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CATALYTIC MULTI-STAGE LIQUEFACTION OF COAL

SIXTH QUARTERLY REPORT FOR THE PERIOD

1 JANUARY 1994 - 31 MARCH 1994

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R.H. Stalzer

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MAY 1994

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ABSTRACT

This quarterly report covers the activities of Catalytic Multi-Stage Liquefaction of Coal during the Period January 1 - March 31, 1994, at Hydrocarbon Research, Inc. in Lawrenceville and Princeton, New Jersey. This DOE Contract Period is from December 8, 1992 to December 7, 1994.

The overall objective of this program is to produce liquid fuels from coal by direct liquefaction at a cost that is competitive with conventional fuels. Specifically, this continuous bench-scale program contains provisions to examine new ideas in areas such as: low temperature pretreatments, more effective catalysts, on-line hydrotreating, new coal feedstocks, other hydrogen sources, more concentrated coal feeds and other highly responsive process improvements while assessing the design and economics of the bench-scale results.

This quarterly report covers work on Laboratory Scale Studies, Continuous Bench-Scale Operations, Technical Assessment and Project Management.

SUMMARY

During this reporting period a short series of microautoclave tests were carried out to evaluate the activity of both dispersed and supported catalysts. The dispersed catalyst, ammonium heptamolybdate, as well as the two supported catalyst, Shell 317 (Ni/Mo) and Amocat 1A (Co/Mo), in the presence of coal derived solvent and DMDS were active in promoting the water-gas shift reaction

During this reporting period CMSL-04 (Run 227-81) catalytic two-stage liquefaction of Black Thunder Coal, was completed. Under mild liquefaction conditions, first and second stage temperatures of 388 and 427 °C respectively, the performance of using syngas vs hydrogen as the reducing gas in the first stage of the CTSL operation in catalytic/catalytic mode was compared. The syngas system resulted in an initial coal conversion 3.1% higher than with pure hydrogen and a final steady state value 0.8% higher than the pure hydrogen condition. The residuum conversion stayed constant for these two conditions even though the catalyst has aged by the time of the syngas condition indicating that the effect of syngas balances the effect due to catalyst aging at this time. Syngas can be used as an effective reducing gas in the first stage of the CTSL process operating in the catalytic/catalytic mode.

Comparing syngas used under both low and high severity conditions shows that the high severity condition results in an increase in coal conversion, by 3.2 wt% maf coal, an increase in residuum conversion, by 2.7 wt% maf coal, and a shift in distillate selectivity toward the lighter products.

An initial comparison between Periods 12 and 13, performed under identical operating conditions except for Period 13 being "starved" of CO due to a low flow rate, shows the effect of CO feed on process performance. Coal conversion increases by 1.1 wt% maf coal, residuum conversion increases by 3.9 wt% maf coal, light gas yield increases by 2.2 wt% maf coal, and distillate selectivity does not change. This fortuitous result should be further investigated to determine why the lower CO flow resulted in improved performance.

TASK 2 - LABORATORY SCALE STUDIES

Laboratory Support for Bench Run CMSL-04

During this reporting period, a short microautoclave test program was designed to evaluate the activity of catalysts, both dispersed and supported, for promoting the water gas shift reaction. The Microautoclave tests were carried out, in the absence of coal, at a temperature of 399°C (750°F), 5.5MPa (800 psig) of cold CO pressure with and without solvent. The amount of water for the shift reaction was 0.85 gm, equivalent to a one to one molar ratio with the carbon monoxide. Results of the tests are summarized in Table 1.

One dispersed catalyst (ammonium heptamolybdate, AHM) and two supported catalysts, Shell 317 (Ni/Mo) and Amocat 1A (Co/Mo), were evaluated with or without the presence of solvent. CO conversion for each test was calculated according to the production of hydrogen and carbon dioxide. The amount of hydrogen yield was adjusted to account for reactions with dimethyl disulfide (DMDS) and carbon dioxide (to form methane). In the presence of a coal derived solvent (L-769) and DMDS, AHM (300 wppm) and both supported catalysts were active in promoting the shift reaction. With the exception of one anomalous result (CL-101), higher CO conversions were observed for tests without the solvent, probably due to better contact between the reactant (CO & water) with the catalyst.

Under the test conditions, both supported catalysts were more active than AHM in promoting the water-gas shift reaction in the absence of solvent and were similar in activity with the solvent. The supported catalyst also yielded 2 to 3 times more methane.

TASK 3 - BENCH-SCALE STUDIES

CMSL-03 Impact of Syngas on Liquefaction - Dispersed Catalyst

In CMSL-03 the effect of hydrogen and syngas as reducing gas was evaluated using Black Thunder coal in a thermal/catalytic two-stage configuration. The first was a slurry reactor while the second stage was charged with a Ni.Mo (Shell 317) supported catalyst in a ebullated bed configuration. Ammonium hetamolybdate equivalent to 1500 ppm of molybdenum (on dry coal basis) was added along with the coal feed. The first stage reducing gas was either hydrogen (Condition 1, Periods 1 to 5) or a 25/75 mixture of hydrogen in carbon monoxide (Conditions 2 thru 4, Periods 6 to 15), while the second stage was also under a hydrogen atmosphere. The first stage gases and low boiling materials were removed as overheads from an interstage separator prior to the second catalytic stage. Results of this run had been reported in the Fourth Quarterly Report. The first stage, second stage overheads and Atmospheric Still overheads were combined proportionably according to the rates at which each stream was generated. The composite was fractionated into four true boiling (TBP) fractions. A series of analyses, including API gravity, elemental, compound class type (PONA) on the lighter fractions, bromine number, aniline point and flash point, was performed. The results of these analyses are reported in *Tables 5 thru 8*.

The initial and final boiling point of the composite Overheads samples are ranging from 46° to 57°C and 414° to 438°C, respectively. The Period 5 sample had the highest the highest initial boiling point, while the Period 15 sample had the highest final boiling point. The distribution of the four TBP fractions from these samples are compared in *Figure 1*. Over 87 W% of the materials in these samples boiled below 343°C.

With the exception of the 260-343°C fraction in which hydrogen content decreased gradually as the run progressed, process conditions seem to have minor effect on hydrogen content in the other fractions, as illustrated in *Figure 2*. Surprisingly, the sulfur content of the IBP-177° and 177-260° fractions decreased significantly when hydrogen was replaced by syngas as the first stage reducing gas, as shown in *Figure 3a*. Also, it seems that the sulfur addition rate to the second stage reactor had a positive impact on the sulfur content in these two lighter fractions. In spite of increasing space velocity, the decline continuously from Periods 9 thru 15. Due to equipment problem the H₂S addition rate dropped off from 2.7 W% dry coal to 1.7 and 0.4 W% in Periods 12 and 15, respectively. Process conditions had a different impact on the nitrogen level in these two fractions. The type of reducing gas had no significant impact on the nitrogen content in these two TBP fraction, Period 5 vs Period 9, as shown in *Figure 3b*. However, there was considerable increase in nitrogen level in Periods 12 and 15.

The compound class type distribution in the two lightest fractions was presented in *Figures 4a and 4b*. The largest component in the IBP-177°C fraction was naphthenics (50 to 60 V%) and was followed paraffins, which constituted 25 to 30 V%, while the remaining shared equally between olefins and aromatics (less than 10 V% each). The compound class type distribution was some what different in the 177-260°C fraction.

Although the naphthenics compound was still the largest component, its concentration had dropped to 45-50 V%, while the second largest component was shifted from paraffins to aromatics (25 to 40 V%). There was no detectable olefins in the sample from Periods 5 and 9.

RUN CMSL-04 (227-81) IMPACT OF SYNGAS ON LIQUEFACTION IN A CATALYTIC/CATALYTIC MODE OF OPERATION

Objectives and Background

The main objective of this run was to evaluate the effect of reaction severity on the dissolution of Black Thunder Mine coal using syngas as a reducing gas in the first stage of a two-stage ebullated bed configuration. Additional technical objectives included the determination of the relative performance of using CO/H₂ vs H₂ in the first stage of a fully ebullated two-stage system. Finally, the effect of reaction severity on CTSL process performance with synthesis gas was determined.

The production of hydrogen constitutes one of the major operating cost components of a coal liquefaction process. Alternative sources of hydrogen such as synthesis gas, a mixture of carbon monoxide and hydrogen coming directly out of a gasifier or reformer, can potentially reduce the operating cost by 10-15%. It is known that in the presence of an alkali salt, a mixture of CO/H₂ and water is very effective in solubilizing high oxygen containing low rank coals at relatively mild severity conditions, usually below 400°C.

A successful operation with promising results such as improved liquefaction yields, kinetics, and hydrogen efficiency of the CTSL process was demonstrated during CMSL-03 (227-79). The interstage separation used during CMSL-03 was instrumental in removing CO_x off gases, thereby reducing further reforming of these gases to methane/water at the expense of the valuable hydrogen. CMSL-03 operating conditions were identical to those in the run plan for CMSL-04 except that CMSL-03 used a slurry catalyst (ammonium molybdate) instead of a supported catalyst in the first stage reactor.

Scope of Work

The dissolution of Black Thunder Mine coal was evaluated using H₂ in the first stage at a base condition and the syngas in the first stage at the same base condition as well as at a higher reactor severity in a catalytic/catalytic two-stage operation. The impact of using syngas on liquefaction was evaluated in terms of product quality and overall process performance.

Summary

Under mild liquefaction conditions, first and second stage temperatures of 388 and 427 °C respectively, the performance of using syngas vs hydrogen as the reducing gas in the CTSL operation in catalytic/catalytic mode was compared. A mixture of syngas and water for the coal solubilization step in stage 1 resulted in an initial coal conversion 3.1% higher than with pure hydrogen and a final steady state value 0.8% higher than the pure hydrogen condition. The residuum conversion is constant for these two conditions even

though the catalyst has aged by the time of the syngas condition indicating that the effect of syngas balances the effect of catalyst aging at this time. The syngas condition also has a slightly higher distillate yield than the hydrogen condition by 0.9 wt% maf coal. Syngas can be used as an effective reducing gas in the first stage of the CTSL process operating in the catalytic/catalytic mode.

Comparing syngas used under both low and high severity conditions shows that the high severity condition results in an increase in coal conversion, by 3.2 wt% maf coal, an increase in residuum conversion, by 2.7 wt% maf coal, and a shift in distillate selectivity toward the lighter products.

An initial comparison between Periods 12 and 13, performed under identical operating conditions except for Period 13 being "starved" of CO due to a low flow rate, shows the effect of CO feed on process performance. Coal conversion increases by 1.1 wt% maf coal, residuum conversion increases by 3.9 wt% maf coal, light gas yield increases by 2.2 wt% maf coal, and distillate selectivity does not change. This fortuitous result should be further investigated to determine why the lower CO flow resulted in improved performance.

Run Plan

This bench run was performed using two equal volume ebullated beds. A simplified process flow diagram is presented in *Figure 5*. Coal was dissolved in the first-stage in the presence of syngas and water, while the primary coal liquids were further upgraded in the second stage in a hydrogen rich atmosphere. An interstage separator was installed to remove gases and light distillates (as a separator overhead, SOH) generated from the first stage and to recover any unreacted CO/H₂O. Fresh hydrogen was injected into the second stage. Products from the second stage were recovered using a hot separator (as a separator overhead, SOH) and a continuous atmospheric still (CAS) which has atmospheric still overheads (ASOH) and CAS bottoms as the outlets. The CAS bottoms were pressure filtered in a batch mode into pressure filter liquid (PFL) and pressure filter cake (PFC). The PFL was used as recycle solvent and as buffer liquids for the ebullating pump operations. The first stage SOH, (oil fraction only), second stage ASOH, and second stage SOH were combined as the feed to the hydrotreater.

The run plan is given in *Table 6*. Four process conditions were chosen for this run, based upon the conditions used in CMSL-03 so that these two runs could be directly compared. The actual operating conditions are presented in *Table 7*. Condition 1 extended from Periods 1 to 6 using pure hydrogen as reducing gas for both reaction stages. This condition served as a base case for comparison. Condition 2 was identical to Condition 1 except the reducing gas was changed from pure hydrogen to a 86% / 14% mixture of CO and H₂. Also for this condition, and all subsequent conditions, water was injected upstream of the preheater in order to obtain an overall water content of 40 wt% of the dry coal in the feed slurry. Condition 3 was at a higher overall system severity than the initial two conditions, the space velocity was changed from 722 to 899 Kg dry coal/hr/m³ catalyst and the first and second stage reactor temperatures were raised from

388/427 °C to 399/438 °C. Condition 4 was to vary from Condition 3 by the CO/H₂ balance in the reducing gas used in the first stage. This was to have changed from 90% CO to 75 % CO.

A problem was experienced during the run with the CO gas compressor. An internal seal leaked allowing oil to enter the CO flow stream which then proceeded to plug the mass flow meter being used to control the CO flow. It did not plug the entire meter, just the main flow stream, the slip stream that is used to calculate the mass flow through the meter was still open. From the end of Period 12 on, this is the only flow of CO into the system resulting in approximately 1/3 of the desired flow.

Tri-nonyl polysulfide (TNPS) and hydrogen sulfide was added to the first and second stage, respectively. The target sulfur addition rate to the first stage was 1 wt% of dry coal and 3 wt% to the second stage.

Coal

Wyoming sub-bituminous coal, Black Thunder Mine, was used as the feedstock. The coal analyses of the feed (HRI-5828) are presented in *Table 8*. This batch of coal was previously evaluated in CC-13, CC-15, CMSL-01 and CMSL-03. This coal has approximately 25 wt% moisture in the raw form and was pulverized without further drying.

Start-up/Makeup Oil

The start-up/make-up oil used during this run was the coal-derived recycle solvent generated during POC-01 (PDU RUN 260-04). The elemental analysis and boiling point distribution is shown in *Table 9*.

Catalyst

Fresh Shell S-317 1/32" extrudates were charged to both the first and second stage ebullated beds. The hydrotreater was charged with Criterion C-411 trilobe catalyst. The catalyst was presulfided during normal start-up operations.

PERFORMANCE AND RESULTS

The overall operation performance is presented in *Figures 6-9*. The daily material balance average 97.5 wt% with the greatest variation at the beginning and end of the run, *Figure 6*. The reactor temperatures are presented in *Figure 7* and show a very small variation from target values. The space velocity is presented in *Figure 8* and shows smooth operation for the later part of Condition 1 and Condition 2 at 660 kg dry coal/hr/m³ catalyst and smooth operations for Condition 3 & 4 at 830 kg dry

coal/hr/m³ catalyst. The initial part of Condition 1 shows the original target space velocity of 980 but during Period 3 it was decided to lower this to 660. The overall process severity is presented in *Figure 9* showing an initial relative severity of 1.1 for Conditions 1 & 2 and a severity of 1.6 for Conditions 3 & 4. The first 3 periods show an low severity of 0.8 due to the initial higher space velocity which was later lowered. A summary of performance is given in *Table 10*. The product inspection of the various flow streams are given in *Tables 11-15*.

Coal Conversion

Coal conversion varied during Condition 1 from 82.8 to 86.8 wt% maf coal, *Figure 10*. It initially began low due to the high space velocity that the unit was being operated at, when the space velocity was reduced at the end of period 3 the coal conversion increased to a steady value of 86.8 wt% maf coal. This is 2-3 wt% lower than would normally be expected for Black Thunder coal at these processing condition. In Condition 2 the coal conversion initially jumped to a high value of 89.9 and then slowly decreased for the remainder of the condition to 87.6 wt% maf coal. This initial increase just as the CO is introduced to the reactor is probably due to the direct interaction between CO and the coal. The later decrease in coal conversion could be due to a decrease in the quality of the recycle solvent. During Condition 3 the coal conversion increases to 92 wt% as the temperature of both reactors is raised. This conversion level only changes at the very end of the run when it falls off, possibly due to the poor material balance experienced during the end of the run.

524°C⁺ Residuum Conversion and Yield

The residuum conversion, calculated by dividing the sum of 524 °C residuum yield plus the unreacted coal by the value of the maf coal, is normally a strong function of catalyst age (*Figure 11*). Due to batch aging of the catalyst in a Bench Unit, the resid conversion is expected to decrease over the course of the run as the residuum yield increases. Going from Condition 1 to Condition 2 there is no decrease in the resid conversion indicating, that the effect of the CO nearly balances the effect due to catalyst aging for this time. During Condition 3 the residuum conversion increased due to the effect of raising the process severity. However, in comparing Periods 12 & 13, at identical process severities, the resid conversion increases from 87.2 to 91.1 wt% maf coal. During Condition 4 the residuum conversion decreased sharply, part of this decrease could be due to the poor material balance experienced at the end of the run.

C₁-C₃ Gas Make

The gas make during Condition 1, pure hydrogen used as the reducing gas, was typical for Black Thunder coal, a light gas make of 6.3 to 6.6 wt% (maf coal basis) and a total gas make of 9.9 to 10.0 wt% (*Figure 12*). During Condition 2 the gas make increased slightly to a light gas make of 7.3 to 7.6 wt% and a total gas make of 11.8 wt%. This

is also the gas make during the initial part of Condition 3 (Period 12) even though the temperature was increased from 388 °C to 399 °C in the first reactor and from 426 °C to 438 °C in the second reactor. During Period 13 the CO input stream was approximately 1/3 of the flow during Period 12 and the gas make increased. During Condition 4, also with the low CO flow, the light gas make was 8.9 to 9.1 wt% and the total gas make was 13.7 to 14.6 wt%. Since reaction severity in Conditions 1 & 2 were essentially the same, the increase in methane yield in Condition 2 to 4 was assumed to be from the reforming of hydrogen and carbon oxides.

C₄-524°C Distillate Yield

Condition 1, with hydrogen being used in both stages, resulted in a distillate yield of 57.6 wt% maf coal (*Figure 13* and *Table 10*). This is lower than the more typical value of 61 wt% normally found with Black Thunder coal. Under identical processing condition, the syngas (Condition 2) gave a 0.9% higher distillate yield of 58.5 wt% maf coal. This total yield, however, is shifted more toward the heavy end. As can be easily seen in *Figure 13*, Condition 2 gives a 2.7 wt% lower IBP-177 °C yield and a 4.4 wt % higher 343-454 °C yield. Condition 3, at a higher reactor temperature and higher space velocity, gave a distillate yield of 58.9 wt% maf coal, 0.4 wt% higher than Condition 2. Condition 3 also shows an even better yield distribution than Condition 1, giving 1.3 wt% more IBP-177 °C, 1.6 wt% more 177-260°C, identical 260-343 °C, 0.8 wt% less 343-454 °C, and 0.5 wt% less 454-524 °C. Condition 4 has an extremely low distillate yield of only 53.3 wt maf coal; however, this is most likely due to the poor material balance experienced at the end of the run.

Inspections of Products

Analyses of the liquid products are presented in *Tables 11-15* and *Figures 14-18*.

The first stage SOH and second stage ASOH were actually two of the three feed streams to the hydrotreater, the other being the second stage Hot Separator Overheads, which were not sampled. The second stage SOH is the product from the hydrotreater (see *Figure 5* for the simplified flow scheme). The sulfur, nitrogen and H/C atomic ratios for the first stage SOH and second stage SOH and ASOH are presented in *Figures 14-16*.

The sulfur content of the first stage SOH rises sharply from Condition 1, H₂ to the first stage, to Condition 2, CO to the first stage, from 147 to 428 ppm. During Conditions 3 and 4 the sulfur content stays high at 410-471 ppm. The ASOH sulfur shows a smoother rise from an initial value of 38 ppm to 115 ppm. The second stage SOH, the hydrotreater product, has a low sulfur content ranging from 31 ppm to 45 ppm, showing that the hydrotreater is doing a good job removing the sulfur.

The nitrogen content of the first stage SOH shows a smooth rise from an initial value of 747 ppm to 2941 ppm in Condition 3 where it plateaus, rising in Condition 4 to 2965 ppm. The ASOH shows a smooth rise throughout the run from an initial value of

435 ppm to a final value of 3412 ppm. The second stage SOH shows a difference in values from the H₂ condition of 22 ppm to the CO conditions from 32-36 ppm. This shows the nitrogen removal being performed by the hydrotreater.

The H/C atomic ratio is consistently higher for the second stage SOH, ranging from 1.73 to 1.79, as compared to either the first stage SOH, ranging from 1.59 to 1.76, or the ASOH, ranging from 1.52 to 1.63. The hydrotreater is increasing the hydrogen content of the product.

The pressure filter liquid (PFL) properties are summed up in *Table 14* and *Figure 17-18*. As the catalyst ages the resid content of the PFL increases, from 22 wt% to 39.7 wt%, as the H/C atomic ratio decreases, from 1.37 to 0.98. The cyclohexane and toluene insolubles also increase, from 0.29 to 17.63 and 0.10 to 1.98 respectively, as the run progresses and the PFL becomes richer in asphaltenes and pre-asphaltenes.

The pressure filter cake (PFC) analysis is presented in *Table 10*.

Analyses of TBP fractions and interstage samples will be discussed in the next quarterly report.

CO Conversion & Water Balance

CO interacts with coal either directly or indirectly via nascent hydrogen formed from the water-gas shift reaction. The WGS reaction :

is slightly exothermic, hence low temperature tends to shift the reaction towards the right. The equilibrium conversion for a syngas feed containing 25 vol% of H₂ is estimated as 45% under the process condition planned for the first reactor.

In calculating the amount of CO and CO₂ in the off gas, a small correction was made based on the amount of CO and CO₂ produced from the coal when only hydrogen was used, Condition 1. This correction was based on the average values for Periods 5 & 6.

Table 16 provides a summary of the CO conversion and water balance for all four conditions. The water consumed in the Water gas shift reaction is taken as the average of two different calculations. The amount of CO that is calculated to have reacted and the stoichiometric equation for the water-gas shift reaction is used to calculate the amount of water consumed. Similarly, the CO₂ that is calculated to have been produced is used to calculate the amount of water that was consumed. The average of these two calculations is then taken as the "true" amount of water consumed by the shift reaction.

The water balance is presented as the feed water and the collected water. The feed water includes the water injected into the first reactor, the moisture content of the coal (25.38 wt%) and the water formed by the deoxygenation of the coal. The collected water is just the water collected in the SOH outlet streams. The difference between these two

values gives the amount of water that was consumed in the water-gas shift reaction. The assumed level of deoxygenation was chosen so as that this calculated water consumed would exactly match that calculated from the stoichiometric equation and the CO and CO₂ balance. This results in a reasonable value for the deoxygenation, ranging from 84 to 96 %.

The CO conversion calculated increasing from 68.5% in Condition 2 to 93.2% in Condition 4. This is due to the system being "starved" of CO from the lower than planned CO feed rate. As would be expected with an excess of one reactant (H₂O) the equilibrium is shifted.

COMPARISON OF CMSL-04 WITH CMSL-03

Basis for Comparison

The operating conditions for CMSL-04 were chosen so as to facilitate a comparison with CMSL-03. ALL pertinent reactor conditions and flows were chosen as identical to CMSL-03. The significant difference between these two runs is that CMSL-03 used a slurry catalyst (1500 ppm Mo as ammonium heptamolybdate) as a first stage catalyst while CMSL-04 used the same supported catalyst in the first stage as was used in the second stage. As can be seen from *Figures 5-7* the four conditions in each test program provide a reasonable basis for comparison. The space velocity, *Figure 19*, is directly comparable for Conditions 1 and 2 and slightly higher for CMSL-04 in Condition 3 and slightly lower in Condition 4. The overall system severity, *Figure 20*, is comparable for Conditions 1 and 2 and CMSL-04 is slightly higher in Condition 3 and lower in Condition 4. The catalyst age, *Figure 21*, is slightly higher for CMSL-04 for all four conditions.

Process Performance

The comparison of process performance is presented in *Figures 22-25*. The coal conversion for CMSL-04 is always lower than it is for CMSL-03, possibly due to the larger thermal volume of the first stage reactor in CMSL-03 due to the lack of a supported catalyst. This difference in coal conversions, *Figure 22*, must be kept in mind when making other comparisons of performance and yield for many of these results are strongly dependent on changes in coal conversion. The resid conversions, *Figure 23*, follow the trend seen in the coal conversions; for CMSL-04 Conditions 1 and 4 are mildly lower, Condition 2 is much lower and Condition 3 is slightly lower. This indicates that the differences in resid conversions between CMSL-04 and CMSL-03 are predominantly due to the differences in coal conversions. The hydrogen consumption presented in *Figure 20* shows that except for Condition 1 CMSL-04 had approximately a 1.5 higher consumption. The CO conversion, *Figure 25*, shows that for the CO feed periods, while CMSL-03 conversion stayed the same and then decreased at the end, CMSL-04 increased continually over the course of the run. This increase in the CO conversion as a direct result of the system being "starved" of CO feed was previously discussed.

Distillate Yield

The distillate yields is presented in *Figures 26 & 27*. *Figure 26* shows the total C_4 to 524°C yield for the two runs. This figure shows a similar trend as does the coal conversion and resid conversion figures. Condition 3 for these three figures shows a small difference between the two runs while the other three conditions show a larger difference. The overall distillate yield is also a strong function of the coal conversion. A better basis for comparison would be the distillate selectivity. This would put the comparison on an equal yield basis. *Figure 27* shows a comparison of distillate selectivity for these two runs. As can be seen from this figure, for CMSL-03 the heavy oil fraction ($343^\circ\text{C}+$) increases over the course of the run at the expense of the light oil

fraction (343C*-). For CMSL-04 this takes the opposite trend, the light oil fraction actually increases as the run progresses at the expense of the heavy oil. Since the effect of catalyst aging is to increase the yield of the heavy cut, this result is very surprising and warrants further investigation.

RUN CMSL-05 (227-82) RUN PLAN

Bench Run CMSL-05 (227-82) will be carried out using Illinois No. 6 bituminous coal, two-stage ebullated bed reactors with Shell-317 extrudate catalyst and Criterion C-411 NiMo/Alumina catalyst in the hydrotreater. The technical objectives are:

1. To investigate the effect of interstage separation of light gases and hydrocarbons (650°F distillates) from the heavier product stream going from the first stage to the second stage reactor on the overall process performance.
2. To determine the effect of an in-line hydrotreater at two different operating conditions and two space velocities.
3. To determine the effect of using synthesis gas as a reducing gas (first time at HRI with a bituminous coal) in the first stage ebullated bed reactor on the process performance.

The Run Plan for CMSL-05 is given in *Table 17*.

TASK 4 TECHNICAL ASSESSMENT

The economic incentive for increasing the catalyst replacement rate in a commercial CTSL facility was evaluated. The basis was a 9,800 T/D complex, using washed coal and a catalyst replacement rate of 2.7 lb/ton of coal. Assuming equivalent process performance and yields, for an increased catalyst replacement rate 4.5 lb/ton, the equivalent crude price increases from \$32.50/bbl to \$33.38/bbl (2.7% increase), at the same coal feed rate. A twelve percent higher coal feed rate, to 11,000 T/D, is required to bring the equivalent crude price back to \$32.50/bbl. Table 18 shows the results of this comparison.

TABLE 1

Evaluation of Water Gas Shift Reaction: Effect of Catalyst

Condition: 399C, 30 min. 5.52MPa Cold CO pressure

CO/H₂O molar ratio = 1/1; Solvent = 8 g; DMDS = 0.32 g; Supported catalyst = 2 g; AHM = 0.0024 g

AHM: Ammonium Heptamolybdate; CoMo: Amocat 1A.; NiMo: Shell 317

Test No.	CL-97	CL-99	CL-80*	CL-96	CL-98	CL-102	CL-81*	CL-101	CL-89*	CL-100	CL-103
Catalyst	None	CoMo	AHM	AHM	CoMo	NiMo	AHM	AHM	CoMo	CoMo	NiMo
Additive	DMDS	None	DMDS	DMDS	DMDS	DMDS	DMDS	DMDS	DMDS	DMDS	DMDS
Solvent	L-769	L-769	L-769	L-769	L-769	L-769	None	None	L-769	None	None
Gas Analysis, V%											
H ₂	4.63	18.12		11.91	18.93	15.92		6.35		16.91	18.21
CH ₄	0.91	0.16	1.65	2.45	2.16	1.61	0.62	0.79	1.29	2.02	1.51
C1-C3	1.03	0.32		2.59	2.47	2.07		0.79		2.55	2.27
C4-C7	0.03	0.06		0.07	0.04	0.07		0.01		0.02	0.03
CO	86.79	67.90		71.06	66.95	66.57		84.27		53.18	51.73
CO ₂	6.49	13.57		12.24	11.51	15.34		7.47		27.31	27.73
H ₂ S	1.03	0.04		2.13	0.10	0.03		1.11		0.03	0.03
CO Conversion, V%											
Based on H ₂	6.47	21.10		18.00	25.40	21.90		8.21		27.80	28.60
Based On CO ₂	6.96	16.70	20.20	14.70	16.01	19.40	28.00	8.14	30.00	34.80	35.40

* Previously Reported.

TABLE 2

DETAILED ANALYSES OF TBP FRACTIONS
Run No. 227 - 79 Period No. 05 CMSL-03

TBP Distillation, % IBP = 57°C EP = 415°C

	<u>W%</u>			
IBP-177°C	21.77			
177-260°C	27.70			
260-343°C	47.67			
343°C+	2.86			

TBP FRACTION [°C]	<u>IBP-177</u>	<u>177-260</u>	<u>260-343</u>	<u>343+</u>
API Gravity	52.3	30.8	23.1	11.9
Elemental Analysis [W%]				
Carbon	85.10	86.68	87.68	87.42
Hydrogen	14.15	12.41	12.07	10.71
Sulfur	0.0241	0.0350	0.0179	0.0874
Nitrogen	0.08	0.15	0.11	0.47
Antek N, ppm	214	932	N/A	N/A
Bromine No. [g/100g]	20.58	11.22	10.45	
Aniline Point, [°C]	44	31	44	
Flash Point, [°C]	<-7	73	132	
PONA [V%]				
Paraffins	29.51	13.54		
Olefins	9.60	0.00		
Naphthenics	49.97	50.71		
Aromatics	10.95	35.75		
Aromatics (ASTM D2549)			47.82	

TABLE 3

DETAILED ANALYSES OF TBP FRACTIONS

Run No. 227 - 79 Period No. 09 CMSL-03

TBP Distillation, %	IBP = 48°C		EP = 421°C	
			<u>W%</u>	
IBP-177°C			23.73	
177-260°C			28.70	
260-343°C			37.30	
343°C+			10.27	
TBP FRACTION [°C]	<u>IBP-177</u>	<u>177-260</u>	<u>260-343</u>	<u>343+</u>
API Gravity	51.1	29.4	21.1	14.1
Elemental Analysis [W%]				
Carbon	84.96	86.69	87.79	87.55
Hydrogen	14.00	12.21	11.67	10.69
Sulfur	0.0202	0.0268	0.0170	0.0357
Nitrogen	0.06	0.23	0.20	0.40
Antek N, ppm	198	1224	N/A	N/A
Bromine No. [g/100g]	21.82	16.36	9.47	
Aniline Point, [°C]	42	25	33	
Flash Point, [°C]	<-7	76	129	
PONA [V%]				
Paraffins	28.14	10.48		
Olefins	7.40	0.00		
Naphthenics	56.36	50.95		
Aromatics	8.10	38.57		
Aromatics (ASTM D2549)			55.63	

TABLE 4

DETAILED ANALYSES OF TBP FRACTIONS

Run No. 227 - 79 Period No. 12 CMSL-03

TBP Distillation, %	IBP = 46 °C EP = 438 °C			
			<u>W%</u>	
IBP-177 °C			21.77	
177-260 °C			27.70	
260-343 °C			47.67	
343 °C+			2.86	
TBP FRACTION [°C]	<u>IBP-177</u>	<u>177-260</u>	<u>260-343</u>	<u>343+</u>
API Gravity	50.6	28.0	20.2	13.8
Elemental Analysis [W%]				
Carbon	85.08	86.80	87.83	88.18
Hydrogen	13.96	12.09	11.57	10.73
Sulfur	0.0194	0.0156	0.0183	0.0342
Nitrogen	0.08	0.25	0.25	0.42
Antek N, ppm	326	2688	N/A	N/A
Bromine No. [g/100g]	16.74	15.54	10.92	
Aniline Point, [°C]	43	24	28	
Flash Point, [°C]	<-7	76	141	
PONA [V%]				
Paraffins	28.52	9.55		
Olefins	6.80	4.80		
Naphthenics	58.11	45.38		
Aromatics	6.57	40.27		
Aromatics (ASTM D2549)			61.42	

TABLE 5

DETAILED ANALYSES OF TBP FRACTIONS

Run No. 227 - 79 Period No. 15 CMSL-03

TBP Distillation, %	IBP = 54°C EP = 438°C			
			W%	
IBP-177°C			23.30	
177-260°C			37.40	
260-343°C			29.40	
343°C+			9.90	
TBP FRACTION [°C]	<u>IBP-177</u>	<u>177-260</u>	<u>260-343</u>	<u>343+</u>
API Gravity	50.4	28.0	18.1	12.1
Elemental Analysis [W%]				
Carbon	85.19	86.86	88.34	87.70
Hydrogen	14.11	12.30	11.27	10.43
Sulfur	0.0174	0.0152	0.0210	0.0325
Nitrogen	0.07	0.27	0.33	0.52
Antek N, ppm	273	2823	N/A	N/A
Bromine No. [g/100g]				
Aniline Point, [°C]	85.19	86.86	88.34	87.70
Flash Point, [°C]	14.11	12.30	11.27	10.43
	0.0174	0.0152	0.0210	0.0325
	0.07	0.27	0.33	0.52
	273	2823	N/A	N/A
PONA [V%]				
Paraffins	26.44	7.00		
Olefins	6.80	6.20		
Naphthenics	60.29	48.46		
Aromatics	6.47	38.34		
Aromatics (ASTM D2549)			64.35	

TABLE 6

RUN PLAN FOR BENCH RUN CMSL-04 (227-81)

BLACK THUNDER MINE COAL
 1ST & 2ND STAGE CATALYST : SHELL S-317 NI/MO 1/32" EXTRUDATES
 HYDROTREATER CATALYST: CRITERION C-411 TRILOBE

CONDITION	1	2	3	4
PERIOD	1-5	6-9	10-12	13-15
FEED GAS				
1 ST STAGE	H ₂	CO/H ₂	CO/H ₂	CO/H ₂
2 ND STAGE	H ₂	H ₂	H ₂	H ₂
TEMPERATURE (°C)				
1 ST STAGE	388	388	399	399
2 ND STAGE	427	427	438	438
HYDROTREATER	357	357	377	377
SPACE VELOCITY				
LB DRY COAL/HR/FT ³ CAT	45	45	56	56
KG DRY COAL/HR/M ³ CAT	720	720	900	900
SOLVENT/COAL RATIO	1.2	1.2	1.2	1.2
1ST STAGE FEED (% SCF)				
H ₂	100	10	10	25
CO	0	90	90	75
OVERALL PROCESS SEVERITY (1.00 AT 388/427 °C AND 720 SV)				
STTU	1.00	1.00	1.45	1.45

TABLE 7

ACTUAL OPERATING CONDITIONS FOR CMSL-04 (227-81)

BLACK THUNDER MINE COAL
1ST & 2ND STAGE CATALYST : SHELL S-317 NI/MO 1/32" EXTRUDATES
HYDROTREATER CATALYST: CRITERION C-411 TRILOBE

CONDITION	1	2	3	4
PERIOD	1-6	7-10	11-13	14-17
FEED GAS				
1 ST STAGE	H ₂	CO/H ₂	CO/H ₂	CO/H ₂
2 ND STAGE	H ₂	H ₂	H ₂	H ₂
TEMPERATURE (°C)				
1 ST STAGE	388	388	399	399
2 ND STAGE	427	427	438	437
HYDROTREATER	359	363	377	374
SPACE VELOCITY				
LB DRY COAL/HR/FT ³ CAT	43.4	42.1	51.4	52.3
KG DRY COAL/HR/M ³ CAT	700	680	820	840
SOLVENT/COAL RATIO	1.2	1.2	1.2	1.2
1ST STAGE FEED (%SCF)				
H ₂	100	14	16	70
CO	0	86	84	30
OVERALL PROCESS SEVERITY (1.00 AT 388/427 °C AND 720 SV)				
STTU	1.06	1.06	1.59	1.52

TABLE 8
ANALYSIS OF FEED COAL FOR CMSL-04

HRI No.	5828
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MOISTURE CONTENT	25.38
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PROXIMATE ANALYSIS, WT% Dry Basis

Volatile Matter	43.48
Fixed Carbon	50.49
Ash	6.03

ULTIMATE ANALYSIS, WT% Dry Basis

Carbon	70.4
Hydrogen	3.88
Sulfur	0.44
Nitrogen	0.87
Ash	6.03
Oxygen (by diff.)	18.38

H/C RATIO	0.66
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TABLE 9

ANALYSIS OF START-UP SOLVENT FOR CMSL-04

HRI No.	L-809
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°API Gravity	6.3
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ELEMENTAL ANALYSIS, Wt%

Carbon	88.62
--------	-------

Hydrogen	9.32
----------	------

Sulfur	0.58
--------	------

Nitrogen	0.18
----------	------

H/C RATIO	1.26
-----------	------

ASTM D-1160 DISTILLATION, °C

IBP	311
-----	-----

5 V%	335
------	-----

10 V%	343
-------	-----

20 V%	360
-------	-----

30 V%	378
-------	-----

40 V%	387
-------	-----

50 V%	406
-------	-----

60 V%	425
-------	-----

70 V%	445
-------	-----

80 V%	488
-------	-----

88 V%	524
-------	-----

Weight Percents

IBP-343°C	9.71
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343-454°C	59.51
-----------	-------

454-524°C	15.24
-----------	-------

524°C+	15.15
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Loss	0.39
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TABLE 10

PROCESS PERFORMANCE FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	12	16
HOURS OF RUN	120	240	288	384
CATALYST AGE (Lb Dry Coal/Lb Cat)	185	315	377	507
REACTOR TEMPERATURES, °C				
K-1 TEMPERATURE	388	388	399	399
K-2 TEMPERATURE	427	427	438	437
SPACE VELOCITY (2 ND Stage Catalyst) (Lb Dry Coal/Hr/Ft ³ Cat)	43.4	42.1	51.4	52.3
(Kg Dry Coal.Hr/M ³ Cat)	697	676	825	839
RELATIVE SEVERITY IND	1.06	1.06	1.59	1.52
SOLVENT/COAL RATIO	1.2	1.2	1.2	1.2
1 ST STAGE GAS FEED, %SCF				
H ₂	100	14	16	70
CO	0	86	84	30
PROCESS PERFORMANCE, WT% (MAF COAL)				
COAL CONVERSION	86.82	87.63	90.80	86.81
524+ °C CONVERSION	84.63	84.54	87.25	82.00
C ₄ -524 °C DISTILLATE YIELD	57.64	58.52	58.93	53.26
HYDROGEN CONS	8.10	9.61	9.51	8.35
DENITROGEN , WT%	96.90	95.15	89.53	90.02
DESULFUR (ORG), WT%	99.03	97.20	95.43	97.28
NORMALIZED YIELDS, WT% (MAF COAL)				
C ₁ -C ₃ GASES	6.3	7.65	7.96	9.17
C ₄ -C ₇ GASES	3.64	4.11	3.27	5.45
IBP-177 °C	12.78	10.07	14.07	14.36
177-260 °C	9.95	9.11	11.59	9.76
260-343 °C	21.26	20.63	21.35	15.95
343-454 °C	8.28	12.67	7.44	6.56
454-524 °C	1.73	1.93	1.21	1.18
524+ °C	2.19	3.09	3.55	4.81
UNCONV COAL	13.18	12.37	9.20	13.19
WATER	22.34	21.49	23.00	21.27
CO	0.51	0.51	0.52	0.51
CO ₂	4.85	5.02	5.15	5
NH ₃	0.90	0.83	0.94	0.91
H ₂ S	0.20	0.14	0.26	0.23

TABLE 11

ANALYSIS OF 1ST STAGE SOH FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	13	17
YIELD, W% DRY COAL	17.6	16.8	14.3	14.0
GRAVITY, °API	31.3	25.2	25.8	30.5
IBP, °C	88.3	95.0	81.7	67.2
FBP, °C	427	431	433	426

ASTM D-86 DISTILLATION, WT%

IBP-177 °C	34.52	23.1	25.8	40.0
177-260 °C	9.43	10.5	11.2	11.3
260-343 °C	26.70	27.5	26.6	22.0
343+ °C	28.65	38.3	36.0	26.1
Loss	0.70	0.6	0.4	0.6

ELEMENTAL ANALYSIS, WT%

Carbon	86.62	86.84	86.59	85.83
Hydrogen	12.67	11.67	11.5	11.78
Sulfur	0.0147	0.0428	0.0471	0.0410
Nitrogen	0.0747	0.1960	0.2941	0.2965

H/C RATIO	1.76	1.61	1.59	1.65
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TABLE 12
ANALYSIS OF ASOH FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	13	17
YIELD, W% DRY COAL	21.6	18.8	19.0	15.5
GRAVITY, °API	24.4	22.8	22.6	21.6
IBP, °C	160	143	137.2	132
FBP, °C	366	361	362	364

ASTM D-86 DISTILLATION, WT%

IBP-177 °C	4.85	5.1	5.9	6.9
177-260 °C	10.13	21.7	24.1	23.3
260-343 °C	70.48	64.8	62.1	60.8
343+ °C	12.45	8.1	7.7	8.5
Loss	2.09	0.3	0.2	0.5

ELEMENTAL ANALYSIS, WT%

Carbon	87.75	87.59	87.61	87.14
Hydrogen	11.89	11.59	11.24	11.06
Sulfur	0.0038	0.0079	0.0106	0.0115
Nitrogen	0.0435	0.1345	0.3067	0.3412

H/C RATIO	1.63	1.59	1.54	1.52
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TABLE 13

ANALYSIS OF 2ND STAGE SOH (HYDROTREATER OUTLET) FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	13	17
YIELD, W% DRY COAL	41.3	41.6	50.4	41.0
GRAVITY, °API	31.9	31.1	32.1	33.5
IBP, °C	84.4	77.2	54.4	60.0
FBP, °C	389	401	407	396

ASTM D-86 DISTILLATION, WT%

IBP-177 °C	30.48	25.7	28.3	35.0
177-260 °C	19.63	18.2	20.2	20.7
260-343 °C	41.11	39.9	37.6	33.1
343+ °C	7.97	15.7	13.2	10.6
Loss	0.81	0.5	0.7	0.6

ELEMENTAL ANALYSIS, WT%

Carbon	86.65	87.19	87.07	86.31
Hydrogen	12.92	12.73	12.55	12.60
Sulfur	0.0037	0.0045	0.0041	0.0031
Nitrogen	0.0022	0.0036	0.0035	0.0032
H/C RATIO	1.79	1.75	1.73	1.75

TABLE 14
ANALYSIS OF PFL FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	13	17
GRAVITY, °API	10.0	5.5	1.0	-4.7
IBP, °C	299	295	293	301
ASTM D-1160 DISTILLATION, WT%				
IBP-343 °C	6.2	6.87	5.14	6.09
343-454 °C	53.0	49.95	46.73	39.25
454-524 °C	18.5	17.52	17.01	14.70
524+ °C	22.0	25.36	30.65	39.70
Loss	0.3	0.30	0.47	0.26
ELEMENTAL ANALYSIS, WT%				
Carbon	89.83	89.93	90.64	91.2
Hydrogen	10.22	9.16	8.39	7.46
Sulfur	0.01	0.047	0.06	0.047
Nitrogen	0.21	0.29	0.43	0.57
H/C RATIO	1.37	1.22	1.11	0.98
CCR, WT%	29.22	40.72	55.98	65.42
CYCLOHEXANE INS, WT%	0.29	1.92	8.94	17.63
TOLUENE INS, WT%	0.10	0.18	0.88	1.98

TABLE 15
ANALYSIS OF PFC FOR CMSL-04

CONDITION	1	2	3	4
PERIOD	5	10	13	17
ELEMENTAL ANALYSIS, WT%				
Carbon	70.71	70.33	65.55	68.03
Hydrogen	6.09	5.46	4.69	4.65
Sulfur	0.841	1.01	1.06	0.999
Nitrogen	0.59	0.73	0.60	0.61
H/C RATIO	1.03	0.93	0.86	0.82
COMPOSITION, WT%				
Quinoline Insolubles	54.78	54.59	54.26	51.42
Ash (quinoline filtration)	18.29	19.26	24.36	16.18
ASTM Ash	17.85	18.55	23.99	22.73
S in Ash	4.15	4.13	3.80	3.87

TABLE 16
CO CONVERSION AND WATER BALANCE

PERIOD	5	10	12	16
WATER BALANCE, BM				
WATER INJECTED	22670	16498	20351	19849
MOISTURE IN COAL	5087	4936	6046	6133
COAL OXYGEN WT%	18.74	18.74	18.74	18.74
ASSUMED DEOXYGEN (%)	85.6	96.0	86.0	84.0
WATER FROM OXYGEN	2699	2937	3223	3193
TOTAL FEED WATER	30456	24371	29620	29175
SOH H2O COLLECTED	30456	17171	21544	27500
WATER CONSUMED	0	7200	8076	1675
WATER CONSUMED IN CO SHIFT REACTION				
CO REACTED	0	11173	11834	3227
EQUIVALENT WATER	0	7182	7607	2074
CO2 PRODUCED	0	17637	20892	3167
EQUIVALENT WATER	0	7215	8547	1296
AVERAGE WATER CONS	0	7199	8077	1685
CO CONVERSION, WT%	NA	68.5	74.5	93.2

TABLE 17

CMSL-05 RUN PLAN: EFFECT OF INTERSTAGE SEPARATION

ILLINOIS NO. 6 CROWN II MINE COAL (HRI-6107)

Back Pressure: 17.2 MPa

Catalyst: Shell-317, HRI-5394 Both Stages

Condition		1	2	3	4	4-EXT
Period		1-5	6-8	9-11	12-15	16A
Feed Gas:	1st Stage	H2	H2	H2	CO/H2	CO/H2
	2nd Stage	H2	H2	H2	CO/H2	CO/H2
Temperature [F]	1st Stage	415	415	402	402	402
	2nd Stage	043	443	429	429	429
	Hydrotreater	357	379	379	379	379
Space Velocity [Kg coal/h/m ³ cat.]		1070	1070	720	720	720
	[lb coal/h/ft ³ cat.]	67	67	45	45	45
Solvent/Coal Ratio		0.9	0.9	0.9	0.9	No Coal
Sulfur Additives to 2nd Stage, w% dry coal		0	2	0	0	0

TABLE 18

Effect of Increased Catalyst Replacement Rate

	Base Case	Alternate
Coal Feed Rate, T/D	9,800	11,000
Catalyst Replacement Rate, lb/T	2.7	4.5
Product, B/D	49,962	48,199
Total Plant Investment, \$MM (1)	2,223	2,416
Product Cost, \$/Bbl		
	4.68	4.68
Coal	6.41	6.41
Gas, Electric, etc.	2.79	3.80
Catalyst & Chemicals	3.11	2.76
Fixed Costs	23.74	23.08
Capital Related	2.16	2.16
By-product Credit		
	38.57	38.57
Total		
Equivalent Crude Price	32.50	32.50

(1) Note: 1991 US Gulf Coast

CMSL-03 Inspection of TBP Fractions

Weight Distribution [W%]

Period 5 Period 9 Period 12 Period 15

FIGURE 1

CMSL-03 Inspection of TBP Fractions

Hydrogen Content [W%]

FIGURE 2

FIGURE 3a

FIGURE 3b

FIGURE 4a

FIGURE 4b

HRI SYNGAS - H₂ TWO-STAGE COAL LIQUEFACTION SCHEME

CMSL-04 (RUN 227-81) DAILY MATERIAL BALANCE

FIGURE 6

CMSL-04 (RUN 227-81) TEMPERATURE VS PERIOD

—■— REACTOR 1 —▲— REACTOR 2

FIGURE 7

FIGURE 8

CMSL-04 (RUN 227-81) SPACE VELOCITY

CMSL-04 (RUN 227-81) OVERALL PROCESS SEVERITY

FIGURE 9

FIGURE 10

CMSL-04 (RUN 227-81) COAL CONVERSION

CMSL-04 (RUN 227-81) RESID CONVERSION & YIELD

FIGURE 11

FIGURE 12

CMSL-04 (RUN 227-81) GAS YIELDS

CMSL-4 (RUN 227-81) DISTILLATE YIELD

FIGURE 13

CMSL-04 (RUN 227-81) SULFUR CONTENT

FIGURE 14

CMSL-04 (RUN 227-81) NITROGEN CONTENT

FIGURE 15

CMSL-04 (RUN 227-81) H/C RATIO

FIGURE 16

CMSL-04 (RUN 227-81) PFL PROPERTIES

FIGURE 17

CMSL-04 (RUN 227-81) PFL INSOLUBLES

FIGURE 19

COMPARISON CMSL-3 AND CMSL-4 SPACE VELOCITY

COMPARISON CMSL-3 AND CMSL-4 OVERALL SYSTEM SEVERITY

FIGURE 20

FIGURE 21

COMPARISON CMSL-3 AND CMSL-4 CATALYST AGE

COMPARISON CMSL-3 AND CMSL-4 COAL CONVERSION

FIGURE 22

COMPARISON CMSL-3 AND CMSL-4 524+C CONVERSION

FIGURE 23

COMPARISON CMSL-3 AND CMSL-4 H2 CONSUMPTION

FIGURE 25

COMPARISON CMSL-3 AND CMSL-4 CO CONVERSION

FIGURE 26

COMPARISON CMSL-3 AND CMSL-4 C4-524C DISTILLATE YIELD

COMPARISON OF CMSL-3 AND CMSL-4 DISTRIBUTION OF DISTILLATE YIELD

FIGURE 27