THERMOPHYSICAL PROPERTIES OF HCFC ALTERNATIVES

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THERMOPHYSICAL PROPERTIES OF HCFC ALTERNATIVES**ARTI MCLR Project Number 660-50800****W.M. Haynes****Physical and Chemical Properties Division
National Institute of Standards and Technology****ABSTRACT**

Numerous fluids and fluid mixtures have been identified as promising alternatives to the HCFC refrigerants, but, for many of them, reliable thermodynamic data do not exist. In particular, reliable thermodynamic properties data and models are needed to predict the performance of the new refrigerants in heating and cooling equipment and to design and optimize equipment to be reliable and energy efficient. The objective of this project is to measure, with high accuracy, selected thermodynamic properties data for two pure refrigerants and nine refrigerant blends; these data will be used to fit equations of state and other property models which can be used in equipment design. The new data will fill in gaps in the existing data and resolve problems and differences that exist in and between existing data sets. Most of the studied fluids and blends are potential replacements for HCFC-22 and/or R502; in addition, one pure fluid and one blend are potential replacements for CFC-13 in low temperature refrigeration applications.

SCOPE

This project involves the measurement of primary thermodynamic data for eight binary mixtures and one ternary mixture (R125/134a; R32/143a; R125/143a; R143a/134a; R32/290; R125/290; R290/134a; R41/744; and R32/125/134a). These data will consist of simultaneous measurements of the coexisting liquid and vapor densities and coexisting liquid and vapor compositions as well as the bubble point pressures over wide ranges of temperature and composition. These data will be used to determine the mixing parameters, including any temperature and/or composition dependence, for the Carnahan-Starling-DeSantis (CSD) equation of state and the extended corresponding states (ECS) model. Primary data will also be measured for the pure fluid HFC-41. These measurements will include the determination of vapor pressure; liquid density; liquid, two-phase, and ideal gas heat capacities; vapor sound speeds and virial coefficients; and a determination of the critical point parameters. These data will cover wide ranges of temperature and pressure and will be used (together with available literature data) to fit equations of state and other models of the thermodynamic properties. The original proposal called for similar measurements of primary data for HFC-245ca; since work on this fluid has been funded by another sponsor, comprehensive mixture data for the binary system R125/143a have been substituted as recommended by the MCLR Advisory Committee. During the past quarter, we have carried out the following tasks: (a) measurements of coexisting densities, bubble point pressures, and liquid and vapor compositions of R32/125 and R32/143a systems with new VLE apparatus, (b) development of vapor pressure equation for HFC-41, (c) completion of five PVT isochores from 192 to 288 K (-114 to 59°) for R41/744 system, (d) initiation of vapor-liquid equilibrium and coexisting density measurements on the R134a/143a system, and (e) determination of saturated liquid densities and critical point parameters for HFC-41.

SIGNIFICANT RESULTS

Primary mixture data for binary mixtures of relatively volatile components (R125/134a; R32/143a; R125/143a; R143a/134a; R32/290; R125/290; R134a/290)

Vapor-liquid equilibrium data were obtained for three binary systems during the past few months. Measurements were made at temperatures from 280 to 340 K (44 to 152°F) in 15 K (27°F), or less, increments on the following systems: R32/143a, R32/125, and R32/134a. Although the R32/125 and R32/134a data are not part of the current project (they are funded by DOE), the data for these binary systems are essential information in the study of the ternary R32/125/134a system that is included in the current project.

The vapor-liquid equilibrium data for the R32/134a system were finalized, and the results are presented in Table 1. Comparisons of the data with literature results were presented in the previous progress report. The bubble point pressure and composition data taken in the new liquid-bath vapor-liquid equilibrium apparatus were of much higher quality than the data that had been obtained in the air-bath apparatus. Near-saturation (p,p,T) data for this system were also presented in a previous progress report.

The vapor-liquid equilibrium data for the R32/125 and R32/143a systems are presented in Tables 2 and 3. In Table 4 the near-saturation (p,p,T) results for the R32/143a system are presented. The near-saturation (p,p,T) data for the R32/125 system were presented in an earlier progress report.

Figures 1 and 2 show the results for the R32/125 system from the new apparatus compared to literature data discussed in a previous report. The improvement in the vapor compositions and the bubble point pressures (shown as crosses) is significant. The data are in agreement with the bubble point pressures of Kleemiss [University of Hannover, private communication]. The bubble point pressures of Fujiwara *et al.* [Fujiwara *et al.*, 13th Japanese Symposium on Thermophysical Properties, A116:61 (1992)] are higher than those of the other data sets. The bubble point pressures of Widiatmo *et al.* [Widiatmo, J.V., Sato, H., and Watanabe, K., High Temperature and Pressure 25:677 (1993)] and of Higashi [Iwaki Meisei University, private communication] exhibit more scatter than our results. The bubble point pressures of Defibaugh and Morrison [NIST, Gaithersburg, MD, private communication] are lower than the other data sets. In Figure 2 the deviations of the vapor compositions are examined. The data from the new apparatus show scatter an order of magnitude less than those from the air-bath apparatus and are in good agreement with the other data sets.

Figures 3 and 4 present similar comparisons for the R32/143a system. The amount of data available for this system is not as extensive as for the R32/134a and R32/125 systems. The comparisons show that the scatter in the data from the new oil-bath apparatus is significantly less than that from the previous air-bath apparatus and from the data of Fujiwara *et al.* The model in REFPROP (which is used as the baseline for comparisons) was not optimized for this system. This is the reason that both the bubble point pressures and vapor compositions are not in as good agreement with the model as for the R32/125 and R32/134a systems. Also, some of the higher temperature data could not be plotted because the model fails to converge in that region.

During the past few months, progress on these binary mixture measurements has been delayed by a number of experimental difficulties. The experimental measurements were delayed by unexpected electrical problems that damaged some of our equipment. The programmable power supply that controls the heaters in the oil-bath and the integrator for the gas chromatograph were damaged by electrical noise and power fluctuations. The integrator was repaired locally, but the programmable power supply was returned to the manufacturer for repairs. We continued to take data by balancing the heaters manually. It takes much longer to balance the heaters manually than it does for the computer to balance the heaters using the programmable power supply. Also, with manual control, the temperature will drift overnight as the room temperature changes. Each morning the apparatus must be returned to the set point temperature and the heaters rebalanced.

The unexpected electrical problems in the laboratory were due to overloading of the power circuits and to the noisy power in the building. To address these problems, the apparatus is being moved to a new laboratory with better room temperature control, more electrical power, and more stable power.

Another problem we have encountered during the past quarter is an apparent drift in the densimeters. The densimeters in the new apparatus were reported to perform significantly better than older models from the same supplier that were used in the air-bath vapor-liquid equilibrium apparatus. The older models of the densimeters have been out of production for several years. From the analysis of the performance of the new densimeters, it appears that they are not as stable as the old densimeters. We are currently discussing the problems with the manufacturer. Meanwhile, we are obtaining vapor-liquid equilibrium data on the new apparatus and (p,ρ,T) data on the previous apparatus until this problem is resolved.

The R134a/143a system is currently being studied in the new apparatus. It will be completed in three weeks. We will then study the R125/143a and R134a/125 systems. These systems have not been studied previously in the air-bath apparatus; therefore, we will measure the near-saturation densities in this apparatus using the prepared standard mixtures. We will then complete measurements on the R32/290, R134a/290, and R125/290 systems, as needed. If we find that the air-bath apparatus performed satisfactorily for these systems, we will only need a few additional measurements.

Primary mixture data for a ternary mixture of relatively volatile components (R32/125/134a)

Upon completion of the measurements for the R134a/R125 system, we will perform vapor-liquid equilibrium measurements on the R32/125/134a system in the new apparatus. Three standard mixtures will be used in the air-bath apparatus for near saturation (p,ρ,T) measurements unless the problems with the new densimeters can be resolved. Results for a single composition (0.346 HFC-32 + 0.300 HFC-125 + 0.354 HFC-134a) are presented in Tables 5 and 6. Comparisons with other data were presented in the last progress report.

Primary mixture data for binary mixture of very low boiling components (R41/744)

Isochoric PVT measurements for an equimolar mixture of R41/744 have been completed for five liquid isochores at temperatures from 192 K to 288 K (-114 to 59°F) at pressures to 35 MPa (5100 psi). The temperatures, pressures, and densities are presented in Tables 7 and 8, and the locations of the measurements are shown in Figure 5. During the course of charging this mixture to a pre-cooled PVT cell at 180 K (-136°F), it was established that the solidification temperature is below this temperature. Future charging processes will be carried out at 170 and 160 K (-154 and -172°F) in an effort to determine the lower and upper bounds for the solidification temperature of this composition. The PVT apparatus is presently receiving a new data-acquisition computer which replaces a computer which had served that purpose for 11 years, but had recently experienced disk failures. When this modification has been completed, ten additional isochores for the equimolar R41/744 mixture will be measured in the PVT apparatus, thereby extending the temperature coverage to 400 K (260°F).

Comprehensive data for the R143a/125 mixture

Comprehensive isochoric PVT measurements were completed earlier for an equimolar mixture of R143a/125. C_p measurements will be initiated in the near future.

Primary data for a very low-boiling fluid (HFC-41)

The vapor pressures of HFC-41 were measured from 218 to 312 K (-67 to 102°F) in the new vapor-liquid equilibrium apparatus. The vapor pressure data are reported in Table 9. These data agree within $\pm 0.3\%$ with the vapor pressures obtained in the isochoric PVT apparatus. Near-saturation (p, ρ , T) data were taken in the air-bath vapor-liquid equilibrium apparatus. The density data are reported in Table 10. The data in Tables 9 and 10 were used to estimate the critical parameters using a method developed by Van Poolen [Holcomb, C.D. and Van Poolen, L.J., *Fluid Phase Equilibria*, 100:222 (1994)]. The critical parameters are presented in Table 11. The critical pressure agrees within the experimental uncertainty with the critical pressure reported by Bominaar *et al.* [Bominaar, S.A.R.C., Biswas, S.N., Trappeniers, N.J., and Ten Seldam, C.A., *J. Chem. Thermodynamics* 19:959 (1987)] and of Biswas *et al.* [Biswas, S.N., Ten Seldam, C.A., Bominaar, S.A.R.C., and Trappeniers, N.J., *Fluid Phase Equilibria* 49:1 (1989)]. The critical temperature is within 0.14 K (0.25°F) of the values of Bominaar *et al.* and of Biswas *et al.* Our critical density is approximately 1.5% higher than the literature values. (The results of Bominaar *et al.* and Biswas *et al.* are based on the same data set with some minor differences in the fitting procedure.)

Vapor pressure measurements on HFC-41 from this project have been combined with selected literature data [Oi, T., Shulman, J., Popowicz, A. and Ishida, T., *J. Phys. Chem.* 87: 3153-3160 (1983)] to develop a vapor pressure correlation valid from the triple point temperature of 131.35 K (-223.24°F) to the critical temperature of 317.28 K (111.43 °F). This equation is of the form:

$$P_{SAT} = P_c \exp \left[\left(\frac{a_1\tau + a_2\tau^{1.5} + a_3\tau^3 + a_4\tau^6}{1 - \tau} \right) \right]$$

with:

$$\tau = 1 - T/T_c$$

The critical temperature T_c used in this correlation is 317.28 K (571.10°R) and the critical pressure P_c is 5.897 MPa (855.3 psia). The critical pressure was determined by extrapolating the present vapor pressure data to the critical temperature. The coefficients a_i are dimensionless and are:

$$a_1 = -7.01707106$$

$$a_2 = 1.33436478$$

$$a_3 = -1.78330695$$

$$a_4 = -1.84394112$$

Figure 6 compares the available experimental data with this correlation as the baseline. The present work of Magee (isochoric PVT apparatus, data given in last progress report) and of Holcomb (liquid-bath VLE apparatus, data in this report) fill a large gap between 211 and 253 K (-80 and -4°F) that exists in the literature data. The NIST datasets are in good agreement except for the two lowest temperature points of Holcomb, which are at the very low end of validity for the pressure transducer used. The NIST data indicate that the previous vapor pressure measurements of Demiriz *et al.* [Demiriz, A.M., Kohlen, R., Koopmann, C., Moeller, D., Sauermann, P., Iglesias-Silva, G.A., and Kohler, F., *Fluid Phase Equilibria* 85:313-333 (1993)] and Bominaar *et al.* [Bominaar, S.A.R.C., Trappeniers, N.J., and Biswas, S.N., *Physica B* 140:83-86 (1986)] were systematically low. This correlation represents the experimental data with a RMS deviation of 0.17%. Vapor pressures calculated with this correlation are given in Table 12.

Saturated liquid densities have been determined from the isochoric PVT measurements for fluoromethane (HFC-41) reported previously. Curve-fitted equations for temperature-pressure, density-temperature, and vapor pressure (P_v) - temperature were solved simultaneously for the

properties of the saturated liquid state at temperatures from 131 to 309 K (-224 to 96°F). The vapor pressure equation was described above. The temperatures, pressures, and densities of the saturated liquid are given in Tables 13 and 14.

Heat capacity at constant volume (C_v) will be measured for HFC-41 with an adiabatic calorimeter in a temperature range from 132 to 345 K (-222 to 161°F) at pressures to 35 MPa (5100 psi). This activity will start in the coming quarter once new calibration measurements of the system volume have been completed and analyzed. The volume of the pressure transducer has already been determined by weighing the quantity of distilled water needed to fill this volume. The volume of the calorimeter is being determined because an impurity in a sample of HFC-236fa previously measured in the calorimeter obstructed the capillary tubing which leads from the cell to the pressure transducer. This resulted in an overpressure condition, not detected by the transducer, which stretched the cell body by a few percent. This impurity was determined, by mass spectrometry, to consist of aliphatic hydrocarbon species with molecular masses in a range from 286 to 428 g·mol⁻¹. Due to the impurity's large molecular mass, it had evaded detection by earlier gas chromatography tests. Since the impurity was found to be soluble in HFC-32, the system volume was cleaned out with a high purity supply of HFC-32, until no traces were found in the effluent.

COMPLIANCE WITH AGREEMENT

NIST has complied with all terms of the grant agreement modulo small shifts in the estimated level of effort from one property and/or fluid to another.

PRINCIPAL INVESTIGATOR EFFORT

Dr. W.M. Haynes is the NIST Principal Investigator for the MCLR program. During the first quarter of calendar year 1996, Dr. Haynes devoted approximately one week to planning this project, monitoring and reviewing the research, and preparing the report. The project involves multiple researchers and capabilities in Gaithersburg, MD and Boulder, CO.

Figure 1: Comparison of bubble point pressures for the R32/125 system; the baseline is from an extended corresponding states (ECS) model in REFPROP.

Figure 2: Comparison of vapor compositions for the R32/125 system; the baseline is from an extended corresponding states (ECS) model in REFPROP.

Figure 3: Comparison of bubble point pressures for the R32/143a system; the baseline is from an extended corresponding states (ECS) model in REFPROP.

Figure 4: Comparison of vapor compositions for the R32/143a system; the baseline is from an extended corresponding states (ECS) model in REFPROP.

Figure 5: Locus of (p,p,T) data for 0.49982 HFC-41 + 0.50018 R744 mixture.

Figure 6. Comparison of experimental vapor pressure data for HFC-41 with correlation.

APPENDIX A:
TABLES OF THERMOPHYSICAL PROPERTIES DATA

Table 1: Vapor-Liquid Equilibrium Data for R32/134a Mixtures from 280 to 340 K (44 to 152°F).

Temperature K	x(R32) Liquid Mole Fraction	y(R32) Vapor Mole Fraction	Pressure MPa	Temperature °F	x(R32) Liquid Mass Fraction	y(R32) Vapor Mass Fraction	Pressure psia
280.390	0.000	0.000	0.3790	45.01	0.000	0.000	55.0
279.977	0.299	0.489	0.5549	44.27	0.179	0.328	80.5
280.054	0.478	0.676	0.6736	44.41	0.318	0.515	97.7
279.983	0.604	0.757	0.7390	44.28	0.437	0.614	107.2
279.978	0.642	0.797	0.7715	44.27	0.478	0.667	111.9
280.089	0.642	0.793	0.7743	44.47	0.478	0.661	112.3
279.985	0.650	0.805	0.7782	44.28	0.486	0.678	112.9
280.190	0.650	0.796	0.7777	44.65	0.486	0.665	112.8
280.128	0.737	0.869	0.8516	44.54	0.588	0.772	123.5
280.000	1.000	1.000	1.0060	44.31	1.000	1.000	145.9
283.020	0.302	0.483	0.6092	49.75	0.181	0.323	88.4
283.056	0.632	0.782	0.8395	49.81	0.467	0.647	121.8
283.010	0.640	0.795	0.8458	49.73	0.475	0.664	122.7
283.076	0.640	0.795	0.8466	49.85	0.475	0.664	122.8
289.912	0.283	0.466	0.7461	62.15	0.168	0.308	108.2
289.786	0.637	0.782	1.0302	61.93	0.472	0.647	149.4
292.926	0.282	0.457	0.8114	67.58	0.167	0.300	117.7
292.917	0.634	0.773	1.1238	67.56	0.469	0.635	163.0
294.690	0.000	0.000	0.5962	70.75	0.000	0.000	86.5
294.949	0.275	0.448	0.8558	71.22	0.162	0.293	124.1
294.987	0.452	0.626	1.0180	71.29	0.296	0.460	147.6
295.127	0.529	0.696	1.0914	71.54	0.364	0.539	158.3
294.937	0.584	0.748	1.1485	71.20	0.417	0.602	166.6
294.960	0.632	0.773	1.1867	71.24	0.467	0.635	172.1
294.943	0.720	0.838	1.2706	71.21	0.567	0.725	184.3
295.480	1.000	1.000	1.5747	72.18	1.000	1.000	228.4
299.958	0.619	0.761	1.3569	80.24	0.453	0.619	196.8
303.871	0.262	0.403	1.0760	87.28	0.153	0.256	156.1
303.069	0.609	0.756	1.4530	85.84	0.443	0.612	210.7
309.310	0.000	0.000	0.9131	97.07	0.000	0.000	132.4
309.978	0.181	0.292	1.1673	98.27	0.101	0.174	169.3
309.995	0.225	0.370	1.2250	98.30	0.129	0.230	177.7
309.971	0.228	0.363	1.2153	98.26	0.131	0.225	176.3
309.984	0.360	0.515	1.3924	98.28	0.223	0.351	202.0
309.992	0.681	0.793	1.8264	98.30	0.521	0.661	264.9

Table 1: Vapor-Liquid Equilibrium Data for R32/134a Mixtures from 280 to 340 K (44 to 152°F) (continued).

Temperature K	x(R32) Liquid Mole Fraction	y(R32) Vapor Mole Fraction	Pressure MPa	Temperature °F	x(R32) Liquid Mass Fraction	y(R32) Vapor Mass Fraction	Pressure psia
309.990	1.000	1.000	2.2939	98.29	1.000	1.000	332.7
324.500	0.000	0.000	1.3669	124.41	0.000	0.000	198.3
325.004	0.162	0.256	1.6432	125.32	0.090	0.149	238.3
324.987	0.294	0.422	1.8887	125.29	0.175	0.271	273.9
324.998	0.595	0.710	2.4234	125.31	0.428	0.555	351.5
324.988	0.733	0.809	2.6975	125.29	0.583	0.683	391.2
324.992	0.783	0.860	2.8052	125.30	0.648	0.758	406.9
324.990	1.000	1.000	3.2790	125.29	1.000	1.000	475.6
340.160	0.000	0.000	1.9862	152.60	0.000	0.000	288.1
339.865	0.466	0.569	3.0362	152.07	0.308	0.402	440.4
339.933	0.577	0.669	3.3151	152.19	0.410	0.508	480.8
339.895	0.763	0.820	3.8065	152.12	0.621	0.699	552.1
339.980	1.000	1.000	4.5610	152.28	1.000	1.000	661.5

Table 2: Vapor-Liquid Equilibrium Data for R32/125 Mixtures from 280 to 340 K (44 to 152 °F).

Temperature K	x(R32) Liquid Mole Fraction	y(R32) Vapor Mole Fraction	Pressure MPa	Temperature °F	x(R32) Liquid Mass Fraction	y(R32) Vapor Mass Fraction	Pressure psia
280.29	0.000	0.000	0.8302	44.85	0.000	0.000	120.4
279.83	0.346	0.398	0.9191	44.03	0.187	0.223	133.3
280.02	0.350	0.387	0.9239	44.36	0.189	0.215	134.0
280.06	0.482	0.520	0.9546	44.43	0.287	0.320	138.5
279.95	0.650	0.676	0.9854	44.25	0.446	0.475	142.9
280.07	0.948	0.947	1.0166	44.46	0.888	0.886	147.4
280.00	1.000	1.000	1.0076	44.33	1.000	1.000	146.1
295.23	0.000	0.000	1.2720	71.74	0.000	0.000	184.5
295.01	0.347	0.387	1.4148	71.34	0.187	0.215	205.2
295.03	0.470	0.508	1.4608	71.39	0.278	0.309	211.9
295.06	0.653	0.676	1.5159	71.44	0.449	0.475	219.9
295.03	0.914	0.915	1.5585	71.38	0.822	0.824	226.0
294.94	0.948	0.947	1.5555	71.22	0.888	0.886	225.6
293.99	1.000	1.000	1.5152	69.51	1.000	1.000	219.8
310.61	0.000	0.000	1.8904	99.43	0.000	0.000	274.2
310.01	0.339	0.373	2.0801	98.34	0.182	0.205	301.7
310.00	0.465	0.498	2.1497	98.32	0.274	0.301	311.8
309.96	0.646	0.672	2.2267	98.26	0.442	0.470	323.0
310.08	0.915	0.914	2.3032	98.47	0.824	0.822	334.1
310.51	1.000	1.000	2.3288	99.25	1.000	1.000	337.8
324.89	0.000	0.000	2.6390	125.13	0.000	0.000	382.8
324.91	0.334	0.358	2.9026	125.18	0.179	0.195	421.0
325.07	0.462	0.489	3.0581	125.45	0.271	0.293	443.5
324.97	0.644	0.661	3.1740	125.28	0.439	0.458	460.3
325.17	0.919	0.919	3.2966	125.63	0.831	0.831	478.1
325.49	1.000	1.000	3.3179	126.21	1.000	1.000	481.2
339.38	0.000	0.000	3.6291	151.21	0.000	0.000	526.4
340.18	0.740	0.749	4.4817	152.65	0.552	0.564	650.0
340.17	0.885	0.890	4.5716	152.64	0.769	0.778	663.1
340.17	0.919	0.919	4.5815	152.64	0.831	0.831	664.5
340.08	1.000	1.000	4.5717	152.47	1.000	1.000	663.1

Table 3: Vapor-Liquid Equilibrium Data for R32/143a Mixtures from 280 to 340 K (44 to 152 °F).

Temperature K	x(R32) Liquid Mole Fraction	y(R32) Vapor Mole Fraction	Pressure MPa	Temperature °F	x(R32) Liquid Mass Fraction	y(R32) Vapor Mass Fraction	Pressure psia
280.23	0.000	0.000	0.7673	44.73	0.000	0.000	111.3
279.88	0.357	0.439	0.9056	44.11	0.256	0.326	131.3
279.88	0.377	0.434	0.9125	44.12	0.273	0.322	132.3
279.90	0.605	0.652	0.9708	44.15	0.487	0.537	140.8
279.86	0.674	0.692	0.9802	44.09	0.561	0.582	142.2
279.85	0.735	0.759	0.9925	44.06	0.632	0.661	143.9
280.00	1.000	1.000	1.0073	44.33	1.000	1.000	146.1
294.11	0.000	0.000	1.1345	69.73	0.000	0.000	164.6
294.96	0.414	0.479	1.4067	71.27	0.304	0.363	204.0
295.03	0.638	0.665	1.4932	71.39	0.522	0.551	216.6
294.96	0.706	0.761	1.5088	71.23	0.598	0.663	218.8
294.92	0.719	0.756	1.5171	71.18	0.613	0.657	220.0
293.99	1.000	1.000	1.5155	69.51	1.000	1.000	219.8
309.37	0.000	0.000	1.6768	97.18	0.000	0.000	243.2
310.10	0.635	0.676	2.1906	98.51	0.519	0.564	317.7
310.02	0.706	0.753	2.2162	98.34	0.598	0.654	321.4
310.05	0.811		2.2663	98.43	0.726		328.7
310.07	0.875		2.2811	98.45	0.812		330.8
310.06	0.912	0.936	2.2883	98.43	0.865	0.901	331.9
310.51	1.000	1.000	2.3290	99.25	1.000	1.000	337.8
326.02	0.000	0.000	2.4669	127.15	0.000	0.000	357.8
325.06	0.818	0.841	3.2094	125.43	0.736	0.766	465.5
325.05	0.746	0.762	3.1559	125.43	0.645	0.665	457.7
325.07	0.604	0.638	3.0672	125.46	0.486	0.522	444.9
324.94	0.547	0.575	3.0127	125.22	0.428	0.456	437.0
325.49	1.000	1.000	3.3177	126.19	1.000	1.000	481.2
340.07	0.000	0.000	3.3177	152.44	0.000	0.000	481.2
340.03	0.543	0.570	4.1691	152.38	0.424	0.451	604.7
340.01	0.499	0.546	4.1357	152.34	0.381	0.427	599.8
340.02	0.698	0.718	4.3449	152.37	0.589	0.612	630.2
340.08	1.000	1.000	4.5712	152.47	1.000	1.000	663.0

Table 4: Near-Saturation (p, ρ, T) Data for R32/143a Mixtures from 280 to 350 K (44 to 170°F).

x (R32) Liquid Mole Fraction	Temperature K	Pressure MPa	Density kg/m ³	x (R32) Liquid Mass Fraction	Temperature °F	Pressure psia	Density lb/ft ³
0.271	279.76	0.9298	1003.2	0.187	43.87	134.9	62.57
0.271	279.73	0.9293	1003.2	0.187	43.83	134.8	62.57
0.542	279.40	0.9867	1009.8	0.423	43.23	143.1	62.99
0.952	279.22	1.0172	1026.5	0.925	42.90	147.5	64.02
0.861	293.04	1.5273	972.9	0.793	67.78	221.5	60.68
0.713	293.09	1.4913	962.9	0.606	67.87	216.3	60.06
0.619	293.27	1.4693	957.8	0.501	68.19	213.1	59.74
0.619	293.35	1.4802	957.6	0.501	68.33	214.7	59.73
0.436	293.66	1.4169	950.9	0.324	68.89	205.5	59.31
0.203	293.35	1.2679	948.9	0.136	68.33	183.9	59.18
0.203	293.14	1.2487	949.3	0.136	67.96	181.1	59.21
0.362	309.97	2.0219	874.6	0.260	98.26	293.3	54.55
0.263	309.22	1.8873	877.4	0.181	96.91	273.7	54.72
0.635	309.76	2.1973	882.6	0.519	97.88	318.7	55.05
0.736	309.22	2.2236	892.2	0.633	96.91	322.5	55.65
0.89	308.65	2.3428	903.6	0.834	95.88	339.8	56.36
0.172	325.59	2.7533	782.1	0.114	126.37	399.3	48.78
0.450	325.90	3.1204	786.6	0.336	126.94	452.6	49.06
0.581	326.01	3.2036	789.2	0.462	127.12	464.6	49.22
0.596	326.03	3.2108	789.2	0.477	127.17	465.7	49.22
0.664	326.01	3.2722	793.4	0.550	127.12	474.6	49.49
0.668	326.01	3.2690	793.0	0.555	127.12	474.1	49.46
0.728	325.98	3.3267	797.8	0.624	127.08	482.5	49.76
0.731	325.98	3.3269	797.7	0.627	127.08	482.5	49.75
0.805	325.80	3.3308	801.6	0.719	126.75	483.1	50.00
0.787	325.93	3.3486	801.1	0.696	126.98	485.7	49.97
0.787	325.93	3.3413	801.0	0.696	126.98	484.6	49.96
0.259	339.51	3.7959	660.0	0.178	151.43	550.6	41.17
0.267	339.51	3.7958	659.6	0.184	151.43	550.5	41.14
0.394	339.49	4.0427	665.0	0.287	151.39	586.3	41.48
0.394	339.49	4.0437	664.9	0.287	151.39	586.5	41.47
0.467	339.46	4.2711	671.0	0.352	151.34	619.5	41.85
0.887	339.25	4.5893	704.8	0.829	150.96	665.6	43.96
0.811	339.15	4.5545	689.9	0.726	150.78	660.6	43.03
0.792	339.12	4.5499	689.6	0.702	150.73	659.9	43.01
0.715	339.23	4.4695	678.0	0.608	150.92	648.2	42.29

Table 4: Near-Saturation (p, ρ, T) Data for R32/143a Mixtures from 280 to 350 K (44 to 170°F) (continued).

x (R32) Liquid Mole Fraction	Temperature K	Pressure MPa	Density kg/m ³	x (R32) Liquid Mass Fraction	Temperature °F	Pressure psia	Density lb/ft ³
0.633	339.25	4.4695	677.8	0.516	150.96	648.2	42.28
0.633	339.25	4.3982	672.5	0.516	150.96	637.9	41.95
0.636	339.25	4.3963	672.0	0.520	150.96	637.6	41.91
0.640	347.07	5.1181	500.7	0.524	165.04	742.3	31.23
0.775	347.02	5.3140	548.1	0.681	164.95	770.7	34.18
0.802	347.07	5.3141	543.4	0.715	165.04	770.7	33.89
0.802	347.05	5.3138	546.8	0.715	165.00	770.7	34.11
0.862	347.00	5.3762	597.7	0.795	164.90	779.8	37.28
0.864	347.05	5.3746	595.3	0.797	165.00	779.5	37.13
0.864	347.02	5.3751	597.2	0.797	164.95	779.6	37.25
0.855	349.99	5.6074	559.7	0.785	170.29	813.3	34.91
0.855	349.14	5.6138	559.3	0.785	168.77	814.2	34.88
0.882	348.99	5.6311	557.5	0.822	168.49	816.7	34.77
0.880	348.99	5.6312	557.6	0.819	168.49	816.7	34.78
0.423	280.57	0.9298	34.5	0.312	45.35	134.9	2.15
0.423	280.57	0.9293	34.3	0.312	45.35	134.8	2.14
0.658	294.47	1.4693	47.7	0.544	70.36	213.1	2.98
0.560	294.86	1.4169	48.4	0.441	71.05	205.5	3.02
0.231	294.34	1.2487	52.5	0.157	70.13	181.1	3.28
0.429	310.26	2.0219	82.6	0.317	98.79	293.3	5.15
0.289	309.44	1.8873	82.1	0.201	97.29	273.7	5.12
0.700	309.95	2.1973	78.2	0.591	98.23	318.7	4.88
0.771	309.46	2.2236	75.4	0.676	97.34	322.5	4.70
0.231	326.66	2.7410	134.9	0.157	128.29	397.5	8.41
0.231	326.58	2.7254	134.4	0.157	128.15	395.3	8.38
0.56	326.79	3.1204	133.5	0.441	128.53	452.6	8.32
0.628	326.89	3.2036	133.9	0.511	128.71	464.6	8.35
0.693	326.89	3.2690	132.5	0.583	128.71	474.1	8.26
0.522	340.23	4.2711	219.2	0.403	152.72	619.5	13.67
0.921	340.04	4.5893	185.8	0.878	152.39	665.6	11.59
0.787	339.83	4.5545	199.0	0.696	152.01	660.6	12.41
0.817	339.78	4.5499	198.9	0.734	151.92	659.9	12.40
0.654	339.96	4.4695	209.8	0.539	152.25	648.2	13.09
0.656	339.96	4.4695	210.4	0.541	152.25	648.2	13.12
0.645	339.96	5.1181	317.1	0.529	152.25	742.3	19.78
0.807	339.96	5.3141	279.5	0.721	152.25	770.7	17.43

Table 4: Near-Saturation (p, ρ, T) Data for R32/143a Mixtures from 280 to 350 K (44 to 170°F) (continued).

x (R32) Liquid Mole Fraction	Temperature K	Pressure MPa	Density kg/m ³	x (R32) Liquid Mass Fraction	Temperature °F	Pressure psia	Density lb/ft ³
0.863	347.74	5.3762	282.3	0.796	166.24	779.8	17.61
0.865	347.71	5.3746	286.2	0.799	166.20	779.5	17.85
0.865	347.71	5.3751	283.7	0.799	166.20	779.6	17.70
0.860	349.79	5.6074	374.9	0.792	169.93	813.3	23.38
0.860	349.76	5.6138	385.0	0.792	169.88	814.2	24.01
0.877	349.73	5.6311	405.3	0.815	169.83	816.7	25.28
0.886	349.73	5.6312	402.7	0.828	169.83	816.7	25.11

Table 5: Bubble Point Pressures for a Ternary Mixture of R32/125/134a from 221 to 345 K (-62 to 161°F). (The composition in mole fraction is 0.346 R32/0.300 R125/0.354 R134a and in mass fraction is 0.200 R32/0.400 R125/ 0.400 R134a.)

Temperature K	Bubble Point Pressure MPa	Temperature °F	Bubble Point Pressure psia
220.93 ¹	0.0731	-62.02	10.6
230.45 ¹	0.1164	-44.88	16.9
239.53 ¹	0.1722	-28.53	25.0
250.06 ¹	0.2657	-9.58	38.5
259.92 ¹	0.3834	8.17	55.6
269.95 ¹	0.5411	26.23	78.5
279.94 ¹	0.7469	44.20	108.3
289.88 ¹	0.9997	62.10	145.0
284.50 ²	0.8516	52.41	123.5
293.40 ²	1.0990	68.43	159.4
305.28 ²	1.5044	89.82	218.2
317.09 ²	2.0133	111.07	292.0
326.18 ²	2.4896	127.44	361.1
335.01 ²	3.0198	143.33	438.0
335.33 ²	3.0464	143.91	441.8
345.21 ²	3.7336	161.69	541.5

¹Data from new liquid-bath apparatus.

²Data from previous air-bath apparatus.

Table 6: Near Saturation (p, ρ ,T) Data for a Ternary Mixture of R32/125/134a from 244 to 346 K (-20 to 163°F). (The composition in mole fraction is 0.346 R32/0.300 R125/0.354 R134a and in mass fraction is 0.200 R32/0.400 R125/ 0.400 R134a.)

Temperature K	Pressure MPa	Density kg/m ³	Temperature °F	Pressure psia	Density lb/ft ³
244.44	0.2291	1350.3	-19.70	33.2	84.22
244.72	0.2316	1349.3	-19.19	33.6	84.16
254.15	0.3292	1320.2	-2.22	47.7	82.34
265.42	0.4854	1277.2	18.07	70.4	79.66
275.23	0.6637	1240.5	35.73	96.3	77.37
283.44	0.8486	1207.6	50.50	123.1	75.32
296.82	1.2312	1149.6	74.59	178.6	71.70
306.88	1.5952	1101.6	92.70	231.4	68.71
316.62	2.0188	1051.2	110.23	292.8	65.57
326.55	2.5309	991.3	128.10	367.1	61.83
336.32	3.1213	919.5	145.69	452.7	57.35
346.02	3.8000	823.6	163.15	551.1	51.37

Table 7. Compressed Liquid-Phase (P, ρ , T) Data for 0.49982 HFC-41 +
0.50018 R744 in SI Units.

T	P	ρ
K	MPa	mol·dm ⁻³
192.000	3.4374	27.5818
193.001	5.1455	27.5750
194.001	6.8244	27.5664
195.000	8.4192	27.5546
196.001	9.8510	27.5369
198.000	12.9219	27.5126
200.001	16.3609	27.4998
202.000	19.8917	27.4901
204.001	23.4505	27.4818
206.000	27.0106	27.4741
207.999	30.5707	27.4670
210.001	34.1270	27.4601
206.002	3.4172	26.6417
206.999	4.9531	26.6355
208.001	6.4485	26.6282
209.001	7.9136	26.6185
210.002	9.2734	26.6041
211.999	11.9340	26.5771
214.000	14.9304	26.5635
216.000	18.0241	26.5538
218.000	21.1517	26.5456
220.001	24.2884	26.5382
221.999	27.4204	26.5313
223.999	30.5507	26.5248
225.998	33.6762	26.5185
220.001	3.3521	25.6434
221.000	4.7067	25.6379
222.000	6.0293	25.6316
223.000	7.3398	25.6239
224.000	8.6005	25.6132
225.999	10.9508	25.5851
227.999	13.5215	25.5696
229.999	16.2066	25.5595
232.002	18.9311	25.5514
234.001	21.6669	25.5442
236.001	24.4034	25.5375
238.000	27.1387	25.5313
240.001	29.8764	25.5252
242.001	32.6057	25.5194
244.001	35.3201	25.5138
236.002	3.4541	24.4192
237.002	4.6060	24.4143
238.000	5.7333	24.4090
239.001	6.8704	24.4027
240.000	7.9802	24.3947
242.001	10.0655	24.3699
243.999	12.1880	24.3515
246.000	14.4469	24.3407
248.001	16.7442	24.3324
250.000	19.0570	24.3253
252.001	21.3712	24.3189
254.001	23.6900	24.3129
256.000	26.0080	24.3072
258.000	28.3185	24.3017
260.001	30.6312	24.2963
261.999	32.9384	24.2911

Table 7. Compressed Liquid-Phase (P, ρ , T) Data for 0.49982 HFC-41 + 0.50018 R744 in SI Units (continued).

T	P	ρ
K	MPa	mol·dm ⁻³
264.000	35.2414	24.2860
253.999	3.4254	22.8577
255.002	4.3640	22.8538
256.000	5.2941	22.8493
257.001	6.2155	22.8444
258.000	7.1369	22.8387
259.998	8.9215	22.8226
262.002	10.6372	22.8006
264.000	12.4228	22.7871
266.000	14.2623	22.7780
268.000	16.1238	22.7707
270.000	17.9903	22.7643
272.000	19.8611	22.7585
273.998	21.7344	22.7531
275.999	23.6078	22.7478
278.001	25.4787	22.7428
280.001	27.3503	22.7379
282.000	29.2200	22.7332
284.002	31.0882	22.7285
286.000	32.9518	22.7239
287.999	34.8110	22.7194

Table 8. Compressed Liquid-Phase (P, ρ ,T) Data for 0.49982 HFC-41 + 0.50018 R744 in PI Units.

T °F	P psia	ρ lb-ft ⁻³
-114.07	498.554	67.193
-112.27	746.293	67.177
-110.47	989.796	67.156
-108.67	1221.098	67.127
-106.87	1428.770	67.084
-103.27	1874.160	67.025
-99.67	2372.955	66.993
-96.07	2885.046	66.970
-92.47	3401.202	66.950
-88.87	3917.554	66.931
-85.27	4433.900	66.914
-81.67	4949.703	66.897
-88.87	495.616	64.903
-87.07	718.387	64.888
-85.27	935.279	64.870
-83.47	1147.766	64.847
-81.67	1344.990	64.811
-78.07	1730.884	64.746
-74.47	2165.471	64.712
-70.87	2614.171	64.689
-67.27	3067.794	64.669
-63.67	3522.739	64.651
-60.07	3976.992	64.634
-56.47	4431.006	64.618
-52.87	4884.321	64.603
-63.67	486.177	62.471
-61.87	682.651	62.458
-60.07	874.478	62.442
-58.27	1064.554	62.423
-56.47	1247.394	62.398
-52.87	1588.274	62.329
-49.27	1961.133	62.291
-45.67	2350.575	62.267
-42.07	2745.729	62.247
-38.47	3142.522	62.229
-34.87	3539.409	62.213
-31.27	3936.135	62.198
-27.67	4333.201	62.183
-24.07	4729.051	62.169
-20.47	5122.750	62.155
-34.87	500.974	59.489
-33.07	668.039	59.477
-31.27	831.548	59.464
-29.47	996.464	59.448
-27.67	1157.425	59.429
-24.07	1459.880	59.369
-20.47	1767.725	59.324
-16.87	2095.347	59.297
-13.27	2428.539	59.277
-9.67	2763.983	59.260
-6.07	3099.627	59.244
-2.47	3435.940	59.230
1.13	3772.143	59.216
4.73	4107.252	59.202
8.33	4442.673	59.189
11.93	4777.308	59.177

Table 8. Compressed Liquid-Phase (P, ρ ,T) Data for 0.49982 HFC-41 + 0.50018 R744 in PI Units (continued).

T	P	ρ
$^{\circ}\text{F}$	psia	lb-ft ⁻³
15.53	5111.327	59.164
-2.47	496.810	55.685
-0.67	632.938	55.675
1.13	767.838	55.664
2.93	901.475	55.652
4.73	1035.122	55.638
8.33	1293.947	55.599
11.93	1542.798	55.546
15.53	1801.780	55.513
19.13	2068.574	55.490
22.73	2338.559	55.473
26.33	2609.275	55.457
29.93	2880.602	55.443
33.53	3152.306	55.430
37.13	3424.020	55.417
40.73	3695.378	55.405
44.33	3966.829	55.393
47.93	4238.004	55.381
51.53	4508.954	55.370
55.13	4779.253	55.359
58.73	5048.910	55.348

Table 9: Vapor Pressures for HFC-41 from 219 to 312 K (-65 to 102 °F) using the New Vapor-Liquid Equilibrium Apparatus.

Temperature K	Vapor Pressure MPa	Temperature °F	Vapor Pressure psia
218.69	0.330	-66.05	47.8
231.96	0.554	-42.16	80.4
241.92	0.790	-24.22	114.6
251.86	1.096	-6.35	159.0
251.90	1.097	-6.28	159.1
251.96	1.099	-6.17	159.4
263.62	1.566	14.83	227.1
268.80	1.817	24.14	263.5
270.66	1.916	27.49	277.9
279.93	2.456	44.18	356.2
279.92	2.454	44.17	355.9
284.42	2.751	52.27	399.0
289.88	3.150	62.10	456.9
292.78	3.387	67.32	491.2
292.83	3.391	67.41	491.8
294.85	3.558	71.04	516.1
294.85	3.558	71.04	516.1
294.93	3.561	71.18	516.5
297.02	3.742	74.95	542.8
299.99	4.010	80.29	581.6
304.86	4.480	89.05	649.8
309.93	5.016	98.18	727.4
311.92	5.241	101.77	760.1

Table 10: Near-Saturation (p, ρ, T) Data for HFC-41 from 274 to 296 K (34 to 73°F).

Temperature K	Pressure MPa	Density kg/m ³	Temperature °F	Pressure psia	Density lb/ft ³
274.38	2.195	670.7	34.20	318.4	41.83
276.68	2.339	662.2	38.34	339.3	41.30
278.42	2.460	655.9	41.47	356.8	40.91
280.06	2.564	650.0	44.42	371.9	40.54
281.65	2.681	642.8	47.28	388.9	40.09
282.94	2.775	637.8	49.60	402.5	39.78
284.66	2.906	631.3	52.70	421.5	39.38
286.58	3.053	623.7	56.16	442.7	38.90
288.28	3.186	615.3	59.22	462.1	38.38
290.44	3.349	606.9	63.10	485.8	37.85
292.21	3.497	598.1	66.29	507.1	37.30
293.86	3.644	590.2	69.26	528.5	36.81
294.89	3.728	584.3	71.11	540.7	36.44
275.93	2.195	45.5	36.99	318.4	2.84
278.32	2.339	50.0	41.29	339.3	3.12
280.14	2.460	53.3	44.56	356.8	3.33
281.76	2.564	56.7	47.48	371.9	3.53
283.4	2.681	60.7	50.43	388.9	3.79
284.71	2.775	63.2	52.79	402.5	3.94
286.46	2.906	66.8	55.94	421.5	4.17
288.36	3.053	71.0	59.36	442.7	4.43
290.12	3.186	76.7	62.53	462.1	4.78
292.07	3.349	81.5	66.04	485.8	5.09
294.11	3.497	88.5	69.71	507.1	5.52
295.35	3.644	97.4	71.94	528.5	6.07
295.89	3.728	106.4	72.91	540.7	6.64

Table 11: Estimated Critical Parameters for HFC-41 from Bominaar *et al.*, from Biswas *et al.*, and from the Data in Tables 9 and 10 using the Method of Van Poolen and the Vapor Pressure Equation in this Report.

	Critical Temperature	Critical Pressure	Critical Density
This Work	317.28 ± 0.08 K (111.43 ± 0.14 °F)	5.897 ± 0.01 MPa (855.3 ± 1.4 psia)	316.5 ± 1.5 kg/m ³ (19.74 ± 0.9 lb/ft ³)
Bominaar <i>et al.</i>	317.38 K (converted to ITS-90) (111.61°F)	5.87 MPa (851 psia)	311.4 kg/m ³ (19.40 lb/ft ³)
Biswas <i>et al.</i>	317.40 K (converted to ITS-90) (111.65°F)	5.875 MPa (852.1 psia)	312.3 kg/m ³ (19.48 lb/ft ³)

Table 12: Calculated Vapor Pressures for HFC-41.

T (K)	P _{calc} (MPa)
131.35	0.00043
140.00	0.00126
150.00	0.00367
160.00	0.00923
170.00	0.02056
180.00	0.04143
190.00	0.07691
200.00	0.13332
210.00	0.21823
220.00	0.34027
230.00	0.50910
240.00	0.73523
250.00	1.03002
260.00	1.40572
270.00	1.87558
280.00	2.45416
290.00	3.15794
300.00	4.00640
310.00	5.02477
317.28	5.89700

T (°F)	P _{calc} (psia)
-223.24	0.062
-200.00	0.294
-180.00	0.889
-160.00	2.278
-140.00	5.118
-120.00	10.338
-100.00	19.153
-80.00	33.046
-60.00	53.746
-40.00	83.197
-20.00	123.537
0.00	177.089
20.00	246.381
40.00	334.197
60.00	443.691
80.00	578.614
100.00	743.912
111.43	855.288

Table 13. Saturated Liquid Data for Fluoromethane (HFC-41) in SI Units.

T	P	ρ
K	MPa	mol·dm ⁻³
130.785	0.0004	29.6135
140.784	0.0014	29.0599
152.713	0.0048	28.3897
166.519	0.0158	27.6038
181.998	0.0472	26.6759
197.962	0.1198	25.7135
217.537	0.3063	24.4261
236.983	0.6603	23.0421
256.759	1.2743	21.4523
275.799	2.1968	19.6684
285.861	2.8503	18.5469
297.612	3.7895	16.9509
308.959	4.9099	14.7304

Table 14. Saturated Liquid Data for Fluoromethane (HFC-41) in PI Units.

T	P	ρ
$^{\circ}\text{F}$	psia	$\text{lb}\cdot\text{ft}^{-3}$
-224.26	0.057	62.918
-206.26	0.200	61.741
-184.79	0.694	60.317
-159.94	2.287	58.648
-132.07	6.846	56.676
-103.34	17.376	54.631
-68.10	44.425	51.896
-33.10	95.768	48.956
2.50	184.822	45.578
36.77	318.619	41.788
54.88	413.401	39.405
76.03	549.620	36.014
96.46	712.121	31.297