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1220 S. St. Francis Dr., Santa Fe, NM
87505

State of New Mexico
Energy, Minerals and Natural Resources

Form C-103
Revised July 18, 2013

OIL CONSERVATION DIVISION
1220 South St. Francis Dr.
Santa Fe, NM 87505

SUNDRY NOTICES AND REPORTS ON WELLS (DO NOT USE THIS FORM FOR PROPOSALS TO DRILL OR TO DEEPEN OR PLUG BACK TO A DIFFERENT RESERVOIR. USE "APPLICATION FOR PERMIT" (FORM C-101) FOR SUCH PROPOSALS.)		WELL API NO. 30-025-40448
1. Type of Well: Oil Well <input type="checkbox"/> Gas Well <input checked="" type="checkbox"/> Other Acid Gas Injection		5. Indicate Type of Lease STATE <input type="checkbox"/> FEE <input checked="" type="checkbox"/>
2. Name of Operator Targa Northern Delaware, LLC		6. State Oil & Gas Lease No. NMLC063798
3. Address of Operator 3100 McKinnon Street, Suite 800, Dallas, TX 75201		7. Lease Name or Unit Agreement Name Red Hills AGI
4. Well Location Unit Letter <u>I</u> : <u>1600</u> feet from the <u>South</u> line and <u>150</u> feet from the <u>East</u> line Section <u>13</u> Township <u>24S</u> Range <u>33E</u> NMPM County <u>Lea</u>		8. Well Number <u>1</u>
11. Elevation (Show whether DR, RKB, RT, GR, etc.) 3580 ft GL		9. OGRID Number 331548
		10. Pool name or Wildcat Exploratory Cherry Canyon

12. Check Appropriate Box to Indicate Nature of Notice, Report or Other Data

NOTICE OF INTENTION TO:

PERFORM REMEDIAL WORK ☐ PLUG AND ABANDON ☐
 TEMPORARILY ABANDON ☐ CHANGE PLANS ☐
 PULL OR ALTER CASING ☐ MULTIPLE COMPL ☐
 DOWNHOLE COMMINGLE ☐
 CLOSED-LOOP SYSTEM ☐
 OTHER: ☐

SUBSEQUENT REPORT OF:

REMEDIAL WORK ☐ ALTERING CASING ☐
 COMMENCE DRILLING OPNS. ☐ P AND A ☐
 CASING/CEMENT JOB ☐
 OTHER: TAG Gas concentration & injection volume per R-13507F ☒

13. Describe proposed or completed operations. (Clearly state all pertinent details, and give pertinent dates, including estimated date of starting any proposed work). SEE RULE 19.15.7.14 NMAC. For Multiple Completions: Attach wellbore diagram of proposed completion or recompletion.

Six month report of TAG composition and injection volumes from the Red Hills Plant being injected into the Red Hills AGI #1 as required by NMOCC Order R-13507 item F and agreements with NMOCD staff.

During the period of July - December 2023 the measured H₂S concentrations in the TAG ranged from 4.15% to 9.31% with an average value of 6.75% as derived from direct sampling and analysis of the TAG entering the well. Appendix A table 1 details the gas analysis of twelve TAG samples Targa Northern Delaware had taken during the report period to measure H₂S concentration directly. Average daily TAG volume injected is about 1371.6 MSCFD for the reporting period.

This report is submitted to fulfill the reporting requirement established by NMOCD for sampling of TAG concentrations every six-months beginning in June 2018. The following information is contained herein:

1. Measured TAG concentrations and volumes for each of the thirteen TAG sampling events (Appendix A, Table 1)
2. Graph of TAG volumes July 1, 2023 – December 31, 2023 (Appendix A, Figure 1)
3. C6+ Gas/Vapor Fractional Analysis report for each sample date (Appendix B)
4. Anticipated range of H₂S concentrations in TAG under normal operating conditions.
5. Technical Explanations for compositions. (Appendix C)

Attachment A to this C-103 includes all supporting analyses and data. NMOCD requested that sampling be done and reported any time a major source change occurs and every six months normally. These results will be submitted to Santa Fe and the Hobbs District office on a C-103 form to be incorporated into the well file by NMOCD upon receipt.

Based on an analysis of the data attached herein, Targa anticipates the H₂S concentrations being injected into the Red Hills AGI #1 to range between 4.43% and 15.79%. Targa will notify the NM OCD if concentrations differ substantially based on inlet gas changes or gathering system updates.

Spud Date:

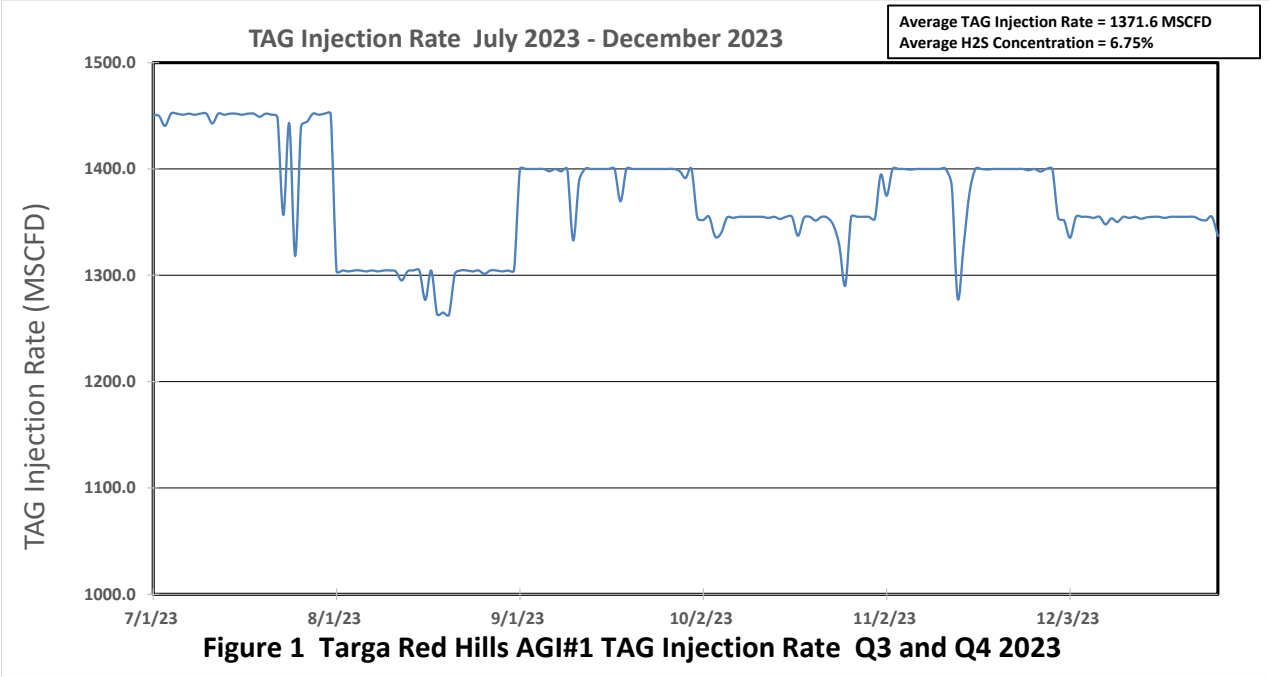
Rig Release Date:

I hereby certify that the information above is true and complete to the best of my knowledge and belief.SIGNATURE Matt Eales TITLE VP of Regulatory DATE January 30, 2024Type or print name Matt Eales E-mail address: meales@targaresources.com PHONE: 832.496.7513For State Use Only

APPROVED BY: _____ TITLE _____ DATE _____

Conditions of Approval (if any):

Appendix A: Summarized TAG Concentrations and Injection Volumes for Red Hills AGI #1



TAG Concentration

Date	H ₂ S %	CO ₂ %
7/12/2023	5.74	85.63
7/26/2023	6.59	85.90
8/9/2023	5.74	89.44
8/23/2023	7.98	90.87
9/6/2023	6.49	92.10
9/20/2023	6.62	92.12
10/4/2023	4.15	83.86
10/18/2023	9.31	89.63
11/1/2023	6.80	91.83
11/29/2023	7.91	90.74
12/13/2023	7.49	91.21
12/27/2023	6.15	69.29
Average	6.75	87.72

Appendix B: Red Hills AGI #1 C6+ Gas/Vapor Fractional Analysis by Date

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	85 F
Site Type	Plant	Collection Date	07/12/2023
Sample Point	Inlet to Compressor	Collection Time	8:02 AM
Spot/Comp	Spot	Collection By	Mike McKinney
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS013

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	7.267	4.811	0.801
CARBON DIOXIDE	CO2	85.632	89.055	14.675
HYDROGEN SULFIDE	H2S	5.739	4.622	0.777
METHANE	C1	0.463	0.176	0.079
ETHANE	C2	0.130	0.092	0.035
PROPANE	C3	0.063	0.066	0.017
I-BUTANE	iC4	0.453	0.622	0.149
N-BUTANE	nC4	0.029	0.040	0.009
I-PENTANE	iC5	0.013	0.022	0.005
N-PENTANE	nC5	0.018	0.031	0.007
HEXANES PLUS	C6+	0.193	0.463	0.086
TOTALS:		100.000	100.000	16.640

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.308	0.273	0.256	0.098	0.140	0.109

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	73.28	1.469	0.995	42.318	60.47
WATER SATURATED	72.91	1.455	0.994	41.583	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	75 F
Site Type	Plant	Collection Date	07/26/2023
Sample Point	Inlet to Compressor	Collection Time	7:44 AM
Spot/Comp	Spot	Collection By	Mike McKinney
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS007

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	7.020	4.658	0.774
CARBON DIOXIDE	CO2	85.903	89.557	14.720
HYDROGEN SULFIDE	H2S	6.586	5.317	0.892
METHANE	C1	0.254	0.097	0.043
ETHANE	C2	0.060	0.043	0.016
PROPANE	C3	0.025	0.026	0.007
I-BUTANE	iC4	0.052	0.072	0.017
N-BUTANE	nC4	0.009	0.012	0.003
I-PENTANE	iC5	0.002	0.003	0.001
N-PENTANE	nC5	0.002	0.003	0.001
HEXANES PLUS	C6+	0.087	0.212	0.039
TOTALS:		100.000	100.000	16.513

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.084	0.068	0.061	0.041	0.059	0.046

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	53.63	1.465	0.995	42.214	44.31
WATER SATURATED	53.60	1.451	0.994	41.481	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	78 F
Site Type	Plant	Collection Date	08/09/2023
Sample Point	Inlet to Compressor	Collection Time	8:38 AM
Spot/Comp	Spot	Collection By	Mike McKinney
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS027

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	3.830	2.505	0.422
CARBON DIOXIDE	CO2	89.438	91.895	15.331
HYDROGEN SULFIDE	H2S	5.743	4.569	0.778
METHANE	C1	0.424	0.159	0.072
ETHANE	C2	0.112	0.079	0.030
PROPANE	C3	0.054	0.056	0.015
I-BUTANE	iC4	0.170	0.231	0.056
N-BUTANE	nC4	0.024	0.033	0.008
I-PENTANE	iC5	0.006	0.010	0.002
N-PENTANE	nC5	0.016	0.027	0.006
HEXANES PLUS	C6+	0.183	0.436	0.081
TOTALS:		100.000	100.000	16.801

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.198	0.168	0.153	0.089	0.129	0.100

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	62.06	1.487	0.994	42.833	50.90
WATER SATURATED	61.89	1.472	0.994	42.088	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	72 F
Site Type	Plant	Collection Date	08/23/2023
Sample Point	Inlet to Compressor	Collection Time	8:23 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS011

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.387	0.251	0.043
CARBON DIOXIDE	CO2	90.868	92.678	15.581
HYDROGEN SULFIDE	H2S	7.976	6.299	1.081
METHANE	C1	0.330	0.123	0.056
ETHANE	C2	0.087	0.061	0.023
PROPANE	C3	0.050	0.051	0.014
I-BUTANE	iC4	0.140	0.189	0.046
N-BUTANE	nC4	0.022	0.030	0.007
I-PENTANE	iC5	0.010	0.017	0.004
N-PENTANE	nC5	0.012	0.020	0.004
HEXANES PLUS	C6+	0.118	0.281	0.052
TOTALS:		100.000	100.000	16.911

Value of "0.000" in fractional interpreted as below detectable limit.
 If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.150	0.127	0.113	0.060	0.087	0.067

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	70.26	1.498	0.994	43.150	57.40
WATER SATURATED	69.95	1.484	0.994	42.400	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	16 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	84 F
Site Type	Plant	Collection Date	09/06/2023
Sample Point	Inlet to Compressor	Collection Time	9:47 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS004

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.350	0.226	0.039
CARBON DIOXIDE	CO2	92.099	93.506	15.793
HYDROGEN SULFIDE	H2S	6.491	5.103	0.880
METHANE	C1	0.334	0.124	0.057
ETHANE	C2	0.069	0.048	0.019
PROPANE	C3	0.034	0.035	0.009
I-BUTANE	iC4	0.472	0.633	0.155
N-BUTANE	nC4	0.014	0.019	0.004
I-PENTANE	iC5	0.019	0.032	0.007
N-PENTANE	nC5	0.009	0.015	0.003
HEXANES PLUS	C6+	0.109	0.259	0.048
TOTALS:		100.000	100.000	17.014

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.245	0.226	0.217	0.058	0.082	0.063

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	70.41	1.505	0.994	43.347	57.39
WATER SATURATED	70.10	1.491	0.994	42.594	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	15 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	77 F
Site Type	Plant	Collection Date	09/20/2023
Sample Point	Inlet to Compressor	Collection Time	9:39 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS017

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.472	0.305	0.052
CARBON DIOXIDE	CO2	92.121	93.661	15.796
HYDROGEN SULFIDE	H2S	6.616	5.209	0.897
METHANE	C1	0.307	0.114	0.052
ETHANE	C2	0.049	0.034	0.013
PROPANE	C3	0.025	0.025	0.007
I-BUTANE	iC4	0.279	0.375	0.092
N-BUTANE	nC4	0.021	0.028	0.007
I-PENTANE	iC5	0.009	0.015	0.003
N-PENTANE	nC5	0.008	0.013	0.003
HEXANES PLUS	C6+	0.093	0.221	0.042
TOTALS:		100.000	100.000	16.964

Value of "0.000" in fractional interpreted as below detectable limit.
If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.167	0.154	0.147	0.048	0.070	0.054

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	62.90	1.503	0.994	43.287	51.30
WATER SATURATED	62.71	1.488	0.994	42.534	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	76 F
Site Type	Plant	Collection Date	10/04/2023
Sample Point	Inlet to Compressor	Collection Time	10:09 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.730 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS027

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	3.116	2.135	0.343
CARBON DIOXIDE	CO2	83.862	90.286	14.370
HYDROGEN SULFIDE	H2S	4.152	3.461	0.562
METHANE	C1	8.192	3.215	1.396
ETHANE	C2	0.369	0.271	0.099
PROPANE	C3	0.063	0.068	0.017
I-BUTANE	iC4	0.027	0.038	0.009
N-BUTANE	nC4	0.016	0.023	0.005
I-PENTANE	iC5	0.005	0.009	0.002
N-PENTANE	nC5	0.005	0.009	0.002
HEXANES PLUS	C6+	0.193	0.485	0.085
TOTALS:		100.000	100.000	16.890

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.219	0.120	0.103	0.089	0.109	0.099

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	130.86	1.418	0.995	40.879	109.87
WATER SATURATED	129.52	1.405	0.994	40.168	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	78 F
Site Type	Plant	Collection Date	10/18/2023
Sample Point	Inlet to Compressor	Collection Time	10:50 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.696 psi
Purchaser		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS011

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.469	0.306	0.052
CARBON DIOXIDE	CO2	89.631	91.805	15.334
HYDROGEN SULFIDE	H2S	9.306	7.381	1.258
METHANE	C1	0.325	0.121	0.055
ETHANE	C2	0.072	0.050	0.019
PROPANE	C3	0.028	0.029	0.008
I-BUTANE	iC4	0.081	0.110	0.027
N-BUTANE	nC4	0.009	0.012	0.003
I-PENTANE	iC5	0.002	0.003	0.001
N-PENTANE	nC5	0.002	0.003	0.001
HEXANES PLUS	C6+	0.075	0.180	0.033
TOTALS:		100.000	100.000	16.791

Value of "0.000" in fractional interpreted as below detectable limit.
 If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.092	0.073	0.065	0.035	0.051	0.040

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	72.26	1.492	0.994	42.968	59.16
WATER SATURATED	71.92	1.478	0.994	42.219	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	41 F
Site Type	Plant	Collection Date	11/01/2023
Sample Point	Inlet to Compressor	Collection Time	9:39 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.696 psi
Regulatory ID		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS032

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.420	0.272	0.046
CARBON DIOXIDE	CO2	91.832	93.474	15.710
HYDROGEN SULFIDE	H2S	6.797	5.357	0.919
METHANE	C1	0.457	0.170	0.078
ETHANE	C2	0.129	0.090	0.035
PROPANE	C3	0.069	0.070	0.019
I-BUTANE	iC4	0.050	0.067	0.016
N-BUTANE	nC4	0.043	0.058	0.014
I-PENTANE	iC5	0.022	0.037	0.008
N-PENTANE	nC5	0.022	0.037	0.008
HEXANES PLUS	C6+	0.159	0.368	0.070
TOTALS:		100.000	100.000	16.923

Value of "0.000" in fractional interpreted as below detectable limit.

If Onsite H2S testing is performed, its resulting value is used in fractional table

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.170	0.135	0.116	0.086	0.118	0.094

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	65.81	1.501	0.994	43.237	53.71
WATER SATURATED	65.57	1.487	0.994	42.484	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	53 F
Site Type	Plant	Collection Date	12/13/2023
Sample Point	Inlet to Compressor	Collection Time	8:35 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.696 psi
Regulatory ID		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS002

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.394	0.256	0.043
CARBON DIOXIDE	CO2	91.214	93.195	15.603
HYDROGEN SULFIDE	H2S	7.494	5.929	1.013
METHANE	C1	0.641	0.239	0.109
ETHANE	C2	0.067	0.047	0.018
PROPANE	C3	0.032	0.033	0.009
I-BUTANE	iC4	0.003	0.004	0.001
N-BUTANE	nC4	0.020	0.027	0.006
I-PENTANE	iC5	0.067	0.112	0.025
N-PENTANE	nC5	0.006	0.010	0.002
HEXANES PLUS	C6+	0.062	0.148	0.028
TOTALS:		100.000	100.000	16.857

Value of "0.000" in fractional interpreted as below detectable limit. Onsite H2S value is used in fractional table if performed.

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.089	0.071	0.062	0.055	0.065	0.051

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	63.71	1.496	0.994	43.074	52.10
WATER SATURATED	63.51	1.481	0.994	42.324	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	50 F
Site Type	Plant	Collection Date	11/29/2023
Sample Point	Inlet to Compressor	Collection Time	10:03 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.696 psi
Regulatory ID		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS041

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	0.479	0.311	0.053
CARBON DIOXIDE	CO2	90.735	92.508	15.523
HYDROGEN SULFIDE	H2S	7.906	6.242	1.069
METHANE	C1	0.340	0.126	0.058
ETHANE	C2	0.059	0.041	0.016
PROPANE	C3	0.033	0.034	0.009
I-BUTANE	iC4	0.136	0.183	0.045
N-BUTANE	nC4	0.024	0.032	0.008
I-PENTANE	iC5	0.218	0.364	0.080
N-PENTANE	nC5	0.008	0.013	0.003
HEXANES PLUS	C6+	0.062	0.146	0.027
TOTALS:		100.000	100.000	16.891

Value of "0.000" in fractional interpreted as below detectable limit. Onsite H2S value is used in fractional table if performed.

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.188	0.172	0.163	0.110	0.138	0.050

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	73.80	1.499	0.994	43.167	60.28
WATER SATURATED	73.43	1.484	0.994	42.415	

SAMPLE ID		COLLECTION DATA	
Operator	Targa Resources Inc	Pressure	12 psig
Location	Red Hills Processing Complex	Sample Temp	N/A
Site	AGI Plant	Atm Temp	37 F
Site Type	Plant	Collection Date	12/27/2023
Sample Point	Inlet to Compressor	Collection Time	9:42 AM
Spot/Comp	Spot	Collection By	Dakota Kiser
Meter ID		Pressure Base	14.696 psi
Regulatory ID		Temperature Base	60 F
Fluid	Gas	Container(s)	PLS041

GPA 2261 Gas Fractional Analysis

COMPOUND	FORMULA	MOL%	WT%	GPM
NITROGEN	N2	20.778	14.512	2.283
CARBON DIOXIDE	CO2	69.288	76.030	11.836
HYDROGEN SULFIDE	H2S	6.147	5.223	0.830
METHANE	C1	1.303	0.521	0.221
ETHANE	C2	0.662	0.496	0.177
PROPANE	C3	0.545	0.599	0.151
I-BUTANE	iC4	0.094	0.136	0.031
N-BUTANE	nC4	0.292	0.423	0.092
I-PENTANE	iC5	0.096	0.173	0.035
N-PENTANE	nC5	0.107	0.192	0.039
HEXANES PLUS	C6+	0.688	1.695	0.300
TOTALS:		100.000	100.000	15.995

Value of "0.000" in fractional interpreted as below detectable limit. Onsite H2S value is used in fractional table if performed.

LIQUID YIELD	C2+	C3+	C4+	C5+	26# Liquid	10# Liquid
GAL/MSCF (GPM)	0.825	0.648	0.497	0.374	0.510	0.406

GPA 2172/ASTM D3588 CALCULATED PROPERTIES

WATER CONTENT	BTU/CF	Specific Gr.	Z Factor	Mol Weight	Wobbe IDX
DRY	136.14	1.391	0.995	40.109	115.45
WATER SATURATED	134.70	1.378	0.995	39.410	

Appendix C: Technical Explanations for Compositions

2. a. High N₂ concentration indicates insufficient purging of the sample bottle by the 3rd party technician, resulting in small quantities of air contamination. It's an issue that can be difficult to prevent given that the sample bottles need to be purged to atmosphere with lethal concentrations of acid gas.

- See attached for normalized estimates of the sample compositions, adjusting for removal of N₂ contamination.

	Sample: 7-12-2023			Sample: 12-27-2023	
	mol-%	corrected mol-%		mol-%	corrected mol-%
Nitrogen	7.267	0.000		20.778	0.000
Carbon Dioxide	85.632	92.343		69.288	87.461
Hydrogen Sulfide	5.739	6.189		6.147	7.759
Methane	0.463	0.499		1.303	1.645
Ethane	0.130	0.140		0.662	0.836
Propane	0.063	0.068		0.545	0.688
I-Butane	0.453	0.488		0.094	0.119
N-Butane	0.029	0.031		0.292	0.369
I-Pentane	0.013	0.014		0.096	0.121
N-Pentane	0.018	0.019		0.107	0.135
Hexanes Plus	0.193	0.208		0.688	0.868
Totals	100.000	100.000		100.000	100.000

2. b. Regarding low H₂S concentrations, we've generally observed a decline in the inlet H₂S concentration delivered from the field. We've also reduced our CO₂ slip, as treating and well capacity allows, which results in a higher concentration of CO₂.

2. c. The addition of small quantities of H₂S to CO₂ do not significantly enlarge the P–T region of fluid immiscibility. Consequently, H₂S is of lesser concern with regard to phase equilibrium effects on compression requirements. The documentation below supports this statement. If you would like more information, we should be able to use CMG-Winprop to create PVT data with different ratio of H₂S content.

perfect solution laws. That this is true, even for those members, namely, ethane and propane, which have about the same boiling point or volatility as hydrogen sulfide, is shown by the fact that an azeotrope is formed in these systems. However, it is known that in binary systems of this kind the formation of an azeotrope is limited to those members of the series which are within a certain characteristic boiling point range or volatility range of the common component.

The work of Reamer, Sage, and Lacey on the methane-hydrogen sulfide system (4) showed that an azeotrope is not formed between methane and hydrogen sulfide. It is presumed, therefore, that the difference in volatility of the components is greater than the characteristic range. Figure 7 shows the critical loci of the methane-, ethane-, and propane-hydrogen sulfide systems together with a portion of the vapor pressure curves of the components. The critical locus of the methane-hydrogen sulfide system shows a maximum pressure point which is considerably above the critical pressure of either methane or hydrogen sulfide. In this respect the behavior is similar to that observed in methane-hydrocarbon systems (5). As the volatility of the hydrocarbon approaches that of hydrogen sulfide the nature of the critical locus is greatly affected. The maximum pressure point disappears and the curve appears to swing around the critical point of hydrogen sulfide as an axis, becoming almost linear for the ethane-hydrogen sulfide system, then looping to form a minimum point in critical temperature for the propane-hydrogen sulfide system. While the *n*-butane- and higher member-hydro-

gen sulfide systems have not been studied, it seems reasonable to suppose that their critical loci would vary in such a way that the minimum critical temperature point would disappear and a maximum pressure point reappear, probably, in the *n*-pentane- or *n*-hexane-hydrogen sulfide system. This is the behavior exhibited in *n*-paraffin hydrocarbon-carbon dioxide systems as shown by Poettmann and Katz (3), and may be considered as a characteristic pattern of behavior for binary systems formed by a compound termed an "azeotropic agent" with a series of homologous compounds.

ACKNOWLEDGMENT

Grateful acknowledgment is made to the Procter & Gamble Co. for financial aid in the form of a fellowship to one of the authors and to the Phillips Petroleum Co. for furnishing the sample of ethane.

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Phase-Equilibrium Properties of System Carbon Dioxide-Hydrogen Sulfide

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HYDROGEN sulfide and carbon dioxide, which are present in many natural petroleum reservoirs, tend to concentrate with the light hydrocarbon fractions during the refining of the crudes. The recovery of the hydrogen sulfide is becoming a matter of increasing commercial importance because traditional sources of sulfur are no longer adequate to meet the expanding demands for this element. For this reason, it is desirable to have available phase and equilibrium data for mixtures of hydrogen sulfide with the various volatile components from which it is to be separated.

A number of recent investigations have been devoted to the determination of the behavior of binary systems of hydrogen sulfide with the lower paraffin hydrocarbons under high pressure. Gilliland and Scheeline (4) have made *P-T-x-y* measurements on the system propane-hydrogen sulfide, using an equilibrium still; their experiments were confined to propane-rich mixtures and to constant pressures of 400, 500, and 600 pounds per square inch. Kay and Brice (7) and Kay and Rambosek (8) obtained complete *P-V-T-x-y* data for saturated mixtures of hydrogen sulfide with ethane and with propane in the temperature region above the ice point. Reamer, Sage, and Lacey (10) made similar measurements in the methane-hydrogen sulfide system, including extensive compressibility data for superheated states. These researches comprise advances toward the accumulation of a reliable fund of data for the processing of multicomponent systems containing hydrogen sulfide.

The present investigation of the hydrogen sulfide-carbon dioxide system is intended to contribute further to the attainment of this goal. The only information heretofore available on the system in question is due to Steckel (11), who determined the isothermal dew and bubble point curves at 0° C. and two lower temperatures; these temperatures are below the range that is ordinarily of practical interest to the chemical industry.

In the present work, the pressure-volume-temperature behavior was determined for the saturated states of eight mixtures of hydrogen sulfide and carbon dioxide at temperatures between the ice point and the eiccondentherm. These data suffice to define the equilibrium-state properties of the system to a degree of precision suitable for most scientific and engineering purposes.

MATERIALS

The hydrogen sulfide and carbon dioxide used in the experiments were commercial products which were carefully purified and sealed as gases in 50-ml. glass ampoules until used.

In brief, the purification procedure consisted in introducing the dry crude gas into a highly evacuated all-glass system communicating via a stopcock (initially closed) to an evacuated manifold onto which were sealed the ampoules intended for the storage of the end product. The gas was condensed as a solid in a cold finger by use of liquid nitrogen. The finger was then allowed to warm up slowly, evaporating the solid. The first and last portions of the gas to evaporate were pumped away. The middle fraction was condensed in a second cold finger. The sublimation between the two cold fingers was continued until the light and heavy contaminants were believed to be thoroughly stripped out. The stopcock to the manifold was then opened and a small quantity of the pure gas was con-

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March 1953

INDUSTRIAL AND ENGINEERING CHEMISTRY

619

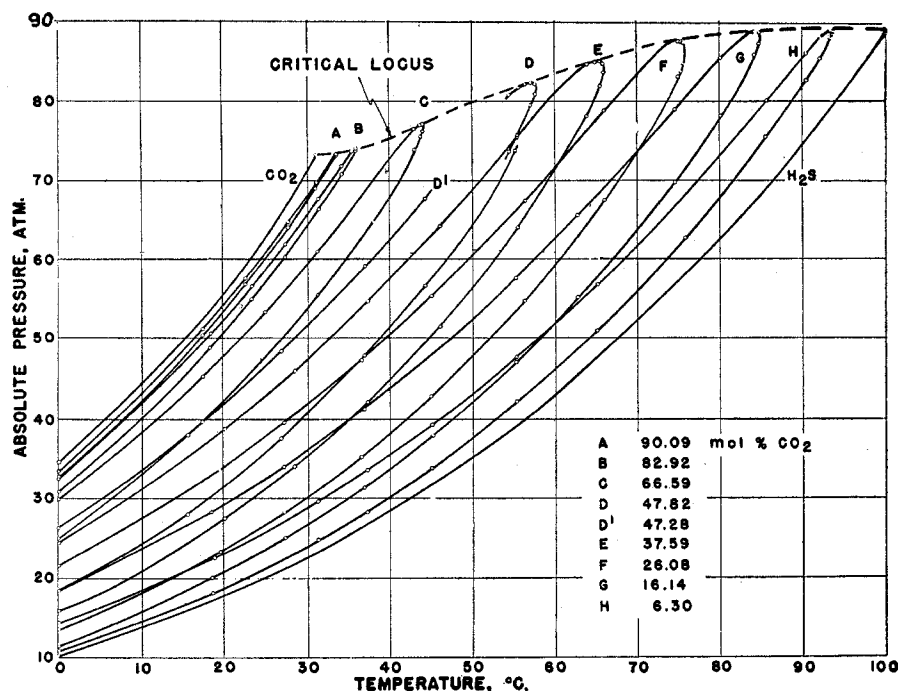


Figure 1. Pressure-Temperature Diagram

densed in the tip of each storage ampoule; the gas was allowed to evaporate and was pumped away in order to scavenge residual air from the manifold and ampoules. After several scavengings, each of the storage ampoules was charged with pure gas by condensing a small quantity of solid in the ampoule and then sealing and pulling off the ampoule with a hand torch. A sufficient mass of pure substance was condensed in each ampoule to provide a gas pressure of about 2 atmospheres (absolute) at room temperature.

Prior to use in the present investigation, a sample of each purified gas was tested for contamination by measuring, in the experimental apparatus, the difference between the bubble and dew point pressures at the temperature of melting ice. The critical pressures and temperatures were also determined for comparison with the accepted values reported in the literature. Table I summarizes the results of these tests.

TABLE I. PROPERTIES OF MATERIALS

	Hydrogen Sulfide		Carbon Dioxide	
	Present work	Literature (8)	Present work	Literature (8)
Vapor pressure at ice point, atm.	10.38 (b.p.)	10.33	34.42 (b.p.)	34.28
Pressure rise on condensation, atm.	10.23 (d.p.)		34.35 (d.p.)	
	0.15 (0° C.)		0.07 (0° C.)	
Critical pressure, atm.	88.92	88.87	73.02	72.95
Critical temperature, °C.	100.38	100.4	31.10	31.1

EXPERIMENTAL METHODS

The experimental procedure consisted of confining a known mass of sample, of definite composition, over mercury in a glass capillary tube. The sample was thermostated, and a known pressure was applied. After equilibrium was attained, the volume was determined by measuring the length of the column of sample; this length was related to the volume by a prior calibration of the tube. In this method, the dew and bubble point phenomena can be directly observed through the tube walls as the pressure is changed, and it is possible to bracket the *P-V-T* values for the saturated states of the fluid as exactly as the experimenter desires. An adequate description of the apparatus is available (8).

In the present work, the temperature was measured with a seven-junction copper-constantan thermocouple and a potentiometer reading to 1 microvolt. This equipment was sensitive to a temperature change of 0.003° C.; it was calibrated at the ice point and against the saturated vapors of acetone, water, naphthalene, and benzophenone condensing at known pressures near 1 atmosphere. The temperatures indicated by the thermel are believed to deviate not more than 0.02° from the true thermodynamic temperature within the range in which the thermel was used (0° to 100° C.). Pressures were determined with a dead-weight piston gage, suitable corrections being applied for the various hydrostatic heads existing between the sample and the gage. The gage was calibrated by comparison with a master instrument of the same type, which had been tested against the vapor pressure of pure carbon dioxide at the ice point, as recommended by Bridgeman (1), to ensure an accuracy of 1 part in 10,000. A cathetometer was used to measure the sample volume and to determine the hydrostatic heads which contributed to the measured pressure. The reproducibility of the length measurements was about ±0.1 mm., corresponding to a precision in volume measurement of about 0.5% at the smallest volumes observed and about 0.02% at the largest ones.

The gas samples were introduced into the experimental tube by pneumatic trough techniques which have been described in detail by Kay (6). Each pure gas was injected separately into the mercury-filled tube from a microburet in which the mass of gas was determined by simultaneous observations of its temperature, pressure, and volume; the experimental compressibility factors reported by Cooper and Maass (2) for carbon dioxide and by Wright and Maass (12) for hydrogen sulfide were used to modify the ideal gas law. Enough of each gas

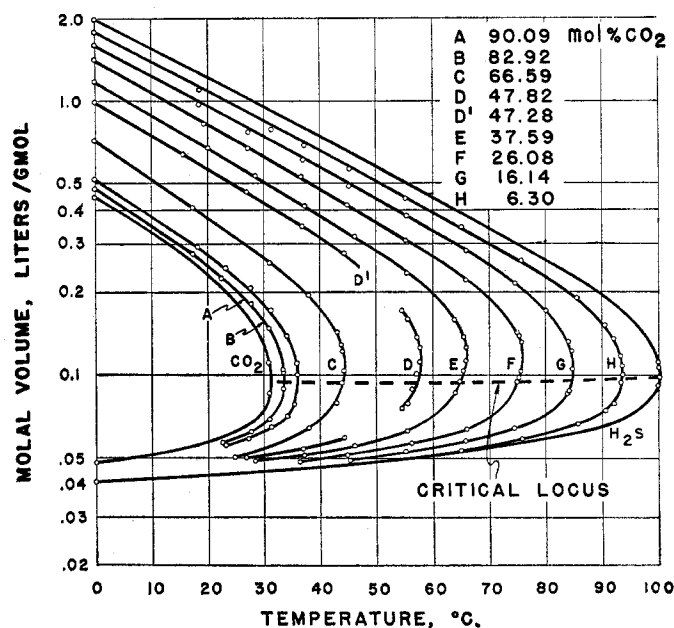


Figure 2. Volume-Temperature Diagram

was added to the experimental tube to make a mixture of the approximate composition desired; when possible, the composition of the sample was altered by diluting the mixture already present in the tube with one of the pure components. Experience accumulated by previous workers indicates that the uncertainty in the composition of mixtures prepared by these techniques does not exceed 0.0005 mole fraction.

TABLE II. UNIQUE STATES OF HYDROGEN SULFIDE-CARBON DIOXIDE SYSTEM

Composi- tion, Mole Fr. CO ₂	Critical Point			Cricondenbar Point			Cricondentherm Point		
	T _c , ° C.	P _c , Atm.	V _c , liter/ gram-mole	T _c , ° C.	P _c , Atm.	V _c , liter/ gram-mole	T _c , ° C.	P _c , Atm.	V _c , liter/ gram-mole
0.0000	100.38	88.87	0.0977	100.38	88.87	0.0977	100.38	88.87	0.0977
0.0630	93.50	88.79	0.0954	93.54	88.81	0.0956	97.2	88.5	0.100
0.1614	84.16	88.60	0.0942	84.30	88.61	0.0952	84.9	87.9	0.110
0.2608	74.48	87.86	0.0936	75.00	87.41	0.0967	76.1	86.6	0.118
0.3759	64.74	84.74	0.0931	65.43	84.82	0.0993	66.0	84.0	0.122
0.4728	56.98	82.12	0.0930	57.59	82.20	0.1005	57.7	81.2	0.118
0.6659	43.72	76.83	0.0931	44.07	76.87	0.1008	44.2	75.9	0.108
0.8292	35.96	73.85	0.0935	36.00	73.87	0.0966	36.1	73.7	0.099
0.9009	33.53	73.19	0.0938	33.55	73.20	0.0954	33.6	73.0	0.096
1.0000	31.10	72.95	0.0946	31.10	72.95	0.0946	31.10	72.95	0.0946

until a change in the number of phases was observed. Dew and bubble point data for each mixture were determined for approximate 10° intervals of temperature, except near the critical point, where the temperature increment was much reduced in order to locate the unique states of the fluid precisely.

EQUILIBRIUM DATA

The pressures and specific volumes of the saturated states

In making a determination, the temperature of the sample was brought to the desired level and the pressure in the system was adjusted to the approximate saturation value. It was found convenient to approach both the dew and bubble points in the direction of increasing pressure; the equilibria appeared to be established somewhat more quickly when approached in this way. The customary procedure was to begin measurements at about 0.05 atmosphere less than saturation pressure and to increase the pressure in increments of 0.01 atmosphere

of eight mixtures of hydrogen sulfide and carbon dioxide were determined for temperatures ranging between the ice point and the cricondentherm. In the case of the pure components, measurements were not extended beyond the purity checks already described. Inasmuch as satisfactory agreement was obtained with the values cited in the literature, the data of Reamer *et al.* (9) and of the International Critical Tables (5) were used for

TABLE III. PROPERTIES OF SATURATED STATES

Composi- tion, Mole Fr. CO ₂	Saturated Liquid			Saturated Vapor			Composi- tion, Mole Fr. CO ₂	Saturated Liquid			Saturated Vapor		
	Pressure, Atm.	T _c , ° C.	V _c , liter/ gram-mole	T _c , ° C.	V _c , liter/ gram-mole			Pressure, Atm.	T _c , ° C.	V _c , liter/ gram-mole	T _c , ° C.	V _c , liter/ gram-mole	
0.0000	15	13.28	0.0419	13.28	1.307		0.3759	20	8.16	0.930	
	20	24.50	0.0434	24.50	0.978			25	1.10	0.0424	16.47	0.758	
	25	33.64	0.0449	33.64	0.790			30	8.76	0.0435	23.37	0.606	
	30	41.55	0.0463	41.55	0.674			35	15.61	0.0447	29.34	0.510	
	35	48.58	0.0477	48.58	0.579			40	21.81	0.0461	35.35	0.437	
	40	54.99	0.0490	54.99	0.499			45	27.53	0.0476	40.33	0.378	
	45	60.83	0.0504	60.83	0.431			50	32.84	0.0492	44.88	0.331	
	50	66.28	0.0520	66.28	0.375			55	37.80	0.0509	49.01	0.291	
	55	71.39	0.0536	71.39	0.330			60	42.46	0.0528	52.78	0.257	
	60	76.20	0.0554	76.20	0.290			65	46.89	0.0550	56.25	0.277	
	65	80.76	0.0575	80.76	0.256			70	51.14	0.0576	59.45	0.200	
0.0630	15	1.66	0.0407	11.67	1.301		0.4728	20	3.30	0.913	
	20	13.63	0.0421	22.79	0.974			25	11.70	0.726	
	25	23.52	0.0435	32.00	0.773			30	5.52	0.0438	18.42	0.595	
	30	32.02	0.0449	39.85	0.636			35	12.11	0.0450	24.31	0.501	
	35	39.58	0.0463	46.92	0.537			40	18.05	0.0464	29.58	0.429	
	40	46.30	0.0476	53.17	0.461			45	23.49	0.0479	34.40	0.372	
	45	52.46	0.0490	58.89	0.400			50	28.53	0.0496	38.86	0.326	
	50	58.14	0.0505	64.16	0.350			55	33.27	0.0516	43.04	0.287	
	55	63.45	0.0521	69.04	0.309			60	37.78	0.0539	46.93	0.253	
	60	68.43	0.0540	73.60	0.274			65	42.11	0.0565	50.45	0.221	
	65	73.11	0.0562	77.81	0.244			70	46.29	0.0591	53.06	0.192	
0.1614	15	7.72	1.288		0.6659	25	0.12	0.680	
	20	3.36	0.0412	18.62	0.967			30	0.38	0.0446	6.69	0.559	
	25	13.04	0.0425	27.65	0.767			35	6.50	0.0459	12.52	0.469	
	30	21.44	0.0438	35.44	0.631			40	12.10	0.0474	17.77	0.399	
	35	28.90	0.0450	42.12	0.530			45	17.24	0.0490	22.51	0.344	
	40	35.65	0.0463	47.97	0.454			50	22.00	0.0508	26.86	0.297	
	45	41.79	0.0476	53.25	0.393			55	26.44	0.0529	30.91	0.258	
	50	47.42	0.0491	58.20	0.343			60	30.61	0.0554	34.70	0.222	
	55	52.64	0.0507	62.87	0.303			65	34.54	0.0586	38.22	0.190	
	60	57.59	0.0524	67.21	0.268			70	38.21	0.0630	41.00	0.162	
	65	62.34	0.0544	71.20	0.238			75	42.16	0.0731	43.61	0.132	
0.2608	15	3.56	1.271		0.8292	35	3.08	0.0468	4.90	0.445	
	20	13.94	0.952			40	8.40	0.0484	10.14	0.377	
	25	6.26	0.0422	22.51	0.753			45	13.29	0.0501	14.93	0.324	
	30	14.36	0.0434	29.96	0.619			50	17.80	0.0519	19.32	0.279	
	35	21.54	0.0446	36.42	0.522			55	22.00	0.0542	23.37	0.241	
	40	28.07	0.0459	42.16	0.447			60	25.95	0.0570	27.14	0.208	
	45	34.08	0.0473	47.35	0.389			65	29.67	0.0612	30.63	0.176	
	50	39.62	0.0488	52.13	0.341			70	33.12	0.0678	33.65	0.145	
	55	44.76	0.0504	56.56	0.300		0.9009	35	1.92	0.0472	2.70	0.438	
	60	49.60	0.0523	60.64	0.265			40	7.21	0.0488	7.97	0.372	
	65	54.23	0.0544	64.35	0.233			45	12.04	0.0505	12.75	0.318	
	70	58.67	0.0569	68.59	0.205			50	16.48	0.0525	17.13	0.274	
	75	62.96	0.0598	72.82	0.180			55	20.61	0.0548	21.19	0.236	
	80	67.18	0.0636	77.00	0.156			60	24.43	0.0579	25.00	0.203	
	85	71.84	0.0705	81.15	0.135			65	28.00	0.0625	28.46	0.171	
				83.76	0.125			70	31.42	0.0700	31.71	0.136	
							1.0000	35	0.65	0.0478	0.65	0.440	
								40	5.80	0.0495	5.80	0.379	
								45	10.51	0.0514	10.51	0.327	
								50	14.85	0.0536	14.85	0.282	
								55	18.82	0.0562	18.82	0.243	
								60	22.52	0.0593	22.52	0.210	
								65	26.01	0.0635	26.01	0.182	
								70	29.26	0.0705	29.26	0.149	
								72.95	31.10	0.0946	31.10	0.0946	

March 1953

INDUSTRIAL AND ENGINEERING CHEMISTRY

621

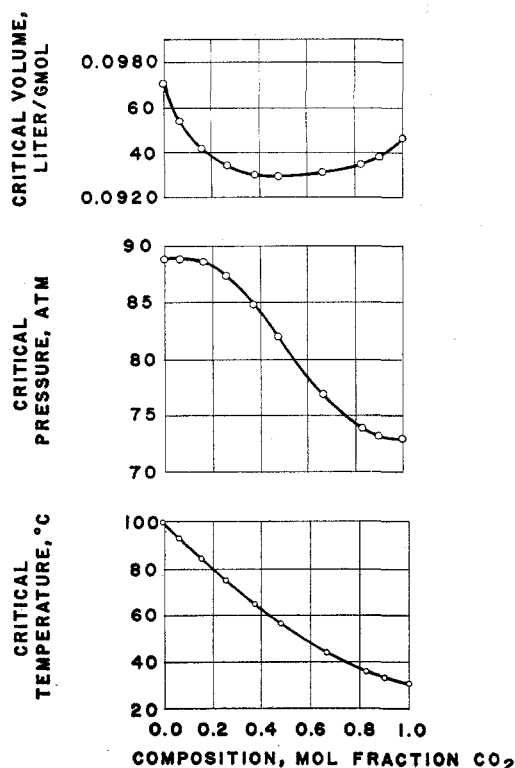


Figure 3. Critical States of System

hydrogen sulfide and carbon dioxide, respectively, in subsequent representations of the properties of the system.

Figures 1 and 2, plotted directly from the experimental data, show the pressure and volumetric characteristics of the saturated phases as a function of temperature. Large scale plots similar to these were used to evaluate the unique states of the systems—i.e., the critical, cricondenbar, and cricondentherm points. The pertinent pressures and temperatures were read from a large P - T graph; the critical envelope curve was used to locate the critical pressures and temperatures. The corresponding volumes were found from a V - T graph. The unique states of the system thus found are listed in Table II. The unique states for the mixture containing 0.4728 mole fraction of carbon dioxide were not directly observed, because of breakage of the experimental tube as the critical region was approached; subsequent observations of unique states were made on a mixture of approximately equal composition—0.4782 mole fraction—and the values for 0.4728 mole fraction were obtained graphically from plots against composition. Figure 3 shows such plots against composition for the critical properties of the system; the cricondenbar and cricondentherm properties show a similar general behavior.

Table III summarizes the saturation properties of the various mixtures in uniform intervals of pressure. The table was constructed from the experimental data by a combination of analytical and graphical procedures.

The relation between pressure and temperature was formulated by fitting a Clausius-Clapeyron equation ($\log P = a + b/T$) to the experimental dew and bubble points for each composition investigated; points in the region of retrograde condensation were excluded in making the fit. The equations thus obtained (two for each mixture) were used to calculate the saturation pressures for each experimental temperature. The calculated pressures were then compared with the measured values and graphical residual functions were constructed. These residuals could be plotted very precisely, as the maximum difference between the experimental pressure and that calculated from the appropriate equation never exceeded 2.5 atmospheres for any

mixture. The scatter of the residuals about a smooth curve did not exceed 0.01 atmosphere in any case; this serves as an indication of the consistency of the data. The equations and the graphical residues were used to compute the temperatures corresponding to the pressure arguments of Table III.

The molal volumes shown in Table III were also determined by means of graphical methods. Different procedures were used for the liquid and vapor phases. The range of variation of the molal volume for the saturated liquids was sufficiently small to permit the construction of accurate V - P graphs on Cartesian coordinates; the molal volumes were read directly from these graphs. In the case of the saturated vapors, the range of variation of volume was so great as to make it impractical to plot directly to a sufficiently large scale; hence a residual function was computed for graphing. The residual employed was

$$\alpha = RT/P - V \quad (1)$$

all the state variables being experimental dew point values. The parameter α for each mixture was plotted against temperature. The molal volume (equal to $RT/P - \alpha$) was then computed for each pressure-temperature point in Table III.

Figure 4 is a plot of saturation pressure and temperature against composition. The ice point data of Steckel (11), discussed later, are shown for comparison on this figure.

Table IV presents the liquid-vapor equilibria of the system in terms of the familiar y , x , and K parameters; Figures 5 and 6 show some of the data graphically. The table was constructed from values read directly from a large scale plot of temperature-composition isobars. No attempt was made to smooth the data thermodynamically or to represent the results generally by any of the integral forms of the Gibbs-Duhem equation. However, for the sake of completeness, some y - x values for temperatures below the ice point were included in the tabulation. These values, indicated by parentheses, were estimated by application of the van Laar solution of the Gibbs-Duhem relation to the liquid phase. In order to do this, the van Laar constants A and B were determined for three sets of isothermal data (0° , 10° , and 20°), read at 2-atmosphere intervals from pressure-composition graphs similar to Figure 4; no fugacity corrections were applied to the vapor. To evaluate the constants, the data were plotted in the form $(T \ln \gamma_{\text{CO}_2})^{-1/2}$ against $x/(1-x)$. Such a plot should be linear if the van Laar equation were to apply rigorously, as may be seen by inspection of the expression for the activity coefficient of carbon dioxide:

$$\ln \gamma_{\text{CO}_2} = \frac{B}{T} \left[\frac{Ax}{1-x} + 1 \right]^{-2} \quad (2)$$

According to this expression, the slope of such a plot is equal to $AB^{-1/2}$, while the intercept equals $B^{-1/2}$. The actual experimental data approximated straight lines when so plotted, but showed a slight upward concavity. However, the best straight lines were fitted by least squares, and the slopes and intercepts were used to determine A and B . Both van Laar constants were found to be nearly linear in temperature, and by graphical extrapolation to -20° it was possible to estimate values of the constants for use below the ice point. By use of these constants, the activity coefficient of hydrogen sulfide is also calculable:

$$\ln \gamma_{\text{H}_2\text{S}} = \frac{AB}{T} \left[\frac{1-x}{x} + A \right]^{-2} \quad (3)$$

From the two activity coefficients, x and y were computed for temperatures below the ice point in order to complete Table IV for the lower values of pressure.

The parenthetical (calculated) equilibrium values given in Table IV, although probably reasonably accurate, are of course less reliable than the remainder of the data—not only because of the subjective element in the extrapolation of the van Laar

TABLE IV. LIQUID-VAPOR EQUILIBRIA

P , Atm.	T , °C.	x	y	K_{CO_2}	K_{H_2S}	P , Atm.	T , °C.	x	y	K_{CO_2}	K_{H_2S}
20	24.50	0.000	0.000	...	1.00	60	76.20	0.000	0.000	...	1.00
	20	0.020	0.131	6.50	0.89		75	0.009	0.035	3.90	0.97
	15	0.052	0.240	4.62	0.80		70	0.050	0.122	2.44	0.92
	10	0.092	0.341	3.71	0.72		65	0.092	0.194	2.11	0.88
	5	0.142	0.441	3.11	0.65		60	0.137	0.271	1.98	0.84
	0	0.211	0.531	2.52	0.60		55	0.190	0.345	1.81	0.81
-5	(0.331) ^a	(0.665)	(1.66)	(0.50)	(0.56)		50	0.256	0.423	1.65	0.78
-10	(0.431)	(0.715)	(1.27)	(0.56)	(0.58)		45	0.330	0.496	1.50	0.75
-15	(0.618)	(0.785)	(1.22)	(0.64)	(0.68)		40	0.425	0.583	1.37	0.73
-16	(0.669)	(0.810)	(1.12)	(0.68)	(0.72)		35	0.540	0.662	1.23	0.74
-17	(0.748)	(0.840)	(1.06)	(0.68)	(0.72)		30	0.687	0.759	1.10	0.77
-18	(0.836)	(0.889)	1.00	1.00	1.00		25	0.873	0.901	1.03	0.78
-19.08	1.000	1.000	1.00	1.00	1.00		24	0.922	0.938	1.02	0.80
							22.52	1.000	1.000	1.00	1.00
30	41.55	0.000	0.000	...	1.00	70	85.13	0.000	0.000	...	1.00
	40	0.010	0.056	5.60	0.95		80	0.042	0.090	2.14	0.95
	35	0.042	0.169	4.03	0.87		75	0.086	0.160	1.86	0.92
	30	0.077	0.261	3.39	0.80		70	0.131	0.226	1.73	0.89
	25	0.122	0.353	2.90	0.74		65	0.184	0.300	1.62	0.86
	20	0.179	0.445	2.48	0.68		60	0.243	0.369	1.52	0.83
	15	0.251	0.531	2.12	0.63		55	0.312	0.442	1.42	0.81
	10	0.349	0.611	1.75	0.60		50	0.398	0.532	1.34	0.78
	5	0.484	0.698	1.44	0.59		45	0.489	0.601	1.23	0.78
	0	0.684	0.808	1.18	0.60		40	0.615	0.687	1.12	0.81
-1	(0.732)	(0.837)	(1.14)	(0.61)	(0.65)		35	0.765	0.796	1.04	0.87
-2	(0.793)	(0.866)	(1.09)	(0.65)	(0.68)		30	0.967	0.969	1.01	0.94
-3.01	1.000	1.000	1.00	1.00	1.00		29.26	1.000	1.000	1.00	1.00
40	54.99	0.000	0.000	...	1.00	80	93.29	0.000	0.000	...	1.00
	50	0.035	0.124	3.54	0.91		90	0.025	0.050	2.00	0.98
	45	0.074	0.216	2.92	0.85		85	0.070	0.127	1.81	0.94
	40	0.117	0.300	2.56	0.79		80	0.116	0.176	1.52	0.93
	35	0.168	0.382	2.28	0.74		75	0.167	0.240	1.44	0.91
	30	0.232	0.467	2.02	0.70		70	0.225	0.308	1.37	0.89
	25	0.313	0.546	1.74	0.66		65	0.292	0.373	1.23	0.88
	20	0.420	0.630	1.50	0.64		60	0.366	0.438	1.20	0.89
	15	0.563	0.720	1.28	0.64		55	0.456	0.508	1.11	0.91
	10	0.756	0.835	1.10	0.67		52.08	0.544 ^b	0.544 ^b	1.00	1.00
	9	0.797	0.862	1.08	0.68						
	8	0.852	0.899	1.06	0.68						
	7	0.916	0.942	1.03	0.69						
	5.80	1.000	1.000	1.00	1.00						
50	66.28	0.000	0.000	...	1.00						
	65	0.010	0.045	4.50	0.97						
	60	0.048	0.132	2.75	0.91						
	55	0.089	0.216	2.43	0.86						
	50	0.135	0.295	2.18	0.81						
	45	0.190	0.374	1.97	0.77						
	40	0.255	0.455	1.78	0.73						
	35	0.336	0.534	1.59	0.70						
	30	0.438	0.614	1.40	0.69						
	25	0.572	0.701	1.23	0.70						
	20	0.739	0.811	1.10	0.72						
	17	0.874	0.908	1.04	0.73						
	16	0.932	0.950	1.02	0.74						
	14.85	1.000	1.000	1.00	1.00						

^a Values in parentheses are extrapolated by van Laar equation.^b Critical point.

TABLE V. TEST OF VAN LAAR EQUATION AT ICE POINT

P , Atm.	x , Exptl.	y , Exptl.	y , Calcd.
10.33	0.000	0.000	0.000
15	0.075	0.338	0.345
20	0.211	0.531	0.540
25	0.407	0.669	0.664
30	0.684	0.808	0.810
34.38	1.000	1.000	1.000

constants but also because the van Laar equation is not entirely adequate in representing the experimental data. Some idea of the degree of confidence which should attach to the calculated (parenthetical) values in Table IV may be gained from Table V, which compares experimental data at the ice point with those predicted by van Laar's equation for the same temperature.

DISCUSSION

In general, only the last significant figure of any value appearing in Tables II and III is subject to uncertainty. The probable limits of error have been indicated previously.

The most important source of error in fixing the composition of coexisting phases (Table IV) is that connected with the precise construction and reading of the P - x and T - x graphs from which the equilibrium data were computed. It is difficult to assign any quantitative index to the errors introduced in drawing and reading the graphs. With the relatively large number of compositions investi-

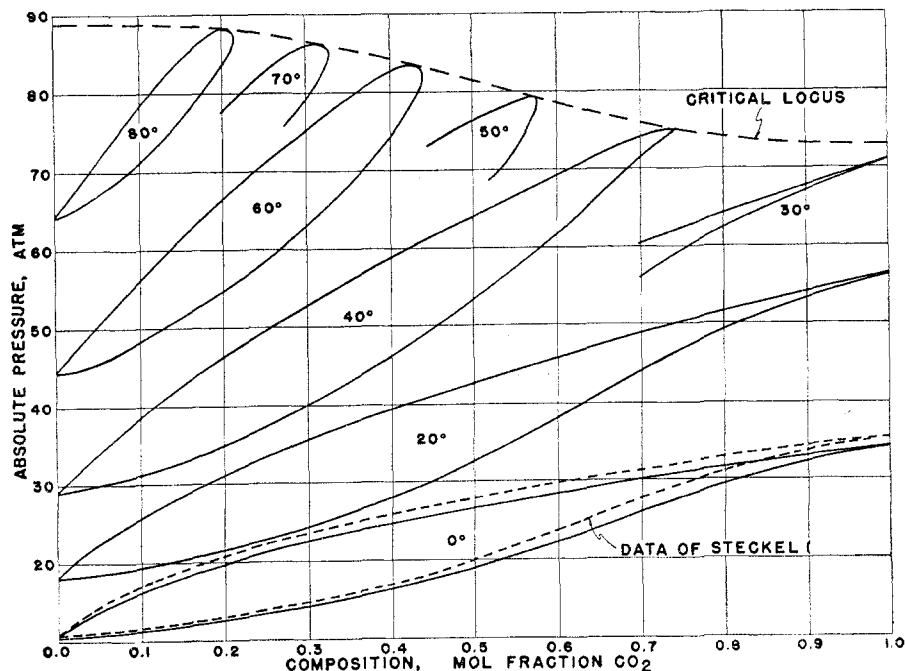


Figure 4. Pressure-Composition Diagram

March 1953

INDUSTRIAL AND ENGINEERING CHEMISTRY

623

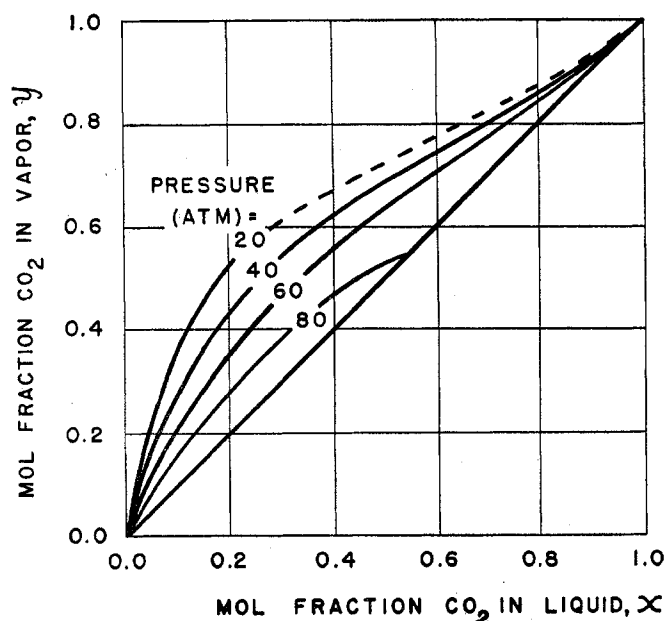


Figure 5. Compositions of Coexisting Phases

gated, there was a minimum of subjectivity in placing the curves, and there was no observable scatter in the P - x plots; however, the flat slope of the curves at the carbon dioxide-rich end was undoubtedly conducive to reading errors of 0.001 or 0.002 mole fraction in both the dew and bubble point curves.

In reporting low-pressure liquid-vapor equilibrium determinations, it is customary to smooth the data by calculating one liquid-phase activity-coefficient curve from the other and thereafter making any adjustments necessary to achieve a pair of thermodynamically consistent curves which best represents the measurements. This is a relatively simple matter if (as is true at low pressures) the vapor phase is ideal, so that partial pressures and fugacities can be considered as identical. In the case of high-pressure equilibria, however, attempts to smooth data thermodynamically usually create more uncertainty than they eliminate. This is due to the fact that neither phase is ideal and the Gibbs-Duhem relation must be simultaneously applied to both fluids, instead of to the liquid alone. A necessary step in such a procedure is the calculation of the activity coefficients of the components in the vapor phase (3):

$$\gamma_i = f/p_i = \exp \left[\frac{1}{RT} \int_{p_0}^p \left(\frac{\partial v}{\partial n_i} \right) - \frac{RT}{p} \right] dp \quad (4)$$

The evaluation of the partial molal volume which appears in the integrand requires very accurate volumetric data on the superheated vapors near their saturation points and involves a tedious and exacting graphical calculation. The present experiments do not include observations appreciably removed from the saturated states, and such a rigorous treatment is impossible. Empirical rules for evaluating component fugacities in a mixture, although convenient for engineering estimation when no data are available at all, are of no utility in evaluating experimental data, as one has no way of knowing how far in error these rules may be.

In the absence of reliable methods of test, the liquid-vapor equilibrium data were recorded directly in Table IV without smoothing; although the maximum error in these values is unknown, they are certainly superior to any data which might be estimated for the system on the basis of the properties of the pure components.

The comparison between the present measurements and those of Steckel (11) is interesting; Figure 4 shows the unmistakable

qualitative similarity between the experimental saturation pressures at the ice point. The present data are almost certainly more reliable than Steckel's. He performed his determinations in a metal bomb, which contained a vigorously stirred mixture of the components and was immersed in a constant-temperature bath. Data were taken at 0°, -26.8°, and -52.0° C.; the temperatures were measured with a liquid-in-glass thermometer graduated to 0.1°. The total pressure on the system was measured with Bourdon-tube gages, calibrated periodically against a manometer. When equilibrium had been established, both phases were sampled through valves and analyzed. The withdrawal of the samples caused pressure decreases as large as 2% at the lower pressures; Steckel states that corrections were applied for the pressure decrease but does not describe the correction procedure. The end data were not tabulated, but the experimental points were shown on his graphs; the scatter about the mean curve reproduced in Figure 4 was as great as 0.5 atmosphere for some points. Steckel's value for the vapor pressure of pure carbon dioxide deviates almost 1 atmosphere from that generally accepted.

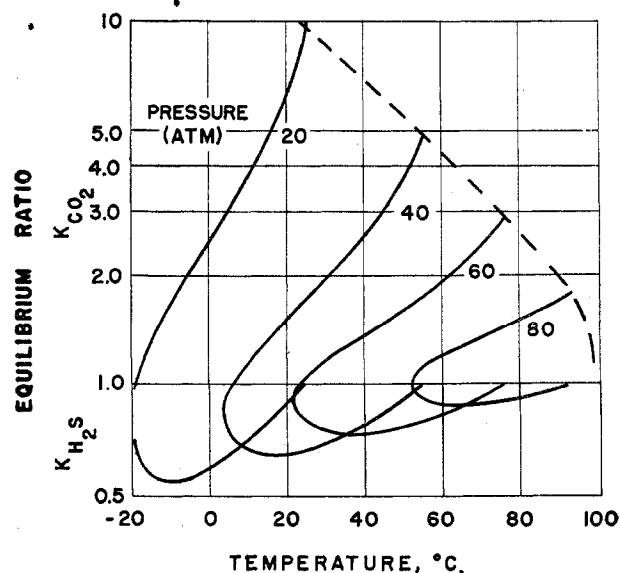


Figure 6. Vapor-Liquid Equilibrium Ratios

The present measurements, together with those of Steckel, reveal that no azeotropism exists in this system. However, there is some evidence that intermolecular forces of the kind causing azeotrope formation are strongly developed. The inflection in the P - x curves at the carbon dioxide-rich end, together with the very flat terminal slope, suggests a strong tendency toward formation of a minimum-boiling mixture; but the incipient azeotrope never quite forms, at least within the temperature range above -52° (the lowest temperature used by Steckel).

None of the critical properties obeys an additive law with respect to composition; this is not surprising in view of the strong intermolecular forces which are in evidence even at the lowest pressures investigated. The behavior of the critical pressure with changing composition is unusual in that it shows abrupt curvatures near its end points but displays neither the maximum nor the minimum which is generally characteristic of highly nonideal systems.

The separation of pure hydrogen sulfide from mixtures rich in this component should be relatively easy, as carbon dioxide is strongly squeezed out. Above about 0.8 mole fraction of carbon dioxide, separation becomes difficult, so that a large number of transfer units would be needed if it were desired to extract carbon dioxide in a high state of purity. However,

because hydrogen sulfide is the component of the greater commercial interest, this is actually of little practical consequence.

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Vapor-Liquid Equilibria for Ethyl Alcohol Binary Systems

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VAPOR-liquid equilibria at 760 mm. of mercury are reported for the systems: ethyl alcohol-*n*-butanol, ethyl alcohol-*sec*-butanol, ethyl alcohol-*n*-pentanol, ethyl alcohol-acetone, ethyl alcohol-methyl ethyl ketone, and ethyl alcohol-methyl *n*-propyl ketone.

The ethyl alcohol-alcohol systems exhibited no azeotropes, and the activity coefficients calculated from the experimental data show these systems deviate only slightly from ideal solution behavior. The ethyl alcohol-*n*-butanol data compare well with those reported by Brunjes and Bogart (1).

The ethyl alcohol-ketone systems showed the following azeotropes: ethyl alcohol-acetone: none; ethyl alcohol-methyl ethyl ketone: 50.1 mole % ethyl alcohol at 165.2° F. (74.0° C.); ethyl alcohol-methyl *n*-propyl ketone: 96.2 mole % ethyl alcohol at 172.5° F. (78.0° C.). The ethyl alcohol-acetone data check those given by Perry (6).

ETHYL ALCOHOL-ALCOHOL SYSTEMS AT 760 MM. OF MERCURY

MATERIALS. The source and physical constants of the purified materials are presented in Table I. Absolute ethyl alcohol as received was used without purification. The *n*-butanol and *sec*-butanol samples were purified as outlined by Hill (2), by taking the middle cut of 500 ml. from a charge of 1500 ml. The *n*-pentanol was the middle 300-ml. cut from an 800-ml. charge. All the materials employed had a boiling range of less than 0.1° C. The alcohols contacted only a dry atmosphere during storage, since suitable desiccating agents were provided for both air and the liquid samples.

During the experimental work, frequent checks were made on the refractive indices of the compounds to detect possible contamination.

APPARATUS. The method of analysis was the same as that used by Hill (2). The equilibrium still thermocouple was an iron and constantan couple calibrated by a Cottrell boiling point apparatus operated at 760 mm. of mercury. Boiling temperatures were cross-checked with a set of National Bureau of Standards thermometers. The still system was provided with connections to a course of filtered air to maintain the total pressure on the system at 760 mm. of mercury. The vapor-liquid equilibrium still arrangement was the same as that employed by Hill (2) with the exception that *n*-tetradecane was used as a manometer fluid.

PROCEDURE. Samples were prepared by the addition of ethyl alcohol to a known volume of the second alcohol in the binary system so that a total volume of 25 ml. was obtained. The operating and analytical procedures were those reported by Hill (2). It was necessary to employ a special stopcock grease composed of silicone grease and carnauba wax to prevent leakage when the ethyl alcohol-*sec*-butanol system was investigated.

DATA. Of the systems investigated, only the ethyl alcohol-*n*-butanol system (1) has been previously reported in the literature (4). The experimental data were obtained at a constant pressure of 760 mm. of mercury for the binary systems existing in two-phase vapor-liquid equilibria. Figures 1 and 2 and Tables II, III, and IV show the results.

The compositions of the liquid and vapor phases were obtained by refractive index measurements, and equilibrium temperatures were determined from the thermocouple measurements. The maximum experimental error is believed to be 0.6 % mole fraction based on the 50% (composition point)—i.e., an absolute error of ± 0.003 mole fraction—and $\pm 0.1^\circ$ F.

Activity coefficients from the experimental data were calculated by Equation 1 and are shown in Tables II, III, and IV.

$$\gamma = \frac{y\pi}{xP} \quad (1)$$

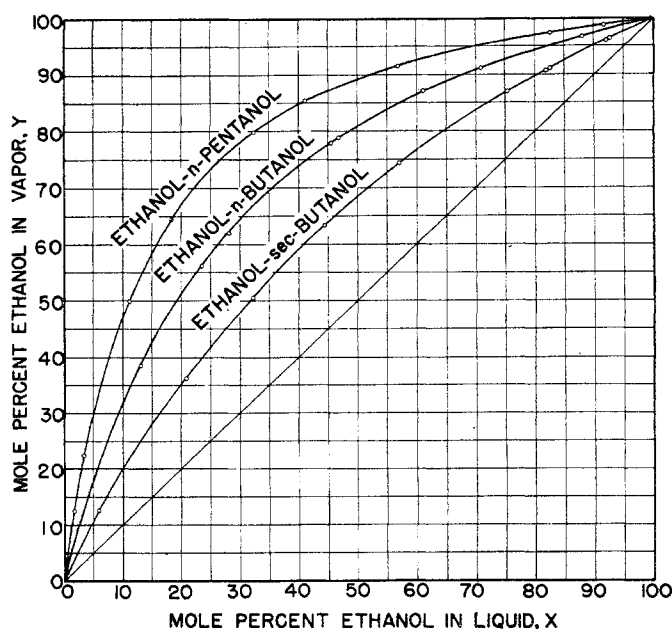


Figure 1. Vapor-Liquid Equilibrium Diagram for Ethyl Alcohol Systems at 760 Mm.

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