

**The Development of the KENTORT II Process
for Eastern U.S. Oil Shale**

Final Report

**S.D. Carter
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T.L. Robl
D.N. Taulbee**

May 1990

Work Performed Under Contract No.: DE-FC21-86LC11086

**For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia**

**By
University of Kentucky
Center for Applied Energy Research
Lexington, Kentucky**

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Lexington, Kentucky 40511-8433**

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ABSTRACT

This report summarizes the development of the KENTORT II process during the period of September 1, 1986 through December 31, 1989. The KENTORT II development program was conducted during this period under Cooperative Agreement No. DE-FC21-86LC11086 between the University of Kentucky Center for Applied Energy Research and the Laramie Project Office of the Morgantown Energy Technology Center, U.S. Department of Energy. The purpose of this program was to develop on a small scale a process to improve efficiency, process economics, and resolve environmental problems involved in the utilization of eastern U.S. oil shale as an energy source and chemical feedstock. The process includes fluidized bed sections of pyrolysis, gasification, and combustion with combined solid/gas heat transfer among the stages.

The operation of each zone of the KENTORT II process was demonstrated both separately in a bench-scale unit and as an integrated system in a 5-lb/hr mini-plant. There were no technological barriers encountered or foreseen that would hinder operation at a larger scale. Sulfur removal of up to 91% was achieved following integral pyrolysis-gasification processing, and no more than 8.3% of the sulfur was converted to SO_2 when the combustion zone was in operation. Carbon gasification proceeded more slowly than sulfur removal, but a significant amount of H_2 -rich syngas was produced. Under most conditions, sufficient hydrogen was produced to completely upgrade the crude shale oil. Heat transfer to the pyrolysis zone by solid recirculation was shown to be practical and did not seriously reduce oil yields. Oil yields averaged 111% of modified Fischer assay when the mini-plant was operated with solid recirculation and with the combustor in place.

This study was successful in experimentally demonstrating the KENTORT II process at a miniature scale, and it is recommended that development of the process continue. Due to the geometry of the KENTORT II design, a commercially realistic heat transfer situation could be not achieved in the 5-lb/hr system. Therefore, further testing at a larger scale is the next necessary step in the development of the KENTORT process.

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The work described herein was performed at the University of Kentucky Center for Applied Energy Research under Cooperative Agreement DE-FC21-86LC11086 with the Laramie Project Office of the Morgantown Energy Technology Center (METC), U.S. Department of Energy.

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1. EXECUTIVE SUMMARY

1.1 ABSTRACT

This report summarizes the development of the KENTORT II process during the period of September 1, 1986 through December 31, 1989. The KENTORT II development program was conducted during this period under Cooperative Agreement No. DE-FC21-86LC11086 between the University of Kentucky Center for Applied Energy Research and the Laramie Project Office of the Morgantown Energy Technology Center, U.S. Department of Energy. The purpose of this program was to develop on a small scale a process to improve efficiency, process economics, and resolve environmental problems involved in the utilization of eastern U.S. oil shale as an energy source and chemical feedstock. The process includes fluidized bed sections of pyrolysis, gasification, and combustion with combined solid/gas heat transfer among the stages.

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This study was successful in experimentally demonstrating the KENTORT II process at a miniature scale, and it is recommended that development of the process continue. Due to the geometry of the KENTORT II design, a commercially realistic heat transfer situation could be not achieved in the 5-lb/hr system. Therefore, further testing at a larger scale is the next necessary step in the development of the KENTORT process.

1.2 INTRODUCTION AND BACKGROUND

Fluidized bed pyrolysis of eastern U.S. oil shale can enhance oil yields up to 50% above Fischer assay.(1,2,3,4,5) Fluidized bed technology also provides rapid pyrolysis kinetics (short solid residence time), total use of mined shale, and thermal efficiency. Therefore, fluidized bed technology applied to eastern U.S. oil shale has the potential to increase oil yield, and enhance process efficiency while decreasing capital

and operating costs. The University of Kentucky Center for Applied Energy Research (CAER) and its forerunner organizations have investigated fluidized bed pyrolysis of eastern U.S. oil shales for the past several years, and emphasis has recently been broadened to include the development of a commercial retorting concept based on fluidized bed pyrolysis. This effort focuses on improving process efficiency and mitigating environmental considerations for an Eastern oil shale process. Improvement in environmental performance is accomplished by incorporating waste stream reduction into the design of the reactor. This approach is inherently more efficient than tail-end cleaning of waste materials.

While fluidized bed pyrolysis significantly increases carbon conversion to liquid products, approximately 40% of the original carbon remains in the spent shale. This material is ideal for on-site combustion since it has already been heated in excess of 500°C and is of no value once removed from the process. Additionally, it is a good heat carrier because of its small size, relatively low rates of decrepitation, and high heat capacity. Accordingly, char combustion and the use of char as a heat transfer medium are integral features of the KENTORT II process.

Several processes have been proposed for other shales that utilize char combustion with hot solids recirculation for generating and transferring heat in the retort. However, direct combustion of the char does not represent the best approach for processing Eastern shale. Since only a portion of the carbon in the char from pyrolysis is required by the combustor to meet process heat requirements, total carbon utilization would require the removal of heat from the reactor. This excess heat could be used in the plant or power could be exported. However, the economics of cogeneration are highly site-specific, and this is generally not considered advantageous for an oil shale process. In addition, rapid pyrolysis concentrates sulfur in the spent shale compared to carbon, so direct combustion would generate substantial SO₂ emissions. Environmental considerations make it imperative that these emissions be minimized. Eastern U.S. shales contain substantial amounts of sulfur that, with an appropriate process configuration, can actually contribute favorably to process economics and reduce SO₂ emission simultaneously.

Steam pyrolysis studies indicated that up to 90% of the sulfur in Eastern shale can be removed under the proper conditions.(6) These results provided the impetus to include a mild gasification stage between pyrolysis and combustion in the KENTORT II process. The removal of sulfur from the shale as H₂S prior to combustion creates a relatively concentrated stream of H₂S which is easier to scrub than a dilute stream of SO₂ from a combustor. The additional H₂S also serves as feedstock for

elemental sulfur production. Moreover, the generation of syngas eliminates the need to export power. The syngas has potential applications ranging from in-plant use to feedstock for alcohol production to shale oil upgrading. Environmentally, the spent shale is safe for disposal since the gasification process removes up to 90% of the original sulfur and the combustor removes the balance. Sulfur removal and the oxidation of iron during the combustion step greatly reduce the potential for acid drainage conditions in the disposal area.(7)

1.3 OBJECTIVES

The main objectives of this program were to experimentally examine the untested elements of the KENTORT II concept, demonstrate the operation of the process in an integrated reactor, modify the process concept if necessary, and demonstrate the feasibility of the concept. There were two critical areas of the KENTORT II design that required experimental confirmation before the process could be considered feasible for continued development. The first was the chemistry and kinetics associated with the gasification section. The concern here was that the conditions required for substantial sulfur removal and carbon gasification may be too severe. Lengthy solids residence times would make gasification economically unattractive, and high temperatures would impede optimal combustor performance. The second area of concern was the use of hot, recycled shale to transfer heat to the pyrolysis zone. It has been reported that substantially reduced oil yields result when hot, combusted shale was mixed with raw shale during pyrolysis.(8,9) The KENTORT II concept offered theoretical, but unproven, characteristics that would reduce oil losses in a solid-recycle system, so demonstration of high oil yields was an important goal of the program.

The final objectives were to make modifications to the KENTORT concept, if necessary, and evaluate the viability of the process. The purpose of this step was to determine whether further testing of the concept was warranted. It was beyond the scope of this project to perform an economic evaluation of the process, but it was possible to make general conclusions regarding commercial-scale operation and feasibility.

1.4 APPROACH

The experiments in this study were performed at two scales: bench-scale and mini-plant. The bench-scale system was a 1.5-inch diameter fluidized bed reactor which was used to investigate the kinetics and chemistry of shale char gasification and combustion. These results were then applied to the design and operation of the mini-plant. The mini-plant was a 3-inch diameter, 5-lb/hr system which simulated the three major zones of the KENTORT process. The mini-plant was developed and operated in two phases which correspond with two of the major tasks of the program:

- o The development and integration of fluidized bed oil shale char gasification technology with fluidized bed pyrolysis technology.
- o The development of oil shale combustion technology and integration with pyrolysis and gasification stages.

Each of these tasks contained two main subtasks:

- o Fundamental chemical and kinetic investigations of char gasification and combustion processes to be performed in the 1.5-inch diameter bench-scale unit.
- o Integral operation of gasification and combustion zones with a fluidized bed pyrolyzer in the 3-inch diameter reactor system.

A third task involved the examination of the leaching characteristics of the processed materials. Characterization of the liquid and gaseous products generated from the mini-plant was also performed.

1.5 RESULTS

1.5.1 Gasification. The steam-carbon reaction was found to be the dominant gasification reaction in both reactor systems. This was not surprising since water vapor was substantially in excess of all other components allowing the water gas shift reaction proceed readily. Bench-scale experiments showed decreases in water gas shift conversion when the steam partial pressure was decreased. Temperature was the dominant parameter influencing carbon gasification. Temperatures ranging from 700 to 850°C were investigated in the bench-scale reactor and a slightly narrower range (735-840°C) was achieved in the integrated mini-plant reactor. Mean solid residence time, τ , was the other major parameter of influence, and this was varied from 10 to 60 minutes in the bench-scale reactor and from 20 to 30 minutes in the integrated reactor. Little gasification occurred

at 700°C in the bench-scale unit, but temperatures over 800°C affected a significant increase in carbon conversion. Nearly 50% of the carbon in the char was converted at 850°C and $\tau=60$ minutes, but the incremental difference between $\tau=60$ and 30 minutes was small. Carbon gasification ranged from 10 to 20% in the mini-plant at 30 minutes holding time and temperatures between 800 and 840°C.

The carbon gasification kinetics of the spent shales in the two systems were nearly identical. This was somewhat of an unexpected result because it was believed that the integrated unit would result in higher char reactivity from the lack of a cool-down step following pyrolysis. Previous studies with other oil shales have shown an intermediate cooling and storage period to have a detrimental effect on char reactivity. The gasification reactivity of Eastern oil shale chars does not appear to be highly influenced by the method of pyrolysis. Results from the bench-scale unit were very similar to gasification studies where chars were produced by Fischer assay.(10,11) The reactivity of eastern U.S. oil shale char as a function of pyrolysis conditions is an area that certainly deserves additional investigation.

Sulfur removal in the bench-scale unit and the mini-plant proceeded more quickly and nearer to completion compared to carbon gasification. At 800°C and $\tau=30$ minutes, nearly 80% of the original sulfur in the char fed to the bench-scale reactor was converted to H_2S . Sulfur removal in the mini-plant in all modes of operation was similar to the results from the bench-scale unit. However, some recapture of H_2S by recirculating shale was evident, which lowered the overall conversion to H_2S , but this effect was minimized by the large partial pressure of steam in the system. Since substantial sulfur removal was achieved at temperatures as low as 700°C, the conditions of the gasification section of KENTORT will be dictated primarily by the level of syngas production that is desired. Without tight processing constraints for sulfur removal, the KENTORT process has considerable processing flexibility in terms of the gas product slate that is possible.

1.5.2 Combustion. Combustion of raw, pyrolyzed, gasified, and blended shales was performed in the bench-scale unit. The gasified shale was combusted without difficulty, and nearly complete removal of carbon, nitrogen, and sulfur was achieved at $\tau = 2$ minutes and 700°C. Shorter residence times were not possible in the bench-scale reactor. SO_2 and NO_x emissions were low due to the reduced levels of S and N in the gasified shale. Combustion of raw and pyrolyzed shales was hindered by sintering. The shales that sintered were found to contain iron hydroxysulfate, an oxidation product which probably formed during storage and which has also been linked to sintering during coal combustion. While fresh shale feedstocks may eliminate this

problem from the onset, it appears that gasification removes the tendency for sintering. This represents another, albeit unexpected, advantage of the gasification section of KENTORT.

Combustion operations were trouble-free in the mini-plant although autothermal operation of the mini-plant was not quite achieved because the solid-recycle loop to the combustor was not large enough. Therefore, the rate of shale recycled to the combustor was established at its maximum level for each run, and the only means of temperature control was to starve the combustor of oxygen by blending N_2 with the fluidizing air. Relatively high levels of CO were attributed to this method of operation. There was sufficient carbon remaining after the pyrolysis and gasification steps to fuel the system, so an increase in recycle capability would resolve this problem. The generation of SO_2 and NO_x were low during operation of the mini-plant with a maximum of 8.3% of the original sulfur being converted to SO_2 , and no more than 1.4% of the nitrogen was detected as NO_x . Therefore, solid recirculation between the gasifier and combustor simply affected greater sulfur removal from the char and did not significantly alter H_2S production.

1.5.3 Solid Recycle Pyrolysis. Studies were performed in the mini-plant to examine the interactions of the gasifier and combustor with the pyrolyzer. No bench-scale pyrolysis was performed because of our previous experience in the area. Runs were performed in the mini-plant in the pyrolysis-gasification mode to establish baseline oil yields where the pyrolysis zone was heated by hot gases and the hot walls of the reactor. These conditions are expected to minimize oil losses due to cracking and coking. These runs resulted in an average of 45.9% carbon conversion to oil, which represents a 29% increase compared to the modified Fischer assay.(12) These yields represent the amount of oil that was actually collected and measured, as opposed to a 'by-difference' yield. Moreover, the organic carbon dissolved in the water was not classified as 'oil.' Thus, the reported oil yields in this study are most conservative.

The first phase of solid-recycle tests was designed to investigate the effects of recycle rate and solid temperature on the yield and composition of pyrolysis products. As expected, both increased rate and temperature of the solids recycled to the pyrolyzer resulted in oil yield losses. Under the most severe solid-recycle conditions, less than 15% oil loss was recorded. In the KENTORT reactor, a given amount of heat may be transferred by a wide range of solid recycle rates and temperatures. These parameters are inversely proportional when the raw shale throughput is held constant. By comparing runs of constant raw shale feed rate in the first phase of operation, it appears that solid temperature was more detrimental to oil yield than recycle rate. This sort of relationship would have important implications for the KENTORT process because it would provide a

simple engineering solution for reducing oil yield losses; i.e., operating at high recycle rate with low temperature solids.

The second phase of mini-plant operation incorporated an integral combustion zone. A key objective in operating the mini-plant with and without combustion was to distinguish the catalytic activity of gasified versus partially combusted/gasified shale. The two independent variables in this portion of the study were gasification and combustion temperature. The solid recycle rate to the pyrolyzer was inversely related to the gasification temperature and was a dependent parameter. While gasifier temperature was determined to have a slight negative effect on oil yield during non-combustion operation, no such trend was discernable in the second phase of operation. There was also no correlation between combustor temperature or carbon burn-off and oil yield. While oil yields were not perceptibly affected, it does appear that some product properties were influenced by altering the rate, temperature, and composition of the recycle solids. These include changes in the aromatic content, density, average molecular weight, and viscosity of the product oil as well as differences in the quantity and composition of the hydrocarbon gases. In general, oil properties were improved for the solids recycle runs relative to the non-recycle runs.

It appears that combustion caused slightly lower oil yields based on a comparison between non-combustion and combustion operation. However, the difference in the median carbon conversion to oil was only 0.7 percent which was statistically significant at only an 80 percent confidence level. While somewhat difficult to explain at this time, these results are nevertheless encouraging. Oil yields averaged 111% of the modified Fischer assay for the combustion phase of this work, and it appears that more extensive combustion will not cause greater oil yield reduction.

Because of the geometry of the KENTORT design, heat transfer by the fluidizing gases from the gasifier is significant, thus only 60% of the heat of pyrolysis could be provided by solids recirculation in the current integrated system. Further experiments on larger reactors are required to more adequately investigate the effects of a more commercially realistic regime of solid-recycle conditions.

1.5.4 Environmental. Improved environmental performance is a major factor in the KENTORT concept. The strategy is to reduce or eliminate process waste streams and improve the environmental characteristics of the spent shale. Waste stream reduction is achieved by recycling sour water and by the minimization of SO₂ scrubbing requirements (and associated sludges). KENTORT also recovers substantial nitrogen in the form of ammonia which results in low NO_x emissions.

This study found that only 8.3% of the original sulfur was ultimately converted to SO_2 , and the majority of the balance was converted to H_2S . Nitrogen oxides from combustion were also very low, and are not considered to be a problem for the technology. The high CO levels generated in the combustor can be lowered in future designs by using excess air to enhance freeboard combustion. In general, from an air emission standpoint, the research demonstrated KENTORT to be, potentially, a very clean technology.

The spent shales produced in the study have very low levels of sulfur, indicating complete conversion and oxidation of sulfide minerals in the process. Mossbauer spectrometry also found the iron in the spent shale to be in the ferric form (Fe^{3+}). Iron sulfide hydrolysis and oxidation are the principal chemical reactions responsible for acid generation from the spent shale. Thus, the final KENTORT spent shale will have little or no acid generating capacity and greatly lowered elemental leaching characteristics compared to solids from older technologies.

1.6 CONCLUSIONS AND RECOMMENDATIONS

The operation of each zone of the KENTORT II process was demonstrated both separately in the bench-scale unit and as an integrated system in the 5-lb/hr mini-plant. There were no technological barriers encountered or foreseen that would hinder operation at a larger scale. Autothermal operation was not reached because the solid recycle loop between the gasifier and combustor was slightly undersized. However, considering the small size of the system, remarkably realistic conditions were achieved in the mini-plant. The operation of larger scale systems will actually be easier. For example, the solid transfer lines were so small that literally fractions of an inch made major differences in the solid transfer rates that were achieved.

Demonstrating operability of the process and proving the viability of the process concept are not entirely the same. From the onset of the program, there were two fundamental aspects of the KENTORT II concept that required experimental confirmation before the process could be considered for further development. First, demonstration of substantial sulfur removal under mild gasification conditions was necessary. Sulfur removal in the gasification zone proceeded quickly and under relatively mild conditions. Additionally, syngas production under these conditions was significant which could play an important role in the economics of the process. Secondly, heat transfer to the pyrolysis zone by solid recirculation needed to be accomplished without seriously reducing oil yields. This result was achieved during the study, as oil yields averaged 111% of modified Fischer assay when the mini-plant was operated with the combustor in place.

This study was successful in experimentally demonstrating the KENTORT II process at a miniature scale, and it is recommended that development of the process continue. Due to the geometry of the KENTORT II design, a commercially realistic heat transfer situation could be not achieved in the 5-lb/hr system. Therefore, further testing at a larger scale is the next necessary step in the development of the KENTORT process.

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2. PROCESS DESCRIPTION

2.1 INTRODUCTION

There are two characteristics of eastern U.S. oil shales which make efficient and environmentally acceptable oil production particularly troublesome. First, the generally low H/C ratio of the kerogen coupled with a reactive mineral matrix promotes coking and cracking of the product oil during pyrolysis. This results in low Fischer assay oil yields compared to Western oil shales of similar carbon contents. Second, the sulfur content of Eastern shale is comparatively high, posing potential environmental problems in terms of solid disposal and flue gas emissions. Retorting studies at the CAER over the past several years have revealed information to overcome these obstacles and have enabled the conception of an efficient processing technology for Kentucky oil shales called KENTORT II.

By rapidly heating the shale, and quickly sweeping the pyrolysis products from the hot reaction zone, it was observed that secondary reactions were reduced and oil yield was increased.(1) When fluidized bed technology was used to simultaneously achieve rapid heating rate and fast sweep gas velocity, the highest oil yields were recorded. Oil yields of up to 155% of Fischer assay have been observed using the fluidized bed technique.(2,3)

While fluidized bed pyrolysis is an efficient method of quickly generating high yields of crude shale oil, substantial amounts of residual carbon and sulfur remain in the spent shale. Some residual carbon is desirable in order to fuel a combustion step, but the residual sulfur creates solid disposal problems. Spent shale leaching studies have indicated that incomplete conversion of pyrite during pyrolysis results in elevated levels of acid drainage.(4) However, prohibitively lengthy pyrolysis reaction times are required to complete the pyrite conversion. During steam fluidization studies it was observed that sulfur removal was achieved faster compared to other fluidizing media and that 90% sulfur removal could be achieved at residence times greater than 30 minutes and temperatures less than 600°C.(3) The exploitation of the steam/FeS_x reaction is, therefore, potentially a useful method for controlling the speciation of sulfur during and after pyrolysis.

Combustion of the char from pyrolysis has traditionally been reserved as the method for utilizing residual carbon. This technique presents two problems when applied to Eastern shales. First, char from fluidized bed pyrolysis contains more residual carbon than required to provide heat for the process. It would be necessary to remove substantial amounts of energy from the reactor to deplete all of the carbon in this manner. Second, the

S:C weight ratio of pyrolysis char is on the order of 1:3, which if combusted would produce large, if not prohibitive, amounts of SO_2 .

Therefore, steam gasification of the excess carbon in the pyrolysis char represents an attractive alternative to complete combustion. An intermediate gasification step was included in the KENTORT II design to take advantage of these attributes.

2.2 PROCESS OVERVIEW

The heart of the KENTORT II process (Figure 2-1) is a two-stage fluidized bed reactor system (pyrolysis followed by mild gasification). The reactor maximizes the removal of oil, gas, and sulfur from the shale using steam at atmospheric pressure as the fluidizing medium. While improved oil yield from Eastern shale was the main attraction to fluidized bed technology, there are resource-independent considerations of fluidized bed technology that make shale processing more efficient. For example, fluid beds can process fines, are thermally efficient, improve mass transfer, and enhance reaction kinetics.

A combustor burns undersized raw shale fines and partially gasified oil shale char to provide the energy to drive the process. Shale recirculation among the zones provides an efficient heat transfer mechanism. Sensible heat from the processed shale is recovered through a quench from recycled process water. Following oil condensation, the remaining vapor stream is routed to light oil recovery, and gas separation units. Elemental sulfur produced by a Claus reactor and ammonia are recovered as marketable by-products.

2.3 PYROLYSIS ZONE

The first stage of the KENTORT II reactor is a bubbling fluidized bed where pyrolysis takes place. The fluidizing medium consists of unreacted steam and product gases from the gasification section below. Heat is provided to the pyrolysis zone by an optimum combination of fluidizing gas and recirculating solids from the hotter, second-stage. The pyrolysis zone is baffled in order to narrow the residence time distribution of the shale. Raw shale is fed over-bed while the hot recirculating solids are injected into the bottom of the bed to prevent initial direct contact between the hot and cold shales. Temperatures between 500 and 550°C and mean shale residence times of 1-2 minutes are required to liberate greater than 95% of the oil product.

The baffled, dense-phase fluidized bed pyrolysis reactor offers advantages over other types of small particle technologies

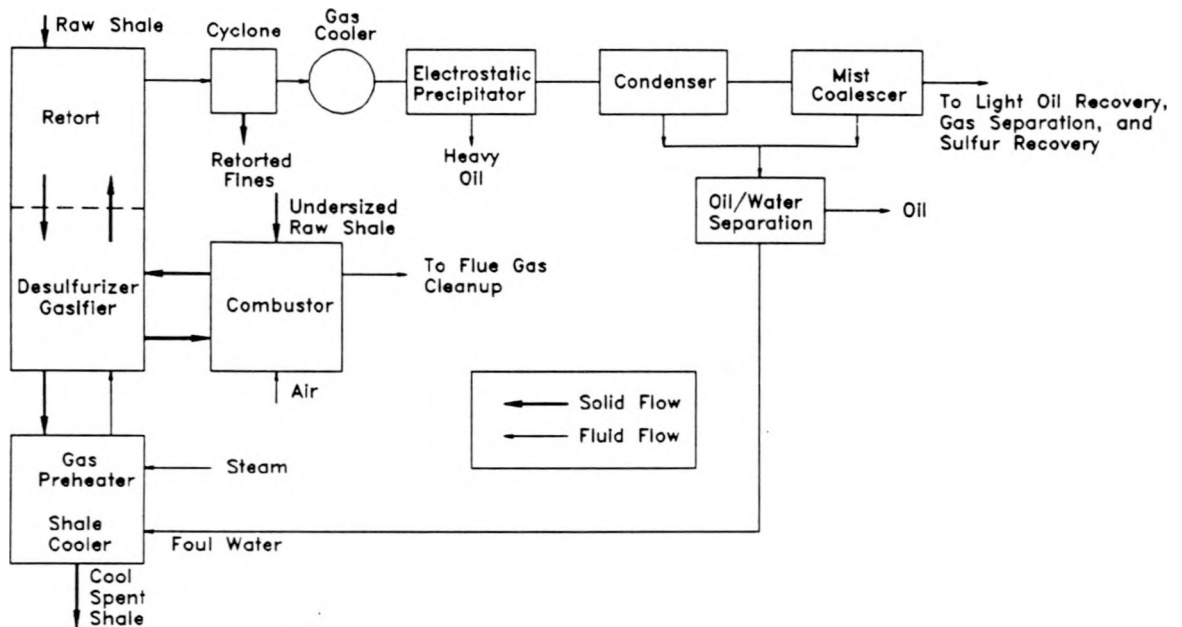


Figure 2-1. *Simplified Schematic of the Original KENTORT II Process Concept.*

such as circulating fluidized beds and entrained reactors. Dense phase fluidized beds have a low void volume, and the solids residence time is not dependent on the gas flow rate. Additionally, relatively low gas velocities are required, compared to a lean-phase reactor, resulting in less gas and/or steam handling, lower cyclone loading, reduced particle attrition, lower erosion rates, and ultimately, reduced fines content of the oil.

2.4 GASIFICATION ZONE

Shale is fed to the gasifier, a multi-staged bubbling fluidized bed, via a downcomer from the pyrolysis zone. Temperatures of 700-1000°C are used to gasify a portion of the residual carbon following pyrolysis. The reaction rate of steam with the iron sulfides is quickened at these temperatures, and the original pyrite is reduced almost completely to iron oxides in this zone. The conversion of the majority of the sulfur to H_2S in this zone is advantageous in itself, since the removal of concentrated H_2S is much easier than clean-up of a dilute stream of SO_2 . The additional advantage of this scheme is that H_2S is easily convertible to elemental sulfur which contributes to the economics of the process. Heat for the gasification zone is provided by recirculating solids from the combustor. Hot combusted solids enter the lower stage and create a thermal gradient in the unit resulting in efficient countercurrent heat transfer. Staging in the gasification section also serves to narrow the residence time distribution of the char.

2.5 COMBUSTION ZONE

The combustor provides all the energy required for the pyrolysis and gasification sections of the KENTORT II process. Exchange of solids between the combustor and the gasifier occurs without mixing the gas streams. As a result, the gasification and pyrolysis products are not diluted with combustion flue gases. The combustor is fueled by an optimum ratio of recirculating, partially gasified spent shale and raw shale fines.

2.6 PRODUCT RECOVERY

Pyrolyzed shale fines are removed from the product stream by cyclones located at the top of the retort. Subsequently, the product stream is cooled to approximately 150°C and the resulting oil aerosol is trapped by electrostatic precipitation. Light oil vapor and the fluidizing steam are then condensed and separated, and any resulting mist is removed by coalescers. The remaining gas stream is sent to light oil recovery and gas separation units. Foul water from the oil/water separator is used to quench the spent shale, so there is no discharge of process water.

Optimization of operating parameters will dictate the volume and heating value of the gas from the gasification section, but it is expected that at minimum a medium-Btu gas will be produced. In addition to the synthesis gas, a significant amount of hydrocarbon gases are produced in the pyrolysis zone. The ultimate processing scheme for the various gaseous products from KENTORT II will depend on prevailing market conditions and site-specific factors. However, regardless of economic conditions,

H₂S will be scrubbed, concentrated, and routed to a Claus reactor for elemental sulfur production. Since most of the sulfur in the process is converted to H₂S in the gasifier, a much smaller amount of SO₂ is produced by the combustor for scrubbing. Due to the high ratio of H₂S to SO₂ in the process, concentrated SO₂ recovered from a scrubbing solution could be utilized as a Claus reactant for total sulfur recovery. Ammonia is also an economically important byproduct, and is recovered from the water and gas streams.

2.7 SOLID DISPOSAL

The KENTORT II process reduces the potential problems of spent shale disposal for three reasons. First, the gasification and combustion zones remove essentially all of the inorganic sulfur from the shale. This significantly reduces the acid leaching potential of the disposed material. Second, all of the mined shale is processed which means that no raw shale must be returned to the mine. Not only does this enhance process economics, but it is an environmental advantage because raw shales have been shown to produce more acid drainage than processed shales.(4) Third, the smaller particles required for fluidized bed technology are more easily compacted and are more suitable for reclamation. The mitigation of environmental and waste stream clean-up problems at the reactor design level makes the process inherently more efficient.

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3. PROCUREMENT AND PROCESSING OF THE MASTER SAMPLE

3.1 OIL SHALE SELECTION

Oil Shale from the Cleveland Member of the Ohio Shale was selected for use in the KENTORT II study based upon the judgment that it represented one of the shale types in the eastern U.S. that has the highest probability for initial commercial development. This assessment was based on two factors:

First, during the early 1980's, proposals for commercial development were submitted to the U.S. Synthetic Fuels Corporation, for projects in Fleming County, Kentucky, from the American Syn-Crude Corporation;(1) and in Montgomery County, Kentucky, from Breckinridge Minerals and Dravo Engineers, Inc.(2) The latter project met all of the technical criteria and much of the economic maturity criteria for selection before the SFC was retired. Both of these commercial projects utilized the upper portion of the Cleveland Shale as their primary feedstock, along with the overlying Sunbury Shale, which constituted approximately 1/3 of the feedstock.

Secondly, resource assessment studies have indicated that the oil shales located in East-Central Kentucky (i.e. from Fleming County south to approximately Powell County) are in an optimal configuration for commercial development.(3,4,5) Factors such as overburden configuration, total thickness and quality of the organic rich stratigraphic zone and geomorphic configuration of the deposits favor exploitation in this region.

3.2 SAMPLE SITE DESCRIPTION

3.2.1 Site Selection. The sample site selection was based on accessibility, background information on the area, and previous experience in working at the location. Oil shale from the Cleveland Member at the site had been used in a pilot plant test of Eastern oil shale at Petrobras' Petrosix Plant in Brazil and several U.S. DOE sponsored Eastern oil shale studies.(6,7,8)

3.2.2 Geographic Setting. The sample site is located in southeastern Fleming County (Figure 3-1) near the Fleming-Rowan County line. In this area the western edge of the black shale outcrop runs roughly along the county line (Figure 3-2). The Rowan-Fleming County area is one of transition from the physiographic region of rolling hills in the outer Bluegrass through the Knobs, to the broad flat-topped ridges of the Pottsville escarpment, and within a short distance to the narrow ridges and valleys of the Cumberland Plateau (Figure 3-1). The black shale outcrop begins in the outer Bluegrass and extends up into the sides of the knobby hills and ridges of the Knobs and into the ridges of the Pottsville escarpment.

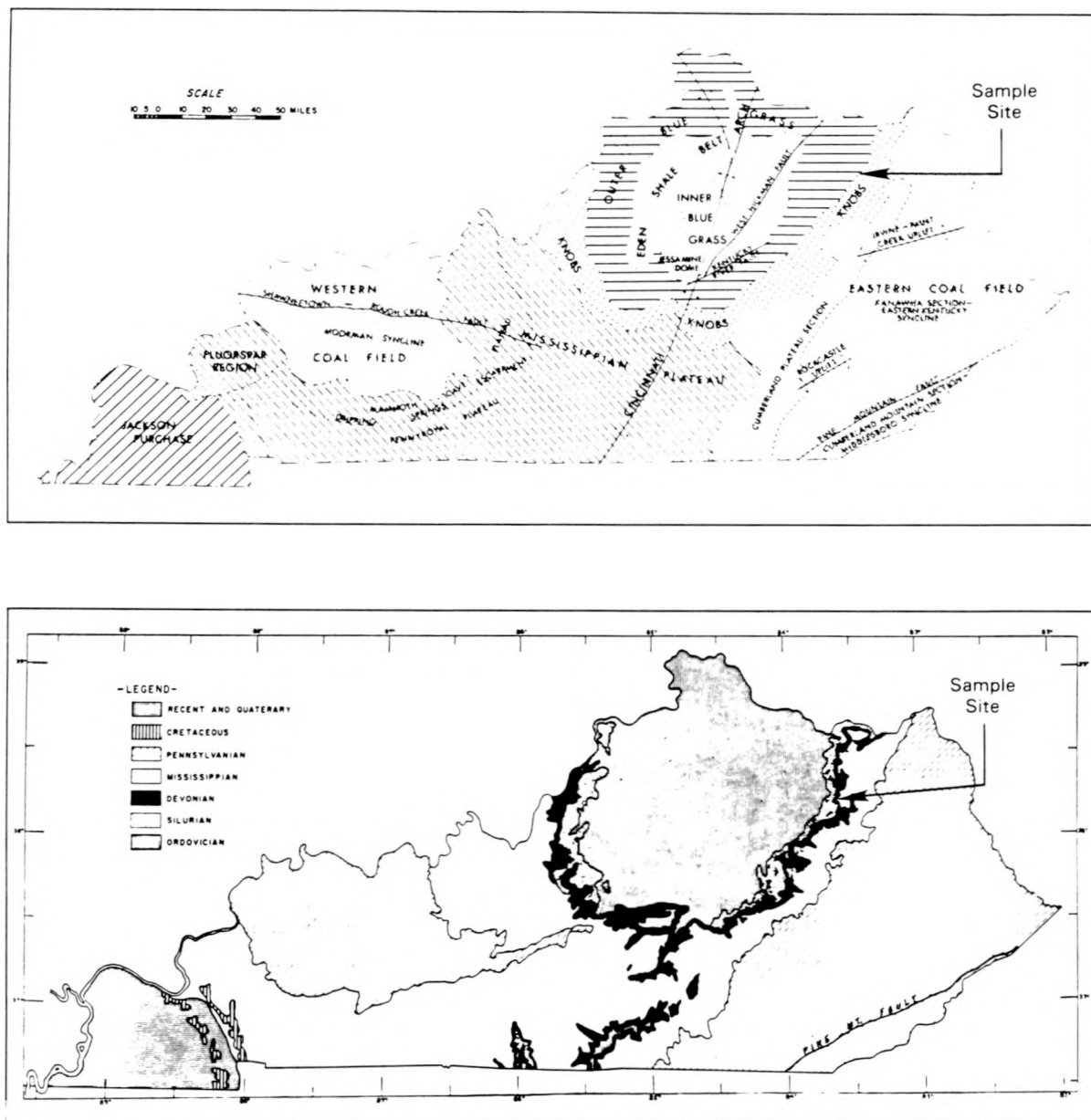


Figure 3-1. Physiographic (upper) and Geologic (lower) Maps of Kentucky.

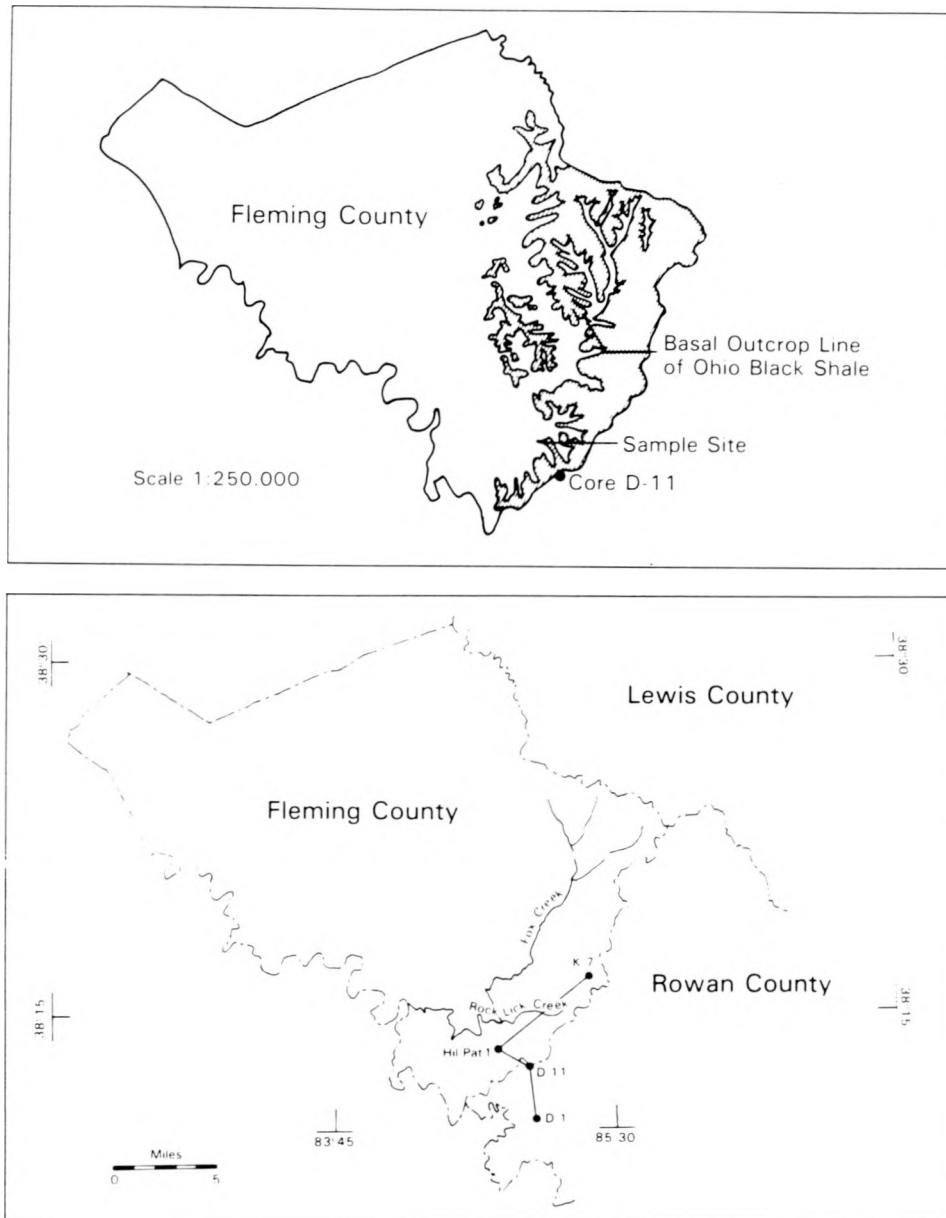


Figure 3-2. Maps of Fleming County Kentucky Showing the Distribution of Oil Shale (upper) and the Location of Cores Drilled Near the Sample Site. Core Hil-Pat #1 is at the Site.

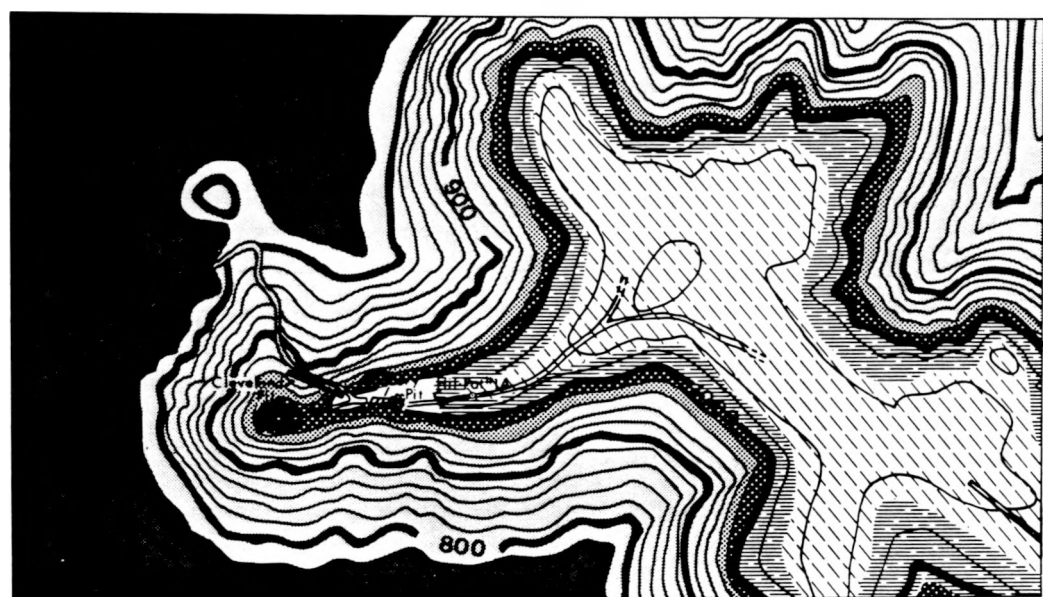
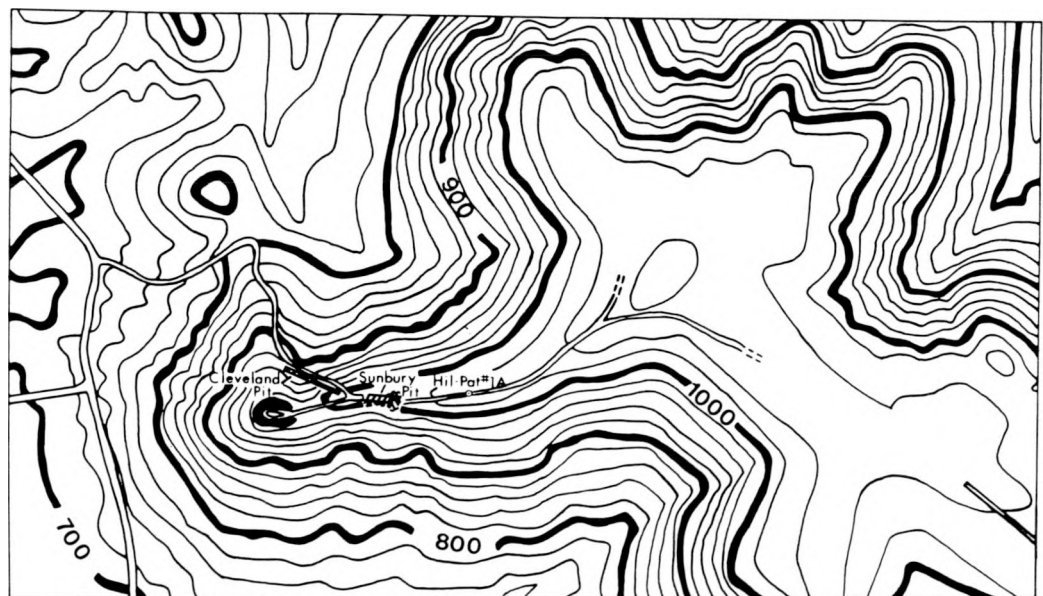
The site itself is located on the extreme western nose of a narrow ridge in the northwestern corner of the Farmers Quadrangle. The local relief between the ridge crest and the nearest drainage in Fox Creek is approximately 400 feet. The ridge top and slope are primarily forested in hard woods. Broader portions of the ridge to the southwest are used for pasture and hayfields.

3.2.3 Geologic Setting. The bedrock (Figure 3-3) exposed in Fox Creek is the Preachersville Member of the Upper Ordovician Drakes Formation, which consists of interbedded greenish-gray shale and gray-to-reddish-brown dolomite. The Silurian Brassfield Dolomite and the greenish-gray shale of the Crab Orchard Formation crop out above the Drakes and below the Ohio Shale. The latter occupies nearly all of the ridge slope. A small break in slope occurs at the position of the Bedford Shale. The Sunbury Shale overlies the Bedford and in turn is overlain by the Henley Bed and the Farmers Member of the Mississippian Borden Formation.(9) The more erosion-resistant siltstones of the Farmers cap this ridge and nearby ridges.

The oil shale overburden consists of the gray shales of the Nancy Member of the Mississippian Borden Formation and the interbedded siltstones and gray shales of the Farmers Member. The Henley Bed (a clay-shale) occurs between the base of the siltstones and the top of the Sunbury. The broad table-topped ridges to the east are also held up by the Farmers siltstones, but a series of knobs 4.5 miles east of the sample site exhibit an additional 300 feet of relief held up by the massive siltstone of the Cowbell Member of the Borden.

The definition of oil shale in the area is based on organic carbon content, with a lower limit of 8% C used as a cutoff. The Sunbury Shale and a high grade zone (HGZ) in the upper portion of the Cleveland Member of the Ohio Shale meet this criteria.

3.2.4 Previous Work. Four cores, near the sample site, have been drilled and analyzed. One of them (Hil-Pat 1, Figure 3-4) was drilled, by American Syn-Crude at the sample site itself, just over the ridge from the sample pit. The data from these cores indicate the thickness of the oil shale units and their composition to be highly consistent. The total thickness of the Cleveland Member is approximately 50 feet with ~30 feet of HGZ. Organic carbon in the Cleveland generally decreases from top to bottom (Figure 3-5), although the weighted mean for organic carbon in HGZ is very consistent among the cores (Table 3-1).



-  Nancy Member
-  Farmers Member
-  Sunbury Shale
-  Bedford Shale
-  Ohio Shale
-  Upper Part of Crab Orchard Formation

Figure 3-3. Maps of Topography (upper) and Geology (lower) of the Sample Site.

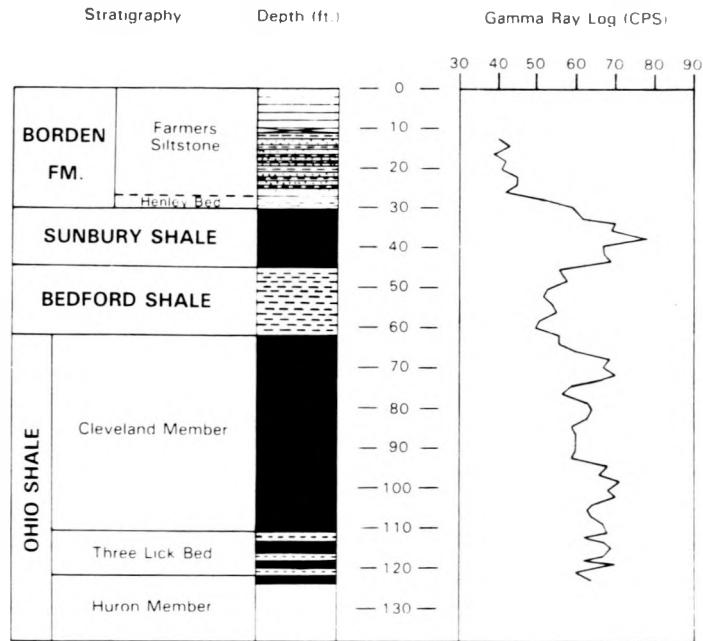


Figure 3-4. *Stratigraphic Column of the Sample Site from the Hil-Pat #1 Core.*

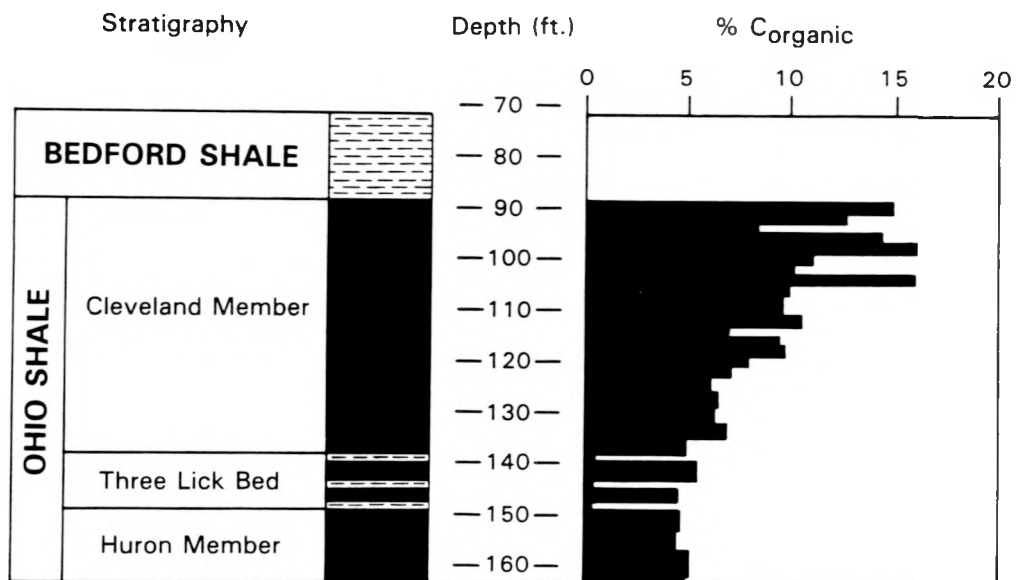


Figure 3-5. *Stratigraphic Column and Distribution of Organic Carbon for Core D-11.*

Table 3-1. *Thickness of Oil Shale Units and Associated Strata from Cores Drilled Near the KENTORT Sample Site.*

Core	Sunbury Shale	Bedford Shale	Cleveland Member	Cleveland HGZ	%C in HGZ
K-7	17.9	16.8	54.1	31.8	10.9
Hil-Pat#1	14.7	17.2	50.1	32.6	10.8
D-11	16.4	16.8	49.6	32.2	11.2
D-1	16.7	18.2	49.4	32.0	10.9

3.3 SAMPLE COLLECTION AND PROCESSING

The principal objective of the sampling operation was to collect as fresh and un-weathered a sample as possible. The Cleveland was collected from the side of the ridge along the farm road leading up the ridge at a slope of approximately 10 degrees (Figure 3-3). A tracked front-end loader was used to remove approximately six-feet of overburden and spoil material that had been used to reclaim the site in 1983. Approximately two to four feet of weathered oil shale was removed in order to expose fresh shale. Due to the nature of the cut, an unweathered composite section of the entire Cleveland HGZ was not obtainable. The fresh shale was ripped up with the bucket of the loader and dumped into a 5-ton truck and transported to a covered pad at the CAER. Approximately 10 to 12 tons of shale were recovered. The sample cut was backfilled and the site regraded to its original condition.

Although care was taken in sample mining and loading, the sample pit was narrow and some contamination occurred. As a consequence the entire 10-ton sample was hand picked to remove weathered shale, soil, and overburden. Large pieces of oil shale were broken to approximately six-inch top size with a sledge hammer. The shale was piled and repiled several times with a front-end loader to achieve good mixing. All of the shale was then loaded into twenty 55-gallon drums. Sixteen drums of the broken shale were fed through a jaw crusher producing an approximately three-quarter-inch product. The minus-three-quarter-inch product was fed through a smaller jaw crusher to produce minus-one-eighth inch material.

The product from the jaw crusher was then initially screened using U.S. No. 20 and 60 screens. The fines were saved and the top size was recycled to the small jaw crusher and/or a "coffee

mill" grinder for finer grinding. The final result was five 55-gallon drums of finished product at 20x60-mesh, stored in permalon drum liners in steel drums; fifty pounds of 18x20-mesh shale screened from the top size material; and fifty pounds of 60x80-mesh shale screened from the fines. These products were stored under argon. The remainder of the top size and fines were stored in 55-gallon drums.

Characterization of the 20x60 mesh shale sample is shown in Table 3-2. The Fischer assay oil yield and the elemental composition are within the range of what has been previously found for other shales from the Cleveland HGZ from this region of Kentucky.

Table 3-2. *KENTORT II Master Sample Characteristics.*

				Modified Fischer Assay*	
Carbon	16.7	wt%		Oil	17.7 gal/ton
Hydrogen	2.0	wt%		Oil	7.0 wt%
Nitrogen	0.7	wt%		Water	4.2 wt%
Sulfur	1.7	wt%		Gas+Loss	2.6 wt%
Ash	74.5	wt%			

* Modified Fischer Assay developed at CAER for Eastern shale. Coburn, T.T., A.M. Rubel, Fuel Processing Technology, 8, 1, 1983.

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4. SHALE CHAR GASIFICATION KINETICS

4.1 OBJECTIVES

The characterization of eastern U.S. oil shale char gasification kinetics was essential to the design of the integrated pyrolysis/gasification reactor system for the KENTORT II mini-plant. A 1.5-inch diameter bench-scale fluidized bed reactor was used to investigate fluid-bed pyrolysis char gasification under conditions applicable to the process. The parameters studied included: bed temperature, mean solid residence time (MSRT), and fluidizing gas composition. Of particular importance was the understanding of residual carbon and sulfur removal from the char during the gasification.

4.2 METHODOLOGY

4.2.1 Apparatus. A 1.5-inch diameter bench-scale fluidized bed reactor (described in detail elsewhere) was used for the study (Figure 4-1).(1) The reactor, constructed of 316 stainless steel, was contained within a split tube furnace. A 100 micron stainless steel filter was used as a distributor plate. Char height in the bed was controlled by a stand-pipe which was baffled to prevent solids from short-circuiting. The fluidizing gas was preheated in heat exchangers filled with 0.25-inch diameter steel balls located within furnaces. The spent shale was collected in a bin consisting of two chambers which allowed the separation of gasified char produced during start-up from that produced during steady state. A gas sampling train consisting of two 10-liter acrylic gas collection syringes connected in parallel to a bellows pump was used to collect the entire gas volume leaving the reactor. The bellows pump prevented back pressure to the reactor when switching from one cylinder to the next.

4.2.2 Materials Used. The shale char used for the gasification study was produced from an 18 x 20 mesh raw shale master sample (designated as CLE82C-S29) prepared from a bulk sample of the Cleveland Member of the Ohio Shale mined in northeastern Kentucky. The prepared raw sample was stored under inert gas until pyrolyzed for this study. Char was produced from an 8 kg sample in one run in a 3-inch diameter fluidized reactor (2) under the following pyrolysis conditions: mean bed temperature = $535 \pm 4^{\circ}\text{C}$, mean shale feed rate = 26.2 g/minute, mean solid residence time (MSRT) = 6.9 ± 0.8 minutes, fluidizing medium = N_2 , and mean superficial fluidization velocity = 42.6 m/minute. These conditions have been reported to be near optimum for maximum oil yields from eastern U.S. oil shales.(3) Organic carbon and sulfur removed from the shale by pyrolysis was 53.9% and 29.8%, respectively (Table 4-1). The char was blended and

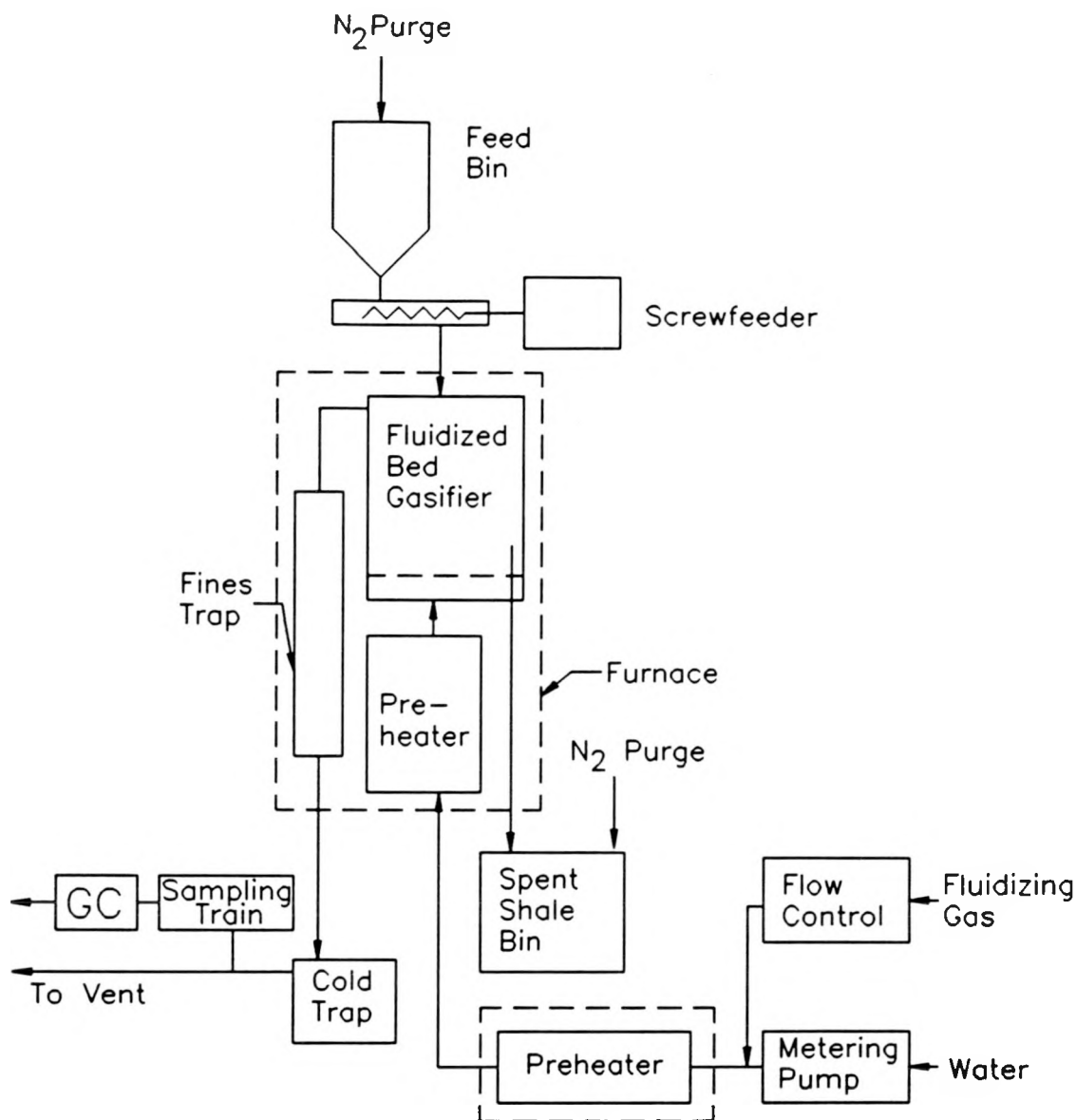


Figure 4-1. *Schematic of Fluidized Bed Bench-Scale Gasifier.*

riffled into 100 g aliquots and stored in airtight jars under argon until used for gasification runs.

4.2.3 Gasification Procedures. The entire apparatus was brought to operating temperature before each run with N₂ as a purge gas. Once steady temperatures were achieved, a weighed amount of char was placed in the feed bin and the flow of the appropriate fluidizing gas (steam, N₂/steam, or CO₂/steam) was started. After temperature equilibration, char feed was initiated and the product gas diverted from the vent to the collection train.

Table 4-1. *Ultimate and Proximate Analyses for Shale Materials.*

	Raw Shale	Pyrolysis Char	Gasified Char
%C	12.1	6.2	3.8
%H	1.5	0.4	0.1
%N	0.6	0.4	0.3
%S	2.2	1.9	0.6
%Mois.	1.2	0.2	0.0
%VM	13.4	4.2	0.9
%FC	4.6	5.3	3.0
%Ash	80.7	90.3	96.1
C/H	0.7	1.3	4.5
C/N	22.0	18.6	14.2
C/S	14.6	8.7	16.8

Steady state bed temperature was typically achieved 2 to 3 minutes after the char feed was initiated. Ten minutes after steady state was reached, the gasified char was collected in a separate portion of the overflow bin. Steady state run duration ranged from 1 to 2 hours in which 50 to 100 g of char were gasified. Bed temperature could typically be held within $\pm 5^{\circ}\text{C}$ of the set point.

During a run, the entire gas stream was collected and measured by plunger displacement in 10-liter gas collection syringes. Depending upon residence time, between 6 and 26 cylinders were filled. The GC required 30 minutes to cycle and typically 6 to 8 cylinders could be analyzed during and shortly after a run. Dry gas compositions were calculated as the volume-weighted average of GC determinations over the run.

All gasification runs were at atmospheric pressure and a mean fluidization velocity of 31.2 m/s, which was 140% of minimum for the particle size range. The experimental matrix included bed temperature, MSRT, and fluidizing gas composition (Tables 4-2 and 4-3) as variables.

4.2.4 Analytical Procedures. The raw and spent chars from each gasification run were analyzed with LECO equipment, using standard procedures, for carbon, hydrogen, nitrogen, and sulfur. Proximate analysis was performed in duplicate by thermogravimetric analysis (TGA) on a Perkin-Elmer TGS-2.(4) Gases were analyzed with a Hache-Carle gas chromatograph (GC) using Hache application number 158-B. Gas volume measurements

Table 4-2. *Experimental Matrix for 100% Steam Fluidized Gasification Runs.*

Fluidizing gas = 100 kPa steam Fluidizing gas velocity = 31.2 m/min Pressure = 101 kPa		
Bed Temp, C	MSRT, min*	
700	10	30
750	10	30
800	10	30
850	10	30

*Char residence time was varied by keeping the bed height constant and adjusting the char feed rate from 0.4 to 2.6 g/min to attain the desired residence time.

and dried gas compositions were corrected for N₂ and O₂ (purge gas and air contamination). The heating value of the product gas was calculated using ASTM D3588-81.

The CHN analyses of the char were reproducible within $\pm 0.04\%$, ± 0.02 , and ± 0.03 weight percent for C, H, and N, respectively. The determination of total sulfur was somewhat lower in precision, since replicate determinations were within ± 0.2 weight percent. Because the analysis of the solids was both more precise and probably more accurate than the gases, which were highly diluted by the fluidizing media, it was used as the primarily basis for data interpretation.

Table 4-3. *Experimental Matrix to Study the Effect of CO₂ and Steam Partial Pressure on Char Gasification.*

MSRT = 30 min Fluidizing gas velocity = 31.2 m/min Pressure = 101 kPa	
Fluidizing Gas	Bed Temp, C
50:50 N ₂ :Steam	750 850
50:50 CO ₂ :Steam*	750 850

*Actual reactant gas composition adjusted for feed and spent overflow tube purges was 47:3:50 for CO₂:N₂:Steam.

Standard deviation for CO, CH₄, CO₂ and H₂ analysis was typically $\pm 5.0\%$ of the mean value. In general the precision of the gas determinations improved at higher temperatures with increased gas production. Considering that the carbon-bearing gases were diluted by a factor of between 16 and 500 to one, this was considered to be good precision for this type of analysis. The sulfur gas analysis (H₂S) was much lower in precision compared to the other gas components with standard deviations typically 12 to 19% of the mean value.

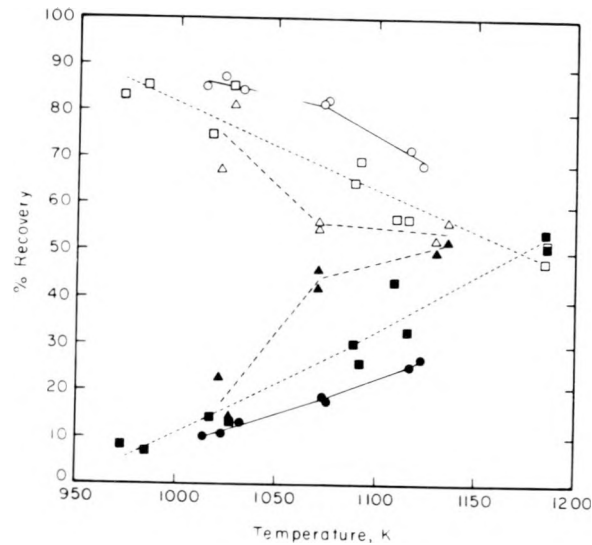
Total gas phase carbon was calculated by summing CO, CH₄, and CO₂. Although the GC was set up to monitor ethane, ethylene and propane, none were measured above detection limits (0.02% by volume). Hydrogen sulfide was the only sulfur gas detected.

4.3 RESULTS AND DISCUSSION

4.3.1 Run Balances. The mean and standard deviation for the material, carbon, and sulfur balances for the 42 gasifier runs were $101.9 \pm 2.2\%$, $96.7 \pm 4.8\%$, and $99.3 \pm 5.7\%$ respectively. These balances indicated the experimental and analytical procedures were adequately accounting for the gasification products. The balance components consisted entirely of the char and gases as proportionation to the steam condensate was negligible. For example, total organic carbon in the condensate ranged from 9 to 110 ppm, which accounts for only 0.2 to 0.4% of the starting char carbon. The amount of sulfur in the water was not determined, however, sulfur removed from the shale was nearly balanced by H₂S detected in the gas.

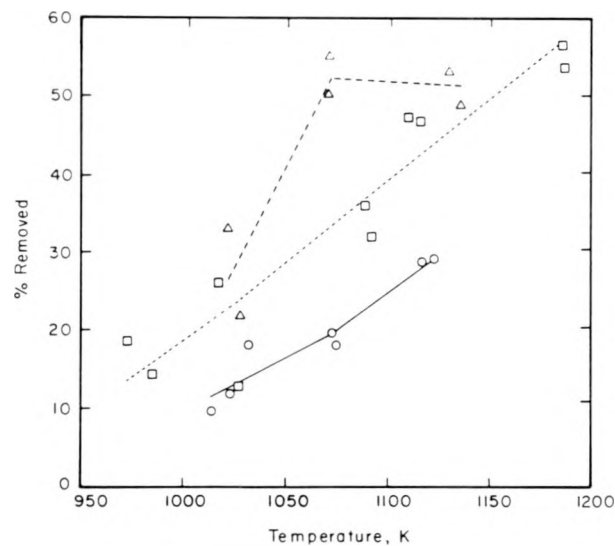
4.3.2 Steam Gasification: Carbon and Sulfur Distributions. Gasification was measured by the removal of total and fixed carbon from the char and by carbon bearing-gas production. The measurement of gasification by carbon oxides production is possible for eastern U.S. shales because of the lack of carbonate minerals. Total carbon oxides production corresponded to the carbon removed from the char in terms of time-temperature dependence (Figure 4-2). Little gasification was observed at 700°C. At 60 minutes MSRT, the increase in total carbon removal was marginal between 800°C and 850°C. Since total carbon conversion does not exclude organic carbon removal due to any additional primary or secondary pyrolysis, proximate analysis was used to monitor the conversion of fixed carbon by the carbon gasification reactions (Eqs. 1-3).





○— 10 min MSRT; □..... 30 min MSRT; Δ--- 60 min MSRT; open symbols = spent char; closed symbols = gas.

Figure 4-2. *Recovery of Total Carbon in the Spent Char and Gas as a Function of Bed Temperature and MSRT for 100 Percent Steam Fluidization.*

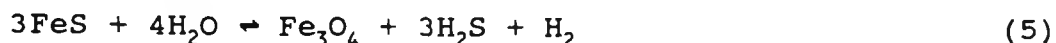


○— 10 min MSRT; □..... 30 min MSRT; Δ--- 60 min MSRT.

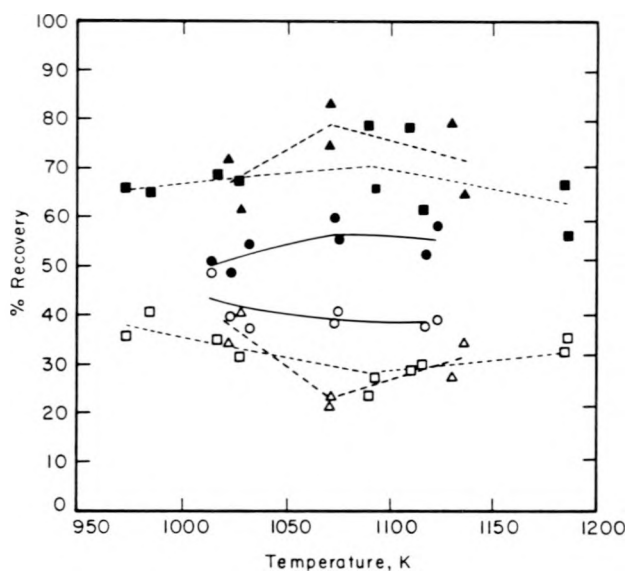
Figure 4-3. *Fixed Carbon Removed from Shale Char by Steam Gasification.*

The removal of fixed carbon from the char (Figure 4-3) was essentially identical in terms of the time-temperature relationship to the removal of total carbon based on elemental analysis.

Sulfur removed from the char (Figure 4-4) was balanced by the production of H_2S in the gas stream. Since 80-90% of the total sulfur in eastern U.S. shale is associated with iron sulfides, the use of total sulfur analyses can be appropriately utilized to monitor FeS_2 chemistry. Therefore, both sulfur removal from the char and H_2S concentration in the gas can be used to measure the decomposition and conversion of pyrite to iron oxides (Equations 4 and 5).



4.3.3 Total Carbon and Sulfur Conversion Rates. While the bench-scale reactor is not particularly suited for the determination of rate constants, pseudo first-order expressions were applied to the data to allow comparisons to be made with values for similar shales available in the literature. From previous solids mixing studies in this reactor, a perfectly mixed residence time distribution is a valid assumption. Since the bed of solids is not microscopically homogenous and is more akin to a



○— 10 min MSRT; □----- 30 min MSRT; △--- 60 min MSRT; open symbols = spent char; closed symbols = gas.

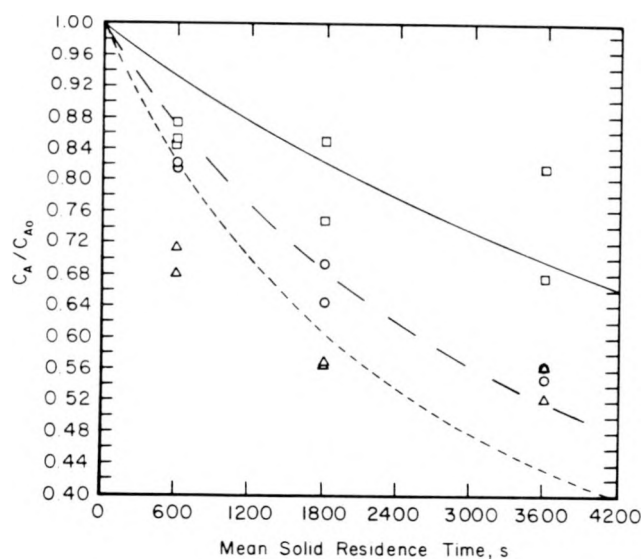
Figure 4-4. Recovery of Sulfur in the Spent Char and Gas as a Function of Bed Temperature and MSRT for 100 Percent Steam Fluidization.

macrofluid, the residence time distribution must be incorporated into the design equation for a perfectly mixed reactor.(5) For a first-order reaction the expressions for a well-mixed microfluid and macrofluid are relatively simple and equivalent:

$$C_A/C_{A0} = 1/(1 + kt) \quad (6)$$

where k is the first order rate constant, t is the MSRT, C_{A0} is the carbon content of the pyrolyzed shale and C_A is the carbon content at t .

The rate of carbon gasification was assumed to be first-order with respect to the fraction of carbon remaining in the shale. First-order kinetics were applied to all three measures of carbon gasification in the study and the goodness-of-fit was only fair in each case. The mean sum of squares of the residuals ranged from 0.0063 to 0.0116 for the data in Figure 4-5. It is apparent that first-order kinetics does not precisely fit the data. Conversion is under-predicted at short residence times, and over-predicted at the most lengthy residence times. Application of other rate expressions was not attempted because they become very complex for a mixed reactor and this type of analysis was beyond the scope of this work. All three measures of carbon gasification responded similarly to the model. The average rate constant for 100 kPa steam at 850°C was $3.8E-4 \text{ s}^{-1}$. A similar rate constant ($3.1E-4 \text{ s}^{-1}$) was determined during the



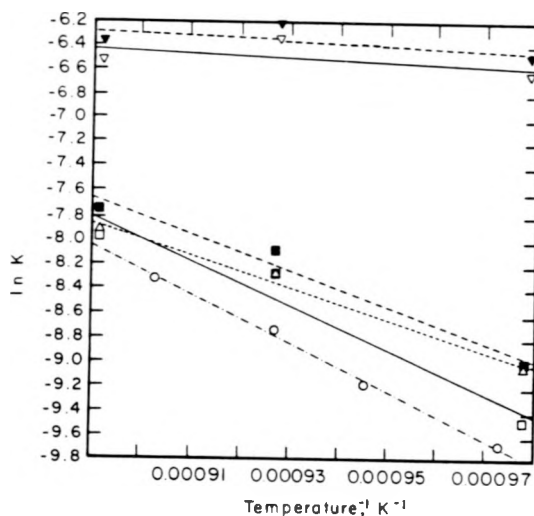
□ — 750 °C; ○ — 800 °C; Δ --- 850 °C

Figure 4-5. Determination of First Order Rate Constants for Shale Char Gasification on Total Carbon Removal Basis.

mini-plant pyrolysis/gasification studies using the same shale and reactor conditions.(6) These values compare well to the rate constant reported by VanTuyt and Thomson of $6.6\text{E-}4 \text{ s}^{-1}$ for Michigan Antrim oil shale char at 850°C and 76 Kpa steam.(7) An Arrhenius plot of the derived rate constants (Figure 4-6) indicates a dependence on the method in which the extent of gasification was determined. However, the range of activation energies (106-148.5 kJ/mol) is in line with the values that have been reported for steam gasification of various carbonaceous materials. A slightly higher activation energy (160.8 kJ/mol) was obtained from the mini-plant integrated pyrolysis/gasification runs.(6)

First-order kinetics were also applied to the removal of sulfur with similar success. Like carbon gasification, the experimental results deviated from first-order behavior at the short and lengthy residence times. An Arrhenius plot of the sulfur removal rate constants (Figure 4-6) indicates that pyrite conversion occurs much more quickly than carbon gasification. Since little temperature dependence was observed for sulfur removal, it appears to be a diffusion limited process.

4.3.4 Steam Gasification: Gas Composition. Hydrogen (Figure 4-7) was the major constituent of the gas. Under nearly all conditions the molar ratio of $\text{H}_2:\text{CO}:\text{CO}_2$ remained constant at approximately 3.7:1:1. The ratio of H_2 to the carbon oxides was somewhat high for H_2 to be generated solely from the steam-carbon



■--- fixed carbon; ▲..... carbon in spent char; □—— carbon in gas; ○--- carbon in spent from pilot-scale integrated pyrolysis/ gasification; ▼--- sulfur in spent char; ▼—— sulfur in gas.

Figure 4-6. Arrhenius Plot for Gasification Data.

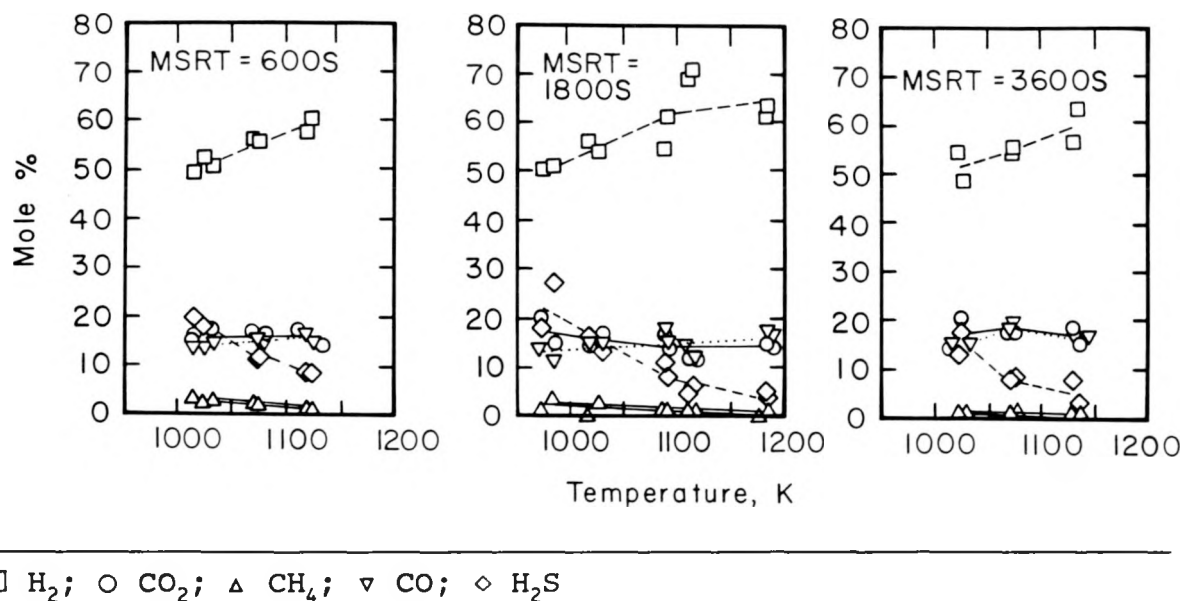


Figure 4-7. Composition of Offgas for Steam Fluidization.

and the water gas shift reactions (Equations 1 and 2). Stoichiometrically, a 1:1 ratio of CO to CO_2 would produce 3 H_2 molecules. The additional H_2 is likely the result of high temperature coking and charring in the gasifier and from steam mediated pyrite and pyrrhotite decomposition as described by Levy and White.(8) Mössbauer spectra of the feed char indicated a pyrite to pyrrhotite ratio of 2:1. Gasified chars had an abundance of magnetite (Fe_3O_4), little pyrrhotite (FeS), and no pyrite or marcasite (FeS_2). (9) Since Fe_3O_4 appears to be the predominant iron oxide resulting from pyrrhotite conversion, a net production of H_2 can be expected.

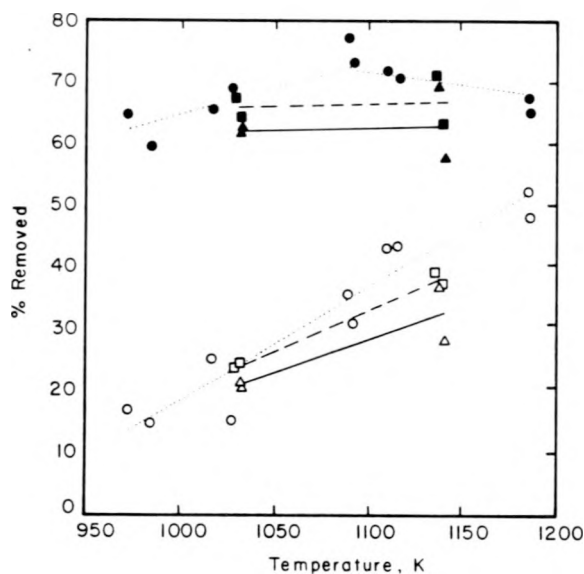
The amount of H_2S produced in proportion to the gasifier feed remained nearly constant throughout the test matrix. Therefore, its concentration declined at higher temperature and MSRT as a result of dilution from increased carbon gasification products, CO , H_2 , and CO_2 (Figure 4-7). The small quantities of methane detected indicated that methanation of carbon can be discounted as a significant reaction. Higher pressure is normally required for methanation to occur. The methane observed is probably the result of high temperature pyrolytic removal of methyl groups in the char.(10)

Total gas production increased with bed temperature and was maximum (200 l_{stp}/kg of char, 6400 $ft^3/short\ ton$) at 850°C and MSRT above 30 minutes. The calculated heating value of the gas decreased with both bed temperature and MSRT. The heating value of the offgas for the entire test matrix was 11-14 MJ/m^3_{stp} (308-

373 Btu/ft³_{stp}. At 850°C and MSRT above 30 minutes, 2000 MJ/metric tonne of char (1.7E6 Btu/short ton) were produced.

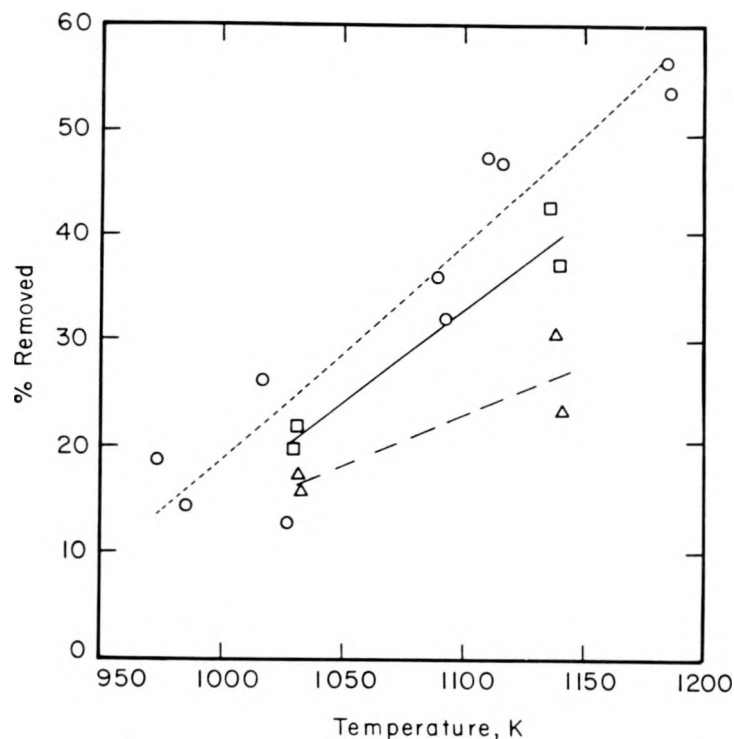
4.3.5 Steam/N₂ and Steam/CO₂ Gasification: Carbon and Sulfur Conversions. Steam gasification removed more carbon from the char than either steam/N₂ or steam/CO₂ gasification at 850°C (Figure 4-8). Based on total carbon determination, there was not a statistical difference (at the 0.05 significance level for the Student's t-test) between the steam/N₂ or the steam/CO₂ runs at 850°C. However, fixed carbon analysis (Figure 4-9), indicated that steam/N₂ had a statistically significant higher degree of conversion compared to steam/CO₂ gasification. The conversion of sulfur did not show a significant dependance on either the steam or CO₂ partial pressures of the fluidizing medium.

4.3.6 Steam/N₂ and steam/CO₂ gasification: Gas composition. Altering the steam partial pressure of the fluidizing gas changed the ratio of CO to CO₂ in the offgas (Figure 4-10). The measured gas composition indicated that the water-gas shift reaction was not at equilibrium at reactor conditions. However, at equilibrium an increase in the ratio is predicted. For steam only runs, CO and CO₂ were consistently produced in a 1:1 ratio (Figure 4-7) whereas decreasing the steam partial pressure by one half resulted in an increase in the ratio with temperature to 2:1 at 850°C. Identical to steam only fluidization, the



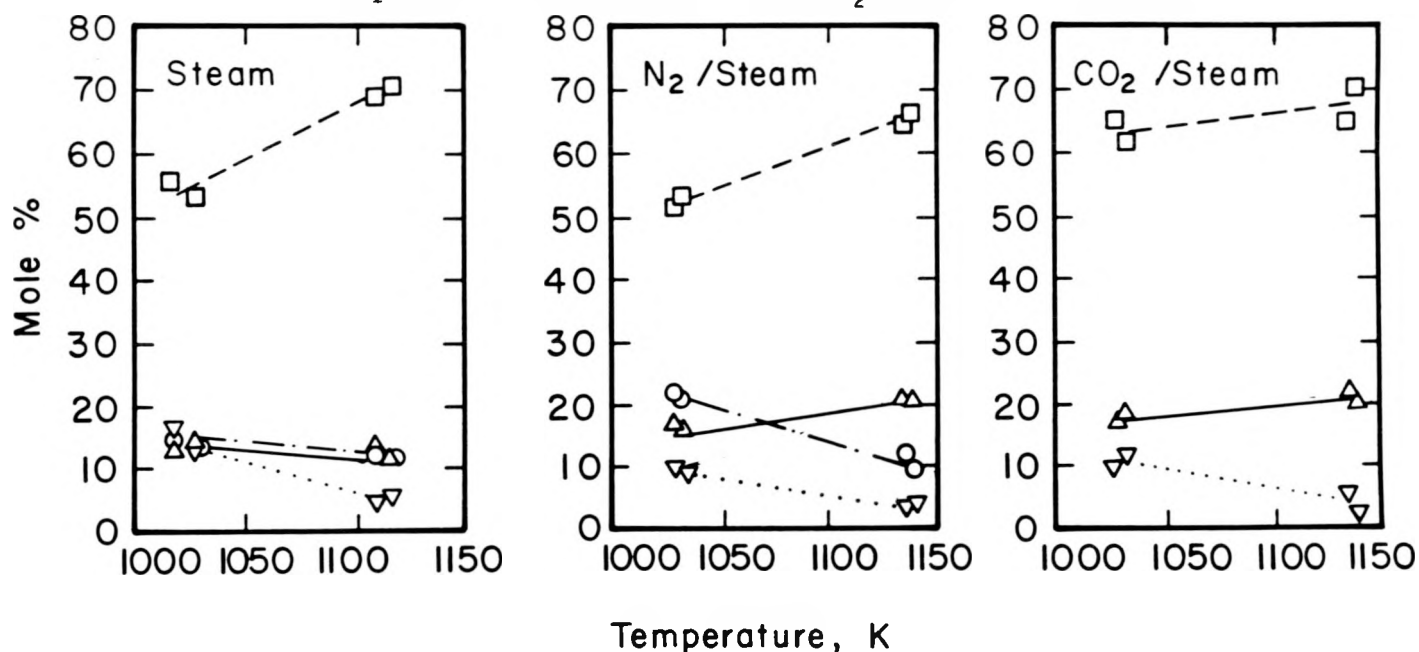
○..... steam; □- - - N₂/steam; ▲—— CO₂/steam; open symbols = carbon; closed symbols = sulfur.

Figure 4-8. Total Carbon and Sulfur Conversion as a Function of Bed Temperature and Steam and CO₂ Partial Pressure.



○..... steam; □—— N₂/steam; △—— — CO₂/steam.

Figure 4-9. Fixed Carbon Removed from Shale Char vs. Bed Temperature and Steam and CO₂ Partial Pressure.



□ H₂; ○ CO₂; △ CO; ▽ H₂S

Figure 4-10. Offgas Composition of Steam, Steam/N₂, and Steam/CO₂ Gasification of Shale Char.

concentration of H_2S in the offgas decreased as the rate of carbon gasification increased at $850^\circ C$.

The reduction in carbon conversion observed with reduced steam partial pressure was reflected in a statistically significant decrease in the amount of offgas produced. At a gasifier bed temperature of $850^\circ C$, total gas production decreased from 200 to $140\text{ m}^3_{\text{stp}}/\text{kg char}$.

4.4 CONCLUSIONS

The gas composition indicated that neither the steam-char nor the water-gas shift reactions were at equilibrium at reactor conditions. At atmospheric pressure, H_2 and CO are the thermodynamically favored products in an equilibrium steam-carbon system at the temperature range of this study. The high proportion of steam to carbon (and its products) forced the water-gas shift reaction to the right causing a relatively low $CO:CO_2$ ratio of 1:1 for the 100 kPa steam experiments. Additionally, an increase in this ratio to 2:1 for the 50 kPa steam experiments at $850^\circ C$ is consistent with this hypothesis.

The dominant primary gasification reaction was the steam char reaction. Since carbon conversion decreased with increased CO_2 partial pressure, the Boudouard reaction (Equation 3) must not proceed quickly under the conditions studied. High CO_2 levels coupled with decreased steam partial pressure inhibited the water/gas shift reaction (Equation 2) resulting in the higher CO levels observed which in turn has the potential of decreasing the steam/char reaction rate (Equation 1). Sulfur reactions did not exhibit a strong dependence on steam partial pressure. This is probably due to the fact that the initial removal of sulfur from the char is the result of thermal decomposition (Equation 4) which is not dependent on steam partial pressure. The subsequent conversion of pyrrhotite to magnetite (Equation 5) may also be less dependent on steam partial pressure since the lack of significant temperature effects suggest that this reaction might be diffusion and not chemical reaction limited.

The carbon and sulfur present in Kentucky oil shale fluid-bed pyrolysis char was found to be sufficiently reactive in steam to make the gasification step of the KENTORT II Process feasible. Carbon reactions were dependent on bed temperature, steam and CO_2 partial pressure of the fluidizing medium, and were not complete at the longest MSRT used (60 min). Sulfur reactions, on the other hand, exhibited little temperature dependence and occurred more quickly. The differences in the rates of carbon and sulfur removal have important implications to the development of the KENTORT II Process. By exploiting the rapid removal of sulfur in the intermediate gasification step prior to combustion, the production of SO_2 would be lessened in exchange for higher H_2S production. In this manner, an efficient method of removing

sulfur from the gas stream is achieved, from which elemental sulfur can be produced and sold to defray some of the clean-up costs. During the bench-scale work, conversions of 80-85% sulfur and 70-75% carbon (raw shale basis) for pyrolysis/ gasification were achieved. These conversions would produce high liquid yields (1,11,12,13); 200 m³_{stp} of a H₂ rich gas per metric tonne of char to use for heavy shale oil upgrading; a combustor feedstock with sufficient carbon for process heat but with a much improved C/S ratio; and removal of sulfur as H₂S for elemental sulfur production.

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5. OIL SHALE MATERIALS COMBUSTION KINETICS

5.1 OBJECTIVES

The investigation of the combustion kinetics of raw and processed shales in the 1.5-inch fluidized bed reactor was undertaken to provide design data and a range of operating conditions for the KENTORT II combustion section. The variables identified as important were: feedstock, bed temperature, and MSRT. The emphasis of the study was the determination of the carbon combustion efficiency of various blends of raw and processed shales.

5.2 METHODOLOGY

5.2.1 Apparatus. The reactor used for the gasification kinetics study previously discussed was modified for fluidization by a mixture of N_2 and O_2 , 75/25% (Figure 5-1). This required the addition of a mass flow controlling/gas blending system. The blended fluidizing gas was passed through two heat exchangers to bring it to the targeted temperature before entering the bottom of the reactor below the distributor plate. Thermocouples at three heights within the reactor monitored combustion temperature. Product gas left the combustor at the top of the reactor, passed into a gas cooling system, and was sampled by on-line gas analyzers.

5.2.2 Materials Used. The feedstocks for the combustion study came from two similar raw shale samples of the Cleveland Member of the Ohio Shale (CLE 003 and CLE S29) mined in Fleming County, Kentucky. Four neat feedstocks (Table 5-1) and one blend were produced from these shales. The feeds used were raw CLE S29 (18x20 mesh), raw CLE 003 (20x60 mesh), pyrolyzed/gasified CLE003, pyrolyzed CLE S29, and a 50:50 blend of raw CLE 003 and gasified/pyrolyzed CLE 003. The pyrolyzed/gasified feedstock, which will be referred to as gasified shale, was produced in the KENTORT II pyrolysis-gasification system. Pyrolysis and gasification conditions were 530°C and 10 minutes MSRT and 837°C and 30 minutes MSRT, respectively. Under these conditions, 78 percent of the starting carbon and 86 percent of the sulfur were removed from the shale (Table 5-1). The pyrolyzed sample was produced in the single-stage, 3-inch diameter fluidized bed pyrolyzer from CLE S29. The pyrolysis conditions were 535°C and 7 minutes MSRT using N_2 as the fluidizing gas. Pyrolysis alone removed 54 percent of the carbon and 22 percent of the sulfur (Table 5-1). Both the gasified and the pyrolyzed feedstocks were produced under pyrolysis conditions which were near optimum for enhanced liquid yields from eastern U.S. oil shales.

5.2.3 Operating Procedure. The combustion system was preheated to temperature with purge gas (N_2) flowing through the

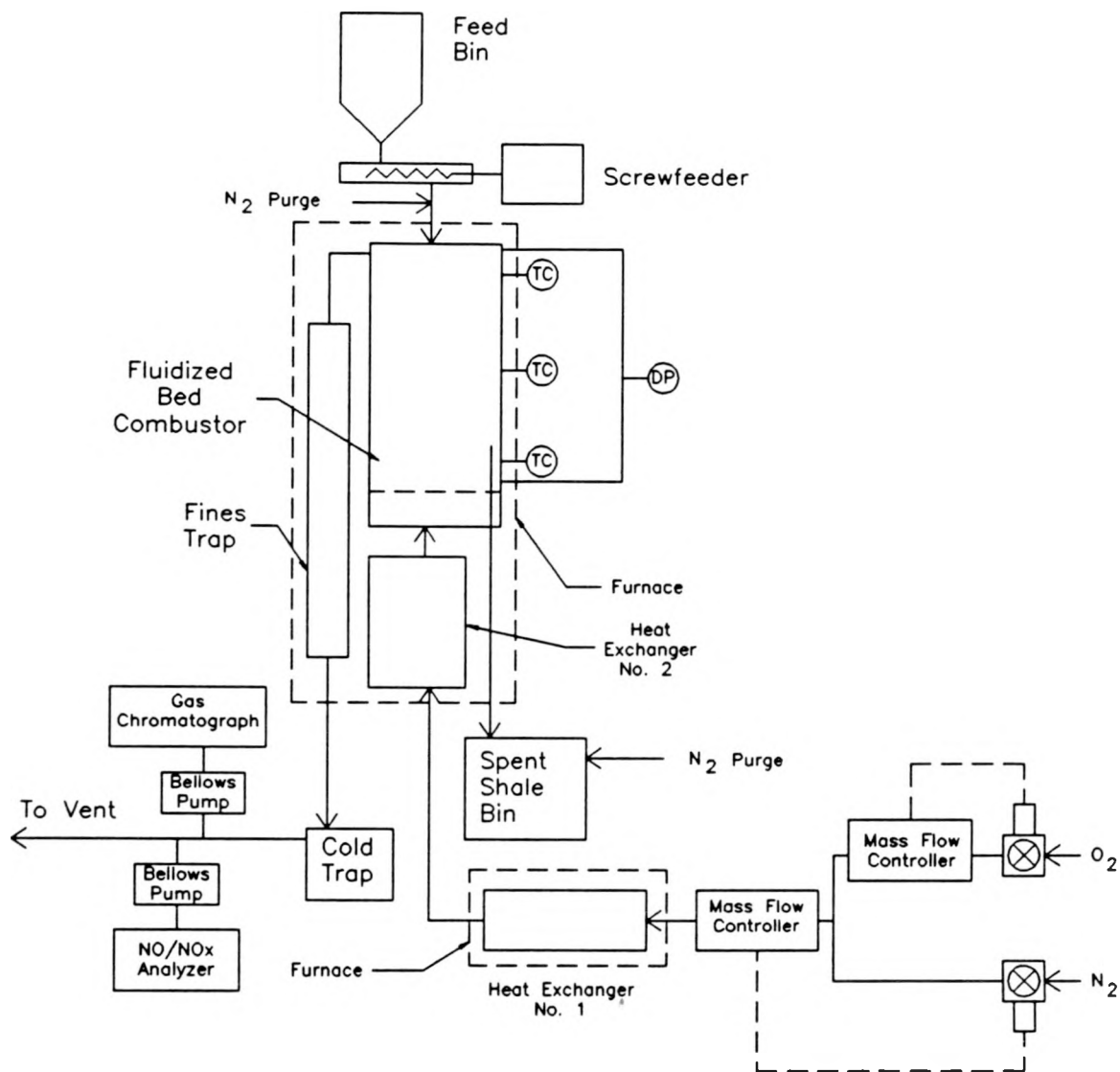


Figure 5-1. *Schematic of Fluidized Bed Bench-Scale Combustor.*

feed and spent overflow tubes. Once steady temperatures were achieved, a weighed amount of feed material was placed in the feed bin and a fluidizing gas velocity of 150 percent of U_{mf} was established. Again, temperatures were allowed to stabilize and a sample of the fluidizing gas was pumped to a gas collection syringe for post run analysis. For all feeds except the raw shales, the screw feeder was turned on and steady state temperatures were typically obtained within 2 minutes. For raw feeds, an intermittent startup procedure was followed for the first two minutes with the screw feeder on for 15 seconds and off for 15 seconds. This prevented a start up temperature spike which if allowed sintered the bed. Steady state run duration

Table 5-1. *Elemental Composition of Feedstocks Used for Combustion Study.*

	% C	% H	% N	% S
RAW CLE 003	16.7	2.0	0.7	1.7
GASIFIED 003	4.6	0.1	0.5	0.3
RAW CLE S29	12.1	1.5	0.6	2.2
PYROLYZED S29	6.2	0.4	0.4	1.9

ranged from 15 to 120 minutes during which 100 to 200g of material were combusted. Run length was decreased as MSRT was decreased. Since solids holding time could only be decreased by increasing the feed rate, the number of times the bed was displaced was greater for shorter MSRT runs in spite of their short duration (3 to 4 turnovers for 30 minutes MSRT vs. 8 turnovers for 2 minutes MSRT runs). The first gas sample was taken downstream from the ice trap 10 minutes from start-up and subsequent samples were analyzed every 27 minutes. For shorter duration experiments, combustion gases were analyzed on-line at 10 minutes and then collected for post-run analysis.

5.2.4 Analytical Procedures. The feed and combusted materials from each run were subjected to carbon, hydrogen, nitrogen, sulfur, and proximate analyses by procedures described in the gasification kinetics section. Selected feed and combusted materials were analyzed for total nitrogen by a micro-Kjeldahl procedure (1) and for non-exchangeable NH_4^+ by the method of Silva and Bremner.(2) Combustion gas composition (CO , CO_2 and $\text{C}_1\text{-C}_3$ hydrocarbon gases) was determined by gas chromatography as previously described. Gas compositions were corrected for N_2 and O_2 . NO_2 and SO_2 emissions were calculated from the amount of nitrogen and sulfur removed from the feed.

Iron distribution was determined by Mössbauer spectrometry on representative solids including magnetically separated sinters from problem runs. Mössbauer analyses were performed on Plexiglas mounted solids using a constant acceleration spectrometer. Velocity calibrations were carried out with iron foil; isomer shifts were measured with respect to metallic Fe at room temperature. The spectra obtained with each fit were a sum of Lorentzians. The Mössbauer parameters determined from the fits include the isomer shift, quadrupole splittings, magnetic hyperfine field, and relative percentage of different Fe species.

Table 5-2. *Experimental Matrix for Combustion of Gasified Feed.*

Fluidizing gas = N ₂ /O ₂ mixture (75/25) Fluidizing gas velocity = 33.4 m/min Pressure = 101 kPa			
Bed Temp, C	MSRT, min*		
700	2	10	30
800	2	10	30
900	2	10	30

*Solid residence time was varied by keeping the bed height constant and adjusting the feed rate from 0.9 to 13 g/min to attain the desired residence time.

Table 5-3. *Experimental Matrix for Combustion of Raw and Blended Feeds.*

Fluidizing gas = N ₂ /O ₂ mixture (75/25) Fluidizing gas velocity = 33.4 m/min Pressure = 101 kPa Feed rate = 2.6 g/min	
Bed Temp, C	MSRT, min
700	10
850	10

Table 5-4. *Average Wt% Recovered in Spent and Gas.*

Feed	Spent	Gas
Gasified	95.9	4.3
Gasified + Raw	85.1	14.8
Raw	76.1	23.9

5.2.5 Test Matrix. Combustion kinetics experiments were conducted within a three parameter matrix comprised of bed temperature, MSRT, and feed material. Gasified feed, considered the most likely primary fuel to the KENTORT II combustor, was tested under the entire range of bed temperatures and MSRT (Table 5-2). Raw and blended CLE 003 feeds were tested using an abbreviated test matrix (Table 5-3). Raw and pyrolyzed CLE S29 shale were tested only under selected conditions. Each matrix element was performed in triplicate.

5.3 RESULTS AND DISCUSSION

5.3.1 Material Balances. The material balances for all bench-scale combustion runs averaged 100.1 ± 0.3 percent. The average weight percent recovered in the spent and gas varied as a function of the feed (Table 5-4).

5.3.2 Combustion of Gasified Feed. No significant differences were found for carbon and nitrogen conversions (Figure 5-2) vs. bed temperature or MSRT within the range of experimental variables studied. Both carbon and sulfur oxidation were nearly complete under conditions as mild as 700°C and 2 minutes MSRT. There was a small temperature dependency for sulfur removal. Based on total nitrogen determined by Leco instrumentation, nitrogen appeared to partition to the combusted solids under all conditions studied. The relative proportions of nitrogen (0.3%) and carbon (0.2%) in the combusted shale was unexpected. A linear regression of nitrogen and organic carbon for the D11 core (Figure 5-3) has a good correlation coefficient ($r=0.94$) and an intercept close to the origin (0.019) suggesting that the nitrogen in the feedstock is entirely in the organic phase.

Two possible explanations for the higher than expected nitrogen content of the combusted shale are nitrogen fixation and/or analytical error. Small errors in the elemental analysis would have a substantial effect on the results. The low nitrogen content of the solid materials used in this study is near or below the detection limits ($0.1 \pm 0.03\%$) of the automated Dumas combustion procedure using Leco instrumentation routinely used for nitrogen analysis in our laboratory. At low nitrogen levels, air contamination can also lead to high analytical values by the Dumas procedure. To examine both alternatives, representative solids were subjected to non-exchangeable NH_4^+ (fixed nitrogen) determinations and to total nitrogen by a micro-Kjeldahl procedure. Fixed nitrogen was very low in the raw shale and did not increase with processing (Table 5-5). A comparison of the total nitrogen of process solids by micro-Kjeldahl and Leco instrumentation, indicated substantially lower nitrogen values by micro-Kjeldahl (Table 5-5). The negligible amount of fixed nitrogen in the process solids coupled with the micro-Kjeldahl nitrogen values which are compatible with the carbon content of

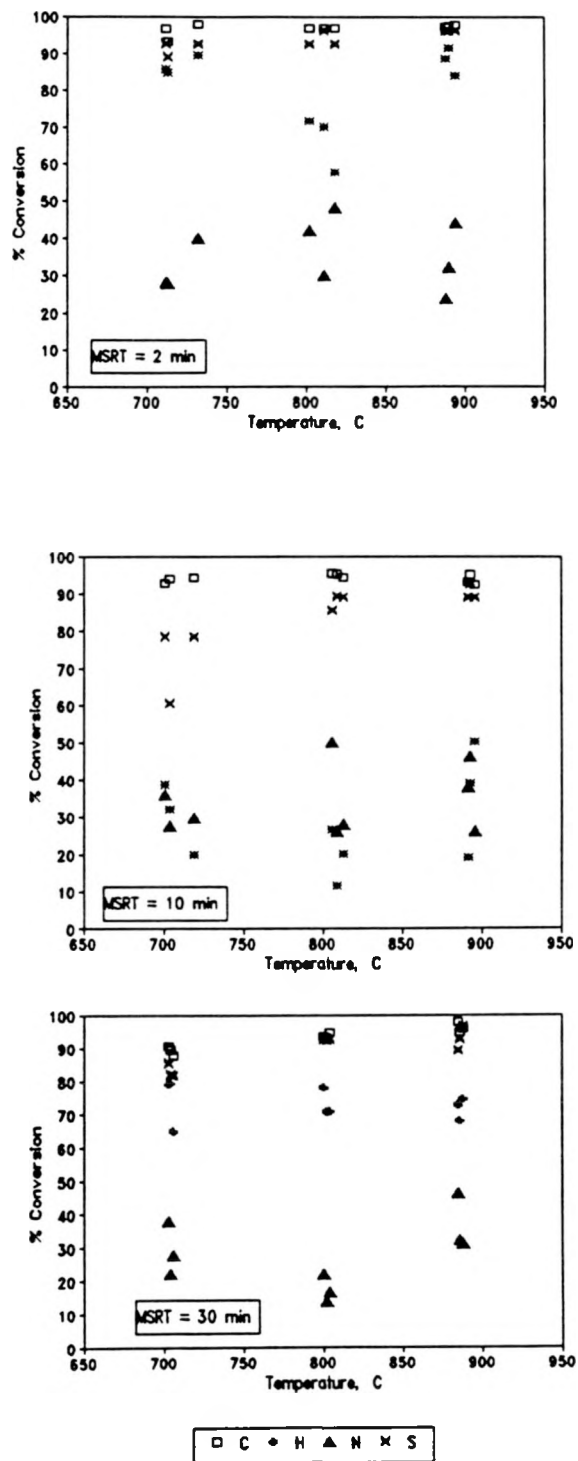


Figure 5-2. *C, H, N, and S Conversions for the Combustion of Gasified Shale.*

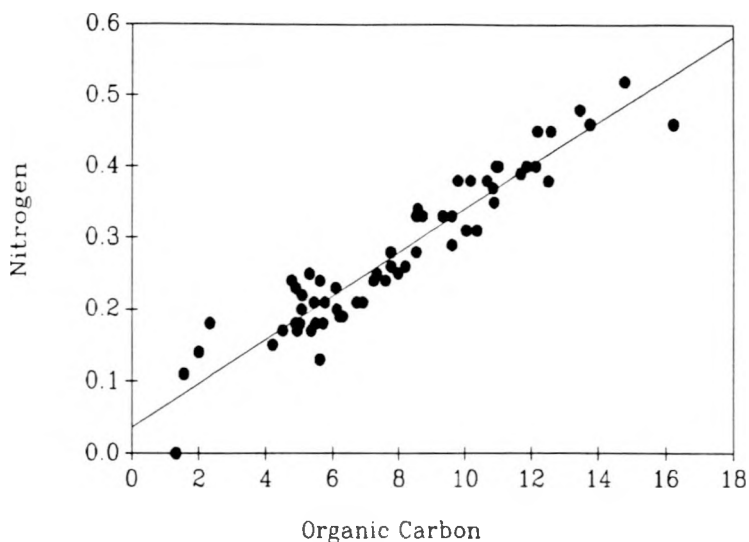
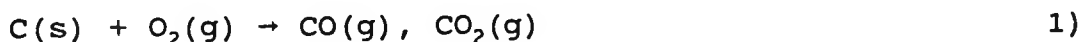


Figure 5-3. Relationship of Nitrogen Concentration to Organic Carbon in the D11 Core (See Figs. 5-2 and 5-5).

the gasified and combusted shales suggests that most of the shale nitrogen is removed during gasification. The removal of most of the nitrogen (probably as NH_3) in the reducing atmosphere of the gasifier would reduce the need for NO_x gas clean-up from combustion of gasified shale. Additionally, ammonia could be recovered as another process by-product. The solid recycle conditions to be utilized for KENTORT should not induce more oxidation of nitrogen than found in this study.

The composition of the combustion gas on an N_2 - and O_2 -free basis was predominated by carbon oxides and contained relatively low levels of NO_2 and SO_2 (Figure 5-4). NO_2 was calculated based on the greater relative nitrogen oxidization indicated by the Leco total nitrogen determinations. Therefore, NO_2 levels in Figure 5-4 would be substantially lower using the micro-Kjeldahl total nitrogen values. As indicated in Figure 5-2 and 5-4 process parameters affected only small variations in NO_2 or SO_2 concentrations. In contrast, the carbon oxides were significantly affected by temperature and mean solid residence time. At 700 and 800°C, both CO_2 and CO were produced, whereas at 900°C, only CO_2 was detected. The CO/CO_2 molar ratio decreased with both bed temperature and MSRT (Table 5-6). For coal combustion, the carbon-oxygen surface reaction:



can produce both CO and CO_2 . The CO can then further oxidize in the gas phase:



Table 5-5. Fixed and Total Nitrogen for Process Solids.

Cle 003 Shale	Nitrogen		
	Fixed	Kj	Leco
Raw	0.03	0.6	0.7
Gasified	0.003	0.1	0.5
Combusted	0.001	0.06	0.3

to generate CO₂. (24) Assuming a similar combustion mechanism for oil shale, detectable CO at 700°C probably reflects incomplete gas phase conversion of CO at lower temperatures. The observed decrease in the average CO/CO₂ ratio at higher MSRT is difficult to explain since similar amounts of carbon were released at all residence times. An explanation is complicated because shorter MSRT in the reactor could only be achieved by increasing the feed rate to the combustor. Therefore, the shale feed rate at 2 minutes MSRT was 15 times greater than the feed rate at 30 minutes. The increased concentration of products in the main gas stream at the lower residence times may have had an effect. However, it is not clear whether solid residence time, increased shale throughput, or some other factor caused the CO/CO₂ ratio to be affected. Additional experiments may be required to clear up this matter.

As indicated in Figure 5-2 and 5-4, the amounts of gaseous carbon, sulfur, and nitrogen species generated from the combustion of gasified shale were not affected greatly by the range of conditions studied. The amounts of these elements released were 37.0 ± 1.9 , 2.4 (0.42 by Kjeldahl) ± 0.2 , and 1.6 ± 0.4 g/kg, respectively. The very small variations in the gas emissions compared to the wide range of combustion operating conditions studied indicated that control of nitrogen and sulfur

Table 5-6. CO/CO₂ Ratio in Product Gas from the Combustion of Gasified Shale.

Combustor Temp., C	Mean Solid Residence Time, Minutes (MSRT)		
	30	10	2
700	0.5	0.4	0.3
800	0.2	0.04	0.02
900	0.0	0.0	0.0

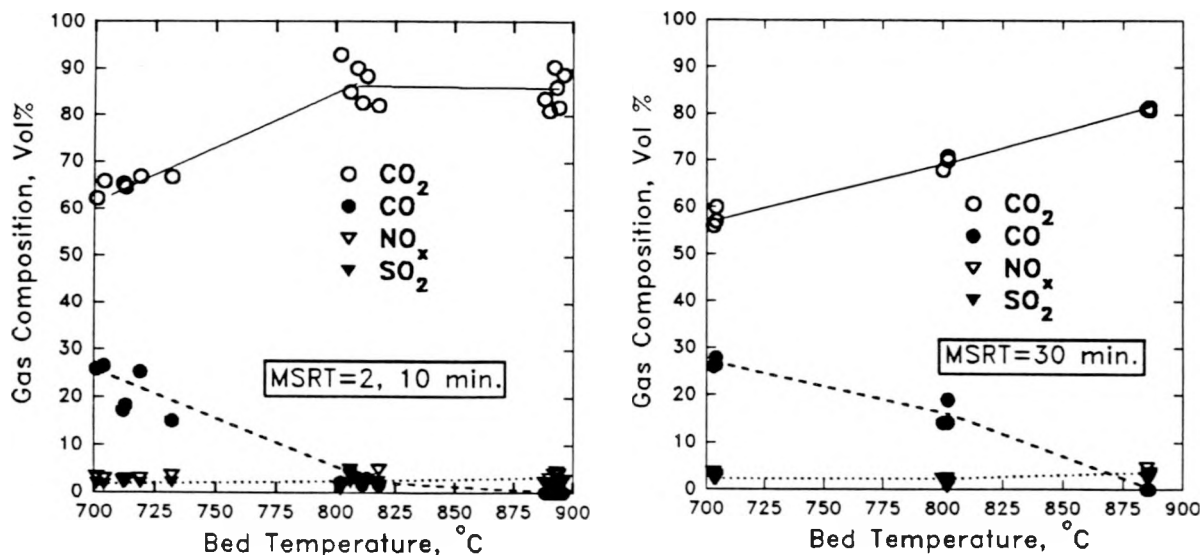


Figure 5-4. *Composition of Product Gas from the Combustion of Gasified Shale.*

emissions will not be possible with combustor operating parameters. This emphasizes the importance of the gasification interstage in the KENTORT II process which provides a relatively low-sulfur combustor fuel.

5.3.3 Combustion of Raw and Blended Feeds. C, H, N, and S conversion for raw and blended feeds were similar to those obtained for gasified feed except that the small temperature dependant increase in sulfur conversion was not observed (Figures 5-5 and 5-6). Nitrogen determinations based on the Kjeldahl procedure are plotted on these graphs.

The N₂-, O₂-free product gas composition for the blended feed was similar to that for the gasified feed alone (Figure 5-7 and 5-8). CO₂ comprised 85-94% of the gas. The ratio of CO:CO₂ in the product gas for the blended feed decreased with temperature from 0.1 to 0.01. The product gas composition for raw feed combustion did not vary significantly within the temperature range studied. Again the dominant gas was CO₂ which comprised $82.5 \pm 3.0\%$ of the total product. CO ($2.0 \pm 0.7\%$) and SO₂ ($1.6 \pm 0.5\%$) were also detected. The CO:CO₂ ratio was 0.02. Bed temperature control was very difficult for the combustion of raw shale due to the richness of the feed. Overheating occurred easily. The actual mean temperature range studied (720-827°C) may not have included sufficiently low or high temperatures to

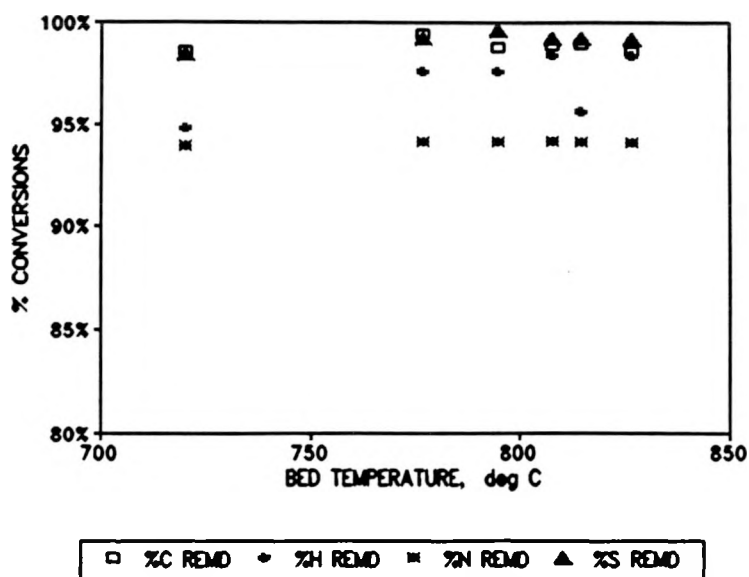


Figure 5-5. C, H, N, S Conversions for Raw Feed.

see this effect. However, the increased carbon available for oxidation may also have contributed to the lack of variation in the carbon oxides in this temperature range at constant MRST.

The amount of C, N, and S released to gaseous products, of course, increased as the proportion of raw shale in the feed increased. A comparison of the release of these elements for the various feeds is given in Table 5-7. Note that the use of gasified feed does significantly limit the release of sulfur and nitrogen oxides from the process.

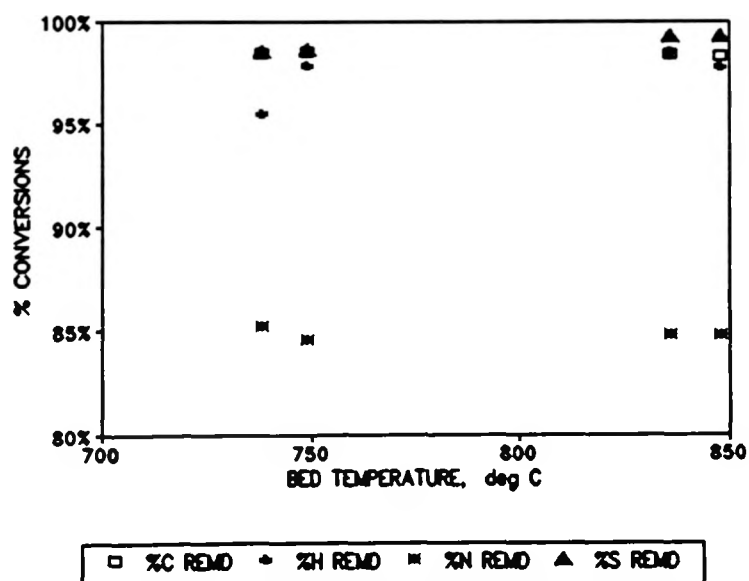


Figure 5-6. C, H, N, S Conversions for Combustion of Blended Raw and Gasified Feed.

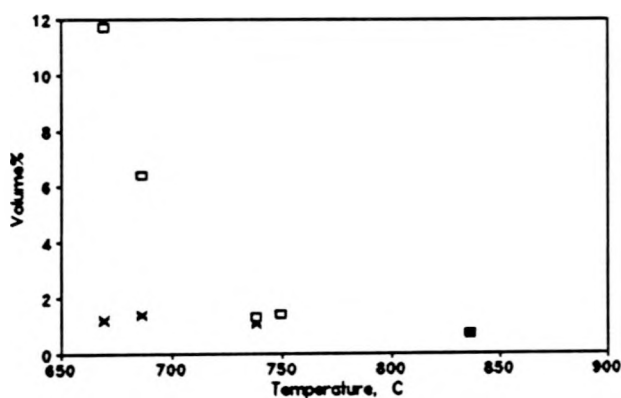
Table 5-7. The Amount of C,H,N Released in Combustion Gas.

	g C,N,S released / kg feed		
	Gasified	Blended	Raw
C	37.0 $\pm$ 1.9	107.1 $\pm$ 1.2	166.3 $\pm$ 0.5
N*	0.4 $\pm$ 0.2	2.8 $\pm$ 0.5	7.2 $\pm$ 0.1
S	1.6 $\pm$ 0.4	10.7 $\pm$ 0.2	18.8 $\pm$ 0.1

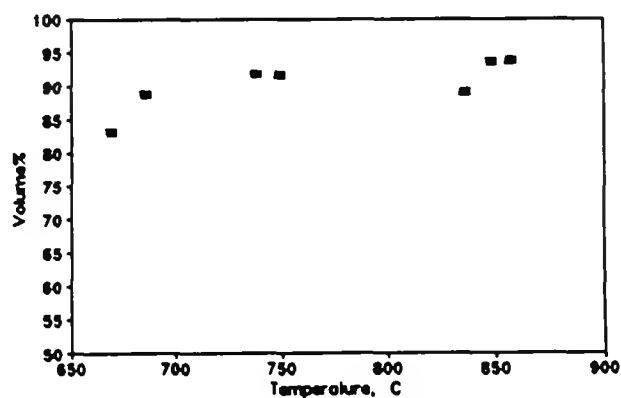
* Kjeldahl determined.

5.3.4 Composition of the KENTORT II Process Solids. Shale materials have been processed in each step of the KENTORT II process. Pyrolysis-gasification removed 78, 87, and 86 percent (Table 5-8) of the raw shale carbon, nitrogen, and sulfur, respectively, with subsequent combustion decreasing the carbon, nitrogen, and sulfur content by 99, 93, and 99 percent on a raw shale basis. These materials are expected to be representative of the composition of KENTORT II solid waste and should not present significant disposal problems.

5.3.5 Sintering. Shale particle agglomerates formed during combustion of both the CLE S29 raw and pyrolyzed feedstocks. The agglomerates were magnetic which implicated iron compounds and allowed for their separation from the combusted materials for Mössbauer analyses. For comparison, thin sections were also prepared from a clinker of fused shale material from a run where the targeted temperature of 900°C was exceeded.



□ CO x SO2



■ CO2

Figure 5-7. Volume Percent CO and SO₂ in Combustion Gas from Blended Feed.

Figure 5-8. Volume Percent CO₂ in Combustion Gas from Blended Feed.

Table 5-8. *Mean Elemental Composition of KENTORT II Process Solids.*

	%Carbon	%Hydrogen	%Nitrogen*	%Sulfur
Raw (CLE 003)	16.7	2.0	0.6	1.7
Pyrolyzed- Gasified	4.6	0.1	0.1	0.3
Combusted	0.2	0.07	0.06	0.03

* Kjeldahl determined.

Thin sections of the magnetic particles indicated that the agglomerates were primarily formed by sintering and not by melting. Agglomerates consisted of intact shale particles bound together by darker appearing magnetic sintered material. In contrast, the shale particles showed clear evidence of melting and fusion in the clinker material, including vesicle formation from gas evolution.

The distributions of iron compounds in the feedstocks, gasified, combusted and sintered materials (Table 5-9) as determined by Mössbauer analyses indicated that the feedstocks which sintered (raw and pyrolyzed CLE S29) contained an iron hydroxysulfate. Hydroxysulfate is an oxidation product which probably formed in the S29 sample during storage. The CLE 003 gasified feedstock did not contain an iron hydroxysulfate and did not agglomerate. The combusted pyrolyzed shale (PYRO S29) and the separated sinters contained iron as magnetite, hematite and silicates (Table 5-9). The magnetite was predominantly associated with the sinters since 48 percent of the iron in the agglomerates was present as magnetite vs. only 4 percent in the unseparated material. Combusted material from gasified CLE 003 which did not sinter did not contain magnetite. During coal combustion, iron hydroxysulfates are known to sinter as they are oxidized to hematite in the temperature range of this study.(3) Since magnetite is a possible intermediate oxidation state in the conversion of an iron hydroxysulfate to hematite, a similar sintering mechanism is suspected.

The initial presence of magnetite did not result in sintering since 48 percent of the iron in gasified CLE 003 was as detected as magnetite. This is important since during gasification, all the iron hydroxysulfate in CLE S29 was converted to either magnetite or iron silicates (Table 5-9). Since gasified shale is

Table 5-9. *Distribution of Iron in the Feed and Combusted Materials.*

	Distribution of Iron, %							
	Feedstocks					Combusted Solids		
Compound	CLE S29	PYRO S29	CLE 003	P/G 003	P/G S29	PYRO* S29	S29** Sinters	P/G* 003
Pyrite	82	45	86					
Pyrrhotite		20			5			
Illite	12	19	14					
Iron Hydroxy-sulfate	6	16						
Magnetite				48	45	4	48	
Hematite						58	32	48
Fe Silicates				52	50	38	19	52

* Combusted at 700°C and 30 minutes MSRT.

** Sinters separated from combusted PYRO-S29.

the likely primary feedstock to the KENTORT II combustor, sintering should not be a problem.

5.4 CONCLUSIONS

All feeds tested exhibited similar combustion behavior except for the CLE S29 feeds. These feeds contained an iron hydroxysulfate which is known to sinter during coal combustion. Sinters were magnetic being composed of the iron oxides, magnetite and hematite. During gasification, iron hydroxysulfate present in the shale was converted to magnetite and/or iron silicates. Therefore, sintering of gasified combustor feeds should not be a problem.

Carbon, nitrogen and sulfur were nearly completely removed by the combustion of gasified shale at 700°C and 2 minutes MSRT. With minimal quantities of carbon and sulfur in the disposal solid, little acid generating potential is expected. Nitrogen appears to be removed in the gasifier reducing the need for NO_x gas clean-up of the combustor gases. Since essentially all the sulfur in the feed was released at all combustor conditions tested, the importance of a gasification interstage to manage SO₂ emissions was confirmed.

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6. COLD-FLOW STUDIES

From previous oil shale studies, the CAER gained experience with the fluidization properties (i.e., distributor design, minimum fluidization velocity, attrition, etc.) of eastern U.S. oil shales, but additional information was necessary to integrate two or more fluidized beds. Data was essential in the areas of pneumatic transport, aerated downflow, and slip-stick downflow. To obtain these data and to confirm the preliminary design parameters, cold-flow modelling became an integral part of the design effort for the 5-lb/hr integrated system.

The cold-flow modelling study, like the reactor development effort in general, was performed in 2 stages. The first stage involved a configuration consisting of stacked fluidized beds which represented the pyrolysis and gasification zones. The second stage of study added a third fluidized bed (combustor) to the stage 1 system.

The main objective of the cold-flow modelling work was to develop a full scale, working prototype of each reactor configuration. Critical design parameters were then directly transferred to the reactor system. The stable recirculation of char from the gasifier to the pyrolysis and combustion zones reactor is especially important in the KENTORT II design, so a major effort involved the development of a device to control the rate of char recirculation.

6.1 METHODOLOGY

6.1.1 Apparatus. Shown in Figure 6-1 is a schematic of the acrylic model for the pyrolysis-gasification configuration of the KENTORT II reactor from the first stage of the study. The model included two dense phase fluidized beds which were aligned vertically and shared a common fluidizing gas. There were transfer lines connecting the fluidized beds so that solids could circulate (counterclockwise as viewed in Figure 6-1) between the two. The upper fluidized bed represented the pyrolysis reactor, and the lower fluidized bed represented the gasification reactor.

The body of the model was constructed of 3" O.D. x 2.75" I.D. clear, extruded acrylic tubing. The body was flanged at both gas distributors, each consisting of a perforated aluminum plate. A screw feeder above the reactor metered shale into a 3/8" O.D. acrylic tube which extended into the freeboard region of the upper bed. The gravity-fed standpipe connecting the upper and lower beds (SP-1 in Figure 6-1) was constructed of 0.43" O.D. x 0.36" I.D. glass tubing with a section of curved copper tubing at the bottom. Shale exited the system from the lower fluidized bed by spilling into one of three 3/8" O.D. x 1/4" I.D. acrylic tubes

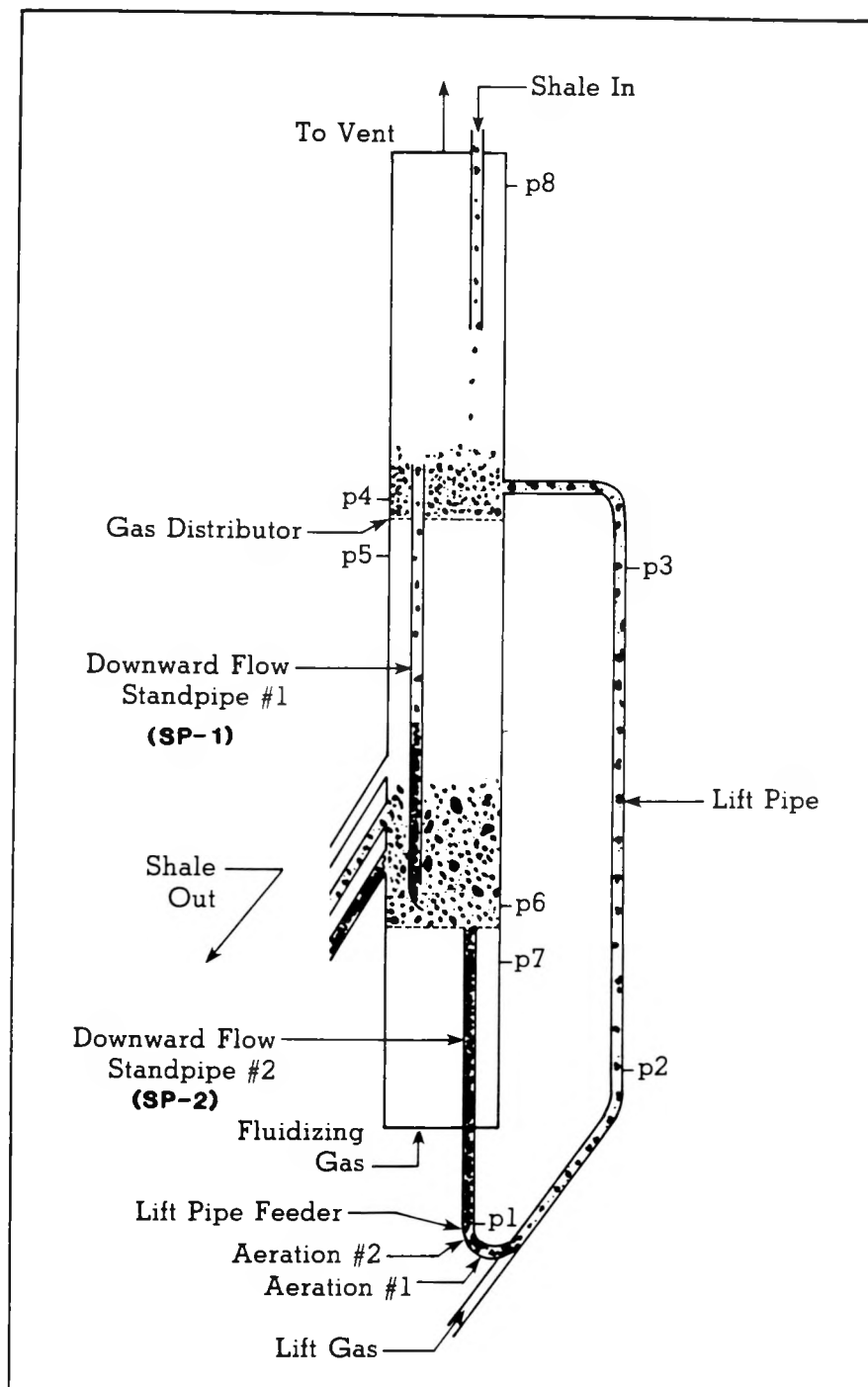


Figure 6-1. *Schematic of the Cold-Flow Model Developed for the Pyrolysis-Gasification Configuration of the KENTORT II Reactor.*

angled to 30° with the wall of the model. The bed height was determined by the position of the exit tube that was selected. The gravity-fed standpipe which connected the lower fluidized bed to the lift pipe feeder (SP-2 in Figure 6-1) was constructed of 1/2" O.D. copper tubing at the distributor and was reduced to 3/8" O.D. stainless steel tubing at the lift pipe feeder (aerated J-valve). The main body of the J-valve was constructed of 3/8" O.D. tubing and contained 4 points of aeration. The lift pipe consisted of 3/8" O.D. Tygon tubing supported by stainless steel tubing at each end. All metal portions of the model were grounded to reduce the buildup of static charge.

The model of the complete KENTORT II reactor is shown in Figure 6-2. This system was identical to the first with the addition of another recirculation loop. In this loop, solids were withdrawn from the lower bed and lifted to an adjacent fluidized bed which represented the combustion zone of the KENTORT II reactor. An identical J-valve and lift pipe arrangement was used to meter shale to the third fluidized bed. A gravity-fed standpipe (SP-3 in Figure 6-2) returned the solids from the third bed to the lower bed.

Pressure taps were located throughout each system, and the various pressure measurements were performed with water-filled manometers. The flow rates of the fluidizing, transport, and aeration gases (air) were measured by rotameters.

6.1.2 Material Studied. The material used for the study was a 20x60 mesh New Albany shale pyrolyzed in a 3-inch diameter fluidized bed reactor using recycle-gas for fluidization as described by Carter and Taulbee.(1) A pyrolyzed shale was used since it most represented the type of material which would be present in the reactor system.

6.1.3 Procedure. The most extensive cold-flow development and testing was performed with the unit as configured in Figure 6-1. The procedures employed to develop this configuration were conducted in a stepwise fashion. Initially, the portions of the model required to move solids between the fluidized beds (i.e., lift pipes, standpipes, and a solids metering device) were designed and tested separately. Subsequently, the unit was then tested as a complete system, and once the proper operating parameters were achieved, the pressure profile of the system was determined for the range of proposed reactor conditions. The cold-flow model development program can be categorized into 4 separate phases:

- 1) Design and testing of Standpipe SP-1. The position, length, and diameter of this standpipe were adjusted so that the height of the upper fluidized bed remained stable and steady transfer of solids from the upper to the lower fluidized bed was achieved.

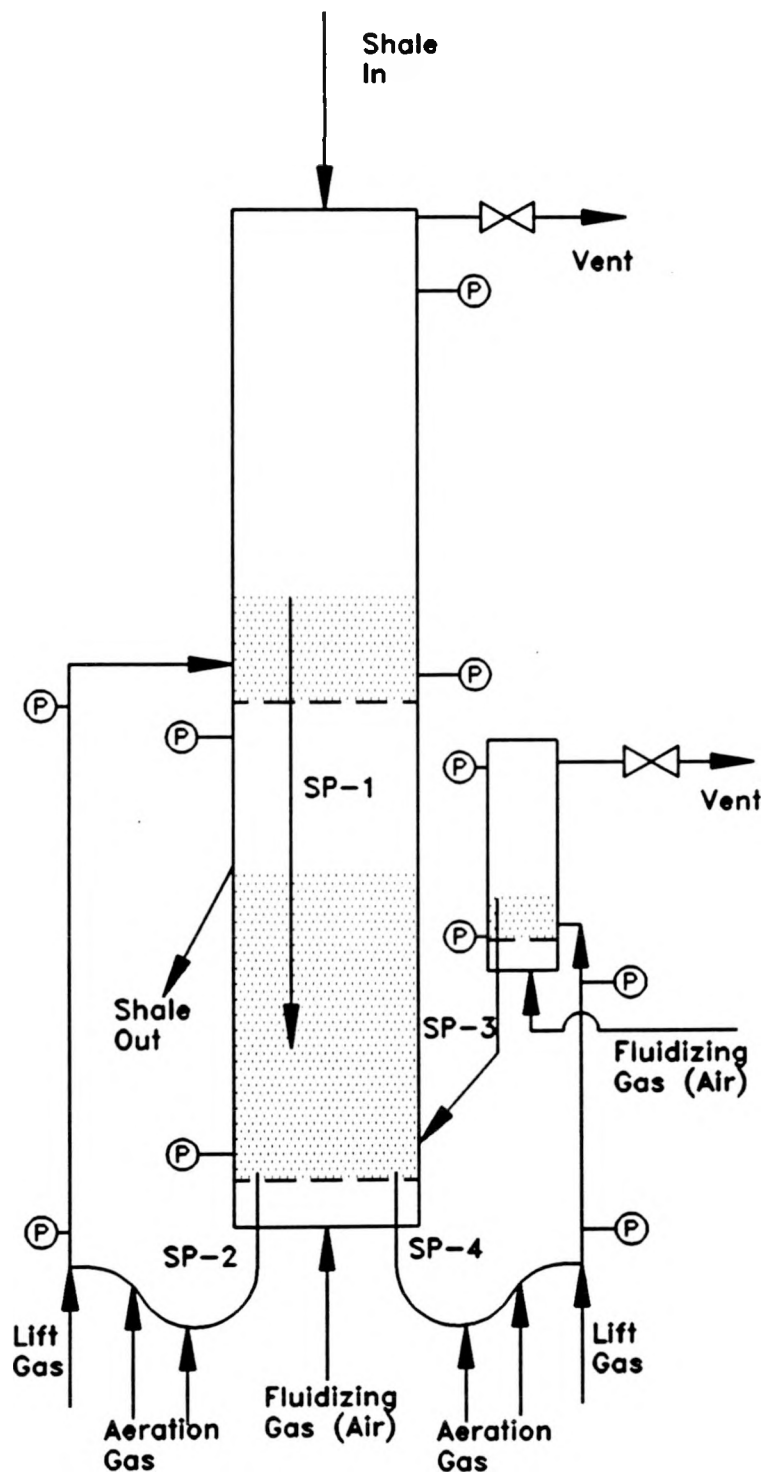


Figure 6-2. Schematic of the Cold-Flow Model Representing the Complete KENTORT II Reactor System.

2) Development of Lift Pipe Feeder. This phase was performed in two parts. The first part involved the testing of 'J-type' and 'L-type' aerated feeding devices. Each was tested to determine its ability to control the discharge rate of shale char as a function of aeration gas flow rate. A sealed container full of shale was attached to the feeder, and the shale discharged was collected during a measured time period over a range of aeration gas flow rates. The feeder most able to achieve the range of design conditions was chosen for continued testing.

The second part of this phase concerned the testing of the successful feeder while attached to the model under simulated recirculation conditions. A cyclone was attached to the lift pipe and was maintained at the same pressure as the point of injection to the upper fluidized bed. A char feedrate of approximately 20 lb/hr was established for these tests in order to replenish the char leaving the system through the modified recirculation system. The rate of shale collected in the cyclone was then measured as a function of aeration gas flow rate and position. The pressure drop across the lift pipe (p_2-p_3 , see Figure 6-1) was recorded for each test.

3) Demonstration of Shale Recirculation. Shale was continuously recirculated between the upper and lower fluidized beds over an 18 hour period. All important process conditions were recorded by a computerized data acquisition system.

4) Determination of Pressure Profile of the Model. At nominal recirculation rates of 0:1, 1:1, 2:1, and 3:1, the static pressure at each of the 8 pressure taps (Figure 6-1) was measured with all other conditions held constant. Pressure measurements with no shale in the model were also performed in order to determine the contribution of the gas distributors to the recorded pressure drop.

The remaining parameter which needed to be tested to complete the KENTORT II model (see Figure 6-2) was the design of standpipe SP-3. This standpipe was tested by operating the completed system under recirculating conditions, while monitoring the system for stability and adjustability.

6.2 RESULTS AND DISCUSSION

The following sections primarily describe the design and operation of the cold-flow model in its first stage of development. Most of the results from the first stage were directly transferred to the final configuration, making extensive retesting unnecessary.

6.2.1 Design and Testing of SP-1. The purpose of standpipe SP-1 (see Figure 6-1) in the KENTORT II reactor is to maintain a constant level in the upper fluidized bed and deliver shale to the lower fluidized bed. This standpipe is the only exit for solids from the upper fluidized bed, so the solids flow through the tube is the sum of the raw shale feedrate and recycle rate of the char. Therefore, for a recycle ratio of 3:1, this tube needed the capacity to handle up to 20 lb/hr of shale char.

The initial design of the standpipe called for a straight 3/8" O.D. x 1/4" I.D. acrylic tube which was immersed into the lower fluidized bed. This design was found to be unacceptable for three reasons. First, excessive slugging and backflow of solids from the lower to the upper bed was evident. Second, shale would only flow downward through the tube when the level of the top bed well exceeded the height of the standpipe. Finally, the flow rate of solids capable through this standpipe was determined to be well below 20 lb/hr.

Since the first design indicated the mass flow rate to be low, it was essential that the modified design include a standpipe of increased diameter. A 0.43" O.D. x 0.36" I.D. glass tube was next considered. The increased diameter proved capable of providing the required 20 lb/hr mass flow rate. Additionally, the larger standpipe effectively moved solids from the upper to the lower bed while maintaining the level of the upper bed at a constant height slightly above the standpipe. However, excessive slugging and backflow of solids was witnessed when there was no net shale throughput. While zero throughput is not expected during normal operation, if an upset in raw shale feed or char circulation were to occur the levels of the beds must remain constant for process stability. Excessive backflow of char could empty the lower bed below SP-1 and cause gas bypassing which would prevent subsequent downward solid flow. Once the standpipe became uncovered in this fashion, it was found necessary to repeatedly collapse and refluidize the beds to drain the shale from the upper bed.

In order to alleviate the slugging and backflow condition of the standpipe, the upward flow of gas through the tube had to be reduced. By raising the bottom of SP-1 in the lower bed, the upward flow of gas decreased in the standpipe and reduced the unstable condition at zero char throughput. However, this increased the potential for the standpipe to become uncovered during normal operation. It was determined that this situation was also unacceptable, and another method of reducing the slugging in SP-1 was required. A portion of a bend of 1/2" O.D. copper tubing was attached to bottom of the standpipe so that shale was discharged radially and gas had to also enter radially rather than axially. This design was found to restrict the upward flow of gas while allowing sufficient downward transfer of shale char. Operation of this standpipe remained stable at zero

throughput while being capable of the required 20 lb/hr mass flow rate. This modification was incorporated into the final design of the reactor system.

6.2.2 Development of Lift Pipe Feeder. An important criterion for the design of a lift pipe feeder was that it have no moving parts. This was determined to be important because of the extremely high temperatures and the requirement that the feeder be totally enclosed in a furnace. The small size of the transport lines was also a consideration, since it was doubtful that something such as a rotary valve could operate satisfactorily with such a small pipe diameter to particle diameter ratio. For these reasons the use of an aerated feeder was deemed most appropriate since the solid mass flow rate is controlled by aeration gas flow rate and there are no mechanical problems associated with moving parts. However, a disadvantage of the aerated feeder design is that a positive seal between sections is not achieved and the use of purge gas and pressure balancing becomes most critical. Additionally, control and measurement of solid mass flow rates are subject to empirical correlations and heat balance calculations.

A 'J-type' feeder was tested which had four points of aeration and delivered shale vertically to the transport tube. Also, an 'L-type' feeder was tested which had only one point of aeration that was parallel to the horizontal discharge of the solids. Figure 6-3 shows the results of tests conducted to determine the discharge rate of each feeder design as a function of total aeration rate. These results were obtained with no flow of gas in the standpipe, and the solid flow down the standpipe must, therefore, be characterized as slip-stick.(2) The J-type feeder displayed nearly linear discharge as a function of aeration rate from 2.5 to 7 ft³/hr (SCFH), above which, the discharge rate remained constant at approximately 11.5 lb/hr. A minimum gas flow rate of approximately 2.5 SCFH was required to obtain discharge of char from the feeder; therefore, 4.5 lb/hr is the lowest practical discharge rate to be expected. The L-type feeder displayed a constant discharge rate of approximately 15 lb/hr regardless of aeration gas flow rate above 2.5 SCFH. The rate of freefall through the standpipe was found to be 15.9 lb/hr, and this indicates that the aeration gas in the L-type feeder was simply blowing the shale away from the end of the standpipe as fast as it could fall through the standpipe. The L-type feeder was therefore rejected from further consideration. It was determined to proceed with the testing of the J-type feeder even though its maximum discharge rate was somewhat lower than desired. Since these determinations were made with no net flow of gas in the standpipe, it was expected that the gas flow under normal operating conditions would increase the mobility of the solids and reduce resistance to downward flow.

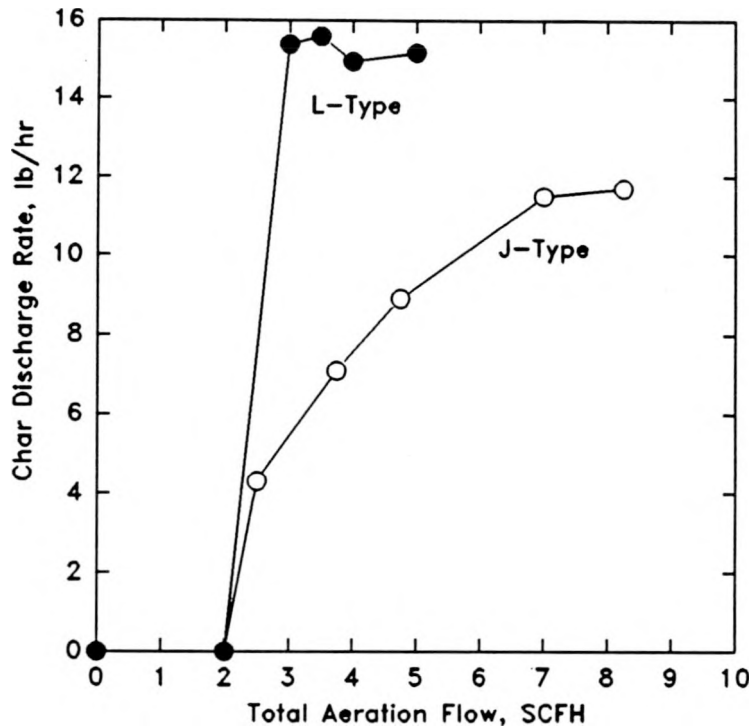


Figure 6-3. *Effect of Total Aeration Gas Flowrate on the Discharge Rate of Shale Char from the 'J-type' and 'L-type' Valves.*

6.2.3 Electrostatic Charge Problems. Throughout the testing of the recirculation loop, intermittent stoppage of shale recycle had occurred. Vibration of the unit was found to usually remedy the problem. The reason for this problem was unknown until it was discovered that downward flow of shale through the acrylic standpipe which fed the lift pipe feeder could be stopped by touching the outside of the tube with one's hand. Build up of electrostatic charge was causing shale char to cling to the wall of the tube and completely shut down the flow of solids. The acrylic tube was then replaced with a metal tube, and all metal parts in contact with shale char were subsequently grounded. As a result, no such interruption of solids flow was subsequently observed.

6.2.4 Calibration of Lift Pipe Feeder. The J-valve was attached to the cold-flow model via SP-2 for additional testing and calibration. The desired range of recirculation rates was achieved in this configuration (Figure 6-4). As found during stand-alone testing, it appears that a minimum total aeration flow rate of 2.5 SCFH is required to propagate solid recirculation. The position and flow rate of aeration gas were found to be important considerations for the discharge rate of shale char. With aeration at the AR-1 position only, the

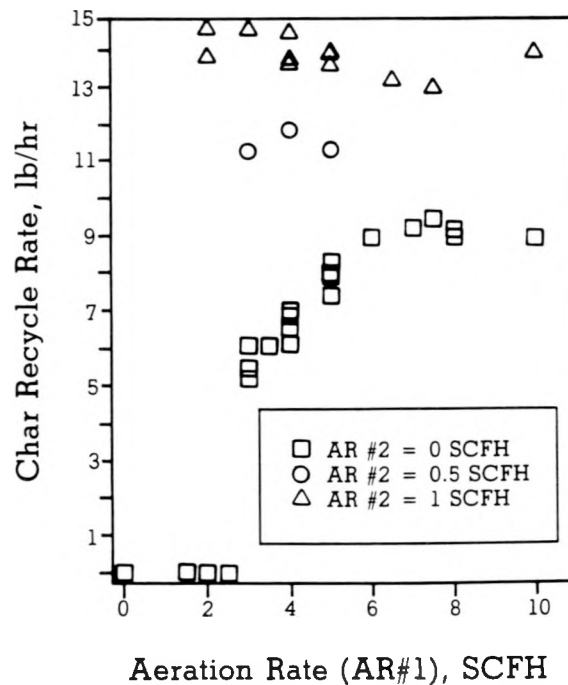


Figure 6-4. *Effect of Aeration Gas Flowrate and Position on the Rate of Shale Char Recirculation for the J-Valve.*

behavior was similar to that observed from stand-alone testing of the feeder as shown in Figure 6-3. However, the addition of 0.5 SCFH of air at the AR-2 position caused a jump in the shale char flow rate which was independent of total aeration gas flow rate. Additionally, an air flow rate of 1 SCFH or greater at the AR-2 position caused the solids mass rate to reach a maximum regardless of the AR-1 flow rate (above 2 SCFH). This type of behavior was not experienced during stand-alone testing of the feeder because there was no flow of gas in the standpipe, and the aeration gas flow rate was not controlled at individual positions.

Pressure drop measurements across SP-2 indicated that the position and rate of aeration greatly influenced the direction and magnitude of gas flow. For example, injection at AR-1 favored a downward flow of gas from the lower fluidized bed to the feeder, while aeration at the AR-2 position preferred the upward flow of gas against the direction of solid flow. Therefore, the AR#2 position is expected to create more of a fluidized or aerated condition in the lower portion of the standpipe and produce more solid mobility and less resistance to flow.

It appears that the most convenient method of controlling the recirculation rate continuously over the entire range of rates is to hold constant the flow rate of aeration at AR-1 and closely meter the flow rate of gas to AR-2. The flow meter available at the time of testing was accurate to only 0.5 SCFH, so continuous adjustment from 0 to 1 SCFH was not possible. While it will probably not be possible to extrapolate this correlation to actual process conditions, this exercise did illustrate that the solids recirculation rate can be effectively controlled by the flow rate and positioning of aeration gas.

It was determined that an independent method of measurement was required to verify the aeration correlation. Shown in Figure 6-5 is the solid recirculation rate plotted as a function of pressure drop (p_2-p_3) across a vertical 22.25 inch section of the lift pipe. For a given flow rate of lift gas, the frictional resistance and the static head of the flowing solid/gas mixture increase causing the pressure drop to rise accordingly. The data shows, for a given shale flow rate, an increase in lift gas velocity caused the pressure drop to increase. This indicates that the range of lift gas velocities studied was sufficiently high that frictional resistance was the controlling factor. If static head were the controlling factor, an increase in gas velocity would decrease the pressure drop because of reduced

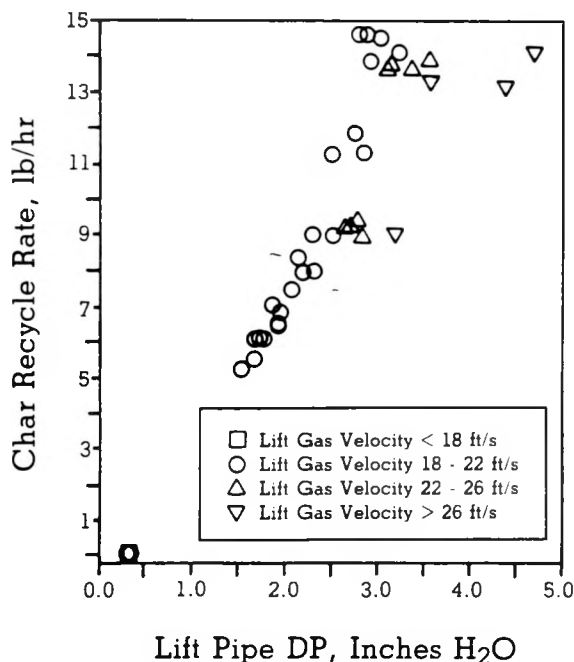


Figure 6-5. Relationship Between Lift Pipe Pressure Drop, Lift Gas Flow Rate, and Char Recirculation Rate.

static head of the mixture. The variations in gas velocity indicated in Figure 6-5 were a reflection of the aeration flow rates as the lift gas flow rate was held nominally constant at 0.35 SCFM. The lift gas velocities were not corrected for the addition or loss of gas flow due to leakage into or out of the lower fluidized bed. Once again, this correlation is probably not applicable at actual reactor conditions, but it does indicate that measurement of solid flow rate can be effectively related to pressure drop.

6.2.5 Demonstration of Shale Recirculation. The operation of the model was performed over an 18-hour period in order to confirm its capability for sustained, stable transfer of solids

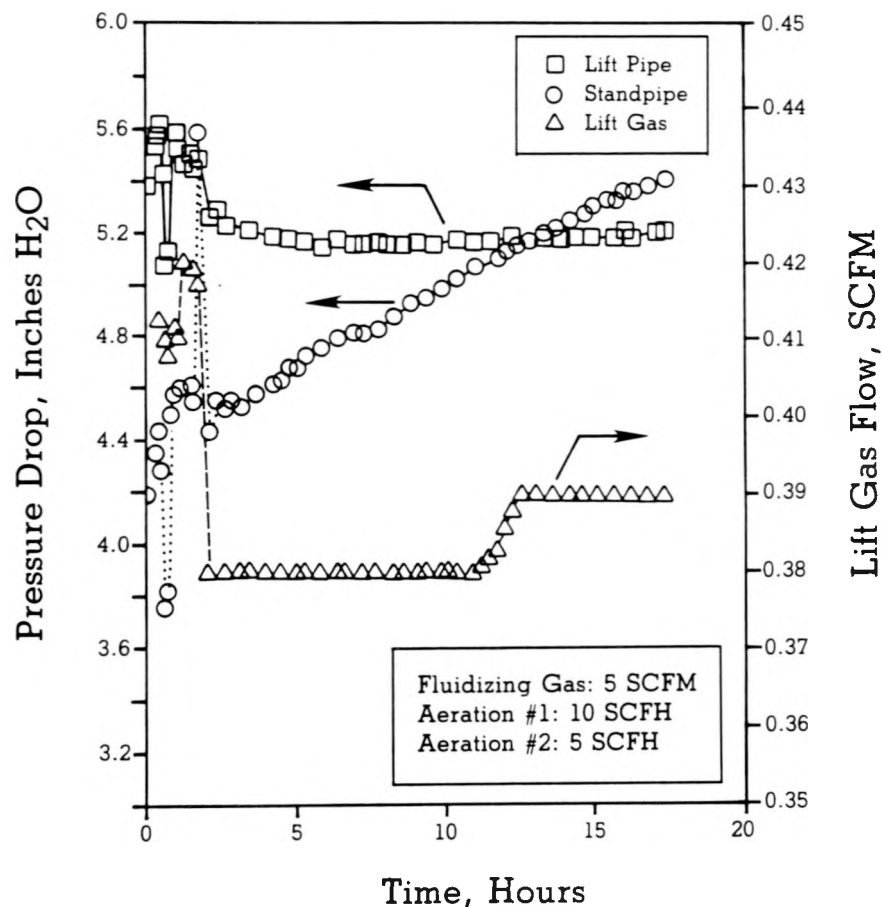


Figure 6-6. Log of 18-hour Run of the Cold-Flow Model.

and the maintenance of steady fluidized bed levels during an extended period. The run was performed with no net throughput of solids in the system. Continuous feeding and withdrawal was not necessary since it was the stability of solid recirculation which was to be tested. Figure 6-6 shows the conditions recorded during the run. The first few hours were spent adjusting the parameters to test the operability of the system, thus the upsets indicated by the run log during this period were intentional. Following the first two hours of operation, the pressure drop across the lift pipe (p2-p3) remained remarkably constant. According to the recorded pressure drop and the aeration gas flow rate settings, the solid recirculation rate remained at 14.5 lb/hr throughout the period. Approximately 11 hours into the run a gradual rise and leveling of the lift gas flow rate was evident. As a result, a slight rise in the lift pipe pressure drop was observed.

The pressure drop across SP-2 (p1-p6) rose steadily throughout the run. Since there was no net solid throughput, there was a gradual decrease in solid inventory due to decrepitation and subsequent elutriation. The decrease in inventory was reflected by decreased bed height in the lower fluidized bed and a corresponding decrease in absolute pressure in the lower portion of the bed. Therefore, as the height of the lower fluidized bed decreased, the pressure drop across the standpipe increased.

6.2.6 Pressure Profile of Model. Pressure profiles were determined for the model at various solid recycle rates with an overall throughput of 5.5 lb/hr (Figures 6-7 - 6-10). The pressure profile of the system with no solid recirculation (Figure 6-7) indicated some pressure drop across the lift pipe which was due to the friction of the lift gas. With no solid recycle there was some leakage of gas from the lower fluidized bed to the lift pipe feeder and the lift pipe. The only significant differences compared with non-recycle operation (see Figure 6-8) were that the lift pipe pressure drop increased markedly and the pressure drop across the lower standpipe decreased. The leakage of gas from the lower fluidized bed (Fig. 6-8) indicated that slip-stick solid flow was definitely occurring. The leakage of gas was reduced at a recirculation rate of 11 lb/hr as shown in Figure 6-9. Under these conditions the pressure drop across the lower standpipe was nearly zero. Figure 6-10 shows the pressure profile at the highest recirculation rate of 14.5 lb/hr. Lift gas in this situation was leaking into the lower fluidized bed, and gas was flowing against the downward flow of solids in the standpipe. However, the pressure drop was so low that it is doubtful that the entire length of the standpipe was aerated. The level of shale in the upper standpipe was significantly above the level of the fluidized bed indicating that the throughput limit of the standpipe was nearly reached.

Figure 6-7. Pressure Profile for Recycle Ratio of 0:1.

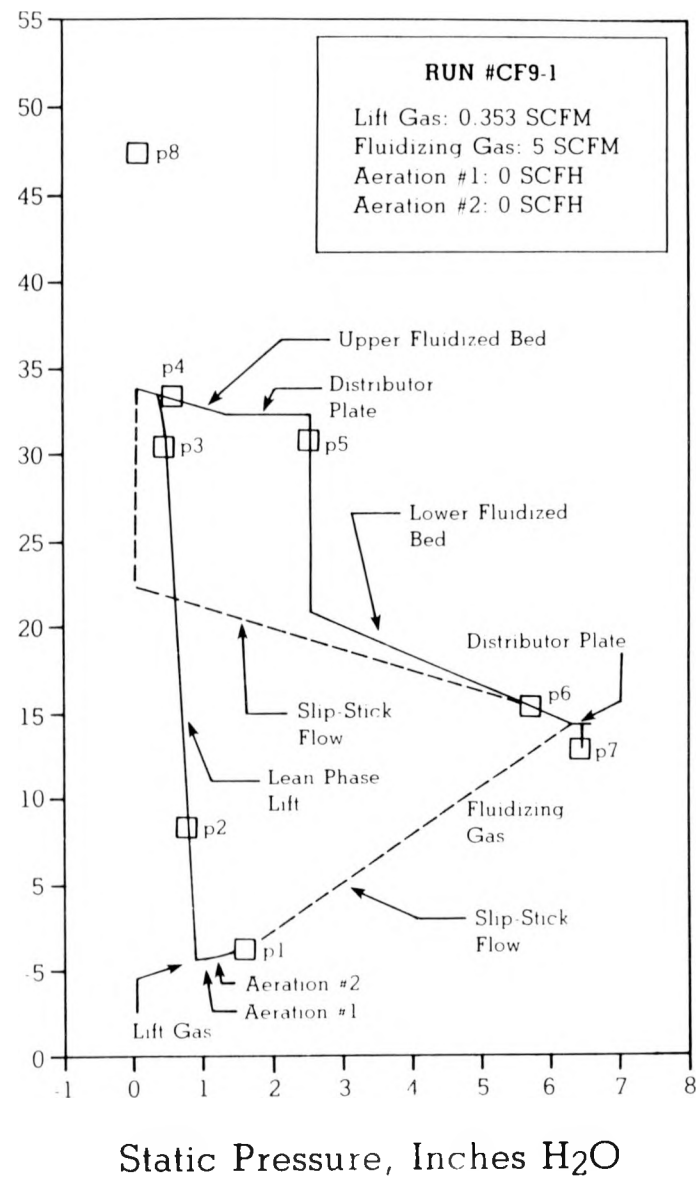
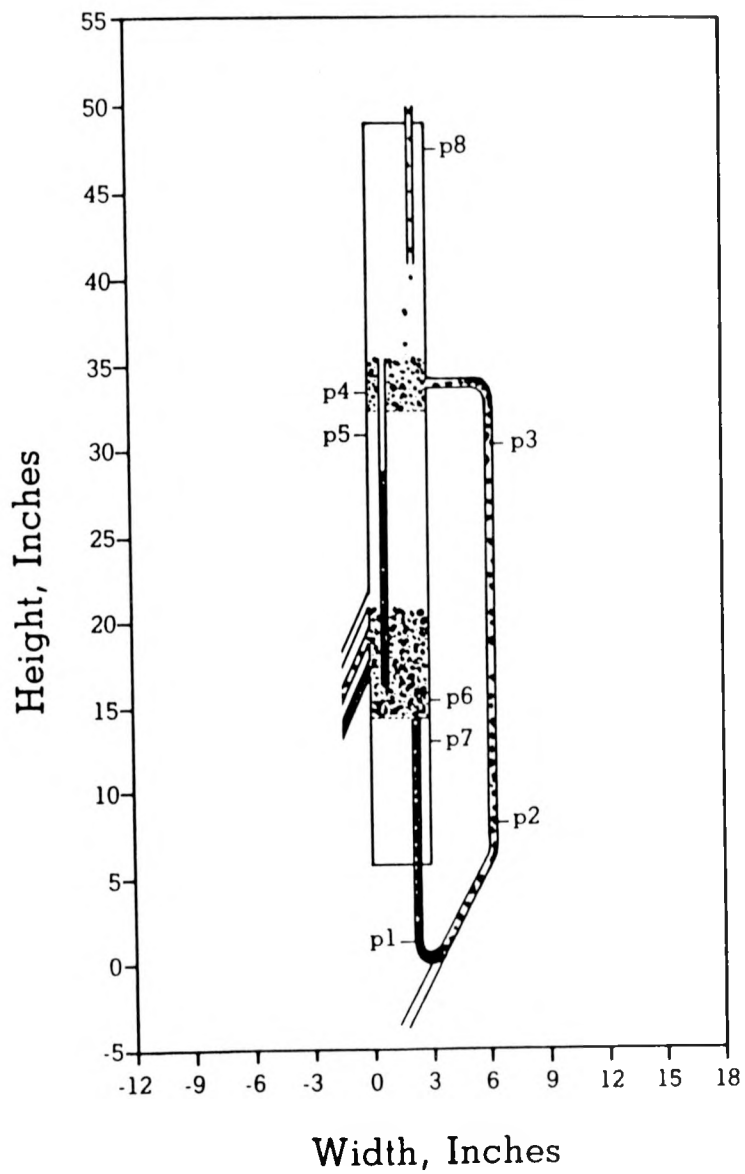


Figure 6-8. Pressure Profile for Recycle Ratio of 1:1.

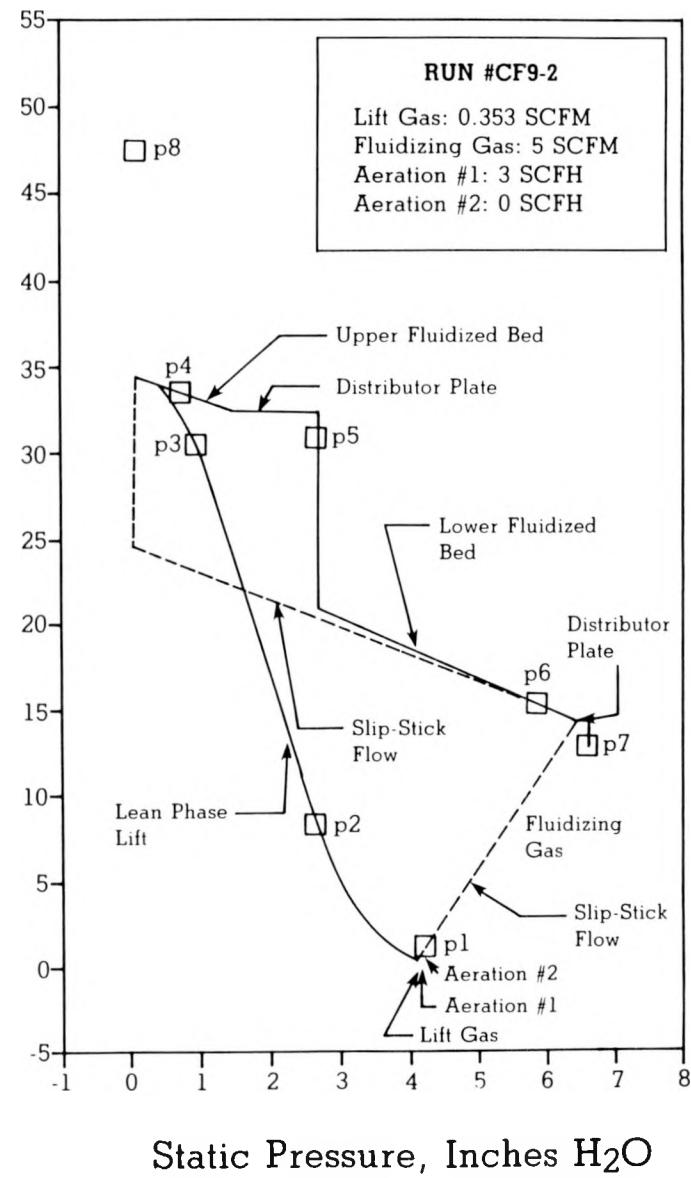
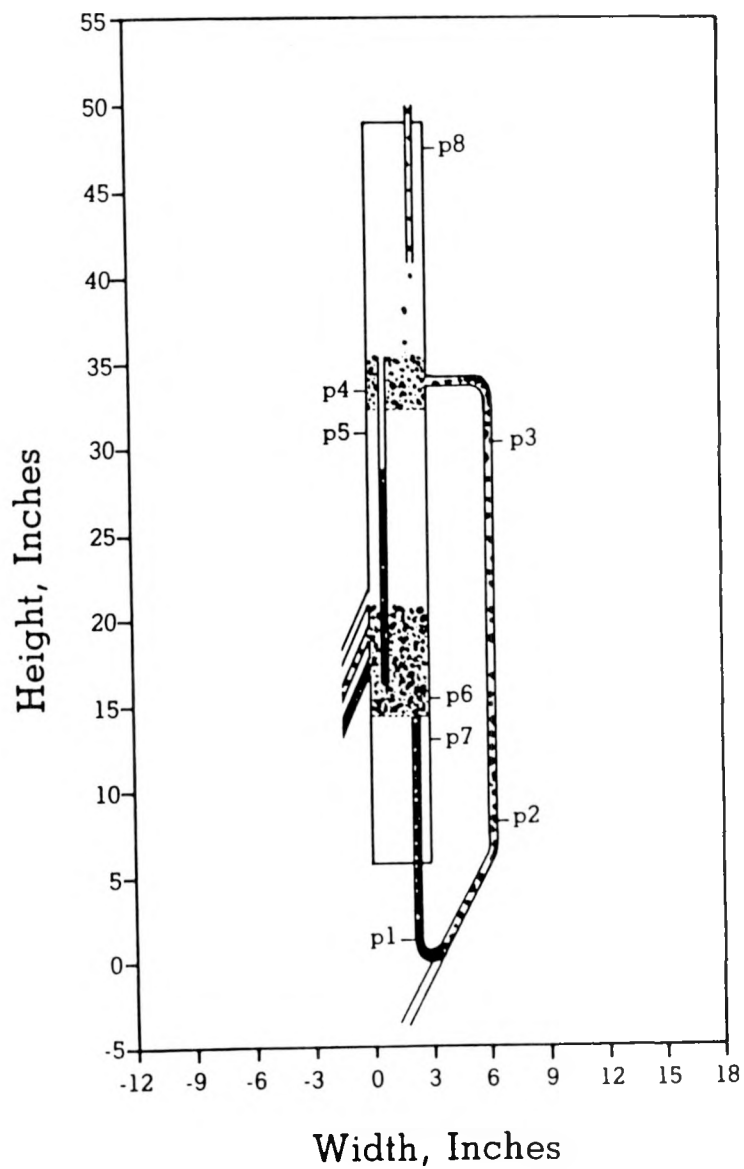


Figure 6-9. Pressure Profile for Recycle Ratio of 2:1.

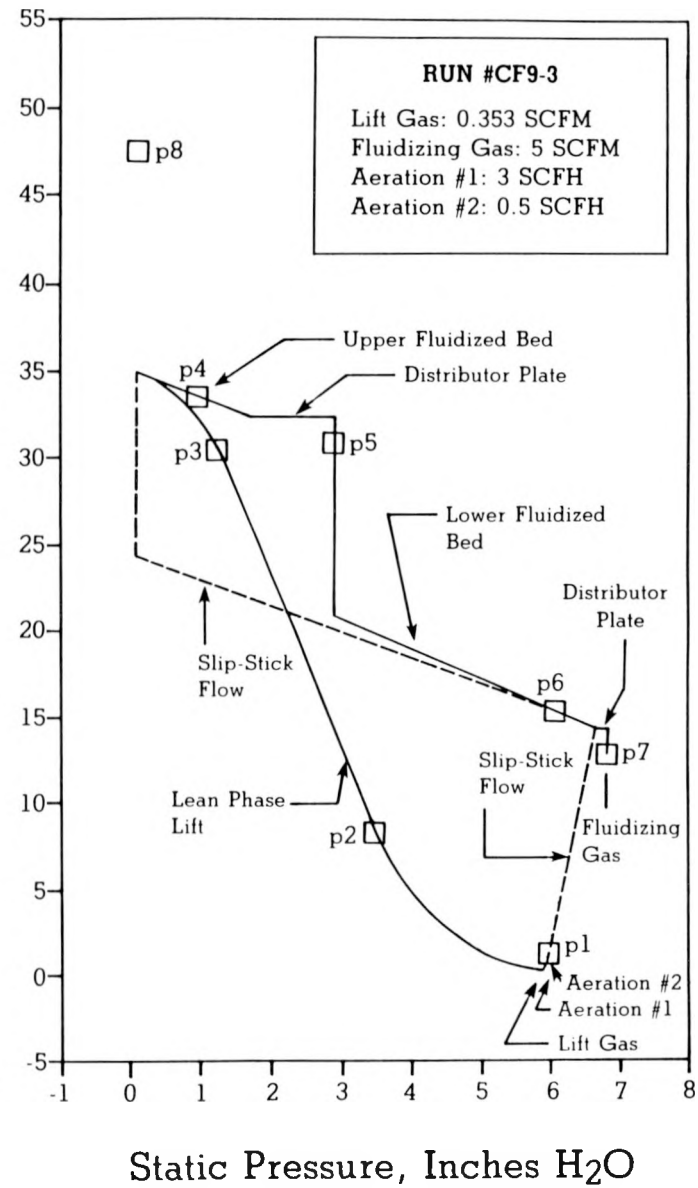
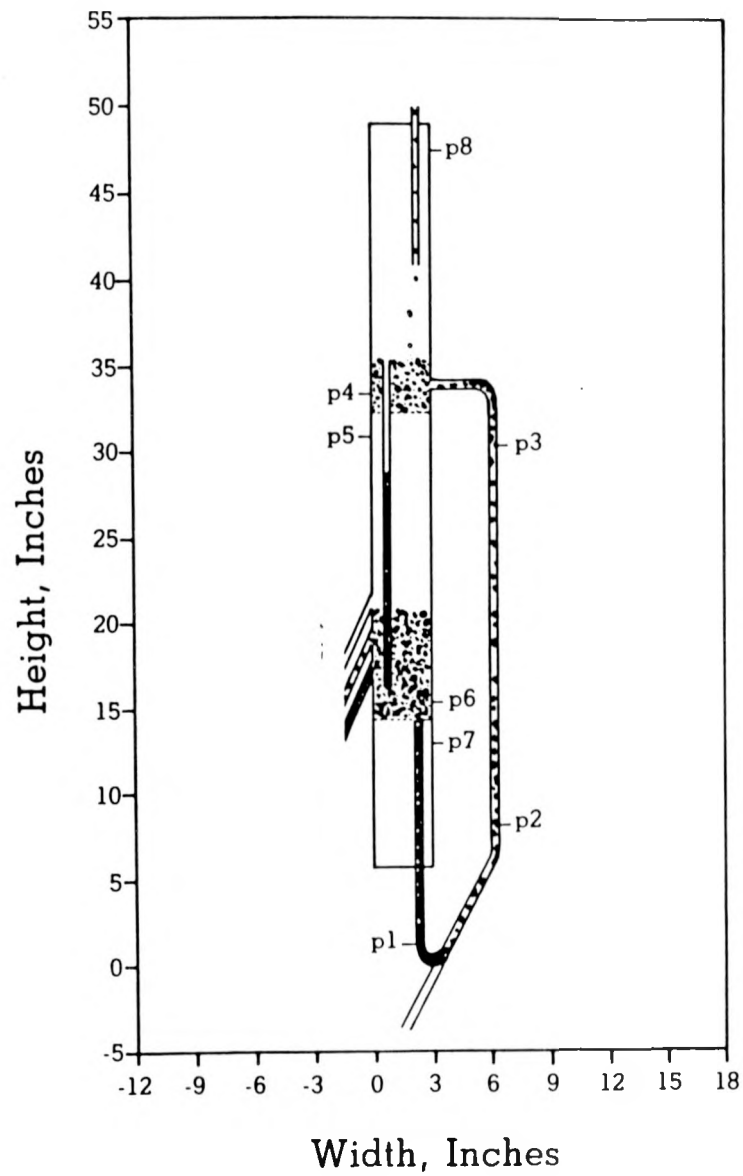
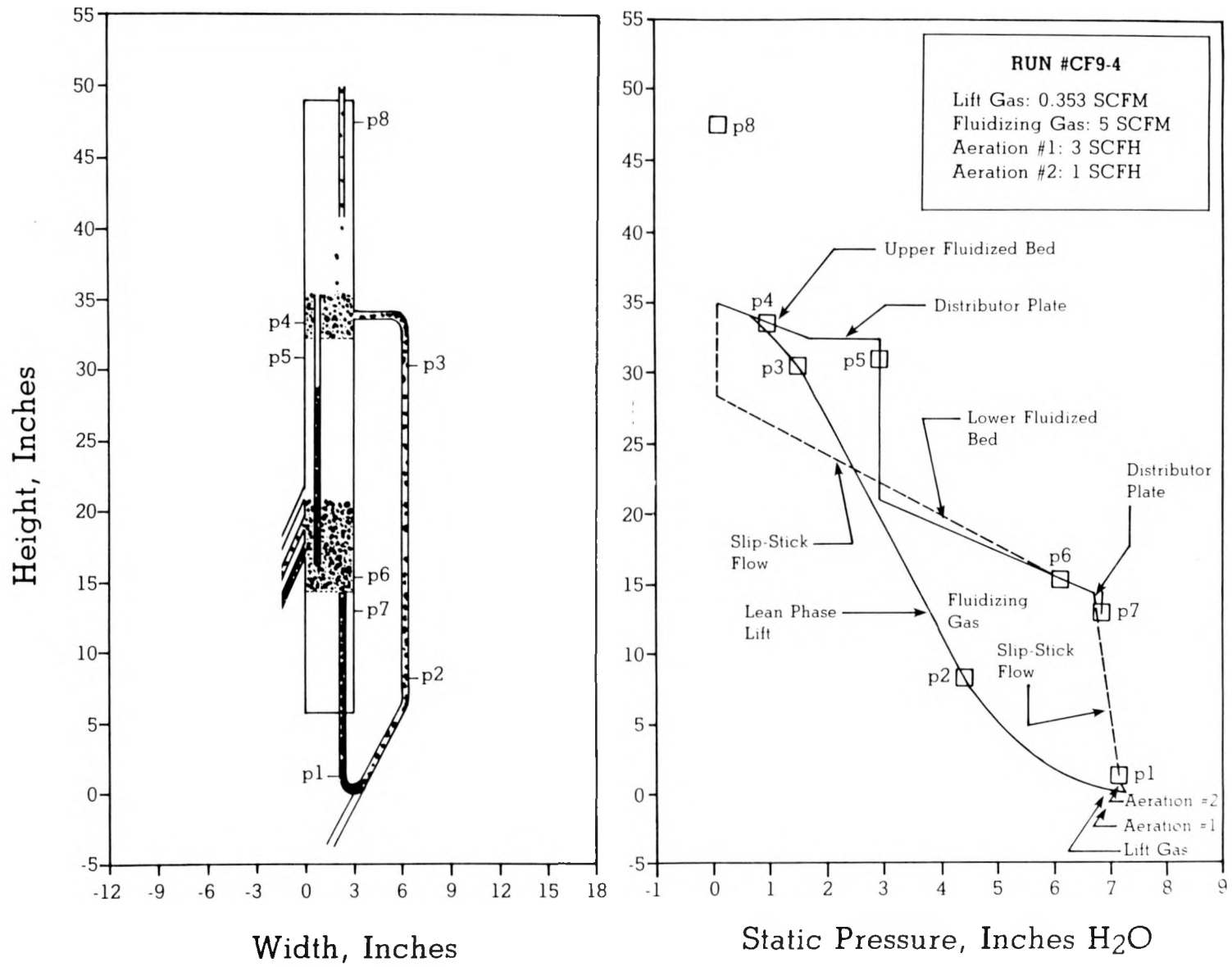


Figure 6-10. Pressure Profile for Recycle Ratio of 3:1.



6.2.7 Final Design of Cold-Flow Model. A third fluidized bed, which emulated the combustion zone of the KENTORT II reactor, was mounted adjacent to the previously discussed cold-flow system (Figure 6-2). A second J-valve and lift pipe arrangement was used to transport shale from the gasifier to the combustor. A standpipe extending through the gas distributor of the combustor maintained the bed level constant and transferred shale back to gasification section. The pressure differential between the combustor and gasifier was balanced by adjusting the back-pressure in each bed in order to keep gas flow through SP-3 to a minimum. Only the size and orientation of the standpipe connection to the gasification section remained uncertain. However, this was not as critical as the design of SP-1 since little gas flow occurred through the standpipe.

6.3 CONCLUSIONS

The cold-flow model developed during the study was found to operate smoothly and within the targeted range of parameters of the KENTORT II reactor. The design feedrate of 5 lb/hr was achieved along with the targeted solid recirculation ratio of 3:1. The solid recycle system was very stable as demonstrated by an 18-hour run. The aerated lift pipe feeder developed to control the rate of solids recirculation in the system adequately metered shale according to the flow rate and point of injection of aeration gas. Measuring the pressure drop across the lift pipe was found to be a reliable method of monitoring the rate of solid recirculation. The correlations developed at room temperature to control and monitor the rate of shale char recycle may not be extrapolatable to actual retorting conditions, but it should be possible to recalibrate at operating conditions. Operation of the unit was successful considering that the information obtained was ultimately incorporated in the mini-plant design.

6.4 REFERENCES

1. Carter, S.D. and Taulbee, D.N., Fuel Processing Technology, 11, 251-272, 1985.
2. Kunii, D. and Levenspiel, O., Fluidization Engineering, John Wiley and Sons, New York, 1969.

7. OPERATION OF INTEGRATED PROCESS

A 5-lb/hr fluidized bed system was utilized to test all three zones of the KENTORT II process in an integrated configuration. The operation of the integrated process was divided into three phases: hot-flow shakedown, pyrolysis-gasification tests, and pyrolysis-gasification-combustion tests. The hot-flow shakedown phase was a necessary extension of the cold-flow work to resolve flow problems caused by high-temperature operation. Once the reactor was capable of stable operation at gasification temperatures, an experimental program integrating the pyrolysis and gasification zones was performed. Finally, a combustion zone was integrated into the system, and a series of runs was performed with the entire KENTORT II configuration intact. This stepwise addition of processing zones served to isolate the effects of both the gasifier and combustor, and provided a means of incrementally attaining operating experience as the reactor became more complex.

7.1 HOT-FLOW SHAKEDOWN

All of the hot-flow shakedown procedures were performed with the reactor in its pyrolysis-gasification configuration. Like the cold-flow studies, once the problems were resolved in this mode, it was trivial to extend the results to the combustion configuration. While we would normally not place a lot of emphasis on a series of shake-down tests, a few operating problems were experienced that should be documented to avoid future duplication.

7.1.1 Apparatus. The pyrolysis-gasification reactor system used in this work is shown in Figure 7-1, and was essentially a replica of the cold-flow model (Figure 6-1). The upper (pyrolysis) fluidized bed region was 18-inches in height and was constructed of 3-inch O.D. tubing (type 304 SS). The gas distributor plates were perforated 11-gauge (type 310 SS) sheet with an open area of 7%. The lower (gasification) fluidized bed zone was 22 inches in height and was constructed of nominal 3-inch, schedule 40 pipe (type 310 SS).

The bed height of the gasification section was controlled by the selection of one of five overflow pipes in the side of the reactor. Additional staging was incorporated into the KENTORT design by placing a vertical baffle in each of the fluidized beds. This did not introduce any significant complexity to the reactor and improved the residence time distribution of the solids. The system adopted for recycling hot solids from the gasification zone to the pyrolysis zone was identical to that tested in the cold-flow model. An overflow standpipe maintained the upper bed at a constant height and delivered char to the gasification section. The char recirculation rate was controlled

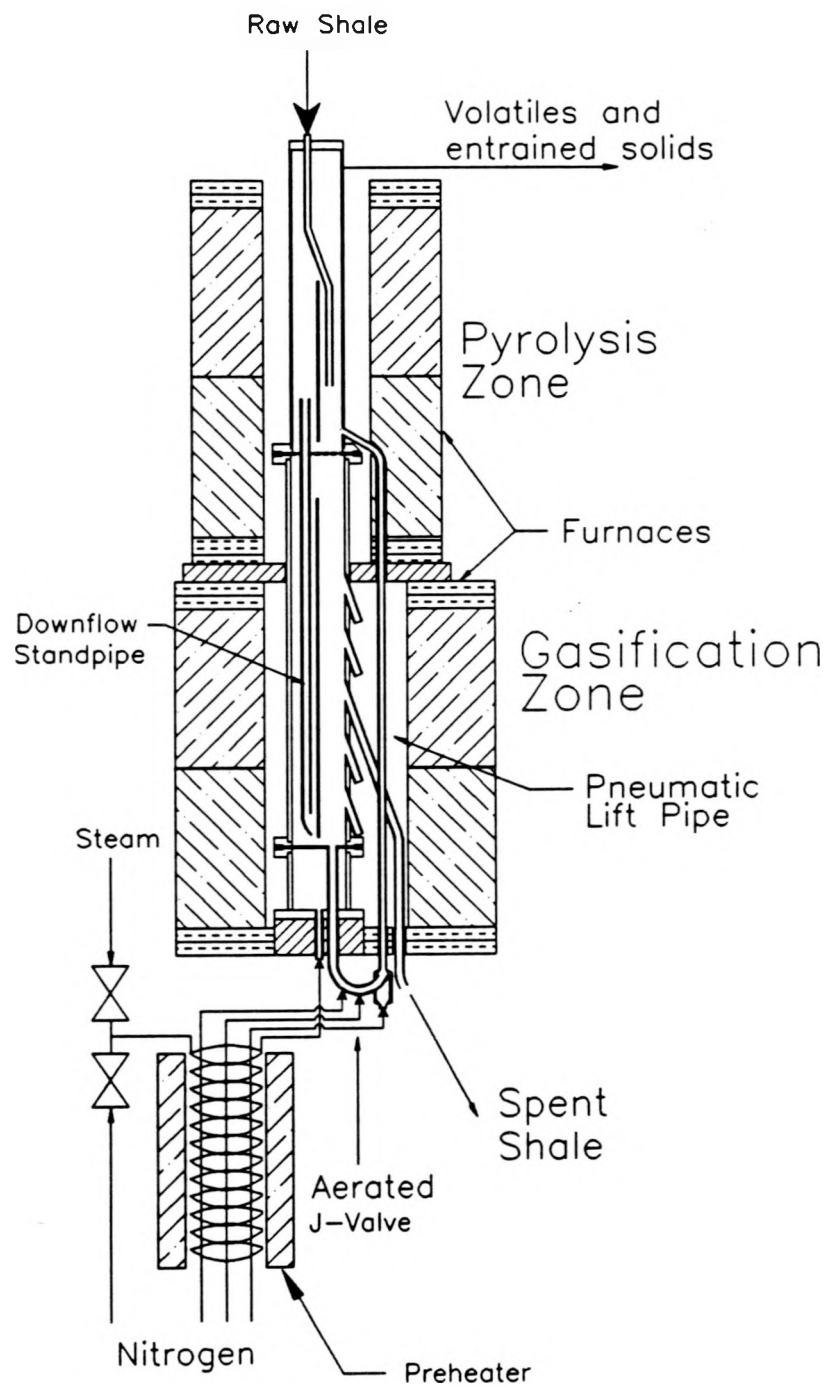


Figure 7-1. *Schematic of Integrated Reactor in the Pyrolysis-Gasification Configuration.*

by adjusting the flow rate of aeration gas to the J-valve which, in turn, regulated the amount of char entering the pneumatic lift pipe. The hot gases exited the reactor system through the top of the pyrolysis zone and were cooled and vented.

Pyrolysis temperature was controlled by the recirculation ratio and temperature of the char, the flow rate and temperature of the fluidizing gas stream, and the temperature of the surrounding furnace. The gasification section was heated by its surrounding furnace.

7.1.2 Materials Used. Since the purpose of the shakedown runs was to evaluate the physical operation of the reactor, a previously pyrolyzed shale (20 x 60 mesh) was used for the shakedown tests.(1) This material simulated the real bed material closely, and it was not necessary to collect the shale oil from pyrolysis. Nitrogen was used for fluidization since producing and condensing steam is cumbersome and was unnecessary for these tests.

7.1.3 Procedure. The reactor was preheated prior to each test while N₂ flowed through the system. Once the temperature set-points were reached, the screw feeder introduced char continuously to the reactor for the duration of the run. Process conditions were then adjusted to meet the requirements of the particular test. The reactor was allowed to cool gradually overnight with flowing N₂ before material balances were performed.

Material balances were used to accurately determine the shale feed rate. The pressure drops across each bed and the lift pipe were recorded, as were temperature readings throughout the reactor system.

7.1.4 Reactor Corrosion. Following a relatively short period of exposure (<50 hours) to temperatures above 850°C, scale formation and spalling became apparent. This corrosion only occurred in areas of the gasification section that were in direct contact with the fluidized bed of shale. The spalled scale particles were identified as roughly 50% sulfide and 50% base metal (310 SS), and measured up to 0.5 inches in width. These scale particles became lodged in various solid transfer lines throughout the system, particularly in the lift pipe feeder (J-valve), intermittently ceasing operation of the reactor. The original J-valve, having an inside diameter of 0.245 inches, was replaced with one of 0.43 inches to reduce the likelihood of blockage. This modification did diminish the problem, but reactor operation remained unsatisfactory. The installation of the larger J-valve also increased the maximum recirculation rate of the system, but the amount of increase was not determined through additional cold-flow study.

Four metallurgical coupons were immersed in the bed region of the gasifier to more fully evaluate the performance of the 310 SS alloy and to search for a better material. It should be noted that material selection for a commercial reactor would include superior refractory and ceramic materials. However, the requirement of through-the-wall heat transfer in the experimental reactor requires the use of metal. Following approximately 100 hours of exposure at temperatures exceeding 800°C, the two bare metal coupons (310ss and 304ss) were severely corroded and scale spalling was apparent. However, the two coated coupons (aluminized and chromized) fared much better, showing little or no evidence of scaling.

"Aluminizing" is a process in which aluminum diffuses into the surface of metals for a minimum depth of 100 microns at high temperature, producing an aluminum-rich alloy on the surface. Aluminized Incoloy 802 tubes in an ethylene pyrolysis furnace at 1038°C have been reported to remain virtually intact after 3-1/2 years of industrial service.(2) Considering the favorable results we achieved with the aluminized test coupon and published reports of extended industrial service, the decision was made to fabricate and install a new, aluminized gasification section.

7.1.5 Gas Distributor Modification. In addition to corrosion, the gasification section was found to create another problem unforeseen by cold-flow modelling. Coke formation and entrained fines from the gasifier restricted the flow of gas through the holes (0.0625" dia.) of the pyrolysis gas distributor. During the course of a run, the pressure drop across the distributor increased until the fluidizing gas was forced to mainly flow through the standpipe connecting the pyrolyzer and gasifier (see Fig. 7-1). When this occurred, spent shale from the pyrolyzer could no longer spill into the standpipe, and the upper fluidized bed grew uncontrollably, forcing run termination. By fabricating a new gas distributor with fewer, larger diameter holes (0.1875" dia.), entrained fines could pass freely through the distributor, and coke did not form in the holes.

An additional operating benefit was gained by enlarging the holes of the gas distributor. The holes were made large enough that shale particles were able to pass through the distributor plate when the bed was slumped. Slumping the beds was found to be an important method of transferring shale from the upper bed to the lower bed following a process upset where the dipleg of the standpipe became uncovered. Once the dipleg was no longer immersed in the bed, shale could no longer pass downward through the standpipe because of excessive gas bypassing through the standpipe.

7.1.6 Operation. A log of typical run conditions during hot-flow operation is shown in Figure 7-2. This run was performed with the original lift pipe feeder and at a low gasifier temperature to minimize scale spalling. The step change in the lift pipe differential pressure in Fig. 7-2 represents the initiation of char recycle. The general temperature drop and the rapid rise of the gasifier bed pressure drop following the first 5 minutes of the run log indicates the initiation of char feed to the reactor.

The pyrolyzer temperature closely followed transitions from char recycle to non-recycle. Even short periods of char recirculation intended only to check the operability of the lift pipe feeder registered spikes in the pyrolyzer temperature log. The gasifier temperature was affected by char recycle, but not as significantly as the pyrolyzer. The pressure drop across the gasifier bed was only affected transiently by the step changes in char recycle. The bed inventory dropped slightly at the onset of char recycle and increased slightly following a period of sustained recirculation. The recorded lift pipe temperature did not appear to be affected by changes in char recirculation. It is not clear whether this temperature reading is a measure of the flowing contents of the lift pipe or if it is more a measure of the tube wall temperature.

7.1.7 Replacement of Gasification Section. A duplicate of the original gasification reactor was fabricated of 310 SS and aluminized inside and out. All surfaces in contact with the fluidized bed of solids in the gasification section were aluminized, including thermowells, baffles, standpipes, etc. The pyrolysis and gas preheating zones of the reactor did not require alteration. Hot-flow testing of the new gasification zone revealed no corrosion of the type experienced previously with the bare 310 SS alloy. Moreover, no corrosion related problems were experienced with aluminized materials for the remainder of the program.

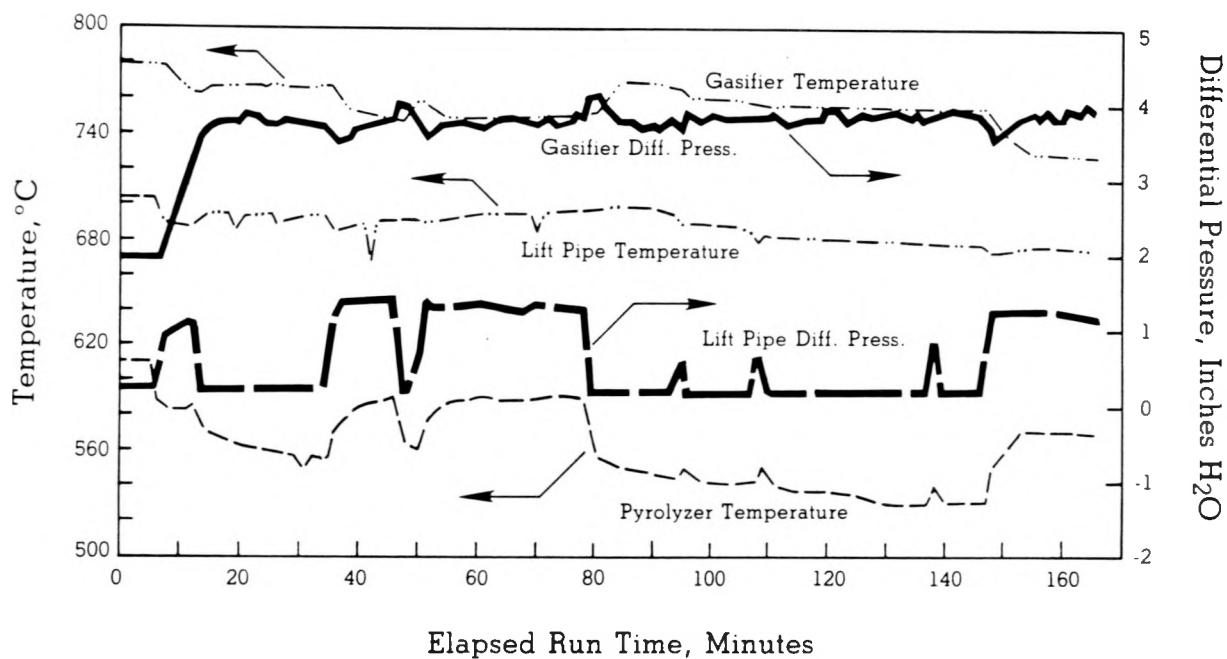


Figure 7-2. Log of Conditions for Hot-Flow Run#4 in the Integrated Pyrolysis-Gasification Reactor with a Char Feed Rate of 3.8 lb/hr and N₂ as the Fluidizing Gas.

7.1.8 Summary. The gasification section of the reactor was aluminized in order to alleviate the corrosion problems experienced with the bare 310 SS alloy. Additionally, the holes of the gas distributor for the pyrolysis zones were enlarged to prevent blockage by coke and shales fines. Following these modifications, the integrated pyrolysis-gasification reactor system demonstrated the capability for achieving the design reactor conditions established for Task 1 of the KENTORT II development program.

7.2 PYROLYSIS-GASIFICATION STUDY

The reactor was configured with integral pyrolysis and gasification zones for this phase of the study. The combustion zone was not in place so that the effects of combusted shale could be more easily isolated later in the program. This stepwise approach also aided the operation of the unit since experience could be gained with the relatively simple two-stage system before running the system with 3 beds. There were two major experimental objectives for this part of the program:

- o To investigate the gasification kinetics of nascent oil shale char. In order to obtain a realistic evaluation, shale char should enter the gasification section directly from the pyrolyzer without cool-down. Information regarding carbon and sulfur conversion as a function of mean char residence time and temperature is essential to the optimization of the KENTORT II process.
- o To investigate the effects of char recirculation between the pyrolysis and gasification sections on pyrolysis performance. By understanding the effects of char temperature and recirculation ratio it was expected that oil coking and cracking losses could be minimized. It was also important to study this without shale combustion so that the effects of combusted shale in the system could be more easily discerned.

7.2.1 Apparatus. The integrated pyrolysis-gasification reactor system was configured as shown in Figure 7-3 for this phase of operation. Process heat for the reactor was provided through a combination of superheated steam and external heat to the gasification section. The pyrolysis zone was heated by a combination of hot gases and recirculating hot char from the gasification section. The liquid collection system was developed from similar systems used in the past for pyrolysis studies.(1) The main features of the system include: 1) an electrostatic precipitator which traps the majority of the oil from its aerosol suspension, 2) a water-cooled condenser which traps the more volatile oil components, and 3) on-line gas analysis capable of quantitating vapor phase C_5+ materials.

7.2.2 Oil Shale Sample. The Cleveland oil shale sample described in Section 3 was utilized for this study. See Table 7-1 for the average analyses of the sample.

7.2.3 Procedure. The reactor system was brought to operating temperature with the electrical resistance furnaces while nitrogen flowed through the system to properly distribute heat. Following the preheating period, the nitrogen flow was replaced with steam, and the system temperatures and pressures

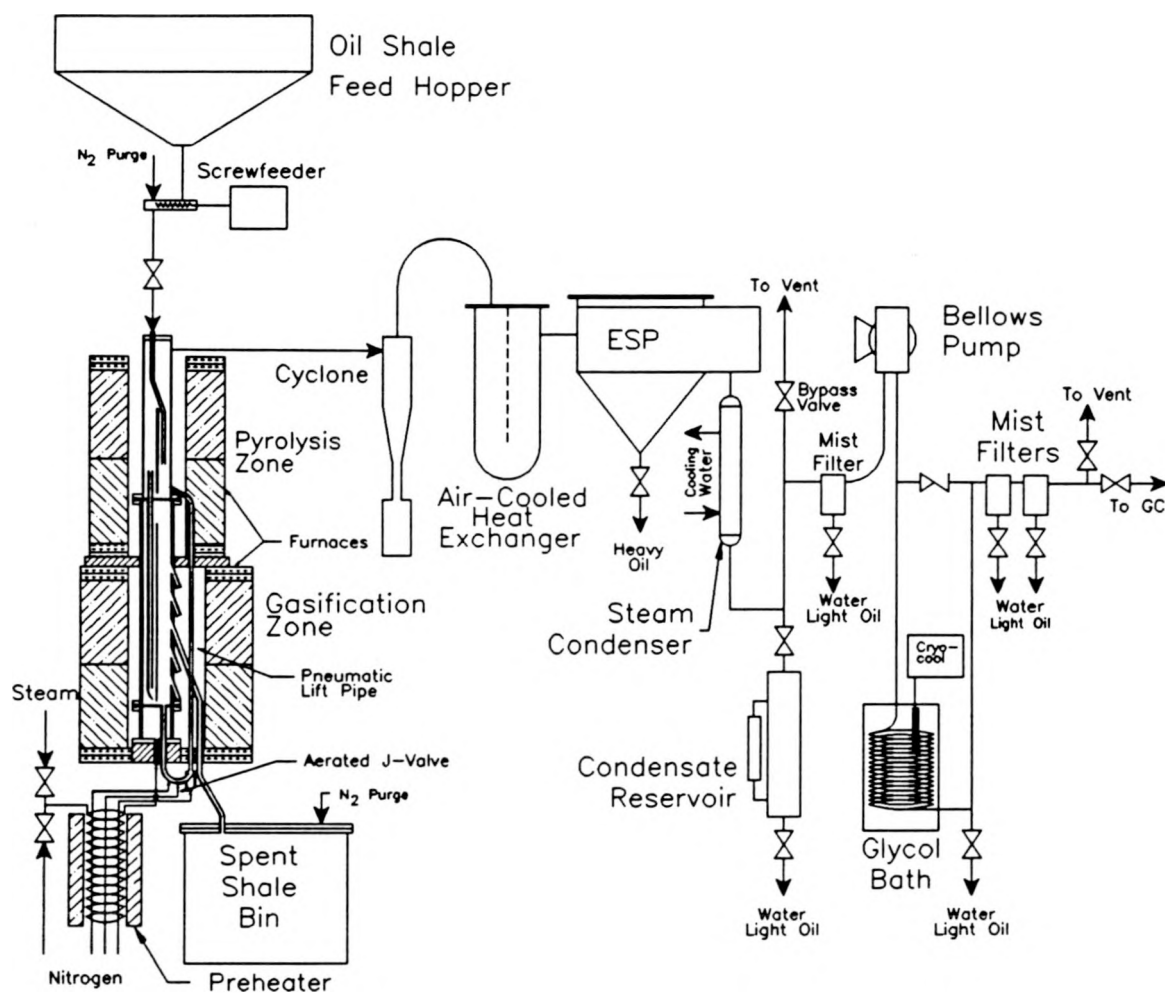


Figure 7-3. *Simplified Flow Diagram of the Integrated Pyrolysis-Gasification Reactor System.*

Table 7-1. *KENTORT II Master Sample Characteristics.*
Modified Fischer Assay*

Carbon	16.7 wt%	Oil	17.7 gal/ton
Hydrogen	2.0 wt%	Oil	7.0 wt%
Nitrogen	0.7 wt%	Water	4.2 wt%
Sulfur	1.7 wt%	Gas+Loss	2.6 wt%
Ash	74.5 wt%		

* Modified Fischer Assay developed at CAER for Eastern shale. (Coburn, T.T. and Rubel, A.M., Fuel Processing Technology, 8 , 1, 1983)

were allowed to stabilize. Subsequently, raw shale feed was initiated and final system adjustments were made. Shale recirculation, when applicable, was initiated when shale had filled the pyrolyzer and gasifier to near capacity (typically, 10-20 minutes into the run). Steam and light oil condensation was periodically drained from the system. On-line gas samples were taken every 35 minutes for gas chromatography. Otherwise, conditions were maintained as constant as practicable, and recorded by a computerized data acquisition system for the duration of the run (typically 2 to 5 hours).

The flow of steam was halted 10 minutes following the termination of shale feed, and the system was cooled overnight with flowing nitrogen. Liquids were drained from the various traps in the system following the termination of steam flow. The collection of solid samples and the quantitation of liquid hold-up in the system were performed on the following day. Oil held up in the system was rinsed from the internals with CH_2Cl_2 , filtered to remove solids, and then recovered by distillation. Solids recovered in the spent shale bin and cyclone were weighed, and sampled for ultimate analyses. The recovered oil fractions were analyzed separately and as a whole according to production ratios.

7.2.4 Experimental Design. The parametric study during this phase of the program was performed in two steps. The first step evaluated the oil yields generated with no solids recirculation. These runs (KEN4-11 in Table 7-2) provided baseline data for the solid recycle runs to follow. Additionally, the effects of gasifier temperature and mean char residence time on carbon and sulfur removal were investigated. This range of conditions was chosen from the data produced by the bench-scale unit (see Section 4) as being appropriate for practical gasifier operation. The mean shale holding times in the pyrolyzer and gasifier were controlled by the raw shale throughput because the volumes of the beds were held constant. Pyrolyzer temperature was maintained as constant as possible so that the char feed to the gasification section was similar for each run. The average holding times in the pyrolysis zone were more than sufficient for complete pyrolysis (>98%) to occur. Recent findings by Coburn et al. (3) have indicated that 2 minutes is ample time for the majority of oil to be generated. The pyrolysis zone for these runs was heated exclusively by the upward flowing gases from the gasifier and external heat through the walls of the reactor.

The second part of the test matrix (KEN16-36 in Table 7-2) was designed to evaluate the effects of char recirculation as a method of heat transfer between the gasification and pyrolysis zones. Two major variables were studied: char recirculation rate and char temperature. The temperature of the recycle shale was varied through control of the gasification temperature, and

Table 7-2. *Average Run Conditions for the 5-lb/hr Integrated Pyrolysis-Gasification Reactor.*

Run #	Mean Bed* Temp., °C		Shale Feed Rate, g/min.	Mean Solid Res. Time, min.		Solid Rec. Temp., °C		Heat Provided by Steam for Pyrolysis %	Solid Recirc Ratio
	Pyrol	Gasif		Pyrol	Gasif	Meas.	Avg.		
KEN 4	526	802	19.72	14.8	30.4	-	-	100	0
KEN 5	543	840	20.64	14.1	29.1	-	-	100	0
KEN 6	526	839	15.92	18.3	37.7	-	-	100	0
KEN 7	533	802	17.82	16.4	33.7	-	-	100	0
KEN 8	532	803	34.84	8.4	17.2	-	-	100	0
KEN 9	535	831	33.38	8.7	18.0	-	-	100	0
KEN10	541	832	30.50	9.6	19.7	-	-	100	0
KEN11	539	805	30.00	9.7	20.0	-	-	100	0
KEN16	524	735	32.14	9.1	25.5	689	712	54.6	1.58
KEN18	526	756	31.42	9.3	26.0	706	731	50.3	1.57
KEN20	518	747	29.46	9.9	27.8	689	718	62.3	1.15
KEN24	529	812	30.30	9.6	27.0	734	773	60.5	1.00
KEN25	526	780	38.61	7.6	21.2	709	745	60.7	1.21
KEN27	533	771	39.88	7.3	20.5	673	722	52.9	1.91
KEN28	544	783	51.71	5.6	15.8	689	736	40.2	2.44
KEN29	546	796	46.73	6.2	17.5	701	749	43.9	2.24
KEN30	536	754	48.54	6.0	16.9	678	716	41.2	2.52
KEN31	537	827	48.30	6.0	16.9	727	777	49.7	1.50
KEN32	527	787	48.28	6.0	16.9	705	746	42.6	1.83
KEN33	530	760	47.25	6.2	17.3	687	723	38.8	2.22
KEN34	530	763	47.49	6.1	17.2	687	725	39.7	2.21
KEN35	536	839	47.28	6.2	17.3	745	792	46.7	1.48
KEN36	532	795	46.18	6.3	17.2	710	753	45.0	1.76

* Reactor operated at 0 psig $\pm$ 5 inches water column.

the solid recycle ratio (rate of recycled char/raw shale feed rate, w/w) was adjusted to achieve the targeted pyrolysis temperature. The mean shale residence time in the pyrolyzer did not change significantly due to shale recycle, but the time distribution within the reactor was changed. Because of solid recycle the time a newly introduced shale particle spent in the pyrolyzer before going to the gasifier averaged approximately 2 minutes. The average particle was then recycled to the pyrolyzer about two times, but this additional residence time in the pyrolyzer did not contribute to oil generation.

Solid recycle ratios were determined by performing a heat balance around the pyrolyzer. The enthalpy relations for eastern U.S. oil shale derived by Camp were applied to the specific characteristics of the oil shale sample used in this study.(4) It was assumed that no net heat loss occurred through the walls of the pyrolyzer because of heat compensation from the external furnace. The temperatures of the gas and the solids in the pyrolyzer were assumed to be equal and were measured by thermocouples immersed in the bed. The measured solid recycle temperature was consistently lower than the gasifier temperature (see Table 7-2). The measured solid recycle temperature was somewhat low because of heat loss through the sheath and tube fitting securing the thermocouple. However, some temperature

drop in the recycle shale temperature was expected because of heat loss at the lift pipe feeder which was located outside the furnace (see Figure 7-3). Therefore, the true solid temperature fell somewhere between, so the average of these two readings will be used for the later discussions.

A complete listing of measured process parameters and calculated heat balance values for all runs that achieved steady state are shown in Appendix A.

7.2.5 Mass and Component Balances. The absolute and relative amounts of solid, liquid, and gas produced from all runs that achieved steady state conditions are included in Appendix B. The ultimate analysis of each collected fraction and the calculated carbon, nitrogen, and sulfur balances are also shown in Appendix B. Only runs that had 90+% carbon balances were considered for further analysis. The carbon, nitrogen and sulfur balances for these runs are summarized in Table 7-3. It was determined that 1-3% of the carbon was not measured from each run because of nonsteady state gasification and degassing of spent shale. Therefore, if a run was slightly below 90% carbon balance, but was superior in other aspects such as run length and constancy of run conditions, it was included for analysis.

Table 7-3. Carbon, Nitrogen and Sulfur Balances for Pyrolysis-Gasification Runs.

KEN#	Spent Shale			Oil			Gas			Water			Total		
	C	N	S	C	N	S	C	N	S	C	N	S	C	N	S
4	28.3	46.4	11.4	53.5	15.6	9.2	13.4	NA	55.3	2.9	NA	0.5	98.1	NA	76.3
5	28.7	46.1	13.5	43.8	12.8	8.4	18.4	NA	69.9	2.8	NA	9.7	93.7	NA	101.4
6	23.8	46.8	11.2	46.5	9.8	8.2	22.6	NA	72.7	3.5	NA	9.5	96.3	NA	101.6
7	29.7	54.1	11.1	49.1	10.1	8.4	14.8	NA	64.9	3.6	NA	0.1	97.2	NA	84.5
8	35.3	60.5	19.9	42.0	10.6	9.4	12.4	NA	54.1	1.8	NA	8.4	91.5	NA	91.7
9	30.1	52.4	15.6	42.8	9.1	7.1	15.6	NA	73.8	2.0	NA	0.8	90.5	NA	97.2
10	29.0	50.7	12.0	46.7	10.5	8.6	15.5	NA	64.0	1.9	NA	1.0	93.0	NA	85.4
11	31.5	58.7	14.7	42.2	11.7	6.7	14.7	NA	71.2	2.0	NA	0.4	90.5	NA	93.2
16	42.9	77.2	25.8	43.5	12.3	9.1	8.5	NA	64.6	2.0	NA	3.2	97.0	NA	102.7
18	39.0	67.7	33.0	44.0	13.3	6.8	8.8	NA	57.4	1.3	NA	2.3	93.1	NA	99.5
20	39.2	47.4	28.0	49.0	9.9	7.7	8.6	NA	57.7	1.8	NA	2.5	98.6	NA	96.0
24	36.6	53.3	20.8	42.0	12.9	5.6	14.0	NA	53.8	1.7	NA	2.3	94.1	NA	82.5
25	39.4	61.0	22.9	43.3	18.5	6.0	9.6	NA	51.1	2.0	NA	2.4	94.3	NA	82.4
27	39.3	59.2	25.8	40.7	12.9	6.9	9.0	NA	56.1	1.8	NA	2.2	90.7	NA	91.0
28	36.8	61.3	25.8	41.2	11.4	6.8	9.0	NA	48.6	1.9	NA	2.6	89.0	NA	83.9
29	35.6	43.9	24.4	43.4	11.2	7.8	11.4	NA	65.7	1.8	NA	2.4	92.2	NA	100.3
30	40.7	85.6	27.4	40.3	7.6	8.0	8.7	NA	68.8	2.0	NA	2.3	91.7	NA	106.5
31	33.4	20.3	22.1	39.6	9.9	6.4	13.1	NA	55.0	1.7	23.0	1.9	87.9	NA	85.4
32	41.8	28.2	32.2	40.2	10.8	6.2	10.2	NA	50.8	1.6	18.1	1.4	93.8	NA	90.6
33	40.4	30.2	34.9	39.9	10.7	6.7	9.1	NA	48.5	1.5	15.5	1.6	90.9	NA	91.8
34	38.0	26.7	31.1	40.6	11.0	7.1	9.4	NA	46.3	1.6	16.1	1.6	89.6	NA	86.0
35	32.0	19.3	25.5	39.5	10.7	6.4	15.9	NA	45.2	1.2	21.4	1.5	88.6	NA	78.6
36	36.8	24.4	28.8	40.6	10.9	6.7	11.4	NA	49.8	1.8	17.7	1.7	90.6	NA	87.0

Nitrogen and sulfur balances were generally less precise than the carbon balances. Nitrogen was difficult to track because the levels in the raw and spent shale were in most cases at the detection limits of the Leco analyzer. Moreover, nitrogen species in the gas stream were not monitored, and the nitrogen in the water phase was performed on only selected samples. Nitrogen in the spent shale and the water for KEN#31-36 was determined by the Kjeldahl procedure. The nitrogen balance deficiencies for these runs indicate that a significant portion of the nitrogen exited in the form of ammonia. Sulfur partitioning was easier to monitor than nitrogen, but some difficulties remained. The sulfur content of the spent shales was low and again near the detection limits of the sulfur analyzer, causing these determinations to exhibit more scatter than the raw shale. Additionally, hydrogen sulfide measurements in the gas phase were somewhat erratic. Hydrogen sulfide eluted the GC column as a very broad peak that was difficult to accurately integrate, and the calibration of this peak varied more than the other peaks on the column.

7.2.6 Char Gasification-Nonrecycle Runs. The extent of carbon gasification for Eastern shales can be conveniently represented by the yields of CO plus CO₂ since only barely detectable levels carbonates are present in the raw shale. A fluidized bed pyrolysis study of eastern shale showed that only small amounts of CO and CO₂ were produced under similar high oil yield conditions using steam for fluidization.(1) Therefore, the measured carbon oxides were primarily generated from gasification reactions. The yield of carbon oxides was found to increase as a function of gasifier temperature and to a lesser extent as a function of mean solid residence time (see Figure 7-4). The increase in carbon oxide production at increasing gasifier severity was reflected by a corresponding decrease in carbon content of the spent shale. The latter relationship was possible because a similar amount of carbon was released upon pyrolysis by maintaining a constant run to run pyrolysis temperature.

The most abundant gas component, by volume, was H₂, followed by lesser quantities of CO and CO₂ which were present in nearly equal proportions (Figure 7-5). It is emphasized that the vapor phase products of pyrolysis and gasification were combined as a result of the design of the reactor, so it was sometimes difficult to clearly distinguish the pyrolysis zone reactions from those occurring in the gasifier. It was assumed that changes in gas make were due to the variation in gasifier conditions because the pyrolysis zone conditions were held constant throughout the study. Only H₂, CO, and CO₂ were significantly affected by gasification conditions. Methane yields were higher than those obtained from pyrolysis at 530°C, but were little changed over the gasifier temperatures studied. Otherwise, the balance of vapor phase components from pyrolysis

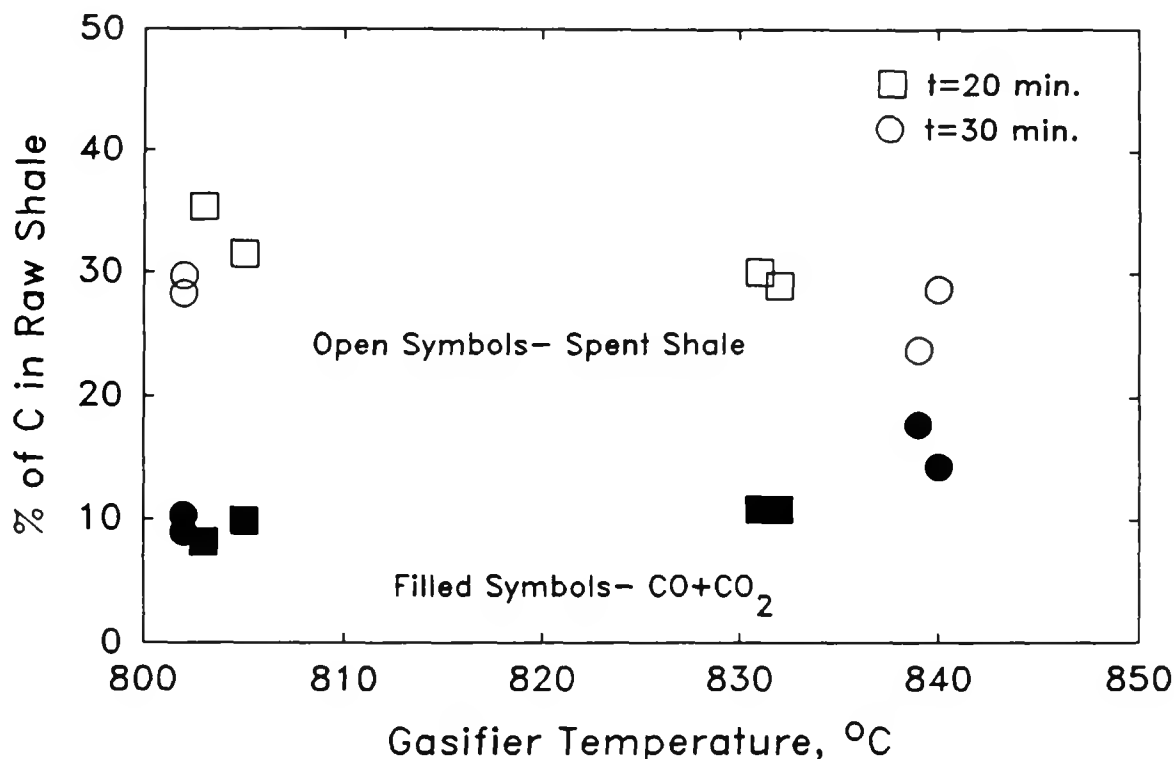
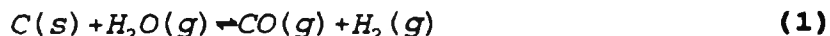


Figure 7-4. *Effect of Gasification Temperature and Residence Time on the Production of Carbon Oxides and the Removal of Carbon from the Pyrolysis Char.*

(C₂+ hydrocarbons) were similar in yield and composition to previous pyrolysis investigations.(1)

The predominant carbon gasification reaction for this study was the steam-carbon reaction (Eq. 1) because of its thermodynamic favorability and the large partial pressure of



water vapor utilized. Although thermodynamically favorable, it was improbable that the Boudouard reaction (Eq. 2) proceeded significantly because of the low partial pressure of CO₂.



Hydrogasification (Eq. 3) was discounted as a primary gasification reaction because of the relatively low hydrogen partial pressure and its thermodynamic unfavorability at the temperature range studied.



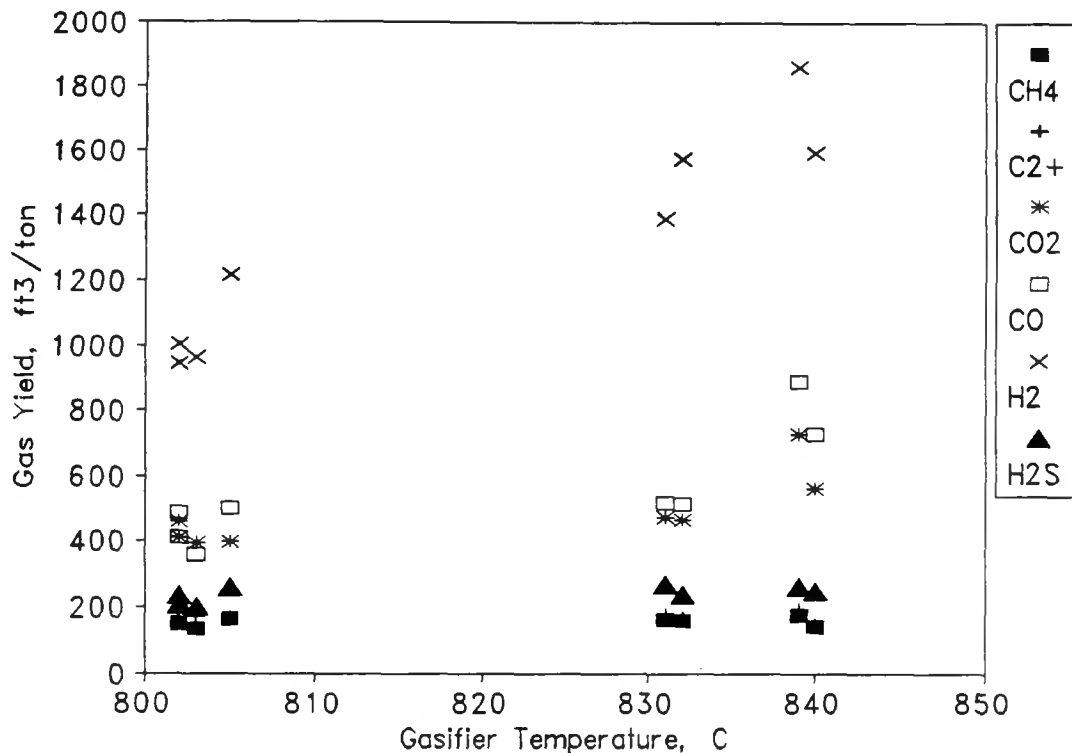
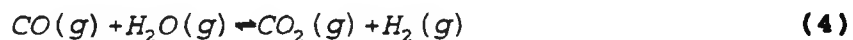


Figure 7-5. Effect of Gasifier Conditions on the Production of Major Gas Components. Mean Char Residence Time, τ , was 20-30 Minutes.

The main sources of hydrogen, which increased in yield with gasifier temperature (Figure 7-5), are considered to be the carbon/steam reaction (Eq. 1) coupled with the water gas shift reaction:



The proportions of H_2 , CO_2 , and CO that were observed correspond well with the hypothesis that Eqs. 1 and 4 are the dominant reactions under the conditions of the study. The $[\text{H}_2]:[\text{CO}_2]$ and $[\text{H}_2]:[\text{CO}]$ ratios ranged from 2.06-3.37 and 1.94-3.05, respectively (see Table 7-4). Stoichiometrically, these are approximately the ratios that would result from 50% conversion of CO by water gas shift, and the observed $[\text{CO}_2]:[\text{CO}]$ ratios (0.77 to 1.1) indicate that indeed roughly 50% conversion took place. As shown in Table 7-4, the gas compositions were well below the equilibrium values at gasifier temperature for Eqs. 1 or 4, or the combination of the two. Moreover, in no part of the system was the average bulk composition of the gas found to correspond to equilibrium conversion. Therefore, it appears that the reactions are kinetically controlled.

The H_2 to CO_x ratios for this study were below those observed from the bench-scale study in Section 4. In that study the proportionately high hydrogen production was primarily attributed to high temperature pyrolysis of the char. It is somewhat surprising that the same did not occur in the integrated reactor. With the added complexity of the integral pyrolysis zone, it is difficult to elucidate this difference between the two studies. It is possible that some hydrogen scavenging occurred in the pyrolysis zone.

Higher levels of methane production were measured in the integrated reactor compared to previous fluidized bed pyrolysis studies. High temperature secondary pyrolysis of char is the most likely source of this observed increase. Methanation reactions (Eqs. 5,6) are not considered a likely source of methane because the high water vapor concentration forces these reactions to the left.

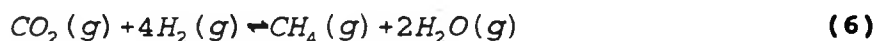
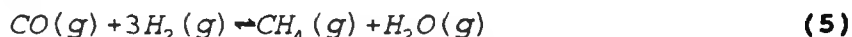


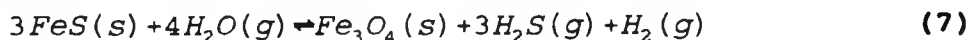
Table 7-4. Gas Compositions and Selected Ratios for Non-Recycle Gasification Runs.

KEN#	Partial Pressure, atm.						
	CH ₄	CO ₂	CO	H ₂	H ₂ S	N ₂	H ₂ O
4	0.0015	0.0042	0.0042	0.0102	0.0021	0.1199	0.8558
5	0.0015	0.0061	0.0079	0.0172	0.0027	0.1273	0.8357
6	0.0017	0.0070	0.0085	0.0178	0.0025	0.1461	0.8146
7	0.0014	0.0042	0.0045	0.0087	0.0021	0.1375	0.8399
8	0.0024	0.0070	0.0064	0.0171	0.0036	0.1470	0.8135
9	0.0026	0.0077	0.0085	0.0226	0.0044	0.1298	0.8217
10	0.0025	0.0075	0.0083	0.0252	0.0038	0.1267	0.8234
11	0.0025	0.0062	0.0078	0.0190	0.0041	0.1276	0.8303
KEN#	$\frac{[CO_2][H_2]^2}{[H_2O]^2}$		$\frac{[CO_2][H_2]}{[CO][H_2O]}$		$\frac{[CO][H_2]}{[H_2O]}$		
4	5.98E-7		0.0120		0.00005		
5	2.59E-6		0.0159		0.00016		
6	3.34E-6		0.0180		0.00019		
7	4.54E-7		0.0098		0.00005		
8	3.10E-6		0.0231		0.00013		
9	5.81E-6		0.0250		0.00023		
10	7.01E-6		0.0277		0.00025		
11	3.25E-6		0.0182		0.00018		
Equil, 727 C	3.606E0		1.378		2.617		

The kinetics of carbon gasification from integrated processing were similar to the results found during the bench-scale study (see Section 4).(5) No enhanced conversion was observed for the direct transfer of char from the pyrolyzer to the gasifier without cool-down. It also appears that the method of pyrolysis of Eastern shales has little or no effect on char reactivity.(6) Studies regarding the structure and reactivity of Eastern shale chars and the influence of mineral matter are needed to shed some light on this issue.

The carbon conversion in the gasifier was somewhat low (5-20%) because of the relatively mild conditions employed during this study. Even so, sufficient H_2 gas was produced to completely upgrade the crude shale oil under most circumstances. The reactor was operated at the low end of what are generally considered 'gasification conditions'. There were two reasons for this. The first is that short residence times and lower temperatures make the reactor smaller and less expensive. Secondly, substantial carbon conversion is not the function of the gasification zone in the KENTORT process. Not only must sufficient carbon remain to fuel the system, but there are beneficial attributes of the gasifier besides carbon conversion that can be attained under mild conditions.

Sulfur Removal. In contrast to carbon conversion, sulfur removal proceeded more quickly and nearly to completion in the gasifier. Pyrrhotite conversion by water vapor (Eq. 7) is the final reaction in the sulfur removal sequence.



In terms of inorganic sulfur removal, water vapor is much more effective than hydrogen because hydrogen cannot reduce iron sulfides further than pyrrhotite. Mossbauer spectrometry of the spent shales confirmed the occurrence of Eq. 7 by revealing substantial quantities of Fe_3O_4 . For the conditions studied, very little temperature or solid residence time dependence was observed for sulfur removal (see Figure 7-6). Both H_2S production and residual sulfur can be used as a measure of sulfur removal because relatively small and constant amounts of sulfur are distributed to liquid phase products. Hydrogen sulfide was the only gas phase sulfur species detected.

Even under the mildest conditions studied, 70% of the sulfur in the raw shale was removed and most was converted to H_2S . This has beneficial implications for the KENTORT II process since it relaxes the process parameter requirements for one constraint

(i.e. sulfur removal) and allows gasifier conditions to be regulated by the targeted carbon removal requirements.

7.2.7 Oil Yield-Nonrecycle Runs. While this report has dealt almost exclusively with gaseous and solid products of the KENTORT process, oil remains the most important product of the process. In the nonrecycle mode, oil constituted up to 50% of the carbon in the raw shale (see Table 7-3). These yields are comparable to the maximum yields achieved during previous fluidized bed pyrolysis studies.(1) Five percent of the carbon weight was attributed to C_1 - C_4 hydrocarbons produced in the pyrolyzer, and up to 3% of the carbon was collected in the aqueous phase. Only the fraction of carbon distributed between carbon oxides and spent shale changed significantly with the addition of the gasification section.

No correlations to pyrolysis or gasification conditions have been discerned for oil yields from nonrecycle runs. This is not surprising since the pyrolysis temperature was held nominally constant among the runs. The average collected oil yield for the gas-heated runs, expressed as a percentage of the carbon in the raw shale, was 46.5%. This corresponds to 129% of the modified Fischer assay oil yield on a carbon basis. These runs represent the baseline oil yields for comparison to the solid-recycle runs

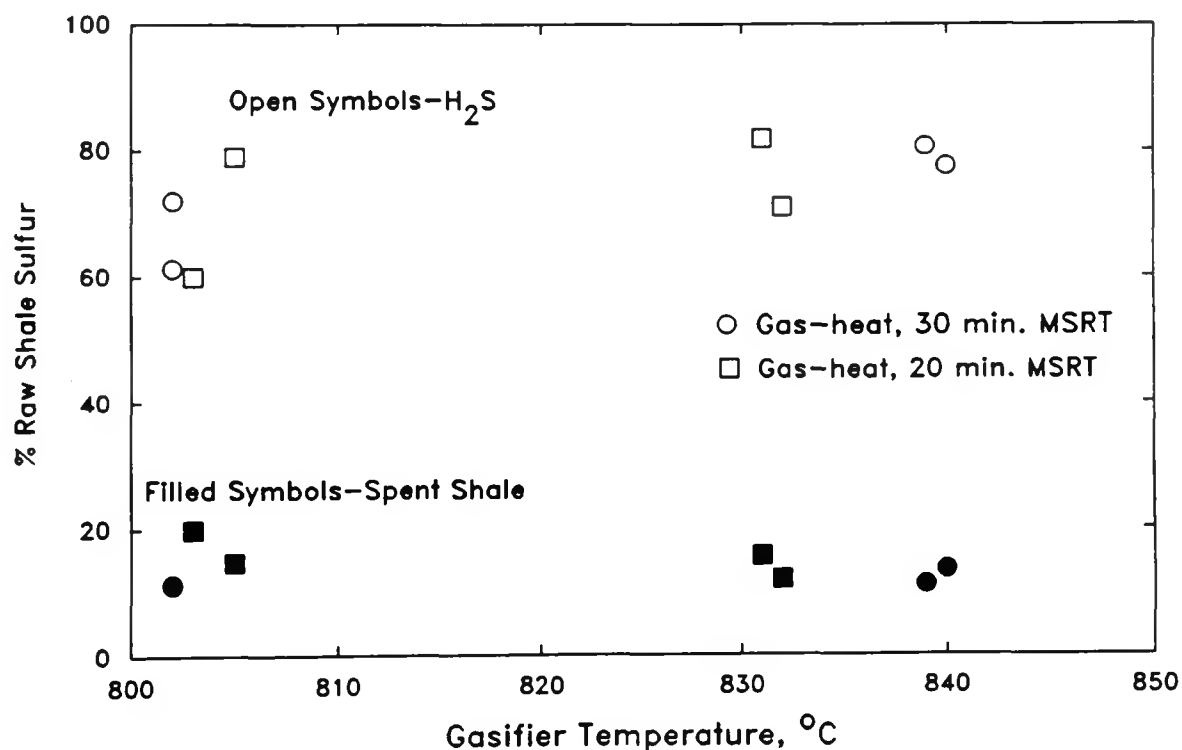


Figure 7-6. Effect of Gasifier Conditions on Sulfur Removal.

to follow. Detailed analyses of the nonrecycle oils are presented in Section 8.

7.2.8 Oil Yield-Solid Recycle Runs. The heat of pyrolysis for non-recycle runs was provided by a combination of hot gases from the gasification section and external heat through the walls of the reactor. Under these conditions coking losses were relatively small, but were not completely absent since oil vapors must pass through the bed of pyrolyzing shale with the potential to react with shale surfaces and crack to form gases and coke. A situation in which all oil vapor/solid contact is eliminated could be defined as the level of zero coking, but this is not practical. The non-recycle mode of operation provides the lowest realistic level of oil coking in a fluidized bed pyrolyzer, so we define it as the baseline carbon conversion to oil, F_{A0} . The fractional oil loss, X_A , due to vapor phase coking is defined as:

$$X_A = (F_{A0} - F_A) / F_{A0}$$

where F_A is carbon conversion to oil in the solid-recycle mode.

If the usual mass-action laws are applied to oil coking, the rate of oil loss is proportional to the product of the oil vapor concentration and the concentration of active surface sites. Therefore, at a given oil vapor concentration and solid temperature the amount of coking should increase as the number of active surface sites increases assuming that the activity of the shale remains constant. The number of surface sites per unit of oil vapor is proportional to the solid-recycle ratio (gasifier shale/raw shale, w/w). In Figure 7-7, fractional oil loss, X_A , is shown to increase as the solid-recycle ratio is increased for a given temperature range. Generally, the amount of oil loss also increased at higher solid temperatures for a given solid recycle range.

Under the least severe solid-recycle conditions (low temperature/low recycle ratio), oil yields similar to those from non-recycle operation were obtained. However, only 40% of the pyrolyzer heat was provided by solid recycle for these runs, and this does not represent a realistic scenario on the commercial-scale. A better simulation of large-scale conditions was achieved by increasing the shale feed rate to the operating limits of the reactor, which resulted in 60% heat transfer by solids. Under these conditions, approximately 15% oil loss was observed.

In the KENTORT reactor, a wide range of recycle ratio and recycle-solid temperature combinations may be utilized to transport the same quantity of energy. If oil loss is related to this combination, then the recycle-ratio and recycle-solid temperature could be adjusted to reduce oil losses, providing a simple engineering solution to oil yield reduction. Knowledge of

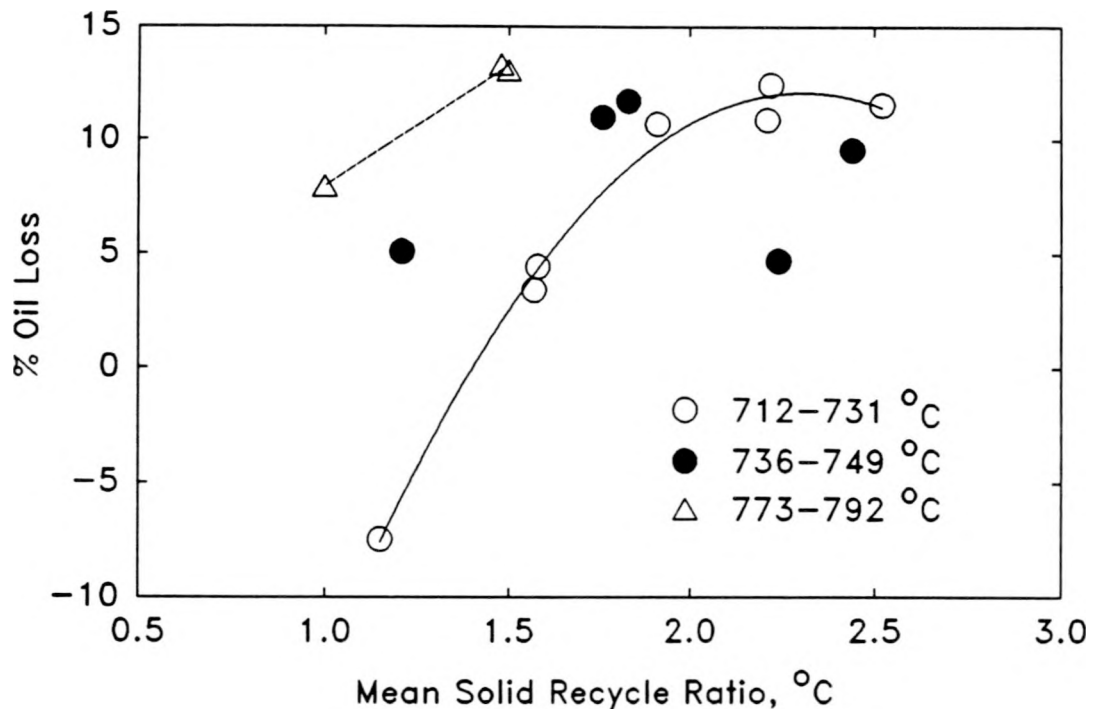


Figure 7-7. *Effect of Solid-Recycle Ratio and Temperature on Oil Yield Loss in the Pyrolysis Zone.*

these relationships is essential for the eventual scale-up and design of a solid-recycle system like KENTORT.

By focusing on runs with similar throughputs (KEN#31-36), the examination of recycle-ratio and recycle-solid temperature combinations was facilitated. The inverse relationship between recycle-solid temperature and recycle-ratio, and the direct proportionality between gasifier temperature and recycle-solid temperature are shown in Figure 7-8. Since these parameters are interrelated, it is sufficient to characterize the conditions for these runs with one parameter. The average recycle-solid temperature was chosen as the classifying condition for the discussions to follow.

Oil loss appeared to increase slightly at increased recycle-shale temperature (and lower recycle ratio) for the gasified material, but this trend occurred within a relatively narrow range of oil yield losses (11-14%) making it difficult to derive confident conclusions about the relationship (Figure 7-9). The observed trend is not unreasonable since a linear increase in coking rate is expected with recycle ratio while an exponential increase in coking rate is expected with a rise in temperature, making temperature the dominant parameter. The temperature range of this study needs to be extended somewhat to substantiate these

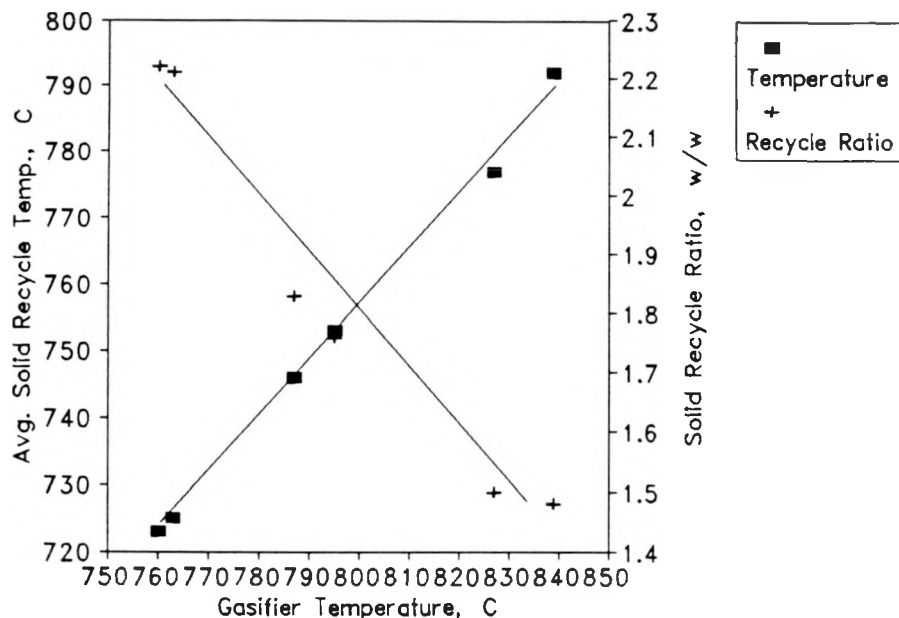


Figure 7-8. Relationship of Gasifier Temperature to the Average Solid Recycle Temperature and Recycle Ratio for Runs: KEN#31-36.

results, but there are practical limitations for the combinations of temperature and recycle rate that can be used. For example, it appears that operation at a low recycle temperature is least detrimental to oil yield, but it may not be feasible to operate the gasifier below 700°C without staging. Similarly, an extremely low recycle rate would require a high recycle temperature, forcing the combustor temperature to be excessive.

7.2.9 Product Distribution-Solid Recycle Runs. Besides the reduction in oil yield caused by recirculating solids, some differences in the product slate were affected. The major elemental distribution for solid recycle runs is displayed with the nonrecycle runs in Table 7-3. When the effects of gasifier temperature and solid residence time are considered, it appears that slightly greater carbon gasification occurred according to CO_x production (see Figure 7-10). Carbon gasification is expected to proceed more quickly in the recycle mode since fresh surface coke on the recirculated shale is accessible and not highly structured making it reactive.

While carbon gasification was slightly improved in the solid recycle mode of operation, it appears that sulfur removal was somewhat retarded (Figure 7-11). More sulfur was retained in the spent shale and less was converted to H₂S compared to nonrecycle operation. Capture of H₂S in the pyrolyzer by the recirculated shale is considered the most likely cause. Iron oxides generated

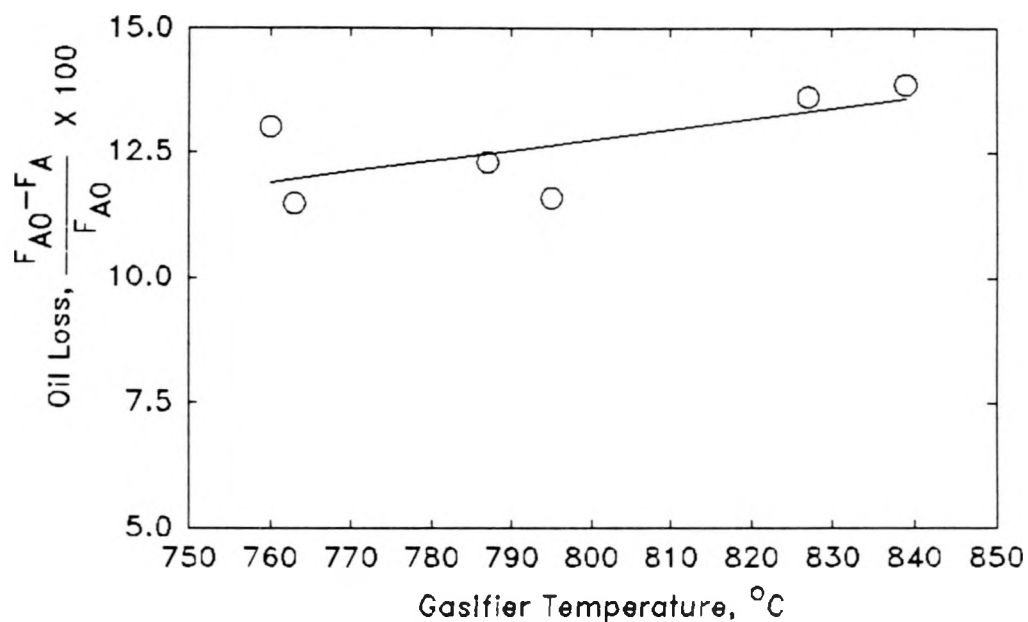


Figure 7-9. *Effect of Gasifier Temperature on the Fraction of Oil Losses in the Pyrolyzer for Runs: KEN#31-36.*

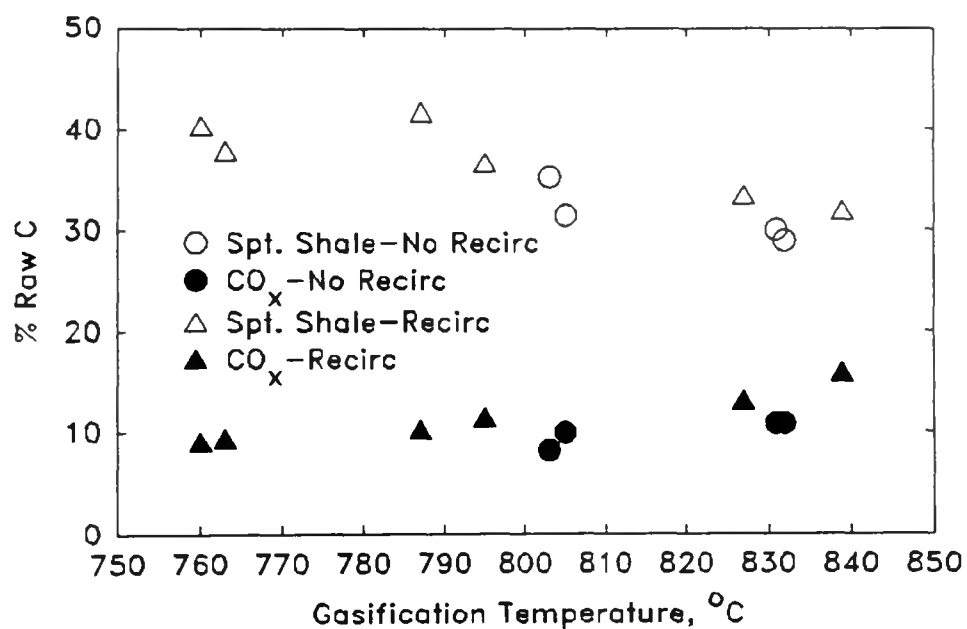


Figure 7-10. *Effect of Solid Recirculation on Carbon Gasification in the Pyrolysis-Gasification Configuration.*

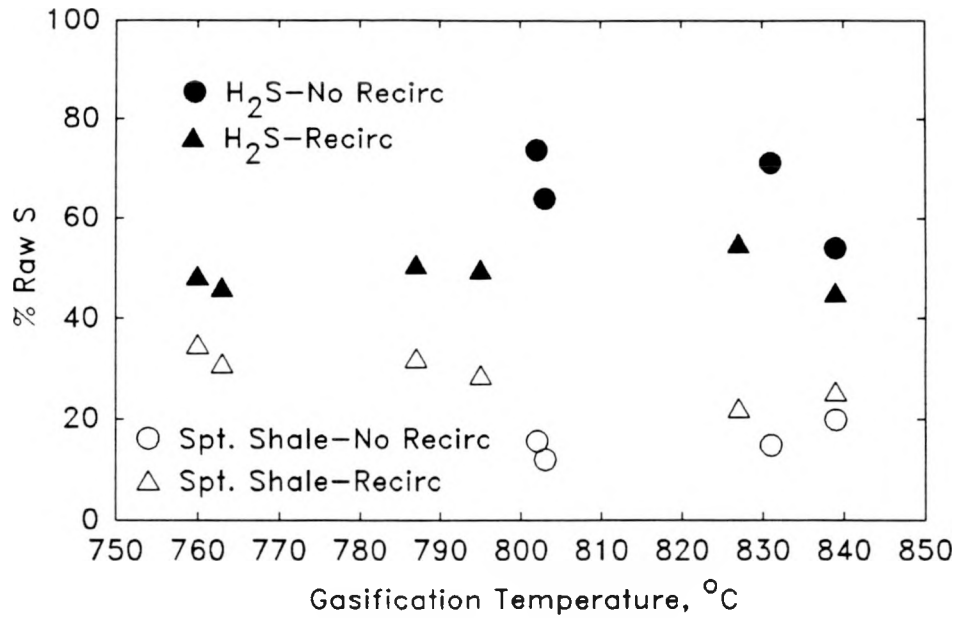


Figure 7-11. *Effect of Shale Recirculation on the Removal of Sulfur in the Pyrolysis-Gasification Configuration.*

in the gasification zone can react with H_2S to reform FeS . This situation has been encountered primarily when combusted shale is mixed with raw shale during pyrolysis, affecting nearly complete H_2S capture.(7) The large excess of steam in the KENTORT process prohibits significant net H_2S recapture.

7.3 INTEGRATED PYROLYSIS-GASIFICATION-COMBUSTION OPERATION

In this phase of the study, the integrated reactor was operated in the complete KENTORT II configuration. A major objective was to demonstrate the operation of the KENTORT II concept in an integrated reactor. While the system was much too small to truly simulate the dynamics of a commercial-scale module, a valid demonstration of the process could be obtained because: 1) all of the vital reaction zones were intact, 2) near-adiabatic conditions were achieved, and 3) the majority of heat transfer was achieved through shale recirculation.

Besides demonstration of the process as a whole, another major objective of this portion of the study was to examine to shale combustion and its effects on the other zones of the reactor. Of great interest is the effect on oil yield and quality of recycling partially combusted shale to the pyrolyzer.

7.3.1 Apparatus. The integrated reactor system was configured as shown in Figure 7-12 for the final portion of the study. The reactor was identical to the previous configuration except that a combustion zone with a solid recirculation loop was added. The oil collection and gas analysis system was identical to the previous system with the addition of another gas conditioning stream for the analysis of combustion flue gases.

7.3.2 Oil Shale Sample. The oil shale sample described in Section 3 was also used in this part of the study.

7.3.3 Procedure. The procedure was identical to that used in Section 7.2 with the addition of the combustion zone. The combustor and its liftpipe were purged with N_2 during heatup and the initial stage of each run. Once shale was steadily recirculating between the gasifier and pyrolyzer, air was blended into the fluidizing gas and shale recycle to the combustor was initiated. The shale recirculation rate was set at the maximum level and the targeted combustion temperature was maintained by adjusting the amounts of air and nitrogen in the fluidizing gas. It required 20-30 minutes from the start of each run to get both solid recirculation loops operating and to begin a period of steady state operation.

7.3.4 Experimental Design. Gasification and combustion temperatures were the two independent process parameters that were varied during this portion of the study. Changes in these temperatures also affected the required solid recycle rate to the pyrolyzer which was considered a dependent parameter. The raw shale throughput was held constant, so the temperature and rate of recycle solids were related as described in Section 7.2. Therefore, the gasifier and combustion temperatures were sufficient to comparatively characterize the conditions. The gasification temperature was nominally varied from 775-825°C at 3

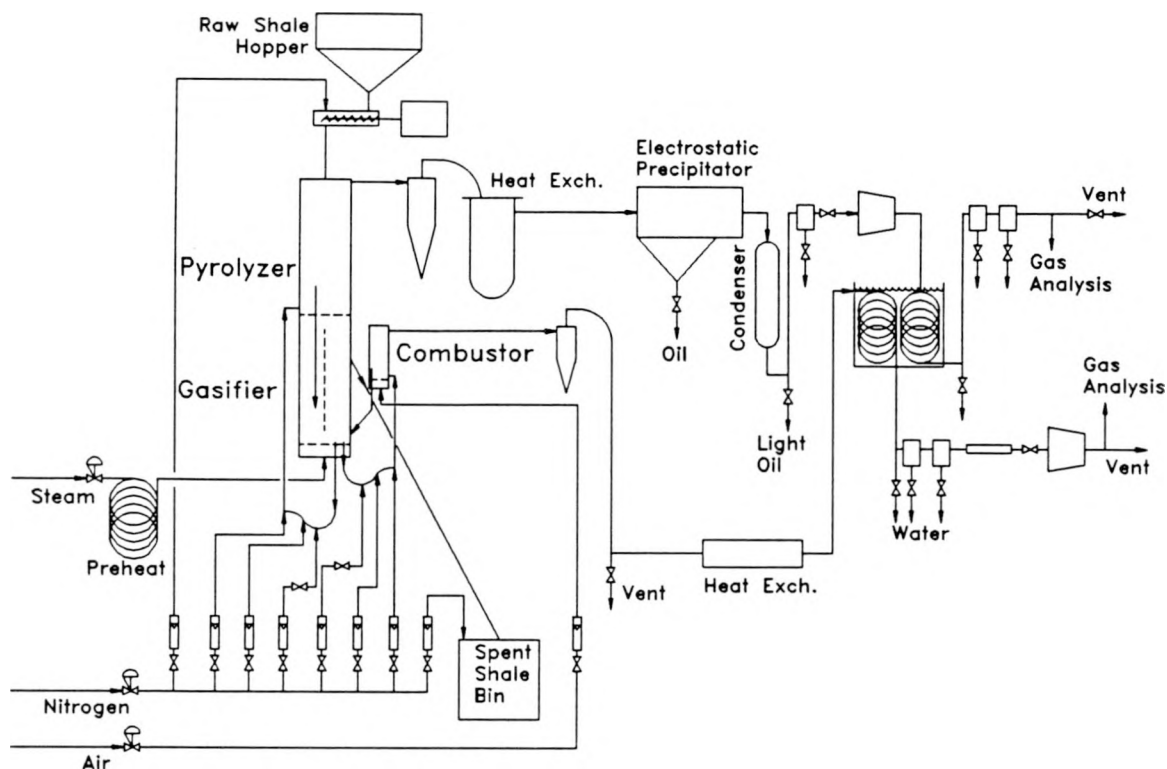


Figure 7-12. *Schematic of the Integrated 5-lb/hr KENTORT II Experimental System Including the Oil Collection and Gas Analysis Systems.*

combustor temperatures (860, 880, and 900°C). The average run conditions that were achieved are shown in Table 7-5. The method of operation required that steady state conditions be reached quickly, so no tweaking of conditions was performed. Therefore, the actual run conditions sometimes varied from the target conditions set forth prior to each run.

7.3.5 Mass and Component Balances. The absolute and relative amounts of solid, liquid, and gas produced from all runs that achieved steady state conditions are included in Appendix B. The ultimate analysis of each collected fraction and the calculated carbon, nitrogen, and sulfur balances are also shown. Analogous to non-combustion operation, only runs that had 90+% carbon balances were considered for further analysis, but if a run was slightly below 90% and was superior in other aspects, it was included for examination (see Table 7-6).

Table 7-5. Average Run Conditions for the Integrated 5-lb/hr KENTORT II Reactor System.

Run #	Mean Bed* Temp., °C		Shale Feed Rate, g/min.	Mean Solid Res Time, min		Solid Recycle Temp., °C		Shale Recirc Ratio	Combustor Temp, C
	Pyrol	Gasif		Pyrol	Gasif	Meas.	Avg.		
KEN46	537	774	39.80	8.7	20.6	647	711	2.53	861
KEN47	541	788	40.86	8.5	20.0	658	723	2.64	851
KEN48	537	754	40.00	8.7	20.4	699	727	2.29	849
KEN49	537	761	41.14	8.4	19.9	698	730	2.11	853
KEN50	538	773	41.76	8.3	19.6	691	732	2.05	855
KEN52	536	806	39.64	8.8	20.6	708	757	1.60	866
KEN53	531	823	38.21	9.1	21.4	711	767	1.37	868
KEN54	533	784	40.86	8.5	20.0	694	739	1.97	880
KEN55	534	813	39.89	8.7	20.5	709	761	1.84	869
KEN56	531	757	40.29	8.6	20.3	676	717	2.16	878
KEN57	534	813	40.01	8.7	20.4	713	763	1.52	883
KEN58	531	811	40.06	8.7	20.4	710	761	1.32	893
KEN59	533	786	38.92	8.9	21.0	696	741	1.73	886
KEN60	541	837	39.20	8.9	20.9	731	784	1.29	888
KEN61	533	808	38.76	9.0	21.1	711	760	1.50	884
KEN62	535	810	37.97	9.1	21.5	710	760	1.34	909
KEN63	534	762	39.70	8.8	20.6	684	723	2.15	897
KEN64	537	811	39.11	8.9	20.9	719	765	1.48	905

* Reactor operated at 0 psig + 5 inches water column.

7.3.6 Combustor Operation. The combustor operated smoothly in all aspects during this run sequence. When the pressures in the combustion and gasification zones were properly balanced, shale moved steadily between the two zones. Space limitations required the disengaging height of the combustor to be too low,

Table 7-6. Carbon, Nitrogen, and Sulfur Balances for Pyrolysis-Gasification-Combustion Runs.

KEN#	Spent Shale			Oil			P-G Gases		C Gases			Water			Total		
	C	N	S	C	N	S	C	S	C	N	S	C	N*	S	C	N	S
46	28.0	46.0	20.7	39.8	10.1	8.6	10.9	57.1	13.0	1.4	7.6	2.9	21.9	5.4	94.6	NA	99.5
47	29.5	45.7	21.2	38.3	9.7	7.9	11.5	58.4	10.2	0.7	5.4	1.9	18.2	2.5	91.3	NA	95.4
48	29.6	47.9	22.6	39.5	11.5	7.8	11.1	72.2	12.2	0.9	6.1	4.0	23.4	1.4	96.3	NA	110.1
49	30.1	67.3	21.7	43.6	13.2	8.9	11.4	56.3	10.8	0.9	5.9	2.4	20.8	0.9	98.3	NA	93.7
50	27.1	56.3	18.1	38.6	10.4	7.8	11.4	62.3	10.5	0.7	5.2	2.4	24.3	1.4	90.1	NA	94.9
52	28.1	51.9	20.5	39.2	10.6	8.0	12.3	63.1	10.7	1.0	6.2	1.9	22.8	0.2	92.3	NA	97.8
53	24.8	42.0	17.8	38.0	12.5	7.5	13.0	58.1	9.9	0.8	5.1	2.5	22.6	9.5	88.1	NA	98.0
54	27.1	48.6	21.7	38.8	12.6	7.8	10.7	60.7	12.0	0.9	7.5	1.8	19.8	8.4	90.3	NA	106.1
55	25.9	47.9	19.9	40.2	14.1	8.3	12.1	52.1	9.8	0.7	5.5	1.8	NA	5.3	89.8	NA	91.1
56	26.9	49.9	20.0	38.0	13.8	8.0	9.7	53.9	14.4	1.1	8.3	1.8	21.2	0.9	90.8	NA	91.1
57	25.1	42.2	17.0	41.2	13.3	7.8	14.2	59.0	11.3	0.9	6.0	1.7	23.6	0.9	93.5	NA	90.6
58	27.3	48.8	19.9	39.6	13.6	8.4	14.1	80.2	12.2	0.9	5.6	1.6	NA	8.2	94.8	NA	122.2
59	26.2	36.8	21.6	40.1	14.2	7.3	12.1	66.0	13.8	1.0	6.9	2.3	22.3	2.2	94.4	NA	104.0
60	26.8	42.1	22.0	38.8	14.3	7.2	15.1	58.8	8.6	0.5	3.9	1.9	24.6	2.4	91.2	NA	94.3
61	27.5	35.9	22.5	38.8	12.3	7.3	13.5	57.7	10.1	0.8	4.7	1.4	24.6	2.1	91.4	NA	94.2
62	24.8	30.4	18.1	40.0	13.1	7.4	14.6	66.9	13.9	1.1	6.2	1.2	22.9	1.4	94.5	NA	100.0
63	27.6	27.6	22.5	40.2	12.9	7.4	11.3	50.1	11.7	0.8	6.6	1.1	16.0	0.9	91.8	NA	87.4
64	27.0	42.0	23.9	37.1	11.4	7.1	12.3	55.0	10.4	0.8	5.7	1.3	19.9	9.7	88.1	NA	101.3

* NH3 only.

and this caused excessive particle carryover early in the run matrix. To reduce this problem, a grid after Blyakher and Pavlov (8) was installed in the freeboard region of the combustor. This perforated stainless steel plate served to dampen pressure fluctuations caused by bursting bubbles which, in turn, reduced particle carryover.

The system was not maintained under adiabatic conditions because sufficient heat could not be transferred from the combustor to the gasifier. While limits of both combustor temperature and hot shale recycle rate were employed during the run series, autothermal operation was not achieved. Some thermal deformation of the thermowell in the combustor was evident following operation at 900°C, so higher temperatures were not attempted. Moreover, combustor temperatures much in excess of 900°C would cause elevated levels of NO_x and is not considered an ideal operating regime at the commercial scale. Calculated recycle ratios ranged from 4.5 to 7.2. The recirculation rate between the gasifier and the combustor was limited by the downflow standpipe connecting the combustor to the gasifier because the combustor was overfilled on a few occasions when the flow of solids from the gasifier to the combustor was maximized. The lack of sufficient heat production in the combustor was compensated by the furnace surrounding the gasification zone.

Carbon oxides production from combustion (see Figure 7-13) confirmed the non-adiabatic operation. Carbon burn-off in the combustor ranged from 8.5 to 14.5% of the raw shale carbon, which was not sufficient to drive the pyrolysis and gasification reactions. With nearly 25% of the carbon remaining in the spent shale, there was abundant fuel for heat production, but the solid recycle loop could not transfer the solids quickly enough. The combustor could have handled faster throughput since it was starved of oxygen to control temperature. The mean solid residence time in the combustor was less than a minute in all cases.

The amount of carbon burned in the combustor was inversely related to the temperature of the gasifier since the latter was heated in part by an external furnace (see Figure 7-13). This is in contrast to truly adiabatic operation where more carbon burn-off would be required to achieve a higher gasifier temperature. Runs utilizing nominal combustor temperatures of 880°C produced more CO_x compared to nominal combustor temperatures of 860°C, as expected. A lot of scatter was observed in the 900°C runs making interpretation of these results difficult. Total carbon oxides production in the gasifier was seemingly unaffected by the operation of the combustion zone. Carbon gasification was similar in magnitude compared to non-combustion operation and increased with higher gasifier temperatures.

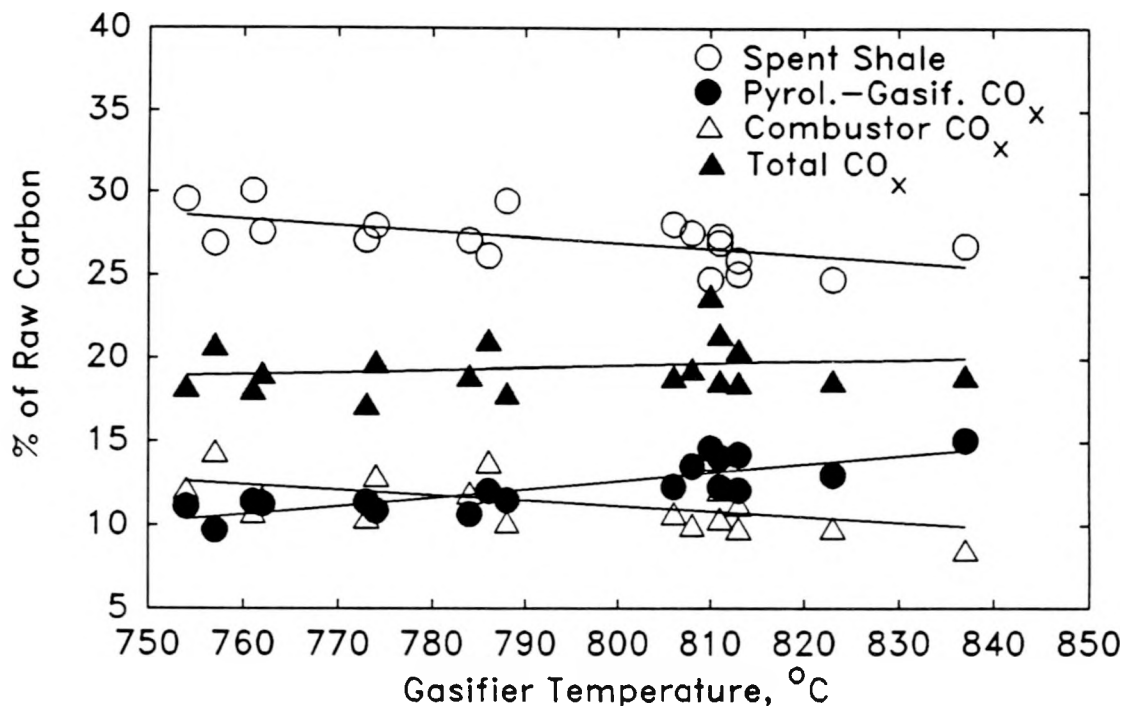


Figure 7-13. *Effect of Gasifier Conditions on the Production of Carbon Oxides in the Gasifier and Combustor.*

Sulfur dioxide production in the combustor was much lower than CO_x production, but it shared a similar trend of decreasing yield with increased gasifier temperature (see Figure 7-14). Again, this is attributed to the partial heating of the gasifier by the external furnace. The production of SO₂ was low compared to H₂S by quite a margin. Hydrogen sulfide production remained essentially constant over the range of conditions studied with the combustion zone in place. No more than 8.3% of the sulfur in the raw shale was observed as SO₂, while over 70% was converted to H₂S in some cases. This result is significant because it demonstrates a major purpose of the gasification zone in the KENTORT II process under realistic conditions. By shifting the sulfur distribution toward H₂S production, significant amounts of by-product sulfur are produced while minimizing unwanted SO₂ emissions. It should be noted, however, that SO₂ generation should increase when more complete combustion is attained during autothermal operation, but this should not upset the favored production of H₂S over SO₂.

Nitrogen oxide emissions from the combustion zone were very low. The measured NO_x levels represented no more 1.5% of the original nitrogen in the shale. Since this is such a minor component of the overall nitrogen balance, it is difficult to determine whether fuel or thermal nitrogen contributed to the NO_x.

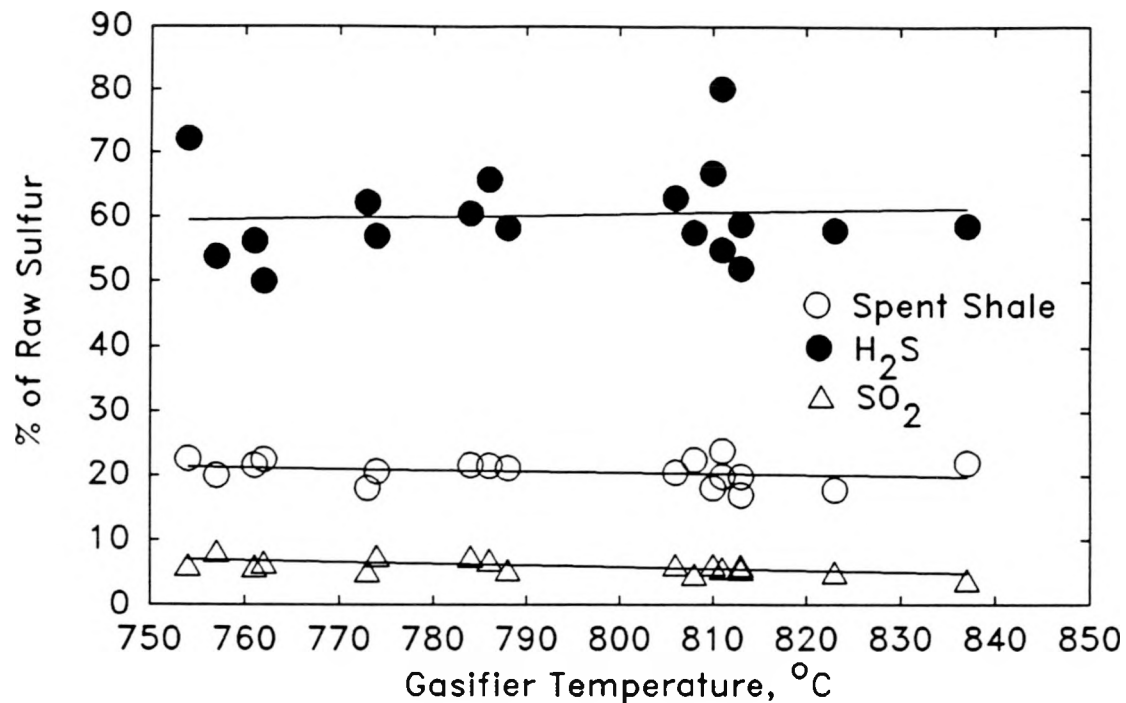


Figure 7-14. *Effect of Gasifier Temperature on the Production of H₂S and SO₂ in the Gasifier and Combustor, Respectively.*

production. However, the combustor temperature was held at 900°C or below throughout the study so thermal NO_x production is unlikely.

7.3.7 Oil Yield. The run conditions in this test series were maintained as similar as possible to those in KEN31-36 to facilitate comparisons between oil yields with and without combusted particles under solid recycle conditions. The raw shale throughput was slightly lower for the combustion runs, but otherwise the conditions were essentially identical. In addition, the raw shale throughput was held constant for the combustion runs series thereby coupling solid recycle temperature and rate. Therefore, as for non-combustion runs, the gasifier temperature was sufficient to characterize the recycle conditions to the pyrolyzer on a relative basis.

While gasifier temperature was determined to have a slight negative effect on oil yield during non-combustion operation, no such trend was discernable with the combustor in operation (see Figure 7-15). There was also no correlation between combustor temperature or carbon burn-off in the combustor and oil yield. It appears that the range of conditions was not wide enough to significantly affect oil yield in this situation. Because of the

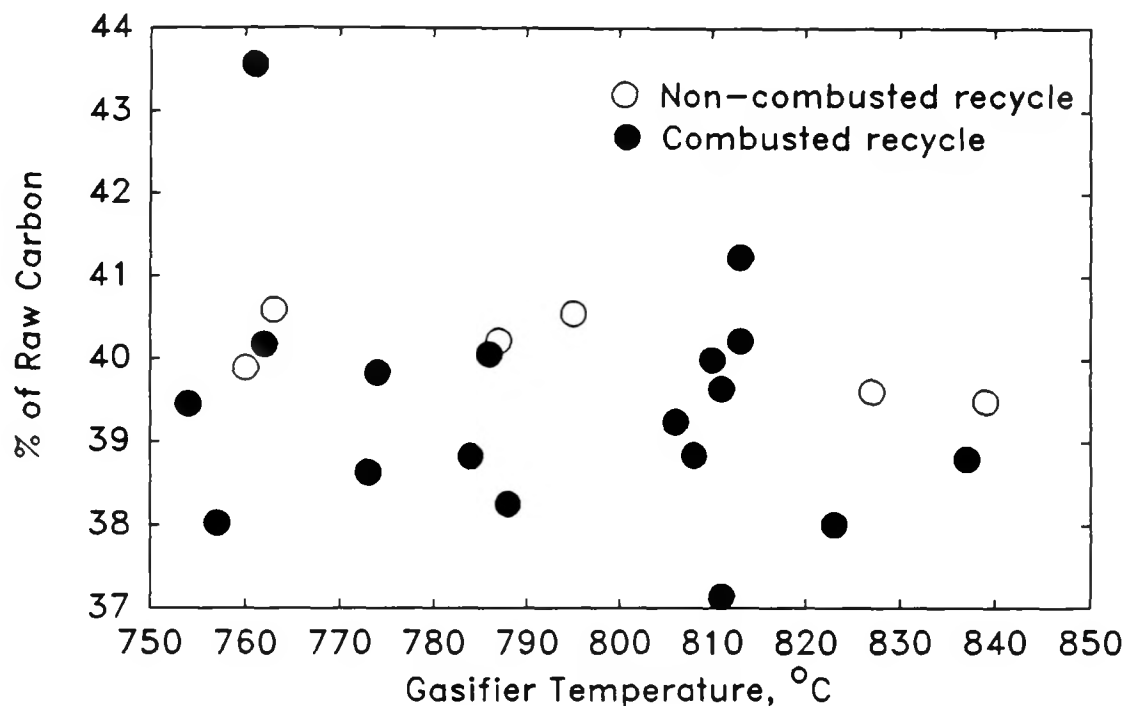


Figure 7-15. *Comparison of Solid Recycle Oil Yields with and without Combustion Operation.*

inverse relationship between solid recycle rate and temperature imposed by the design of the reactor, there are offsetting factors to consider when the gasifier temperature is varied. The benefits of reduced surface area may be completely countered by increased particle temperature, and vice-versa.

While oil yields were not perceptibly affected, it does appear that some oil properties were influenced by altering the rate and temperature of the recycle solids. These results, which are discussed in detail in Section 8 of this report, also seem to indicate some differences in activity between gasified shale and partially combusted gasified shale. A comparison of oil yields from combustion and non-combustion operation is shown in Figure 7-15. While it is difficult to compare data sets of such unequal sizes, it appears that non-combusted recycle shale affected slightly higher oil yields than the partially combusted shale. The difference in the median carbon conversion to oil was only 0.7 percent, and was found statistically significant at only the 80 percent level.

The chemistry and dynamics that control these results are difficult to explain at this time, but the results are nevertheless encouraging. Oil yields greater than Fischer assay have been achieved throughout the solid-recycle phase of this

work and combusted shale, regardless of the extent of combustion, has affected very little oil yield reduction compared to non-combusted shale. Admittedly, a low level of combustion was achieved in this study, and it is possible that more complete combustion of the char-containing ash could change the situation completely. However, this is unlikely since the combustion that took place in this study undoubtedly occurred on or near the surfaces of the particles which is where the coking and cracking chemistry occurs. More extensive combustion would have less effect than the initial surface oxidation.

7.3.8 Particle Size Analysis. The size distribution of all solid streams was determined for selected runs from each major portion of the study: non-recycle, gasified-solid recycle, and combusted-solid recycle. Figure 7-16 shows the average size distributions for the system during combustion operation. This is particularly important when considering the design of cyclones and other particle collection equipment for larger scale systems.

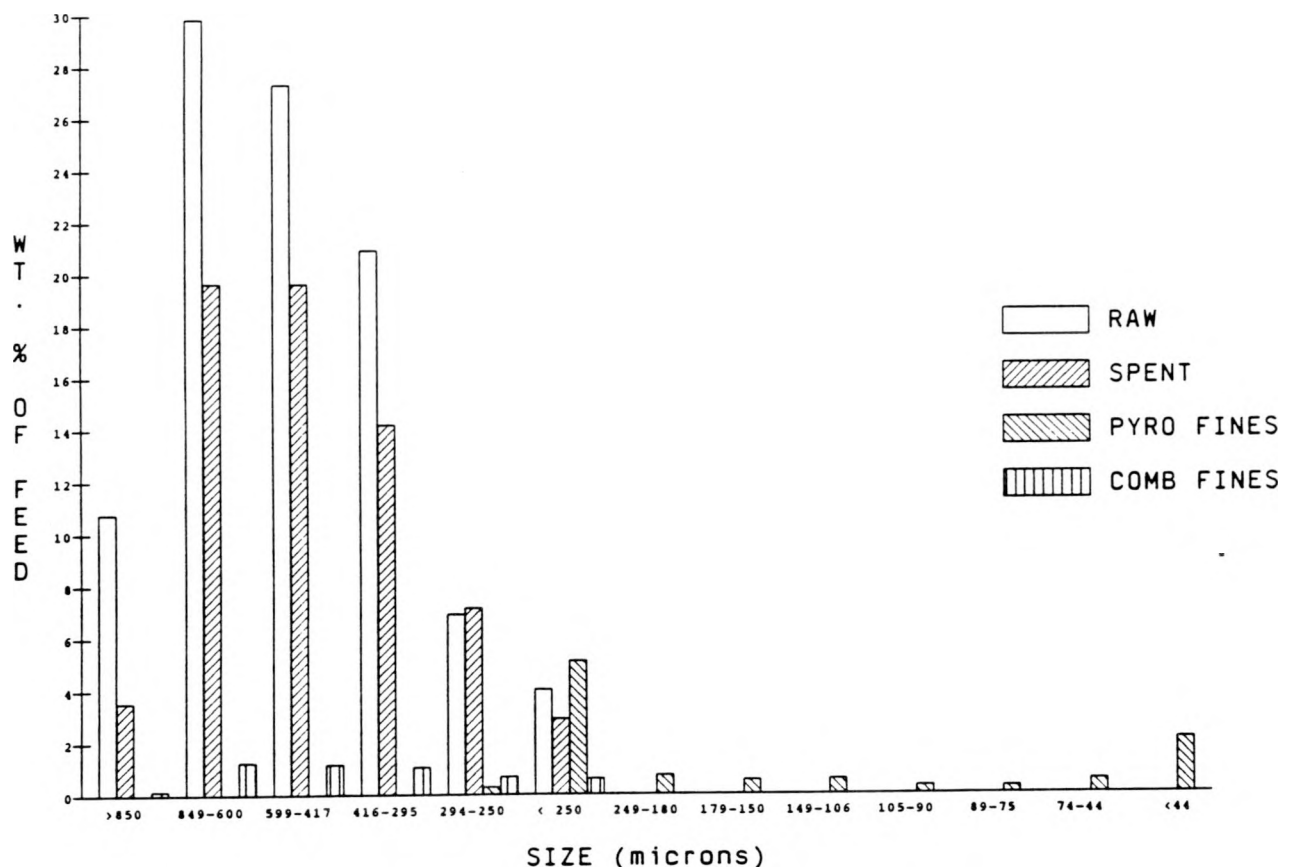


Figure 7-16. Average Particle Size Distribution for Combustion Runs.

The shale maintained its integrity regardless of process conditions. The majority of the solids was taken from the bottom of the process and little from overhead streams. High carryover values for combustion runs reflect overfilling the combustor rather than large amounts of particle attrition. Size distributions for selected runs throughout the test series are included in Appendix C. Only a small amount of fines generation was observed, and swelling of the shale did not occur. These results indicate that spent shale disposal should not be a problem because the fines content will be small.

7.4 CONCLUSIONS

The operation of the integrated reactor system, following modifications mainly associated with corrosion problems, was stable and reliable in each configuration. Design process conditions were achieved in all instances with the exception of insufficient solid recycle between the gasification and combustion zones. Therefore, true autothermal conditions were not attained in the combustion mode at gasification temperatures exceeding 775°C. However, sufficient carbon remained in the spent shale to heat balance the system if the solid recycle loop were larger, and this is considered a minor design change. With this exception noted, it is concluded that the physical operation of the KENTORT II concept was demonstrated in this program.

While the physical operation of the KENTORT II process was achieved at the 5-lb/hr scale, the demonstration of viability of the concept is an entirely different issue. Two basic objectives of the KENTORT II concept needed to be experimentally demonstrated under 'reasonable' conditions before the process, as originally proposed, could be considered feasible. The first objective of KENTORT is to utilize hot solids recycle as the primary method of heat transfer to the pyrolyzer while maintaining oil yields in excess of Fischer assay. This objective was experimentally confirmed in this study at the 5-lb/hr scale in a completely integrated system. There are scale factors that cannot be avoided in the KENTORT design, so testing at a larger scale is essential for confirmation of the commercial scale KENTORT concept.

The second objective of the KENTORT process is to shift the disposition of sulfur toward the production of H_2S by a mild gasification step. There are, of course, other benefits of gasification, but without nearly complete sulfur removal the gasification step loses a lot of its attractiveness. The stipulation of *mild* gasification conditions is important because: 1) long residence times have an adverse effect on economics, and 2) high gasifier temperatures would require high combustor temperatures causing increased NO_x production and would possibly create ash agglomeration difficulties. Sulfur removal as H_2S

ranged from 50 to 80% during integral testing with no more than 8.3% of the starting S going toward SO₂. These results were achieved at mean solid residence times of 20 minutes and temperatures as low as 750°C. While no economics have yet been performed on the KENTORT process, it is believed that these conditions make the gasification zone viable. Certainly, the results of this study indicate that continued investigation of the gasification zone in the KENTORT process, as presently configured, is warranted.

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8. CHARACTERIZATION OF THE MINI-PLANT PYROLYSIS PRODUCT STREAM

8.1 OBJECTIVES

The objectives of the KENTORT product characterization program were threefold. First, weight measurements and major element analyses for the various product streams were utilized to measure or confirm material and elemental balances around the reactor. These balances were the primary criterion used to evaluate the integrity of a given run and thus its inclusion or rejection from further consideration. Second, certain analyses were directed at gaining a better understanding of the process chemistry, particularly with regard to oil loss reactions. This information is often used to more efficiently devise reactor modifications aimed at suppressing these detrimental reactions. Third, the analytical data served as a measure of the relative quality (economic value) and physical properties of the oil and HC gas products.

8.2 ANALYTICAL PROCEDURES

Ultimate, heating value, water content, and physical testing were conducted according to ASTM procedures for the recovered hydrocarbon liquids. Proximate analyses were obtained by TGA, average molecular weights were determined by vapor phase osmometry, oil density was measured with an oscillating densitometer or a picnometer, and oil aromaticity was determined by ^{13}C -NMR on a Varian XL-200.

Retort off-gases were analyzed on a Carle Series Sx-AGC gas analyzer equipped with both FID and TC detectors using Carle Application #397. A split of the non-condensed gas stream was automatically sampled at 35 minute intervals during steady state portions of a given run and averaged. Primary standards from Matheson, Inc. were used for calibration. Additional details of the analytical procedures are presented in Appendix D and have been reported in the literature.(1)

8.3 SAMPLING

Oil streams were collected or measured in one of five locations (see reactor schematic-Figures 7-3, 7-12). These include the air-cooled heat exchanger, the electrostatic precipitator (ESP), a water cooled heat exchanger, a pair of filter mist coalescers, and by an on-line gas chromatograph used to measure C_5+ hydrocarbon vapors. Oil collected in the air cooled heat exchanger was quantitatively removed with CH_2Cl_2 which rendered this product fraction unsuitable for subsequent analyses. Likewise, additional analyses of the C_5+ vapor phase hydrocarbons were not feasible since these materials were measured with an FID detector and not physically collected. Two

of the remaining three oil streams, i.e., the water cooled heat exchanger and the filter coalescers, were composited into one fraction termed the light oil (LO). A portion of the third oil stream, the ESP oil, was then combined with the light oil fractions according to the weight production of oil in the air condenser plus ESP (using ESP oil), and the water cooled heat exchanger plus C_5+ vapors (using the light oil fraction). This combination ESP/LO material was thoroughly blended and subsequently referred to as the composite oil (CO).

Though not all oil streams were collected and combined into the composite, this sample is presumed to be representative of the total oil product since the two streams that are omitted, the air cooled heat exchanger oil and the C_5+ vapors, are produced in similar quantities and their respective properties should have offsetting effects on the composite properties. That is, the omitted streams represent the extremes in properties such as H/C ratio, average molecular weight, density, etc. Thus, the net effect of omitting both from the composite oil should be reasonably minor.

The products of pyrolysis and gasification exited the reactor collectively. Following condensation of the liquid components, a split of the non-condensed gas stream was passed through the sampling loops of a Carle refinery gas analyzer which cycled at 35 minute intervals. A minimum of four refinery gas analyses were conducted and averaged for each reactor run. Combustion gases exited the integrated reactor in the combustion gas stream where selected components (CO , CO_2 , SO_2 , O_2 , and NO_x) were continuously monitored with on-line Beckman industrial gas analyzers. The analyses of this latter stream is reported in Chapter 7 and will not be discussed further.

Process waters were sampled from the water cooled heat exchanger where the bulk of the fluidizing steam was condensed (>95%). An aliquot of each process water was analyzed without treatment for suspended/dissolved solids and pH. A base treated split (10 g NaOH/L) was analyzed for total sulfur and organic carbon. An acid treated split (10 mL/L concentrated HNO_3) was tested for heavy metals and ammonia.

8.4 CHEMICAL/PHYSICAL PRODUCT PROPERTIES

Analytical data for the various product streams are shown for the mini-plant runs in Tables 8-1 to 8-5. In addition to individual run data, analytical data for the non-recycle runs (KEN #'s 4-11), gasifier recycle runs ('s 16-36), the final six gasifier recycle runs ('s 31-36), and the combustor runs ('s 46-64) were averaged and presented along with the grand average in the last five rows of each table. The final six gasifier runs are grouped separately since the heat load to the retort was held constant for these runs. A constant heat load, i.e., constant

raw shale feed rate, implies that solids recycle rate and solids recycle temperature were inversely coupled in order to maintain a constant pyrolysis temperature. Likewise, the heat load for all combustor runs was essentially constant. This was not the case for gasifier runs 16-30 for which recycle rate and recycle temperature were varied in a near independent manner. This is an important consideration since comparison of data from the constant heat load runs makes it possible to more clearly discern the effect of recycling larger amounts of lower temperature solids relative to fewer higher temperature solids. Further, since the heat load and pyrolysis temperature for the final six gasifier runs was similar to that of the combustor runs, direct comparisons can be made between product properties for these two data sets. Additional discussion of this point is presented in Section 7.

Table 8-1. *Light Oil Analyses for Mini-Plant Runs.*

0 RUN #	1 Density (g/mL)	2 % LO & C5+ Production	3 % C	4 % H	5 % N	6 % S	7 % H ₂ O	8 H/C (dry)	9 N/C X100	10 S/C X100
1 KEN 4	0.896	26.89	83.51	11.38	0.45	1.88	0.50	1.63	0.46	0.84
2 KEN 5	0.894	32.92	85.03	11.80	0.57	2.02	0.20	1.66	0.57	0.89
3 KEN 6	0.894	26.62	83.67	11.56	0.45	2.05	0.34	1.65	0.46	0.92
4 KEN 7	0.900	27.58	84.26	11.36	0.47	2.07	0.40	1.61	0.48	0.92
5 KEN 8	0.885	28.09	85.06	11.48	0.64	2.06	0.13	1.62	0.65	0.91
6 KEN 9	0.884	25.41	84.96	11.63	0.46	2.31	0.24	1.64	0.46	1.02
7 KEN 10	0.887	23.48	84.66	11.22	0.46	1.99	0.14	1.58	0.47	0.88
8 KEN 11	0.889	26.75	84.20	11.09	0.53	2.19	0.18	1.58	0.54	0.97
9										
10 KEN 16	0.889	25.39	85.54	11.16	0.59	2.29	0.38	1.56	0.59	1.00
11 KEN 18	0.870	22.30	85.40	11.50	0.52	2.12	0.22	1.61	0.52	0.93
12 KEN 20	0.881	26.19	85.23	11.35	0.61	2.00	0.22	1.60	0.61	0.88
13 KEN 23	0.888	30.78	85.29	11.84	0.56	1.93	0.22	1.66	0.56	0.85
14 KEN 24	0.882	28.34	86.30	11.84	0.64	1.79	0.18	1.65	0.64	0.78
15 KEN 25	0.886	30.03	85.28	12.01	0.55	1.95	0.27	1.69	0.55	0.86
16 KEN 27	0.883	27.80	85.19	12.28	0.61	2.14	0.17	1.73	0.61	0.94
17 KEN 29	0.883	31.02	85.08	12.00	0.57	2.04	0.21	1.69	0.57	0.90
18 KEN 30	0.896	33.17	83.43	11.62	0.50	2.42	0.24	1.67	0.51	1.09
19 KEN 31	0.884	30.82	83.62	11.45	0.44	2.06	0.19	1.64	0.45	0.92
20 KEN 32	0.882	30.54	83.70	11.52	0.49	1.95	0.17	1.65	0.50	0.87
21 KEN 33	0.873	28.15	85.36	11.62	0.42	1.97	0.13	1.63	0.42	0.86
22 KEN 34	0.887	31.30	83.98	11.32	0.47	2.05	0.32	1.61	0.48	0.91
23 KEN 35	0.881	27.84	84.57	11.70	0.48	1.82	0.19	1.66	0.49	0.81
24 KEN 36	0.875	27.08	84.01	11.99	0.42	2.14	0.17	1.71	0.43	0.95
25										
26 KEN 46	0.918	37.08	84.74	11.30	0.63	2.21	0.53	1.59	0.64	0.98
27 KEN 47	0.904	31.76	85.09	11.37	0.65	2.18	0.20	1.60	0.65	0.96
28 KEN 48	0.895	29.22	85.71	11.63	0.68	2.00	0.27	1.62	0.68	0.88
29 KEN 49	0.908	30.69	85.74	11.55	0.67	2.10	0.25	1.61	0.67	0.92
30 KEN 50	0.905	32.15	84.86	11.37	0.51	2.04	0.29	1.60	0.52	0.90
31 KEN 51	0.907	27.42	85.36	11.42	0.56	2.06	0.21	1.60	0.56	0.90
32 KEN 52	0.908	32.74	84.80	11.06	0.54	1.96	0.70	1.55	0.55	0.87
33 KEN 53	0.906	32.25	82.89	11.10	0.51	1.80	0.30	1.60	0.53	0.81
34 KEN 54	0.909	27.60	84.20	11.09	0.55	1.69	0.23	1.58	0.56	0.75
35 KEN 55		32.82	83.39	10.28	0.79	1.67	0.48	1.47	0.81	0.75
36 KEN 56		30.41	83.69	10.96	0.60	2.16	0.16	1.57	0.61	0.97
37 KEN 57	0.911	31.69	83.14	11.07	0.51	1.88	0.28	1.59	0.53	0.85
38 KEN 58	0.914	32.06	85.97	11.52	0.70	1.72	0.33	1.60	0.70	0.75
39 KEN 59	0.903	32.94	85.87	11.54	0.68	1.92	0.26	1.61	0.68	0.84
40 KEN 60	0.898	29.38	84.50	11.79	0.95	1.89	0.26	1.67	0.96	0.84
41 KEN 61	0.903	29.88	85.49	11.78	0.74	1.91	0.23	1.65	0.74	0.84
42 KEN 62	0.897	28.88	84.03	11.44	0.50	2.00	0.33	1.63	0.51	0.89
43 KEN 63	0.897	27.95	85.13	11.34	0.53	1.85	0.23	1.59	0.53	0.87
44 KEN 64	0.897	29.46	83.36	11.69	0.51	2.01	0.28	1.68	0.52	0.90
45 AVERAGES										
46 non-recyc	0.891	27.22	84.42	11.44	0.50	2.07	0.27	1.62	0.51	0.92
47 gasifier	0.883	28.72	84.80	11.68	0.52	2.04	0.22	1.65	0.53	0.90
48 gasif-final6	0.880	29.29	84.21	11.60	0.45	2.00	0.20	1.65	0.46	0.89
49 combustor	0.905	30.82	84.63	11.33	0.62	1.95	0.31	1.60	0.63	0.87
50 all	0.894	29.40	84.65	11.48	0.56	2.01	0.27	1.62	0.57	0.89

Table 8-2. ESP Oil Analyses for Mini-Plant Runs.

0 RUN #	1 % ESP + A/C PRODUCT	2 % C	3 % H	4 % N	5 % S	6 H ₂ O (KF)	7 % ASH	8 H/C (dry)	9 N/C	10 S/C
1 KEN 4	73.11	78.35	8.55	1.31	1.650	1.24		1.29	0.014	0.0079
2 KEN 5	67.08	70.95	10.07	1.27	1.540	17.22		1.38	0.015	0.0081
3 KEN 6	73.38	63.81	9.50	0.82	1.320	26.75	1.00	1.23	0.011	0.0077
4 KEN 7	72.42	66.41	9.52	0.81	1.260	21.15	2.65	1.30	0.010	0.0071
5 KEN 8	71.91	67.92	9.18	1.09	1.930	17.11	2.75	1.29	0.014	0.0106
6 KEN 9	74.59	68.63	9.22	0.84	1.210	16.06	3.50	1.30	0.010	0.0066
7 KEN 10	76.52	70.38	9.64	0.94	1.540	14.25	2.25	1.37	0.011	0.0082
8 KEN 11	73.25	69.73	9.50	1.12	1.500	15.59	2.50	1.34	0.014	0.0081
9										
10 KEN 16	74.61	62.75	9.20	1.17	1.560	22.55	5.00	1.28	0.016	0.0093
11 KEN 18	77.80	69.16	8.33	1.18	1.430	16.35	3.00	1.13	0.015	0.0077
12 KEN 20	73.79	70.55	8.54	1.12	1.590	11.55	3.90	1.23	0.014	0.0084
13 KEN 23	69.22	72.05	8.78	1.01	1.640	11.11	3.10	1.26	0.012	0.0085
14 KEN 24	71.66	71.07	9.32	1.30	1.400	12.15	3.80	1.35	0.016	0.0074
15 KEN 25	69.97	75.77	8.68	1.28	1.610	9.13	4.62	1.21	0.014	0.0080
16 KEN 27	72.20	75.10	8.91	1.38	1.690	8.14	3.75	1.28	0.016	0.0084
17 KEN 28	72.15	75.92	8.81	1.13	1.590	7.39	4.50	1.26	0.013	0.0078
18 KEN 29	68.98	76.61	8.64	1.50	1.700	8.97	3.50	1.20	0.017	0.0083
19 KEN 30	66.83	78.59	8.68	1.10	1.630	2.01	3.80	1.29	0.012	0.0078
20 KEN 31	69.18	76.76	8.75	1.17	1.700	2.94	4.10	1.32	0.013	0.0083
21 KEN 32	69.46	78.37	9.00	1.25	1.730	3.72	3.30	1.31	0.014	0.0083
22 KEN 33	71.85	77.44	9.09	1.24	1.760	4.43	3.30	1.33	0.014	0.0085
23 KEN 34	68.70	79.32	8.94	1.30	1.930	1.13	3.80	1.33	0.014	0.0091
24 KEN 35	72.16	78.49	8.99	1.28	1.890	2.41	3.60	1.33	0.014	0.0090
25 KEN 36	72.92	76.18	9.15	1.23	1.730	4.67	4.00	1.36	0.014	0.0085
26										
27 KEN 46	62.90	78.97	8.52	1.03	2.120	2.25	3.70	1.26	0.011	0.0101
28 KEN 47	68.24	79.88	8.71	1.00	2.005	3.29	2.40	1.25	0.011	0.0094
29 KEN 48	70.78	78.06	8.67	1.15	1.970	4.44	2.60	1.26	0.013	0.0095
30 KEN 49	69.31	81.72	8.77	1.24	2.030	3.76	1.40	1.23	0.013	0.0093
31 KEN 50	67.85	76.56	8.81	1.10	1.950	6.19	4.40	1.27	0.012	0.0096
32 KEN 51	72.58	82.44	9.03	1.05	2.080	3.60	3.10	1.26	0.011	0.0095
33 KEN 52	67.26	80.26	9.04	1.13	2.050	4.93	3.60	1.27	0.012	0.0096
34 KEN 53	67.75	80.23	8.61	1.42	2.010	4.28	3.55	1.22	0.015	0.0094
35 KEN 54	72.40	75.16	8.83	1.24	1.920	8.22	5.47	1.26	0.014	0.0096
36 KEN 55	67.18	70.43	8.71	1.26	1.900	8.85	8.40	1.32	0.015	0.0101
37 KEN 56	69.59	74.39	8.48	1.42	1.870	6.61	5.31	1.25	0.016	0.0094
38 KEN 57	68.31	76.27	8.80	1.34	1.830	4.97	3.24	1.30	0.015	0.0090
39 KEN 58	67.94	81.03	8.86	1.44	2.260	5.19	1.27	1.23	0.015	0.0105
40 KEN 59	67.06	78.88	9.28	1.50	1.810	3.66	3.38	1.35	0.016	0.0086
41 KEN 60	70.62	80.02	8.70	1.44	1.820	3.31	5.08	1.25	0.015	0.0085
42 KEN 61	70.12	80.20	9.55	1.25	1.820	3.94	2.94	1.36	0.013	0.0085
43 KEN 62	71.12	80.56	9.66	1.38	1.810	4.45	3.00	1.37	0.015	0.0084
44 KEN 63	72.05	78.54	9.23	1.29	1.770	6.05	1.57	1.31	0.014	0.0085
45 KEN 64	70.54	78.28	9.54	1.24	1.810	4.02	3.04	1.39	0.014	0.0087
46 AVERAGES										
47 non-recyc	72.78	69.52	9.40	1.03	1.490	16.17	2.50	1.31	0.012	0.0080
48 gasifier	71.34	74.63	8.86	1.23	1.660	8.04	3.80	1.28	0.014	0.0083
49 gasif-final6	70.71	77.76	8.99	1.25	1.790	3.22	3.70	1.33	0.014	0.0086
50 combustor	69.14	78.52	8.94	1.26	1.940	4.84	3.60	1.28	0.014	0.0093
51 all	70.64	75.40	9.00	1.20	1.750	8.14	3.70	1.29	0.014	0.0087

Table 8-3. Composite Oil Analyses for Mini-Plant Runs.

0 RUN #	1 RS CARBON TO OIL	2 DENSITY	3 % C	4 % H	5 % N	6 % S	7 % H ₂ O (KF)	8 % ASH	9 BTU/LB	10 H/C DRY	11 W/C	12 S/C
1 KEN 4	53.41		77.31	9.400	0.840	1.750	1.56		15780	1.42	0.0093	0.0085
2 KEN 5	43.78	1.005	73.88	10.150	0.770	1.720	12.65	1.140	12880	1.41	0.0089	0.0087
3 KEN 6	46.21	1.000	70.70	10.470	0.550	1.390	20.11	0.500	13730	1.39	0.0067	0.0074
4 KEN 7	50.03	1.006	73.40	9.950	0.560	1.300	17.38		14290	1.30	0.0065	0.0066
5 KEN 8	41.96	1.006	75.82	9.170	0.760	1.800	9.73	1.300	14950	1.27	0.0086	0.0089
6 KEN 9	42.82	1.003	75.27	9.940	0.730	1.700	12.06	0.250	14540	1.36	0.0083	0.0085
7 KEN 10	46.70	1.018	77.74	9.870	0.900	1.530	11.50	0.300	14840	1.32	0.0099	0.0074
8 KEN 11	42.23	1.014	73.69	9.890	0.880	1.570	13.17	0.200	14800	1.36	0.0102	0.0080
9												
10 KEN 16	45.14	1.029	69.47	9.430	0.830	1.460	16.60	3.000	13800	1.30	0.0102	0.0079
11 KEN 18	43.92	1.000	74.70	9.540	0.780	1.570	14.32	1.100		1.27	0.0090	0.0079
12 KEN 20	48.97	1.014	75.88	9.710	0.740	1.820	7.25	1.000		1.40	0.0084	0.0090
13 KEN 23	40.99	1.010	79.84	9.790	0.800	1.670	7.83	1.000	14510	1.33	0.0086	0.0078
14 KEN 24	43.05	1.012	76.48	10.010	0.830	1.550	9.35	3.000	13600	1.40	0.0093	0.0076
15 KEN 25	43.96	1.007	79.29	9.880	0.910	1.660	6.45	4.400	15480	1.38	0.0098	0.0078
16 KEN 27	41.38	0.999	79.12	9.830	0.900	1.700	5.93	4.800	16110	1.38	0.0098	0.0080
17 KEN 29	43.49	1.005	77.45	9.980	0.930	1.650	6.17	3.100	16020	1.43	0.0103	0.0080
18 KEN 30	40.31	1.005	81.47	10.010	0.810	1.800	0.32	4.900	16830	1.46	0.0085	0.0083
19 KEN 31	39.59	1.010	80.74	9.820	0.890	1.900	2.75	5.000	16470	1.40	0.0095	0.0088
20 KEN 32	40.22	1.008	83.69	9.910	0.790	1.850	0.32	2.000	17220	1.41	0.0081	0.0083
21 KEN 33	39.51	0.998	82.33	9.880	0.770	1.880	3.84	1.500	16780	1.37	0.0080	0.0086
22 KEN 34	40.71	1.017	83.55	10.280	0.780	1.980	0.96	2.250	16990	1.45	0.0080	0.0089
23 KEN 35	37.62	1.022	83.32	10.260	0.740	1.780	0.55	1.000	16830	1.46	0.0076	0.0080
24 KEN 36	40.47	1.022	81.16	9.960	0.760	1.740	3.54	3.000	16370	1.40	0.0080	0.0080
25												
26 KEN 46	39.84	1.025	80.46	9.340	0.900	1.940	1.86	3.730		1.36	0.0096	0.0090
27 KEN 47	38.26	1.024	83.31	9.510	1.000	1.960	2.61	0.810		1.33	0.0103	0.0088
28 KEN 48	39.45	1.020	83.44	9.670	0.990	2.020	3.77	0.300		1.33	0.0102	0.0091
29 KEN 49	43.56	1.027	84.08	9.865	1.035	2.095	2.18	1.500		1.37	0.0106	0.0093
30 KEN 50	38.64	1.024	83.38	9.860	0.730	2.330	2.52	2.015	17010	1.38	0.0075	0.0105
31 KEN 51	43.15	1.031	84.16	9.980	0.710	1.980	2.82	0.210	17270	1.38	0.0072	0.0088
32 KEN 52	39.24	1.045	81.92	9.760	0.820	2.050	3.76	1.790	16910	1.37	0.0086	0.0094
33 KEN 53	38.01	1.005	82.81	10.040	0.790	1.600	2.62	1.800	15970	1.41	0.0082	0.0072
34 KEN 54	38.83	1.027	78.25	9.500	0.790	1.750	6.52	1.620	15370	1.35	0.0087	0.0084
35 KEN 55	40.22	1.047	77.37	9.100	0.810	1.850	5.09	5.180	15360	1.32	0.0090	0.0090
36 KEN 56	38.03	1.040	79.39	9.650	0.810	1.640	4.58	1.060	16050	1.38	0.0087	0.0077
37 KEN 57	41.23	1.020	80.94	9.700	0.880	1.710	3.66	3.380	16490	1.38	0.0093	0.0079
38 KEN 58	39.64	1.027	81.25	9.510	0.990	2.040	3.73	1.230	16150	1.34	0.0104	0.0094
39 KEN 59	40.06	1.016	83.15	9.730	0.940	1.810	2.75	2.170	15910	1.36	0.0097	0.0082
40 KEN 60	38.80	1.015	83.02	10.120	0.920	1.850	0.73	0.120	17070	1.45	0.0095	0.0084
41 KEN 61	38.84	1.008	84.17	10.240	0.910	1.880	0.45	0.240	16980	1.45	0.0093	0.0084
42 KEN 62	39.99	1.018	83.30	10.130	0.870	1.870	3.41	0.380	16020	1.40	0.0090	0.0084
43 KEN 63	40.18	1.015	83.63	9.950	0.890	1.830	4.87	0.440	15970	1.35	0.0091	0.0082
44 KEN 64	37.14	1.018	83.87	10.230	0.870	1.810	3.07	0.520	16390	1.41	0.0089	0.0081
45 averages												
46 non-recycle	45.89	1.007	74.70	9.860	0.750	1.600	12.30	0.620	14500	1.35	0.0086	0.0080
47 gasifier	41.96	1.011	79.20	9.890	0.820	1.730	5.70	2.700	15900	1.39	0.0089	0.0082
48 gasif-final6	39.69	1.013	82.50	10.020	0.790	1.860	2.00	2.500	16800	1.42	0.0082	0.0084
49 combustion	39.64	1.024	82.20	9.780	0.880	1.900	3.20	1.500	16350	1.37	0.0091	0.0086
50 all	41.66	1.106	79.70	9.830	0.830	1.780	5.80	1.900	15800	1.38	0.0089	0.0084

TABLE 8-3
COMPOSITE OIL ANALYSES FROM 3" REACTOR RUNS

0 RUN #	13 % CHLORIDES	14 As ppm	15 Ni ppm	16 Fe ppm	17 V ppm	18 Pour Pt (F)	19 Viscosity 40 C (cps)	20 Conradson Carbon	21 Ave Mol Weight	22 % Arom Carbon
1 KEN 4								25.20	383.0	
2 KEN 5	0.12					10	164.0	13.60	323.0	39.10
3 KEN 6						20	303.0	10.70	377.0	40.70
4 KEN 7	0.13					30	359.0	14.70	340.0	38.90
5 KEN 8	0.15					20	191.0	14.70	362.0	39.10
6 KEN 9	0.15					10	277.0	15.00	373.0	40.40
7 KEN 10	0.13					10	381.0	18.10	372.0	
8 KEN 11						5		22.60	342.0	41.00
9										
10 KEN 16						0	304.0	17.30	533.0	42.90
11 KEN 18	0.16					0	236.0	17.40	459.0	41.60
12 KEN 20	0.12					0	304.0	18.40		42.80
13 KEN 23	0.20					-10	160.0	16.60	421.0	41.10
14 KEN 24	0.11					-10	168.0	16.10	369.0	40.20
15 KEN 25	0.16					-10	220.0	16.90	372.0	41.20
16 KEN 27	0.18					-5	256.0	16.20	382.0	40.10
17 KEN 29	0.12	25.3	76.8	383	42.5	-10	148.0	17.90	360.0	40.20
18 KEN 30	0.25					0	172.0	17.80	309.0	39.60
19 KEN 31	0.13	26.1	71.7	388	36.5	0	180.0	17.30	294.0	40.00
20 KEN 32	0.23					5	144.0	17.20	332.0	40.20
21 KEN 33		15.0	69.8	328	33.4	-5	116.0	16.40	306.0	39.40
22 KEN 34						5	144.0	17.30	291.0	39.70
23 KEN 35		21.9	74.1	390	36.6	0	122.0	17.40		40.60
24 KEN 36		15.6	80.3	392	38.2	0	140.0	17.40	305.0	38.30
25										
26 KEN 46		28.0	45.2	775	68.8	0	153.6	19.23	333.0	42.30
27 KEN 47		27.1	46.2	771	58.4	-5	156.8	17.60	321.0	
28 KEN 48						0	142.4	15.39	336.0	40.25
29 KEN 49						0	174.4	16.27	387.0	41.10
30 KEN 50						-10	111.0	16.53	287.0	42.85
31 KEN 51						0	260.0	15.36	289.0	41.80
32 KEN 52						-5	129.6	15.09	320.0	40.70
33 KEN 53	0.10					-5	163.0	14.70	358.0	40.60
34 KEN 54	0.16					0	266.0	15.70	371.0	41.45
35 KEN 55	0.13									
36 KEN 56	0.08					-10	174.4	15.30	511.0	41.15
37 KEN 57	0.12					-5	136.0	15.00	362.0	40.38
38 KEN 58	0.31									
39 KEN 59	0.14					-15	119.2	14.90	416.0	40.21
40 KEN 60	0.16					0	208.8	19.60	354.0	40.09
41 KEN 61	0.14					0	190.7	16.50	397.0	40.09
42 KEN 62	0.14					-5	184.0	16.80	508.5	40.13
43 KEN 63	0.12	35.0	41.7	1070	65.8	0	154.4	14.40	417.4	40.36
44 KEN 64	0.20	30.7	48.4	779	76.4	-10	168.8	15.50	485.1	39.70
45 averages										
46 non-recycle	0.14					15	280.0	16.80	359.0	39.90
47 gasifier	0.17	20.8	75.0	376	37.0	-3	188.0	17.20	364.0	40.50
48 gasif-final6	0.18	19.7	74.0	375	36.0	1	141.0	17.20	306.0	39.70
49 combustion	0.15	30.2	45.4	849	67.4	-4	170.0	16.10	380.0	40.80
50 all	0.15	25.0	61.3	585	50.0	0	213.0	16.70	370.0	40.60

TABLE 8-3
COMPOSITE OIL ANALYSES FROM 3" REACTOR RUNS

0 RUN #	23 % Naptha (IBP-400F)	24 % Mid Dist (400-650 F)	25 % Heavy Dist (650-1000F)	26 % Residue (>1000 F)
1 KEN 4	6.0	21.0	14.0	59.0
2 KEN 5	6.0	21.0	13.0	58.0
3 KEN 6	5.0	21.0	12.0	62.0
4 KEN 7				
5 KEN 8				
6 KEN 9				
7 KEN 10				
8 KEN 11	5.0	21.0	28.0	46.0
9				
10 KEN 16	5.0	24.0	27.0	44.0
11 KEN 18	9.0	25.0	33.0	33.0
12 KEN 20	5.0	57.0	0.0	38.0
13 KEN 23	5.0	30.0	38.0	27.0
14 KEN 24	5.0	30.0	32.0	33.0
15 KEN 25	5.0	25.0	36.0	34.0
16 KEN 27	5.0	25.0	37.0	33.0
17 KEN 29	5.0	30.0	30.0	35.0
18 KEN 30	5.0	25.0	32.0	38.0
19 KEN 31	4.0	31.0	44.0	21.0
20 KEN 32	5.0	26.0	39.0	30.0
21 KEN 33	5.0	25.0	32.0	38.0
22 KEN 34	5.0	26.0	37.0	32.0
23 KEN 35	5.0	31.0	43.0	21.0
24 KEN 36	5.0	30.0	40.0	25.0
25				
26 KEN 46	5.0	23.0	33.0	39.0
27 KEN 47	5.0	24.0	35.0	36.0
28 KEN 48	4.0	25.0	37.0	34.0
29 KEN 49	5.0	22.0	34.0	39.0
30 KEN 50	5.0	25.0	36.0	34.0
31 KEN 51	4.0	23.0	41.0	32.0
32 KEN 52	5.0	25.0	40.0	30.0
33 KEN 53	4.0	26.0	39.0	31.0
34 KEN 54	4.0	21.0	37.0	38.0
35 KEN 55				
36 KEN 56	5.0	25.0	39.0	31.0
37 KEN 57	5.0	25.0	38.0	32.0
38 KEN 58				
39 KEN 59	5.0	27.0	38.0	30.0
40 KEN 60	4.0	29.0	37.0	33.0
41 KEN 61	5.0	25.0	37.0	33.0
42 KEN 62	5.0	26.0	37.0	32.0
43 KEN 63	5.0	24.0	37.0	34.0
44 KEN 64	5.0	28.0	42.0	25.0
45 averages				
46 non-recycle	5.5	21.0	16.8	56.3
47 gasifier	5.2	29.3	33.3	32.1
48 gasif-final6	4.8	28.2	39.2	27.8
49 combustion	4.7	24.9	37.5	33.1
50 all	5.0	26.3	33.4	35.3

Table 8-4. Hydrocarbon Gas Analyses for Mini-Plant Runs.

0 KEN#	1 % CARBON TO HC GAS	2 CH4 % C	3 C2'S % C	4 C3'S % C	5 C4'S % C	6 ETHENE/ ETHANE RATIO	7 PROPENE/ PROPANE RATIO	8 1-BUTENE/ n-BUTANE RATIO	9 C2+C3 + C4	10 C1/C4 RATIO
1 KEN#4	4.46	1.610	1.06	0.91	0.87		1.21	1.31	2.84	1.84
2 KEN#5	4.08	1.550	0.99	0.84	0.70		1.26	1.33	2.53	2.21
3 KEN#6	4.85	1.940	1.14	0.98	0.80		1.23	1.31	2.92	2.43
4 KEN#7	4.48	1.650	1.07	0.94	0.82		1.21	1.32	2.83	2.01
5 KEN#8	4.21	1.440	1.01	0.91	0.84	0.51	1.25	1.20	2.76	1.71
6 KEN#9	4.77	1.760	1.13	0.96	0.92	0.36	1.22	1.23	3.01	1.91
7 KEN#10	4.64	1.720	1.09	0.93	0.90	0.49	1.21	1.31	2.92	1.91
8 KEN#11	4.82	1.790	1.16	0.97	0.90	0.37	1.22	1.31	3.03	1.98
9										
10 KEN#16	4.70	1.670	1.16	0.98	0.90	0.44	1.18	1.26	3.04	1.85
11 KEN#18	4.17	1.480	0.99	0.86	0.84	0.41	1.15	1.17	2.69	1.77
12 KEN#20	4.98	1.720	1.24	1.04	0.98	0.40	1.18	1.26	3.26	1.75
13 KEN#23	5.43	1.920	1.36	1.12	1.04	0.41	1.14	1.30	3.52	1.85
14 KEN#24	5.26	1.900	1.29	1.08	0.99	0.42	1.17	1.31	3.36	1.91
15 KEN#25	4.37	1.580	1.06	0.93	0.80	0.37	1.15	1.19	2.79	1.97
16 KEN#27	4.20	1.500	1.01	0.86	0.84	0.43	1.17	1.30	2.71	1.79
17 KEN#29	5.00	1.730	1.25	1.05	0.98	0.45	1.21	1.30	3.28	1.76
18 KEN#30	4.94	1.810	1.19	0.99	0.96	0.47	1.20	1.28	3.14	1.89
19 KEN#31	4.52	1.660	1.12	0.92	0.82	0.52	1.32	1.45	2.86	2.03
20 KEN#32	5.01	1.830	1.25	1.02	0.92	0.48	1.26	1.27	3.19	1.99
21 KEN#33	4.89	1.730	1.22	1.00	0.94	0.47	1.23	1.29	3.16	1.84
22 KEN#34	5.03	1.810	1.23	1.02	0.97	0.47	1.23	1.36	3.22	1.86
23 KEN#35	5.59	2.120	1.34	1.11	1.02	0.52	1.32	1.43	3.47	2.07
24 KEN#36	5.40	1.890	1.25	1.03	1.24	0.48	1.24	1.24	3.52	1.53
25										
26 KEN#46	4.07	1.420	1.05	0.86	0.74	0.52	1.33	1.52	4.07	1.92
27 KEN#47	3.83	1.400	0.98	0.78	0.67	0.60	1.41	1.45	3.83	2.09
28 KEN#48	5.00	1.720	1.29	1.10	0.89	0.53	1.32	1.28	5.00	1.93
29 KEN#49	4.07	1.370	1.05	0.86	0.79	0.47	1.38	1.47	4.07	1.73
30 KEN#50	4.65	1.600	1.21	0.98	0.86				4.65	1.86
31 KEN#51	4.22	1.520	1.07	0.87	0.76	0.53	1.40	1.46	4.22	2.00
32 KEN#52	4.15	1.470	1.08	0.89	0.71	0.51	1.37	1.44	4.15	2.07
33 KEN#53	4.21	1.500	1.08	0.89	0.74			1.42	4.21	2.03
34 KEN#54	3.66	1.240	0.93	0.78	0.71		1.34	1.35	3.66	1.75
35 KEN#55	3.44	1.270	0.88	0.71	0.58		1.35	1.36	3.44	2.19
36 KEN#56	3.34	1.180	0.88	0.72	0.56		1.30	1.39	3.34	2.11
37 KEN#57	4.97	1.790	1.25	1.01	0.92	0.52	1.38	1.49	4.97	1.95
38 KEN#58	4.75	1.680	1.19	0.99	0.89	0.47	1.36	1.44	4.75	1.89
39 KEN#59	4.74	1.660	1.24	1.01	0.83			1.46	4.74	2.00
40 KEN#60	4.66	1.700	1.16	0.95	0.85	0.59	1.40		4.66	2.00
41 KEN#61	4.22	1.470	1.06	0.89	0.80	0.56	1.33	1.39	4.22	1.84
42 KEN#62	4.71	1.690	1.24	1.00	0.78	0.55	1.36	1.47	4.71	2.17
43 KEN#63	3.87	1.400	0.99	0.81	0.67	0.52	1.29	1.38	3.87	2.09
44 KEN#64	4.10	1.500	1.04	0.85	0.71	0.53	1.35	1.44	4.10	2.11
45 AVERAGES										
46 non-recycle	4.54	1.683	1.08	0.93	0.85	0.43	1.23	1.29	2.86	2.00
47 gasified	4.90	1.757	1.20	1.00	0.95	0.45	1.21	1.29	3.15	1.86
48 gasif-6	5.07	1.840	1.24	1.02	0.98	0.49	1.27	1.34	3.24	1.89
49 combusted	4.25	1.504	1.09	0.89	0.76	0.53	1.35	1.42	4.25	1.99
50 all	4.53	1.628	1.13	0.94	0.84	0.48	1.27	1.35	3.59	1.94

Table 8-5. Water Analyses for Mini-Plant Runs.

0 RUN #	1 Ph	2 % Suspended SolidsX100	3 Dissolved Solids (%)	4 % C total	5 S total (ppm)	6 NH3 (ppm)	7 Cr ppm	8 Mo ppm	9 As ppm	10 Cd ppm	11 Pb ppm
1 KEN 4	8.67	0.09	0.067	0.22	79		<.1	<.2	<.3	0.03	<.3
2 KEN 5	8.63	0.03	0.110	0.27	1515		<.1	<.2	<.3	0.03	<.3
3 KEN 6	8.59	0.10	0.079	0.26	1775		<.1	<.2	<.3	0.03	<.3
4 KEN 7	8.73	0.09	0.107	0.28	109		<.1	<.2	<.3	<.01	<.3
5 KEN 8	6.53	0.19	0.410	0.28	3035		<.1	<.2	<.3	<.01	<.3
6 KEN 9	8.57	0.21	0.123	0.23	2805		<.1	<.2	<.3	<.01	<.3
7 KEN 10	8.78	0.13	0.113	0.25	2940		<.1	<.2	<.3	<.01	<.3
8 KEN 11	8.81	0.16	0.111	0.26	319		<.1	<.2	<.3	<.01	<.3
9											
10 KEN 16	8.73	0.21	0.144	0.29	600		<.1	<.2	<.3	<.01	0.7
11 KEN 18	8.66	0.12	0.163	0.22	1700		<.1	<.2	<.3	<.01	<.4
12 KEN 20	7.95	0.18	0.070	0.27	1200		<.1	<.2	<.3	<.01	0.5
13 KEN 23	8.23	0.15	0.160	0.24	1200		<.1	0.5	<.4	0.05	<.4
14 KEN 24	9.08	0.08	0.069	0.27	1600		<.1	0.5	<.4	0.05	<.4
15 KEN 25	8.82	0.05	0.193	0.31	1200		<.1	0.5	<.4	<.01	<.4
16 KEN 27	9.51	0.05	0.219	0.31	1400		<.1	0.4	<.4	<.01	<.4
17 KEN 28	9.39	0.09	0.183	0.30	1400		<.1	0.5	<.4	0.03	<.4
18 KEN 29	8.70	0.12	0.177	0.30	1200		<.1	<.3	<.4	0.02	<.4
19 KEN 30	9.11	0.22	0.330	0.40	1395		<.1	<.2	<.4	<.01	<.4
20 KEN 31	8.60	0.08	0.401	0.34	1900		<.1	<.2	<.4	<.01	<.4
21 KEN 32	8.76	0.20	0.342	0.41	2175	2610	<.1	<.2	<.4	<.01	<.4
22 KEN 33	8.70	0.16	0.448	0.37	2855	2320	<.1	<.2	<.4	<.01	<.4
23 KEN 34	8.66	0.12	0.476	0.38	2890	2490	<.1	<.2	<.4	<.01	<.4
24 KEN 35	8.64	0.13	0.303	0.31	3195	3940	<.1	<.2	<.4	<.01	<.4
25 KEN 36	8.67	0.20	0.190	0.42	2140	3560	<.1	<.2	<.4	<.01	<.4
26											
27 KEN 46	8.52	0.15	0.380	0.58	1200	2270	<.01	<.1	<.1	<.1	<.1
28 KEN 47	8.54	0.18	0.420	0.40	555	2040	<.1	<.1	<.1	<.1	<.1
29 KEN 48	8.43	0.15	0.370	0.77	269	2330	<.1	<.1	<.1	<.1	<.1
30 KEN 49	8.51	0.31	0.360	0.49	208	2160	<.1	<.1	<.1	<.1	<.1
31 KEN 50	8.30	0.40	0.400	0.47	324	2470					
32 KEN 51	8.56	0.14	0.340	0.65	497	2570					
33 KEN 52	8.47	0.26	0.280	0.39	48	2430					
34 KEN 53	8.61	0.30	0.360	0.49	2070	2323					
35 KEN 54	8.43	0.21	0.340	0.38	2005	2200					
36 KEN 55	8.51	0.44	0.370	0.36	1160						
37 KEN 56	8.30	0.11	0.340	0.34	243	2140					
38 KEN 57	8.45	0.25	0.420	0.35	223	2510					
39 KEN 58	8.39	0.08	0.360	0.31	1825		<.5	<.5		<.5	<.5
40 KEN 59	8.40	0.13	0.410	0.46	512	2360	<.5	<.5		<.5	<.5
41 KEN 60	8.62	0.35	0.240	0.41	578	2760	<.5	<.5		<.5	<.5
42 KEN 61	8.53	0.29	0.400	0.30	512	2700	<.5	<.5		<.5	<.5
43 KEN 62	8.57	0.22	0.410	0.23	287	2260	<.5	<.5		<.5	<.5
44 KEN 63	8.68		0.460	0.23	161	1720	<.5	<.5		<.5	<.5
45 KEN 64	8.63	0.53	0.460	0.25	2135	2020	<.5	<.5		<.5	<.5
46 AVERAGES											
47 non-recycle	8.41	0.13	0.140	0.26	1572						
48 gasifier-all	8.82	0.13	0.170	0.29	1290						
49 gasif, final-6	8.67	0.15	0.360	0.37	2526	2984					
50 combusted	8.50	0.25	0.370	0.41	780	2310					
51 all runs	8.58	0.18	0.280	0.35	1289	2463					

TABLE 8-5
WATER ANALYSES FOR 3" INTEGRATED REACTOR RUNS

0 RUN #	12 Se ppm	13 Ni ppm	14 Fe ppm	15 Mn ppm	16 Cu ppm	17 Mg ppm	18 Na ppm	19 Ca ppm	20 V ppm	21 Zn ppm	22 Al ppm	23 K ppm
1 KEN 4	1.60	<.1	0.37	<.01	<.01				<.01	0.03	0.50	
2 KEN 5	1.50	<.1	0.25	<.01	<.01				<.01	0.05	0.35	
3 KEN 6	1.40	<.1	0.15	0.02	<.01				<.01	<.03	0.50	
4 KEN 7	1.00	<.1	0.14	<.01	<.01				<.01	0.02	0.31	
5 KEN 8	0.90	<.1	1.00	<.01	<.01				<.01	0.05	0.40	
6 KEN 9	0.90	<.1	0.24	0.05	<.01				<.01	0.04	0.44	
7 KEN 10	2.80	<.1	0.09	0.01	<.01				<.01	<.03	0.20	
8 KEN 11	2.40	<.1	0.12	<.01	<.01				<.01	<.05	0.06	
9												
10 KEN 16	<.3	0.23	0.75	<.01	<.01	0.220	1.01	0.25	0.04	<.1	1.33	1.30
11 KEN 18	0.90	<.1	0.23	0.10	<.01	0.060	0.31	0.18	<.01	<.1	0.34	0.50
12 KEN 20	0.50	<.1	0.26	<.01	<.01	0.050	0.27	0.18	<.01	<.1	0.26	0.17
13 KEN 23	0.50	0.14	0.19	0.04	<.01		0.52	0.59	<.01	<.1	0.29	0.26
14 KEN 24	1.70	0.15	0.15	0.04	0.01		0.22	0.16	<.01	<.1	0.13	0.13
15 KEN 25	1.20	<.1	0.27	0.03	0.01		0.92	0.24	<.01	0.14	0.53	0.35
16 KEN 27	0.80	<.1	0.21	0.02	<.01		0.34	0.10	<.01	<.1	0.32	0.13
17 KEN 28	0.70	<.1	0.34	0.02	<.01		0.54	0.25	<.01	<.1	0.24	0.10
18 KEN 29	1.60	<.1	0.21	<.01	0.01		0.27	0.17	<.01	<.1	0.17	0.32
19 KEN 30	<.3	<.1	0.14	<.01	<.01	0.070	0.88	0.31	<.1	0.27	<.01	
20 KEN 31	2.80	<.1	0.18	<.01	0.02	0.040	0.30	0.14	<.01	<.1	0.19	<.01
21 KEN 32	2.90	<.1	0.14	<.01	0.01	0.040	0.23	0.15	<.01	<.1	0.19	<.01
22 KEN 33	<.3	<.1	0.27	<.01	<.01	0.050	0.35	0.16	<.01	<.1	0.33	<.01
23 KEN 34	2.40	<.1	0.20	<.01	<.01	0.040	0.26	0.16	<.01	<.1	0.23	<.01
24 KEN 35	2.70	<.1	0.25	<.01	0.01	0.040	0.24	0.16	<.01	<.1	0.20	<.01
25 KEN 36	2.00	<.1	0.26	<.01	0.02	0.050	0.48	0.32	<.01	<.1	0.24	<.01
26												
27 KEN 46	2.57	<0.1	0.20	<0.1	<0.1	0.060	0.97	0.25	<0.1	<0.1	0.70	0.41
28 KEN 47	0.67	<0.1	1.27	<0.1	<0.1	0.110	0.99	0.24	<0.1	<0.1	0.76	0.57
29 KEN 48	1.01	<0.1	0.25	<0.1	<0.1	0.070	1.08	0.34	<0.1	<0.1	0.66	0.59
30 KEN 49	1.60	<0.1	0.21	<0.1	<0.1	0.040	0.70	0.19	<0.1	<0.1	0.44	0.28
31 KEN 50												
32 KEN 51												
33 KEN 52												
34 KEN 53												
35 KEN 54												
36 KEN 55												
37 KEN 56												
38 KEN 57												
39 KEN 58	0.93	<.1	2.64	<.1	<.1	0.100	1.23	0.29	0.17	<.1	1.57	0.38
40 KEN 59	<.5	<.1	0.31	<.1	<.1	0.120	1.32	0.33	<.1	<.1	1.99	0.50
41 KEN 60	1.61	<.1	0.41	<.1	<.1	0.130	0.74	0.27	<.1	<.1	2.18	0.57
42 KEN 61	3.82	<.1	0.20	<.1	<.1	0.110	0.29	0.20	<.1	<.1	1.21	0.48
43 KEN 62	0.83	<.1	0.32	<.1	<.1	0.130	1.05	0.55	<.1	<.1	2.15	0.68
44 KEN 63	1.01	<.1	0.75	<.1	<.1	0.140	1.22	0.30	<.1	<.1	2.19	0.57
45 KEN 64	0.06	<.1	0.41	<.1	<.1	0.180	2.11	0.82	<.1	<.1	2.55	0.64
46 AVERAGES												
47 non-recycle			0.30									
48 gasifier-all			0.25			0.067	0.45	0.22				
49 gasif. final-6			0.22			0.043	0.31	0.18				
50 combusted			0.63			0.108	1.06	0.34				
51 all runs			0.38			0.088	0.70	0.27				

8.5 LIGHT OIL ANALYSES

Light oil properties and production were relatively constant for all runs (Table 8-1) exhibiting only minimal differences in the averages for the non-recycle, gasified solids recycle, and combusted solids recycle runs. The scatter that is present is attributed more to run-to-run differences in the ESP trapping efficiency (resulting in a variable portion of the total oil partitioning to the light oil fraction) than to inherent differences in the light oil product. This is supported by a direct correlation between light oil production and light oil density for all runs since partitioning of larger oil components to the LO fraction results in an increase in density.

However, for those runs in which solids recycle rate and temperature were coupled (combustor and final 6 gasifier), there was also a correlation between the rate of solids recycle from the gasifier and the amount of light oil production. This is consistent with additional cracking of vapor phase hydrocarbons on the surface of the hot recycle particles within the pyrolyzer. Comparisons between gasifier-only and gasifier/combustor runs show that both modes resulted in enhanced light oil production relative to the non-recycle runs. However, the oils from the integrated combustor runs were of higher density and lower H/C molar ratio than those from otherwise similar, gasifier-only runs. This suggests a propensity of the combusted particles to favor cracking reactions relative to the gasified particles which, as to be shown later, appear to be more prone to selectively coke olefinic and aromatic components. For the combustion runs, it was also noted that total production, density, and the H/C molar ratio of the light oils correlated with either combustor temperature or the rate of recycle of combusted solids to the gasifier. This also suggests differences in catalytic activity between gasified and combusted particle surfaces. This latter observation should be treated with caution, however, since the combustor operating parameters were not truly independent of gasifier operation, i.e., the temperature and solids recycle rate to the pyrolyzer from the gasifier were correlated with combustor temperature. Thus, based solely on analyses of the light oils, it is not totally clear that combusted particles enhanced cracking vapor phase hydrocarbons relative to gasified particles which appeared to favor coking reactions. This subject be addressed in more detail in the composite oils analyses section that follows.

8.6 ESP OIL ANALYSES

Analogous analytical data for the ESP oils (Table 8-2) generally exhibit little or no correlation with the processing variables of gasifier-, pyrolysis-, or solids recycle temperature, solids recycle rate, or raw shale feed rate. This lack of correlation is attributed to several factors including

the previously mentioned run-to-run variation of oil partitioning between the ESP and light oil condenser; the difficulty in obtaining representative samples for analyses due to the viscous nature and high water content of the ESP oils; the fact that the single most important variable in terms of oil properties, pyrolysis temperature, was varied over a relatively narrow range; and the complex interaction of processing variables in this system which often have opposing effects on a given oil parameter and thus tend to mask correlations. Consequently, the ESP properties will not be discussed further except to note that the ash contents of the gasifier-only and gasifier/combustion runs were essentially the same. This similarity in ash content provides evidence that solid-solid heat transfer using gasified versus combusted particles does not result in measurable differences in particle attrition and carryover to the oil collection train.

8.7 COMPOSITE OIL AND HYDROCARBON GAS ANALYSES

Tables 8-3 and 8-4 present analytical data for the composite oils and non-condensed gases from the gasifier/pyrolyzer section. Again, average values for the non-recycle (runs 4-11), gasifier recycle (runs 16-36), final 6 gasifier recycle (31-36), combustor recycle (46-64), and all runs are shown at the bottom of each table.

A major concern in this program was the potential for oil yield loss due to the recycling of hot solids to the retort. Figure 8-1 shows a weak inverse relation between the portion of raw shale carbon that partitioned to the oil fraction and the temperature of the particles that were recycled from the gasifier to the retort. Though there is significant scatter in this data, attributable to a large extent to variations in the solids recycle rate, there still appears to be a correlation between oil loss and solids recycle temperature.

The most obvious explanation for this oil loss would be enhanced cracking and coking of vapor phase hydrocarbons on the recycle solids surfaces. Since it is known that HC cracking reactions result in an increase in the 1-alkene/n-alkane ratio(2) (coking reactions tend to decrease this ratio), the ethene/ethane ratio versus oil yield plot (Figure 8-2) is congruent with cracking of vapor phase hydrocarbons on the surface of the hot recycle solids. Additional evidence of the recycle solids activity is given in Figure 8-3 which shows an inverse relation between the average molecular weight of the composite oil and the rate of solids recycle to the retort. Though not plotted, composite oil viscosity and solids recycle rate exhibited a similar inverse relation. It is interesting to note that composite oil average molecular weight and viscosity values were also inversely related to recycle solids temperature for gasifier run #'s 16-30 in which recycle rate and temperature

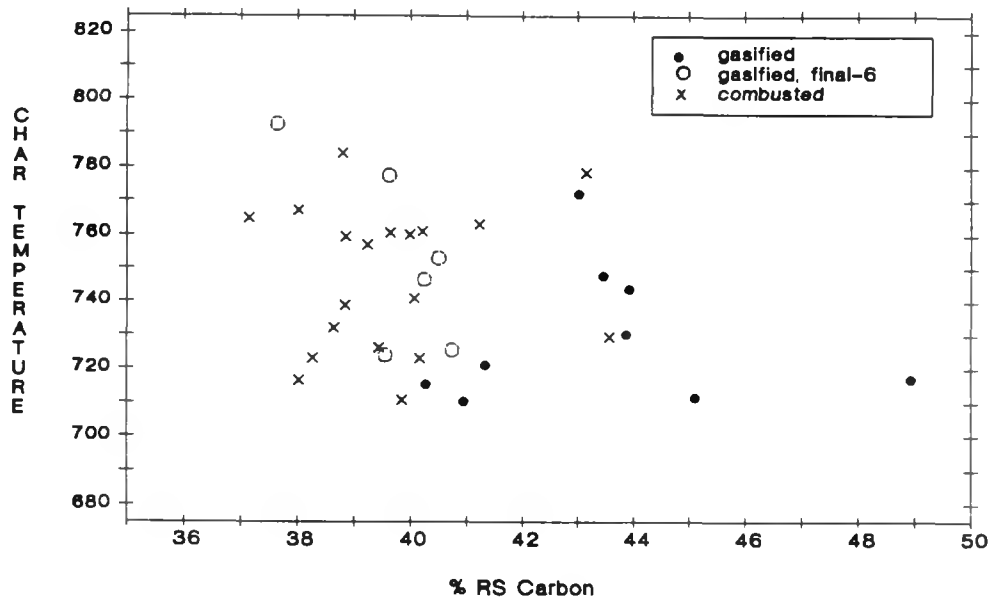


Figure 8-1. Percent Raw carbon Partitioning to the Oil Phase as a Function of Recycle Char Temperature.

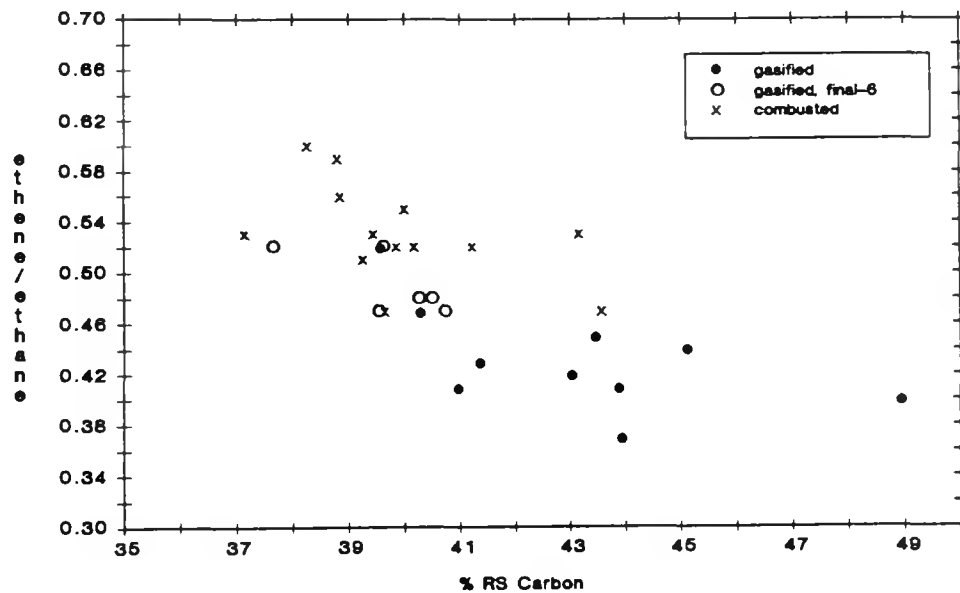


Figure 8-2. Percent Raw Shale Carbon Partitioning to the Oil Phase vs. the Ethene/Ethane Ratio.

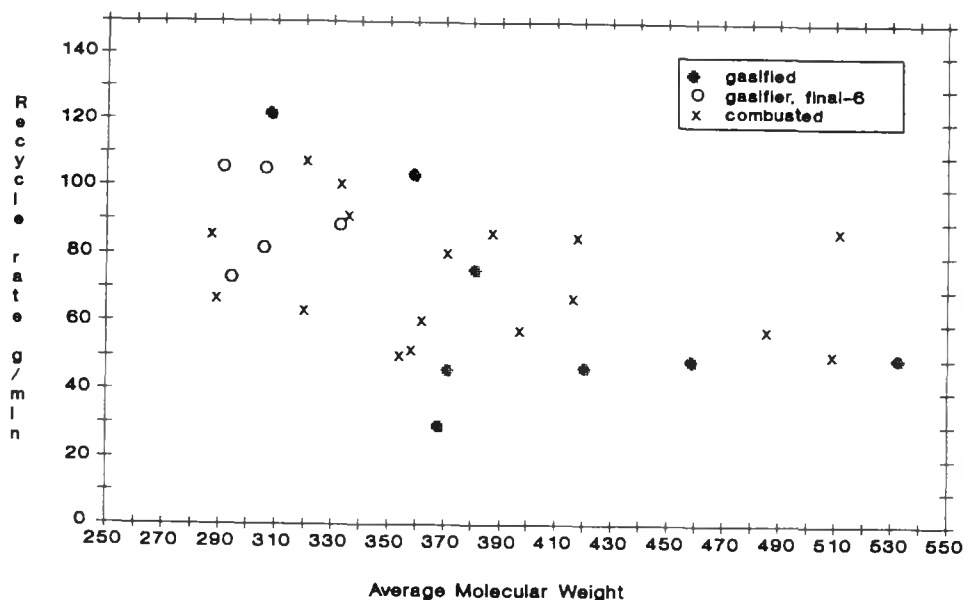


Figure 8-3. Average Molecular Weight of the Composite Oil as a Function of Pyrolysis Recycle Rate.

were essentially independent variables. However, for the runs in which recycle temperature and rate were inversely coupled, i.e., the final six gasifier runs (31-36) and all combustor runs, viscosity and molecular weight were more directly related to solids recycle temperature.

Since the introduction of hot recycle solids to the retort resulted in modest oil losses, the next question to be addressed was whether there was a measurable difference in the activity or mechanism of oil loss induced by the recycled gasified particles relative to that by recycled gasified/combusted particles. The relation between the composite oil aromatic content and solids recycle rate for gasified (Fig. 8-4a) and combination gasified/combusted (Fig. 8-4b) particles exhibit opposing trends suggesting perhaps a difference in the mechanism of loss. Figures 8-5a and 8-5b likewise show opposite trends in the relation between aromatic carbon content and average molecular weight of the product oil produced from recycling gasified versus combined gasified/combusted particles. Scrutiny of the composite oil density and H/C ratios versus recycle rate confirm these trends as real and not an aberration of the analytical scatter. Thus, even though both solids recycle streams tend to produce an overall lighter product, the relative composition of this product exhibits measurable differences that relate to the composition of the recycle particles utilized for heat transfer.

Though there is little doubt that a combination of coking and cracking vapor phase reactions are induced by both types of recycle solids streams, one explanation for the above findings (Figures 8-4 and 8-5) is that gasified shale surfaces tend to favor coking reactions relative to combusted particles which appear to favor cracking reactions. A preference for coking over cracking reactions by the gasified recycle particles relative to a combination

gasified/combusted recycle stream should be reflected by a lower 1-alkene/n-alkane ratio. This is shown in Figures 8-6a and 8-6b which exhibit propene/propane ratios as a function of recycle temperature (6a) and recycle rate (6b).

These ratios follow parallel trends for both gasifier solids and gasifier/combustor solids recycle. However, at equivalent recycle temperatures or rates, the alkene/alkane ratio is lower in both plots for gasified-only recycle. Plots of the ethene/ethane and 1-butene/n-butane ratios exhibited similar trends though the ethene/ethane relation was not as distinct due perhaps to

variable primary production of these components within the gasifier and/or the beta-scission mechanism exhibited by free radical chains. These observations can be interpreted as preferential removal of alkenes by gasified solids and/or preferential cracking induced by recycling gasified/combusted particles

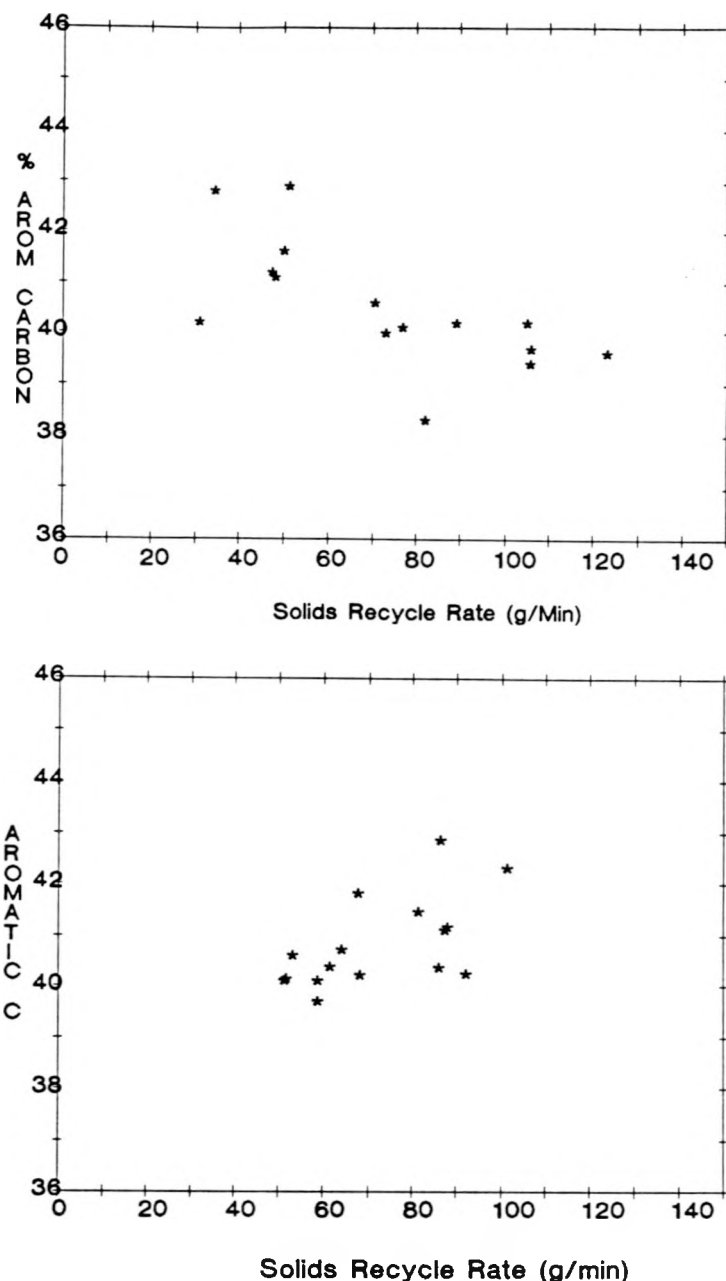


Figure 8-4. a) Composite Oil Aromatic Carbon Content as a Function of Solids Recycle Rate (Top); b) CO Aromatic Carbon Content vs. Solids Recycle Rate in Non-Combustion Mode (Bottom).

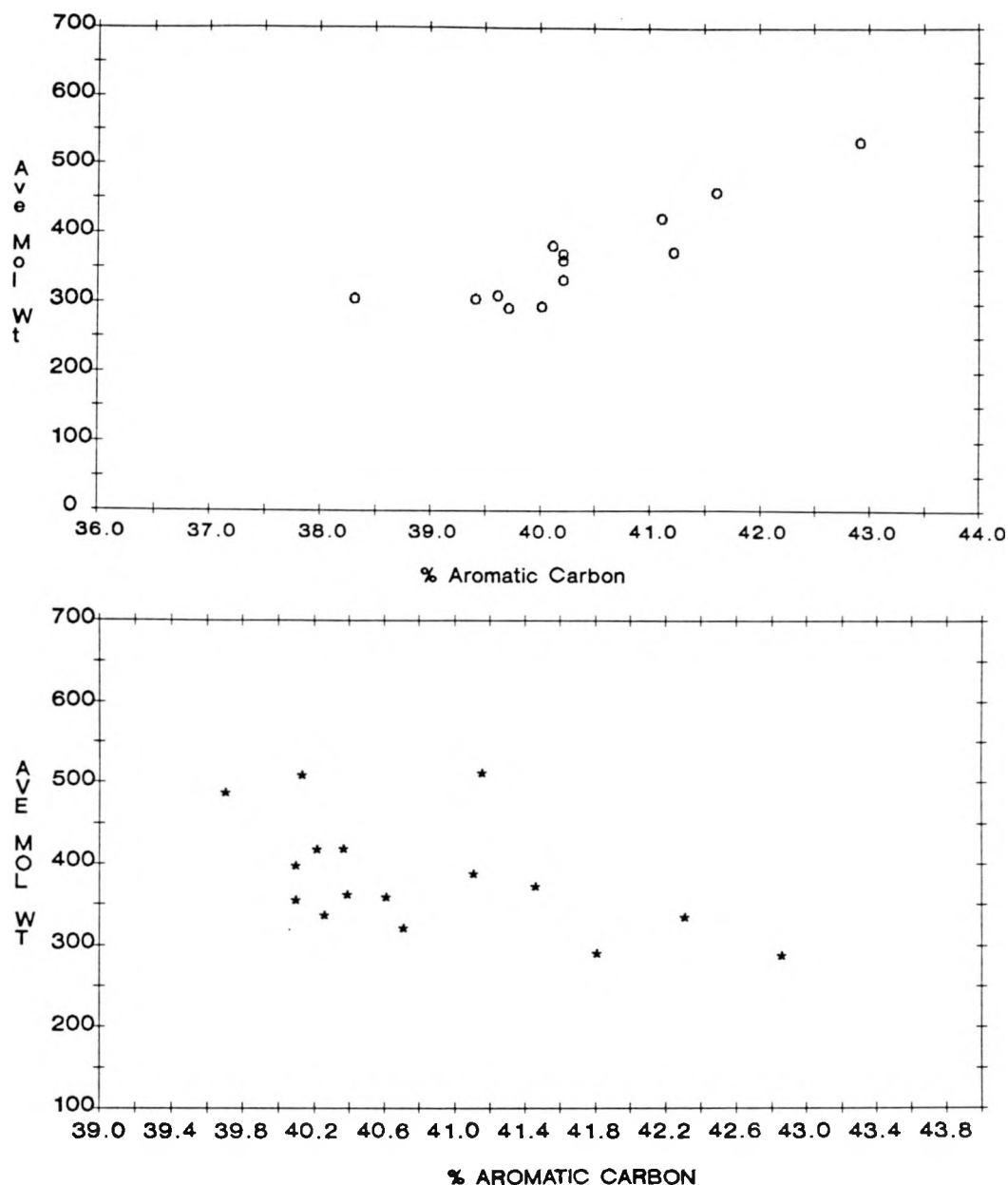


Figure 8-5. a) Average Molecular Weight vs. Aromatic Carbon in Non-Combustion Mode (Top); b) Aromatic Carbon in the Composite Oil vs. Average Mol. Wt. in the Combustion Mode (Bottom).

Preferential coking of alkenes (and to a lesser extent aromatics) should produce additional stable hydrocarbon gases relative to cracking reactions. This is because cracking reactions tend to produce a higher proportion of vapor phase free radicals which rapidly undergo addition and/or cyclization/dehydrogenation reactions.(3) This ultimately yields a product with a relatively higher average molecular weight and aromatic content. This proposed pathway is supported by the plot of Figure 8-7 showing little correlation between the rate of solids recycle and the portion of the raw shale carbon

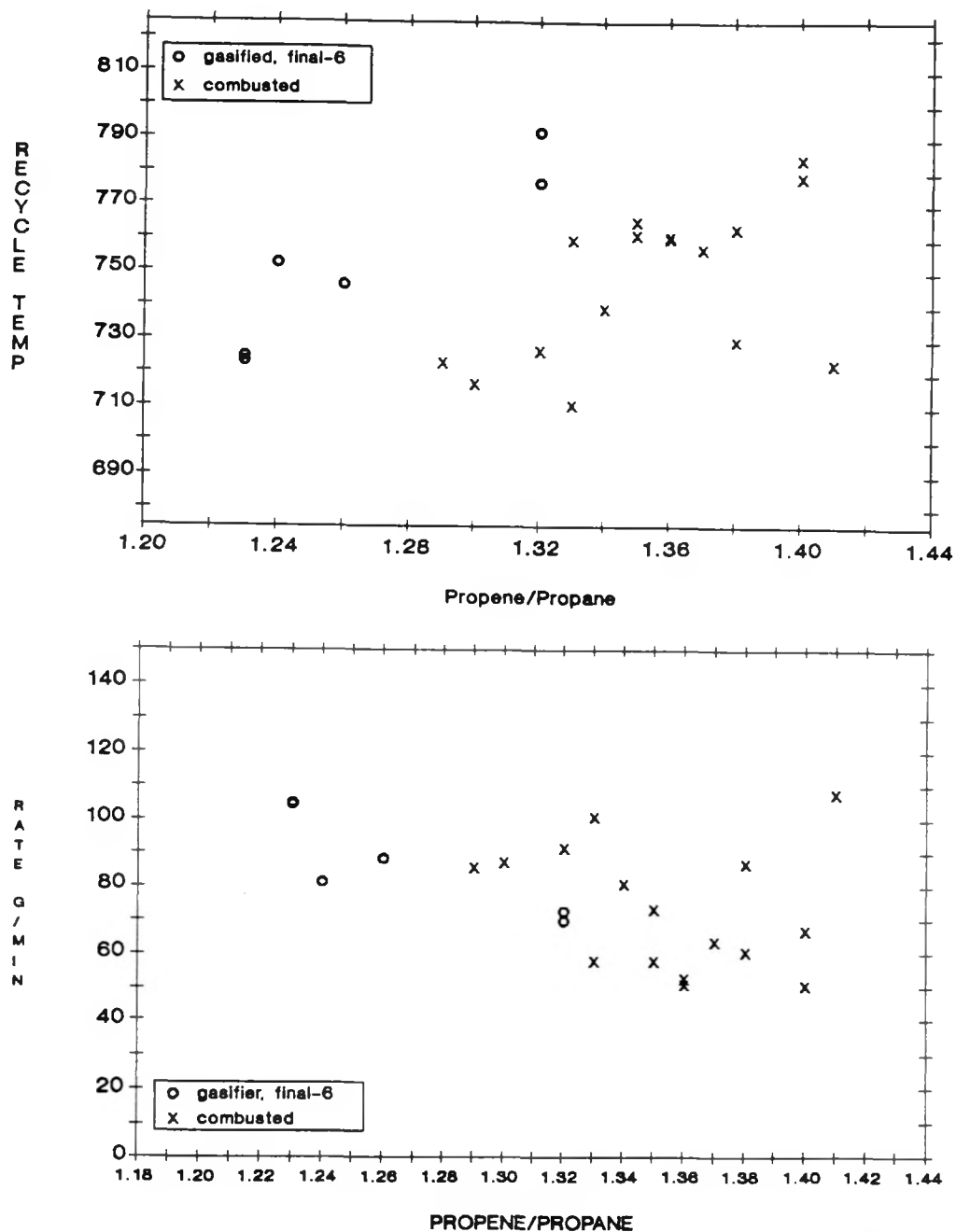


Figure 8-6. a) Propene/Propane Ratio vs. Solids Recycle Temperature in Non-Combustion Mode (Top); b) Propene/Propane Ratio vs. Solids Recycle Rate in Combustion Mode (Bottom).

that partitions to the gas phase, but, displaying a clear increase in HC gas production for recycled gasified-only particles. It is also in agreement with Figures 8-4a and 8-4b showing a reverse relationship between the aromatic content of the product oil and the recycle of gasified versus gasified/combusted solids. This relation is shown again in Figure 8-8 where the aromatic content of the composite oil is

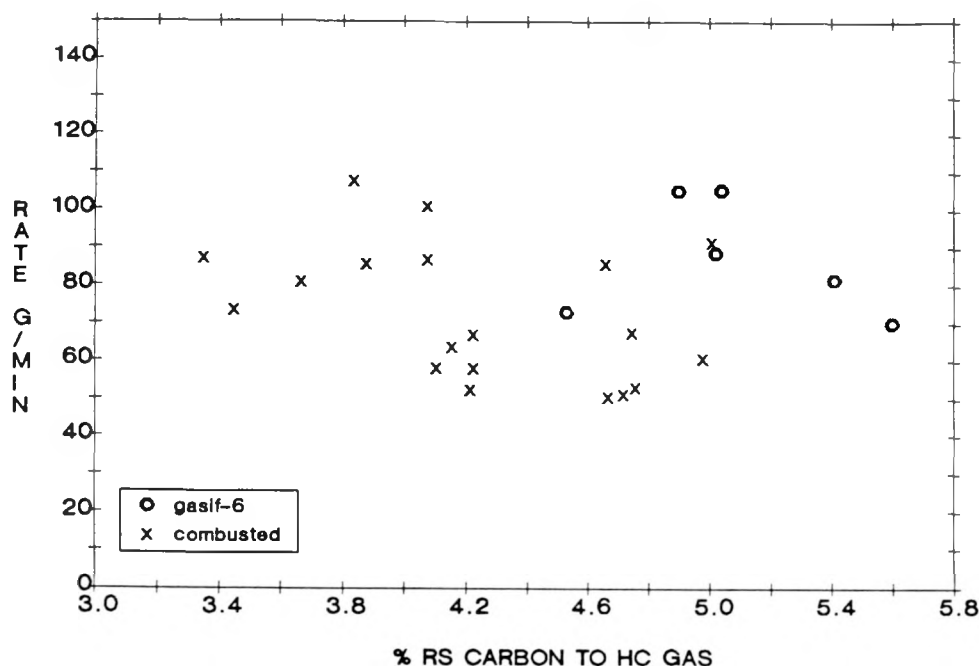


Figure 8-7. *HC Gas Production vs. Solids Recycle Rate in Non-Combustion Mode.*

plotted against the product of the gasifier/pyrolyzer recycle rate and the combustor/gasifier recycle ratio. This latter parameter is a rough measure of the proportion of combusted particles in the gasifier/pyrolyzer recycle stream. Figure 8-8 demonstrates that when the proportion of gasified versus combusted particles in the solids recycle stream is factored in, the aromatic content-recycle rate plot exhibits a somewhat improved correlation relative to Figure 8-4b where only the recycle rate is plotted. The average values (Table 8-3) for density and average molecular weights, which are greater for the combustor recycle runs, and the H/C molar ratios, which are greater for the final 6 gasifier runs, are all consistent with the reaction pathways proposed above.

In summary, the recycling of hot solids to the retort had a qualitatively predictable effect on the product yield and properties. Though there appear to be subtle differences in the mechanism of yield reduction induced by the recycling of gasified-only versus gasified/combusted solids, it should be noted that these differences are relatively minor and that oil yields were essentially the same for both modes of operation. Further, though oil losses for solids recycle versus non-recycle were measurable, the oil product from solids recycling was a somewhat improved product as evidenced by a higher average H/C

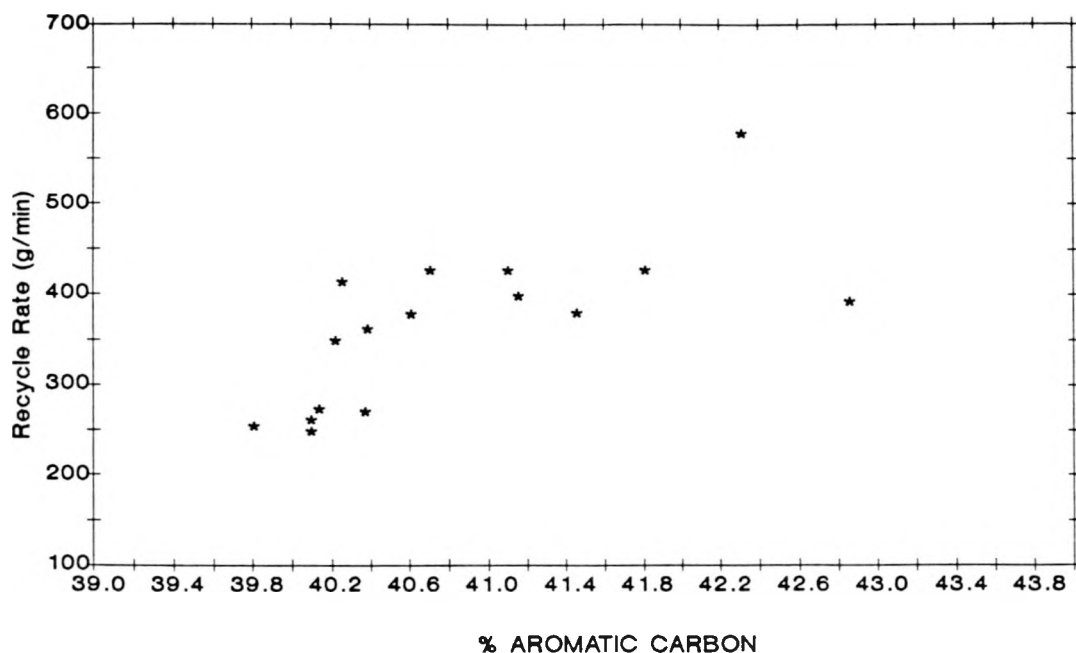


Figure 8-8. *CO Aromatic Carbon Content vs. Product of Pyrolyzer Recycle Rate and the Combustor Recycle Ratio.*

ratio, lower viscosity, lower average molecular weight, and an improved boiling range distribution.

8.8 REFERENCES

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9. MINERALOGIC CHANGES INDUCED BY KENTORT II PROCESSING

9.1 INTRODUCTION

The objective of this phase of the project was to determine the nature of the mineralogy of Eastern oil shale and the extent of mineralogic transformations which take place during processing in the KENTORT II technology. These transformations are significant as they affect physical and chemical properties of the spent shale and perhaps the surface activity which, in turn, may affect the rate and mechanism of oil loss during processing.

In addition to shales from the process steps of KENTORT II, other materials

examined included: a series of unprocessed oil shales from various locations in the oil shale outcrop belt in Kentucky; shales which have been systematically pyrolyzed in a fixed bed retort; and, materials which have been processed in the Dravo and PETROSIX's oil shale retort pilot plants. (1,2)

TABLE 9-1. *Normalized Silicate Composition. Cleveland Value is Average of 3 Samples.*

	Il	Chl	Kao	Qtz
Clegg Creek	71	10	7	11
Blocher	68	3	3	26
Sunbury	61	11	6	20
Cleveland	63	10	3	23

The oil shales examined include the Sunbury Shale, recovered from a core drilled in Fleming County Kentucky; three samples of the Cleveland Member of the Ohio Shale, from bulk samples collected in Lewis, Fleming and Montgomery Counties of Kentucky; and the Blocher and Clegg Creek Member of the New Albany Shale, collected from a core in Marion County, Kentucky (Table 9-1). The Cleveland sample from Montgomery County was the feedstock for the Petrobras, KENTORT II, and fixed bed experiments. The feedstock for the Dravo retort test consisted of 1/3 Sunbury and 2/3 Cleveland Shales also mined in Montgomery County, Kentucky.

9.2 METHODS AND PROCEDURES

9.2.1 Clay Analysis. One of the objectives of the study was to quantitatively determine the clay composition of the shales. A brief description of the procedure used follows; more detailed discussions can be found in Rich and Barnhisel. (3)

The samples were crushed to approximately 6.4 mm with a jaw crusher, then pulverized in a ring and puck mill for 5 minutes.

The powdered samples were low temperature ashed and/or treated with hydrogen peroxide to remove the organic matter. Unoxidized sample splits were also processed. The samples were then separated into size fractions using an elutriation technique.(4) Each size fraction was vacuum oriented on clay tiles. Five sample size fractions were separated for the materials from the fixed bed retort, and two sizes for the rest of the samples. Size separation is necessary to keep larger particles from masking smaller particles in X-ray diffraction analysis, as the clay minerals typically have smaller particle size distributions and would be the last to settle out. Separate tiles were saturated with either potassium or magnesium and heat treated. The various treatments are necessary to determine the nature and abundance of mixed layered clays.

The samples were subjected to Cu K- α radiation. The K-saturated samples were scanned after heat treatments at 25, 110, 300 and 550 °C. The Mg-saturated and glycerated fractions were scanned following 25 and 110 °C heat treatments. The proportions of the clay minerals in the feedstock samples were computed from peak height measurements of the basal reflections (001) of the clays and the 4.3 Å quartz peak (Figure 9-1).(3)

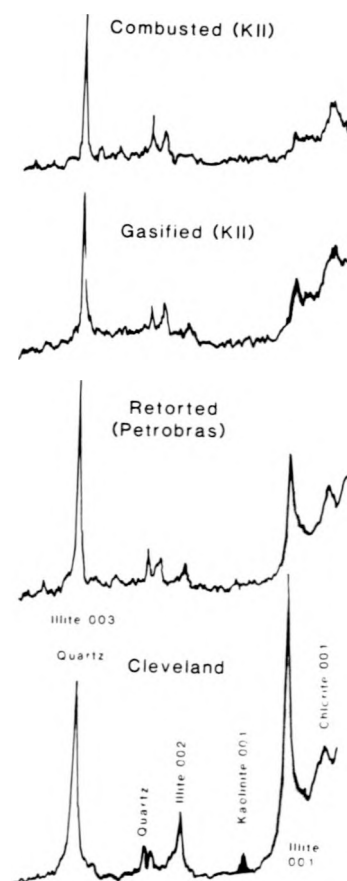


Figure 9-1. X-Ray Diffraction Patterns from KENTORT II Materials.

9.2.2 Fixed Bed Pyrolysis. Samples (140 g) of the Cleveland shale were retorted in a fixed bed apparatus.(5) The shale was heated at a rate of 4 °C/minute to the maximum or "soak" temperature which was held for thirty minutes. A nitrogen sweep, at a rate of 22 cm³/minute, as measured at room temperature and pressure, was used throughout the runs. Sweep gas flow rates at maximum temperature ranged from 50 to 84 cm³/min. The soak temperature was incremented for each run by approximately 50 °C, spanning the temperature range from 400 to 850 °C. The soak temperature was briefly exceeded during temperature ramping by up to 14 °C, and the final soak temperature varied from the target, by 2 to 7°C.

Table 9-2. *Elemental Composition of Sulfuric Acid Extracts in g/Kg of Shale for KENTORT II Materials.*

	Raw	Retorted	Gasified	Combus.
Fe	4.92	26.37	23.31	9.51
Al	5.76	11.10	13.68	20.30
K	2.20	5.67	8.47	8.99
Mg	0.64	2.84	1.66	0.40
Na	0.10	0.67	0.57	0.68
Zn	0.33	0.52	0.48	0.33
V	0.07	0.11	0.22	0.52
Se	0.09	0.14	0.11	0.11
Cu	0.09	0.09	0.11	0.08
Ni	0.09	0.11	0.07	0.08
Mn	0.03	0.20	0.06	0.04

9.2.3 Extractions. Samples of the KENTORT II and fixed bed materials were extracted with H_2SO_4 . One gram of sample was mixed with 100 ml of 6N H_2SO_4 . The samples were agitated periodically and left to react for approximately 24 hours at room temperature. The extracts were diluted 1:1 with distilled-deionized water and analyzed with an inductively couple plasma spectrometer (ICP).

9.3 RESULTS

9.3.1 Mineralogy of the Oil Shale Samples. Eastern oil shale comprises a small component of the Devonian and Mississippian Black Shales of the Eastern U.S. which comprises a large and extensive body of sedimentary rocks. Discussion of the regional variation of the mineralogy of these rocks may be found elsewhere. (6)

In general, the inorganic component of the oil shale samples consisted of clay minerals. The most abundant was illite, a 2:1 layer clay; i.e. two layers of tetrahedrally coordinated silica sandwiching an octahedrally coordinated alumina layer, with the interlayer consisting of potassium. In general the illite was found to be comparatively well crystallized and represented about 2/3 of the minerals (Table 9-1) in the study samples. Second in abundance was chlorite, a 2:1:1 clay, which comprised approximately 10% of the minerals. Kaolinite, a 1:1 clay, was always present, but in generally low abundances, i.e. 5% or less.

Mixed-layer clays were indicated in some of the samples. Vermiculite, a 2:1 clay which has a limited expansion characteristic, appeared to be the major interlayered mineral, with smectite, a 2:1 clay capable of undergoing extensive interlayer expansion, also present. Mixed-layered clays were, however, only a small component of the shale (i.e. <4%).

Quartz (SiO_2) was the most abundant non-clay mineral present (about 25%), and pyrite and marcasite (FeS_2) were ubiquitous, present at abundances of 2 to 5 percent.

Many other minerals have been found in the shale, and are either locally abundant or more generally distributed in low abundance. These constitute a long list and include such minerals as calcite, dolomite, galena, sphalerite, chalcopryrite, apatite, and several forms of feldspar.

9.3.2 Changes in Mineralogy Induced by Retorting. In the KENTORT II technology the pyrolyzer is operated at approximately 525 °C, the gasifier approximately 750 to 800 °C, and the combustor at approximately 850 °C. Thus materials from the various process steps have been incrementally heated and the final product also oxidized. Because the gasification and retort zone are integrated, a fraction which had only been pyrolyzed could not be collected. A Petrobras sample, which used the same feedstock as KENTORT II, and was pyrolyzed at approximately same temperature, was substituted.

X-ray diffraction patterns of samples from the KENTORT II technology (Figure 9-1) indicate illite to exhibit a pattern of decrystallization as a function of processing temperature as reflected in the attenuation of the 001 and 002 illite reflections at 10 Å and 5 Å (Figure 9-2). Almost complete illite decrystallization is indicated for the combusted sample. Kaolinite is decrystallized at lower temperatures, as indicated by the loss of the kaolinite reflection at 7 Å (001) and is essentially absent in the pyrolyzed sample. Chlorite appears to be relatively stable. X-ray diffraction patterns from the samples which were processed in the fixed bed retort show similar patterns, although some indication of kaolinite was found in the pyrolyzed sample from the 455 and 496 °C runs.

The X-ray diffraction patterns did not indicate the development of any new silicate mineral phases such as Richterite or Mullite. Neoformed minerals either are not present; or are present in abundances below that which could be measured by X-ray (about 2-5%); or are not sufficiently crystallized to give reflections discernable above background noise.

9.3.3 The Effect of Mineralogic Changes on the Acid Solubility of the Shales. The KENTORT II and fixed bed materials were extracted with sulfuric acid to determine if changes in

mineralogy could be related to changes in the acid solubility of the materials. The silicates originally present in the raw shale are only sparingly soluble in sulfuric acid. The change in their solubility provides some measure of the change in the chemistry of the minerals induced by retorting. In addition, because sulfuric acid is the most inexpensive of common industrial acids, the composition of the extract may have economic implications for processing of the spent shales after retorting.

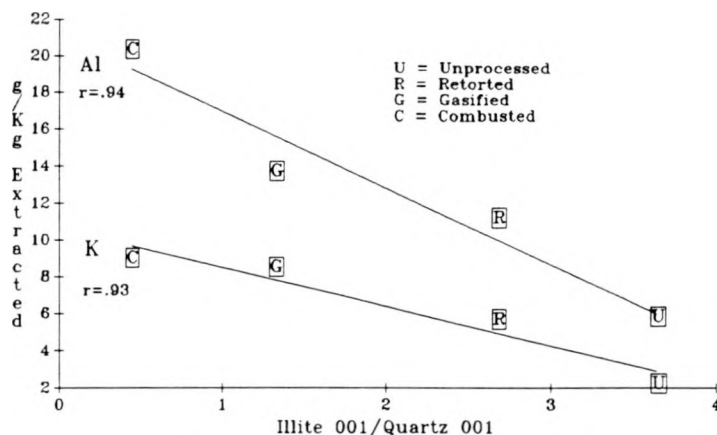


Figure 9-2. Al and K Extracted from KENTORT II Samples as a Function of Degree of Decrystallization of Illite.

The concentration of Al and K was highest in the combusted shale extracts (Table 9-2) and was relatable to the decrystallization of illite. Changes in the crystallinity of the illite were measured by ratioing the height of the illite 001 peak to that of the 001 quartz reflection at 4.3 Å (the much stronger 101 quartz reflection at 3.3 Å coincides with the 003 illite reflection). Quartz is stable at the retorting temperature studied, thus a decrease in this ratio is a reflection of the decrystallization of the illite. A strong correlation between the amount K and Al which could be extracted and the degree of illite decrystallization is found (Figure 9-2).

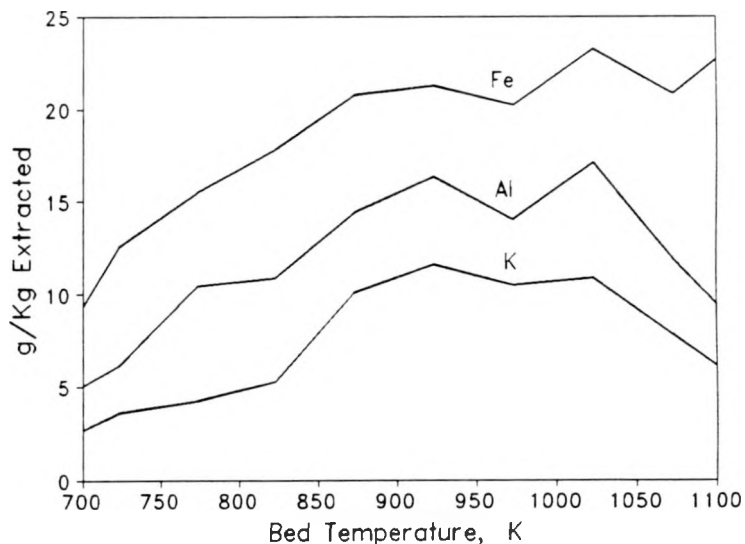


Figure 9-3. Fe, Al and K Extracted from the Fixed Bed Retorted Shale as a Function of Temperature.

Thus the solubility of these elements are a function of clay decrystallization, in turn a function of temperature, and do not appear to be sensitive to oxidation.

The extracts from the fixed bed materials showed a somewhat different pattern. In this case most of the elements, including K and Al, increased in concentration with temperature up to approximately 750 °C and then decreased (Figure 9-3). The exception being Fe which shows a maxima at the highest temperature.

As the fixed bed samples were pyrolyzed under reducing conditions the decline in the solubility K and Al, at the highest temperatures cannot be explained by oxidation. The explanation may lie in the examination of the particle size distribution. A comparison of the 849 °C sample with those from lower temperatures indicated a high proportion of the shale to be fractionated to the coarsest particle size, suggesting that the high temperature sample may have been partially melted. Thus the decrease in elemental solubility may be due to the development of a vitreous phase.

Iron does not follow this trend. Its solubility is most likely controlled by changes in the sulfide phase. Although the abundance of iron sulfides are too low in the Cleveland shale to be accurately measured by the X-ray technique used, the conversion of FeS_2 to $\text{Fe}_{(1-x)}\text{S}$ (pyrrhotite) could easily be calculated stoichiometrically. The acid solubility of Fe correlated well with the degree of pyrite conversion.

9.4 DISCUSSION

9.4.1 Effect of Mineral Changes on the Disposal Characteristics of the KENTORT II shale. One of the stated objectives of the KENTORT II design strategy is to produce a spent shale with improved disposal characteristics. Previous research has indicated that retorted Eastern oil shale is capable of producing highly acidic leachates with high sulfate concentrations. Sulfur is primarily present as pyrite and marcasite (FeS_2) in the raw unretorted shale. In conventional processing some of the pyrite is converted to pyrrhotite, Fe_{1-x}S , an iron deficient sulfide, where x can vary from 0 to about 0.2. Slightly more than two thirds of the pyrite was found converted to pyrrhotite in pilot plant tests of Eastern oil shale in the PETROSIX retort and Dravo's traveling grate retort (Table 9-3). The amount pyrite and pyrrhotite left in the shale has a strong impact on the acid generating capacity of spent shale. Pyrite oxidation may be written as:(5)

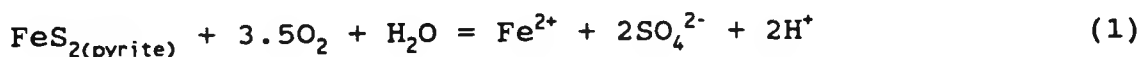
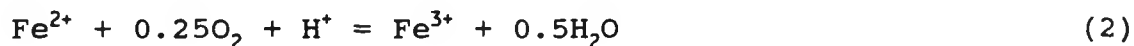
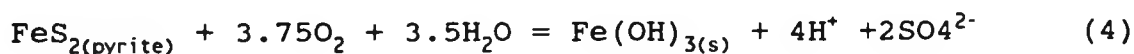


Table 9-3. *Composition of Raw and Spent Shale from the KENTORT II, Petrobras' PETROSIX and Dravo's Traveling Grate Retorts.*

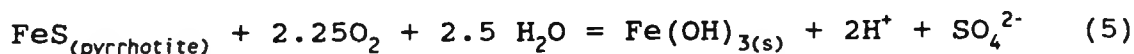
KENTORT II	%C	%H	%N	%S
Raw Feed Shale	16.7	2.0	0.7	1.7
Spent Shale	0.2	0.07	0.06	0.03
Dravo	%C	%H	%N	%S
Raw Feed Shale	11.4	1.38	0.8	3.35
Spent Shale	6.7	0.20	0.3	2.38
PETROBRAS	%C	%H	%N	%S
Raw Feed Shale	13.1	1.63	0.6	2.10
Spent Shale	7.9	0.48	0.45	1.42



Summing the above,



thus, for every mole of pyrite which is weathered, four moles of protons are generated. Pyrrhotite is not quite as strong a proton generator as pyrite,



producing slightly more^a than two protons per mole. Retorting should reduce the acid generating capacity of the shale. This has been observed in comparative leaching studies of raw and retorted shale from the Dravo pilot plant test.(6)

The combusted shale from KENTORT II is very low in sulfur indicating that the pyrite and pyrrhotite from the spent shale has been oxidized in the gasification and combustion processes. Mössbauer spectroscopy indicated that most of the iron, which is

^aDue to the hydrolysis of excess (non-stoichiometric) sulfur.

in the ferrous (Fe^{2+}) state in the raw and pyrolyzed shale, was converted to the ferric state (Fe^{3+}) in the combusted shale (Table 5-12).(7) Thus from these considerations we expect that the final spent shale from KENTORT II to have little or no acid generating capacity. Shale leachates with pH close to neutral will also result in much lower concentrations of dissolved metals.

The increased solubility of K and Al in the combusted shale compared to the retorted shale (Figure 9-4) also may have economic implications, which will be explored in further research on the spent shale from the KENTORT II technology.

9.4.2 Summary. The major minerals present in the oil shales of Kentucky include: illite, chlorite, kaolinite, quartz, and iron sulfides. The principal effects of retorting on these minerals was the decrystallization of kaolinite, the partial decrystallization of illite, and the transformation of pyrite to pyrrhotite. The decrystallization of illite appeared to be a function of processing temperature and was directly correlatable to the acid solubility of Al and K. The solubility of Fe was proportional to the degree of pyrite conversion to pyrrhotite. Fe solubility was reduced by the conversion of the iron sulfides to oxides.

Extracts for retorted and gasified KENTORT II materials were higher in Fe, compared to extracts from the shale after combustion. This was due to oxidation of the iron to the ferrous state, which are less soluble in acid. The mineralogic changes induced by processing in the KENTORT II, particularly the conversion of sulfides and oxidation of iron, will result in a spent shale which has lowered acid generating capacity and metals leaching compared to technologies which do not have a combustion step.

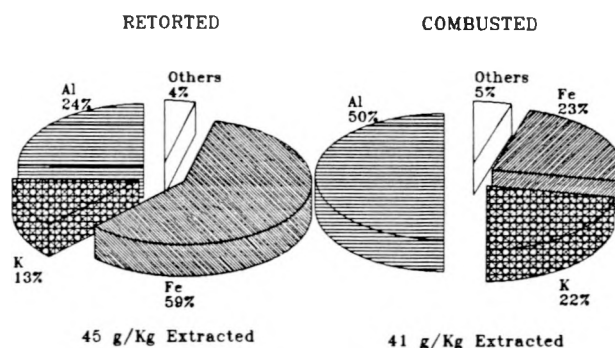


Figure 9-4. Graph of Relative Proportions of Major Elements Extracted from the Retorted and Combusted Shale Fractions of the KENTORT II Materials.

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10. CONCLUSIONS AND RECOMMENDATIONS

The operation of each zone of the KENTORT II process was demonstrated both separately in the bench-scale unit and as an integrated system in the 5-lb/hr mini-plant. There were no technological barriers encountered that would hinder operation at a larger scale. Autothermal operation was not reached because the solid recycle loop between the gasifier and combustor was slightly undersized. However, remarkably realistic conditions were achieved in the mini-plant for such a small system, and it is believed that operation of the system at a larger scale will actually be easier. The solid transfer lines were so small that literally fractions of an inch made major differences in the solid transfer rates that were achieved. From a physical standpoint, it is concluded that the operation of the KENTORT II concept was demonstrated in this study.

However, the demonstration of operability and the question of viability are entirely different issues. From the onset of the program, there were two fundamental aspects of the KENTORT II concept that required experimental confirmation before the process could be considered for further development. First, demonstration of substantial sulfur removal under mild gasification conditions was necessary. The rate of sulfur removal was significantly greater than carbon gasification and resulted in approximately 80% removal under conditions as mild as 750°C and less than 20 minutes holding time. During combustion runs in the mini-plant, no more than 8.3% of the original sulfur was converted to SO₂. Additionally, syngas production under these conditions was significant and could play an important role in the economics of the process. For example, under most conditions, sufficient H₂ was produced to completely upgrade the crude shale oil. From these results, it is concluded that the gasification zone is beneficial, and it should remain as an integral part of the KENTORT process.

The second area of concern was that heat transfer to the pyrolysis zone by solid recirculation would seriously reduce oil yields. Under the most realistic large-scale conditions that could be achieved in the mini-plant, oil yields averaged 111% of modified Fischer assay. While these yields are reduced compared to the original expectations for fluidized bed pyrolysis of Eastern shale, it is nevertheless significant that oil yields remained above Fischer assay under these conditions. Furthermore, as with most emerging technologies, additional reactor modifications may be reasonably anticipated to improve on the oil yields obtained in this study. These results indicate that solid-recycle pyrolysis is a technique that deserves continued development. Once more understanding is gained regarding the mechanisms of oil losses in solid-recycle systems, greater oil yields should be attainable.

This program showed the KENTORT II process to be viable in its original configuration, but there are a few modifications and refinements that should be incorporated into an updated design. The modified KENTORT II configuration is shown in Figure 10-1. The original configuration specified the utilization of raw shale fines as a fuel for the combustor, but the modified process feeds all the raw shale (1/4" minus) to the pyrolyzer. Pyrolysis of fines occurs quickly, so some oil and gas is produced even though the residence time in the reactor is short. The entrained fines are then separated from the oil and gas stream and fed directly to the combustor.

The commercial reactor configuration specified a cyclonic combustor located in the freeboard section of the gasification zone. This design was thermally efficient and made the reactor very compact and modular. However, for operating simplicity, an external fluidized bed combustor was utilized for experimentation in the bench-scale reactor and the mini-plant systems. In a

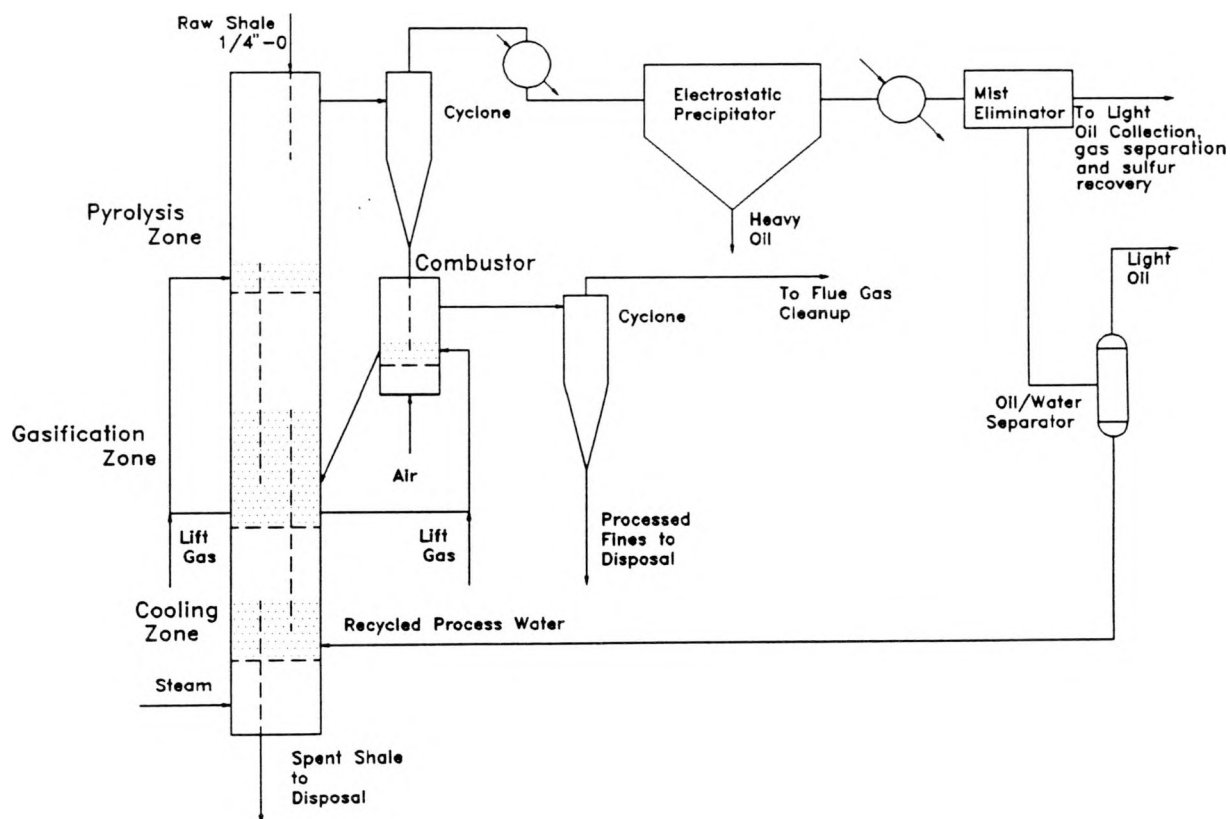


Figure 10-1. Modified KENTORT II Process Schematic.

similar configuration, an external fluidized bed combustor was included in the modified KENTORT II design. While some thermal efficiency may be lost through the use of an adjacent fluidized bed combustor, the technology is well known and will not pose unnecessary technical risks.

Results from mini-plant operation indicated that temperature staging is an important consideration for the gasification zone. This feature was included in the original KENTORT II design, and it is an element that should remain. The data is inconclusive at this point, but it appears to be less detrimental to oil yield to utilize a high rate of low temperature solids as the heat transfer combination to the pyrolysis zone. Through staging, the temperature of the recycled gasifier shale may be optimized for high oil yield, while maintaining high levels of carbon and sulfur conversion.

This study was successful in experimentally demonstrating the KENTORT II process at a miniature scale, and it is recommended that development of the process continue. Because of the geometry of the KENTORT II design, a commercially realistic heat transfer situation could not be achieved in the 5-lb/hr system. Therefore, further testing at a larger scale is recommended as the next necessary step in the development of the KENTORT process.

APPENDIX A. Average Run Conditions and Heat Balance
Calculations for the Mini-Plant Run Series

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	1 IN. LEFT PYROLYZER ,C	1 IN. RIGHT PYROLYZER,C	2 IN. LEFT PYROLYZER,C	2 IN. RIGHT PYROLYZER,C	3 IN. LEFT PYROLYZER,C	4 IN. RIGHT PYROLYZER,C
KEN#2 3/1/88	538	521	534	518	532	517
KEN#3 3/8/88	548	532	544	528	543	529
KEN#4 3/10/88	540	521	535	516	534	519
KEN#5 3/16/88	555	536	551	534	548	533
KEN#6 3/22/88	539	521	534	517	531	517
KEN#7 3/28/88	545	527	542	524	540	526
KEN#8 4/6/88	547	520	543	521	542	523
KEN#9 4/14/88	550	525	546	524	547	528
KEN#10 4/19/88	557	532	552	530	552	530
KEN#11 4/25/88	554	529	550	528	550	531
KEN#14 5/19/88	550	507	545	507	549	513
KEN#15 5/24/88	508	510	508	511	507	508
KEN#16 8/31/88	526	523	520	527	523	521
KEN#17 9/8/88	531	523	523	529	528	523
KEN#18 9/15/88	533	523	523	528	530	523
KEN#19 9/19/88	557	518	521	518	540	520
KEN#20 9/22/88	525	518	513	522	521	518
KEN#21 9/26/88	520	523	520	533	520	522
KEN#22 9/28/88	536	527	532	530	532	521
KEN#23 10/3/88	527	524	525	530	526	523
KEN#24 10/5/88	529	524	525	532	527	520
KEN#25 10/10/88	531	522	528	523	529	520
KEN#26 10/13/88	552	546	545	542	555	543
KEN#27 10/17/88	541	513	538	527	528	511
KEN#28 10/19/88	550	518	547	540	540	521
KEN#29 10/24/88	558	516	554	538	549	525
KEN#30 11/9/88	544	518	541	531	539	514
KEN#31 12/5/88	542	517	536	537	531	520
KEN#32 12/7/88	528	508	525	529	517	508
KEN#33 12/12/88	529	509	527	532	519	509
KEN#34 12/22/88	532	511	529	531	522	513
KEN#35 1/4/89	541	505	536	536	527	511
KEN#36 1/9/89	535	512	531	533	524	514

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	1 IN. LEFT GASIFIER, C	1 IN. RIGHT GASIFIER, C	9 IN. LEFT GASIFIER, C	9 IN. RIGHT GASIFIER, C	L-PIPE DP, IN. H2O	GASIFIER DP, IN. H2O
KEN#2 3/1/88	832	835	908	834	NA	3.71
KEN#3 3/8/88	832	835	785	834	NA	3.37
KEN#4 3/10/88	799	802	720	800	NA	3.06
KEN#5 3/16/88	837	840	NA	838	NA	3.22
KEN#6 3/22/88	837	839	NA	839	NA	3.42
KEN#7 3/28/88	800	802	NA	802	NA	3.21
KEN#8 4/6/88	800	803	NA	801	NA	4.77
KEN#9 4/14/88	828	831	829	826	NA	3.56
KEN#10 4/19/88	829	832	830	717	NA	4.23
KEN#11 4/25/88	803	805	804	660	NA	4.38
KEN#12 5/19/88	797	809	806	NA	NA	7.00
KEN#13 5/24/88	747	755	742	NA	1.10	6.17
KEN#14 8/31/88	727	735	723	647	1.36	5.24
KEN#15 9/8/88	NA	761	749	650	1.21	5.65
KEN#16 9/15/88	NA	756	740	643	0.97	6.03
KEN#17 9/19/88	NA	771	769	659	0.42	4.67
KEN#18 9/22/88	NA	747	734	NA	1.15	5.40
KEN#19 9/26/88	NA	824	806	730	1.78	4.94
KEN#20 9/28/88	NA	838	825	NA	0.96	5.58
KEN#21 10/3/88	NA	747	723	732	2.07	5.70
KEN#22 10/5/88	NA	812	794	802	1.23	5.74
KEN#23 10/10/88	NA	780	770	773	0.93	5.56
KEN#24 10/13/88	NA	777	760	769	0.35	4.53
KEN#25 10/17/88	NA	771	753	763	1.07	6.26
KEN#26 10/19/88	NA	783	761	777	1.42	6.13
KEN#27 10/24/88	NA	796	777	785	1.26	6.41
KEN#28 11/9/88	NA	754	737	745	1.44	6.47
KEN#29 12/5/88	NA	827	812	816	1.26	6.28
KEN#30 12/7/88	NA	787	765	773	1.39	6.66
KEN#31 12/12/88	NA	760	733	747	1.53	6.95
KEN#32 12/22/88	NA	763	742	752	1.39	6.95
KEN#33 1/4/89	NA	839	818	824	1.46	6.97
KEN#34 1/9/89	NA	795	775	782	1.43	6.88

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	PYROLYSIS DP, IN. H2O	N2 FLOW, G/MIN.	SCREWFEEPER RPM	SHALE FEED- RATE, G/MIN	STEAM FLOW, G/MIN.	RECYCLE CHAR TEMP, C	PYROLYZER TEMP., K
KEN#2 3/1/88	0.68	4.68	62	23.04	NA		807
KEN#3 3/8/88	0.63	4.68	61	19.39	NA		817
KEN#4 3/10/88	0.27	4.68	54	19.72	41.58		808
KEN#5 3/16/88	0.77	5.04	54	20.64	40.07		824
KEN#6 3/22/88	0.64	5.04	52	15.92	34.02		807
KEN#7 3/28/88	0.47	5.78	51	17.82	40.82		815
KEN#8 4/6/88	0.70	6.49	78	34.84	40.07		816
KEN#9 4/14/88	1.14	5.78	76	33.38	42.34		819
KEN#10 4/19/88	0.78	4.97	79	30.50	39.31		825
KEN#11 4/25/88	0.49	5.11	78	30.00	40.07		823
KEN#14 5/19/88	NA	5.14	130	46.29	25.89	778	818
KEN#15 5/24/88	NA	6.35	126	45.76	28.34	725	781
KEN#16 8/31/88	1.86	5.14	80	32.14	30.36	689	793
KEN#17 9/8/88	1.62	5.78	80	29.72	27.64	709	796
KEN#18 9/15/88	1.76	6.13	80	31.42	25.28	706	796
KEN#19 9/19/88	1.07	5.78	80	35.30	37.55	NA	794
KEN#20 9/22/88	3.43	6.03	80	29.46	28.69	689	786
KEN#21 9/26/88	5.22	5.96	80	31.52	27.80	762	793
KEN#22 9/28/88	2.47	4.83	80	30.92	25.13	761	805
KEN#23 10/3/88	3.86	6.17	80	29.42	27.70	675	798
KEN#24 10/5/88	3.99	6.31	80	30.30	23.73	734	798
KEN#25 10/10/88	3.91	6.56	100	38.61	34.91	709	801
KEN#26 10/13/88	5.79	6.10	100	37.39	30.75	672	818
KEN#27 10/17/88	0.00	6.84	100	39.88	34.66	673	811
KEN#28 10/19/88	0.00	7.02	120	51.71	34.27	689	820
KEN#29 10/24/88	0.00	6.63	120	46.73	33.35	701	827
KEN#30 11/9/88	0.00	6.42	120	48.54	36.30	678	814
KEN#31 12/5/88	0.00	6.45	120	48.30	31.21	727	809
KEN#32 12/7/88	0.00	6.63	120	48.28	29.22	705	798
KEN#33 12/12/88	0.00	6.84	120	47.25	29.54	687	800
KEN#34 12/22/88	0.00	6.56	120	47.49	30.38	687	802
KEN#35 1/4/89	0.00	6.45	120	47.28	27.54	745	809
KEN#36 1/9/89	0.00	6.74	120	46.18	29.66	710	804

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	H_MIN, J/G MIN	H KER, J/G KER	H CHAR, J/G CHAR	H H2O, J/G H2O	H TOTAL, J/G RS	H TOTAL, J/MIN	GASIFIER TEMP, K	H STEAM, J/G STEAM
KEN#2 3/1/88	498.6	1126	907.1	3463	840.5	19365	1108	671.7
KEN#3 3/8/88	509.9	1155	930.4	3542	858.1	16638	1108	650.3
KEN#4 3/10/88	499.7	1129	909.4	3471	842.2	16609	1075	593.0
KEN#5 3/16/88	517.8	1175	946.9	3598	870.4	17965	1113	646.9
KEN#6 3/22/88	498.6	1126	907.1	3463	840.5	13381	1112	681.0
KEN#7 3/28/88	507.6	1149	925.8	3526	854.5	15228	1075	578.0
KEN#8 4/6/88	508.7	1152	928.1	3534	856.3	29833	1076	578.2
KEN#9 4/14/88	512.1	1161	935.1	3558	861.6	28760	1104	636.7
KEN#10 4/19/88	518.9	1178	949.3	3606	872.2	26601	1105	626.1
KEN#11 4/25/88	516.6	1172	944.6	3590	868.6	26059	1078	567.8
KEN#14 5/19/88	511.0	1158	932.8	3550	859.8	39801	1082	587.8
KEN#15 5/24/88	469.5	1051	847.0	3259	795.3	36391	1028	543.0
KEN#16 8/31/88	482.9	1085	874.6	3353	816.0	26228	1008	472.0
KEN#17 9/8/88	486.3	1094	881.5	3376	821.3	24408	1034	524.7
KEN#18 9/15/88	486.3	1094	881.5	3376	821.3	25804	1029	513.3
KEN#19 9/19/88	484.0	1088	876.9	3361	817.8	28868	1044	551.8
KEN#20 9/22/88	475.1	1065	858.4	3298	803.9	23683	1020	514.1
KEN#21 9/26/88	482.9	1085	874.6	3353	816.0	25722	1097	676.0
KEN#22 9/28/88	496.4	1120	902.4	3447	837.0	25880	1111	683.0
KEN#23 10/3/88	488.5	1100	886.1	3392	824.8	24264	1020	488.5
KEN#24 10/5/88	488.5	1100	886.1	3392	824.8	24990	1085	637.5
KEN#25 10/10/88	491.9	1108	893.1	3415	830.0	32046	1053	557.4
KEN#26 10/13/88	511.0	1158	932.8	3550	859.8	32149	1050	514.1
KEN#27 10/17/88	503.1	1137	916.4	3495	847.5	33799	1044	515.4
KEN#28 10/19/88	513.2	1164	937.5	3566	863.3	44644	1056	523.6
KEN#29 10/24/88	521.1	1184	954.0	3623	875.7	40922	1069	538.4
KEN#30 11/9/88	506.5	1146	923.4	3518	852.8	41394	1027	470.2
KEN#31 12/5/88	500.9	1132	911.7	3479	844.0	40765	1100	648.8
KEN#32 12/7/88	488.5	1100	886.1	3392	824.8	39819	1060	579.9
KEN#33 12/12/88	490.7	1106	890.8	3408	828.2	39135	1033	513.8
KEN#34 12/22/88	493.0	1111	895.4	3423	831.7	39499	1036	516.4
KEN#35 1/4/89	500.9	1132	911.7	3479	844.0	39904	1112	676.8
KEN#36 1/9/89	495.2	1117	900.1	3439	835.2	38571	1068	585.5

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	H STEAM, J/MIN	H NIT, J/G N2	H NIT, J/MINUTE	H GAS, J/MINUTE	H DIFF, J/MINUTE	H RECIRC, J/G @TGAS	PYRO. RECIRC. RATE G/MIN @ GASIF. TEMP
KEN#2 3/1/88	NA					344.6	
KEN#3 3/8/88	NA					333.3	
KEN#4 3/10/88	24658					305.8	
KEN#5 3/16/88	25920					331.1	
KEN#6 3/22/88	23169					349.1	
KEN#7 3/28/88	23597					297.9	
KEN#8 4/6/88	23167					297.9	
KEN#9 4/14/88	26953					326.5	
KEN#10 4/19/88	24613					320.8	
KEN#11 4/25/88	22749					292.3	
KEN#14 5/19/88	15219	258.8	1331	16550	23251	302.5	76.87
KEN#15 5/24/88	15389	239.6	1520	16909	19481	282.1	69.05
KEN#16 8/31/88	14331	186.3	958	15289	10939	245.7	44.52
KEN#17 9/8/88	14501	205.3	1187	15688	8720	272.2	32.03
KEN#18 9/15/88	12976	202.0	1239	14215	11590	266.5	43.49
KEN#19 9/19/88	20719					286.0	
KEN#20 9/22/88	14751	193.9	1169	15920	7763	267.3	29.04
KEN#21 9/26/88	18792	268.1	1597	20389	5333	347.8	15.33
KEN#22 9/28/88	17163	253.8	1225	18388	7492	350.2	21.39
KEN#23 10/3/88	13532	165.2	1019	14552	9713	253.9	38.26
KEN#24 10/5/88	15129	231.1	1459	16588	8402	328.4	25.58
KEN#25 10/10/88	19460	199.9	1311	20771	11275	288.4	39.09
KEN#26 10/13/88	15811	140.1	854	16665	15484	265.8	58.25
KEN#27 10/17/88	17865	148.8	1018	18883	14916	266.8	55.90
KEN#28 10/19/88	17942	156.8	1100	19043	25601	270.5	94.66
KEN#29 10/24/88	17958	162.5	1077	19036	21887	277.4	78.89
KEN#30 11/9/88	17070	151.1	970	18039	23355	243.9	95.75
KEN#31 12/5/88	20247	211.3	1363	21611	19154	333.2	57.49
KEN#32 12/7/88	16944	198.7	1317	18261	21558	299.8	71.90
KEN#33 12/12/88	15177	176.4	1207	16384	22751	266.6	85.34
KEN#34 12/22/88	15690	174.2	1143	16833	22667	267.8	84.64
KEN#35 1/4/89	18639	231.5	1494	20132	19772	346.9	57.00
KEN#36 1/9/89	17363	197.7	1332	18695	19876	302.2	65.76

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	SOLID RECYCLE TEMP, K	H RECIRC, J/G @TRECIRC	PYRO. RECYIRC. RATE G/MIN @ TRECIRC	RECIRC RATIO @ GASIF. TEMP.	RECIRC RATE @ TRECIRC
KEN#2 3/1/88					
KEN#3 3/8/88					
KEN#4 3/10/88					
KEN#5 3/16/88					
KEN#6 3/22/88					
KEN#7 3/28/88					
KEN#8 4/6/88					
KEN#9 4/14/88					
KEN#10 4/19/88					
KEN#11 4/25/88					
KEN#14 5/19/88	1051	267.0	87.09	1.66	1.88
KEN#15 5/24/88	998	247.6	78.67	1.51	1.72
KEN#16 8/31/88	962	192.7	56.75	1.39	1.77
KEN#17 9/8/88	982	212.4	41.06	1.08	1.38
KEN#18 9/15/88	979	208.9	55.47	1.38	1.77
KEN#19 9/19/88					
KEN#20 9/22/88	962	200.6	38.70	0.99	1.31
KEN#21 9/26/88	1035	276.7	19.27	0.49	0.61
KEN#22 9/28/88	1034	262.1	28.58	0.69	0.92
KEN#23 10/3/88	948	171.0	56.79	1.30	1.93
KEN#24 10/5/88	1007	238.9	35.17	0.84	1.16
KEN#25 10/10/88	982	206.8	54.53	1.01	1.41
KEN#26 10/13/88	945	145.0	106.76	1.56	2.86
KEN#27 10/17/88	946	154.1	96.80	1.40	2.43
KEN#28 10/19/88	962	162.3	157.70	1.83	3.05
KEN#29 10/24/88	974	168.2	130.10	1.69	2.78
KEN#30 11/9/88	951	156.4	149.28	1.97	3.08
KEN#31 12/5/88	1000	218.5	87.66	1.19	1.81
KEN#32 12/7/88	978	205.5	104.89	1.49	2.17
KEN#33 12/12/88	960	182.6	124.61	1.81	2.64
KEN#34 12/22/88	960	180.3	125.70	1.78	2.65
KEN#35 1/4/89	1018	239.2	82.65	1.21	1.75
KEN#36 1/9/89	983	204.6	97.17	1.42	2.10

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION

RUN # - DATE	AVERAGE SOLID RECIRC RATIO TO PYROLYZER	AVG. REC. TEMP., C	CONSTANTS ENTHALPY	FOR SHALE CALCULATIONS

KEN#2 3/1/88			F_MIN	0.7600
KEN#3 3/8/88			F_KER	0.1900
KEN#4 3/10/88			F_H2O	0.0500
KEN#5 3/16/88			F_PYRITE	0.0337
KEN#6 3/22/88				
KEN#7 3/28/88				
KEN#8 4/6/88			X_CHAR	0.4500
KEN#9 4/14/88			X_PYRITE	0.5000
KEN#10 4/19/88				
KEN#11 4/25/88				
KEN#12 5/19/88	1.77	794	H_PYRITE	740.0000
KEN#13 5/24/88	1.61	740	H_PYROLYSIS	425.0000
KEN#16 8/31/88	1.58	712		
KEN#17 9/8/88	1.23	735		
KEN#18 9/15/88	1.57	731		
KEN#19 9/19/88				
KEN#20 9/22/88	1.15	718		
KEN#21 9/26/88	0.55	793		
KEN#22 9/28/88	0.81	800		
KEN#23 10/3/88	1.62	711		
KEN#24 10/5/88	1.00	773		
KEN#25 10/10/88	1.21	745		
KEN#26 10/13/88	2.21	725		
KEN#27 10/17/88	1.91	722		
KEN#28 10/19/88	2.44	736		
KEN#29 10/24/88	2.24	749		
KEN#30 11/9/88	2.52	716		
KEN#31 12/5/88	1.50	777		
KEN#32 12/7/88	1.83	746		
KEN#33 12/12/88	2.22	724		
KEN#34 12/22/88	2.21	725		
KEN#35 1/4/89	1.48	792		
KEN#36 1/9/89	1.76	753		

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	1 IN. LEFT PYROLYZER ,C	1 IN. RIGHT PYROLYZER,C	2 IN. LEFT PYROLYZER,C	2 IN. RIGHT PYROLYZER,C	3 IN. LEFT PYROLYZER,C	4 IN. RIGHT PYROLYZER,C
KEN#46 9/13/89	526	526	537	537	507	532
KEN#47 9/15/89	531	531	544	537	533	538
KEN#48 9/18/89	528	528	539	534	526	533
KEN#49 9/20/89	528	528	539	534	508	534
KEN#50 9/27/89	528	528	540	533	544	534
KEN#51 10/2/89	531	531	541	535	543	535
KEN#52 10/9/89	530	530	539	533	542	533
KEN#53 10/11/89	527	527	537	525	538	531
KEN#54 10/23/89	524	524	536	529	539	528
KEN#55 10/25/89	526	526	537	531	539	530
KEN#56 11/1/89	522	522	534	528	537	527
KEN#57 11/6/89	524	524	537	530	540	530
KEN#58 11/20/89	514	514	531	NA	535	526
KEN#59 11/29/89	519	519	533	NA	538	526
KEN#60 12/6/89	529	529	541	NA	545	534
KEN#61 12/6/89	520	520	533	NA	538	527
KEN#62 12/14/89	523	523	535	NA	539	529
KEN#63 12/18/89	520	520	534	NA	539	527
KEN#64 12/20/89	523	523	537	NA	542	531

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	1 IN. LEFT GASIFIER,C	1 IN. RIGHT GASIFIER,C	9 IN. LEFT GASIFIER,C	9 IN. RIGHT GASIFIER,C	L-PIPE DP, IN. H2O	GASIFIER DP, IN. H2O
KEN#46 9/13/89	771	774	773	777	1.08	6.51
KEN#47 9/15/89	785	788	787	791	1.02	5.79
KEN#48 9/18/89	752	754	754	758	1.16	5.62
KEN#49 9/20/89	758	761	761	765	1.13	6.19
KEN#50 9/27/89	769	773	772	776	1.45	5.27
KEN#51 10/2/89	824	829	827	834	1.11	5.90
KEN#52 10/9/89	802	806	805	810	1.15	5.76
KEN#53 10/11/89	820	823	824	828	0.95	5.63
KEN#54 10/23/89	782	784	784	789	1.44	5.90
KEN#55 10/25/89	812	813	815	820	0.98	5.66
KEN#56 11/1/89	756	757	759	764	1.46	5.68
KEN#57 11/6/89	812	813	815	819	1.19	6.04
KEN#58 11/20/89	812	811	815	818	0.43	5.82
KEN#59 11/29/89	786	786	789	795	0.98	5.59
KEN#60 12/6/89	837	837	841	846	0.83	5.45
KEN#61 12/6/89	808	808	812	818	0.59	5.43
KEN#62 12/14/89	811	810	815	821	0.88	5.58
KEN#63 12/18/89	762	762	765	773	1.33	5.78
KEN#64 12/20/89	811	811	816	824	1.07	5.54

APPENDIX A
(cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	PYROLYSIS DP, IN. H2O	N2 FLOW, G/MIN.	SCREWFEEEDER RPM	SHALE FEED- RATE, G/MIN	STEAM FLOW, G/MIN.	RECYCLE CHAR TEMP, C	PYROLYZER TEMP, K
KEN#46 9/13/89	1.24	7.82	120	39.80	29.44	647	810
KEN#47 9/15/89	1.19	7.82	120	40.86	27.58	658	817
KEN#48 9/18/89	1.28	7.29	120	40.00	28.36	699	812
KEN#49 9/20/89	0.99	7.47	120	41.14	31.10	698	812
KEN#50 9/27/89	1.25	7.82	120	41.76	31.57	691	813
KEN#51 10/2/89	1.18	7.82	120	44.10	28.96	728	814
KEN#52 10/9/89	1.22	9.24	120	39.64	28.42	708	812
KEN#53 10/11/89	1.11	7.47	120	38.21	28.19	711	810
KEN#54 10/23/89	1.22	7.29	120	40.86	28.04	694	809
KEN#55 10/25/89	1.18	7.65	120	39.89	23.60	709	810
KEN#56 11/1/89	1.28	6.76	120	40.29	31.53	676	807
KEN#57 11/12/89	1.09	7.12	120	40.01	28.48	713	810
KEN#58 11/20/89	1.14	8.00	120	40.06	29.95	710	804
KEN#59 11/29/89	1.29	8.18	120	38.92	28.37	696	806
KEN#60 12/6/89	1.22	9.42	120	39.20	27.37	731	814
KEN#61 12/6/89	0.92	8.71	120	38.76	26.92	711	806
KEN#62 12/14/89	1.44	8.00	120	37.97	29.37	710	808
KEN#63 12/18/89	1.27	7.12	120	39.70	28.81	684	807
KEN#64 12/20/89	0.73	8.18	120	39.11	27.73	719	810

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	H MIN, J/G MIN	H KER, J/G KER	H CHAR, J/G CHAR	H H2O, J/G H2O	H TOTAL, J/G RS	H TOTAL, J/MIN	GASIFIER TEMP, K	H STEAM, J/G STEAM
KEN#46 9/13/89	502.0	1134	914.1	3487	845.8	33661	1047	524.4
KEN#47 9/15/89	509.9	1155	930.4	3542	858.1	35060	1061	541.5
KEN#48 9/18/89	504.2	1140	918.7	3502	849.3	33970	1027	474.5
KEN#49 9/20/89	504.2	1140	918.7	3502	849.3	34939	1034	490.4
KEN#50 9/27/89	505.4	1143	921.1	3510	851.0	35539	1046	515.7
KEN#51 10/2/89	506.5	1146	923.4	3518	852.8	37607	1102	642.7
KEN#52 10/9/89	504.2	1140	918.7	3502	849.3	33665	1079	593.7
KEN#53 10/11/89	502.0	1134	914.1	3487	845.8	32316	1096	637.4
KEN#54 10/23/89	500.9	1132	911.7	3479	844.0	34486	1057	549.5
KEN#55 10/25/89	502.0	1134	914.1	3487	845.8	33737	1086	614.2
KEN#56 11/1/89	498.6	1126	907.1	3463	840.5	33863	1030	492.0
KEN#57 11/5/89	502.0	1134	914.1	3487	845.8	33838	1086	614.2
KEN#58 11/20/89	495.2	1117	900.1	3439	835.2	33460	1084	622.4
KEN#59 11/29/89	497.5	1123	904.7	3455	838.7	32644	1059	560.5
KEN#60 12/6/89	506.5	1146	923.4	3518	852.8	33429	1110	661.4
KEN#61 12/6/89	497.5	1123	904.7	3455	838.7	32509	1081	611.2
KEN#62 12/14/89	499.7	1129	909.4	3471	842.2	31980	1083	611.5
KEN#63 12/18/89	498.6	1126	907.1	3463	840.5	33367	1035	503.4
KEN#64 12/20/89	502.0	1134	914.1	3487	845.8	33077	1084	609.5

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	H STEAM, J/MIN	H NIT, J/G N2	H NIT, J/MINUTE	H GAS, J/MINUTE	H DIFF, J/MINUTE	H RECIRC, J/G @TGAS	PYRO. RECYCLE RATE G/MIN @GASIF. TEMP.
KEN#46 9/13/89	15438	121.0	947	16385	17276	271.4	63.65
KEN#47 9/15/89	14934	125.6	983	15916	19144	279.6	68.47
KEN#48 9/18/89	13455	176.7	1289	14744	19226	246.2	78.10
KEN#49 9/20/89	15253	175.6	1312	16565	18374	254.2	72.27
KEN#50 9/27/89	16280	166.7	1304	17585	17954	266.9	67.27
KEN#51 10/2/89	18615	206.9	1619	20234	17373	329.8	52.67
KEN#52 10/9/89	16871	186.8	1726	18597	15068	305.8	49.27
KEN#53 10/11/89	17968	192.3	1437	19405	12912	327.5	39.42
KEN#54 10/23/89	15407	174.4	1272	16679	17807	284.0	62.70
KEN#55 10/25/89	14494	190.1	1454	15948	17789	316.1	56.28
KEN#56 11/1/89	15513	156.5	1059	16572	17291	255.3	67.74
KEN#57 11/5/89	17491	194.5	1385	18876	14962	316.1	47.34
KEN#58 11/20/89	18640	197.7	1582	20223	13237	320.6	41.29
KEN#59 11/29/89	15901	179.9	1471	17372	15272	289.7	52.72
KEN#60 12/6/89	18102	210.3	1980	20082	13347	338.9	39.38
KEN#61 12/6/89	16453	196.7	1713	18165	14344	314.9	45.56
KEN#62 12/14/89	17960	193.4	1547	19507	12473	314.9	39.61
KEN#63 12/18/89	14503	165.4	1177	15681	17687	261.0	67.76
KEN#64 12/20/89	16902	201.2	1646	18548	14529	313.8	46.30

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	PYRO. SOLID RECYC. TEMP.,K (TRECYC.)	H RECIRC, J/G @TRECIRC	PYRO RECIRC RATE /MIN @TRECYC	PYRO RECYC RATIO GASIF. TEMP.	PYRO. RECYC RATIO @TRECYC
KEN#46 9/13/89	920	125.3	137.85	1.60	3.46
KEN#47 9/15/89	931	130.1	147.19	1.68	3.60
KEN#48 9/18/89	972	182.9	105.13	1.95	2.63
KEN#49 9/20/89	971	181.7	101.11	1.76	2.46
KEN#50 9/27/89	964	172.5	104.06	1.61	2.49
KEN#51 10/2/89	1001	214.0	81.18	1.19	1.84
KEN#52 10/9/89	981	193.2	77.97	1.24	1.97
KEN#53 10/11/89	984	198.9	64.90	1.03	1.70
KEN#54 10/23/89	967	180.5	98.65	1.53	2.41
KEN#55 10/25/89	982	196.6	90.46	1.41	2.27
KEN#56 11/1/89	949	162.0	106.71	1.68	2.65
KEN#57 11/15/89	986	201.3	74.35	1.18	1.86
KEN#58 11/20/89	983	204.6	64.71	1.03	1.62
KEN#59 11/29/89	969	186.2	82.02	1.35	2.11
KEN#60 12/6/89	1004	217.5	61.37	1.00	1.57
KEN#61 12/6/89	984	203.5	70.50	1.18	1.82
KEN#62 12/14/89	983	200.1	62.35	1.04	1.64
KEN#63 12/18/89	957	171.2	103.28	1.71	2.60
KEN#64 12/20/89	992	208.2	69.80	1.18	1.78

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	AVERAGE SOLID RECIRC RATIO TO PYROLYZER	AVG. REC. TEMP., C	1-inch Comb., K	2-inch Comb., K	Airflow SCFH	Comb. DP, in. H ₂ O	H _{min} C, J/g
KEN#46 9/13/89	2.53	711	1137	1131	11.73	0.76	102.7
KEN#47 9/15/89	2.64	723	1131	1116	9.16	0.70	79.8
KEN#48 9/18/89	2.29	727	1129	1114	10.99	0.81	116.6
KEN#49 9/20/89	2.11	730	1131	1121	11.36	0.71	110.8
KEN#50 9/27/89	2.05	732	1133	1122	10.26	0.83	99.3
KEN#51 10/2/89	1.52	779	1144	1127	7.33	0.67	47.7
KEN#52 10/9/89	1.60	757	1142	1136	9.53	0.79	71.7
KEN#53 10/11/89	1.37	767	1149	1133	8.06	0.68	60.1
KEN#54 10/23/89	1.97	739	1159	1147	10.63	0.67	116.0
KEN#55 10/25/89	1.84	761	1153	1130	8.43	0.62	76.1
KEN#56 11/1/89	2.16	717	1157	1145	11.73	0.80	144.8
KEN#57 11/5/89	1.52	763	1161	1150	9.16	0.74	85.1
KEN#58 11/20/89	1.32	761	1171	1160	8.79	0.80	98.6
KEN#59 11/29/89	1.73	741	1165	1153	9.89	0.76	120.5
KEN#60 12/6/89	1.29	784	1173	1149	5.86	0.70	71.2
KEN#61 12/6/89	1.50	760	1167	1147	8.06	0.75	97.5
KEN#62 12/14/89	1.34	760	1189	1175	9.89	0.77	119.9
KEN#63 12/18/89	2.15	723	1180	1159	11.73	0.69	164.9
KEN#64 12/20/89	1.48	765	1190	1165	8.43	0.70	119.9

APPENDIX A
(cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN # - DATE	H _{N2C} , J/g	H _{N2C} , J/minute	CO g/min	CO ₂ g/min	SO ₂ g/min	H produced, J/minute	Comb. Rec., g/min.	Comb. Rec. Ratio
KEN#46 9/13/89	102.1	1596	0.29	2.61	0.11	24997	227.9	5.73
KEN#47 9/15/89	79.4	1038	0.30	1.88	0.08	18369	217.1	5.31
KEN#48 9/18/89	115.5	1691	0.34	2.34	0.09	22687	180.1	4.50
KEN#49 9/20/89	109.9	1659	0.29	2.51	0.10	24056	202.1	4.91
KEN#50 9/27/89	98.6	1398	0.31	2.09	0.08	20286	190.2	4.56
KEN#51 10/2/89	47.8	538	0.30	1.40	0.05	13937	281.1	6.37
KEN#52 10/9/89	71.6	1013	0.29	2.06	0.09	19985	264.7	6.68
KEN#53 10/11/89	60.4	712	0.27	1.78	0.07	17310	275.9	7.22
KEN#54 10/23/89	115.9	1655	0.24	2.51	0.11	23905	191.8	4.69
KEN#55 10/25/89	76.3	934	0.25	1.93	0.08	18618	232.4	5.83
KEN#56 11/1/89	144.1	2177	0.25	3.04	0.12	28731	183.4	4.55
KEN#57 11/6/89	85.4	1086	0.30	2.20	0.09	21277	237.3	5.93
KEN#58 11/20/89	99.1	1268	0.32	2.38	0.08	22919	219.6	5.48
KEN#59 11/29/89	120.5	1686	0.29	2.72	0.10	25934	201.3	5.17
KEN#60 12/6/89	71.9	760	0.29	1.46	0.05	14434	192.1	4.90
KEN#61 12/6/89	98.0	1216	0.27	1.87	0.07	18114	173.2	4.47
KEN#62 12/14/89	120.9	1680	0.24	2.76	0.08	26002	202.9	5.34
KEN#63 12/18/89	164.8	2519	0.18	2.46	0.09	23129	125.0	3.15
KEN#64 12/20/89	120.9	1514	0.23	2.32	0.08	22027	171.1	4.38

APPENDIX A (cont.)

AVERAGE RUN CONDITIONS AND HEAT BALANCE
CALCULATIONS FOR THE KENTORTII RUN MATRIX
PYROLYSIS/GASIFICATION/COMBUSTION

RUN #	DATE	CONSTANTS ENTHALPY	FOR SHALE CALCULATIONS
KEN#46	9/13/89	F_MIN	0.7600
KEN#47	9/15/89	F_KER	0.1900
KEN#48	9/18/89	F_H2O	0.0500
KEN#49	9/20/89	F_PYRITE	0.0337
KEN#50	9/27/89		
KEN#51	10/2/89		
KEN#52	10/9/89	X_CHAR	0.4500
KEN#53	10/11/89	X_PYRITE	0.5000
KEN#54	10/23/89		
KEN#55	10/25/89		
KEN#56	11/1/89	H_PYRITE	740.0000
KEN#57	11/6/89	H_PYROLYSIS	425.0000
KEN#58	11/20/89		
KEN#59	11/29/89		
KEN#60	12/6/89		
KEN#61	12/6/89		
KEN#62	12/14/89		
KEN#63	12/18/89		
KEN#64	12/20/89		

APPENDIX A (cont.)

APPENDIX B. Overall Mass and Selected Component Balances
for the Mini-Plant Test Matrix

OVERALL AND COMPONENT BALANCES											
KEN#2	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4262.0	1.000	16.71	2.14	0.62	1.84	75.39	1.69	1.000	1.000	1.000
SPENT SHALE	3343.2	0.784	0.00	0.00	0.00	0.00	0.00	0.00	0.272	0.483	0.301
-BIN	3007.8	0.706	5.24	0.08	0.36	0.66	94.19	0.00	0.221	0.410	0.253
-TUBE	10.2	0.002	10.41	0.75	0.55	1.11	0.00	0.00	0.001	0.002	0.001
-CYCLONE	305.9	0.072	10.41	0.75	0.55	1.11	85.97	0.00	0.045	0.064	0.043
-A/C	19.3	0.005	16.38	1.50	1.06	1.23	78.22	0.00	0.004	0.077	0.003
OIL	383.6	0.090	0.00	0.00	0.00	0.00	0.00	0.00	0.415	0.142	0.080
-A/C	24.0	0.006	87.31	9.43	1.37	1.89	0.00	0.00	0.029	0.012	0.056
-ESP	256.7	0.060	72.51	7.83	1.14	1.57	6.75	3.99	0.261	0.111	0.051
-LP OIL	74.4	0.017	85.38	11.35	0.66	2.38	0.00	0.37	0.089	0.019	0.023
	0.0	0.000	85.38	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	28.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
GAS	444.5	0.104	0.00	0.00	0.00	0.00	0.00	0.00	0.195	0.000	0.810
-CH4	16.9	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-C2'S	9.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	7.6	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	6.6	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.078	0.000	0.000
-CO2	168.5	0.040	27.30	0.00	0.00	0.00	0.00	0.00	0.065	0.000	0.000
-CO	141.1	0.033	42.90	0.00	0.00	0.00	0.00	0.00	0.085	0.000	0.000
-H2	27.0	0.006	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	67.5	0.016	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.810
WATER	7286.7	1.710	0.00	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.147
-LIQUID	7286.7	1.709	0.17	0.00	0.00	0.16	0.00	100.00	0.017	0.000	0.147
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11458.0	2.688	0.00	0.00	0.00	0.00	0.00	0.00	0.899	0.625	1.337

OVERALL AND COMPONENT BALANCES											
KEN#3	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5350.6	1.000	16.80	2.32	0.75	1.80	75.38	1.16	1.000	1.000	1.000
SPENT SHALE	4243.9	0.793	0.00	0.00	0.00	0.00	0.00	0.00	0.268	0.508	0.119
-BIN	3780.7	0.707	5.00	0.10	0.43	0.16	94.59	0.56	0.210	0.405	0.063
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.51	0.000	0.000	0.000
-CYCLONE	426.2	0.080	10.64	0.92	0.88	1.17	86.09	0.60	0.050	0.093	0.052
-A/C	37.0	0.007	17.85	1.56	0.99	1.07	77.75	0.00	0.007	0.009	0.004
OIL	516.2	0.096	0.00	0.00	0.00	0.00	0.00	0.00	0.421	0.143	0.079
-A/C	23.1	0.004	85.97	10.35	1.78	1.90	0.00	0.00	0.022	0.010	0.005
-ESP	334.0	0.062	66.47	8.00	1.38	1.47	13.35	10.01	0.247	0.115	0.051
-LP OIL	114.1	0.021	85.72	11.44	0.63	1.99	0.00	0.03	0.109	0.018	0.024
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	45.0	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.044	0.000	0.000
GAS	480.4	0.090	0.00	0.00	0.00	0.00	0.00	0.00	0.163	0.000	0.834
-CH4	17.8	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	11.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	9.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	8.4	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	179.9	0.034	27.30	0.00	0.00	0.00	0.00	0.00	0.055	0.000	0.000
-CO	140.3	0.026	42.90	0.00	0.00	0.00	0.00	0.00	0.067	0.000	0.000
-H2	27.8	0.005	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	85.4	0.016	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.834
WATER	10181.8	1.903	0.00	0.00	0.00	0.00	0.00	0.00	0.031	0.000	0.015
-LIQUID	10181.8	1.903	0.27	0.00	0.00	0.01	0.00	100.00	0.031	0.000	0.015
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	15422.3	2.882	0.00	0.00	0.00	0.00	0.00	0.00	0.884	0.651	1.047

OVERALL AND COMPONENT BALANCES											
KEN#4	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	2504.1	1.000	16.94	2.34	0.74	1.84	75.71	1.69	1.000	1.000	1.000
SPENT SHALE	1911.2	0.763	0.00	0.00	0.00	0.00	0.00	0.00	0.283	0.464	0.114
-BIN	1710.0	0.683	5.84	0.12	0.43	0.18	93.76	0.00	0.235	0.397	0.067
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	179.5	0.072	9.36	1.04	0.56	1.10	87.00	0.00	0.040	0.054	0.043
-A/C	21.7	0.009	15.10	1.14	1.07	0.89	78.92	0.00	0.008	0.013	0.004
OIL	275.2	0.110	0.00	0.00	0.00	0.00	0.00	0.00	0.529	0.156	0.092
-A/C	41.6	0.017	87.33	9.37	1.46	1.84	0.00	0.00	0.086	0.033	0.017
-ESP	159.6	0.064	78.35	8.41	1.31	1.65	1.50	1.24	0.295	0.113	0.057
-LP OIL	43.9	0.018	83.51	11.38	0.45	1.88	0.00	0.50	0.086	0.011	0.018
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	30.1	0.012	87.00	13.00	0.00	0.00	0.00	0.00	0.062	0.000	0.000
GAS	156.6	0.063	0.00	0.00	0.00	0.00	0.00	0.00	0.122	0.000	0.502
-CH4	8.3	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	5.0	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	4.2	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	4.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	63.3	0.025	27.30	0.00	0.00	0.00	0.00	0.00	0.041	0.000	0.000
-CO	40.2	0.016	42.90	0.00	0.00	0.00	0.00	0.00	0.041	0.000	0.000
-H2	7.0	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	24.6	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.502
WATER	5683.6	2.270	0.00	0.00	0.00	0.00	0.00	0.00	0.029	0.000	0.010
-LIQUID	5683.6	2.270	0.22	0.00	0.00	0.01	0.00	100.00	0.029	0.000	0.010
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	8026.6	3.205	0.00	0.00	0.00	0.00	0.00	0.00	0.963	0.620	0.718

OVERALL AND COMPONENT BALANCES											
KEN#5	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	3962.7	1.000	16.69	2.25	0.75	1.75	75.61	1.69	1.000	1.000	1.000
SPENT SHALE	3051.4	0.770	0.00	0.00	0.00	0.00	0.00	0.00	0.287	0.461	0.135
-BIN	2728.6	0.689	4.96	0.13	0.40	0.16	94.65	0.00	0.205	0.367	0.063
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	250.1	0.063	11.19	1.21	0.64	1.52	85.67	0.00	0.042	0.054	0.055
-A/C	72.7	0.018	36.52	2.39	1.65	1.63	53.68	0.00	0.040	0.040	0.017
OIL	365.4	0.092	0.00	0.00	0.00	0.00	0.00	0.00	0.434	0.128	0.084
-A/C	65.3	0.016	86.61	9.96	1.55	1.88	0.00	0.00	0.086	0.034	0.018
-ESP	179.8	0.045	70.95	8.16	1.27	1.54	0.00	17.22	0.193	0.077	0.040
-LF OIL	91.0	0.023	85.03	11.80	0.57	2.02	0.00	0.20	0.117	0.017	0.027
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	29.3	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.039	0.000	0.000
GAS	345.2	0.087	0.00	0.00	0.00	0.00	0.00	0.00	0.167	0.000	0.635
-CH4	12.4	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	7.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	6.0	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	5.0	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.006	0.000	0.000
-CO2	137.2	0.035	27.30	0.00	0.00	0.00	0.00	0.00	0.057	0.000	0.000
-CO	112.9	0.028	42.90	0.00	0.00	0.00	0.00	0.00	0.073	0.000	0.000
-H2	17.6	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	46.8	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.635
WATER	6980.9	1.762	0.00	0.00	0.00	0.00	0.00	0.00	0.028	0.000	0.153
-LIQUID	6980.9	1.762	0.27	0.00	0.00	0.15	0.00	100.00	0.028	0.000	0.153
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	10742.9	2.711	0.00	0.00	0.00	0.00	0.00	0.00	0.916	0.590	1.007

OVERALL AND COMPONENT BALANCES											
KEN#6	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4488.4	1.000	16.81	2.32	0.78	1.77	75.05	1.69	1.000	1.000	1.000
SPENT SHALE	3554.9	0.792	0.00	0.00	0.00	0.00	0.00	0.00	0.238	0.468	0.112
-BIN	3176.8	0.708	4.31	0.14	0.43	0.14	95.35	0.00	0.181	0.390	0.056
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	302.4	0.067	10.12	1.04	0.62	1.25	74.78	0.00	0.041	0.054	0.048
-A/C	75.7	0.017	16.11	1.15	1.10	0.86	78.20	0.00	0.016	0.024	0.008
OIL	465.9	0.104	0.00	0.00	0.00	0.00	0.00	0.00	0.459	0.098	0.082
-A/C	94.2	0.021	88.04	9.01	1.13	1.82	0.00	0.00	0.110	0.030	0.022
-ESP	247.7	0.055	63.81	6.53	0.82	1.32	1.00	26.75	0.209	0.058	0.041
-LP OIL	76.0	0.017	83.67	11.56	0.45	2.05	0.00	0.34	0.084	0.010	0.020
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	48.0	0.011	87.00	13.00	0.00	0.00	0.00	0.00	0.055	0.000	0.000
GAS	477.6	0.106	0.00	0.00	0.00	0.00	0.00	0.00	0.205	0.000	0.661
-CH4	17.7	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-C2'S	9.6	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	8.0	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	6.5	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	201.1	0.045	27.30	0.00	0.00	0.00	0.00	0.00	0.073	0.000	0.000
-CO	155.6	0.035	42.90	0.00	0.00	0.00	0.00	0.00	0.088	0.000	0.000
-H2	23.3	0.005	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	55.8	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.661
WATER	10047.2	2.238	0.00	0.00	0.00	0.00	0.00	0.00	0.035	0.000	0.225
-LIQUID	10047.2	2.238	0.26	0.00	0.00	0.18	0.00	100.00	0.035	0.000	0.225
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	14545.6	3.241	0.00	0.00	0.00	0.00	0.00	0.00	0.937	0.566	1.080

OVERALL AND COMPONENT BALANCES											
KEN#7	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5916.1	1.000	16.93	2.27	0.78	1.76	75.28	1.69	1.000	1.000	1.000
SPENT SHALE	4688.9	0.793	0.00	0.00	0.00	0.00	0.00	0.00	0.297	0.541	0.111
-BIN	4342.4	0.734	5.95	0.18	0.52	0.20	93.26	0.00	0.258	0.489	0.083
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	276.3	0.047	10.10	1.00	0.58	0.82	86.27	0.00	0.028	0.035	0.022
-A/C	70.2	0.012	16.11	1.15	1.10	0.86	78.20	0.00	0.011	0.017	0.006
OIL	644.6	0.109	0.00	0.00	0.00	0.00	0.00	0.00	0.486	0.101	0.085
-A/C	119.2	0.020	87.79	9.48	1.07	1.67	0.00	0.00	0.104	0.028	0.019
-ESP	347.6	0.059	66.41	7.17	0.81	1.26	2.65	21.15	0.230	0.061	0.042
-LF OIL	119.7	0.020	84.26	11.36	0.47	2.07	0.00	0.40	0.101	0.012	0.024
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	58.1	0.010	87.00	13.00	0.00	0.00	0.00	0.00	0.050	0.000	0.000
GAS	410.9	0.069	0.00	0.00	0.00	0.00	0.00	0.00	0.134	0.000	0.590
-CH4	20.0	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	11.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	10.2	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	8.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	166.7	0.028	27.30	0.00	0.00	0.00	0.00	0.00	0.045	0.000	0.000
-CO	112.3	0.019	42.90	0.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
-H2	15.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	65.3	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.590
WATER	13054.6	2.207	0.00	0.00	0.00	0.00	0.00	0.00	0.036	0.000	0.014
-LIQUID	13054.6	2.207	0.28	0.00	0.00	0.01	0.00	100.00	0.036	0.000	0.014
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	18799.0	3.178	0.00	0.00	0.00	0.00	0.00	0.00	0.954	0.642	0.800

OVERALL AND COMPONENT BALANCES											
KEN#8	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9162.6	1.000	16.95	2.23	0.83	1.82	74.76	1.69	1.000	1.000	1.000
SPENT SHALE	7066.4	0.771	0.00	0.00	0.00	0.00	0.00	0.00	0.353	0.605	0.199
-BIN	6680.4	0.729	7.19	0.15	0.63	0.42	92.37	0.28	0.309	0.553	0.168
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	253.1	0.028	14.62	0.99	0.87	1.33	81.12	1.20	0.024	0.029	0.020
-A/C	132.9	0.015	23.60	1.60	1.28	1.27	69.41	1.23	0.020	0.022	0.010
OIL	850.7	0.093	0.00	0.00	0.00	0.00	0.00	0.00	0.415	0.106	0.094
-A/C	125.5	0.014	86.83	9.31	1.39	2.47	0.00	0.00	0.070	0.023	0.019
-ESP	486.2	0.053	67.92	7.28	1.09	1.93	2.75	17.11	0.213	0.070	0.056
-LP CIL	153.5	0.017	85.06	11.48	0.64	2.06	0.00	0.13	0.084	0.013	0.019
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	85.5	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
GAS	536.2	0.059	0.00	0.00	0.00	0.00	0.00	0.00	0.113	0.000	0.491
-CH4	27.1	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-C2'S	17.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	15.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	14.1	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	222.1	0.024	27.30	0.00	0.00	0.00	0.00	0.00	0.039	0.000	0.000
-CO	128.3	0.014	42.90	0.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
-H2	24.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	87.1	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.491
WATER	10011.9	1.093	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.183
-LIQUID	10011.9	1.093	0.28	0.00	0.00	0.30	0.00	100.00	0.018	0.000	0.183
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	18465.2	2.015	0.00	0.00	0.00	0.00	0.00	0.00	0.899	0.710	0.966

OVERALL AND COMPONENT BALANCES											
KEN#9	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4572.7	1.000	16.93	2.16	0.77	1.78	74.58	2.07	1.000	1.000	1.000
SPENT SHALE	3379.4	0.739	0.00	0.00	0.00	0.00	0.00	0.00	0.301	0.524	0.156
-BIN	3057.0	0.669	6.41	0.13	0.53	0.33	92.89	0.42	0.253	0.460	0.124
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	289.5	0.063	10.20	0.58	0.64	0.75	86.62	0.93	0.038	0.053	0.027
-A/C	32.9	0.007	22.04	1.41	1.22	1.31	71.14	1.20	0.009	0.011	0.005
OIL	431.7	0.094	0.00	0.00	0.00	0.00	0.00	0.00	0.424	0.091	0.071
-A/C	68.2	0.015	87.86	9.52	1.08	1.55	0.00	0.00	0.077	0.021	0.013
-ESP	253.8	0.056	68.63	7.44	0.84	1.21	3.50	16.06	0.225	0.061	0.038
-LP CIL	70.6	0.015	84.96	11.63	0.46	2.31	0.00	0.24	0.077	0.009	0.020
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	39.1	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.044	0.000	0.000
GAS	343.3	0.075	0.00	0.00	0.00	0.00	0.00	0.00	0.142	0.000	0.671
-CH4	16.5	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	9.7	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	8.1	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	7.7	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	132.8	0.029	27.30	0.00	0.00	0.00	0.00	0.00	0.047	0.000	0.000
-CO	92.8	0.020	42.90	0.00	0.00	0.00	0.00	0.00	0.051	0.000	0.000
-H2	17.7	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	58.0	0.013	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.671
WATER	6867.5	1.502	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.237
-LIQUID	6867.5	1.502	0.23	0.00	0.00	0.28	0.00	100.00	0.020	0.000	0.237
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11021.9	2.410	0.00	0.00	0.00	0.00	0.00	0.00	0.886	0.615	1.134

OVERALL AND COMPONENT BALANCES											
KEN#10	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9211.7	1.000	16.73	2.11	0.80	1.83	74.70	2.26	1.000	1.000	1.000
SPENT SHALE	6759.6	0.734	0.00	0.00	0.00	0.00	0.00	0.00	0.290	0.507	0.120
-BIN	6500.9	0.706	6.17	0.14	0.54	0.27	93.75	0.09	0.260	0.476	0.104
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	211.1	0.023	17.24	1.40	0.82	1.04	77.78	1.52	0.024	0.023	0.013
-A/C	47.6	0.005	18.83	1.20	1.13	0.85	75.81	0.98	0.006	0.007	0.002
OIL	922.0	0.100	0.00	0.00	0.00	0.00	0.00	0.00	0.463	0.105	0.086
-A/C	191.4	0.021	86.98	9.96	1.16	1.90	0.00	0.00	0.108	0.030	0.022
-ESP	514.1	0.056	70.38	8.06	0.94	1.54	2.25	14.25	0.235	0.066	0.047
-LP OIL	144.1	0.016	84.72	10.60	0.49	1.99	0.00	0.14	0.079	0.010	0.017
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	72.4	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.041	0.000	0.000
GAS	674.9	0.073	0.00	0.00	0.00	0.00	0.00	0.00	0.140	0.000	0.582
-CH4	32.2	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	18.6	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	15.5	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	15.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	263.5	0.029	27.30	0.00	0.00	0.00	0.00	0.00	0.047	0.000	0.000
-CO	185.5	0.020	42.90	0.00	0.00	0.00	0.00	0.00	0.052	0.000	0.000
-H2	40.4	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	104.2	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.582
WATER	11489.7	1.247	0.00	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.200
-LIQUID	11489.7	1.247	0.25	0.00	0.00	0.29	0.00	100.00	0.019	0.000	0.200
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	19846.2	2.154	0.00	0.00	0.00	0.00	0.00	0.00	0.912	0.612	0.987

OVERALL AND COMPONENT BALANCES											
KEN11	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4979.6	1.000	16.69	2.06	0.75	1.79	74.73	2.05	1.000	1.000	1.000
SPENT SHALE	3865.8	0.776	0.00	0.00	0.00	0.00	0.00	0.00	0.315	0.587	0.147
-BIN	3717.1	0.746	6.58	0.13	0.56	0.32	93.11	0.20	0.294	0.557	0.133
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	127.5	0.026	10.50	0.82	0.69	0.81	85.17	2.20	0.016	0.024	0.012
-A/C	21.2	0.004	17.88	1.18	1.09	0.96	76.85	0.98	0.005	0.006	0.002
OIL	459.5	0.092	0.00	0.00	0.00	0.00	0.00	0.00	0.418	0.117	0.080
-A/C	49.6	0.010	87.03	9.70	1.40	1.87	0.00	0.00	0.052	0.019	0.010
-ESP	287.0	0.058	69.73	7.77	1.12	1.50	2.50	15.59	0.241	0.086	0.048
-LP OIL	86.2	0.017	84.20	11.09	0.53	2.19	0.00	0.18	0.087	0.012	0.021
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	36.7	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.038	0.000	0.000
GAS	342.5	0.069	0.00	0.00	0.00	0.00	0.00	0.00	0.134	0.000	0.647
-CH4	18.0	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	10.7	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	8.8	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	8.1	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	121.6	0.024	27.30	0.00	0.00	0.00	0.00	0.00	0.040	0.000	0.000
-CO	97.1	0.019	42.90	0.00	0.00	0.00	0.00	0.00	0.050	0.000	0.000
-H2	16.9	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	61.3	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.647
WATER	6531.8	1.312	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.023
-LIQUID	6531.8	1.312	0.26	0.00	0.00	0.03	0.00	100.00	0.020	0.000	0.023
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11199.6	2.249	0.00	0.00	0.00	0.00	0.00	0.00	0.888	0.704	0.898

OVERALL AND COMPONENT BALANCES											
KEN#12	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6168.6	1.000	16.77	2.19	0.69	1.79	74.65	2.41	1.000	1.000	1.000
SPENT SHALE	4674.7	0.758	0.00	0.00	0.00	0.00	0.00	0.00	0.319	0.545	0.218
-BIN	4557.5	0.739	6.87	0.24	0.49	0.50	91.90	0.42	0.303	0.525	0.206
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	101.7	0.016	14.24	1.30	0.67	1.17	80.30	2.52	0.014	0.016	0.011
-A/C	15.5	0.003	18.69	1.25	1.12	0.94	75.87	0.98	0.003	0.004	0.001
OIL	477.6	0.077	0.00	0.00	0.00	0.00	0.00	0.00	0.355	0.092	0.066
-A/C	55.4	0.009	87.21	9.72	1.15	1.92	0.00	0.00	0.047	0.015	0.010
-ESP	300.7	0.049	71.88	8.01	0.95	1.58	5.20	8.11	0.209	0.067	0.043
-LP OIL	71.3	0.012	83.16	11.90	0.60	2.11	0.00	1.09	0.057	0.010	0.014
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	50.2	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
GAS	262.4	0.043	0.00	0.00	0.00	0.00	0.00	0.00	0.088	0.000	0.522
-CH4	23.1	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	13.7	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	12.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	11.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-CO2	85.9	0.014	27.30	0.00	0.00	0.00	0.00	0.00	0.023	0.000	0.000
-CO	42.7	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-H2	11.5	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	61.3	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.522
WATER	4429.5	0.718	0.00	0.00	0.00	0.00	0.00	0.00	0.012	0.000	0.183
-LIQUID	4429.5	0.718	0.28	0.00	0.00	0.46	0.00	100.00	0.012	0.000	0.183
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	9844.2	1.596	0.00	0.00	0.00	0.00	0.00	0.00	0.774	0.637	0.990

OVERALL AND COMPONENT BALANCES											
KEN#14	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9675.5	1.000	16.83	2.27	0.77	1.82	74.36	2.36	1.000	1.000	1.000
SPENT SHALE	7662.0	0.792	0.00	0.00	0.00	0.00	0.00	0.00	0.352	0.571	0.229
-BIN	7459.8	0.771	7.31	0.17	0.55	0.51	91.93	0.35	0.335	0.551	0.216
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	163.5	0.017	12.42	1.08	0.68	1.19	82.37	2.24	0.012	0.015	0.011
-A/C	38.7	0.004	19.47	1.24	1.01	0.96	75.30	0.97	0.005	0.005	0.002
OIL	822.0	0.085	0.00	0.00	0.00	0.00	0.00	0.00	0.409	0.115	0.082
-A/C	99.5	0.010	87.35	9.19	1.41	2.04	0.00	0.00	0.053	0.019	0.012
-ESP	426.6	0.044	75.34	7.93	1.22	1.76	4.50	7.29	0.197	0.070	0.043
-LP OIL	237.5	0.025	87.06	10.72	0.81	2.04	0.00	0.27	0.127	0.026	0.028
0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	58.4	0.006	87.00	13.00	0.00	0.00	0.00	0.00	0.031	0.000	0.000
GAS	537.1	0.056	0.00	0.00	0.00	0.00	0.00	0.00	0.109	0.000	0.481
-CH4	33.6	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	20.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	16.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	15.8	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	206.2	0.021	27.30	0.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
-CO	121.0	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.032	0.000	0.000
-H2	33.8	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	90.1	0.009	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.481
WATER	6091.7	0.630	0.00	0.00	0.00	0.00	0.00	0.00	0.010	0.000	0.171
-LIQUID	6091.7	0.630	0.28	0.00	0.00	0.49	0.00	100.00	0.010	0.000	0.171
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	15112.8	1.562	0.00	0.00	0.00	0.00	0.00	0.00	0.880	0.685	0.963

OVERALL AND COMPONENT BALANCES											
KEN#15	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9975.9	1.000	16.65	1.74	1.02	1.76	74.75	2.36	1.000	1.000	1.000
SPENT SHALE	7596.4	0.761	0.00	0.00	0.00	0.00	0.00	0.00	0.397	0.686	0.365
-BIN	6123.2	0.614	8.35	0.32	0.96	0.79	88.74	1.43	0.308	0.578	0.276
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	1378.3	0.138	9.26	0.60	0.73	1.06	87.26	0.79	0.077	0.099	0.083
-A/C	94.9	0.010	20.75	1.50	1.00	1.21	73.62	0.93	0.012	0.009	0.007
OIL	793.6	0.080	0.00	0.00	0.00	0.00	0.00	0.00	0.360	0.090	0.067
-A/C	84.4	0.008	88.06	8.41	1.72	1.81	0.00	0.00	0.045	0.014	0.009
-ESP	496.2	0.050	68.71	6.56	1.34	1.41	7.10	11.30	0.205	0.065	0.040
-LP OIL	162.0	0.016	85.43	11.48	0.63	2.00	0.00	0.15	0.083	0.010	0.018
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	51.0	0.005	87.00	13.00	0.00	0.00	0.00	0.00	0.027	0.000	0.000
GAS	309.2	0.031	0.00	0.00	0.00	0.00	0.00	0.00	0.064	0.000	0.434
-CH4	26.9	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C2'S	18.4	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	15.8	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	14.4	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	94.2	0.009	27.30	0.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-CO	46.7	0.005	42.90	0.00	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-H2	11.9	0.001	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	80.9	0.008	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.434
WATER	7289.0	0.731	0.00	0.00	0.00	0.00	0.00	0.00	0.010	0.000	0.032
-LIQUID	7289.0	0.731	0.22	0.00	0.00	0.08	0.00	100.00	0.010	0.000	0.032
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	15988.2	1.603	0.00	0.00	0.00	0.00	0.00	0.00	0.830	0.776	0.898

OVERALL AND COMPONENT BALANCES											
KEN#16	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4242.0	1.000	16.68	1.81	0.84	1.76	74.72	2.46	1.000	1.000	1.000
SPENT SHALE	3435.2	0.810	0.00	0.00	0.00	0.00	0.00	0.00	0.429	0.772	0.258
-BIN	3207.6	0.756	8.48	0.08	0.80	0.53	90.47	0.56	0.384	0.720	0.228
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	198.8	0.047	13.37	0.75	0.77	0.99	82.94	1.42	0.038	0.043	0.026
-A/C	28.8	0.007	16.53	1.09	1.14	1.04	76.81	2.10	0.007	0.009	0.004
OIL	425.3	0.100	0.00	0.00	0.00	0.00	0.00	0.00	0.431	0.123	0.092
-A/C	53.1	0.013	86.95	9.27	1.62	2.16	0.00	0.00	0.065	0.024	0.015
-ESP	264.2	0.062	62.75	6.69	1.17	1.56	5.00	22.55	0.234	0.087	0.055
-LP CIL	70.7	0.017	85.54	11.16	0.59	2.29	0.00	0.38	0.085	0.012	0.022
0.0	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	37.3	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.046	0.000	0.000
GAS	172.1	0.041	0.00	0.00	0.00	0.00	0.00	0.00	0.077	0.000	0.587
-CH4	14.3	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	9.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	7.5	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	6.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	55.2	0.013	27.30	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-CO	22.1	0.005	42.90	0.00	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-H2	10.4	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	46.6	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.587
WATER	4934.5	1.163	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.040
-LIQUID	4934.5	1.163	0.29	0.00	0.00	0.06	0.00	100.00	0.020	0.000	0.040
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	8967.1	2.114	0.00	0.00	0.00	0.00	0.00	0.00	0.957	0.895	0.977

OVERALL AND COMPONENT BALANCES											
KEN#17	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5616.5	1.000	16.46	1.74	0.88	1.75	74.50	2.34	1.000	1.000	1.000
SPENT SHALE	4540.5	0.808	0.00	0.00	0.00	0.00	0.00	0.00	0.363	0.613	0.257
-BIN	4228.6	0.753	7.01	0.08	0.65	0.53	90.87	0.55	0.321	0.556	0.228
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	264.7	0.047	11.99	0.47	0.85	0.89	84.72	1.53	0.034	0.046	0.024
-A/C	47.2	0.008	15.60	0.85	1.21	1.00	78.47	1.47	0.008	0.012	0.005
OIL	482.0	0.086	0.00	0.00	0.00	0.00	0.00	0.00	0.404	0.077	0.093
-A/C	69.1	0.012	87.86	8.56	1.12	2.46	0.00	0.00	0.066	0.016	0.017
-ESP	286.1	0.051	71.17	6.93	0.91	1.99	3.50	12.95	0.220	0.053	0.058
-LP OIL	82.3	0.015	85.88	11.19	0.54	2.08	0.00	0.22	0.076	0.009	0.017
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	44.5	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
GAS	225.8	0.040	0.00	0.00	0.00	0.00	0.00	0.00	0.076	0.000	0.555
-CH4	16.9	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	11.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	9.2	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	7.8	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	76.8	0.014	27.30	0.00	0.00	0.00	0.00	0.00	0.023	0.000	0.000
-CO	31.4	0.006	42.90	0.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-H2	14.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	58.0	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.555
WATER	5799.8	1.033	0.00	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.071
-LIQUID	5799.8	1.033	0.33	0.00	0.00	0.12	0.00	100.00	0.021	0.000	0.071
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11048.1	1.967	0.00	0.00	0.00	0.00	0.00	0.00	0.864	0.691	0.976

OVERALL AND COMPONENT BALANCES											
KEN#18	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6126.1	1.000	16.46	1.66	0.73	2.03	74.33	2.61	1.000	1.000	1.000
SPENT SHALE	4737.1	0.773	0.00	0.00	0.00	0.00	0.00	0.00	0.390	0.677	0.330
-BIN	4532.7	0.740	7.99	0.06	0.63	0.84	90.74	0.62	0.359	0.639	0.306
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	167.4	0.027	14.84	0.92	0.75	1.48	79.88	1.40	0.025	0.028	0.020
-A/C	37.0	0.006	15.97	0.98	1.21	1.23	76.99	2.17	0.006	0.010	0.004
OIL	573.4	0.094	0.00	0.00	0.00	0.00	0.00	0.00	0.436	0.133	0.068
-A/C	113.5	0.019	88.35	8.32	1.51	1.83	0.00	0.00	0.099	0.038	0.017
-ESP	334.2	0.055	69.16	6.51	1.15	1.43	3.00	16.35	0.229	0.086	0.038
-LP CIL	76.3	0.012	85.40	11.50	0.52	2.12	0.00	0.22	0.065	0.009	0.013
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	49.4	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.043	0.000	0.000
GAS	265.4	0.043	0.00	0.00	0.00	0.00	0.00	0.00	0.080	0.000	0.522
-CH4	18.1	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-C2'S	11.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	9.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	9.1	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	90.2	0.015	27.30	0.00	0.00	0.00	0.00	0.00	0.024	0.000	0.000
-CO	42.1	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-H2	16.4	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	69.0	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.522
WATER	6023.5	0.983	0.00	0.00	0.00	0.00	0.00	0.00	0.013	0.000	0.082
-LIQUID	6023.5	0.983	0.22	0.00	0.00	0.17	0.00	100.00	0.013	0.000	0.082
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11599.4	1.893	0.00	0.00	0.00	0.00	0.00	0.00	0.919	0.810	1.002

Runs

OVERALL AND COMPONENT BALANCES											
KEN#19	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6671.0	1.000	16.51	1.66	0.82	2.15	74.28	2.54	1.000	1.000	1.000
SPENT SHALE	2379.7	0.357	0.00	0.00	0.00	0.00	0.00	0.00	0.179	0.270	0.109
-BIN	2138.8	0.321	7.60	0.05	0.59	0.60	91.47	0.55	0.148	0.231	0.089
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	201.5	0.030	13.45	0.72	0.86	1.20	81.36	1.73	0.025	0.032	0.017
-A/C	39.4	0.006	18.66	1.37	1.06	1.15	72.57	3.56	0.007	0.008	0.003
OIL	627.3	0.094	0.00	0.00	0.00	0.00	0.00	0.00	0.435	0.118	0.066
-A/C	58.0	0.009	88.23	8.33	1.53	1.91	0.00	0.00	0.046	0.016	0.008
-ESP	388.5	0.058	69.69	6.58	1.21	1.51	3.10	15.84	0.246	0.086	0.041
-LF OIL	127.5	0.019	86.74	9.83	0.69	1.97	0.00	0.22	0.100	0.016	0.018
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	53.3	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
GAS	335.6	0.050	0.00	0.00	0.00	0.00	0.00	0.00	0.092	0.000	0.558
-CH4	20.3	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	13.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	10.9	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	10.2	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	113.7	0.017	27.30	0.00	0.00	0.00	0.00	0.00	0.028	0.000	0.000
-CO	61.1	0.009	42.90	0.00	0.00	0.00	0.00	0.00	0.024	0.000	0.000
-H2	21.1	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	85.1	0.013	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.558
WATER	8130.3	1.219	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.062
-LIQUID	8130.3	1.219	0.27	0.00	0.00	0.11	0.00	100.00	0.020	0.000	0.062
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11472.9	1.720	0.00	0.00	0.00	0.00	0.00	0.00	0.725	0.388	0.796

Appendix B (Cont.) - Mass and Component Balances for Mini-Plant

OVERALL AND COMPONENT BALANCES											
KEN#20	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	3594.0	1.000	16.56	1.76	1.02	2.12	75.47	2.61	1.000	1.000	1.000
SPENT SHALE	2859.1	0.796	0.00	0.00	0.00	0.00	0.00	0.00	0.392	0.474	0.280
-BIN	2533.8	0.705	7.89	0.08	0.60	0.71	91.07	0.56	0.336	0.415	0.236
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	299.0	0.083	9.93	0.47	0.62	1.00	85.90	1.93	0.050	0.051	0.039
-A/C	26.3	0.007	13.95	2.32	1.16	1.45	65.39	15.63	0.006	0.008	0.005
OIL	368.9	0.103	0.00	0.00	0.00	0.00	0.00	0.00	0.485	0.099	0.077
-A/C	78.8	0.022	87.62	9.02	1.39	1.97	0.00	0.00	0.116	0.030	0.020
-ESP	193.4	0.054	70.55	7.26	1.12	1.59	3.90	11.55	0.229	0.059	0.040
-LP OIL	62.8	0.017	85.23	11.35	0.61	2.00	0.00	0.22	0.090	0.010	0.016
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	33.9	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.050	0.000	0.000
GAS	145.9	0.041	0.00	0.00	0.00	0.00	0.00	0.00	0.079	0.000	0.525
-CH4	12.4	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	8.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	6.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	6.3	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	43.0	0.012	27.30	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO	18.8	0.005	42.90	0.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-H2	8.0	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	42.5	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.525
WATER	4015.2	1.117	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.025
-LIQUID	4015.2	1.117	0.27	0.00	0.00	0.12	0.00	100.00	0.018	0.000	0.025
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	7389.1	2.056	0.00	0.00	0.00	0.00	0.00	0.00	0.973	0.573	0.908

Runs

OVERALL AND COMPONENT BALANCES											
KEN#21	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6147.3	1.000	16.43	1.83	1.02	2.01	74.59	2.50	1.000	1.000	1.000
SPENT SHALE	4965.3	0.808	0.00	0.00	0.00	0.00	0.00	0.00	0.476	0.481	0.325
-BIN	2006.1	0.326	6.74	0.41	0.46	0.48	88.62	3.86	0.134	0.147	0.078
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	1649.1	0.268	8.18	0.17	0.49	0.81	90.07	0.40	0.134	0.129	0.108
-A/C	1310.1	0.213	16.09	1.00	0.98	1.31	78.79	0.91	0.209	0.205	0.139
OIL	506.3	0.082	0.00	0.00	0.00	0.00	0.00	0.00	0.387	0.064	0.061
-A/C	57.1	0.009	87.86	9.07	1.25	1.82	0.00	0.00	0.050	0.011	0.008
-ESP	280.7	0.046	70.16	7.24	1.00	1.45	5.30	11.39	0.195	0.045	0.033
-LF OIL	112.7	0.018	84.12	11.59	0.44	2.14	0.00	0.21	0.094	0.008	0.020
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	55.8	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
GAS	406.2	0.066	0.00	0.00	0.00	0.00	0.00	0.00	0.128	0.000	0.481
-CH4	22.6	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	13.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	11.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	9.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	170.5	0.028	27.30	0.00	0.00	0.00	0.00	0.00	0.046	0.000	0.000
-CO	86.5	0.014	42.90	0.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
-H2	28.3	0.005	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	63.2	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.481
WATER	6421.9	1.045	0.00	0.00	0.00	0.00	0.00	0.00	0.016	0.000	0.047
-LIQUID	6421.9	1.045	0.25	0.00	0.00	0.09	0.00	100.00	0.016	0.000	0.047
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12299.7	2.001	0.00	0.00	0.00	0.00	0.00	0.00	1.007	0.545	0.914

OVERALL AND COMPONENT BALANCES											
KEN#22	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5844.6	1.000	16.53	1.84	0.76	2.02	74.89	2.24	1.000	1.000	1.000
SPENT SHALE	4544.1	0.777	0.00	0.00	0.00	0.00	0.00	0.00	0.316	0.453	0.189
-BIN	4279.8	0.732	6.47	0.05	0.43	0.45	93.09	0.34	0.287	0.414	0.163
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	217.1	0.037	10.19	0.48	0.58	1.11	86.59	0.90	0.023	0.028	0.020
-A/C	47.2	0.008	14.01	0.90	1.01	1.28	80.67	1.04	0.007	0.011	0.005
OIL	428.3	0.073	0.00	0.00	0.00	0.00	0.00	0.00	0.353	0.066	0.052
-A/C	0.0	0.000	88.02	9.11	1.07	1.80	0.00	0.00	0.000	0.000	0.000
-ESP	277.8	0.048	76.81	7.95	0.93	1.57	3.00	6.11	0.221	0.058	0.037
-LP CIL	93.9	0.016	83.73	11.91	0.39	1.91	0.00	0.26	0.081	0.008	0.015
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	56.6	0.010	87.00	13.00	0.00	0.00	0.00	0.00	0.051	0.000	0.000
GAS	504.8	0.086	0.00	0.00	0.00	0.00	0.00	0.00	0.162	0.000	0.614
-CH4	24.6	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-C2'S	14.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	11.8	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	10.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	217.0	0.037	27.30	0.00	0.00	0.00	0.00	0.00	0.061	0.000	0.000
-CO	113.3	0.019	42.90	0.00	0.00	0.00	0.00	0.00	0.050	0.000	0.000
-H2	36.8	0.006	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	77.0	0.013	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.614
WATER	5109.8	0.874	0.00	0.00	0.00	0.00	0.00	0.00	0.014	0.000	0.082
-LIQUID	5109.8	0.874	0.27	0.00	0.00	0.19	0.00	100.00	0.014	0.000	0.082
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	10587.0	1.811	0.00	0.00	0.00	0.00	0.00	0.00	0.846	0.520	0.937

OVERALL AND COMPONENT BALANCES											
KEN#23	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5295.9	1.000	16.49	1.81	0.74	2.10	75.10	1.96	1.000	1.000	1.000
SPENT SHALE	4435.6	0.838	0.00	0.00	0.00	0.00	0.00	0.00	0.392	0.616	0.278
-BIN	3554.6	0.671	7.17	0.17	0.53	0.63	89.71	1.13	0.292	0.481	0.201
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	850.0	0.161	9.63	0.21	0.59	0.95	87.44	0.80	0.094	0.128	0.073
-A/C	31.0	0.006	17.59	1.03	0.97	1.28	77.07	0.75	0.006	0.008	0.004
OIL	454.8	0.086	0.00	0.00	0.00	0.00	0.00	0.00	0.317	0.097	0.064
-A/C	46.4	0.009	87.60	9.18	1.23	1.99	0.00	0.00	0.047	0.015	0.008
-ESP	268.4	0.051	72.05	7.55	1.01	1.64	3.10	11.11	0.221	0.069	0.040
-LP OIL	90.6	0.017	0.00	11.84	0.56	1.93	0.00	0.22	0.000	0.013	0.016
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	49.4	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.049	0.000	0.000
GAS	222.6	0.042	0.00	0.00	0.00	0.00	0.00	0.00	0.085	0.000	0.507
-CH4	20.3	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	13.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	10.6	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	9.8	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	65.8	0.012	27.30	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-CO	30.7	0.006	42.90	0.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-H2	12.3	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	59.9	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.507
WATER	5886.2	1.111	0.00	0.00	0.00	0.00	0.00	0.00	0.016	0.000	0.064
-LIQUID	5886.2	1.111	0.24	0.00	0.00	0.12	0.00	100.00	0.016	0.000	0.064
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	10999.2	2.077	0.00	0.00	0.00	0.00	0.00	0.00	0.810	0.713	0.911

Runs

OVERALL AND COMPONENT BALANCES											
KEN#24	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	4121.6	1.000	16.30	1.87	0.73	2.10	74.85	2.22	1.000	1.000	1.000
SPENT SHALE	3248.7	0.788	0.00	0.00	0.00	0.00	0.00	0.00	0.366	0.533	0.208
-BIN	2809.1	0.682	7.16	0.11	0.48	0.49	91.80	0.71	0.299	0.448	0.159
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	416.6	0.101	9.86	0.42	0.56	0.95	87.31	0.82	0.061	0.078	0.046
-A/C	23.0	0.006	15.04	1.22	0.95	1.21	77.91	2.04	0.005	0.007	0.003
OIL	358.5	0.087	0.00	0.00	0.00	0.00	0.00	0.00	0.415	0.129	0.057
-A/C	50.6	0.012	86.95	9.75	1.59	1.71	0.00	0.00	0.065	0.027	0.010
-ESP	206.3	0.050	71.07	7.97	1.30	1.40	3.80	12.15	0.218	0.089	0.033
-LP CIL	64.0	0.016	86.30	11.84	0.64	1.79	0.00	0.18	0.082	0.014	0.013
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	37.6	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.049	0.000	0.000
GAS	265.9	0.065	0.00	0.00	0.00	0.00	0.00	0.00	0.127	0.000	0.489
-CH4	15.5	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	9.6	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	7.9	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	7.2	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	107.1	0.026	27.30	0.00	0.00	0.00	0.00	0.00	0.044	0.000	0.000
-CO	55.6	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
-H2	18.0	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	45.0	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.489
WATER	4118.9	0.999	0.00	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.076
-LIQUID	4118.9	0.999	0.27	0.00	0.00	0.16	0.00	100.00	0.017	0.000	0.076
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	7992.0	1.939	0.00	0.00	0.00	0.00	0.00	0.00	0.924	0.662	0.830

Runs

OVERALL AND COMPONENT BALANCES											
KEN#25	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	3861.2	1.000	16.49	1.90	0.74	2.10	74.81	2.22	1.000	1.000	1.000
SPENT SHALE	3232.3	0.837	0.00	0.00	0.00	0.00	0.00	0.00	0.394	0.610	0.229
-BIN	2814.3	0.729	7.48	0.04	0.53	0.50	91.66	0.47	0.331	0.522	0.174
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	391.8	0.101	9.14	0.41	0.58	1.08	87.37	1.13	0.056	0.080	0.052
-A/C	26.2	0.007	18.12	1.14	0.97	1.14	76.61	0.76	0.007	0.009	0.004
OIL	340.7	0.088	0.00	0.00	0.00	0.00	0.00	0.00	0.428	0.185	0.065
-A/C	34.4	0.009	87.77	11.06	1.86	1.86	0.00	0.00	0.047	0.022	0.008
-ESP	204.0	0.053	75.77	7.67	1.61	1.61	4.62	9.13	0.243	0.115	0.041
-LF OIL	69.6	0.018	85.28	12.01	1.95	1.95	0.00	0.27	0.093	0.047	0.017
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	32.7	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.045	0.000	0.000
GAS	176.6	0.046	0.00	0.00	0.00	0.00	0.00	0.00	0.088	0.000	0.464
-CH4	12.2	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	7.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	6.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	5.5	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	63.1	0.016	27.30	0.00	0.00	0.00	0.00	0.00	0.027	0.000	0.000
-CO	30.8	0.008	42.90	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-H2	11.1	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	40.0	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.464
WATER	4105.6	1.063	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.061
-LIQUID	4105.6	1.063	0.31	0.00	0.00	0.12	0.00	100.00	0.020	0.000	0.061
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	7855.2	2.034	0.00	0.00	0.00	0.00	0.00	0.00	0.930	0.795	0.820

OVERALL AND COMPONENT BALANCES											
KEN#26	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% Ash	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6318.7	1.000	16.58	1.86	0.78	2.06	74.68	2.40	1.000	1.000	1.000
SPENT SHALE	5191.1	0.822	0.00	0.00	0.00	0.00	0.00	0.00	0.419	0.661	0.314
-BIN	1175.6	0.186	7.37	0.31	0.56	0.53	89.48	1.28	0.083	0.134	0.048
-A/C-2	2174.5	0.344	8.90	0.24	0.64	0.85	88.70	0.70	0.185	0.282	0.142
-CYCLONE	1580.2	0.250	8.31	0.20	0.65	0.85	89.54	0.42	0.125	0.208	0.103
-A/C	260.8	0.041	10.47	0.36	0.70	1.07	86.00	0.98	0.026	0.037	0.021
OIL	509.1	0.081	0.00	0.00	0.00	0.00	0.00	0.00	0.386	0.107	0.061
-A/C	61.6	0.010	87.38	9.13	1.59	1.90	0.00	0.00	0.051	0.020	0.009
-ESP	276.5	0.044	73.44	7.67	1.34	1.60	6.50	8.92	0.194	0.075	0.034
-LP OIL	121.3	0.019	86.13	11.99	0.49	1.89	0.00	0.15	0.100	0.012	0.018
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	49.7	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.041	0.000	0.000
GAS	278.7	0.044	0.00	0.00	0.00	0.00	0.00	0.00	0.087	0.000	0.474
-CH4	20.3	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	14.6	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	12.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	11.1	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	93.0	0.015	27.30	0.00	0.00	0.00	0.00	0.00	0.024	0.000	0.000
-CO	44.8	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-H2	17.0	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	65.5	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.474
WATER	5894.9	0.933	0.00	0.00	0.00	0.00	0.00	0.00	0.015	0.000	0.050
-LIQUID	5894.9	0.933	0.26	0.00	0.00	0.11	0.00	100.00	0.015	0.000	0.050
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11873.8	1.879	0.00	0.00	0.00	0.00	0.00	0.00	0.907	0.768	0.898

Runs

OVERALL AND COMPONENT BALANCES											
KEN#27	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6859.7	1.000	16.77	1.81	0.75	2.05	75.09	1.93	1.000	1.000	1.000
SPENT SHALE	5548.0	0.809	0.00	0.00	0.00	0.00	0.00	0.00	0.393	0.592	0.258
-BIN	5330.4	0.777	7.87	0.07	0.54	0.63	91.57	0.45	0.365	0.559	0.239
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	180.1	0.026	14.00	0.83	0.73	1.31	81.69	1.20	0.022	0.026	0.017
-A/C	37.5	0.005	19.15	1.14	1.02	0.94	75.50	1.21	0.006	0.007	0.003
OIL	584.2	0.085	0.00	0.00	0.00	0.00	0.00	0.00	0.403	0.129	0.069
-A/C	65.4	0.010	87.14	9.29	1.60	1.96	0.00	0.00	0.050	0.020	0.009
-ESP	356.4	0.052	75.10	8.01	1.38	1.69	3.75	8.14	0.233	0.096	0.043
-LP OIL	114.2	0.017	85.19	12.28	0.61	2.14	0.00	0.17	0.085	0.014	0.017
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	48.2	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
GAS	305.0	0.044	0.00	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.510
-CH4	20.9	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	12.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	10.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	10.4	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	104.8	0.015	27.30	0.00	0.00	0.00	0.00	0.00	0.025	0.000	0.000
-CO	50.4	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-H2	18.7	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	76.2	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.510
WATER	6523.5	0.951	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.065
-LIQUID	6523.5	0.951	0.31	0.00	0.00	0.14	0.00	100.00	0.018	0.000	0.065
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12960.7	1.889	0.00	0.00	0.00	0.00	0.00	0.00	0.896	0.722	0.902

Appendix B (Cont.) - Mass and Component Balances for Mini-Plant
Runs

OVERALL AND COMPONENT BALANCES											
KEN#28	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	3878.1	1.000	16.58	1.87	0.74	1.95	74.97	2.05	1.000	1.000	1.000
SPENT SHALE	3218.4	0.830	0.00	0.00	0.00	0.00	0.00	0.00	0.368	0.613	0.258
-BIN	2931.5	0.756	7.28	0.07	0.55	0.60	91.79	0.50	0.332	0.562	0.233
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	264.6	0.068	7.36	0.31	0.46	0.66	90.23	1.11	0.030	0.042	0.023
-A/C	22.3	0.006	16.51	1.04	1.10	0.84	79.10	1.16	0.006	0.009	0.002
OIL	327.9	0.085	0.00	0.00	0.00	0.00	0.00	0.00	0.409	0.114	0.068
-A/C	32.7	0.008	87.64	9.22	1.30	1.84	0.00	0.00	0.045	0.015	0.008
-ESP	203.9	0.053	75.92	7.99	1.13	1.59	4.50	7.39	0.241	0.080	0.043
-LP OIL	65.8	0.017	86.83	11.23	0.83	2.00	0.00	0.00	0.089	0.019	0.017
-Av analysis	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	25.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
GAS	158.8	0.041	0.00	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.442
-CH4	12.4	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	8.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	6.5	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	6.3	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	53.6	0.014	27.30	0.00	0.00	0.00	0.00	0.00	0.023	0.000	0.000
-CO	26.8	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-H2	9.6	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	35.5	0.009	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.442
WATER	4167.0	1.074	0.00	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.077
-LIQUID	4167.0	1.074	0.30	0.00	0.00	0.14	0.00	100.00	0.019	0.000	0.077
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	7872.1	2.030	0.00	0.00	0.00	0.00	0.00	0.00	0.878	0.727	0.845

OVERALL AND COMPONENT BALANCES											
KEN#29	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7193.7	1.000	16.53	1.79	0.94	1.90	74.86	1.88	1.000	1.000	1.000
SPENT SHALE	5692.1	0.791	0.00	0.00	0.00	0.00	0.00	0.00	0.356	0.439	0.244
-BIN	5288.0	0.735	7.32	0.07	0.51	0.57	92.11	0.44	0.326	0.399	0.221
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	363.7	0.051	8.19	0.43	0.62	0.77	87.27	2.70	0.025	0.033	0.020
-A/C	40.4	0.006	16.66	0.91	1.06	0.95	79.08	0.79	0.006	0.006	0.003
OIL	636.0	0.088	0.00	0.00	0.00	0.00	0.00	0.00	0.431	0.112	0.078
-A/C	66.1	0.009	87.60	8.74	1.72	1.94	0.00	0.00	0.049	0.017	0.009
-ESP	372.6	0.052	76.61	7.64	1.50	1.70	3.50	8.97	0.240	0.083	0.046
-LP CIL	147.8	0.021	85.08	12.00	0.57	2.04	0.00	0.21	0.106	0.012	0.022
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	49.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
GAS	391.8	0.054	0.00	0.00	0.00	0.00	0.00	0.00	0.103	0.000	0.597
-CH4	24.9	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	16.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	13.5	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	12.6	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	142.5	0.020	27.30	0.00	0.00	0.00	0.00	0.00	0.033	0.000	0.000
-CO	69.6	0.010	42.90	0.00	0.00	0.00	0.00	0.00	0.025	0.000	0.000
-H2	25.5	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	86.7	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.597
WATER	6971.2	0.969	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.061
-LIQUID	6971.2	0.969	0.30	0.00	0.00	0.12	0.00	100.00	0.018	0.000	0.061
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	13691.1	1.903	0.00	0.00	0.00	0.00	0.00	0.00	0.908	0.550	0.980

Runs

OVERALL AND COMPONENT BALANCES											
KEN#30	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9772.7	1.000	16.32	1.87	0.92	1.70	74.37	2.52	1.000	1.000	1.000
SPENT SHALE	7580.8	0.776	0.00	0.00	0.00	0.00	0.00	0.00	0.407	0.856	0.274
-BIN	7163.0	0.733	8.46	0.11	1.00	0.59	90.50	0.45	0.380	0.797	0.254
-TUBE	0.0	0.000	0.00	0.00	0.95	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	367.4	0.038	9.43	0.37	1.30	0.76	87.53	1.22	0.022	0.053	0.017
-A/C	50.4	0.005	16.88	0.93	1.06	0.90	77.85	1.15	0.005	0.006	0.003
OIL	781.7	0.080	0.00	0.00	0.00	0.00	0.00	0.00	0.399	0.076	0.080
-A/C	88.7	0.009	87.51	9.44	1.23	1.82	0.00	0.00	0.049	0.012	0.010
-ESP	433.7	0.044	78.39	8.46	1.10	1.63	3.80	2.01	0.213	0.053	0.043
-LP OIL	188.1	0.019	83.43	11.62	0.50	2.42	0.00	0.24	0.098	0.010	0.027
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	71.2	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.039	0.000	0.000
GAS	392.4	0.040	0.00	0.00	0.00	0.00	0.00	0.00	0.079	0.000	0.626
-CH4	34.9	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	21.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	17.1	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	16.5	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	117.7	0.012	27.30	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO	53.5	0.005	42.90	0.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-H2	21.1	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	110.5	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.626
WATER	7843.4	0.803	0.00	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.066
-LIQUID	7843.4	0.803	0.40	0.00	0.00	0.14	0.00	100.00	0.020	0.000	0.066
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	16598.3	1.698	0.00	0.00	0.00	0.00	0.00	0.00	0.905	0.931	1.046

OVERALL AND COMPONENT BALANCES											
KEN#31	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	10061.7	1.000	16.22	1.81	0.74	2.05	74.88	1.95	1.000	1.000	1.000
SPENT SHALE	7811.1	0.776	0.00	0.00	0.00	0.00	0.00	0.00	0.334	0.203	0.221
-BIN	7366.9	0.732	6.87	0.06	0.18	0.57	92.84	0.28	0.310	0.178	0.204
-TUBE	0.0	0.000	0.00	0.00	0.95	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	380.8	0.038	7.66	0.31	0.37	0.75	89.87	1.07	0.018	0.019	0.014
-A/C	63.4	0.006	16.49	0.84	0.75	1.04	78.74	0.98	0.006	0.006	0.003
OIL	794.5	0.079	0.00	0.00	0.00	0.00	0.00	0.00	0.392	0.099	0.064
-A/C	108.6	0.011	87.18	9.56	1.33	1.93	0.00	0.00	0.058	0.019	0.010
-ESP	441.0	0.044	76.76	8.42	1.17	1.70	4.10	2.94	0.207	0.069	0.036
-LP OIL	174.8	0.017	83.62	11.45	0.44	2.06	0.00	0.19	0.090	0.010	0.017
0.0	0.000	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	70.1	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
GAS	629.5	0.063	0.00	0.00	0.00	0.00	0.00	0.00	0.119	0.000	0.500
-CH4	32.9	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	20.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	16.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	14.4	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	264.8	0.026	27.30	0.00	0.00	0.00	0.00	0.00	0.044	0.000	0.000
-CO	127.1	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.033	0.000	0.000
-H2	44.1	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	109.6	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.500
WATER	8328.7	0.828	0.00	0.00	0.00	0.00	0.00	0.00	0.017	0.230	0.077
-LIQUID	8328.7	0.828	0.34	0.00	0.21	0.19	0.00	100.00	0.017	0.230	0.077
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	17563.8	1.746	0.00	0.00	0.00	0.00	0.00	0.00	0.863	0.533	0.861

Runs

OVERALL AND COMPONENT BALANCES											
KEN#32	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9173.0	1.000	16.46	1.77	0.74	2.15	74.61	2.36	1.000	1.000	1.000
SPENT SHALE	7954.1	0.867	0.00	0.00	0.00	0.00	0.00	0.00	0.418	0.282	0.322
-BIN	7732.2	0.843	7.73	0.06	0.23	0.78	91.74	0.28	0.396	0.262	0.306
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	164.2	0.018	13.56	0.72	0.54	1.46	82.27	0.98	0.015	0.013	0.012
-A/C	57.7	0.006	18.81	0.97	0.87	1.22	76.09	1.03	0.007	0.007	0.004
OIL	740.4	0.081	0.00	0.00	0.00	0.00	0.00	0.00	0.398	0.108	0.062
-A/C	80.6	0.009	87.14	9.55	1.39	1.92	0.00	0.00	0.047	0.017	0.008
-ESP	433.7	0.047	78.37	8.59	1.25	1.73	3.30	3.72	0.225	0.080	0.038
-LP OIL	159.6	0.017	83.70	11.52	0.49	1.95	0.00	0.17	0.088	0.012	0.016
	0.0	0.000	0.00	0.00	0.00	1.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	66.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.038	0.000	0.000
GAS	435.3	0.047	0.00	0.00	0.00	0.00	0.00	0.00	0.093	0.000	0.462
-CH4	33.5	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	20.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	16.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	15.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	161.9	0.018	27.30	0.00	0.00	0.00	0.00	0.00	0.029	0.000	0.000
-CO	62.9	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-H2	27.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	96.8	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.462
WATER	5922.8	0.646	0.00	0.00	0.00	0.00	0.00	0.00	0.016	0.181	0.065
-LIQUID	5922.8	0.646	0.41	0.00	0.21	0.22	0.00	100.00	0.016	0.181	0.065
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	15052.6	1.641	0.00	0.00	0.00	0.00	0.00	0.00	0.925	0.571	0.911

OVERALL AND COMPONENT BALANCES											
KEN#33	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	9135.1	1.000	16.39	1.88	0.74	2.00	74.62	2.17	1.000	1.000	1.000
SPENT SHALE	7339.5	0.803	0.00	0.00	0.00	0.00	0.00	0.00	0.404	0.302	0.349
-BIN	7098.3	0.777	8.04	0.11	0.26	0.85	91.07	0.29	0.381	0.273	0.330
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	181.3	0.020	13.11	0.75	0.76	1.52	83.07	0.70	0.016	0.020	0.015
-A/C	59.9	0.007	17.42	1.05	0.99	1.25	77.73	0.67	0.007	0.009	0.004
OIL	730.7	0.080	0.00	0.00	0.00	0.00	0.00	0.00	0.395	0.107	0.067
-A/C	88.8	0.010	86.97	9.66	1.39	1.98	0.00	0.00	0.052	0.018	0.010
-ESP	436.2	0.048	77.44	8.60	1.24	1.76	3.30	4.43	0.226	0.080	0.042
-LP CIL	141.0	0.015	85.36	11.62	0.42	1.97	0.00	0.13	0.080	0.009	0.015
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	64.7	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.038	0.000	0.000
GAS	368.4	0.040	0.00	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.441
-CH4	31.4	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	20.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	16.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	15.2	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	125.3	0.014	27.30	0.00	0.00	0.00	0.00	0.00	0.023	0.000	0.000
-CO	52.4	0.006	42.90	0.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-H2	21.9	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	85.6	0.009	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.441
WATER	6232.3	0.682	0.00	0.00	0.00	0.00	0.00	0.00	0.015	0.155	0.098
-LIQUID	6232.3	0.682	0.37	0.00	0.17	0.29	0.00	100.00	0.015	0.155	0.098
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	14670.9	1.606	0.00	0.00	0.00	0.00	0.00	0.00	0.897	0.564	0.955

OVERALL AND COMPONENT BALANCES											
KEN#34	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	8785.6	1.000	16.40	1.83	0.74	2.06	75.04	1.87	1.000	1.000	1.000
SPENT SHALE	6974.8	0.794	0.00	0.00	0.00	0.00	0.00	0.00	0.380	0.267	0.311
-BIN	6752.5	0.769	7.63	0.08	0.23	0.79	91.71	0.28	0.358	0.239	0.295
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	158.9	0.018	12.68	1.10	0.73	1.36	83.66	0.65	0.014	0.018	0.012
-A/C	63.4	0.007	18.83	1.05	1.04	1.30	76.02	0.95	0.008	0.010	0.005
OIL	706.8	0.080	0.00	0.00	0.00	0.00	0.00	0.00	0.402	0.110	0.070
-A/C	87.4	0.010	86.82	9.64	1.42	2.11	0.00	0.00	0.053	0.019	0.010
-ESP	398.2	0.045	79.32	8.81	1.30	1.93	3.80	1.13	0.219	0.080	0.042
-LP OIL	157.3	0.018	83.98	11.32	0.47	2.05	0.00	0.32	0.092	0.011	0.018
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	63.9	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.039	0.000	0.000
GAS	365.5	0.042	0.00	0.00	0.00	0.00	0.00	0.00	0.086	0.000	0.421
-CH4	31.6	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	19.7	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	16.0	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	15.1	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	124.4	0.014	27.30	0.00	0.00	0.00	0.00	0.00	0.024	0.000	0.000
-CO	55.1	0.006	42.90	0.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-H2	22.7	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	80.9	0.009	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.421
WATER	6019.5	0.685	0.00	0.00	0.00	0.00	0.00	0.00	0.016	0.161	0.096
-LIQUID	6019.5	0.685	0.38	0.00	0.17	0.29	0.00	100.00	0.016	0.161	0.096
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	14066.6	1.601	0.00	0.00	0.00	0.00	0.00	0.00	0.884	0.538	0.898

OVERALL AND COMPONENT BALANCES											
KEN#35	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	8778.2	1.000	16.38	1.87	0.74	2.03	74.89	1.95	1.000	1.000	1.000
SPENT SHALE	6873.0	0.783	0.00	0.00	0.00	0.00	0.00	0.00	0.320	0.193	0.255
-BIN	6651.4	0.758	6.39	0.08	0.16	0.64	93.36	0.27	0.296	0.164	0.239
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-CYCLONE	155.6	0.018	14.94	0.88	0.79	1.38	81.07	1.22	0.016	0.019	0.012
-A/C	66.0	0.008	16.92	0.95	0.98	1.22	78.71	0.54	0.008	0.010	0.005
OIL	688.5	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.391	0.107	0.066
-A/C	84.7	0.010	86.84	9.65	1.42	2.09	0.00	0.00	0.051	0.019	0.010
-ESP	394.4	0.045	78.49	8.72	1.28	1.89	3.60	2.41	0.215	0.078	0.042
-LP OIL	140.3	0.016	84.57	11.70	0.48	1.82	0.00	0.19	0.083	0.010	0.014
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	69.1	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
GAS	624.3	0.071	0.00	0.00	0.00	0.00	0.00	0.00	0.144	0.000	0.411
-CH4	36.9	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-C2'S	21.4	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	17.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	15.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	271.9	0.031	27.30	0.00	0.00	0.00	0.00	0.00	0.052	0.000	0.000
-CO	140.7	0.016	42.90	0.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
-H2	42.4	0.005	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	77.8	0.009	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.411
WATER	5644.1	0.643	0.00	0.00	0.00	0.00	0.00	0.00	0.012	0.214	0.101
-LIQUID	5644.1	0.643	0.31	0.00	0.25	0.32	0.00	100.00	0.012	0.214	0.101
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	13829.9	1.575	0.00	0.00	0.00	0.00	0.00	0.00	0.867	0.513	0.834

Runs

OVERALL AND COMPONENT BALANCES											
KEN#36	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7665.1	1.000	16.21	1.84	0.74	2.03	74.89	1.75	1.000	1.000	1.000
SPENT SHALE	6143.4	0.801	0.00	0.00	0.00	0.00	74.95	0.00	0.368	0.244	0.288
-BIN	5897.2	0.769	7.37	0.09	0.21	0.73	93.36	0.21	0.350	0.218	0.277
-TUBE	0.0	0.000	0.00	0.00	0.00	0.00	92.21	0.00	0.000	0.000	0.000
-CYCLONE	189.7	0.025	6.30	0.06	0.51	0.58	81.07	0.18	0.010	0.017	0.007
-A/C	56.5	0.007	18.02	1.07	0.89	1.25	93.73	0.57	0.008	0.009	0.005
OIL	624.8	0.082	0.00	0.00	0.00	0.00	77.48	0.00	0.401	0.109	0.067
-A/C	73.9	0.010	86.80	9.83	1.40	1.97	0.00	0.00	0.052	0.018	0.009
-ESP	381.7	0.050	76.18	8.63	1.23	1.73	4.00	4.67	0.234	0.083	0.042
-LP OIL	111.2	0.015	84.01	11.99	0.42	2.14	0.00	0.17	0.075	0.008	0.015
	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-C5+ VAPOR	58.0	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.041	0.000	0.000
GAS	392.5	0.051	0.00	0.00	0.00	0.00	0.00	0.00	0.104	0.000	0.453
-CH4	28.4	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	17.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	13.9	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	16.6	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-CO2	153.1	0.020	27.30	0.00	0.00	0.00	0.00	0.00	0.034	0.000	0.000
-CO	61.8	0.008	42.90	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-H2	26.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	74.9	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.453
WATER	5360.7	0.699	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.177	0.074
-LIQUID	5360.7	0.699	0.42	0.00	0.19	0.21	0.00	100.00	0.018	0.177	0.074
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12521.4	1.634	0.00	0.00	0.00	0.00	0.00	0.00	0.891	0.531	0.882

OVERALL AND COMPONENT BALANCES											
KEN#46	WEIGHT,G	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7521.4	1.000	16.10	2.00	0.65	1.79	74.20	2.71	1.000	1.000	1.000
SPENT SHALE	5783.9	0.769	0.00	0.00	0.00	0.00	0.00	0.00	0.280	0.460	0.207
-BIN	5567.1	0.740	5.63	0.16	0.38	0.46	93.50	0.54	0.259	0.433	0.190
-PYR FINES	166.5	0.022	12.56	0.91	0.64	1.13	81.86	1.24	0.017	0.022	0.014
-COMB FINES	28.1	0.004	5.75	0.22	0.42	0.56	92.13	1.12	0.001	0.002	0.001
-A/C FINES	22.2	0.003	12.56	0.91	0.64	1.13	81.86	1.24	0.002	0.003	0.002
OIL	584.7	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.398	0.101	0.086
-A/C	89.5	0.012	87.12	9.40	1.14	2.34	0.00	0.00	0.064	0.021	0.016
-ESP	278.4	0.037	78.97	8.52	1.03	2.12	3.70	2.25	0.181	0.059	0.041
-LITE OIL	164.7	0.022	84.74	11.30	0.63	2.21	0.00	0.53	0.115	0.021	0.021
-C5+ VAPOR	52.1	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
PYR GAS	430.0	0.057	0.00	0.00	0.00	0.00	0.00	0.00	0.109	0.000	0.571
-CH4	23.0	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	15.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	12.5	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	10.7	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	211.6	0.028	27.30	0.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
-CO	58.0	0.008	42.90	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-H2	17.0	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	81.7	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.571
COMB GAS	578.2	0.077	0.00	0.00	0.00	0.00	0.00	0.00	0.130	0.014	0.076
-CO	54.6	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-CO2	490.9	0.065	27.27	0.00	0.00	0.00	0.00	0.00	0.111	0.000	0.000
-SO2	20.4	0.003	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.076
-O2	10.8	0.001	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	1.5	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.014	0.000
WATER	6058.5	0.806	0.00	0.00	0.00	0.00	0.00	0.00	0.029	0.219	0.054
-LIQUID	6058.5	0.806	0.58	0.00	0.18	0.12	0.00	100.00	0.029	0.219	0.054
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	13435.3	1.786	0.00	0.00	0.00	0.00	0.00	0.00	0.946	0.794	0.995

OVERALL AND COMPONENT BALANCES											
KEN#47	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7803.6	1.000	16.11	2.02	0.65	1.79	74.07	2.91	1.000	1.000	1.000
SPENT SHALE	6037.3	0.774	0.00	0.00	0.00	0.00	0.00	0.00	0.295	0.457	0.212
-BIN	5163.1	0.662	5.89	0.12	0.38	0.48	93.51	0.42	0.242	0.387	0.177
-PYR FINES	168.3	0.022	16.00	1.17	0.67	1.09	78.73	1.98	0.021	0.022	0.013
-COMB FINES	679.6	0.087	5.19	0.12	0.33	0.40	94.23	0.43	0.028	0.044	0.020
-A/C FINES	26.3	0.003	16.00	1.17	0.67	1.09	78.73	1.98	0.003	0.003	0.002
OIL	581.0	0.075	0.00	0.00	0.00	0.00	0.00	0.00	0.383	0.097	0.079
-A/C	86.6	0.011	87.21	9.51	1.09	2.19	0.00	0.00	0.060	0.019	0.014
-ESP	309.9	0.040	79.88	8.71	1.00	2.01	2.40	3.29	0.197	0.061	0.045
-LITE OIL	133.0	0.017	85.09	11.37	0.65	2.18	0.00	0.20	0.090	0.017	0.021
-C5+ VAPOR	51.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
PYR GAS	474.9	0.061	0.00	0.00	0.00	0.00	0.00	0.00	0.115	0.000	0.584
-CH4	23.5	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	15.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	11.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	10.0	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	228.0	0.029	27.30	0.00	0.00	0.00	0.00	0.00	0.050	0.000	0.000
-CO	80.3	0.010	42.90	0.00	0.00	0.00	0.00	0.00	0.027	0.000	0.000
-H2	19.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	86.7	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.584
COMB GAS	462.3	0.059	0.00	0.00	0.00	0.00	0.00	0.00	0.102	0.007	0.054
-CO	59.6	0.008	42.86	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO2	376.9	0.048	27.27	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.000
-SO2	15.1	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.054
-O2	9.9	0.001	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.8	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.007	0.000
WATER	5817.7	0.746	0.00	0.00	0.00	0.00	0.00	0.00	0.019	0.182	0.025
-LIQUID	5817.7	0.746	0.40	0.00	0.16	0.06	0.00	100.00	0.019	0.182	0.025
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	13373.2	1.714	0.00	0.00	0.00	0.00	0.00	0.00	0.913	0.743	0.954

Runs

OVERALL AND COMPONENT BALANCES											
KEN#48	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	8199.1	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	6383.0	0.779	0.00	0.00	0.00	0.00	0.00	0.00	0.296	0.479	0.226
-BIN	5596.9	0.683	5.93	0.15	0.40	0.50	93.29	0.38	0.251	0.420	0.191
-PYR FINES	185.6	0.023	10.74	0.85	0.50	1.04	84.11	2.26	0.015	0.017	0.013
-COMB FINES	573.0	0.070	6.21	0.12	0.36	0.52	93.23	0.27	0.027	0.039	0.020
-A/C FINES	27.5	0.003	10.74	0.85	0.50	1.04	84.11	2.26	0.002	0.003	0.002
OIL	635.6	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.395	0.115	0.078
-A/C	112.6	0.014	86.88	9.65	1.28	2.19	0.00	0.00	0.074	0.027	0.017
-ESP	337.3	0.041	78.06	8.67	1.15	1.97	2.60	4.44	0.199	0.073	0.045
-LITE OIL	119.4	0.015	85.71	11.63	0.68	2.00	0.00	0.27	0.078	0.015	0.016
-C5+ VAPOR	66.3	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.044	0.000	0.000
PYR GAS	474.6	0.058	0.00	0.00	0.00	0.00	0.00	0.00	0.111	0.000	0.722
-CH4	30.3	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	20.9	0.003	81.80	18.20	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-C3'S	17.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C4'S	14.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	202.3	0.025	27.30	0.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
-CO	59.7	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-H2	17.4	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	112.6	0.014	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.722
COMB GAS	580.8	0.071	0.00	0.00	0.00	0.00	0.00	0.00	0.122	0.009	0.061
-CO	70.1	0.009	42.86	0.00	0.00	0.00	0.00	0.00	0.023	0.000	0.000
-CO2	479.2	0.058	27.27	0.00	0.00	0.00	0.00	0.00	0.099	0.000	0.000
-SO2	17.8	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.061
-O2	12.7	0.002	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	1.0	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.009	0.000
WATER	6866.3	0.837	0.00	0.00	0.00	0.00	0.00	0.00	0.040	0.234	0.014
-LIQUID	6866.3	0.837	0.77	0.00	0.18	0.03	0.00	100.00	0.040	0.234	0.014
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	14940.3	1.822	0.00	0.00	0.00	0.00	0.00	0.00	0.963	0.836	1.101

OVERALL AND COMPONENT BALANCES											
KEN#49	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6911.9	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5388.1	0.780	0.00	0.00	0.00	0.00	0.00	0.00	0.301	0.673	0.216
-BIN	5193.5	0.751	6.09	0.17	0.56	0.48	93.36	0.42	0.284	0.647	0.202
-PYR FINES	137.3	0.020	10.46	0.90	0.60	1.08	84.32	2.12	0.013	0.018	0.012
-COMB FINES	40.7	0.006	6.09	0.17	0.56	0.48	93.23	0.42	0.002	0.005	0.002
-A/C FINES	16.6	0.002	10.46	0.90	0.60	1.08	84.32	2.12	0.002	0.002	0.001
OIL	577.8	0.084	0.00	0.00	0.00	0.00	0.00	0.00	0.436	0.132	0.089
-A/C	95.0	0.014	87.16	9.35	1.32	2.17	0.00	0.00	0.074	0.028	0.017
-ESP	305.5	0.044	81.72	8.77	1.24	2.03	1.40	3.76	0.224	0.084	0.050
-LITE OIL	131.8	0.019	85.74	11.55	0.67	2.10	0.00	0.25	0.102	0.020	0.022
-C5+ VAPOR	45.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
PYR GAS	418.3	0.061	0.00	0.00	0.00	0.00	0.00	0.00	0.114	0.000	0.563
-CH4	20.3	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	14.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	11.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	10.5	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	224.9	0.033	27.30	0.00	0.00	0.00	0.00	0.00	0.055	0.000	0.000
-CO	48.3	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-H2	14.6	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	74.0	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.563
COMB GAS	432.9	0.063	0.00	0.00	0.00	0.00	0.00	0.00	0.108	0.009	0.059
-CO	43.6	0.006	42.86	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-CO2	371.5	0.054	27.27	0.00	0.00	0.00	0.00	0.00	0.091	0.000	0.000
-SO2	14.7	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.059
-O2	2.2	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.9	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.009	0.000
WATER	5553.3	0.803	0.00	0.00	0.00	0.00	0.00	0.00	0.024	0.208	0.009
-LIQUID	5553.3	0.803	0.49	0.00	0.17	0.02	0.00	100.00	0.024	0.208	0.009
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12370.4	1.790	0.00	0.00	0.00	0.00	0.00	0.00	0.983	1.022	0.937

OVERALL AND COMPONENT BALANCES											
KEN#50	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	5136.1	1.000	16.11	2.03	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	3885.5	0.756	0.00	0.00	0.00	0.00	0.00	0.00	0.271	0.563	0.181
-BIN	3048.2	0.594	5.51	0.18	0.47	0.39	93.88	0.30	0.203	0.429	0.129
-PYR FINES	168.3	0.033	10.58	0.72	0.68	0.97	85.54	1.39	0.022	0.034	0.018
-COMB FINES	643.0	0.125	5.57	0.13	0.50	0.45	93.81	0.43	0.043	0.096	0.032
-A/C FINES	26.0	0.005	10.58	0.72	0.47	0.97	85.54	1.39	0.003	0.004	0.003
OIL	395.9	0.077	0.00	0.00	0.00	0.00	0.00	0.00	0.386	0.104	0.078
-A/C	52.5	0.010	86.59	9.96	1.24	2.21	0.00	0.00	0.055	0.020	0.013
-ESP	216.1	0.042	76.56	8.81	1.10	1.95	4.40	6.19	0.200	0.071	0.046
-LITE OIL	88.8	0.017	84.86	11.37	0.51	2.04	0.00	0.29	0.091	0.014	0.020
-C5+ VAPOR	38.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.040	0.000	0.000
PYR GAS	300.5	0.059	0.00	0.00	0.00	0.00	0.00	0.00	0.114	0.000	0.623
-CH4	17.6	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-C2'S	12.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	9.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	8.5	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	125.8	0.025	27.30	0.00	0.00	0.00	0.00	0.00	0.042	0.000	0.000
-CO	50.4	0.010	42.90	0.00	0.00	0.00	0.00	0.00	0.026	0.000	0.000
-H2	15.4	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	60.9	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.623
COMB GAS	313.6	0.061	0.00	0.00	0.00	0.00	0.00	0.00	0.105	0.007	0.052
-CO	38.5	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO2	257.6	0.050	27.27	0.00	0.00	0.00	0.00	0.00	0.085	0.000	0.000
-SO2	9.6	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.052
-O2	7.4	0.001	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.5	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.007	0.000
WATER	4228.7	0.823	0.00	0.00	0.00	0.00	0.00	0.00	0.024	0.243	0.014
-LIQUID	4228.7	0.823	0.47	0.00	0.19	0.03	0.00	100.00	0.024	0.243	0.014
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	9124.2	1.777	0.00	0.00	0.00	0.00	0.00	0.00	0.901	0.918	0.949

Runs

OVERALL AND COMPONENT BALANCES											
KEN#51	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6791.1	1.000	16.11	2.03	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5260.8	0.775	0.00	0.00	0.00	0.00	0.00	0.00	0.289	0.524	0.175
-BIN	5082.4	0.748	5.79	0.17	0.43	0.38	93.94	0.38	0.269	0.495	0.159
-PYR FINES	157.2	0.023	12.58	0.97	0.71	1.12	82.82	1.62	0.018	0.025	0.014
-COMB FINES	2.7	0.000	5.79	0.12	0.43	0.38	93.94	0.38	0.000	0.000	0.000
-A/C FINES	18.5	0.003	12.58	0.97	0.71	1.12	82.82	1.62	0.002	0.003	0.002
OIL	558.7	0.082	0.00	0.00	0.00	0.00	0.00	0.00	0.432	0.112	0.089
-A/C	133.0	0.020	87.15	9.55	1.11	2.20	0.00	0.00	0.106	0.033	0.024
-ESP	272.5	0.040	82.44	9.03	1.05	2.08	3.10	3.60	0.205	0.065	0.047
-LITE OIL	106.1	0.016	85.36	11.42	0.56	2.06	0.00	0.21	0.083	0.014	0.018
-C5+ VAPOR	47.1	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
PYR GAS	465.1	0.069	0.00	0.00	0.00	0.00	0.00	0.00	0.133	0.000	0.553
-CH4	22.1	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	14.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	11.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	9.9	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	214.1	0.032	27.30	0.00	0.00	0.00	0.00	0.00	0.053	0.000	0.000
-CO	95.5	0.014	42.90	0.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
-H2	26.3	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	71.5	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.553
COMB GAS	283.1	0.042	0.00	0.00	0.00	0.00	0.00	0.00	0.072	0.005	0.033
-CO	46.2	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-CO2	216.4	0.032	27.27	0.00	0.00	0.00	0.00	0.00	0.054	0.000	0.000
-SO2	8.1	0.001	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.033
-O2	11.9	0.002	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.5	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.005	0.000
WATER	4968.8	0.732	0.00	0.00	0.00	0.00	0.00	0.00	0.029	0.225	0.020
-LIQUID	4968.8	0.732	0.65	0.00	0.20	0.05	0.00	100.00	0.029	0.225	0.020
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11536.5	1.699	0.00	0.00	0.00	0.00	0.00	0.00	0.956	0.866	0.871

OVERALL AND COMPONENT BALANCES											
KEN#52	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6817.8	1.000	16.11	1.90	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5278.3	0.774	0.00	0.00	0.00	0.00	0.00	0.00	0.281	0.519	0.205
-BIN	4707.1	0.690	5.50	0.12	0.42	0.42	94.07	0.42	0.236	0.446	0.162
-PYR FINES	316.3	0.046	10.70	0.92	0.64	1.22	84.23	1.22	0.031	0.046	0.032
-COMB FINES	217.3	0.032	5.28	0.13	0.45	0.42	94.22	0.48	0.010	0.022	0.007
-A/C FINES	37.6	0.006	10.70	0.92	0.64	1.22	84.23	1.22	0.004	0.005	0.004
OIL	519.6	0.076	0.00	0.00	0.00	0.00	0.00	0.00	0.392	0.106	0.080
-A/C	79.3	0.012	86.79	9.78	1.22	2.22	0.00	0.00	0.063	0.022	0.014
-ESP	270.2	0.040	80.26	9.04	1.13	2.05	3.60	4.93	0.197	0.069	0.045
-LITE OIL	123.0	0.018	84.80	11.06	0.54	1.96	0.00	0.70	0.095	0.015	0.020
-C5+ VAPOR	47.1	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
PYR GAS	445.8	0.065	0.00	0.00	0.00	0.00	0.00	0.00	0.123	0.000	0.631
-CH4	21.5	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	14.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	11.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	9.3	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	206.0	0.030	27.30	0.00	0.00	0.00	0.00	0.00	0.051	0.000	0.000
-CO	78.4	0.012	42.90	0.00	0.00	0.00	0.00	0.00	0.031	0.000	0.000
-H2	22.6	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	81.8	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.631
COMB GAS	421.5	0.062	0.00	0.00	0.00	0.00	0.00	0.00	0.107	0.010	0.062
-CO	49.4	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-CO2	354.6	0.052	27.27	0.00	0.00	0.00	0.00	0.00	0.088	0.000	0.000
-SO2	15.0	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.062
-O2	1.6	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.9	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.010	0.000
WATER	5335.0	0.783	0.00	0.00	0.00	0.00	0.00	0.00	0.019	0.228	0.002
-LIQUID	5335.0	0.783	0.39	0.00	0.19	0.01	0.00	100.00	0.019	0.228	0.002
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12000.2	1.760	0.00	0.00	0.00	0.00	0.00	0.00	0.923	0.862	0.979

OVERALL AND COMPONENT BALANCES											
KEN#53	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6228.1	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	4604.2	0.739	0.00	0.00	0.00	0.00	0.00	0.00	0.248	0.420	0.178
-BIN	4050.5	0.650	5.00	0.10	0.36	0.37	94.84	0.30	0.202	0.360	0.134
-PYR FINES	297.4	0.048	10.68	0.86	0.52	1.18	84.76	1.32	0.032	0.038	0.032
-COMB FINES	223.6	0.036	4.83	0.08	0.32	0.45	94.64	0.35	0.011	0.018	0.009
-A/C FINES	32.7	0.005	10.68	0.86	0.52	1.18	84.76	1.32	0.003	0.004	0.003
OIL	462.6	0.074	0.00	0.00	0.00	0.00	0.00	0.00	0.380	0.125	0.075
-A/C	66.3	0.011	86.95	9.33	1.54	2.18	0.00	0.00	0.058	0.025	0.013
-ESP	247.1	0.040	80.23	8.61	1.42	2.01	3.55	4.28	0.198	0.087	0.045
-LITE OIL	105.2	0.017	82.89	11.10	0.51	1.80	0.00	0.30	0.087	0.013	0.017
-C5+ VAPOR	44.0	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.038	0.000	0.000
PYR GAS	425.8	0.068	0.00	0.00	0.00	0.00	0.00	0.00	0.130	0.000	0.581
-CH4	20.1	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	13.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	10.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	8.8	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	209.0	0.034	27.30	0.00	0.00	0.00	0.00	0.00	0.057	0.000	0.000
-CO	72.9	0.012	42.90	0.00	0.00	0.00	0.00	0.00	0.031	0.000	0.000
-H2	22.2	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	68.8	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.581
COMB GAS	372.7	0.060	0.00	0.00	0.00	0.00	0.00	0.00	0.099	0.008	0.051
-CO	44.4	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-CO2	293.0	0.047	27.27	0.00	0.00	0.00	0.00	0.00	0.080	0.000	0.000
-SO2	11.3	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.051
-O2	23.3	0.004	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.7	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.008	0.000
WATER	5054.2	0.812	0.00	0.00	0.00	0.00	0.00	0.00	0.025	0.226	0.095
-LIQUID	5054.2	0.812	0.49	0.00	0.18	0.21	0.00	100.00	0.025	0.226	0.095
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	10919.5	1.753	0.00	0.00	0.00	0.00	0.00	0.00	0.881	0.779	0.980

Runs

OVERALL AND COMPONENT BALANCES											
KEN#54	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6660.6	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5000.2	0.751	0.00	0.00	0.00	0.00	0.00	0.00	0.271	0.486	0.217
-BIN	4658.6	0.699	5.44	0.12	0.41	0.47	94.01	1.05	0.236	0.441	0.184
-PYR FINES	286.1	0.043	11.41	0.97	0.58	1.23	83.63	1.33	0.030	0.038	0.029
-COMB FINES	28.8	0.004	5.44	0.12	0.41	0.47	94.01	1.05	0.002	0.003	0.001
-A/C FINES	26.7	0.004	11.41	0.97	0.58	1.23	83.63	1.33	0.003	0.004	0.003
OIL	524.0	0.079	0.00	0.00	0.00	0.00	0.00	0.00	0.388	0.126	0.078
-A/C	78.8	0.012	86.24	10.13	1.42	2.20	0.00	0.00	0.063	0.026	0.015
-ESP	300.6	0.045	75.16	8.83	1.24	1.92	5.47	8.22	0.211	0.086	0.048
-LITE OIL	108.3	0.016	84.20	11.09	0.55	1.69	0.00	0.23	0.085	0.014	0.015
-C5+ VAPOR	36.3	0.005	87.00	13.00	0.00	0.00	0.00	0.00	0.029	0.000	0.000
PYR GAS	383.8	0.058	0.00	0.00	0.00	0.00	0.00	0.00	0.107	0.000	0.607
-CH4	17.8	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C2'S	12.2	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	10.0	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	9.1	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	190.3	0.029	27.30	0.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
-CO	54.5	0.008	42.90	0.00	0.00	0.00	0.00	0.00	0.022	0.000	0.000
-H2	13.0	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	76.9	0.012	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.607
COMB GAS	466.8	0.070	0.00	0.00	0.00	0.00	0.00	0.00	0.120	0.009	0.075
-CO	39.1	0.006	42.86	0.00	0.00	0.00	0.00	0.00	0.016	0.000	0.000
-CO2	409.0	0.061	27.27	0.00	0.00	0.00	0.00	0.00	0.104	0.000	0.000
-SO2	17.8	0.003	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.075
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.9	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.009	0.000
WATER	5003.1	0.751	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.198	0.084
-LIQUID	5003.1	0.751	0.38	0.00	0.17	0.20	0.00	100.00	0.018	0.198	0.084
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11377.9	1.708	0.00	0.00	0.00	0.00	0.00	0.00	0.903	0.819	1.061

OVERALL AND COMPONENT BALANCES											
KEN#55	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	3230.9	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	2515.9	0.779	0.00	0.00	0.00	0.00	0.00	0.00	0.259	0.479	0.199
-BIN	2341.3	0.725	4.81	0.05	0.38	0.41	94.49	0.41	0.216	0.424	0.166
-PYR FINES	164.9	0.051	12.74	0.94	0.66	1.08	83.21	0.66	0.040	0.052	0.031
-COMB FINES	0.0	0.000	4.52	0.21	0.58	0.53	93.43	0.81	0.000	0.000	0.000
-A/C FINES	9.7	0.003	12.74	0.94	0.66	1.08	83.21	0.66	0.002	0.003	0.002
OIL	273.3	0.085	0.00	0.00	0.00	0.00	0.00	0.00	0.402	0.141	0.083
-A/C	30.4	0.009	85.58	10.58	1.53	2.31	0.00	0.00	0.050	0.022	0.012
-ESP	153.2	0.047	70.43	8.71	1.26	1.90	8.40	8.85	0.207	0.092	0.050
-LITE OIL	72.4	0.022	83.39	10.28	0.79	1.67	0.00	0.48	0.116	0.027	0.021
-C5+ VAPOR	17.3	0.005	87.00	13.00	0.00	0.00	0.00	0.00	0.029	0.000	0.000
PYR GAS	206.6	0.064	0.00	0.00	0.00	0.00	0.00	0.00	0.121	0.000	0.521
-CH4	8.8	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-C2'S	5.6	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	4.4	0.001	83.70	16.30	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-C4'S	3.6	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.006	0.000	0.000
-CO2	105.7	0.033	27.30	0.00	0.00	0.00	0.00	0.00	0.055	0.000	0.000
-CO	37.8	0.012	42.90	0.00	0.00	0.00	0.00	0.00	0.031	0.000	0.000
-H2	8.7	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	32.0	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.521
COMB GAS	183.0	0.057	0.00	0.00	0.00	0.00	0.00	0.00	0.098	0.007	0.055
-CO	20.0	0.006	42.86	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-CO2	156.3	0.048	27.27	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.000
-SO2	6.4	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.055
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.3	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.007	0.000
WATER	2535.5	0.785	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.000	0.053
-LIQUID	2535.5	0.785	0.36	0.00	0.00	0.12	0.00	100.00	0.018	0.000	0.053
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	5714.3	1.769	0.00	0.00	0.00	0.00	0.00	0.00	0.898	0.627	0.911

OVERALL AND COMPONENT BALANCES											
KEN#56	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7251.9	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5682.5	0.784	0.00	0.00	0.00	0.00	0.00	0.00	0.269	0.499	0.200
-BIN	4878.3	0.673	5.22	0.13	0.41	0.41	94.27	0.37	0.218	0.424	0.154
-PYR FINES	309.3	0.043	11.11	0.99	0.60	1.21	83.92	1.22	0.029	0.039	0.029
-COMB FINES	471.7	0.065	4.80	0.10	0.32	0.42	94.64	0.32	0.019	0.032	0.015
-A/C FINES	23.2	0.003	11.11	0.99	0.60	1.21	83.92	1.22	0.002	0.003	0.002
OIL	560.4	0.077	0.00	0.00	0.00	0.00	0.00	0.00	0.380	0.138	0.080
-A/C	85.0	0.012	86.34	9.84	1.65	2.17	0.00	0.00	0.063	0.030	0.014
-ESP	305.0	0.042	74.39	8.48	1.42	1.87	5.31	6.61	0.194	0.092	0.044
-LITE OIL	128.5	0.018	83.69	10.96	0.60	2.16	0.00	0.16	0.092	0.016	0.021
-C5+ VAPOR	41.9	0.006	87.00	13.00	0.00	0.00	0.00	0.00	0.031	0.000	0.000
PYR GAS	381.0	0.053	0.00	0.00	0.00	0.00	0.00	0.00	0.097	0.000	0.539
-CH4	18.4	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C2'S	12.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C3'S	10.0	0.001	83.70	16.30	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-C4'S	7.8	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.006	0.000	0.000
-CO2	199.5	0.027	27.30	0.00	0.00	0.00	0.00	0.00	0.047	0.000	0.000
-CO	47.1	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-H2	11.4	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	74.3	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.539
COMB GAS	614.3	0.085	0.00	0.00	0.00	0.00	0.00	0.00	0.144	0.011	0.083
-CO	45.3	0.006	42.86	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-CO2	546.4	0.075	27.27	0.00	0.00	0.00	0.00	0.00	0.127	0.000	0.000
-SO2	21.5	0.003	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.083
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	1.1	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.011	0.000
WATER	6001.4	0.828	0.00	0.00	0.00	0.00	0.00	0.00	0.018	0.212	0.009
-LIQUID	6001.4	0.828	0.34	0.00	0.17	0.02	0.00	100.00	0.018	0.212	0.009
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	13239.6	1.826	0.00	0.00	0.00	0.00	0.00	0.00	0.908	0.859	0.911

OVERALL AND COMPONENT BALANCES											
KEN#57	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6482.4	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	4826.8	0.745	0.00	0.00	0.00	0.00	0.00	0.00	0.251	0.422	0.170
-BIN	4314.1	0.666	5.08	0.07	0.36	0.37	94.57	0.30	0.210	0.369	0.138
-PYR FINES	327.7	0.051	9.58	0.65	0.46	0.92	87.24	0.88	0.030	0.036	0.026
-COMB FINES	136.5	0.021	5.00	0.07	0.40	0.40	94.43	0.34	0.007	0.013	0.005
-A/C FINES	48.5	0.007	9.58	0.07	0.36	0.37	87.24	0.88	0.004	0.004	0.002
OIL	537.7	0.083	0.00	0.00	0.00	0.00	0.00	0.00	0.412	0.133	0.078
-A/C	65.1	0.010	86.43	9.97	1.52	2.07	0.00	0.00	0.054	0.024	0.012
-ESP	302.2	0.047	76.27	8.80	1.34	1.83	3.24	4.97	0.221	0.096	0.048
-LITE OIL	114.5	0.018	83.14	11.07	0.51	1.88	0.00	0.28	0.091	0.014	0.019
-C5+ VAPOR	55.9	0.009	87.00	13.00	0.00	0.00	0.00	0.00	0.047	0.000	0.000
PYR GAS	467.9	0.072	0.00	0.00	0.00	0.00	0.00	0.00	0.142	0.000	0.590
-CH4	24.9	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.018	0.000	0.000
-C2'S	15.9	0.003	81.80	18.20	0.00	0.00	0.00	0.00	0.013	0.000	0.000
-C3'S	12.6	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	11.4	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	219.1	0.034	27.30	0.00	0.00	0.00	0.00	0.00	0.057	0.000	0.000
-CO	85.7	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
-H2	25.6	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	72.7	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.590
COMB GAS	418.4	0.065	0.00	0.00	0.00	0.00	0.00	0.00	0.113	0.009	0.060
-CO	48.0	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO2	355.6	0.055	27.27	0.00	0.00	0.00	0.00	0.00	0.093	0.000	0.000
-SO2	14.0	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.060
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.8	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.009	0.000
WATER	5090.0	0.785	0.00	0.00	0.00	0.00	0.00	0.00	0.017	0.236	0.009
-LIQUID	5090.0	0.785	0.35	0.00	0.20	0.02	0.00	100.00	0.017	0.236	0.009
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11340.8	1.750	0.00	0.00	0.00	0.00	0.00	0.00	0.935	0.800	0.906

OVERALL AND COMPONENT BALANCES											
KEN#58	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	7051.0	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5603.0	0.795	0.00	0.00	0.00	0.00	0.00	0.00	0.273	0.488	0.199
-BIN	5072.5	0.719	5.03	0.08	0.38	0.40	94.46	0.29	0.225	0.421	0.161
-PYR FINES	345.1	0.049	12.63	0.89	0.66	1.10	83.42	0.59	0.038	0.050	0.030
-COMB FINES	147.7	0.021	4.78	0.06	0.37	0.38	94.71	0.25	0.006	0.012	0.004
-A/C FINES	37.7	0.005	12.63	0.89	0.66	1.10	83.42	0.59	0.004	0.005	0.003
OIL	540.0	0.077	0.00	0.00	0.00	0.00	0.00	0.00	0.396	0.136	0.084
-A/C	66.3	0.009	86.58	9.47	1.54	2.41	0.00	0.00	0.051	0.022	0.013
-ESP	300.6	0.043	81.03	8.86	1.44	2.26	1.27	5.19	0.214	0.094	0.054
-LITE OIL	125.1	0.018	85.97	11.52	0.70	1.72	0.00	0.33	0.095	0.019	0.017
-C5+ VAPOR	48.0	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.037	0.000	0.000
PYR GAS	538.9	0.076	0.00	0.00	0.00	0.00	0.00	0.00	0.141	0.000	0.802
-CH4	25.4	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	16.5	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	13.4	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	12.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	246.1	0.035	27.30	0.00	0.00	0.00	0.00	0.00	0.059	0.000	0.000
-CO	90.2	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.034	0.000	0.000
-H2	27.8	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	107.5	0.015	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.802
COMB GAS	489.2	0.069	0.00	0.00	0.00	0.00	0.00	0.00	0.122	0.009	0.056
-CO	55.8	0.008	42.86	0.00	0.00	0.00	0.00	0.00	0.021	0.000	0.000
-CO2	418.3	0.059	27.27	0.00	0.00	0.00	0.00	0.00	0.100	0.000	0.000
-SO2	14.2	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.056
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.9	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.009	0.000
WATER	5721.4	0.811	0.00	0.00	0.00	0.00	0.00	0.00	0.016	0.000	0.082
-LIQUID	5721.4	0.811	0.31	0.00	0.00	0.18	0.00	100.00	0.016	0.000	0.082
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12892.5	1.829	0.00	0.00	0.00	0.00	0.00	0.00	0.948	0.633	1.222

OVERALL AND COMPONENT BALANCES											
KEN#59	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6966.3	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5296.0	0.760	0.00	0.00	0.00	0.00	0.00	0.00	0.262	0.368	0.216
-BIN	4806.5	0.690	5.15	0.08	0.30	0.47	94.28	0.25	0.221	0.318	0.181
-PYR FINES	262.4	0.038	11.98	0.90	0.52	1.21	83.52	0.73	0.028	0.030	0.026
-COMB FINES	196.1	0.028	4.64	0.06	0.29	0.41	94.52	0.34	0.008	0.013	0.006
-A/C FINES	31.0	0.004	18.67	1.39	0.93	1.17	74.93	0.70	0.005	0.006	0.003
OIL	546.1	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.401	0.142	0.073
-A/C	76.5	0.011	86.24	10.15	1.64	1.98	0.00	0.00	0.059	0.028	0.012
-ESP	289.7	0.042	78.88	9.28	1.50	1.81	3.38	3.66	0.204	0.096	0.042
-LITE OIL	124.9	0.018	85.87	11.54	0.68	1.92	0.00	0.26	0.096	0.019	0.019
-C5+ VAPOR	55.0	0.008	87.00	13.00	0.00	0.00	0.00	0.00	0.043	0.000	0.000
PYR GAS	438.7	0.063	0.00	0.00	0.00	0.00	0.00	0.00	0.121	0.000	0.660
-CH4	24.9	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	17.0	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	13.6	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	11.0	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	196.8	0.028	27.30	0.00	0.00	0.00	0.00	0.00	0.048	0.000	0.000
-CO	66.1	0.010	42.90	0.00	0.00	0.00	0.00	0.00	0.025	0.000	0.000
-H2	21.9	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	87.4	0.013	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.660
COMB GAS	556.5	0.080	0.00	0.00	0.00	0.00	0.00	0.00	0.138	0.010	0.069
-CO	52.0	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO2	486.2	0.070	27.27	0.00	0.00	0.00	0.00	0.00	0.118	0.000	0.000
-SO2	17.3	0.003	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.069
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	1.0	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.010	0.000
WATER	5507.8	0.791	0.00	0.00	0.00	0.00	0.00	0.00	0.023	0.223	0.022
-LIQUID	5507.8	0.791	0.46	0.00	0.18	0.05	0.00	100.00	0.023	0.223	0.022
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12345.1	1.772	0.00	0.00	0.00	0.00	0.00	0.00	0.944	0.744	1.040

OVERALL AND COMPONENT BALANCES											
Ken#60	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6939.1	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5388.9	0.777	0.00	0.00	0.00	0.00	0.00	0.00	0.268	0.421	0.220
-BIN	5047.4	0.727	5.14	0.07	0.34	0.46	94.57	0.22	0.232	0.381	0.187
-PYR FINES	291.3	0.042	12.21	0.76	0.54	1.24	83.45	0.66	0.032	0.035	0.029
-COMB FINES	20.8	0.003	4.96	0.09	0.35	0.55	94.07	0.50	0.001	0.002	0.001
-A/C FINES	29.4	0.004	12.21	0.76	0.54	1.24	83.45	0.66	0.003	0.003	0.003
OIL	525.8	0.076	0.00	0.00	0.00	0.00	0.00	0.00	0.388	0.144	0.072
-A/C	70.9	0.010	87.00	9.46	1.57	1.98	0.00	0.00	0.055	0.025	0.011
-ESP	300.4	0.043	80.02	8.70	1.44	1.82	5.08	3.31	0.215	0.096	0.044
-LITE OIL	108.9	0.016	84.50	11.79	0.95	1.89	0.00	0.26	0.082	0.023	0.017
-C5+ VAPOR	45.6	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.036	0.000	0.000
PYR GAS	540.1	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.151	0.000	0.588
-CH4	25.4	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	15.8	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	12.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	11.3	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-CO2	263.1	0.038	27.30	0.00	0.00	0.00	0.00	0.00	0.064	0.000	0.000
-CO	104.9	0.015	42.90	0.00	0.00	0.00	0.00	0.00	0.040	0.000	0.000
-H2	29.3	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	77.6	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.588
COMB GAS	331.1	0.048	0.00	0.00	0.00	0.00	0.00	0.00	0.086	0.005	0.039
-CO	52.6	0.008	42.86	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-CO2	268.4	0.039	27.27	0.00	0.00	0.00	0.00	0.00	0.066	0.000	0.000
-SO2	9.6	0.001	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.039
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.5	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.005	0.000
WATER	5162.8	0.744	0.00	0.00	0.00	0.00	0.00	0.00	0.019	0.246	0.025
-LIQUID	5162.8	0.744	0.41	0.00	0.21	0.06	0.00	100.00	0.019	0.246	0.025
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11948.7	1.722	0.00	0.00	0.00	0.00	0.00	0.00	0.912	0.815	0.943

Appendix B (Cont.) - Mass and Component Balances for Mini-Plant
Runs

OVERALL AND COMPONENT BALANCES											
KEN#61	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6628.5	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5163.3	0.779	0.00	0.00	0.00	0.00	0.00	0.00	0.275	0.359	0.224
-BIN	4779.6	0.721	5.21	0.09	0.28	0.47	94.40	0.26	0.233	0.311	0.189
-PYR FINES	309.3	0.047	12.49	0.94	0.57	1.15	83.56	0.64	0.036	0.041	0.030
-COMB FINES	38.2	0.006	5.02	0.07	0.30	0.52	94.26	0.31	0.002	0.003	0.002
-A/C FINES	36.2	0.006	12.49	0.94	0.57	1.15	83.56	0.64	0.004	0.005	0.003
OIL	502.3	0.076	0.00	0.00	0.00	0.00	0.00	0.00	0.388	0.123	0.073
-A/C	73.5	0.011	86.40	10.29	1.35	1.96	0.00	0.00	0.060	0.023	0.012
-ESP	278.7	0.042	80.20	9.55	1.25	1.82	2.94	3.94	0.209	0.081	0.043
-LITE OIL	112.8	0.017	84.49	11.78	0.74	1.91	0.00	0.23	0.089	0.019	0.018
-C5+ VAPOR	37.3	0.006	87.00	13.00	0.00	0.00	0.00	0.00	0.030	0.000	0.000
PYR GAS	467.0	0.071	0.00	0.00	0.00	0.00	0.00	0.00	0.135	0.000	0.577
-CH4	21.0	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	13.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.011	0.000	0.000
-C3'S	11.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	10.1	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	229.4	0.035	27.30	0.00	0.00	0.00	0.00	0.00	0.059	0.000	0.000
-CO	85.8	0.013	42.90	0.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
-H2	22.8	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	72.7	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.577
COMB GAS	378.6	0.057	0.00	0.00	0.00	0.00	0.00	0.00	0.100	0.008	0.047
-CO	46.7	0.007	42.86	0.00	0.00	0.00	0.00	0.00	0.019	0.000	0.000
-CO2	320.1	0.048	27.27	0.00	0.00	0.00	0.00	0.00	0.082	0.000	0.000
-SO2	11.1	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.047
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.7	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.008	0.000
WATER	5055.3	0.763	0.00	0.00	0.00	0.00	0.00	0.00	0.014	0.246	0.021
-LIQUID	5055.3	0.763	0.30	0.00	0.21	0.05	0.00	100.00	0.014	0.246	0.021
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	11566.5	1.745	0.00	0.00	0.00	0.00	0.00	0.00	0.914	0.736	0.942

OVERALL AND COMPONENT BALANCES											
KEN#62	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6530.3	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5102.1	0.781	0.00	0.00	0.00	0.00	0.00	0.00	0.248	0.304	0.181
-BIN	4619.1	0.707	4.57	0.07	0.23	0.36	95.23	0.19	0.201	0.250	0.142
-PYR FINES	312.3	0.048	11.94	1.06	0.49	1.19	83.64	0.79	0.035	0.036	0.032
-COMB FINES	137.9	0.021	4.27	0.06	0.31	0.32	95.15	0.42	0.006	0.010	0.004
-A/C FINES	32.8	0.005	19.24	1.53	0.98	1.12	75.14	0.78	0.006	0.008	0.003
OIL	508.3	0.078	0.00	0.00	0.00	0.00	0.00	0.00	0.400	0.131	0.074
-A/C	83.3	0.013	86.24	10.34	1.48	1.94	0.00	0.00	0.068	0.029	0.014
-ESP	278.2	0.043	80.56	9.66	1.38	1.81	3.00	4.45	0.213	0.090	0.043
-LITE OIL	98.5	0.015	84.03	11.44	0.50	2.00	0.00	0.33	0.079	0.012	0.017
-C5+ VAPOR	48.3	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.040	0.000	0.000
PYR GAS	498.5	0.076	0.00	0.00	0.00	0.00	0.00	0.00	0.146	0.000	0.669
-CH4	23.7	0.004	75.00	25.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-C2'S	15.9	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-C3'S	12.6	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C4'S	9.8	0.002	84.20	15.80	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-CO2	235.1	0.036	27.30	0.00	0.00	0.00	0.00	0.00	0.061	0.000	0.000
-CO	92.5	0.014	42.90	0.00	0.00	0.00	0.00	0.00	0.038	0.000	0.000
-H2	25.8	0.004	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	83.1	0.013	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.669
COMB GAS	529.9	0.081	0.00	0.00	0.00	0.00	0.00	0.00	0.139	0.011	0.062
-CO	40.5	0.006	42.86	0.00	0.00	0.00	0.00	0.00	0.017	0.000	0.000
-CO2	473.9	0.073	27.27	0.00	0.00	0.00	0.00	0.00	0.123	0.000	0.000
-SO2	14.5	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.062
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	1.0	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.011	0.000
WATER	5527.2	0.846	0.00	0.00	0.00	0.00	0.00	0.00	0.012	0.229	0.014
-LIQUID	5527.2	0.846	0.23	0.00	0.18	0.03	0.00	100.00	0.012	0.229	0.014
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12166.0	1.863	0.00	0.00	0.00	0.00	0.00	0.00	0.945	0.675	1.000

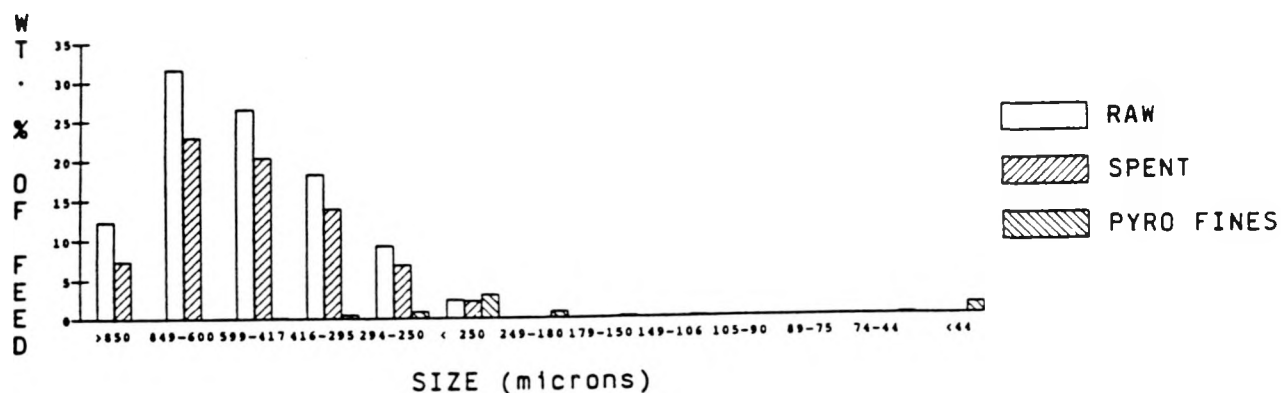
OVERALL AND COMPONENT BALANCES											
KEN#63	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6827.6	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5337.8	0.782	0.00	0.00	0.00	0.00	0.00	0.00	0.276	0.276	0.225
-BIN	5005.9	0.733	5.26	0.14	0.21	0.47	94.03	0.41	0.239	0.237	0.193
-PYR FINES	268.6	0.039	11.01	0.89	0.47	1.25	84.59	0.81	0.027	0.028	0.027
-COMB FINES	18.7	0.003	4.40	0.15	0.21	0.50	94.10	0.58	0.001	0.001	0.001
-A/C FINES	44.6	0.007	21.54	1.63	1.00	1.25	72.79	0.76	0.009	0.010	0.005
OIL	541.7	0.079	0.00	0.00	0.00	0.00	0.00	0.00	0.402	0.129	0.074
-A/C	72.2	0.011	86.47	10.16	1.42	1.95	0.00	0.00	0.057	0.023	0.012
-ESP	318.1	0.047	78.54	9.23	1.29	1.77	1.57	6.05	0.227	0.092	0.046
-LITE OIL	110.3	0.016	85.13	11.34	0.53	1.85	0.00	0.23	0.085	0.013	0.017
-C5+ VAPOR	41.1	0.006	87.00	13.00	0.00	0.00	0.00	0.00	0.033	0.000	0.000
PYR GAS	401.3	0.059	0.00	0.00	0.00	0.00	0.00	0.00	0.113	0.000	0.501
-CH4	20.5	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.014	0.000	0.000
-C2'S	13.3	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	10.7	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.008	0.000	0.000
-C4'S	8.8	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	220.3	0.032	27.30	0.00	0.00	0.00	0.00	0.00	0.055	0.000	0.000
-CO	50.5	0.007	42.90	0.00	0.00	0.00	0.00	0.00	0.020	0.000	0.000
-H2	12.2	0.002	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	65.0	0.010	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.501
COMB GAS	469.9	0.069	0.00	0.00	0.00	0.00	0.00	0.00	0.117	0.008	0.066
-CO	30.4	0.005	42.86	0.00	0.00	0.00	0.00	0.00	0.012	0.000	0.000
-CO2	422.7	0.062	27.27	0.00	0.00	0.00	0.00	0.00	0.105	0.000	0.000
-SO2	16.0	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.066
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.8	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.008	0.000
WATER	5301.3	0.777	0.00	0.00	0.00	0.00	0.00	0.00	0.011	0.160	0.009
-LIQUID	5301.3	0.777	0.23	0.00	0.13	0.02	0.00	100.00	0.011	0.160	0.009
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12052.0	1.765	0.00	0.00	0.00	0.00	0.00	0.00	0.918	0.573	0.874

OVERALL AND COMPONENT BALANCES											
KEN#64	Mass, g	FRACTION OF RAW SHALE	% C	% H	% N	% S	% ASH	% H2O	FRACT OF RAW C	FRACT OF RAW N	FRACT OF RAW S
RAW SHALE	6883.7	1.000	16.11	1.98	0.65	1.79	74.18	2.78	1.000	1.000	1.000
SPENT SHALE	5303.6	0.771	0.00	0.00	0.00	0.00	0.00	0.00	0.270	0.420	0.239
-BIN	4692.6	0.682	5.23	0.14	0.34	0.42	94.37	0.46	0.221	0.357	0.160
-PYR FINES	291.7	0.042	12.83	1.01	0.55	1.39	82.77	0.81	0.034	0.036	0.033
-COMB FINES	275.7	0.040	2.57	0.08	0.27	1.84	97.31	0.29	0.006	0.017	0.041
-A/C FINES	43.6	0.006	20.38	1.57	1.15	1.27	73.88	0.77	0.008	0.011	0.005
OIL	507.7	0.074	0.00	0.00	0.00	0.00	0.00	0.00	0.371	0.114	0.071
-A/C	68.9	0.010	86.15	10.50	1.36	1.99	0.00	0.00	0.054	0.021	0.011
-ESP	292.1	0.042	78.28	9.54	1.24	1.81	3.04	4.02	0.206	0.081	0.043
-LITE OIL	102.2	0.015	83.36	11.69	0.51	2.01	0.00	0.28	0.077	0.012	0.017
-C5+ VAPOR	44.5	0.007	87.00	13.00	0.00	0.00	0.00	0.00	0.035	0.000	0.000
PYR GAS	440.1	0.064	0.00	0.00	0.00	0.00	0.00	0.00	0.123	0.000	0.550
-CH4	22.2	0.003	75.00	25.00	0.00	0.00	0.00	0.00	0.015	0.000	0.000
-C2'S	14.1	0.002	81.80	18.20	0.00	0.00	0.00	0.00	0.010	0.000	0.000
-C3'S	11.3	0.002	83.70	16.30	0.00	0.00	0.00	0.00	0.009	0.000	0.000
-C4'S	9.4	0.001	84.20	15.80	0.00	0.00	0.00	0.00	0.007	0.000	0.000
-CO2	210.8	0.031	27.30	0.00	0.00	0.00	0.00	0.00	0.052	0.000	0.000
-CO	78.0	0.011	42.90	0.00	0.00	0.00	0.00	0.00	0.030	0.000	0.000
-H2	22.3	0.003	0.00	100.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-H2S	72.0	0.011	0.00	5.90	0.00	94.10	0.00	0.00	0.000	0.000	0.550
COMB GAS	433.0	0.063	0.00	0.00	0.00	0.00	0.00	0.00	0.104	0.008	0.057
-CO	10.2	0.002	42.86	0.00	0.00	0.00	0.00	0.00	0.004	0.000	0.000
-CO2	408.0	0.059	27.27	0.00	0.00	0.00	0.00	0.00	0.100	0.000	0.000
-SO2	14.0	0.002	0.00	0.00	0.00	50.00	0.00	0.00	0.000	0.000	0.057
-O2	0.0	0.000	0.00	0.00	0.00	0.00	0.00	0.00	0.000	0.000	0.000
-NOx	0.8	0.000	0.00	0.00	46.67	0.00	0.00	0.00	0.000	0.008	0.000
WATER	5670.2	0.824	0.00	0.00	0.00	0.00	0.00	0.00	0.013	0.199	0.097
-LIQUID	5670.2	0.824	0.25	0.00	0.16	0.21	0.00	100.00	0.013	0.199	0.097
-VAPOR	0.0	0.000	0.00	0.00	0.00	0.00	0.00	100.00	0.000	0.000	0.000
TOTAL	12354.6	1.795	0.00	0.00	0.00	0.00	0.00	0.00	0.881	0.741	1.012

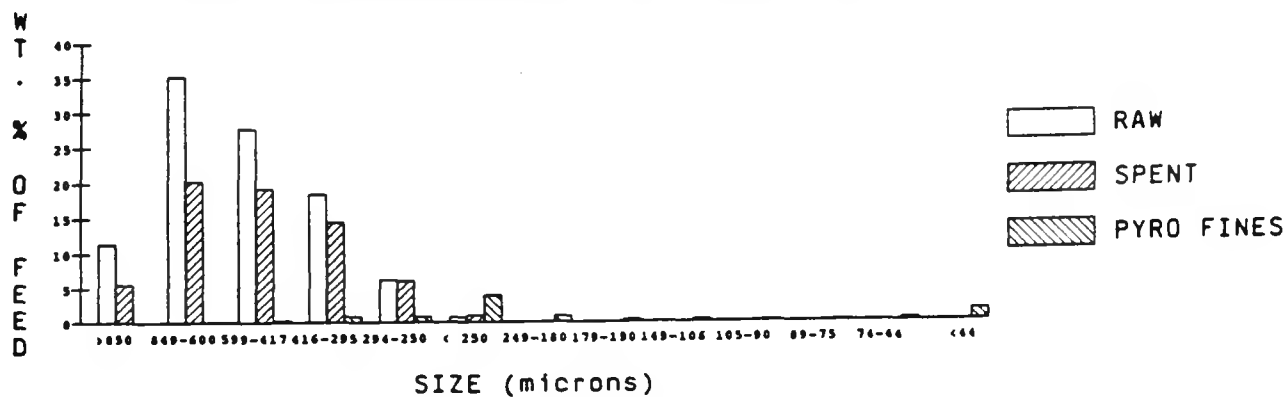
APPENDIX C. Particle Size Analyses from Mini-Plant Test
Matrix

Appendix C (Cont.)

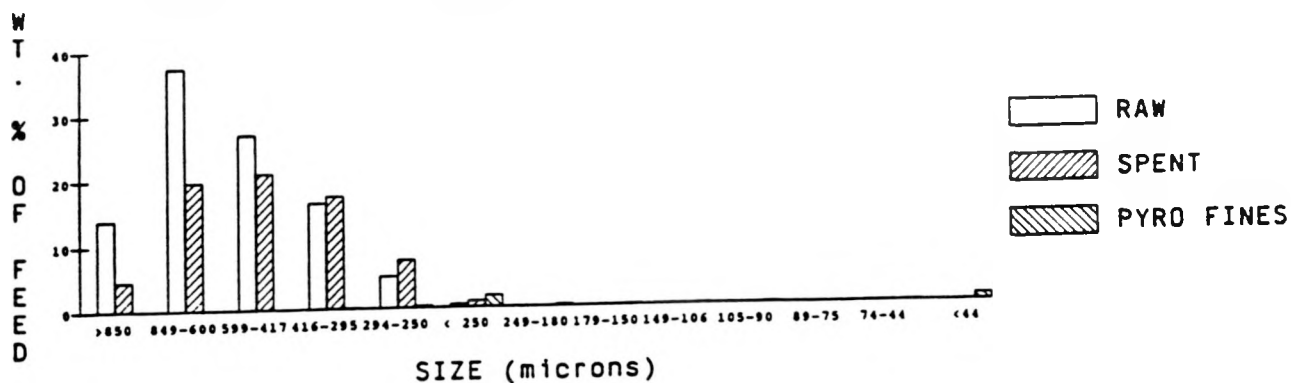
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #7



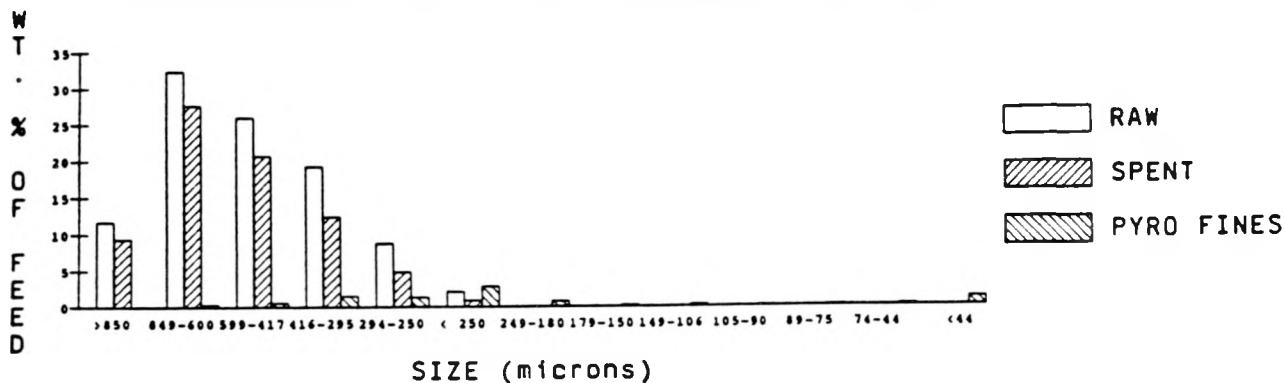
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #9



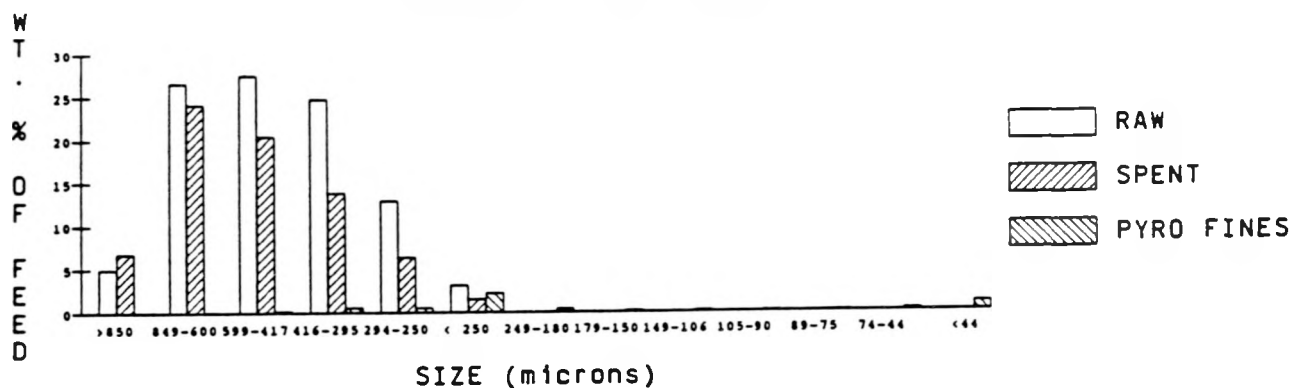
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #10



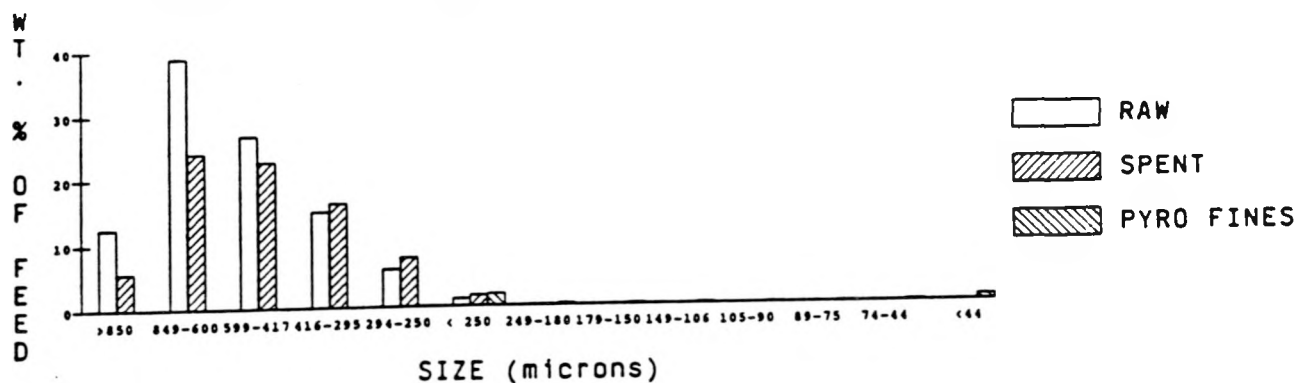
Appendix C (Cont.)
 PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #28



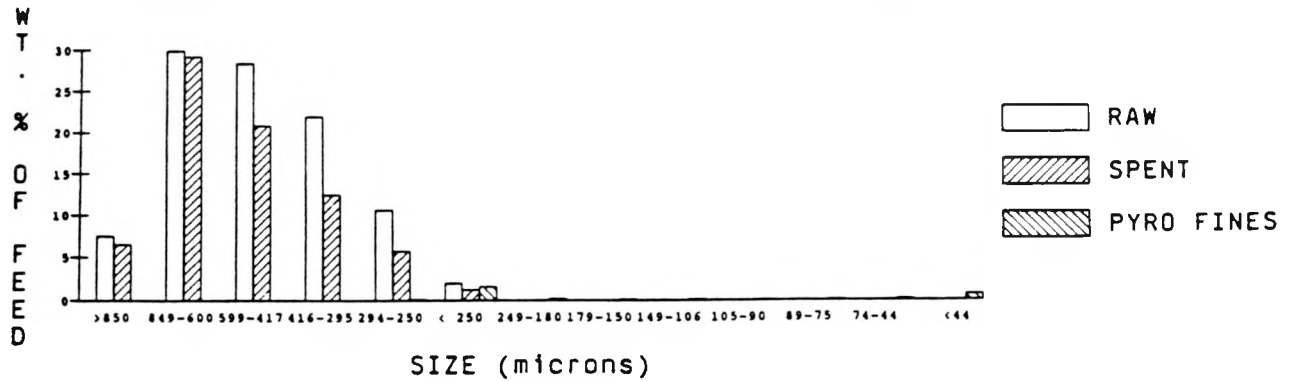
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #31



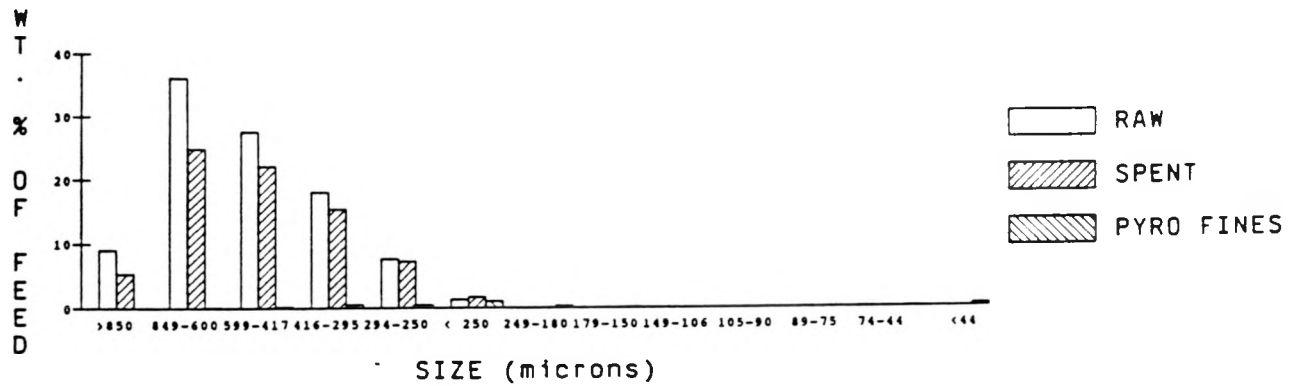
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #33



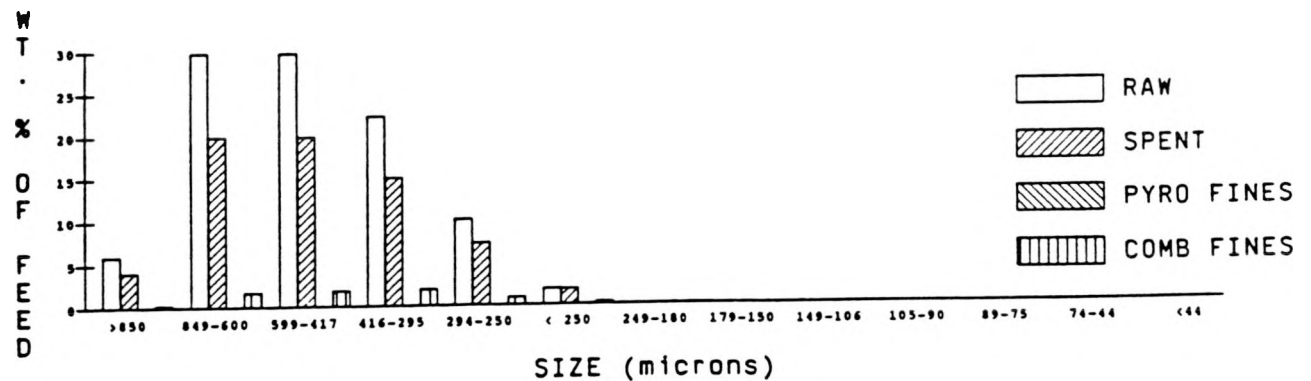
Appendix C (Cont.)
 PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #35



PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #36

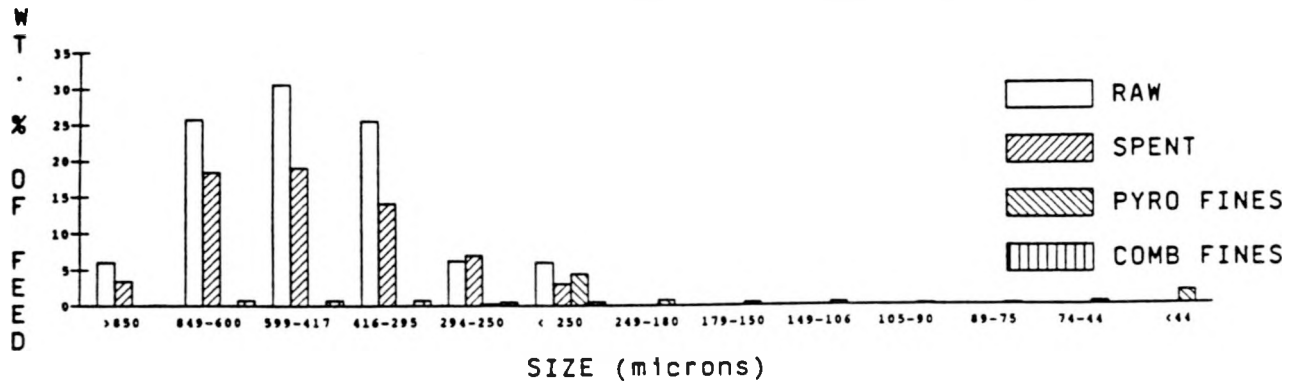


PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #48

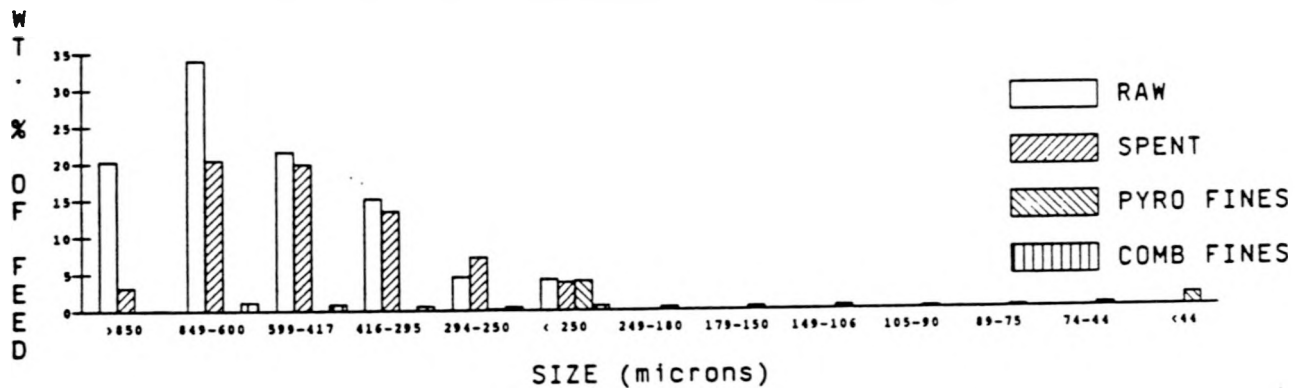


Appendix C (Cont.)

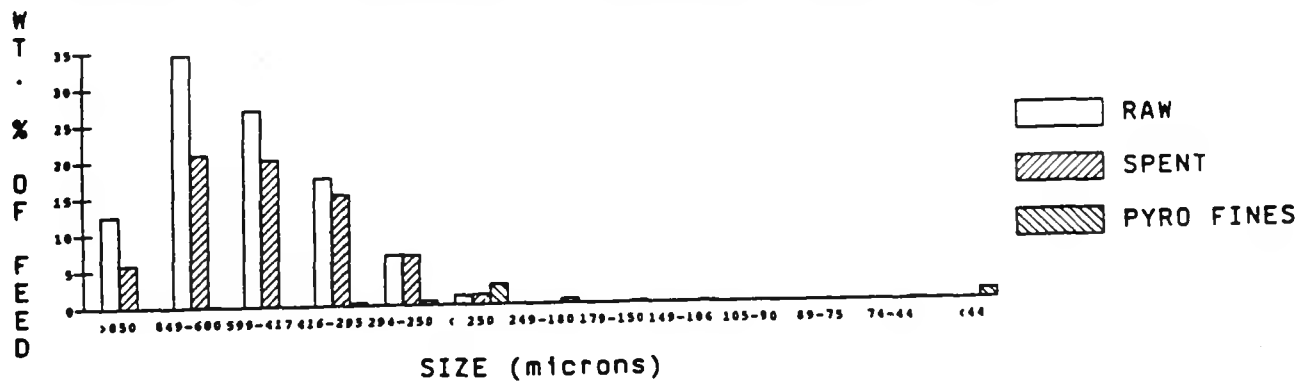
PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #53



PARTICLE SIZE DISTRIBUTION FOR KENTORT RUN #64

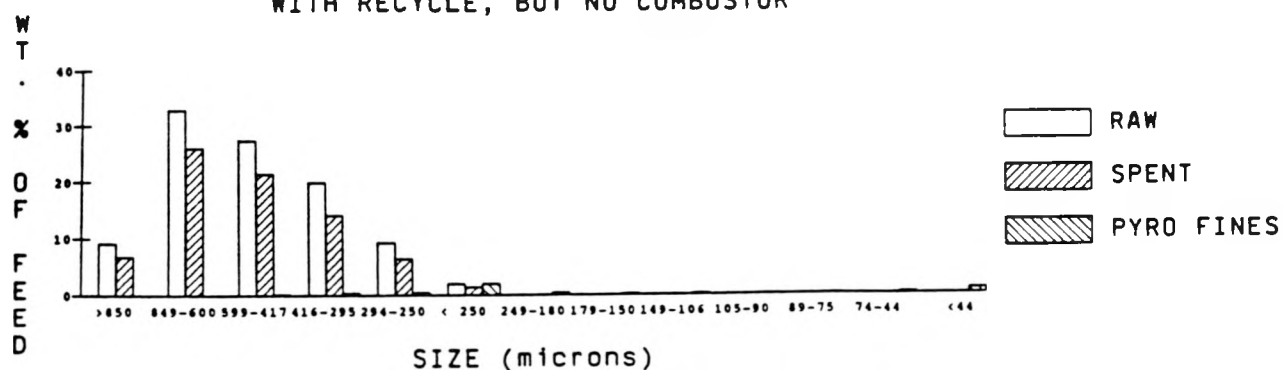


AVERAGE SIZE DISTRIBUTION FOR NON-RECYCLE KENTORT RUNS

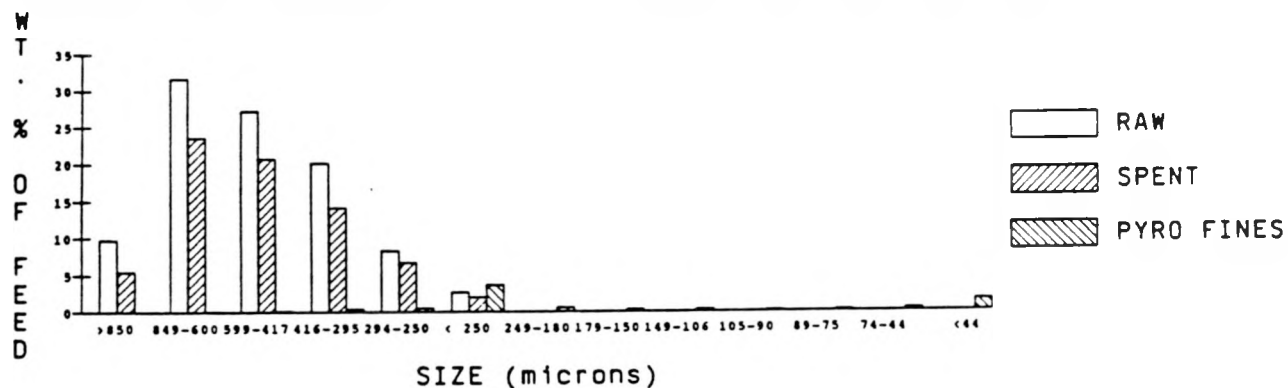


Appendix C (Cont.)

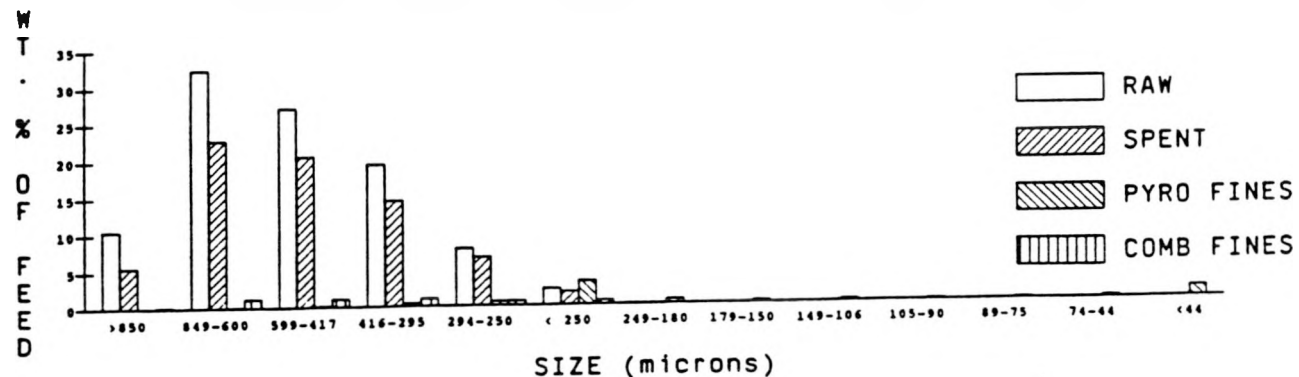
AVERAGE PARTICLE SIZE DISTRIBUTION FOR ALL KENTORT RUNS WITH RECYCLE, BUT NO COMBUSTOR



AVERAGE SIZE DISTRIBUTION FOR ALL KENTORT RUNS WITH RECYCLE



AVERAGE PARTICLE DISTRIBUTION FOR ALL KENTORT RUNS



APPENDIX D. Analytical Procedure for Mini-Plant Reactor Products

Ammonia Ammonia content of the process waters was determined by selective ion ammonium electrode (Orion 95-10). Acid treated samples were first distilled to remove interferences according to test number HCL-224 in Methods for Chemical Analyses of Water and Wastes, EPA, 1974.

Aromatic Carbon Determined by ^{13}C -NMR on a Varian XL-200. CDCl_3 used as solvent, $\text{Cr}(\text{AcAc})_2$ used for relaxation. At least 5000 transients taken with 1 sec delay. Reported values represent integration of aromatic (100-150ppm) vs aliphatic region (0-50 ppm).

Average Molecular Weight Measured by vapor phase osmometry (Wesken Model 233) at three sample concentrations in pyridine. Benzil used as calibration standard.

Capillary GC Composite oil samples were dissolved in CH_2Cl_2 then injected on an HP-5880A GC equipped with FID detectors and a 30M DB-5 fused silica capillary column.

CHN Both solids and oil fractions were analyzed in duplicate (~50 mg each) on a Leco model 600 using acetanilide as standard. Base preserved process waters were analyzed for carbon content only.

Chlorine Composite oils were combusted in an oxygen bomb with the Chlorides extracted and measured by ion chromatography.

Conradsen Carbon Determined according to ASTM D189-76

Density Light oil density was determined with a Mettler DMA-40 oscillating densitometer. Composite oil densities were determined with a 25-cc glass picnometer.

Dissolved/suspended solids Duplicate 100 mL aliquots of process water were filtered through tared 0.45 μm Metrical membrane filters to determine suspended solids. Dissolved solids were determined following evaporation of filtrate.

Fischer Assay Procedure similar to that of ASTM D3904-80. Modifications are described in "Rubel & Coburn, 1981 EOSS Proceedings, p.21".

Appendix D (Cont.)

Fixed Gases Following condensation of the liquid products, pyrolysis/gasifier offgases from the 3" integrated reactor were analyzed on a Carle Series Sx-AGC Gas analyzer using Carle App. #397-B using both FID and TC detection. A minimum of four injections were made and averaged for each retort run. CO, CO₂, NO_x, O₂, and SO₂ from the combustor were continuously monitored with on-line Beckman industrial gas analyzers (models 865, 864, 951A, 755, and 865, respectively). Gasifier and combustion gases in the bench scale studies were measured on a Hache/Carle Gas analyzer (App. #158-B). Matheson primary standards were employed for daily calibration of all analyzers.

Heating Value Composite oil samples were analyzed in duplicate according to D3286 using a Parr adiabatic calorimeter model 1242. Standardized with benzoic acid.

Metals Reported heavy metal values were obtained on a Spectra-Span IIIB ICP unit (+-3% relative RSD). High purity standards from VHG employed for calibration. Acid treated retort waters and perchloric acid digested composite oils were tested.

Oxygen Measured in duplicate for the composite oils by conversion of O₂ to CO₂ followed by coulometric determination of the CO₂. Acetanilide used daily as standard.

Pour Point Determined according to ASTM D97-66

Proximate Conducted on a Perkin-Elmer TGS2, details given in Elder, Fuel, 62, 580 (1983); the average of duplicate runs reported.

Simulated Distillation Composite oils analyzed on an HP3351B lab automation system equipped with an HP5711 GC and dual FID detection according to proposed ASTM method D2887.

Sulfur Solids, oils, and both acid and base preserved process waters run in duplicate on a Leco SC132 sulfur analyzer according to D4239-84. Leco supplied standards were used for calibration.

Water content of oils Analyzed by Karl Fischer Titration in triplicate according to D1744-64 using a Fischer model 392. Calibrated daily with distilled water.

Viscosity Composite oils analyzed at 40 °C with a Brookfield HBT viscometer equipped with thermocell.