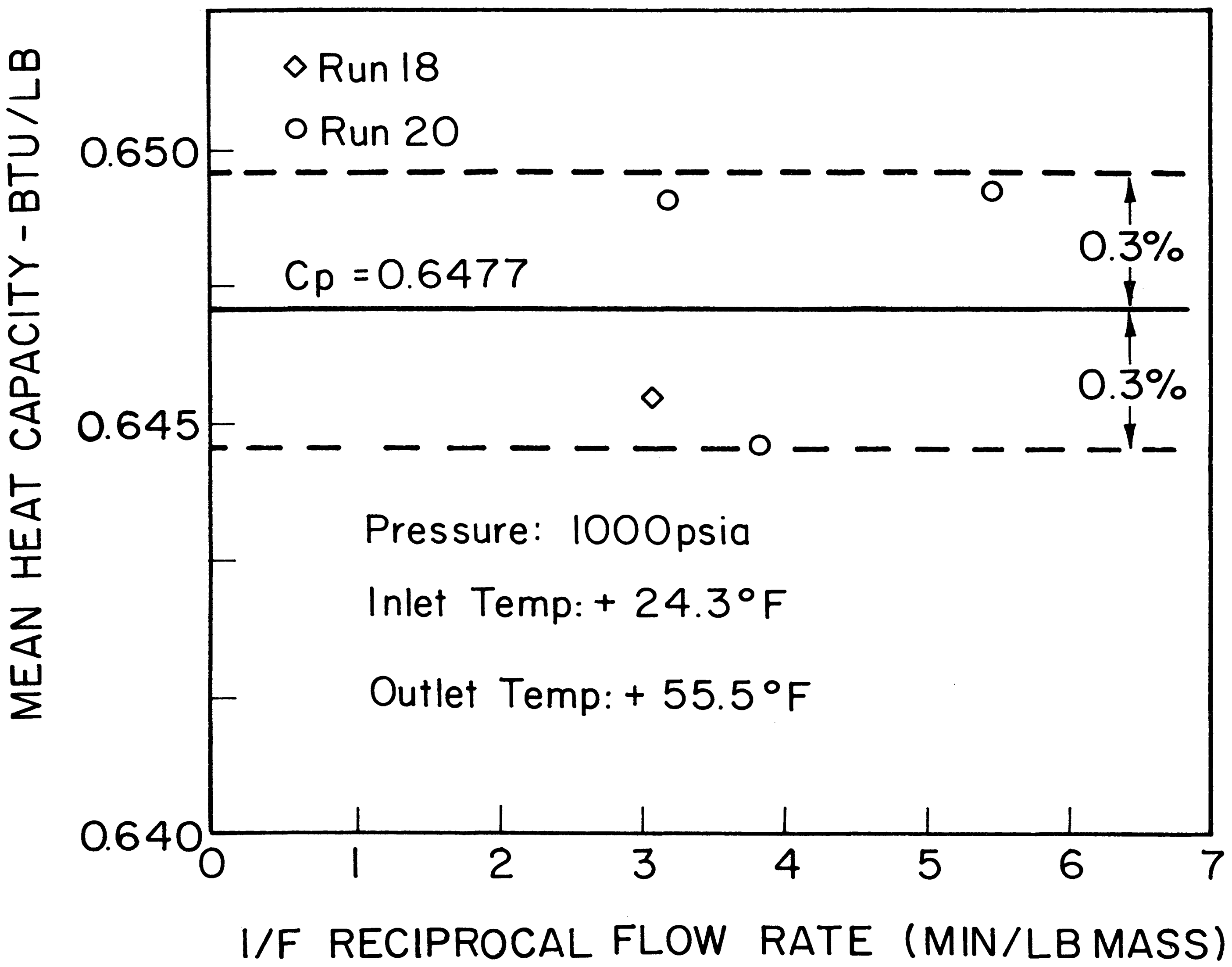


Figure 31. Heat Capacity as a Function of Reciprocal Flow Rate for the Nominal 77 Percent Mixture

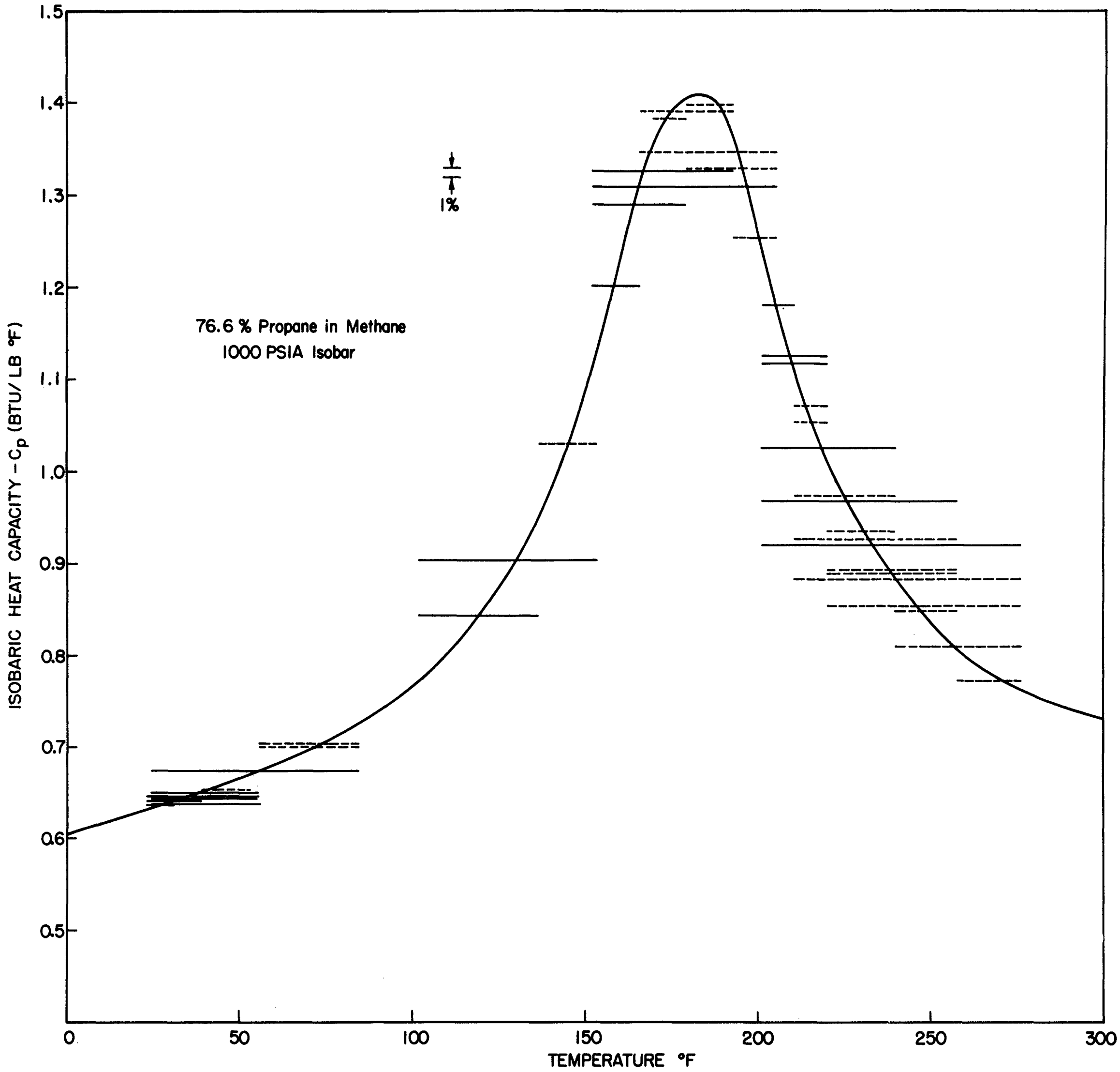


Figure 32. Isobaric Heat Capacity at 1000 psia in the Upper Temperature Range for the Nominal 77 Percent Mixture

TABLE (XV)

TABULATED VALUES OF ISOBARIC HEAT CAPACITIES

FOR A NOMINAL 77 MOL PERCENT PROPANE IN METHANE MIXTURE

C_p (Btu/lb - °F)

Temperature °F	Pressure, psia 0	250	500	1000	1500	2000
-280	0.485**	0.490	0.492	0.492	0.486	0.484
-270	0.487**		0.493	0.494	0.488	0.486
-260	0.489**	0.493	0.495	0.495	0.489	0.487
-250			0.497	0.496	0.491	0.490
-240		0.496	0.499	0.498	0.494	0.493
-230		0.498	0.501	0.500	0.497	0.495
-220		0.501	0.504	0.502	0.500	0.499
-210		0.503	0.506	0.504		0.501
-200		0.507	0.509	0.506	0.504	0.504
-190		0.510	0.512	0.509		0.506
-180		0.513	0.515	0.511	0.508	0.508
-170		0.517	0.518	0.513		0.510
-160		0.521	0.521	0.515	0.512	0.512
-150		0.525	0.524	0.518		0.514
-140		0.529	0.527	0.522	0.520	0.517
-130		0.533	0.533	0.527		0.520
-120		0.538	0.534	0.532	0.528	0.523
-110		0.543	0.538	0.537		0.527
-100		0.549	0.542	0.542	0.538	0.531
- 90		0.555	0.546	0.546		0.535
- 80		0.561	0.551	0.550	0.545	0.540

TABLE (XV) - (Cont.)

Temperature °F	Pressure, psia 0	250	500	1000	1500	2000
- 70			0.557	0.555		0.546
- 60			0.564	0.559	0.553	0.551
- 50			0.572	0.564	0.558	0.557
- 40			0.580	0.570	0.565	0.563
- 30			0.588	0.576	0.571	0.569
- 20			0.597	0.584	0.577	0.576
- 10			0.607	0.593	0.584	0.583
0			0.617	0.604	0.592	0.591
10				0.616	0.601	0.597
20				0.628	0.611	0.604
30				0.640	0.621	0.611
40				0.652	0.630	0.619
50				0.665	0.642	0.627
60				0.680	0.653	0.636
70				0.696	0.664	0.645
80				0.716	0.676	0.655
90				0.738	0.689	0.665
100	0.425*	0.601		0.766	0.704	0.676
110	0.431*	0.567		0.801	0.721	0.687
120	0.437*	0.541		0.847	0.744	0.700
130	0.443*	0.530		0.904	0.768	0.714
140	0.449*	0.525		0.980	0.794	0.728
150	0.455*	0.523	0.738	1.088	0.819	0.743

TABLE (XV) - (Cont.)

Temperature °F	Pressure, psia					
	0	250	500	1000	1500	2000
160	0.461*	0.523	0.718	1.233	0.845	0.758
170	0.467*	0.523	0.679	1.360	0.870	0.773
180	0.473*	0.523	0.652	1.407	0.899	0.786
190	0.479*	0.525	0.633	1.388	0.922	0.800
200	0.485*	0.528	0.621	1.259	0.948	0.813
210	0.491*	0.531	0.613	1.115	0.976	0.827
220	0.497*	0.535	0.608	1.011	1.001	0.840
230	0.503*	0.539	0.603	0.940	1.002	0.854
240	0.509*	0.544	0.601	0.884	0.984	0.868
250	0.514*	0.548	0.601	0.839	0.960	0.881
260	0.520*	0.553	0.602	0.801	0.940	0.886
270	0.526*	0.559	0.604	0.776	0.916	0.875
280	0.532*	0.565	0.606	0.756	0.896	0.855
290	0.538*	0.571	0.609	0.741	0.879	0.834
300	0.543*	0.577	0.612	0.729	0.864	0.812

*Ideal Gas Values of Rossini et. al. (118)

**Experimental Data of Cutler and Morrison (23)

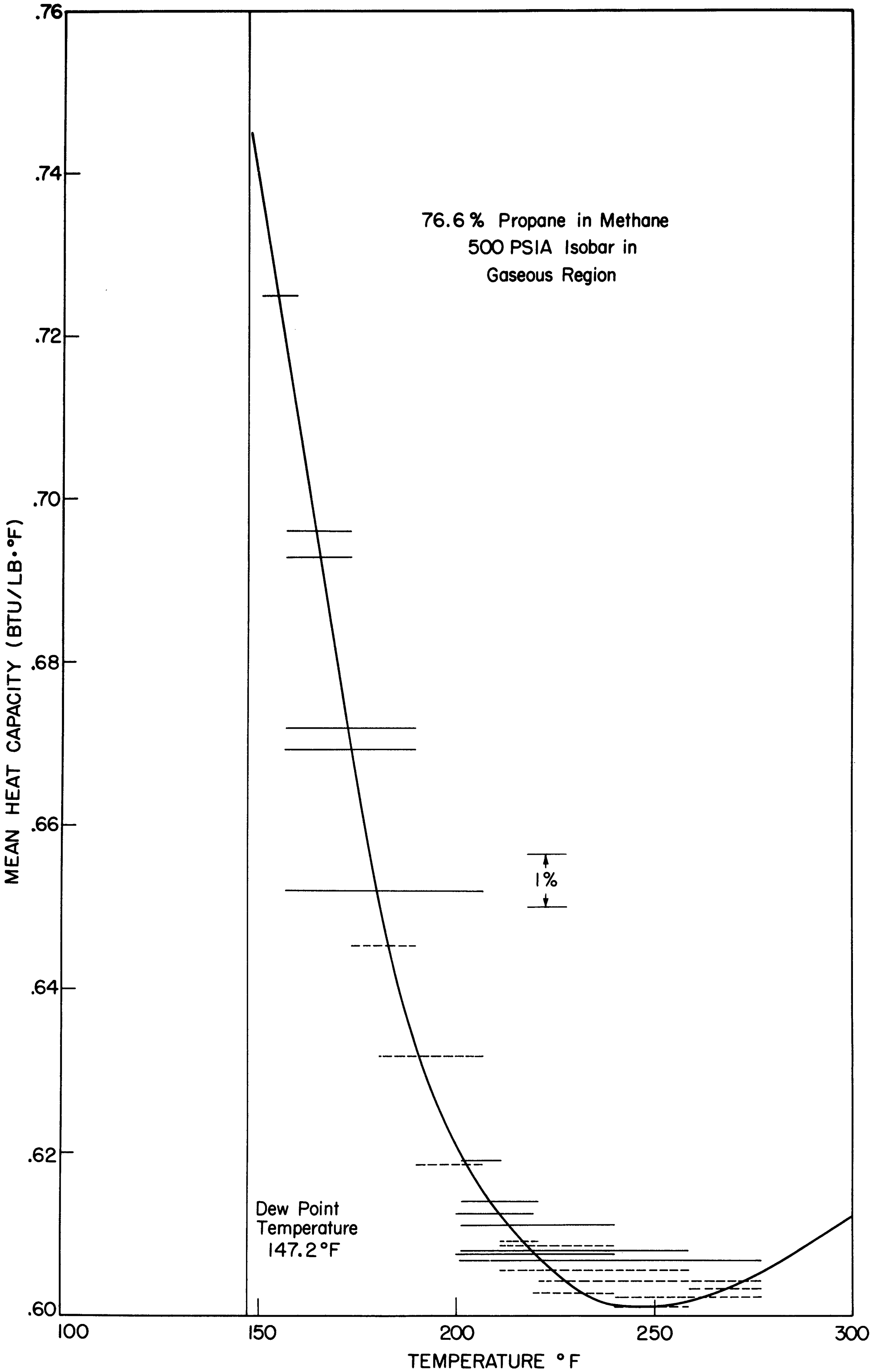


Figure 33. Isobaric Heat Capacity at 500 psia in the Gaseous Region for the 77 Percent Mixture

TABLE XVI

Supplemental Table of Experimental Values of Isobaric Heat Capacity, C_p (Btu/lb)

Pressure (psia)										
	Temperature (°F)									
400	-45	-40	-35	-30	-25	-24.0	131.2	135	140	145
	C_p (Btu/lb)									
	0.579	0.586	0.594	0.601	0.609	0.610	0.736	0.701	0.655	0.610
	Temperature (°F)									
700	50	55	60	65	70	75	170.1	175	180	185
	C_p (Btu/lb)									
	0.678	0.688	0.698	0.708	0.719	0.729	1.276	1.153	1.028	0.906
	Temperature (°F)									
1200	190	200	210	220	230	240	245	250	255	260
	C_p (Btu/lb)									
	1.169	1.166	1.153	1.118	1.061	0.983	0.943	0.903	0.834	0.764
	Temperature (°F)									
1700	220	225	230	235	240	245	250	260	270	280
	C_p (Btu/lb)									
	0.915	0.925	0.933	0.937	0.936	0.933	0.927	0.915	0.901	0.889

change such as near the maxima in the heat capacity and near the saturation curves.

The results of all heat capacity determinations are summarized on Figure 34.

A typical enthalpy traverse of the two-phase region at constant pressure is illustrated in Figure 35. Note that the traverse was made as two runs. Run 17 had an inlet temperature of about 50°F and was terminated with the two-phase region at about 140°F. In Run 16 a two-phase mixture at about 100°F was fed to the calorimeter and the run was terminated after the exiting fluid was totally vaporized at about 190°F. The results of the two runs are consistent in the region of overlap as illustrated in Figure 35. This procedure was followed so that enthalpy traverses could be made across the two-phase regions at larger flow rates than would be possible if the entire change were experienced in one run. This is necessary due to a limitation on the power available from the constant voltage power supply at low temperatures and the possibility of overheating and burning up the insulation of the nicrome heating wire in the calorimeter capsule at higher temperatures.

Determination of the points of discontinuity in slope of the curve yields values of the bubble point and dew point. For mixtures containing a majority of propane it is relatively difficult to determine the bubble point whereas the dew point is the more difficult determination for mixtures containing high mole fractions of methane.[81] Experimentally determined values of the isobaric enthalpy change on vaporization are listed in Table XVII together with the experimentally determined values of the bubble point and the dew point for the mixture.

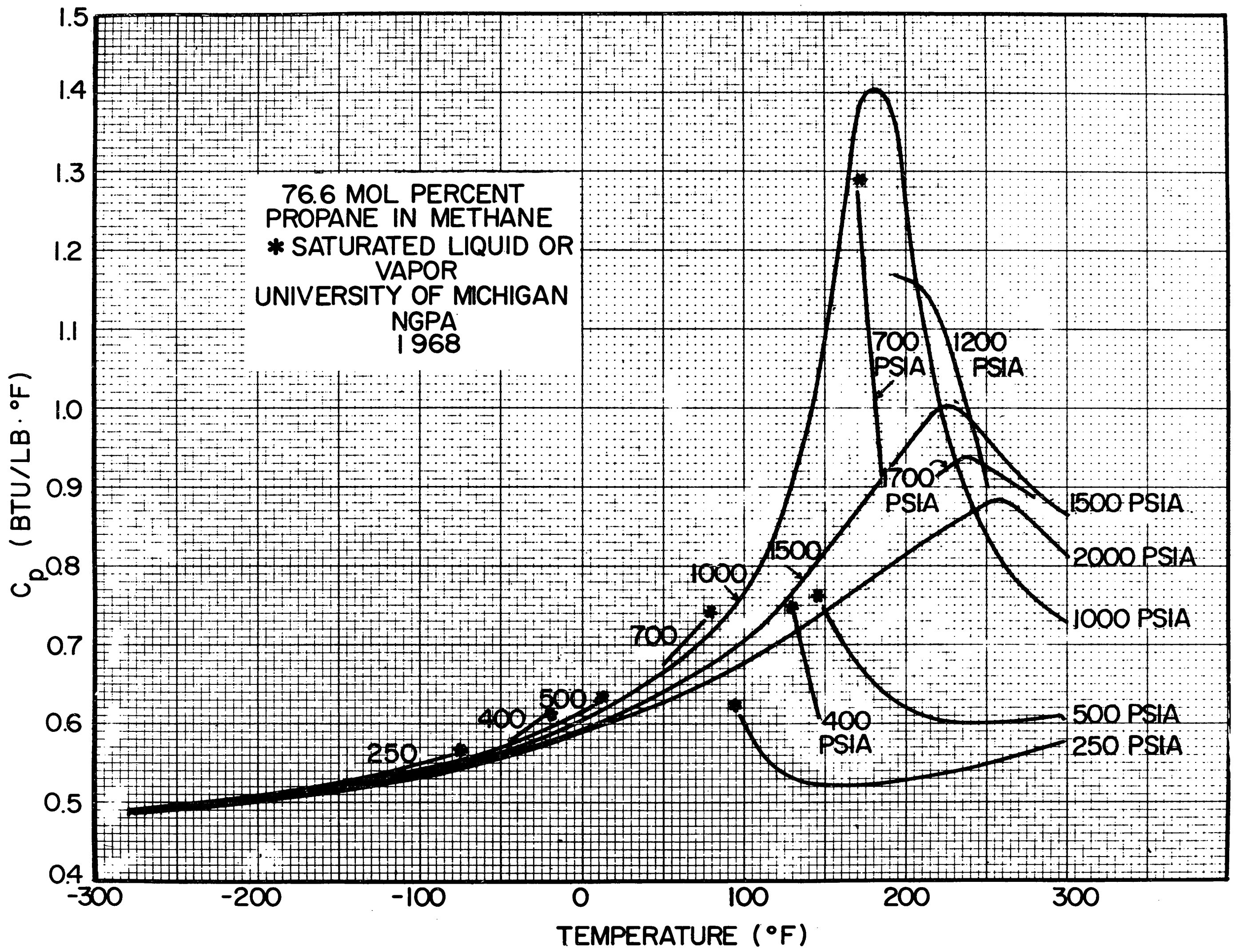


Figure 34. Isobaric Heat Capacity for the Nominal 77 Percent Mixture

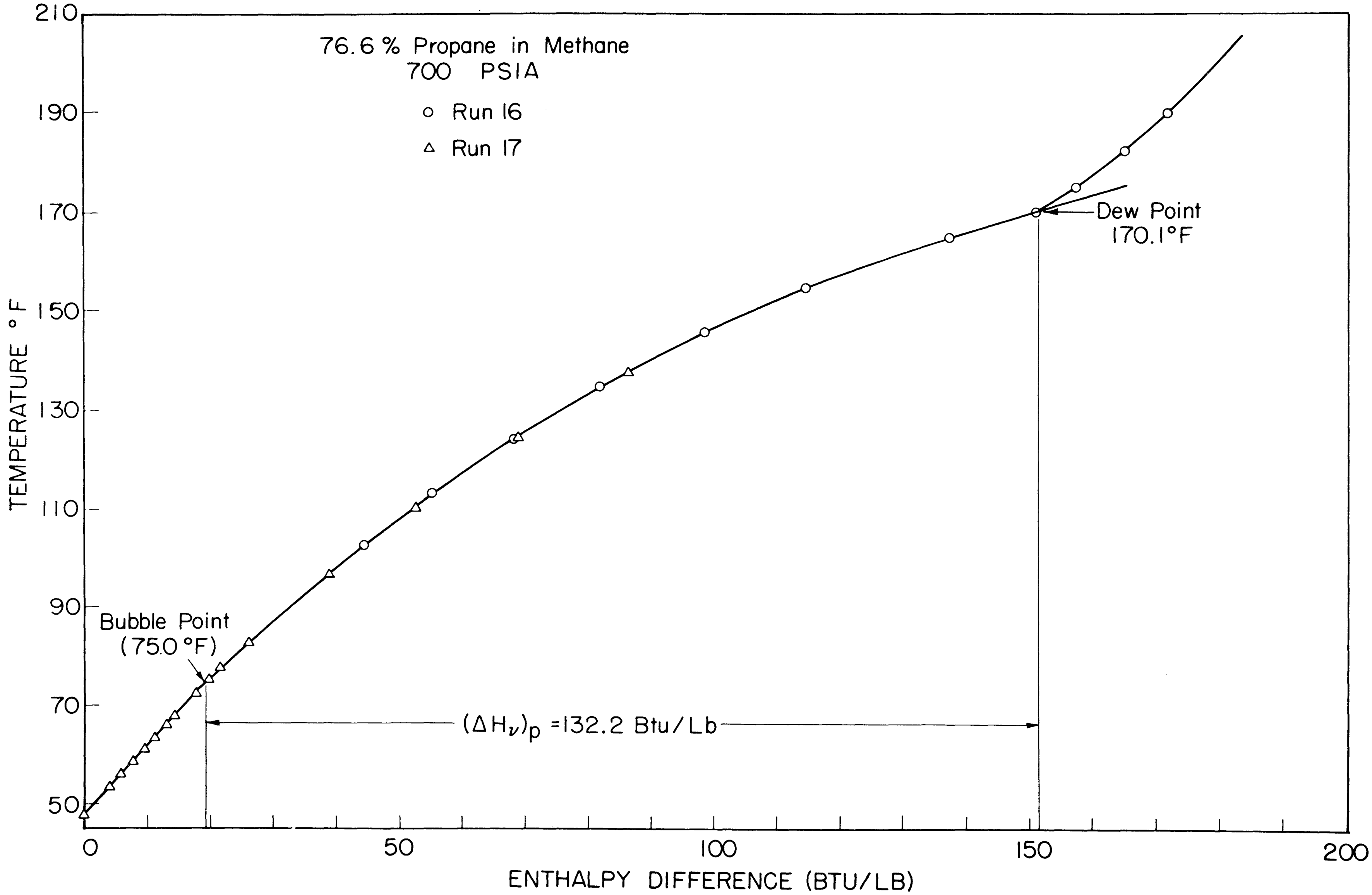


Figure 35. Enthalpy Differences for the 77 Percent Mixture in the Two-Phase Region

TABLE XVII

Properties of the Nominal 77 Mole Percent
Propane in Methane Mixture at the Phase Boundaries

Pressure (psia)

	250		400		500		700	
$(\Delta \underline{H}_{vap})_P$ (Btu/lb)	232.1		208.0		183.2		132.2	
Phase Boundaries	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)
This Investigation	-81.0	95.7	-24.0	131.2	12.2	147.0	75.0	170.1
Akers, Burns and Fairchild(1)	-73.2	--	-23.5	--	11.0	--	--	--
Price and Kobayashi (109)	-79.0	--	-21.0	--	11.7	--	--	--
Reamer, Sage and Lacey (111)	--	100.8	--	131.0	--	147.1	75.8	173.0
Sage, Lacey and Schaafsma (126)	--	97.0	--	132.0	--	148.0	64.0	175.8

Isothermal

The isothermal data obtained in the single phase region were interpreted in accordance with Equation (3) and plots made of the average value of the isothermal throttling coefficient as a function of pressure. Typical results are shown in Figure 36. Point values of the isothermal coefficient were obtained graphically.

As the lower limit on pressure is about 100 psia, it was necessary to estimate $\varphi = f(P)$ at low pressures. To aid in this estimation, values were calculated using the BWR equation of state with the original constants for methane and propane[7] together with mixing rules as originally suggested.[8] Typical results are presented as a center line on Figure 36. In addition, Equations (32) to (34) were used with published values of the second virial coefficient for methane and propane and the interaction term[58] to estimate φ. The resulting value at 201°F is plotted on Figure 36. As another check, PVT data for the mixture[111] as interpreted[37] using Equation (24) yielded values of $\left[\frac{(\underline{H}^0 - \underline{H})}{P}\right]_T$ between zero pressure and 200 psia. A typical value is presented as a dashed line on Figure 36. A solid line (such as shown on Figure 36) was drawn to be consistent with the data obtained at elevated pressures and the estimates based on data from the literature in the low pressure range. Values of the isothermal throttling coefficient, φ, obtained by interpreting the data (including values of φ^0 established as outlined above) are reported for each of the experimental isotherms in Table XVIII. These values are summarized in Figure 37. As previously mentioned one experimental isothermal run was made across the two-phase region. The results of this run are presented in Figure 38. Breaks in the curve indicate both the upper pressure at which vaporization started and the lower pressure

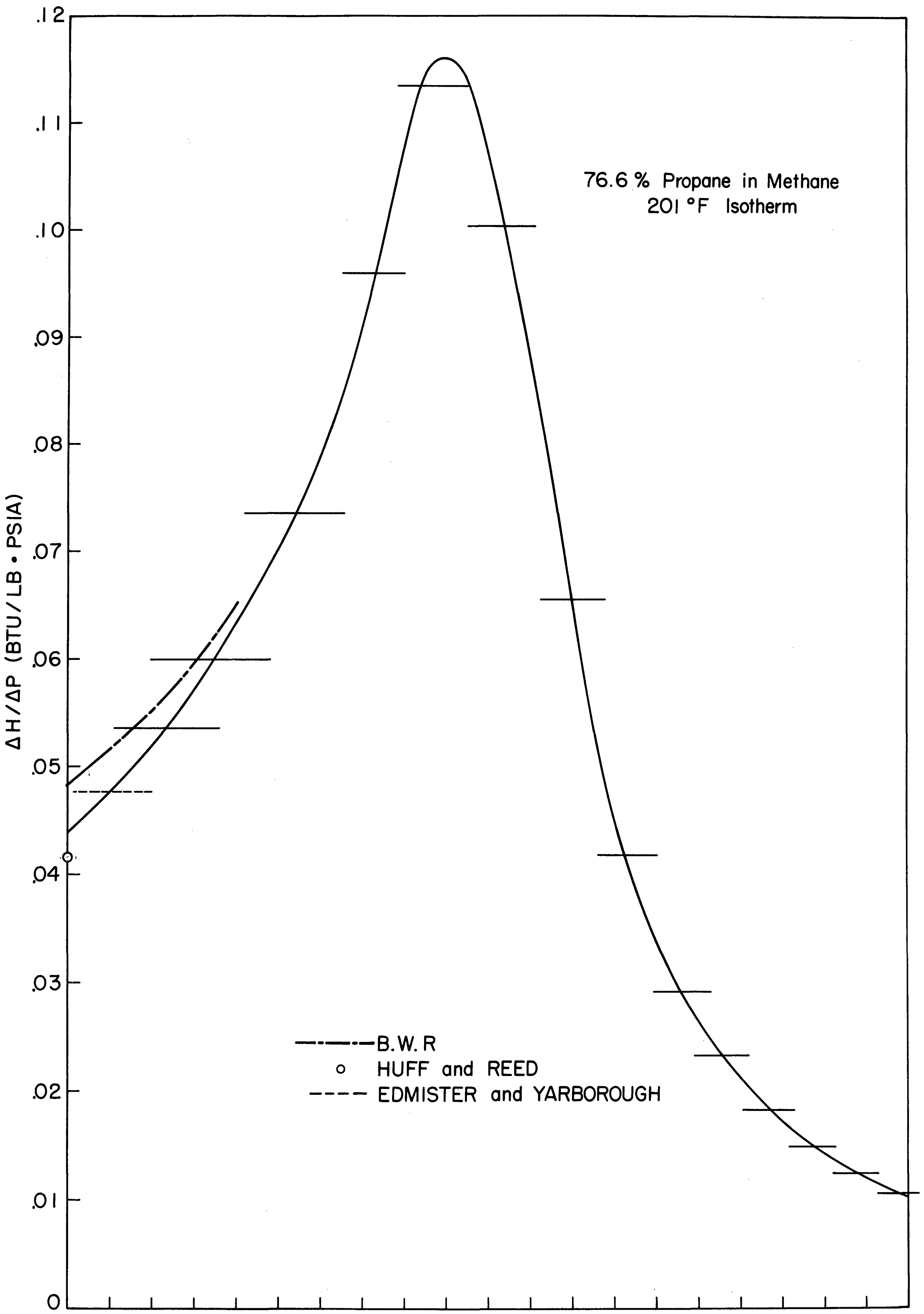


Figure 36. Isothermal Throttling Coefficient for the 77 Percent Mixture at 201°F

TABLE XVIII

Experimental Values of the Isothermal Throttling Coefficient, Ø, for a Nominal 77 Mole Percent Propane in Methane Mixture

Ø x 10^2 (Btu/lb. psia)

Pressure	Temperature -°F			
psia	-96.8[a]	99.9	201.0	251.0
0[b]	--	--	-4.39	-3.76
100	--	--	-4.76	-4.03
200	--	--	-5.19	-4.30
300	+0.301	--	-5.70	-4.56
400	+0.302	--	-6.33	-4.83
500	+0.303	--	-7.01	-5.11
600	+0.304	--	-7.86	-5.43
700	+0.304	--	-9.13	-5.79
800	+0.305	-0.734	-10.81	-6.18
900	+0.306	-0.596	-11.60	-6.58
1000	+0.307	-0.483	-10.66	-6.77
1100	+0.308	-0.396	-8.77	-6.70
1200	+0.308	-0.325	-6.45	-6.36
1300	+0.309	-0.265	-4.54	-5.89
1400	+0.310	-0.214	-3.39	-5.16
1500	+0.311	-0.170	-2.62	-4.39
1600	+0.313	-0.130	-2.10	-3.70
1700	+0.316	-0.101	-1.72	-3.05
1800	+0.321	-0.071	-1.42	-2.53
1900	+0.328	-0.044	-1.19	-2.22
2000	+0.339	-0.017	-1.00	-1.98

a Calculated using Equation (8)

b Extrapolation to zero pressure based primarily on PVT data

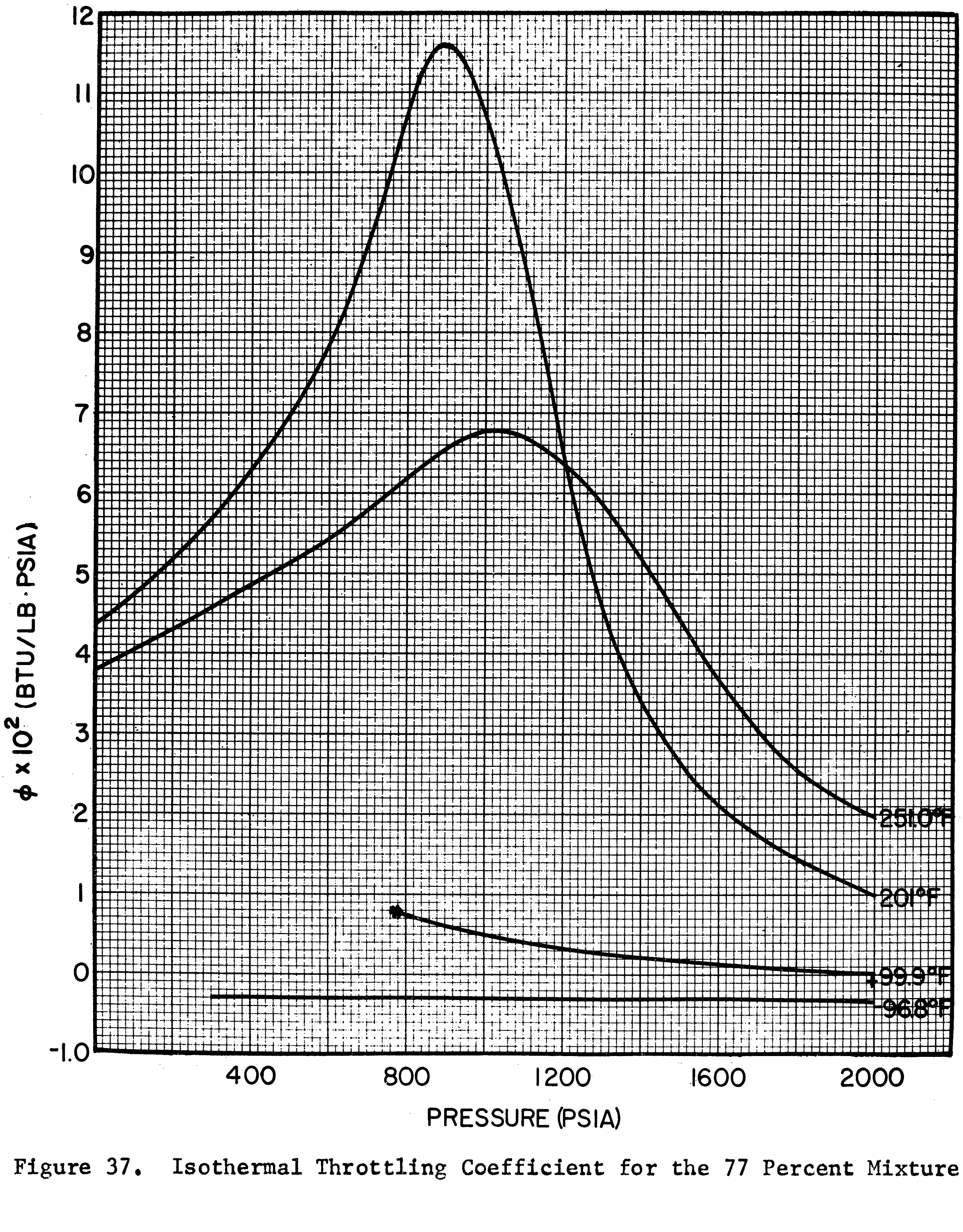


Figure 37. Isothermal Throttling Coefficient for the 77 Percent Mixture

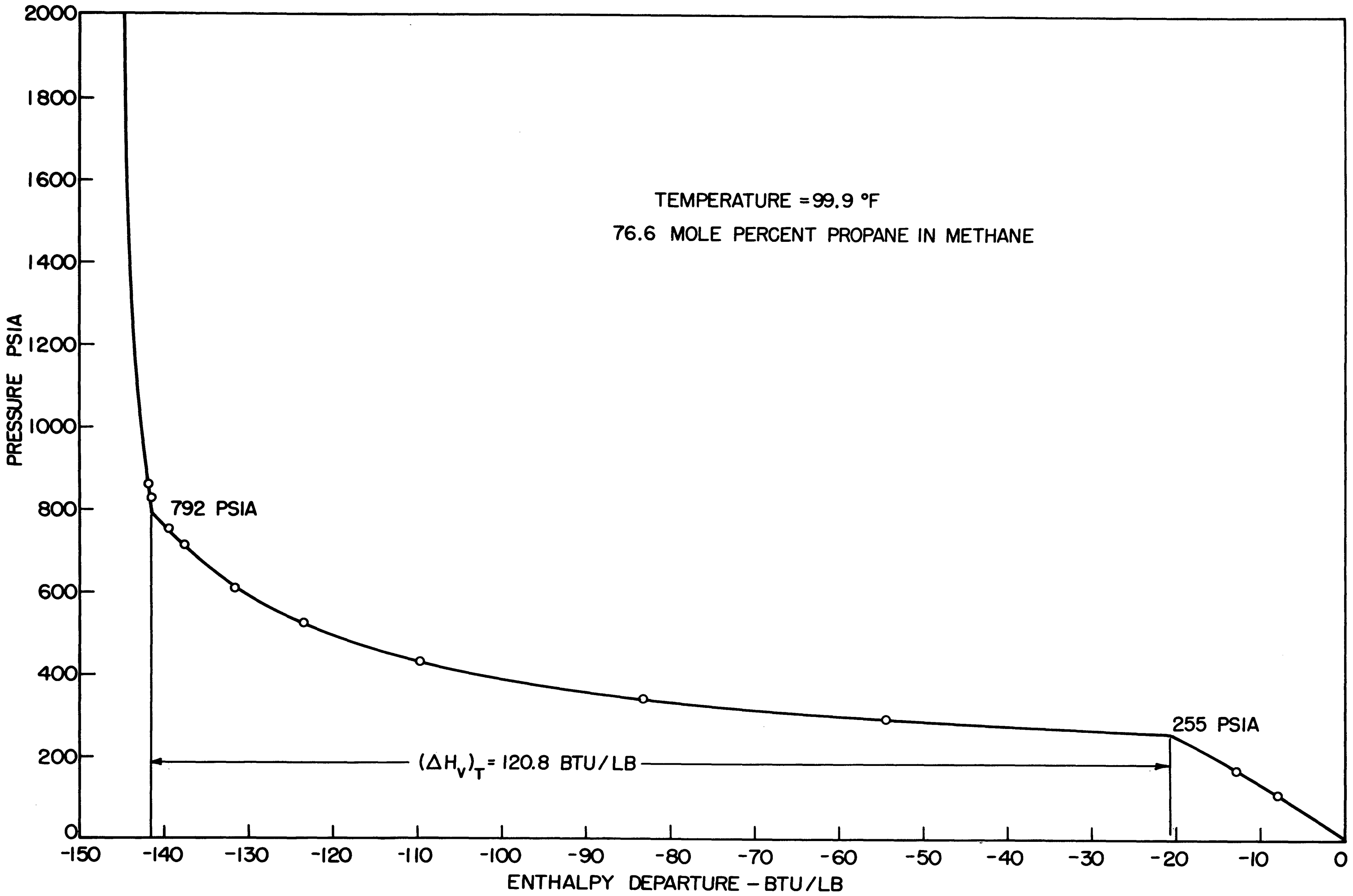


Figure 38. Isothermal Enthalpy Differences Through the Two-Phase Region at 100°F for the 77 Percent Mixture

at which vaporization was complete.

Isenthalpic

Seven determinations were made with the throttling calorimeter under conditions such that a drop in pressure resulted in an increase in temperature (Run 4R on Figure 29). The inlet temperature was constant at -96.8°F in all cases. The data as interpreted using Equations (4) and (11) are illustrated in Figure 39. The values of the Joule-Thomson coefficient determined from these data are summarized in Table XIX. These data are used in conjunction with Equation (17) to obtain isothermal enthalpy differences.

The 50.6 Mole Percent Propane in Methane Mixture

After completion of the 76.6 mole percent propane in methane mixture the fluid in the system was diluted further with methane to obtain a 50.6 mole percent propane in methane mixture.

Composition of Gas

The composition of the system as obtained by a chromatographic analysis is reported in Table XX. Again any small changes in composition which occurred were balanced by adding quantities of the deficient component. The chromatographic analyses on a day to day basis are summarized in Figure 40.

Regions of Measurement

The ranges of pressures and temperatures covered in the experimental investigation of this mixture are indicated by lines drawn on a pressure versus temperature diagram in Figure 41. The isobaric results for individual runs in the single and two-phase region are given in Table LIV of Appendix B. The isothermal and Joule-Thomson data are presented

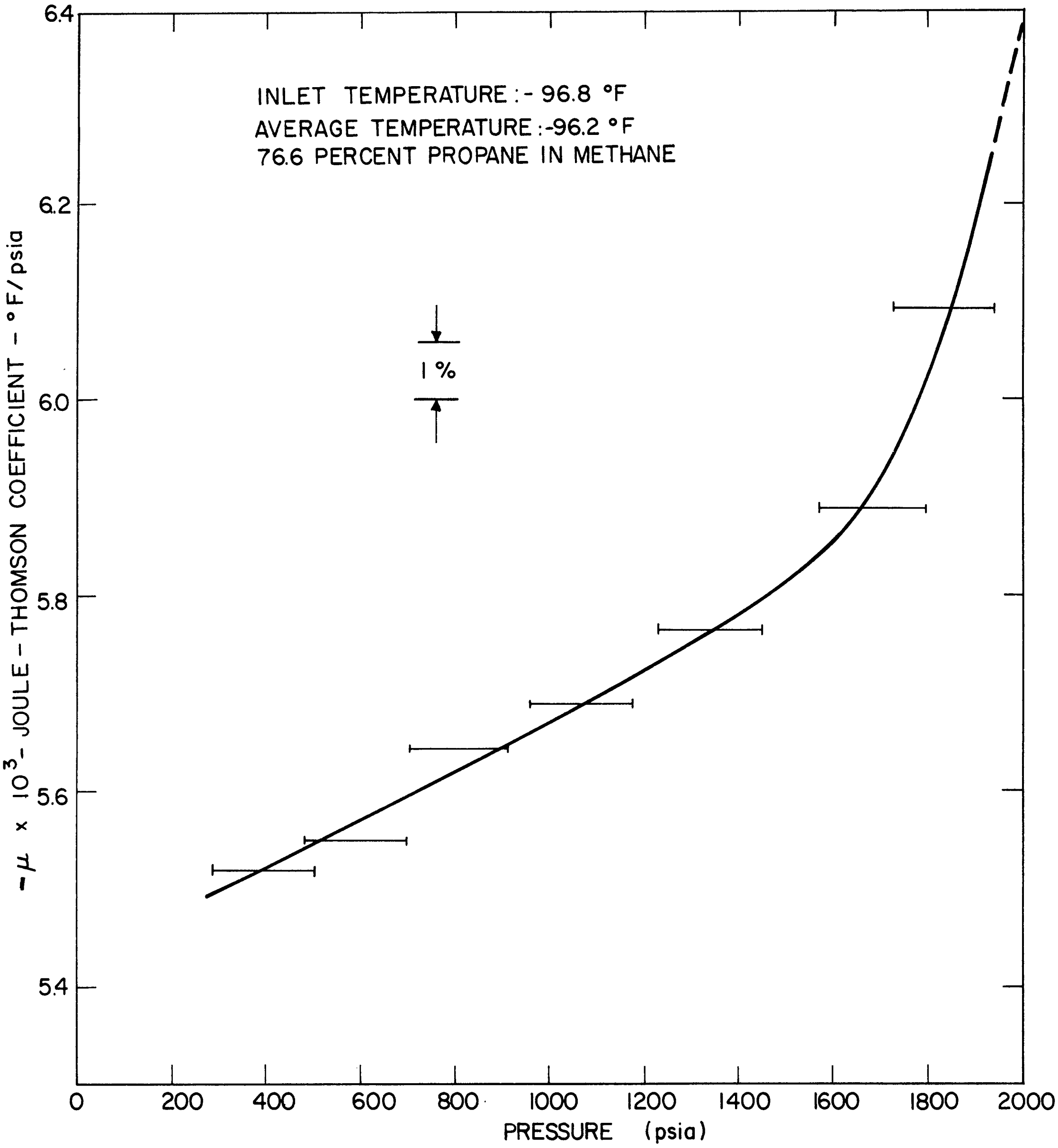


Figure 39. Joule-Thomson Coefficient for the 77 Percent Mixture at -96.2°F

TABLE XIX

Experimental Values of the Joule-Thomson Coefficient, μ, at -96.2°F, for the Nominal 77 Mole Percent Propane in Methane Mixture

Pressure psia	$-\mu \times 10^3$ °F/psi
300	5.50
400	5.52
600	5.57
800	5.62
1000	5.67
1200	5.72
1400	5.78
1600	5.85
1800	6.02

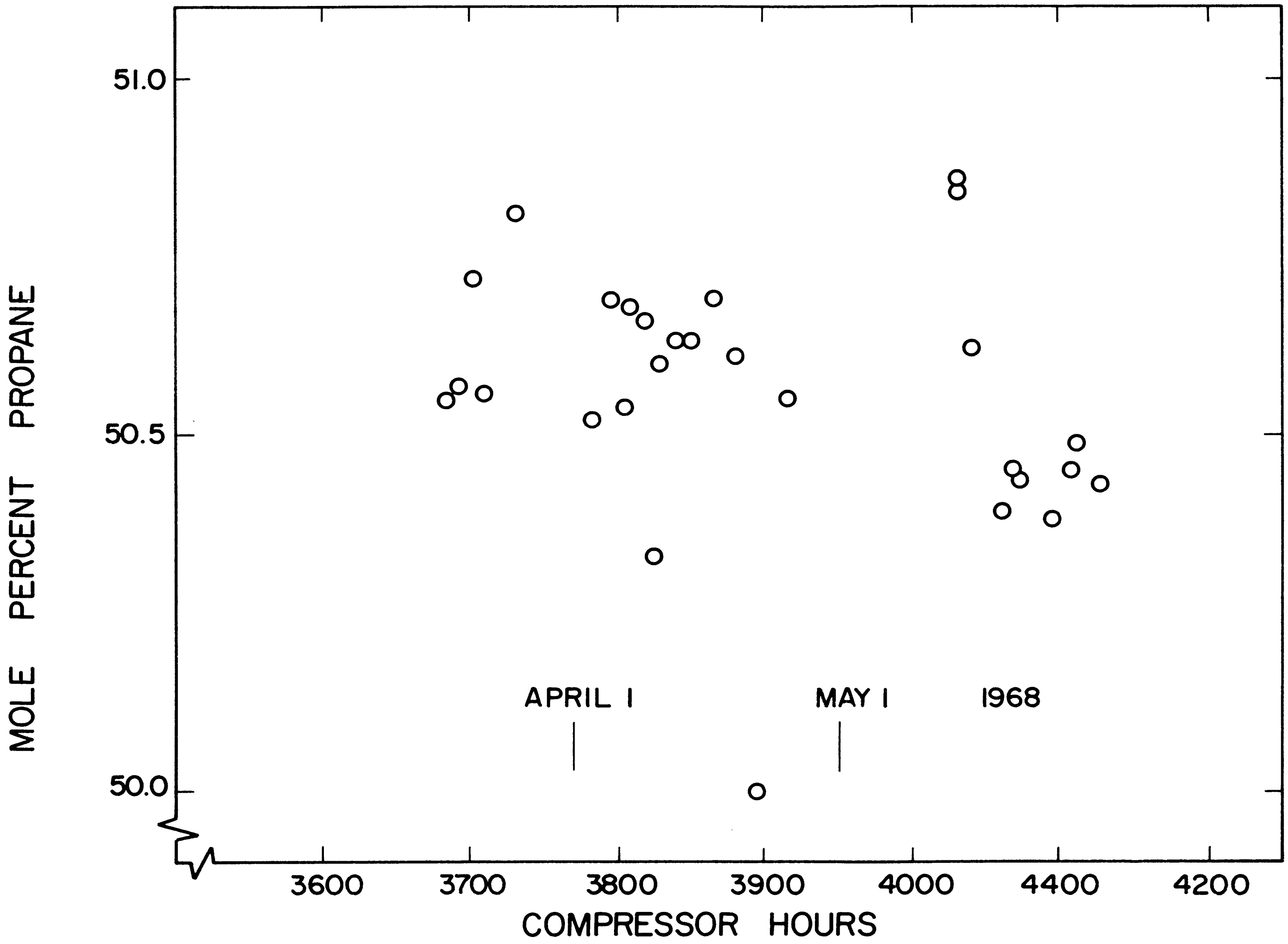


Figure 40. Composition of the Nominal 51 Percent Mixture as a Function of Time

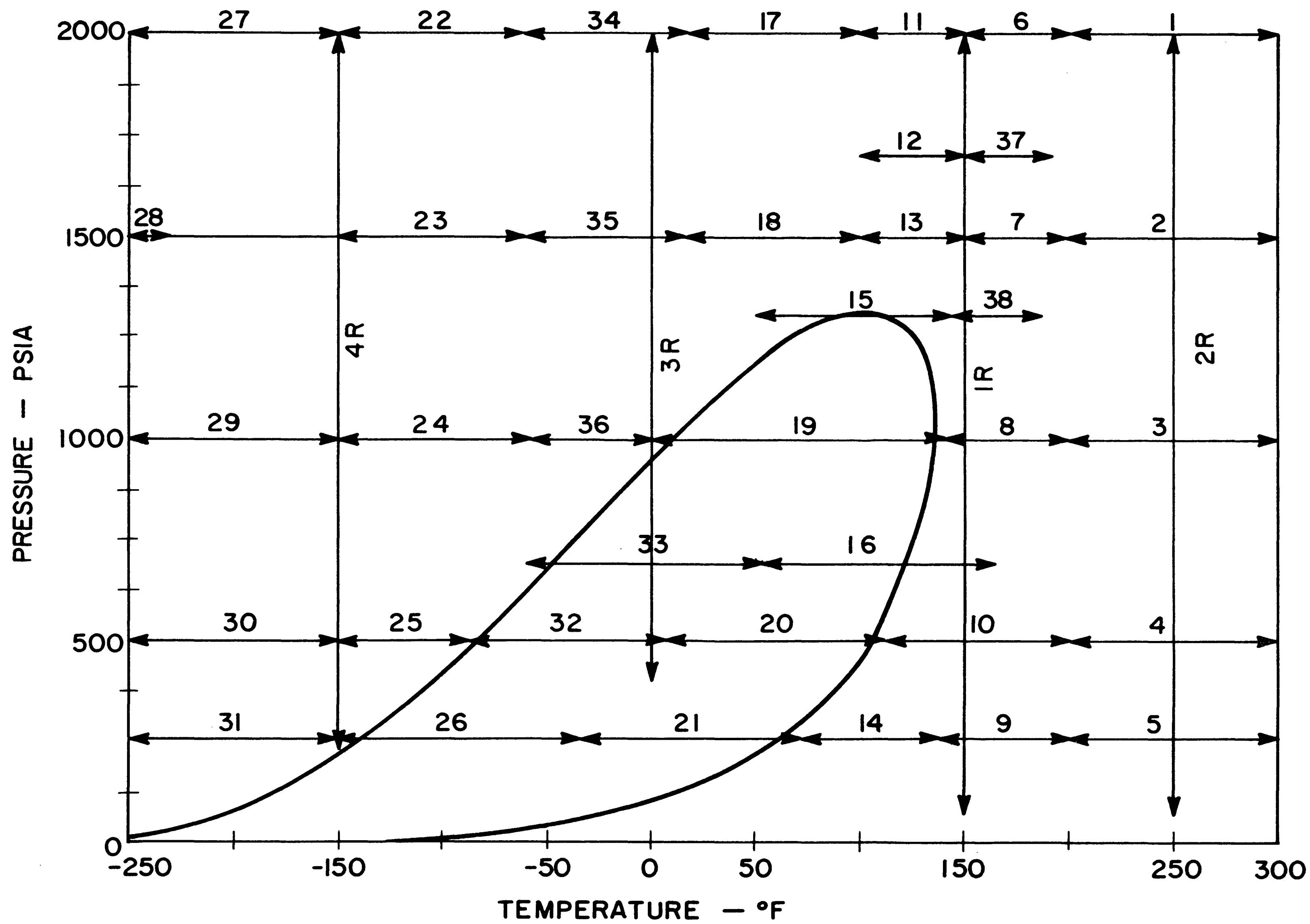


Figure 41. Temperatures and Pressures of Measurement for the Nominal 51 Percent Mixture

in Tables LV and LVI, respectively.

TABLE XX

COMPOSITION OF NOMINAL 51 MOLE PERCENT PROPANE IN METHANE MIXTURE AS DETERMINED BY CHROMATOGRAPHIC ANALYSES

	Mole Fraction
Nitrogen	<0.05
Methane	49.4
Ethane	<0.05
Carbon Dioxide	<0.05
Propane	50.6
Butane	<0.05
	<100.2

Flowmeter Calibrations

Again there were shifts in flowmeter calibration from series to series. The first two calibrations made were very successful and yielded results lying essentially on a single curve. These results are illustrated as the solid line on Figure 42. Again the flow is not laminar in the flowmeter. The average deviations of the experimental calibrations points from the correlating equation for these two runs is ±0.20 percent for the 20 experimental points.

The third calibration was found to differ by about 0.5 percent from the results of the first two. The curve for this calibration is shown as a center line of Figure 42. The flowmeter was removed from the system and ultrasonically cleaned. In addition the entire section around the flowmeter was cleaned. The flowmeter was replaced and three sets of calibrations made. These gave reproducible results. The results are represented by the dashed line on Figure 42. The increased curvature of this last line made it necessary to fit the single curve with two

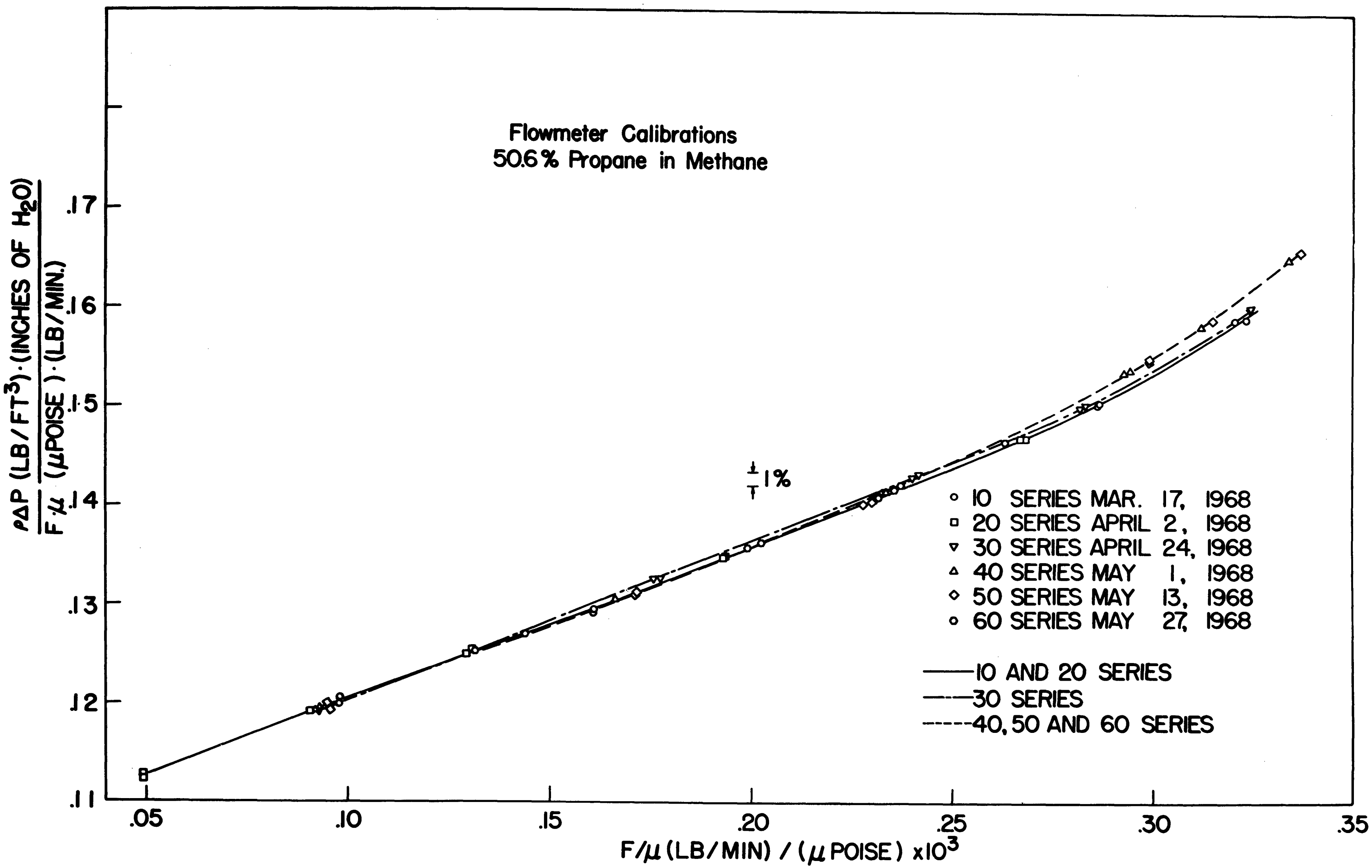


Figure 42. Results of Flowmeter Calibrations for the Nominal 51 Percent Mixture

calibration equations, one for low and the other for high flow rates. The average deviations of the experimental calibrations points from the correlating equation for these three runs is ±0.13 percent for 17 points in the low flow rate region and ±0.18 percent for the 20 points at higher flow rates.

In using the results of flowmeter calibrations to establish the experimental values of the flow rate, results are used which most adequately represent the flow rate at the time of an experimental run. In cases where the flowmeter calibrations changed between runs (in this case between series 20 and 30 or runs 10 and 21) a somewhat arbitrary decision must be made in order to determine when the calibration most likely changed. This is accomplished by carefully investigating the interpreted experimental runs in light of all the data obtained. The calibrations runs used to interpret specific data for this mixture are indicated in Table XXI, together with values of the total number of points for each set from the calibration equation. For the case of this mixture the correlating equations do not yield identical values of the flow rate in the region of overlap. For example, Runs 37 and 38 were obtained in the region of overlap and both correlating equations were used to calculate the flow rate. Table XXII shows the results of calculating flow rates by both equations for these runs. It shows that the effect of the choice of equation in this case is not insignificant (0.3 percent). All of the flowmeter calibration constants used in Equation (76) are presented in Table LVII of Appendix B.

After completion of the three mixtures it now appers that the flowmeter is one of the main sources of uncertainty and trouble in the modified recycle system. It is affected by both physical upsets and impurities.

TABLE XXI

Calibration Data Used in
Interpreting Experimental Results

Experimental Runs	Calibration Runs	Number of Calibration Points	Average Deviation (percent)
1 - 21	10,20		
1 - 3R		20	.20
22 - 38	40,50,60	Low: 17	.13
4R		High: 20	.18

TABLE XXII

Effect of Calibration Equation
on Isobaric Heat Capacity Results

Run	Pressure (psia)	Inlet Temperature (°F)	Outlet Temperature (°F)	C_p(Btu/lb°F) High Flow Equation	C_p(Btu/lb°F) Low Flow Equation
37	1700	131.0	144.6	0.965	0.965
		131.1	158.3	0.961	0.962
		131.2	172.0	0.964	0.963
38	1300	131.2	144.8	1.088	1.085
		131.4	158.7	1.050	1.047
		131.3	171.7	1.017	1.014

The flowmeter calibration can be changed by either small amounts of oil or solid materials, overpressurization, removing it from the system for any length of time, and mechanical handling.

Check on Assumption of Adiabaticity

For this mixture a special flowmeter calibration was made to increase the range of possible flow rates in order to test the assumption of adiabaticity. Normally, the range of possible flow rates is determined by the pressure drop across the water manometer. The reading is usually allowed to vary between 2 and 10 inches of water. For this test a special calibration was made with only 1 inch of water pressure drop. This essentially doubles the range of reciprocal flow rate. Four of the points including the point at highest flow rate and the one at lowest flow rate are plotted on Figure 43. A fifth point differed significantly from the other four indicating the probability of a recording error and is not plotted. The figure as well as the results for the 77 percent mixture indicates that the heat capacity obtained is essentially independent of flow rate within the limits of precision of the measurements (±0.3 percent). These results are consistent with those of other studies of this effect.[80,85]

Interpretation of Results

Isobaric

The isobaric data for this mixture in the single phase region were interpreted in the same way as the data for propane and the 76.6 mole percent mixture. Results right above the critical region at 1500 psia are presented in Figure 44. This curve illustrates the broad maximum which occurs in the heat capacity in the region right above the critical

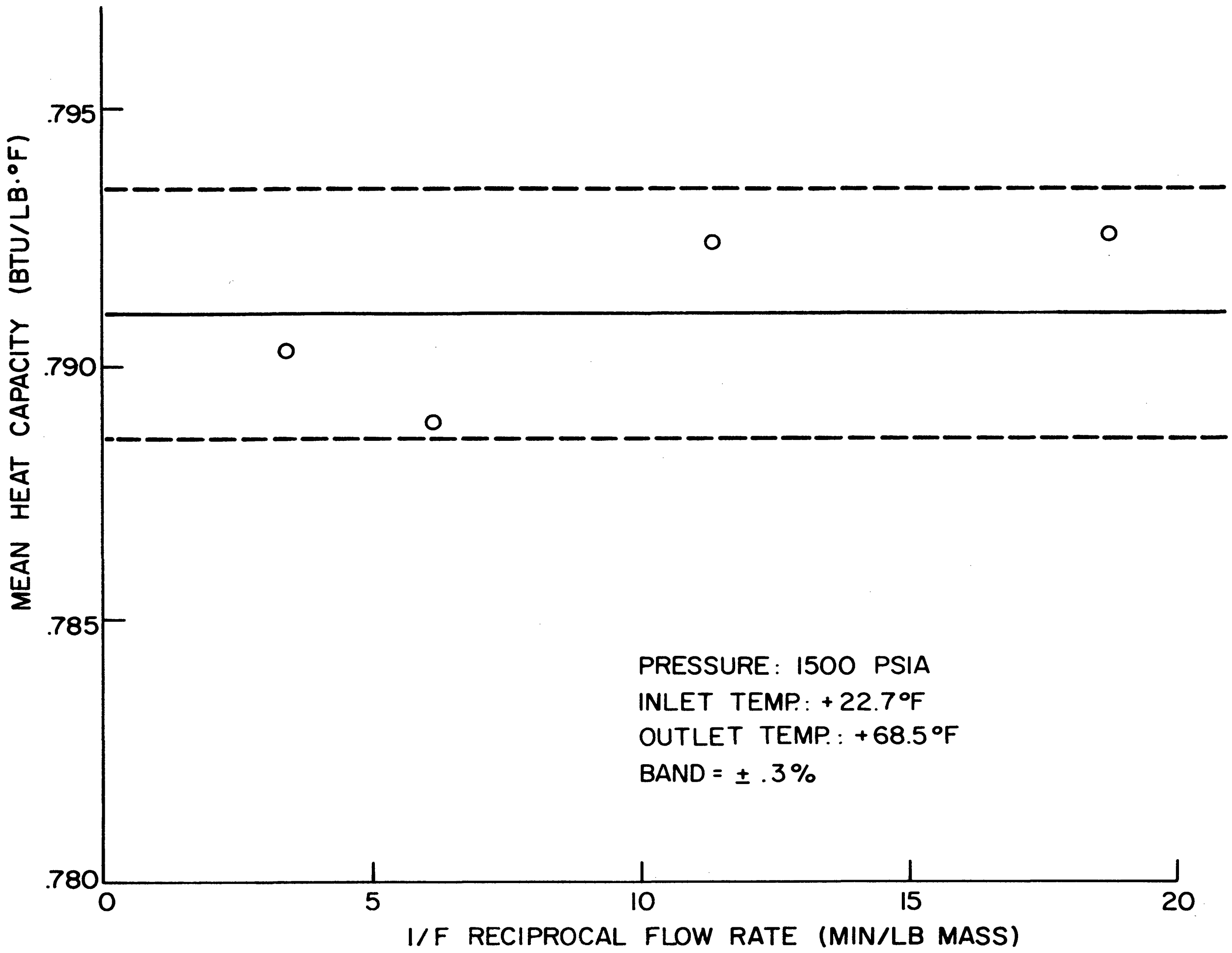


Figure 43. Heat Capacity as a Function of Reciprocal Flow Rate for the Nominal 51 Percent Mixture

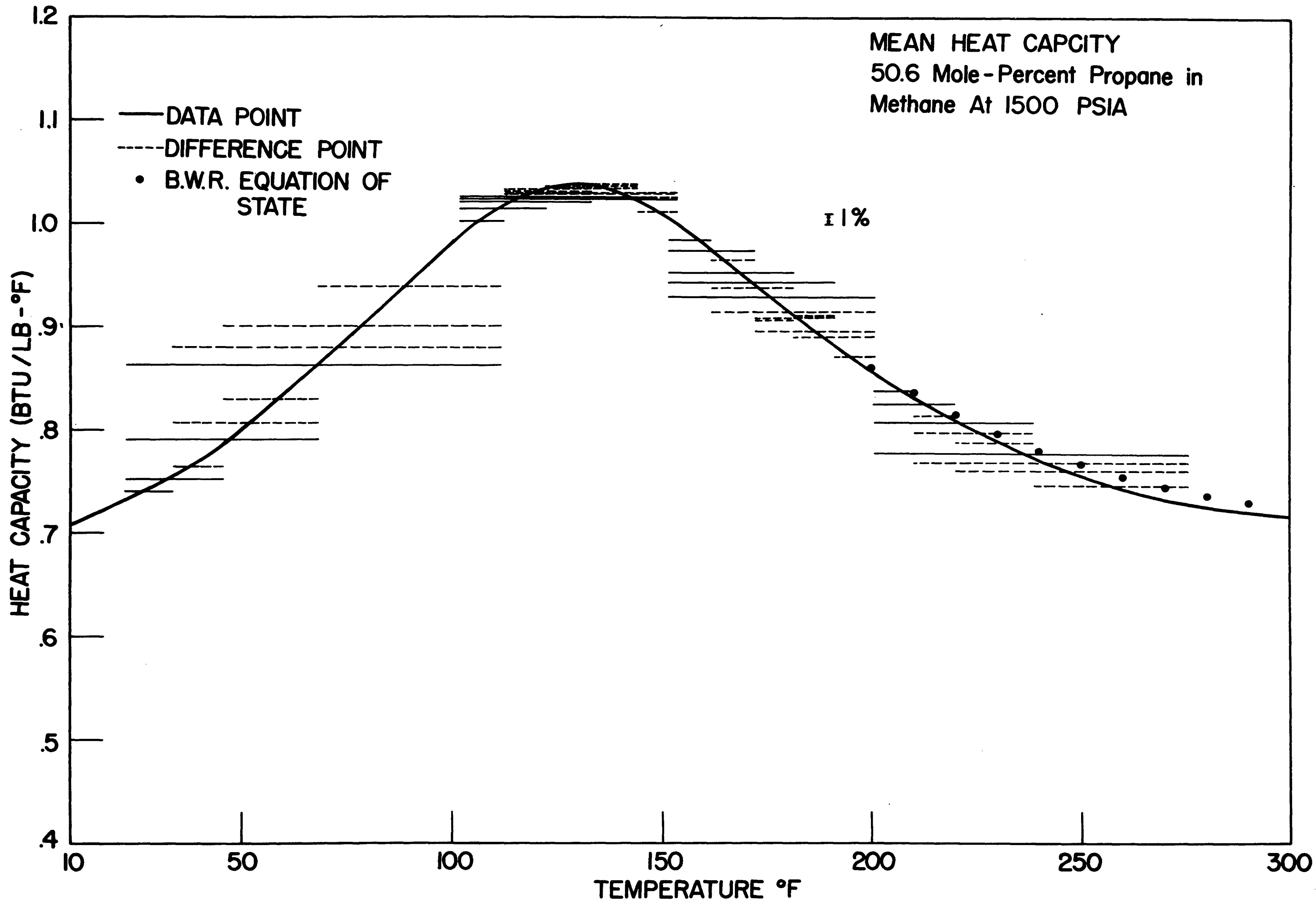


Figure 44. Isobaric Heat Capacity at 1500 psia in the Upper Temperature Range for the Nominal 51 Percent Mixture

point for the mixture. This peak is much less sharp than the peaks either for the 76.6 percent mixture (Figure 32) or for pure propane (Figure 12) both at 1000 psia. It resembles the peaks at higher pressures for these systems (Figures 34 and 15). This suggests that the two-phase envelope for a mixture tends to "cover up" regions where there are drastic changes in the physical properties of a single component. It further suggests that it may actually be easier to represent the behavior of a mixture than of a pure component because the regions most difficult to reproduce or predict for the pure component do not exist for the mixture.

A majority of the values of heat capacity, C_p, obtained from interpretation of isobaric data in the single phase regions are summarized in Table XXIII for equal intervals of temperature and pressure. Table XXIV lists supplementary values of heat capacity in the regions of significant change such as near the maxima in the heat capacity and near the saturation curves. All of the experimental heat capacity results are summarized in Figure 45.

A typical enthalpy traverse of the two-phase region for this mixture at constant pressure is illustrated in Figure 46. Note the large temperature change between dew point and bubble point (almost 200°F). Experimentally determined values of the isobaric enthalpy change on vaporization are listed in Table XXV together with the experimentally determined values of the bubble point and the dew point for the mixture.

Isothermal

Again it was necessary to estimate $\varphi = f(P)$ at low pressures. This procedure is illustrated for this mixture in Figure 47. Values were calculated using the BWR equation of state with the original constants

TABLE XXIII

Tabulated Values of Isobaric Heat Capacities
for a Nominal 51 Mole Percent Propane in Methane Mixture

C_p (Btu/lb - °F)

Temperature °F	Pressure, psia 0	250	500	1000	1500	2000
-280	0.546**	0.545	0.545	0.543	0.543	0.544
-270	0.548**	0.547	0.548	0.544	0.545	0.546
-260	0.551**	0.549	0.550	0.546	0.546	0.548
-250		0.552	0.553	0.548	0.548	0.549
-240		0.554	0.555	0.550	0.550	0.551
-230		0.557	0.558	0.553	0.552	0.554
-220		0.560	0.560	0.555	0.555	0.556
-210		0.563	0.563	0.559	0.557	0.558
-200		0.566	0.565	0.562	0.561	0.561
-190		0.569	0.568	0.565	0.564	0.563
-180		0.572	0.571	0.568	0.567	0.566
-170		0.576	0.574	0.572	0.570	0.568
-160		0.581	0.577	0.576	0.573	0.571
-150		0.586	0.581	0.579	0.577	0.574
-149.0		0.586	0.581	0.580	0.577	0.574
-140		0.592	0.585	0.584	0.581	0.577
-130			0.590	0.588	0.585	0.581
-120			0.595	0.593	0.589	0.585
-110			0.602	0.598	0.594	0.589
-100			0.611	0.604	0.599	0.593
-90			0.627	0.610	0.605	0.597
-80				0.618	0.611	0.602
-70				0.627	0.618	0.608
-60				0.637	0.626	0.613
-50				0.649	0.635	0.620
-40				0.661	0.643	0.627
-30				0.675	0.653	0.635
-20				0.690	0.664	0.643
-10				0.708	0.677	0.652
0				0.730	0.692	0.662

Table XXIII continued

Temperature °F	Pressure, psia 0	250	500	1000	1500	2000
3.5				0.741	0.697	0.666
10				0.771	0.708	0.673
20					0.726	0.686
30					0.746	0.700
40					0.770	0.715
50	0.419*				0.800	0.732
60	0.425*				0.833	0.749
70	0.430*	0.507			0.870	0.768
80	0.435*	0.508			0.907	0.787
90	0.441*	0.509			0.944	0.805
100	0.446*	0.510			0.979	0.821
110	0.451*	0.511	0.664		1.012	0.837
120	0.457*	0.512	0.621		1.031	0.852
130	0.462*	0.514	0.605		1.037	0.866
140	0.468*	0.516	0.596	1.009	1.030	0.879
150	0.473*	0.518	0.589	0.904	1.009	0.889
151.2	0.475*	0.518	0.588	0.887	1.003	0.891
160	0.479*	0.520	0.583	0.837	0.980	0.895
170	0.485*	0.523	0.580	0.790	0.948	0.891
180	0.490*	0.525	0.577	0.756	0.916	0.882
190	0.496*	0.529	0.575	0.730	0.885	0.873
200	0.502*	0.537	0.575	0.710	0.857	0.862
210	0.507*	0.536	0.575	0.695	0.831	0.850
220	0.513*	0.540	0.576	0.684	0.808	0.837
230	0.519*	0.545	0.578	0.675	0.788	0.825
240	0.524*	0.549	0.579	0.669	0.771	0.814
250	0.530*	0.554	0.582	0.664	0.756	0.804
251.3	0.531*	0.555	0.582	0.663	0.754	0.803
260	0.536*	0.560	0.585	0.661	0.743	0.795

Table XXIII continued

Temperature °F	Pressure, psia					
	0	250	500	1000	1500	2000
270	0.541*	0.565	0.589	0.658	0.733	0.787
280	0.547*	0.570	0.592	0.656	0.726	0.780
290	0.553*	0.574	0.596	0.655	0.720	0.773
300	0.558*	0.579	0.600	0.655	0.718	0.767

* Ideal gas values of Rossini et al (118)

** Experimental data of Cutler and Morrison (23)

TABLE XXIV

Supplementary Table of Experimental Values
of Isobaric Heat Capacity, C_p (Btu/lb)
Pressure

700 psia Temp. (°F)	C_p	1300 psia Temp. (°F)	C_p	1700 psia Temp. (°F)	C_p
-50 (ℓ)	.659	50	.8623	110	.9247
135 (g)	.740	60	.9013	120	.9489
140	.724	70	.9390	130	.9581
145	.707	80	.9889	140	.9609
150	.692	90	1.0635	150	.9602
155	.676	100	1.1320	160	.9570
		110	1.1597	170	.9532
		120	1.1486		
		140	1.1171		
		150	1.0208		
		160	.9678		
		170	.9141		

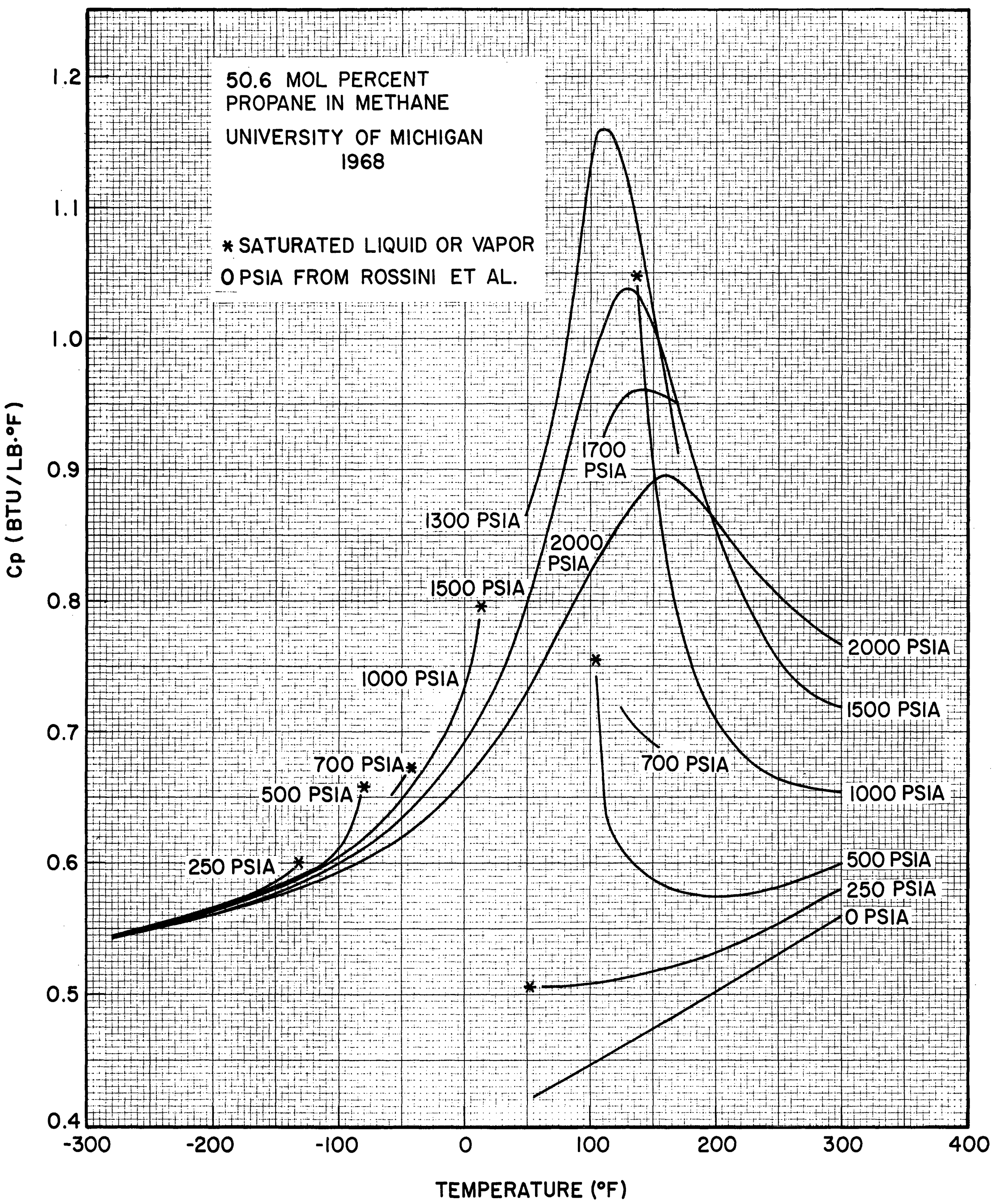


Figure 45. Isobaric Heat Capacity for the Nominal 51 Percent Mixture

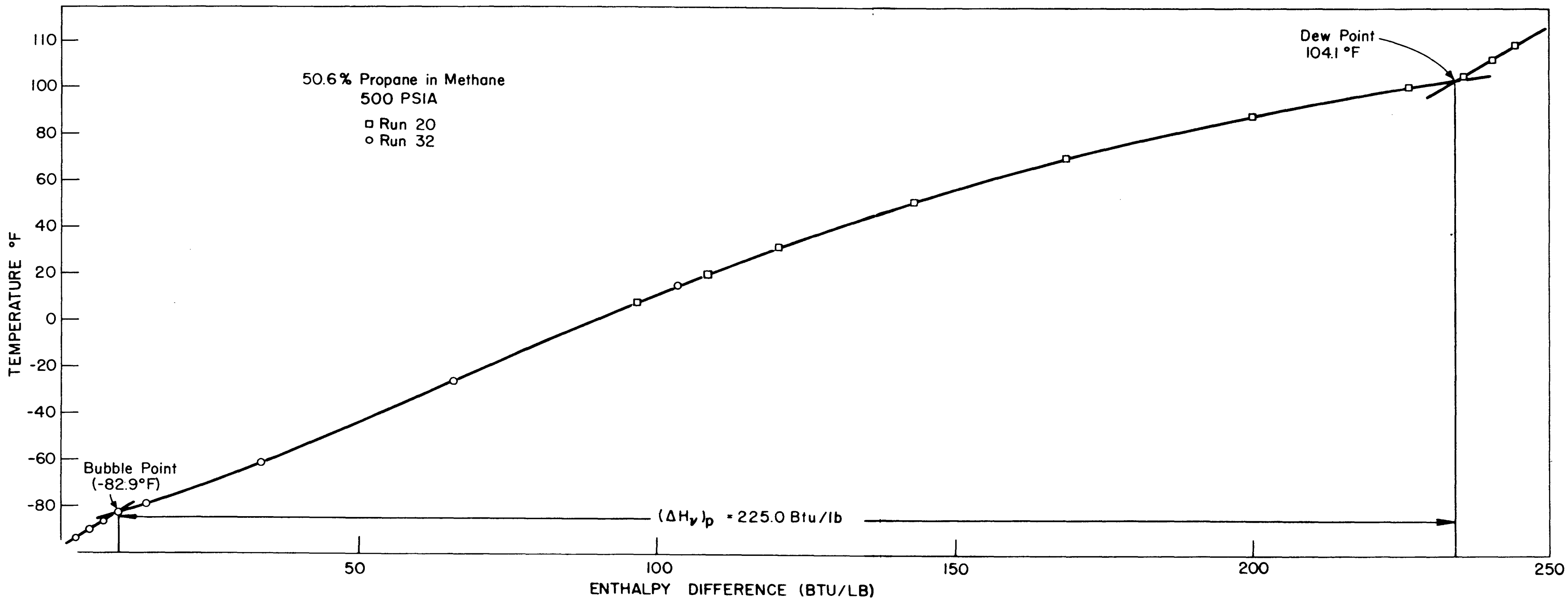


Figure 46. Enthalpy Differences for the 51 Percent Mixture in the Two-Phase Region

TABLE XXV

Properties of the Nominal 51 Mole Percent
Propane in Methane Mixture at the Phase Boundaries

Pressure (psia)	250		500		700		1000	
$(\Delta \underline{H}\ vap)_p$ (Btu/lb)	253.4		225.0		197.5		138.2	
Phase Boundaries	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)	Bubble Point (°F)	Dew Point (°F)
This Investigation	-136.0	62.0	-82.9	104.1	-46.4	123.6	11.1	132.7
Akers, Burns and Fairchild	-125.0	—	-80.9	—	-46.2	—	11.1	—
Price and Kobayashi	-136.3	—	-83.4	—	-46.1	—	11.1	—
Reamer, Sage and Lacey	—	63.8	—	104.6	—	123.5	—	140.5
Sage, Lacey and Schaafsma	—	60.3	—	103.9	—	123.5	—	137.7

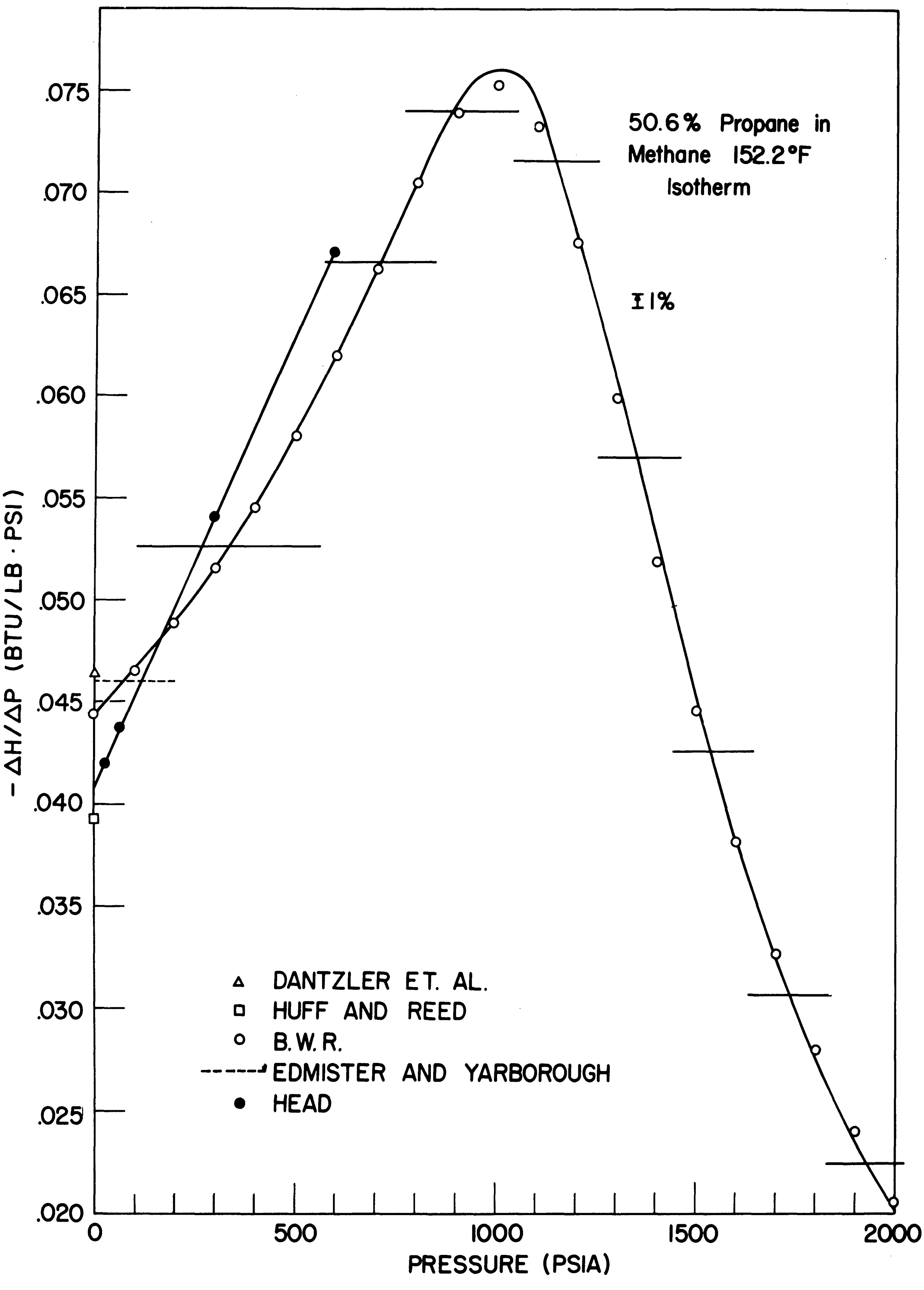


Figure 47. Isothermal Throttling Coefficient for the 51 Percent Mixture at 152.2°F

for methane and propane[7] together with mixing rules as originally suggested.[8] Typical results presented as open circles on Figure 47. Note that the agreement between the BWR equation and the experimental values is good throughout the entire region of pressure.

Again PVT data for the mixture[111] as interpreted[37] using Equation (43) yielded values of $\left[\frac{(\underline{H}^0 - \underline{H})}{P}\right]_T$ between zero pressure and 200 psia. This value is presented as a dashed line on Figure 47. Also Equations (32) to (34) were used with published values of the second virial coefficient for methane and propane and the interaction term[27,58] to estimate φ^0. The resulting values at 152.5°F are plotted on Figure 47. Values from the two sources disagree by over 15 percent. Finally values of μ, the Joule-Thomson coefficient, have been measured by Head[50] for a 51.1 mole percent propane in methane mixture at low pressures. By combining experimentally measured heat capacities from this investigation and Joule-Thomson coefficient data as indicated by Equation (8) values of φ can be calculated. These results are presented as solid circles on Figure 47. There is considerable disagreement between the results of the various methods. The solid line on Figure 47 was drawn to be consistent with the data obtained at elevated pressures and the estimates based on data from the literature in the low pressure range. In the low pressure region the line was drawn to agree with the BWR equation of state because of its excellent agreement with the high pressure data. This line was also reasonably consistent with volumetric data. Values of the isothermal throttling coefficient, φ, obtained by interpreting the data (including values of φ^0 and other values at low pressure determined as outlined above) are reported for each of the experimental isotherms in Table XXVI. These results are summarized in Figure 48.

TABLE XXVI

Experimental Values of the Isothermal Throttling Coefficient, Ø, for a 51 Mole Percent Propane in Methane Mixture

Ø x 10^2 (Btu/lb. psia)

Pressure psia	Temperature (°F) -149.0[a]	3.5	152.2	251.3
0[b]	--	--	-4.44	-3.21
100	--	--	-4.67	-3.30
200	--	--	-4.90	-3.38
300	+0.305	--	-5.16	-3.46
400	+0.306	--	-5.46	-3.53
500	+0.307	--	-5.81	-3.60
600	+0.308	--	-6.20	-3.67
700	+0.309	--	-6.62	-3.73
800	+0.310	--	-7.05	-3.79
900	+0.312	--	-7.45	-3.83
1000	+0.313	-0.1904	-7.61	-3.85
1100	+0.314	-0.1463	-7.44	-3.83
1200	+0.315	-0.1069	-6.80	-3.80
1300	+0.317	-0.0745	-6.08	-3.75
1400	+0.318	-0.0474	-5.31	-3.67
1500	+0.319	-0.0240	-4.52	-3.56
1600	+0.321	-0.0030	-3.82	-3.42
1700	+0.322	+0.0156[a]	-3.23	-3.26
1800	+0.324	+0.0320[a]	-2.75	-3.09
1900	+0.325	+0.0466[a]	-2.35	-2.90
2000	+0.326	+0.0593[a]	-2.01	-2.72

[a] Calculated using Equation (8)

[b] Extrapolated to zero pressure based primarily on PVT data

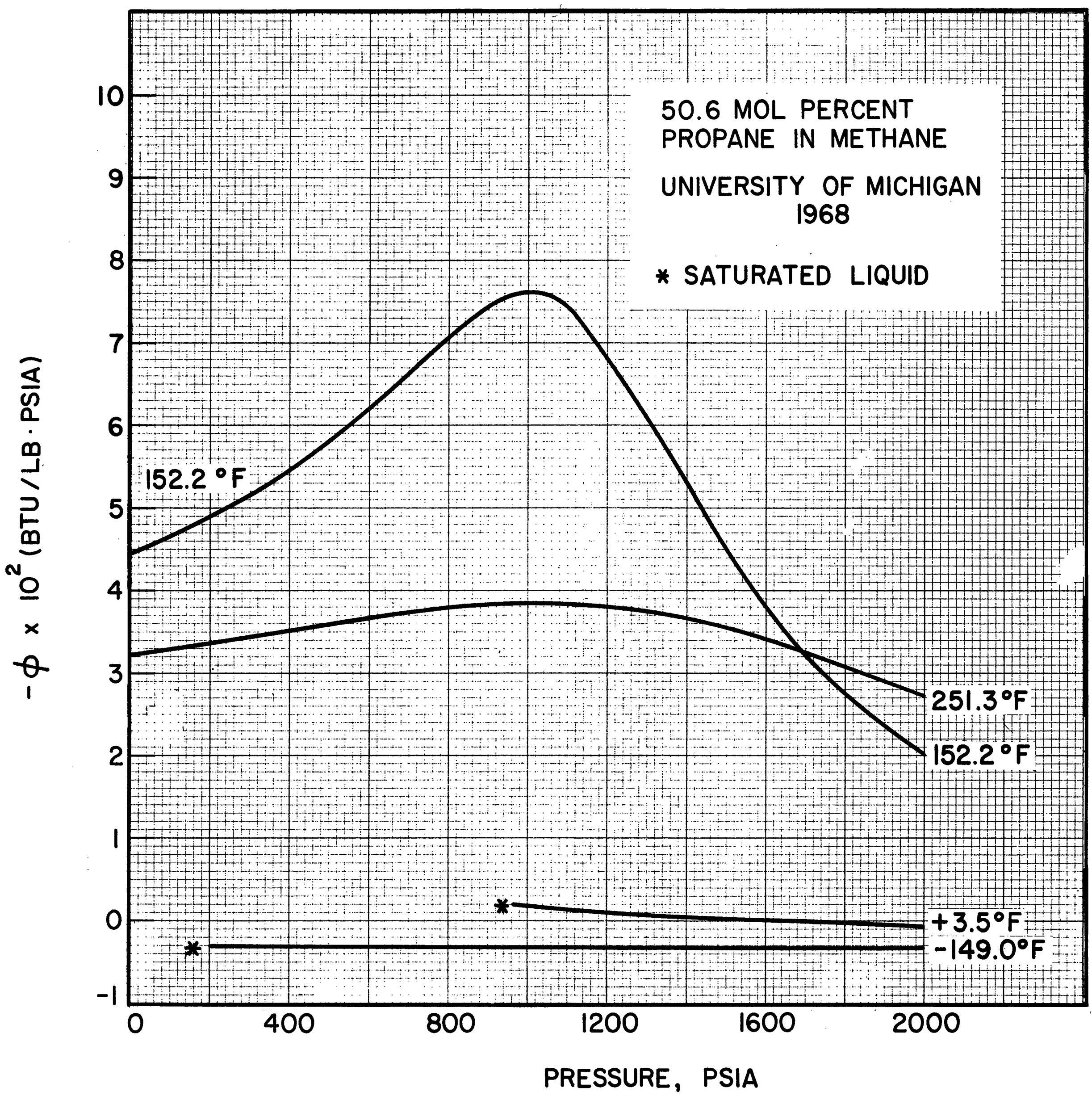


Figure 48. Isothermal Throttling Coefficient for the 51 Percent Mixture

One experimental isothermal run was made into the two-phase region at 3.5°F. A break in the curve was obtained at 966 psia. This break indicated the upper pressure at which vaporization starts.

Isenthalpic

Twelve determinations were made with the throttling calorimeter under conditions such that a drop in pressure resulted in an increase in temperature (Run 4R on Figure 41). The inlet temperature was constant at -149.0°F in all cases. The scatter in the data of greater than 1 percent as illustrated in Figure 49 is the result of the measurement of unusually small temperature differences due to the small Joule-Thomson effect at these conditions. When this effect is related to an enthalpy difference by Equation (17) and used in the preparation of a PT$\underline{H}$ diagram or table the error involved in this scatter becomes extremely insignificant. The values of the Joule-Thomson coefficient determined from these data are summarized in Table XXVII.

TABLE XXVII

EXPERIMENTAL VALUES OF THE JOULE-THOMSON COEFFICIENT, μ, AT -149.0°F FOR THE 51 MOLE PERCENT PROPANE IN METHANE MIXTURE

Pressure psia	$-\mu \times 10^3$ °F/psi
300	5.24
400	5.26
600	5.30
800	5.35
1000	5.40
1200	5.44
1400	5.48
1600	5.53
1800	5.57
2000	5.62

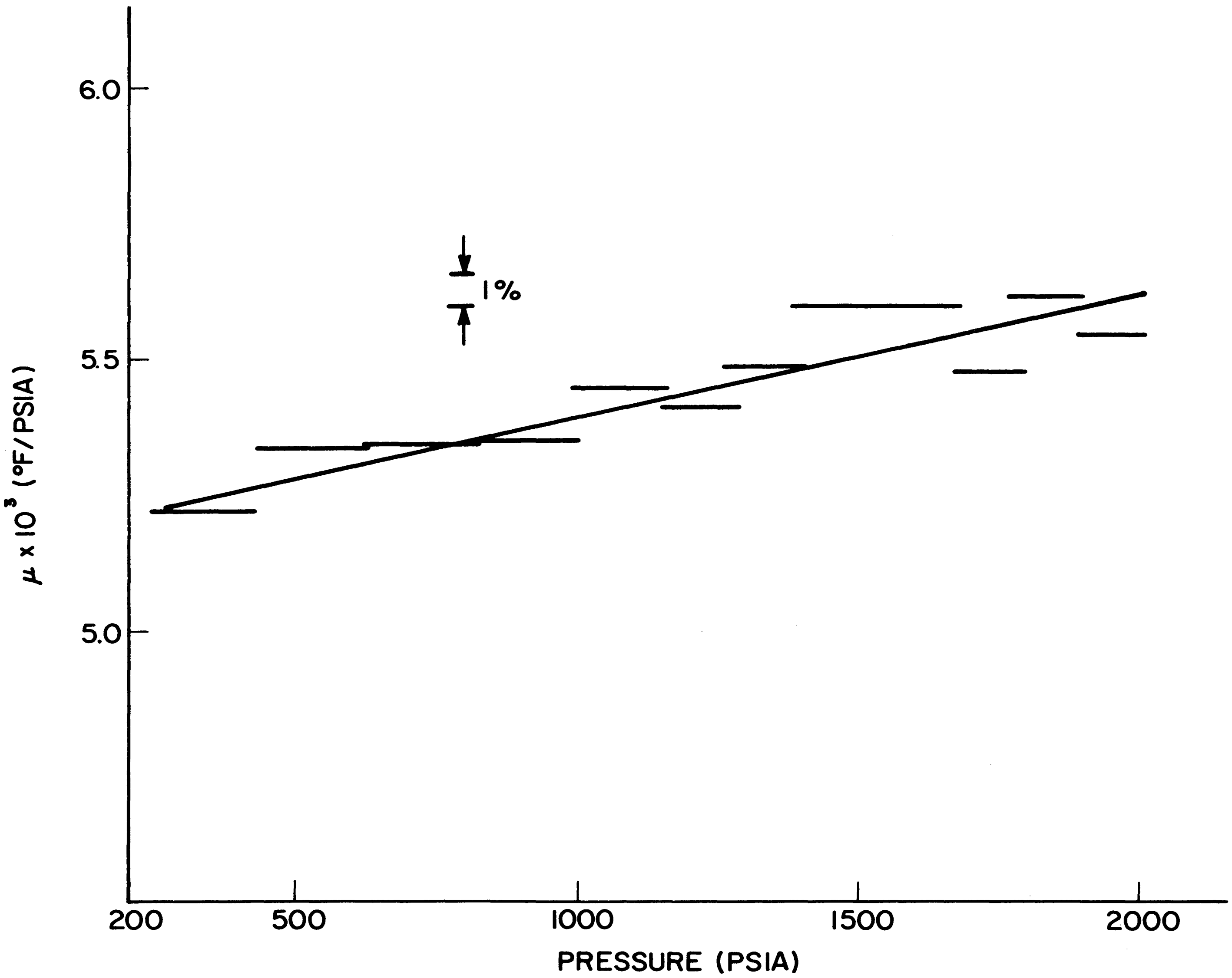


Figure 49. Joule-Thomson Coefficient for the 51 Percent Mixture at $-149.0^{o}F$

Consistency Checks

As illustrated for propane, if isobaric, isothermal, and isenthalpic data are obtained for a system at properly selected values of pressure and temperature, it is possible to check the thermodynamic consistency of such data. For the 77.6 mole percent propane in methane mixture experimentally determined isobars and isotherms intersect forming closed loops which are shown in Figure 50. As can be seen, the largest percentage deviation is 0.47 percent for a loop which included both isobaric and isothermal data within the two-phase region. The average absolute deviation for the 14 loops was found to be 0.18 percent. This should give some indication of the accuracy of the data presented. Figure 51 shows the results of consistency checks for the 50.6 mole percent propane in methane mixture. The largest percentage deviation is 0.74 percent for a loop which included both isobaric and isothermal data within the two-phase region. This is the largest deviation obtained using the present recycle flow facility. The average absolute deviation for the 13 loops was found to be 0.23 percent.

Enthalpy Diagrams

The data reported in previous sections have been used to prepare skeleton tables of values of the enthalpy for these mixtures at selected values of pressure and temperature as reported in Tables XXVIII and XXIX. In addition, enthalpy - pressure - temperature diagrams have been prepared and are presented as Figures 52a and 52b, and 53a and 53b.

The 76.6 Mole Percent Propane in Methane Mixture

The following procedure was used in preparing the skeleton table and diagram for the 76.6 mole percent propane in methane mixture.

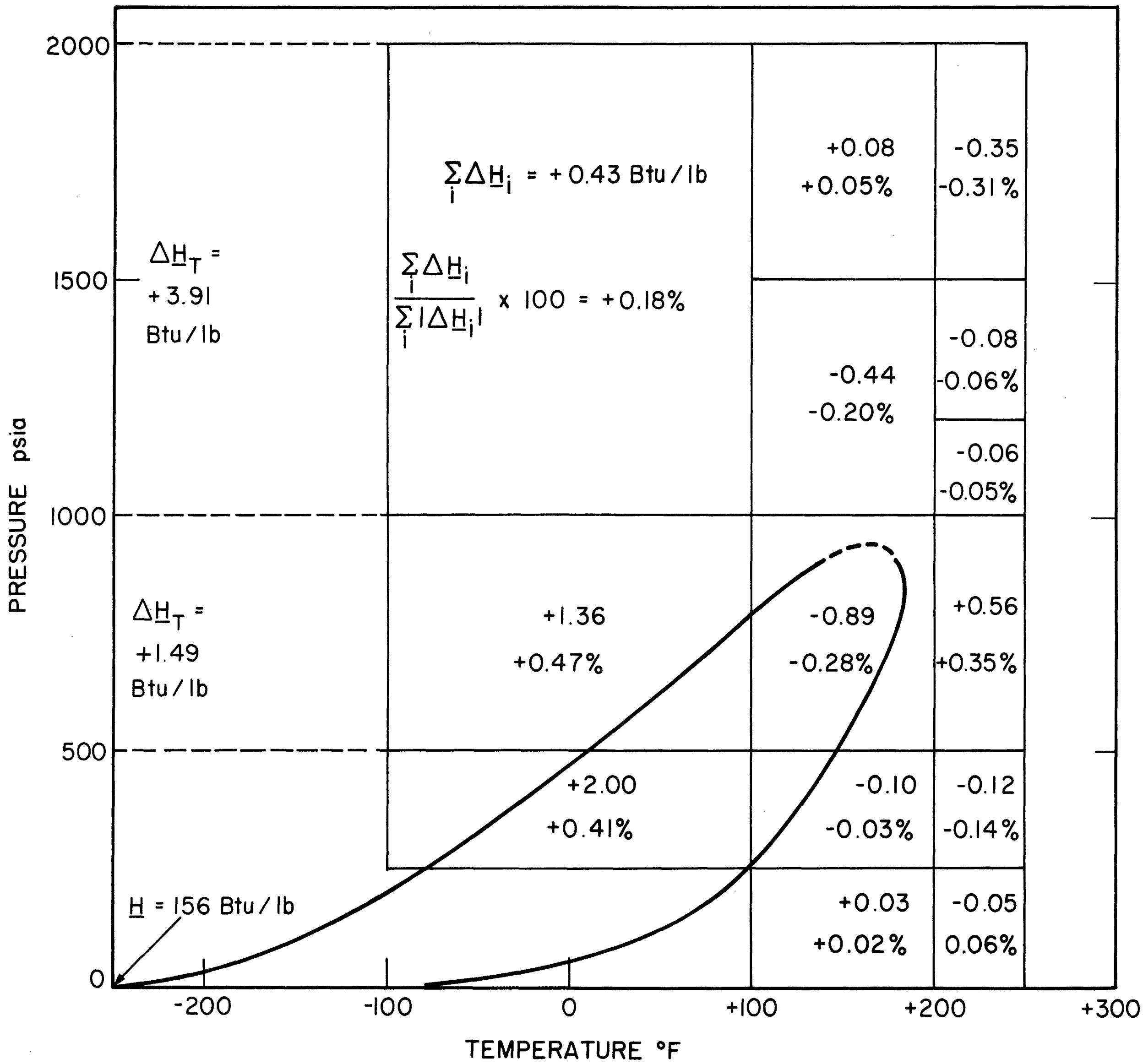


Figure 50. Checks of Thermodynamic Consistency of Thermal Data for the 77 Percent Mixture

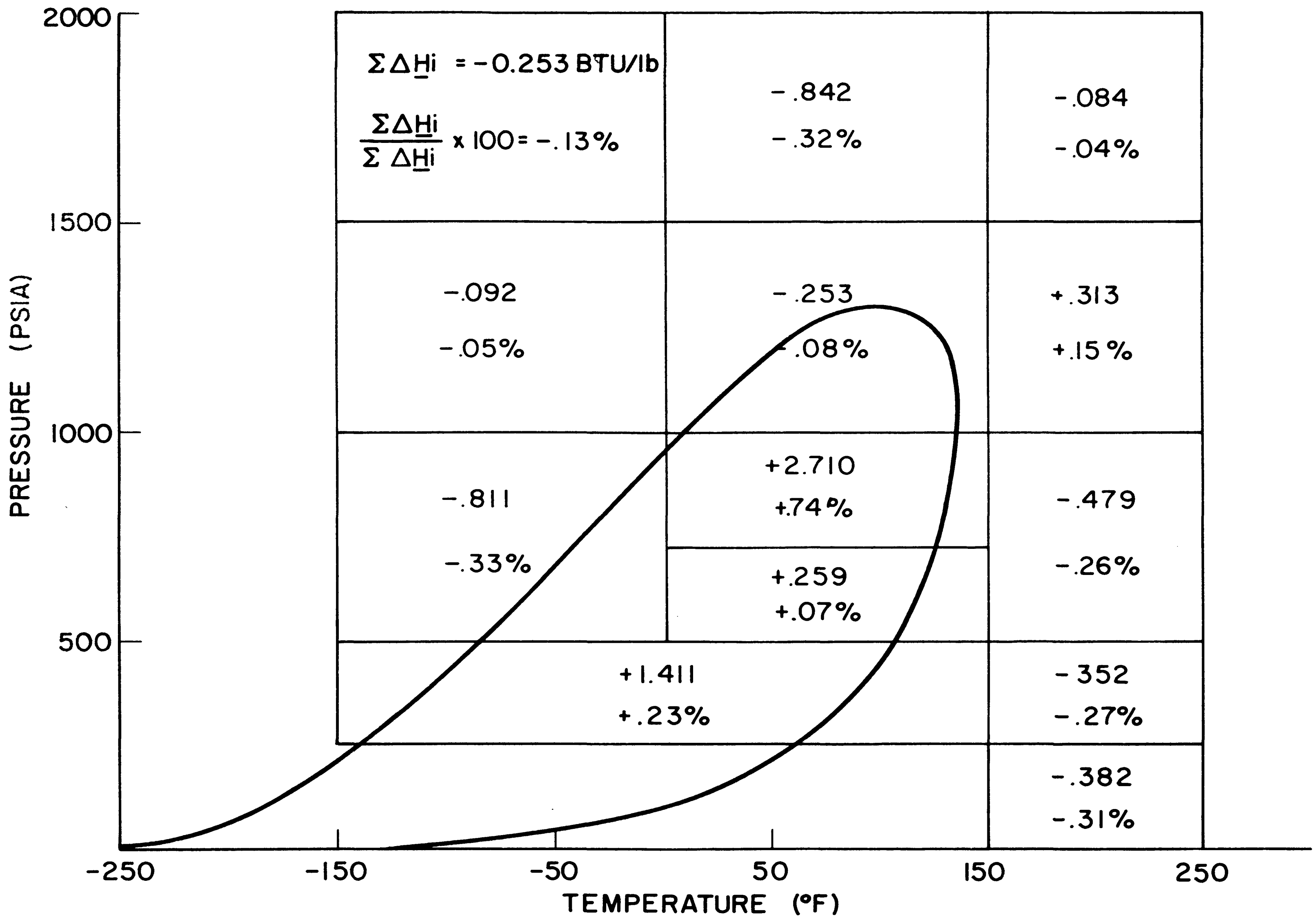


Figure 51. Checks of Thermodynamic Consistency of Thermal Data for the 51 Percent Mixture

TABLE XXVIII

TABULATED VALUES OF ENTHALPY
FOR THE NOMINAL 77 MOL PERCENT PROPANE
IN METHANE MIXTURE

Pressure (psia)	Temperature, °F Bubble Point	Dew Point	Saturated Liquid Enthalpy (Btu/lb)	Saturated Vapor Enthalpy (Btu/lb)	Latent Heat Of Vaporization (Btu/lb)
100	-144	+36	69.0	325.0	256.0
200	-100	+83	93.6	335.7	242.1
300	-62	+110	115.3	342.6	227.3
400	-24	+132	137.2	344.7	207.5
500	12	147	159.2	342.7	183.5
600	47	160	181.3	338.7	157.4
700	76	170	201.6	333.4	131.8
800	102	178	221.6	326.5	104.9

H (Btu/lb)

Temperature (°F)	Pressure, psia 0	250	500	750	1000	1250	1500	1750	2000
-280	234.9	1.2	2.1	3.1	4.2	5.2	5.9	6.9	8.0
-270	237.7	6.1	7.1	8.2	9.2	10.0	10.7	11.6	12.9
-260	240.4	11.1	12.0	13.3	14.2	15.0	15.6	16.5	17.8
-250	243.1	16.1	17.1	18.3	19.2	20.0	20.5	21.4	22.7
-240	245.9	21.1	22.0	23.2	24.0	24.9	25.5	26.5	27.7
-230	248.6	26.0	27.0	28.2	29.0	30.0	30.7	31.6	32.6
-220	251.4	30.9	32.1	33.2	34.0	34.9	35.7	36.5	37.6
-210	254.2	35.9	37.0	38.1	39.0	39.9	40.7	41.5	42.6
-200	257.1	40.9	42.0	43.2	44.0	44.9	45.8	46.6	47.6
-190	259.9	45.9	47.1	48.4	49.2	50.0	50.8	51.5	52.6
-180	262.8	50.9	52.3	53.5	54.4	55.2	55.9	56.7	57.6
-170	265.7	56.1	57.5	58.5	59.4	60.4	61.0	61.9	62.7
-160	268.7	61.3	62.7	63.8	64.6	65.5	66.1	67.0	67.8
-150	271.7	66.6	68.0	67.0	69.8	70.6	71.4	72.2	73.0
-140	274.7	71.8	73.2	74.2	75.0	75.7	76.6	77.4	78.2
-130	277.7	77.0	78.5	79.4	80.2	80.8	81.7	82.5	83.3
-120	280.8	82.3	83.8	84.8	85.5	86.0	86.9	87.6	88.4
-110	284.0	87.9	89.0	90.2	90.8	91.5	92.3	93.0	93.6
-100	287.1	93.7	94.4	95.5	96.2	97.0	97.7	98.4	98.9
- 96.8	288.2	95.7	96.1	97.2	98.0	98.6	99.3	100.0	100.5
- 90	290.4	99.4	100.0	100.9	101.6	102.2	103.0	103.6	104.2
- 80	293.6	105.0	105.5	106.3	107.0	107.6	108.3	109.0	109.5
- 70	296.9	112.3	111.0	111.9	112.5	113.0	113.7	114.5	114.9
- 60	300.3	119.3	116.7	117.4	118.0	118.6	119.5	120.1	120.4
- 50	303.7	126.5	122.4	123.0	123.6	124.4	125.1	125.7	125.9
- 40	307.1	133.8	128.2	128.6	129.3	130.1	130.8	131.3	131.4
- 30	310.6	140.9	134.0	134.3	134.9	135.7	136.4	136.9	137.1
- 20	314.2	148.2	139.9	140.1	140.6	141.4	142.2	142.8	143.0
- 10	317.8	156.2	145.8	146.0	146.3	147.1	148.0	148.5	148.8
0	321.4	163.7	152.0	152.1	152.1	152.9	153.9	154.3	154.7
10	325.2	172.5	158.3	158.2	158.3	159.0	159.9	160.3	160.7
20	328.9	181.2	165.0	164.4	164.5	165.2	165.9	166.4	166.8
30	332.7	191.1	173.1	170.6	170.7	171.5	172.0	172.6	173.0
40	336.6	201.8	181.4	176.9	177.3	177.9	178.4	178.7	179.0
50	340.5	214.7	189.7	183.4	183.9	184.4	184.9	185.0	185.1
60	344.5	229.4	198.8	190.5	190.8	191.0	191.4	191.5	191.4
70	348.6	249.1	207.9	197.6	197.7	197.8	197.9	197.9	197.8
80	352.7	276.2	218.1	204.8	204.6	204.9	204.5	204.4	204.2
90	356.9	312.0	228.6	212.3	212.0	211.7	211.4	211.0	210.7
100	361.1	340.7	240.5	221.9	219.3	218.7	218.3	217.7	217.5
110	365.4	346.9	254.2	231.6	227.2	226.0	225.4	224.8	224.2
120	369.7	352.3	271.1	242.0	235.4	233.9	233.0	232.0	231.2
130	374.1	357.5	292.0	253.4	243.6	241.9	240.6	239.5	238.3
140	378.6	362.8	317.0	266.5	253.4	250.3	248.4	246.9	245.5
150	383.1	368.1	345.0	281.5	263.6	259.6	256.4	254.3	253.0
160	387.7	373.3	352.6	299.0	275.1	268.9	264.5	262.0	260.5
170	392.3	378.5	359.2	318.8	288.2	278.9	273.0	269.6	268.1
180	397.0	383.8	366.0	337.7	302.6	288.9	282.0	278.3	276.0
190	401.8	389.1	372.6	347.7	316.1	298.8	291.0	286.7	283.8
200	406.6	394.2	378.7	356.6	329.5	309.0	300.0	295.0	291.9
201.0	407.1	394.8	379.4	357.5	330.7	310.2	301.0	295.8	292.8
210	411.5	399.6	384.8	365.1	341.2	320.3	309.4	303.6	300.2
220	416.4	404.9	391.0	372.8	351.5	331.5	319.4	312.9	308.8
230	421.4	410.3	397.0	380.5	361.4	342.2	329.4	322.1	317.3
240	426.4	415.7	403.0	387.9	370.6	352.7	339.6	331.3	326.0
250	431.5	421.3	409.0	395.2	378.9	362.6	349.4	340.1	334.7
251.0	432.0	421.9	409.8	396.1	379.8	363.6	350.3	341.2	335.6
260	436.7	426.7	415.0	402.0	387.2	372.4	359.2	349.6	343.4
270	441.9	432.2	421.2	408.7	395.0	381.0	368.0	358.3	352.2
280	447.2	437.7	427.3	415.4	402.4	389.0	376.9	367.9	360.8
290	452.5	443.5	433.2	421.6	409.4	397.2	385.8	376.8	369.3
300	457.9	449.3	440.2	427.7	416.0	404.8	394.5	385.6	377.6

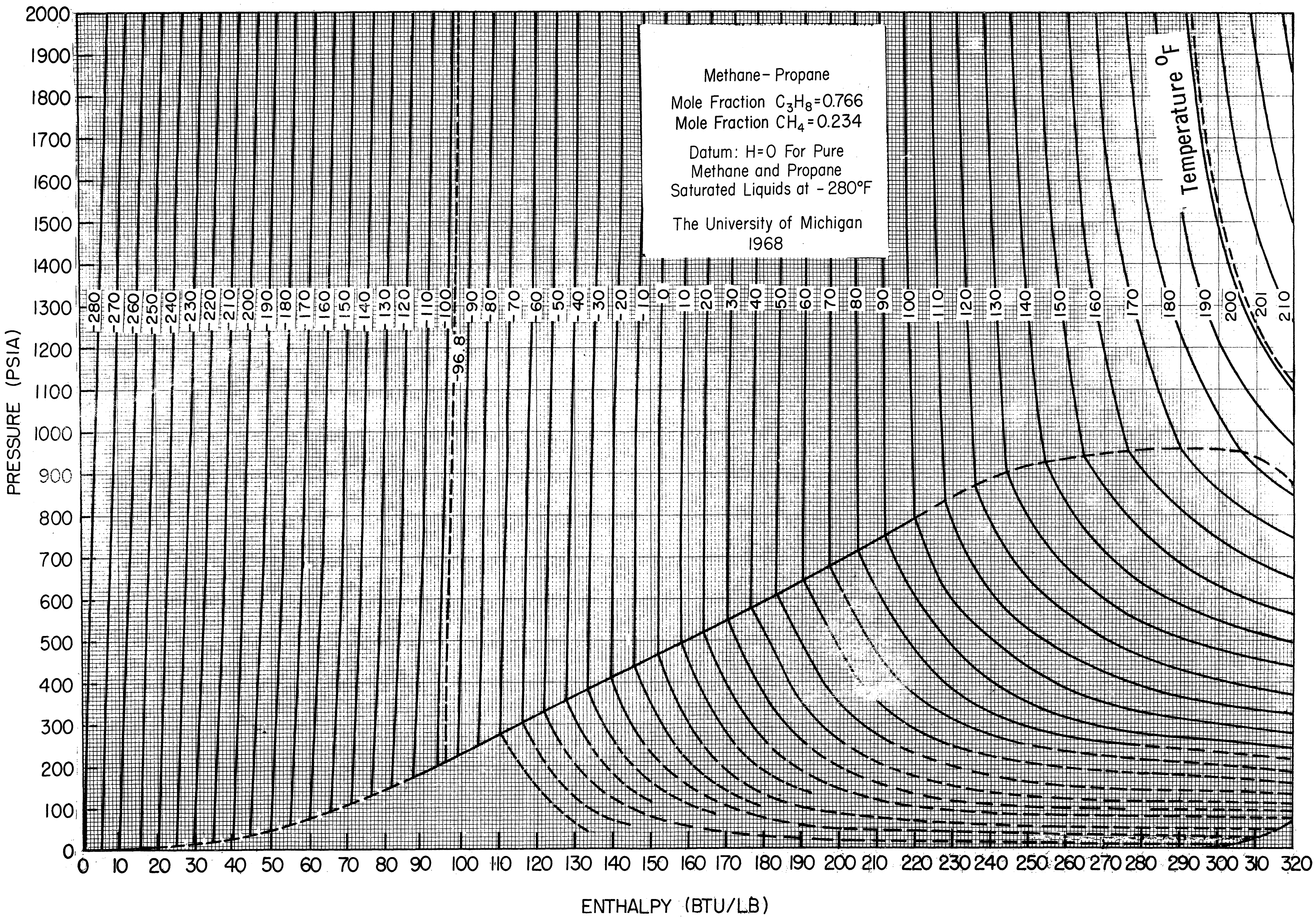


Figure 52a. Pressure-Temperature-Enthalpy Diagram for the 77 Percent Propane in Methane Mixture

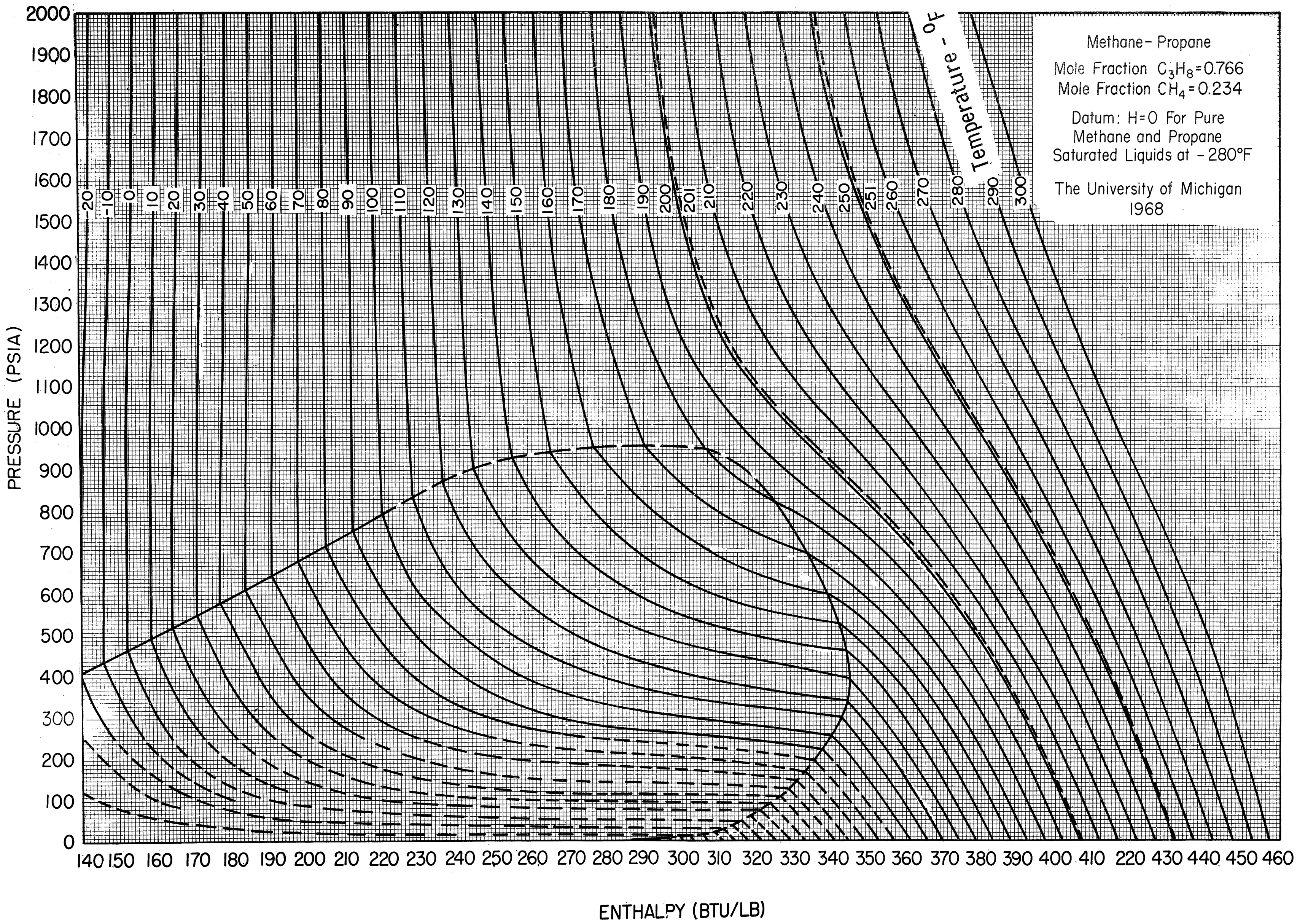


Figure 52b. Pressure-Temperature-Enthalpy Diagram for the 77 Percent Propane in Methane Mixture

TABLE XXIX

TABULATED VALUES OF ENTHALPY
FOR THE NOMINAL 51 MOL PERCENT PROPANE
IN METHANE MIXTURE

Pressure (psia)	Temperature, °F Bubble Point	Temperature, °F Dew Point	Saturated Liquid Enthalpy (Btu/lb)	Saturated Vapor Enthalpy (Btu/lb)	Latent Heat of Vaporization (Btu/lb)
100	-180	13	56.6	327.0	270.4
200	-149	50	74.8	335.5	260.7
300	-125	73	89.7	339.6	249.9
400	-103	91	102.9	341.3	238.4
500	-83	104	115.4	341.5	226.1
600	-66	116	127.8	340.6	212.8
700	-46	124	140.0	338.4	198.4
800	-27	132	152.7	335.6	182.9
900	-9	137	165.8	330.4	164.6
1000	11	137	180.0	322.2	142.2
1100	31	136	194.5	312.1	117.6
1200	51	131	210.2	299.6	89.4
1300	79	114	235.4	276.5	41.1

H (Btu/lb)

Temperature (°F)	Pressure, psia 0	250	500	750	1000	1250	1500	1750	2000
-280	234.0	1.5	2.1	2.9	3.9	4.9	6.0	6.7	7.5
-270	237.1	7.0	7.5	8.4	9.6	10.4	11.4	12.1	13.1
-260	240.2	12.3	12.9	13.9	15.1	16.0	16.7	17.6	18.5
-250	243.4	17.7	18.5	19.5	20.4	21.3	22.2	23.0	24.0
-240	246.5	23.4	24.0	24.9	26.0	26.8	27.7	28.5	29.5
-230	249.7	28.9	29.6	30.5	31.5	32.3	33.3	34.1	34.9
-220	252.9	34.6	35.4	36.3	37.0	37.6	38.7	39.6	40.4
-210	256.1	40.2	40.9	41.8	42.5	43.3	44.4	45.2	46.1
-200	259.3	45.9	46.5	47.4	48.2	49.0	49.9	50.7	51.7
-190	262.5	51.4	52.3	52.9	53.6	54.6	55.5	56.4	57.2
-180	265.8	57.0	57.9	58.8	59.6	60.4	61.3	62.1	62.8
-170	269.1	62.8	63.6	64.3	65.2	65.9	66.9	67.7	68.6
-160	272.4	68.6	69.5	70.2	71.0	71.7	72.5	73.3	74.4
-150	275.7	74.4	75.1	75.8	76.7	77.5	78.3	79.0	80.0
-149.0	276.0	75.0	75.7	76.5	77.2	78.0	78.9	79.7	80.5
-140	279.1	80.4	80.8	81.4	82.6	83.5	84.3	85.0	85.8
-130	282.5	93.5	86.7	87.4	88.4	89.3	90.0	90.7	91.6
-120	285.9	107.0	92.7	93.6	94.3	95.0	95.8	96.5	97.4
-110	289.4	118.2	98.6	99.2	100.2	101.0	101.2	102.4	103.3
-100	292.9	127.1	104.8	105.5	106.2	107.0	107.6	108.3	109.2
-90	296.4	136.0	110.8	111.5	112.3	113.0	113.6	114.3	115.1
-80	300.0	144.2	119.0	118.1	118.5	118.9	119.6	120.3	121.2
-70	303.6	152.7	130.5	124.6	124.5	125.1	125.9	126.4	127.2
-60	307.2	160.7	140.6	131.3	131.0	131.3	132.0	132.5	133.3
-50	310.9	168.6	149.9	137.6	137.4	137.6	138.2	138.8	139.5
-40	314.7	177.4	159.1	143.9	143.9	144.2	144.7	145.2	145.7
-30	318.4	186.4	168.3	153.3	150.6	150.7	151.1	151.4	151.9
-20	322.3	195.7	176.7	163.4	157.5	157.6	157.9	158.0	158.4
-10	326.1	205.7	185.5	173.3	164.6	164.2	164.4	164.5	165.0
0	330.1	217.5	194.7	182.1	171.5	171.3	171.3	171.2	171.6
3.5	331.4		197.9	185.1	174.1	173.9	173.7	173.6	173.8
10	334.0	230.0	204.2	191.5	179.2	178.5	178.2	178.2	178.4
20	338.0	244.3	213.7	200.8	188.3	185.6	185.2	185.0	185.0
30	342.1	260.4	224.0	210.6	198.1	193.0	192.5	192.3	192.0
40	346.2	278.2	234.9	220.2	208.1	201.1	200.1	199.3	199.0
50	350.4	301.9	247.0	229.7	217.6	209.3	208.2	207.2	206.5
60	354.6	330.5	260.0	239.8	227.6	218.3	216.3	214.6	213.7
70	358.9	341.7	274.2	251.1	237.6	228.1	224.8	222.8	221.5
80	363.2	347.0	290.8	263.0	248.4	238.4	233.8	231.0	229.1
90	367.6	352.0	309.4	276.2	259.5	248.5	242.9	239.3	237.0
100	372.0	357.5	330.5	290.2	271.7	259.9	252.2	247.8	245.1
110	376.5	362.4	345.1	306.2	284.7	271.5	262.3	256.7	253.5
120	381.0	367.7	351.3	323.8	298.7	283.4	272.5	266.0	262.0
130	385.6	372.9	357.6	338.5	313.6	295.4	282.7	275.3	270.5
140	390.3	377.9	363.7	346.1	325.8	306.7	293.1	284.7	279.5
150	395.0	383.2	369.5	353.7	335.3	317.2	303.3	294.0	288.2
152.2	396.0	384.5	371.0	355.0	337.2	319.2	305.5	296.1	290.1
160	399.8	388.5	375.6	360.8	344.1	327.1	313.2	303.6	297.1
170	404.6	393.7	381.4	367.3	351.8	336.1	323.0	313.2	306.0
180	409.5	398.9	387.3	374.2	359.8	345.0	332.0	322.5	314.9
190	414.4	404.2	393.2	380.8	367.1	353.5	341.1	331.4	323.7
200	419.4	409.6	398.7	387.1	374.5	361.7	350.1	340.6	332.5
210	424.4	414.8	404.7	393.6	381.6	369.8	358.5	349.0	341.0
220	429.5	420.2	410.3	399.9	388.7	377.3	366.5	357.3	349.5
230	434.7	425.7	416.1	406.0	395.3	384.7	374.5	365.7	357.8
240	439.9	431.2	421.9	412.4	402.3	392.1	382.2	373.7	366.0
250	445.1	436.6	427.9	418.2	408.7	399.0	390.0	381.6	374.0
251.3	445.8	437.2	428.5	419.2	409.6	399.9	390.8	382.5	375.0
260	450.5	442.2	433.7	424.9	415.4	407.2	397.4	389.3	382.1
270	455.8	447.9	439.5	431.0	422.0	413.3	404.7	397.1	390.1
280	461.3	453.4	445.5	437.3	428.5	420.1	411.9	404.6	397.7
290	466.8	459.2	451.3	443.3	435.2	427.1	419.3	412.3	405.5
300	472.3	465.1	457.4	449.7	441.8	434.1	426.5	419.7	413.2

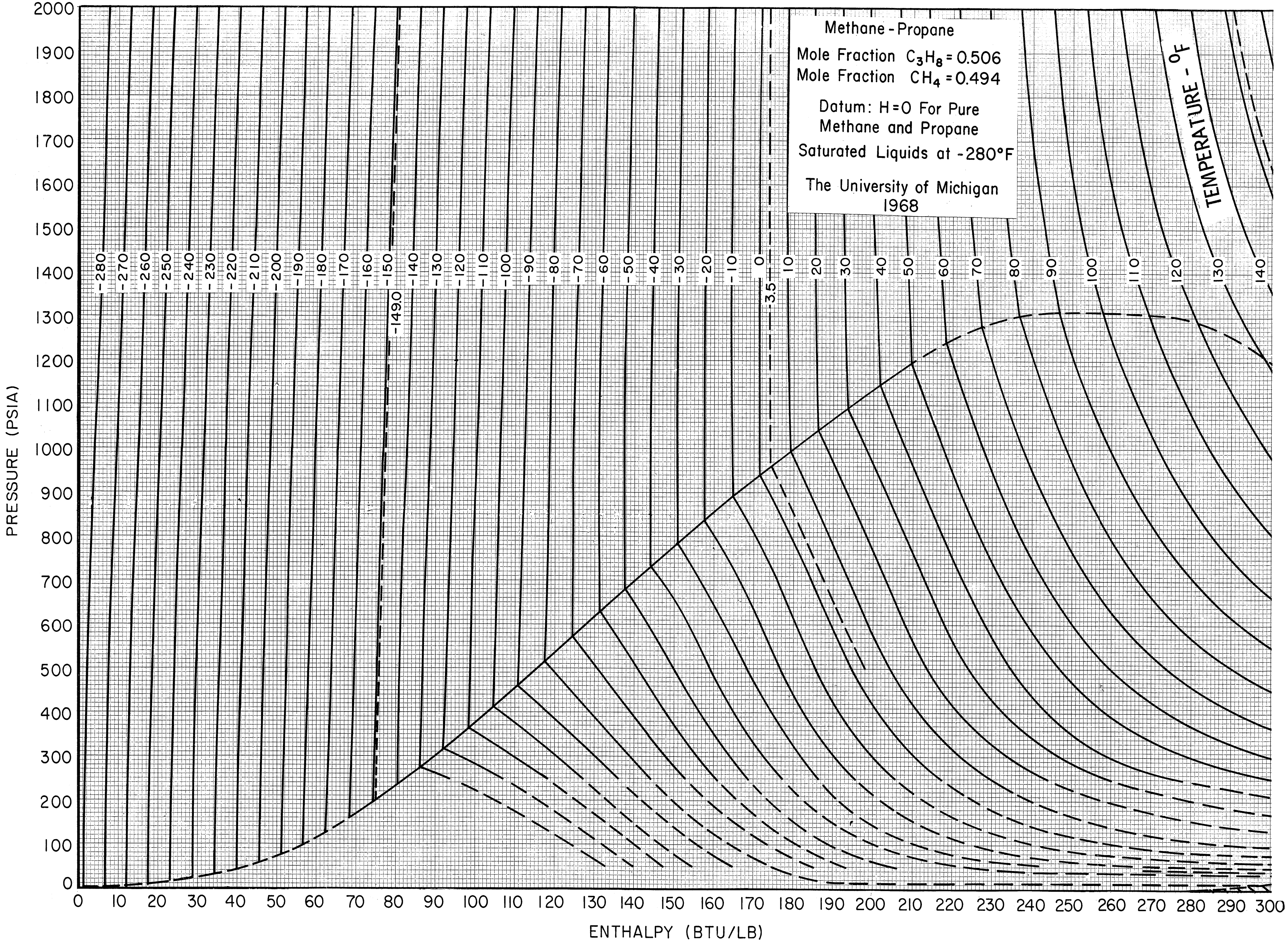


Figure 53a. Pressure-Temperature-Enthalpy Diagram for the 51 Percent Propane in Methane Mixture

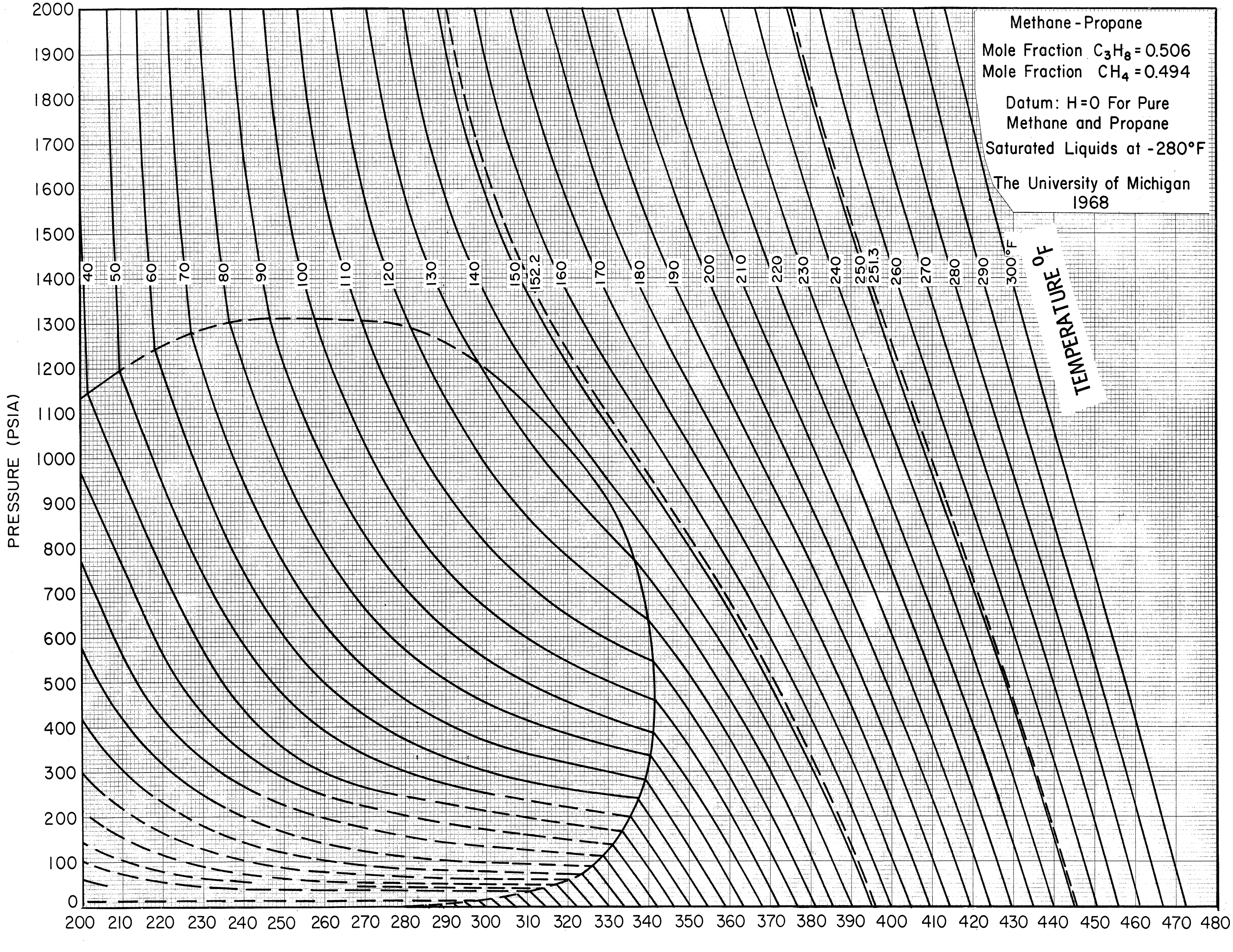


Figure 53b. Pressure-Temperature-Enthalpy Diagram for the 51 Percent Propane in Methane Mixture

1. Reference states were taken to be $\underline{H}$ = 0 Btu/lb for the pure components as saturated liquids at -280°F. This choice is consistent with that previously used for pure methane,[63] pure propane and for other mixtures of propane and methane which have been investigated.[80,82,85,87,88]

2. The enthalpy of pure methane as a gas at zero pressure and +201.0°F was calculated using published data on the latent heat of vaporization at 5 psia,[43] the BWR equation of state with the original constants[7] to correct from 5 psia to zero pressure at -280°F and published values of the ideal gas heat capacity[118] between -280 and +201.0°F. These calculations are summarized below.

	$\underline{H}$(Btu/lb)
Enthalpy change on vaporization at -280°F	+228.27
Effect of pressure on enthalpy (5 to 0 psia)	+1.43
Effect of temperature on zero pressure enthalpy (-280 to +201.0°F)	+248.67
$\underline{H}^0_{C_1}$ (Pure methane at zero pressure and +201.0°F)	478.37

3. The enthalpy of pure propane as a gas at zero pressure and 201.0°F was calculated using data on the liquid phase heat capacity,[66] the latent of heat of vaporization at 1 atmosphere,[66] the BWR equation of state[7] to correct from 1 atmosphere to zero pressure at the normal boiling point, and values of the ideal heat capacity[118] to account for a change in temperature from the normal boiling point of propane to 201.0°F. These calculations are summarized below.

	$\underline{H}$(Btu/lb)
Saturated liquid (-280 to -43.7°F)	115.30
Enthalpy change on vaporization at -43.7°F	183.17
Effect of pressure on enthalpy (14.7 to 0 psia)	2.70

Effect of temperature on zero pressure enthalpy (-43.7°F to +201.0°F) 97.99

$\underline{H}^{0}_{C_3}$ (Pure propane at zero pressure and +201.0°F) 399.16

4. The enthalpy of the propane-methane mixture at zero pressure and 201.0°F was calculated assuming negligible heat of mixing under these conditions. The molecular weight of methane was taken to be 16.042 and that of propane as 44.094. The resulting value for the mixture is 407.08 Btu/lb.

5. The isothermal effect of pressure on the enthalpy of the mixture at 201.0°F was established from the basic experimental data obtained at this temperature (see Figure 36). As indicated previously, extrapolation of the experimental data to zero pressure was necessary and involved application of data from the literature.[7,8,37,58,111]

6. Isobaric data reported in this manuscript were used at various pressures to determine the isobaric effect of temperature on enthalpy in both the gaseous and liquid regions as well as within the two-phase envelope. The limits of the two-phase region were determined using results from the traverses of the two-phase regions which were made during the course of this investigation (Table XVII and Figures 35 and 38) supplemented by data from the literature.[1,109,111,117,123,126]

7. A skeleton table of values determined in this manner was prepared. Slight adjustments were made in the values such that all deviations reported in Figure 50 were reduced to zero. The final results are presented as Table XXVIII.

8. Values from the skeleton table were plotted on graph paper and a smooth plot of the results was prepared by graphical methods. The PT$\underline{H}$ diagram is presented as Figures 52a and 52b.

The 50.6 Mole Percent Propane in Methane Mixture

Essentially the same procedure was used for this mixture as for the 76.6 mole percent mixture. However, the enthalpy at zero pressure and 152.2°F was calculated from the data of the pure components. The resulting value for the enthalpy of the mixture under these conditions is 396.05 Btu/lb.

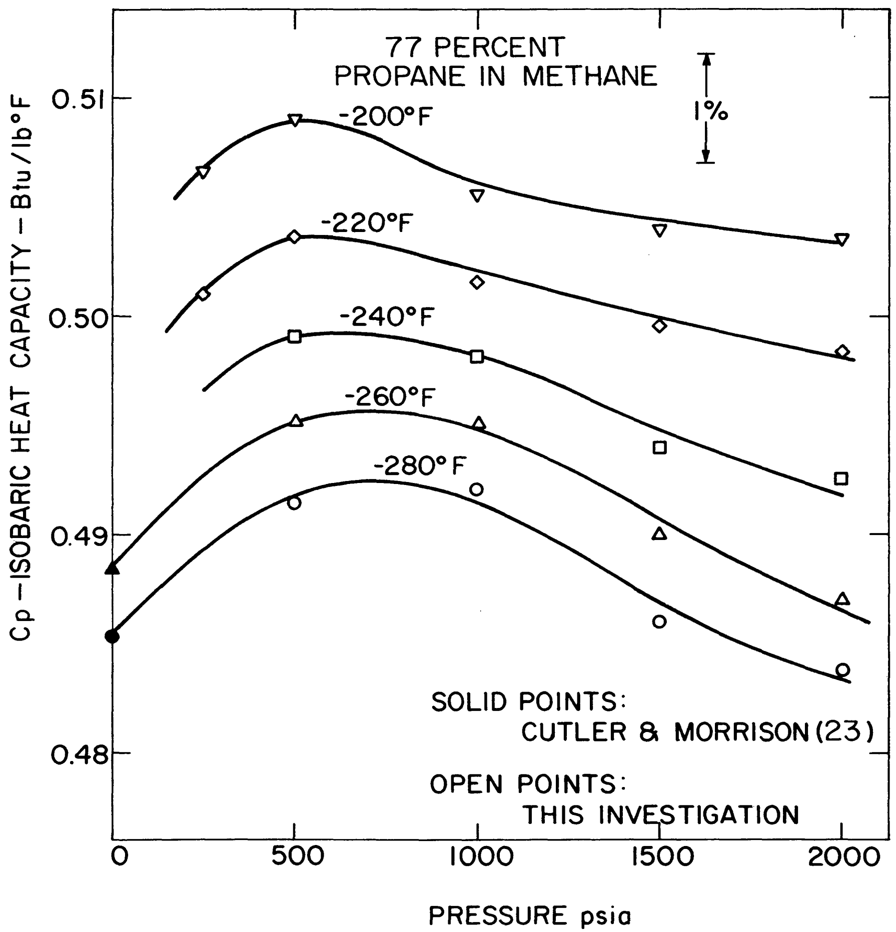
The isothermal effect of pressure on the enthalpy at 152.2°F was established from the basic experimental data and extrapolated to zero pressure by application of data from the literature[7,8,27,33,50,58,111] for this mixture.

A skeleton table of values determined for this mixture is given in Table XXIX. Values from the skeleton table are presented as a smooth plot on Figures 53a and 53b.

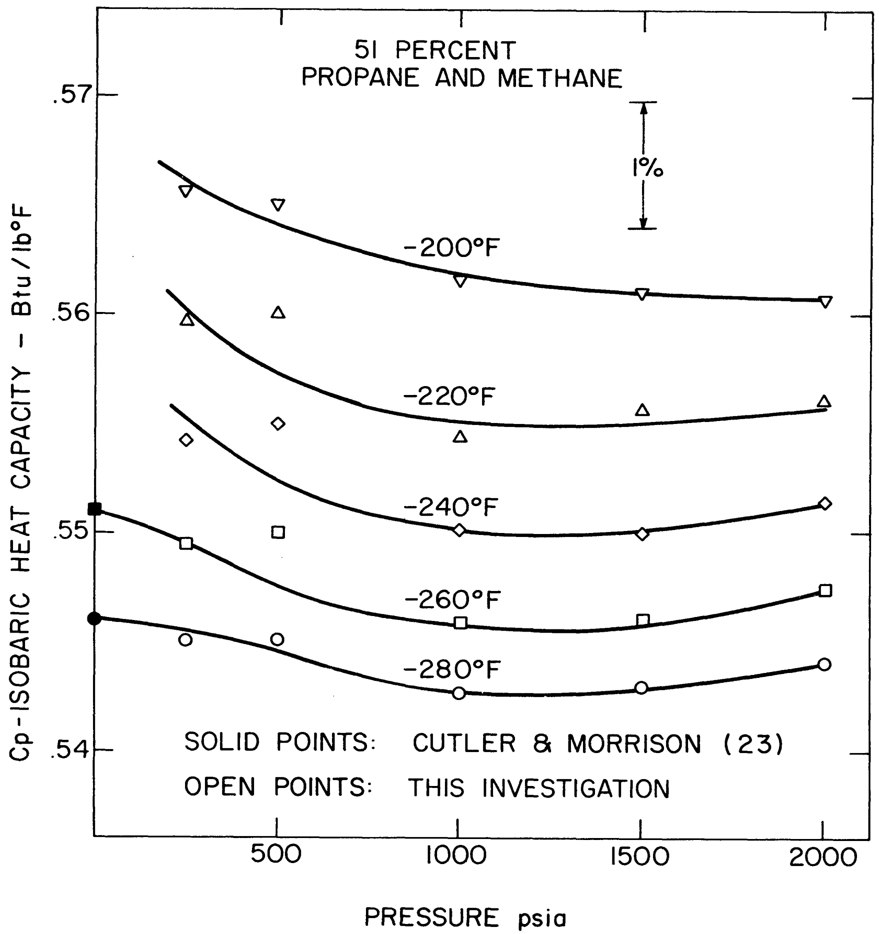
Comparison with Other Published Data

Enthalpy Data

There are some data in the literature which permit direct comparison with the data reported here. Cutler and Morrison[23] report data on the heat capacity of liquid mixtures of methane and propane at temperatures around -280°F as well as data on the heat of mixing of liquid methane and propane at -280°F. The values of the heat capacity at -260 and -280°F for the saturated liquid based on data reported by Cutler and Morrison are plotted on Figure 54 together with values obtained by extrapolation to lower temperatures of isobaric determinations at elevated pressures. The data from these independent investigations illustrate the effect of pressure on the isobaric heat capacity in the dense fluid region. There may be some question regarding the difference in the curvature



(a)



(b)

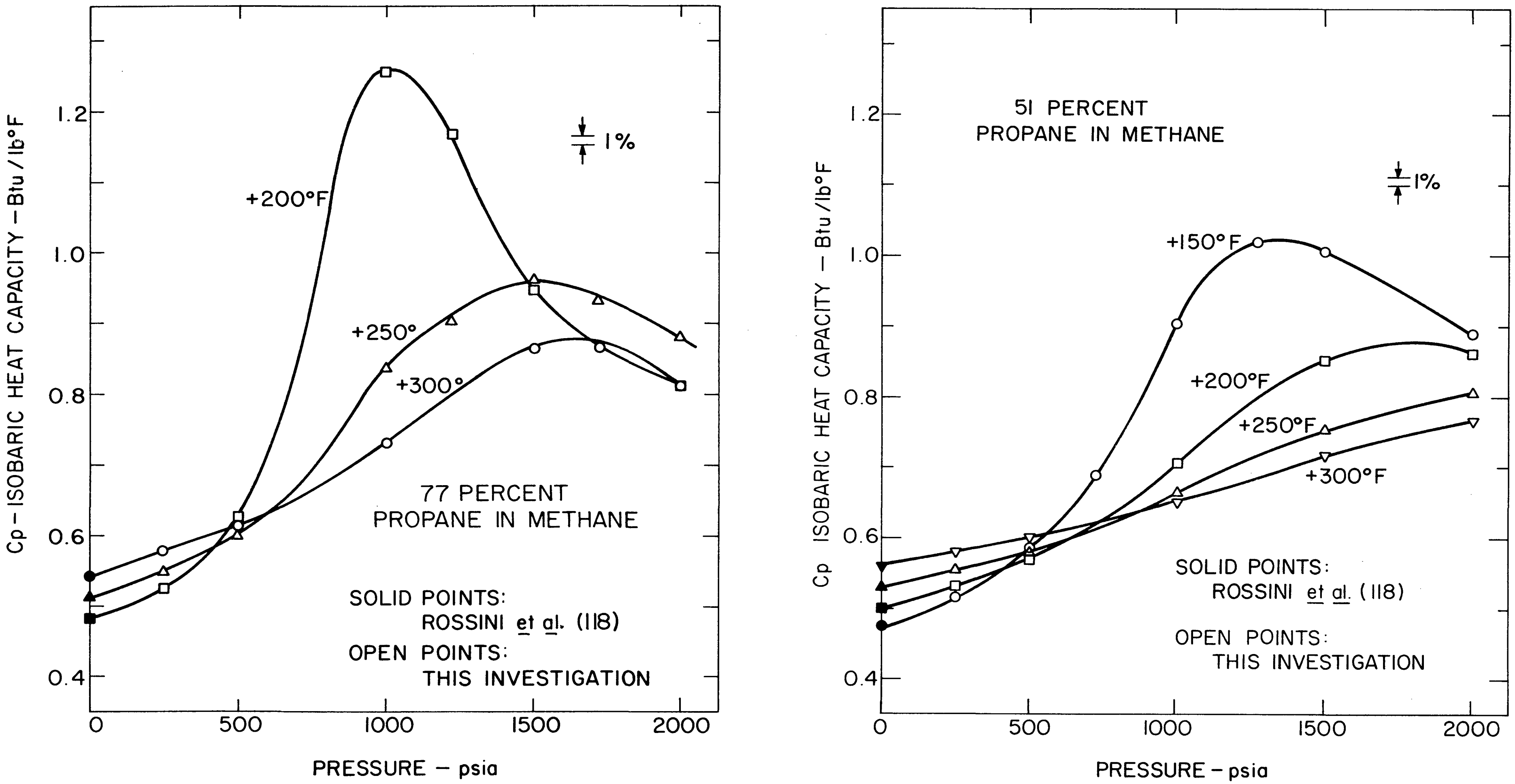
Figure 54. The Effect of Pressure on C_p at Low Temperatures Including Comparison with Data of Cutler and Morrison (23)

between the results for the two mixtures. The curvature, however, is difficult to determine because of the small variation in heat capacity with respect to pressure (note the 1 percent band).

Values of heat capacity of the mixtures can be calculated at zero pressure from published values of ideal heat capacities.[118] Values for both mixtures thus determined are plotted on Figure 55 together with values obtained during the course of this investigation. These values from independent sources are consistent and illustrate the effect of pressure on the isobaric heat capacity in the gaseous region at temperatures just above the two-phase envelope.

Data on the Joule-Thomson coefficient, μ, have been published for a 51.1 mole percent propane in methane mixture at 152°F at pressures up to 600 psia by Head.[50] In addition, data on the Joule-Thomson coefficient have been published for several binary mixtures of methane and propane over a wide temperature range and up to 1500 psia.[11] These latter values were interpolated with respect to composition to establish the values reported in Table XXX. In addition to both sets of μ values, values of C_p and φ for the 50.6 mole percent mixture and the ratio of $\frac{-\mu C_p}{\varphi} \equiv 1.00$ are also listed. The values of μ obtained by both experimental investigators are reasonably self consistent, however, they differ from the experimental results of this investigation by as much as 10 percent. The data of Head[50] when combined with experimental heat capacities from this investigation and ideal gas heat capacities reported in the literature by Rossini[118] give the results as shown in Figure 47. These results do not agree well with values of isothermal throttling coefficient obtained from $P\underline{V}T$ data or the BWR equation of state.

The Joule-Thomson results[11] mentioned above must be used with



(a) (b)

Figure 55. The Effect of Pressure on C_p at High Temperatures Including Comparison with Published Values of Rossini et al. (118)

TABLE XXX

Test of Consistency of Data Based on Equation (8)

Temperature (°F)	Pressure (psia)	C_p (Btu/lb°F)	ϕ (Btu/lbpsi)	μ^{a} (°F/psi)	$\frac{-\mu^{a} C_p}{\phi}$	μ^{b} (°F/psi)	$\frac{\mu^{b} C_p}{\phi}$
152.2	0	.475	-.0440	.0859	.927	.0856	.924
	200	.509	-.0490	.0979	1.017	.0978	1.016
	400	.558	-.0546	.1030	1.053	.1038	1.061
	600	.635	-.0620	.1039	1.064	.1068	1.093
251.3	0	.530	-.0321	.0592	.977		
	200	.547	-.0338	.0614	.994		
	400	.570	-.0353	.0627	1.012		
	600	.597	-.0367	.0633	1.030		
	800	.628	-.0379	.0627	1.039		
	1000	.664	-.0385	.0608	1.049		
	1250	.712	-.0378	.0569	1.072		
	1500	.754	-.0356	.0525	1.112		

a Interpolated with respect to composition using values of μ from Budenholzer, et. al (11).

b Values for μ for a 51.1 mole percent propane in methane mixture from Head (50).

Joule-Thomson results for propane[125] in order to obtain interpolated results for the 77 percent propane in methane mixture. The results for propane[125] extend only to moderate pressures (600 psia) but can be extended to elevated pressures by applying Equation (8) with values of C_p and φ from this investigation. These values were interpolated with respect to composition to establish the values reported for a 76.6 mole percent propane in methane mixture in Table XXXI. Also listed in this table are values of C_p and φ for the 76.6 mole percent mixture from this investigation and the ratio $-\mu C_p/\varphi$. These interpolated values of μ are consistent with experimental results of this investigation to ±3 percent.

Phase Behavior

A number of independent investigators have reported data on the vapor-liquid equilibrium of the methane-propane system.[1,109,111,126] Data from these sources were used to estimate bubble and dew points corresponding to the pressures of investigation for both the 76.6 and the 50.6 mole percent propane in methane mixtures. These results are included in Tables XVII and XXV for the 76.6 and 50.6 mole percent mixtures, respectively. The results indicate not only the variation among the various investigators, but give some indication of the accuracy of the vapor-liquid equilibrium measurements made in the course of the present investigation.

Enthalpy of the Methane-Propane System

The isobaric effect of temperature and the isothermal effect of pressure on the enthalpy have been obtained for propane, a 76.6 mole percent and, a 50.6 mole percent propane in methane mixture. Skeleton enthalpy tables and diagrams have been presented for these mixtures.

TABLE XXXI

Test of Consistency of Data Based on Equation (8)

Temperature (°F)	Pressure (psia)	C_P (Btu/lb °F)	$\emptyset$ (Btu/lb psi)	μ (°F/psi)	$\frac{-\mu C_P}{\emptyset}$
	0	0.486	-0.0438	--	--
201°F	200	0.517	-0.0519	0.1034[a]	1.030
	400	0.579	-0.0633	0.1115[a]	1.02
	0	0.514	-0.0376	--	--
250°F	200	0.541	-0.0430	0.0807[b]	1.015
	400	0.577	-0.0483	0.0837[b]	1.000
	600	0.631	-0.0543	0.0874[b]	1.016
	800	0.705	-0.0618	0.0874[b]	0.997

a - Interpolated with respect to composition using values of μ for propane from (125) and for methane-propane mixtures from (11)

b - Interpolated with respect to composition using values of μ for the methane-propane mixtures from (11) and values of μ for propane calculated using Equation (8) and data for C_p and $\emptyset$.

8

In addition both isothermal and isobaric effects on enthalpy have been measured by Manker[80] and Mather[85] for a 5.1 mole percent propane in methane mixture, and an enthalpy diagram has been constructed by Mather.[85] The isobaric effect of temperature on the enthalpy for a 12 and 28 mole percent mixture has been measured also by Mather.[85] In constructing the enthalpy diagram for the 11.7 mole percent mixture Mather[85] used the averaged data of Dillard[32] and the BWR equation of state with the original constants[7,8] and mixing rules to determine the effect of pressure on enthalpy. These results were in agreement with each other for the 11.7 mole percent mixture, and for the 5.1 mole percent propane in methane mixture, both results agreed with the data of Mather.

For the 28 mole percent mixture, however, no experimental data were available and Mather used the BWR equation of state alone above the two-phase region to establish the effect of pressure on enthalpy. His justification was based on the agreement of the equation with experimental data for the 5.2 and 11.7 mole percent mixtures. Figure 56 shows the smoothed enthalpy departures obtained above the two-phase region for the 50.6 mole percent mixture at 152.2 and 251.3°F. The points are values of the enthalpy departure obtained using the BWR equation of state. The agreement of the BWR equation with the experimental data is excellent for both isotherms (less than 0.5 Btu/lb or 1 percent). Since the BWR equation with the original mixing rules agrees with experimental data for both an 11.7 and 50.6 mole percent mixture, the use of the equation by Mather is most probably valid for the effect of pressure on enthalpy for the 28 percent mixture. Thus, accurate results for enthalpy as a function of pressure and temperature are available for a 5.2, 11.7, 28.0, 50.6, and a 76.6 mole percent propane in methane mixture and complete the

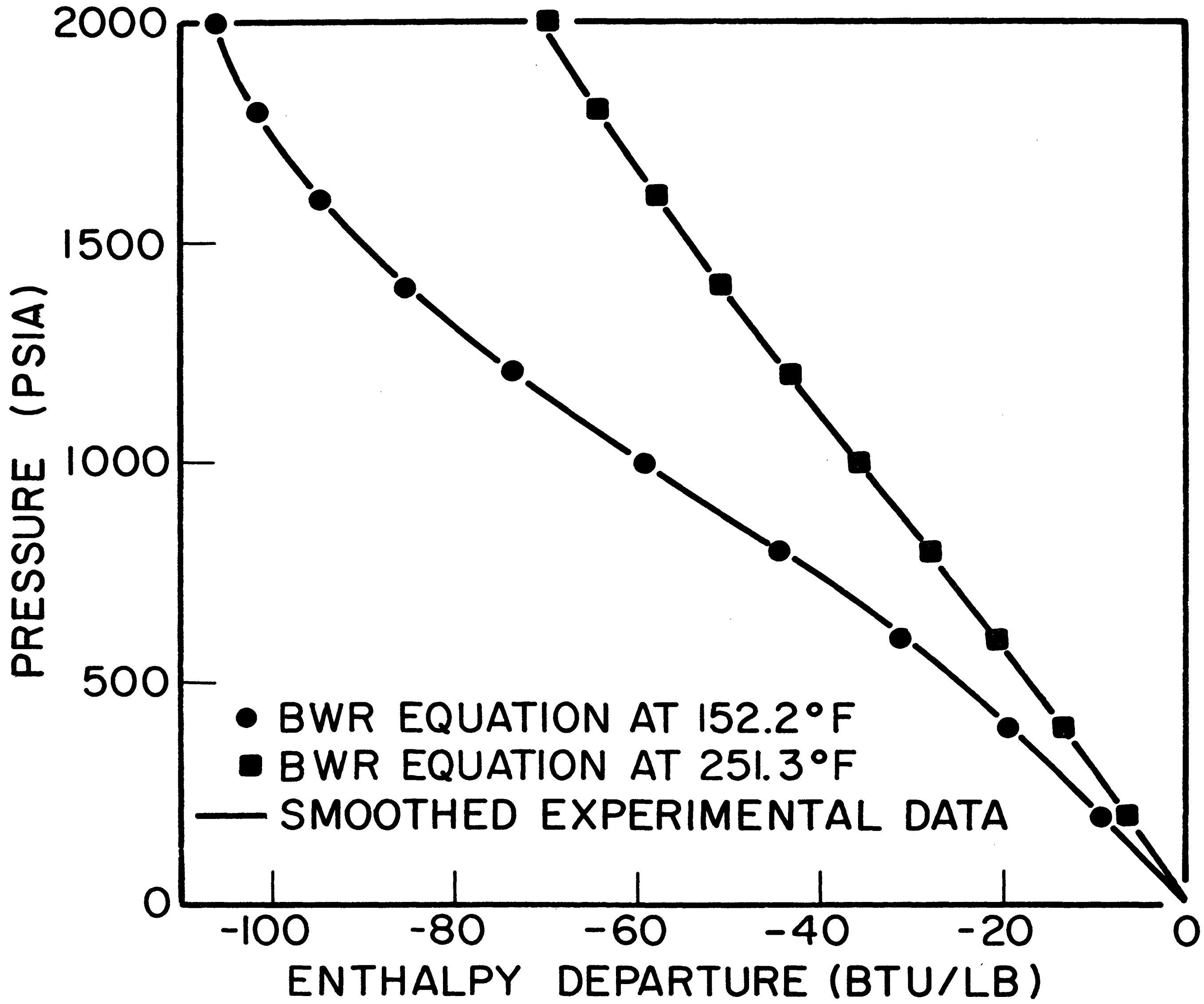


Figure 56. Comparison of Experimental Isothermal Enthalpy Departures with the BWR Equation of State for the 50.6 Percent Mixture

work on the methane-propane system.

SECTION VI - EVALUATION AND EXTENSION OF METHODS OF PREDICTION

The thermal data obtained for the methane-propane system and other data obtained at the Thermal Properties of Fluids Laboratory permits a direct evaluation and an extension of the current methods of prediction of the enthalpy of fluid mixtures at elevated pressures, which were discussed in Section II. This section presents the results of comparison studies and an extension of the corresponding states principle which was suggested from the results of these studies.

The various methods of prediction were compared with experimentally determined enthalpy data. The results of these comparisons indicated that the method of corresponding states looked most promising for representing enthalpy behavior of fluid mixtures. Empirical mixing rules were obtained for a three-parameter corresponding states correlation which accurately represented the enthalpy of the methane-propane binary system. The mixing rules were compared with generalized rules available in the literature and the technique extended to a mixture of nitrogen in methane.

Comparisons of Methods of Prediction

There have been several comparison studies of methods of prediction of mixture enthalpies. Most have been based mainly on volumetric data. The most extensive of these is the one conducted by the American Petroleum Institute.[140] The recommendations of the authors, however, are restricted for mixtures due to the lack of direct experimental data available when the study was conducted. Since that time there have been several papers which have compared methods of prediction of enthalpies of mixtures which are based on the more recent thermal data. These include Mather et al.,[87] Barner and Schreiner,[6] Wiener,[144] Findley et al.,[41] Sehgal et al.,[132] and Yesavage et al.[151] All of the above comparisons have used experimental data obtained at the Thermal Properties of Fluids Laboratory. The last two of the above which are the most extensive, were conducted as a part of the present investigation and are summarized in the following pages.

Two separate techniques were used to make comparisons of the data with methods of prediction. The techniques used were determined by the data available at the time comparisons were made. The first set of comparisons were based on enthalpy differences resulting from 100°F isobaric temperature differences. This comparison was based on isobaric data made available for mixtures before operation of the isothermal calorimeter. In the second set of comparisons results for isothermal enthalpy departures were utilized, basing the comparisons on the most recent data.

Isobaric Enthalpy Differences

The data available and used in the first set of comparisons includes the results of isobaric determinations made on the following mixtures:

Nominal 5 percent propane in methane; Manker et al.,[80,81] Mather.[85]

Nominal 12 percent propane in methane; Mather,[85] Mather et al.[86]

Nominal 28 percent propane in methane; Mather,[85] Mather et al.[87]

Nominal 43 percent nitrogen in methane; Mather,[85] Mather et al.[87]

Nominal 25 percent helium in nitrogen; Mage,[77] Mage and Katz.[79]

Nominal 50 percent helium in nitrogen; Mage,[77] Mage and Katz.[79]

Most of the data from the references cited above have been obtained with an isobaric flow calorimeter in the temperature range from -250°F to about 250°F. Some data have been extended to 250°F from about 100°F using low pressure C_p data and corrections for pressure calculated from the BWR equation. This extrapolation was made only when experimental values of C_p at 100°F agreed with predicted values within 1/2 percent and showed the same trend with temperature.

Six methods of prediction were compared with the above data. These included three three-parameter corresponding states methods. The correlations of Curl and Pitzer,[24] and modified by Yarborough,[147] the correlation of Yen[149] and that of Lydersen, Greenkorn and Hougen.[76] The BWR equation of state with the original constants[7] and mixing rules[8] was applied[96] as an example of an equation of state which was compared with experimental data. Finally, comparisons were made with the empirical methods of Peters[99] and Canjar and Peterka[15] for hydrocarbon mixtures.

In applying these correlations to predict enthalpy changes, pseudocritical properties and values of the correlating parameters for the mixtures must be estimated. A variety of mixing rules have already been suggested. The simplest of these are the linear mixing rules based on mole fraction as originally suggested by Kay[65] for estimation of pseudocritical temperatures and pressures. This rule has been extended to estimation of the third parameter for mixtures. Pitzer and Hultgren[101] make use of a quadratic term in predicting correlating parameters for mixtures. Gunn, Chueh and Prausnitz have suggested mixing rules applicable to mixtures containing quantum gases.[48]

For the helium-nitrogen mixtures, the mixing rules of Gunn, Chueh and Prausnitz[48] were used for all mixtures. In applying the methods of Lydersen, Greenkorn and Hougen,[76] and Yen[149] for the remaining mixtures linear mixing rules[65] were applied. In connection with the Curl and Pitzer method[24] the mixing rules suggested by Pitzer and Hultgren[101] were applied for the methane-propane mixtures and those of Prausnitz and Gunn[106] were used with the methane-nitrogen mixture.

As mentioned previously, several of the procedures use a combination of experimental enthalpy data at low pressures with estimations of the effect of pressure on enthalpy. In order to make these comparisons as meaningful as possible, common sets of low pressure data were used for all methods. For the light hydrocarbons, values were taken from the compilation of API Project 44.[118] Values for nitrogen are those of Goff and Gratch.[46] For helium, calculations were made using $C_p^0 = (5/2)R$.

To estimate enthalpy changes within and in the vicinity of the two-phase region of a mixture, it is essential that the limits of this region be established accurately. For purposes of obtaining the most meaningful comparison the experimental data of Price and Kobayashi[109] were used to establish these limits for the methane-propane mixtures, the data of Bloomer and Rao[9] were used for the methane-nitrogen mixture and those of DeVaney et al.,[29] and Mage[77] for the helium-nitrogen mixture.

For purposes of comparison isobaric enthalpy changes at pressures of 500, 1000, 1500 and 2000 psia were calculated for five temperature intervals of 100°F and compared with experimental data. The pressures selected were those for which actual experimental enthalpy determinations were made. The large temperature interval of 100°F was selected to dampen out regions of rapid change in the heat capacity and to reduce round off error involved in reading charts. A percentage deviation was defined as:

$$\frac{\left[\underline{H}(T_2)-\underline{H}(T_1)\right]_{P(\text{predicted})}-\left[\underline{H}(T_2)-\underline{H}(T_1)\right]_{P(\text{experimental})}}{\left[\underline{H}(T_2)-\underline{H}(T_1)\right]_{P(\text{experimental})}}.$$

Results of the comparison are presented in graphical form for the six mixtures in Figure 57. The figures represent the percentage deviation on a pressure-temperature diagram. Lines corresponding to zero percent deviation between calculated and experimental values are sketched in much the same manner one might draw a contour line from survey determinations on a topographic map. In a similar manner contour lines corresponding to ± 5 percent, ± 10 percent and ±20 percent deviation were sketched in. Suitable coding was developed to distinguish these regions and to identify regions in which comparison was not possible.

From the results of the comparison several observations can be noted.

The procedures based on use of molal average boiling point by Canjar and Peterka[15] and partial molal diagrams by Peters[99] yield results which are comparable to other methods. However, they are limited to a smaller region of pressure and temperature and to fewer systems than the other methods as was mentioned in Section II.

The BWR equation of state (using the original constants derived from a variety of data) is adequate in the gaseous and critical region but predicts erroneous results in the liquid region. This results, in part, from the fact that limited data were available in this region when the original constants were evaluated and illustrates that empirical equations of state should not be used for purposes of extrapolation. Again it is

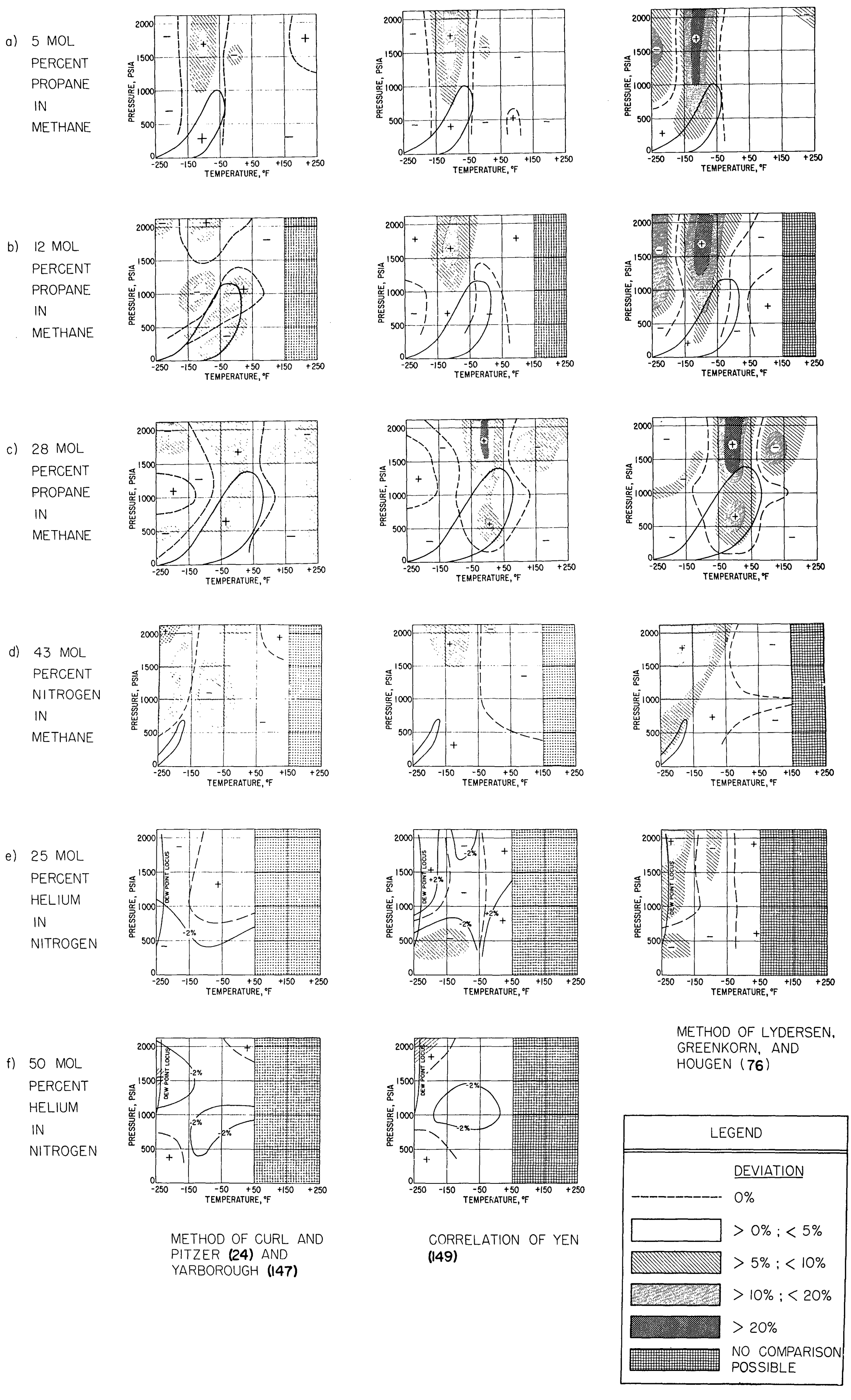


Figure 57. Comparison of Isobaric Enthalpy Differences with Numerous Methods of Prediction

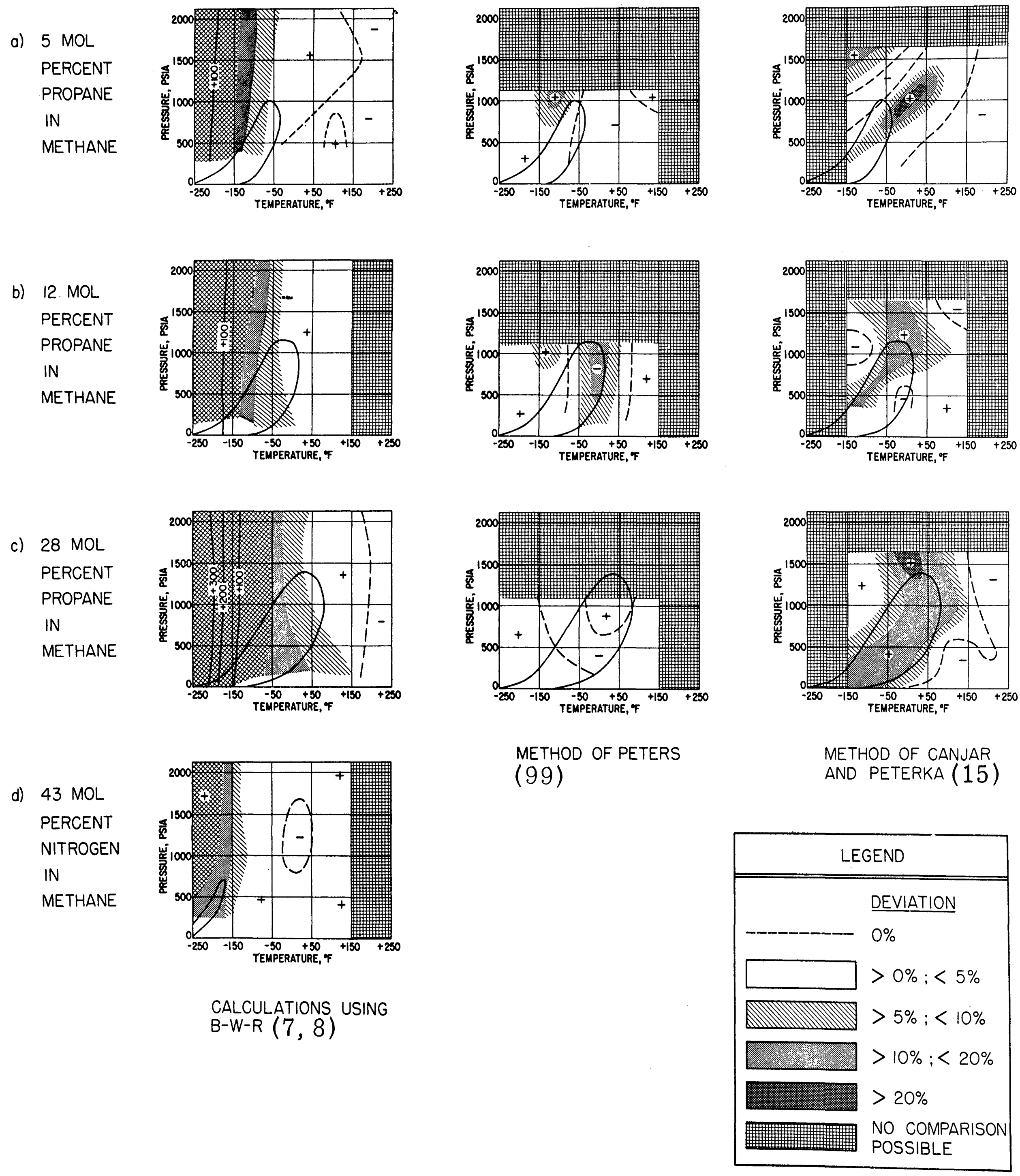


Figure 57. Comparison of Isobaric Enthalpy Differences with Numerous Methods of Prediction

more limited with regard to the number of systems which can be predicted than the corresponding states methods.

Procedures based on the principle of corresponding states can be applied with some degree of confidence to predict enthalpy departures for mixtures of nonpolar compounds over a wide range of temperatures and pressures. The recent correlation of Yen[149] yields results which are slightly better than those obtained from the correlation of Curl and Pitzer[24] and represent a considerable improvement over that of Lydersen, Greenkorn, and Hougen.[76] This is not too surprising since the latter two correlations are based primarily on PVT data whereas the former incorporates enthalpy data.

Enthalpy Departure Comparisons

In this set of comparisons the results of the investigation of the 76.6 mole percent propane in methane system of Table XXVIII was used as an example to determine enthalpy departures. These departures were compared with various methods of prediction with more emphasis on corresponding states correlations. Again the methods compared with include the BWR equation, the methods of Canjar and Peterka[15] and Peters.[99] The same three corresponding states methods were studied in this comparison as in the previous one. For the correlations of Yen,[149] and Yarborough's[147] extension of Curl and Pitzer,[24] two-mixing rules, Kay's rule[65] and the Pitzer-Hultgren rule,[101] were used to permit comparison with each other. In addition an equation of state (Hirschfelder, Bueller, McGee, Sutton[53,54]) which is based on an extended theorem of corresponding states was also studied.

In making comparisons in the two-phase region it is necessary to

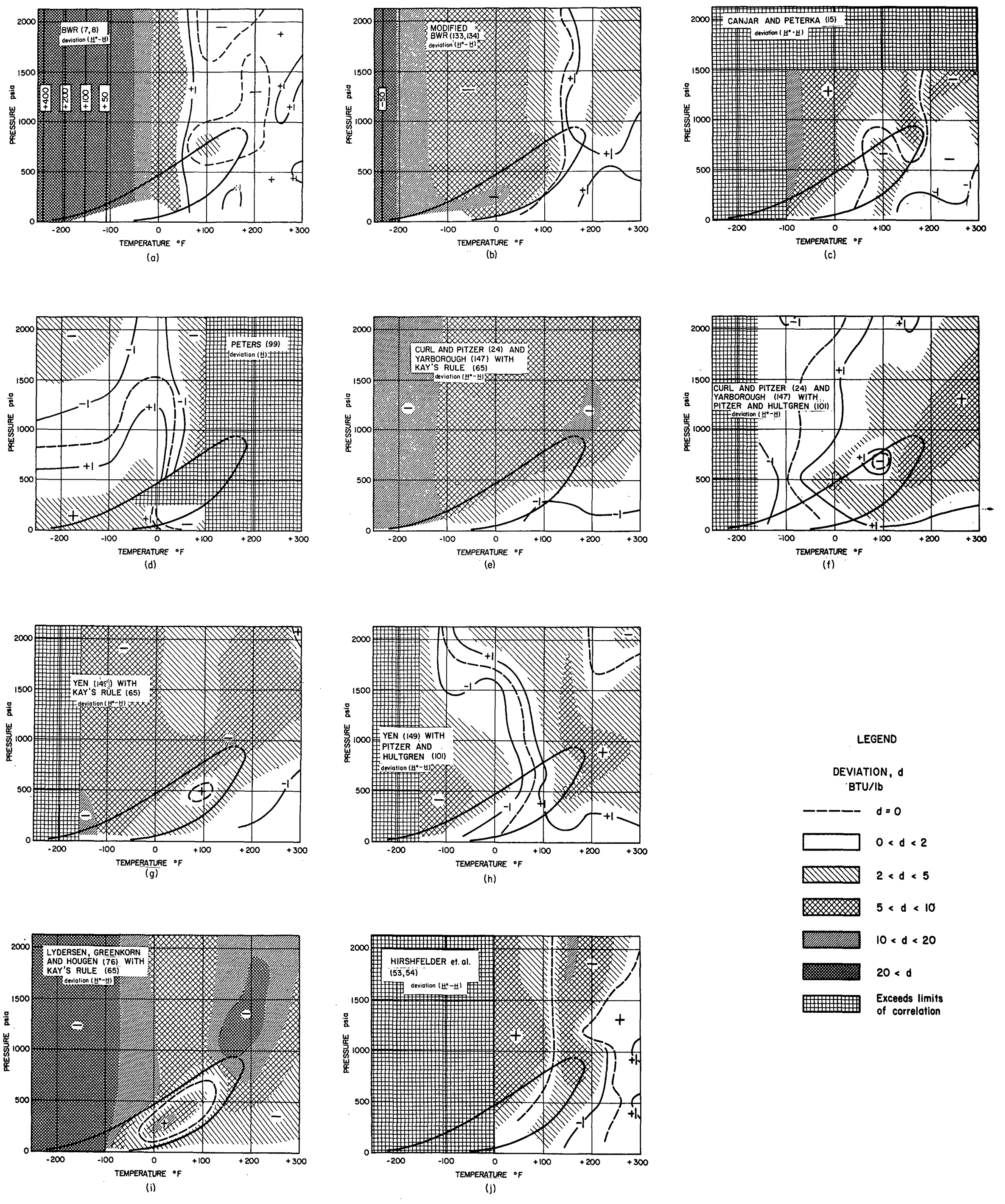


Figure 58. Comparison of Isothermal Enthalpy Departures with Numerous Methods of Prediction for the 77 Percent Mixture

make some sort of flash calculation, and the NGPA Chao-Seader program[145] and data in the literature were used to generate the vapor-liquid equilibrium data. In using the method of Peters[99] which is not based on an enthalpy departure it was necessary to determine the departure by substraction.

Results of the comparisons are presented in graphical form in Figure 58. These graphs were obtained by plotting the difference between the experimental and calculated enthalpy departures.

$$(\underline{H} - \underline{H}^0)_{exp} - (\underline{H} - \underline{H}^0)_{calc} \equiv \text{deviation } (\underline{H})(\text{Btu/lb}) \qquad (78)$$

on a pressure-temperature diagram. Percentage errors between any two points on the pressure-temperature diagram can be obtained by referring to the enthalpy table for the 76.6 percent mixture (Table XXVIII).

Figures 58a and 58b show the results of the comparisons of the data with the original BWR equation[7,8] and a recently modified BWR equation.[133,134] Both results indicate good agreement in the gaseous and critical region but much poorer agreement for the liquid.

Figures 58c and 58d show the results of the comparisons for the correlations of Canjar and Peterka[15] and Peters.[99] Again both methods show fair agreement but are limited in their region of application. Figures 58e and 58f are the results for the Curl and Pitzer[24] and Yarborough[147] method using two different mixing rules. Figures 58g and 58h are the results for the Yen correlation[149] using two different rules. It appears that both of these correlations show a distinct improvement when used with the nonlinear Pitzer-Hultgren mixing rule as opposed to Kay's rule. For this mixture the method of Curl and Pitzer, and Yarborough is somewhat better than that of Yen and both are better than that of Lydersen, Greenkorn, and Hougen[76] (Figure 58i). Figure 58j is a comparison of the data with the equation of state of Hirschfelder et al.[53,54] It is an empirical corresponding states equation with up to five parameters. It is fit by three analytical functions which are forced to conform at their boundaries. By comparing Figure 58j with Figures 58f and 58h it appears that a five parameter corresponding states analytical equation of state cannot reproduce the data as accurately as a three parameter corresponding states tabular function. This implies that the reduced enthalpy is (to a reasonable degree of accuracy) a function of pseudoreduced temperature, pressure, and a third parameter. However, when trying to fit the tabulated function to one or a set of analytical equations it is at present necessary to add additional parameters. Even with these additional parameters the results of the equation of state do not compare as favorably as the tabulated correlations with experimental data. This is also demonstrated by Figure 58a and 58b where the eight constant BWR equations very poorly reproduce the enthalpy departure in the liquid region.

Conclusions

A goal of this research was to extend methods of prediction to the point of accurately reproducing the enthalpy data obtained in this investigation with the hope of using an approach that would result in considerable generalization. In light of this goal and the results of the comparison studies it was decided to develop a three parameter corresponding states tabular correlation to describe the behavior of the enthalpy departures of the methane-propane system.

There were many considerations which influenced this decision. In the first place, methods such as Peters[99] and Canjar and Peterka,[15] although reproducing the data quite effectively, do not lend themselves easily to extension and generalization. This is in part due to the fact that there is no fundamental justification for such an approach other than success in their region of application.

Equations of state also can be made to reproduce the data accurately over their region of application. However, they also cannot be readily generalized to systems for which data are not available. A more fruitful approach appears to be a generalized equation of state based on the corresponding states principle. However, at present, the use of an analytical form appears to require the use of additional parameters.

A corresponding states graphical correlation has numerous advantages. It can represent the results to a reasonable degree of accuracy and at the same time has some grounds for fundamental validity. It can be used to represent the enthalpy departure of systems for which data, other than a knowledge of the parameters, is not available. It is generally valid over a wide range of conditions, and can be improved for mixtures by the use of improved mixing rules. Finally, if an analytical form is desirable, the tabular functions themselves can eventually be fit by one or a series of equations. For these advantages a three parameter corresponding states tabulation was utilized.

Application of the Corresponding States Principle to Fit Experimental Enthalpy Data

In applying the corresponding states principle it was first necessary to develop a tabular function of the reduced enthalpy departure. This was accomplished using Equation (44) with reference substance enthalpy departure functions, $\left[\frac{\underline{H}^0-\underline{H}}{RT_c}\right]_1$ and $\left[\frac{\underline{H}^0-\underline{H}}{RT_c}\right]_2$, developed from enthalpy data for methane and propane. Next, the corresponding states principle was tested for pure components by comparison of calculated results with experimental data for nitrogen. The reference substance tables were interpolated with respect to reduced temperature and pressure to determine the two reduced enthalpies. Next, Equation (44) was used to determine the departure for nitrogen, and the results compared with experimental data. After proving to be successful for pure components, the results were extended to the methane-propane system by developing mixing rules which best fit the experimental data. First an optimization technique was developed which would search to find a set of optimum values over a wide range of conditions of the three parameters for a given mixture. The individual optimum values of each parameter for each mixture were then fit with respect to composition to obtain smooth mixing rules.

Development of the Reference Substance Enthalpy Departures

Reduced enthalpy departure tables are available from most three parameter corresponding states correlations. These tables have, however, been generated in the most part from volumetric data and may be of questionable accuracy. Since accurate enthalpy data have been obtained at the Thermal Properties of Fluids Laboratory for methane and propane, it was decided to develop reference reduced enthalpy functions from these results.

In generating the tabular reduced enthalpy function for propane the values of enthalpy, temperature, and pressure presented in Table XII were used to determine reduced quantities: $\frac{\underline{\underline{H}}^0 - \underline{\underline{H}}}{RT_c}$, T/T_c and P/P_c. The resulting reduced table is presented as Table LXII of Appendix D.

The second substance reduced enthalpy departure table was based on values of enthalpy for methane as reported by Jones.[62] In obtaining Jones' table his isobaric heat capacity data were used to determine the effect of temperature on enthalpy at numerous pressures. The effect of pressure on enthalpy was determined by the BWR equation of state with the original constants.[7] Recently accurate volumetric results for methane by Douslin et al.[35] have been used in Equation (38) to determine the effect of pressure on enthalpy for methane at 32°F.[34,48a] The values for enthalpy departure at 32°F obtained from both the BWR and Douslin are plotted on Figure 59. This figure shows a significant difference between the two results especially at high pressure. The differences are reported in Table XXXII. Since it is believed that the differentiated volumetric data are likely to be more accurate than the BWR equation with the original constants, the Jones' table was adjusted. This was accomplished by adjusting the enthalpy at all pressures in the table by the difference between the results of Douslin[34,48a] and the BWR results. The resulting corrected table amounts essentially to the table that Jones would have obtained had he used the departures of Douslin[34,48a] instead of the BWR. This table was put in reduced form and the results are presented in Table LXIII of Appendix D.

Values of the three parameters used in this investigation for methane and propane as well as those for nitrogen are the ones presented in the NGPSA data book and are listed in Table XXXIII. The range of the methane table is for $0.52 \leq T_r \leq 1.49$ and $P_r \leq 2.99$ and the propane

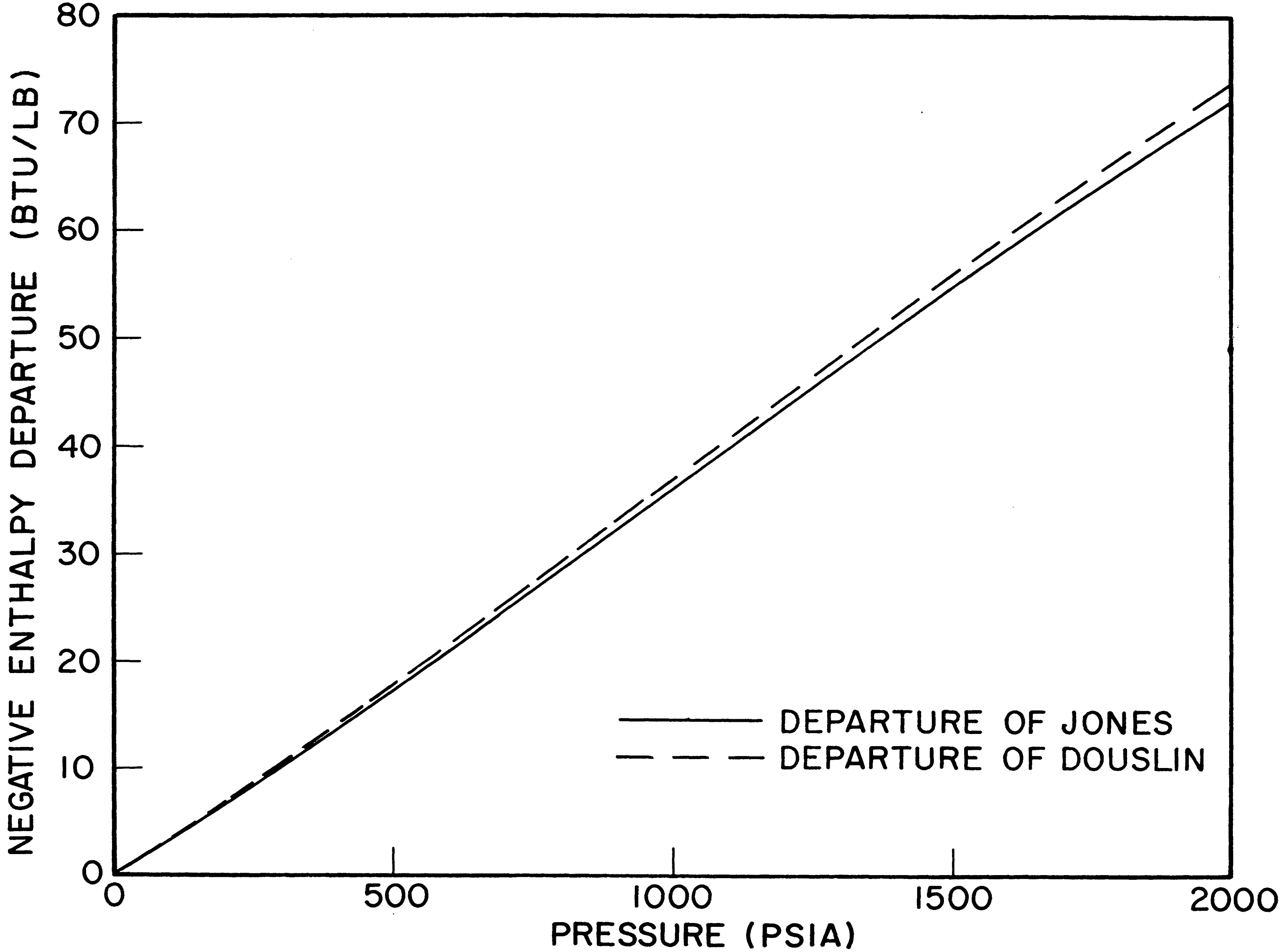


Figure 59. Comparison of Enthalpy Departure of Methane at 32°F Obtained Using Original BWR Equation (7) with Results Calculated by Douslin (34,35)

TABLE XXXII

CORRECTION MADE IN JONES' TABLE TO AGREE WITH DOUSLIN'S ENTHALPY DEPARTURE

Pressure psia	[H(Jones)-H(Douslin)] Btu/lb
50	0.00
100	0.18
150	0.27
200	0.35
250	0.39
300	0.42
350	0.49
400	0.54
450	0.52
500	0.40
550	0.62
600	0.58
625	0.64
680	0.67
700	0.64
800	0.56
900	0.70
1000	0.82
1200	0.94
1500	1.45
2000	1.64

table for $0.27 \leq T_r \leq 1.44$ and $P_r \leq 3.24$. In order to obtain departures for a third substance both tables must be applied and, therefore, the range of application of the present method is $0.52 \leq T_r \leq 1.44$ and $P_r \leq 2.99$. This includes much of the liquid and gaseous region as well as the region around the critical. The method can be applied in the two-phase region for mixtures if vapor-liquid equilibrium data is generated from some other source.

Application of Reference Substance Equations to Pure Components

If the three parameter corresponding states principle is valid for non-polar fluids then it should be possible to predict the enthalpy departure for a substance such as nitrogen from the tabulated departures developed for methane and propane and the three additional parameters only. An enthalpy table developed from experimental data has been presented by Mage *et al.*[77,78] and from it enthalpy departures were obtained for comparison. This procedure permitted a check of the three parameter corresponding states theory and the accuracy of the experimentally determined reduced functions.

A third order interpolating polynomial was used to determine reduced enthalpy departures from the methane table and the propane table for a given reduced temperature and pressure. The values of reduced enthalpy from the two functions along with a value of the third parameter, W, were used in Equation (44) to obtain enthalpy departures for nitrogen. The values of the three parameters used for nitrogen are listed in Table XXXIII. Equation (48) was used to determine the value of W from ω. As can be seen from Table XXXIII the third parameter for nitrogen falls between methane and propane about 1/3 of the way from methane.

TABLE XXXIII

PARAMETERS USED IN CORRESPONDING STATES CALCULATIONS

Component	Critical Temperature (°R)	Critical Pressure (psia)	Accentric Factor
Methane	343.3	673.1	0.010
Propane	666.0	617.4	0.152
Nitrogen	226.9	492.9	0.047

Enthalpy departures were obtained for nitrogen throughout the region of application of the reduced tables. Calculations were obtained at pressures of 500, 1000, and 1500 psia and at 20°F temperature intervals between -300°F and -120°F.

A computer program subroutine (DEV) was used to perform the numerical calculations. The subroutine searches the reduced enthalpy tables to find the grid points which center a given data point. After selecting the points on which the interpolation is to be based a second subroutine is called which uses Newton's third order interpolating polynomial. The program is called four times to interpolate with respect to reduced temperature and once with respect to reduced pressure for each table to determine enthalpy departures. This subroutine was essentially the one given by Carnahan et al.[16] and converted to the FORTRAN language. The calling subroutine then calculates the enthalpy departure for the points and also the deviation and percent deviation between the calculated and experimental value.

The subroutine repeats the process for the next input condition. When the departures and deviations are calculated for all of the conditions desired, the root mean square difference between the experimental and calculated enthalpy departures is determined. A simplified main program can be used to transmit data and print results. As will be discussed later, however, this subroutine was used as part of the optimization scheme applied for mixtures. The complete program is given in Table LX of Appendix C.

A major disadvantage of the calculation procedure is its inability to correctly interpolate (or extrapolate) the tables in the vicinity of two-phase region. The search procedure arrives at a series of grid points on which the interpolation is based which contain both gaseous and liquid points. Since there is a natural discontinuity at the two-phase region which the interpolation does not consider, the interpolation is likely tc be in error. This is, however, a limitation only for pure components, since for mixtures the two-phase envelope, in which the mixture is not stable as a single phase, extends to prohibit the possibility of calculations near the single component reduced two-phase region.

For the 32 conditions which were compared for nitrogen, the sum of the squares deviation is 0.52 Btu/lb. The maximum deviation is 1.5 Btu/lb but this occurred at -120°F (T_r = 1.50) slightly outside of the region of application of the tabulated functions. The results of this comparison are presented in Table XXXIV. The agreement is good especially when considering the possible uncertainties of 1 Btu/lb present in each of the tables for the three pure components. Thus, in this instance it appears that the three component principle of corresponding states can be used to predict enthalpy departures for non-polar pure components almost to the degree of uncertainty of the measured quantities. The parameters that are used are quantities that have definite physical meaning and thus knowledge of these quantities alone is enough to

TABLE XXXIV

TEST OF THREE PARAMETER CORRESPONDING STATES PRINCIPLE USING DATA FOR NITROGEN

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	159.5999908	-81.6999817	-82.0645905	0.3646088	-0.4462774
500.0000000	179.5999908	-76.6000061	-76.9572449	0.3572388	-0.4663691
500.0000000	199.5999908	-70.3999939	-70.9122925	0.5122986	-0.7276967
500.0000000	219.5999908	-60.8000031	-61.5042419	0.7042389	-1.1582870
500.0000000	239.5999908	-21.7000122	-22.2260132	0.5260010	-2.4239655
500.0000000	259.5998535	-16.3000031	-16.1213684	-0.1786346	1.0959177
500.0000000	279.5998535	-12.6999969	-13.0271521	0.3271551	-2.5760241
500.0000000	299.5998535	-10.6000061	-10.9802246	0.3802185	-3.5869646
500.0000000	319.5998535	-9.0000000	-9.4181461	0.4181461	-4.6460676
500.0000000	339.5998535	-7.8999939	-6.4066486	-1.4933453	18.9031067
1000.0000000	159.5999908	-80.9999847	-80.7055817	-0.2944031	0.3634606
1000.0000000	179.5999908	-76.0000000	-75.8509827	-0.1490173	0.1960754
1000.0000000	199.5999908	-70.8000031	-70.5439453	-0.2560577	0.3616634
1000.0000000	219.5999908	-64.3999939	-64.1284637	-0.2715302	0.4216306
1000.0000000	229.5999908	-60.1000061	-60.2589264	0.1589203	-0.2644264
1000.0000000	239.5999908	-55.8000031	-55.4396057	-0.3603973	0.6458733
1000.0000000	259.5998535	-41.3000031	-41.4095459	0.1095428	-0.2652369
1000.0000000	279.5998535	-30.1000061	-30.1832428	0.0832367	-0.2765338
1000.0000000	299.5998535	-23.6000061	-23.8399963	0.2399902	-1.0169067
1000.0000000	319.5998535	-19.6999969	-19.8444977	0.1445007	-0.7335063
1000.0000000	339.5998535	-16.8999939	-17.2372894	0.3372955	-1.9958315
1500.0000000	159.5999908	-80.5999756	-79.6302338	-0.9697418	1.2031536
1500.0000000	179.5999908	-75.9000092	-75.1092224	-0.7907867	1.0418787
1500.0000000	199.5999908	-70.5999908	-70.2011871	-0.3988037	0.5648777
1500.0000000	219.5999908	-65.3999939	-64.7862701	-0.6137238	0.9384155
1500.0000000	229.5999908	-62.1000061	-61.7879486	-0.3120575	0.5025077
1500.0000000	239.5999908	-58.8000031	-58.7319031	-0.0681000	0.1158162
1500.0000000	259.5998535	-51.1000061	-50.9752350	-0.1247711	0.2441704
1500.0000000	279.5998535	-42.1999969	-42.7993927	0.5993958	-1.4203682
1500.0000000	299.5998535	-35.0000000	-35.2981567	0.2981567	-0.8518763
1500.0000000	319.5998535	-29.0999908	-29.4218597	0.3218689	-1.1060781
1500.0000000	339.5998535	-24.8999939	-23.7418671	-1.1581268	4.6511126

interpolate accurate reference substance functions to accurately predict enthalpies of fluids under pressure.

Application of the Correlation to Mixtures

When attempting to apply this technique to mixtures, the critical properties of the mixtures cannot be used in the reduced enthalpy function, since mixture critical properties do not have the same physical and thermodynamic significance as those for the pure components. In addition for nonconformal mixtures, such as mixtures of methane-propane, the principle of corresponding states has at present no funcamental justification. This also eliminates the possibility of developing mixing rules from relatively rigorous theoretical analysis. Thus, an empirical approach is essential. It was, therefore, assumed that the reduced enthalpy of methane-propane mixtures does behave the same as that of a pure substance. Empirical mixing rules containing a total of six constants were determined which represent the methane-propane data to within ±1Btu/lb which is of the order of the uncertainty of the data. The success of these results justified the validity of the corresponding states principle when applied to mixtures of this type.

The pure component reference substance enthalpy departures obtained from the methane and the propane enthalpy data were used to calculate mixture enthalpy data for values of the three parameters of the given mixture. These parameters were determined by a direct search procedure which found values of the three parameters which minimized the root mean square difference between the calculated and experimental departures for each mixture. Calculations were made with the aid of the computer program previously mentioned.

The program assumes initial values of T_{cx}, P_{cx}, and W_x. The enthalpy departures are calculated using the subroutine described above for a wide variety of conditions including the gaseous, liquid, and critical regions. Input points were selected at pressures of 500, 1000, 1500, and 2000 psia where actual experimental determinations were usually made. Temperatures at 20°F intervals were used when the point fell in the single phase region. The range of temperature was determined usually by the range of the reference tables or occasionally by the limits of the experimental data. The root mean square deviation for all points was calculated. Root mean square deviations were then calculated with each parameter incremented in both the positive and negative direction. The base values of P_{cx}, T_{cx}, and W_x would shift to the point which had the lowest root mean square deviation. The procedure would continue until the root mean square deviation of the base value was lower than that of any of the six points surrounding it.

Due to limits on computer execution time minimum step sizes of $\Delta P_x = 3.0$ psia, $\Delta T_x = 1.5$°F and $\Delta W_x = 0.002$ were used. If initial conditions far away from reasonable values (Kay's rule is a normal "reasonable" initial value) were used the calculation would not necessarity converge to the optimum. This indicated that the function was not unimodal. However, reasonable but different initial conditions would usually converge to the same optimum. Due to the large computer execution times involved a more complete analysis of the different optimums was not undertaken. It was, however, noted that the minimum in the temperature direction was in a relatively steep valley. A gradient search was originally used, instead of the direct search, but it would not converge to the optimum. The valley in the temperature direction had too much of an effect on the gradient when using reasonable step sizes.

The technique was used to determine optimum parameters for all five experimentally determined methane-propane mixtures. The original optimum values, the sum of the squares deviations, and the number of data points used for each mixture are given in Table XXXV. The actual comparisons for each data point are given in Table LXIV of Appendix D for all five mixtures. The Δ's represent departures from Kay's rule as for example pseudo critical temperature.

$$\Delta T_{cx} = T_{cx} - T_{cKay} \tag{79}$$

TABLE XXXV

ORIGINAL OPTIMUM VALUES FOR THE THREE PSUEDOPARAMETERS IN THE SEARCH CALCULATIONS

Mole Fraction Propane	P_{cx} psia	ΔP_{cx} psi	T_{cx} (°R)	ΔT_{cx} (°F)	W_x	ΔW_x	F Btu/lb
0.052	664.2	-6.0	361.5	1.7	0.032	-0.020	0.964
0.117	667.5	+0.9	388.0	7.0	0.175	0.058	0.776
0.280	654.0	-3.4	447.0	13.4	0.359	0.019	0.966
0.506	632.4	-12.5	521.5	14.6	0.618	0.108	0.709
0.766	6.914	-11.0	596.5	6.0	0.855	0.089	0.644

The three parameters are plotted on Figure 60 as deviations from Kay's rule. The figure indicates that the results for critical temperature are the smoothest and most regular. This is again due to the fact that the optimum pseudocritical temperature lies in a very steep valley and has the strongest effect on the magnitude of the error. It can also be noted that these points are not symmetrical and thus cannot be represented by a single empirical correction term as was suggested by Pitzer and Hultgren.[101] In order to fit the points to a smooth curve to within the value of the minimum step size, $\Delta T_{cx} = 1.5°F$, it was necessary to use a three constant empirical equation.

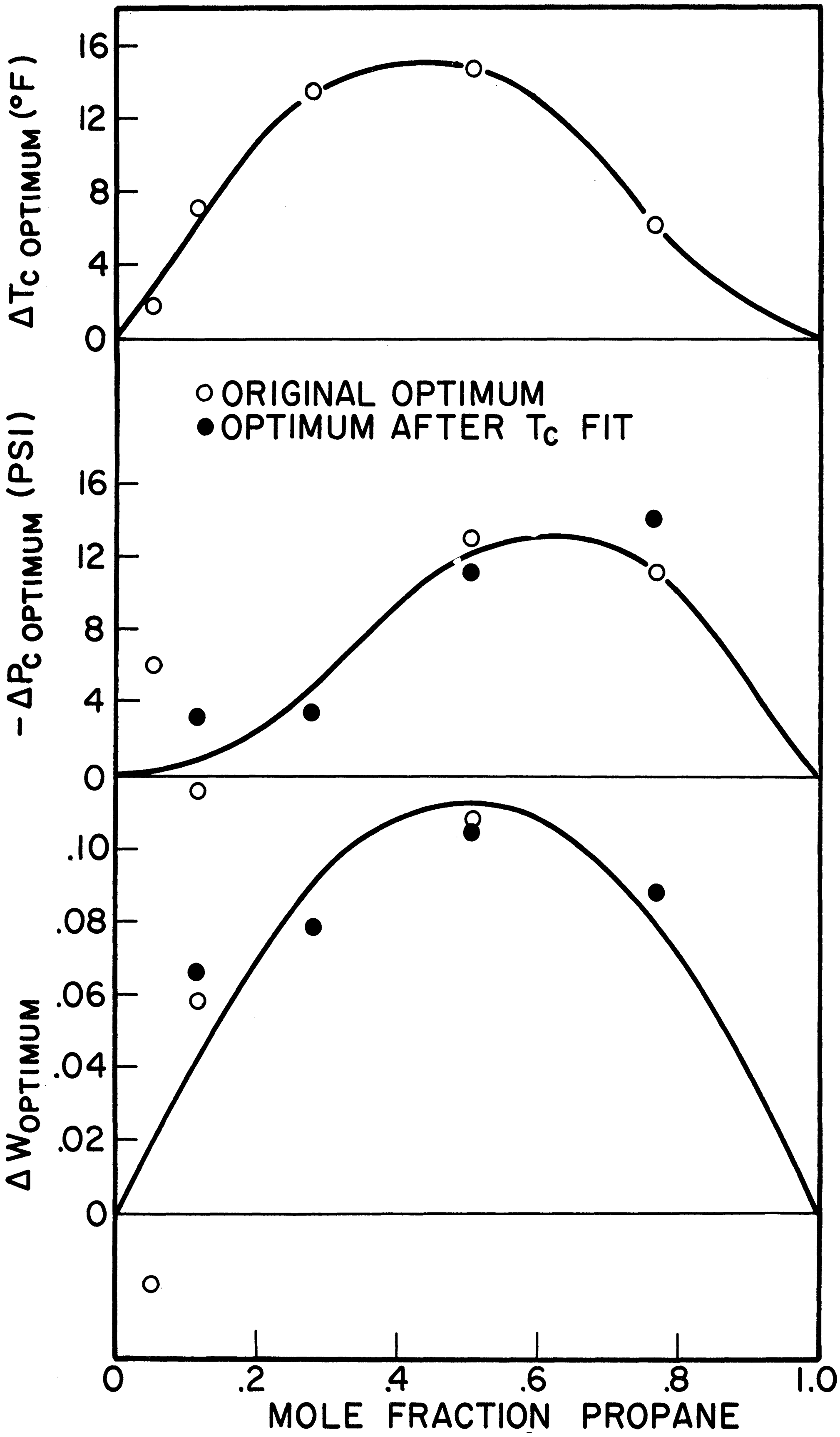


Figure 60. Optimum Mixing Rules for the Methane-Propane System

$$T_{cx} = \sum x_i T_{ci} + x_i x_j [A' + (1-2x_i)B' + (1-2x_i)^2 C'] \quad (54)$$

The values of the constants obtained by a least squares fit were:

$A' = 59.454$

$B' = 50.868$

$C' = -35.880$

The resulting fitting curve is shown on Figure 60. Since the root mean square deviations were so sensitive to small temperature changes it was felt that changes in the optimum critical temperature caused by the fit could significantly effect the optimum values of the remaining two parameters. Thus, values of T_{cx} were calculated by Equation (54) for the five mixtures and a second search procedure performed. Using these values for the pseudocritical temperature, P_{cx} and W_x for each mixture were allowed to vary and new optimum values found. The resulting optimum parameters along with the root mean square deviations are presented in Table XXXVI. The enthalpies calculated and compared with experimental results for each tested condition are presented in Table LXV of Appendix D.

The results of this procedure are shown as the solid points on Figure 60. Since there is considerable uncertainty in the results obtained for these two parameters, there was some question regarding the choice of fitting equation. Because these two parameters do not significantly effect the root mean square deviation it was decided to use simple equations for the fit. For P_{cx} it appears that the results show a definite asymmetry. Therefore, a two constant equation was used.

$$P_{cx} = \sum x_i P_{ci} + x_i x_i [D' + E' (1-2x_i)] \quad (55)$$

For W_x, however, any asymmetry, if present, cannot be observed due to scatter in the results. Therefore, the simple one constant equation was used.

$$W_x = \sum x_i W_i + x_i x_j F' \qquad (56)$$

The values of the constants were obtained by least squares fitting. The points for the 5.2 percent mixture differed significantly from the trend of the data for the other mixtures and were not used in determining the constants. The values of these constants are given as

$D' = 43.448$

$E' = -46.590$

$F' = 0.44250$

The parameters for each mixture were calculated using Equations (55) and (56) with T_{cx} from Equation (54). The fit parameters were then used to calculate enthalpy departures and compared with experimental results. The root mean square of the deviations are presented in Table XXXVII.

Fitting the calculated optimum values for the parameters does cause some increase in the average deviation for all of the mixtures. This is illustrated in Table XXXVIII where columns A, B, and C present respectively the root mean square deviations obtained using the original parameters, using the smooth pseudocritical temperature values with adjusted values of the other parameters, and using the final smooth mixing rules for each parameter. The worst case (except for the 5.1 percent mixture which was not used in the final fit) was for the 11.7 percent mixture where the root mean square deviation for all data points increased from 0.776 to 0.937.

TABLE XXXVI

OPTIMUM VALUES OF PARAMETERS AFTER FITTING PSEUDOCRITICAL TEMPERATURE AND HOLDING IT CONSTANT IN OPTIMIZATION

Mole Fraction Propane	P_{cx} psia	ΔP_{cx} psi	T_{cx} (°R)	ΔT_{cx} (°F)	W_x	ΔW_x	F Btu/lb
0.052	674.2	+14.4	362.6	2.8	0.012	-0.040	0.999
0.117	663.5	-3.1	387.5	6.4	0.183	0.066	0.703
0.280	645.0	-3.4	447.0	13.3	0.359	0.079	0.966
0.506	634.4	-10.5	521.7	14.8	0.610	0.104	0.720
0.766	617.4	-13.0	596.4	5.9	0.855	0.089	0.648

TABLE XXXVII

FINAL FIT OPTIMUM PARAMETER VALUES

Mole Fraction Propane	P_{cx} psia	ΔP_{cx} psi	T_{cx} (°R)	ΔT_{cx} (°F)	W_x	ΔW_x	F Btu/lb
0.052	670.1	-0.1	362.6	2.8	0.072	+0.021	2.100
0.177	665.7	-0.8	387.5	6.4	0.163	0.046	0.937
0.280	652.9	-4.6	447.0	13.3	0.370	0.089	1.010
0.506	633.9	-11.0	521.7	14.8	0.617	0.111	0.744
0.766	618.2	-12.2	596.4	5.9	0.845	0.079	0.709

The results of the final smooth mixing rules are also presented on topographic pressure-temperature charts for each mixture in Figure 61. Deviations are represented as Btu/lb as obtained from Equation (79). The tabulated, calculated, and experimental departures for each data point are presented in Table LXVI of Appendix D.

The root mean square deviations are on the order of 1 Btu/lb or less for all mixtures with the exception of the 5.2 mole percent propane in methane mixture. These deviations are in the range of uncertainty of the experimental data and thus, the proposed mixing rules and departure function represent the data well.

TABLE XXXVIII

ROOT MEAN SQUARE DEVIATIONS OF THE RESULTS OF THIS STUDY AND NUMEROUS MIXING RULES FOR THE METHANE-PROPANE SYSTEM

	A	B	C	D	E	F	G	H	I	J	K
	Optimum Parameters									Optimum 2 Parameters	
Mole Fraction Propane	original	smooth T_{cx}	all smooth	Kay (65)	PH (101)	J-JBV (136)	LM (74)	PG (106)	SPG (106)	original	smooth
5.1	0.964	0.999	2.100	1.79	2.87	1.31	2.78	4.46	1.72	1.01	2.70
11.7	0.776	0.703	0.937	6.59	1.39	3.45	1.62	3.10	6.42	1.25	1.32
28.0	0.966	0.966	1.010	10.11	3.34	5.04	2.34	2.52	10.00	1.53	1.48
50.6	0.709	0.720	0.744	10.04	4.63	5.01	2.21	1.94	9.93	1.44	1.75
76.6	0.644	0.648	0.709	5.37	5.27	0.91	2.63	2.87	5.30	1.05	1.06

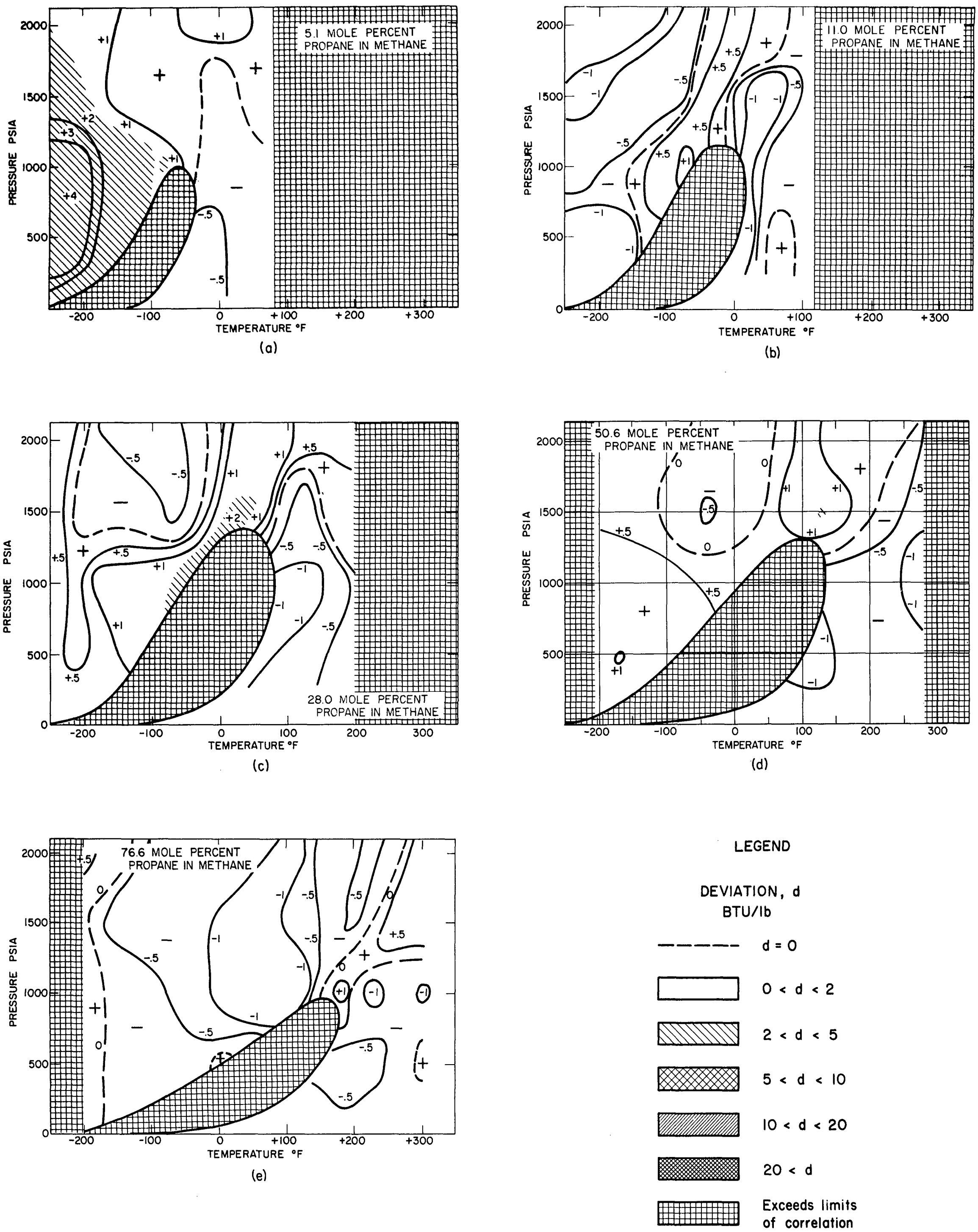


Figure 61. Comparison of Experimental Enthalpy Departures for the Methane-Propane System with the Results from the Corresponding States Correlation and Mixing Rules of This Investigation

By looking at Figure 61a, the comparison chart for the 5.2 mole percent mixture, it appears that a significant part of the deviations for this mixture is present in the liquid region and especially at lower pressures. There are several possible explanations for these large uncertainties in this region. By looking at the range of measurement for this mixture (Figure 3 in Mather's thesis (85)) it can be seen that at 500 psia it was necessary to interpolate heat capacity data to obtain enthalpy data at very low temperatures since there is a region where data was not taken for the liquid. In addition the pressure-temperature-enthalpy diagram (Figure 34 in Mather's thesis) appears to be constructed inconsistently in the low temperature liquid region. Table XXXIX presents the results at 500 and 1000 psia for the average heat capacities = $\Delta \underline{H}/\Delta T$ obtained from this diagram. The strange behavior suggests the possibility of error in this region. In addition the -80°F isotherm appears to be drawn inconsistently. This suggests the large deviation obtained at -80 and 1000 psia on Figure 61a. Thus, the large deviations in this region may well be caused by errors made in constructing the pressure-temperature-enthalpy diagram for the mixture.

Figures 61b, 61c, 61c, and 61e show that for the other mixtures the mixing rule proposed predicts enthalpy departures almost universally to within 2 Btu/lb or better. Thus, the proposed equations containing a total of six constants appear to be sufficient to predict the enthalpy of the entire methane-propane system over a wide range of conditions.

Comparison with Other Mixing Rules

The mixing rules determined in the present investigation although truly representative of the mixture behavior are obtained by fitting a

TABLE XXXIX

DIFFERENCED RESULTS FOR 5.1 MOLE PERCENT MIXTURE

Average Temperature (°F)	$\frac{\Delta H}{\Delta T}$ (500 psia) (Btu/lb-°F)	$\frac{\Delta H}{\Delta T}$ (1000 psia) (Btu/lb-°F)
-275	7.5	7.6
-265	7.5	7.7
-255	9.1	8.2
-245	8.0	8.1
-235	7.5	7.7
-225	7.8	7.6
-215	8.0	8.0
-205	8.2	8.1
-195	7.9	7.8
-185	7.8	7.7
-175	9.9	9.5
-165	7.3	7.0
-155	9.4	8.8

large amount of experimental data. The generalized mixing rules which are available in the literature can be used to estimate mixture properties from a knowledge of pure component properties and minimal amount of mixture results if any. Thus, a comparison of the results of this investigation with the generalized mixing rules of Kay (K),[65] Joffe-Stewart, Burkhardt, Voo (S-SBV),[136] Leland, Mueller (LM),[74] Prausnitz, Gunn (PG),[106] and a simplified rule of Prausnitz, Gunn (SPG)[106] was undertaken.

The rules were used in conjunction with the reference enthalpy tables of this investigation to determine enthalpy departures for the mixtures at the same conditions as were calculated using the empirical mixing rules of this investigation. Both the rules of Leland, Muller[74] and Prausnitz, Gunn[106] show a dependence of the pseudoparameters on temperature and pressure. It was found that for the system studied in this investigation the effects were minor. Therefore, the variations were not considered.

Table XXXVIII presents the results of these calculations. It gives the root mean square deviation obtained for each mixture for each rule in columns D through I. As can be seen the empirical rule is a considerable improvement over the other available generalized rule even when the new reference enthalpy departure tables are used.

The generalized mixing rules which give the best agreement with experimental results for the methane-propane system are those due to Leland and Mueller[74] and Prausnitz and Gunn.[106] The results of the comparisons for the best rule, that of Leland and Mueller,[74] is illustrated graphically on Figure 62. As can be seen the discrepancies between the calculations using this rule and the experimental data is as large as 10 Btu/lb in some regions.

It should be remembered from Section II that all of the mixing rules tested with the exception of the Pitzer-Hültgren rule[101] assume that the third parameter is a linear function of composition. In addition it has been shown that if a two parameter corresponding states principle is valid these mixing rules have some theoretical justification. Thus, in order to more realistically compare these mixing rules with an optimum rule, the original optimization program was modified to search only for an optimum pseudocritical temperature and pressure with the third parameter held constant. The program was run for each mixture using a linear variation of the third parameter, W, with composition to obtain optimum values for the remaining two parameters. This approach assumed that in effect the mixing rules for pseudocritical temperature and pressure are not altered by mixing two nonconformal substances. Table XL lists the

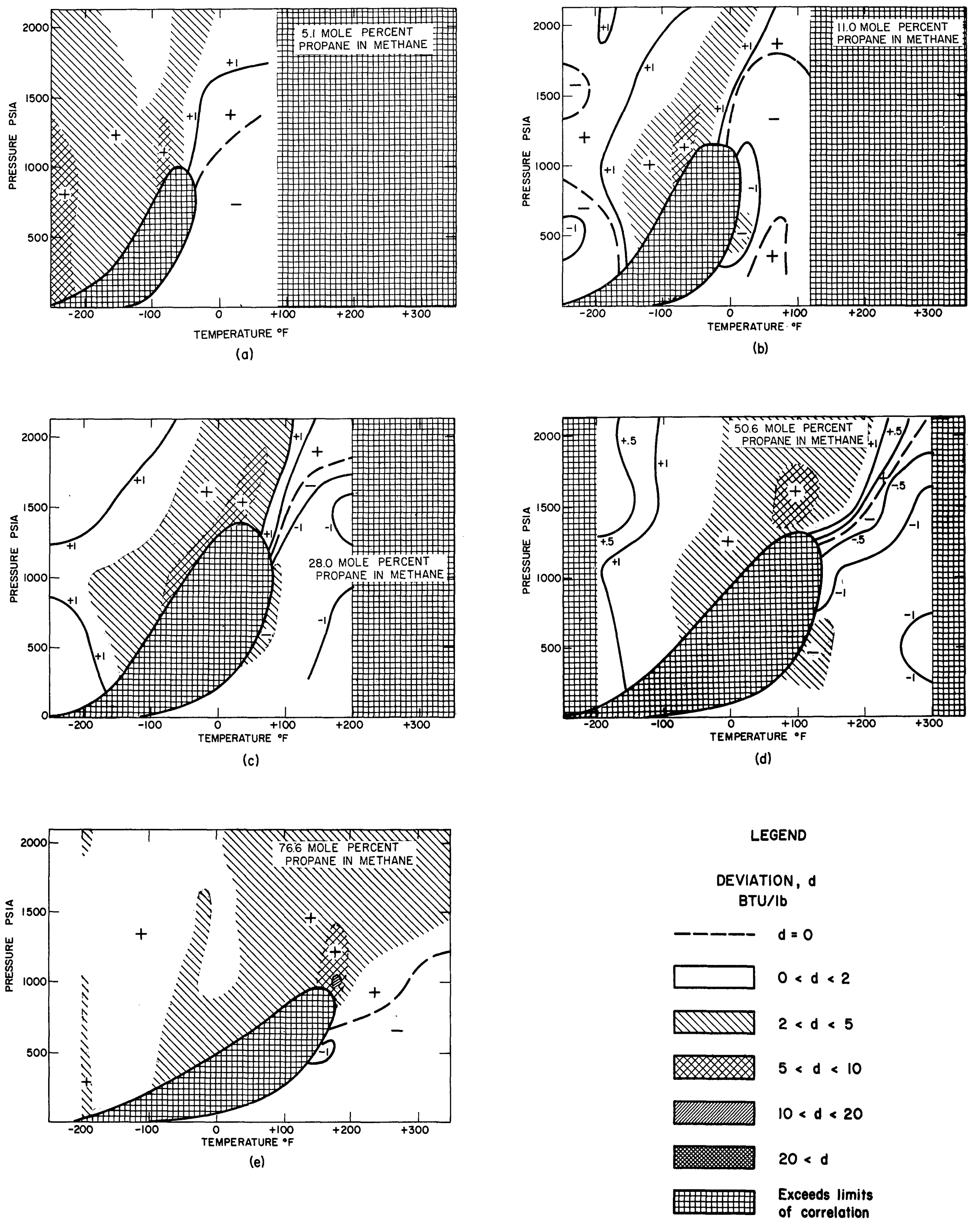


Figure 62. Comparison of Experimental Enthalpy Departures for the Methane-Propane System with Results from the Corresponding States Correlation of This Investigation and Mixing Rules of Leland-Mueller (74)

TABLE XL

OPTIMUM VALUES OF PSEUDOCRITICAL PARAMETERS
WITH THIRD PARAMETER, W, HELD CONSTANT

Mole Fraction Propane	P_c	ΔP_c	T_c	ΔT_c	F
	Original Optimum Values				
0.052	662.0	-8.0	361.1	1.3	1.01
0.117	689.0	22.4	390.5	9.5	1.25
0.280	675.0	17.5	450.0	16.3	1.53
0.506	659.4	14.5	526.2	19.3	1.44
0.766	643.4	13.0	600.9	10.4	1.05
	Smoothed Optimum Values				
0.052	676.6	6.4	364.3	4.6	2.70
0.117	679.3	12.8	390.5	9.5	1.32
0.280	678.5	21.0	450.6	16.9	1.48
0.506	664.1	19.2	525.0	18.4	1.75
0.766	638.5	8.1	601.5	11.0	1.06

optimum temperatures and pressures obtained for each mixture along with the root mean square deviations.

Again the results obtained for the 5.2 mole percent propane in methane mixture did not agree with the other results and were not used to obtain the empirical equations. The two parameters could each be fit in this instance with two constant equations,

$$T_{cx} = \sum T_{ci}x_i + A''x_ix_j[1 + (1 - 2x_i)B''] \tag{80}$$

$$P_{cx} = \sum P_{ci}x_i + D''x_ix_j[1 + (1 - 2x_i)E''] \tag{81}$$

The constants obtained by a least squares fit of the optimization results are given as

$A'' = 73.749$

$B'' = 23.289$

$D'' = 77.417$

$E'' = 60.490$

Values of pseudocritical temperature and pressure were calculated for each mixture from the above equations and these results used to obtain enthalpy departures. These results were compared with experimental values and the root mean square deviations for each mixture are listed in column K of Table XXXVIII. Although these results are not as good a fit as the results for the three parameter optimization they are still considerably more in agreement with experimental data than the generalized mixing rules.

Next the actual values of the parameters were compared. Figure 63 shows the pseudocritical temperatures as a function of composition for the various mixing rules including both the linear and nonlinear third parameter mixing rules of this investigation. All of the mixing rules show a positive deviation in temperature with respect to Kay's rule. Inclusion of a variable third parameter causes the departure from Kay's rule to decrease.

Figure 64 shows the pseudocritical pressure as a function of composition. Inclusion of a nonlinear third parameter causes a change in sign in the departure from Kay's rule. The Pitzer-Hultgren mixing rule[101] seems to overdo the deviation in pressure from Kay's rule. Also the optimum curve obtained with the two parameter search does qualitatively agree with the generalized mixing rules in magnitude as well as in shape.

Finally, Figure 65 presents the third parameter deviation from

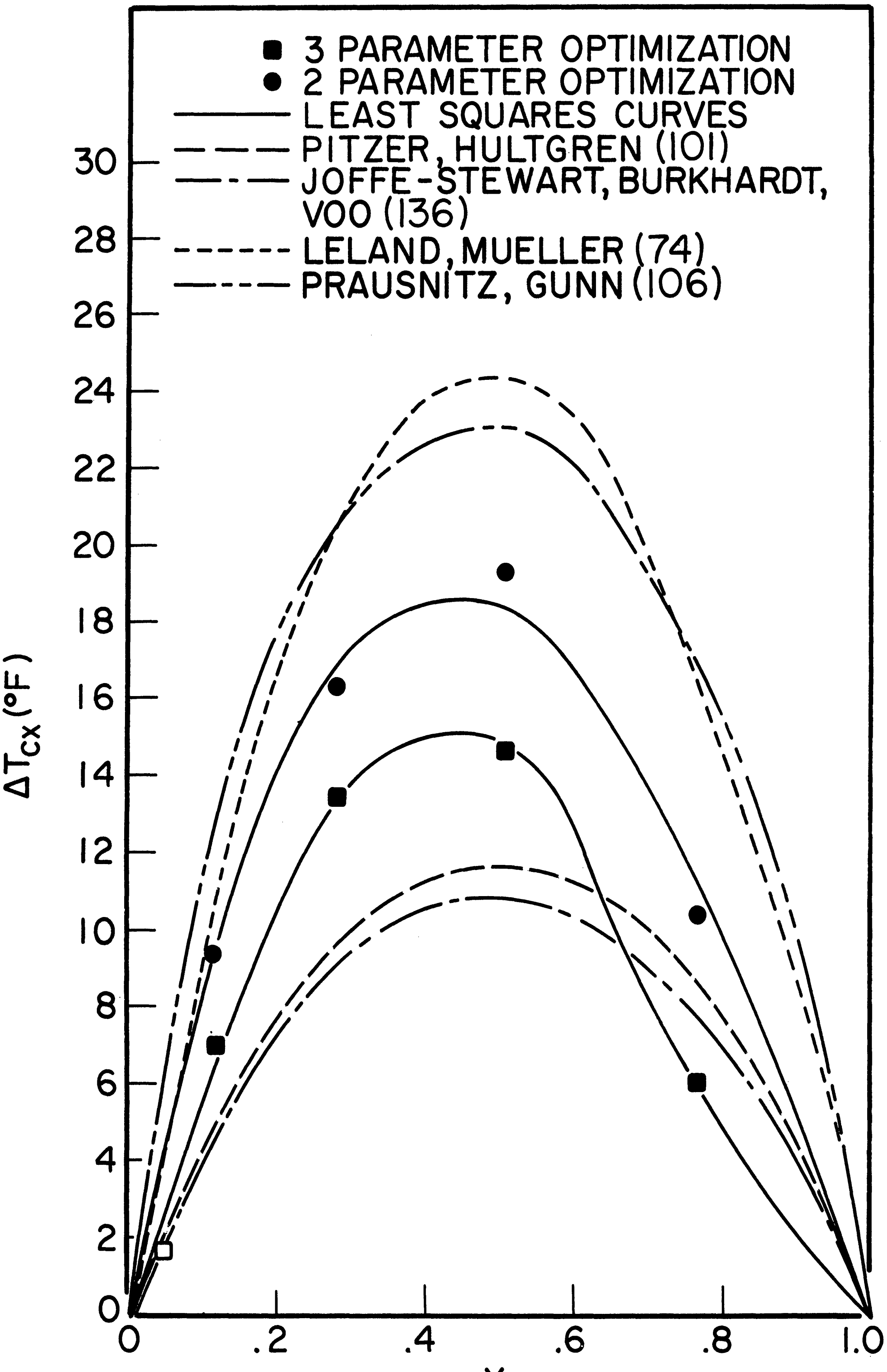


Figure 63. Mixing Rules for Pseudo-Critical Temperature for the Methane-Propane System

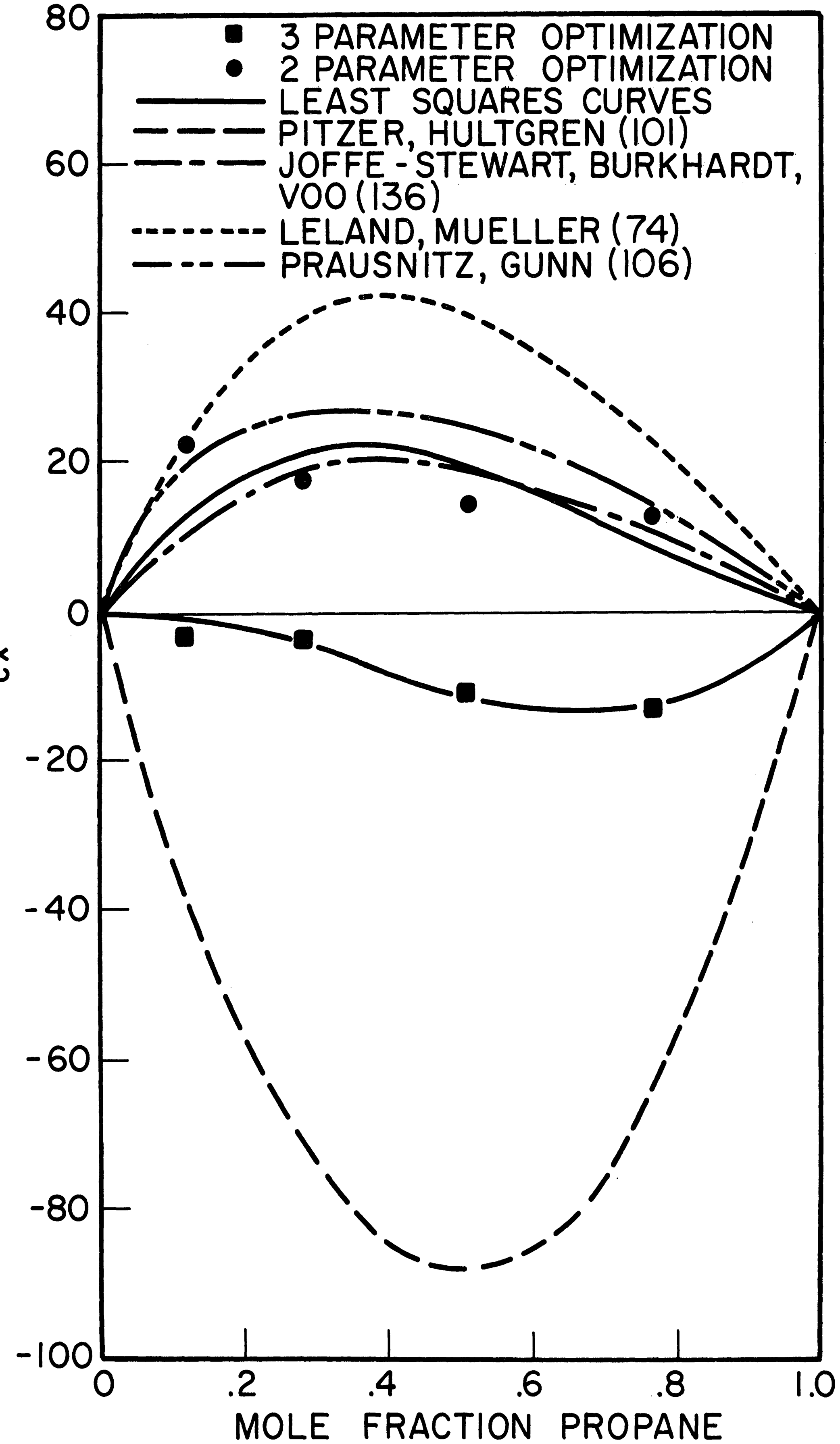


Figure 64. Mixing Rules for the Pseudo-Critical Pressure for the Methane-Propane System

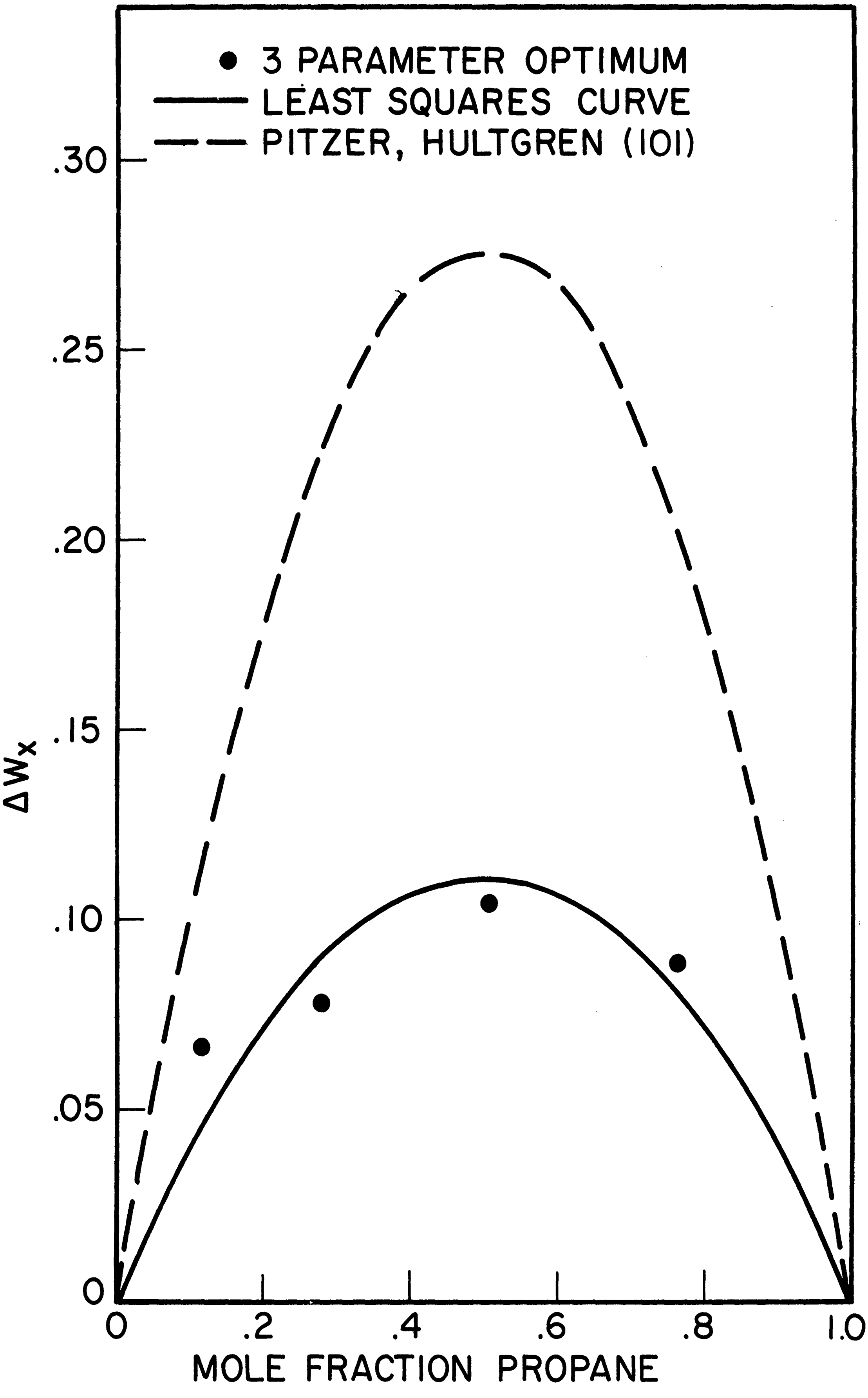


Figure 65. Mixing Rules for the Third Parameter for the Methane-Propane System

linearity as a function of Kay's rule. Again it appears that that the rule due to Pitzer and Hultgren overemphasizes the diviation.

Extension to a Mixture of Nitrogen in Methane

As a final step both the two parameter and three parameter optimization techniques were applied to correlate data for a mixture containing 43.3 mole percent nitrogen in methane. An enthalpy diagram and table based on isobaric enthalpy determinations for this mixture are found in the thesis of Mather.[85] In addition departures were obtained using the various mixing rules described above. The results of the optimization and mixing rule enthalpy departure comparisons in terms of a root-mean-square deviation are given in Table XXXXI. It can be noted that in this case the deviation of pseudocritical temperature is negative. Also there is hardly any effect on varying the third parameter. This seems reasonable in view of the relative similarity between methane and nitrogen. Also Table XXXXI is arranged in such a way that the root-mean-square deviation increases from left to right for the various mixing rules. At the same time the difference between the optimum pseudocritical temperature also increases in the same fashion regardless of the effect of the other parameters. This shows the strong sensitivity of the deviations to the value of the temperature parameter.

Figure 66 presents the results of estimations based on the various mixing rules. All of the rules appear to differ from the experimental results mainly in the liquid region. The results with the optimized parameters are in much better agreement than the generalized rules. It must, however, be pointed out that the diagram of Mather[85] can be in error in the compressed liquid region since it is based entirely on isobaric data with the effect of pressure on enthalpy estimated from the

TABLE XXXXI

ROOT MEAN SQUARE DEVIATIONS OF THE RESULTS OF THIS STUDY AND NUMEROUS MIXING RULES FOR A METHANE-NITROGEN MIXTURE

	optimum parameters	optimum 2 parameters	J-SBV (136)	PG (106)	Kay (65)	SPG (106)	LM (74)
P_{cx}(psia)	558.0	558.0	596.4	599.5	549.9	599.9	602.3
T_{cx}(°R)	286.1	286.1	291.2	292.8	292.8	293.5	294.0
W_x	.0981	.1131	.1131	.1131	.1131	.1131	.1131
F	0.649	0.675	2.54	3.53	3.60	3.99	4.30

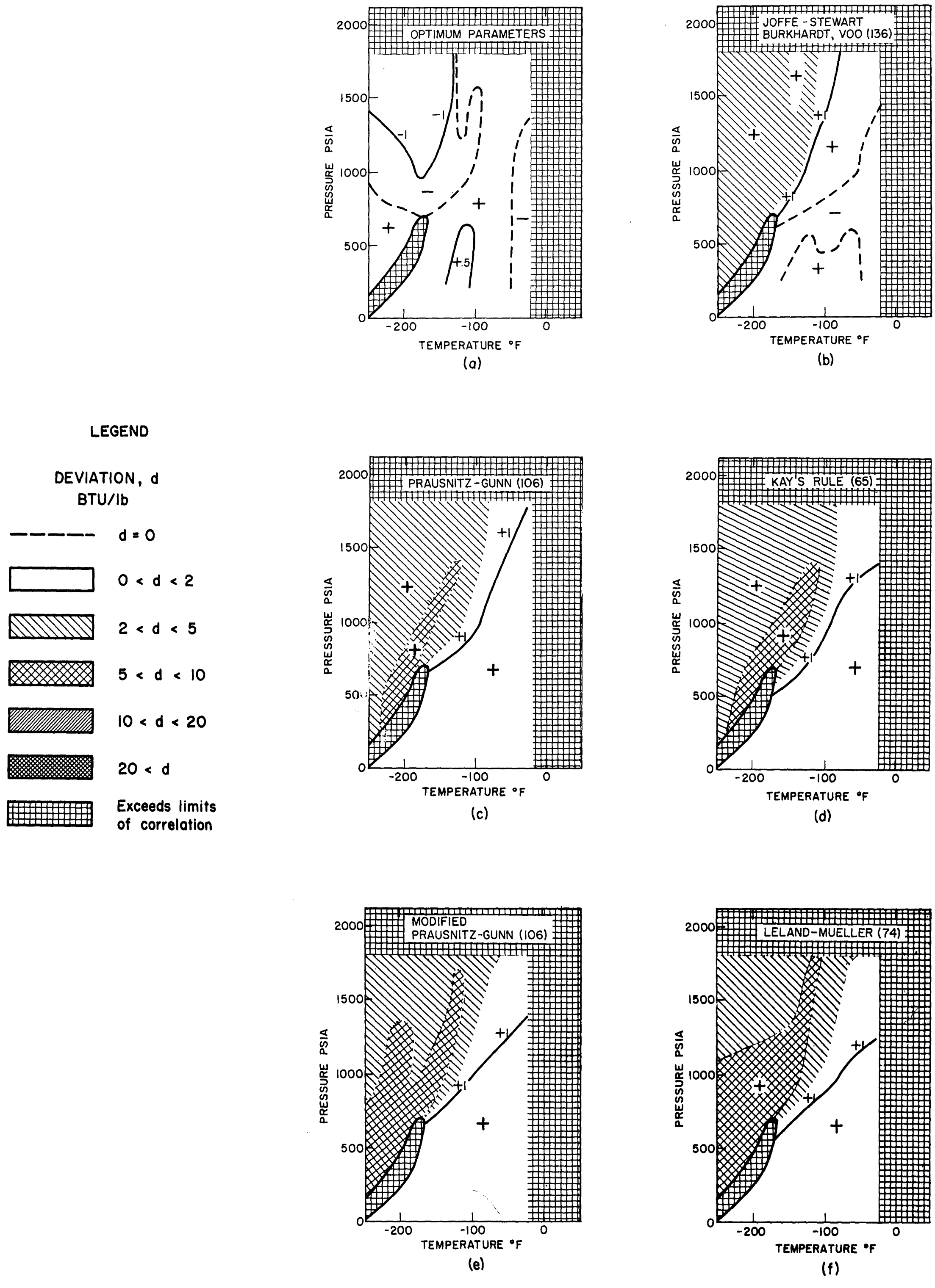


Figure 66. Comparison of Experimental Enthalpy Departures for a Methane-Nitrogen Mixture with Results from the Corresponding States Correlation of This Investigation Using Several Mixing Rules

BWR equation at room temperature. Cumulative errors could produce results in the liquid region which are considerably in error. This may in part account for the reversal in the deviation from Kay's rule of the optimum pseudocritical temperature.

It is somewhat disappointing that the deviation from Kay's rule of the parameters obtained from the mixing rules in general differ in sign from the optimum rule for this mixture. It must be remembered, however, that for the methane-nitrogen system departures from Kay's rule for all parameters are quite small in all cases. Thus, the effect of even an incorrect sign in the departure from Kay's rule is not necessarily that important. In addition the mixing rules which apply most successfully to the methane-propane system are not the best for the methane-nitrogen system. If this were not the case, it might have been possible to make a more definite recommendation of a mixing rule which could be used for systems not experimentally investigated.

Discussion of Results

As has been illustrated the theorem of corresponding state can be used to accurately fit the enthalpy of simple fluid mixtures. From a table of reference substance functions and empirically fit mixing rules the enthalpy of a binary mixture can be calculated almost to within the experimental uncertainty of the experimental data.

Although this method was applied to the methane-propane system using data for pure methane and propane it appears that the procedure could be applied to other systems using the same methane-propane reference functions. This is in part justified by the ability of the method to calculate the enthalpy of nitrogen under pressure and of a nitrogen-methane mixture. In fact although a smaller range of temperature and pressure was considered, the root-mean-square deviations for the methane-nitrogen mixture was smaller than those obtained for any of the methane-propane mixtures.

One could therefore accurately represent the enthalpy behavior of a wide variety of mixtures from two reference substance reduced enthalpy functions and a maximum of six constants for each mixture. In addition, by use of the reference substance function constants for these mixtures could probably be developed from a much smaller amount of new experimental data.

The above procedure need not be limited to the enthalpy function but can be extended to other thermodynamic properties as well. If a thermodynamically consistent and accurate series of reference substance functions were developed for all of the thermodynamic properties of non-polar fluids under pressure then mixing rules could be obtained by an optimization using all of the available thermodynamic data.

Of course actual data cannot be obtained for every mixture and in some cases generalized mixing rules must be used. At present these rules, although not extremely accurate, can be applied with a knowledge that no gross errors will result. If more empirical mixing rules were available, however, such results could be used to test mixing rules developed from theoretical considerations.

SUMMARY AND CONCLUSIONS

1. The recycle flow facility as described by Mather was modified to allow for measurements of fluids of lower volatility than that for which the equipment was originally designed.

2. The effects of pressure and temperature on enthalpy were determined experimentally for propane, a nominal 77 percent and a 51 percent mixture of propane in methane. Measurements were made in the liquid, two-phase, critical, and gaseous regions at temperatures from -250 to +300°F at pressures from 100 to 2000 psia. The data are self consistent to about 0.2 percent.

3. An enthalpy-pressure-temperature table is presented for propane between -280 and +500°F at pressures up to 2000 psia. The table was prepared with the aid of supplementary data in the literature.

4. From the data for the two mixtures, enthalpy-pressure-temperature tables in the regions between -280 and +300°F and pressures up to 2000 psia were developed. These results plus those obtained by Jones, Manker, and Mather adequately represent the enthalpy behavior of the methane-propane binary system.

5. The data obtained in the course of this investigation plus data in the literature were used to compare several of the available methods of prediction. This comparison study indicated that the corresponding states principle would be a most fruitful approach for extending methods of prediction to represent the available data.

6. A three parameter corresponding states correlation was developed which used reference tables derived from enthalpy data for methane and propane. The correlation is valid between T_r of 0.5 and 1.5 at values of P_r up to 3.0. The three parameter corresponding states principle was tested for enthalpy departures by comparing the results of the correlation with data for nitrogen. The correlation predicted the enthalpies of nitrogen to within the experimental uncertainty of the data.

7. The correlation was extended and justified for mixtures by developing a set of mixing rules containing six empirical constants to represent the behavior of the methane-propane binary system. The correlation predicted departures for the mixture to almost within experimental uncertainty in the single phase region over the entire region of its validity.

APPENDIX A

CALIBRATIONS

TABLE XLII

THERMOPILE M-3 CALIBRATION FOR ISOBARIC CALORIMETER

Electromotive Force as a Function of Temperature of Measuring Junction (reference junctions at 0°C)

Degrees C (Int. 1948)	Absolute Microvolts	Degrees C (Int. 1948)	Absolute Microvolts
-196	-33328	20	4774
-183	-31987	40	9761
-100	-20336	60	14946
- 80	-16778	80	20320
- 60	-12954	100	25877
- 40	- 8876	120	31600
- 20	- 4556	140	37486
0	0	160	43533

TABLE XLIII

THERMOPILE M-4 CALIBRATION FOR ISOBARIC CALORIMETER

Electromotive Force as a Function of Temperature of Measuring Junction (reference junctions at 0°C)

Degrees C (Int. 1948)	Absolute Microvolts	Degrees C (Int. 1948)	Absolute Microvolts
-196	-33332	20	4774
-183	-31990	40	9762
-100	-20330	60	14947
- 80	-16778	80	20322
- 60	-12956	100	25882
- 40	- 8874	120	31606
- 20	- 4551	140	37494
0	0	160	43543

TABLE XLIV

THERMOPILE M-5 CALIBRATION FOR THROTTLING CALORIMETER

Electromotive Force as a Function of Temperature of Measuring Junction (reference junctions at 0°C)

Degrees C (Int. 1948)	Absolute Microvolts	Degrees C (Int. 1948)	Absolute Microvolts
-196	-33331	20	4774
-183	-31990	40	9760
-100	-20338	60	14946
- 80	-16779	80	20326
- 60	-12955	100	25872
- 40	- 8878	120	31596
- 20	- 4556	140	37486
0	0	160	43530

TABLE XLV

THERMOPILE M-6 CALIBRATION FOR THROTTLING CALORIMETER

Electromotive Force as a Function of Temperature of Measuring Junction (reference junctions at 0°C)

Degrees C (Int. 1948)	Absolute Microvolts	Degrees C (Int. 1948)	Absolute Microvolts
-196	-33330	20	4775
-183	-31986	40	9761
-100	-20338	60	14944
- 80	-16781	80	20322
- 60	-12956	100	25868
- 40	- 8877	120	31593
- 20	- 4557	140	37484
0	0	160	43530

APPENDIX B

EXPERIMENTAL DATA

TABLE XLVI

TABULATED EXPERIMENTAL ISOBARIC DATA FOR PROPANE

RUN	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER / FLOW (Btu/lb)	ΔH_P (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_P$ (Btu/lb-°F)
1.010	400.6	158.06	158.92	.257	.3406	-.003	.756	.753	.8703
1.020	399.4	158.06	159.87	.559	.3454	-.003	1.618	1.616	.8921
1.030	400.6	158.07	160.80	.850	.3466	-.003	2.451	2.449	.8964
1.040	400.2	158.07	161.21	1.216	.3460	-.003	3.516	3.513	*
1.050	399.8	158.07	161.42	1.704	.3474	-.003	4.906	4.903	*
1.060	399.9	158.07	161.58	2.188	.3441	-.003	6.359	6.356	*
1.070	399.7	158.07	161.79	2.875	.3444	-.003	8.346	8.343	*
1.080	399.7	158.07	162.23	4.659	.3448	-.004	13.514	13.510	*
1.090	399.1	158.07	162.55	7.232	.3447	-.005	20.977	20.973	*
1.100	399.0	158.07	162.80	10.852	.3443	-.006	31.515	31.510	*
1.110	399.1	158.07	163.17	16.486	.3440	-.007	47.930	47.923	*
1.120	398.8	158.07	163.26	21.649	.3424	-.009	63.223	63.214	*
1.130	398.5	158.07	163.27	21.641	.3155	-.008	68.588	68.581	*
1.140	397.8	158.07	163.31	24.629	.3141	-.009	78.410	78.401	*
1.150	397.7	158.07	163.30	26.563	.3136	-.009	84.697	84.688	*
2.010	399.5	158.07	163.12	4.896	.2003	-.003	24.449	24.446	*
2.020	400.0	158.06	163.38	9.120	.1998	-.000	45.637	45.637	*
2.030	398.8	157.98	163.39	15.830	.1995	-.003	79.358	79.355	*
2.040	399.4	158.09	163.65	18.078	.1994	-.004	90.671	90.667	*
2.050	399.3	158.09	163.59	19.197	.1996	-.004	96.174	96.170	*
2.060	399.4	158.06	164.76	19.767	.1994	-.004	99.150	99.145	*
2.070	398.5	158.06	166.54	20.064	.1995	-.005	100.576	100.571	*
2.080	399.6	158.01	170.12	20.663	.1998	-.005	103.402	103.397	*
2.090	400.8	158.09	178.67	21.957	.2002	-.005	109.691	109.686	*
3.012	2000.3	102.10	112.51	2.183	.3353	.001	6.511	6.512	.6253
3.011	2000.3	102.11	112.79	2.205	.3304	.001	6.672	6.673	.6246
3.010	2001.7	102.16	112.89	2.205	.3294	.001	6.694	6.694	.6238
3.020	2001.9	102.15	122.83	4.319	.3326	.001	12.988	12.988	.6280
3.030	2002.2	102.09	142.43	8.595	.3341	.001	25.726	25.727	.6378
3.041	2002.8	102.09	184.41	17.831	.3288	.001	54.231	54.232	.6588
3.040	1998.7	102.09	184.61	17.832	.3268	.001	54.569	54.570	.6613
4.010	999.1	102.02	112.47	2.360	.3460	.000	6.819	6.819	.6523
4.020	999.8	102.02	122.90	4.834	.3498	.000	13.819	13.819	.6619
4.030	999.9	101.93	145.24	9.846	.3354	.000	29.352	29.352	.6778
4.040	999.8	102.02	182.94	19.174	.3322	.000	57.723	57.724	.7134
5.010	1498.9	102.05	113.08	2.340	.3313	-.000	7.066	7.065	.6406
5.020	1498.5	102.05	124.79	4.822	.3283	-.000	14.687	14.686	.6456
5.030	1499.1	102.06	145.27	9.341	.3289	-.000	28.404	28.404	.6573
5.040	1499.5	102.06	185.32	18.359	.3234	-.000	56.765	56.764	.6817
6.010	499.2	102.04	112.24	2.340	.3342	-.001	7.002	7.001	.6868
6.020	501.9	101.97	124.06	4.966	.3232	-.001	15.366	15.365	.6955
6.030	498.9	102.17	144.47	9.710	.3173	-.001	30.603	30.602	.7234
6.040	500.4	102.29	183.12	21.630	.3212	-.002	67.342	67.340	.8331
7.010	251.6	102.16	111.89	2.305	.3325	-.000	6.933	6.933	.7125
7.040	250.7	101.94	118.87	3.932	.3233	-.000	12.160	12.160	.7183
7.050	251.2	102.27	116.00	3.172	.3233	-.000	9.811	9.811	.7147
7.021	249.3	102.06	120.32	4.715	.3289	-.000	14.334	14.334	*
7.020	249.3	102.03	119.87	4.713	.3250	-.000	14.499	14.499	*
7.030	251.2	102.00	121.28	6.285	.3251	-.001	19.333	19.333	*
7.060	250.0	102.08	120.59	5.643	.3209	-.001	17.585	17.585	*
7.070	250.8	101.85	121.71	8.399	.3224	-.001	26.054	26.054	*
7.080	250.7	102.08	121.75	5.647	.1906	-.000	29.630	29.630	*
7.090	252.4	102.08	122.85	8.400	.1912	-.000	43.927	43.926	*
7.100	251.2	102.11	122.71	11.832	.1906	-.000	62.094	62.094	*
7.110	250.1	101.96	122.16	15.510	.1894	-.001	81.888	81.887	*
7.120	249.4	102.08	122.28	19.724	.1887	-.001	104.519	104.518	*
7.130	248.9	102.08	122.15	22.161	.1880	-.001	117.888	117.887	*
7.140	249.0	102.11	122.21	23.340	.1880	-.001	124.156	124.155	*
7.150	248.7	102.08	122.23	25.366	.1876	-.001	135.201	135.199	*
7.160	247.8	102.11	125.79	26.020	.1871	-.001	139.095	139.094	*
7.170	248.6	102.14	129.49	26.513	.1879	-.001	141.090	141.089	*
7.180	249.3	102.11	133.52	27.038	.1886	-.001	143.346	143.345	*
8.010	246.5	201.59	212.22	2.028	.3363	-.300	6.032	5.732	.5393
8.020	246.6	201.54	222.03	3.822	.3368	-.310	11.350	11.039	.5388
8.031	245.9	201.53	241.05	7.230	.3315	-.317	21.810	21.493	.5438
8.030	245.7	201.53	241.12	7.230	.3309	-.317	21.850	21.533	.5439
8.040	245.6	201.67	277.55	13.898	.3309	-.337	41.999	41.662	.5490
9.010	500.2	201.48	210.71	2.622	.3385	-.295	7.746	7.451	.8068
9.020	499.5	201.54	221.28	5.188	.3358	-.313	15.451	15.137	.7670
9.031	499.5	201.44	239.53	9.483	.3360	-.331	28.220	27.889	.7321
9.030	499.2	201.46	239.60	9.482	.3359	-.340	28.227	27.887	.7312
9.041	498.7	201.46	277.26	17.662	.3349	-.389	52.732	52.343	.6905
9.040	498.9	201.48	277.13	17.659	.3356	-.394	52.618	52.225	.6904
10.011	1502.6	201.41	210.89	2.375	.3201	-.002	7.420	7.418	.7823
10.010	1502.1	201.45	211.09	2.354	.3184	-.002	7.393	7.391	.7661
11.010	246.2	201.35	211.37	1.978	.3454	-.317	5.727	5.410	.5404
11.020	243.9	201.40	221.19	3.781	.3429	-.320	11.024	10.704	.5410
11.030	245.1	201.33	240.14	7.405	.3439	-.334	21.532	21.198	.5463
11.040	243.4	201.44	279.16	14.727	.3405	-.351	43.235	42.884	.5518
12.011	999.1	201.48	212.11	2.954	.3106	-.007	9.511	9.504	.8947
12.010	1000.1	201.57	212.24	2.954	.3089	-.007	9.564	9.558	.8957
12.020	999.4	201.50	221.93	5.935	.3141	-.008	18.893	18.885	.9245
12.031	1000.3	201.52	242.85	13.523	.3203	-.009	42.214	42.205	1.0213
12.030	1002.1	201.55	243.08	13.519	.3188	-.009	42.410	42.401	1.0209
13.011	1500.9	201.49	212.10	2.597	.3169	-.002	8.193	8.191	.7721
13.010	1500.2	201.49	212.06	2.597	.3191	-.002	8.140	8.138	.7704
13.020	1502.1	201.51	222.91	5.441	.3255	-.002	16.714	16.712	.7811
13.021	1500.7	201.51	223.12	5.442	.3219	-.002	16.906	16.904	.7824
13.030	1498.3	201.48	242.20	11.138	.3397	-.003	32.788	32.785	.8051
14.010	700.2	201.49	210.52	4.044	.3205	-.015	12.619	12.605	1.3956
14.011	700.0	201.53	210.54	4.045	.3217	-.015	12.573	12.558	1.3948
14.020	701.2	201.64	220.75	14.327	.3184	-.020	44.997	44.977	2.3545
15.011	699.3	201.53	231.21	13.596	.1943	-.009	69.977	69.968	2.3577
15.010	700.3	201.52	230.70	13.597	.1955	-.009	69.539	69.530	2.3822
15.021	702.3	201.54	242.85	16.658	.2004	-.009	83.136	83.127	2.0122
15.020	701.2	201.49	244.25	16.656	.1969	-.009	84.607	84.598	1.9785
15.030	701.1	201.56	257.71	18.040	.1842	-.009	97.912	97.902	1.7437

TABLE XLVI
(CONTINUED)

RUN	INLET PRESSURE (psia)	INLET TEMPERATURE (^oF)	OUTLET TEMPERATURE (^oF)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER/FLOW (Btu/lb)	ΔH_P (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_P$ (Btu/lb-^oF)
16.010	2000.0	201.50	211.32	2.289	.3227	+.001	7.096	7.095	.7219
16.020	2000.6	201.46	221.49	4.567	.3122	-.001	14.629	14.628	.7304
16.030	2001.2	201.58	240.49	8.863	.3056	-.001	28.998	28.997	.7454
16.042	2000.2	201.46	277.15	17.582	.3025	-.001	58.123	58.122	.7679
16.041	1999.7	201.54	277.38	17.580	.3016	-.001	58.288	58.287	.7686
16.040	2001.8	201.58	277.68	17.580	.2973	-.001	59.134	59.134	.7771
17.010	617.4	202.15	203.44	1.008	.3417	-.039	2.950	2.911	2.2572
17.020	616.8	202.14	204.75	2.437	.3429	-.029	7.107	7.078	2.7151
18.010	517.5	203.79	205.09	1.503	.3272	-.073	4.595	4.522	3.4593
18.020	618.1	203.83	206.18	7.353	.3309	-.092	22.223	22.131	9.4276
18.030	617.8	203.79	206.30	10.118	.3335	-.112	30.342	30.231	12.0242
18.040	618.5	203.84	207.30	11.834	.3062	-.097	38.651	38.553	11.1487
18.050	618.1	203.84	208.48	11.831	.2712	-.097	43.625	43.528	9.3640
18.060	618.0	203.80	210.22	11.687	.2414	-.078	48.420	48.342	7.5280
18.070	618.4	203.82	211.74	11.432	.2227	-.078	51.336	51.258	6.4771
19.010	498.2	203.88	213.11	2.223	.3013	-.170	7.378	7.208	.7814
19.020	499.2	203.87	223.23	4.478	.3015	-.179	14.853	14.674	.7583
19.030	499.5	203.87	242.53	8.485	.3010	-.197	28.186	27.989	.7240
19.040	499.6	203.94	260.87	12.057	.3003	-.206	40.146	39.940	.7016
19.051	499.2	203.77	280.50	11.884	.2255	-.134	52.693	52.558	.6850
19.050	499.4	203.79	280.33	11.884	.2257	-.134	52.661	52.526	.6862
20.010	617.9	212.22	218.17	2.597	.2928	-.070	8.871	8.800	1.4787
20.020	517.5	212.20	227.83	5.694	.2934	-.130	19.409	19.279	1.2335
20.030	617.1	212.14	253.11	11.910	.2925	-.163	40.717	40.554	.9898
21.010	995.3	251.29	260.23	3.258	.1309	-.527	24.883	24.357	2.7268
21.021	999.8	251.34	270.00	6.074	.1250	-.656	48.579	47.923	2.5686
21.020	1000.1	251.35	270.10	6.074	.1245	-.689	48.804	48.115	2.5667
21.030	998.6	251.34	282.52	8.468	.1158	-.661	73.124	72.463	2.3243
22.011	1199.8	251.32	262.11	2.027	.1078	-.003	18.807	18.804	1.7416
22.010	1199.8	251.28	262.96	2.028	.1027	-.003	19.750	19.747	1.6915
22.021	1201.3	251.28	276.94	4.550	.1020	-.003	44.628	44.626	1.7391
22.020	1201.4	251.28	276.93	4.549	.1020	-.004	44.622	44.618	1.7397
22.030	1200.9	251.31	294.55	7.975	.1034	-.004	77.111	77.107	1.7833
23.010	1500.4	251.34	264.43	2.115	.1095	-.001	19.320	19.319	1.4760
23.020	1502.4	251.34	276.63	3.999	.1071	-.001	37.331	37.330	1.4761
23.030	1499.2	251.41	294.50	6.804	.1053	-.001	64.635	64.634	1.4998
24.010	1001.4	251.34	293.85	8.543	.1008	-.010	84.798	84.788	1.9943
25.010	998.9	231.39	241.43	2.886	.1322	-.005	21.821	21.816	2.1743
25.020	1000.8	231.46	250.42	5.612	.1304	-.005	43.038	43.033	2.2696
25.030	999.2	231.44	260.47	7.764	.1185	-.005	65.505	65.500	2.2559
26.010	498.5	171.47	174.22	.640	.1301	-.001	4.924	4.923	1.7895
26.020	498.8	171.39	176.43	1.193	.1300	-.001	9.177	9.176	1.8203
26.030	499.3	171.46	178.71	1.749	.1302	-.001	13.439	13.438	1.8533
26.040	499.8	171.46	180.94	2.337	.1299	-.001	17.986	17.984	1.8962
26.050	500.6	171.44	180.58	2.243	.1302	-.001	17.233	17.232	1.8856
26.060	498.6	171.42	180.05	2.091	.1294	-.001	16.156	16.155	1.8721
26.070	500.0	171.43	181.60	2.530	.1301	-.001	19.443	19.441	1.9114
26.080	500.1	171.47	184.00	3.203	.1299	-.001	24.659	24.657	1.9690
26.090	500.5	171.62	185.11	3.928	.1299	-.001	30.233	30.232	*
26.100	500.9	171.27	185.38	4.787	.1302	-.002	36.776	36.774	*
26.110	500.9	171.36	185.36	5.631	.1299	-.002	43.345	43.343	*
26.120	500.7	171.34	185.26	4.773	.1298	-.002	36.774	36.772	*
26.130	500.6	171.37	185.35	6.540	.1298	-.003	50.405	50.402	*
26.140	501.0	171.56	185.63	7.972	.1298	-.003	61.419	61.416	*
26.150	500.8	171.31	185.44	7.970	.1170	-.003	68.136	68.133	*
26.160	500.0	171.33	185.34	7.968	.1058	-.001	75.286	75.285	*
26.170	501.4	171.33	185.65	9.986	.1061	-.003	94.127	94.124	*
26.180	501.0	171.35	185.62	12.130	.1059	-.002	114.500	114.498	*
26.190	501.1	171.33	185.59	14.761	.1059	-.003	139.430	139.427	*
26.200	501.0	171.33	188.30	15.366	.1061	-.003	144.886	144.883	*
26.210	499.2	171.34	192.78	16.254	.1055	-.003	154.097	154.095	*
26.220	498.9	171.37	202.78	17.786	.1056	-.003	168.498	168.495	*
27.050	250.2	136.13	176.50	5.457	.1305	-.124	41.799	41.674	1.0322
27.060	249.5	136.16	176.60	3.184	.0940	-.042	33.857	33.814	.8361
27.010	250.3	136.05	146.49	1.880	.3120	-.216	6.026	5.809	.5564
27.020	249.3	136.36	157.00	3.637	.3107	-.225	11.706	11.481	.5563
27.030	250.1	136.26	177.14	7.123	.3117	-.233	22.852	22.619	.5533
27.040	248.3	136.36	177.09	8.294	.3644	-.324	22.762	22.438	.5509
27.071	248.0	136.17	203.75	10.759	.2896	-.220	37.158	36.937	.5466
27.070	248.1	136.17	203.44	10.760	.2902	-.217	37.079	36.862	.5480
28.010	1498.4	22.71	34.40	2.069	.3079	.000	6.722	6.722	.5749
28.020	1498.5	22.72	47.00	4.233	.3016	.000	14.035	14.036	.5779
28.030	1499.7	22.93	68.13	8.070	.3055	.000	26.413	26.413	.5844
28.041	1498.5	22.55	104.71	15.325	.3114	.000	49.210	49.211	.5990
28.040	1497.7	22.53	104.95	15.327	.3103	.000	49.389	49.389	.5993
29.010	2000.9	22.64	34.74	2.124	.3077	.000	6.903	6.903	.5704
29.020	1998.9	22.62	46.31	4.183	.3080	.000	13.583	13.583	.5735
29.030	1998.8	22.61	67.27	7.938	.3075	.000	25.816	25.816	.5781
29.041	1999.4	22.61	107.59	15.338	.3040	.000	50.455	50.455	.5937
29.040	1999.3	22.62	108.11	15.339	.3027	.000	50.666	50.666	.5927
30.010	1001.0	22.63	34.13	2.102	.3159	.000	6.656	6.656	.5790
30.020	1002.0	22.60	45.20	4.173	.3164	.000	13.191	13.191	.5836
30.030	1001.5	22.63	68.59	7.962	.2922	.000	27.246	27.247	.5927
30.040	998.4	22.57	108.41	15.169	.2894	.000	52.421	52.422	.6107
31.010	502.2	22.55	33.73	2.023	.3094	.000	6.539	6.539	.5847
31.020	501.1	22.58	45.63	4.178	.3062	.000	13.645	13.645	.5918
31.030	499.9	22.60	67.70	8.245	.3038	.000	27.142	27.143	.6019
31.040	501.0	22.61	103.78	15.340	.3033	.000	50.572	50.572	.6230
32.010	251.8	22.32	33.64	2.035	.3016	.000	6.747	6.747	.5965
32.020	248.4	22.32	45.76	4.214	.2999	.000	14.049	14.050	.5993
32.030	248.5	22.38	67.88	8.255	.2973	.000	27.762	27.762	.6102
32.040	250.1	22.36	104.97	15.485	.2953	.000	52.445	52.446	.6348
33.010	1001.9	22.33	67.93	10.205	.3769	.000	27.075	27.075	.5937
33.020	1000.4	22.32	67.47	8.277	.3095	.000	26.747	26.747	.5924
33.030	1000.0	22.29	68.46	6.600	.2421	.000	27.263	27.263	.5905
33.040	1002.0	22.24	67.18	4.056	.1543	.000	26.286	26.286	.5848
34.010	999.6	-236.75	-223.33	2.305	.3641	.002	6.332	6.334	.4720
35.010	1000.7	-236.25	-209.59	4.345	.3461	.001	12.553	12.554	.4708
35.021	998.5	-236.23	-186.89	8.429	.3603	.001	23.394	23.395	.4741
35.020	998.4	-236.26	-186.74	8.418	.3588	.001	23.465	23.466	.4738
35.031	1001.9	-236.24	-186.00	6.883	.2913	.001	23.628	23.629	.4703
35.030	998.9	-236.22	-185.73	6.879	.2887	.001	23.825	23.826	.4719
35.040	1001.4	-236.21	-142.60	14.799	.3307	.001	44.752	44.753	.4781

TABLE XLVI
(CONTINUED)

RUN	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER FLOW (Btu/lb)	ΔH_P (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_P$ (Btu/lb-°F)
36.010	2001.5	-236.22	-222.89	2.134	.3406	.001	6.266	6.267	.4701
36.024	2003.9	-236.23	-209.11	4.139	.3219	.001	12.859	12.860	.4741
36.023	2002.2	-236.24	-209.20	4.139	.3248	.001	12.744	12.745	.4713
36.022	1999.4	-236.21	-209.21	4.139	.3231	.001	12.810	12.811	.4745
36.021	2000.6	-236.22	-208.69	4.139	.3172	.001	13.047	13.048	.4741
36.020	1998.7	-236.23	-208.10	4.139	.3098	.001	13.360	13.361	.4749
36.031	2001.3	-236.20	-185.15	7.727	.3184	.001	24.267	24.268	.4754
36.030	2002.6	-236.20	-184.92	7.727	.3182	.001	24.286	24.286	.4736
36.041	2001.5	-236.20	-141.02	14.090	.3081	.001	45.730	45.731	.4805
36.040	2003.2	-236.20	-138.95	14.090	.3012	.001	46.775	46.775	.4810
37.010	1000.9	-136.72	-126.74	1.697	.3470	.001	4.890	4.890	.4902
37.020	1001.7	-136.72	-116.40	3.495	.3488	.001	10.022	10.022	.4931
37.031	1001.4	-136.72	-96.02	6.940	.3436	.001	20.201	20.201	.4964
37.030	1001.4	-136.72	-96.03	6.940	.3433	.001	20.214	20.214	.4968
37.042	1001.1	-136.73	-60.90	13.611	.3557	.001	38.263	38.263	.5046
37.041	999.4	-136.73	-60.60	13.609	.3542	.001	38.428	38.429	.5048
37.040	1000.5	-136.73	-60.37	13.610	.3534	.001	38.510	38.511	.5044
38.010	2000.7	-136.73	-126.39	1.761	.3456	.001	5.096	5.097	.4932
38.020	1999.3	-136.72	-115.79	3.532	.3380	.001	10.450	10.451	.4994
38.030	2001.1	-136.68	-96.69	7.013	.3541	.001	19.804	19.804	.4952
38.040	2000.9	-136.69	-59.13	13.689	.3501	.001	39.099	39.099	.5041
39.010	1000.5	-56.94	-48.15	1.640	.3533	.001	4.641	4.642	.5282
39.020	999.3	-56.97	-39.48	3.257	.3518	.001	9.260	9.260	.5294
39.030	1000.7	-57.06	-22.02	6.554	.3508	.001	18.685	18.686	.5333
39.040	999.0	-56.90	11.12	12.757	.3451	.001	36.962	36.963	.5434
40.010	2000.5	-56.98	-48.09	1.587	.3420	.001	4.639	4.640	.5218
40.020	1999.3	-56.92	-38.84	3.243	.3422	.001	9.477	9.477	.5242
40.030	1999.9	-56.98	-21.75	6.275	.3374	.001	18.601	18.602	.5279
40.040	2001.1	-56.96	11.36	12.513	.3415	.001	36.641	36.642	.5363
43.010	1201.3	267.58	269.45	.514	.2423	-.001	2.122	2.121	1.1352
43.020	1201.5	267.59	272.19	1.259	.2385	-.001	5.278	5.277	1.1460
43.030	1200.9	267.17	274.55	1.976	.2343	-.001	8.433	8.432	1.1406
43.040	1200.5	267.29	277.48	2.686	.2324	-.001	11.557	11.556	1.1344
43.050	1197.6	267.49	280.25	3.326	.2289	-.001	14.532	14.532	1.1389
43.060	1199.7	267.53	282.96	4.119	.2333	-.001	17.656	17.655	1.1443
44.010	700.5	211.94	212.87	.481	.2522	-.076	1.907	1.831	1.9577
44.020	702.0	211.94	213.87	1.054	.2535	-.076	4.157	4.081	2.1108
44.030	700.2	211.94	214.77	1.720	.2581	-.080	6.662	6.583	2.3226
44.040	701.8	212.02	215.92	2.418	.2505	-.080	9.653	9.574	2.4526
44.050	701.4	212.00	216.74	3.188	.2513	-.080	12.685	12.605	2.6602
44.060	700.2	211.98	217.73	4.143	.2499	-.080	16.578	16.498	2.8696
44.070	701.3	212.03	218.68	5.115	.2495	-.080	20.503	20.423	3.0709
44.080	702.1	212.09	219.76	6.319	.2496	-.080	25.319	25.240	3.2931
45.010	400.6	181.84	191.73	1.992	.2830	-.193	7.040	6.847	.6916
45.020	399.8	181.80	201.37	3.857	.2881	-.202	13.388	13.186	.6736
45.030	397.5	181.93	211.13	5.589	.2864	-.214	19.513	19.298	.6608
46.010	449.0	191.46	201.25	2.109	.2810	-.277	7.504	7.227	.7389
46.020	448.9	191.38	210.85	4.058	.2816	-.292	14.411	14.119	.7252
47.010	701.1	218.37	219.30	1.275	.2696	-.055	4.729	4.674	5.0303
47.020	700.6	218.40	220.26	2.524	.2704	-.072	9.335	9.263	4.9774
47.030	701.2	218.40	221.24	3.601	.2693	-.079	13.374	13.295	4.6680
47.040	699.4	218.40	222.26	4.569	.2692	-.085	16.972	16.887	4.3663

TABLE XLVII

TABULATED EXPERIMENTAL ISOTHERMAL DATA FOR PROPANE

RUN	INLET TEMPERATURE (°F)	INLET PRESSURE (psia)	PRESSURE DROP (psi)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	ΔH_T (Btu/lb)	$(\Delta H/\Delta P)_T$ (Btu/lb-psi)
3.020	250.0	2005.6	339.2	.985	.3540	.013	2.769	-.008163
3.040	249.9	1821.3	360.4	1.473	.3607	-.010	4.094	-.011358
3.060	250.0	1581.2	365.0	2.475	.3543	-.028	7.014	-.019219
3.080	249.9	1333.7	402.3	7.294	.3499	-.105	20.951	-.052084
3.010	250.1	2010.2	181.5	.338	.2564	.023	1.296	-.007141
3.030	250.0	1809.4	186.7	.474	.2546	.003	1.858	-.009954
3.050	250.0	1586.4	213.8	.884	.2689	-.005	3.292	-.015398
3.070	250.0	1347.8	232.9	1.890	.2723	-.040	6.982	-.029980
3.090	249.9	1046.8	297.4	10.751	.2524	-.039	42.635	-.143372
3.100	249.9	829.4	333.4	8.046	.1936	-.011	41.566	-.124686
3.110	249.9	659.8	316.6	3.879	.1463	.001	26.518	-.083768
3.120	249.9	654.6	536.6	6.311	.1625	-.012	38.847	-.072402
4.010	201.0	2024.8	170.9	.101	.2608	-.023	.409	-.002393
4.030	201.0	1826.4	192.2	.175	.2750	-.028	.662	-.003446
4.050	201.0	1544.9	181.2	.238	.2558	.022	.909	-.005016
4.070	201.0	1284.7	188.7	.428	.2666	-.026	1.632	-.008650
4.020	201.0	2025.0	319.2	.315	.3623	-.006	.876	-.002744
4.040	201.0	1828.2	333.6	.472	.3688	-.003	1.283	-.003846
4.060	201.0	1530.4	340.6	.834	.3772	-.004	2.216	-.006507
4.080	201.0	1273.7	345.1	1.288	.3621	-.048	3.605	-.010448
5.020	201.0	1018.5	354.5	2.896	.3642	.042	7.911	-.022317
5.010	201.0	1021.2	160.4	.472	.2395	-.037	2.007	-.012512
5.030	201.0	720.6	77.9	.565	.1568	.020	3.584	-.046012
5.040	201.0	721.6	89.4	.739	.1682	.041	4.354	-.048707
5.050	201.0	822.7	104.5	1.047	.1834	.039	5.669	-.054254
5.060	201.0	719.3	119.9	1.487	.1940	-.033	7.696	-.064189
5.070	201.0	726.0	176.6	13.530	.1968	.052	68.707	-.389091
5.080	200.9	722.1	190.6	14.381	.2001	.036	71.844	-.376971
5.090	201.0	717.7	205.2	15.203	.2022	-.031	75.235	-.366674
5.100	201.0	555.3	352.6	5.575	.1421	.006	39.241	-.111301
5.110	201.0	551.3	440.9	6.652	.1479	-.019	45.004	-.102084
6.010	161.7	2019.8	202.9	.011	.2911	-.021	.059	-.000293
6.020	160.6	1877.5	201.5	.039	.2912	-.003	.135	-.000672
6.030	160.6	1710.4	203.0	.063	.2889	.004	.216	-.001062
6.040	160.6	1567.5	202.0	.087	.2900	-.002	.301	-.001493
6.050	160.4	1386.4	210.5	.126	.2914	.007	.427	-.002027
6.060	160.5	1217.9	220.0	.199	.3025	-.000	.657	-.002987
6.070	160.5	1024.4	223.0	.290	.3012	.026	.938	-.004205
6.080	160.5	827.9	232.4	.452	.3053	.001	1.479	-.006364
6.090	160.6	638.3	210.8	.589	.2874	.013	2.037	-.009665
7.010	200.6	562.8	163.7	4.844	.1814	-.024	26.725	-.163271
7.020	200.5	470.6	256.2	4.129	.1733	.010	23.820	-.092981
7.030	200.5	447.8	331.2	4.762	.1751	.010	27.186	-.082091

TABLE XLVIII

TABULATED EXPERIMENTAL JOULE-THOMSON DATA FOR PROPANE

Run	Inlet Temperature (°F)	Inlet Pressure (psia)	Temperature Difference (°F)	Pressure Drop (psi)	$-\left(\frac{\Delta T}{\Delta P}\right)_H$ (°F/psi)x10^3
0.011	21.2	1992.1		381.2	4.668
0.010	21.3	1992.5	1.524	352.0	4.371
2.010	2.13	2026.0	3.334	773.9	4.357
2.020	21.3	2020.5	4.281	1016.6	4.262
2.030	21.2	1082.2	1.837	481.5	3.853
2.040	21.1	1071.7	2.580	697.7	3.738
2.050	21.2	1099.6	3.372	941.9	3.622

TABLE XLIX

FLOWMETER CALIBRATION EQUATION CONSTANTS USED FOR PROPANE

Flow Meter Series		B	A	C	D
10 - 20	low	0.045462835	94.606418	-34925.356	5060329.4
	high	0.4390828	-270.13652	26375.322	-6141762.4
30 high		0.86077077	-610.80764	166847.2	-13993813.0
40 - 50	low	0.045305256	93.5971	-34102.569	4941925.4
	high	0.50686614	-320.16109	88793.802	-7156188.0
50 - 60	low	0.057971521	81.50558	30315.252	4554907.9
	high	0.49223719	-308.87943	86038.184	-6942817.9
60 - 70	low	0.050889514	89.836006	-33417.221	4928000.5
	high	0.50108158	-316.88119	88371.262	-7159948.1

TABLE L

TABULATED EXPERIMENTAL ISOBARIC DATA FOR THE NOMINAL 77 PERCENT MIXTURE

RUN	MOLE FRACTION C_3H_8	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER/FLOW (Btu/lb)	ΔH_p (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_p$ (Btu/lb-°F)
1.0100	.767	1499.8	199.49	215.81	3.869	.2454	-.004	15.764	15.759	.9659
1.0200	.767	1501.6	200.72	233.44	8.091	.2496	-.005	32.420	32.415	.9905
1.0300	.767	1502.1	200.98	249.09	11.598	.2441	-.005	47.504	47.499	.9872
2.0100	.767	2001.2	200.99	216.69	3.114	.2410	-.001	12.917	12.916	.8228
2.0200	.767	2001.0	200.99	234.89	6.420	.2252	-.001	28.514	28.512	.8412
2.0310	.767	1999.5	201.02	252.32	8.974	.2049	-.001	43.792	43.791	.8537
2.0300	.767	2000.9	200.99	252.61	8.974	.2041	-.001	43.971	43.970	.8517
3.0100	.767	1199.6	201.02	216.85	4.045	.2202	-.009	18.375	18.366	1.1601
3.0200	.766	1199.4	201.12	234.14	7.967	.2158	-.012	36.921	36.909	1.1179
3.0310	.767	1202.0	201.06	249.54	11.123	.2137	-.011	52.056	52.045	1.0735
3.0300	.767	1201.6	200.94	249.92	11.123	.2121	-.012	52.432	52.419	1.0701
4.0100	.767	999.9	201.04	210.36	2.302	.2085	-.019	11.043	11.024	1.1819
4.0210	.767	1001.8	200.84	220.02	4.440	.2069	-.021	21.456	21.435	1.1173
4.0200	.767	1002.0	200.82	219.94	4.440	.2060	-.021	21.550	21.530	1.1264
4.0300	.768	998.9	200.56	239.72	7.926	.1972	-.021	40.196	40.175	1.0260
4.0400	.768	1001.8	201.05	257.72	10.338	.1884	-.020	54.887	54.867	.9682
4.0500	.766	998.9	200.99	276.17	13.043	.1885	-.015	69.201	69.186	.9202
5.0100	.766	497.8	201.00	210.83	1.267	.2073	-.026	6.112	6.086	.6192
5.0200	.766	498.2	201.18	220.77	2.503	.2075	-.032	12.060	12.028	.6142
5.0300	.766	498.4	200.92	240.09	4.968	.2072	-.032	23.975	23.943	.6114
5.0400	.766	498.1	201.02	258.49	7.244	.2071	-.037	34.989	34.952	.6081
5.0500	.766	498.0	201.39	277.07	9.496	.2066	-.036	45.971	45.935	.6069
6.0100	.770	246.8	199.22	209.08	1.132	.2128	-.082	5.317	5.235	.5305
6.0200	.770	247.5	200.61	220.54	2.280	.2138	-.084	10.662	10.579	.5307
6.0300	.767	247.7	201.03	239.95	4.478	.2137	-.088	20.952	20.864	.5362
6.0400	.767	247.3	201.04	259.02	6.705	.2135	-.089	31.402	31.314	.5401
6.0500	.767	247.8	201.03	276.47	8.780	.2133	-.092	41.160	41.068	.5444
7.0110	.767	1499.0	251.54	261.21	1.986	.2165	-.007	9.169	9.162	.9475
7.0100	.766	1498.6	251.02	260.60	1.996	.2185	-.007	9.134	9.126	.9529
7.0210	.766	1499.8	251.04	269.93	3.832	.2158	-.007	17.754	17.747	.9394
7.0200	.766	1498.2	251.02	270.09	3.832	.2134	-.007	17.950	17.944	.9409
7.0300	.766	1499.0	251.12	282.32	6.032	.2099	-.007	28.745	28.738	.9210
7.0400	.766	1498.5	251.02	299.64	6.859	.1556	-.005	44.086	44.081	.9067
8.0100	.767	1697.9	250.61	263.00	2.499	.2190	-.004	11.413	11.409	.9208
8.0200	.767	1701.2	251.17	276.97	4.994	.2149	-.005	23.236	23.232	.9007
8.0300	.767	1698.1	250.39	293.12	8.124	.2105	-.005	38.598	38.593	.9032
9.0100	.767	2000.9	251.02	264.72	2.412	.2012	-.002	11.989	11.987	.8751
9.0210	.767	2000.9	251.02	275.70	4.530	.2120	-.002	21.374	21.371	.8657
9.0200	.767	1999.1	251.02	278.40	4.530	.1883	-.001	24.062	24.061	.8788
10.0100	.766	1001.9	151.73	165.64	3.713	.2220	-.004	16.727	16.723	1.2019
10.0200	.766	999.9	151.70	178.86	7.685	.2192	-.004	35.055	35.051	1.2902
10.0300	.766	1001.9	151.67	192.58	11.871	.2188	-.006	54.262	54.256	1.3264
10.0400	.765	999.6	151.61	205.28	15.196	.2163	-.006	70.271	70.265	1.3092
11.0100	.765	501.5	156.63	173.26	2.539	.2195	-.045	11.566	11.521	.6930
11.0200	.765	501.8	156.59	189.69	4.884	.2198	-.058	22.215	22.157	.6694
11.0300	.765	501.2	156.58	206.68	7.182	.2195	-.054	32.727	32.673	.6522
12.0100	.766	1499.4	102.05	112.94	2.036	.2610	-.000	7.803	7.803	.7166
12.0200	.766	1500.9	102.03	123.61	4.010	.2561	-.000	15.656	15.656	.7256
12.0300	.765	1498.1	102.08	144.61	8.046	.2500	-.000	32.186	32.186	.7569
12.0400	.765	1498.4	101.93	183.53	16.446	.2506	-.000	65.636	65.636	.8043
13.0100	.764	1998.4	101.65	112.00	1.831	.2587	.000	7.078	7.078	.6866
13.0200	.764	1998.7	101.74	123.49	3.548	.2364	.000	15.011	15.011	.6900
13.0300	.764	1998.2	101.82	144.27	7.404	.2478	.000	29.882	29.882	.7040
13.0400	.764	1998.2	101.67	184.37	14.766	.2424	.000	60.921	60.921	.7367
14.0100	.765	998.3	101.71	119.43	3.588	.2615	-.001	13.719	13.718	.7742
14.0200	.765	999.0	101.67	136.51	7.519	.2564	-.001	29.330	29.330	.8419
14.0300	.765	1000.1	101.68	153.35	12.130	.2600	-.001	46.646	46.645	.9028
15.0100	.768	246.4	112.28	122.71	1.557	.2714	-.224	5.736	5.512	.5281
15.0200	.768	246.4	112.25	133.18	3.060	.2711	-.228	11.290	11.063	.5285
15.0300	.768	246.5	112.26	154.76	6.126	.2712	-.201	22.586	22.385	.5266
15.0400	.768	246.6	112.26	180.77	9.823	.2713	-.239	36.203	35.963	.5250
16.0100	.766	700.0	102.47	113.25	2.175	.1933	-.000	11.254	11.254	*
16.0200	.767	698.5	102.48	124.22	4.604	.1933	-.000	23.820	23.820	*
16.0300	.767	698.7	102.43	134.81	7.284	.1939	-.000	37.571	37.571	*
16.0400	.767	697.4	102.46	145.87	10.491	.1930	-.000	54.365	54.365	*
16.0500	.766	699.1	102.44	154.92	13.609	.1927	-.000	70.633	70.633	*
16.0600	.766	698.8	102.49	164.82	13.597	.1459	-.000	93.193	93.193	*
16.0700	.766	700.9	102.44	170.06	15.608	.1460	-.000	106.912	106.912	*
16.0800	.766	701.0	102.45	175.10	16.498	.1460	-.000	113.033	113.033	*
16.0900	.766	698.4	102.46	182.22	17.480	.1449	-.000	120.622	120.622	*
16.1000	.766	699.4	102.46	189.96	18.529	.1454	-.000	127.405	127.405	*
17.0100	.769	698.7	47.35	51.14	.543	.2195	.000	2.473	2.473	.6539
17.0200	.769	698.8	47.49	53.59	.856	.2131	.000	4.016	4.016	.6588
17.0300	.769	699.4	47.48	55.98	1.236	.2126	.000	5.815	5.815	.6842
17.0400	.769	698.7	47.48	58.61	1.613	.2115	.000	7.626	7.626	.6854
17.0500	.768	698.6	47.50	61.36	2.011	.2111	.000	9.526	9.526	.6875
17.0600	.768	699.2	47.50	63.62	2.356	.2112	.000	11.155	11.155	.6921
17.0700	.768	699.0	47.50	66.14	2.718	.2102	.000	12.931	12.931	.6937
17.0800	.768	699.8	47.48	68.16	3.044	.2119	.000	14.366	14.366	.6947
17.0900	.768	698.6	47.50	72.55	3.727	.2116	.000	17.609	17.609	*
17.1000	.769	699.6	47.50	77.72	4.580	.2123	.000	21.570	21.570	*
17.1100	.769	699.4	47.51	75.42	4.177	.2121	.000	19.697	19.697	*
17.1200	.769	699.1	47.51	82.81	5.406	.2073	.000	26.083	26.083	*
17.1300	.769	699.3	47.51	96.49	5.420	.1396	.000	38.822	38.822	*
17.1400	.769	698.7	47.52	110.28	7.321	.1394	.000	52.528	52.528	*
17.1500	.769	700.7	47.51	124.88	9.566	.1390	.000	68.812	68.812	*
17.1600	.769	700.2	47.50	137.86	11.991	.1387	.000	86.431	86.431	*
17.1700	.769	699.1	47.53	155.03	15.891	.1382	.000	114.962	114.962	*
18.0100	.764	998.5	22.91	30.80	1.478	.2945	.000	5.019	5.019	.6363
18.0200	.764	1000.1	22.90	38.93	3.037	.2958	.000	10.265	10.265	.6405
18.0300	.764	1000.0	22.82	53.07	6.390	.3266	.000	19.564	19.564	.6467
19.0100	.766	998.2	24.26	55.02	6.144	.3106	.000	19.777	19.777	.6431
20.0100	.766	998.3	24.26	55.45	6.354	.3136	.000	20.261	20.261	.6496
20.0200	.766	999.4	24.31	55.49	5.268	.2618	.000	20.117	20.117	.6452
20.0500	.766	998.2	24.32	84.48	11.852	.2926	.000	40.507	40.507	.6733
20.0300	.766	998.7	24.31	55.62	3.737	.1837	-.000	20.344	20.344	.6497
20.0400	.766	999.7	24.27	56.19	2.293	.1128	-.000	20.326	20.326	.6367

TABLE L
(CONTINUED)

RUN	MOLE FRACTION C_3H_8	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER / FLOW (Btu/lb)	ΔH_P (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_P$ (Btu/lb-°F)
21.0100	.766	1502.0	24.30	32.24	1.461	.2960	.000	4.935	4.935	.6223
21.0200	.766	1498.7	24.28	39.82	2.823	.2905	.000	9.717	9.718	.6253
21.0300	.766	1501.7	24.31	54.78	5.555	.2883	.000	19.269	19.270	.6324
21.0400	.766	1501.1	24.32	84.32	11.026	.2837	.000	38.867	38.868	.6478
22.0100	.766	1999.7	24.29	32.02	1.371	.2907	.000	4.716	4.716	.6098
22.0200	.766	1999.1	24.28	40.45	2.789	.2812	.000	9.919	9.920	.6132
22.0300	.765	1999.2	24.29	55.53	5.237	.2708	.000	19.343	19.343	.6193
22.0400	.765	1999.2	24.31	85.06	10.381	.2705	.000	38.378	38.378	.6317
23.0100	.766	499.4	-.09	3.61	.495	.2162	.000	2.292	2.292	.6203
23.0200	.766	500.2	-.09	6.06	.821	.2151	.000	3.817	3.817	.6204
23.0300	.766	499.8	-.09	8.69	1.170	.2149	.000	5.446	5.446	.6206
23.0400	.766	498.6	-.06	11.23	1.513	.2143	.000	7.058	7.058	.6250
23.0500	.766	498.7	-.07	14.00	1.928	.2142	.000	9.002	9.002	*
23.0600	.765	500.9	-.08	16.92	2.402	.2147	.000	11.189	11.189	*
23.0700	.765	499.5	-.08	19.77	2.873	.2142	.000	13.415	13.415	*
23.0800	.765	498.3	-.08	31.89	4.887	.2132	.001	22.921	22.922	*
23.0900	.766	499.7	-.09	47.39	7.583	.2134	.001	35.540	35.541	*
23.1000	.766	498.5	-.09	69.78	11.815	.2132	.001	55.408	55.409	*
23.1100	.766	498.3	-.07	91.41	16.564	.2129	.000	77.798	77.799	*
24.0100	.766	251.1	-.09	15.98	2.319	.1697	.001	13.663	13.664	*
24.0200	.766	248.1	-.07	31.02	4.722	.1676	.001	28.181	28.182	*
24.0300	.766	248.9	-.09	46.53	7.686	.1675	.001	45.879	45.880	*
24.0400	.766	249.8	-.07	61.77	11.422	.1678	.001	68.053	68.054	*
24.0500	.766	249.6	-.05	76.45	16.483	.1674	.001	98.472	98.473	*
24.0600	.767	247.8	-.05	88.62	23.385	.1658	.001	141.005	141.006	*
24.0700	.767	248.4	-.14	91.82	19.053	.1228	.001	155.196	155.197	*
24.0800	.767	251.0	-.08	94.90	20.724	.1239	.001	167.296	167.296	*
24.0900	.767	250.0	.01	102.15	21.781	.1234	.001	176.548	176.549	*
24.1000	.767	249.7	.01	113.42	22.561	.1232	.001	183.066	183.067	*
24.1100	.767	248.9	.00	118.18	22.820	.1227	.001	185.920	185.921	*
25.0100	.765	399.0	58.70	69.97	1.602	.1345	-.000	11.913	11.913	*
25.0200	.765	398.5	58.65	80.93	3.346	.1358	-.000	24.648	24.648	*
25.0300	.765	399.9	58.66	91.95	5.381	.1365	-.000	39.411	39.411	*
25.0400	.765	399.4	58.66	102.68	7.799	.1364	.000	57.178	57.178	*
25.0500	.766	398.8	58.70	113.32	10.838	.1359	-.000	79.731	79.731	*
25.0600	.766	399.6	58.61	124.29	15.309	.1359	-.000	112.693	112.692	*
25.0700	.766	398.3	58.62	129.87	18.457	.1354	-.000	136.288	136.288	*
25.0800	.766	399.7	58.63	130.87	18.819	.1360	-.000	138.385	138.385	*
25.0900	.766	399.4	58.63	135.38	19.514	.1359	-.000	143.592	143.592	*
25.1000	.767	398.8	58.73	141.26	19.965	.1355	-.000	147.391	147.391	*
25.1100	.767	398.9	58.67	145.84	20.418	.1358	-.000	150.333	150.333	*
26.0100	.767	502.1	82.72	93.51	1.611	.1344	-.001	11.983	11.983	*
26.0200	.767	498.6	82.52	104.26	3.434	.1327	-.000	25.868	25.868	*
26.0300	.767	498.9	82.52	114.94	5.540	.1335	-.000	41.493	41.493	*
26.0400	.765	498.6	82.49	125.63	8.164	.1328	-.001	61.475	61.474	*
26.0500	.765	499.1	82.49	135.60	11.197	.1324	-.001	84.549	84.548	*
26.0600	.765	500.7	82.61	145.99	15.504	.1330	-.001	116.574	116.574	*
26.0700	.765	501.2	82.63	147.50	16.168	.1327	.000	121.806	121.806	*
26.0800	.765	501.0	82.71	150.61	16.572	.1324	-.001	125.133	125.132	*
26.0900	.765	500.2	82.57	154.79	16.922	.1320	-.001	128.213	128.212	*
26.1000	.765	499.5	82.56	159.26	17.369	.1320	-.001	131.580	131.579	*
27.0100	.764	1302.1	139.72	153.36	2.872	.1864	-.002	15.406	15.404	1.1297
27.0200	.765	1300.0	140.06	168.20	5.717	.1793	-.002	31.886	31.884	1.1332
27.0300	.765	1300.0	140.18	181.49	8.715	.1767	-.002	49.328	49.326	1.1940
27.0400	.765	1300.1	139.99	193.18	11.676	.1755	-.003	66.549	66.547	1.2511
28.0100	.766	1499.8	157.32	167.77	1.822	.2088	-.017	8.722	8.705	.8337
28.0210	.766	1497.9	158.13	178.64	3.496	.1988	-.016	17.583	17.566	.8563
28.0200	.766	1497.9	158.13	178.45	3.496	.1973	-.016	17.716	17.699	.8711
29.0100	.766	1500.4	181.71	191.65	1.820	.2001	-.016	9.098	9.081	.9134
29.0200	.766	1498.6	181.70	200.90	3.671	.2067	-.020	17.760	17.740	.9241
30.0100	.766	1198.4	191.74	198.40	1.506	.1924	-.078	7.827	7.749	1.1629
30.0200	.765	1199.2	191.75	204.94	2.910	.1891	-.081	15.387	15.306	1.1599
31.0100	.765	495.5	199.94	219.23	2.420	.1983	-.423	12.205	11.781	.6107
31.0200	.765	493.1	199.98	238.16	4.650	.1971	-.459	23.585	23.126	.6057
32.0100	.765	1499.6	213.80	220.43	1.263	.1906	-.037	6.626	6.589	.9936
32.0200	.765	1497.7	213.73	226.75	2.408	.1855	-.039	12.980	12.941	.9941
32.0300	.765	1499.7	213.75	232.85	3.746	.2004	-.047	18.688	18.641	.9762
32.0400	.765	1498.4	213.73	239.47	5.050	.1975	-.050	25.565	25.516	.9915
33.0100	.765	1701.1	221.24	230.68	1.765	.2008	-.027	8.790	8.763	.9280
33.0200	.765	1697.9	221.20	240.10	3.509	.2000	-.029	17.539	17.510	.9268
33.0300	.765	1698.6	221.23	249.65	5.279	.1994	-.031	26.482	26.451	.9307
34.0100	.765	1998.5	231.07	240.55	1.746	.2133	-.016	8.183	8.167	.8607
34.0200	.766	1998.6	231.02	249.95	3.430	.2096	-.018	16.368	16.350	.8638
34.0300	.766	2001.0	230.97	259.33	5.105	.2079	-.019	24.557	24.538	.8652
34.0400	.766	1997.3	231.04	268.58	6.569	.2012	-.019	32.644	32.624	.8689
34.0500	.766	1999.2	231.07	277.43	8.067	.2002	-.020	40.293	40.273	.8687
34.0600	.766	2000.1	230.95	286.43	9.670	.2007	-.021	48.185	48.164	.8682
35.0100	.766	1697.9	221.09	258.56	7.285	.2092	-.029	34.817	34.788	.9282
35.0210	.766	1697.7	220.32	267.08	9.048	.2100	-.038	43.075	43.037	.9203
35.0200	.766	1700.5	220.33	267.18	9.047	.2094	-.038	43.207	43.169	.9215
36.0300	.766	1998.9	-57.38	-21.58	4.218	.2090	.000	20.183	20.183	.5638
36.0400	.766	2002.1	-57.32	9.51	7.241	.1885	.000	38.405	38.406	.5747
36.0100	.766	1999.3	-57.40	-48.51	1.350	.2731	.000	4.943	4.943	.5560
36.0200	.766	2001.9	-57.38	-38.74	2.476	.2388	.000	10.370	10.370	.5563
37.0100	.765	1498.2	-57.43	-48.95	1.309	.2771	.000	4.724	4.725	.5573
37.0200	.765	1498.8	-57.41	-40.31	2.719	.2839	.000	9.580	9.580	.5603
37.0300	.765	1499.5	-57.39	-23.16	5.369	.2771	.000	19.374	19.375	.5660
37.0400	.765	1500.2	-57.30	9.98	10.627	.2738	.000	38.818	38.818	.5770
38.0200	.765	999.2	-57.37	-40.38	2.986	.3130	.000	9.542	9.542	.5616
38.0300	.765	1000.0	-57.38	-22.93	6.020	.3075	.000	19.578	19.578	.5683
38.0400	.765	998.7	-57.39	10.04	11.907	.3040	.000	39.167	39.167	.5809
38.0100	.765	999.0	-57.29	-47.08	1.201	.2077	.000	5.781	5.781	.5662
39.0100	.765	499.3	-57.55	-48.83	1.338	.2695	.000	4.967	4.967	.5699
39.0200	.765	499.0	-57.40	-39.05	2.845	.2690	.000	10.577	10.578	.5765
39.0300	.765	499.3	-57.51	-22.68	5.452	.2689	.000	20.278	20.278	.5821
40.0100	.766	2002.4	-147.56	-137.51	1.411	.2710	.000	5.206	5.206	.5176
40.0200	.765	1999.8	-147.56	-126.99	2.868	.2680	.000	10.700	10.700	.5202
40.0300	.765	1998.8	-147.56	-107.33	5.712	.2721	.000	20.996	20.996	.5219
40.0400	.765	2000.5	-147.56	-68.43	11.050	.2646	.000	41.759	41.760	.5277
41.0100	.766	998.5	-147.60	-137.68	1.494	.2904	.000	5.145	5.145	.5185
41.0200	.766	999.5	-147.60	-126.72	3.119	.2878	.000	10.836	10.836	.5190
41.0300	.766	1001.1	-147.51	-107.98	6.128	.2947	.000	20.796	20.796	.5261
41.0400	.766	1001.3	-147.52	-69.52	12.258	.2940	.000	41.690	41.690	.5345

TABLE L
(CONTINUED)

RUN	MOLE FRACTION C_3H_8	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER/FLOW (Btu/lb)	$\Delta \underline{H}_p$ (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_p$ (Btu/lb-°F)
42.0100	.765	500.8	-147.60	-137.48	1.511	.2846	.000	5.310	5.311	.5249
42.0200	.765	500.8	-147.52	-126.14	3.217	.2847	.001	11.299	11.300	.5285
42.0300	.765	501.8	-147.59	-107.52	6.045	.2841	.000	21.276	21.277	.5310
42.0400	.765	498.3	-147.54	-70.08	11.815	.2833	.000	41.704	41.705	.5384
43.0100	.765	248.5	-217.80	-206.15	1.770	.2999	.000	5.904	5.904	.5068
43.0200	.765	248.3	-217.82	-193.59	3.843	.3097	.000	12.410	12.411	.5122
43.0300	.765	249.6	-217.88	-170.05	7.349	.3029	.000	24.263	24.263	.5073
43.0400	.765	251.0	-217.78	-128.55	13.702	.2977	.000	46.020	46.020	.5158
44.0100	.766	2001.7	-236.16	-223.40	1.792	.2797	.000	6.407	6.408	.5021
44.0200	.766	2000.9	-236.12	-209.98	3.597	.2764	.000	13.012	13.012	.4978
44.0300	.766	1999.3	-236.10	-186.05	6.737	.2684	.000	25.101	25.101	.5015
44.0400	.766	2000.3	-236.07	-142.71	12.458	.2631	.000	47.358	47.358	.5073
45.0100	.766	998.5	-236.11	-222.48	1.943	.2855	.000	6.806	6.806	.4996
45.0200	.766	998.7	-236.07	-210.40	3.637	.2836	.000	12.825	12.825	.4996
45.0300	.766	999.2	-236.11	-187.54	6.971	.2851	.000	24.451	24.451	.5035
45.0400	.766	1000.3	-236.07	-141.99	13.633	.2850	.000	47.837	47.838	.5084
46.0100	.766	499.0	-236.07	-223.02	1.884	.2884	.000	6.532	6.533	.5009
47.0100	.765	1000.3	-57.14	-48.82	1.390	.2975	.000	4.672	4.672	.5621
47.0200	.765	998.7	-57.22	-39.16	3.058	.3015	.000	10.142	10.142	.5615
47.0300	.765	998.7	-57.19	-20.91	5.971	.2908	.000	20.530	20.531	.5659
48.0100	.765	1006.2	-147.07	-135.85	1.631	.2789	.000	5.847	5.847	.5213
48.0200	.766	1004.2	-147.05	-126.77	2.912	.2749	.000	10.594	10.595	.5224
48.0300	.766	1006.2	-146.91	-106.53	5.825	.2731	.000	21.329	21.329	.5281
48.0400	.766	1009.2	-146.91	-68.85	11.484	.2730	.000	42.058	42.058	.5388
49.0100	.765	254.2	-107.37	-104.34	.265	.1623	.000	1.632	1.632	.5400
49.0200	.765	254.2	-107.38	-101.46	.535	.1622	.000	3.302	3.302	.5574
49.0300	.765	254.2	-107.38	-98.75	.761	.1618	.000	4.704	4.704	.5446
49.0400	.765	257.2	-107.38	-95.79	1.040	.1637	.000	6.353	6.353	.5484
49.0500	.765	256.2	-107.37	-93.05	1.284	.1636	.000	7.847	7.847	.5481
49.0600	.765	257.2	-107.38	-89.95	1.562	.1636	.000	9.548	9.548	.5477
49.0700	.765	256.2	-107.36	-86.79	1.848	.1627	.000	11.358	11.358	.5521
49.0800	.767	256.2	-107.39	-83.92	2.108	.1641	.000	12.842	12.842	.5471
49.0900	.767	256.1	-107.39	-81.12	2.353	.1624	.000	14.490	14.490	.5516
49.1000	.767	256.0	-107.62	-78.51	2.658	.1630	.001	16.311	16.312	*
49.1100	.767	254.0	-107.54	-75.77	2.957	.1612	.001	18.338	18.338	*
49.1200	.767	255.0	-107.38	-71.94	3.381	.1617	.001	20.910	20.910	*
49.1300	.763	254.9	-107.38	-60.47	4.726	.1606	.001	29.426	29.426	*
49.1400	.763	253.9	-107.36	-34.39	7.718	.1607	.001	48.036	48.037	*
49.1500	.768	254.0	-107.37	-9.65	10.658	.1608	.001	66.293	66.294	*
49.1600	.768	256.9	-107.38	13.81	13.874	.1623	.001	85.504	85.505	*
50.0100	.764	406.1	-47.83	-42.46	.514	.1656	.000	3.103	3.103	.5781
50.0200	.764	406.2	-47.76	-37.45	.990	.1653	.000	5.989	5.990	.5810
50.0300	.764	407.1	-47.71	-32.11	1.518	.1651	.000	9.197	9.197	.5892
50.0400	.764	407.1	-47.77	-27.04	2.015	.1651	.000	12.204	12.204	.5885
50.0500	.764	408.1	-47.75	-22.15	2.541	.1660	.000	15.309	15.310	*
50.0600	.764	408.1	-47.73	-17.20	3.141	.1661	.000	18.904	18.905	*
50.0700	.764	408.1	-47.71	-11.84	3.803	.1659	.000	22.927	22.928	*
50.0800	.764	408.1	-47.73	2.58	5.567	.1655	.000	33.649	33.649	*
50.0900	.764	408.1	-47.69	18.72	7.614	.1655	.000	45.995	45.995	*
50.1000	.764	409.1	-47.69	40.74	10.665	.1665	.000	64.054	64.054	*
50.1100	.764	408.1	-47.74	65.75	14.599	.1652	.000	88.379	88.379	*
51.0100	.765	501.0	200.63	211.78	1.446	.2094	-.022	6.905	6.884	.6175
51.0200	.765	500.7	200.64	220.20	2.501	.2086	-.016	11.989	11.973	.6121
51.0300	.765	498.7	200.60	239.17	4.879	.2075	-.025	23.512	23.487	.6089
51.0400	.765	498.6	200.74	275.97	9.351	.2049	-.032	45.631	45.599	.6061

TABLE LI

TABULATED EXPERIMENTAL ISOTHERMAL DATA FOR THE NOMINAL 77 PERCENT MIXTURE

Run	Mole Fraction C_3H_8	Inlet Temperature (OF)	Inlet Pressure (psi)	Pressure Drop (psi)	Power (Btu/min)	Flow (lb/min)	Corr. (Btu/lb)	$\Delta \underline{H}_T$ (Btu/lb)	$(\Delta \underline{H}/\Delta P)_T$ (Btu/lb-psi)
1.010	.766	200.9	2029.7	105.6	.3229	.29048	.0097	1.112	-.010437
1.020	.766	200.9	1893.1	112.1	.4105	.29500	.0137	1.392	-.012292
1.030	.767	200.9	1830.9	117.3	.5209	.29970	.0077	1.738	-.014754
1.040	.767	200.9	1731.1	127.1	.7158	.30957	.0052	2.312	-.018154
1.050	.767	200.9	1622.1	131.9	.9523	.31073	.0091	3.065	-.023168
1.060	.767	200.9	1529.9	137.1	1.2340	.31175	-.0232	3.958	-.029044
1.070	.767	200.9	1403.2	144.3	1.8854	.31189	.0240	6.045	-.041729
1.080	.767	200.9	1277.0	158.7	3.2092	.31026	-.0315	10.343	-.065380
1.090	.767	201.2	1111.9	161.0	4.5689	.28263	.0072	16.166	-.100374
1.100	.767	201.2	950.7	169.4	4.6841	.24350	.0215	19.236	-.113439
1.110	.767	200.9	801.2	149.5	2.7615	.19260	-.0022	14.338	-.095927
1.120	.767	200.9	657.3	239.3	3.4539	.19618	.0017	17.605	-.073570
1.130	.767	201.0	482.2	285.0	2.6800	.15830	-.0194	16.929	-.059475
1.140	.767	200.9	361.0	251.5	1.6074	.11916	.0108	13.490	-.053599
2.010	.765	251.3	2022.3	144.3	.9540	.31595	.0278	3.019	-.020733
2.020	.765	251.5	1904.7	154.4	1.2602	.32102	.0157	3.926	-.025325
2.030	.765	251.3	1747.2	158.2	1.6519	.31360	-.0069	5.268	-.033344
2.040	.765	250.7	1632.3	176.0	2.3237	.32310	.0182	7.192	-.040763
2.050	.765	252.1	1474.4	179.1	2.8521	.30320	-.0012	9.407	-.052533
2.060	.765	250.9	1315.7	204.2	3.8209	.29632	.0001	12.894	-.063151
2.070	.765	250.9	1112.3	233.4	4.2942	.27570	-.0070	15.576	-.066770
2.080	.765	250.8	893.7	201.5	2.6726	.21480	.0178	12.442	-.061665
2.090	.766	250.9	660.4	187.5	1.6425	.16383	.0200	10.026	-.053368
2.100	.766	250.9	506.7	212.8	1.4416	.14012	.0132	10.289	-.048291
2.110	.766	251.0	390.2	283.2	1.5539	.12321	.0223	12.612	-.044459
3.010	.764	99.8	1998.7	179.7	.0200	.21569	.0158	.093	-.000427
3.020	.764	99.8	1856.1	178.5	.0338	.21338	.0159	.158	-.000799
3.030	.764	99.8	1701.9	192.0	.0558	.21979	.0121	.254	-.001259
3.040	.764	99.8	1550.1	205.0	.0891	.22275	.0094	.400	-.001906
3.050	.764	100.0	1388.2	204.6	.1265	.22348	-.0012	.566	-.002773
3.060	.764	100.0	1206.3	211.2	.1889	.22575	-.0040	.837	-.003980
3.070	.764	99.9	1005.1	145.0	.1464	.17960	-.0154	.815	-.005728
3.080	.764	99.8	990.5	177.6	.2078	.20150	-.0169	1.031	-.005904
3.090	.764	99.9	1005.7	254.2	.7678	.24986	-.0080	3.073	-.012122
3.100	.764	99.8	990.6	291.6	1.3169	.26841	.0146	4.906	-.016777
3.110	.764	99.8	1006.1	396.2	3.5425	.32087	.0269	11.040	-.027800
3.120	.764	100.0	640.0	140.0	1.4012	.12647	-.0080	11.079	-.079202
3.130	.764	99.9	640.9	236.2	3.8363	.15380	-.0164	24.943	-.105682
3.140	.764	99.9	641.4	324.8	8.7386	.17022	.0079	51.336	-.158044
3.150	.764	99.9	519.2	242.8	8.5695	.12230	.0125	70.068	-.288559
3.160	.764	99.8	474.8	307.5	12.1448	.11650	.0010	104.244	-.339032
3.170	.764	99.8	476.7	369.4	13.7217	.12532	.0042	109.493	-.296424

TABLE LII

TABULATED EXPERIMENTAL JOULE-THOMSON DATA
FOR THE NOMINAL 77 PERCENT MIXTURE

Run	Mole Fraction Propane	Inlet Temperature (°F)	Inlet Pressure (psia)	Temperature Difference (°F)	Pressure Drop (psi)	$-\left(\frac{\Delta T}{\Delta P}\right)_H$ (°F/psi)x10^3
4.010	0.761	-96.8	1942.3	1.285	221.0	6.00
4.020	0.761	-96.7	1796.2	1.331	226.0	5.89
4.030	0.761	-96.7	1451.8	1.269	220.2	5.77
4.040	0.761	-96.8	1173.6	1.229	216.0	5.69
4.050	0.761	-96.8	912.3	1.213	214.8	5.65
4.060	0.761	-97.0	694.5	1.183	213.2	5.55
4.070	0.761	-96.7	602.1	1.211	219.4	5.52

TABLE LII

FLOWMETER CALIBRATION EQUATION CONSTANTS
USED FOR THE NOMINAL 77 PERCENT MIXTURE

Flow Meter Series		B	A	C	D
10,20,30	low	0.074556366	59.519280	-21616.839	3479232.9
	high	0.12003127	16.100944	-8125.2457	2118787.9
40 - 50	low	0.10684014	15.203628	-1462.5168	535593.14
	high	-0.16666467	315.17891	-110295.06	1360895.0
60 - 70	low	0.087080753	43.093003	-14088.483	2339258.0
	high	0.15376994	-13.008022	7.5220947	1395307.2

TABLE LIV

TABULATED EXPERIMENTAL ISOBARIC DATA FOR THE NOMINAL 51 PERCENT MIXTURE

RUN	MOLE FRACTION C_3H_8	INLET PRESSURE (psia)	INLET TEMPERATURE (°F)	OUTLET TEMPERATURE (°F)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER / FLOW (Btu/lb)	ΔH_p (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_p$ (Btu/lb-°F)
1.0100	.505	2000.1	200.54	209.92	1.641	.2051	-.001	8.004	8.003	.8529
1.0200	.505	2002.1	200.51	220.02	3.408	.2063	-.001	16.518	16.516	.8465
1.0300	.505	1998.3	200.54	238.76	6.559	.2039	-.002	32.166	32.165	.8414
1.0400	.505	1998.6	200.57	275.92	12.500	.2025	-.002	61.713	61.711	.8190
2.0100	.506	1502.0	200.58	209.96	1.564	.1985	-.003	7.877	7.874	.8393
2.0200	.506	1499.9	200.56	220.06	3.186	.1976	-.003	16.120	16.117	.8265
2.0300	.506	1501.3	200.57	238.69	6.099	.1980	-.002	30.800	30.798	.8080
2.0400	.506	1500.4	200.56	275.69	11.571	.1979	-.002	58.453	58.450	.7780
3.0100	.506	1000.1	200.54	210.69	1.425	.1991	-.019	7.156	7.137	.7032
3.0200	.506	998.6	200.58	220.29	2.740	.1988	-.021	13.780	13.759	.6980
3.0300	.506	998.0	200.46	238.91	5.246	.1989	-.022	26.378	26.356	.6856
3.0400	.506	998.0	200.48	275.62	10.065	.1986	-.022	50.694	50.672	.6744
4.0100	.506	498.1	200.55	210.26	1.108	.1983	-.018	5.588	5.570	.5734
4.0200	.506	498.4	200.57	220.10	2.236	.1987	-.019	11.252	11.233	.5752
4.0300	.506	499.9	200.60	238.86	4.398	.1988	-.019	22.115	22.096	.5775
4.0400	.506	498.8	200.60	275.96	8.628	.1970	-.021	43.798	43.777	.5810
5.0110	.507	245.7	200.76	210.20	1.080	.1990	-.116	5.426	5.309	.5625
5.0100	.507	251.0	200.74	210.68	1.067	.2018	-.034	5.287	5.252	.5284
5.0200	.507	251.3	200.77	220.57	2.144	.2016	-.032	10.632	10.599	.5355
5.0300	.507	250.9	200.69	238.71	4.129	.2014	-.031	20.503	20.472	.5385
5.0400	.507	251.0	200.77	275.77	8.270	.2012	-.024	41.107	41.082	.5478
6.0100	.506	1999.2	151.51	161.22	1.795	.2075	-.001	8.652	8.651	.8912
6.0200	.506	1999.0	151.53	171.21	3.643	.2070	-.001	17.598	17.597	.8940
6.0300	.506	2001.1	151.56	181.08	5.442	.2070	-.001	26.290	26.289	.8903
6.0400	.506	2001.8	151.51	191.19	7.290	.2070	-.001	35.214	35.213	.8874
6.0500	.506	2002.1	151.54	201.44	9.097	.2061	-.001	44.133	44.132	.8844
7.0100	.506	1500.8	151.52	161.76	1.810	.1793	-.005	10.095	10.090	.9850
7.0200	.506	1498.7	151.57	172.30	3.572	.1766	-.006	20.227	20.220	.9753
7.0300	.506	1499.4	151.57	181.52	5.048	.1765	-.008	28.597	28.589	.9544
7.0400	.506	1498.5	151.58	191.33	6.606	.1760	-.008	37.538	37.530	.9440
7.0500	.506	1498.9	151.56	201.05	8.084	.1756	-.009	46.030	46.021	.9299
8.0100	.505	999.3	141.96	158.71	2.982	.1979	-.011	15.068	15.057	.8991
8.0200	.505	1001.7	141.97	175.30	5.639	.1977	-.011	28.521	28.510	.8549
8.0300	.505	999.1	141.99	193.24	8.199	.1961	-.013	41.822	41.809	[illegible]
9.0100	.508	250.5	141.96	159.30	1.792	.1993	-.038	8.990	8.952	.5162
9.0200	.508	249.7	142.00	175.84	3.482	.1984	-.039	17.549	17.510	.5175
9.0300	.508	248.4	141.89	193.52	5.293	.1965	-.039	26.934	26.894	.5209
10.0100	.505	501.4	112.16	122.39	1.267	.1996	-.022	6.349	6.328	.6187
10.0200	.505	501.5	112.22	133.06	2.537	.1977	-.024	12.833	12.810	.6148
10.0300	.505	501.3	112.19	153.97	4.952	.1969	-.024	25.146	25.123	.6013
10.0400	.505	501.2	112.35	193.73	9.433	.1957	-.027	48.200	48.174	.5919
11.0100	.507	2002.3	101.72	111.92	1.663	.1962	-.000	8.478	8.478	.8317
11.0200	.507	2000.0	101.76	122.71	3.407	.1932	-.000	17.632	17.631	.8413
11.0300	.507	2001.2	101.73	133.54	5.173	.1916	-.000	26.992	26.992	.8486
11.0400	.507	1999.3	101.79	143.24	6.733	.1899	-.000	35.448	35.448	.8552
11.0500	.507	2000.2	101.85	154.56	8.498	.1864	-.000	45.589	45.589	.8649
12.0100	.506	1699.4	101.78	112.02	1.707	.1822	-.001	9.367	9.366	.9146
12.0200	.506	1699.0	101.75	122.83	3.531	.1806	-.001	19.551	19.550	.9273
12.0300	.506	1698.5	101.75	133.19	5.293	.1794	-.001	29.499	29.498	.9382
12.0400	.506	1699.5	101.74	143.09	6.977	.1789	-.001	38.994	38.993	.9430
12.0500	.506	1698.8	101.73	154.13	8.851	.1781	-.001	49.704	49.704	.9487
13.0100	.507	1499.4	101.79	112.36	1.879	.1775	-.001	10.588	10.587	1.0020
13.0200	.507	1498.4	101.84	122.64	3.712	.1758	-.002	21.111	21.109	1.0149
13.0300	.507	1498.7	101.80	133.30	5.642	.1753	-.002	32.179	32.177	1.0215
13.0400	.507	1499.8	101.89	144.09	7.563	.1747	-.002	43.290	43.288	1.0257
13.0500	.507	1498.8	101.74	153.85	9.281	.1741	-.002	53.317	53.315	1.0231
14.0100	.503	249.4	72.26	84.05	1.083	.1792	-.040	6.042	6.002	.5089
14.0200	.503	251.0	72.26	93.87	1.996	.1804	-.040	11.068	11.028	.5104
14.0300	.503	251.5	72.30	115.88	4.023	.1804	-.042	22.303	22.261	.5107
14.0400	.503	251.1	72.30	156.38	7.692	.1782	-.042	43.160	43.118	.5128
15.0100	.506	1298.4	52.19	66.77	2.445	.1871	-.000	13.068	13.067	.8960
15.0200	.506	1298.5	52.11	81.96	5.253	.1873	-.001	28.052	28.051	.9400
15.0300	.506	1299.2	52.21	96.54	7.644	.1808	-.001	42.284	42.283	.9538
15.0400	.506	1299.6	52.15	110.33	10.259	.1761	-.001	58.272	58.271	1.0016
15.0500	.506	1299.7	52.25	124.57	13.144	.1768	-.001	74.354	74.353	1.0281
15.0600	.506	1300.1	52.15	138.43	15.968	.1776	-.001	89.897	89.896	1.0419
16.0100	.506	701.1	52.29	63.07	2.127	.1787	-.003	11.900	11.897	1.1033
16.0200	.506	698.3	52.39	73.00	4.314	.1850	-.002	23.312	23.310	1.1312
16.0300	.506	698.3	52.45	81.92	4.439	.1280	-.000	34.670	34.670	1.1765
16.0400	.506	698.3	52.49	92.22	6.230	.1276	-.000	48.832	48.832	1.2292
16.0500	.506	698.7	52.47	102.04	8.148	.1276	-.000	63.871	63.871	1.2885
16.0600	.506	699.1	52.47	109.70	9.801	.1277	-.003	76.737	76.734	1.3409
16.0700	.506	699.5	52.43	117.26	11.578	.1277	-.003	90.656	90.653	1.3982
16.0800	.506	699.5	52.43	125.70	13.370	.1278	-.004	104.656	104.652	1.4283
16.0900	.506	700.1	52.44	134.61	14.193	.1279	-.009	110.956	110.947	1.3502
16.1000	.506	699.3	52.51	146.19	15.163	.1272	-.000	119.249	119.249	1.2728
16.1100	.506	699.8	52.50	155.79	16.003	.1271	-.000	125.866	125.866	1.2185
17.0100	.506	1999.3	22.72	34.91	1.686	.1971	.000	8.553	8.553	.7014
17.0200	.506	1998.6	22.51	46.12	3.251	.1941	.000	16.755	16.755	.7097
17.0300	.506	1998.8	22.72	68.60	6.399	.1921	.000	33.321	33.321	.7263
17.0400	.506	1999.9	22.80	111.91	12.962	.1903	.000	68.105	68.105	.7643
18.0100	.506	1500.8	22.76	34.03	1.571	.1883	-.000	8.347	8.347	.7404
18.0200	.506	1501.4	22.78	45.90	3.275	.1882	-.000	17.403	17.403	.7527
18.0300	.506	1499.7	22.78	68.80	1.939	.0532	-.000	36.468	36.468	.7925
18.0400	.506	1500.3	22.78	68.46	3.178	.0878	-.000	36.197	36.197	.7924
18.0500	.506	1499.8	22.71	68.16	5.803	.1618	-.000	35.859	35.859	.7889
18.0600	.506	1499.6	22.71	68.71	8.253	.2286	-.000	36.097	36.097	.7847
18.0700	.506	1498.5	22.73	68.53	10.596	.2927	-.000	36.196	36.196	.7903
18.0800	.506	1499.9	22.82	111.89	14.631	.1903	-.000	76.886	76.886	.8632
19.0100	.507	1000.8	2.69	6.03	.477	.1927	-.000	2.476	2.476	.7406
19.0200	.507	1000.2	2.76	9.65	.990	.1921	-.000	5.155	5.155	.7473
19.0300	.507	1000.3	2.69	10.86	1.182	.1916	-.000	6.166	6.166	.7549
19.0400	.507	998.6	2.69	12.89	1.565	.1918	-.000	8.161	8.161	*
19.0500	.507	999.8	2.71	16.08	2.140	.1920	-.000	11.146	11.146	*
19.0600	.507	998.8	2.68	19.45	2.781	.1910	-.000	14.560	14.559	*
19.0700	.507	1000.8	2.74	26.66	4.077	.1915	-.000	21.284	21.284	*
19.0800	.507	998.4	2.74	42.08	3.883	.1068	-.000	36.356	36.356	*
19.0900	.507	999.8	2.75	64.68	6.284	.1065	-.000	58.994	58.994	*
19.1000	.507	1000.4	2.75	86.37	8.739	.1064	-.000	82.145	82.145	*
19.1100	.507	999.8	3.91	102.02	10.554	.1057	-.000	99.851	99.851	*
19.1200	.507	999.2	3.97	115.20	12.341	.1053	-.000	117.145	117.145	*
19.1300	.507	998.8	3.93	122.07	13.320	.1049	-.000	126.931	126.931	*

TABLE LIV
(CONTINUED)

RUN	MOLE FRACTION C_3H_8	INLET PRESSURE (psia)	INLET TEMPERATURE (^oF)	OUTLET TEMPERATURE (^oF)	POWER (Btu/min)	FLOW (lb/min)	CORR. (Btu/lb)	POWER/FLOW (Btu/lb)	ΔH_p (Btu/lb)	$\left(\frac{\Delta H}{\Delta T}\right)_P$ (Btu/lb-^oF)
19.1400	.507	998.3	3.87	129.41	14.460	.1046	-.000	138.177	138.177	*
19.1500	.507	998.3	3.90	137.45	15.596	.1045	-.000	149.219	149.218	*
19.1600	.507	998.7	3.90	145.53	16.394	.1042	-.000	157.273	157.273	*
19.1700	.507	999.3	3.83	148.03	16.609	.1041	-.000	159.596	159.596	*
20.0100	.506	498.9	8.10	20.19	1.431	.1214	-.000	11.785	11.784	.9749
20.0200	.506	498.7	8.11	31.88	2.895	.1215	-.001	23.829	23.828	1.0023
20.0300	.506	502.0	8.13	51.49	5.720	.1230	-.001	46.486	46.486	1.0721
20.0400	.506	501.8	8.16	70.32	8.877	.1234	-.001	71.940	71.940	*
20.0500	.506	501.5	8.16	88.71	12.726	.1227	-.001	103.732	103.732	*
20.0600	.506	501.0	8.24	105.79	17.077	.1226	-.002	139.294	139.292	*
20.0700	.506	500.7	8.23	113.02	17.687	.1227	-.002	144.202	144.200	*
20.0800	.506	500.5	8.18	101.10	15.938	.1226	-.003	130.041	130.038	*
20.0900	.506	501.0	8.29	119.04	18.132	.1226	-.003	147.935	147.932	*
21.0100	.500	251.5	-26.52	-14.49	1.608	.1352	-.074	11.892	11.818	.9826
21.0200	.500	251.4	-26.46	-2.15	3.417	.1345	-.059	25.408	25.349	1.0424
21.0300	.500	250.3	-26.49	17.87	6.404	.1248	-.059	51.328	51.269	1.1557
21.0400	.500	250.1	-26.49	37.19	10.470	.1248	-.059	83.890	83.831	*
21.0500	.500	250.2	-26.44	55.83	16.049	.1246	-.089	128.848	128.759	*
21.0600	.500	250.5	-26.33	68.18	18.859	.1245	-.074	151.465	151.391	*
21.0700	.500	249.7	-26.42	78.87	19.516	.1242	-.089	157.082	156.994	*
21.0800	.500	250.4	-26.40	81.45	19.758	.1248	-.096	158.312	158.216	*
21.0900	.500	250.0	-26.39	59.57	17.404	.1246	-.089	139.731	139.642	*
21.1000	.500	250.6	-26.34	92.88	20.084	.1221	-.096	164.459	164.363	*
22.0100	.509	2001.0	-147.77	-137.02	1.499	.2419	.000	6.196	6.196	.5768
22.0200	.509	1998.3	-147.77	-125.85	3.047	.2413	.000	12.628	12.628	.5763
22.0300	.509	1999.5	-147.71	-106.09	5.776	.2393	.000	24.133	24.134	.5798
22.0400	.509	2001.7	-147.66	-69.32	11.063	.2402	.000	46.052	46.052	.5878
23.0100	.507	1499.2	-147.71	-136.91	1.542	.2470	.000	6.242	6.242	.5781
23.0200	.507	1499.8	-147.72	-126.50	3.045	.2470	.000	12.328	12.328	.5810
23.0300	.507	1499.3	-147.70	-107.06	5.872	.2471	.000	23.768	23.768	.5848
23.0400	.507	1499.6	-147.67	-69.57	11.472	.2471	.000	46.431	46.432	.5945
24.0100	.509	1000.0	-147.73	-137.24	1.534	.2513	.000	6.106	6.106	.5824
24.0200	.509	1000.5	-147.67	-126.85	3.052	.2512	.000	12.149	12.149	.5834
24.0300	.509	1000.2	-147.69	-107.00	6.024	.2513	.000	23.971	23.971	.5891
24.0400	.509	998.6	-147.68	-69.52	11.769	.2510	.000	46.887	46.887	.5999
25.0100	.508	499.8	-147.62	-131.85	2.151	.2319	.000	9.275	9.275	.5883
25.0200	.508	499.5	-147.57	-116.82	4.204	.2320	.000	18.125	18.126	.5895
25.0300	.508	500.5	-147.56	-87.81	8.336	.2323	.000	35.883	35.883	.6006
26.0100	.506	249.1	-147.61	-144.08	.253	.1212	.000	2.085	2.085	.5910
26.0200	.506	249.9	-147.59	-141.12	.466	.1217	.000	3.827	3.827	.5919
26.0300	.506	249.7	-147.57	-137.56	.718	.1214	.000	5.914	5.914	.5909
26.0400	.506	250.0	-147.58	-135.29	1.007	.1216	.000	8.284	8.284	*
26.0500	.506	250.2	-147.57	-132.14	1.685	.1214	.001	13.881	13.881	*
26.0600	.506	249.5	-147.56	-116.61	4.216	.1210	.001	34.852	34.853	*
26.0700	.506	248.7	-147.58	-88.18	7.409	.1208	.001	61.353	61.354	*
26.0800	.506	251.1	-147.49	-52.36	10.998	.1218	.001	90.295	90.296	*
26.0900	.506	250.5	-147.51	-19.74	14.492	.1213	.001	119.523	119.524	*
27.0100	.504	1998.8	-241.75	-227.71	2.060	.2648	.000	7.781	7.781	.5542
27.0200	.504	1998.1	-242.11	-215.85	3.877	.2681	.000	14.460	14.461	.5507
27.0300	.504	1999.7	-242.11	-191.38	7.561	.2691	.000	28.102	28.103	.5539
27.0400	.504	1999.0	-242.12	-147.25	14.182	.2660	.000	53.314	53.314	.5619
28.0100	.504	1499.9	-242.13	-228.78	1.836	.2496	.001	7.356	7.357	.5509
28.0200	.504	1499.3	-242.12	-216.20	3.573	.2494	.001	14.324	14.325	.5528
29.0100	.504	1000.2	-242.12	-229.13	1.836	.2556	.001	7.185	7.186	.5530
29.0200	.504	998.5	-242.11	-215.23	3.790	.2551	.001	14.859	14.860	.5528
29.0300	.504	999.5	-242.09	-191.39	7.174	.2549	.001	28.149	28.150	.5552
29.0400	.504	998.1	-242.08	-146.51	13.491	.2502	.001	53.912	53.913	.5641
30.0100	.504	498.8	-242.13	-229.53	1.802	.2553	.000	7.057	7.057	.5601
30.0200	.504	498.4	-242.14	-215.79	3.755	.2556	.000	14.696	14.696	.5577
30.0300	.504	498.1	-241.30	-191.21	7.193	.2555	.000	28.158	28.158	.5621
30.0400	.504	500.0	-242.10	-146.99	13.833	.2565	.000	53.929	53.930	.5671
31.0100	.501	248.0	-242.12	-227.90	1.982	.2503	.000	7.920	7.920	.5568
31.0200	.501	249.4	-242.11	-214.68	3.868	.2519	.000	15.355	15.355	.5597
31.0300	.501	249.2	-242.11	-190.77	7.260	.2511	.000	28.910	28.910	.5631
31.0400	.501	250.5	-242.08	-146.64	13.734	.2530	.000	54.283	54.283	.5687
32.0100	.504	498.5	-97.84	-93.81	.338	.1361	.000	2.481	2.481	.6158
32.0200	.504	498.5	-97.83	-89.97	.664	.1362	.000	4.879	4.879	.6206
32.0300	.504	498.8	-97.83	-86.30	.982	.1362	.000	7.208	7.209	.6250
32.0400	.504	498.5	-97.80	-82.62	1.330	.1363	.000	9.757	9.757	.6428
32.0500	.504	499.0	-97.82	-78.83	1.967	.1367	.001	14.384	14.385	*
32.0600	.504	499.1	-97.82	-74.46	2.684	.1369	.001	19.606	19.606	*
32.0700	.504	498.7	-97.81	-61.13	4.614	.1364	.000	33.819	33.819	*
32.0800	.504	499.5	-97.79	-25.78	8.934	.1354	.000	65.996	65.996	*
32.0900	.504	499.5	-97.74	15.44	14.035	.1355	.001	103.593	103.593	*
33.0100	.504	700.0	-57.29	-53.60	.344	.1395	.000	2.468	2.468	.6686
33.0200	.504	699.1	-57.24	-50.33	.634	.1394	.000	4.546	4.546	.6579
33.0300	.504	700.3	-57.32	-46.30	1.013	.1395	.000	7.265	7.265	.6593
33.0400	.504	700.3	-57.29	-43.57	1.389	.1382	.000	10.046	10.046	.7318
33.0500	.504	699.1	-57.33	-39.17	2.034	.1391	.000	14.616	14.616	*
33.0600	.504	701.1	-57.26	-24.01	4.099	.1382	.000	29.663	29.663	*
33.0700	.504	700.4	-57.26	8.21	8.222	.1387	.000	59.276	59.277	*
33.0800	.504	700.8	-57.22	40.96	12.519	.1378	.000	90.851	90.852	*
34.0100	.505	2000.8	-57.25	-48.11	1.391	.2444	.000	5.691	5.692	.6224
34.0200	.505	1998.4	-57.23	-39.19	2.712	.2421	.000	11.204	11.204	.6208
34.0300	.505	1998.2	-57.27	-22.94	5.370	.2477	.000	21.678	21.679	.6315
34.0400	.505	1999.1	-57.25	9.91	10.633	.2469	.000	43.067	43.067	.6413
35.0100	.505	1499.0	-57.28	-48.45	1.375	.2472	.000	5.564	5.564	.6306
35.0200	.505	1501.3	-57.28	-40.03	2.700	.2472	.000	10.921	10.921	.6333
35.0300	.505	1499.6	-57.25	-22.56	5.268	.2364	.000	22.286	22.287	.6424
35.0400	.505	1498.0	-57.41	7.49	11.394	.2558	.000	44.548	44.548	.6863
36.0100	.505	1001.1	-57.42	-43.99	2.134	.2454	.000	8.695	8.695	.6472
36.0200	.505	1000.1	-57.43	-30.12	4.404	.2458	.000	17.914	17.914	.6561
36.0300	.505	1000.2	-57.42	-5.62	8.725	.2495	.000	34.968	34.968	.6750
37.0100	.504	1698.6	131.04	144.64	2.896	.2206	-.002	13.127	13.125	.9649
37.0200	.504	1699.7	131.13	158.35	5.793	.2214	-.003	26.162	26.159	.9612
37.0300	.504	1698.1	131.21	171.99	8.146	.2071	-.003	39.326	39.323	.9642
38.0100	.504	1301.2	131.18	144.84	2.906	.1954	-.005	14.874	14.869	1.0882
38.0200	.504	1298.3	131.38	158.75	5.604	.1950	-.004	28.740	28.736	1.0500
38.0300	.504	1300.6	131.27	171.68	8.014	.1950	-.006	41.095	41.089	1.0166

TABLE LV

TABULATED EXPERIMENTAL ISOTHERMAL DATA FOR THE NOMINAL 51 PERCENT MIXTURE

RUN	MOLE FRACTION C_3H_8	INLET TEMPERATURE (°F)	INLET PRESSURE (psia)	PRESSURE DROP (psi)	POWER (Btu/min)	FLOW lb/min)	CORR. (Btu/lb)	ΔH_T (Btu/lb)	$(\Delta H/\Delta P)_T$ (Btu/lb-psi)
1.010	.505	152.2	2026.1	198.0	.805	.1796	.022	4.458	-.022519
1.020	.505	152.3	1839.0	207.0	1.115	.1744	.033	6.361	-.030732
1.030	.505	152.2	1642.3	202.8	1.396	.1610	.019	8.653	-.042670
1.040	.505	152.2	1460.7	210.4	1.811	.1506	.018	12.006	-.057069
1.050	.505	152.2	1253.7	218.2	2.094	.1341	-.027	15.648	-.071721
1.060	.505	152.2	1049.3	283.2	2.732	.1301	-.003	21.007	-.074184
1.070	.505	152.2	845.8	278.2	1.938	.1042	.011	18.574	-.066772
1.080	.505	152.2	655.5	420.0	2.116	.0906	-.003	23.358	-.055619
1.090	.505	152.2	560.9	460.2	1.960	.0807	-.002	24.281	-.052766
2.010	.507	251.3	1984.6	195.0	.852	.1503	.001	5.670	-.029078
2.020	.507	251.3	1796.8	196.6	.909	.1415	.003	6.419	-.032654
2.030	.507	251.3	1622.9	206.4	.993	.1364	.003	7.279	-.035268
2.040	.507	251.3	1450.8	212.2	1.020	.1295	-.011	7.889	-.037179
2.050	.507	251.3	1285.8	219.2	1.009	.1213	.011	8.308	-.037906
2.060	.507	251.3	1087.3	229.4	.983	.1114	-.004	8.826	-.038480
2.070	.507	251.3	878.6	231.4	.951	.1005	.009	9.455	-.040864
2.080	.507	251.3	666.2	299.4	.952	.0881	.007	10.792	-.036048
2.090	.507	251.2	594.5	491.6	1.475	.0857	.000	17.206	-.035003
3.010	.505	3.4	1996.5	188.2	.000	.1818	.083	-.083	.000442
3.020	.505	3.5	1813.8	198.0	.000	.1851	.035	-.035	.000179
3.030	.505	3.5	1633.3	207.6	.000	.1887	-.051	.051	-.000248
3.040	.505	3.5	1432.9	207.6	.027	.1875	.004	.141	-.000681
3.050	.505	3.5	1231.1	212.0	.062	.1891	.017	.313	-.001475
3.060	.505	3.5	1098.6	113.2	.028	.1323	.014	.199	-.001754
3.070	.505	3.5	1100.2	159.6	.243	.1597	-.019	1.543	-.009669
3.080	.505	3.5	1100.1	211.6	.758	.1833	-.014	4.151	-.019617
3.090	.505	3.5	1100.0	262.7	1.374	.2074	.008	6.617	-.025191
3.100	.505	3.6	1100.4	346.6	2.663	.2459	-.015	10.846	-.031295
3.110	.505	3.5	1099.7	434.6	4.263	.2800	-.005	15.226	-.035037
3.120	.505	3.6	698.5	199.2	1.062	.1049	-.030	10.153	-.050971

TABLE LVI

TABULATED EXPERIMENTAL JOULE-THOMSON DATA
FOR THE NOMINAL 51 PERCENT MIXTURE

Run	Mole Fraction Propane	Inlet Temperature (°F)	Inlet Pressure (psia)	Temperature Difference (°F)	Pressure Drop (psi)	$-\left(\frac{\Delta T}{\Delta P}\right)_H$ (°F/psi)x10^3
4.010	0.507	-149.0	2011.1	0.713	128.4	5.55
4.020	0.507	-149.1	1899.3	0.753	134.0	5.62
4.030	0.507	-149.0	1800.6	0.725	132.2	5.48
4.040	0.507	-149.0	1681.0	0.956	170.7	5.60
4.050	0.507	-149.0	1540.9	0.933	166.7	5.60
4.060	0.507	-149.0	1403.5	0.759	138.3	5.49
4.070	0.507	-149.1	1291.6	0.763	140.8	5.41
4.080	0.507	-149.0	1166.1	0.961	176.4	5.45
4.090	0.507	-149.0	1005.6	0.995	186.0	5.35
4.100	0.507	-149.0	827.2	1.050	196.6	5.35
4.110	0.507	-149.0	630.7	1.075	201.2	5.34
4.120	0.507	-149.0	433.8	0.998	191.2	5.22

TABLE LVII

FLOWMETER CALIBRATION EQUATION CONSTANTS
USED FOR THE NOMINAL 51 PERCENT MIXTURE

Flow Meter Series		B	A	C	D
10 - 20		0.10331343	20.513388	-4137.6569	962875.55
40,50,60	low	0.10390242	18.303396	-2433.9077	630217.66
	high	0.043650561	98.539793	-37002.100	5443353.1

APPENDIX C

COMPUTER PROGRAMS

TABLE LVIII

PROGRAM FOR FITTING ISOBARIC DATA

```
            INTEGER N,I,J,K,L,M
            DIMENSION CPBAR(200),TIN(200),TOUT(200),CPARR(200),
          1 EX((1...200)*(0...3)),X((1...8)*(1...3)),Y(8),B(3)
            BOOLEAN IPO
 BEGIN      READ DATA N,TO,TF,IPO
            TO=TO+460.
            TF=TF+460.
            THROUGH DO,FOR I=1,1,I.G.N
            READ FORMAT INPUT,CPBAR(I),TIN(I),TOUT(I)
            TIN(I)=TIN(I)+460.
            TOUT(I)=TOUT(I)+460.
 DO         EX(I,0)=TIN(I)+TOUT(I)
            VECTOR VALUES INPUT=$3E14.8*$
            THROUGH GO,FOR I=1,1,I.G.N
            K=1
            THROUGH ORDER,FOR J=2,1,J.G.N
            WHENEVER EX(K,0).G.EX(J,0)
            K=J
 ORDER      END OF CONDITIONAL
            EX(I,1)=EX(K,0)
            CPARR(I)=CPBAR(K)
            EX(I,2)=TIN(K).P.2+TIN(K)*TOUT(K)+TOUT(K).P.2
            EX(I,3)=TIN(K).P.3+(TIN(K).P.2)*TOUT(K)+TIN(K)*(TOUT(K).P.2)
          1 +TOUT(K).P.3
 GO         EX(K,0)=5000.
            WHENEVER IPO,PRINT RESULTS CPARR(1)...CPARR(N),EX(1,1)...
          1 EX(N,1)
            L=0
            M=1
            DELH=0.0
            EXECUTE ZERO.(B(0)...B(3))
            THROUGH ALL,FOR T=TO,10.,T.G.TF
            WHENEVER L.E.1, TRANSFER TO SKIP
            HO=0.0
            THROUGH TIL,FOR K=3,-1,K.L.0
 TIL        HO=(HO+B(K))*EX(M-1,1)/2.
            HO=HO+DELH
            THROUGH RANGE,FOR I=M,1,T.L.EX(I,1)/2. .OR.I.G.N
 RANGE      CONTINUE
            WHENEVER I.L.5,I=5
            WHENEVER (I+3).G.N,I=N-3
            K=1
            THROUGH HORSE, FOR J=I-4,1,J.G.I+3
            X(K,1)=EX(J,1)
            X(K,2)=EX(J,2)
            X(K,3)=EX(J,3)
            Y(K)=CPARR(J)
 HORSE      K=K+1
            S=REGR.(X,Y,A,B,BAD,IPO)
            B(0)=A
            HON=0.0
            THROUGH KIQ, FOR K=3,-1,K.L.0
 KIQ        HON=(HON+B(K))*EX(M-1,1)/2.
```

TABLE LVIII

(CONTINUED)

```
        DELH=HO-HON
SKIP    H=0.0
        THROUGH DIM,FOR K=3,-1,K.L.0
DIM     H=(H+B(K))*T
        H=H+DELH
        CP=B(0)+2.*B(1)*T+3.*B(2)*T.P.2+4.*B(3)*T.P.3
        TFAR=T-460.
        PRINT RESULTS TFAR,H,CP,S
        TRANSFER TO JUMP
BAD     PRINT RESULTS T
        PRINT COMMENT$1FOR THE ABOVE TEMPERATURE, COEFFICIENT MATRIX
      1 IS SINGULAR OR NEARLY SINGULAR$
JUMP    M=I+1
        L=0
ALL     WHENEVER T.L.(EX(I,1)/2.-10.).OR.(I+3).E.N,L=1
        TRANSFER TO BEGIN
        END OF PROGRAM
```

TABLE LVIII

(CONTINUED)

```
            EXTERNAL FUNCTION (N,A,X,EPS)
            NORMAL MODE IS INTEGER
            FLOATING POINT DETER,A,X,EPS,BIGA,AJCK
            DIMENSION IR(3),JC(3)
            ENTRY TO SIMUL.
            MAX=N+1
            DETER=1.0
            THROUGH L1, FOR K=1,1,K.G.N
            BIGA=0.0
            THROUGH L2,FOR I=1,1,I.G.N
            THROUGH L2,FOR   J=1,1,J.G.N
            THROUGH L3,FOR I1=1,1,I1.E.K
            THROUGH L3,FOR J1=1,1,J1.E.K
  L3        WHENEVER I.E.IR(I1).OR.J.E.JC(J1),TRANSFER TO L2
            WHENEVER.ABS.A(I,J).G.BIGA
            BIGA=.ABS.A(I,J)
            IR(K)=I
            JC(K)=J
  L2        END OF CONDITIONAL
            WHENEVER BIGA.L.EPS, FUNCTION RETURN 0.
            BIGA=A(IR(K),JC(K))
            DETER=DETER*BIGA
            THROUGH L4,FOR J=1,1,J.G.MAX
   L4       A(IR(K),J)=A(IR(K),J)/BIGA
            THROUGH L1, FOR I=1,1,I.G.N
            AJCK=A(I,JC(K))
            WHENEVER I.NE.IR(K)
            THROUGH L5, FOR J=1,1,J.G.MAX
L5          WHENEVER J.NE.JC(K),A(I,J)=A(I,J)-AJCK*A(IR(K),J)
L1          END OF CONDITIONAL
            THROUGH L7, FOR I=1,1,I.G.N
L7          X(JC(I))=A(IR(I),MAX)
            FUNCTION RETURN DETER
            END OF FUNCTION
```

TABLE LVIII

(CONTINUED)

```
          EXTERNAL FUNCTION(X,Y,A,B,BAD,BOOL)
          STATEMENT LABEL BAD
          BOOLEAN BOOL
          INTEGER I,J,L
          DIMENSION SX(3),SYX(3),CYX(3),C((1...3)*(1...4))
          ENTRY TO REGR.
          EXECUTE ZERO.(SY,SYY,SX(1)...SX(3),SYX(1)...SYX(3))
          THROUGH OWN,FOR I=1,1,I.G.8
          SY=SY+Y(I)
          SYY=SYY+Y(I).P.2
          THROUGH OWN,FOR J=1,1,J.G.3
          SX(J)=SX(J)+X(I,J)
 OWN      SYX(J)=SYX(J)+X(I,J)*Y(I)
          THROUGH COEF,FOR J=1,1,J.G.3
          THROUGH COEF,FOR I=1,1,I.G.3
          C(I,J)=-SX(I)*SX(J)/8.0
          THROUGH COEF, FOR L=1,1,L.G.8
COEF      C(I,J)=C(I,J)+X(L,I)*X(L,J)
          THROUGH LIP, FOR I=1,1,I.G.3
          CYX(I)=-SX(I)*SY/8.0
          THROUGH BID,FOR J=1,1,J.G.8
 BID      CYX(I)=CYX(I)+X(J,I)*Y(J)
LIP       C(I,4)=CYX(I)
          WHENEVER BOOL,PRINT RESULTS C(1,1)...C(3,4)
          CYY=SYY-((SY).P.2/8.0)
          DET=SIMUL.(3,C,B,1.0E-20)
          WHENEVER DET.E.0.0, TRANSFER TO BAD
          A=SY/8.0
          TEMP=CYY
          THROUGH GIN,FOR I=1,1,I.G.3
          A=A-B(I)*SX(I)/8.0
GIN       TEMP=TEMP-B(I)*CYX(I)
          S=SQRT.(.ABS.(TEMP/4.0))
          WHENEVER BOOL, PRINT RESULTS A,B(1)...B(3), DET
          FUNCTION RETURN S
          END OF FUNCTION
```

TABLE LIX

LIST OF VARIABLES FOR ISOBARIC DATA FITTING PROGRAM

Main Program

Program Symbol	Definition
A, B(I)	Regression coefficients
CP	C_p
CPARR	Array of ordered values of C_{pm}
CPBAR	Array of C_{pm} variables
DELH	Enthalpy base correction
EX	Data input storage array
H	$\underline{H}$
HO	Original enthalpy base value
HON	Enthalpy base value for a new set of constants
I, J, K	Counter variables
IPO	Boolean conditional variable
L, M	Defined in program
N	Number of input data points
REGR	Regression subroutine name
S	Standard deviation
SIMUL	Simultaneous equation subroutine name
T	$T(^oF)$
TO	Lower temperature interval limit
TF	Upper temperature interval limit
TFAR	$T(^oF)$

TABLE LIX CONTINUED

TIN	Inlet temperature of data point
TOUT	Outlet temperature of data point
X	Array containing regression independent variables, x_i
Y	Array containing dependent variable, y

Subroutine REGR*

BAD	Error return dummy statement label
BOOL	Boolean conditional variable
C	Coefficient matrix, C_{ij}
CYX	C_{iy}
CYY	C_{yy}
DET	Determinant of matrix C_{ij}
I, J, L	Counter variables
SX(I)	$\sum x_i$
SYX(I)	$\sum x_{i.y}$
SY, SYY	$\sum y, \sum y^2$
TEMP	Temporary iteration variable

Subroutine SIMUL*

A	Coefficient matrix, A_{ij}
AJCK	$A_{i,j_{ck}}$
BIGA	Pivot element, A_{IR_k, JC_k}
DETER	Value of determinant of A_{ij}

TABLE LIX CONTINUED

EPS	Tolerance value,
I	Row index, i
I1	Index on the array IR during pivot element search
IR	Array containing row subscripts of pivot elements, in order
J	Column index, j
J1	Index of the array JC during element search
JC	Array containing column subscripts of pivot elements, in order
MAX	Number of columns in the matrix A_{ij}
N	Number of equations
TEMP	Temporary location used in ordering CJ array
X	Vector containing the ordered solution values

* Subroutines are explained more fully in Carnahan et al.[16] from which they were obtained with minor modifications.

TABLE LX

PROGRAM FOR OPTIMIZATION OF PARAMETERS IN CORRESPONDING STATES CORRELATION

```
FORTRAN IV G COMPILER       MAIN          10-09-68         12:36.37        PAGE 0001

0001              DIMENSION TABLE(11,11),
                 1H1(30,40),T1(40),P1(30),HLIQ1(20),HGAS1(20),KEQ1(20),PLIQ1(20),
                 2H2(20,79),T2(79),P2(20),F(7),PMIX(100),TMIX(100),HMIX(100),
                 3HOMIX(100),HEXP(100),HMIXC(100,7)
0002              COMMON TABLE,H1,T1,P1,H2,T2,P2,PMIX,TMIX,HMIX,HOMIX,HEXP,HMIXC,
                 1DELH,M1,N1,M2,N2,F
0003              NAMELIST/NL/PEC,TEC,WGA,IND,F
0004                M2=20
0005                N2=79
0006              READ(8) (P2(J),J=1,10)
0007              READ(8)(P2(J),J=11,20)
0008              READ(8)(T2(J),J=1,16)
0009              READ(8)(T2(J),J=17,32)
0010              READ(8)(T2(J),J=33,48)
0011              READ(8)(T2(J),J=49,64)
0012              READ(8)(T2(J),J=65,79)
0013              DO 122 J=1,79
0014              READ(8)(H2(I,J),I=1,10)
0015        122   READ(8)(H2(I,J),I=11,20)
0016              READ(8)(P1(J),J=1,19)
0017              READ(8)(P1(J),J=20,22)
0018              READ(8)(T1(J),J=1,19)
0019                READ(8)(T1(J),J=20,38)
0020              READ(8)(HLIQ1(J),J=1,19)
0021              READ(8)(HGAS1(J),J=1,19)
0022              READ(8)(KEQ1(J),J=1,19)
0023              READ(8)(PLIQ1(J),J=1,19)
0024              READ(8)M1,N1,LIQVA1
0025              DO 121 J=1,38
0026              READ(8)(H1(I,J),I=1,19)
0027        121   READ(8)(H1(I,J),I=20,22)
0028          1   READ(5,110)PC,TC,W,WM,DPC,DTC,DW,N,INDEX,NPR
0029              WRITE(6,110)PC,TC,W,WM,DPC,DTC,DW,N,INDEX,NPR
0030        110   FORMAT(2F6.1,F6.4,F7.3,2F6.1,F6.4,I4,I4,I3)
0031              DO 202 I=1,N
0032        201   READ (5,111)PMIX(I),TMIX(I),HMIX(I),HOMIX(I)
0033        202   HEXP(I)=HMIX(I)-HOMIX(I)
0034        111    FORMAT (F6.0,3F6.1)
0035              PEC=PC
0036              TEC=TC
0037              WGA=W
0038              DR=DPC/200.
0039              IND=1
0040              CALL DEV(PEC,TEC,WGA,WM,N,1)
0041              IF(INDEX-1)52,3,52
0042          3   DO 2 I=2,7
0043          2   F(I)=F(1)
0044              GO TO 50
0045         52   PEC=PEC+DPC
0046              CALL DEV(PEC,TEC,WGA,WM,N,2)
0047              PEC=PEC-DPC
0048              TEC=TEC+DTC
0049              CALL DEV(PEC,TEC,WGA,WM,N,3)
0050              TEC=TEC-DTC
0051              WGA=WGA+DW
```

TABLE LX

(CONTINUED)


```
          CALL DEV(PEC,TEC,WGA,WM,N,4)
          WGA=WGA-DW
          PEC=PEC-DPC
          CALL DEV(PEC,TEC,WGA,WM,N,5)
          PEC=PEC+DPC
          TEC=TEC-DTC
          CALL DEV(PEC,TEC,WGA,WM,N,6)
          TEC=TEC+DTC
          WGA=WGA-DW
          CALL DEV(PEC,TEC,WGA,WM,N,7)
          WGA=WGA+DW
          II=1
          DO 11 I=2,7
     11   IF(F(I).LT.F(II))II=I
          IF(II.EQ.1) GO TO 50
          IF(II.EQ.2) PEC=PEC+DPC
          IF(II.EQ.3) TEC=TEC+DTC
          IF(II.EQ.4) WGA=WGA+DW
          IF(II.EQ.5) PEC=PEC-DPC
          IF(II.EQ.6) TEC=TEC-DTC
          IF(II.EQ.7) WGA=WGA-DW
          IF (NPR.GT.0) WRITE(6,NL)
          F(1)=F(II)
          DO 53 I=1,N
     53   HMIXC(I,1)=HMIXC(I,II)
          IND=IND+1
          IF (IND.GT.INDEX) GO TO 50
          GO TO 52
     50   WRITE(6,NL)
          WRITE(6,120)
    120   FORMAT(30X, ' OPTIMUM CALCULATED ENTHALPY DEPARTURES'/'        PRES
         1SURE          TEMPERATURE          EXPER. DEPARTURE        CALC. DEPARTU
         2RE         DIFFERENCE         PERCENT DIFFERENCE'/)
          DO 54 I=1,N
          DELH=HEXP(I)-HMIXC(I,1)
          DELHP=(HEXP(I)-HMIXC(I,1))/HEXP(I)*100.
     54   WRITE(6,123)PMIX(I),TMIX(I),HEXP(I),HMIXC(I,1),DELH,DELHP
    123   FORMAT(1H ,6F20.7)
          GO TO 1
          END
```

TABLE LX

(CONTINUED)

```
FORTRAN IV G COMPILER          DEV              08-03-68           20:25.26         PAGE 0001

0001                SUBROUTINE DEV(PEC,TEC,W,WM,N,EF)
0002                INTEGER EF
0003                DIMENSIONTABLE(11,11),PTAB(11),HVEC(11),HTAB(4,4),HOUT(11),
                   1TTAB(11),F(7),PMIX(100),TMIX(100),HMIX(100),HOMIX(100),HEXP(100),
                   2HMIXC(100,7),H1(30,40),T1(40),P1(30),H2(20,79),T2(79),P2(20)
0004                COMMON TABLE,H1,T1,P1,H2,T2,P2,PMIX,TMIX,HMIX,HOMIX,HEXP,HMIXC,
                   1DELH,M1,N1,M2,N2,F
0005                F(EF)=0.0
0006                DO 416 L=1,N
0007                PR=PMIX(L)/PEC
0008                TR=TMIX(L)/TEC
0009                DO1 I=1,M1,1
0010                IY=I
0011                IF(PR-P1(I))3,1,1
0012          1     CONTINUE
0013          3     DO 5 J=1,N1,1
0014                JAY=J
0015                IF(TR-T1(J))401,5,5
0016          5     CONTINUE
0017        401     IF(JAY.GT.N1-1)JAY=N1-1
0018                IF(IY.GT.M1-1)IY=M1-1
0019                IF(JAY.LT.3)JAY=3
0020                CALL INT4(H1,T1,P1,HTAB,TTAB,PTAB,IY,JAY,30,40)
0021                DO 30 J=1,4
0022                DO 21 I=1,4
0023         21     HVEC(I)=HTAB(I,J)
0024                 CALL DTABLE(PTAB,HVEC,TABLE,3,3,&15)
0025         30        CALL NEWTON(PTAB,PR,TABLE,3,3,HOUT(J),3,&15)
0026                 CALL DTABLE(TTAB,HOUT,TABLE,3,3,&15)
0027                 CALL NEWTON(TTAB,TR,TABLE,3,3,HINT1,3,&15)
0028                DO 10 I=1,M2,1
0029                IY=I
0030                IF(PR-P2(I))11,10,10
0031         10      CONTINUE
0032         11     DO12 J=1,N2,1
0033                JAY=J
0034                IF(TR-T2(J))402,12,12
0035         12     CONTINUE
0036        402     IF(JAY.GT.N2-1)JAY=N2-1
0037                IF(JAY.LT.3)JAY=3
0038                IF(IY.GT.M2-1)IY=M2-1
0039         13     CALL INT4(H2,T2,P2,HTAB,TTAB,PTAB,IY,JAY,M2,N2)
0040                DO 22 J=1,4
0041                DO 23 I=1,4
0042         23     HVEC(I)=HTAB(I,J)
0043                 CALL DTABLE(PTAB,HVEC,TABLE,3,3,&15)
0044         22        CALL NEWTON(PTAB,PR,TABLE,3,3,HOUT(J),3,&15)
0045                 CALL DTABLE(TTAB,HOUT,TABLE,3,3,&15)
0046                 CALL NEWTON(TTAB,TR,TABLE,3,3,HINT2,3,&15)
0047                R=1.987/WM
0048                HMIXC(L,EF)=((1.-W)*HINT1+W*HINT2)*R*TEC
0049                DELH=HEXP(L)-HMIXC(L,EF)
0050                F(EF)=F(EF)+DELH**2
0051        416     CONTINUE
0052                EN=N
```

TABLE LX

(CONTINUED)

```
FORTRAN IV G COMPILER        DEV          07-27-68        13:40.12       PAGE 0002

0053                  F(EF)=SQRT(F(EF)/EN)
0054            15    RETURN
0055                  END
```

```
FORTRAN IV G COMPILER        INT4         07-16-68        22:38.32       PAGE 0001

0001                  SUBROUTINE INT4(H,T,P,HTAB,TTAB,PTAB,IY,JAY,M,N)
0002                  DIMENSION H(M,N),T(N),P(M),HTAB(4,4),TTAB(11),PTAB(11)
0003                  ILOW=IY-3
0004                  JLOW=JAY-3
0005                  DO 1 J=1,4
0006             1    TTAB(J)=T(JLOW+J)
0007                  DO 2 I=1,4
0008             2    PTAB(I)=P(ILOW+I)
0009                  DO 3 J=1,4
0010                  DO 3 I=1,4
0011             3    HTAB(I,J)=H(ILOW+I,JLOW+J)
0012                  RETURN
0013                  END
```

TABLE LX

(CONTINUED)

FORTRAN IV G COMPILER DTABLE 07-16-68 22:38.16 PAGE 0001

```
      SUBROUTINE DTABLE(X,Y,TABLE,N,M,*)
      INTEGER DEGREE
      DIMENSION X(11),Y(11),TABLE(11,11)
      IF(M-N)1,1,2
    2 WRITE(6,3)
    3 FORMAT('       SCREWED UP      ')
      RETURN 1
    1 DO 4 J=1,M
      DO 4 I=J,N
      IF(J-1)5,6,5
    6 TABLE(I,1)=(Y(I+1)-Y(I))/(X(I+1)-X(I))
      GO TO 4
    5 TABLE(I,J)=(TABLE(I,J-1)-TABLE(I-1,J-1))/(X(I+1)-X(I-J+1))
    4 CONTINUE
      RETURN
       ENTRY NEWTON(X,XARG,TABLE,N,M,YEST,DEGREE,*)
      IF(DEGREE-M)7,7,8
    8 WRITE(6,9)
    9 FORMAT('    SCREWED UPPP    ')
      RETURN 1
    7 L=N+1
      DO 10 I=1,L
      K=I
      IF (XARG.LE.X(K)) GO TO 11
   10 CONTINUE
   11 MAX=K+DEGREE/2
      IF  (MAX.LT.DEGREE+1)MAX=DEGREE+1
       IF(MAX.GT.N+1)MAX=N+1
      YEST=TABLE(MAX-1,DEGREE)
      L=DEGREE-1
      IF(L.EQ.0) GO TO 13
      DO 12 I=1,L
   12 YEST=YEST*(XARG-X(MAX-I))+TABLE(MAX-I-1,DEGREE-I)
   13 YEST=YEST*(XARG-X(MAX-DEGREE))+Y(MAX-DEGREE)
      RETURN
      END
```

TABLE LXI

LIST OF VARIABLES FOR CORRESPONDING STATES CALCULATION AND PARAMETER OPTIMIZATION PROGRAM

Main Program

Program Symbol	Definition
DELH	Difference between experimental and calculated enthalpy
DELHP	Percent difference between experimental and calculated enthalpy
DEV	Enthalpy departure calculation and comparison subroutine name
DPC, DTC, DW	Step size of parameters in optimization procedure
F	Root mean square deviation between experimental and calculated enthalpy departures
HOMIX	Ideal gas enthalpy
H1	Reduced enthalpy departure for first reference substance
H2	Reduced enthalpy departure for second reference substance
HEXP	Experimental enthalpy
HMIX	Experimental enthalpy departure
HMIXC	Calculated enthalpy departure
I, J	Counter variables
II	Variable indicating direction of maximum reduction in standard deviation
IND	Counter variable of search iterations
INDEX	Maximum number of search iterations
M1	Number of reduced pressure values for first reference substance table

TABLE LXI CONTINUED

M2	Number of reduced pressure values for second reference substance table
N1	Number of reduced temperature values for first reference substance table
N2	Number of reduced temperature values for second reference substance table
NPR	Conditional print index
P1	Reduced pressure of first reference substance
P2	Reduced pressure of second reference substance
PC	Initial critical pressure
PEC	Critical pressure
PMIX	Mixture pressure
T1	Reduced temperature of first reference substance
T2	Reduced temperature of second reference substance
TABLE	Storage matrix for difference table
TC	Initial critical temperature
TEC	Critical temperature
TMIX	Mixture temperature
W	Initial third parameter
WGA	Third parameter
WM	Molecular Weight of mixture

Subroutines DEV and INT4

DTABLE	Difference table calculation subroutine name
EN	Number of data points
EF	Subscript of F

TABLE LXI CONTINUED

HINT1	Final interpolated enthalpy using first reference table
HINT2	Final interpolated enthalpy using second reference table
HOUT	Vector containing enthalpies interpolated with respect to pressure
HTAB	Matrix containing reduced enthalpies upon which interpolation is based
HVEC	Vector containing enthalpies from HTAB at the same reduced temperature
I, J, L	Counter variables
ILOW, JLOW	Defined in program
IY, JAY	Defined in program
NEWTON	Interpolation subroutine name
PR	Reduced pressure of mixture
PTAB	Vector containing reduced pressures corresponding to HTAB values of reduced enthalpy
R	Gas constant
TR	Reduced temperature
TTAB	Vector containing reduced temperatures corresponding to HTAB values of reduced enthalpy

Subroutines DTABLE* and NEWTON*

DEGREE	Degree of desired interpolating polynomial
I, J	Counter variables
K, L	Defined in program
M	Maximum order of divided differences to be calculated by DTABLE

TABLE LXI CONTINUED

MAX	Subscript of largest X value used in constructing the interpolating polynomial
N	Maximum subscript on X and Y
TABLE	Matrix containing divided difference table
X	An array containing abscissa values in ascending order
XARG	Interpolation argument
Y	An array containing ordinate values
YEST	Variable used to hold partially computed value of interpolant

* Subroutines are more fully explained in Carnahan et al.[16] from where they were obtained with minor modification.

APPENDIX D

DATA AND RESULTS OF CORRESPONDING STATES CORRELATION

TABLE LXII

REDUCED ENTHALPY AS A FUNCTION OF REDUCED TEMPERATURE AND REDUCED PRESSURE FOR PROPANE

TR \ PR	0.0	0.16198	0.32395	0.40494	0.48593	0.56691	0.64790	0.72889	0.80988	0.89086	0.97185	0.99939	1.13382	1.29580	1.45778	1.61975	2.02469	2.42963	2.83456	3.23950
0.26969	0.0	-7.84080	-7.83747	-7.83414	-7.83080	-7.82747	-7.82414	-7.82081	-7.81747	-7.81414	-7.81081	-7.80748	-7.79748	-7.78748	-7.78082	-7.76083	-7.73084	-7.71084	-7.69751	-7.69085
0.28471	0.0	-7.76749	-7.76416	-7.76083	-7.76083	-7.76083	-7.75749	-7.75749	-7.75416	-7.74750	-7.74416	-7.74416	-7.72750	-7.71751	-7.70418	-7.69085	-7.66086	-7.63420	-7.62087	-7.61754
0.29972	0.0	-7.70085	-7.69418	-7.69085	-7.68752	-7.69085	-7.68752	-7.68418	-7.67752	-7.67419	-7.67085	-7.66752	-7.66086	-7.64753	-7.63087	-7.62087	-7.59088	-7.56756	-7.55423	-7.54756
0.31474	0.0	-7.62754	-7.62420	-7.62087	-7.61754	-7.61754	-7.61421	-7.61087	-7.61087	-7.60421	-7.59755	-7.59421	-7.59088	-7.57755	-7.56422	-7.55089	-7.52090	-7.49425	-7.47758	-7.47425
0.32975	0.0	-7.56089	-7.55756	-7.55089	-7.54756	-7.54423	-7.54090	-7.53757	-7.53090	-7.52757	-7.52090	-7.51757	-7.51091	-7.49425	-7.48425	-7.47425	-7.44426	-7.42427	-7.40761	-7.40094
0.34477	0.0	-7.49091	-7.48758	-7.48425	-7.48092	-7.48092	-7.47758	-7.47092	-7.46759	-7.46426	-7.45426	-7.45093	-7.44093	-7.42760	-7.41427	-7.40427	-7.37095	-7.34763	-7.34096	-7.33430
0.35979	0.0	-7.41427	-7.41094	-7.40761	-7.40427	-7.40094	-7.39761	-7.39428	-7.39095	-7.38428	-7.38095	-7.38095	-7.36762	-7.35762	-7.34763	-7.33430	-7.30097	-7.28098	-7.26432	-7.26099
0.37480	0.0	-7.33763	-7.33430	-7.33430	-7.33097	-7.33430	-7.33097	-7.32763	-7.32097	-7.31764	-7.31097	-7.31097	-7.30097	-7.28431	-7.27098	-7.25766	-7.23100	-7.20434	-7.19434	-7.18768
0.38982	0.0	-7.27098	-7.26765	-7.26765	-7.26432	-7.26099	-7.25766	-7.25432	-7.25432	-7.24766	-7.23766	-7.23766	-7.23100	-7.21767	-7.20434	-7.19434	-7.16102	-7.14103	-7.12103	-7.11437
0.40483	0.0	-7.20100	-7.19767	-7.19434	-7.19100	-7.19100	-7.18434	-7.18101	-7.17434	-7.17434	-7.16435	-7.16101	-7.15435	-7.14102	-7.13102	-7.11769	-7.08437	-7.06438	-7.04772	-7.04105
0.41985	0.0	-7.13436	-7.13103	-7.12770	-7.12436	-7.12103	-7.11437	-7.11104	-7.10770	-7.10104	-7.09771	-7.09438	-7.08438	-7.07771	-7.06438	-7.05106	-7.01773	-6.99774	-6.98108	-6.97108
0.43487	0.0	-7.06438	-7.06105	-7.05772	-7.05439	-7.05439	-7.05105	-7.04772	-7.04106	-7.03772	-7.03106	-7.02773	-7.01773	-7.00440	-6.99441	-6.98441	-6.94775	-6.92443	-6.91110	-6.90110
0.44988	0.0	-6.98774	-6.98441	-6.98108	-6.97774	-6.97774	-6.97441	-6.97108	-6.96441	-6.96108	-6.95442	-6.95109	-6.94442	-6.93109	-6.92109	-6.91110	-6.87111	-6.85112	-6.83446	-6.82779
0.46490	0.0	-6.92110	-6.91776	-6.91776	-6.91443	-6.91443	-6.91110	-6.90443	-6.90110	-6.89111	-6.88777	-6.88777	-6.87444	-6.86778	-6.85445	-6.84112	-6.80447	-6.78781	-6.76781	-6.75448
0.47992	0.0	-6.84779	-6.84446	-6.84446	-6.84112	-6.83779	-6.83446	-6.83113	-6.82780	-6.82446	-6.81780	-6.81447	-6.80447	-6.79447	-6.78114	-6.76781	-6.73782	-6.71783	-6.69784	-6.68784
0.49493	0.0	-6.77780	-6.77447	-6.77114	-6.76781	-6.76447	-6.76114	-6.75781	-6.75448	-6.74781	-6.74448	-6.74115	-6.73115	-6.72116	-6.70783	-6.69783	-6.67117	-6.64785	-6.63118	-6.61786
0.50995	0.0	-6.70783	-6.70450	-6.70117	-6.69783	-6.69450	-6.69117	-6.69117	-6.68784	-6.68784	-6.68117	-6.68117	-6.66784	-6.65452	-6.64452	-6.63452	-6.60120	-6.57787	-6.55788	-6.54788
0.52496	0.0	-6.63785	-6.63452	-6.63452	-6.63119	-6.62785	-6.62452	-6.62452	-6.62119	-6.61786	-6.61786	-6.61452	-6.60120	-6.59120	-6.57787	-6.56787	-6.53455	-6.50456	-6.48790	-6.48457
0.53998	0.0	-6.56787	-6.56121	-6.56121	-6.55788	-6.55788	-6.55454	-6.55121	-6.55121	-6.55121	-6.54788	-6.54455	-6.53455	-6.52455	-6.51122	-6.50123	-6.46791	-6.44125	-6.42459	-6.41792
0.55500	0.0	-6.50123	-6.49456	-6.49456	-6.49123	-6.49123	-6.48790	-6.48790	-6.48457	-6.48457	-6.48123	-6.47790	-6.46791	-6.45458	-6.44458	-6.43458	-6.40126	-6.37460	-6.35794	-6.34461
0.57001	0.0	-6.43792	-6.43459	-6.43459	-6.43125	-6.43125	-6.42792	-6.42459	-6.42126	-6.41792	-6.41126	-6.40793	-6.40126	-6.38793	-6.37794	-6.36794	-6.33129	-6.30463	-6.29130	-6.27797
0.58503	0.0	-6.36460	-6.36127	-6.35794	-6.35794	-6.35794	-6.35794	-6.35794	-6.35461	-6.34794	-6.34128	-6.34128	-6.33128	-6.32129	-6.31129	-6.30129	-6.26797	-6.24131	-6.22132	-6.20799
0.60004	0.0	-6.30463	-6.30130	-6.29797	-6.29797	-6.29797	-6.29463	-6.29130	-6.28464	-6.28130	-6.27797	-6.27464	-6.26464	-6.25131	-6.24132	-6.23132	-6.19800	-6.17467	-6.15468	-6.14135
0.61506	0.0	-6.23132	-6.23132	-6.22799	-6.22465	-6.22799	-6.22465	-6.22132	-6.21799	-6.21133	-6.20466	-6.20466	-6.19466	-6.18467	-6.17467	-6.16134	-6.12469	-6.10803	-6.09136	-6.07470
0.63008	0.0	-6.16467	-6.16467	-6.16467	-6.16467	-6.15801	-6.15468	-6.15468	-6.14801	-6.14468	-6.14135	-6.13802	-6.13135	-6.11802	-6.10803	-6.09803	-6.06138	-6.04138	-6.02472	-6.01472
0.64509	0.0	-6.09136	-6.09136	-6.09136	-6.09136	-6.08803	-6.08803	-6.08469	-6.07803	-6.07470	-6.07470	-6.07137	-6.06137	-6.05137	-6.04138	-6.02805	-5.99806	-5.97473	-5.95807	-5.94141
0.66011	0.0	-6.02139	-6.02139	-6.02139	-6.02139	-6.01805	-6.01472	-6.01472	-6.01472	-6.00806	-6.00473	-6.00473	-5.99473	-5.98473	-5.97474	-5.96141	-5.93475	-5.91475	-5.89476	-5.88143
0.67513	0.0	-5.95474	-5.95141	-5.95141	-5.95141	-5.94807	-5.94807	-5.94807	-5.94474	-5.94141	-5.93808	-5.93474	-5.92475	-5.91808	-5.90475	-5.89143	-5.86810	-5.84811	-5.82811	-5.81478
0.69014	0.0	-5.89476	-5.88809	-5.88476	-5.88476	-5.88143	-5.88143	-5.87810	-5.87476	-5.87143	-5.86810	-5.86810	-5.85810	-5.84477	-5.83478	-5.82811	-5.80145	-5.78146	-5.76813	-5.75147
0.70516	0.0	-5.81812	-5.81479	-5.81145	-5.81145	-5.81145	-5.81145	-5.81145	-5.80812	-5.80479	-5.80146	-5.79813	-5.78813	-5.77813	-5.76814	-5.75814	-5.73481	-5.71815	-5.70482	-5.68816
0.72017	0.0	-5.76147	-5.75813	-5.75480	-5.75147	-5.74814	-5.74814	-5.74480	-5.73814	-5.73481	-5.73148	-5.72814	-5.71815	-5.70815	-5.69482	-5.69149	-5.67149	-5.65817	-5.64150	-5.62151
0.73519	0.0	-5.68150	-5.68150	-5.68150	-5.67817	-5.67483	-5.67150	-5.66817	-5.66484	-5.66150	-5.65484	-5.65484	-5.64484	-5.63485	-5.62818	-5.62152	-5.60152	-5.58486	-5.56487	-5.55154
0.75021	0.0	-5.60485	-5.60485	-5.60485	-5.60152	-5.59819	-5.59819	-5.59819	-5.59486	-5.59153	-5.58819	-5.58819	-5.57486	-5.56820	-5.55820	-5.54821	-5.52488	-5.51155	-5.49156	-5.48156
0.76522	0.0	-5.51488	-5.52155	-5.52488	-5.53155	-5.53488	-5.53488	-5.53821	-5.53488	-5.53488	-5.53155	-5.52821	-5.51822	-5.50489	-5.49489	-5.48156	-5.45490	-5.43824	-5.42491	-5.41825
0.78024	0.0	-0.35655	-5.44490	-5.44823	-5.44823	-5.44823	-5.44823	-5.44823	-5.44490	-5.44157	-5.43824	-5.43490	-5.42824	-5.42158	-5.41158	-5.40491	-5.37826	-5.36826	-5.36160	-5.35160
0.79525	0.0	-0.30990	-5.35493	-5.36160	-5.36493	-5.36826	-5.36826	-5.36826	-5.36826	-5.36826	-5.36493	-5.36493	-5.35827	-5.35160	-5.34161	-5.33161	-5.30828	-5.29829	-5.29829	-5.28829
0.81027	0.0	-0.28990	-5.27829	-5.28496	-5.28829	-5.29496	-5.29829	-5.29829	-5.29829	-5.29496	-5.29496	-5.29162	-5.28496	-5.27496	-5.26830	-5.25830	-5.24164	-5.22831	-5.22164	-5.22164
0.82529	0.0	-0.26992	-5.17166	-5.19165	-5.20165	-5.20832	-5.21498	-5.21498	-5.21498	-5.21165	-5.20832	-5.20832	-5.20165	-5.19832	-5.18832	-5.18499	-5.17499	-5.15833	-5.14500	-5.14500
0.84030	0.0	-0.25658	-5.09835	-5.10834	-5.11834	-5.12500	-5.12834	-5.13167	-5.13500	-5.13500	-5.13500	-5.13500	-5.12834	-5.12167	-5.11168	-5.10834	-5.09835	-5.08835	-5.08169	-5.07835
0.85532	0.0	-0.23993	-0.59981	-5.00838	-5.01838	-5.02838	-5.03837	-5.04504	-5.05170	-5.05503	-5.05503	-5.05503	-5.04837	-5.04504	-5.03504	-5.03171	-5.03171	-5.01505	-5.01505	-5.01505
0.87034	0.0	-0.23992	-0.55649	-4.90508	-4.91508	-4.92507	-4.93840	-4.94507	-4.95506	-4.95506	-4.95840	-4.95840	-4.95506	-4.95173	-4.95173	-4.94840	-4.94507	-4.93840	-4.93507	-4.93840
0.88535	0.0	-0.22660	-0.52650	-0.73977	-4.79512	-4.81511	-4.83177	-4.84510	-4.85510	-4.86176	-4.86510	-4.86843	-4.87176	-4.87176	-4.86843	-4.86843	-4.86510	-4.86510	-4.86510	-4.87176
0.90037	0.0	-0.21992	-0.50650	-0.68978	-0.88638	-4.68181	-4.71180	-4.73513	-4.74846	-4.75512	-4.76179	-4.76179	-4.76845	-4.77512	-4.77845	-4.77512	-4.77512	-4.78178	-4.78845	-4.79511
0.91538	0.0	-0.21993	-0.48984	-0.65645	-0.84306	-4.52187	-4.57185	-4.60851	-4.62850	-4.63850	-4.64850	-4.65183	-4.66183	-4.67515	-4.68515	-4.68848	-4.69848	-4.70514	-4.71514	-4.72514
0.93040	0.0	-0.21660	-0.46319	-0.61647	-0.78642	-0.98968	-4.37525	-4.45856	-4.49854	-4.51187	-4.52854	-4.53187	-4.54853	-4.56519	-4.58518	-4.59185	-4.62184	-4.63183	-4.64516	-4.65849
0.94542	0.0	-0.20327	-0.43319	-0.57315	-0.74643	-0.93969	-1.16963	-4.27529	-4.34859	-4.38192	-4.40524	-4.41191	-4.44190	-4.46522	-4.47855	-4.48855	-4.51521	-4.54186	-4.55853	-4.57519
0.96043	0.0	-0.19994	-0.42986	-0.55981	-0.70644	-0.87638	-1.08631	-1.36955	-4.16532	-4.23197	-4.27195	-4.28195	-4.31527	-4.34526	-4.37192	-4.39191	-4.42523	-4.45522	-4.47521	-4.49521
0.97545	0.0	-0.19661	-0.41320	-0.52983	-0.66312	-0.81640	-1.00301	-1.23961	-1.56283	-3.98205	-4.04203	-4.06203	-4.13533	-4.19532	-4.23197	-4.26529	-4.33193	-4.37525	-4.40191	-4.41858
0.99046	0.0	-0.19327	-0.40321	-0.50984	-0.63313	-0.76975	-0.92970	-1.12964	-1.39288	-1.75943	-3.74546	-3.78211	-3.92874	-4.03537	-4.10201	-4.15199	-4.23863	-4.28528	-4.31527	-4.34193
1.00548	0.0	-0.18327	-0.37988	-0.48985	-0.60980	-0.74309	-0.89305	-1.06632	-1.27626	-1.54617	-1.93938	-2.10599	-3.61217	-3.81211	-3.92874	-4.01204	-4.12867	-4.18532	-4.22530	-4.26196
1.02050	0.0	-0.17660	-0.36655	-0.46984	-0.58314	-0.70644	-0.84305	-0.99634	-1.18628	-1.40287	-1.68279	-1.78608	-2.73578	-3.50553	-3.72879	-3.85542	-4.01203	-4.08534	-4.13200	-4.17531
1.03551	0.0	-0.16661	-0.34989	-0.44985	-0.56316	-0.67979	-0.81308	-0.95303	-1.11964	-1.30624	-1.51951	-1.59616	-2.08932	-3.14565	-3.50553	-3.67882	-3.89874	-3.97872	-4.04203	-4.09535
1.05053	0.0	-0.16661	-0.33989	-0.43652	-0.54316	-0.64979	-0.76975	-0.89637	-1.04300	-1.20627	-1.39288	-1.45286	-1.85606	-2.51918	-3.20563	-3.46221	-3.74545	-3.86542	-3.94872	-3.99871
1.06555	0.0	-0.15995	-0.32656	-0.41653	-0.51983	-0.62647	-0.73642	-0.85972	-0.99301	-1.13297	-1.29624	-1.35289	-1.67279	-2.18929	-2.79576	-3.20896	-3.59217	-3.75212	-3.84209	-3.90874
1.08056	0.0	-0.15661	-0.31656	-0.40653	-0.49651	-0.59648	-0.70644	-0.81974	-0.94637	-1.08965	-1.23960	-1.29625	-1.56616	-1.96270	-2.43255	-2.90573	-3.40556	-3.62883	-3.74546	-3.81543
1.09558	0.0	-0.15328	-0.30989	-0.39320	-0.47651	-0.56981	-0.67311	-0.77974	-0.89971	-1.02966	-1.16962	-1.21960	-1.47286	-1.82274	-2.20595	-2.61915	-3.20896	-3.49553	-3.63882	-3.71547
1.11059	0.0	-0.14662	-0.29658	-0.37655	-0.44319	-0.55649	-0.65313	-0.75643	-0.86306	-0.98302	-1.10964	-1.15630	-1.37956	-1.66613	-1.98603	-2.37590	-3.00903	-3.35559	-3.51887	-3.61550
1.12561	0.0	-0.13663	-0.28658	-0.36322	-0.44653	-0.53316	-0.62313	-0.72643	-0.82974	-0.93969	-1.04966	-1.08632	-1.28292	-1.54284	-1.83940	-2.11597	-2.83575	-3.21896	-3.40890	-3.51220
1.14063	0.0	-0.12663	-0.27325	-0.34989	-0.42987	-0.51650	-0.60314	-0.69645	-0.79308	-0.89305	-0.99968	-1.03300	-1.21627	-1.45620	-1.71945	-2.01269	-2.69580	-3.07235	-3.27228	-3.41556
1.15564	0.0	-0.12329	-0.26658	-0.34322	-0.42320	-0.50650	-0.59314	-0.67978	-0.77308	-0.86639	-0.96302	-0.99634	-1.16295	-1.37955	-1.61281	-1.87939	-2.52585	-2.93239	-3.15565	-3.31226
1.17066	0.0	-0.12995	-0.26658	-0.32989	-0.40986	-0.48650	-0.56648	-0.64979	-0.73642	-0.82307	-0.91304	-0.94302	-1.09964	-1.29958	-1.52284	-1.77276	-2.38256	-2.79910	-3.04234	-3.21229
1.18567	0.0	-0.11996	-0.25325	-0.32323	-0.39654	-0.47318	-0.54982	-0.63313	-0.71310	-0.79641	-0.88305	-0.91637	-1.06632	-1.25959	-1.45952	-1.67612	-2.23927	-2.66580	-2.92571	-3.11566
1.20069	0.0	-0.12329	-0.24992	-0.31323	-0.38321	-0.45652	-0.52983	-0.60647	-0.68978	-0.76975	-0.85306	-0.88305	-1.02634	-1.20627	-1.39621	-1.59615	-2.12598	-2.54251	-2.81242	-3.01903
1.21571	0.0	-0.11663	-0.23993	-0.30657	-0.37322	-0.44653	-0.51984	-0.59314	-0.66979	-0.74643	-0.82974	-0.85640	-0.99635	-1.16296	-1.33957	-1.52285	-2.00602	-2.41922	-2.70913	-2.92572
1.23072	0.0	-0.10997	-0.22992	-0.29324	-0.35989	-0.42986	-0.49984	-0.57315	-0.64313	-0.71976	-0.79308	-0.81973	-0.95302	-1.11631	-1.28292	-1.45619	-1.91271	-2.31258	-2.60582	-2.82242
1.24574	0.0	-0.11329	-0.22660	-0.28990	-0.35322	-0.41653	-0.48651	-0.55316	-0.62647	-0.69311	-0.76642	-0.78974	-0.91304	-1.06966	-1.22960	-1.40288	-1.83940	-2.21595	-2.50253	-2.73579
1.26075	0.0	-0.10330	-0.21992	-0.27991	-0.34322	-0.40653	-0.47318	-0.53982	-0.60980	-0.67645	-0.74642	-0.76975	-0.88638	-1.03300	-1.18295	-1.34290	-1.75276	-2.11598	-2.40255	-2.64581
1.27577	0.0	-0.10664	-0.21993	-0.27658	-0.33656	-0.39987	-0.45985	-0.52650	-0.59314	-0.65646	-0.72643	-0.74976	-0.86306	-1.00301	-1.14630	-1.29292	-1.67946	-2.03268	-2.31592	-2.55917
1.29079	0.0	-0.10663	-0.21660	-0.26992	-0.33323	-0.39321	-0.45652	-0.51984	-0.58648	-0.64313	-0.71310	-0.72976	-0.84306	-0.97635	-1.11297	-1.25293	-1.61614	-1.95603	-2.23261	-2.47587
1.30580	0.0	-0.09997	-0.20660	-0.25991	-0.31656	-0.37654	-0.43319	-0.49317	-0.55649	-0.61647	-0.67644	-0.69977	-0.80640	-0.93636	-1.06632	-1.19628	-1.54283	-1.87605	-2.13597	-2.38589
1.32082	0.0	-0.09663	-0.19994	-0.25325	-0.30989	-0.36655	-0.42320	-0.48318	-0.53982	-0.59981	-0.66312	-0.68311	-0.78641	-0.91304	-1.03633	-1.15962	-1.48285	-1.80275	-2.06600	-2.30592
1.33584	0.0	-0.09997	-0.19994	-0.24992	-0.30324	-0.35655	-0.41320	-0.46652	-0.52650	-0.57981	-0.63646	-0.65979	-0.75642	-0.87305	-0.99301	-1.11297	-1.42953	-1.73944	-1.99602	-2.23927
1.35085	0.0	-0.09330	-0.19327	-0.23993	-0.29324	-0.34656	-0.39987	-0.45985	-0.51650	-0.56648	-0.62314	-0.64313	-0.73643	-0.84973	-0.96303	-1.07966	-1.37956	-1.67613	-1.93271	-2.15598
1.36587	0.0	-0.08997	-0.18660	-0.23659	-0.28657	-0.33655	-0.38987	-0.43986	-0.49650	-0.54982	-0.59981	-0.61979	-0.71310	-0.81973	-0.93303	-1.04632	-1.33290	-1.61947	-1.86939	-2.08599
1.38088	0.0	-0.08331	-0.17994	-0.22659	-0.27657	-0.32655	-0.37654	-0.42986	-0.48318	-0.53316	-0.58647	-0.60313	-0.68644	-0.79307	-0.89971	-1.00967	-1.28958	-1.56283	-1.80941	-2.01934
1.39590	0.0	-0.08664	-0.17661	-0.22327	-0.27325	-0.31990	-0.36988	-0.41653	-0.46652	-0.51650	-0.56316	-0.57982	-0.66645	-0.76642	-0.87306	-0.97969	-1.24627	-1.51285	-1.74277	-1.95604
1.41092	0.0	-0.08663	-0.16994	-0.21660	-0.25991	-0.30657	-0.35321	-0.39987	-0.44985	-0.49650	-0.54315	-0.55981	-0.64312	-0.74309	-0.84639	-0.94636	-1.20294	-1.45952	-1.68279	-1.89272
1.42593	0.0	-0.08331	-0.16995	-0.21327	-0.25658	-0.30324	-0.34989	-0.39654	-0.44319	-0.48984	-0.53650	-0.55316	-0.63313	-0.72977	-0.82307	-0.92303	-1.16629	-1.41621	-1.63614	-1.83940
1.44095	0.0	-0.07997	-0.16328	-0.20326	-0.24658	-0.29324	-0.33989	-0.38654	-0.43319	-0.47651	-0.52650	-0.53982	-0.61647	-0.70977	-0.80307	-0.89971	-1.13629	-1.37289	-1.58615	-1.78608

TABLE LXIII

REDUCED ENTHALPY AS A FUNCTION OF REDUCED TEMPERATURE AND REDUCED PRESSURE FOR

METHANE

TR \ PR	0.0	0.07474	0.14948	0.22422	0.29895	0.37369	0.44843	0.52317	0.59791	0.67265	0.74738	0.82212	0.89686	0.93423	1.01644	1.04634	1.19581	1.34529	1.49477	1.79372	2.24215	2.98954
0.52361	0.0	-5.32907	-5.33331	-5.33543	-5.33731	-5.33825	-5.33896	-5.33825	-5.33943	-5.33896	-5.33378	-5.33660	-5.33331	-5.33472	-5.33072	-5.32766	-5.31636	-5.30553	-5.29188	-5.24998	-5.20079	-5.14406
0.55277	0.0	-5.25610	-5.26034	-5.26246	-5.26434	-5.26528	-5.26599	-5.26528	-5.26646	-5.26599	-5.26081	-5.26363	-5.26034	-5.26175	-5.25775	-5.25469	-5.24339	-5.23256	-5.21891	-5.17701	-5.12782	-5.07109
0.58192	0.0	-5.18313	-5.18737	-5.18949	-5.19373	-5.19467	-5.19537	-5.19467	-5.19584	-5.19537	-5.19020	-5.19302	-5.18972	-5.19114	-5.18713	-5.18408	-5.17278	-5.16195	-5.14830	-5.10640	-5.05720	-5.00048
0.61108	0.0	-5.10546	-5.10969	-5.11417	-5.11605	-5.11699	-5.11770	-5.11699	-5.11817	-5.11770	-5.11252	-5.11534	-5.11205	-5.11346	-5.10946	-5.10640	-5.09745	-5.08663	-5.07297	-5.03343	-4.98423	-4.92751
0.64023	0.0	-5.02543	-5.02966	-5.03414	-5.03602	-5.03696	-5.03767	-5.03696	-5.03814	-5.03767	-5.03249	-5.03531	-5.03202	-5.03343	-5.02943	-5.02637	-5.01978	-5.00895	-4.99530	-4.95811	-4.90891	-4.85219
0.66939	0.0	-4.94069	-4.94493	-4.94940	-4.95128	-4.95222	-4.95293	-4.95222	-4.95340	-4.95293	-4.94775	-4.95293	-4.94963	-4.95105	-4.94940	-4.94634	-4.93975	-4.92892	-4.91762	-4.88043	-4.83359	-4.77686
0.69854	0.0	-0.12240	-4.85783	-4.86231	-4.86419	-4.86513	-4.86584	-4.86749	-4.86866	-4.86819	-4.86301	-4.86819	-4.86725	-4.86866	-4.86937	-4.86631	-4.86207	-4.85124	-4.83994	-4.80275	-4.75591	-4.70154
0.72770	0.0	-0.12004	-4.76368	-4.76815	-4.77003	-4.77098	-4.77168	-4.77568	-4.77686	-4.77639	-4.77592	-4.78110	-4.78016	-4.78157	-4.78227	-4.77921	-4.77498	-4.76415	-4.75521	-4.71801	-4.67353	-4.62386
0.75685	0.0	-0.11769	-0.24903	-4.66223	-4.66646	-4.66976	-4.67517	-4.67917	-4.68035	-4.67988	-4.67941	-4.68694	-4.68600	-4.68741	-4.68812	-4.68741	-4.68553	-4.67470	-4.66576	-4.63092	-4.58879	-4.54383
0.78601	0.0	-0.11063	-0.23727	-4.55395	-4.56054	-4.56384	-4.56925	-4.57325	-4.57679	-4.57631	-4.58055	-4.59044	-4.59185	-4.59326	-4.59397	-4.59326	-4.59138	-4.58055	-4.57396	-4.54148	-4.50170	-4.46145
0.81516	0.0	-0.10592	-0.22550	-0.37356	-4.44756	-4.45556	-4.46098	-4.46733	-4.47557	-4.47510	-4.48169	-4.48922	-4.49063	-4.49205	-4.49511	-4.49440	-4.49252	-4.48405	-4.47745	-4.44968	-4.41225	-4.37906
0.84431	0.0	-0.09886	-0.21138	-0.34295	-0.49077	-4.32610	-4.33387	-4.34022	-4.34846	-4.35270	-4.35929	-4.37153	-4.37530	-4.37906	-4.38683	-4.38612	-4.38424	-4.37812	-4.37388	-4.35082	-4.31810	-4.29197
0.87347	0.0	-0.09180	-0.19961	-0.31471	-0.44605	-0.59529	-4.19029	-4.20606	-4.21430	-4.21853	-4.22512	-4.23972	-4.24584	-4.25196	-4.26443	-4.26373	-4.26655	-4.26985	-4.26561	-4.24490	-4.21924	-4.20017
0.90262	0.0	-0.08474	-0.18784	-0.29352	-0.41075	-0.53879	-0.69250	-0.88009	-4.04247	-4.06318	-4.07448	-4.09142	-4.10461	-4.11308	-4.12556	-4.12956	-4.13709	-4.14745	-4.15027	-4.13426	-4.11567	-4.10367
0.93178	0.0	-0.08003	-0.17842	-0.27705	-0.38249	-0.49407	-0.62424	-0.77417	-0.96601	-3.86310	-3.88617	-3.91018	-3.93042	-3.93890	-3.95843	-3.96950	-3.98880	-4.00622	-4.01610	-4.01657	-4.00739	-4.00716
0.96093	0.0	-0.07532	-0.16900	-0.25821	-0.35425	-0.45876	-0.56774	-0.69649	-0.85067	-1.03144	-1.26400	-3.65361	-3.69739	-3.71528	-3.75365	-3.76707	-3.81226	-3.83909	-3.85839	-3.87534	-3.88735	-3.90359
0.97551	0.0	-0.07297	-0.16665	-0.25115	-0.34483	-0.44464	-0.54891	-0.66825	-0.81066	-0.96554	-1.16279	-1.44336	-3.51379	-3.55051	-3.62654	-3.64231	-3.71104	-3.74729	-3.77130	-3.80237	-3.82144	-3.84945
0.99009	0.0	-0.07297	-0.16194	-0.24409	-0.33306	-0.42816	-0.52773	-0.63765	-0.76829	-0.91140	-1.08041	-1.28566	-1.59307	-1.65803	-3.43117	-3.47048	-3.58158	-3.63431	-3.67244	-3.72234	-3.75788	-3.79531
1.00000	0.0	-0.07297	-0.18077	-0.23938	-0.32365	-0.41639	-0.50654	-0.61411	-0.73534	-0.89493	-1.04274	-1.22682	-1.47302	-1.63214	-3.19814	-3.31513	-3.48978	-3.57076	-3.61830	-3.68468	-3.68727	-3.73882
1.01924	0.0	-0.07061	-0.15488	-0.23232	-0.31424	-0.40227	-0.49007	-0.59057	-0.70238	-0.82195	-0.95095	-1.10441	-1.29178	-1.40147	-1.75289	-2.02288	-3.14612	-3.34714	-3.44177	-3.54580	-3.61666	-3.68233
1.03382	0.0	-0.06826	-0.15252	-0.22526	-0.30482	-0.39050	-0.47359	-0.56939	-0.67413	-0.78665	-0.90151	-1.04086	-1.19998	-1.29790	-1.54576	-1.67922	-2.59297	-3.11176	-3.29583	-3.44459	-3.53662	-3.62348
1.04840	0.0	-0.06826	-0.14782	-0.22055	-0.29540	-0.37873	-0.45711	-0.54820	-0.64824	-0.75370	-0.86150	-0.98908	-1.12936	-1.21552	-1.42100	-1.50268	-2.04453	-2.80341	-3.12164	-3.32925	-3.45659	-3.56228
1.06297	0.0	-0.06826	-0.14311	-0.21584	-0.28834	-0.36931	-0.44534	-0.53173	-0.62706	-0.72310	-0.82384	-0.94436	-1.07052	-1.14490	-1.32450	-1.39205	-1.79738	-2.41738	-2.89568	-3.19744	-3.36480	-3.49873
1.07755	0.0	-0.06590	-0.14075	-0.20878	-0.28128	-0.35990	-0.43122	-0.51289	-0.60587	-0.69720	-0.79089	-0.90434	-1.01638	-1.08606	-1.24211	-1.29790	-1.64438	-2.15140	-2.61792	-3.05150	-3.27535	-3.43517
1.10670	0.0	-0.06355	-0.13370	-0.19937	-0.26716	-0.34107	-0.40768	-0.48465	-0.56821	-0.65012	-0.73439	-0.83137	-0.93165	-0.98719	-1.11501	-1.15902	-1.42312	-1.76772	-2.13774	-2.70549	-3.06115	-3.30101
1.13586	0.0	-0.06120	-0.12899	-0.18995	-0.25539	-0.32223	-0.38885	-0.45876	-0.53526	-0.61011	-0.68731	-0.77253	-0.86103	-0.90952	-1.01615	-1.05546	-1.27483	-1.55117	-1.84116	-2.34770	-2.82577	-3.15507
1.16501	0.0	-0.05885	-0.12428	-0.18289	-0.24362	-0.30576	-0.37002	-0.43522	-0.50466	-0.57480	-0.64495	-0.72545	-0.80218	-0.84596	-0.93847	-0.97542	-1.16185	-1.39347	-1.62696	-2.05583	-2.57862	-2.99972
1.19417	0.0	-0.05649	-0.11957	-0.17347	-0.23185	-0.29164	-0.35119	-0.41404	-0.47876	-0.54420	-0.60964	-0.68308	-0.75275	-0.79183	-0.87492	-0.90716	-1.07240	-1.26636	-1.46219	-1.83928	-2.34323	-2.83730
1.22332	0.0	-0.05413	-0.11251	-0.16406	-0.22008	-0.27751	-0.33471	-0.39285	-0.45523	-0.51596	-0.57668	-0.64307	-0.70803	-0.74004	-0.82078	-0.85067	-0.99943	-1.16043	-1.33509	-1.66981	-2.13375	-2.67253
1.25248	0.0	-0.05178	-0.10545	-0.15700	-0.20831	-0.26339	-0.31823	-0.37167	-0.43169	-0.48771	-0.54608	-0.60776	-0.67037	-0.70238	-0.77370	-0.80124	-0.93588	-1.08276	-1.23387	-1.53328	-1.95721	-2.50777
1.28163	0.0	-0.04708	-0.10074	-0.14994	-0.19890	-0.25162	-0.30411	-0.35519	-0.41051	-0.46417	-0.51784	-0.57716	-0.63506	-0.66472	-0.72898	-0.75652	-0.87939	-1.01685	-1.14914	-1.42030	-1.80892	-2.35477
1.31079	0.0	-0.04708	-0.09839	-0.14523	-0.19184	-0.24221	-0.29235	-0.34107	-0.39403	-0.44534	-0.49666	-0.55127	-0.60447	-0.63177	-0.69132	-0.71651	-0.83232	-0.95801	-1.08088	-1.32615	-1.68417	-2.21119
1.33994	0.0	-0.04708	-0.09368	-0.13817	-0.18478	-0.23279	-0.28058	-0.32695	-0.37755	-0.42652	-0.47547	-0.52538	-0.57622	-0.60352	-0.66072	-0.68356	-0.78995	-0.90388	-1.01968	-1.24377	-1.57825	-2.07938
1.36909	0.0	-0.04472	-0.08897	-0.13346	-0.17771	-0.22338	-0.26881	-0.31283	-0.36108	-0.40768	-0.45429	-0.50184	-0.55033	-0.57528	-0.63012	-0.65060	-0.75228	-0.85915	-0.96554	-1.17550	-1.48880	-1.95932
1.39825	0.0	-0.04237	-0.08427	-0.12640	-0.17066	-0.21396	-0.25704	-0.29870	-0.34460	-0.39121	-0.43546	-0.48065	-0.52443	-0.54938	-0.60188	-0.62000	-0.71698	-0.81443	-0.91611	-1.11195	-1.40641	-1.85105
1.42740	0.0	-0.04001	-0.08191	-0.12169	-0.16594	-0.20690	-0.24763	-0.28929	-0.33283	-0.37708	-0.41898	-0.46182	-0.50560	-0.52820	-0.57598	-0.59411	-0.68402	-0.77677	-0.87139	-1.05545	-1.33344	-1.75219
1.45656	0.0	-0.03766	-0.07956	-0.11698	-0.16124	-0.19984	-0.24056	-0.27987	-0.32106	-0.36296	-0.40250	-0.44299	-0.48677	-0.50701	-0.55245	-0.57056	-0.65342	-0.74146	-0.83137	-1.00367	-1.26753	-1.66275
1.48571	0.0	-0.03531	-0.07720	-0.11463	-0.15653	-0.19277	-0.23114	-0.27046	-0.30929	-0.34883	-0.38603	-0.42651	-0.46794	-0.48583	-0.52890	-0.54703	-0.62518	-0.70850	-0.79371	-0.95660	-1.20633	-1.58036

TABLE LXIV

RESULTS OF CORRESPONDING STATES CORRELATION FOR METHANE-PROPANE MIXTURES

USING OPTIMUM VALUES OF THE PARAMETERS

A. 5.1 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	179.5999908	-223.5000000	-223.8465271	0.3465271	-0.1550456
500.0000000	199.5999908	-217.6999969	-217.6114197	-0.0885773	0.0406878
500.0000000	219.5999908	-209.6999969	-211.7310486	2.0310516	-0.9685509
500.0000000	239.5999908	-203.7999573	-205.2352753	1.4353180	-0.7042776
500.0000000	259.5998535	-196.9000092	-198.4780273	1.5780182	-0.8014310
500.0000000	279.5998535	-190.3999634	-190.9170227	0.5170593	-0.2715648
500.0000000	299.5998535	-182.3998566	-182.7281189	0.3282623	-0.1799685
500.0000000	319.5998535	-172.3000031	-172.2226105	-0.0773926	0.0449173
500.0000000	419.5998535	-27.8000488	-26.9967194	-0.8033295	2.8896685
500.0000000	439.5998535	-24.9001465	-24.2286530	-0.6714935	2.6967449
500.0000000	459.5998535	-22.5000000	-21.8375854	-0.6624146	2.9440641
500.0000000	479.5998535	-20.3999023	-20.0686493	-0.3312531	1.6237965
500.0000000	499.5998535	-18.6999512	-18.4226685	-0.2772827	1.4827986
500.0000000	519.5998535	-17.3000488	-17.0725708	-0.2274780	1.3148975
1000.0000000	179.5999908	-221.5000000	-221.9775238	0.4775238	-0.2155863
1000.0000000	199.5999908	-215.4000092	-215.8345032	0.4344940	-0.2017149
1000.0000000	219.5999908	-208.1999969	-210.0275879	1.8275909	-0.8778054
1000.0000000	239.5999908	-202.2999573	-203.8572693	1.5573120	-0.7698033
1000.0000000	259.5998535	-195.5000000	-197.5049744	2.0049744	-1.0255613
1000.0000000	279.5998535	-189.1999512	-190.4310455	1.2310944	-0.6506841
1000.0000000	299.5998535	-182.0998535	-182.7608032	0.6609497	-0.3629600
1000.0000000	319.5998535	-173.6000061	-174.2459564	0.6459503	-0.3720911
1000.0000000	339.5998535	-163.3999023	-163.9692383	0.5693359	-0.3484310
1000.0000000	359.5998535	-149.7998047	-150.3798218	0.5800171	-0.3871948
1000.0000000	379.5998535	-125.4998169	-128.1374054	2.6375885	-2.1016665
1000.0000000	399.5998535	-88.8999023	-89.5020294	0.6021271	-0.6773088
1000.0000000	419.5998535	-68.5000000	-68.5833588	0.0833588	-0.1216916
1000.0000000	439.5998535	-56.8000488	-56.5252686	-0.2747803	0.4837673
1000.0000000	459.5998535	-49.2001953	-48.7428894	-0.4573059	0.9294797
1000.0000000	479.5998535	-43.5000000	-43.3065186	-0.1934814	0.4447848
1000.0000000	499.5998535	-39.0998535	-38.9842224	-0.1156311	0.2957328
1000.0000000	519.5998535	-35.6000977	-35.4225464	-0.1775513	0.4987381
1500.0000000	179.5999908	-219.3999939	-218.2074738	-1.1925201	0.5435370
1500.0000000	199.5999908	-213.1999969	-212.1310272	-1.0689697	0.5013928
1500.0000000	219.5999908	-206.3999939	-206.4019775	0.0019836	-0.0009611
1500.0000000	239.5999908	-200.8999634	-200.4232788	-0.4766846	0.2372746
1500.0000000	259.5998535	-194.3000031	-194.1630096	-0.1369934	0.0705061
1500.0000000	279.5998535	-187.8999634	-187.4026794	-0.4972839	0.2646535
1500.0000000	299.5998535	-181.0998535	-180.2934113	-0.8064423	0.4453022
1500.0000000	319.5998535	-173.7000122	-172.5238342	-1.1761780	0.6771315
1500.0000000	339.5998535	-165.4999084	-164.0117950	-1.4881134	0.8991625
1500.0000000	359.5998535	-155.3998108	-153.5645447	-1.8352661	1.1809959
1500.0000000	379.5998535	-143.0998077	-142.0413971	-1.0584106	0.7396311
1500.0000000	399.5998535	-127.4999084	-126.5802460	-0.9196625	0.7213041
1500.0000000	419.5998535	-108.8000031	-108.0916443	-0.7083588	0.6510649
1500.0000000	439.5998535	-91.1999512	-90.3188019	-0.8811493	0.9661726
1500.0000000	459.5998535	-77.6000977	-76.8830719	-0.7170258	0.9240009
1500.0000000	479.5998535	-67.5998535	-67.1749878	-0.4248657	0.6285008
1500.0000000	499.5998535	-60.0000000	-59.9863129	-0.0136871	0.0228119
1500.0000000	519.5998535	-54.0000000	-54.1292572	0.1292572	-0.2393652
2000.0000000	179.5999908	-217.2999878	-215.9124298	-1.3875580	0.6385446
2000.0000000	199.5999908	-211.1000061	-209.8416138	-1.2583923	0.5961116
2000.0000000	219.5999908	-204.7999878	-204.1183624	-0.6816254	0.3328249
2000.0000000	239.5999908	-199.5999603	-198.1295013	-1.4704590	0.7367030
2000.0000000	259.5998535	-193.1000061	-192.0975647	-1.0024414	0.5191303
2000.0000000	279.5998535	-186.6999512	-185.7134094	-0.9865417	0.5284101
2000.0000000	299.5998535	-180.1998596	-179.1232758	-1.0765839	0.5974386
2000.0000000	319.5998535	-173.2000122	-171.8867340	-1.3132782	0.7582433
2000.0000000	339.5998535	-165.3999023	-164.2259216	-1.1739807	0.7097829
2000.0000000	359.5998535	-156.9998169	-155.5579681	-1.4418488	0.9183760
2000.0000000	379.5998535	-147.4998169	-146.7535095	-0.7463074	0.5059715
2000.0000000	399.5998535	-136.6999054	-136.7344360	0.0345306	-0.0252602
2000.0000000	419.5998535	-124.5000000	-124.3732452	-0.1267548	0.1018110
2000.0000000	439.5998535	-111.5999603	-111.8323975	0.2324371	-0.2082770
2000.0000000	459.5998535	-99.0000000	-99.3701172	0.3701172	-0.3738557
2000.0000000	479.5998535	-87.5998535	-88.2673035	0.6674500	-0.7619302
2000.0000000	499.5998535	-78.0998535	-78.8437347	0.7438812	-0.9524744
2000.0000000	519.5998535	-70.3000488	-70.9891663	0.6891174	-0.9802516

TABLE LXIV
(CONTINUED)

B. 11.7 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (^{o}R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	199.5999908	-223.6999969	-222.9557190	-0.7442780	0.3327125
500.0000000	219.5999908	-218.0999908	-217.0829010	-1.0170898	0.4663408
500.0000000	239.5999908	-212.0999146	-211.0576172	-1.0422974	0.4914179
500.0000000	259.5998535	-205.7000122	-204.6004181	-1.0995941	0.5345616
500.0000000	279.5998535	-198.7998657	-197.9979553	-0.8019104	0.4033755
500.0000000	299.5998535	-191.2999115	-190.7028503	-0.5970612	0.3121073
500.0000000	319.5998535	-181.8999634	-182.8739929	0.9740295	-0.5354751
500.0000000	479.5998535	-24.5998535	-22.6493683	-1.9504852	7.9288483
500.0000000	499.5998535	-20.9001465	-20.6938171	-0.2063293	0.9872146
500.0000000	519.5998535	-18.6999512	-19.0550690	0.3551178	-1.8990297
500.0000000	539.5998535	-17.3999023	-17.5503082	0.1504059	-0.8644063
500.0000000	559.5998535	-16.8000488	-16.3127899	-0.4872589	2.9003410
1000.0000000	199.5999908	-220.8999939	-221.2119751	0.3119812	-0.1412318
1000.0000000	219.5999908	-215.0999908	-215.3538208	0.2538300	-0.1180055
1000.0000000	239.5999908	-209.2999115	-209.4137573	0.1138458	-0.0543936
1000.0000000	259.5998535	-203.1000061	-203.2781982	0.1781921	-0.0877361
1000.0000000	279.5998535	-196.2998657	-196.9353180	0.6354523	-0.3237150
1000.0000000	299.5998535	-189.2999115	-190.0705261	0.7706146	-0.4070863
1000.0000000	319.5998535	-181.6999512	-182.6600189	0.9600677	-0.5283806
1000.0000000	339.5998535	-172.9000092	-174.5535736	1.6535645	-0.9563703
1000.0000000	359.5998535	-163.8000031	-165.1951599	1.3951569	-0.8517440
1000.0000000	379.5998535	-150.2999115	-152.9467773	2.6468658	-1.7610559
1000.0000000	479.5998535	-52.8999023	-51.6362762	-1.2636261	2.3887110
1000.0000000	499.5998535	-45.9001465	-45.4912415	-0.4089050	0.8908577
1000.0000000	519.5998535	-40.8000488	-40.7847748	-0.0152740	0.0374363
1000.0000000	539.5998535	-37.0998535	-36.9780884	-0.1217651	0.3282092
1000.0000000	559.5998535	-34.1999512	-33.8241577	-0.3757935	1.0988121
1500.0000000	199.5999908	-218.0000000	-217.8072357	-0.1927643	0.0884240
1500.0000000	219.5999908	-212.1999969	-211.9319611	-0.2680359	0.1263129
1500.0000000	239.5999908	-206.3999023	-206.1060028	-0.2938995	0.1423932
1500.0000000	259.5998535	-200.5000000	-200.1881104	-0.3118896	0.1555559
1500.0000000	279.5998535	-193.8998566	-194.0115509	0.1116943	-0.0576041
1500.0000000	299.5998535	-187.3999023	-187.3063507	-0.0935516	0.0499208
1500.0000000	319.5998535	-180.3999634	-180.4412384	0.0412750	-0.0228797
1500.0000000	339.5998535	-173.0000000	-172.9879761	-0.0120239	0.0069502
1500.0000000	359.5998535	-165.0000000	-164.9716949	-0.0283051	0.0171546
1500.0000000	379.5998535	-155.5999146	-155.7413025	0.1413879	-0.0908663
1500.0000000	399.5998535	-144.3000031	-145.0418549	0.7418518	-0.5141035
1500.0000000	419.5998535	-130.1998596	-131.7047424	1.5048828	-1.1558247
1500.0000000	439.5998535	-113.4998169	-114.8953552	1.3955383	-1.2295504
1500.0000000	459.5998535	-97.5000000	-97.2541504	-0.2458496	0.2521534
1500.0000000	479.5998535	-83.5000000	-82.4559479	-1.0440521	1.2503614
1500.0000000	499.5998535	-72.5000000	-71.5027771	-0.9972229	1.3754787
1500.0000000	519.5998535	-64.1999512	-63.2568207	-0.9431305	1.4690514
1500.0000000	539.5998535	-57.3999023	-56.9259644	-0.4739380	0.8256770
1500.0000000	559.5998535	-52.1000977	-51.6867065	-0.4133911	0.7934555
2000.0000000	199.5999908	-215.2999878	-215.7004395	0.4004517	-0.1859970
2000.0000000	219.5999908	-209.3999939	-209.8628693	0.4628754	-0.2210484
2000.0000000	239.5999908	-203.8999023	-204.0389252	0.1390228	-0.0681819
2000.0000000	259.5998535	-198.0000000	-198.0725403	0.0725403	-0.0366365
2000.0000000	279.5998535	-191.5998535	-192.1285858	0.5287323	-0.2759565
2000.0000000	299.5998535	-185.4999084	-185.7765656	0.2766571	-0.1491413
2000.0000000	319.5998535	-179.2999573	-179.2699585	-0.0299988	0.0167311
2000.0000000	339.5998535	-172.6000061	-172.3524933	-0.2475128	0.1434025
2000.0000000	359.5998535	-165.4000092	-165.0437469	-0.3562622	0.2153943
2000.0000000	379.5998535	-157.3999023	-157.0870209	-0.3128815	0.1987812
2000.0000000	399.5998535	-148.5000000	-148.4530792	-0.0469208	0.0315965
2000.0000000	419.5998535	-138.3998566	-138.9698334	0.5699768	-0.4118330
2000.0000000	439.5998535	-127.0998077	-128.0904541	0.9906464	-0.7794238
2000.0000000	459.5998535	-115.1999512	-116.2451935	1.0452423	-0.9073287
2000.0000000	479.5998535	-103.3000488	-104.2760315	0.9759827	-0.9448037
2000.0000000	499.5998535	-92.5000000	-93.0516510	0.5516510	-0.5963791
2000.0000000	519.5998535	-82.8999023	-83.1849213	0.2850189	-0.3438109
2000.0000000	539.5998535	-74.6999512	-74.8662415	0.1662903	-0.2226110
2000.0000000	559.5998535	-67.8000488	-67.8438416	0.0437927	-0.0645910

TABLE LXIV
(CONTINUED)

C. 28.0 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	219.5999908	-219.6999969	-219.7361450	0.0361481	-0.0164534
500.0000000	239.5999908	-214.3999939	-213.9856720	-0.4143219	0.1932471
500.0000000	259.5998535	-209.0000000	-208.8074036	-0.1925964	0.0921513
500.0000000	279.5998535	-202.8000031	-203.2437439	0.4437408	-0.2188071
500.0000000	299.5998535	-197.2998657	-197.4878082	0.1879425	-0.0952572
500.0000000	319.5998535	-190.5998077	-191.6642914	1.0644836	-0.5584914
500.0000000	339.5998535	-184.1998596	-185.4840851	1.2842255	-0.6971911
500.0000000	519.5998535	-27.0998535	-25.2987976	-1.8010559	6.6459951
500.0000000	539.5998535	-24.3000488	-23.0268860	-1.2731628	5.2393417
500.0000000	559.5998535	-22.1000977	-21.0830688	-1.0170288	4.6019192
500.0000000	579.5998535	-20.3000488	-19.5655518	-0.7344971	3.6182022
500.0000000	599.5998535	-18.7998047	-18.1147308	-0.6850739	3.6440468
500.0000000	619.5998535	-17.3000488	-16.7906189	-0.5094299	2.9446726
500.0000000	639.5998535	-16.1000977	-15.7055397	-0.3945580	2.4506550
1000.0000000	219.5999908	-217.4000092	-217.8384857	0.4384766	-0.2016911
1000.0000000	239.5999908	-212.0999908	-212.2733154	0.1733246	-0.0817183
1000.0000000	259.5998535	-206.8000031	-207.0347900	0.2347870	-0.1135333
1000.0000000	279.5998535	-201.0000000	-201.5647888	0.5647888	-0.2809894
1000.0000000	299.5998535	-195.5998535	-196.0337372	0.4338837	-0.2218220
1000.0000000	319.5998535	-189.4998169	-190.3890381	0.8892212	-0.4692461
1000.0000000	339.5998535	-183.6998596	-184.4388123	0.7389526	-0.4022605
1000.0000000	359.5998535	-177.0000000	-178.0918274	1.0918274	-0.6168514
1000.0000000	379.5998535	-170.3000031	-171.3118591	1.0118561	-0.5941607
1000.0000000	399.5998535	-161.6998596	-163.9705963	2.2707367	-1.4042902
1000.0000000	539.5998535	-55.8000488	-53.8077698	-1.9922791	3.5703888
1000.0000000	559.5998535	-48.6999512	-47.3797913	-1.3201599	2.7108030
1000.0000000	579.5998535	-43.8000488	-42.6066895	-1.1933594	2.7245607
1000.0000000	599.5998535	-39.7998047	-38.6299438	-1.1698608	2.9393625
1000.0000000	619.5998535	-36.3000488	-35.3710785	-0.9289703	2.5591431
1000.0000000	639.5998535	-33.3999023	-32.6143951	-0.7855072	2.3518238
1500.0000000	219.5999908	-215.1000061	-214.9495850	-0.1504211	0.0699307
1500.0000000	239.5999908	-209.8000031	-209.3151093	-0.4848938	0.2311219
1500.0000000	259.5998535	-204.7000122	-204.0634766	-0.6365356	0.3109602
1500.0000000	279.5998535	-199.4000092	-198.6932678	-0.7067413	0.3544339
1500.0000000	299.5998535	-194.0998535	-193.4026489	-0.6972046	0.3591989
1500.0000000	319.5998535	-188.4998169	-187.9442444	-0.5555725	0.2947336
1500.0000000	339.5998535	-182.6998596	-182.0074005	-0.6924591	0.3790146
1500.0000000	359.5998535	-176.5000000	-175.9835968	-0.5164032	0.2925797
1500.0000000	379.5998535	-170.5000000	-169.7929535	-0.7070465	0.4146900
1500.0000000	399.5998535	-163.8998566	-163.0111237	-0.8887329	0.5422413
1500.0000000	419.5998535	-156.3999023	-155.8802948	-0.5196075	0.3322300
1500.0000000	439.5998535	-147.6998138	-147.9449615	0.2451477	-0.1659769
1500.0000000	459.5998535	-137.3998108	-138.6822968	1.2824860	-0.9333968
1500.0000000	479.5998535	-125.3000031	-127.5434113	2.2434082	-1.7904291
1500.0000000	499.5998535	-111.6998138	-114.0764008	2.3765869	-2.1276541
1500.0000000	519.5998535	-98.0998535	-99.8649902	1.7651367	-1.7993259
1500.0000000	539.5998535	-85.9001465	-86.2155304	0.3153839	-0.3671517
1500.0000000	559.5998535	-75.8000488	-75.2431946	-0.5568542	0.7346354
1500.0000000	579.5998535	-67.7001953	-66.7154541	-0.9847412	1.4545612
1500.0000000	599.5998535	-61.0000000	-59.9455566	-1.0544434	1.7285948
1500.0000000	619.5998535	-54.4001465	-54.5367737	0.1366272	-0.2511522
1500.0000000	639.5998535	-49.6999512	-49.9396667	0.2397156	-0.4823253
2000.0000000	219.5999908	-212.8000031	-213.1689758	0.3689728	-0.1733894
2000.0000000	239.5999908	-207.5999908	-207.5888214	-0.0111694	0.0053803
2000.0000000	259.5998535	-202.6000061	-202.2734833	-0.3265228	0.1611662
2000.0000000	279.5998535	-197.7000122	-196.9940338	-0.7059784	0.3570957
2000.0000000	299.5998535	-192.5998535	-191.5930481	-1.0068054	0.5227442
2000.0000000	319.5998535	-187.2998047	-186.3367310	-0.9630737	0.5141880
2000.0000000	339.5998535	-181.8998566	-180.6253052	-1.2745514	0.7006884
2000.0000000	359.5998535	-176.2000122	-175.0912170	-1.1087952	0.6292820
2000.0000000	379.5998535	-170.5000000	-169.1955261	-1.3044739	0.7650871
2000.0000000	399.5998535	-163.7998657	-162.8309326	-0.9689331	0.5915344
2000.0000000	419.5998535	-157.0999146	-156.4238129	-0.6761017	0.4303638
2000.0000000	439.5998535	-149.8998108	-149.4604797	-0.4393311	0.2930831
2000.0000000	459.5998535	-141.8998108	-142.1370544	0.2372437	-0.1671909
2000.0000000	479.5998535	-133.2000122	-134.5644684	1.3644562	-1.0243654
2000.0000000	499.5998535	-123.7998047	-125.3279724	1.5281677	-1.2343855
2000.0000000	519.5998535	-113.7998047	-114.5359497	0.7361450	-0.6468769
2000.0000000	539.5998535	-103.8000488	-104.8703308	1.0702820	-1.0310993
2000.0000000	559.5998535	-94.3000488	-95.3125305	1.0124817	-1.0736799
2000.0000000	579.5998535	-85.8000488	-86.3955078	0.5954590	-0.6940074
2000.0000000	599.5998535	-78.0998535	-78.4010773	0.3012238	-0.3856905
2000.0000000	619.5998535	-71.0000000	-71.4418030	0.4418030	-0.6222576
2000.0000000	639.5998535	-64.8999023	-65.3887939	0.4888916	-0.7533006

TABLE LXIV
(CONTINUED)

D. 50.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE (Btu/lb)
500.0000000	259.5998535	-212.7998047	-213.4365540	0.6367493	-0.2992245
500.0000000	279.5998535	-207.8998108	-208.3826752	0.4828644	-0.2322582
500.0000000	299.5998535	-202.8999023	-203.6472321	0.7473297	-0.3683243
500.0000000	319.5998535	-198.2998657	-198.6600647	0.3601990	-0.1816435
500.0000000	339.5998535	-193.1999054	-193.5580902	0.3581848	-0.1853959
500.0000000	359.5998535	-188.0999146	-188.4100800	0.3101654	-0.1648939
500.0000000	579.5998535	-29.7001953	-27.8946075	-1.8055878	6.0793791
500.0000000	599.5998535	-26.5998535	-25.3979950	-1.2018585	4.5182896
500.0000000	619.5998535	-24.1999512	-23.2492371	-0.9507141	3.9285774
500.0000000	639.5998535	-22.2001953	-21.4089813	-0.7912140	3.5639944
500.0000000	659.5998535	-20.6999512	-19.9047089	-0.7952423	3.8417587
500.0000000	679.5998535	-19.2001953	-18.5703430	-0.6298523	3.2804470
500.0000000	699.5998535	-18.0000000	-17.3711395	-0.6288605	3.4936686
500.0000000	719.5998535	-16.8000488	-16.1923523	-0.6076965	3.6172304
500.0000000	739.5998535	-15.7998047	-15.0905466	-0.7092581	4.4890299
1000.0000000	259.5998535	-211.0998077	-211.5381470	0.4383392	-0.2076454
1000.0000000	279.5998535	-206.1998138	-206.6434326	0.4436188	-0.2151402
1000.0000000	299.5998535	-201.3999023	-201.8343353	0.4344330	-0.2157066
1000.0000000	319.5998535	-196.4998627	-196.8405304	0.3406677	-0.1733679
1000.0000000	339.5998535	-191.5999146	-191.9116974	0.3117828	-0.1627259
1000.0000000	359.5998535	-186.6999054	-186.9467773	0.2468719	-0.1322293
1000.0000000	379.5998535	-181.5000000	-181.8092346	0.3092346	-0.1703772
1000.0000000	399.5998535	-176.1999512	-176.5315399	0.3315887	-0.1881889
1000.0000000	419.5998535	-170.7999573	-170.8943787	0.0944214	-0.0552818
1000.0000000	439.5998535	-164.7998047	-165.0761566	0.2763519	-0.1676894
1000.0000000	459.5998535	-158.5998535	-158.8870087	0.2871552	-0.1810563
1000.0000000	599.5998535	-64.5000000	-63.9630127	-0.5369873	0.8325383
1000.0000000	619.5998535	-55.6999512	-55.1568909	-0.5430603	0.9749744
1000.0000000	639.5998535	-49.7001953	-48.9238434	-0.7763519	1.5620699
1000.0000000	659.5998535	-44.8999023	-44.0820770	-0.8178253	1.8214407
1000.0000000	679.5998535	-40.8000488	-40.1949463	-0.6051025	1.4830923
1000.0000000	699.5998535	-37.6000977	-36.8521881	-0.7479095	1.9891157
1000.0000000	719.5998535	-35.1000977	-34.0541229	-1.0459747	2.9799767
1000.0000000	739.5998535	-32.7998047	-31.5944824	-1.2053223	3.6747847
1500.0000000	259.5998535	-209.3998108	-209.2494812	-0.1503296	0.0717907
1500.0000000	279.5998535	-204.4998169	-204.1678009	-0.3320160	0.1623551
1500.0000000	299.5998535	-199.8999023	-199.3116302	-0.5882721	0.2942833
1500.0000000	319.5998535	-194.7998657	-194.4917755	-0.3080902	0.1581573
1500.0000000	339.5998535	-190.0999146	-189.6732025	-0.4267120	0.2244672
1500.0000000	359.5998535	-185.2999115	-184.8447876	-0.4551239	0.2456147
1500.0000000	379.5998535	-180.4000092	-180.0071259	-0.3928833	0.2177845
1500.0000000	399.5998535	-175.1999512	-174.5782013	-0.6217499	0.3548802
1500.0000000	419.5998535	-169.9999542	-169.2781677	-0.7217865	0.4245803
1500.0000000	439.5998535	-164.3998108	-163.9106445	-0.4891663	0.2975467
1500.0000000	459.5998535	-158.7998657	-158.1855927	-0.6142731	0.3868221
1500.0000000	479.5998535	-152.8000031	-152.1361237	-0.6638794	0.4344757
1500.0000000	499.5998535	-146.0999603	-145.5708466	-0.5291138	0.3621587
1500.0000000	519.5998535	-138.2998657	-138.3458405	0.0459747	-0.0332428
1500.0000000	539.5998535	-129.3999634	-130.5618744	1.1619110	-0.8979220
1500.0000000	559.5998535	-119.8000031	-121.1949463	1.3949432	-1.1643925
1500.0000000	579.5998535	-108.5000000	-110.3257751	1.8257751	-1.6827412
1500.0000000	599.5998535	-97.1999512	-99.1565399	1.9565887	-2.0129519
1500.0000000	619.5998535	-86.5998535	-87.8372803	1.2374268	-1.4289007
1500.0000000	639.5998535	-77.5000000	-77.6985779	0.1985779	-0.2562295
1500.0000000	659.5998535	-69.3000488	-69.4961395	0.1960907	-0.2829589
1500.0000000	679.5998535	-63.0000000	-62.8487396	-0.1512604	0.2400958
1500.0000000	699.5998535	-57.6999512	-57.2648468	-0.4351044	0.7540807
1500.0000000	719.5998535	-53.1000977	-52.5844574	-0.5156403	0.9710718
1500.0000000	739.5998535	-49.3999023	-48.5066986	-0.8932037	1.8081074
2000.0000000	259.5998535	-207.5998077	-207.8436737	0.2438660	-0.1174693
2000.0000000	279.5998535	-202.9998169	-202.9090576	-0.0907593	0.0447090
2000.0000000	299.5998535	-197.9999084	-197.9739075	-0.0260010	0.0131318
2000.0000000	319.5998535	-193.2998657	-193.1004028	-0.1994629	0.1031883
2000.0000000	339.5998535	-188.4999084	-188.2640686	-0.2358398	0.1251140
2000.0000000	359.5998535	-183.6999054	-183.5434113	-0.1564941	0.0851901
2000.0000000	379.5998535	-178.8000031	-178.6508179	-0.1491852	0.0834368
2000.0000000	399.5998535	-173.8999634	-173.6158600	-0.2841034	0.1633717
2000.0000000	419.5998535	-168.9999542	-168.7568359	-0.2431183	0.1438569
2000.0000000	439.5998535	-163.8998108	-163.5118713	-0.3879395	0.2366930
2000.0000000	459.5998535	-158.4998627	-158.0872192	-0.4126434	0.2603431
2000.0000000	479.5998535	-153.0000000	-152.6286469	-0.3713531	0.2427145
2000.0000000	499.5998535	-147.1999512	-146.6492767	-0.5506744	0.3740996
2000.0000000	519.5998535	-140.8998566	-140.9973907	0.0975342	-0.0692223
2000.0000000	539.5998535	-134.0999603	-134.6815338	0.5815735	-0.4336864
2000.0000000	559.5998535	-126.9000092	-128.5209351	1.6209259	-1.2773247
2000.0000000	579.5998535	-119.0000000	-120.7305603	1.7305603	-1.4542513
2000.0000000	599.5998535	-110.7998047	-110.8277435	0.0279388	-0.0252156
2000.0000000	619.5998535	-102.6999512	-102.6545715	-0.0453796	0.0441866
2000.0000000	639.5998535	-94.6000977	-95.1943207	0.5942230	-0.6281421
2000.0000000	659.5998535	-86.8999023	-87.7607880	0.8608856	-0.9906635
2000.0000000	679.5998535	-80.0000000	-80.5902405	0.5902405	-0.7378004
2000.0000000	699.5998535	-73.8999023	-74.2270966	0.3271942	-0.4427530
2000.0000000	719.5998535	-68.4001465	-68.3797455	-0.0204010	0.0298260
2000.0000000	739.5998535	-63.5998535	-63.1242065	-0.4756470	0.7478740

TABLE LXIV

(CONTINUED)

E. 76.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	259.5998535	-215.0998535	-216.0159760	0.9161224	-0.4259053
500.0000000	279.5998535	-210.4998169	-211.0794220	0.5796051	-0.2753471
500.0000000	299.5998535	-205.9999542	-206.1875153	0.1875610	-0.0910490
500.0000000	319.5998535	-201.4999542	-201.6920776	0.1921234	-0.0953466
500.0000000	339.5998535	-196.9998169	-197.3670197	0.3672028	-0.1863975
500.0000000	359.5998535	-192.6998596	-192.8692474	0.1693878	-0.0879024
500.0000000	379.5998535	-188.0998535	-188.2959747	0.1961212	-0.1042644
500.0000000	399.5998535	-183.5998077	-183.8383789	0.2385712	-0.1299408
500.0000000	419.5998535	-178.8998566	-179.2449188	0.3450623	-0.1928801
500.0000000	439.5998535	-174.2999573	-174.4692841	0.1693268	-0.0971467
500.0000000	459.5998535	-169.3999023	-169.8678589	0.4679565	-0.2762437
500.0000000	619.5998535	-35.1000977	-34.1748047	-0.9252930	2.6361542
500.0000000	639.5998535	-31.0000000	-30.2193756	-0.7806244	2.5181427
500.0000000	659.5998535	-27.8999023	-27.2233734	-0.6765289	2.4248428
500.0000000	679.5998535	-25.3999023	-24.8325653	-0.5673370	2.2336187
500.0000000	699.5998535	-23.3999023	-22.8972473	-0.5026550	2.1481066
500.0000000	719.5998535	-21.6999512	-21.3358612	-0.3640900	1.6778374
500.0000000	739.5998535	-19.9001465	-19.8022156	-0.0979309	0.4921112
500.0000000	759.5998535	-17.6999512	-18.6399384	0.9399872	-5.3106766
1000.0000000	259.5998535	-213.0998535	-214.0740051	0.9741516	-0.4571337
1000.0000000	279.5998535	-208.3998108	-209.1317596	0.7319489	-0.3512233
1000.0000000	299.5998535	-204.0999603	-204.4316711	0.3317108	-0.1625236
1000.0000000	319.5998535	-199.6999512	-200.0546722	0.3547211	-0.1776270
1000.0000000	339.5998535	-195.2998047	-195.6636963	0.3638916	-0.1863246
1000.0000000	359.5998535	-190.8998566	-191.1503754	0.2505188	-0.1312305
1000.0000000	379.5998535	-186.5998535	-186.7207184	0.1208649	-0.0647722
1000.0000000	399.5998535	-182.2998047	-182.1741638	-0.1256409	0.0689198
1000.0000000	419.5998535	-177.7998657	-177.7269287	-0.0729370	0.0410220
1000.0000000	439.5998535	-173.5999603	-173.1340179	-0.4659424	0.2684000
1000.0000000	459.5998535	-169.2999115	-168.4081879	-0.8917236	0.5267121
1000.0000000	479.5998535	-164.3999023	-163.4295044	-0.9703979	0.5902666
1000.0000000	499.5998535	-159.2998657	-158.4062347	-0.8936310	0.5609739
1000.0000000	519.5998535	-153.7000122	-153.0404663	-0.6595459	0.4291121
1000.0000000	539.5998535	-148.0999603	-147.2147064	-0.8852539	0.5977407
1000.0000000	559.5998535	-141.7998657	-140.7905579	-1.0093079	0.7117830
1000.0000000	579.5998535	-134.2999573	-133.4449310	-0.8550262	0.6366540
1000.0000000	599.5998535	-125.1998596	-124.8813477	-0.3185120	0.2544028
1000.0000000	619.5998535	-112.6000977	-113.3713989	0.7713013	-0.6849915
1000.0000000	639.5998535	-94.4001465	-96.1907349	1.7905884	-1.8968058
1000.0000000	659.5998535	-77.0998535	-76.7966156	-0.3032379	0.3933053
1000.0000000	679.5998535	-65.3000488	-63.6920471	-1.6080017	2.4624805
1000.0000000	699.5998535	-55.8000488	-55.2490692	-0.5509796	0.9874176
1000.0000000	719.5998535	-49.5000000	-49.2533264	-0.2466736	0.4983302
1000.0000000	739.5998535	-44.8000488	-44.6920166	-0.1080322	0.2411430
1000.0000000	759.5998535	-41.8999023	-40.8411102	-1.0587921	2.5269556
1500.0000000	259.5998535	-211.2998657	-211.9787292	0.6788635	-0.3212796
1500.0000000	279.5998535	-206.8998108	-207.2814789	0.3816681	-0.1844699
1500.0000000	299.5998535	-202.5999603	-202.5853119	-0.0146484	0.0072302
1500.0000000	319.5998535	-198.0999603	-197.9884338	-0.1115265	0.0562981
1500.0000000	339.5998535	-193.8998108	-193.5381317	-0.3616791	0.1865288
1500.0000000	359.5998535	-189.3998566	-189.2272491	-0.1726074	0.0911338
1500.0000000	379.5998535	-185.2998657	-184.8013611	-0.4985046	0.2690259
1500.0000000	399.5998535	-180.7998047	-180.5911407	-0.2086639	0.1154116
1500.0000000	419.5998535	-176.2998657	-176.2270050	-0.0728607	0.0413277
1500.0000000	439.5998535	-171.9999542	-171.7733612	-0.2265930	0.1317401
1500.0000000	459.5998535	-167.4999084	-166.9301605	-0.5697479	0.3401482
1500.0000000	479.5998535	-162.9999084	-162.2540436	-0.7458649	0.4575859
1500.0000000	499.5998535	-158.1998596	-157.5592651	-0.6405945	0.4049271
1500.0000000	519.5998535	-153.1000061	-152.5584564	-0.5415497	0.3537228
1500.0000000	539.5998535	-148.1999512	-147.2889099	-0.9110413	0.6147377
1500.0000000	559.5998535	-142.7998657	-142.0412140	-0.7586517	0.5312688
1500.0000000	579.5998535	-136.6999512	-136.2548065	-0.4451447	0.3256363
1500.0000000	599.5998535	-130.1998596	-129.7807159	-0.4191437	0.3219233
1500.0000000	619.5998535	-123.1999512	-122.8943024	-0.3056488	0.2480916
1500.0000000	639.5998535	-115.0000000	-115.0190887	0.0190887	-0.0165989
1500.0000000	659.5998535	-106.5998535	-105.9700928	-0.6297607	0.5907707
1500.0000000	679.5998535	-97.0000000	-96.5324860	-0.4675140	0.4819732
1500.0000000	699.5998535	-86.8000488	-87.1248169	0.3247681	-0.3741565
1500.0000000	719.5998535	-77.5000000	-78.2465820	0.7465820	-0.9633313
1500.0000000	739.5998535	-70.3000488	-70.6694489	0.3694000	-0.5254619
1500.0000000	759.5998535	-63.3999023	-64.1792145	0.7793121	-1.2292004
2000.0000000	259.5998535	-209.4998627	-210.9630432	1.4631805	-0.6984159
2000.0000000	279.5998535	-205.1998138	-206.0866547	0.8868408	-0.4321839
2000.0000000	299.5998535	-200.8999634	-201.4947968	0.5948334	-0.2960843
2000.0000000	319.5998535	-196.4999542	-197.1403961	0.6404419	-0.3259246
2000.0000000	339.5998535	-192.3998108	-192.5645142	0.1647034	-0.0856047
2000.0000000	359.5998535	-188.1998596	-188.0761566	-0.1237030	0.0657296
2000.0000000	379.5998535	-184.0998535	-183.7904205	-0.3094330	0.1680788
2000.0000000	399.5998535	-179.8998108	-179.4332886	-0.4665222	0.2593233
2000.0000000	419.5998535	-175.6998596	-175.2061157	-0.4937439	0.2810155
2000.0000000	439.5998535	-171.1999512	-170.6735535	-0.5263977	0.3074753
2000.0000000	459.5998535	-166.6999054	-166.2579803	-0.4419250	0.2651021
2000.0000000	479.5998535	-160.0999146	-161.9475708	1.8476562	-1.1540642
2000.0000000	499.5998535	-157.5998535	-157.1962585	-0.4035950	0.2560884
2000.0000000	519.5998535	-153.1000061	-152.5149689	-0.5850372	0.3821275
2000.0000000	539.5998535	-148.4999542	-147.6849060	-0.8150482	0.5488541
2000.0000000	559.5998535	-143.5998535	-142.9332275	-0.6666260	0.4642244
2000.0000000	579.5998535	-138.4999542	-137.5990753	-0.9008789	0.6504539
2000.0000000	599.5998535	-133.0998535	-132.6007233	-0.4991302	0.3750043
2000.0000000	619.5998535	-127.1999512	-126.8971252	-0.3028259	0.2380707
2000.0000000	639.5998535	-121.0000000	-121.1513214	0.1513214	-0.1250589
2000.0000000	659.5998535	-114.6999512	-114.6189880	-0.0809631	0.0705869
2000.0000000	679.5998535	-107.6000977	-106.8897247	-0.7103729	0.6601970
2000.0000000	699.5998535	-100.3999023	-99.7025604	-0.6973419	0.6945640
2000.0000000	719.5998535	-93.3000488	-93.3039856	0.0039368	-0.0042195
2000.0000000	739.5998535	-86.4001465	-86.8682861	0.4681396	-0.5418271
2000.0000000	759.5998535	-80.3000488	-80.8584290	0.5583801	-0.6953667

TABLE LXV

RESULTS OF CORRESPONDING STATES CORRELATION FOR METHANE-PROPANE MIXTURES

USING OPTIMUM PARAMETERS WITH SMOOTH PSEUDOCRITICAL TEMPERATURES

A. 5.1 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	179.5999908	-223.5000000	-223.6318359	0.1318359	-0.0589870
500.0000000	199.5999908	-217.6999969	-217.4304962	-0.2695007	0.1237945
500.0000000	219.5999908	-209.6999969	-211.6576538	1.9576569	-0.9335510
500.0000000	239.5999908	-203.7999573	-205.2513428	1.4513855	-0.7121615
500.0000000	259.5998535	-196.9000092	-198.5638275	1.6638184	-0.8450065
500.0000000	279.5998535	-190.3999634	-191.0784302	0.6784668	-0.3563376
500.0000000	299.5998535	-182.3998566	-183.0291901	0.6293335	-0.3450295
500.0000000	319.5998535	-172.3000031	-172.6023102	0.3023071	-0.1754539
500.0000000	419.5998535	-27.8000488	-26.7770691	-1.0229797	3.6797762
500.0000000	439.5998535	-24.9001465	-24.0481415	-0.8520050	3.4216862
500.0000000	459.5998535	-22.5000000	-21.6835175	-0.8164825	3.6288109
500.0000000	479.5998535	-20.3999023	-19.9320984	-0.4678040	2.2931671
500.0000000	499.5998535	-18.6999512	-18.3113556	-0.3885956	2.0780563
500.0000000	519.5998535	-17.3000488	-16.9819641	-0.3180847	1.8386345
1000.0000000	179.5999908	-221.5000000	-221.8489990	0.3489990	-0.1575616
1000.0000000	199.5999908	-215.4000092	-215.7433777	0.3433685	-0.1594096
1000.0000000	219.5999908	-208.1999969	-210.0404816	1.8404846	-0.8839983
1000.0000000	239.5999908	-202.2999573	-203.9433899	1.6434326	-0.8123741
1000.0000000	259.5998535	-195.5000000	-197.6898346	2.1898346	-1.1201191
1000.0000000	279.5998535	-189.1999512	-190.6936951	1.4937439	-0.7895052
1000.0000000	299.5998535	-182.0998535	-183.1223145	1.0224609	-0.5614836
1000.0000000	319.5998535	-173.6000061	-174.7048492	1.1048431	-0.6364301
1000.0000000	339.5998535	-163.3999023	-164.5803680	1.1804657	-0.7224396
1000.0000000	359.5998535	-149.7998047	-150.9364166	1.1366119	-0.7587537
1000.0000000	379.5998535	-125.4998169	-128.9161835	3.4163666	-2.7222080
1000.0000000	399.5998535	-88.8999023	-89.5667877	0.6668854	-0.7501528
1000.0000000	419.5998535	-68.5000000	-68.3029938	-0.1970062	0.2876003
1000.0000000	439.5998535	-56.8000488	-56.2066345	-0.5934143	1.0447416
1000.0000000	459.5998535	-49.2001953	-48.4727478	-0.7274475	1.4785452
1000.0000000	479.5998535	-43.5000000	-43.0586243	-0.4413757	1.0146561
1000.0000000	499.5998535	-39.0998535	-38.7781219	-0.3217316	0.8228458
1000.0000000	519.5998535	-35.6000977	-35.2352295	-0.3648682	1.0249071
1500.0000000	179.5999908	-219.3999939	-218.0461121	-1.3538818	0.6170835
1500.0000000	199.5999908	-213.1999969	-212.0303650	-1.1696320	0.5486078
1500.0000000	219.5999908	-206.3999939	-206.4029541	0.0029602	-0.0014342
1500.0000000	239.5999908	-200.8999634	-200.5047607	-0.3952026	0.1967161
1500.0000000	259.5998535	-194.3000031	-194.3301086	0.0301056	-0.0154944
1500.0000000	279.5998535	-187.8999634	-187.6579742	-0.2419891	0.1287861
1500.0000000	299.5998535	-181.0998535	-180.6305695	-0.4692841	0.2591299
1500.0000000	319.5998535	-173.7000122	-172.9298706	-0.7701416	0.4433743
1500.0000000	339.5998535	-165.4999084	-164.5115662	-0.9883423	0.5971856
1500.0000000	359.5998535	-155.3998108	-154.4683075	-0.9315033	0.5994234
1500.0000000	379.5998535	-143.0998077	-142.7532196	-0.3465881	0.2422003
1500.0000000	399.5998535	-127.4999084	-127.3916321	-0.1082764	0.0849227
1500.0000000	419.5998535	-108.8000031	-108.2607117	-0.5392914	0.4956718
1500.0000000	439.5998535	-91.1999512	-90.2763367	-0.9236145	1.0127354
1500.0000000	459.5998535	-77.6000977	-76.7010498	-0.8990479	1.1585646
1500.0000000	479.5998535	-67.5998535	-66.9355621	-0.6642914	0.9826816
1500.0000000	499.5998535	-60.0000000	-59.7538757	-0.2461243	0.4102070
1500.0000000	519.5998535	-54.0000000	-53.9082794	-0.0917206	0.1698529
2000.0000000	179.5999908	-217.2999878	-215.7217407	-1.5782471	0.7262986
2000.0000000	199.5999908	-211.1000061	-209.6983643	-1.4016418	0.6639704
2000.0000000	219.5999908	-204.7999878	-204.0653229	-0.7346649	0.3587231
2000.0000000	239.5999908	-199.5999603	-198.1613159	-1.4386444	0.7207636
2000.0000000	259.5998535	-193.1000061	-192.1994476	-0.9005585	0.4663687
2000.0000000	279.5998535	-186.6999512	-185.8863373	-0.8136139	0.4357867
2000.0000000	299.5998535	-180.1998596	-179.3311768	-0.8686829	0.4820660
2000.0000000	319.5998535	-173.2000122	-172.1945801	-1.0054321	0.5805034
2000.0000000	339.5998535	-165.3999023	-164.5603638	-0.8395386	0.5075808
2000.0000000	359.5998535	-156.9998169	-156.0212555	-0.9785614	0.6232880
2000.0000000	379.5998535	-147.4998169	-146.8545380	-0.6452789	0.4374776
2000.0000000	399.5998535	-136.6999054	-136.3205414	-0.3793640	0.2775159
2000.0000000	419.5998535	-124.5000000	-125.2012177	0.7012177	-0.5632270
2000.0000000	439.5998535	-111.5999603	-112.2586365	0.6586761	-0.5902115
2000.0000000	459.5998535	-99.0000000	-99.3920441	0.3920441	-0.3960039
2000.0000000	479.5998535	-87.5998535	-88.1770325	0.5771790	-0.6588809
2000.0000000	499.5998535	-78.0998535	-78.7508850	0.6510315	-0.8335884
2000.0000000	519.5998535	-70.3000488	-70.9214172	0.6213684	-0.8838803

TABLE LXV
(CONTINUED)

B. 11.7 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (^{o}R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	199.5999908	-223.6999969	-223.0068207	-0.6931763	0.3098686
500.0000000	219.5999908	-218.0999908	-217.1113892	-0.9886017	0.4532788
500.0000000	239.5999908	-212.0999146	-211.0460510	-1.0538635	0.4968710
500.0000000	259.5998535	-205.7000122	-204.5604553	-1.1395569	0.5539894
500.0000000	279.5998535	-198.7998657	-197.9233704	-0.8764954	0.4408933
500.0000000	299.5998535	-191.2999115	-190.5954132	-0.7044983	0.3682690
500.0000000	319.5998535	-181.8999634	-182.7186432	0.8186798	-0.4500713
500.0000000	479.5998535	-24.5998535	-22.7081757	-1.8916779	7.6897917
500.0000000	499.5998535	-20.9001465	-20.7657623	-0.1343842	0.6429817
500.0000000	519.5998535	-18.6999512	-19.1062469	0.4062958	-2.1727095
500.0000000	539.5998535	-17.3999023	-17.5858459	0.1859436	-1.0686464
500.0000000	559.5998535	-16.8000488	-16.3319092	-0.4681396	2.7865372
1000.0000000	199.5999908	-220.8999939	-221.2240906	0.3240967	-0.1467164
1000.0000000	219.5999908	-215.0999908	-215.3415680	0.2415771	-0.1123092
1000.0000000	239.5999908	-209.2999115	-209.3639832	0.0640717	-0.0306124
1000.0000000	259.5998535	-203.1000061	-203.1998596	0.0998535	-0.0491647
1000.0000000	279.5998535	-196.2998657	-196.8235321	0.5236664	-0.2667685
1000.0000000	299.5998535	-189.2999115	-189.9216919	0.6217804	-0.3284631
1000.0000000	319.5998535	-181.6999512	-182.4802399	0.7802887	-0.4294376
1000.0000000	339.5998535	-172.9000092	-174.3420105	1.4420013	-0.8340086
1000.0000000	359.5998535	-163.8000031	-164.9284210	1.1284180	-0.6888996
1000.0000000	379.5998535	-150.2999115	-152.6010284	2.3011169	-1.5310163
1000.0000000	479.5998535	-52.8999023	-51.7379303	-1.1619720	2.1965485
1000.0000000	499.5998535	-45.9001465	-45.5902863	-0.3098602	0.6750743
1000.0000000	519.5998535	-40.8000488	-40.8630066	0.0629578	-0.1543080
1000.0000000	539.5998535	-37.0998535	-37.0459137	-0.0539398	0.1453909
1000.0000000	559.5998535	-34.1999512	-33.8877258	-0.3122253	0.9129409
1500.0000000	199.5999908	-218.0000000	-217.8262177	-0.1737823	0.0797166
1500.0000000	219.5999908	-212.1999969	-211.9238281	-0.2761688	0.1301455
1500.0000000	239.5999908	-206.3999023	-206.0637970	-0.3361053	0.1628418
1500.0000000	259.5998535	-200.5000000	-200.1178741	-0.3821259	0.1905864
1500.0000000	279.5998535	-193.8998566	-193.9117432	0.0118866	-0.0061303
1500.0000000	299.5998535	-187.3999023	-187.1718597	-0.2280426	0.1216877
1500.0000000	319.5998535	-180.3999634	-180.2784119	-0.1215515	0.0673789
1500.0000000	339.5998535	-173.0000000	-172.7988586	-0.2011414	0.1162666
1500.0000000	359.5998535	-165.0000000	-164.7543640	-0.2456360	0.1488702
1500.0000000	379.5998535	-155.5999146	-155.5247192	-0.0751953	0.0483260
1500.0000000	399.5998535	-144.3000031	-144.7363434	0.4363403	-0.3023841
1500.0000000	419.5998535	-130.1998596	-131.3392029	1.1393433	-0.8750725
1500.0000000	439.5998535	-113.4998169	-114.7121124	1.2122955	-1.0681028
1500.0000000	459.5998535	-97.5000000	-97.1982269	-0.3017731	0.3095108
1500.0000000	479.5998535	-83.5000000	-82.4818115	-1.0181885	1.2193871
1500.0000000	499.5998535	-72.5000000	-71.5636292	-0.9363708	1.2915449
1500.0000000	519.5998535	-64.1999512	-63.3362732	-0.8636780	1.3452930
1500.0000000	539.5998535	-57.3999023	-56.9979706	-0.4019318	0.7002305
1500.0000000	559.5998535	-52.1000977	-51.7556305	-0.3444672	0.6611637
2000.0000000	199.5999908	-215.2999878	-215.7375946	0.4376068	-0.2032544
2000.0000000	219.5999908	-209.3999939	-209.8743134	0.4743195	-0.2265136
2000.0000000	239.5999908	-203.8999023	-204.0192261	0.1193237	-0.0585207
2000.0000000	259.5998535	-198.0000000	-198.0189514	0.0189514	-0.0095714
2000.0000000	279.5998535	-191.5998535	-192.0462036	0.4463501	-0.2329595
2000.0000000	299.5998535	-185.4999084	-185.6760712	0.1761627	-0.0949664
2000.0000000	319.5998535	-179.2999573	-179.1514740	-0.1484833	0.0828127
2000.0000000	339.5998535	-172.6000061	-172.2033844	-0.3966217	0.2297924
2000.0000000	359.5998535	-165.4000092	-164.8913727	-0.5086365	0.3075190
2000.0000000	379.5998535	-157.3999023	-156.9180145	-0.4818878	0.3061551
2000.0000000	399.5998535	-148.5000000	-148.3316650	-0.1683350	0.1133568
2000.0000000	419.5998535	-138.3998566	-139.0845490	0.6846924	-0.4947204
2000.0000000	439.5998535	-127.0998077	-127.8479462	0.7481384	-0.5886227
2000.0000000	459.5998535	-115.1999512	-115.9387665	0.7388153	-0.6413329
2000.0000000	479.5998535	-103.3000488	-104.1500397	0.8499908	-0.8228365
2000.0000000	499.5998535	-92.5000000	-93.0359802	0.5359802	-0.5794380
2000.0000000	519.5998535	-82.8999023	-83.1945953	0.2946930	-0.3554805
2000.0000000	539.5998535	-74.6999512	-74.8671112	0.1671600	-0.2237753
2000.0000000	559.5998535	-67.8000488	-67.8381500	0.0381012	-0.0561964

TABLE LXV
(CONTINUED)

C. 28.0 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (^{o}R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	219.5999908	-219.6999969	-219.7362061	0.0362091	-0.0164812
500.0000000	239.5999908	-214.3999939	-213.9856720	-0.4143219	0.1932471
500.0000000	259.5998535	-209.0000000	-208.8074036	-0.1925964	0.0921513
500.0000000	279.5998535	-202.8000031	-203.2437439	0.4437408	-0.2188071
500.0000000	299.5998535	-197.2998657	-197.4878082	0.1879425	-0.0952572
500.0000000	319.5998535	-190.5998077	-191.6642914	1.0644836	-0.5584914
500.0000000	339.5998535	-184.1998596	-185.4840851	1.2842255	-0.6971911
500.0000000	519.5998535	-27.0998535	-25.2987976	-1.8010559	6.6459951
500.0000000	539.5998535	-24.3000488	-23.0268860	-1.2731628	5.2393417
500.0000000	559.5998535	-22.1000977	-21.0830688	-1.0170288	4.6019192
500.0000000	579.5998535	-20.3000488	-19.5655518	-0.7344971	3.6182022
500.0000000	599.5998535	-18.7998047	-18.1147308	-0.6850739	3.6440468
500.0000000	619.5998535	-17.3000488	-16.7906189	-0.5094299	2.9446726
500.0000000	639.5998535	-16.1000977	-15.7055397	-0.3945580	2.4506550
1000.0000000	219.5999908	-217.4000092	-217.8384857	0.4384766	-0.2016911
1000.0000000	239.5999908	-212.0999908	-212.2733459	0.1733551	-0.0817327
1000.0000000	259.5998535	-206.8000031	-207.0347900	0.2347870	-0.1135333
1000.0000000	279.5998535	-201.0000000	-201.5647888	0.5647888	-0.2809894
1000.0000000	299.5998535	-195.5998535	-196.0337372	0.4338837	-0.2218220
1000.0000000	319.5998535	-189.4998169	-190.3891144	0.8892975	-0.4692863
1000.0000000	339.5998535	-183.6998596	-184.4388123	0.7389526	-0.4022605
1000.0000000	359.5998535	-177.0000000	-178.0918274	1.0918274	-0.6168514
1000.0000000	379.5998535	-170.3000031	-171.3119202	1.0119171	-0.5941964
1000.0000000	399.5998535	-161.6998596	-163.9705811	2.2707214	-1.4042807
1000.0000000	539.5998535	-55.8000488	-53.8077698	-1.9922791	3.5703888
1000.0000000	559.5998535	-48.6999512	-47.3797913	-1.3201599	2.7108030
1000.0000000	579.5998535	-43.8000488	-42.6066895	-1.1933594	2.7245607
1000.0000000	599.5998535	-39.7998047	-38.6299438	-1.1698608	2.9393625
1000.0000000	619.5998535	-36.3000488	-35.3710785	-0.9289703	2.5591431
1000.0000000	639.5998535	-33.3999023	-32.6143951	-0.7855072	2.3518238
1500.0000000	219.5999908	-215.1000061	-214.9496307	-0.1503754	0.0699095
1500.0000000	239.5999908	-209.8000031	-209.3151093	-0.4848938	0.2311219
1500.0000000	259.5998535	-204.7000122	-204.0634766	-0.6365356	0.3109602
1500.0000000	279.5998535	-199.4000092	-198.6932678	-0.7067413	0.3544339
1500.0000000	299.5998535	-194.0998535	-193.4026489	-0.6972046	0.3591989
1500.0000000	319.5998535	-188.4998169	-187.9442444	-0.5555725	0.2947336
1500.0000000	339.5998535	-182.6998596	-182.0074005	-0.6924591	0.3790146
1500.0000000	359.5998535	-176.5000000	-175.9835968	-0.5164032	0.2925797
1500.0000000	379.5998535	-170.5000000	-169.7929535	-0.7070465	0.4146900
1500.0000000	399.5998535	-163.8998566	-163.0111542	-0.8887024	0.5422227
1500.0000000	419.5998535	-156.3999023	-155.8802948	-0.5196075	0.3322300
1500.0000000	439.5998535	-147.6998138	-147.9449615	0.2451477	-0.1659769
1500.0000000	459.5998535	-137.3998108	-138.6822968	1.2824860	-0.9333968
1500.0000000	479.5998535	-125.3000031	-127.5434113	2.2434082	-1.7904291
1500.0000000	499.5998535	-111.6998138	-114.0764008	2.3765869	-2.1276541
1500.0000000	519.5998535	-98.0998535	-99.8649902	1.7651367	-1.7993259
1500.0000000	539.5998535	-85.9001465	-86.2155304	0.3153839	-0.3671517
1500.0000000	559.5998535	-75.8000488	-75.2431946	-0.5568542	0.7346354
1500.0000000	579.5998535	-67.7001953	-66.7154541	-0.9847412	1.4545612
1500.0000000	599.5998535	-61.0000000	-59.9455566	-1.0544434	1.7285948
1500.0000000	619.5998535	-54.4001465	-54.5367737	0.1366272	-0.2511522
1500.0000000	639.5998535	-49.6999512	-49.9396667	0.2397156	-0.4823253
2000.0000000	219.5999908	-212.8000031	-213.1689758	0.3689728	-0.1733894
2000.0000000	239.5999908	-207.5999908	-207.5888214	-0.0111694	0.0053803
2000.0000000	259.5998535	-202.6000061	-202.2735138	-0.3264923	0.1611511
2000.0000000	279.5998535	-197.7000122	-196.9940338	-0.7059784	0.3570957
2000.0000000	299.5998535	-192.5998535	-191.5930481	-1.0068054	0.5227442
2000.0000000	319.5998535	-187.2998047	-186.3367310	-0.9630737	0.5141880
2000.0000000	339.5998535	-181.8998566	-180.6253357	-1.2745209	0.7006716
2000.0000000	359.5998535	-176.2000122	-175.0912170	-1.1087952	0.6292820
2000.0000000	379.5998535	-170.5000000	-169.1955566	-1.3044434	0.7650692
2000.0000000	399.5998535	-163.7998657	-162.8309326	-0.9689331	0.5915344
2000.0000000	419.5998535	-157.0999146	-156.4238129	-0.6761017	0.4303638
2000.0000000	439.5998535	-149.8998108	-149.4604797	-0.4393311	0.2930831
2000.0000000	459.5998535	-141.8998108	-142.1370239	0.2372131	-0.1671694
2000.0000000	479.5998535	-133.2000122	-134.5644989	1.3644867	-1.0243893
2000.0000000	499.5998535	-123.7998047	-125.3279724	1.5281677	-1.2343855
2000.0000000	519.5998535	-113.7998047	-114.5359497	0.7361450	-0.6468769
2000.0000000	539.5998535	-103.8000488	-104.8703308	1.0702820	-1.0310993
2000.0000000	559.5998535	-94.3000488	-95.3125305	1.0124817	-1.0736799
2000.0000000	579.5998535	-85.8000488	-86.3955078	0.5954590	-0.6940074
2000.0000000	599.5998535	-78.0998535	-78.4010315	0.3011780	-0.3856319
2000.0000000	619.5998535	-71.0000000	-71.4418030	0.4418030	-0.6222576
2000.0000000	639.5998535	-64.8999023	-65.3888245	0.4889221	-0.7533479

TABLE LXV
(CONTINUED)

D. 50.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	259.5998535	-212.7998047	-213.3624878	0.5626831	-0.2644190
500.0000000	279.5998535	-207.8998108	-208.3149261	0.4151154	-0.1996709
500.0000000	299.5998535	-202.8999023	-203.5897675	0.6898651	-0.3400027
500.0000000	319.5998535	-198.2998657	-198.6142731	0.3144073	-0.1585514
500.0000000	339.5998535	-193.1999054	-193.5208435	0.3209381	-0.1661171
500.0000000	359.5998535	-188.0999146	-188.3823090	0.2823944	-0.1501300
500.0000000	579.5998535	-29.7001953	-27.8154144	-1.8847809	6.3460169
500.0000000	599.5998535	-26.5998535	-25.3229980	-1.2768555	4.8002338
500.0000000	619.5998535	-24.1999512	-23.1920776	-1.0078735	4.1647739
500.0000000	639.5998535	-22.2001953	-21.3625031	-0.8376923	3.7733545
500.0000000	659.5998535	-20.6999512	-19.8561401	-0.8438110	4.0763903
500.0000000	679.5998535	-19.2001953	-18.5359192	-0.6642761	3.4597359
500.0000000	699.5998535	-18.0000000	-17.3280334	-0.6719666	3.7331467
500.0000000	719.5998535	-16.8000488	-16.1573486	-0.6427002	3.8255844
500.0000000	739.5998535	-15.7998047	-15.0561409	-0.7436638	4.7067900
1000.0000000	259.5998535	-211.0998077	-211.4761200	0.3763123	-0.1782627
1000.0000000	279.5998535	-206.1998138	-206.5892334	0.3894196	-0.1888554
1000.0000000	299.5998535	-201.3999023	-201.7908478	0.3909454	-0.1941140
1000.0000000	319.5998535	-196.4998627	-196.8105011	0.3106384	-0.1580858
1000.0000000	339.5998535	-191.5999146	-191.8905334	0.2906189	-0.1516801
1000.0000000	359.5998535	-186.6999054	-186.9299469	0.2300415	-0.1232145
1000.0000000	379.5998535	-181.5000000	-181.8016968	0.3016968	-0.1662241
1000.0000000	399.5998535	-176.1999512	-176.5414581	0.3415070	-0.1938178
1000.0000000	419.5998535	-170.7999573	-170.9107666	0.1108093	-0.0648766
1000.0000000	439.5998535	-164.7998047	-165.0961456	0.2963409	-0.1798187
1000.0000000	459.5998535	-158.5998535	-158.9174194	0.3175659	-0.2002308
1000.0000000	599.5998535	-64.5000000	-63.8252411	-0.6747589	1.0461369
1000.0000000	619.5998535	-55.6999512	-55.0355377	-0.6644135	1.1928434
1000.0000000	639.5998535	-49.7001953	-48.8198700	-0.8803253	1.7712708
1000.0000000	659.5998535	-44.8999023	-43.9855652	-0.9143372	2.0363894
1000.0000000	679.5998535	-40.8000488	-40.1219482	-0.6781006	1.6620083
1000.0000000	699.5998535	-37.6000977	-36.7792511	-0.8208466	2.1830959
1000.0000000	719.5998535	-35.1000977	-33.9914856	-1.1086121	3.1584291
1000.0000000	739.5998535	-32.7998047	-31.5339203	-1.2658844	3.8594255
1500.0000000	259.5998535	-209.3998108	-209.1833954	-0.2164154	0.1033503
1500.0000000	279.5998535	-204.4998169	-204.1105042	-0.3893127	0.1903731
1500.0000000	299.5998535	-199.8999023	-199.2641144	-0.6357880	0.3180531
1500.0000000	319.5998535	-194.7998657	-194.4530334	-0.3468323	0.1780454
1500.0000000	339.5998535	-190.0999146	-189.6434021	-0.4565125	0.2401434
1500.0000000	359.5998535	-185.2999115	-184.8253937	-0.4745178	0.2560809
1500.0000000	379.5998535	-180.4000092	-179.9970703	-0.4029388	0.2233585
1500.0000000	399.5998535	-175.1999512	-174.5775757	-0.6223755	0.3552372
1500.0000000	419.5998535	-169.9999542	-169.2830811	-0.7168732	0.4216902
1500.0000000	439.5998535	-164.3998108	-163.9248199	-0.4749908	0.2889242
1500.0000000	459.5998535	-158.7998657	-158.2054443	-0.5944214	0.3743210
1500.0000000	479.5998535	-152.8000031	-152.1599426	-0.6400604	0.4188876
1500.0000000	499.5998535	-146.0999603	-145.6017914	-0.4981689	0.3409781
1500.0000000	519.5998535	-138.2998657	-138.4087219	0.1088562	-0.0787103
1500.0000000	539.5998535	-129.3999634	-130.6119843	1.2120209	-0.9366468
1500.0000000	559.5998535	-119.8000031	-121.2412872	1.4412842	-1.2030745
1500.0000000	579.5998535	-108.5000000	-110.3212433	1.8212433	-1.6785650
1500.0000000	599.5998535	-97.1999512	-99.1135864	1.9136353	-1.9687614
1500.0000000	619.5998535	-86.5998535	-87.7532349	1.1533813	-1.3318510
1500.0000000	639.5998535	-77.5000000	-77.5964966	0.0964966	-0.1245117
1500.0000000	659.5998535	-69.3000488	-69.3953552	0.0953064	-0.1375272
1500.0000000	679.5998535	-63.0000000	-62.7524719	-0.2475281	0.3929015
1500.0000000	699.5998535	-57.6999512	-57.1692352	-0.5307159	0.9197857
1500.0000000	719.5998535	-53.1000977	-52.4980316	-0.6020660	1.1338310
1500.0000000	739.5998535	-49.3999023	-48.4240875	-0.9758148	1.9753370
2000.0000000	259.5998535	-207.5998077	-207.7725830	0.1727753	-0.0832251
2000.0000000	279.5998535	-202.9998169	-202.8385620	-0.1612549	0.0794359
2000.0000000	299.5998535	-197.9999084	-197.9224548	-0.0774536	0.0391180
2000.0000000	319.5998535	-193.2998657	-193.0569916	-0.2428741	0.1256462
2000.0000000	339.5998535	-188.4999084	-188.2287598	-0.2711487	0.1438455
2000.0000000	359.5998535	-183.6999054	-183.5183105	-0.1815948	0.0988541
2000.0000000	379.5998535	-178.8000031	-178.6380920	-0.1619110	0.0905542
2000.0000000	399.5998535	-173.8999634	-173.6020966	-0.2978668	0.1712863
2000.0000000	419.5998535	-168.9999542	-168.7507019	-0.2492523	0.1474866
2000.0000000	439.5998535	-163.8998108	-163.5073700	-0.3924408	0.2394394
2000.0000000	459.5998535	-158.4998627	-158.0904541	-0.4094086	0.2583022
2000.0000000	479.5998535	-153.0000000	-152.6328430	-0.3671570	0.2399718
2000.0000000	499.5998535	-147.1999512	-146.6652832	-0.5346680	0.3632256
2000.0000000	519.5998535	-140.8998566	-140.9657898	0.0659332	-0.0467944
2000.0000000	539.5998535	-134.0999603	-134.6620483	0.5620880	-0.4191558
2000.0000000	559.5998535	-126.9000092	-128.4300385	1.5300293	-1.2056961
2000.0000000	579.5998535	-119.0000000	-120.6824341	1.6824341	-1.4138098
2000.0000000	599.5998535	-110.7998047	-110.8658600	0.0660553	-0.0596168
2000.0000000	619.5998535	-102.6999512	-102.6687012	-0.0312500	0.0304284
2000.0000000	639.5998535	-94.6000977	-95.1734467	0.5733490	-0.6060764
2000.0000000	659.5998535	-86.8999023	-87.6919403	0.7920380	-0.9114370
2000.0000000	679.5998535	-80.0000000	-80.5114899	0.5114899	-0.6393622
2000.0000000	699.5998535	-73.8999023	-74.1491852	0.2492828	-0.3373249
2000.0000000	719.5998535	-68.4001465	-68.3155670	-0.0845795	0.1236539
2000.0000000	739.5998535	-63.5998535	-63.0513611	-0.5484924	0.8624114

TABLE LXV
(CONTINUED)

E. 76.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	259.5998535	-215.0998535	-215.9630890	0.8632355	-0.4013184
500.0000000	279.5998535	-210.4998169	-211.0289612	0.5291443	-0.2513751
500.0000000	299.5998535	-205.9999542	-206.1390533	0.1390991	-0.0675238
500.0000000	319.5998535	-201.4999542	-201.6459351	0.1459808	-0.0724471
500.0000000	339.5998535	-196.9998169	-197.3194733	0.3196564	-0.1622622
500.0000000	359.5998535	-192.6998596	-192.8189240	0.1190643	-0.0617874
500.0000000	379.5998535	-188.0998535	-188.2440186	0.1441650	-0.0766428
500.0000000	399.5998535	-183.5998077	-183.7898560	0.1900482	-0.1035122
500.0000000	419.5998535	-178.8998566	-179.1966705	0.2968140	-0.1659107
500.0000000	439.5998535	-174.2999573	-174.4201050	0.1201477	-0.0689315
500.0000000	459.5998535	-169.3999023	-169.8145447	0.4146423	-0.2447712
500.0000000	619.5998535	-35.1000977	-34.3142700	-0.7858276	2.2388182
500.0000000	639.5998535	-31.0000000	-30.3339081	-0.6660919	2.1486826
500.0000000	659.5998535	-27.8999023	-27.3220062	-0.5778961	2.0713186
500.0000000	679.5998535	-25.3999023	-24.9154816	-0.4844208	1.9071751
500.0000000	699.5998535	-23.3999023	-22.9706421	-0.4292603	1.8344526
500.0000000	719.5998535	-21.6999512	-21.4063873	-0.2935638	1.3528309
500.0000000	739.5998535	-19.9001465	-19.8648071	-0.0353394	0.1775833
500.0000000	759.5998535	-17.6999512	-18.6975403	0.9975891	-5.6361113
1000.0000000	259.5998535	-213.0998535	-214.0125885	0.9127350	-0.4283130
1000.0000000	279.5998535	-208.3998108	-209.0690460	0.6692352	-0.3211304
1000.0000000	299.5998535	-204.0999603	-204.3749390	0.2749786	-0.1347274
1000.0000000	319.5998535	-199.6999512	-199.9977875	0.2978363	-0.1491418
1000.0000000	339.5998535	-195.2998047	-195.6057587	0.3059540	-0.1566586
1000.0000000	359.5998535	-190.8998566	-191.0912170	0.1913605	-0.1002412
1000.0000000	379.5998535	-186.5998535	-186.6613770	0.0615234	-0.0329708
1000.0000000	399.5998535	-182.2998047	-182.1132355	-0.1865692	0.1023419
1000.0000000	419.5998535	-177.7998657	-177.6692047	-0.1306610	0.0734876
1000.0000000	439.5998535	-173.5999603	-173.0776520	-0.5223083	0.3008689
1000.0000000	459.5998535	-169.2999115	-168.3477173	-0.9521942	0.5624301
1000.0000000	479.5998535	-164.3999023	-163.3701935	-1.0297089	0.6263438
1000.0000000	499.5998535	-159.2998657	-158.3547974	-0.9450684	0.5932636
1000.0000000	519.5998535	-153.7000122	-152.9866638	-0.7133484	0.4641171
1000.0000000	539.5998535	-148.0999603	-147.1595001	-0.9404602	0.6350170
1000.0000000	559.5998535	-141.7998657	-140.7395935	-1.0602722	0.7477242
1000.0000000	579.5998535	-134.2999573	-133.4069977	-0.8929596	0.6648991
1000.0000000	599.5998535	-125.1998596	-124.8759003	-0.3239594	0.2587537
1000.0000000	619.5998535	-112.6000977	-113.3898468	0.7897491	-0.7013749
1000.0000000	639.5998535	-94.4001465	-96.4388885	2.0387421	-2.1596804
1000.0000000	659.5998535	-77.0998535	-77.0856628	-0.0141907	0.0184056
1000.0000000	679.5998535	-65.3000488	-63.9251556	-1.3748932	2.1055002
1000.0000000	699.5998535	-55.8000488	-55.4507446	-0.3493042	0.6259925
1000.0000000	719.5998535	-49.5000000	-49.4138489	-0.0861511	0.1740426
1000.0000000	739.5998535	-44.8000488	-44.8408203	0.0407715	-0.0910076
1000.0000000	759.5998535	-41.8999023	-40.9667664	-0.9331360	2.2270594
1500.0000000	259.5998535	-211.2998657	-211.9204712	0.6206055	-0.2937084
1500.0000000	279.5998535	-206.8998108	-207.2250214	0.3252106	-0.1571826
1500.0000000	299.5998535	-202.5999603	-202.5259705	-0.0739899	0.0365202
1500.0000000	319.5998535	-198.0999603	-197.9289093	-0.1710510	0.0863458
1500.0000000	339.5998535	-193.8998108	-193.4789581	-0.4208527	0.2170464
1500.0000000	359.5998535	-189.3998566	-189.1691284	-0.2307281	0.1218206
1500.0000000	379.5998535	-185.2998657	-184.7435455	-0.5563202	0.3002269
1500.0000000	399.5998535	-180.7998047	-180.5327148	-0.2670898	0.1477268
1500.0000000	419.5998535	-176.2998657	-176.1721954	-0.1276703	0.0724165
1500.0000000	439.5998535	-171.9999542	-171.7151031	-0.2848511	0.1656111
1500.0000000	459.5998535	-167.4999084	-166.8757019	-0.6242065	0.3726608
1500.0000000	479.5998535	-162.9999084	-162.2016907	-0.7982178	0.4897043
1500.0000000	499.5998535	-158.1998596	-157.5055847	-0.6942749	0.4388593
1500.0000000	519.5998535	-153.1000061	-152.5071259	-0.5928802	0.3872503
1500.0000000	539.5998535	-148.1999512	-147.2438507	-0.9561005	0.6451420
1500.0000000	559.5998535	-142.7998657	-141.9967957	-0.8030701	0.5623743
1500.0000000	579.5998535	-136.6999512	-136.2226715	-0.4772797	0.3491440
1500.0000000	599.5998535	-130.1998596	-129.7465363	-0.4533234	0.3481750
1500.0000000	619.5998535	-123.1999512	-122.8667450	-0.3332062	0.2704597
1500.0000000	639.5998535	-115.0000000	-115.0187378	0.0187378	-0.0162937
1500.0000000	659.5998535	-106.5998535	-106.0254669	-0.5743866	0.5388249
1500.0000000	679.5998535	-97.0000000	-96.6218567	-0.3781433	0.3898384
1500.0000000	699.5998535	-86.8000488	-87.2585297	0.4584808	-0.5282033
1500.0000000	719.5998535	-77.5000000	-78.3973999	0.8973999	-1.1579351
1500.0000000	739.5998535	-70.3000488	-70.8262329	0.5261841	-0.7484831
1500.0000000	759.5998535	-63.3999023	-64.3321075	0.9322052	-1.4703569
2000.0000000	259.5998535	-209.4998627	-210.9001007	1.4002380	-0.6683714
2000.0000000	279.5998535	-205.1998138	-206.0324097	0.8325958	-0.4057485
2000.0000000	299.5998535	-200.8999634	-201.4397278	0.5397644	-0.2686732
2000.0000000	319.5998535	-196.4999542	-197.0929565	0.5930023	-0.3017824
2000.0000000	339.5998535	-192.3998108	-192.5046082	0.1047974	-0.0544685
2000.0000000	359.5998535	-188.1998596	-188.0214539	-0.1784058	0.0947959
2000.0000000	379.5998535	-184.0998535	-183.7349091	-0.3649445	0.1982318
2000.0000000	399.5998535	-179.8998108	-179.3813171	-0.5184937	0.2882124
2000.0000000	419.5998535	-175.6998596	-175.1452789	-0.5545807	0.3156409
2000.0000000	439.5998535	-171.1999512	-170.6218872	-0.5780640	0.3376542
2000.0000000	459.5998535	-166.6999054	-166.2076416	-0.4922638	0.2952994
2000.0000000	479.5998535	-160.0999146	-161.8976135	1.7976990	-1.1228600
2000.0000000	499.5998535	-157.5998535	-157.1539154	-0.4459381	0.2829559
2000.0000000	519.5998535	-153.1000061	-152.4760895	-0.6239166	0.4075221
2000.0000000	539.5998535	-148.4999542	-147.6445465	-0.8554077	0.5760319
2000.0000000	559.5998535	-143.5998535	-142.8975525	-0.7023010	0.4890680
2000.0000000	579.5998535	-138.4999542	-137.5681305	-0.9318237	0.6727971
2000.0000000	599.5998535	-133.0998535	-132.6048889	-0.4949646	0.3718746
2000.0000000	619.5998535	-127.1999512	-126.8992615	-0.3006897	0.2363913
2000.0000000	639.5998535	-121.0000000	-121.2071686	0.2071686	-0.1712137
2000.0000000	659.5998535	-114.6999512	-114.6870728	-0.0128784	0.0112279
2000.0000000	679.5998535	-107.6000977	-106.9406891	-0.6594086	0.6128326
2000.0000000	699.5998535	-100.3999023	-99.7528687	-0.6470337	0.6444562
2000.0000000	719.5998535	-93.3000488	-93.3932648	0.0932159	-0.0999098
2000.0000000	739.5998535	-86.4001465	-86.9765625	0.5764160	-0.6671469
2000.0000000	759.5998535	-80.3000488	-80.9935608	0.6935120	-0.8636504

TABLE LXVI

RESULTS OF CORRESPONDING STATES CORRELATION FOR METHANE-PROPANE MIXTURES

USING SMOOTHED OPTIMUM PARAMETERS

A. 5.1 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	179.5999908	-223.5000000	-226.9316864	3.4316864	-1.5354290
500.0000000	199.5999908	-217.6999969	-220.4833527	2.7833557	-1.2785273
500.0000000	219.5999908	-209.6999969	-214.4564667	4.7564697	-2.2682257
500.0000000	239.5999908	-203.7999573	-207.8097839	4.0098267	-1.9675303
500.0000000	259.5998535	-196.9000092	-200.9120026	4.0119934	-2.0375786
500.0000000	279.5998535	-190.3999634	-193.2441254	2.8441620	-1.4937820
500.0000000	299.5998535	-182.3998566	-184.9583282	2.5584717	-1.4026709
500.0000000	319.5998535	-172.3000031	-174.3747559	2.0747528	-1.2041512
500.0000000	419.5998535	-27.8000488	-27.0675659	-0.7324829	2.6348257
500.0000000	439.5998535	-24.9001465	-24.2700500	-0.6300964	2.5304928
500.0000000	459.5998535	-22.5000000	-21.8718262	-0.6281738	2.7918825
500.0000000	479.5998535	-20.3999023	-20.0746918	-0.3252106	1.5941763
500.0000000	499.5998535	-18.6999512	-18.4159088	-0.2840424	1.5189466
500.0000000	519.5998535	-17.3000488	-17.0569916	-0.2430573	1.4049511
1000.0000000	179.5999908	-221.5000000	-225.1002960	3.6002960	-1.6254148
1000.0000000	199.5999908	-215.4000092	-218.7625427	3.3625336	-1.5610638
1000.0000000	219.5999908	-208.1999969	-212.7849884	4.5849915	-2.2022047
1000.0000000	239.5999908	-202.2999573	-206.4502258	4.1502686	-2.0515413
1000.0000000	259.5998535	-195.5000000	-199.9441681	4.4441681	-2.2732315
1000.0000000	279.5998535	-189.1999512	-192.7454071	3.5454559	-1.8739195
1000.0000000	299.5998535	-182.0998535	-184.9430542	2.8432007	-1.5613413
1000.0000000	319.5998535	-173.6000061	-176.3085480	2.7085419	-1.5602188
1000.0000000	339.5998535	-163.3999023	-165.9652100	2.5653076	-1.5699558
1000.0000000	359.5998535	-149.7998047	-152.1031342	2.3033295	-1.5376043
1000.0000000	379.5998535	-125.4998169	-129.8916321	4.3918152	-3.4994583
1000.0000000	399.5998535	-88.8999023	-90.4928894	1.5929871	-1.7918873
1000.0000000	419.5998535	-68.5000000	-68.8716431	0.3716431	-0.5425445
1000.0000000	439.5998535	-56.8000488	-56.7012329	-0.0988159	0.1739715
1000.0000000	459.5998535	-49.2001953	-48.8535004	-0.3466949	0.7046614
1000.0000000	479.5998535	-43.5000000	-43.3674469	-0.1325531	0.3047197
1000.0000000	499.5998535	-39.0998535	-38.9932098	-0.1066437	0.2727470
1000.0000000	519.5998535	-35.6000977	-35.4002533	-0.1998444	0.5613587
1500.0000000	179.5999908	-219.3999939	-221.3976746	1.9976807	-0.9105198
1500.0000000	199.5999908	-213.1999969	-215.1211853	1.9211884	-0.9011201
1500.0000000	219.5999908	-206.3999939	-209.2149506	2.8149567	-1.3638353
1500.0000000	239.5999908	-200.8999634	-203.0811768	2.1812134	-1.0857201
1500.0000000	259.5998535	-194.3000031	-196.6875610	2.3875580	-1.2287989
1500.0000000	279.5998535	-187.8999634	-189.7619171	1.8619537	-0.9909279
1500.0000000	299.5998535	-181.0998535	-182.5285339	1.4286804	-0.7888909
1500.0000000	319.5998535	-173.7000122	-174.6191101	0.9190979	-0.5291294
1500.0000000	339.5998535	-165.4999084	-165.9848785	0.4849701	-0.2930334
1500.0000000	359.5998535	-155.3998108	-155.7523804	0.3525696	-0.2268790
1500.0000000	379.5998535	-143.0998077	-143.7893066	0.6894989	-0.4818305
1500.0000000	399.5998535	-127.4999084	-128.1256104	0.6257019	-0.4907466
1500.0000000	419.5998535	-108.8000031	-109.0628204	0.2628174	-0.2415600
1500.0000000	439.5998535	-91.1999512	-90.9673615	-0.2325897	0.2550327
1500.0000000	459.5998535	-77.6000977	-77.2834015	-0.3166962	0.4081130
1500.0000000	479.5998535	-67.5998535	-67.3894501	-0.2104034	0.3112483
1500.0000000	499.5998535	-60.0000000	-60.0946655	0.0946655	-0.1577758
1500.0000000	519.5998535	-54.0000000	-54.1584320	0.1584320	-0.2933925
2000.0000000	179.5999908	-217.2999878	-219.1347809	1.8347931	-0.8443594
2000.0000000	199.5999908	-211.1000061	-212.8544922	1.7544861	-0.8311160
2000.0000000	219.5999908	-204.7999878	-206.9452972	2.1453094	-1.0475140
2000.0000000	239.5999908	-199.5999603	-200.8016663	1.2017059	-0.6020568
2000.0000000	259.5998535	-193.1000061	-194.6253204	1.5253143	-0.7899086
2000.0000000	279.5998535	-186.6999512	-188.0641937	1.3642426	-0.7307138
2000.0000000	299.5998535	-180.1998596	-181.2850037	1.0851440	-0.6021891
2000.0000000	319.5998535	-173.2000122	-173.9479065	0.7478943	-0.4318095
2000.0000000	339.5998535	-165.3999023	-166.1320038	0.7321014	-0.4426248
2000.0000000	359.5998535	-156.9998169	-157.4232025	0.4233856	-0.2696726
2000.0000000	379.5998535	-147.4998169	-148.1235962	0.6237793	-0.4229016
2000.0000000	399.5998535	-136.6999054	-137.5641785	0.8642731	-0.6322410
2000.0000000	419.5998535	-124.5000000	-125.6990814	1.1990814	-0.9631175
2000.0000000	439.5998535	-111.5999603	-112.7542572	1.1542969	-1.0343161
2000.0000000	459.5998535	-99.0000000	-99.9122620	0.9122620	-0.9214766
2000.0000000	479.5998535	-87.5998535	-88.6198578	1.0200043	-1.1643896
2000.0000000	499.5998535	-78.0998535	-79.1157684	1.0159149	-1.3007889
2000.0000000	519.5998535	-70.3000488	-71.1804657	0.8804169	-1.2523699

TABLE LXVI
(CONTINUED)

B. 11.7 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (^oR)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	199.5999908	-223.6999969	-221.9612732	-1.7387238	0.7772568
500.0000000	219.5999908	-218.0999908	-216.1458282	-1.9541626	0.8959938
500.0000000	239.5999908	-212.0999146	-210.1582794	-1.9416351	0.9154338
500.0000000	259.5998535	-205.7000122	-203.7401886	-1.9598236	0.9527579
500.0000000	279.5998535	-198.7998657	-197.1702118	-1.6296539	0.8197457
500.0000000	299.5998535	-191.2999115	-189.8962708	-1.4036407	0.7337380
500.0000000	319.5998535	-181.8999634	-182.0914917	0.1915283	-0.1052932
500.0000000	479.5998535	-24.5998535	-22.6095428	-1.9903107	8.0907402
500.0000000	499.5998535	-20.9001465	-20.6689606	-0.2311859	1.1061440
500.0000000	519.5998535	-18.6999512	-19.0365601	0.3366089	-1.8000517
500.0000000	539.5998535	-17.3999023	-17.5312042	0.1313019	-0.7546127
500.0000000	559.5998535	-16.8000488	-16.2901306	-0.5099182	3.0352182
1000.0000000	199.5999908	-220.8999939	-220.2052917	-0.6947021	0.3144872
1000.0000000	219.5999908	-215.0999908	-214.4022217	-0.6977692	0.3243929
1000.0000000	239.5999908	-209.2999115	-208.5052032	-0.7947083	0.3796983
1000.0000000	259.5998535	-203.1000061	-202.4145660	-0.6854401	0.3374889
1000.0000000	279.5998535	-196.2998657	-196.1113892	-0.1884766	0.0960146
1000.0000000	299.5998535	-189.2999115	-189.2724762	-0.0274353	0.0144930
1000.0000000	319.5998535	-181.6999512	-181.8969421	0.1969910	-0.1084155
1000.0000000	339.5998535	-172.9000092	-173.8210297	0.9210205	-0.5326893
1000.0000000	359.5998535	-163.8000031	-164.4685669	0.6685638	-0.4081584
1000.0000000	379.5998535	-150.2999115	-152.1975861	1.8976746	-1.2625914
1000.0000000	479.5998535	-52.8999023	-51.5134277	-1.3864746	2.6209393
1000.0000000	499.5998535	-45.9001465	-45.4022675	-0.4978790	1.0846996
1000.0000000	519.5998535	-40.8000488	-40.7203217	-0.0797272	0.1954095
1000.0000000	539.5998535	-37.0998535	-36.9300842	-0.1697693	0.4576008
1000.0000000	559.5998535	-34.1999512	-33.7904205	-0.4095306	1.1974592
1500.0000000	199.5999908	-218.0000000	-216.7839050	-1.2160950	0.5578414
1500.0000000	219.5999908	-212.1999969	-210.9642487	-1.2357483	0.5823504
1500.0000000	239.5999908	-206.3999023	-205.1847992	-1.2151031	0.5887128
1500.0000000	259.5998535	-200.5000000	-199.3072205	-1.1927795	0.5949024
1500.0000000	279.5998535	-193.8998566	-193.1636047	-0.7362518	0.3797072
1500.0000000	299.5998535	-187.3999023	-186.5019379	-0.8979645	0.4791699
1500.0000000	319.5998535	-180.3999634	-179.6705322	-0.7294312	0.4043407
1500.0000000	339.5998535	-173.0000000	-172.2507935	-0.7492065	0.4330672
1500.0000000	359.5998535	-165.0000000	-164.2658386	-0.7341614	0.4449461
1500.0000000	379.5998535	-155.5999146	-155.0887146	-0.5112000	0.3285348
1500.0000000	399.5998535	-144.3000031	-144.3593597	0.0593567	-0.0411342
1500.0000000	419.5998535	-130.1998596	-131.0386200	0.8387604	-0.6442096
1500.0000000	439.5998535	-113.4998169	-114.3671722	0.8673553	-0.7641907
1500.0000000	459.5998535	-97.5000000	-96.8395691	-0.6604309	0.6773646
1500.0000000	479.5998535	-83.5000000	-82.1640015	-1.3359985	1.5999975
1500.0000000	499.5998535	-72.5000000	-71.2907867	-1.2092133	1.6678801
1500.0000000	519.5998535	-64.1999512	-63.1142578	-1.0856934	1.6911116
1500.0000000	539.5998535	-57.3999023	-56.8190460	-0.5808563	1.0119457
1500.0000000	559.5998535	-52.1000977	-51.6108551	-0.4892426	0.9390432
2000.0000000	199.5999908	-215.2999878	-214.6734314	-0.6265564	0.2910155
2000.0000000	219.5999908	-209.3999939	-208.8889313	-0.5110626	0.2440605
2000.0000000	239.5999908	-203.8999023	-203.1128387	-0.7870636	0.3860049
2000.0000000	259.5998535	-198.0000000	-197.1883240	-0.8116760	0.4099373
2000.0000000	279.5998535	-191.5998535	-191.2792053	-0.3206482	0.1673530
2000.0000000	299.5998535	-185.4999084	-184.9792328	-0.5206757	0.2806878
2000.0000000	319.5998535	-179.2999573	-178.5186462	-0.7813110	0.4357561
2000.0000000	339.5998535	-172.6000061	-171.6310883	-0.9689178	0.5613658
2000.0000000	359.5998535	-165.4000092	-164.3647919	-1.0352173	0.6258867
2000.0000000	379.5998535	-157.3999023	-156.4539337	-0.9459686	0.6009966
2000.0000000	399.5998535	-148.5000000	-147.8839111	-0.6160889	0.4148744
2000.0000000	419.5998535	-138.3998566	-138.5615234	0.1616669	-0.1168114
2000.0000000	439.5998535	-127.0998077	-127.5689240	0.4691162	-0.3690928
2000.0000000	459.5998535	-115.1999512	-115.7554321	0.5554810	-0.4821882
2000.0000000	479.5998535	-103.3000488	-103.9051056	0.6050568	-0.5857274
2000.0000000	499.5998535	-92.5000000	-92.7783813	0.2783813	-0.3009528
2000.0000000	519.5998535	-82.8999023	-82.9710693	0.0711670	-0.0858468
2000.0000000	539.5998535	-74.6999512	-74.6830292	-0.0169220	0.0226533
2000.0000000	559.5998535	-67.8000488	-67.6909790	-0.1090698	0.1608698

TABLE LXVI
(CONTINUED)

C. 28.0 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	219.5999908	-219.6999969	-220.2863007	0.5863037	-0.2668656
500.0000000	239.5999908	-214.3999939	-214.5048828	0.1048889	-0.0489220
500.0000000	259.5998535	-209.0000000	-209.2912445	0.2912445	-0.1393514
500.0000000	279.5998535	-202.8000031	-203.6928406	0.8928375	-0.4402552
500.0000000	299.5998535	-197.2998657	-197.9077148	0.6078491	-0.3080838
500.0000000	319.5998535	-190.5998077	-192.0535431	1.4537354	-0.7627159
500.0000000	339.5998535	-184.1998596	-185.8499756	1.6501160	-0.8958287
500.0000000	519.5998535	-27.0998535	-25.3669434	-1.7329102	6.3945351
500.0000000	539.5998535	-24.3000488	-23.0831146	-1.2169342	5.0079489
500.0000000	559.5998535	-22.1000977	-21.1312561	-0.9688416	4.3838787
500.0000000	579.5998535	-20.3000488	-19.6109467	-0.6891022	3.3945827
500.0000000	599.5998535	-18.7998047	-18.1495972	-0.6502075	3.4585857
500.0000000	619.5998535	-17.3000488	-16.8185883	-0.4814606	2.7830009
500.0000000	639.5998535	-16.1000977	-15.7279387	-0.3721590	2.3115320
1000.0000000	219.5999908	-217.4000092	-218.3764343	0.9764252	-0.4491374
1000.0000000	239.5999908	-212.0999908	-212.7821503	0.6821594	-0.3216216
1000.0000000	259.5998535	-206.8000031	-207.5072327	0.7072296	-0.3419872
1000.0000000	279.5998535	-201.0000000	-202.0022736	1.0022736	-0.4986435
1000.0000000	299.5998535	-195.5998535	-196.4371185	0.8372650	-0.4280496
1000.0000000	319.5998535	-189.4998169	-190.7607880	1.2609711	-0.6654207
1000.0000000	339.5998535	-183.6998596	-184.7817688	1.0819092	-0.5889546
1000.0000000	359.5998535	-177.0000000	-178.4043274	1.4043274	-0.7934052
1000.0000000	379.5998535	-170.3000031	-171.5980377	1.2980347	-0.7622044
1000.0000000	399.5998535	-161.6998596	-164.2261353	2.5262756	-1.5623236
1000.0000000	539.5998535	-55.8000488	-53.9281158	-1.8719330	3.3547153
1000.0000000	559.5998535	-48.6999512	-47.4826508	-1.2173004	2.4995918
1000.0000000	579.5998535	-43.8000488	-42.6932068	-1.1068420	2.5270329
1000.0000000	599.5998535	-39.7998047	-38.6974792	-1.1023254	2.7696753
1000.0000000	619.5998535	-36.3000488	-35.4270477	-0.8730011	2.4049578
1000.0000000	639.5998535	-33.3999023	-32.6619110	-0.7379913	2.2095604
1500.0000000	219.5999908	-215.1000061	-215.5056000	0.4055939	-0.1885605
1500.0000000	239.5999908	-209.8000031	-209.8356171	0.0356140	-0.0169752
1500.0000000	259.5998535	-204.7000122	-204.5474091	-0.1526031	0.0745496
1500.0000000	279.5998535	-199.4000092	-199.1423035	-0.2577057	0.1292405
1500.0000000	299.5998535	-194.0998535	-193.8211975	-0.2786560	0.1435632
1500.0000000	319.5998535	-188.4998169	-188.3348083	-0.1650085	0.0875377
1500.0000000	339.5998535	-182.6998596	-182.3637543	-0.3361053	0.1839658
1500.0000000	359.5998535	-176.5000000	-176.3096313	-0.1903687	0.1078576
1500.0000000	379.5998535	-170.5000000	-170.0927734	-0.4072266	0.2388425
1500.0000000	399.5998535	-163.8998566	-163.2828827	-0.6169739	0.3764334
1500.0000000	419.5998535	-156.3999023	-156.1258392	-0.2740631	0.1752322
1500.0000000	439.5998535	-147.6998138	-148.1684418	0.4686279	-0.3172840
1500.0000000	459.5998535	-137.3998108	-138.8779907	1.4781799	-1.0758228
1500.0000000	479.5998535	-125.3000031	-127.7117615	2.4117584	-1.9247866
1500.0000000	499.5998535	-111.6998138	-114.2397308	2.5399170	-2.2738771
1500.0000000	519.5998535	-98.0998535	-100.0403748	1.9405212	-1.9781075
1500.0000000	539.5998535	-85.9001465	-86.3804779	0.4803314	-0.5591739
1500.0000000	559.5998535	-75.8000488	-75.3906403	-0.4094086	0.5401164
1500.0000000	579.5998535	-67.7001953	-66.8442688	-0.8559265	1.2642889
1500.0000000	599.5998535	-61.0000000	-60.0529022	-0.9470978	1.5526190
1500.0000000	619.5998535	-54.4001465	-54.6243286	0.2241821	-0.4120983
1500.0000000	639.5998535	-49.6999512	-50.0119781	0.3120270	-0.6278213
2000.0000000	219.5999908	-212.8000031	-213.7352905	0.9352875	-0.4395146
2000.0000000	239.5999908	-207.5999908	-208.1239777	0.5239868	-0.2524021
2000.0000000	259.5998535	-202.6000061	-202.7685394	0.1685333	-0.0831852
2000.0000000	279.5998535	-197.7000122	-197.4569550	-0.2430573	0.1229424
2000.0000000	299.5998535	-192.5998535	-192.0215454	-0.5783081	0.3002640
2000.0000000	319.5998535	-187.2998047	-186.7361603	-0.5636444	0.3009316
2000.0000000	339.5998535	-181.8998566	-180.9917603	-0.9080963	0.4992284
2000.0000000	359.5998535	-176.2000122	-175.4336548	-0.7663574	0.4349358
2000.0000000	379.5998535	-170.5000000	-169.5098724	-0.9901276	0.5807198
2000.0000000	399.5998535	-163.7998657	-163.1153717	-0.6844940	0.4178841
2000.0000000	419.5998535	-157.0999146	-156.6856079	-0.4143066	0.2637217
2000.0000000	439.5998535	-149.8998108	-149.6936951	-0.2061157	0.1375023
2000.0000000	459.5998535	-141.8998108	-142.3596497	0.4598389	-0.3240588
2000.0000000	479.5998535	-133.2000122	-134.8022614	1.6022491	-1.2028894
2000.0000000	499.5998535	-123.7998047	-125.5167084	1.7169037	-1.3868380
2000.0000000	519.5998535	-113.7998047	-114.6369171	0.8371124	-0.7356007
2000.0000000	539.5998535	-103.8000488	-104.9794312	1.1793823	-1.1362057
2000.0000000	559.5998535	-94.3000488	-95.4340515	1.1340027	-1.2025471
2000.0000000	579.5998535	-85.8000488	-86.5139618	0.7139130	-0.8320655
2000.0000000	599.5998535	-78.0998535	-78.5068054	0.4069519	-0.5210660
2000.0000000	619.5998535	-71.0000000	-71.5308990	0.5308990	-0.7477451
2000.0000000	639.5998535	-64.8999023	-65.4637146	0.5638123	-0.8687410

TABLE LXVI
(CONTINUED)

D. 50.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	259.5998535	-212.7998047	-213.6822662	0.8824615	-0.4146907
500.0000000	279.5998535	-207.8998108	-208.6192322	0.7194214	-0.3460423
500.0000000	299.5998535	-202.8999023	-203.8759460	0.9760437	-0.4810467
500.0000000	319.5998535	-198.2998657	-198.8828430	0.5829773	-0.2939877
500.0000000	339.5998535	-193.1999054	-193.7728424	0.5729370	-0.2965513
500.0000000	359.5998535	-188.0999146	-188.6192627	0.5193481	-0.2761022
500.0000000	579.5998535	-29.7001953	-27.8566132	-1.8435822	6.2073059
500.0000000	599.5998535	-26.5998535	-25.3582458	-1.2416077	4.6677227
500.0000000	619.5998535	-24.1999512	-23.2206879	-0.9792633	4.0465498
500.0000000	639.5998535	-22.2001953	-21.3867645	-0.8134308	3.6640701
500.0000000	659.5998535	-20.6999512	-19.8788452	-0.8211060	3.9667044
500.0000000	679.5998535	-19.2001953	-18.5556488	-0.6445465	3.3569784
500.0000000	699.5998535	-18.0000000	-17.3444824	-0.6555176	3.6417637
500.0000000	719.5998535	-16.8000488	-16.1705322	-0.6295166	3.7471113
500.0000000	739.5998535	-15.7998047	-15.0652752	-0.7345295	4.6489782
1000.0000000	259.5998535	-211.0998077	-211.7926483	0.6928406	-0.3282052
1000.0000000	279.5998535	-206.1998138	-206.8906860	0.6908722	-0.3350498
1000.0000000	299.5998535	-201.3999023	-202.0738525	0.6739502	-0.3346328
1000.0000000	319.5998535	-196.4998627	-197.0742798	0.5744171	-0.2923244
1000.0000000	339.5998535	-191.5999146	-192.1378326	0.5379181	-0.2807506
1000.0000000	359.5998535	-186.6999054	-187.1615143	0.4616089	-0.2472464
1000.0000000	379.5998535	-181.5000000	-182.0179138	0.5179138	-0.2853519
1000.0000000	399.5998535	-176.1999512	-176.7419586	0.5420074	-0.3076093
1000.0000000	419.5998535	-170.7999573	-171.0954132	0.2954559	-0.1729836
1000.0000000	439.5998535	-164.7998047	-165.2679443	0.4681396	-0.2840656
1000.0000000	459.5998535	-158.5998535	-159.0746307	0.4747772	-0.2993554
1000.0000000	599.5998535	-64.5000000	-63.8943329	-0.6056671	0.9390186
1000.0000000	619.5998535	-55.6999512	-55.0928802	-0.6070709	1.0898943
1000.0000000	639.5998535	-49.7001953	-48.8696289	-0.8305664	1.6711521
1000.0000000	659.5998535	-44.8999023	-44.0297546	-0.8701477	1.9379711
1000.0000000	679.5998535	-40.8000488	-40.1583405	-0.6417084	1.5728121
1000.0000000	699.5998535	-37.6000977	-36.8083344	-0.7917633	2.1057472
1000.0000000	719.5998535	-35.1000977	-34.0158081	-1.0842896	3.0891342
1000.0000000	739.5998535	-32.7998047	-31.5534515	-1.2463531	3.7998791
1500.0000000	259.5998535	-209.3998108	-209.5099945	0.1101837	-0.0526188
1500.0000000	279.5998535	-204.4998169	-204.4188232	-0.0809937	0.0396057
1500.0000000	299.5998535	-199.8999023	-199.5538025	-0.3460999	0.1731365
1500.0000000	319.5998535	-194.7998657	-194.7259979	-0.0738678	0.0379198
1500.0000000	339.5998535	-190.0999146	-189.8994751	-0.2004395	0.1054389
1500.0000000	359.5998535	-185.2999115	-185.0658569	-0.2340546	0.1263112
1500.0000000	379.5998535	-180.4000092	-180.2243805	-0.1756287	0.0973551
1500.0000000	399.5998535	-175.1999512	-174.7866974	-0.4132538	0.2358755
1500.0000000	419.5998535	-169.9999542	-169.4774017	-0.5225525	0.3073839
1500.0000000	439.5998535	-164.3998108	-164.1054535	-0.2943573	0.1790496
1500.0000000	459.5998535	-158.7998657	-158.3721771	-0.4276886	0.2693255
1500.0000000	479.5998535	-152.8000031	-152.3120880	-0.4879150	0.3193161
1500.0000000	499.5998535	-146.0999603	-145.7399445	-0.3600159	0.2464175
1500.0000000	519.5998535	-138.2998657	-138.5391388	0.2392731	-0.1730103
1500.0000000	539.5998535	-129.3999634	-130.7234344	1.3234711	-1.0227747
1500.0000000	559.5998535	-119.8000031	-121.3415680	1.5415649	-1.2867813
1500.0000000	579.5998535	-108.5000000	-110.4123230	1.9123230	-1.7625093
1500.0000000	599.5998535	-97.1999512	-99.1941223	1.9941711	-2.0516167
1500.0000000	619.5998535	-86.5998535	-87.8310852	1.2312317	-1.4217472
1500.0000000	639.5998535	-77.5000000	-77.6684418	0.1684418	-0.2173442
1500.0000000	659.5998535	-69.3000488	-69.4592896	0.1592407	-0.2297844
1500.0000000	679.5998535	-63.0000000	-62.8094940	-0.1905060	0.3023904
1500.0000000	699.5998535	-57.6999512	-57.2172394	-0.4827118	0.8365895
1500.0000000	719.5998535	-53.1000977	-52.5367432	-0.5633545	1.0609283
1500.0000000	739.5998535	-49.3999023	-48.4553986	-0.9445038	1.9119539
2000.0000000	259.5998535	-207.5998077	-208.1056824	0.5058746	-0.2436777
2000.0000000	279.5998535	-202.9998169	-203.1564789	0.1566620	-0.0771734
2000.0000000	299.5998535	-197.9999084	-198.2191925	0.2192841	-0.1107495
2000.0000000	319.5998535	-193.2998657	-193.3349915	0.0351257	-0.0181716
2000.0000000	339.5998535	-188.4999084	-188.4903412	-0.0095673	0.0050755
2000.0000000	359.5998535	-183.6999054	-183.7644043	0.0644989	-0.0351110
2000.0000000	379.5998535	-178.8000031	-178.8682861	0.0682831	-0.0381896
2000.0000000	399.5998535	-173.8999634	-173.8172760	-0.0826874	0.0475488
2000.0000000	419.5998535	-168.9999542	-168.9519806	-0.0479736	0.0283868
2000.0000000	439.5998535	-163.8998108	-163.6926270	-0.2071838	0.1264088
2000.0000000	459.5998535	-158.4998627	-158.2630310	-0.2368317	0.1494207
2000.0000000	479.5998535	-153.0000000	-152.7920227	-0.2079773	0.1359329
2000.0000000	499.5998535	-147.1999512	-146.8095093	-0.3904419	0.2652459
2000.0000000	519.5998535	-140.8998566	-141.1008606	0.2010040	-0.1426573
2000.0000000	539.5998535	-134.0999603	-134.7816467	0.6816864	-0.5083419
2000.0000000	559.5998535	-126.9000092	-128.5346680	1.6346588	-1.2881460
2000.0000000	579.5998535	-119.0000000	-120.7703705	1.7703705	-1.4877062
2000.0000000	599.5998535	-110.7998047	-110.9399567	0.1401520	-0.1264911
2000.0000000	619.5998535	-102.6999512	-102.7336273	0.0336761	-0.0327908
2000.0000000	639.5998535	-94.6000977	-95.2316284	0.6315308	-0.6675791
2000.0000000	659.5998535	-86.8999023	-87.7489014	0.8489990	-0.9769849
2000.0000000	679.5998535	-80.0000000	-80.5635223	0.5635223	-0.7044028
2000.0000000	699.5998535	-73.8999023	-74.1958618	0.2959595	-0.4004866
2000.0000000	719.5998535	-68.4001465	-68.3526611	-0.0474854	0.0694228
2000.0000000	739.5998535	-63.5998535	-63.0837708	-0.5160828	0.8114524

TABLE LXVI
(CONTINUED)

E. 76.6 PERCENT MIXTURE

PRESSURE (psia)	TEMPERATURE (°R)	EXPER. DEPARTURE (Btu/lb)	CALC. DEPARTURE (Btu/lb)	DIFFERENCE (Btu/lb)	PERCENT DIFFERENCE
500.0000000	259.5998535	-215.0998535	-215.5321350	0.4322815	-0.2009678
500.0000000	279.5998535	-210.4998169	-210.5978241	0.0980072	-0.0465593
500.0000000	299.5998535	-205.9999542	-205.7217865	-0.2781677	0.1350329
500.0000000	319.5998535	-201.4999542	-201.2463074	-0.2536469	0.1258793
500.0000000	339.5998535	-196.9998169	-196.9413757	-0.0584412	0.0296656
500.0000000	359.5998535	-192.6998596	-192.4624481	-0.2374115	0.1232027
500.0000000	379.5998535	-188.0998535	-187.9075470	-0.1923065	0.1022363
500.0000000	399.5998535	-183.5998077	-183.4690399	-0.1307678	0.0712243
500.0000000	419.5998535	-178.8998566	-178.8936615	-0.0061951	0.0034629
500.0000000	439.5998535	-174.2999573	-174.1351166	-0.1648407	0.0945730
500.0000000	459.5998535	-169.3999023	-169.5427094	0.1428070	-0.0843017
500.0000000	619.5998535	-35.1000977	-34.2124023	-0.8876953	2.5290384
500.0000000	639.5998535	-31.0000000	-30.2560120	-0.7439880	2.3999605
500.0000000	659.5998535	-27.8999023	-27.2572327	-0.6426697	2.3034830
500.0000000	679.5998535	-25.3999023	-24.8617401	-0.5381622	2.1187563
500.0000000	699.5998535	-23.3999023	-22.9255219	-0.4743805	2.0272751
500.0000000	719.5998535	-21.6999512	-21.3645477	-0.3354034	1.5456409
500.0000000	739.5998535	-19.9001465	-19.8291168	-0.0710297	0.3569303
500.0000000	759.5998535	-17.6999512	-18.6642609	0.9643097	-5.4480915
1000.0000000	259.5998535	-213.0998535	-213.5743561	0.4745026	-0.2226667
1000.0000000	279.5998535	-208.3998108	-208.6395416	0.2397308	-0.1150340
1000.0000000	299.5998535	-204.0999603	-203.9591370	-0.1408234	0.0689972
1000.0000000	319.5998535	-199.6999512	-199.5998840	-0.1000671	0.0501087
1000.0000000	339.5998535	-195.2998047	-195.2294464	-0.0703583	0.0360258
1000.0000000	359.5998535	-190.8998566	-190.7369385	-0.1629181	0.0853421
1000.0000000	379.5998535	-186.5998535	-186.3264771	-0.2733765	0.1465040
1000.0000000	399.5998535	-182.2998047	-181.7993317	-0.5004730	0.2745329
1000.0000000	419.5998535	-177.7998657	-177.3726959	-0.4271698	0.2402531
1000.0000000	439.5998535	-173.5999603	-172.7976379	-0.8023224	0.4621673
1000.0000000	459.5998535	-169.2999115	-168.0863342	-1.2135773	0.7168207
1000.0000000	479.5998535	-164.3999023	-163.1274719	-1.2724304	0.7739846
1000.0000000	499.5998535	-159.2998657	-158.1249084	-1.1749573	0.7375754
1000.0000000	519.5998535	-153.7000122	-152.7749023	-0.9251099	0.6018929
1000.0000000	539.5998535	-148.0999603	-146.9677582	-1.1322021	0.7644851
1000.0000000	559.5998535	-141.7998657	-140.5634613	-1.2364044	0.8719362
1000.0000000	579.5998535	-134.2999573	-133.2422333	-1.0577240	0.7875830
1000.0000000	599.5998535	-125.1998596	-124.7229614	-0.4768982	0.3809095
1000.0000000	619.5998535	-112.6000977	-113.2555695	0.6554718	-0.5821235
1000.0000000	639.5998535	-94.4001465	-96.2505951	1.8504486	-1.9602175
1000.0000000	659.5998535	-77.0998535	-76.9190216	-0.1808319	0.2345424
1000.0000000	679.5998535	-65.3000488	-63.8059235	-1.4941254	2.2880917
1000.0000000	699.5998535	-55.8000488	-55.3491516	-0.4508972	0.8080587
1000.0000000	719.5998535	-49.5000000	-49.3303070	-0.1696930	0.3428141
1000.0000000	739.5998535	-44.8000488	-44.7638855	-0.0361633	0.0807216
1000.0000000	759.5998535	-41.8999023	-40.9013672	-0.9985352	2.3831434
1500.0000000	259.5998535	-211.2998657	-211.4673004	0.1674347	-0.0792403
1500.0000000	279.5998535	-206.8998108	-206.7797089	-0.1201019	0.0580483
1500.0000000	299.5998535	-202.5999603	-202.0983887	-0.5015717	0.2475675
1500.0000000	319.5998535	-198.0999603	-197.5228424	-0.5771179	0.2913266
1500.0000000	339.5998535	-193.8998108	-193.0941467	-0.8056641	0.4155051
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1500.0000000	379.5998535	-185.2998657	-184.3977661	-0.9020996	0.4868321
1500.0000000	399.5998535	-180.7998047	-180.2046814	-0.5951233	0.3291614
1500.0000000	419.5998535	-176.2998657	-175.8610535	-0.4388123	0.2489010
1500.0000000	439.5998535	-171.9999542	-171.4217529	-0.5782013	0.3361636
1500.0000000	459.5998535	-167.4999084	-166.6016998	-0.8982086	0.5362440
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1500.0000000	499.5998535	-158.1998596	-157.2647552	-0.9351044	0.5910903
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1500.0000000	559.5998535	-142.7998657	-141.8023529	-0.9975128	0.6985388
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1500.0000000	619.5998535	-123.1999512	-122.7189331	-0.4810181	0.3904369
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1500.0000000	659.5998535	-106.5998535	-105.8873901	-0.7124634	0.6683528
1500.0000000	679.5998535	-97.0000000	-96.5073547	-0.4926453	0.5078819
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2000.0000000	439.5998535	-171.1999512	-170.3227081	-0.8772430	0.5124081
2000.0000000	459.5998535	-166.6999054	-165.9267120	-0.7731934	0.4638232
2000.0000000	479.5998535	-160.0999146	-161.6342316	1.5343170	-0.9583493
2000.0000000	499.5998535	-157.5998535	-156.9091644	-0.6906891	0.4382547
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2000.0000000	699.5998535	-100.3999023	-99.6301270	-0.7697754	0.7667090
2000.0000000	719.5998535	-93.3000488	-93.2831726	-0.0168762	0.0180881
2000.0000000	739.5998535	-86.4001465	-86.8853455	0.4851990	-0.5615718
2000.0000000	759.5998535	-80.3000488	-80.9063873	0.6063385	-0.7550910

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