THE UNIVERSITY OF OKLAHOMA
GRADUATE SCHOOL

MULTI-STAGE SEPARATION OF CRUDE GAS-OIL MIXTURES

A THESIS

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ACKNOWLEDGMENT

The idea of analyzing the effects of stage separation on a scale applicable to industry was given by R. L. Huntington, who acted as supervising instructor on the project. The author wishes to thank him for his many suggestions which aided in design, method of attack, and interpretation of results.

L. S. Reid, graduate student in the University of Oklahoma assisted materially in solving mechanical problems and aided in the actual construction of the apparatus.

Henry Black, undergraduate student in the School of Chemical Engineering, devoted his time as a skillful operator and helped obtain continuous data by working several sixteen hour periods. Marvin Owens, senior engineer, also assisted in the construction and in taking some of the preliminary data.

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INTRODUCTION

Qualitative proofs of the advantages of stage separation are all that have been observed in the field, because of the large number of variables involved in its study. In the Petroleum Engineering Laboratory of the University of Oklahoma, quantitative observations have been made on a two stage separator by R. L. Huntington,[2][3][5] W. F. Cloud and undergraduate research students. The effect of stage pressure ratio on the gas-oil ratio was observed for a charging stock of natural gas saturated crude oil, and theoretical calculations, made on several-component mixtures, verified the results.

Advantages of stage separation to the various branches of the petroleum industry have been pointed out by R. L. Huntington.[1] Until the time of this present investigation, no quantitative study had been made using a charging stock similar to crude from deep high pressure fields, nor at high enough pressures to compare with field conditions.

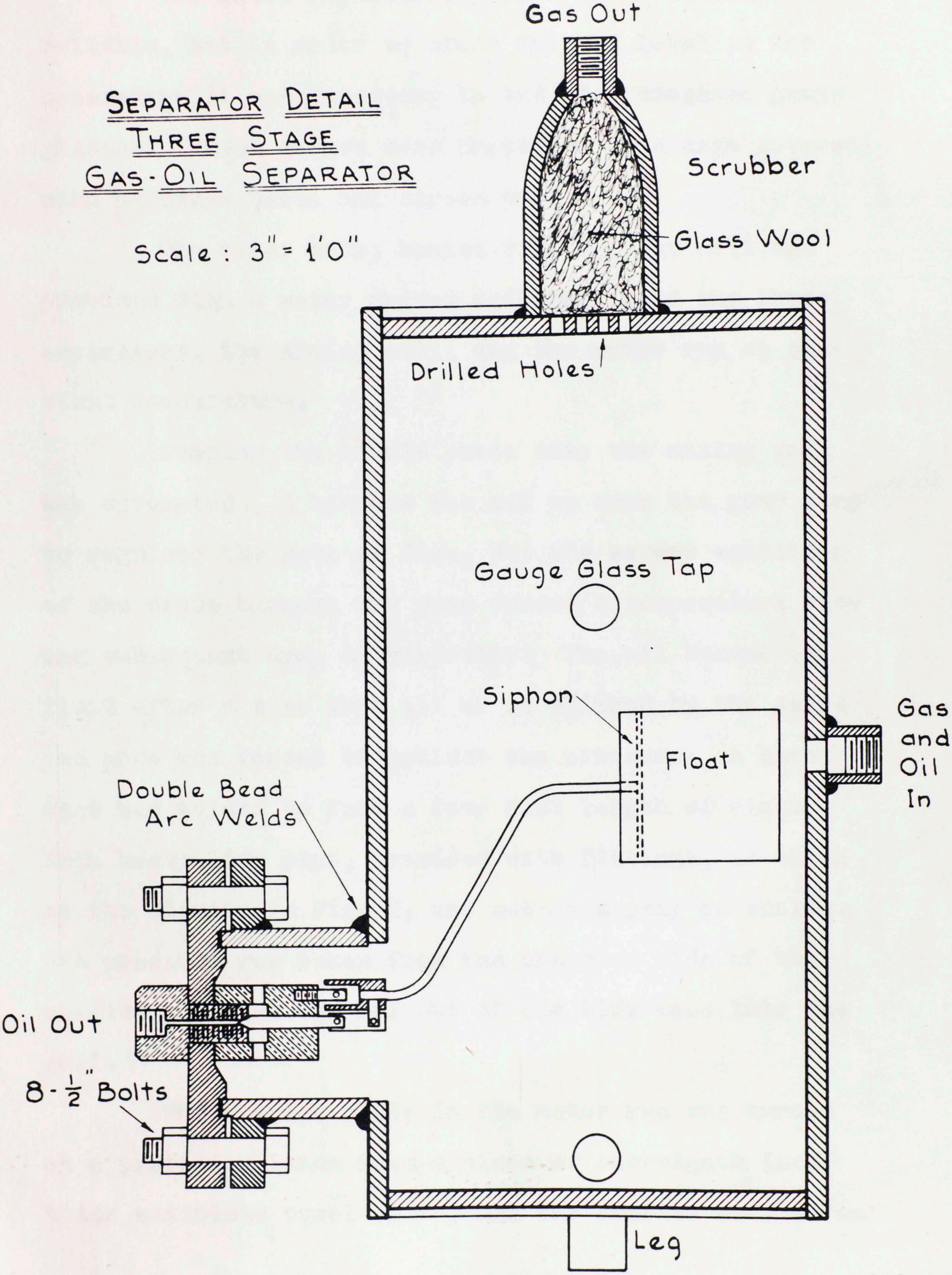
With the promotion of unit operation as a conservation measure, comes the problem of economy of such a program. Here multi-stage separation is bound to play a part. So far the design of separator units has been a trial and error proposition. It is the purpose of this

CHAPTER I

APPARATUS

In any stage separation unit it is necessary to control the flow of both liquid and gaseous phases. The degree of control must be high in the case of an experimental unit because many observed quantities are dependent on steady equilibrium conditions. Control of both phases, manually, would necessitate an operator on every control valve, so the liquid phase was made automatic by incorporating fluid level regulators in the separators. A detail of the float mechanism and separator is shown in Fig. 1.

The separators were made from heavy seamless line pipe, double bead arc welded, and tested to 1200 pounds, to insure an adequate factor of safety. In order to eliminate the danger of blow out, the whole system was designed to stand the maximum pressure. All lines used in the apparatus subjected to pressure, or in such a position to be subjected to pressure by a mistake of the operators, were extra heavy. High pressure stainless steel needle valves were used at all points under strain and where a high degree of control was necessary. Control arms had to be put on the needle valves which controlled the flow of gas from the separators.



Gas Out
Separator Detail
Three Stage
Gas-Oil Separator
Scale : 3" = 1'0"
Scrubber
Glass Wool
Drilled Holes
Gauge Glass Tap
Siphon
Float
Gas and Oil In
Double Bead Arc Welds
Oil Out
8-½" Bolts
Leg

Fig. 1

The fluid regulators proved to be accurate and reliable, but in order to check the oil level in the separators it was necessary to add high pressure gauge glasses. These gauges were protected by a cage covered with hardware cloth and screen wire.

The water bath, heated with a steam coil and provided with a motor driven agitator, held the three separators, the mixing coil, and the meter run at constant temperature.

Pumping the liquid phase into the mixing coil was attempted. A by-pass was set up over the gear pump to regulate the rate of flow, but the severe agitation of the crude through the pump caused a temperature rise and subsequent drop in viscosity. The oil became so fluid after a time that all of it slipped by the gears and none was forced in against the pressure. A blow case was welded up from a four foot length of eight inch heavy line pipe, provided with fittings, as shown on the diagram in Fig. 2, and set on a pair of scales. Gas pressure was taken from the upstream side of the orifice to force the oil out of the blow case into the coil.

The orifice plate in the meter run was turned on a precision lathe from a piece of one-eighth inch thick stainless steel plate, and the orifice drilled to

Fig. 2

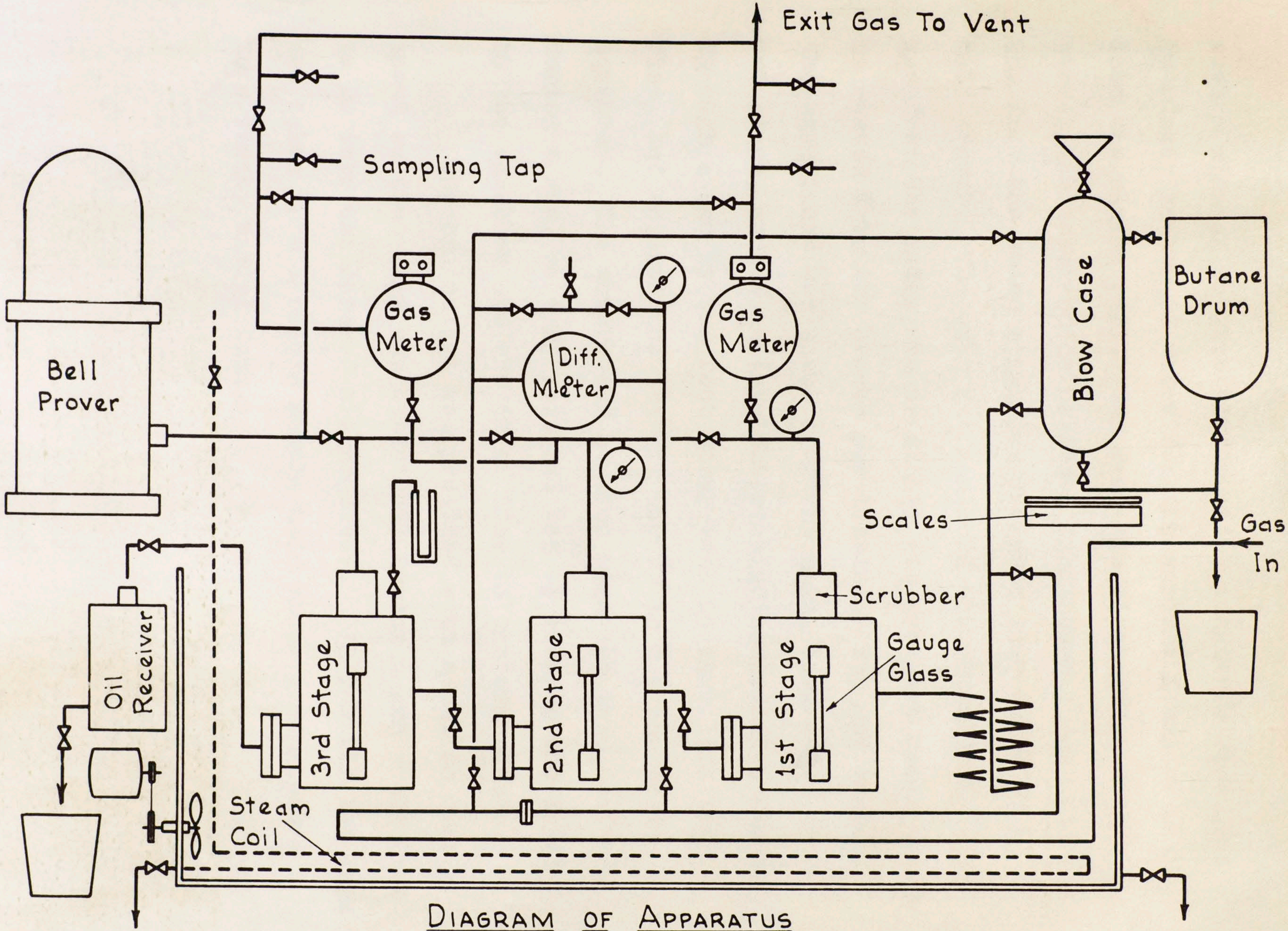


DIAGRAM OF APPARATUS
THREE STAGE GAS-OIL SEPARATOR

such a size as to give differential pressures from ten to ninety inches of water for the various flows of gas into the mixing coil. Gas from the first and second stages was measured by positive displacement gas meters. Gas from the third stage was measured over water in a bell prover. The exit gas manifold was designed so that gas from the meter run or either displacement meter could be routed into the prover for testing. The gas used in the experiment was taken from the main, compressed in a two stage, double acting, intercooled compressor, flowed through a two by ten foot heavy tank to stabilize the flow, and relieved of its oil vapors in a glass wool scrubber. The compressor handled more gas than was used during a run, so a by pass was set up between the inlet to the scrubber and the inlet line to the compressor.

The oil outlet was provided with a switch over spout so that only the outlet oil during a run would have to go into the covered oil receiver. Oil for all the runs was taken from a single sample held under gas pressure in a twenty barrel tank.

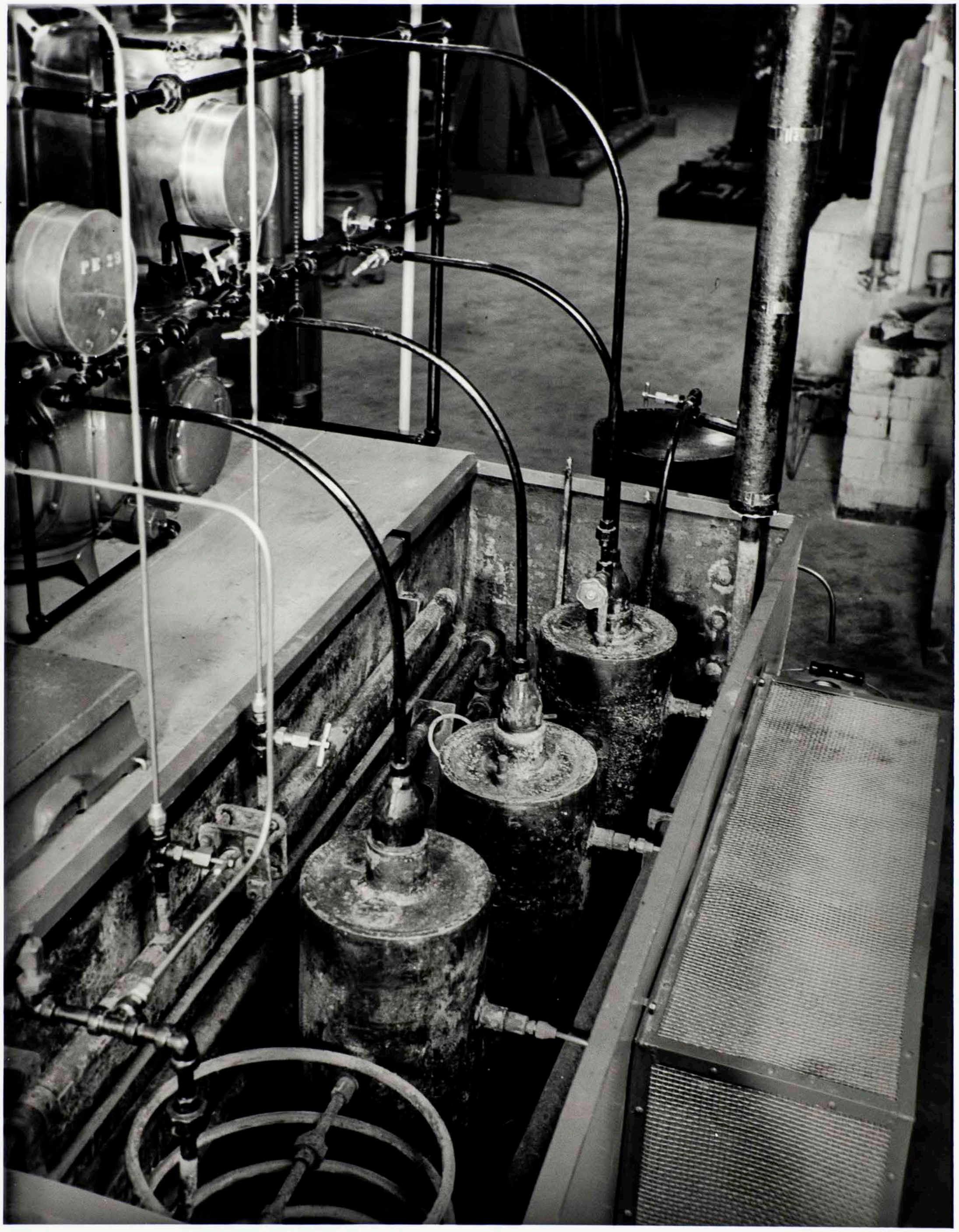


Fig. 3

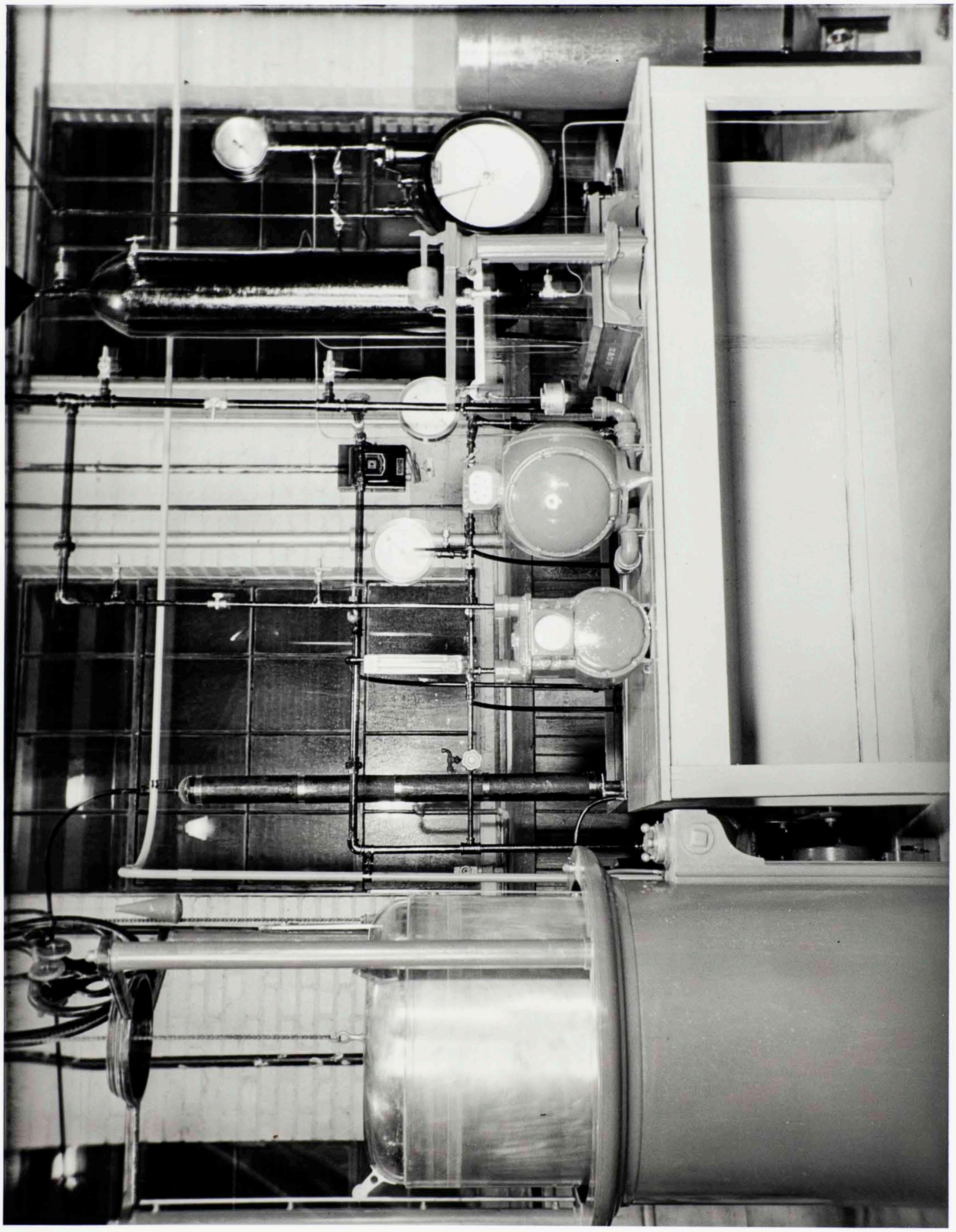


Fig. 4

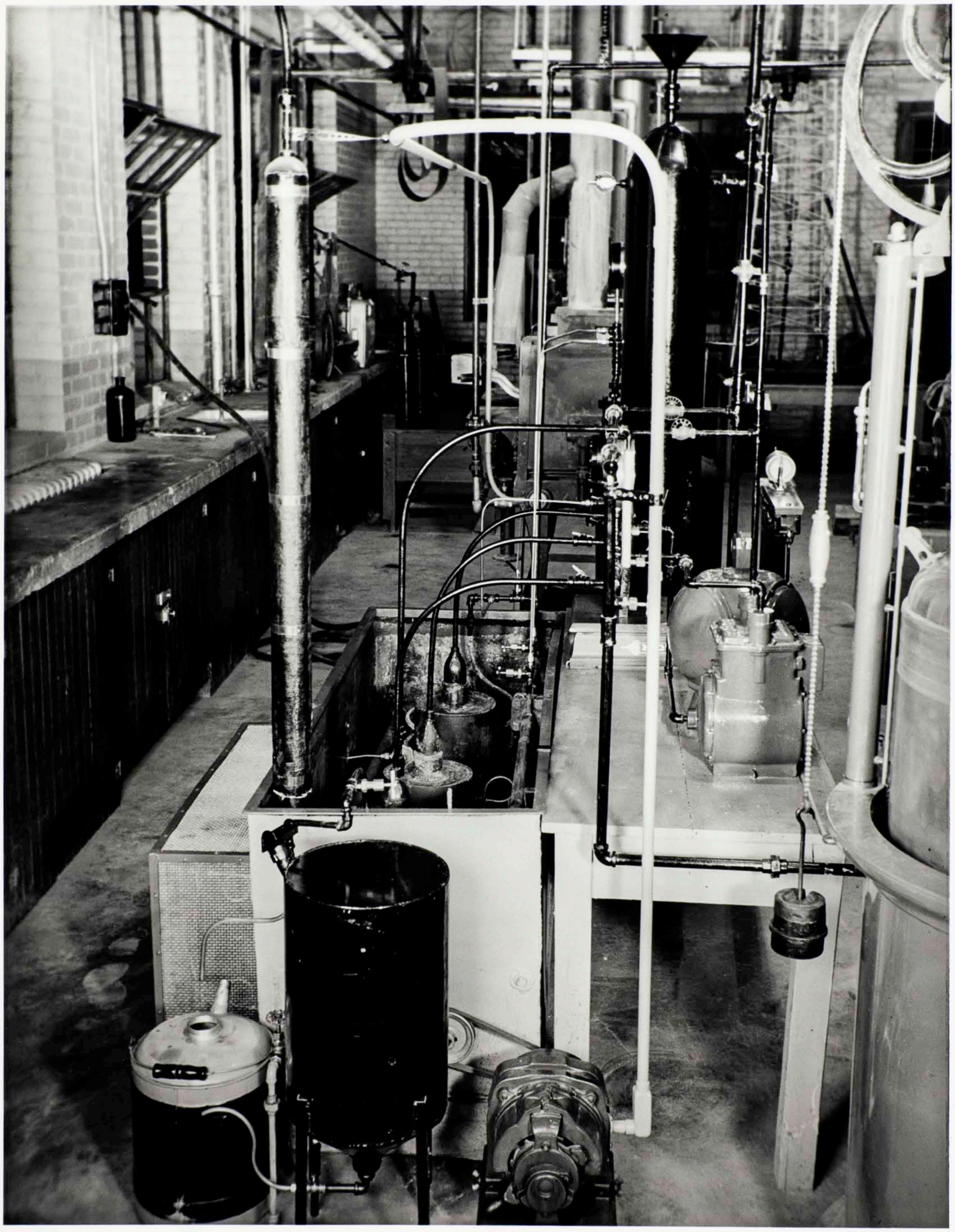


Fig. 5

CHAPTER II

PROCEDURE

The apparatus was tested carefully for leaks by connecting the meter run, blow case, and separators and applying a pressure of 550 pounds. The meters were tested for leaks by connecting them to the third stage through the low pressure manifold and bleeding in a pressure of two inches of mercury. After repairing a few small leaks, the displacement meters were checked against the prover. A coefficient was established for the orifice by flowing into the prover and taking the time for a given volume of gas to pass at various temperatures and rates of flow. The gas from the main, being very constant in composition, eliminated a variable from the orifice formula. To further simplify calculations, the down stream static pressure on the recording differential meter was held at 550 pounds per square inch absolute during all the runs. The orifice formula was thus reduced in form to

$$Q = C\sqrt{\frac{h}{T}}$$

where Q = S.C.F. per hour

C = the orifice coefficient

h = differential pressure in inches of water

T = $^{\circ}$Rankine

With all necessary correction curves plotted for the measuring equipment, a set of trial runs was made to determine the sensitivity and method of attack. It was found that in passing from one set of conditions to another, considerable time was required to put the stages back into a state of equilibrium. It was further found that a longer period of stabilization was required each time the unit was put into operation after a rest of twenty hours or so. The amount of oil that could be stored in the blow case limited the time of stabilization for only one run to a period of eighty minutes, the rate of oil charge being half a pound per minute. At first, runs of one hour were attempted preceded by a stabilization period of twenty minutes. Five of these runs, all following each other over a traverse, still gave unsatisfactory results.

Runs of the traverses appearing in the following chapters were made at one time, the period of stabilization being lengthened to thirty minutes, after which the data were taken around the unit for fifteen minutes. This permitted two runs per charge of the blow case.

A part of the preliminary runs was made with a liquid charge of crude oil only. The drop in gravity was not as high as was expected and the fraction stripped was low. The final traverses were made with a liquid charging stock of a mixture of crude oil and pure butane. One part of butane by weight was mixed with ten parts of crude oil in the blow case. In order to prevent the oil charge from flowing over into the meter run, it was necessary to use only forty-five pounds of oil to four and one-half pounds of butane.

The following brief procedure, accepted after studying twenty-seven trial runs, was perfectly satisfactory. Time prevented any further investigation of various effects of rate of charge, length of stabilization, and duration of runs.

Assuming the apparatus to be in a shut down condition as explained at the end of this procedure, the first step was to charge the blow case. The valve under the funnel was opened and the oil poured in. The bleed valve, and valve under the funnel were closed and butane added by first closing the drain line from the drum, and then opening the two valves in the line from the drum to the bottom of the blow case. After the charge these two valves were closed and the drain line opened. It was necessary to invert the butane drum because in an upright position all the butane that went into the blow case had to boil off, and even though the pressure was raised only slightly in the case by charging gas instead of liquid, the reduction in temperature in the drum lowered the vapor pressure to such a low level that the flow of butane ceased.

The compressor was designed to take gas at atmospheric pressure, so upon starting, it was necessary to throttle the intake valve carefully to prevent the inter-stage from reaching too high a pressure. The gas stabilizing tank was not blown down but was shut off from the compressor and apparatus after each set of runs. After opening the gas circuit from the main to the apparatus, the compressor was lubricated and started, and the pressure allowed to build up to 550 pounds absolute. The compressor by-pass was then regulated to hold this pressure constant. The valve on the blow case, closing the line to the upstream side of the meter, was opened and pressure allowed to bleed in on top of the oil charge. After starting the agitator and heating the bath to the required temperature, valves between stages and on the outlet oil line were opened, and the two in the by-pass on top of the differential meter, closed. Before opening the valve to the manometer on the third stage, any excess gas which might have come out of solution was bled off through the prover. The next step, stabilizing the flow of both phases, was difficult and required the constant watch of two operators. The oil and gas were turned into the mixing coil. The predetermined differential pressure was set on the meter, the oil flow stabilized by taking the time for a small amount to leave the blow case, and the pressure in the three stages set on their respective absolute values. After a stabilization period which was determined by the amount of oil that had left the blow case, the run was started when the beam of the scales reached its mid position on a certain weight. At the instant the run was started, both meters were read, the prover was set on zero, and the exit oil was turned into the closed receiver.

During the run, the gravity of the gas from the first and second stages was taken with an Acme balance. At the end of the run, the outlet valves from the three stages, the oil and gas inlet valves, the oil outlet valve from the last stage, the valves between the stages, the compressor by-pass valve, and the valve on the manometer from the last stage, were closed in order as rapidly as possible. After shutting down the compressor, the meters and prover were read, and the gravity of gas in the prover taken with the balance. The A.P.I. gravity of the oil caught in the receiver was determined with a standardized hydrometer. After closing the valve in the line to the blow case from the upstream side of the differential meter, and opening the meter by-pass, the remaining oil in the case was removed from the bottom, the valve closed, and the gas bled off from the bleed valve at the top. On runs terminating a traverse, the stabilizing gas tank was shut off from the compressor and apparatus.

The floats on the fluid level regulators were made of light sheet brass to eliminate the use of a heavy spring balance. This necessitated a hole in the float to equalize the pressure. In order to make the floats self purging, a small siphon line was run to the bottom of the float. Once when the pressure was relieved suddenly on the separator, a rupture of the float occurred.

CHAPTER III

EXPERIMENTAL STAGE PRESSURE RATIO TRAVERSE

An important variable, over which the producer has control on a multi-stage separator in the field, is the stage pressure ratio. This traverse was made to show the advantage of stage separation over the single separator, and to determine a stage pressure ratio which would yield products from the stages most beneficial to the operator.

The last five runs of the traverse were made at one time. After all runs, involving the use of three stages, were completed, the first stage was disconnected from the last two, and a run made on the single stage under identical conditions to the other five.

Tables I and V show the corrected observed data and calculated quantities for all six runs of the traverse. At a stage pressure ratio of between four and six, five of the curves show a maximum or a minimum point. In this neighborhood of stage pressure ratios, the A.P.I. of the exit oil is highest, and the fraction stripped from the crude oil is lowest.

In the case of the stage pressure ratio traverse, quantities to which the products from the stages are referred, are constant. In Fig. 7 the volume of gas from each of the separators is compared to the volume of gas flowing into the mixing coil.

The irregularity of the points shown on the graph in Fig. 8 are due to the fact that they are obtained by taking small differences between large numbers. A mean curve drawn through the points agrees remarkably well with theoretical calculations shown in Chapter V.

Advantages of stage separation over the single stage are brought out clearly in the graphs. As the stage pressure ratio increases, the advantages become more and more evident up to a ratio of about five. If the pressure in the last stage of a multi-stage separator were predetermined, and the well head pressure were known, the number of stages required to effect the most efficient separation could be easily calculated. R. L. Huntington, (2)(3) in his work on two stage separation at lower pressures, showed that a stage pressure ratio of about five yielded the minimum gas-oil ratio.

Operating a multi-stage separator in the field at a stage pressure ratio of five would give the highest gravity of crude oil and a lean gas from the first stage. This gas could be used to repressure without extracting the natural gasoline, nor having to recompress from atmospheric pressure. At this ratio the third stage gas gravity is a maximum, as shown in Fig. 6, proving that the multi-stage separator also acts as a means of separating heavy and light gas fractions from the crude oil. The gas from the third stage could be processed and its natural gasoline extracted.

TABLE NO. I

STAGE PRESSURE RATIO TRAVERSE

CONSTANTS

Gas-Oil Ratio	1000
Bath Temp. °F.	100
S.C.F. Gas In	28.90
S.G. Gas In	0.610
Lbs. Gas In	1.314
°A.P.I. of Oil In	46.0
Lbs. Oil In	8.05
Length of Runs min.	15

Run No.	28	29	30	31	32
Stage Pres. Ratio	2	3	4	5	6
EXIT FIRST STAGE					
Pres. #/in.^{2} abs.	60	135	240	375	540
S.C.F.	31.80	29.75	28.40	26.40	25.20
Vol. % Gas In	110	103	98.3	91.3	87.2
S.G. Gas	0.817	0.745	0.703	0.684	0.672
Lbs. Gas	1.939	1.651	1.489	1.346	1.253
EXIT SECOND STAGE					
Pres. #/in.^{2} abs.	30	45	60	75	90
S.C.F.	0.263	0.651	1.46	2.57	3.72
Vol. % Gas In	0.91	2.25	5.05	8.89	12.9
S.G. Gas	1.02	0.983	0.953	0.891	0.870
Lbs. Gas	0.020	0.048	0.104	0.171	0.241
EXIT THIRD STAGE					
Pres. #/in.^{2} abs.	15	15	15	15	15
S.C.F.	o.172	0.473	1.22	1.76	2.04
Vol. % Gas In	0.59	1.64	4.22	6.09	7.06
S.G. Gas	1.34	1.42	1.60	1.65	1.65
Lbs. Gas	0.017	0.050	0.146	0.216	0.251
°A.P.I. Oil	42.2	42.5	43.2	43.7	43.7
TOTAL EXIT GAS					
S.C.F.	32.24	30.87	31.08	30.73	30.96
Lbs.	1.976	1.749	1.739	1.733	1.745
GAS FRACTION STRIPPED					
S.C.F.	3.37	2.00	2.21	1.86	2.09
Vol. % Gas In	11.7	6.92	7.65	6.44	7.24
Lbs.	0.662	0.435	0.425	0.419	0.431
Wt. % Oil In	8.3	5.5	5.3	5.2	5.4
S.G.	2.64	2.92	2.58	3.02	2.76

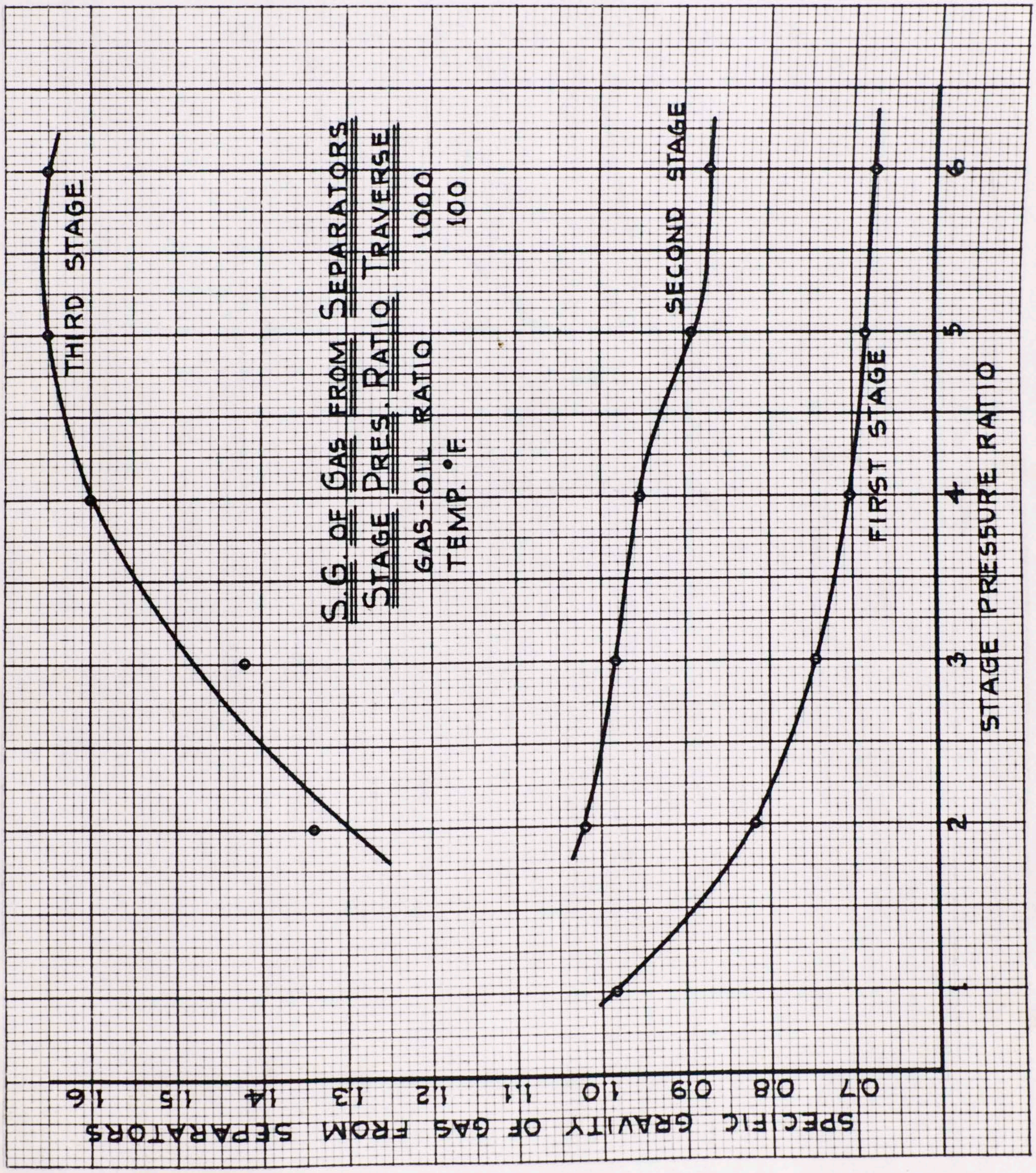

S.G. OF GAS FROM SEPARATORS
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP. °F 100
THIRD STAGE
SECOND STAGE
FIRST STAGE
STAGE PRESSURE RATIO
1
2
3
4
5
6
SPECIFIC GRAVITY OF GAS FROM SEPARATORS
0.7
0.8
0.9
1.0
1.1
1.2
1.3
1.4
1.5
1.6

Fig. 6

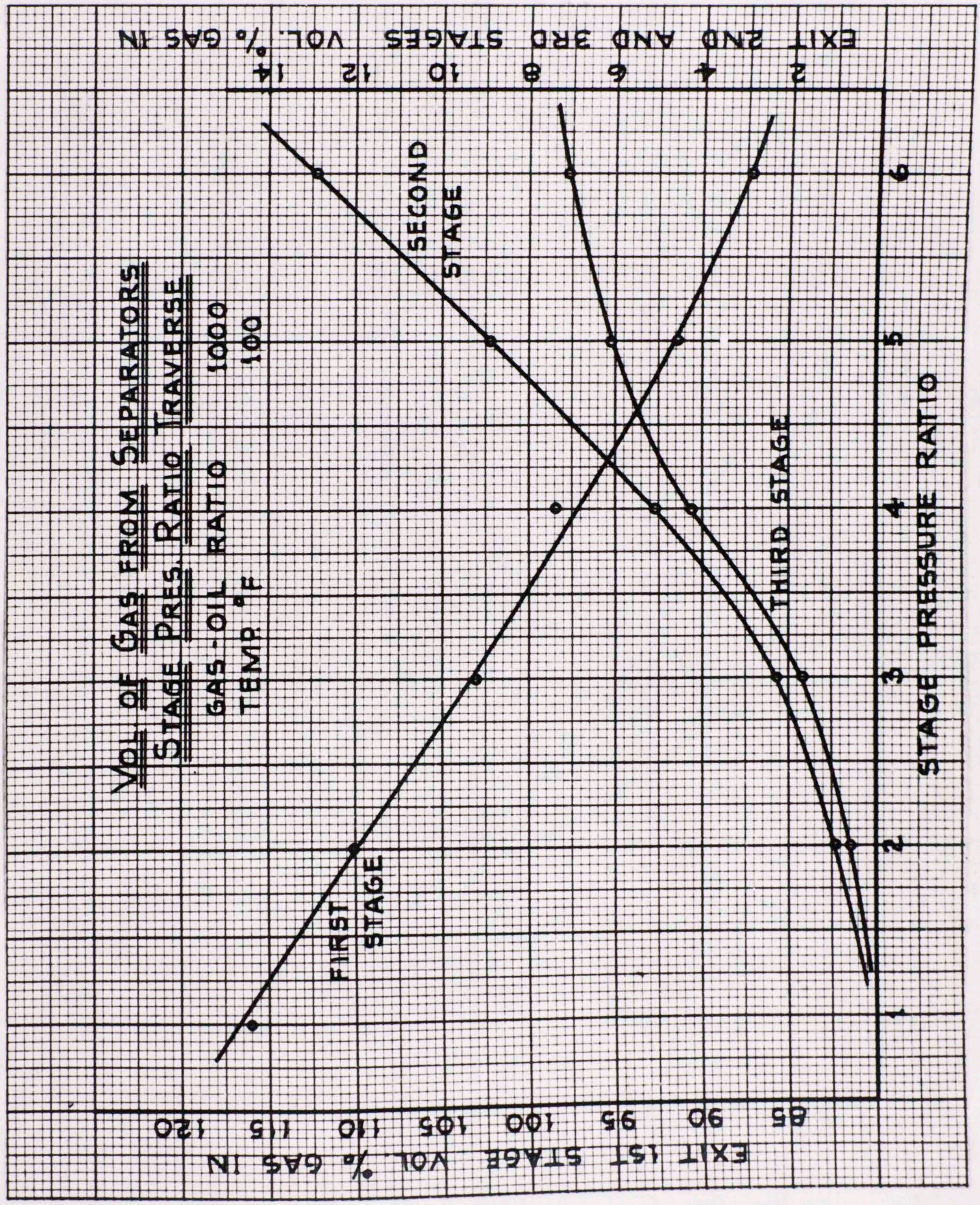

VOL OF GAS FROM SEPARATORS
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP °F 100
FIRST STAGE
SECOND STAGE
THIRD STAGE
STAGE PRESSURE RATIO
1
2
3
4
5
6
EXIT 1ST STAGE VOL. % GAS IN
85
90
95
100
105
110
115
120
EXIT 2ND AND 3RD STAGES VOL. % GAS IN
2
4
6
8
10
12
14

Fig. 7

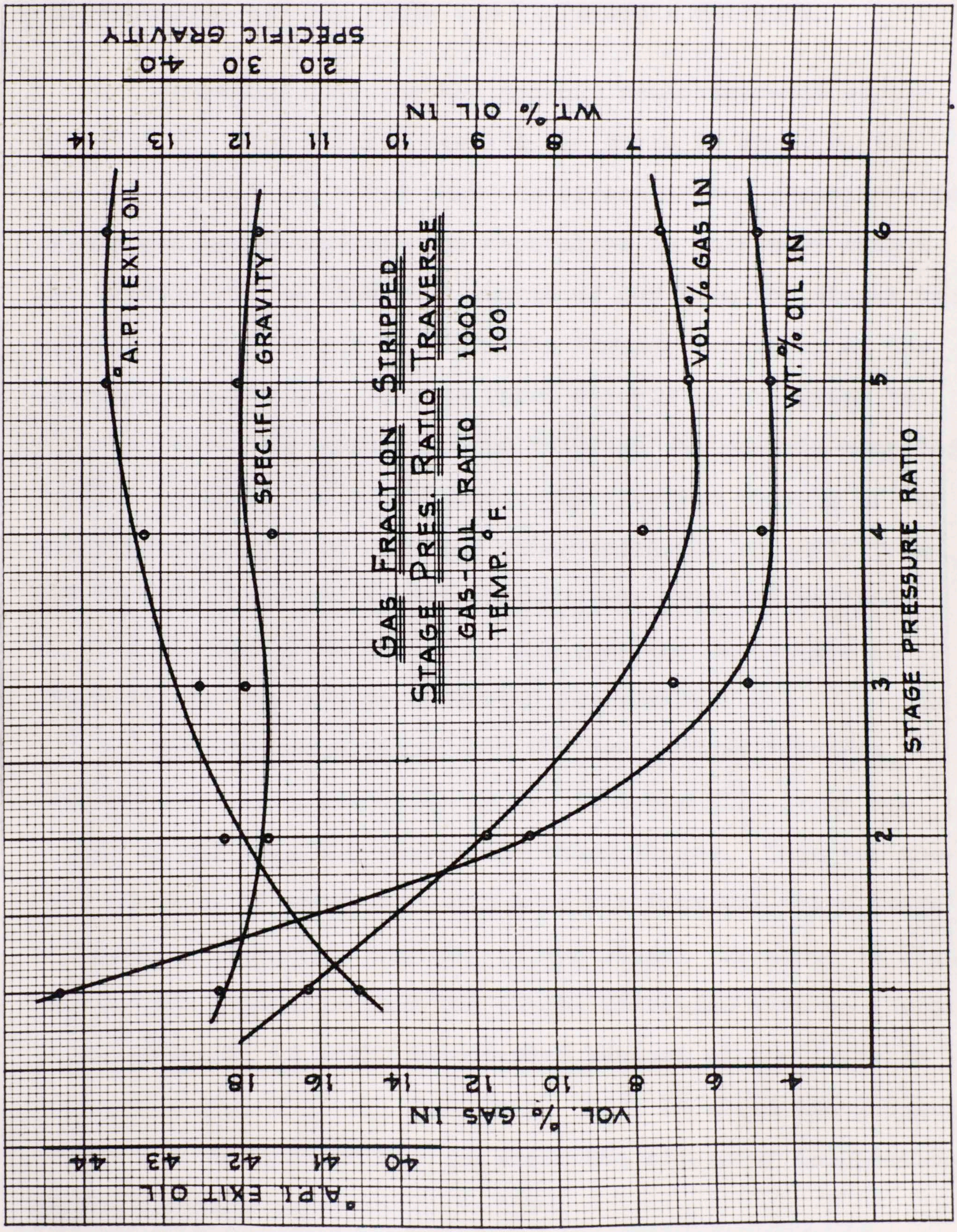

GAS FRACTION STRIPPED
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP. °F 100
STAGE PRESSURE RATIO
VOL. % GAS IN
WT. % OIL IN
°A.P.I. EXIT OIL
SPECIFIC GRAVITY

Fig. 8

CHAPTER IV

EXPERIMENTAL GAS OIL RATIO AND TEMPERATURE TRAVERSES

Corrected data for the gas-oil ratio traverse appear in Table II. The graphed results are shown in Figs. 9, 10, and 11. It will be noted that the effect of the gas-oil ratio on the gas gravities from the separators is only very slight. The gravity of the increased amount of gas from the stages was maintained by a deeper cut into the lighter fractions of the constant oil charge. This is shown clearly on the graph in Fig. 11, where the weight of the fraction stripped is compared to the weight of the oil charge. The specific gravity of the gas fraction stripped from the crude increases steadily from two to four, over the gas-oil ratio range from 500 to 1500. In this traverse, the volume of gas from the separators is compared, in Fig. 10, to a varying gas charge.

The only irregularity of the temperature traverse, the corrected data for which appear in Table III, is in the volume of gas from the second and third stages. This was the only traverse in which the volume of gas from the third stage was greater than that from the second stage. As shown in Fig. 12, the gas gravities from the stages increases with an increase in temperature, as would be expected because of the increase in vapor pressure of all fractions. Fig. 13 shows the irregularity in gas volumes and Fig. 14 shows the effect of temperature on the fraction stripped.

TABLE NO. II

GAS-OIL RATIO TRAVERSE

CONSTANTS

Bath Temp. °F.	100
Stage Pres. Ratio	4
S.G. Gas In	0.610
A.P.I. Oil In	46.0
Lbs. Oil In	8.05
Length of Runs min.	15

Run No.	41	42	43	44	45
Gas-Oil Ratio	500	750	1000	1250	1500
S.C.F. Gas In	14.45	21.7	28.9	36.1	43.4
Lbs. Gas In	0.657	0.987	1.314	1.643	1.971
EXIT FIRST STAGE					
Pres. #/in.^{2} abs.	240	240	240	240	240
S.C.F.	13.7	20.8	28.3	35.2	42.4
Vol. % Gas In	94.8	95.8	97.9	97.5	97.6
S.G. Gas	0.714	0.720	0.715	0.713	0.714
Lbs. Gas	0.729	1.117	1.507	1.870	2.255
EXIT SECOND STAGE					
Pres. #/in.^{2} abs.	60	60	60	60	60
S.C.F.	1.57	1.67	1.47	1.38	1.51
Vol. % Gas In	10.9	7.80	5.09	3.82	3.48
S.G. Gas	0.965	0.975	0.975	0.957	0.959
Lbs. Gas	0.113	0.121	0.107	0.099	0.108
EXIT THIRD STAGE					
Pres. #/in.^{2} abs.	15	15	15	15	15
S.C.F.	1.48	1.59	1.34	1.36	1.26
Vol. % Gas In	10.2	7.33	4.63	3.77	2.90
S.G. Gas	1.71	1.71	1.69	1.68	1.65
Lbs. Gas	0.189	0.202	0.169	0.170	0.155
°A.P.I. Oil	43.9	43.8	43.6	43.5	43.5
TOTAL EXIT GAS					
S.C.F.	16.75	24.06	31.11	37.94	45.17
Lbs.	1.031	1.440	1.783	2.138	2.518
GAS FRACTION STRIPPED					
S.C.F.	2.3	2.4	2.2	1.8	1.8
Vol. % Gas In	15.9	11.1	7.61	4.98	4.15
Lbs.	0.374	0.453	0.469	0.495	0.547
Wt. % Oil In	2.18	2.53	2.86	3.69	4.08

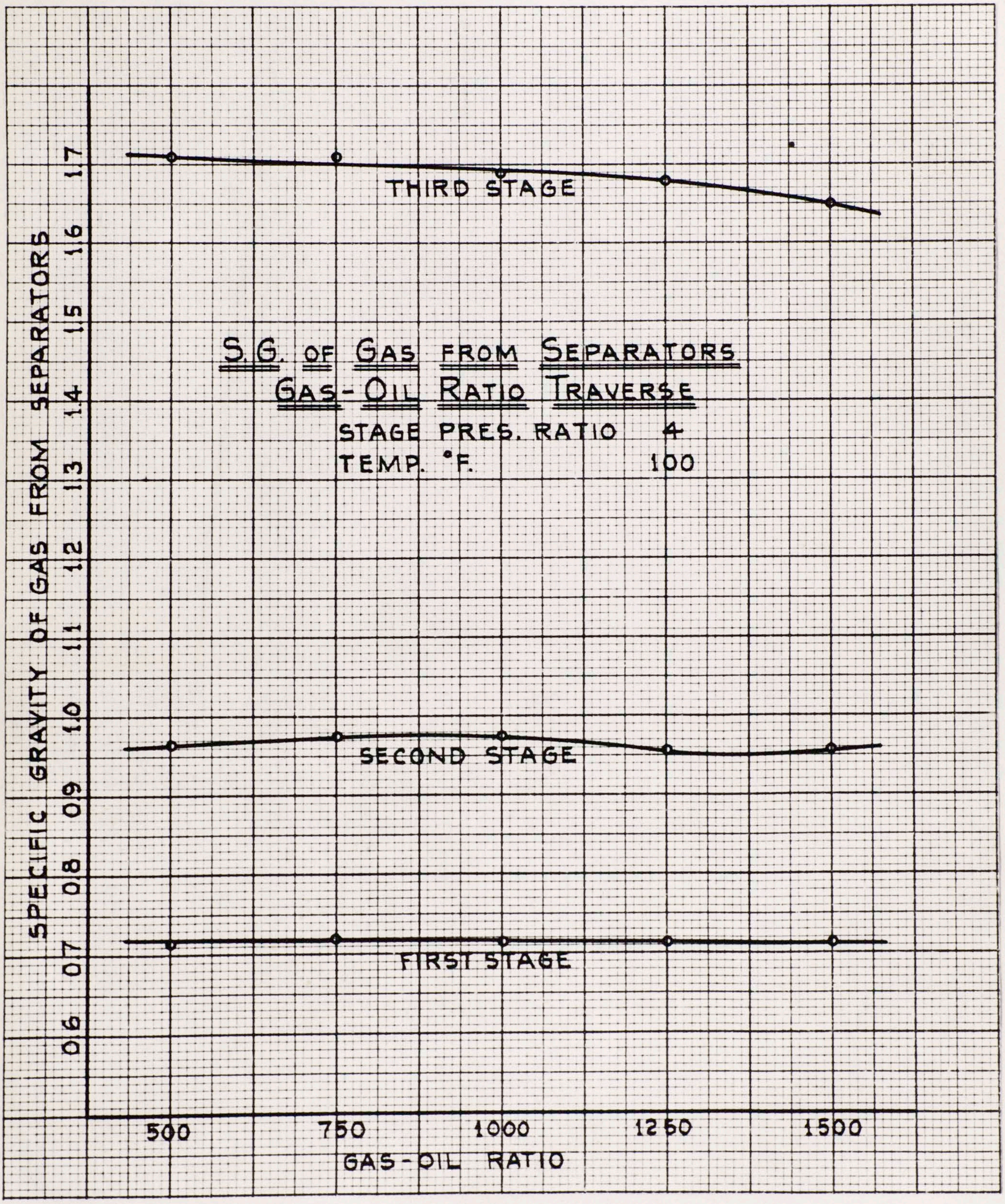

S.G. OF GAS FROM SEPARATORS
GAS-OIL RATIO TRAVERSE
STAGE PRES. RATIO 4
TEMP. °F. 100
THIRD STAGE
SECOND STAGE
FIRST STAGE
SPECIFIC GRAVITY OF GAS FROM SEPARATORS
0.6
0.7
0.8
0.9
1.0
1.1
1.2
1.3
1.4
1.5
1.6
1.7
500
750
1000
1250
1500
GAS-OIL RATIO

Fig. 9

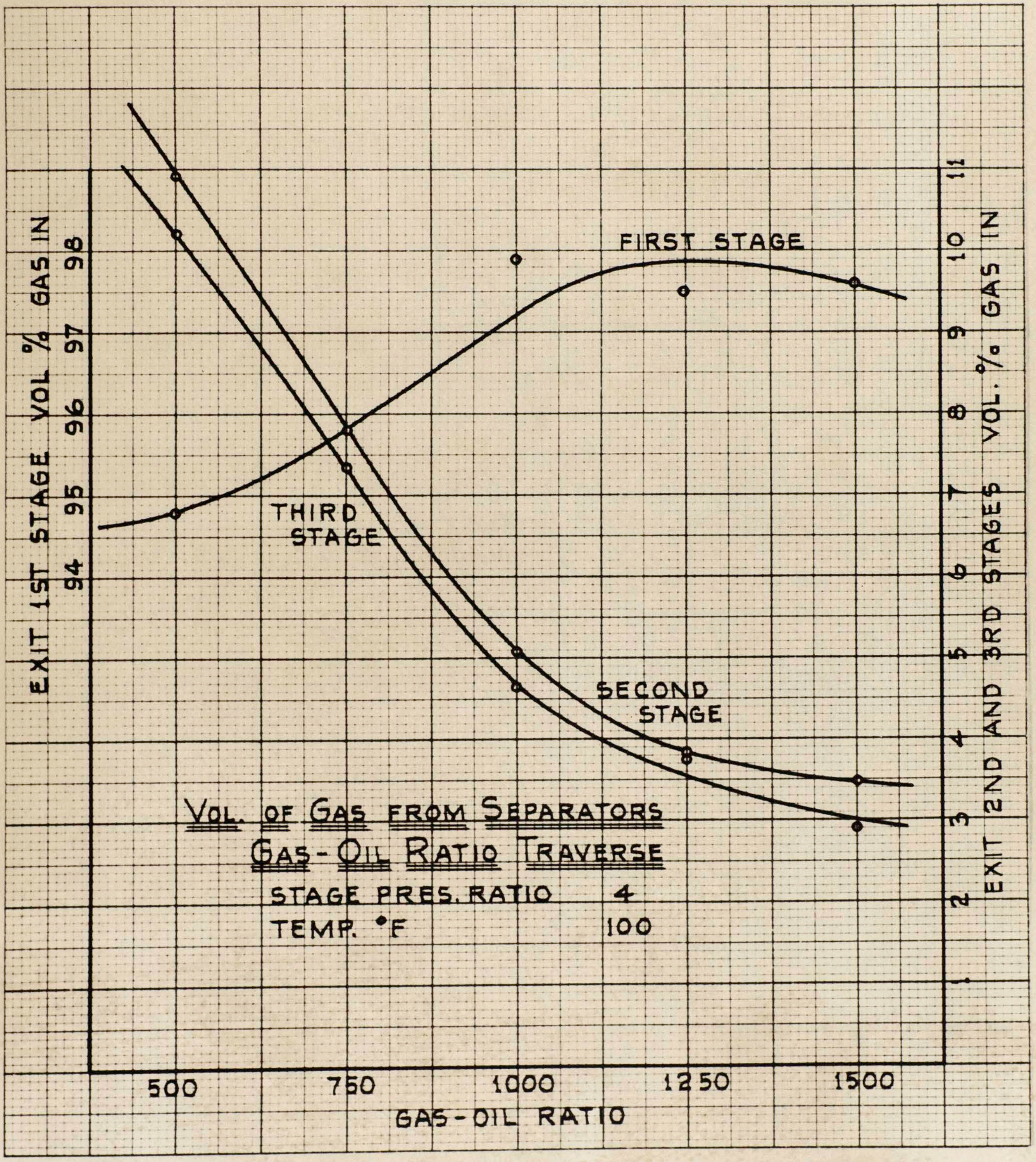


Fig. 10

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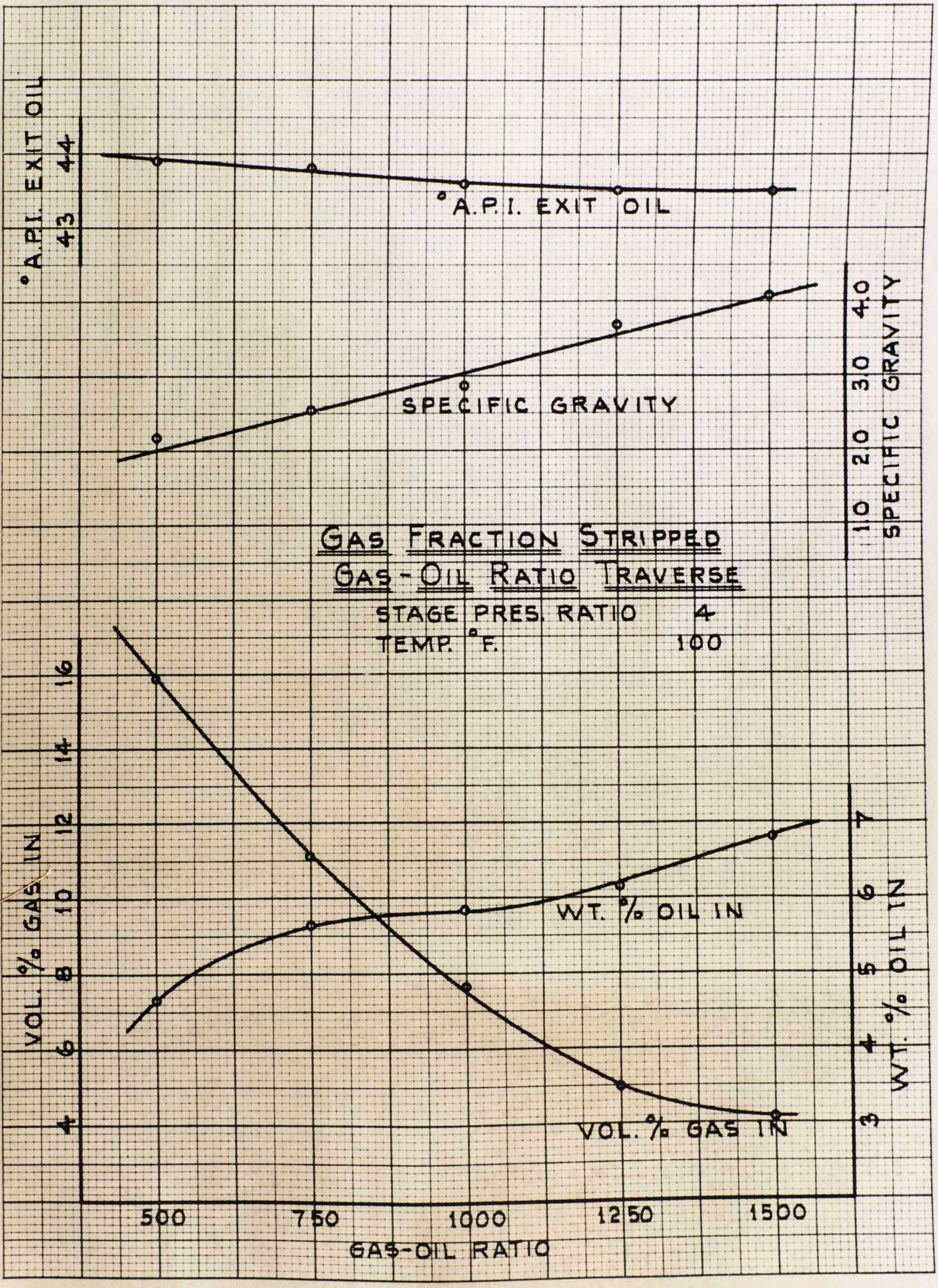


°A.P.I. EXIT OIL
43
44
SPECIFIC GRAVITY
1.0 2.0 3.0 4.0
GAS FRACTION STRIPPED
GAS-OIL RATIO TRAVERSE
STAGE PRES. RATIO 4
TEMP. °F. 100
VOL. % GAS IN
4 6 8 10 12 14 16
WT. % OIL IN
3 4 5 6 7
500 750 1000 1250 1500
GAS-OIL RATIO

Fig. 11

TABLE NO. III

TEMPERATURE TRAVERSE

CONSTANTS

Gas-Oil Ratio	1000
Stage Pres. Ratio	4
S.C. F. Gas In	28.90
S.G. Gas In	0.610
Lbs. Gas In	1.314
°A.P.I. of Oil In	46.0
Lbs. Oil In	8.05
Length of Runs min.	15

Run No.	46	47	48	49	50
Bath Temp. °F.	80	90	100	110	120
EXIT FIRST STAGE					
Pres. #/in.^{2} abs.	240	240	240	240	240
S.C.F.	28.70	28.80	28.50	28.90	29.50
Vol. % Gas In	99.3	99.7	98.6	100	102
S.G. Gas	0.675	0.688	0.702	0.724	0.741
Lbs. Gas	1.444	1.477	1.491	1.560	1.630
EXIT SECOND STAGE					
Pres. #/in.^{2} abs.	60	60	60	60	60
S.C.F.	1.36	1.36	1.41	1.45	1.51
Vol. % Gas In	4.71	4.71	4.88	5.02	5.23
S.G. Gas	0.877	0.905	0.938	0.988	1.04
Lbs. Gas	0.089	0.092	0.099	0.107	0.117
EXIT THIRD STAGE					
Pres. #/in.^{2} abs.	15	15	15	15	15
S.C.F.	0.986	1.01	1.13	1.38	1.58
Vol. % Gas In	3.41	3.49	3.91	4.78	5.47
S.G. Gas	1.53	1.58	1.63	1.69	1.76
Lbs. Gas	0.112	0.119	0.137	0.174	0.207
°A.P.I. Oil	44.2	44.2	43.9	43.3	42.5
TOTAL EXIT GAS					
S.C.F.	31.05	31.17	31.04	31.73	32.59
Lbs.	1.645	1.688	1.727	1.841	1.954
GAS FRACTION STRIPPED					
S.C.F.	2.15	2.27	2.14	2.83	3.69
Vol. % Gas In	7.44	7.85	7.40	9.80	12.8
Lbs.	0.331	0.374	0.413	0.527	0.640
Wt. % Oil In	4.11	4.65	5.13	6.55	7.95
S.G.	2.06	2.21	2.59	2.50	2.32

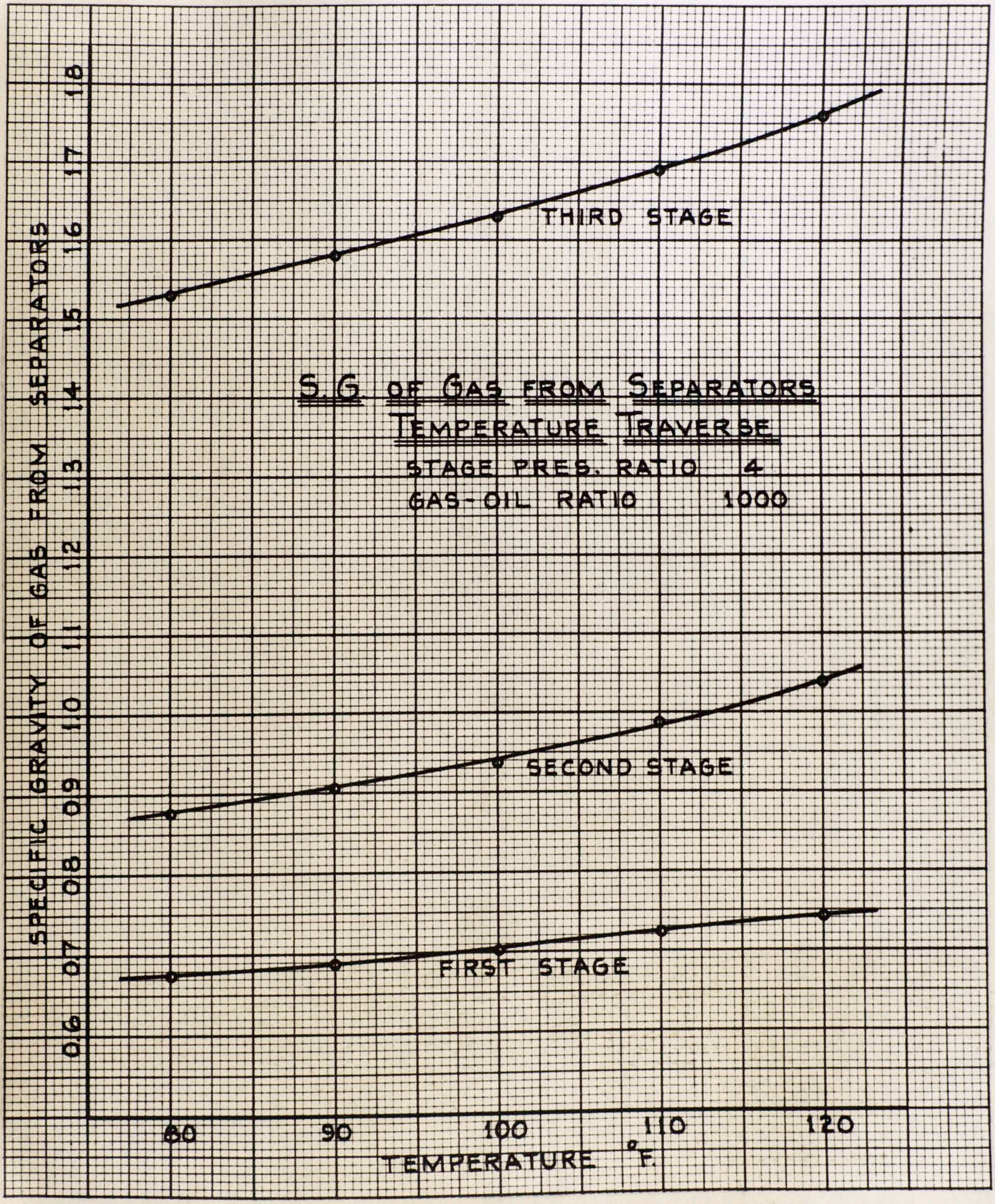

S.G. of Gas from Separators
Temperature Traverse
STAGE PRES. RATIO 4
GAS-OIL RATIO 1000
THIRD STAGE
SECOND STAGE
FIRST STAGE
SPECIFIC GRAVITY OF GAS FROM SEPARATORS
0.6
0.7
0.8
0.9
1.0
1.1
1.2
1.3
1.4
1.5
1.6
1.7
1.8
80
90
100
110
120
TEMPERATURE °F

Fig. 12

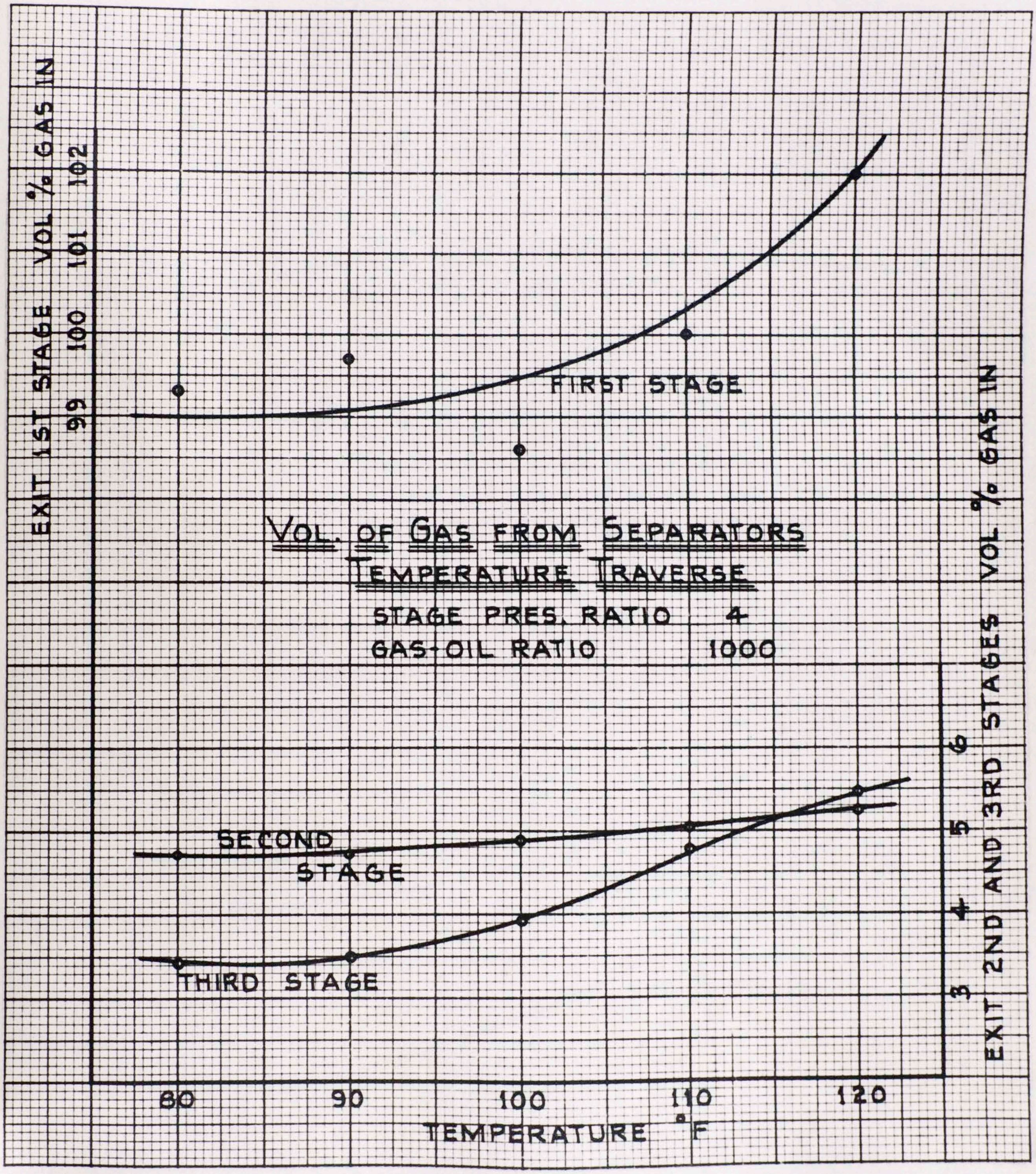
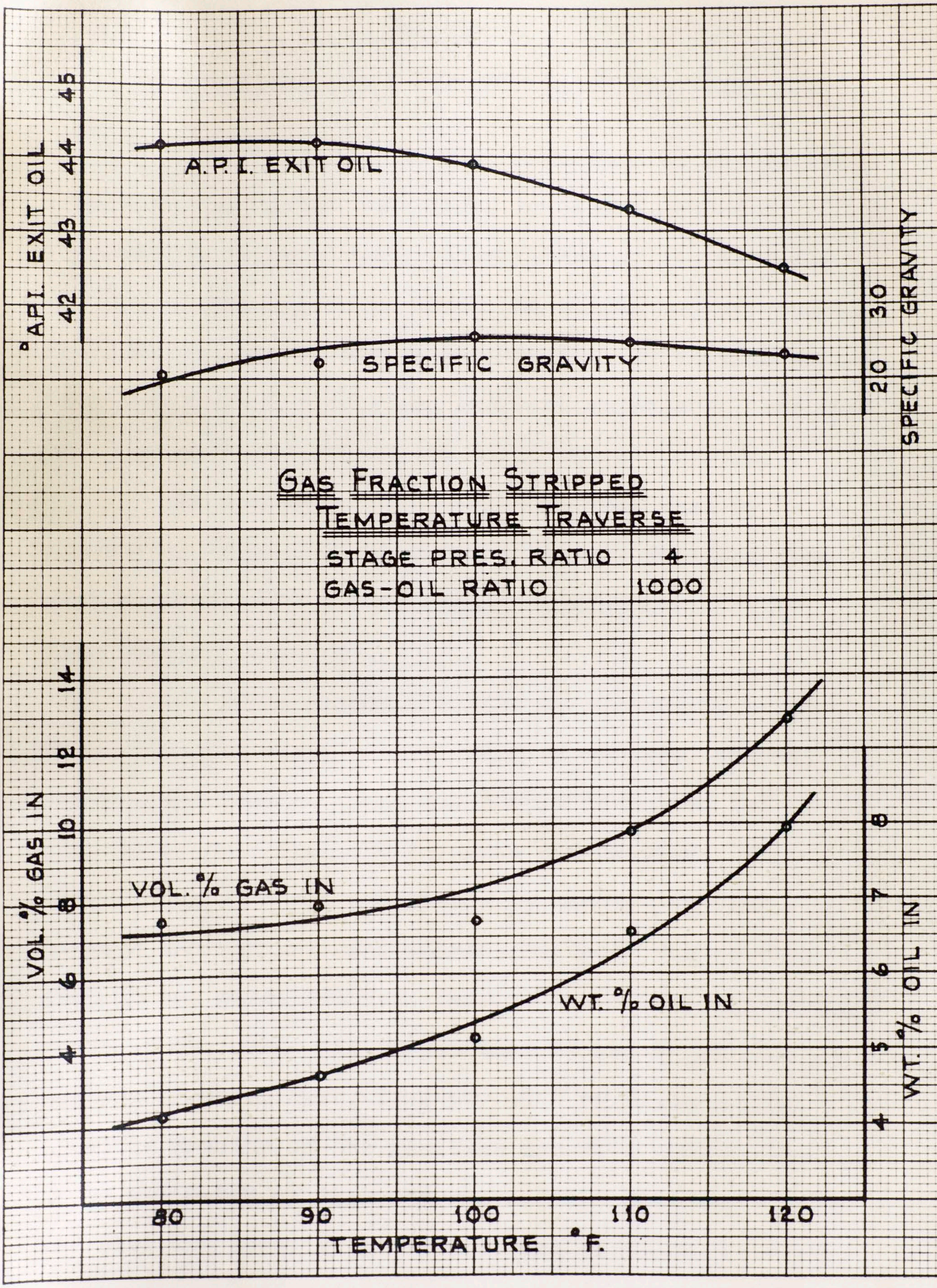

EXIT 1ST STAGE VOL % GAS IN
99
100
101
102
FIRST STAGE
VOL. OF GAS FROM SEPARATORS
TEMPERATURE TRAVERSE
STAGE PRES. RATIO 4
GAS-OIL RATIO 1000
SECOND STAGE
THIRD STAGE
EXIT 2ND AND 3RD STAGES VOL % GAS IN
3
4
5
6
80
90
100
110
120
TEMPERATURE °F

Fig. 13



45
44
43
42
°A.P.I. EXIT OIL
A.P.I. EXIT OIL
SPECIFIC GRAVITY
20
30
SPECIFIC GRAVITY
GAS FRACTION STRIPPED
TEMPERATURE TRAVERSE
STAGE PRES. RATIO 4
GAS-OIL RATIO 1000
14
12
10
8
6
4
VOL. % GAS IN
VOL. % GAS IN
WT. % OIL IN
8
7
6
5
4
WT. % OIL IN
80
90
100
110
120
TEMPERATURE °F.

Fig. 14

CHAPTER V

THEORETICAL STAGE PRESSURE RATIO TRAVERSE

This traverse, the most important of the investigation, was chosen over which to make a set of theoretical calculations. Examination of the data did not reveal the degree to which the stages had approached equilibrium, and it was desired to show by another means, that equilibrium conditions had been approached, and that the data were not subject to large unknown errors.

The method of calculation is that used by R. L. Huntington in his class work, originally obtained from Mott Souders, now with the Shell Development Company, Long Beach, California.

The total charging stock was assumed to be a five component mixture consisting of methane, ethane, butane, and Tonkawa crude oil broken down into two fractions, gasoline, and kerosene and heavier. The natural gas used was substantially ninety per cent methane and ten per cent ethane, so these values were used in the calculations. Examination of the distillation curve of the crude oil showed that it had a gasoline content of about forty-five per cent by volume. The specific gravity of the gasoline was 0.75 and of the crude oil 0.818 or 41.4 degrees A.P.I. From the volume per cent and specific gravity of the gasoline fraction, and the specific gravity of the crude oil, the volume per cent and specific gravity of the kerosene and heavier was calculated.

The mol composition of the exit gas and oil, and the volume overhead were calculated by the formula developed below.

Basis 1 mol

$$X_f = VY_v + (1 - V) X_l \quad (1)$$

or

$$Y_v = \frac{X_f - (1 - V) X_l}{V} \quad (2)$$

$$Y_v = KX_l \quad \text{Henry's Law} \quad (3)$$

Sub. (3) in (1) $X_f = VKX_l + (1 - V) X_l \quad (4)$

Simplifying (4) $X_f = X_l (1 + V (K - 1)) \quad (5)$

Solving (5) for X_l

$$X_l = \frac{X_f}{1 + V (K - 1)}$$

where: X_f = the mol fraction of any one component in the feed

V = mols overhead

Y_v = the mol fraction of any one component in the overhead

X_l = the mol fraction of any one component in the bottoms

K = vaporization equilibrium constant

The following table shows the calculated composition of the feed.

Component	Mol. Wt.	Pounds	Mols	Mol fraction of total charge
CH_4	16	1.087	0.0679	0.514
C_2H_6	30	0.227	0.0076	0.058
C_4H_{10}	58	0.732	0.0126	0.095
Gasoline	110	7.32	0.0265	0.200
Bottoms	250		0.0176	0.133
			0.1322	1.000

In Figs. 15, 16, and 17 are shown the vaporization equilibrium constants for methane, ethane, and butane, within the range of pressures and temperatures used in the calculations. The curves are extrapolated cross plots of data taken by Katz and Hachmuth.[4]

The assumed vapor pressures of the crude oil fractions are shown in Fig. 18. The basis of the curves is an actual vapor pressure curve of the crude oil. The sum of the partial pressures of the two fractions was made equal to the actual vapor pressure.

The calculated values of the traverse are shown in Tables IV and V, and plotted on the graphs in Figs. 19, 20, and 21. In order to determine the volume overhead, values for V were assumed until the sum of the X_1 terms was equal to unity.

TABLE NO. IV

THEORETICAL CALCULATIONS
STAGE PRESSURE RATIO TRAVERSE

CONSTANTS	
Gas-Oil Ratio	1000
Temp.°F.	100
S.C.F. Gas In	29.20
S.G. Gas In	0.605
Lbs. Gas In	1.314
°A.P.I. of Oil In	48.5
Lbs. Oil In	8.05

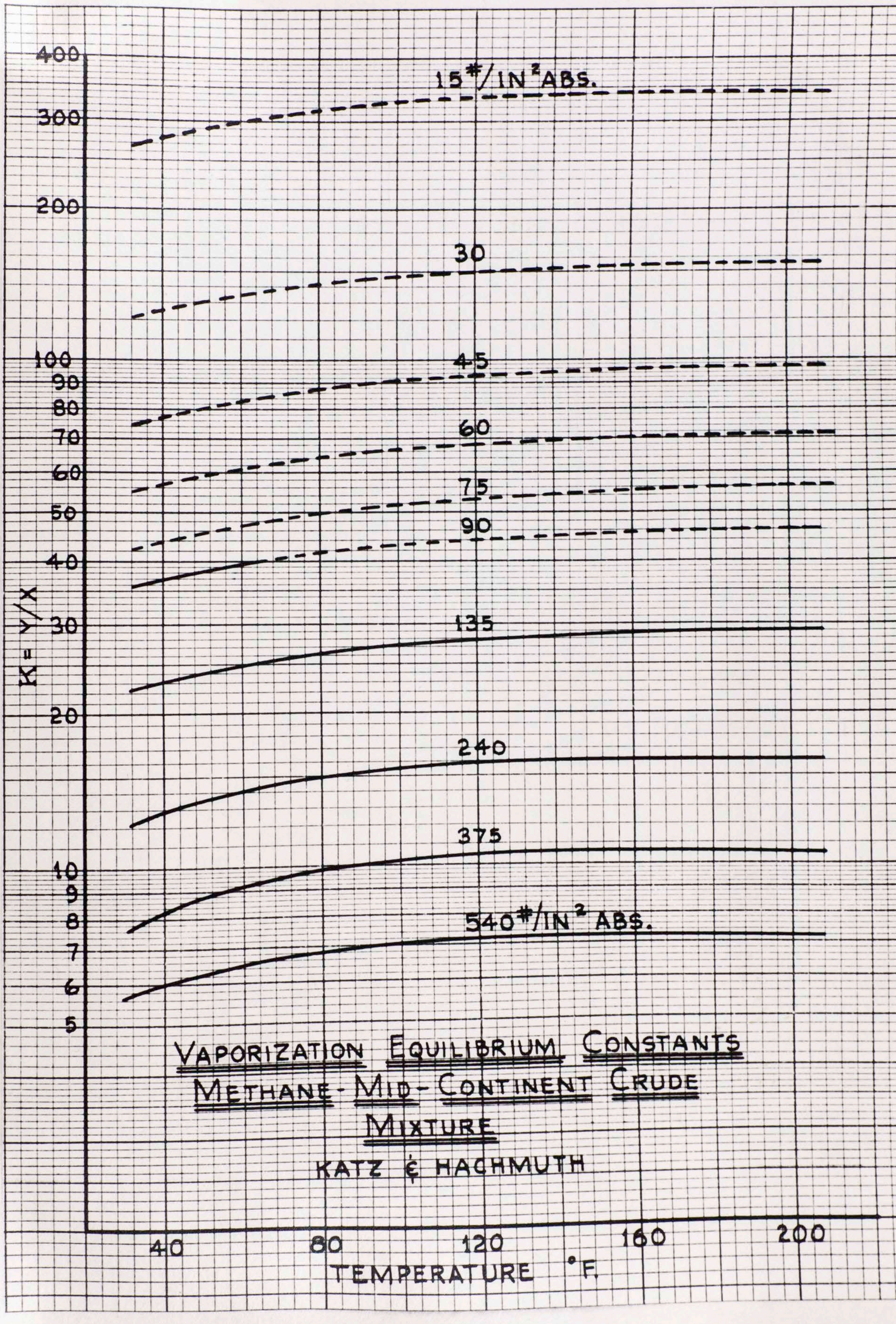
Stage Pres. Ratio	2	3	4	5	6
EXIT FIRST STAGE					
Pres. $\#/in.^2$ abs.	60	135	240	375	540
S.C.F.	32.3	30.3	28.6	27.2	25.6
Vol. % Gas In	111	104	97.9	93.2	87.7
S.G. Gas	0.765	0.705	0.667	0.645	0.631
Lbs. Gas	1.866	1.593	1.423	1.308	1.204
EXIT SECOND STAGE					
Pres. $\#/in.^2$ abs.	30	45	60	75	90
S.C.F.	0.22	0.82	1.58	2.68	4.10
Vol. % Gas In	0.75	2.81	5.41	9.18	14.0
S.G. Gas	0.941	0.912	0.916	0.868	0.826
Lbs. Gas	0.015	0.056	0.108	0.173	0.253
EXIT THIRD STAGE					
Pres. $\#/in.^2$ abs.	15	15	15	15	15
S.C.F.	0.145	0.60	1.02	1.49	1.87
Vol. % Gas In	0.50	2.05	3.49	5.10	6.40
S.G. Gas	1.30	1.51	1.61	1.59	1.56
Lbs. Gas	0.014	0.070	0.122	0.177	0.217
°A.P.I. Oil	44.5	45.8	46.5	46.5	45.9
TOTAL EXIT GAS					
S.C.F.	32.67	31.72	31.20	31.37	31.57
Lbs. Gas	1.895	1.718	1.653	1.658	1.674
GAS FRACTION STRIPPED					
S.C.F.	3.47	2.52	2.00	2.17	2.37
Vol. % Gas In	11.9	8.63	6.85	7.43	8.11
Lbs.	0.581	0.405	0.339	0.344	0.360
Wt. % Oil In	7.22	5.03	4.21	4.27	4.47
S.G.	2.24	2.16	2.27	2.13	1.92

TABLE NO. V

EXPERIMENTAL DATA AND THEORETICAL CALCULATIONS ON SINGLE STAGE

CONSTANTS	EXPERIMENTAL	THEORETICAL
Gas-Oil Ratio	1000	1000
Temp. $^{\circ}$F.	100	100
S.C.F. Gas In	28.90	29.20
S.G. Gas In	0.610	0.605
Lbs. Gas In	1.314	1.314
$^{\circ}$A.P.I. Oil In	46.0	48.5
Lbs. Oil In	8.05	8.05
Length of Run min.	15	
Run No.	51	
Stage Pres. Ratio*	1	1
EXIT SINGLE STAGE		
Pres. #/in.^{2} abs.	15	15
S.C.F.	33.6	34.8
Vol. % Gas In	116	123
S.G. Gas	0.983	0.896
Lbs. Gas	2.460	2.327
GAS FRACTION STRIPPED		
S.C.F.	4.70	5.60
Vol. % Gas In	16.3	19.2
Lbs.	1.15	1.013
Wt. % Oil In	14.3	12.6
S.G.	3.28	2.43

* Flashing gas and taking oil from one separator is equivalent to holding three stages at the same pressure. All the gas would flash from the first stage (there being no pressure drop into the second or third). The oil would simply flow through the second and third separator without changing composition.



15#/IN² ABS.
30
45
60
75
90
135
240
375
540#/IN² ABS.
400
300
200
100
90
80
70
60
50
40
30
20
10
9
8
7
6
5
K = Y/X
40
80
120
160
200
TEMPERATURE °F.
VAPORIZATION EQUILIBRIUM CONSTANTS
METHANE-MID-CONTINENT CRUDE
MIXTURE
KATZ & HACHMUTH

Fig. 15

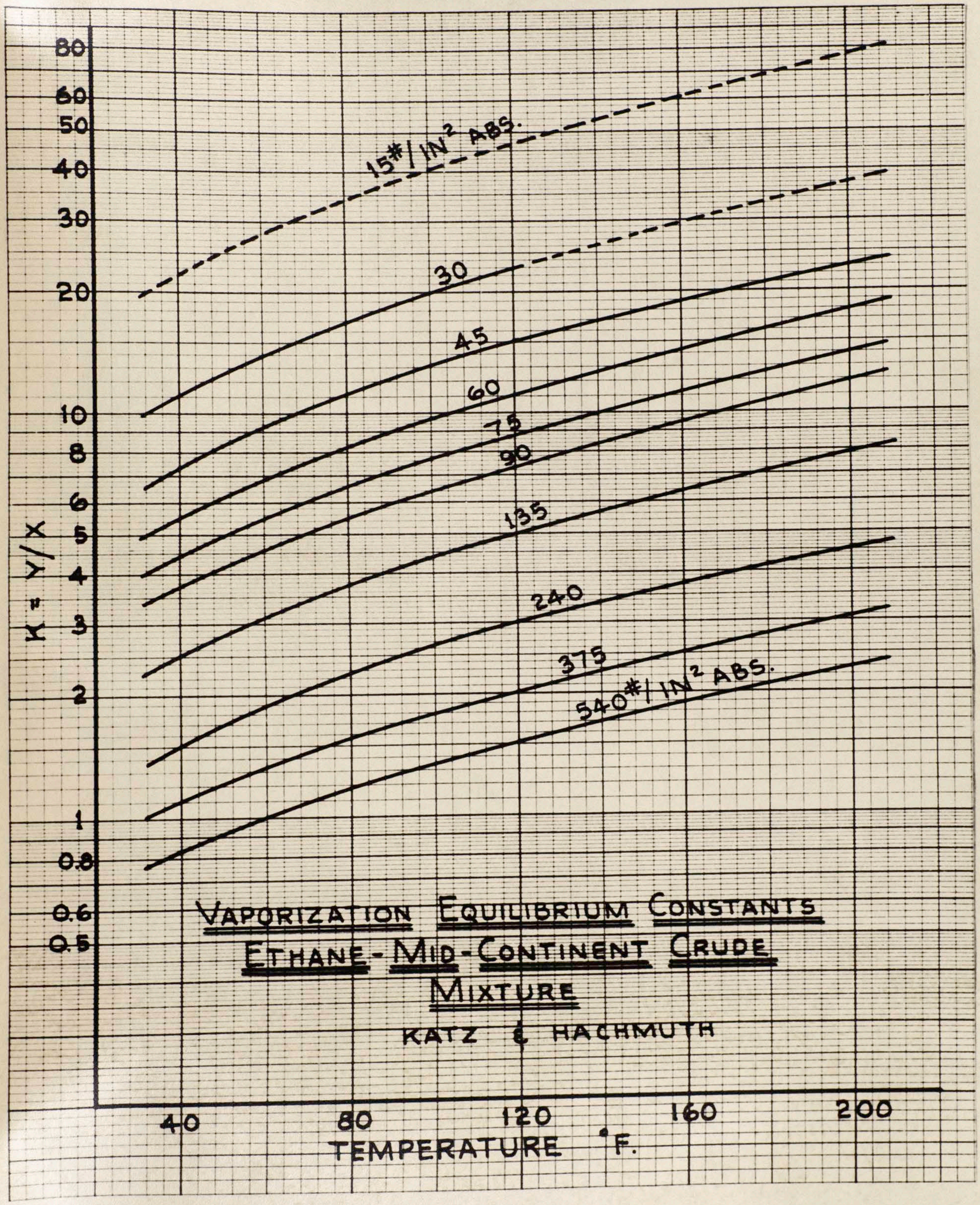

80
60
50
40
30
20
10
8
6
5
4
3
2
1
0.8
0.6
0.5
K = Y/X
15#/IN² ABS.
30
45
60
75
90
135
240
375
540#/IN² ABS.
VAPORIZATION EQUILIBRIUM CONSTANTS
ETHANE - MID - CONTINENT CRUDE
MIXTURE
KATZ & HACHMUTH
40
80
120
160
200
TEMPERATURE °F.

Fig. 16

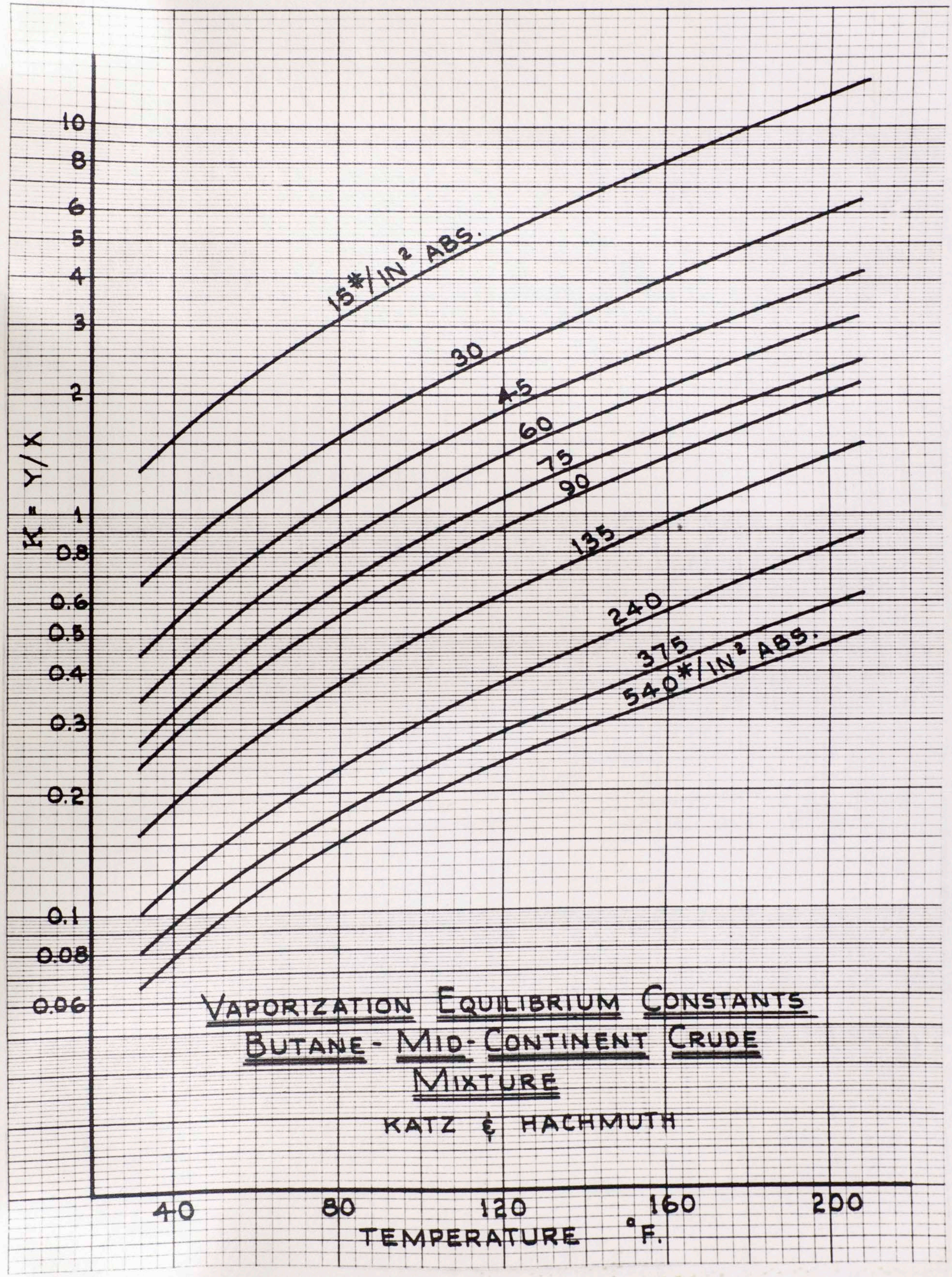

VAPORIZATION EQUILIBRIUM CONSTANTS
BUTANE - MID-CONTINENT CRUDE
MIXTURE
KATZ & HACHMUTH
K = Y/X
TEMPERATURE °F.
10
8
6
5
4
3
2
1
0.8
0.6
0.5
0.4
0.3
0.2
0.1
0.08
0.06
40
80
120
160
200
15#/IN² ABS.
30
45
60
75
90
135
240
375
540#/IN² ABS.

Fig. 17

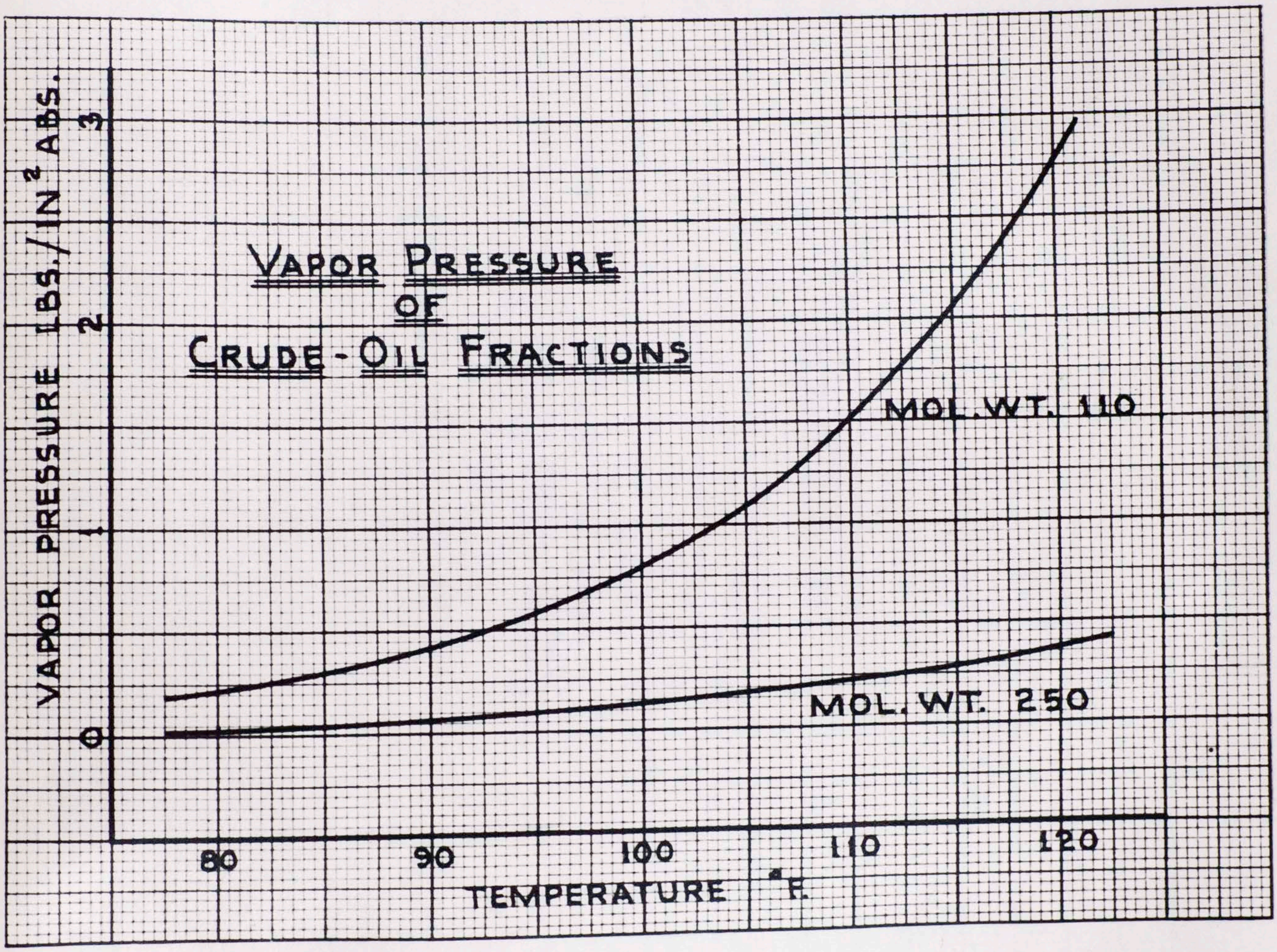

VAPOR PRESSURE
OF
CRUDE-OIL FRACTIONS
MOL. WT. 110
MOL. WT. 250
VAPOR PRESSURE LBS./IN² ABS.
0
1
2
3
80
90
100
110
120
TEMPERATURE °F.

Fig. 18

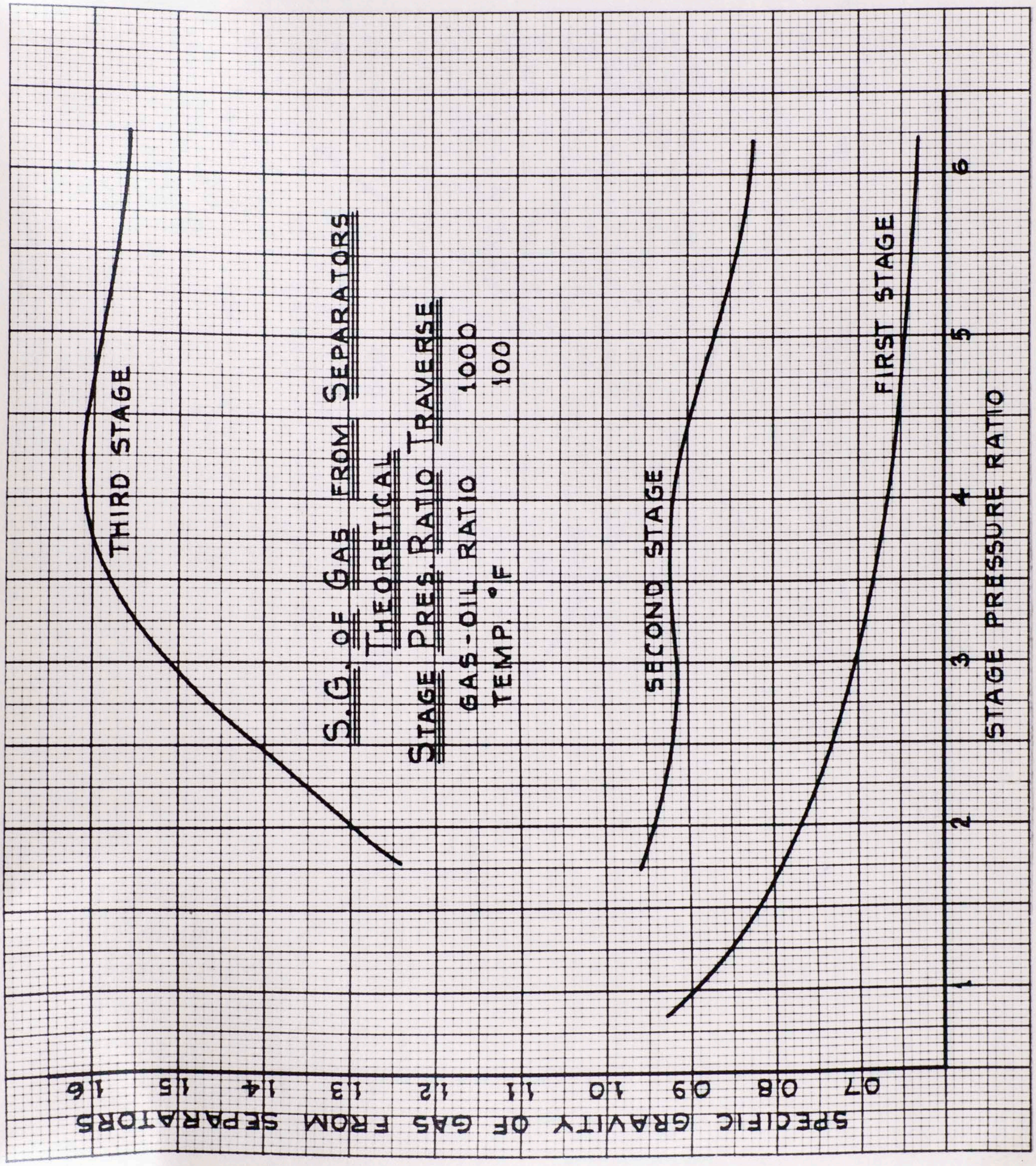

S.G. OF GAS FROM SEPARATORS
THEORETICAL
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP. °F 100
THIRD STAGE
SECOND STAGE
FIRST STAGE
STAGE PRESSURE RATIO
1
2
3
4
5
6
SPECIFIC GRAVITY OF GAS FROM SEPARATORS
0.7
0.8
0.9
1.0
1.1
1.2
1.3
1.4
1.5
1.6

Fig 19

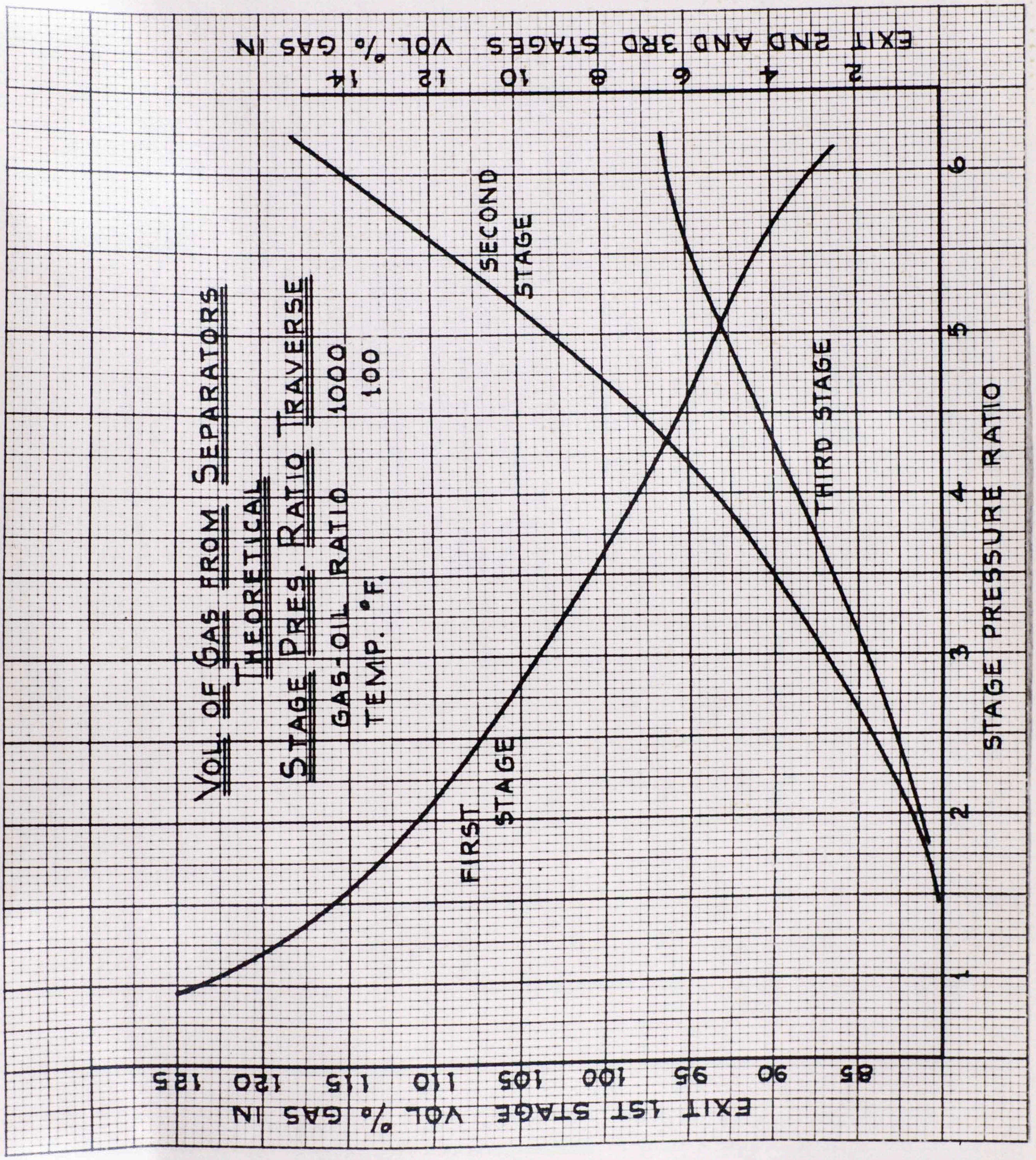

VOL. OF GAS FROM SEPARATORS
THEORETICAL
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP. °F. 100
FIRST STAGE
SECOND STAGE
THIRD STAGE
STAGE PRESSURE RATIO
1
2
3
4
5
6
EXIT 1ST STAGE VOL % GAS IN
85
90
95
100
105
110
115
120
125
EXIT 2ND AND 3RD STAGES VOL % GAS IN
2
4
6
8
10
12
14

Fig. 20

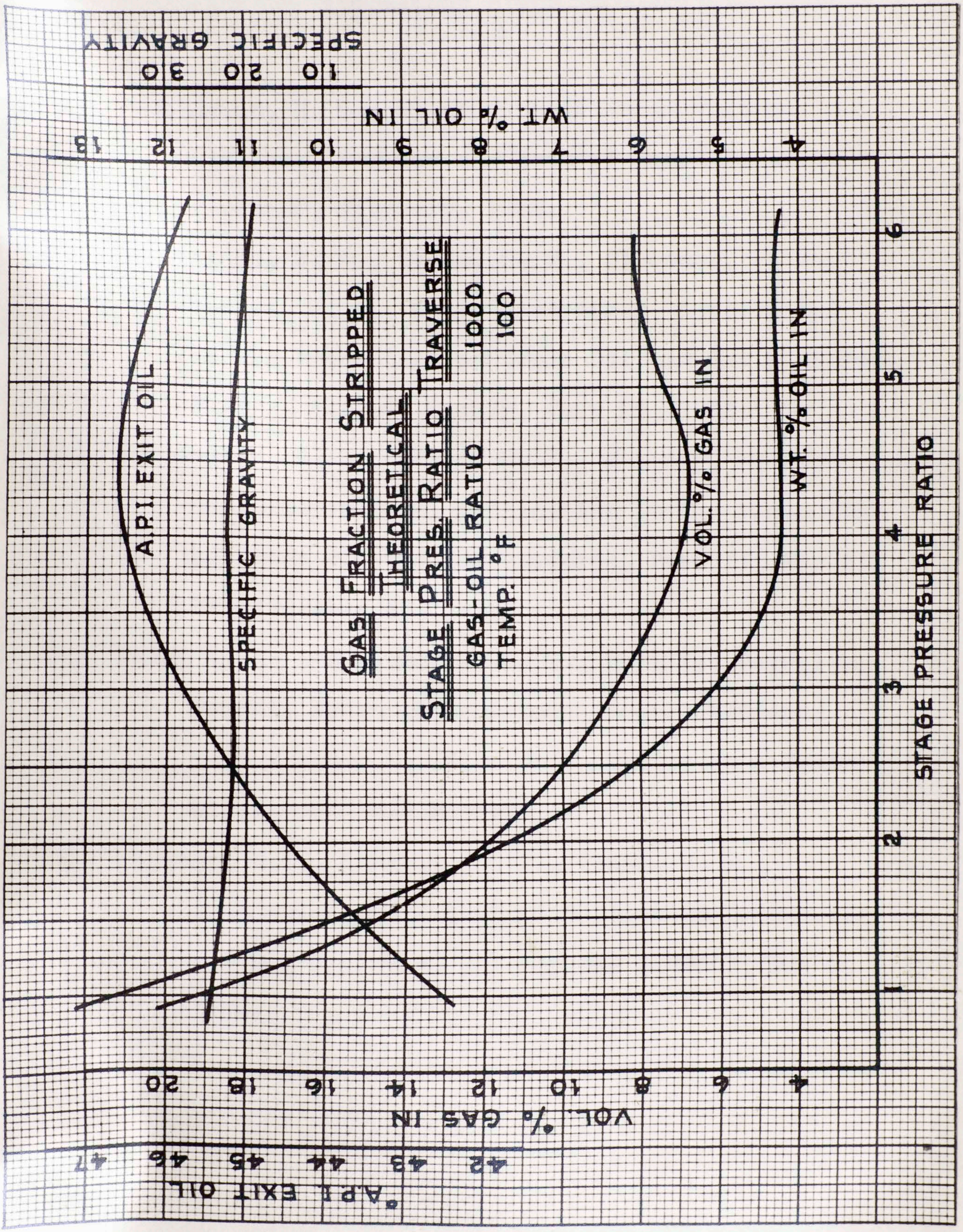

GAS FRACTION STRIPPED
THEORETICAL
STAGE PRES. RATIO TRAVERSE
GAS-OIL RATIO 1000
TEMP. °F 100
API EXIT OIL
SPECIFIC GRAVITY
VOL. % GAS IN
WT. % OIL IN
STAGE PRESSURE RATIO
°API EXIT OIL

Fig. 21

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(5) Matheny, W.F., Langham, J. F., Cloud, W. F., Huntington, R. L., Stage Separation Crude Oil and Gas Mixtures, Oil and Gas Journal, May 2, 1935.