

TWO-STAGE COAL LIQUEFACTION PROCESS DEVELOPMENTS AT WILSONVILLE

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INTRODUCTION

The Advanced Coal Liquefaction R & D Facility at Wilsonville, Alabama has been operating for over 14 years to develop alternate technologies for producing low cost fuels. A recently completed close-coupled integrated (CC-ITSL) processing mode was an important development in the Wilsonville program for making clean distillate fuels. In the CC-ITSL mode the two reactors were directly coupled without any pressure letdown. Interstage cooling was done only to the extent required to control the second reactor temperature. Such close-coupled operation should offer several process benefits such as increased overall thermal efficiency, reduced potential for retrogressive reactions which may take place in the absence of hydrogen at longer residence times, and improved product quality.

This paper is focused on two-stage coal liquefaction process performance with close-coupled reactors. Results are presented for two runs: Run 253 and Run 254. Run 253 processed high ash Illinois No. 6 coal and used Shell 317 1/20" trilobe bimodal catalyst in both reactors. Product yield and product quality data are discussed. Catalyst performance comparisons for Shell 317 (Run 253) and Amocat 1A and 1C (Run 251-I and Run 252) are made in terms of distillate yields, catalyst deactivation and catalyst replacement rates.

Run 254 processed low ash Ohio No. 6 coal and used Shell 317 1/20" trilobe bimodal catalyst in both reactors. The effects on process performance of the low ash feed coal, high resid recycle, high second stage reaction temperature and high space velocity are presented. The effect of the coal type on the distillate yield is discussed in comparisons of Illinois No. 6 and Ohio No. 6 coals. The first stage catalyst equilibrium activity level is discussed. Catalyst performance data at different second stage reaction temperatures are presented in terms of catalyst deactivation and catalyst replacement rates. Optimum and cascading catalyst requirements are discussed.

Catalyst properties used in the close-coupled integrated runs (Run 250-254) are listed in Table 1. Catalyst performance in the close-coupled configuration has been extensively studied and reported in the previous EPRI and DOE contractors' conferences (1,2,3,4). Relative process economics evaluations (5), configuration effects on process performance (6) and developments of ITSL models (7) were reported elsewhere.

PROCESS DESCRIPTION

A block flow diagram of the catalytic-catalytic CC-ITSL process is shown in Figure 1. The process consists of a slurry preparation step and two catalytic reaction stages followed by hydrotreated solvent recovery and critical solvent deashing systems. The system was integrated by the recycle of CSO (Critical Solvent Deashing) resid, hydrotreated solvent, and low-pressure flash bottoms containing ash, unconverted coal, hydrotreated resid, and hydrotreated solvent. Solids recycle allowed an increased concentration of solids in the CSD feed and hence a lower CSD feed rate. The solvent recovery system consists of atmospheric flash and vacuum flash equipment. The recycle distillate was fractionated in a vacuum tower to reduce the light ends (650°F- fraction) in the recycle solvent.

RUN 253 RESULTS

Catalyst Activity and Deactivation

Run 253 was the first CC-ITSL run with a significantly extended operation of catalyst addition to the first stage (2 lb/ton MF coal). The operation time with the catalyst addition was about 27 days. The impact of the catalyst addition was very significant. The C₄+ distillate yield is approximately higher by 21 wt % MAF coal at 3000 lb resid + CI/lb catalyst of the second stage catalyst age, compared to the yield linearly extrapolated from the batch deactivation data.

A slight decline of the slope indicates that 2 lb/ton addition is not sufficient at 385 MF lb/hr coal feed rate to maintain catalyst activity to achieve the "all-distillate" product slate with 69 wt % MAF C₄+ distillate. This observation is further substantiated in the catalyst requirement data for the resid extinction mode. Using the estimated catalyst activity data approximately 4.0 lb/ton catalyst addition is projected for the "all-distillate" product slate with a 69 wt % distillate yield. At 480 MF lb/hr high coal feed rate, the catalyst requirement is doubled to 8.1 lb/ton (Table 2).

These higher catalyst requirements for Run 253 with Shell 317 catalyst (approximately 2 to 3 times higher than those for Run 251-I and 252 with Amocat 1A and 1C catalysts) (Tables 2 and 3) were attributed to less catalyst charges in both reactors by 12 wt % and lower TSL catalyst activity as compared in Figures 2 and 3. Shell 317 catalysts for Run 253B showed a lower C₄+ distillate yield by 10 wt % MAF coal at a similar catalyst volume basis and by 3 wt % MAF at the same catalyst weight basis (Figure 2), compared to Amocat 1A/1C catalysts used in Run 251-IC. In addition, Amocat 1C/1C catalysts for Run 252B showed a higher C₄+ distillate yield by 5 wt % MAF coal at a similar catalyst volume and the same catalyst weight basis, compared to Amocat 1A/1C catalysts used in Run 251-ID and IE (Figure 3).

Each stage catalyst activity was much different from TSL catalyst activity trends discussed before; Run 253 Shell 317 catalyst activity was lower in the first stage and higher in the second stage reactor, compared to Run 251-I and 252 Amocat 1A and 1C catalysts. Run 253 Shell 317 catalyst deactivation rates for both reactors appeared to be very similar to those for Runs 251-I and 252 Amocat 1A and 1C catalysts (2,3).

Catalyst selectivities in TSL hydrogenation and heteroatom removal were discussed and compared in Run 253 Report (8). Selectivities were studied with relation to TSL hydrogen consumption, which is an indication of TSL process severity. Run 253 with Shell 317 and Run 252 with Amocat 1C catalyst showed slightly higher selectivities for the potential liquid yield (C₄+ resid) production by 1 to 3 wt % MAF coal, compared to Run 251-I with Amocat 1A and 1C catalysts. This was primarily attributed to less C₁-C₃ gas make. Overall heteroatom (N, S, O) removal was similar for all three runs. However, Run 253 with Shell 317 and Run 252 with Amocat 1C catalyst showed higher nitrogen removal activity. Analyses of aged Shell 317 catalysts used for Run 253 indicated the first stage as a guard bed by removing catalyst deactivation and poison material before going to the second stage. The second stage aged catalyst showed lower carbon deposits by 6 wt % of catalyst than the first stage and a higher hydrogenation activity value by 50 (m moles of hydrogen consumed) in hydrogenation of naphthalene. This guard bed effect was also commonly observed with other catalysts such as Amocat 1A and 1C used for Run 251-I and Run 252 (2,3).

Product Quality

Run 253 product quality data are listed in Table 4. During Run 253 the recycle distillate was fractionated in a vacuum column to reduce the light ends in the recycle solvent. The naphtha and distillate yields were about the same in all the periods (C, E and F). The gas-oil yield was the lowest in the concentrated coal slurry test. Other than this there were no major differences in product quality among various Run 253 tests. Preliminary laboratory analyses of distillate products indicated a significant reduction of the distillate product end point, by 30 to 80°F. This was achieved by increasing the recycle of 650°F+ by 8 wt % in the recycle process solvent (from 89 to 97 wt %).

RUN 254 RESULTS

Run 254 processed low ash Ohio No. 6 coal in a catalytic-catalytic close-coupled configuration. Unwashed high ash Ohio coal with about 10-12 wt % ash was used for the startup process solvent equilibration at 500 MF lb/hr coal feed rate. Low ash Ohio coal with 6 wt % ash was prepared by washing the run-of-the-mine coal with heavy media. During the operation with 40 wt % resid level, the "all-distillate" product slate could not be achieved because of high potential distillate (C₄+ resid) yields, 76-78 wt % MAF coal, and high excess resid yields, 8-10 wt % MAF processing the low ash Ohio No. 6 coal. In order to achieve one of the run objectives of the resid extinction, it was recognized that a higher catalyst activity level is required, since Ohio No. 6 coal apparently has lower resid conversion activity than Illinois No. 6 coal. The higher level of resid recycle could also improve significantly resid conversion, as observed in the previous runs (8,9), that is, 0.3 wt % MAF coal distillate yield increase per 1 wt % resid content increase in the recycle process solvent.

Low Ash Ohio No. 6 Coal

Advantages of the low ash Ohio No. 6 coal with 6 wt % as opposed to the high ash with 10 wt % are threefold as shown in Tables 5 and 6; (1) an increase of the potential liquid yield (C₄+ resid) by 6 wt % MAF coal, (2) an increase of coal conversion by 3 wt % MAF coal, and (3) a decrease of organic rejection by 8 wt % MAF coal. The increase of coal conversion is probably due to removal of less reactive

coal components such as inertinites by cleaning with heavy media (10). The low UC yield with the low ash coal reduced the organic rejection and consequently increased the potential liquid yield. The ratio of the organic rejection to the UC yield was 2.68, same for both periods, which indicates that the UC yield is a primary variable affecting the organic rejection.

Effect of Second Stage Reaction Temperature

The effect of the second stage reaction temperature change from 760°F to 790°F was studied during periods 254C-D and 254F-G. The observed C₄+ distillate yield increase due to the temperature increase was very significant, about 11 wt % MAF coal during 254C-D. The hydrogen consumption was also increased in a relatively small degree from 6.1 to 6.7 wt % MAF, which gives a significantly better hydrogen efficiency at 790°F (10.3 vs 9.5 lb C₄+ dist/lb H₂ consumed). C₁-C₃ gas make and organic rejection were similar for both periods. The C₄+ distillate yield response during 254F-G due to the second stage reaction temperature change from 760°F to 790°F was 8 to 9 wt % MAF coal, which is slightly lower than that during 254C-D.

Activation energy for the second stage catalyst was calculated based on the C₄+ distillate response during 254C-D. The activation energy was approximately 43,300-53,750 Btu/lb mole for the second stage reaction temperature range of 760°F to 790°F. A similar value of 47,800 Btu/lb-mole was also observed during 254F-G for the periods at a higher level of resid recycle about 50 wt %. This activation energy processing Ohio No. 6 coal is slightly higher than that processing Illinois No. 6 coal (42,300 Btu/lb mole for the temperature range of 720°F to 770°F) (11).

The effect of the second stage reaction temperature change from 790°F to 810°F was studied during period 254J-K. The estimated C₄+ distillate response due to the second stage reaction temperature change by 20°F was 6 wt % MAF coal, which corresponds to the activation energy 69,000 Btu/lb mole for the temperature range of 790°F to 810°F. This activation energy is much higher than that observed for the temperature range of 760°F to 790°F (47,800 Btu/lb mole).

The high second stage reaction temperature at 810°F (254K) lowers hydrogen efficiency and distillate selectivity by producing more C₁-C₃ gas make (2 wt % MAF coal), compared to at 790°F (254J).

Effect of High Resid Recycle

Very careful experimentation for periods 254G-H was conducted to evaluate the effect of 50 wt % resid recycle in the recycle process solvent on process performance. Period 254D was selected for comparisons with 40 wt % resid recycle. Significant advantages of the 50 wt % high resid recycle are found in process performance improvements; (1) 6 wt % MAF coal increase in the C₄+ distillate yield and (2) a higher hydrogen efficiency (11.1 vs 10.3 lb C₄+ dist/lb H₂ consumed). C₁-C₃ gas make and organic rejection were similar for both periods. The "all-distillate" product slate became possible for period 254G-H with more resid conversion through higher recycle of the reactive resid. The C₄+ distillate increase was slightly higher than the projected (5 wt % MAF). The economic impact in the commercial plant design will be slightly compensated due to the higher resid recycle by 40% and the higher slurry rate by 10%, although the production rate of the distillate is increased by about 9%.

During period 254G the C₄+ distillate yield was 78 wt % MAF coal, the highest ever observed at Wilsonville, the lowest organic rejection 7 wt % MAF (254I), and the highest coal conversion 97 wt % MAF. The hydrogen efficiency was very high, 11.3 lb C₄+ dist/lb H₂ consumed. This highly improved distillate production can be achieved by a moderate catalyst addition below 5 lb/ton of MF coal. Three main reasons can be attributed to these excellent process achievements; (1) high coal conversion activity of the low ash Ohio coal, (2) high distillate selectivity at 790°F of the second stage reaction temperature, and (3) high recycle of the reactive resid.

Effect of Space Velocity

Three different coal feed rates, 440, 365 and 300 MF lb/hr were studied during periods 254H, I and J. The first stage catalyst ages were similar approaching the equilibrium catalyst age of 610 lb MF coal/lb catalyst due to the catalyst addition at 3 lb/ton of MF coal, while the second stage catalyst ages were different, gradually increasing, due to the batch mode of operation. The C₄+ distillate yield was 72 wt % MAF coal at 365 MF lb/hr coal feed rate (254H) and increased to 76 wt % MAF at 300 MF lb/hr (254I) and then significantly decreased to 64 wt % MAF at 440 MF lb/hr (254J). The response for 300 MF lb/hr (254I) was relatively smaller than the expected (4 wt % vs 8 wt %). This was due to the catalyst activity decline in the second stage at a low coal feed rate. Similar observations were made during Run 253 processing Illinois No. 6 coal. Catalyst requirements for 440 MF lb/hr will be high, not desirably for the practical operation, more than 5 lb/ton of MF coal, which is suggesting the operation at a higher second stage temperature than 790°F.

The effect of the coal feed rate decrease from 440 to 375 MF lb/hr at 810°F of the second stage reaction temperature was studied during period 254K-L. The estimated C₄+ distillate response due to the coal feed rate change was 8 wt % MAF coal. C₁-C₃ gas make and organic rejection were similar for both periods. Catalyst requirements for 440 MF lb/hr is estimated to be below 5 lb/ton of MF coal (Table 7), much less than at 790°F of the second stage reaction temperature. This was due to the high second stage catalyst activity at 810°F (Figure 6).

Catalyst Activity and Deactivation

Two-stage catalyst activity was compared in Figure 4, based on C₄+ distillate yield trend data. Ohio No. 6 coal with both high (254B) and low (254C) ash contents significantly reduced the C₄+ distillate yield by 7 wt % MAF coal in the batch mode of operation, compared to Illinois No. 6 coal with a high ash content (253D). In addition, in comparisons of 254D and 253EFG Ohio No. 6 coal required a higher catalyst addition (3 vs 2 lb/ton of MF coal) and a higher second stage reaction temperature (790°F vs 760°F) for a similar C₄+ distillate yield (70 wt % MAF coal). These indicate that Ohio No. 6 coal has lower resid conversion activity than Illinois No. 6 coal. Run 254B in the batch operation processing Ohio No. 6 coal showed a similar two-stage catalyst deactivation rate to that for Run 253D processing Illinois No. 6 coal.

Catalyst activity trends for Run 254 operation with a high level of resid recycle (50 wt %) were analyzed for the two-stage system, first stage and second stage, based on two-stage resid make and resid + UC conversion. The first stage equilibrium activity level with 3 lb/ton of MF coal catalyst addition was achieved at the catalyst age near to 550 lb MF coal/lb catalyst (calculated equilibrium age = 610 lb MF coal/lb catalyst), as illustrated in Figure 5. The first stage catalyst ages,

550 and 610 lb MF coal/lb catalyst, are corresponding to the second stage catalyst ages, 2000 and 3000 lb resid + CI/lb catalyst, respectively. In Figure 5 the measured first stage catalyst activity values with the catalyst addition are very close to the theoretically calculated by using the estimated batch deactivation data, which confirms that the first stage batch catalyst activity data in Figure 6 are reasonable for further catalyst requirement calculations.

Because of the continuous catalyst addition in the first stage, the two-stage catalyst deactivation rate was gradually decreased and became primarily dependent on the second stage deactivation rate. Figure 6 shows a graphical presentation of differences in the second stage catalyst activity at different reaction temperatures, 760°F, 790°F and 810°F. Second stage reaction temperature increases from 760°F to 790°F, and further to 810°F significantly increased catalyst activity levels, although the second stage deactivation rate for 810°F was significantly higher (but for 790°F, slightly higher) than for 760°F. At the same reactor temperature of 810°F for both stages, the second stage catalyst activity was higher than that of the first stage and the second stage catalyst deactivation rate was lower. These observations at 810°F of the second stage reaction temperature need to be substantiated because of limited data points at high catalyst ages (1600-2100 lb MF coal/lb catalyst of second stage catalyst ages).

Catalyst requirements at 790°F of the second stage reaction temperature were 3.1 to 4.6 lb/ton of MF coal to achieve the "all-distillate" product slate (77 wt % MAF coal C₄+ distillate yield) for 300 to 370 MF lb/hr coal feed rates (Table 7). Effects of the 810°F of the high second stage reaction temperature on catalyst performance were very substantial (Table 7 and Figure 7): (1) increased coal throughput, (2) increased catalyst activity, (3) lower catalyst requirements and (4) lower distillate selectivity (C₄+ distillate yield decreased to 74 wt % MAF coal because of a high C₁-C₃ gas make) (Figure 8).

Optimum Catalyst Requirements and Cascading

Catalyst replacement rates reported in Table 7 and Figure 7 were calculated, assuming to achieve each stage performance experimentally measured at the Wilsonville pilot plant. Considering the significant differences in each stage catalyst activity due to different reaction temperatures employed for CC-ITSL configuration studies, one might visualize that there would be an optimum catalyst requirement for a fixed coal feed rate with resid extinction only by shifting each stage catalyst requirement loading. For instance, Figure 9 presents graphically optimum catalyst requirements for 300 and 370 MF lb/hr coal feed rates. The variable selected as X-coordinate is the first stage resid + UC conversion. The plots are simply showing that the optimum catalyst requirement is not only a function of the coal feed rate but also of the first stage resid + UC conversion. In addition the operation range for the optimum is very narrow within 1 to 3 wt % variation of the first stage resid + UC conversion. Data in Figure 9 were generated without considering catalyst residual activities and based on 77 wt % C₄+ distillate yield with resid extinction. Organic rejection was assumed to be 8 wt % MAF coal. Operating conditions were 810°F first stage temperature, 790°F second stage temperature, 2.3 solvent/coal ratio, 50 wt % resid and 12 wt % CI in the recycle process solvent.

Other interesting points can be made from plots in Figure 9, that is, (1) catalyst cascading is practically possible with a possible reduction of catalyst addition about 1 to 2 lb/ton of MF coal (this saving can be considered as maximum) and (2) Wilsonville plant experimentations were conducted at operating conditions of optimum catalyst requirements (Run 254I and GH).

Four different cases were studied for optimum catalyst requirements and cascading possibility. Results were generated with considering catalyst residual activities and are summarized below. The catalyst addition for the cascading is listed for the range of interest. The actual amount is dependent on the cascading catalyst activity, which needs to be experimentally determined. The upper limit has only theoretical meaning for an equal amount of the catalyst addition for both stages with zero cascading catalyst activity. This upper limit shows a higher catalyst addition than the optimum.

Case	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>
Base data	254I	254GH	254L	254K
Coal feed, MF lb/hr	300	370	375	440
Second stage T, °F	790	790	810	810
C ₄ + dist with resid extinction wt % MAF coal	77	77	74	73
<u>Catalyst addition</u> lb/ton of MF coal				
<u>experimentally measured</u>				
<u>first stage</u>	2.4	3.3	3.3	3.8
<u>second stage</u>	0.7	1.3	0.2	0.7
<u>total</u>	3.1	4.6	3.5	4.5
<u>optimum</u>				
<u>first stage</u>	2.4	3.3	0.3	1.1
<u>second stage</u>	0.5	1.0	1.2	1.8
<u>total</u>	2.9	4.3	1.5	2.9
<u>cascading</u>				
<u>first stage</u>	1.8	2.6	0.9	1.6
<u>second stage</u>	1.8	2.6	0.9	1.6
<u>total</u>	1.8-3.6	2.6-5.2	0.9-1.8	1.6-3.2

As already pointed out, in Cases 1 and 2 at 790°F of the second stage reaction temperature, optimum catalyst requirements only improve slightly about 0.2-0.3 lb/ton of MF coal, since Wilsonville pilot plant data were generated near at the optimum operating conditions. Another significant improvement can be made by doing cascading from the second stage to the first stage reactor, potentially reducing the catalyst requirement by 1.1 to 1.7 lb/ton of MF coal.

Very unusual striking results were revealed for the operation at 810°F of the second stage reaction temperature. Optimum calculations are suggesting that Wilsonville plant experimentations were not conducted at operating conditions of

optimum catalyst requirements (Run 254L and K) and therefore, if operated at optimum, catalyst savings will be about 1.6 to 2.0 lb/ton of MF coal. Additional reduction of catalyst addition can be made by 0.6 to 1.3 lb/ton of MF coal by doing catalyst cascading (this can be considered as a maximum potential). These observations need to be substantiated by further experiments at young catalyst ages, which data are not available at the present time. The assumption made for the second stage catalyst activity at 810°F should be further investigated and experimentally proved, before making definite conclusions of significant process performance achievements at 810°F of the second stage reaction temperature.

A significant advantage of 810°F of the second stage reaction temperature as opposed to 790°F lies in an increase of the coal throughput by 47% and the C₄+ distillate production rate by 40% (Cases 1 and 4). Optimum catalyst requirements for both Case 1 and 4 were same, 2.9 lb/ton of MF coal. These process performance improvements will impact tremendously the process economic valuation for commercialization and reduce significantly the selling price of the distillate product, if other critical variables in the economic evaluation are similarly affecting the results for both cases.

SUMMARY

Several noteworthy accomplishments in 1987/88 are listed below:

- (1) Successful operations processing Illinois No. 6 and Ohio No. 6 coals were demonstrated by using Shell 317 catalyst in the close-coupled ITSL catalytic-catalytic configuration.
- (2) Interstage separation was eliminated, resulting in significant reduction in capital costs.
- (3) Best process performance was achieved processing low ash Ohio No. 6 coal with 78 wt % MAF coal distillate yield, 97 wt % coal conversion and 7 wt % organic rejection.
- (4) High second stage reaction temperatures, 790°F and 810°F, were investigated, resulting in significantly increased coal throughput and distillate production due to high catalyst activity and low catalyst requirements.
- (5) The increased reactive resid recycle improved the distillate production.
- (6) The increased gas-oil recycle produced lighter and better quality distillates.
- (7) The close-coupled ITSL process data-base was significantly expanded for processing bituminous coals.
- (8) The process economics was significantly improved for future commercialization of the coal liquefaction process.

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Table 1

Wilsonville CC-ITSL catalyst properties

catalyst RUN(s)	Shell 324 250,251	Amocat 1C 250-252	Amocat 1A 251	Shell 317 253,254
shape	(----- cylindrical -----)			trilobe
size	1/32"	(----- 1/16" -----)		1/20"
Ni (wt %)	2.7	2.3		2.7
Co			2.5	
Mo	13.2	10.4	9.8	11.6
surface area, m ² /g	185	190	235	235
pore volume, cc/g	0.48	0.85	0.80	0.75
pore size distribution	unimodal	(----- bimodal -----)		
compactd bulk density, lb/ft ³	54	42	41	36

Table 2. Shell 317 Trilobe Catalyst Replacement Rates

run no.	253D	253B
coal feed (mf lb/hr)	385	480
WHSV (hr ⁻¹)	3.9	5.1
second stage reactor temperature (°F)	760	760
TSL C4+ distillate (% maf coal)	69	68
catalyst addition (lb/ton mf coal)	4.0	8.1

Table 3. Amocat 1A and 1C Catalyst Replacement Rates

run no.	251-IE	251-ID	251-1C	252C
catalyst (1st/2nd stage)	1A/1C	1A/1C	1A/1C	1C/1C
coal feed (mf lb/hr)	300	385	480	385
WHSV (hr ⁻¹)	2.7	3.4	4.2	3.5
second stage reactor temperature (°F)	760	760	760	760
TSL C4+ distillate (% maf coal)	70	70	70	70
catalyst addition (lb/ton mf coal)	0.9	1.6	3.0	1.2

Table 4. Run 253 - Distillate Product Properties

distillation cut	%MAF ¹	elemental (wt%)					*API
		C	H	N ²	S	O (diff)	
Run 253C (40% coal slurry)							
naphtha (IBP-350°F)	19.9	84.80	14.94	0.02	0.06	0.18	46.7
distillate (350-650°F)	27.3	87.74	12.03	0.11	0.01	0.11	18.9
gas oil (650°F+)	18.2	88.84	10.63	0.29	0.01	0.23	5.6
Run 253E (no separator)							
naphtha (IBP-350°F)	14.6	85.35	14.45	0.04	0.07	0.00	44.1
distillate (350-650°F)	26.2	87.97	11.61	0.15	0.05	0.02	17.5
gas oil (650°F+)	27.0	89.57	9.97	0.37	0.02	0.10	2.9
Run 253F (ground catalyst & no separator)							
naphtha (IBP-350°F)	14.7	85.58	14.21	0.03	0.02	0.16	43.8
distillate (350-650°F)	22.4	87.83	11.71	0.13	0.01	0.32	18.1
gas oil (650°F+)	28.5	89.54	10.07	0.34	0.02	0.03	3.8

¹by simulated distillation on GC

*nitrogen by Kjeldahl

Table 5

run no.	TSL operating conditions	
	254B	254CD
first stage		
catalyst replacement (lb/ton MF coal)	none	3
average reactor temperature (°F)	(-----810-----)	(-----)
inlet hydrogen partial pressure (psf)	(-----2600-----)	(-----)
coal feed rate (lb/hr MF)	385	370
space velocity (lb feed/hr-lb cat)	(-----3.9-----)	(-----)
solvent-to-coal ratio	(-----2.0-----)	(-----)
solvent resid content (wt%)	(-----40-----)	(-----)
solvent Cl content (wt%)	(-----12-----)	(-----)
catalyst age [lb (resid + Cl)/lb cat]	850-1300	1150-1200
(lb MF coal/lb cat)	400-650	550-800
second stage		
average reactor temperature (°F)	760	760/790
space velocity (lb feed/hr-lb cat)	(-----3.7-----)	(-----)
feed resid content (wt%)	(-----41-----)	(-----)
catalyst age [lb (resid + Cl)/lb cat]	600-900	1100-1450
(lb MF coal/lb cat)	400-600	700-950

Table 6

run no.	TSL yield structures	
	254B	254CD
10% vs 6% coal ash		
potential liquid yield		
C ₄ + resid (% maf coal)	70	76
coal conversion (% maf coal)	94	97
energy content of feed coal rejected to ash conc. (%)	18	9
organics rejected to ash conc. (% maf coal)	18	8

Table 7. Shell 317 Trilobe Catalyst Replacement Rates

run no.	254I	254GH	254L	254K
coal feed (mf lb/hr)	300	370	375	440
WHSV (hr ⁻¹)	3.5	4.3	4.3	5.1
second stage reactor temperature (°F)	790	790	810	810
TSL C4+ distillate (% mf coal)	77	77	74	73
catalyst addition (lb/ton mf coal)	3.1	4.6	3.5	4.5

Figure 1

**CC-ITSL with solids recycle
catalytic-catalytic
close-coupled mode**

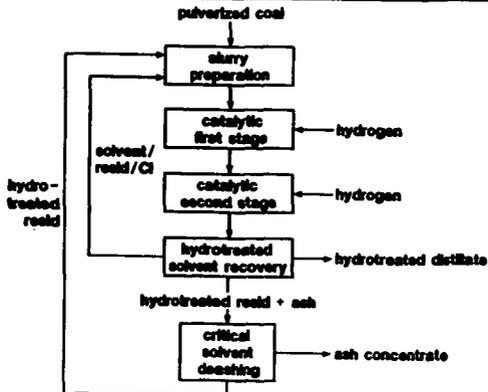


Figure 2. Amocat 1A/1C vs Shell 317 Catalyst

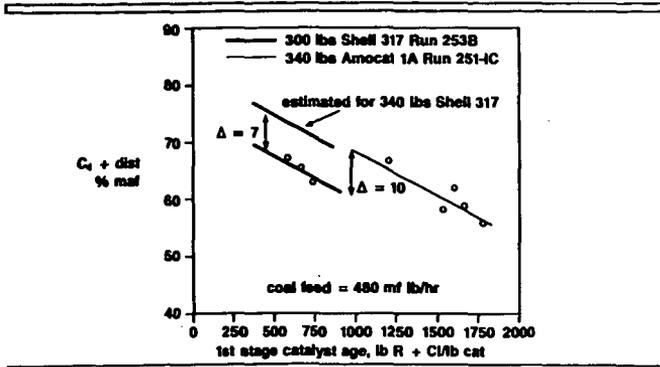


Figure 3. Amocat 1A/1C vs Amocat 1C/1C Catalyst

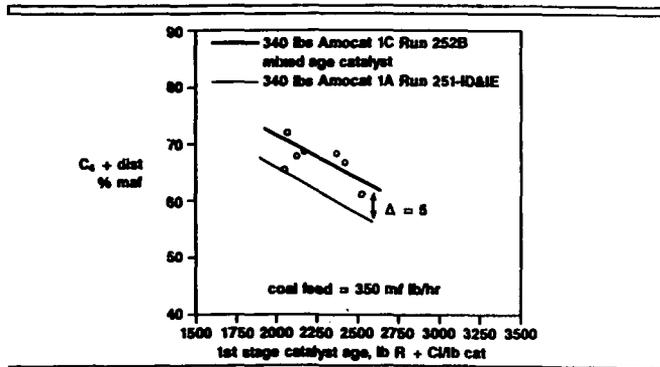


Figure 4. Illinois No. 6 vs Ohio No. 6 Coal

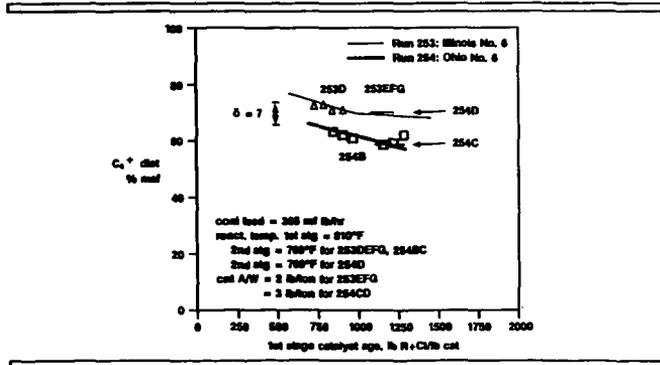
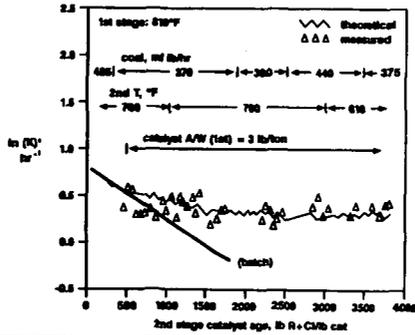


Figure 5

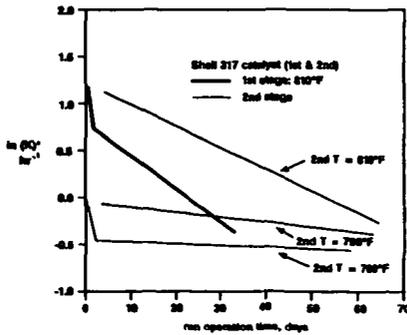
RUN 254 first stage Shell 317 aging with catalyst addition



* (meald + UC) conversion rate constant

Figure 6

catalyst aging - RUN 254



* (meald + UC) conversion rate constant

Figure 7. Run 254 - Catalyst Replacement vs 2nd Stage Temperature

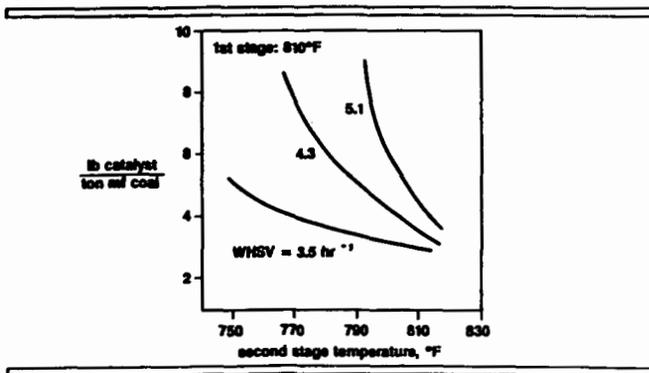


Figure 8. Run 254 - Distillate Selectivity vs 2nd Stage Temperature

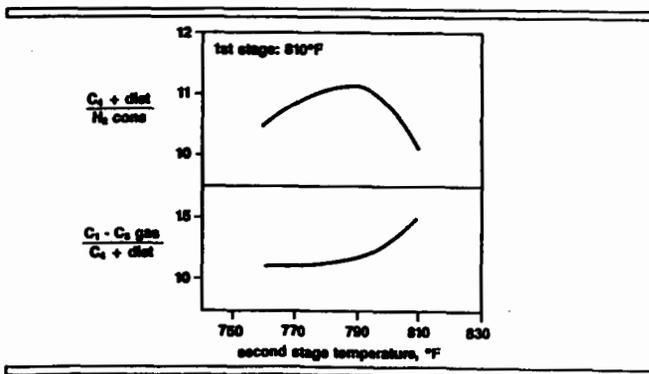


Figure 9. Run 254 - Optimum Catalyst Replacement Rate

