



Feasibility study of coal-hydrocarbon slurries
by James Brent Blazek

A thesis submitted in partial fulfillment of the requirements for the degree of MASTER OF SCIENCE
in Chemical Engineering
Montana State University
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Abstract:

Finely ground coal-hydrocarbon slurries were studied to determine their properties. The research was separated into two parts, an initial phase to determine if the slurries would set up under pipeline pressure conditions and a more detailed look at a crude oil—coal slurry. Brookfield and Stormer viscometers were used to measure viscosities.

It was found that the viscosities of most coal-hydrocarbon slurries did tend to increase after pressurization to 500 and 1000 psig for several weeks. However, it was felt that this viscosity increase was not severe enough to eliminate any of the liquid carriers.

Rheograms obtained from a Brookfield rotational viscometer for a range of coal—crude oil slurries from 0% to 50% by weight showed four distinct rheological models.

Slurries of crude oil and coal were pumped through a one inch schedule 40 pipe to determine their flow characteristics. Various methods were used to compare the effective viscosities obtained from the pipeloop's pressure drop to the apparent viscosities measured by the Brookfield viscometer. The agreement was not good. A method derived for Bingham plastics to predict pipeline pressure drops from viscometer data was tried. Its success in predicting pressure drops was found to be limited.

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June 27, 1975

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SLURRIES

by

JAMES BRENT BLAZEK

A thesis submitted in partial fulfillment
of the requirements for the degree

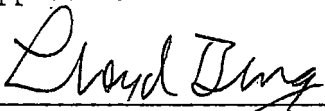
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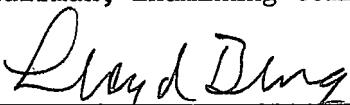
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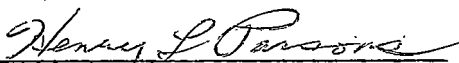
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ABSTRACT

Finely ground coal-hydrocarbon slurries were studied to determine their properties. The research was separated into two parts, an initial phase to determine if the slurries would set up under pipeline pressure conditions and a more detailed look at a crude oil-coal slurry. Brookfield and Stormer viscometers were used to measure viscosities.

It was found that the viscosities of most coal-hydrocarbon slurries did tend to increase after pressurization to 500 and 1000 psig for several weeks. However, it was felt that this viscosity increase was not severe enough to eliminate any of the liquid carriers.

Rheograms obtained from a Brookfield rotational viscometer for a range of coal-crude oil slurries from 0% to 50% by weight showed four distinct rheological models.

Slurries of crude oil and coal were pumped through a one inch schedule 40 pipe to determine their flow characteristics. Various methods were used to compare the effective viscosities obtained from the pipeloop's pressure drop to the apparent viscosities measured by the Brookfield viscometer. The agreement was not good. A method derived for Bingham plastics to predict pipeline pressure drops from viscometer data was tried. Its success in predicting pressure drops was found to be limited.

INTRODUCTION AND BACKGROUND

A major factor in determining which mineral deposits have been first exploited has been the transportation costs of the raw materials. The sources of raw materials which were most accessible are now depleted or already committed to the market. At the same time, industrial expansion has been great worldwide, which has created an intense demand for raw materials. This has meant that the more inaccessible raw materials have begun to be developed.

This situation is especially true in the United States which consumes about 40 percent of the world's energy and is experiencing a rapid growth rate in energy demand while the accessible proven reserves are dropping. The Bureau of Mines has predicted a 3.5 percent yearly increase in energy demand until the year 2000.

At the same time, there is increased pressure within the United States to become energy independent. The recent Arab oil embargo vividly pointed out how undependable foreign importation of raw materials for energy can be. Project Independence which calls on the nation to be energy independent by the year 1985 has been endorsed by both Presidents Nixon and Ford.

Fortunately, the United States has massive reserves of coal, primarily in the Western part of the country. The United States has approximately 40 percent of the world's known coal reserves. Although coal is, of course, another finite resource it can provide a major phase of the nation's energy for a time of about 50 to 100 years. This

time can then be used to develop other long term solutions to the energy problem such as solar energy and nuclear fusion.

The coal in the Eastern part of the country has a high sulfur content and is expensive to mine. Western coal, on the other hand, has a much lower sulfur content, (.5 percent or less compared to 2 to 3 percent for most Eastern coal). It can be burned in present day boilers and meet the 1975 EPA emission requirements. Another attractive feature of Western coal is the fact that it is relatively close to the surface with only about thirty to forty feet of overburden with seams that run as thick as 100 feet. This is opposed to Eastern seams which have more overburden and are not thick. Yet the Eastern reserves have been much more fully developed, mainly because of transportation costs.

One of the major drawbacks of the Western coal reserves is its distance from the major markets. The Midwest offers several large coal markets at a distance of about 500 to 1000 miles from the Western reserves. It can be easily seen that transportation costs are thus a major factor. Increased transportation costs are reflected in an increase in the price of energy, which in turn is reflected in increased production costs for almost all goods or services imaginable. It is therefore necessary to transport the Western coal to its markets by the most economical means possible.

One of the latest approaches to this problem has been to transport the coal as a slurry in a pipeline. This method has recently been

seriously considered as a viable alternative to the other two long distance land transportation alternatives, unit trains and extra high voltage (EHV) transmission.

Several such lines are now in operation and appear to be very promising economically. Two of the largest lines now in existence are the Black Mesa Pipeline and the Gilsonite Pipeline.

The Gilsonite Pipeline went into operation on April 15, 1957. It is a 6 inch line traversing 72 miles and an 8,500 foot mountain pass between American Gilsonite Company's mine at Bonanza, Utah, and its petroleum refinery near Grand Junction, Colorado. It transports 700 tons per day of minus 8-mesh gilsonite as a water slurry. (Gilsonite is a hydrocarbon found only in the Uinta Basin of Utah. It is technically an asphaltite with a specific gravity about 5 percent greater than water).

The Black Mesa pipeline went into operation in 1970. It pumps coal as a slurry with water 273 miles through an 18 inch line. The coal slurry is pumped at a solids' concentration of 45 to 55 percent by weight. The size of the particles is minus 14 mesh. The line handles 4.8 million tons per year.

Coal slurry lines have several advantages over other modes of transportation. A pipeline requires very little labor to operate once it is installed. It is therefore capital intensive rather than labor intensive and thus less sensitive to inflationary pressures. A pipeline is also attractive aesthetically since it is underground.

Another important factor is reliability. Pipelines are not affected by adverse weather conditions and are not greatly affected by labor disputes. Also, pipelines have enabled minerals to be transported over previously inaccessible terrain.

Several factors have been presented by J. G. Montfort, Manager of the Black Mesa Pipeline, indicating when coal slurry lines would be economically feasible. These are 1) when no other transportation presently exists, 2) movements involve long distances and high tonnages, 3) there is a long term requirement, 4) few points of origin and destination, 5) the terrain is rugged and the pipeline can achieve a shorter route since pipelines can be laid on steeper slopes, and 6) a suitable carrier is available. (1)

Although there is a considerable amount of literature available on coal slurries, it is dominated by the use of water as the slurrying material. All the existing coal slurry pipelines use water as the carrier. This, of course, means that the slurry must undergo a separation step at the end of the line before the coal can be fired. It may be possible to bypass this step if the coal was slurried with a liquid that would allow the entire slurry to be fired.

The existing coal-water slurry pipelines combine the separation and grinding steps. If the entire slurry is to be burned the grinding will of necessity take place at the beginning of the lines. This will, therefore, also involve the pumping of a slurry with coal

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particles of significantly smaller size.

RESEARCH OBJECTIVES

PURPOSE

The purpose of this research is to obtain some fundamental data about finely ground coal - hydrocarbon slurries. The current literature focuses almost exclusively on water as the carrier fluid for slurring. Also there is little literature available on coal ground to as fine a mesh size as is used in this research.

The research is intended to be a fundamental look at the feasibility of such a slurry. In addition, it is intended to gain some understanding of the rheology of these slurries and to see what special characteristics they may have or special problems they may present. It is intended to be used as the basis for further research in this area.

OBJECTIVES

The first objective of this research was to look at various types of hydrocarbons to see if there was some reason they would not be suitable as a slurring fluid. The primary concern here was that there would be a tendency for this finely ground coal, under pressure conditions experienced in a pipeline, to absorb the fluid and to set up so that its viscosity would no longer be in an acceptable pumping range. This would eliminate any further testing of unacceptable fluids.

Once this question was resolved, it was hoped that some judgment could be made as to what the most suitable carriers may be. The rheological properties of a coal slurry using one of the fluids was

then to be studied in more detail. This study was to be done in two phases; a viscometer phase and a pipelooop phase.

A third objective of this research was to try to determine how well this data could be correlated to predict pressure drops from either viscometer or pipelooop data. Methods used for other slurries and non-Newtonian fluids were to be tested for their accuracy in predicting pressure drops.

MATERIALS, EQUIPMENT AND PROCEDURES

MATERIALS

The coal used during the majority of this investigation was sub-bituminous coal which is presently being mined at Colstrip, Montana. Some lignite from Savage, Montana was also used during the initial part of the investigation.

A number of petroleum liquids were tested as possible liquid carriers. Their description and origin are given in Table I. Tap water was used as the source of water.

EQUIPMENT

For the initial stages of the investigation, a Thomas-Stormer viscometer was used to measure viscosities. A Brookfield viscometer was used in the latter stages of the research.

A Stormer viscometer basically consists of a rotor that turns inside of a test cup filled with the liquid to be tested. The rotor is powered by a driving weight-pulley system. The driving weight is filled with removable shot so that the rate of rotation can be varied. The rotor is attached to a revolution counter. The test cup sits inside of a water or oil bath to keep the temperature constant.

The viscosity is determined by timing how long it takes for the rotor to make 100 revolutions with a stopwatch. This time is then compared to the time it takes the rotor to make 100 revolutions in a fluid with a known viscosity.

TABLE I. SLURRY CARRIER MATERIALS TESTED

<u>FLUID</u>	<u>ORIGIN</u>	<u>COMMENTS</u>
1,2,3,4 Tetrahydronaphthalene	Aldrich Chemical Company Milwaukee, Wisconsin	Technical Grade
Crude Oil	Amoco Oil Company Mandan, North Dakota	
#4 Cut Anthracene Oil	Reilly Tar and Chemical Corp. Cleveland, Ohio	
Dubb Unit Residual Tar	Union Oil Company Cut Bank, Montana	
Para-xylene		Technical Grade
Medium Aromatic Cut	Exxon Oil Company Billings, Montana	
Aromatic Concentrate	Exxon Oil Company Houston, Texas	
FMC Synthetic Crude Oil	FMC Corporation Princeton, New Jersey	Produced from Utah A-seam coal

The Brookfield viscometer consists of a cone that rotates inside of the test cup. The cone is powered by an electric motor. The rate of rotation is adjustable. The viscometer has a scale that measures the amount of torque exerted on the cone as it rotates in the test material. The viscosity can then be measured by comparing the torque exerted on the cone in the test material with that of a known viscosity. The test cup has a liquid circulating stage attached to a temperature controller to keep the temperature constant.

For the first phase of the investigation, 11 inch sections of two inch schedule 40 pipe were capped on both ends. The pipe was constructed with a valve on one of the caps so that it could be pressurized. The pipe sections were capable of withstanding pressures up to about 1000 psig. These pipe sections were then placed on rubber rollers powered by a motor so that they could be continually agitated while they were under pressure. The roller system and pipe sections are shown in Figure 1.

A drier was also constructed to dry some of the coal samples under a nitrogen atmosphere. It consisted of a steam-jacketed piece of two-inch pipe 23 inches in length placed vertically on a stand. The pipe was capped at each end. Each cap had a valve allowing nitrogen to be blown through. A piece of filter paper was sandwiched between two pieces of screen at the bottom end of the pipe to keep the coal from falling through. Ground coal could thus be placed in the pipe

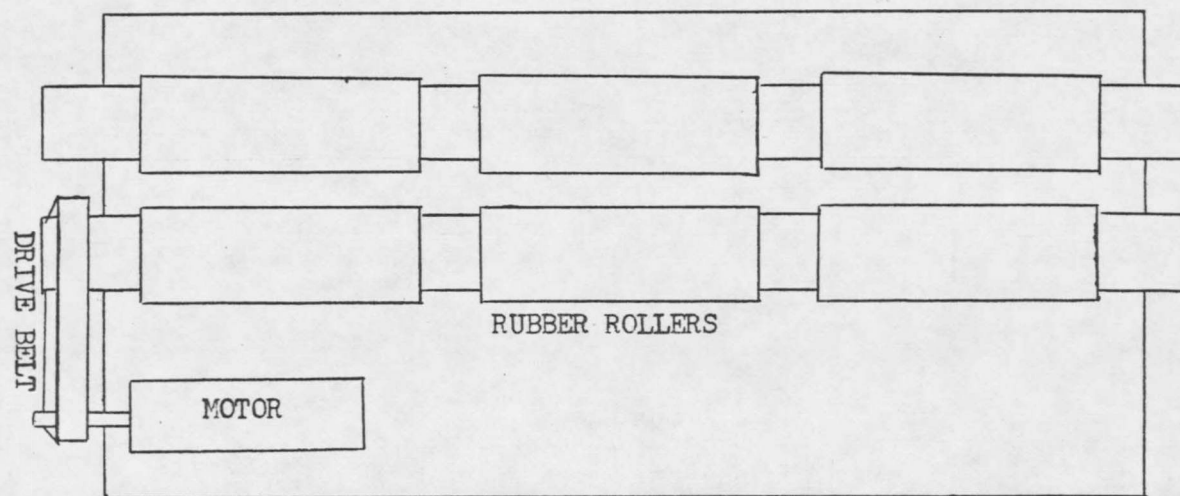
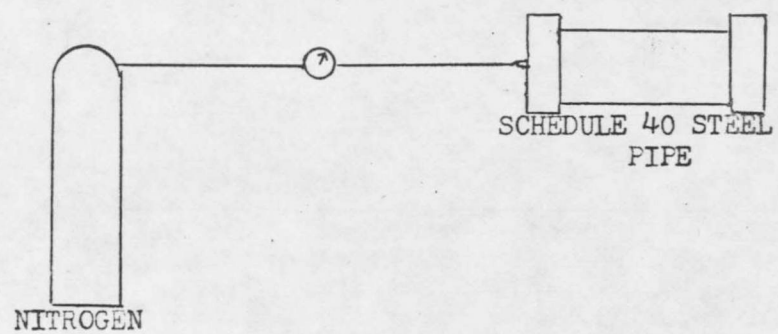


FIGURE 1. The Pipe Sections and Roller Assembly

and nitrogen could be blown through from a standard gas cylinder.

The pipeloop consisted of about 90 feet of one-inch schedule 40 steel pipe. A one horsepower centrifugal pump was used to pump the slurry. Flow rates could be varied by a by-pass valve system. The test section consisted of a horizontal section 30 feet long with standard manometer taps. A mercury filled manometer measured the pressure drops. A bleed valve allowed the lines from the manometer to be bled periodically. Flow rates were determined by measuring the mass velocity of the slurry with a bucket and stopwatch. The loop emptied into a cylinder from which the slurry was once again pumped into the loop. The cylinder was constructed so that it could be covered for long runs. The pipeloop is shown in Figure 2.

A mortar and pestle was used to grind the coal to minus 200 and minus 325 mesh. For the larger amounts of coal required for the pipeloop, a ball mill was used to crush the coal. The coal was analyzed with a standard Tyler screen using a Ro-Tap testing sieve shaker.

PROCEDURE

For the initial phase of the investigation, a slurry of coal or lignite and the chosen liquid was carefully prepared. The coal was ground by a combination of mortar and pestle and ball mill to the desired mesh size. The viscosity of the slurry was measured with a Stormer

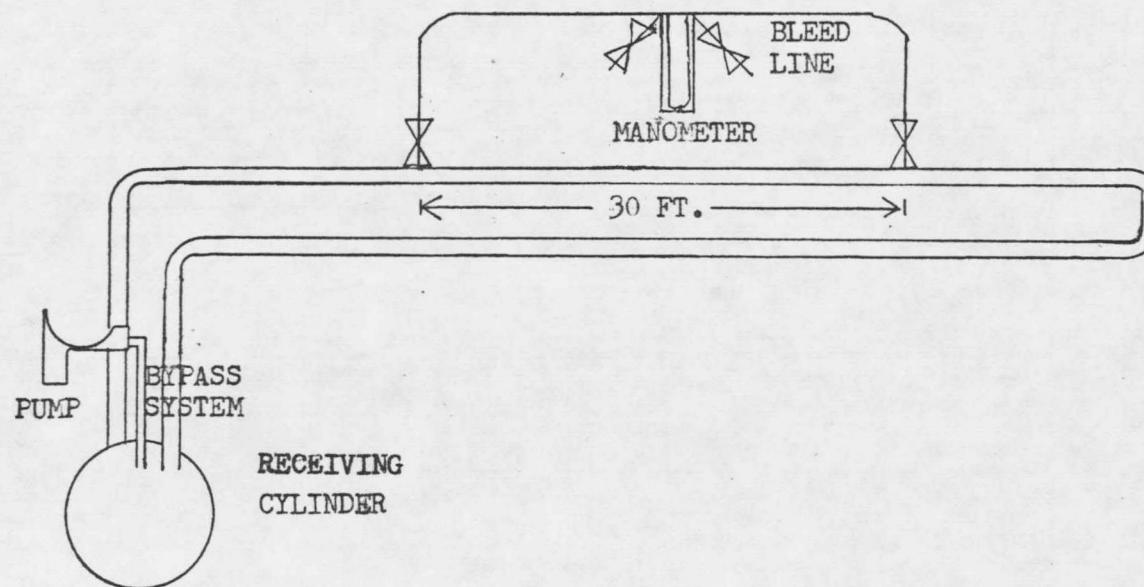


FIGURE 2. The Pipeloo

viscometer following the operating instructions. Care was taken to be certain the slurry was well mixed before any readings were taken. A minimum of four readings were taken to make sure they were consistent. An average of the readings was used.

The viscometer was calibrated periodically with distilled water and glycerol. Since the Stormer viscometer is powered by a driving weight with removable shot, it was important that the calibration and test runs were made under identical conditions. Also, as far as it was possible, to increase the accuracy of the readings, shot was removed if the time for 100 revolutions was less than 20 seconds and added if the movement of the rotor was very slow.

The slurry was then placed in the pipe sections and pressurized with nitrogen. The pipe sections were placed on the roller system which was located on the roof of Ryon Laboratory and rolled for a minimum of 20 days. These runs were made during the winter months and it was felt that in this way any adverse effect of cold temperature on the slurry could be determined.

Where indicated, some of the coal and lignite slurries were dried before the slurries were made up. This was done in the steam jacketed pipe drier. The drier was connected to a nitrogen cylinder and nitrogen was blown slowly through the coal while it was drying. The coal samples were dried for at least 48 hours.

The second phase of the investigation consisted of a more detailed

look at an Amoco crude oil - coal slurry. First the coal was placed in a large ball mill which was allowed to run for 24 hours. The ball mill was emptied and the coal was screened with a 10 mesh screen to separate the coal from the ceramic balls and larger particles. A sample of the coal was then taken and analyzed with standard Tyler screen and a Ro-Tap testing sieve shaker.

The pipe loop was filled with pure crude oil. The weight of the oil necessary to fill the line was noted. Enough ground coal was then weighed and added to the loop to give the desired weight percent slurry. To test the next higher concentration of coal in the slurry, additional coal was merely added to the existing slurry. This was done because of the large amount of slurry that would have been required to refill the loop each time. The concentration was determined by the following formula:

$$C_w = \left(\frac{x}{w + x} \right)$$

where,

C_w = mass fraction of coal

w = known weight of oil in the loop

x = weight of coal that needed to be added

After the addition of coal, the pump was turned on and the by-pass valve was closed so that the slurry was being pumped at its fastest flow rate. The slurry was allowed to pump for approximately

one hour to insure complete mixing of the coal. At this time, the valves to the manometer taps were opened. The bleed valves were also opened to make sure the line from the manometer taps to the manometer were free of air and were not clogged. The bleed valves were then closed and the valves to the manometer were opened. The pressure drop was then measured.

Flow rates were measured by running the slurry into a bucket. The time required to fill the bucket was measured with a stopwatch. The weight of the bucket plus the slurry minus the weight of the bucket was divided by the time required to get the mass flow rate. This process was repeated at least two times at each flow rate to assure consistency. The flow rates were averaged.

The by-pass valve was then opened fully so the slurry would be at its slowest flow rate. The slurry was allowed to pump for approximately 20 minutes so that its flow pattern could become fully developed before the readings for this concentration were taken.

This procedure was then repeated for the other flow rates recorded for each slurry. For a majority of the slurries, data was taken at five different flow rates.

A measured sample of the slurry was taken from the pipe at the completion of each run to determine its density.

Rheological data was also taken for each concentration using a Brookfield viscometer by using the manufacturer's operating instructions.

Best results were obtained when the slurries were made up fresh. Where possible, data was taken at shear rates of 37.5, 75, 150, 300, 750 and 1500 seconds⁻¹. At least four readings were taken at each shear rate and were averaged. The slurry was kept at a constant temperature of 20°C.

RESULTS AND DISCUSSION

PRELIMINARY PHASE

In this phase of the investigation, all viscosity data was taken by a Stormer viscometer. The Stormer viscometer is well suited to handle slurries because of its relatively large gap (0.18 cm) between the rotor and stationary baffled cup. The perforated hollow rotor also causes some pumping action to help keep the slurry well mixed. Its main drawback is that it is difficult to adjust to a given shear rate.

The main concern in this portion of the investigation was that the coal would absorb so much of the slurring liquid that it would set up like a gel. If this happened, there would be no sense in further consideration of this system. In this manner it was hoped that any unacceptable systems could be quickly eliminated.

It was felt that there were several factors in considering which liquid hydrocarbon should be tested. One was its viscosity. As the viscosity of the liquid increases, the power requirements to pump it would also increase. At the same time it was felt that the heavier the liquid was, the less the probability of the coal settling or stratification on standing or falling out in the line would be. The proximity of the liquid to coal was also considered important. It was also felt that the hydrocarbon should be a petroleum liquid so that it could be burned along with the coal.

The stability of the slurry also would depend on the particle size of the coal to a large extent. Smaller particle size would result in

a more stable slurry. Also since the grinding step was to be eliminated at the end of the line, the coal should be ground to a size that could easily be burned.

A slurry was also made up with distilled water to act as a reference.

Table II gives a summary of the viscosity changes.

There are several interesting points about Table II. The first one is that almost all of the slurries tested showed a definite increase in viscosity after pressurization. The most "reactive" combination was Amoco crude oil and coal. The runs made with dried coal and lignite were made because it was felt that this would probably be the system that would most likely absorb the fluid. However, this did not seem to make a great deal of difference. Increasing the pressure from 500 to 1000 psig also did not seem to have much of an effect.

The anthracene oil slurries were rather interesting in that they seemed to exhibit the greatest temperature affects. When it was removed from the pressurized pipe sections it was extremely thick. As it warmed it would begin to bubble noticeably and the viscosity would decrease as expected. No other slurry acted this way.

Further investigation of anthracene oil slurries was conducted by pressurizing pipe sections containing pure anthracene and a slurry of 300 milliliters of anthracene oil with 100 grams of -325 mesh coal to 500 psig. The pipes were placed in a freezer at a temperature of

TABLE II.

VISCOSITY CHANGES AT 70°F

Slurry	Pressure PSIG	Days Rolled	Original Apparent Viscosity lb/ftsec	New Apparent Viscosity lb/ftsec	% Increase
300 ml tetrahydronaphthalene 100 grams -200 mesh coal	500	25	.00097	.00100	3.1
300 ml tetrahydronaphthalene 200 grams -200 mesh coal	500	23	.0019	.0022	15.8
300 ml Amoco crude oil 200 grams -200 mesh coal	500	20	.20	.37	85
300 ml Amoco crude oil 250 grams -200 mesh coal	500	20	.54	.97	80
600 ml #4 cut anthracene oil 200 grams 325 mesh coal	500	20	2.71	2.73	.74
300 ml distilled water 100 grams -200 mesh coal	500	81	.00056	.00062	10.7
50 ml Dubbs unit residual tar 150 ml Tetrahydronaphthalene 200 grams -325 mesh coal	500	21	.59	.67	13.5
300 ml p-xylene 200 grams -325 mesh coal	500	21	.0014	.0013	-7.1
300 ml Amoco crude 300 grams -200 mesh dried coal	1000		2.08	3.77	81.3
600 ml medium Aromatic cut (Exxon) 500 grams -200 mesh dried lignite	500	13	.300	.290 (75°F)	-3.3

TABLE II (cont).

VICOSITY CHANGES AT 70°F

Slurry	Pressure PSIG	Days Rolled	Original Apparent Viscosity lb/ftsec	New Apparent Viscosity lb/ftsec	% Increase
300 ml Amoco crude 200 grams -200 mesh dried coal	500	15	.17	.30	76.5
600 ml synthetic crude (FMC) 500 grams - 200 mesh dried lignite	500	12	.30	.49	63.3
600 ml synthetic crude (FMC) 500 grams -200 mesh dried lignite	1000	26	.30	.50	66.7
600 ml FMC synthetic crude 500 grams dried -200 mesh coal	500	25	.22	.26	18.2
600 ml FMC synthetic crude 500 grams dried -220 mesh coal	1000	28	.22	.29	31.8
600 ml medium Aromatic cut (Exxon) 500 grams -200 mesh dried lignite	1000	13	.31	.30 (75°F)	-3.2
300 ml Amoco crude 250 grams -200 mesh dried coal	1000	21	.54	.93	72.2

-13°F for 24 hours. Both the slurry and pure oil were frozen when the pipes were unloaded. They both had pour points of about 41°F. Bubbling was again noticeable and it seemed to increase until the temperature reached about 59°F. Both the pure oil and slurry returned to their original viscosities at 70°F.

The temperature effect on the viscosity of the other slurries was not examined. None of the slurries exhibited such a dramatic temperature dependence. It was felt that temperature should be held constant for this investigation. This is something that should be studied in future investigations.

Another interesting feature of both the Amoco crude oil and FMC crude oil slurries was that they would set up noticeably when left exposed to air. For instance, a slurry of 250 grams of dried -200 mesh coal and 300 milliliters of Amoco crude oil left uncovered for one week increased in viscosity from 1.93 lb/ftsec to 4.80 lb/ftsec. It was felt that this was due to evaporation of the oil. Crude oil sand slurries exhibited the same trend. Also 90 milliliters oil was placed in a graduated cylinder and let stand open for 72 hours. It was found that the liquid level dropped to 87.5 milliliters in this time.

Additional effects of exposure to air for FMC synthetic crude oil slurries are shown in Figures 18 and 19 in the appendix.

It was felt that although the viscosities of many of the slurries did increase under pressurization, this did not happen to a great

enough degree to cause any of the liquids to be eliminated from further consideration. For this reason, it was decided to end this phase of investigation and to get more detailed rheological and pipelooop data.

RHEOLOGICAL PHASE

Although the Amoco crude oil slurries seemed to have the greatest tendency to set up, it was felt to be the slurring material most likely to be used in a commercial operation. The crude came from the Williston Basin which is in fairly close proximity to the coal fields of Eastern Montana. Also, it is readily available and is not tied to a plant or refinery's capacity to produce it. For these reasons, Amoco crude oil - coal slurries were chose for more detailed studies.

A Wells-Brookfield Syncho-Lectric Viscometer was used to take the viscosity readings for this phase of the investigation. The rate of shear could be controlled much easier and more accurately than it could be with the Stormer viscometer.

The viscometer was calibrated with a standard oil. All readings were taken at 20°C. Care was taken that the slurries were well mixed before they were placed in the test cup.

Slurries of -250 mesh coal in Amoco crude oil were tested at weight concentrations of coal from 0% to 50% in 5% increments.

Table III shows how the apparent viscosity of the slurries varied with the shear rate.

TABLE III. APPARENT VISCOSITIES OF CRUDE OIL-COAL SLURRIES

Weight % Coal	Rate of Shear sec ⁻¹	Apparent Viscosity
		$\frac{\text{lb}}{\text{ft. sec.}}$
0	37.5	.00607
	75	.00476
	150	.00480
	300	.00448
	750	.00418
	1500	.00418
5	37.5	.0133
	75	.00921
	150	.00780
	300	.00712
	750	.00612
	1500	.00517
10	37.5	.0179
	75	.0110
	150	.0109
	300	.0103
	750	.00932
	1500	.00663
15	37.5	.0234
	75	.0150
	150	.0120
	300	.0107
	750	.00935
	1500	.00824
20	37.5	.0398
	75	.0257
	150	.0193
	300	.0148
	750	.0117
25	37.5	.0548
	75	.0292
	150	.0230
	300	.0170
	750	.0136

TABLE III (cont). APPARENT VISCOSITIES OF CRUDE OIL-COAL SLURRIES

Weight % Coal	Rate of Shear sec ⁻¹	Apparent Viscosity ^{lb} ft.sec.
30	37.5	.0537
	75	.0334
	150	.0254
	300	.0206
	750	.0155
35	37.5	.0670
	75	.0336
	150	.0278
	300	.0222
	750	.0171
40	37.5	.0672
	75	.0336
	150	.0284
	300	.0270
45	37.5	.138
	75	.0751
	150	.0618
	300	.0409
50	15	.137
	37.5	.143
	75	.0995
	150	.0832

The readings are not recorded at the higher shear rates for the more concentrated slurries because the viscometer went off scale. The readings at the lower shear rates are not as accurate as those of the higher shear rates because the rheological data at lower rates had more scatter. It was very difficult to get accurate readings below

37.5 seconds⁻¹.

Figures 3 and 4 show the rheograms calculated from the Brookfield viscometer data. Figure 5 is a typical rheogram showing five different classes of fluids.

It can be seen from these rheograms that the pure crude oil appears to closely resemble a Newtonian fluid. Concentrations up to 10% most closely resemble a pseudoplastic. Concentrations from 10% to 40% appear to be real plastics. Above 40% the slurries appear to approximate a Bingham plastic.

PIPELOOP DATA

The pipelooop was located in Ryon Laboratory. All runs were made at room temperature. Special care was taken to keep the receiving cylinder covered when the pipelooop was not in use and during long runs because of crude oil's apparent tendency to evaporate.

Pipelooop data was taken for slurries from 0% to 50% by weight in 5% increments. However, at a 45% slurry, the manometer lines began showing a tendency to plug up. The lines were cleared out before each run but the data still seemed to be erratic. This did not seem to be a problem at concentrations below 45%. For this reason, data is only presented for concentrations up to 40%.

Table IV gives the average size distribution of the coal particles as analyzed with standard Tyler screens and a Ro-Tap testing sieve

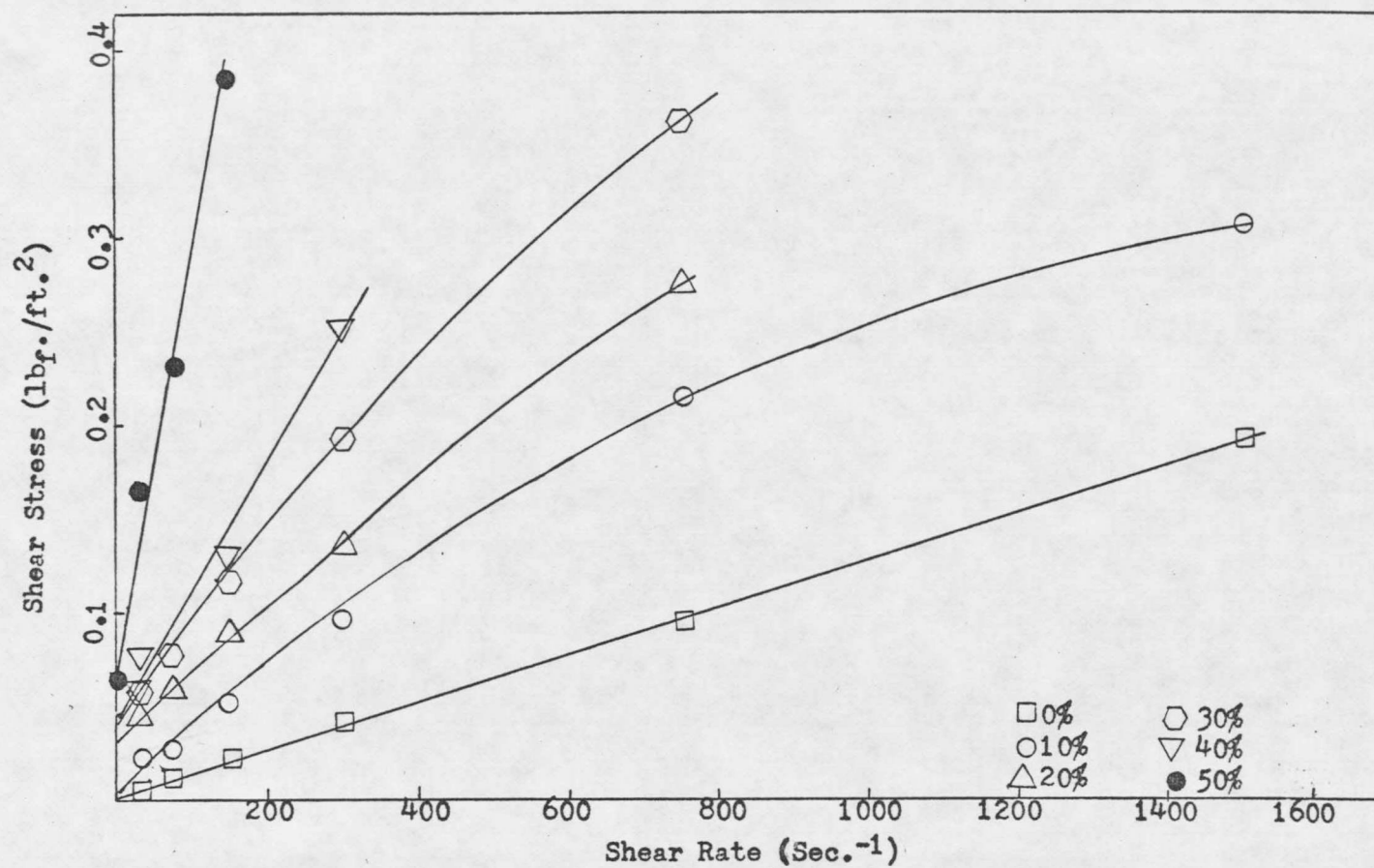


FIGURE 3. Rheogram from Brookfield Viscometer

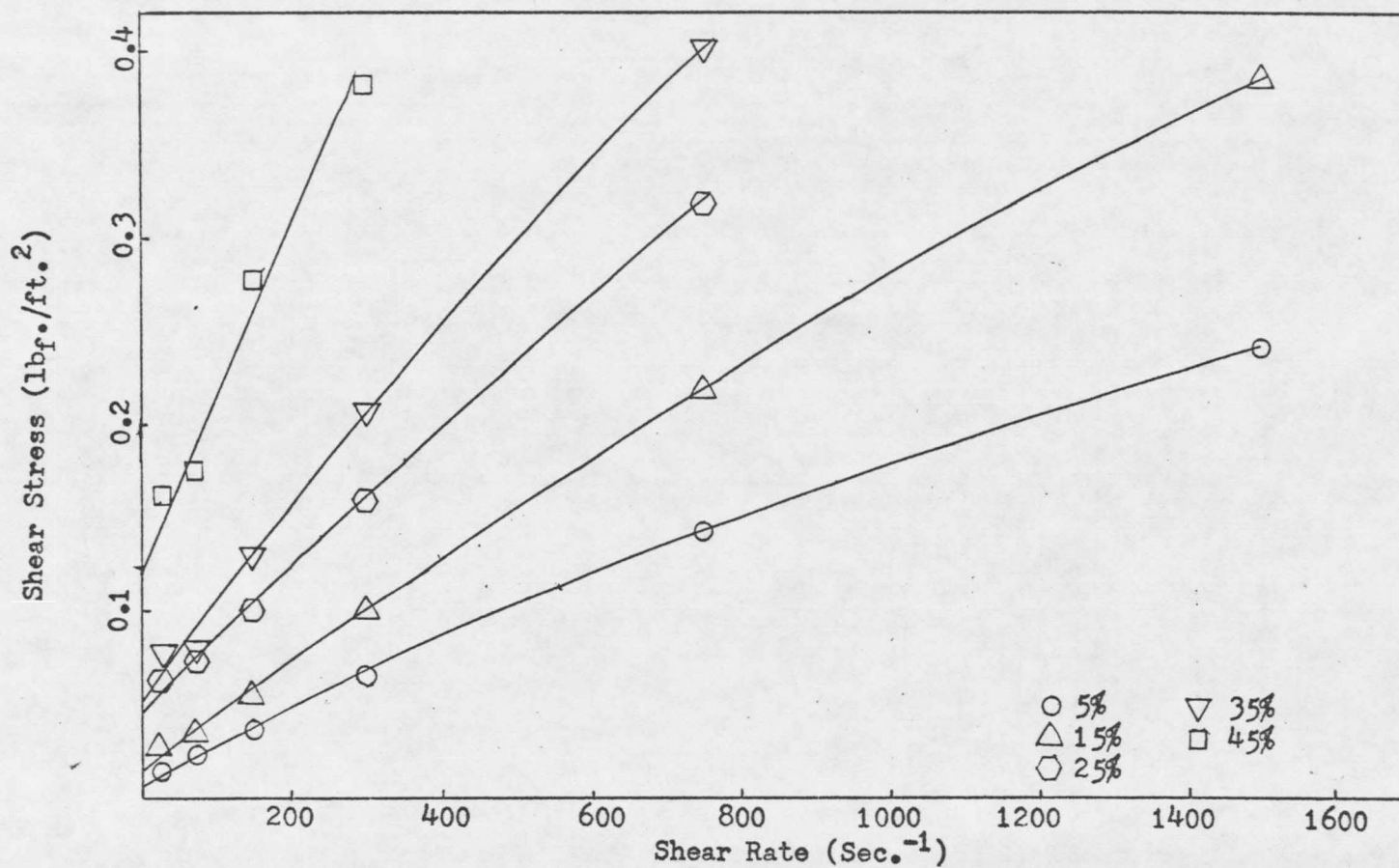


FIGURE 4. Rheogram from Brookfield Viscometer

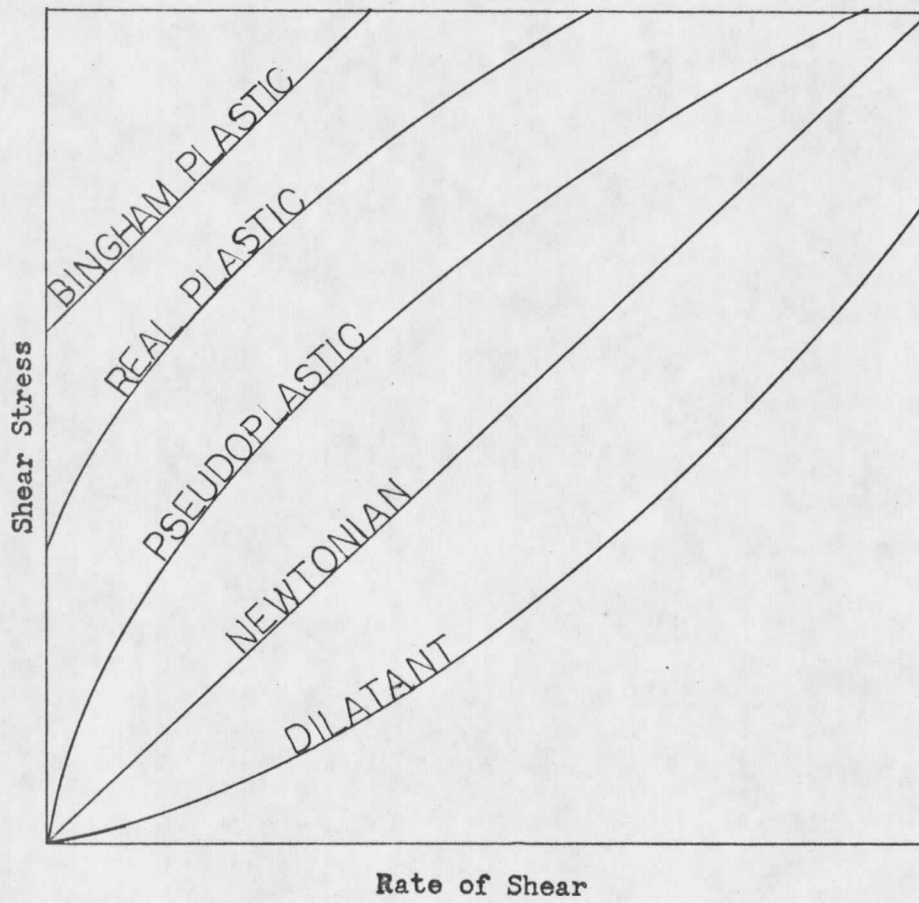


FIGURE 5. Types of Fluids

shaker.

TABLE IV. COAL PARTICLE SIZE DISTRIBUTION

<u>Screen Size</u>	<u>Weight % Coal</u>
-10 to +150 mesh	1.2
-150 to +200 mesh	5.4
-200 to +325 mesh	67.8
-325 mesh	25.5

Figures 6 and 7 show how the pressure drop varied with velocity.

The pressure drops, especially at the lower flow rates, seem to remain fairly constant with changes in concentration up to 30% by weight. At a weight percent of 30 percent there seems to be a definite break in the graphs. This indicates that the viscosity of the slurry begins to increase more rapidly at a 30% concentration. Similar observations have been reported for coal-water slurries (4).

The solid line represents the best curve for the carrier material alone. It is intended to show how the slurry pressure drops deviated from the pure crude oil.

Three theoretical flow types have been presented in the literature for slurries: homogeneous slurries, heterogeneous slurries, and plug flow. (2) (3). Each has its own characteristic flow properties.

In homogeneous slurries, the solid particles are uniformly distributed in the liquid media. Those slurries are usually characterized by high solid concentration and fine particle sizes.

Heterogeneous slurries have concentration gradients along the vertical axis of the pipe. The concentration of solids is greater at

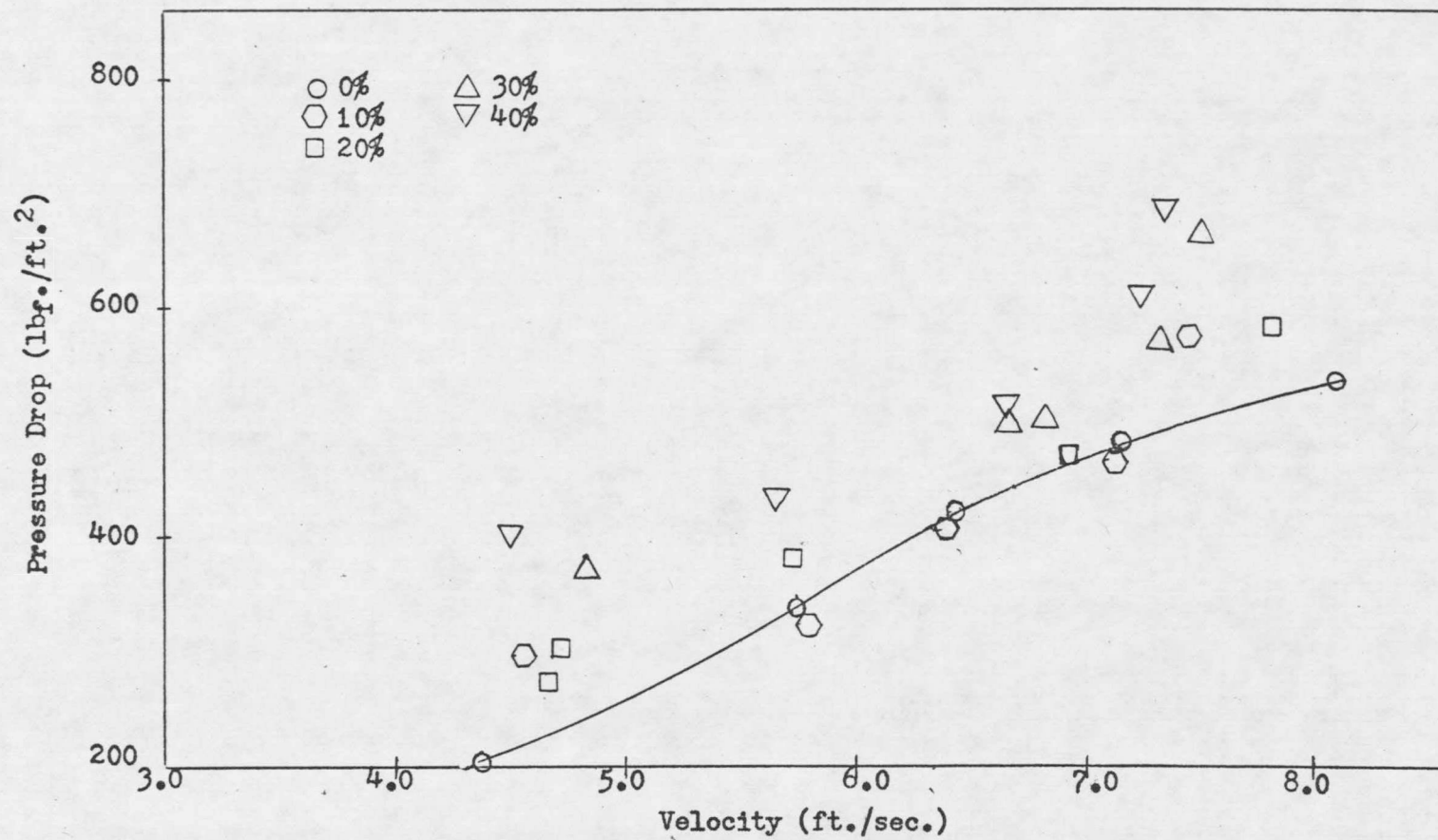


FIGURE 6. Pipeloop Pressure Drops

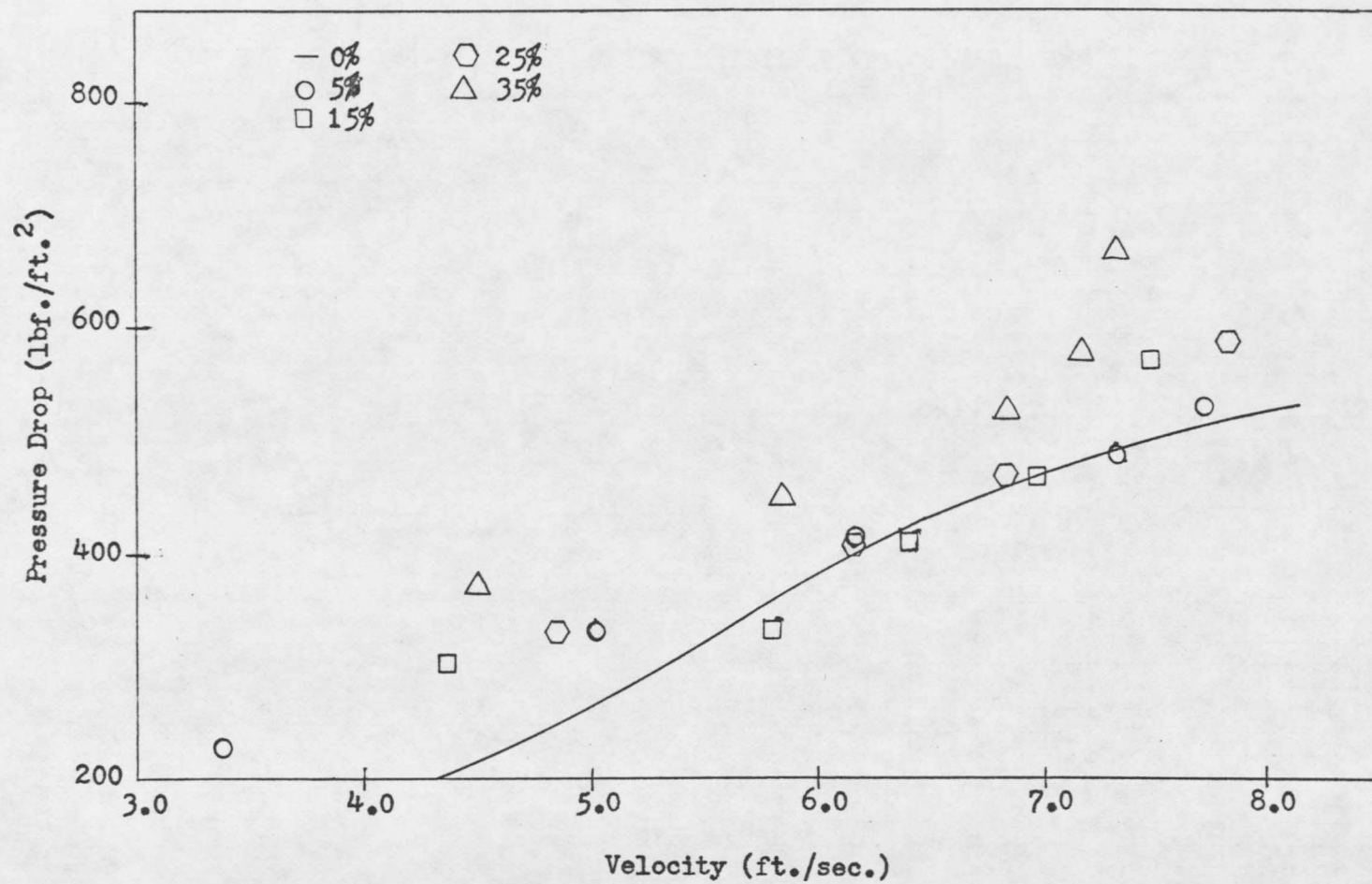


FIGURE 7. PIPELOOP PRESSURE DROPS

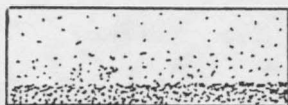
the bottom of the pipe than it is at the top. These slurries are usually of lower solids concentration and have larger particle sizes than homogeneous slurries.

Plug flow theory has been presented to explain the fact that the slurries have on occasion had a lower pressure drop than the pure liquid at the same flow rate. It states that when force is applied, the mixture is compressed and the liquid is squeezed out to the periphery of the plug. Therefore the solids tend to move along the center of the pipe as a plug and liquid serves as a lubricant between the plug and the pipe wall. This can occur in both laminar and turbulent flow. It is very interesting in that if this does indeed occur, slurries can be transported in the laminar flow region which reduces the power requirements. Much of the literature states that the slurries have to be transported in the turbulent region to keep the solid particles in suspension.

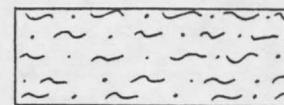
Plug flow has also been postulated to occur in the turbulent flow region. In this case, the solid plug also tends to suppress the turbulence in the middle core. The middle core moves either as a solid plug or shears laminae in a telescopic manner. The outer annulus is in turbulent shear. This accounts for pressure drops for slurries that are lower than for the pure liquid.

The theoretical flow patterns are shown in Figure 8.

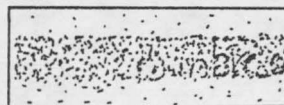
In actual application, the slurries will probably be of a mixed



HETEROGENEOUS SLURRY



HOMOGENEOUS SLURRY



PLUG FLOW

FIGURE 8. THEORETICAL FLOW PATTERNS

character. This increases the difficulty in predicting pressure drops.

Two long runs were made during the investigation to see if any settling was occurring in the line. These runs were made at concentrations of 30% and 50%. The by-pass valve was set wide open so the flow rates were the slowest attainable. The 30% slurry was run for 10 days and the 50% slurry for 11 days. At the end of each run, the pipeloop was taken apart to determine if a layer of coal had formed on the bottom. No evidence of settling was found in either case. Also, the pressure drop remained approximately constant during the 30% run. The manometer lines plugged with the 50% slurry so the pressure drop could not be measured.

The pump had to be repacked once during the investigation. Both the pump and pipe were checked for any signs of wear at the end of the investigation. No sign of excessive wear was found.

PREDICTING PRESSURE DROPS

Pipeloop data was first analyzed using the Mooney-Rabinowitsch-Weissenberg equation (see appendix). A plot of $8V/D$ versus τ_o should be unique regardless of the pipe diameter used. The rate of shear is given by

$$-\left(\frac{dv}{dr}\right)_o = \frac{3}{4} \left(-\frac{8V}{D}\right) + \frac{1}{4} \tau_o \frac{d(8V/D)}{d\tau_o}$$

where $\frac{d(8 V/D)}{d\tau_o}$ is the slope of the curve at the given value of τ_o .

$$\text{and } \tau_o = \frac{D \Delta P}{4L}.$$

In the above equations,

V = bulk velocity

D = inside diameter of the pipe

ΔP = pressure drop

L = length of pipe

τ_o = shear stress at wall

The effective viscosity is then found by

$$\mu_e = \frac{\tau_o g_c}{dv/dr}$$

where

μ_e = effective viscosity

g_c = gravitational constant

The pressure drop can be then calculated using μ_e as the viscosity in the Reynolds number calculations. This method is supposed to be good for a Reynolds number, calculated in this manner, of less than 2100. The friction factor is therefore $16/Re$ and the pressure drop can be calculated using Poiseuille's equation.

$$\Delta P = \frac{2f V^2 L \rho}{g_c D}$$

where,

ΔP = pressure drop

ρ = density of the slurry

Figures 9 through 17 are the plots of $8 V/D$ versus τ_o for slurries from 0% to 40%. Table V shows the predicted pressure drops compared to the measured.

The method seems to be only good to $\pm 25\%$. The predicted pressure drops at higher flow rates are consistently higher than the measured pressure drops. The method only gives appreciably low predictions at the lower velocities.

The change in the shape of the curve between the pure oil and slurries may indicate a transition from turbulent to laminar flow, a damping of the turbulence, or a different fluid characterization. Using the viscosity for pure oil as determined by the Brookfield viscometer the Reynolds number for the pure oil ranges from 7991 to 4316. This indicates that turbulent flow probably does exist in the pipe explaining why the effective viscosities, for pure oil, calculated from the pipeloop data are higher than those calculated from the Brookfield viscometer.

A plot of $D \Delta P/4L$ versus $8 V/D$ on logarithmic coordinates often gives a straight line. The slope of this line is given the symbol n' and is the flow-behavior index. The consistency index k' is the value of $D \Delta P/4L$ for $8 V/D = 1$.

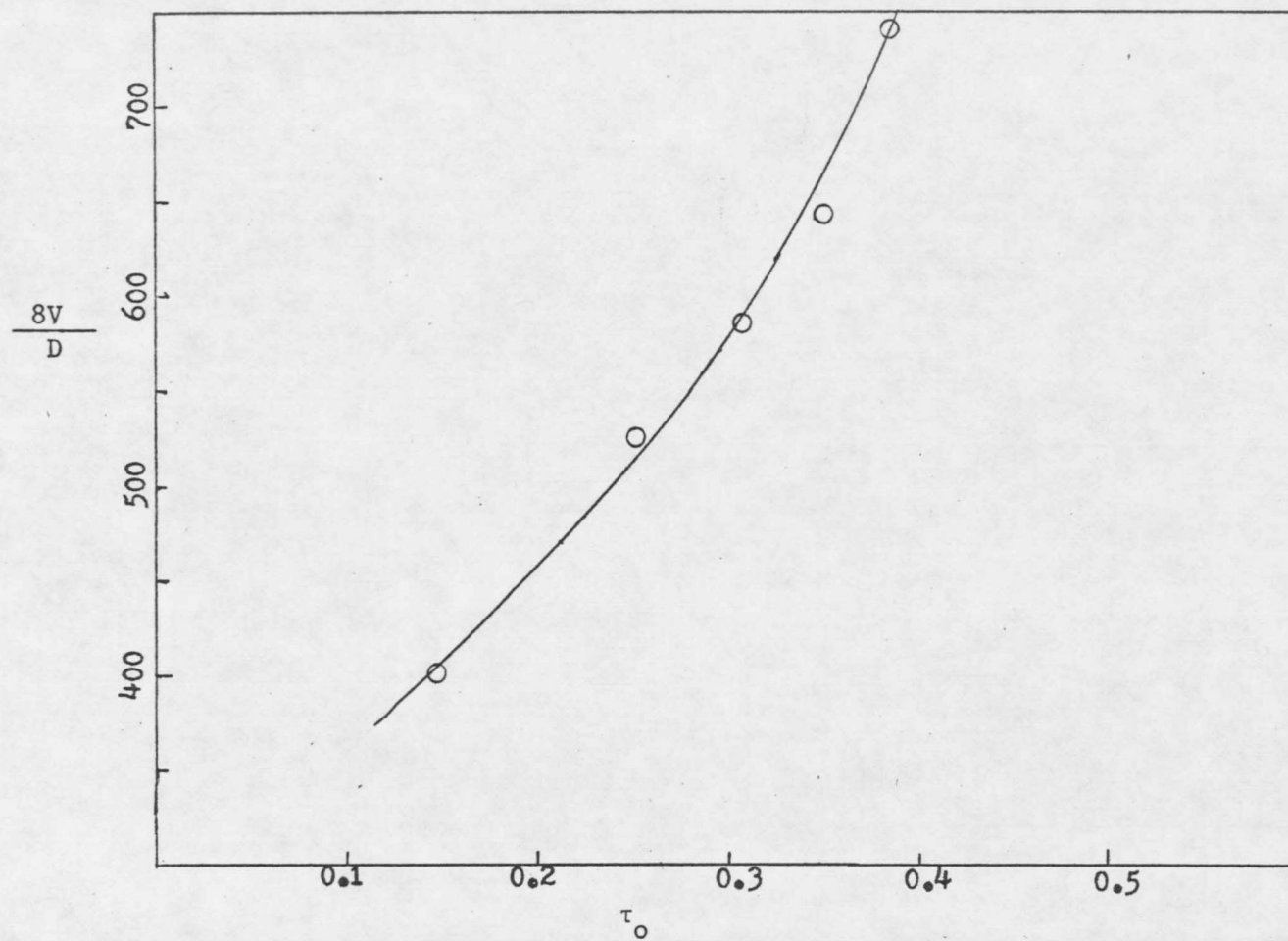


FIGURE 9. 0% Slurry

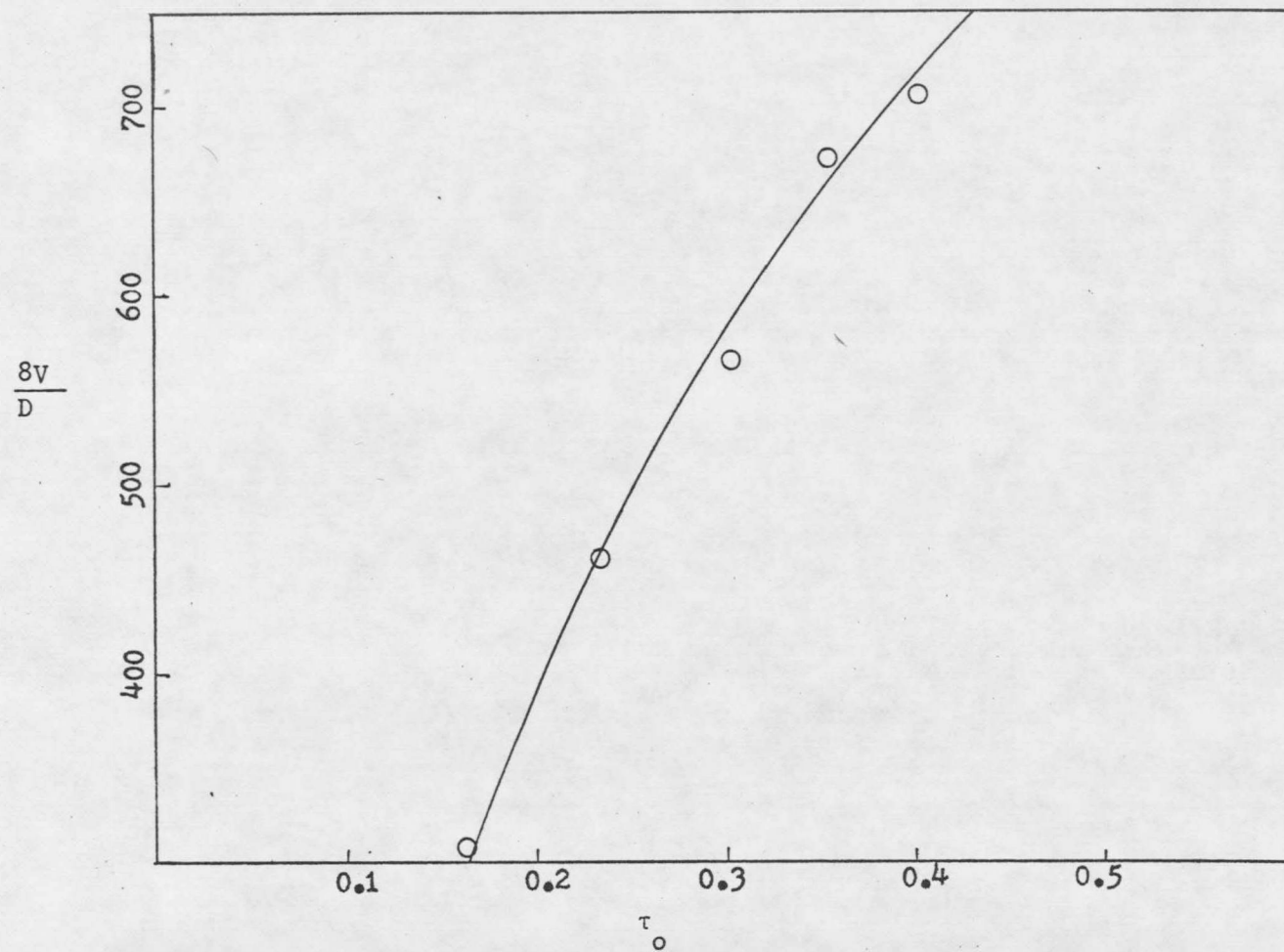


FIGURE 10. 5% Slurry

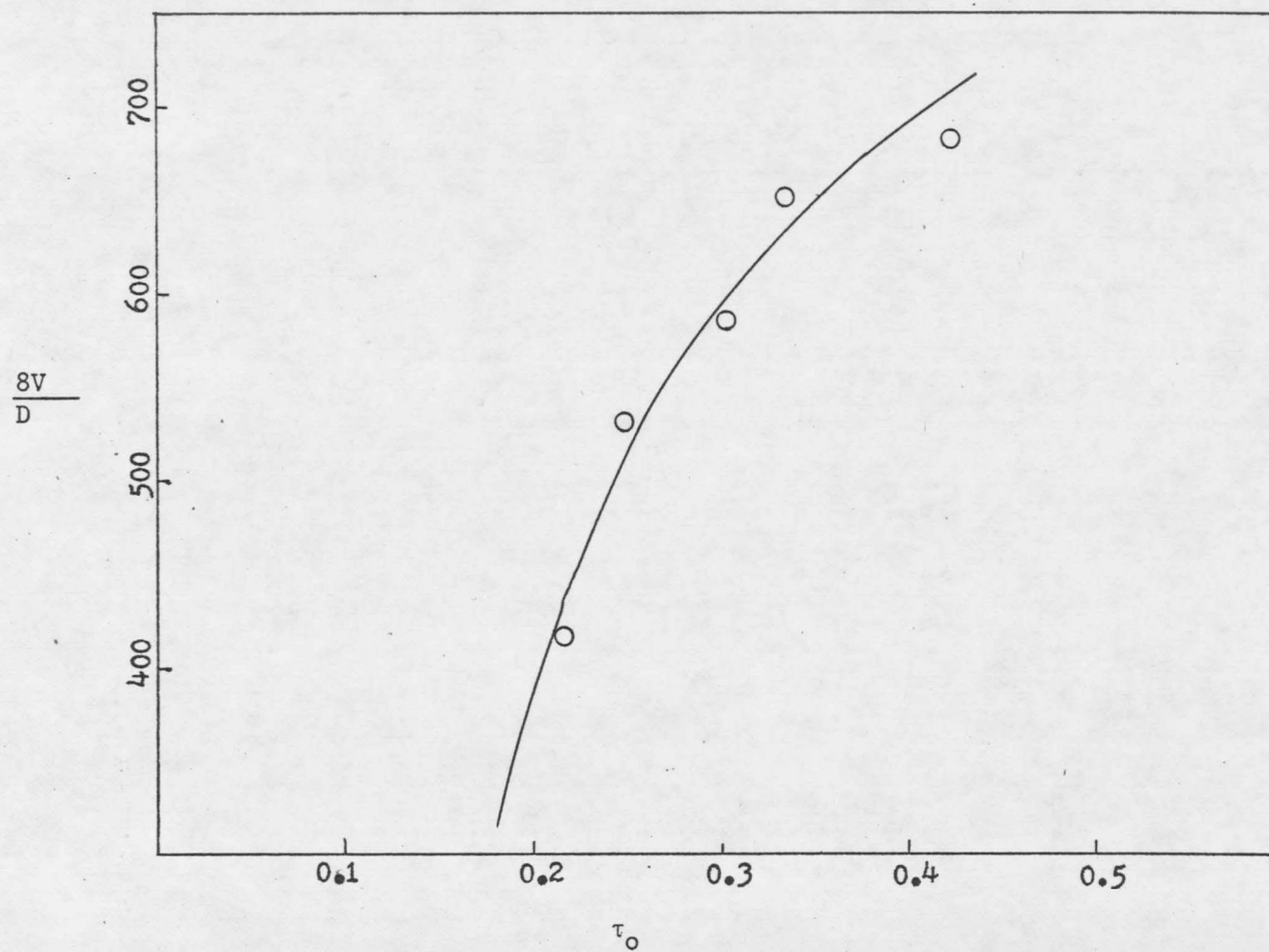


FIGURE 11. 10% Slurry

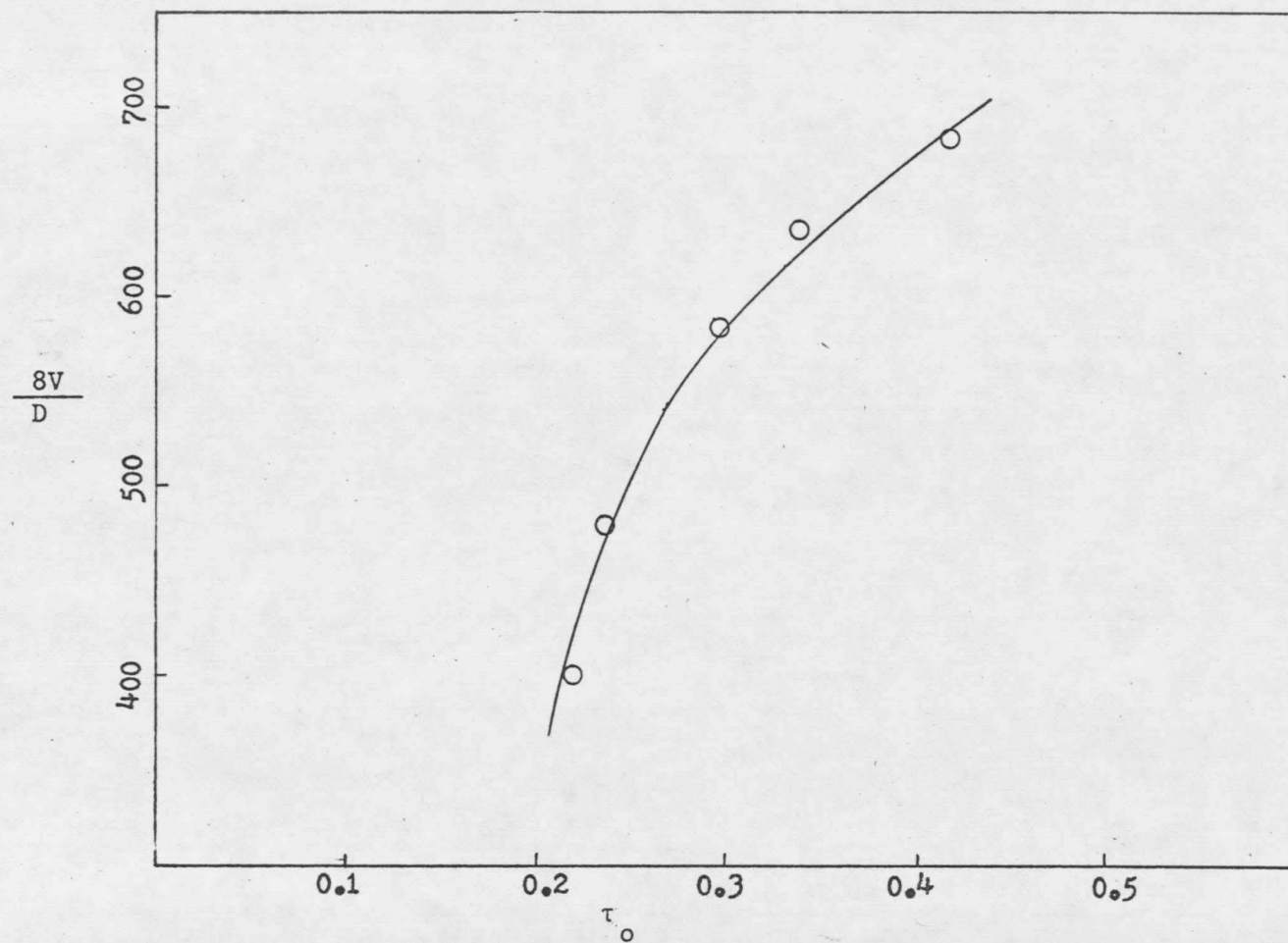


FIGURE 12, 15% Slurry

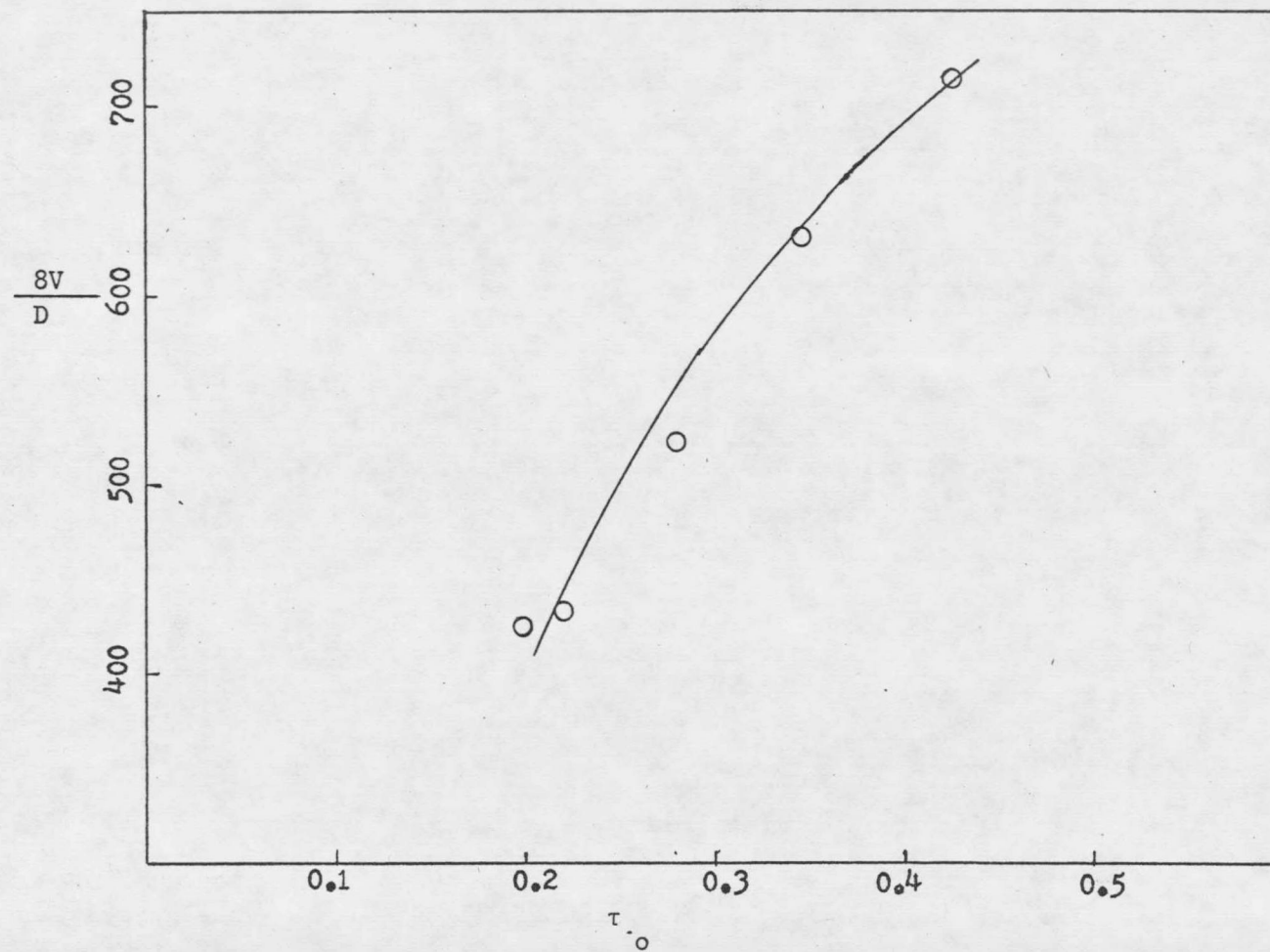


FIGURE 13. 20% Slurry

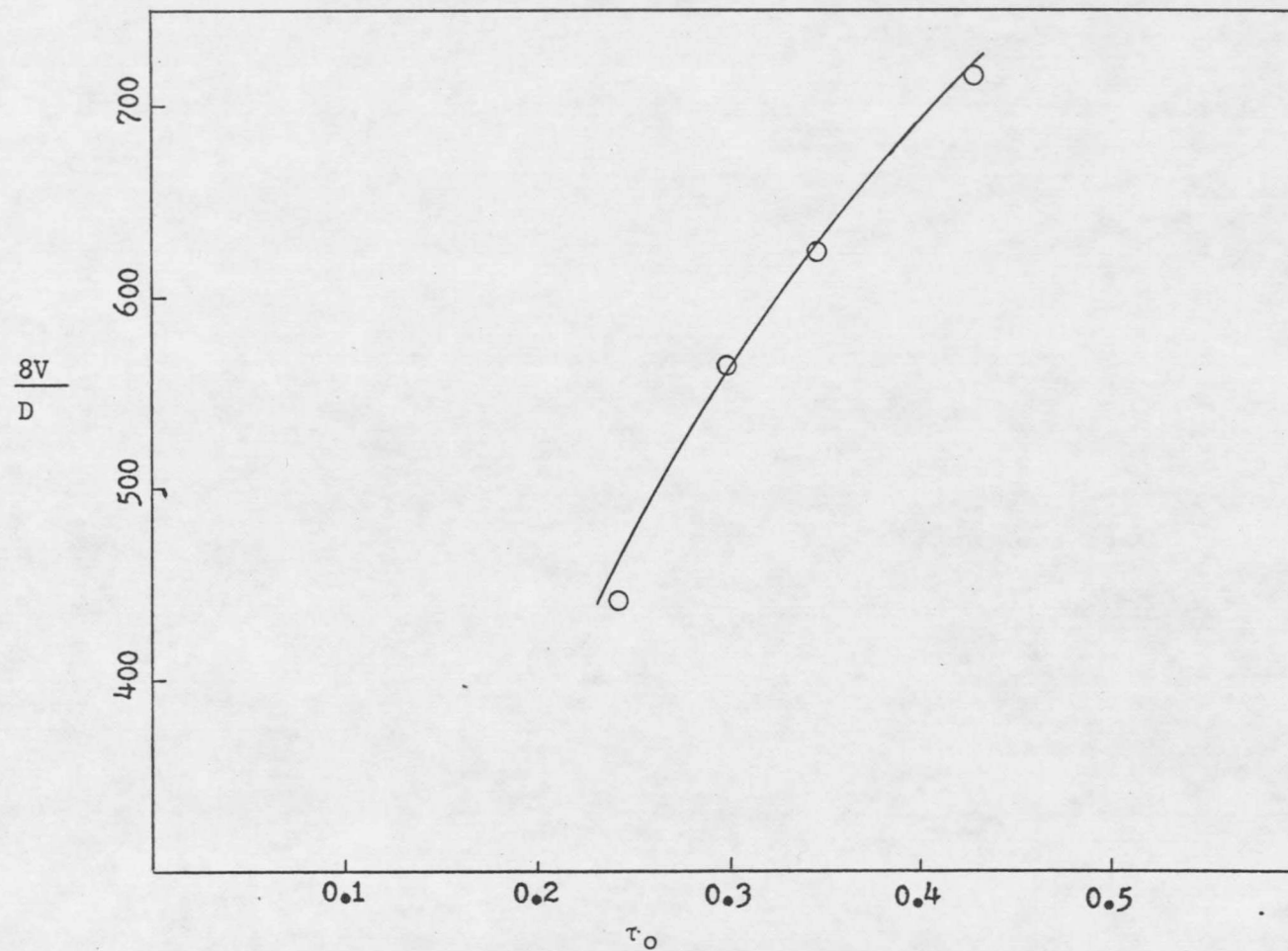


FIGURE 14. 25% Slurry

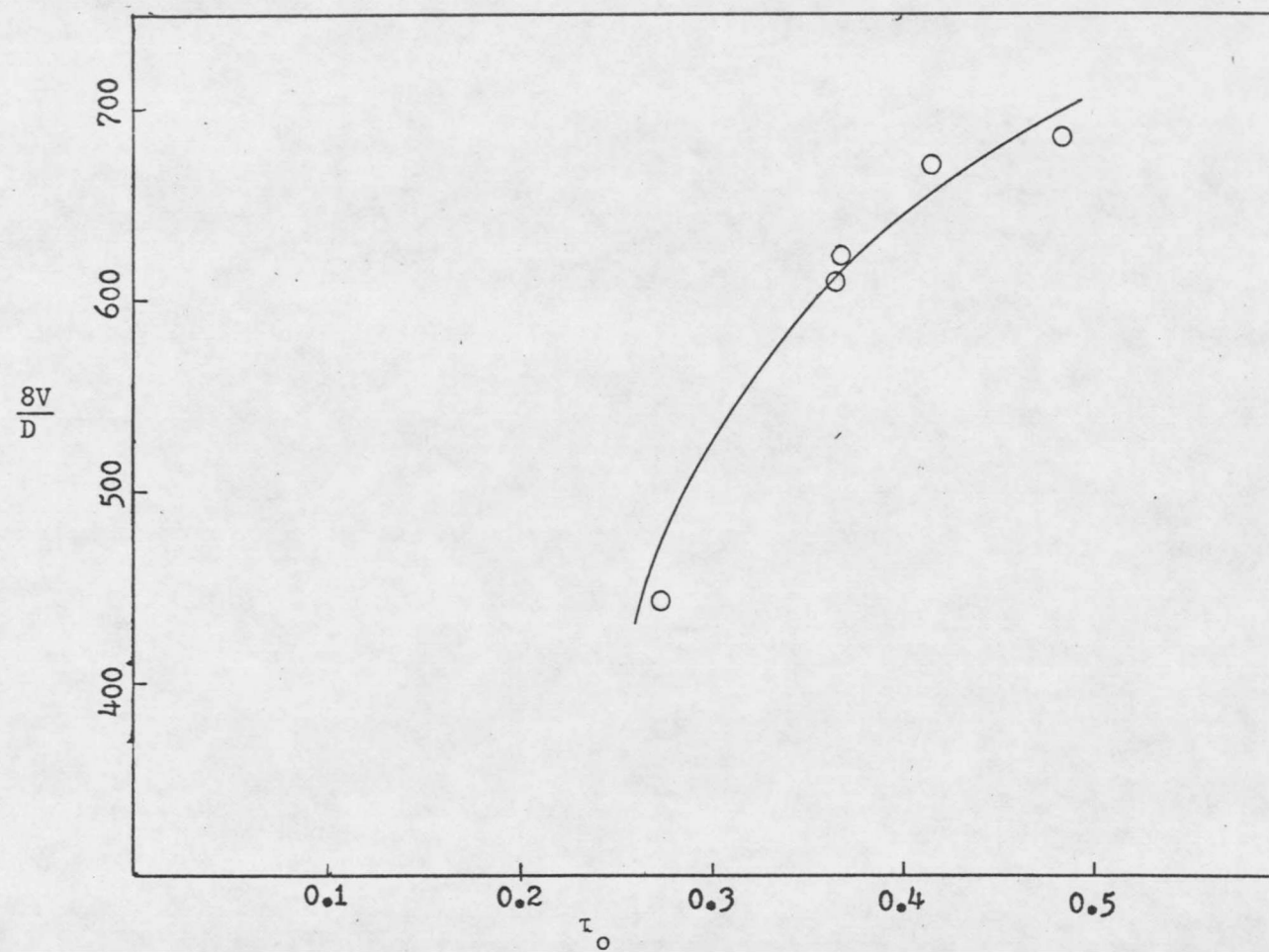


FIGURE 15. 30% Slurry

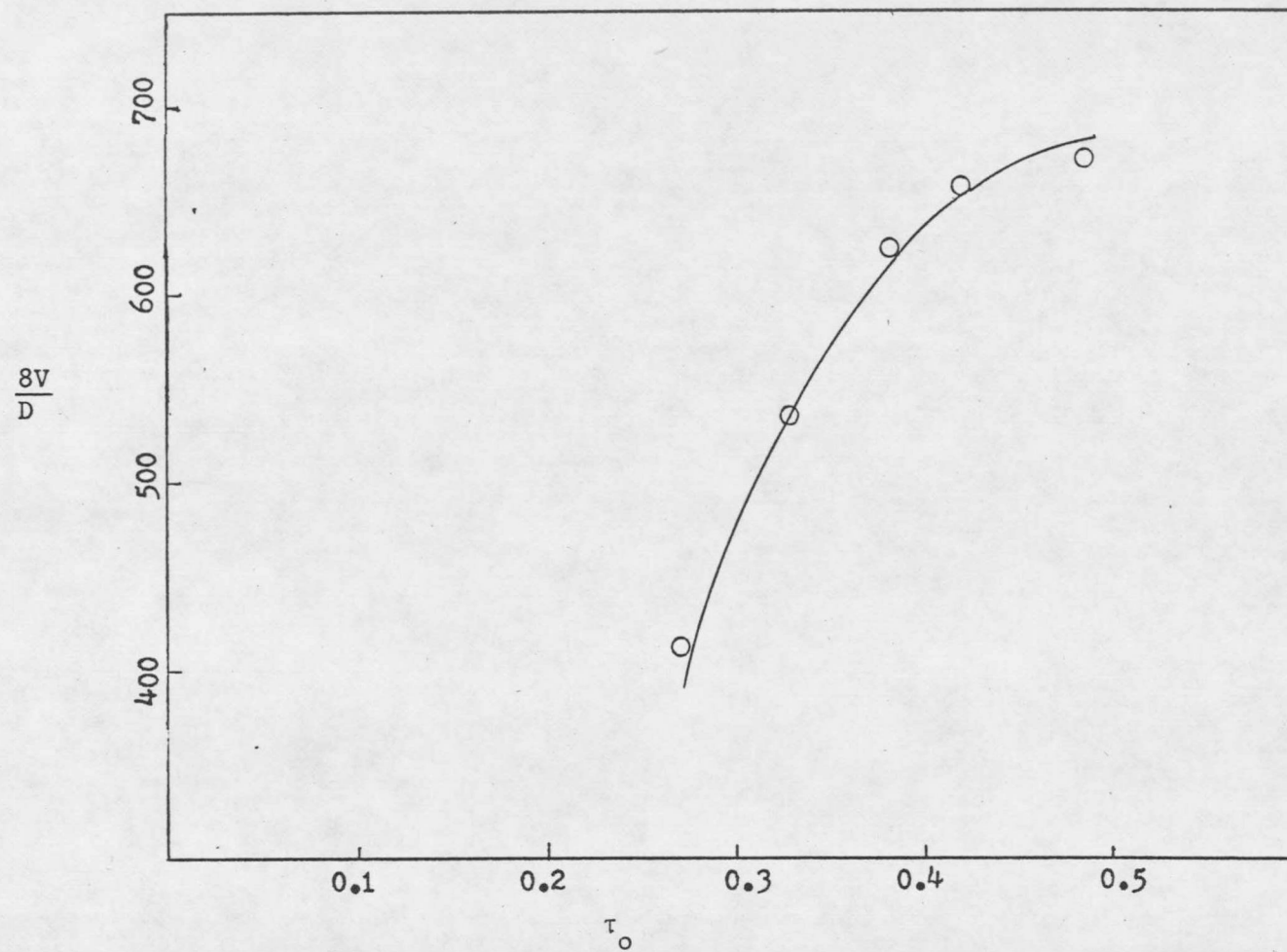


FIGURE 16. 35% Slurry

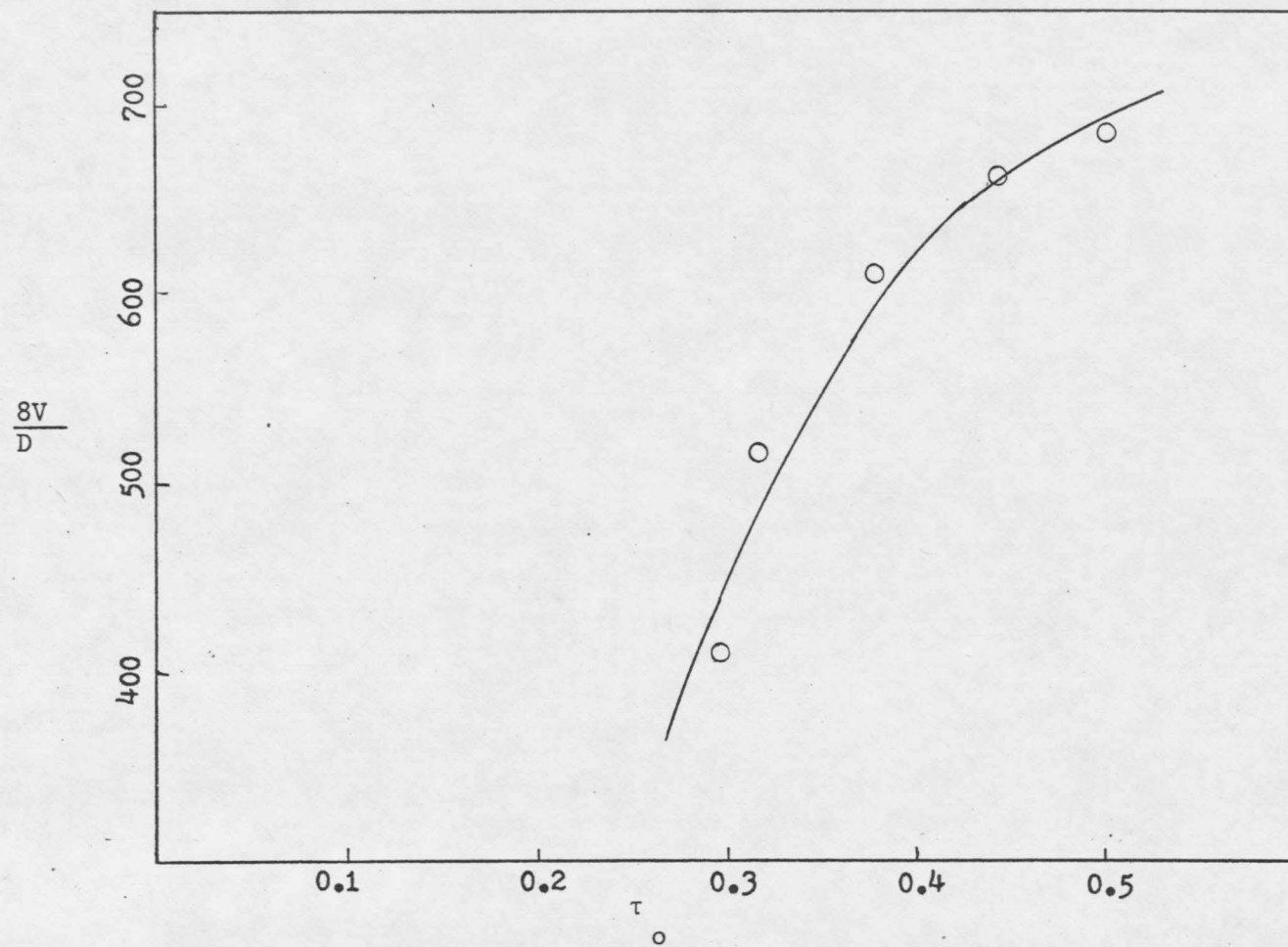


FIGURE 17. 40% Slurry

TABLE V. PRESSURE DROP PREDICTIONS USING THE MOONEY-RABINOWITSCH
WEISSENBERG EQUATION

$8V/D\text{sec}^{-1}$	$\tau_o \frac{lb_f}{ft^2}$	dv/dr	$\mu_e \frac{lb}{ft \cdot sec}$	Re	$\Delta P_p \frac{lb_f}{ft^2}$	$\Delta P_{aft} \frac{lb_f}{ft^2}$	% dev.
<u>0%</u>							
742	.385	797	.0155	2310	-	-	-
655	.345	656	.0169	1870	473	480	-1.5
587	.308	560	.0177	1599	443	423	4.7
526	.262	477	.0177	1434	397	346	14.7
401	.145	334	.0140	1381	239	204	17.1
<u>5%</u>							
706	.391	642	.0196	1797	590	550	7.3
671	.362	615	.0189	1771	541	487	11.1
566	.285	536	.0171	1651	413	416	.72
460	.230	457	.0162	1418	318	331	-3.9
309	.166	344	.0155	996	205	226	-9.3
<u>10%</u>							
682	.389	590	.0212	1634	617	578	6.7
652	.350	577	.0195	1698	542	462	17.3
586	.295	543	.0175	1700	437	414	5.56
531	.248	513	.0156	1729	353	342	3.22
419	.208	462	.0145	1469	259	296	-12.5
<u>15%</u>							
685	.42	584	.0231	1534	676	571	18.4
634	.35	557	.0202	1623	547	465	17.6
586	.302	535	.0182	1664	455	408	11.5
480	.24	484	.0160	1522	328	328	0
400	.215	510	.0136	1520	232	301	-22.9
<u>20%</u>							
717	.424	632	.0216	1735	660	582	13.4
633	.341	572	.0192	1725	519	473	9.7
523	.264	506	.0168	1626	375	384	-2.3
432	.215	433	.0160	1412	295	302	-2.3
428	.210	434	.0156	1435	285	272	4.8

TABLE V. (cont)

$8V/Dsec^{-1}$	$\tau_o \frac{lb_f}{ft^2}$	dy/dr	$\mu_e \frac{lb}{ft \ sec}$	Re	$\Delta P_p \frac{lb_f}{ft^2}$	$\Delta P_{aft} \frac{lb_f}{ft^2}$	% dev.
<u>25%</u>							
717	.421	641	.0211	1816	645	589	9.5
625	.342	580	.0190	1760	507	473	7.2
565	.298	544	.0176	1716	424	409	3.7
442	.232	452	.0165	1433	312	331	-5.7
<u>30%</u>							
686	.459	586	.0252	1489	738	664	11.1
671	.435	576	.0242	1516	693	573	20.9
625	.378	558	.0218	1568	581	506	14.8
610	.365	551	.0213	1567	555	501	10.7
442	.265	505	.0169	1430	319	374	-14.7
<u>35%</u>							
671	.455	561	.0261	1441	747	663	12.7
656	.425	566	.0242	1520	677	575	17.7
625	.392	577	.0219	1600	584	525	11.2
534	.328	554	.0190	1577	433	451	-4.0
412	.274	529	.0167	1382	294	372	-20.9
<u>40%</u>							
674	.483	561	.0277	1383	796	688	15.7
664	.453	573	.0254	1486	719	607	18.5
610	.388	582	.0214	1622	558	518	7.7
517	.333	540	.0198	1491	438	437	.22
412	.286	488	.0189	1239	332	406	-18.23

This method was tried for the slurries tested but it was found that the plot did not give a straight line. However, values for n' and k' could be obtained from the following formulas.

$$n' = \frac{8V}{\tau_o D} \frac{d\tau_o}{d(8V/D)}$$

$$k' = \tau_o / (8V/D)^{n'}$$

From these values a generalized Reynolds number could be calculated by:

$$Re = \frac{D^{n'} v^{2-n'} \rho}{v}$$

where

$$v = g_c k' 8^{n'-1}$$

Table VI gives the pressure drop predictions using the generalized Reynolds number.

This method gives predicted pressure drops that are within $\pm 10\%$ of the measured drops.

If the slurry is not homogeneous, the generalized Reynolds number criteria of 2100 as the transition from laminar to turbulent flow may not be true. This would mean that the assumption that the shear stress is just a function of shear rate may not be true. If it is assumed the slurry is in turbulent flow, Perry (5) recommends computing a turbulent viscosity from the pressure drop data and the friction-factor versus Reynolds number plot. Pressure drops computed

TABLE VI. PRESSURE DROP PREDICTIONS USING THE GENERALIZED REYNOLDS NUMBER

$8V/D$	τ_o	$n' = \frac{8V}{\tau_o D}$	$\frac{d\tau}{d(8V/D)}$	$k' = \frac{\tau_o}{(8V/D)^{n'}}$	$v = g_c k' 8^{n'-1}$	$Re = \frac{D^{n'} U^{2-n'} \rho}{v}$			
<u>WEIGHT % COAL</u>									
<u>0%</u>									
742	.385	.771		2.36×10^{-3}	4.72×10^{-2}	2140			
655	.345	.991		5.58×10^{-4}	1.76×10^{-2}	1869			
587	.308	1.23		1.21×10^{-4}	6.28×10^{-3}	1679			
526	.262	1.59		1.23×10^{-5}	1.35×10^{-3}	1591			
401	.142	3.04		1.73×10^{-9}	3.87×10^{-6}	1702			
	ΔP_p	$\frac{1b_f}{ft^2}$		ΔP_a	$\frac{1b_f}{ft^2}$	% dev.			
	530								
	473			480		-1.45			
	422			423		-0.23			
	358			346		3.47			
	194			204		-4.90			
<u>5%</u>									
706	.391	1.56	1.41×10^{-5}	1.45×10^{-3}	1977	536	550	-2.5	
671	.362	1.50	2.08×10^{-5}	1.89×10^{-3}	1934	496	487	1.8	
566	.285	1.27	9.09×10^{-5}	5.13×10^{-3}	1743	391	416	-6.0	

TABLE VI (Cont).

8V/D	τ_o	n'	k'	v	Re	$\Delta P_p \frac{lb_f}{ft^2}$	$\Delta P_a \frac{lb_f}{ft^2}$	% dev.
<u>5% (cont)</u>								
460	.230	1.03	4.16×10^{-4}	1.42×10^{-2}	1433	315	331	-4.8
309	.166	.687	3.23×10^{-3}	5.42×10^{-2}	894	228	226	.88
<u>10%</u>								
682	.398	2.17	7.75×10^{-7}	1.008×10^{-4}	1894	532	578	-7.9
652	.350	1.84	2.32×10^{-6}	4.28×10^{-4}	1920	479	462	3.7
586	.295	1.42	3.46×10^{-5}	2.67×10^{-3}	1836	405	414	-2.2
531	.248	1.16	1.71×10^{-4}	7.67×10^{-3}	1797	340	342	-.58
419	.208	.707	2.91×10^{-3}	5.09×10^{-2}	1334	286	296	-3.4
<u>15%</u>								
685	.42	2.44	5.06×10^{-8}	3.25×10^{-5}	1796	577	571	1.05
634	.35	1.94	1.28×10^{-6}	2.91×10^{-4}	1848	480	465	3.22
586	.302	1.52	1.87×10^{-5}	1.77×10^{-3}	1835	412	408	.98
480	.240	.964	6.24×10^{-4}	1.86×10^{-2}	1574	324	328	-1.2
400	.215	.475	1.25×10^{-2}	1.35×10^{-1}	1194	295	301	-2.0
<u>20%</u>								
717	.424	1.90	1.59×10^{-6}	3.32×10^{-4}	1975	580	582	-.34
633	.341	1.63	9.26×10^{-6}	1.10×10^{-3}	1917	467	473	-1.27

TABLE VI (Cont).

8V/D	τ_o	n'	k'	v	Re	$\Delta P \frac{1b_f}{p_{ft}^2}$	$\Delta P \frac{1b_f}{a_{ft}^2}$	% dev.
<u>20% (cont)</u>								
523	.264	1.15	1.97×10^{-4}	8.66×10^{-3}	1686	361	384	-5.99
432	.215	.987	5.39×10^{-4}	1.69×10^{-2}	1408	296	302	-1.99
428	.210							
<u>25%</u>								
717	.421	1.74	4.53×10^{-6}	6.79×10^{-4}	2028	578	589	-1.9
625	.342	1.41	3.91×10^{-5}	2.95×10^{-3}	1898	470	473	- .63
565	.298	1.18	1.69×10^{-4}	7.90×10^{-3}	1777	410	409	.24
442	.232	.915	8.81×10^{-4}	3.38×10^{-2}	984	227	331	-3.14
<u>30%</u>								
686	.459	2.39	7.64×10^{-8}	4.42×10^{-5}	1743	630	664	-5.12
671	.435	2.30	1.37×10^{-7}	6.58×10^{-5}	1760	596	573	4.01
625	.378	1.75	4.84×10^{-6}	7.41×10^{-4}	1755	519	506	2.57
610	.365	1.64	9.87×10^{-6}	1.20×10^{-3}	1735	501	501	.0
442	.265	.638	5.44×10^{-3}	8.24×10^{-2}	1253	364	374	-2.67

TABLE VI (Cont).

8V/D	τ_o	n'	k'	ν	Re	$\Delta P \frac{lb_f}{pft^2}$	$\Delta P \frac{lb_f}{a ft^2}$	% dev.
<u>35%</u>								
671	.455	2.91	2.70×10^{-9}	4.61×10^{-6}	1728	623	663	-6.03
656	.425	1.91	1.77×10^{-6}	3.78×10^{-4}	1764	584	575	1.6
625	.392	1.44	3.69×10^{-5}	2.96×10^{-3}	1739	537	525	2.3
534	.328	.868	1.41×10^{-3}	3.45×10^{-2}	1512	452	451	.22
412	.274	.468	1.64×10^{-2}	1.75×10^{-1}	1074	378	372	1.61
<u>40%</u>								
674	.483	3.05	1.14×10^{-9}	2.6×10^{-6}	1666	661	688	-3.9
664	.453	2.22	2.46×10^{-7}	1×10^{-4}	1722	620	607	2.1
610	.388	1.22	1.55×10^{-4}	7.88×10^{-3}	1698	533	518	2.9
517	.333	.849	1.65×10^{-3}	3.88×10^{-2}	1428	459	437	5.03
412	.286	.576	8.92×10^{-3}	1.19×10^{-1}	1047	393	406	-3.20

by this procedure should be accurate within $\pm 25\%$. Table VII shows the pressure drops predicted by this method.

TABLE VII. PRESSURE DROP PREDICTIONS USING A TURBULENT VISCOSITY

Weight Coal	Velocity ft/sec	μ_t $\frac{\text{lb}}{\text{ft sec}}$	Re	ΔP_p $\frac{\text{lb}_f}{\text{ft}^2}$	ΔP_a $\frac{\text{lb}_f}{\text{ft}^2}$	% dev.
0	8.11	.0052	6885	529	529	0
	7.16		6078	486	480	1.3
	6.41		5442	398	423	-5.9
	5.76		4881	331	346	-4.3
	4.38		3718	213	204	4.4
5	7.71	.0054	6521	563	550	2.4
	7.33		6200	484	487	-.61
	6.18		5227	387	416	-7.0
	5.03		4255	274	331	-17.2
	3.38		2859	153	226	-32.3
10	7.45	.0059	5871	554	587	-4.15
	7.12		5611	512	462	10.8
	6.40		5044	423	417	2.17
	5.80		4571	363	342	6.14
	4.58		3609	238	296	-19.6
15	7.49	.0060	5898	568	571	-.53
	6.93		5464	499	465	7.3
	6.40		5046	430	408	5.4
	5.25		4140	312	328	-5.13
	4.37		3446	243	301	-19.3
20	7.83	.0055	6812	601	582	3.3
	6.92		6020	492	473	4.02
	5.71		4977	348	384	-9.4
	4.72		4107	252	302	-16.6
	4.68		4142	251	272	-7.7
25	7.83	.0049	7822	593	589	.68
	6.83		6823	474	473	.21
	6.17		6164	391	409	-4.4
	4.83		4825	259	331	-21.7

TABLE VII (Cont).

Weight Coal	Velocity Ft/sec	μ_t	$\frac{1b}{ft \ sec}$	Re	$\Delta P_p \frac{1b_f}{ft^2}$	$\Delta P_a \frac{1b_f}{ft^2}$	% dev.
30	7.50	.0068		5519	611	664	-7.98
	7.33			5394	591	573	3.14
	6.83			5026	518	506	2.40
	6.67			4908	494	501	-1.4
	4.83			3554	299	374	-20.05
35	7.33	.0074		5081	612	663	-12.4
	7.17			4970	586	575	3.5
	6.83			4734	549	525	6.0
	5.84			4041	417	451	-4.6
	4.50			3119	279	372	-25.0
40	7.36	.0078		4912	626	688	-9.01
	7.25			4838	614	607	1.15
	6.67			4452	543	518	4.80
	5.67			3784	408	437	-6.6
	4.50			3003	283	406	-30.3

In all cases, the average turbulent viscosity is the average value calculated from the friction factors that indicate the flow is turbulent. In all cases, except the pure oil, the slowest flow rates friction factors indicate they are in the laminar flow region. This is probably why they give values that are consistently so low.

It is interesting to note how turbulent viscosities increase up to 20%, begin to decrease and then increase sharply again at 30%. The decrease at 20% may be because the dampening effects of the slurry on the turbulence are overriding the increase in apparent viscosity. The sharp increase at 30% indicates, once again, that the viscosity begins to increase markedly at this concentration.

Comparing the turbulent viscosities with the apparent viscosities listed in Table III points out some interesting features. There is good agreement between the turbulent viscosity and Brookfield viscometer values for pure oil. However, the values from the viscometer increase at a much more rapid rate than the turbulent viscosities do. This once again points out the dampening effects of the slurry.

Methods proposed by Aude, Cowper, Thompson, and Wasp (3) were used to predict the pressure drop from viscometer data for homogeneous and heterogenous flow.

The method for homogeneous flow was developed for slurries exhibiting Bingham plastic characteristics. The Reynolds number is calculated by the equation,

$$Re = DV \rho / \eta$$

where

η = coefficient of rigidity.

The coefficient of rigidity for a Bingham plastic is the slope of a line given by plotting the rate of shear versus the shear stress.

This was approximated by taking the slope of the straight portions of rheograms shown in Figures 3 and 4.

The transition velocity is defined as the velocity where the flow makes its transition from laminar to turbulent flow. It is related

to the Hedstrom number by a curve proposed by Hanks (3). The Hedstrom number is defined as

$$He = \frac{D^2 \tau_y \rho g_c}{\eta^2}; \quad \tau_y = \text{yield stress}$$

Hank's curve is a plot of critical Reynolds number versus the Hedstrom number. Once the value of the Hedstrom number is known the critical Reynolds number can be determined from Hank's curve. The critical velocity is then calculated by:

$$U_c = \frac{Re_c \eta}{D \rho}$$

where

U_c = critical velocity

Re_c = critical Reynolds number

For heterogeneous flow, the Durand correlation is used to correlate the Froude number to volumetric solids concentration. The authors present a graph of the critical Froude number versus the volumetric solids concentration. The desposition velocity is related to the critical Froude number by:

$$U_d = (Fr)_c (g_c D)^{1/2}$$

where

U_d = desposition velocity

Fr_c = critical Froude number

Heterogeneous friction factors are then calculated by Durand's empirical relationship:

$$f = f_e \left[1 + 82 \left(\frac{\partial D}{u^2} - \frac{\rho - \rho_e}{\rho_e} \right)^{3/2} \frac{c_v}{c_d^{3/4}} \right]$$

where,

f_e = friction factor calculated using the viscosity of the pure liquid

ρ_e = density of the pure liquid

ρ = density of the solid

c_v = volumetric concentration

c_d = drag coefficient

The drag coefficient was taken to be constant at .44. The weight concentration was converted to the volumetric concentration by:

$$c_v = \frac{\frac{c_w}{\rho_c}}{\frac{c_w}{\rho_c} + \frac{1-c_w}{\rho_o}}$$

where,

c_w = weight concentration

ρ_c = density of the coal

ρ_o = density of the oil

The density of the coal was taken to be 79 pounds per cubic foot.

The density of the oil was 50.5 pounds per cubic foot.

Table VIII gives the predicted pressure drops calculated by these two methods.

TABLE VIII. PREDICTED PRESSURE DROPS FROM VISCOMETER DATA

Weight % Coal	U	$\frac{\text{ft}}{\text{sec}}$	U _c	$\frac{\text{ft}}{\text{sec}}$	U _d	$\frac{\text{ft}}{\text{sec}}$	Homo- geneous	Hetero- geneous	Measured
							ΔP $\frac{\text{lb}_f}{\text{ft}^2}$	ΔP $\frac{\text{lb}_f}{\text{ft}^2}$	ΔP $\frac{\text{lb}_f}{\text{ft}^2}$
5	7.71		4.06		2.51		629	530	550
	7.33						569	485	487
	6.18						426	366	416
	5.03						388	260	331
	3.38						-	134	226
10	7.45		4.69		2.70		617	511	578
	7.12						575	473	462
	6.40						465	399	414
	5.80						419	337	342
	4.58						286	223	296
15	7.49		4.69		3.02		680	526	571
	6.93						609	457	465
	6.40						544	404	408
	5.25						398	286	328
	4.37						-	211	301
20	7.83		5.43		3.19		716	576	582
	6.92						615	462	473
	5.71						438	335	384
	4.72						325	242	302
	4.68						320	238	272
25	7.83		6.47		3.35		806	591	589
	6.83						657	463	473
	6.17						569	394	409
	4.83						-	261	331

TABLE VIII (Cont)

Weight % Coal							Homo- geneous			Hetero- geneous			Measured
	U	$\frac{\text{ft}}{\text{sec}}$	U_c	$\frac{\text{ft}}{\text{sec}}$	U_d	$\frac{\text{ft}}{\text{sec}}$	ΔP	$\frac{\text{lb}_f}{\text{ft}^2}$	ΔP	$\frac{\text{lb}_f}{\text{ft}^2}$	ΔP	$\frac{\text{lb}_f}{\text{ft}^2}$	$\frac{\text{lb}_f}{\text{ft}^2}$
30	7.50		6.06		3.52		773		559		664		
	7.33						755		541		573		
	6.83						661		478		506		
	6.67						630		457		501		
	4.83						-		270		374		
35	7.33		6.94		3.52		773		554		663		
	7.17						756		531		575		
	6.83						701		491		525		
	5.84						534		382		451		
	4.50						-		261		372		
40	7.36		9.06		3.70		-		573		688		
	7.25						-		557		607		
	6.67						-		483		518		
	5.67						-		378		437		
	4.50						-		273		406		

The values that are not presented for the homogeneous flow case are those that gave a Reynold's number less than 2100. The correlation is not meant to cover laminar flow. The friction factor for pure oil was obtained using a viscosity of .0048 lb/ft.sec.

There are several difficulties in using this method. One, of course, is that for the most part the slurries do not appear to be Bingham plastics. Approximation by using the slope of the straight sections of the rheograms is crude at best. Another is that the yield stress is extremely difficult to measure. It may very well be that slurries do not even have a yield stress, the rheograms of Figures 3

and 4 may simply curve down to a zero value. The best that could be done was to extend the smooth curve to the y-axis.

At best, this method appears only to give a range that the pressure drop will probably fall into. The homogeneous method gives consistently high values for the pressure drop while the heterogeneous values are constantly low.

The study confirms that the slurries have mixed flow characteristics in the range of velocities tested. No ideal flow type was found to correlate all the data.

CONCLUSIONS

The following conclusions can be made from this investigation.

1. The fine coal-crude oil slurries were found to conform to four rheological models. Pure crude oil appears to be Newtonian, concentration up to 10% pseudoplastic, concentrations from 10% to 40% real plastics, and from 40% to 50% Bingham plastic.
2. The flow pattern appears to be mixture of three ideal types, homogeneous flow, heterogeneous flow, and plug flow.
3. Apparent viscosities measured by a Brookfield viscometer increased much more rapidly than do the turbulent viscosities calculated from the pipe loop data.
4. Viscometer data can at best give a rough estimate as to what range the pressure drops will fall into.
5. The viscosity of the slurries increases rapidly at a weight concentration of 30 percent.

RECOMMENDATIONS FOR FUTURE STUDY

1. A larger diameter pipeloop should be constructed to determine how accurately the pipeloop data can be scaled up.
2. A detailed study of the slurry's flow pattern should be conducted.
3. The slurry's abrasiveness should be studied in greater detail.
4. The highest percentage of coal that can be economically carried in the slurry should be determined.
5. Critical velocities, at which the coal settles in the line, should be determined.

APPENDIX

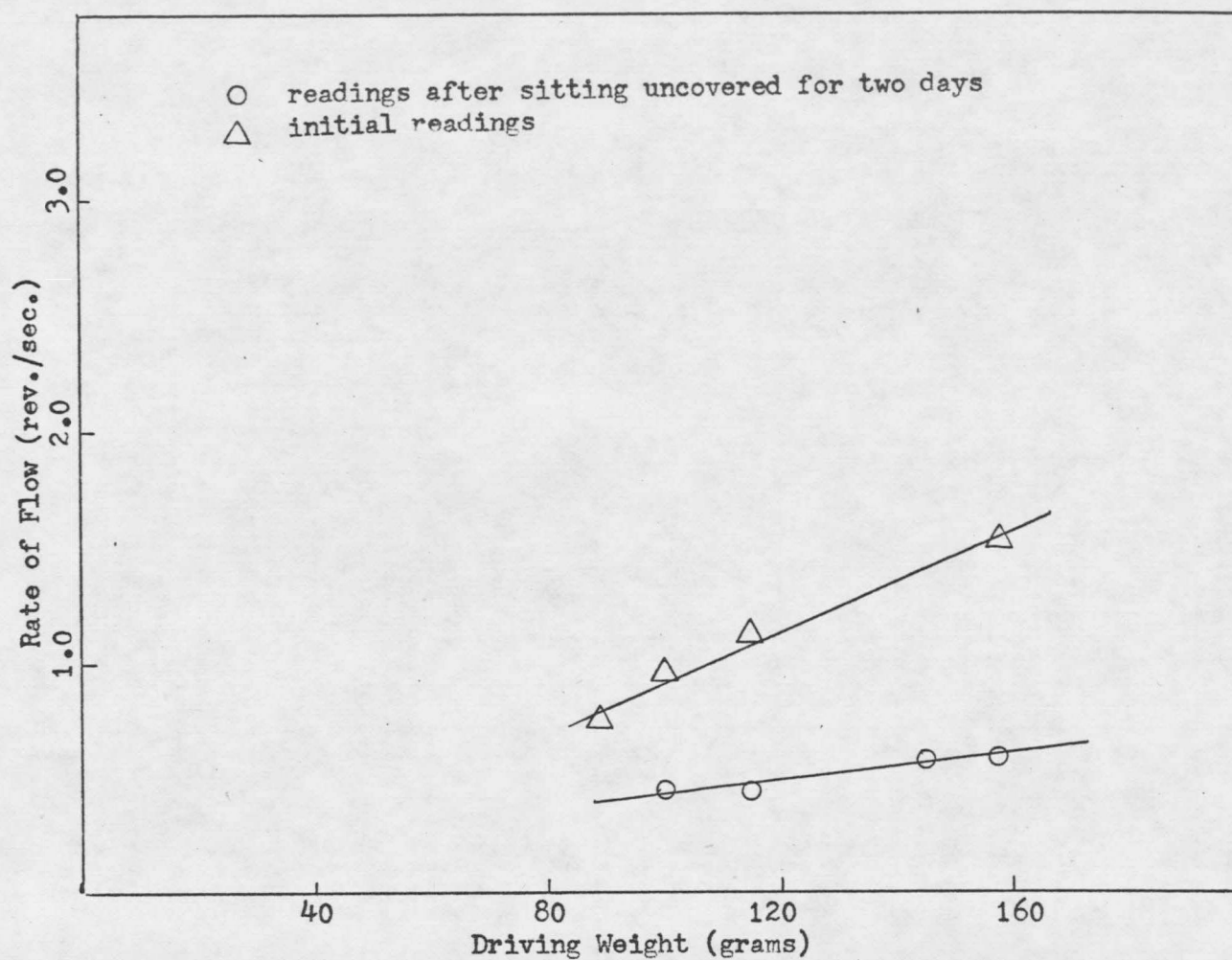


FIGURE 18. Stormer Viscometer Rheogram for 500 Grams Dried Coal-600 Milliliters FMC Synthetic Crude Slurry After Pressurization to 500 psig

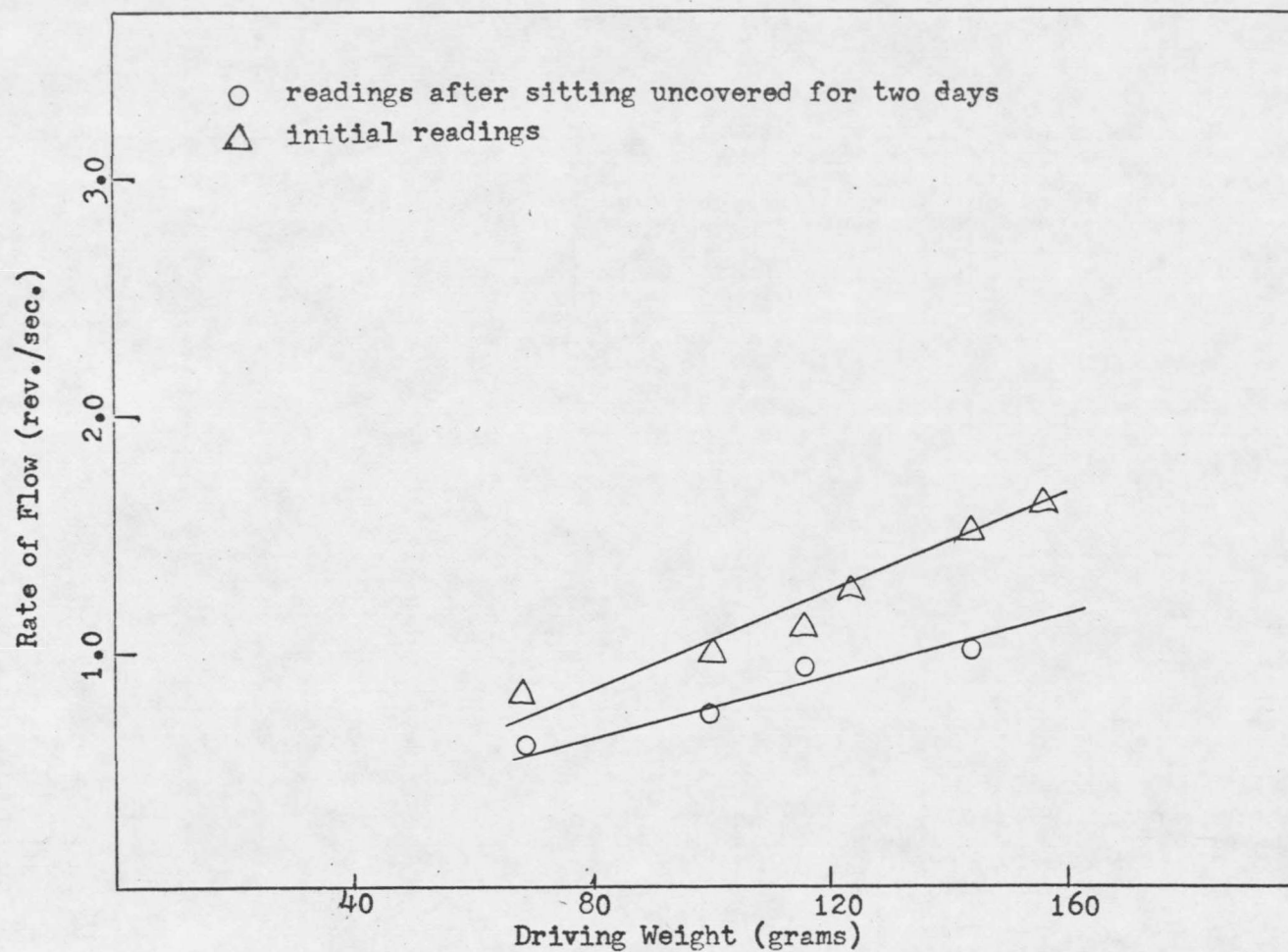


FIGURE 19. Stormer Viscometer Rheogram for 500 Grams Dried Coal-600 Milliliters FMC Synthetic Crude Slurry After Pressurization to 1000 psig

Mooney-Rabinowitsch-Weissenberg Equation:

$$\dot{Q} = \int_0^{r_o} 2\pi r V_x dr = 2\pi \int_0^{r_o} r V_x dr$$

where,

$\dot{Q}$ = volumetric flow rate

Integrate by parts

$$\dot{Q} = 2\pi \left\{ \left[V_x \frac{r^2}{2} \right]_0^{r_o} - \int_0^{r_o} \frac{r^2}{2} dV_x \right\} = -2\pi \int_0^{r_o} \frac{r^2}{2} dV_x$$

But, we know that

$$\tau_{rx} = \frac{\tau_o}{r_o} r$$

so

$$dr = \frac{r_o}{\tau_o} d\tau_{rx}$$

$$\text{If } \gamma = f(\tau) \equiv \frac{dV_x}{dr} \quad (\text{key assumption})$$

$$\dot{Q} = -2\pi \int_0^{r_o} \left(\frac{r_o}{\tau_o} \right)^2 \tau_{rx}^2 \left(\frac{1}{2} \right) f(\tau) dr$$

$$= \frac{-2\pi}{2} \int_0^{\tau_o} \left(\frac{r_o}{\tau_o} \right)^2 \tau_{rx}^2 f(\tau_{rx}) \frac{r_o}{\tau_o} d\tau_{rx}$$

$$= -\pi \left(\frac{r_o}{\tau_o} \right)^3 \int_0^{\tau_o} \tau_{rx}^2 f(\tau_{rx}) d\tau_{rx}$$

$$\frac{\dot{Q}}{\pi r_o^3} = -\frac{1}{\tau_o^3} \int_0^{\tau_o} \tau_{rx}^2 f(\tau_{rx}) d\tau_{rx} \quad \text{which is just a function of } \tau_o$$

So, a plot of $\frac{\dot{Q}}{\pi r_o^3}$ vs τ_o should be a unique function,

even if data are collected from tubes of various sizes.

$$\text{Define } \frac{\dot{Q}}{\pi r_o^3} \equiv -\phi(\tau_o)$$

So, previous equation becomes

$$\phi = \frac{1}{\tau_o^3} \int_0^{\tau_o} \tau_{rx}^2 f(\tau_{rx}) d\tau_{rx}$$

Differentiate ϕ with respect to the upper limit, τ_o :

$$\frac{d\phi}{d\tau_o} = -\frac{3}{\tau_o^4} \int_0^{\tau_o} \tau_{rx}^2 f(\tau_{rx}) d\tau_{rx} + \frac{1}{\tau_o^3} [\tau_o^2 f(\tau_o)]$$

$$\phi' = \frac{3}{\tau_o} (-\phi) + \frac{1}{\tau_o} f(\tau_o)$$

$$f(\tau_o) = \tau_o \phi' + 3\phi$$

the shear rate which corresponds to

In terms of easily measured quantities:

$$v = \text{bulk velocity} = \frac{\dot{Q}}{\pi r_o^2}$$

So

$$\phi(\tau_o) = -\frac{\dot{Q}}{\pi r_o^3} = -\frac{v}{r_o} = -\frac{2v}{D_o}$$

So

$$f(\tau_o) = -\tau_o \frac{d\left(-\frac{2v}{D_o}\right)}{dv_o} = 6 \frac{v}{D_o}$$

So

$$-\left(\frac{dv_x}{dr}\right)_o = \frac{3}{4} \left(-\frac{8v}{D_o}\right) + \frac{\tau_o}{4} \frac{d\left(-\frac{8v}{D_o}\right)}{d\tau_o}$$

and of course,

$$\tau_o = \frac{D_o \Delta P}{4L}$$

NOMENCLATURE

C_d	Drag Coefficient	dimensionless
C_v	Volumetric Concentration	dimensionless
C_w	Weight Concentration	dimensionless
D	Inside diameter	ft.
D_o	Inside diameter in Mooney-Rabinowitsch Weissenberg Equation	ft.
f	Friction factor	dimensionless
f_e	Friction factor of pure liquid	dimensionless
g_c	Gravitational constant	$lb_m/lb_f \quad ft/sec^2$
k'	Consistency index	$lb_f/ft^2(sec^{n'})$
L	Length	ft.
n'	Flow behavior index	dimensionless
ΔP	Pressure drop	lb_f/ft^2
ΔP_a	Experimental pressure drop	lb_f/ft^2
ΔP_p	Predicted pressure drop	lb_f/ft^2
r	Radius	ft
r_o	Inside radius in Mooney-Rabinowitsch- Weissenberg Equation	ft.
U_d	Deposition velocity	ft/sec.
U_c	Critical velocity	ft/sec.
V	Velocity	ft/sec.
v	$g_c k' 8^{n'-1}$	$(lb_m/ft \cdot sec^2) sec^{n'}$
w	weight	lb_m

NOMENCLATURE (Cont).

Fr	Froude Number , V^2/Dg_c	dimensionless
He	Hedstrom Number, $D^2\tau_o\rho g_c/\eta^2$	dimensionless
Re	Reynolds Number, $VD\rho/\mu$	dimensionless
τ_o	Shear Stress at the wall	lb_f/ft^2
τ_y	Yield stress	lb_f/ft^2
μ	Viscosity	$lb_m/ft.sec$
μ_e	Effective viscosity	$lb_m/ft.sec$
μ_t	Turbulent viscosity	$lb_m/ft.sec$
ρ	Density	lb_m/ft^3
ρ_e	Density of pure liquid	lb_m/ft^3
η	Coefficient of rigidity	$lb_m/ft.sec.$

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