The Application of Lummus Antisolvent Deashing Technology to a Solvent Refined Coal Solution

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THE APPLICATION OF LUMMUS ANTISOLVENT DEASHING TECHNOLOGY TO A SOLVENT REFINED COAL SOLUTION

EPRI AF-234
(Research Project 524)

Final Report

September 1976

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I. Introduction

On July 7, 1975, C-E Lummus was authorized by the Electric Power Research Institute (EPRI) to investigate the application of Lummus' antisolvent deashing technology (1) to the solvent refined coal liquefaction process. Briefly, the work was divided into two tasks, namely: a pilot plant test and an economic evaluation.

For the first task, Lummus was to process a feedstock provided by EPRI's Wilsonville SRC pilot plant in its antisolvent deashing pilot unit. A relative economic comparison between the Lummus deashing process and rotary pressure precoat filtration was the objective of the second task on economic evaluation. This latter comparison was to be made on a scale equivalent to a commercial size solvent refined coal liquefaction complex.

II. Summary

An ash containing, Illinois No. 6 coal derived, coal liquefaction product solution prepared by the Wilsonville, Alabama SRC pilot plant has been successfully deashed in the Lummus antisolvent deashing pilot plant. An ash removal efficiency of 99.6% was achieved, which is equivalent to a recovered heavy (+699.9^{o}K or +800^{o}F) SRC product residual ash content of less than 0.1 wt. %.

(1) Sze, M. C. and G. J. Snell, "A New Process for Removing Ash From Coal Liquefied by Hydrogenation", Proc. of Am. Power Conf., 37, pp. 315-321 (1975).

About 77.2% of the total quinoline soluble heavy SRC product and 74.8% of the benzene insoluble/quinoline soluble heavy SRC product present in feedstock were recovered with the substantially ash-free overflow stream in the pilot plant deashing run.

An overall material balance, elemental balances, and deashing yield structure based on experimental data have been developed for the above deashing test run. In addition, bench-scale simulated delayed coking tests have been made with both an ash containing underflow derived feedstock sample and a Wilsonville supplied dry mineral residue sample.

A comparative economic evaluation of the Lummus antisolvent deashing process and rotary pressure precoat filtration deashing was made for a solvent refined coal liquefaction complex in which 9070 MT/D of (mf) coal are liquefied. The economic evaluation was based on an after tax 12% discounted cash flow rate of return and 100% owners equity. For the base case considered, the antisolvent deashing total manufacturing cost was found to be less by 20-26¢/10^6 Btu or \$1.37-1.78/Bbl SRC product than rotary pressure precoat filtration. The capital requirements of the above two deashing processes were found to be significantly different. A total installed cost (including a delayed coker) of \$57 x 10^6 and a total capital requirement of \$73.5 x 10^6 was estimated for antisolvent deashing facilities, whereas a total installed cost and total capital requirement of \$99x10^6 and \$131x10^6 respectively was estimated for filtration deashing. Deashing feedstock in the base case economic comparison was an SRC dissolver effluent liquid product solution in which only about 84% of the moisture and ash free feed coal (Illinois No. 6) to the dissolver was converted to a quinoline soluble form or liquefied. This deashing feedstock corresponds to the actual material successfully tested in the above mentioned antisolvent pilot plant run. As expected, the range of antisolvent deashing cost advantage cited above is a function of the assumed SRC yield for the filtration deashing case.

An underflow coking step was included in the antisolvent deashing case to insure complete recovery of paste solvent boiling range distillate (477.6/699.9°K or 400/800°F). This was considered necessary for the antisolvent deashing case because of the low degree of liquefaction achieved in the SRC dissolver. While a delayed coker was specified for cost estimating purposes, a fluidized-bed carbonizer similar to that operated during the CSF pilot plant development [2,3] work at Cresap, West Virginia would likely be a more economic alternative.

For the base filtration deashing case, optimum conditions were selected, which, in addition to a higher degree of liquefaction, specified also the following parameters: (1) filtration rate, 732 kg/hr/sq.m (150 lbs filtrate/hr. ft^2), (2) 0.10 kg. filter aid/kg. ash, and (3) 95% filter availability.

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(2) U.S. Office of Coal Research. Research and Development Report No.39, Interim Report No. 3, March 1969. (PB-234 129/AS) pp. 8-9, 37-39.

(3) U.S. Office of Coal Research. Research and Development Report No.39, Final Report, Aug. 1973. (PB-234 132/AS) p. 15.

The effect of the above operating parameters were evaluated and the resulting incremental costs were compiled in the form of a sensitivity table (Table 27). If the filter aid usage in the filtration base case was reduced to 0.05 kg filter aid/kg ash and filter reliability maintained at its optimistic 95% level (base case value), the total deashing cost advantage for antisolvent deashing is still about 12¢/10^6 kJ, (11¢/10^6 BTU) or $0.76/Bbl of SRC product lower than filtration deashing. Current Tacoma SRC filtration practice indicates that a filter reliability of the order of 60% might be more realistic at this time. For a filtration case in which filtration rate, filter aid usage, and filter reliability are 732 kg/hr./sq.m. (150 lbs/hr ft^2), 0.05 Kg. filter aid/kg ash, and 60% respectively; total antisolvent deashing costs are estimated to be lower by 35¢/10^6 kJ, (33¢/10^6 BTU) or $2.26/Bbl than filtration deashing.

When the SRC dissolver is operated under optimum conditions, the degree of liquefaction achievable is over 92% of the maf coal feed. Under these conditions, paste solvent self-sufficiency may be achieved without coking of the underflow in the antisolvent deashing case. This allows the elimination of the delayed coker, a substantial reduction in capital and operating costs over the base antisolvent deashing case studied in this report. It cannot be overemphasized that optimizing the dissolver operation can result in significant savings in the ash removal step.

III. Conclusions

Lummus antisolvent deashing technology is the preferred technically and economically viable alternative to rotary pressure precoat filtration when applied to the ash removal step of the Solvent Refined Coal (SRC) process. This is especially so when operability of the two systems are considered. Total deashing costs for antisolvent deashing are expected to be at least $1.37/Bbl SRC less than filtration deashing. Capital requirements associated with antisolvent deashing are significantly lower than those associated with filtration deashing. The relative economic incentives associated with antisolvent deashing would increase significantly if the operating reliability or on-stream time of the filtration deashing alternative were reduced from an optimistic base case value of 95% to 60-80%. In addition to the aforementioned relative economic advantages favoring antisolvent deashing, the recovered SRC product via antisolvent deashing can consistently provide a residual ash content of less than 0.1 wt.%.

IV. Recommendations

- The installation of an antisolvent deashing module should be considered for either or both of the U.S. operational SRC pilot plants.
- Consider applying antisolvent deashing techniques to highly converted or liquefied (over 92% maf coal) SRC derived feedstocks. The need for underflow coking may disappear entirely

in such a case. Antisolvent deashing cost can then be as much as \$2.35/Bbl of SRC product lower than the filtration method for ash removal.

V. Experimental Work

A. Feedstock

Several 55 gallon drums of SRC filter feedstock were shipped to the Lummus antisolvent deashing pilot plant located in Bloomfield, New Jersey in April, 1975. This prospective deashing feedstock was prepared at the EPRI sponsored SRC pilot plant located at Wilsonville, Alabama, from Illinois No. 6 high volatile bituminous coal containing about 10% ash. A dissolver coal paste feed consisting of approximately 1 part by weight of Illinois No. 6 coal and 2 parts by weight of paste solvent was used in the above preparation.

One of the above drums was found to be contaminated with water and was subsequently discarded. The drums actually used in the pilot plant deashing test run were individually melted and individually charged to an agitated blend tank. A composite sample was then prepared from the individual drum samples and designated as the composite feedstock sample for the deashing test run (DA-53-2). Table 1 is a compilation of analytical data obtained on the aforementioned feedstock sample. Table A-1 in Appendix A is a listing of the analytical methods employed.

TABLE 1

FEEDSTOCK INSPECTION DATA (DEASHING RUN DA-53-2)

Specific Gravity 336.0/288.8°K	1.103
Specific Gravity 371.0/288.8°K	1.085
Ash Content, Wt. %	3.55
Quinoline Insolubles, Wt.%	7.87
Benzene Insolubles, Wt. %	18.47
Carbon Content, Wt. %	84.69
Hydrogen Content, Wt. %	6.43
Nitrogen Content, Wt. %	0.92
Sulfur Content, Wt. %	0.62
Organic Oxygen (Via Diff.), Wt. %	3.79

VACUUM DISTILLATION DATA

Ovhd. Temp. Corr. to 760 mm Hg. Abs. Pressure, °F (°K)	Volume % Distilled
400 (477.6)	0.0
445 (502.6)	5.0
461 (511.5)	10.0
500 (533.2)	20.0
532 (551.0)	30.0
571 (572.6)	40.0
628 (504.3)	50.0
650 (616.5)	60.0
740 (666.5)	70.0
800 (699.9)	74.0
Distillation Residue (+800°F or +699.9°K)	32.42 Wt.%

The approximate conversion of maf coal to both a quinoline soluble form and a benzene soluble form achieved during feedstock preparation can be calculated from the inspection data compiled in Table 1 for the case of a feed coal containing about 10% ash. About 86 ± 1% and 52 ± 1% of the maf coal were converted to a quinoline soluble and benzene soluble form respectively. These conversions are significantly lower than those typically observed in liquefaction processes in which an Illinois No. 6 feed coal is employed. In general, at least 92% of maf Illinois No. 6 coal would be converted to a quinoline soluble form and over 75% of the maf Illinois No. 6 coal would be converted to a benzene soluble form. For example, a conversion of maf coal to a quinoline and benzene soluble form of about 95% and 85% respectively has been observed in the Lummus CFFC coal hydroliquefaction process for an Illinois No. 6 feed coal.

An inspection of the vacuum distillation data contained in Table 1 indicates that the initial boiling point of the paste solvent in the deashing feedstock sample received by Lummus is 400°F (477.6°K).

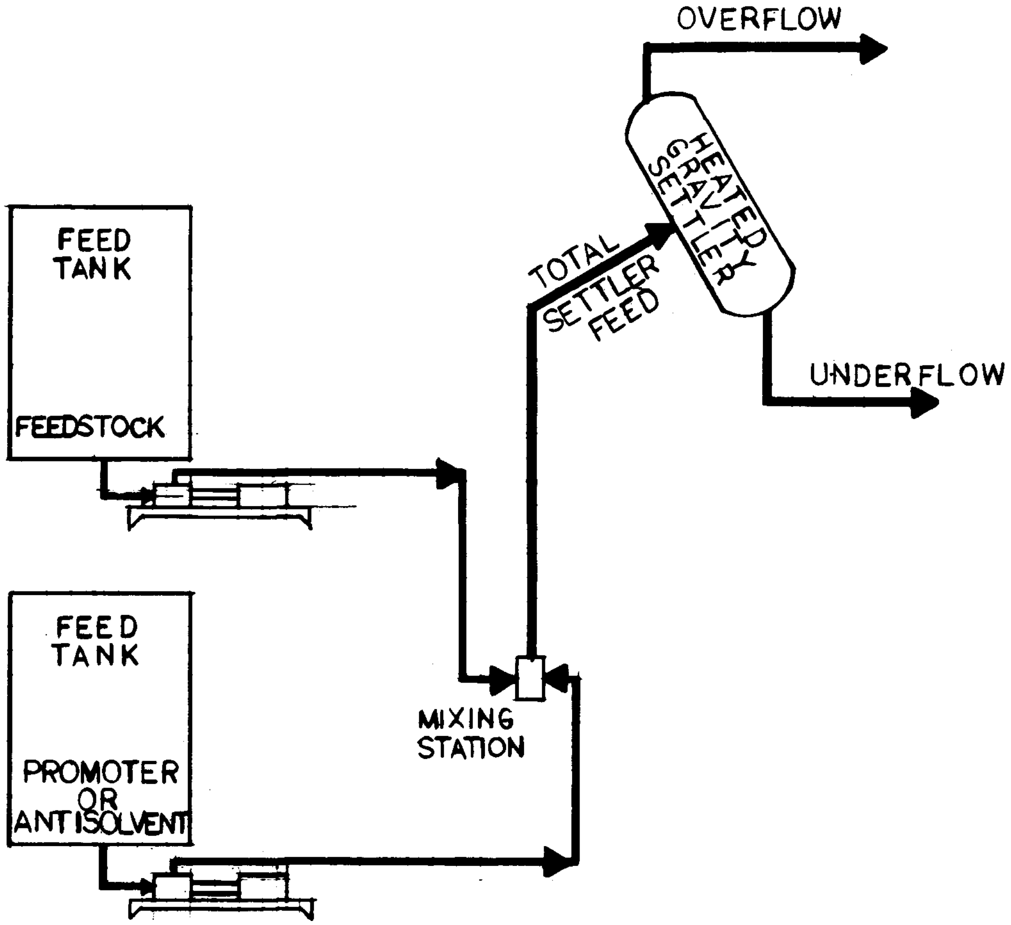
b. Pilot Plant Deashing Operations (Run DA-53-2)

1. Equipment

Figure 1 is a schematic flow diagram of the deashing pilot plant. The basic equipment consists of an electrically heated continuous mixing station and electrically heated continuous $6.0\text{x}10^{-2}$ meter^3 gravity settler. Heat losses from the mixing station and gravity settler are compensated for by the said electric heaters.

FIGURE 1

DEASHING PILOT PLANT SCHEMATIC FLOW DIAGRAM



Heated feedstock and promoter feed tanks, proportioning type feed pumps, various collection tanks, and ancillary piping plus necessary instrumentation are part of the equipment provided to sustain steady continuous pilot plant operation.

2. Procedure

After the unit was heated to the operating temperature, the test run was started. At about 12 hours of cumulative operating time both the feedstock and antisolvent rates were increased, to determine if the pilot plant deashing unit would handle more throughput without compromising overflow product quality. After 20 hours of cumulative operating time, the overflow and underflow product streams were simultaneously diverted to new and empty scale tanks. This operation marked the start of an arbitrarily defined on-stream period of about 40 hours duration. At the end of the on-stream period, the pilot plant deashing unit was voluntarily shut down.

The weights of overflow and underflow products collected in scale tanks were determined for the above on-stream period. After cooling to about $400^{\circ}K$, about 1.5×10^{-2} meter^3 of ash enriched underflow-like material was removed from the bottom of the settler and added to the underflow scale tank.

Overflow product collected during the on-stream deashing period was distilled to separate the antisolvent. Composite underflow product from the appropriate scale tank was vacuum topped to remove antisolvent in a single stage batch still. Bottoms product or a nominal antisolvent free underflow product thus produced, was sampled and packaged in an open top steel drum.

3. Data Aquisition and Test Results

An overall material balance for the on-stream deashing test period was developed. Table 2 is a compilation of the amount of feedstock fed and the quantity of antisolvent free overflow and underflow products obtained during the test period. Product quality data are also included for reference in Table 2.

During the test run small instantaneous samples of overflow and underflow product were withdrawn from the pilot plant equipment and analyzed for ash content. Table 3 is a compilation of the instantaneous ash contents so determined. These data provided an approximate process performance check of the pilot plant during the test run.

Table 4 and Table 5 are compilations of analytical data obtained for the antisolvent free overflow and antisolvent free underflow products. The antisolvent free overflow product was found to contain 0.02 wt.% residual ash (Table 4). An ash removal efficiency or degree of ash removal corresponding to this residual ash content was calculated from the material balance data presented in Table 2 and the feedstock ash content cited in Table 1.

TABLE 2

ANTISOLVENT FREE OVERALL MATERIAL BALANCE AND PRODUCT QUALITY DATA FOR DEASHING RUN DA-53-2

	Amount, kg	Ash Content, Wt. %	Quinoline Insolubles, Wt. %
Feedstock	517.2	3.55	7.87
Clean Product or Overflow	348.8	0.02	< 0.05
Ash-Enriched Bottoms or Underflow	168.4	10.62	-

TABLE 3

INSTANTANEOUS ASH CONTENTS OBSERVED DURING DEASHING RUN DA-53-2

Cumulative Time Hrs.	Overflow (1) Product Ash (2) Content, Wt. %	Underflow(1) Ash(2) Content, Wt. %	Remarks
0	-	-	Started pumping feedstock
1	-	-	Started pumping antisolvent
4	-	-	Overflow first observed. Started underflow withdrawal at about 4.25 hours.
6	0.035	-	
10	0.010	5.96	Feed rates increased at about 12 hours cumulative time
14	<0.01	-	
18	<0.01	-	On stream collection period began @ 20 hours of cumulative time.
22	<0.01	-	
26	0.01	-	
30	0.046	7.92	
34	-	8.37	
38	<0.01	7.59	
42	0.018	-	
46	0.018	10.36	
50	0.046	-	
52	-	12.19	
56	0.040	8.77	
60	0.010	-	Began shutdown procedure

(1) 10-15 gms samples were collected in a porcelain evaporating dish, which was used in the subsequent ash content analysis.

(2) Method cited in Table A located in Appendix A.

TABLE 4

ANTISOLVENT FREE OVERFLOW PRODUCT INSPECTION DATA

(DEASHING RUN DA-53 -2)

Quinoline Insolubles, Wt. %	<0.05 (1)
Ash Content, Wt. %	0.02
Benzene Insolubles, Wt. %	11.79
Carbon Content, Wt. %	88.60
Hydrogen Content, Wt. %	6.72
Nitrogen Content, Wt. %	0.93
Sulfur Content, Wt. %	0.47
Oxygen Content, Wt. % (via difference)	3.26

VACUUM DISTILLATION DATA

Overhead Temp. Corrected to 760 mm Hg. Abs. Pressure, °F (°K)	Volume % Distilled
405 (480.4)	0.0
447 (503.8)	5.0
465 (513.8)	10.0
497 (531.5)	20.0
529 (549.3)	30.0
571 (572.6)	40.0
612 (595.4)	50.0
647 (614.9)	60.0
703 (646.0)	70.0
800 (699.9)	79.0

Distillation Residue (+800°F or +699.9°K) 28.13 WT. %

ELEMENTAL ANALYSIS OF DISTILLATION RESIDUE

Ash Content, WT. %	0.06
Carbon Content, Wt. %	87.31
Hydrogen Content, Wt. %	5.73
Nitrogen Content, Wt. %	1.71
Sulfur Content, Wt. %	0.70
Oxygen Content, Wt. % (via Difference)	4.49

(1) Limits of quinoline insolubles determination

TABLE 5

ANTISOLVENT FREE UNDERFLOW PRODUCT INSPECTION DATA (DEASHING RUN DA-53-2)

Ash Content, Wt. %	10.62
Carbon Content, Wt. %	76.64
Hydrogen Content, Wt. %	5.81
Nitrogen Content, Wt. %	0.87
Sulfur Content, Wt. %	0.94
Oxygen Content, Wt. % (via Difference)	5.12

The result of this calculation is an ash removal efficiency of 99.6% based on feed. Vacuum distillation data presented in Table 1 and 4, and the Table 2 overall material balance data were used in conjunction with the feedstock ash and quinoline insoluble contents (Table 1) to calculate the amount of ash free +699.9°K (+800°F) feedstock derived residual product recovered with the overflow product stream. About 65.7 wt.% of the ash free +699.9°K residual product and about 77.2 wt.% of the quinoline soluble +699.9°K residual product contained in the feedstock were recovered with the overflow product stream. Similarly, the Table 1, 2 and 4 data were used to determine the distribution of the benzene insoluble/quinoline soluble heavy product constituent between the overflow and underflow product streams. About 74.8 wt.% of benzene insoluble/quinoline soluble heavy product present in the feedstock was found in the overflow product.

Carbon, hydrogen, nitrogen, sulfur, organic oxygen, and ash balance closures were checked using the Table 2 antisolvent free overall material balance data reduced to an hourly (3600 secs.) basis, along with the elemental analysis contained in Table 1, Table 4, and Table 5. Table 6-A lists the elemental balance closures thus obtained. Table 6-B contains further detail on the supporting calculations. Specifically, elemental balance closures for carbon, hydrogen, nitrogen, and sulfur were 98.9% or better. An elemental

TABLE 6-A

ELEMENTAL BALANCE CLOSURES (1)

Quantity	Elemental Uncertainty (U_j) kg/hr	Elemental Input (I_j) Kg/Hr	Elemental Balance Closure (C_j)= 100 (1.00 - $\left\|\frac{U_j}{I_j}\right\|$)
Carbon	- 0.0020	10.9504	99.98
Hydrogen	+ 0.0008	0.8314	99.90
Nitrogen	+ 0.0013	0.1190	98.91
Sulfur	- 0.0004	0.0802	99.50
Ash	+ 0.0112	0.4590	97.56
Organic Oxygen	- 0.0099	0.4900	97.92

(1) See Table 6-B for further detail.

TABLE 6-B

ELEMENTAL BALANCE SUMMARY FOR DEASHING RUN DA-53-2

STREAM	FEEDSTOCK	INPUT	ANTISOLVENT FREE OVERFLOW	ANTISOLVENT FREE UNDERFLOW	OUTPUT
Composition					
Carbon Content, Wt. %	84.69		88.60	76.64	
Hydrogen Content, Wt. %	6.43		6.72	5.81	
Nitrogen Content, Wt. %	0.92		0.93	0.87	
Sulfur Content, Wt. %	0.62		0.47	0.94	
Ash Content, Wt. %	3.55		0.02	10.62	
Organic Oxygen, Wt. % (Via Difference)	3.79		3.26	5.12	
Amount					
Carbon, kg/hr	10.9504	10.9504	7.7259	3.2265	10.9524
Hydrogen, kg/hr	0.8314	0.8314	0.5860	0.2446	0.8306
Nitrogen, kg/hr	0.1190	0.1190	0.0811	0.0366	0.1177
Sulfur, kg/hr	0.0802	0.0802	0.0410	0.0396	0.0806
Ash, kg/hr	0.4590	0.4590	0.0017	0.4471	0.4488
Organic Oxygen, kg/hr (via Difference)	0.4900	0.4900	0.2843	0.2156	0.4999
TOTALS, kg/hr	12.93	12.93	8.72	4.21	12.93

Elemental Uncertainty (U_j) $\equiv$ Elemental Input (I_j) - Elemental Output (O_j) --- (1)

Percent Elemental Uncertainity (u_j) $\equiv \frac{U_j}{I_j} \times 100$ --- (2)

Elemental Balance Closure (C_j) $\equiv 100 - |u_j|$ ---(3)

balance closure of 97.5% was obtained for ash and organic oxygen, which was determined by difference (0 = 100 - (C + H + S + N +A)).

C. Yield Structure for the Deashing Operation

An antisolvent free yield structure for deashing run DA-53-2 was developed from experimental data. Table 7 is a compilation of the said yield structure. The constituents of the Table 7 yield structure are a nominal 477.6/699.9°K (400/800°F) boiling range paste oil, a heavy +699.9°K (+800°F) ash free, quinoline soluble SRC product, an ash free quinoline insoluble residue, and ash.

D. Miscellaneous Laboratory Work

A small sample of composite underflow from deashing run DA-53-2 was vacuum stripped at about 6.6 kpascals absolute pressure in a laboratory scale batch still. The purpose of this work was to scope in a preliminary manner a tentative cut point that might be used in the preparation of a coker feedstock for the subsequently reported bench coking work. The contents of the flask began to crack at an overhead cut point of about 650°F or 616.5°K corrected to 101.1 kpascals absolute pressure, which corresponded to an observed final still pot temperature of about 500°F or 533.2°K. A 4-hour time period was used to reach the above cut point. Based on the above observation, an overhead cut point of about 550°F or 561 °K corrected to 101.0 kpascals absolute pressure and operating pressure of 6.6 kpascals absolute

TABLE 7

ANTISOLVENT FREE YIELD STRUCTURE FOR DEASHING RUN DA-53-2

STREAM	FEEDSTOCK		ANTISOLVENT FREE OVERFLOW		ANTISOLVENT FREE UNDERFLOW	
	kg/hr	Wt.%	kg/hr	Wt.%	kg/hr	Wt.%
Components						
Nominal 477.6/+699.9°K (400/800°F) Paste Oil	8.738	67.58	6.267	71.87	2.471	58.69
Heavy (+699.9°K or +800°F) SRC Product (Ash Free, Quin.Sol)	3.174	24.55	2.451	28.11	0.723	17.17
Ash Free Quinoline Ins. Residue	0.559	4.32	Nil	Nil	0.559	13.28
Ash	0.459	3.55	0.002	0.02	0.457	10.86
TOTALS	12.93	100.00	8.72	100.0	4.21	100.00

were chosen as operating conditions for a laboratory scale coker feedstock preparative distillation.

E. Bench-Scale Coking Work

1. SRC Dry Mineral Residue Feedstock

A sample of dry mineral residue supplied by the Wilsonville, Alabama SRC pilot plant was bench coked at the request of EPRI. EPRI in a verbal communique stated that the non-combustible portion of this particular mineral residue sample contained from about 28 to 39 wt.% filter aid on a combustible matter free basis.

Inspection data obtained on the above dry mineral residue are compiled in Table 8. Coker overall yield data and green coke inspection data are presented in Table 9.

2. Underflow Derived Feedstock

A sample of antisolvent free underflow was vacuum distilled in the laboratory to prepare a bench coking feedstock. A nominal cut point of 561.0°K or 550°F corrected to 101.0 kpascals absolute pressure was employed in this preparative distillation, which was run at about 6.6 kpascals absolute pressure. A vacuum bottoms or coker feedstock yield of 69.8 wt.% based on feed (antisolvent free underflow) was obtained in the above laboratory distillation. A coking run was then made in the Lummus bench-scale delayed coking unit.

TABLE 8

WILSONVILLE DRY MINERAL RESIDUE INSPECTION DATA SUMMARY

Ash Content, Wt.%	66.16
Carbon Content, Wt.%	27.73
Hydrogen Content, Wt.%	1.40
Nitrogen Content, Wt.%	0.46
Sulfur Content, Wt.%	3.00

TABLE 9

LABORATORY BENCH COKING OF WILSONVILLE DRY MINERAL RESIDUE SAMPLE

Run No.		
Product Yield		
Gas Yield via Difference,	Wt. %	2.0
Total Liquid Product,	Wt. %	4.0
Green Coke,	Wt. %	93.9
		100.0
Coke Inspection Data		
Sulfur Content,	Wt. %	3.32
Moisture Content,	Wt. %	0.57
Volatile Matter (DB),	Wt. %	7.53
Ash Content (DB,	Wt. %	69.81
Fixed Carbon (DB),	Wt. %	22.66

Table 10 is a compilation of inspection data obtained on the underflow derived coking feedstock. Coker overall yield data and green coke inspection data are presented in Table 11. Inspection data on the total liquid product generated in the bench-coking run are summarized in Table 12.

The experimentally observed coke yield (Table 11) was used in conjunction with the coker feedstock ash content listed in Table 10 to calculate a green coke expected ash content, which was found to be 28.7 wt.%. This calculated ash content was found to be in general agreement with the experimental green coke ash content of 29.8 wt.% (as is basis) compiled in Table 11.

The vacuum distillation data obtained on the total coker liquid product generated during bench coking run LC-133-A, which is presented in Table 12, is of practical interest. Only 3.6 vol.% of the total coker liquid product is lighter than about 360°F or 455.4°K. Thus from a distillation viewpoint as much as 96 vol % of the total coker liquid product could be recycled as paste oil to the dissolver section of an SRC type pilot plant.

F. Additional Analytical Data

Tables B-1 through B-5 in Appendix B contain additional analytical data acquired during the experimental program. Tables B-1 and B-2 contain the elemental analysis of antisolvent free vacuum distillate (800°F- or 699.9°K-) derived from deasher

TABLE 10

UNDERFLOW (DA-53-2) DERIVED COKER FEEDSTOCK INSPECTION DATA SUMMARY

Softening Point (R&B) °K	408
Ash Content, Wt. %	15.43
Carbon Content, Wt. %	71.51
Hydrogen Content, Wt. %	4.71
Nitrogen Content, Wt. %	1.03
Sulfur Content, Wt. %	1.34

TABLE 11

LABORATORY BENCH COKING OF UNDERFLOW (DA-53-2)

Run No.	LC-133-A
Product Yields	
Gas via Difference, Wt. %	5.2
Total Liquid Product, Wt. %	41.1
Green Coke, Wt. %	53.7
Coke Inspection Data	
Sulfur Content, Wt. %	2.06
Moisture Content, Wt. %	0.65
Volatile Matter (DB), Wt. %	9.05
Ash (DB), Wt. %	30.00
Fixed Carbon, Wt. %	60.95

TABLE 12

INSPECTION DATA SUMMARY FOR THE TOTAL COKER LIQUID PRODUCT PRODUCED IN BENCH COKING RUN LC-133-A

Carbon Content, Wt.%	88.36
Hydrogen Content, Wt.%	6.85
Nitrogen Content, Wt.%	0.81
Sulfur Content, Wt.%	0.43

VACUUM DISTILLATION DATA

Overhead Temp. Corrected to 760 mm Hg. Lbs. Pressure, °F (°K)	Volume % Distilled
202 (367.6)	0.0
360 (455.4)	3.6
400 (477.6)	5.0
525 (547.1)	8.0
537 (553.8)	10.0
571 (572.6)	20.0
593 (584.9)	30.0
605 (591.5)	40.0
628 (604.3)	50.0
640 (611.0)	60.0
657 (620.4)	70.0
689 (638.2)	80.0
776 (686.5)	90.0
800 (699.9)	92.2
Distillation Residue (+800°F or +699.9°K)	8.12 Wt.%

overflow and the elemental analysis of antisolvent free vacuum distillate (550^{O}F- or 560^{O}K-) derived from deasher underflow respectively for run DA-53-2. Green coke approximate elemental analysis are compiled in Table B-3. Underflow derived ash compositional analysis are tabulated in Table B-4, and moisture free gas chromatography data obtained on a coker gas sample (withdrawn from a wet gas holder) are compiled in Table B-5.

VI. Economic Evaluation

A. Study Basis

1. Product Slate/Capacity/Yields

The information presented is based on deashing facilities that would be included in a plant capable of processing 9070 MT/D (10,000 ST/D) moisture free Illinois No. 6 Coal via the Solvent Refined Coal liquefaction process.

The scope of this study is limited only to those areas directly or indirectly related to the deashing step and does not include detailed evaluation of other areas including front end processing and liquefaction technology which would, of course, be included in a grassroots coal liquefaction complex.

Each deashing scheme is based on separating the feedstock into the following primary products:

- an ash free SRC product (427°C+ boiling range),
- pasting solvent (204 - 427°C) for recycle to front-end slurrying and pasting facilities,
- an ash rich residue which is subsequently gasified to provide plant hydrogen and energy requirements.

In addition to the above, byproduct gases and liquid distillate are produced in a delayed coking step included in the antisolvent deashing scheme to process the ash concentrated stream.

Properties and ultimate analysis of the deashing feedstock for each process are based on laboratory analyses given in Section V. of a sample provided to Lummus.

In comparing overall manufacturing costs for each process we have considered two cases in which the overall SRC product yield achieved is varied. The basis for each case is as follows:

Case A (Base Case) - The antisolvent deashing yield of SRC is based on Lummus pilot plant tests performed on a low conversion* feedstock sample received from Wilsonville. This is compared to a filtration SRC yield based on cake liquid composition data provided by EPRI for a more highly converted ‡ feedstock. This case forms the basis for determining the overall investments for each process.

Case B (Alternate Base Case) - Antisolvent deashing yields for this case are equivalent to Case A, however, a somewhat lower filtration yield relative to Case A was specified. It is believed that lower coal conversions as shown in the sample provided will result in lower SRC yields for the filtration case.

* Approximately 84% of maf coal fed to the SRC dissolver is liquefied.

‡ Approximately 93% of maf coal fed to the SRC dissolver is liquefied.

The filtration processing scheme selected is similar to that in operation at the SRC demonstration plant (ERDA) located in Tacoma, Washington and operated by the Pittsburgh and Midway Coal Mining Company.

2. Investment/Operating Requirements

Starting with the estimated yield structure for each deashing process, approximate mass and energy balances were made which allowed the estimation of investment and operating requirements reported in sub-section D, Tables 23 and 24.

All hydrogen and energy requirements, for the entire complex, with the exception of purchased electricity, have been assumed to be supplied directly or indirectly from the ashy residue and supplemental coal (where necessary). Make-up water is assumed available as required.

For the most part, the estimate of Total Installed Cost (TIC) for each process given in sub-section D, was developed from purchased equipment costs quoted by U.S. vendors, or from Lummus in-house cost data. As a result of numerous preliminary engineering studies involving coal liquefaction technology, Lummus has developed techniques to translate purchased equipment costs to a Total Installed Cost basis. These techniques are supplemented in part by information provided by EPRI for the filtration process.

3. Economic Evaluation

The economics of each process are evaluated using the Discounted Cash Flow (DCF) technique to calculate manufacturing costs associated with the deashing operation.

For the filtration process, the economics are evaluated for a wide range of potential operating performances and design criteria. This is accomplished by selecting a base set of conditions and then treating variations to the base case in the form of a sensitivity analysis.

The bases, assumptions and results of the economic evaluation are reported and discussed in greater detail in sub-section E of this report.

B. Description of Processing Facilities

1. Overall Process Description

Schematic block flow diagrams illustrating the C-E Lummus Antisolvent Deashing Process and the filtration deashing process are shown in Figure 2.

Note that in addition to including the primary processing areas associated with each deashing method, viz., Antisolvent Deashing, and Underflow Delayed Coking, Filtration/ Fractionation, and Filter Cake Drying and Handling, we have also shown complementary facilities for the production of hydrogen gas for the hydroliquefaction step and fuel gas for in-plant energy requirements that would be included in a commercial-size coal liquefaction complex.

These areas depend on product yields, compositions of ashy residues and energy requirements associated with the two deashing processes, and therefore must be considered in order to present a meaningful comparison.

Primary "front-end" processing areas including the coal pasting and hydroliquefaction sections are identical for either case since they are not affected by differences in the deashing process, and thus are not shown on the flowsheets.

As a basis for comparing overall plant operating requirements, we have assumed the chemical hydrogen requirement in the hydroliquefaction area to be 1.5 wt.% on MAF coal to liquefaction and added 15% as excess to account for solution and purge losses in the liquefaction section.

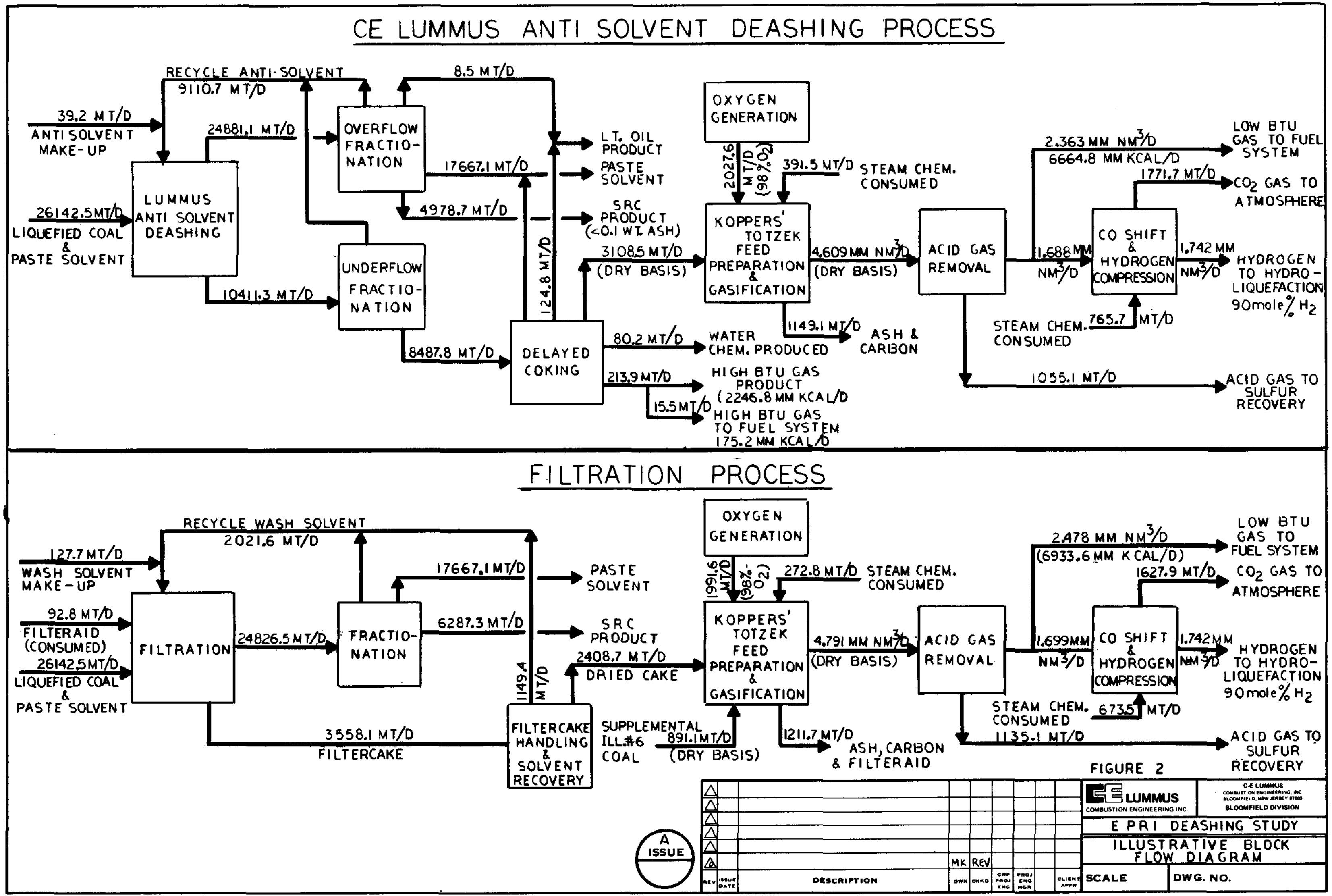


FIGURE 2

106 0575-E3.1-2

We believe the figure for H_2 consumption is representative of the low coal conversion, which characterizes the Wilsonville deashing feedstock actually processed by Lummus in our pilot plant.

An analysis of slurry feed to the deashing areas is given in Table 14. From the data provided, it is apparent that net yields of gases and light distillate oils formed in the hydroliquefaction step have been removed in upstream process area. Thus the feedstock to the deashing areas consists of 204°C-427°C boiling range pasting oil (to be recovered and recycled to the coal pasting area), 427°C+ Solvent Refined Coal product plus ash and unconverted coal in the form of quinoline insolubles.

a. Antisolvent Deashing Process

Feedstock from the hydroliquefaction area at 316°C is continuously fed to the deashing section where it is intimately contacted with an antisolvent stream, which promotes gravity settling. An essentially ash or solids free overflow stream and ash or solids enriched underflow stream is continuously generated in the deashing area. The extent of deashing is such that the final Solvent Refined Coal product will have an ash content of under 0.1 wt.%.

Both the overflow and underflow from deashing are fractionated separately to recover the antisolvent overhead.

TABLE 14

ANALYSIS OF FEEDSTOCK USED IN EPRI COAL LIQUEFACTION DEASHING STUDY

Properties	
Specific Gravity 62.8/15.5°C	1.103
Quinoline Insolubles, wt. % (including ash)	7.87
Benzene Insolubles, wt. %	18.47
Ultimate Analysis	**WT. %**
Carbon	84.25
Hydrogen	6.34
Nitrogen	0.89
Sulfur	0.62
Ash	3.55
Organic Oxygen (by difference)	4.35
	100.00
Component Breakdown	**WT. %**
Nominal 204.4/426.7°C (400-800°F) Paste Oil	67.58
Heavy (426.7°C or +800°F) SRC Product, (Ash Free, Quinoline Soluble)	24.55
Ash Free Quinoline Insoluble Residue	4.32 (1)
Ash	3.55
	100.00

Note: (1) This corresponds to 84% conversion of MAF coal to quinoline soluble material and is inconsistent with the SRC yield assumed for the filtration base case.

The boiling range of the antisolvent employed is sufficiently below the boiling range of the pasting oil, which enables nearly complete antisolvent recovery by conventional fractionation. The essentially antisolvent free overflow stream is then further fractionated via vacuum distillation for recovery of the 204-427°C (400-800°F) boiling range pasting solvent which is recycled to the coal pasting area of the complex. Solvent Refined Coal is withdrawn as the bottoms from the overflow vacuum fractionator, then quenched and solidified to its final product form.

Fresh antisolvent is added to the system to make up for the small quantity that is unrecovered in the fractionation of the overflow and underflow streams.

The stripped underflow stream containing essentially all of the ash and unconverted carbon plus a portion of the pasting oil and SRC is routed to the delayed coking section of the plant to produce an ash-rich coke used as feedstock to the Koppers-Totzek gasification section. Fractionation in the delayed coking plant provides for complete recovery of the recycle paste solvent. In addition, a wet gas stream (comprised of C_1-C_4 gases)and a naphtha boiling range liquid are recovered as byproducts of the coking reaction.

An underflow coking step is included to ensure complete recovery of paste oil boiling range (204-427°C) material.

The primary reason for its inclusion in this study is the nature of the low severity SRC dissolver effluents or its low degree of conversion. The deashing feedstock from low severity SRC processing appears to be characterized by a high percentage of ash free benzene insoluble material. Based on laboratory and pilot plant studies, we have observed that high concentrations of benzene insolubles in the feedstock to the deashing section tend to decrease underflow fluidity. Hence the amount of paste oil and/or SRC product that must remain in stripped underflow to maintain fluidity, would be somewhat higher for the case of a deashing feedstock with a high benzene insolubles content and in which a coking step was not included in the underflow processing sequence.

By contrast, we have found that the deashing feedstock derived from the catalytic hydroliquefaction step of the Lummus CFFC process contains significantly less benzene insolubles when operating on the same type of coal feed. With Illinois No. 6 coal feedstock an underflow coking unit would not be required when producing a boiler fuel product via the CFFC process.

Delayed coking is a common refinery practice to convert heavy liquid petroleum residues. It has been successfully extended to include non-conventional feedstocks such as coal tar pitches and even gilsonites and tar sand bitumens containing mineral matter. In this application sufficient pasting solvent is retained with the underflow to maintain fluidity and reasonably low solids contents in the coking heater. Paste solvent is completely recovered for recycle to the pasting area by conventional fractionation in the coking unit.

Ash-rich coke, after pulverizing and drying is routed to the Koppers-Totzek gasification step for conversion into a H_2/CO rich gas mixture suitable for hydrogen production and fuel gas requirements. The gasifiers operate on the partial oxidation principle whereby the ash-rich coke is combusted with controlled amounts of high purity oxygen in the presence of steam.

The sulfur present in the gasifier feed, (both organic and pyritic), is converted to H_2S in the reducing atmosphere of the gasifiers. The gaseous mixture is cooled in waste heat boilers generating high pressure steam, water scrubbed for particulate removal, compressed and treated for removal of H_2S and CO_2. A portion of the gas is then routed to the hydrogen manufacturing area and the balance is utilized as in-plant fuel gas.

Hydrogen manufacture in the process scheme is based on shifting of the gas followed by purification and compression of the resultant H_2 rich gas to the pressure required for the hydroliquefaction step. The product gas stream on a dry basis contains approximately 90 mole % hydrogen with the balance being carbon oxides and nitrogen.

b. Filtration Process

In the filtration deashing process, feedstock at about 316°C from the hydroliquefaction area is pressured directly to rotary precoat filters.

Rotary precoat filtration is a semi-continuous cyclic operation in which ash and unconverted coal are retained on semi-submerged precoated rotating cylindrical filter drum. The use of precoat or filteraid (usually a suitable grade of diatomaceous earth), in coal liquefaction filtration applications has been found to enhance greatly filter permeability and increase on-stream factors by preventing excessive clogging of the filter screen.

As the precoated rotating drum with retained solids on the outer surface exits the liquid pool in the filter vat, the drum is sprayed with a wash solvent liquid to enhance recovery of SRC product that would otherwise be lost with the filter cake.

Following the washing step, the retained solids and a thin layer of precoat are shaved from the drum by an advancing doctor blade, thereby continuously exposing a fresh, permeable filteraid surface to the slurry being filtered. This continuous removal of the filter cake overcomes the reduction filtration rate that would otherwise accompany an increase in cake thickness.

After the layer of filteraid has been nearly completely removed, the filter is taken off stream and a fresh coat of filteraid (approx. 0.1 meters) is applied in a separate operation.

The driving force for the filtration step is provided by an essentially inert recycle gas stream which maintains a pressure of about 1305 k Pa abs. (175 psig) on the outside of the drum.

Solids-free filtrate is withdrawn from the filters at about 177°C (350°F) and collected in a receiver vessel. The filter recycle gas and a portion of the wash solvent are flashed from the receiver as vapors. The vapors are cooled thereby recovering the wash solvent as condensed liquid and the remaining non-condensible gases are compressed, reheated to 316°C (600°F) and returned to the filters. Liquids from the filtrate receiver consisting of wash solvent, pasting solvent and SRC product are pressured to the fractionation area.

In the fractionation area, the filtrate is first pre-flashed slightly above atmospheric pressure. Vapors from the pre-flash drum are condensed and fractionated into 1) a wash solvent overhead stream which is returned to the filtration area, and 2) a bottoms stream consisting of the light end portion of the recycle pasting solvent.

Liquids from the pre-flash drum are pumped through a fired heater to vaporize and separate the remaining paste solvent in a steam stripped vacuum fractionator. Flashed pasting solvent is condensed and withdrawn as a sidestream from the vacuum fractionator and returned to the coal pasting area. The Solvent Refined Coal product is rejected from the bottom of the tower, then cooled and solidified into its final form.

Wet cake shaved from the filters is diluted with wash solvent to about 40% solids by weight and depressured to a surge drum. The diluted cake slurry is then pumped to indirect rotary gas fired dryers. In the kiln-type dryers, about 90% of the wash solvent present in the washed filter cake plus the additional wash solvent used to slurry the filter cake is vaporized. The vapors are condensed and returned to the filter wash and cake dilution steps previously described. A quantity of wash solvent (10% of wash solvent in washed filter cake) is added as make up to the recycle wash solvent stream to account for losses occuring in the cake drying step.

Cake from the dryers containing unrecovered wash solvent, retained SRC, filteraid, ash and unconverted coal is water cooled in indirect rotary coolers, then pulverized and fed to the Koppers-Totzek gasifiers. Carbonaceous material in the filter cake is converted to a low calorific value gas which after the acid gas removal step is suitable for inplant fuel gas and as a synthesis gas feed for hydrogen production.

Supplemental raw coal, after crushing and drying is also fed to the gasification section. This coal also produces a H_2/CO rich gas primarily to satisfy plant fuel requirements which are not entirely met by the gasification of filter cake alone. The gasification, acid gas removal and hydrogen generation steps operate in the same manner as previously described for the antisolvent deashing process.

2. Overall Material Balance

Overall material balances for the two deashing processes are shown in Tables 15 and 16 for the low coal conversion case.

The quantity of deashing feedstock used as the starting point for material and heat balance calculations is based on a raw coal feed rate to liquefaction of 9,070 MT/D (10,000 ST/D) Illinois No. 6 Coal containing 10.23 wt. % ash on a dry basis. From the feedstock analysis given (see Table 14)and assuming all of the ash appears in the deashing feedstock, the feed rate was readily established.

An inspection of Tables 15 and 16 indicates differences in the yields and product slates of the respective deashing processes. Due to the relatively low coal conversion associated with the deashing feedstock used for this study, the loss of SRC product to the deasher underflow for the antisolvent scheme is larger than it would be for normal SRC conversions in the over 92% maf range. While some of the SRC in the underflow is converted to light naphtha and fuel gas byproducts in the delayed coking step, the major portion appears as coke which is subsequently gasified for plant energy and hydrogen requirements.

Because of the higher SRC yield assumed for the filtration, the quantity of carbonaceous material in the filter cake to gasification is not sufficient to provide all the plant energy and hydrogen requirements and therefore supplemental raw coal must

TABLE 15

OVERALL MATERIAL BALANCE SUMMARY

THE C-E LUMMUS ANTISOLVENT DEASHING PROCESS

BASIS: 9,070 MT/D Illinois No. 6 Coal Feed to Liquefaction (dry basis).

INPUTS:		METRIC TONS/SD
a.	Slurry Feed to Deashing Section	26,142.5
b.	Antisolvent Make-Up	39.2
c.	Oxygen (as 98%)	2,027.6
d.	Steam Chemically Consumed	
	- Koppers-Totzek Gasification	391.5
	- CO Shifting for H_2 Generation	765.7
	TOTAL INPUTS	29,366.5

OUTPUTS:		
a.	Net SRC Product	4,978.7
b.	Coker Naphtha Product (C_5-204°C)	133.3
c.	Paste Oil Recycle (204°C-427°C)	17,667.1
d.	High Calorific Value Fuel Gas (dry basis)	
	- By product to Sale	198.4 (392 MM kJ/Hr LHV)
	- Consumed in Process Heaters	15.5 (30.6 MM kJ/Hr LHV)
e.	Net Water of Reaction Produced (from Underflow Delayed Coking)	80.2
f.	Koppers-Totzek Ash (dry basis)	
	as: Ash	928.0
	as: Unburnt Carbon	221.1
g.	Acid Gases to Sulfur Recovery (dry basis)	1,055.1
h.	Low Calorific Value Fuel Gas (dry basis, consumed in process heaters and boilers)	1,938.7 (1162 MM kJ/Hr LHV)
i.	H_2 Rich Gas to Hydroliquefaction (dry basis)	378.7
j.	CO_2 to Atmosphere (dry basis, from hydrogen generation)	1,771.7
	TOTAL OUTPUTS	29,366.5

TABLE 16

OVERALL MATERIAL BALANCE SUMMARY

FILTRATION DEASHING PROCESS

BASIS: 9,961 MT/SD Illinois No. 6 Coal Feed - Dry Basis (9,070 MT/D to Liquefaction)

INPUTS:		METRIC TONS/SD
a.	Slurry Feed to Filters	26,142.5
b.	Wash Solvent Make-Up (1)	127.7
c.	Filteraid (2)	92.8
d.	Raw Coal to Koppers-Totzek Gasification (dry basis)	891.1
e.	Oxygen (as 98%)	1,991.6
f.	Steam Chemically Consumed	
	- Koppers-Totzek Gasification	272.8
	- CO Shifting for H_2 Generation	673.5
	TOTAL INPUTS	30,192.0

OUTPUTS:		
a.	Net SRC Product	6,287.3
b.	Paste Oil Recycle (204-427°C)	17,667.1
c.	Koppers-Totzek Ash (dry basis)	
	- as: Ash	1,017.2
	- as: Unburnt Carbon	101.7
	- as: Filter Aid	92.8
d.	Acid Gases to Sulfur Recovery (dry basis)	1,135.1
e.	Low Calorific Value Fuel Gas (dry basis consumed in process heaters and boiler)	1,877.9 (1209 MM kJ/hr LHV)
f.	H_2 Rich Gas to Hydroliquefaction (dry basis)	385.0
g.	CO_2 to Atmosphere (dry basis, from hydrogen generation)	1,627.9
	TOTAL OUTPUTS	30,192.0

NOTES:

(1) Based on a 90% recovery of wash solvent contained in the washed filter cake.

(2) Based on a precoat consumption of 0.10 kg precoat per kg of ash removed in the filtration step.

be gasified where none is required for the Antisolvent Deashing Process(1).

Thus, for the poorly converted feed the combined yield of ash free liquid SRC and naphtha products for the antisolvent deashing process is about 6 percentage points less than the SRC yield obtained via the filtration process(2). As stated earlier, the filtration yield in this case is based on filter cake data provided by EPRI for a deashing feed reflecting a higher coal conversion than the feed actually tested in the Lummus pilot plant.

If a more conservative approach were adopted as regards the product yield obtained via filtration, i.e., one in which the amount of SRC liquids retained on the washed filter cake was revised to reflect the higher viscosity and benzene insolubles content of the poorly converted feedstock, the difference in yields between the two processes would be reduced to less than 4 percentage points(3).

(1) The estimated energy requirements for the Antisolvent Deashing process are not entirely met by gasifying the ashy coke. A very small amount of the coker gas must be burned as in-plant fuel (approx. 30.6 MM kJ/Hr) to close the balance.

(2) The overall net liquid product yields (ex. solvent make-up) for each process as a percentage conversion of total coal fed to the plant are 61.8 wt. % for filtration and 56 wt. % for antisolvent deashing.

(3) This assumes that benzene insolubles retained on the filter cake cannot be recovered by washing.

As indicated, oxygen requirements, chemical steam consumptions and overall plant fuel gas requirements are essentially the same for each process. Included in the overall fuel gas requirements shown are rough estimates for process heating requirements in the coal pasting/hydroliquefaction section (the same for each case) and for a plant boiler duty based on production of drive steam for air separation plant and hydrogen plant compression requirements.

C. Design Considerations for Key Process Areas

1. Antisolvent Deashing

a. Antisolvent Deashing/Overflow and Underflow Fractionation (with Process Flow Diagram)

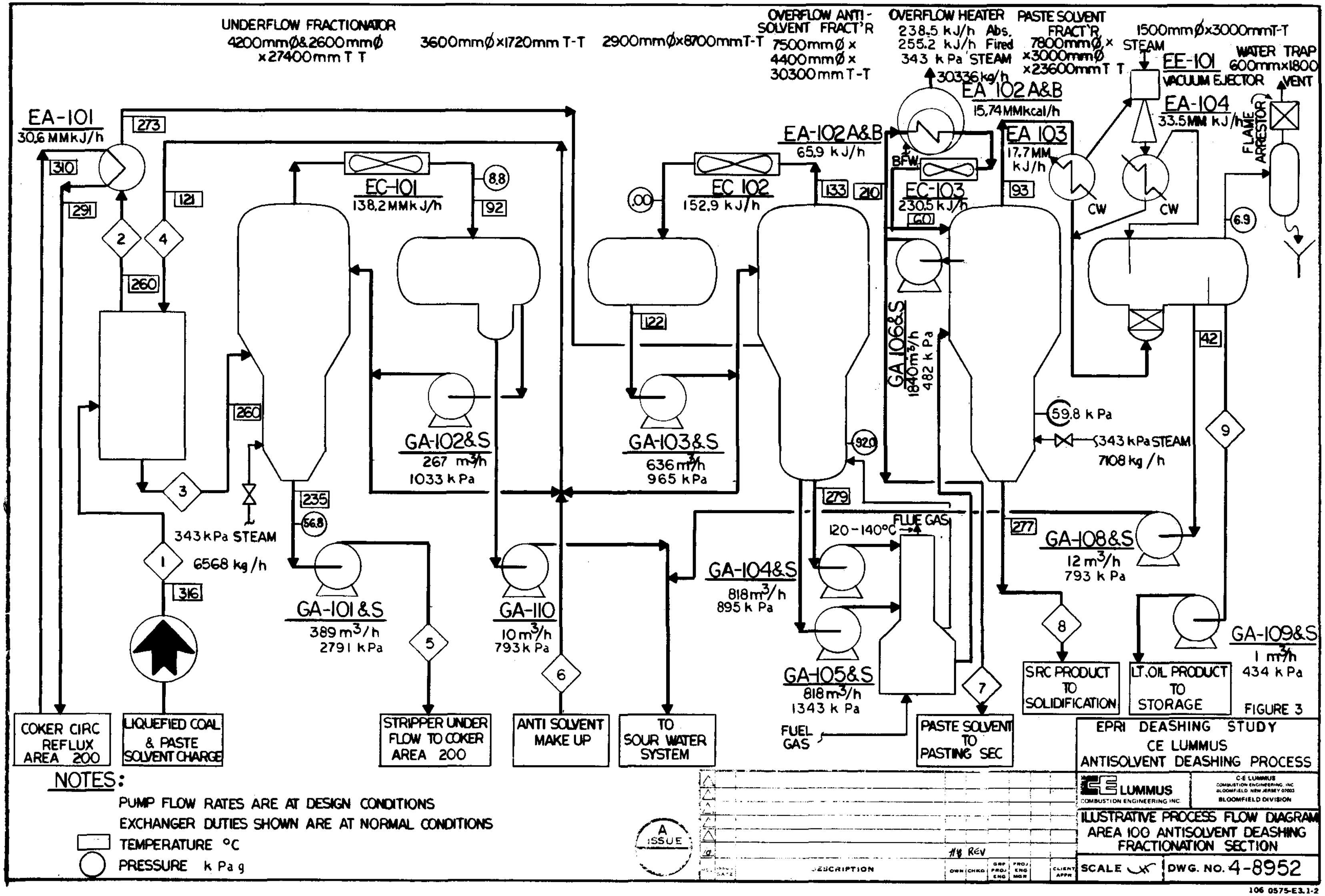
The primary purpose of the Lummus deashing system is to (1) produce an overflow product sufficiently low in ash so that the heavy SRC product (nominal 427°C+) will have an ash content under 0.1 wt.% and (2) recover antisolvent for recycle and reuse. An illustrative process flow diagram is shown in Figure 3. An overall material balance is given in Table 17.

Flash drum liquids from the dissolver section at an assumed 316°C are contacted via an in-line mixing device with the recycled anti-solvent stream. The resulting mixture is introduced into a gravity settler where the antisolvent causes the ash and other quinoline insoluble material (unconverted coal) to settle and concentrate in the bottom portion of the vessel.

The boiling range of the antisolvent is selected with the following considerations in mind:

a) The end point must be sufficiently below the initial boiling point of the dissolver section liquid feed so as to allow essentially complete recovery and minimum contamination.

b) The initial boiling point of the antisolvent should be such as to result in a deasher vessel in which the design pressure is not excessive.



NOTES:

PUMP FLOW RATES ARE AT DESIGN CONDITIONS

EXCHANGER DUTIES SHOWN ARE AT NORMAL CONDITIONS

▭	TEMPERATURE °C
○	PRESSURE kPa g

FIGURE 3

EPRI DEASHING STUDY
CE LUMMUS
ANTISOLVENT DEASHING PROCESS

ILLUSTRATIVE PROCESS FLOW DIAGRAM
AREA 100 ANTISOLVENT DEASHING
FRACTIONATION SECTION

DWG. NO. 4-8952

TABLE 17

ANTISOLVENT DEASHING/OVERFLOW & UNDERFLOW FRACTIONATION

MATERIAL BALANCE

STREAM NO.	DESCRIPTION	KG/HR
Inputs		
1	Liquefied coal & paste solvent	1,089,271
6	Antisolvent makeup	1,633
		1,090,904
Outputs		
5	Stripper Underflow Bottoms	353,658
7	Paste Solvent Recycle	529,445
8	SRC Product	207,446
9	Light Oil Product	355
		1,090,904
Internal Streams		
2	Overflow from Deashing	1,036,711
3	Underflow from Deashing	433,804
4	Antisolvent Recycle	379,613

Note that the pressure in the deashing settlers is set to keep the contents entirely in the liquid phase at the operating temperature.

Both the overflow and underflow streams from the deashing settler are fractionated at a pressure slightly above atmospheric pressure to recover the antisolvent. The overflow stream is preheated against hot circulating reflux from the Delayed Coking section and flashed into the overflow fractionator where with the aid of bottoms reboiler vapors and top reflux the antisolvent is separated from the feed. Reboil heat is provided via a separate coil in the convection section of the paste solvent fractionator feed heater.

In a similar fashion, the underflow stream is also fractionated for removal of antisolvent. This tower has a conical-shaped bottoms section to facilitate the discharge of bottoms product or stripped underflow liquid.

Small amounts of antisolvent are lost to the bottoms of the overflow and underflow fractionators since perfect fractionation can never be fully realized. These losses are made up by bringing fresh antisolvent from outside battery limits. Conservatively, total make-up antisolvent is set at 39 MT/D, representing about 0.4 wt.% of the circulating antisolvent rate to the deasher section. Compared with similar solvent processes used in lube oil refining, this make-up rate could be reduced substantially. The net cost of antisolvent make-up is very small since essentially all of the "lost" antisolvent, with the exception of mechanical leaks, spills, etc., contributes to the production of a light liquids stream produced in the complex.

The bottoms from the overflow tower is preheated and flashed into a vacuum tower for recovery of nominal 204/427°C boiling range paste solvent. Flash zone temperature and pressure are 332°C and 60 k Pa absolute pressure respectively. Heat is removed from the tower by exchanging a circulating draw-off stream against (a) boiler feedwater for production of low pressure steam and (b) air for final trim cooling. Paste solvent liquid at 210°C is withdrawn and recycled to the pasting section. The steam stripped solvent refined coal product is withdrawn at 277°C and sent to product solidification facilities (not included in the scope of this report).

Non-condensable vapors and condensable lighter boiling hydrocarbon are taken overhead and exhausted through the vacuum ejector system. Condensable light oil, consisting primarily of unrecovered antisolvent present in the overflow fractionator bottoms is withdrawn and pumped to storage where it is blended with a naphtha fraction produced in the Delayed Coking section.

b. Underflow Delayed Coking (with Process Flow Diagram)

1. Process Description

The primary purpose of the delayed coking section is to insure paste solvent self-sufficiency of this plant by converting the ash-rich underflow into paste solvent and delayed coke for oxygen blown gasification. In addition to coke, there is also produced a fuel gas and naphtha boiling range material. An illustrative process flow diagram of the delayed coking section is shown in Figure 4, with the corresponding material balance given in Table 18.

Stripped (antisolvent free) deasher underflow at 285°C is pumped at a controlled rate to the three coking heaters in parallel. Because of the potential erosion problem in the underflow coking heater tubes, return headers and U-bends should be of the castable type, with extra metal thickness on the back wall of the bends.

The underflow coking heaters are designed to fire low calorific value gas from the Koppers-Totzek gasification section. The hot flue gases are used to preheat combustion air to the burners to provide an overall fuel efficiency of about 92% based on lower heating value.

The coker heater effluents are combined in a common transfer line and flow through switch valves to the bottom of coke drums in which coke formation occurs. There are four equal-sized coke drums provided. Two drums are on-stream with the other

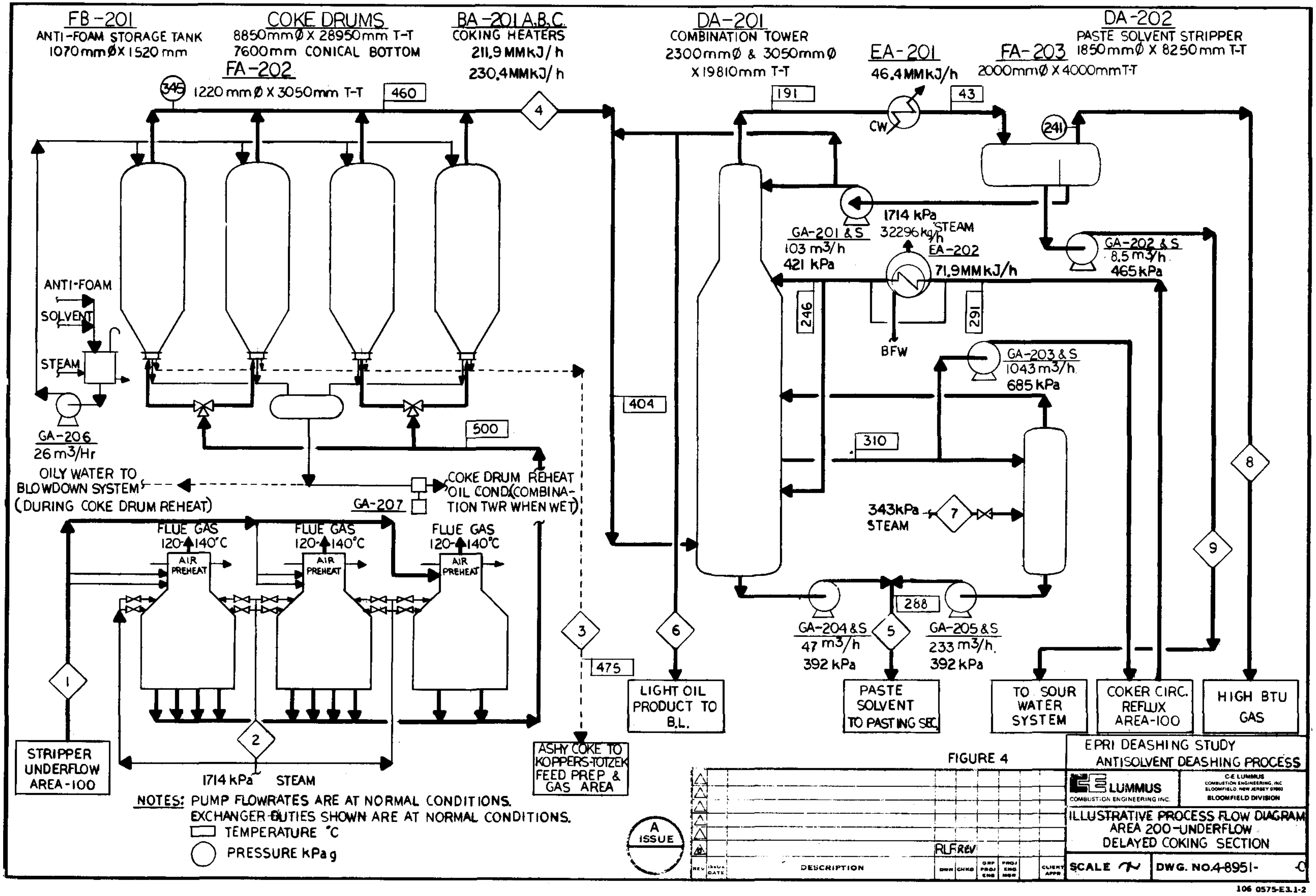


FIGURE 4

NOTES: PUMP FLOWRATES ARE AT NORMAL CONDITIONS.
EXCHANGER DUTIES SHOWN ARE AT NORMAL CONDITIONS.
▭ TEMPERATURE °C
○ PRESSURE kPa g

TABLE 18

UNDERFLOW DELAYED COKING

MATERIAL BALANCE

STREAM NO.	DESCRIPTION	KG/HR
Inputs		
1	Stripper Underflow Bottoms	353,658
2	Coking Heater Injection Stream	907
7	Stripping Steam	1,220
		355,785
Outputs		
3	Ashy Coke	129,521
5	Paste Solvent Recycle	206,684
6	Light Oil Product	5,200
8	High BTU Gas	8,913
9	Sour Water	5,467
		355,785
Internal Streams		
4	Coke Drum Overheads	226,264

two drums being decoked and prepared for the next cycle, thus maintaining continuity of operation. The coke level in the drum is determined by gamma radiation instuments mounted on the outer shell of the coke drum. After it is filled to the proper level with coke, the effluent from the coking heaters is switched to the preheated empty drum by means of the switch valve. The full drum is then steamed, water cooled and decoked.

A typical operating schedule for each drum is as follows:

	Hours
On-stream filling with coke (including switching)	24.0
Steam-Out	8.0
Water Cooling	
Draining Water	2.5
Removing Heads	.5
Hydraulic Decoking	4.0
Replacing Head, Steam Purge, and Testing	2.0
Reheating and Hold	7.0
TOTAL CYCLE TIME	48.0

Vapors leaving the coke drums at 342 k Pa abs. pressure are quenched to 404°C with cold coker naphtha before entering the combination tower.

The vapor then flows upward through a shower deck section where it is cooled by downflowing reflux liquid. This reflux causes the condensation of a small quantity of heavy material which is blended with the paste solvent stream.

Vapor leaving the lower section of the combination tower consists of steam, gaseous coker products (e.g., C_1-C_4 gas, naphtha and gas oil), plus vaporized paste solvent and the vaporized reflux stream. This vapor is cooled by a circulating reflux stream to condense a heavy oil cut for recycle as pasting solvent. The paste solvent cut is steam stripped in a separate 6-tray tower to remove lighter boiling front-end constituents and then pumped back to the pasting area.

A portion of the material drawn off at this section of the tower by-passes the paste solvent stripper and constitutes the circulating reflux stream. The heat contained in the circulating reflux is removed by heat exchange against the following:

- Overflow fractionator feed preheat
- Boiler feed water to produce saturated 1710 k Pa abs. steam.

Vapors leave the combination tower at about 191°C and pass to the overhead system where naphtha for pump-back reflux and coke drum quench is condensed. A vapor stream containing coker off gas and naphtha is withdrawn from the combination tower overhead accumulator drum at 244 k Pa absolute pressure and 43°C. This wet vapor is sent to the fuel gas system.

The delayed coker, as estimated, includes the necessary hydraulic decoking system, coke handling equipment, and auxiliary flow-down system.

2. Filtration Deashing Process

a. Rotary Precoat Filtration and Filter Precoating Area

The design of the filtration deashing system was done based on criteria and data from the following principal sources: (1) Information supplied by the Electric Power Research Institute, Palo Alto, California; (2) Information regarding the basic design and operation of the 50 ST/D SRC pilot plant in Tacoma, Washington, operated by the Pittsburgh and Midway Coal Mining Company.

Pressurized rotary precoat filtration has been a widely proposed and investigated method to separate ash and unconverted coal from coal-derived liquids. An illustrative process flow sheet for rotary precoat filtration including precoating operations is shown in Figure 5. A material balance based on 9,070 MT/SD MF coal to liquefaction is given in Table 19.

Before the actual filtration cycle begins, the filters are initially prepared by building up a thick layer (approx. 0.1m) of precoat on the outer surface of the cylindrical filter drums. The use of precoat (or filteraid) enhances cake permeability, enabling retention of the very fine particulate materials present in the feedstock (less than 10 microns), and preventing excessive clogging of the filter medium.

During the filtration cycle, feedstock at 316°C from the dissolver or hydroliquefaction area is continuously charged to the filters and maintained at a controlled level in the filter vat. An internal mixing device in each filter prevents the solid materials

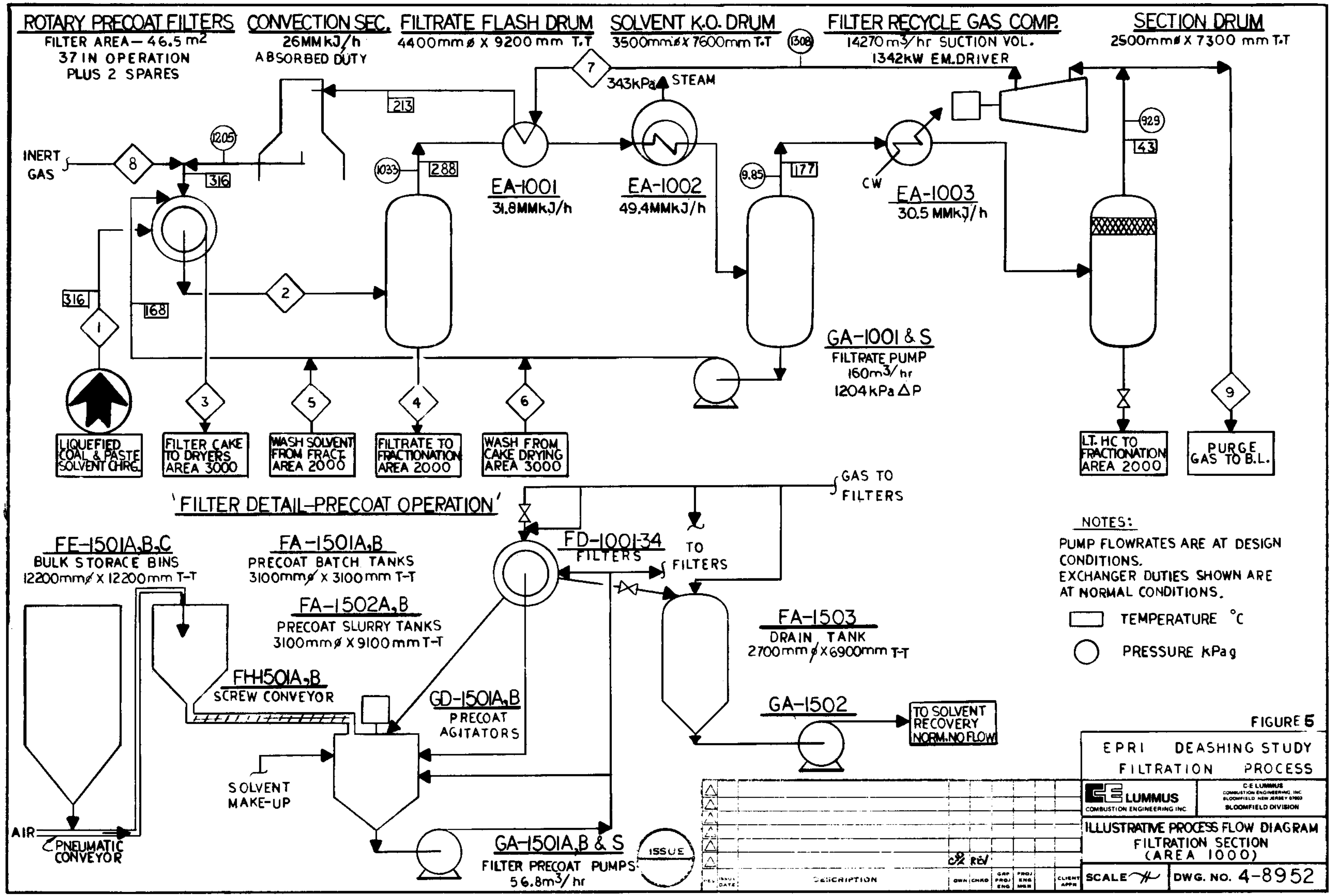


FIGURE 5

EPRI DEASHING STUDY FILTRATION PROCESS

106 0575-E3.1-2

TABLE 19

FILTRATION AREA

MATERIAL BALANCE

STREAM NO.	DESCRIPTION	KG/HR
Inputs		
1	Liquefied Coal & Paste Solvent	1,089,271
5	Wash Solvent from Fractionation	36,338
6	Wash Solvent from Cake Drying	53,212
8	Inert Gas Make-Up	8,115
	TOTAL INPUTS	1,186,936
Outputs		
3	Filter Cake*	144,384
4	Filtrate to Fractionation	1,034,437
9	Purge Gas	8,115
	TOTAL OUTPUTS	1,186,936
Internal Streams		
7	Filter Recycle Gas	175,304

* Filter cake flow rate shown does not include filteraid consumed in process.

from settling out in the vats. Solids are retained on the outer surface of the precoat and clear filtrate passes through the precoat and filter medium as the semi-submerged filter drum continuously rotates through the liquid pool. Prior to cake removal, the surface of the drum is sprayed with clean wash solvent to minimize losses of SRC product to the cake. An advancing doctor blade removes the retained solids and a thin layer of precoat, thus continuously exposing a fresh permeable surface to the slurry liquid. The wet cake is discharged from each filter by screw conveyor and routed to cake drying equipment.

The temperature during filtration is maintained between 316°C and 288°C in order to control the viscosity of the heavy liquids at a reasonable level. In order to suppress vaporization of the lighter portion of the filtrate at these temperatures, the pressure in the filters is maintained at 1202 k Pa abs. pressure by an inert recycle gas stream.

Solids free filtrate including wash solvent and recycle gas are pressured from the filters to the filtrate receiver drum. Recycle gas plus a major portion of the wash solvent are flashed as vapors at the operating conditions of 288°C and 1026 k Pa abs. pressure.

Heat is removed from the flashed vapors by heat exchange against returning recycle gas and boiler feed water for low pressure steam generation.

Recycle gas and condensed wash solvent at 177°C are separated in the wash solvent knock-out drum, with the solvent being recycled to the filters. Since the initial boiling point of the wash solvent is about 175°C, essentially all of the solvent in the original vapors is recovered at the conditions shown.

The recycle gas is further cooled to 43°C (for recovery of any light ends that might have been entrained in the feedstock) and centrifugally compressed to about 1299 k Pa abs. pressure. The compressed vapors are reheated to 316°C by heat exchange against filtrate flash drum vapors and hot flue gases in the convection section of the fired heater located in the fractionation area, and then returned to the filter.

Liquids from the filtrate receiver consisting of pasting solvent, SRC and the heavier portion of the wash solvent are pressured on level control to the fractionation area for separation and recovery.

As the filtration cycle continues, the layer of precoat on the filter drum is gradually diminished by the advancing doctor blade. When only a thin base coat remains (approx. 0.0032 meters), the filter is taken off-stream and a new layer of filteraid is applied.

During the precoat operation, filteraid from storage is slurried with solvent in a well mixed tank. The slurry is then circulated to the filter and back to the slurry tank during which time the filteraid particles bridge over the larger filter medium holes to form a stable precoat cake.

The precoat equipment shown in Figure 5 is designed to enable the simultaneous application of a 0.10 meter layer of precoat to as many as six 46.4 m^2 rotary filters.

Based on information provided by Goslin-Birmingham, the maximum size rotary precoat pressure filter currently feasible would have 46.4 m^2 of filter surface area (2.44 m diameter x 6.10 m length).

For a 9,070 MT/SD coal liquefaction plant, the number of 46.4 m^2 filters shown, (39 including 2 spares assuming filters available 95% of calendar time), was determined based on the following criteria:

a) Average filtration rate - 732 kg/hr wash solvent-free filtrate per m^2 of filter surface.

b) Precoat consumption - 0.10 kg filteraid per kg of ash removed.

c) Wash solvent ratio - approximately 1.6 kg wash solvent per kg wet cake removed.

d) Wet cake composition (including filteraid consumed)

Component	wt. %
Wash Solvent	35.9
SRC	3.7
Unconverted Coal	31.7
Ash	26.1
Filteraid	2.6
	100.0

e) Precoat material - Johns' Manville Co., Hyflo Super Cel (Cake density = 275 kg/m^3).

The number of filters for the same capacity plant was also calculated for various filtration rates, filteraid consumptions and sparing factors. Table 20 summarizes the number of rotary precoat filters required for filtration rates of 488, 732, and 977 $kg/hr/m^2$, filteraid consumption of 0.05, 0.1 and 0.2 kg/kg ash removed, and sparing factors (2) of 40, 20 and 5%.

For example, if the average filtration rate was 488 $kg/hr/m^2$ and the precoat consumption was 0.10 kg/kg ash, the number of 46.4 m^2 filters would range from 54 to 85 (depending on the sparing factor) as compared to 39 for the base case assumed for this study.

b. Filtrate Fractionation

The purpose of the fractionation areas is to separate the heavier portion of the wash solvent, the paste solvent, and the SRC product which comprise the net filtrate liquids from the filtration area.

An illustrative process flow diagram is shown in Figure 6 with corresponding material balance in Table 21.

Filtrate at 288°C is initially preflashed at 147 k Pa abs. pressure. At this pressure, all of the remaining wash solvent and nearly 50% (wt) of the pasting solvent is vaporized at the equilibrium temperature of 265°C. Vaporization of a major portion of the pasting

---

(2) Sparing factor defined as percentage of total number of filters which are not in operation, i.e..,

$$\text{Sparing Factor} = \frac{100\ (\text{Total - Operation})}{\text{Total}}$$

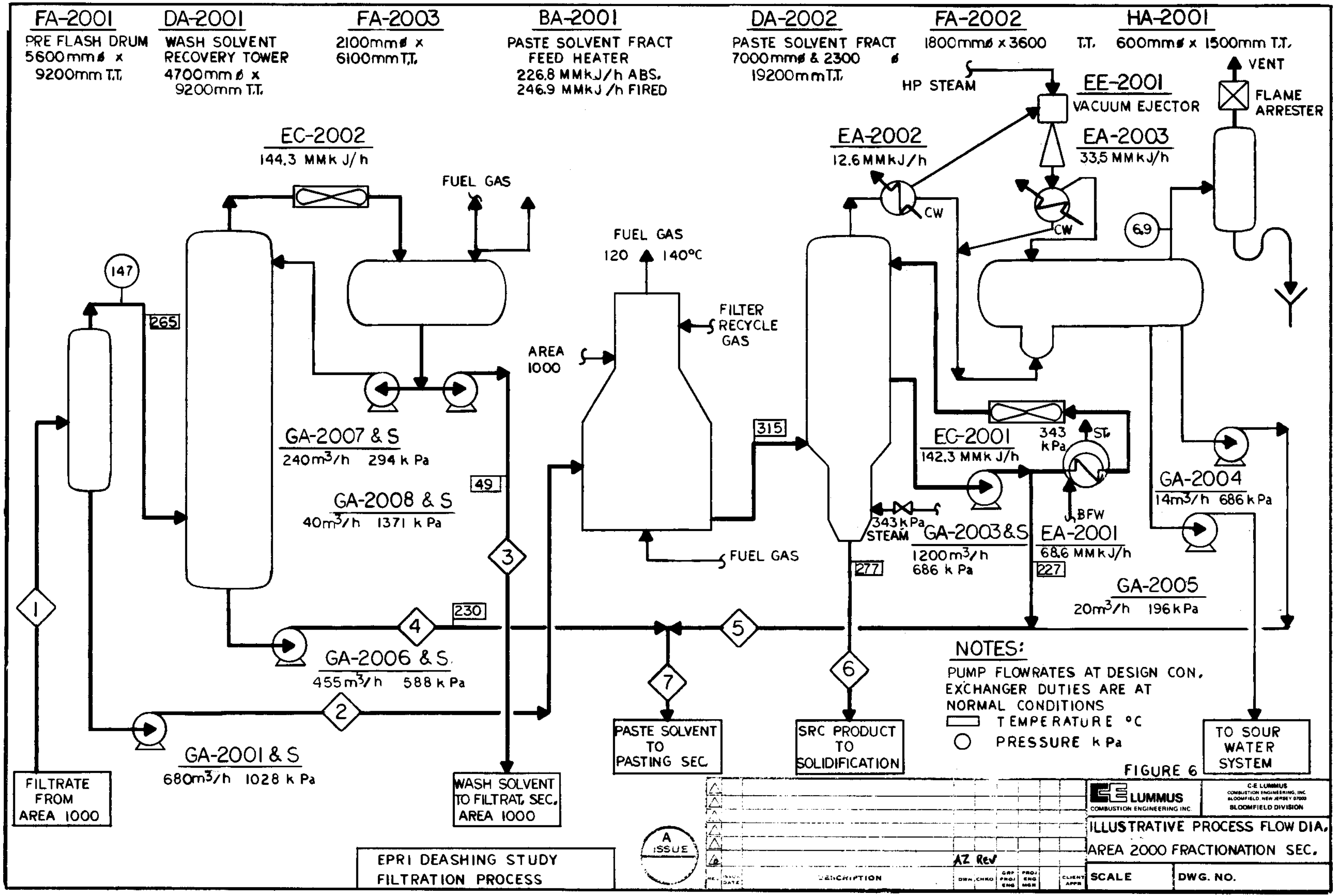
TABLE 20

NUMBER OF 46.4 M^2 ROTARY PRECOAT PRESSURE FILTERS REQUIRED AS A FUNCTION OF FILTRATION RATE, FILTERAID CONSUMPTION, AND SPARING FACTOR

BASIS: 9,070 MT/SD Coal to Liquefaction

FILTERAID CONSUMED kg/kg Ash		FILTRATION RATE - Kg/hr/m^2		
		488	732	977
0.05				
Percent Spares	5	48	34	26
	20	58	40	31
	40	77	53	42
0.1				
Percent Spares	5	54	39 (1)	32
	20	64	46	38
	40	85	62	50
0.2				
Percent Spares	5	63	48	41
	20	75	58	49
	40	100	77	65

(1) Base Case assumed for this study equals 39 - 46.4 m^2 filters, (37 operating + 2 spares)



NOTES:

PUMP FLOWRATES AT DESIGN CON.
EXCHANGER DUTIES ARE AT NORMAL CONDITIONS

▭ TEMPERATURE °C
○ PRESSURE kPa

FIGURE 6

ILLUSTRATIVE PROCESS FLOW DIA.
AREA 2000 FRACTIONATION SEC.

EPRI DEASHING STUDY
FILTRATION PROCESS

TABLE 21

FILTRATE FRACTIONATION AREA

MATERIAL BALANCE

STREAM NO.	DESCRIPTION	KG/HR
Inputs		
1	Filtrate	1,034,438
	TOTAL INPUTS	1,034,438
Outputs		
3	Wash Solvent	36,338
6	SRC Product	261,971
7	Paste Solvent	736,129
	TOTAL OUTPUTS	1,034,438
Internal Streams		
2	Pre-Flash Liquids	648,203
4	Wash Solvent Tower Bottoms	349,897
5	Paste Solvent from Vacuum Fractionator	386,232

solvent in this manner results in an overall savings in investment since downstream vacuum fractionator equipment needed to separate paste solvent and solvent refined coal product is both smaller and less complex.

Vapors from the preflash drum are condensed and separated via conventional fractionation at essentially atmospheric pressure into a wash solvent overhead stream (which is recycled to the filtration area), and a bottoms stream consisting of the front end portion of the recycle paste solvent.

Liquids from the preflash drum are preheated and flashed into a vacuum tower for recovery of the heavier portion of the paste solvent. The flash zone temperature and pressure are 305°C and 37 k Pa abs. pressure respectively.

Heat is removed from the tower by exchanging the circulating drawoff steam against boiler feed water for low pressure steam generation and air for final trim cooling. Paste solvent liquid at 227°C is withdrawn, combined with the bottoms stream from the wash solvent recovery tower and recycled to the coal pasting section. Steam stripped solvent refined coal product is withdrawn at 277°C and sent to the product solidification facilities.

Non-condensible vapors are taken overhead and exhausted through the vacuum ejector system.

c. Filter Cake Drying

The primary purpose of the cake drying area is to recover the wash solvent liquids entrained in the filter cake during the filtration cycle.

The process scheme illustrated in Figure 7 is similar to that currently in operation at the 50 TPD SRC pilot plant located in Tacoma, Washington, scaled up for a 9,070 MTD commercial size coal liquefaction plant. An overall material balance is presented in Table 22.

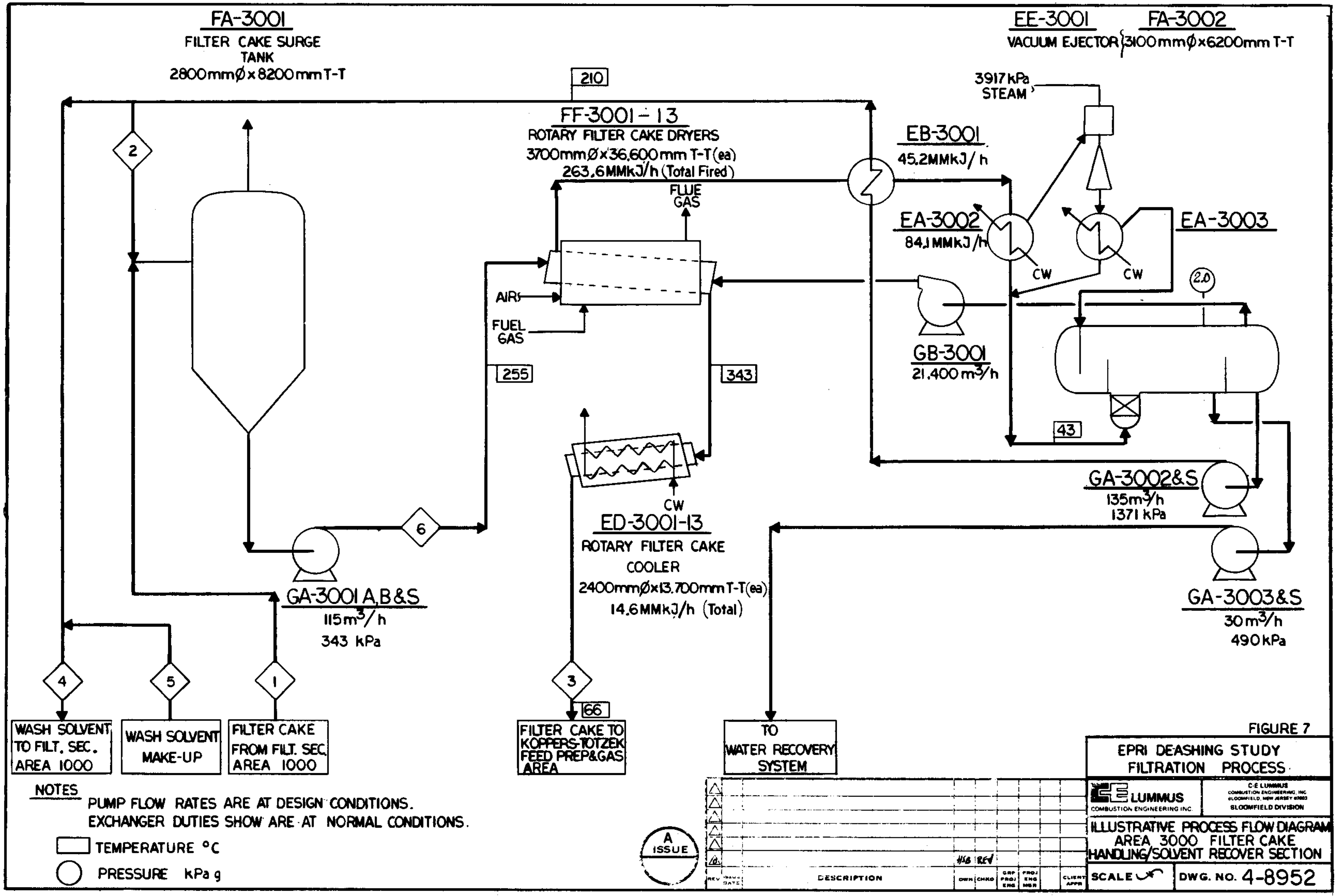
Wet filter cake, continuously discharged by screw conveyor from each filter, is diluted in-line with wash solvent to about 60 wt.% solvent. The slurry is discharged through a back pressure control valve into a surge vessel maintained at slightly above atmospheric pressure.

From the surge tank, the cake slurry is centrifugally pumped to a series of rotating dryers arranged in parallel. In the dryers, the cake is heated and the wash solvent vaporized by indirect heat exchange with hot combustion gases flowing around the outer periphery of the rotating cylinder.

As the cake cascades along the length of the inclined cylinder, the solvent vapors are swept in the opposite direction by a flow of inert purge gas. The counterflow arrangement prevents the possibility of recondensation onto the dried product.

The dried cake at approximately 343°C is discharged from each dryer by screw conveyors into chutes leading to companion water sprayed rotary coolers.

The cake is cooled to 66°C then discharged from the cooler by screw conveyors which dump the cake onto belt conveyors for transport to the gasification area.



NOTES
PUMP FLOW RATES ARE AT DESIGN CONDITIONS.
EXCHANGER DUTIES SHOW ARE AT NORMAL CONDITIONS.

▭ TEMPERATURE °C

○ PRESSURE kPa g

FIGURE 7

EPRI DEASHING STUDY
FILTRATION PROCESS

ILLUSTRATIVE PROCESS FLOW DIAGRAM
AREA 3000 FILTER CAKE
HANDLING/SOLVENT RECOVER SECTION

DWG. NO. 4-8952

TABLE 22

FILTER CAKE DRYING AREA

MATERIAL BALANCE

STREAM NO.	DESCRIPTION	KG/HR
Inputs		
1	Filter Cake (1)	148,254
5	Make-Up Solvent	5,321
	TOTAL INPUTS	153,575
Outputs		
3	Dried Filter Cake	100,363
4	Solvent to Filtration	53,212
	TOTAL OUTPUTS	153,575
Internal Streams		
2	Dilution Solvent	84,513
6	Cake Slurry	232,767

(1) Includes filteraid consumed at 0.10 kg filteraid/kg of ash.

Wash solvent vapors from the dryers are cooled and condensed by heat exchange against (1) recycle wash solvent liquids and (2) cooling water. Non-condensible purge gas is removed by the steam ejector equipment and returned to the dryer by forced draft fan which provides the necessary draft to sweep the solvent vapors out of the cylinder.

The condensed solvent is collected in the receiver vessel then recycled to the filtration area for the cake washing and dilution operations.

Ninety (90) % by weight of the wash solvent present in the washed filter cake plus any wash solvent used for cake slurring is recovered from the dryers. Thus, and amount of wash solvent equal to the wash solvent present in the dried cake fed to the gasification section is made up from storage and combined with the recycle solvent enroute to the filtration area.

Based on estimates quoted by C-E Process Equipment, Bartlett-Snow Division (who supplied the rotary dryer currently in operation at the pilot plant in Tacoma, Washington) thirteen (13) 3.66 m diameter by 36.6 m long rotary dryers coupled with 13 rotary coolers (2.44 m diameter x 13.7 m long) are required.

Each dryer consists principally of a revolving cylinder housed in a refractory lined steel furnace along its active length, inclined to the horizontal and supported in two riding rings each resting on two trunion rolls. The cylinder is rotated by means of a girt gear driven from a hardened steel pinion mounted on the slow speed shaft of a gear reducer. The high speed shaft of this reducer is directly connected to an electric motor.

The indirect rotary coolers consist essentially of a rotating cylinder slightly inclined to the horizontal and provided with a breeching at either end. The cylinder is enclosed with a jacketed section along its entire active length through which the cooling water is sprayed co-currently with the flow of cake.

D. Investment and Operating Requirements

A breakdown of estimated investment and operating requirements by major process area is given in Tables 23 and 24 for the antisolvent deashing process and the filtration deashing process, respectively.

The figures shown reflect only those areas directly associated with each deashing process and by no means represent investment and operating requirements for an entire coal liquefaction complex. In any event, manufacturing costs (both capital and operating) for areas such as coal preparation, slurrying, hydroliquefaction and hydrogen generation would be identical for either case, as these process areas are not affected by the deashing process utilized. In addition, it was determined, based on the overall material balances, that investment and operating requirements for Koppers-Totzek gasification, air separation, gas clean up areas, etc. (which are indirectly affected by deashing conditions), would also be essentially equal for both deashing processes with the exception of the supplemental raw coal requirements associated with the filtration case (see sub-section B).

The total installed costs for each area reflect statistical factors applied to purchased equipment costs for the following items:

- Erection labor required for purchased equipment.
- Commodity materials and labor (piping instruments, foundations/structures, insulation, etc.)

- Field labor benefits; equipment testing, field indirects, e.g. construction supervision, construction equipment and tools.

For the most part, purchased equipment costs were estimated based on Lummus in-house data and cost curves, the major exceptions being (1) the rotary precoat filters, quoted by Goslin-Birmingham Co. at \$600,000 per 46.4 m^2 filter (uninstalled) and (2) the rotary cake dryers and coolers, quoted by C-E Process Equipment, Bartlett-Snow Division, to be \$7.8 x 10^6 total uninstalled. Note that the investment for the filters quoted by Goslin-Birmingham does not include the necessary filter auxiliary equipment such as filter aid slurry systems, flash drums and recycle gas compressor, etc.

TABLE 23

ESTIMATED INVESTMENT AND OPERATING REQUIREMENTS

C-E LUMMUS ANTISOLVENT DEASHING PROCESS

BASIS: 9,070 MT/D (dry basis) Illinois No. 6 Coal to liquefaction section.

A. INVESTMENTS(1), 10^6 $ (1st Quarter, 1976)

1. Antisolvent Deashing Overflow & Underflow Fractionation	32.0
2. Underflow delayed coking	25.0
TOTAL INSTALLED COST	57.0

B. OPERATING REQUIREMENTS

1. Purchased Electric Power, KW

Antisolvent Deashing and Fractionation	2,500
Underflow Delayed Coking	4,400 (2)
TOTAL	6,900

2. Cooling Water, m^3/min

Antisolvent Deashing and Fractionation	10.6
Underflow Delayed Coking	18.6
TOTAL	29.2

3. Steam Consumption, MT/Hr

Antisolvent Deashing and Fractionation	(3.2)
Underflow Delayed Coking	(27.3)
TOTAL (credit)	(30.5)

4. Make-Up Water (Process) for Underflow Delayed Coking, m^3/min.	682

TABLE 23 - cont'd

B. OPERATING REQUIREMENTS - cont'd

5. Process Operator Shift Positions

Antisolvent Deashing/Overflow and Underflow Fractionation	3
Underflow Delayed Coking	8
TOTAL	11

6. Maintenance Labor, Materials and Supplies	3.5% Yearly of Total installed costs
7. Plant Administration and Overhead	150% of process labor cost

8. Catalyst & Chemicals, 10^3\$/yr.

Antisolvent Make-up at \$.30/gal or 97 \$/MT	1,170
Antifoam (for Underflow Delayed Coking)	40
TOTAL	1,210

9. Local Taxes & Insurance	3% yearly of total installed costs

NOTES:

(1) Investment figure specifically excludes the following:

- cost of land, working capital, startup expense, spare parts and catalysts/chemicals.
- return on investment during construction.
- royalty or know-how fees.
- product solidification, transportation and storage facilities.
- pre-operational services and operator training programs, etc.

(2) Includes power estimated for coke crushing and drying, etc. for gasification section.

TABLE 24

ESTIMATED INVESTMENT AND OPERATING REQUIREMENTS

FILTRATION DEASHING PROCESS

BASIS: 9,070 MT/D (dry basis) Illinois No. 6 Coal to liquefaction section, Filtration rate at 732 kg/hr/m^2 filter area, 0.10 kg filteraid/kg ash filteraid consumption and a filter availability of 95%.

A. INVESTMENTS (1), 10^6 \$ (1st Quarter, 1976)

1. Filtration and Precoating	68.6
2. Filtrate Fractionation	10.3
3. Cake Drying & Handling	20.1
TOTAL	99.0

B. OPERATING REQUIREMENTS

1. Purchased Electrical Power, KW

- Filtration and Precoating	4,290
- Filtrate Fractionation	1,400
- Cake Drying & Handling	3,400
- Supplemental Coal Crushing and Drying	630
TOTAL	9,720

2. Cooling Water, M^3/ Min

- Filtration and Precoating	13.7
- Filtrate Fractionation	14.0
- Cake Drying and Handling	55.8
TOTAL	83.5

3. Net Steam Consumption, MT/hr.

- Filtration and Precoating	(27.3)
- Filtrate Fractionation	(12.7)
- Cake Drying and Handling	13.6
TOTAL (Credit)	(26.4)

TABLE 24 - cont'd

B. OPERATING REQUIREMENTS - cont'd

4. Process Operator Shift Positions	
- Filtration and Precoating	13
- Filtrate Fractionation	1
- Cake Drying and Handling	9
- Supplemental Coal Crushing and Drying	1
TOTAL	24
5. Maintenance Labor, Materials and Supplies	3.5% Yearly of Total Installed Cost
6. Plant Administration and Overhead	150% of Process Labor Costs
7. Catalyst & Chemicals Consumption, M$/Yr.	
- Filteraid @ 4.5 ¢/lb. or 9.9 ¢/kg	3,040
- Wash Solvent Make-up at $.30/gal. or 97 $/MT	3,803
TOTAL	6,843
8. Local Taxes and Insurance	3% Yearly of total installed costs

NOTES:

(1) Investment figure specifically excludes the following:

- cost of land, working capital, start-up expense, spare parts and catalyst/chemicals.
- return on investment during construction.
- product solidification transportation and storage facilities.
- Pre-operational services and operator training programs, etc.

(2) Wash solvent makeup requirements based on 90% recovery of solvent in the washed filter cake.

E. Comparative Economics

1. Basis and Assumptions

The economics for each process are evaluated using the Discounted Cash Flow technique to determine manufacturing costs including a 12% DFC return (after tax) on equity, during the plant life. A detailed listing of the basis and assumptions for the economic evaluation is given in Table 25.

Unit costs assigned to utilities and byproducts, and factors given for maintenance, payroll and benefits, plant supervision, administration and overhead, corporate income taxes and local taxes and insurance represent typical present day U. S. conditions.

Applying the ground rules itemized in the table, the following simplified formula was derived to obtain overall deashing costs:

$$M = E + .2615\ I + .1339\ S$$

M = Manufacturing Costs, 10^6\$/Year

E = Net operating expenses, 10^6\$/Year.

I = Total Installed Cost of deashing facilities, 10^6 \$ (including initial catalyst & chemicals inventory)

S = Start-Up Costs, 10^6 \$

In deriving the above equation, capital costs including plant investment and return on investment during construction and start-up expenses are treated as cash flows occurring in year "0", (the year ending with completion start-up).

TABLE 25

BASIS AND ASSUMPTIONS FOR ECONOMICS VIA DCF METHOD

Time Period for Investment:	1st Quarter - 1976/no forward escalation included
Construction Period including start-up:	2 years
Project Life (after start-up):	20 years
Plant Operations:	330 Stream Days/Year - 100% of Nominal Capacity throughout Project Life
Illinois No. 6 Coal Price (wet coal as delivered):	$20/ST or $22 MT
Utilities:	
- Electrical Power	2.5 ¢/KWH
- Cooling Water	3.2 ¢/M gal.or 0.85¢/M^3
- Steam	$5.00/ST or $5.50/MT
- Process Water (Make-Up)	8 ¢/M gal.or 2.11¢/M^3
Process Labor:	20,000 $/Year per person (at 4.5 shift coverage factor, this is equivalent to 90,000 $/shift position/year including 35% payroll benefits)
Maintenance Labor, Materials and Supplies:	3.5% TIC/year
Supervision, Plant Administration, and General Overhead:	150% of Process Labor (including benefits)
Local Taxes & Insurance	3% TIC/year
By-Product Credits:	
- Coker Fuel Gas	$\$2.50/10^6$ BTU
- Coker Distillate	$ .30/gallon or $97/MT (1)

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(1) Based on HHV of approximately 140,000 BTU/Gal. or 45,363 BTU/kg @ $\$2.10/10^6$ BTU.

TABLE 25 - cont'd

Depreciation:	16 years, straight time
Debt Financing:	None
Return on Equity (after tax):	12% (DCF Basis)
Corporate Tax Rate on Earnings	48%

2. Summary of Economics

Capital requirements, operating costs and manufacturing costs for each process are summarized in Table 26.

Investment related items such as return on investment during construction, start-up expenses and initial catalyst/chemicals are itemized separately and added to the installed costs in obtaining the total capital requirements for each process.

The cost of capital during construction equivalent to a 12% return on investment during the construction period was approximated by:

$$I_c = (.01)\ (1.875)\ (12)\ I = .225I \ {}^{*}$$

where I = Total Installed Cost

Start-up expenses were estimated using a 6-month initial operation period averaging 50% of design capacity. Start-up expenses include, 6 months payroll at 150% of normal salary, catalyst and chemical drawdown, raw materials and purchased utilities requirements. No credit is taken for SRC and by-products produced during the start-up period.

In determining overall manufacturing costs we have considered two separate cases for the filtration process. For the base case, the overall operating costs reflect an SRC yield derived from washed filter cake data provided by EPRI (Appendix C).

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* Representative of typical ROI for construction period of 2 years as used by the FPC Synthetic Gas Coal Task Force in their report dated April 1973.

As an alternate base case, we have projected a filtration performance for the same deashing feedstock used for the antisolvent deashing tests performed in the Lummus pilot plant. The SRC filtration yield estimated for this alternate base case is about 5% lower than that determined from the Appendix C data for a more highly converted (circa 92-93%) feedstock. Capital requirements are assumed to be the same for each filtration case.

In order to arrive at a true difference in ¢/10^6 BTU of product between the two deashing methods, we have elected to consider the antisolvent deashing yield as the common denominator and treat any incremental SRC yield from filtration as a byproduct credit toward operating expenses for that scheme. The extra or supplemental coal for the filtration method is treated as an operating cost.

As shown in Table 26, the total manufacturing costs of deashing are estimated to be 21¢/10^6 BTU SRC product for the CE Lummus Antisolvent Deashing Process and 41¢/10^6 BTU SRC product for the base case rotary precoat filtration process, which is equivalent to a savings of 20¢/10^6 BTU in favor of the antisolvent deashing process (ca $1.37/Bbl of SRC). For the lower filtration yield case or alternate base case this advantage is increased to 26 ¢/10^6 BTU ($1.78/Bbl of SRC).

TABLE 26

ECONOMIC COMPARISON OF C-E LUMMUS ANTISOLVENT DEASHING AND FILTRATION DEASHING PROCESSES

BASIS: 9,070 MT/D (dry basis), Illinois No. 6 Coal to liquefaction, Solvent Refined Coal HHV @ 37,216 kJ/kg, 12% DCF After Tax Return, 100% Owner's Equity, 330 Operating Days/Year

A. INVESTMENTS, 10^6\$/1st Quarter 1976 (ex. land and working capital)

	C-E Lummus Antisolvent Deashing	Filtration Deashing
1. Total Installed Cost, Process Units	57.0	99.0
2. Return on Investment during construction	13.1	22.4
3. Start-Up Expenses	2.4	8.9
4. Initial Catalyst/ Chemicals	1.0	0.8
TOTAL CAPITAL REQUIREMENT	73.5	131.1

TABLE 26 - cont'd

	Antisolvent Deashing	Filtration Deashing Base Case (1)	Filtration Deashing Alt. Base Case (2)
B. OPERATING COSTS (10^6\$/YR)			
1. Supplemental Raw Coal @ \$22/MT (\$20/ST) (3)	-	7.2	3.0
2. Utilities Consumption	0.2	1.0	1.0
3. Catalyst/Chemicals	1.2	6.8	6.8
4. Process Labor	1.0	2.1	2.1
5. Administration, Supervision and Plant Overhead	1.8	3.2	3.2
6. Maintenance Labor, Materials & Supplies	2.0	3.5	3.5
7. Local Taxes and Insurance	1.7	3.0	3.0
8. Byproduct Credits:			
Net coker fuel gas @ \$2.50/10^6 BTU	(7.3)	-	-
Net coker distillate @ \$.30/gal or \$97/MT	(4.3)	-	-
Credit for additional SRC product yield @ \$2.10/10^6 BTU (4)	-	(30.9)	(23.5)
TOTAL OPERATING COSTS	(3.7)	(4.1)	(0.9)

(1) Total SRC Yield = 6287 MT/D (70.7 x 10^{12} BTU/Yr) based on Appendix C data for washed filter cake.

(2) Total SRC Yield = 5973 MT/D (67.2 x 10^{12} BTU/Yr) estimated for same feedstock used for antisolvent deashing tests.

(3) Raw coal available @ 10 wt.% moisture.

(4) Determined by subtracting SRC yield via antisolvent deashing (56 x 10^{12} BTU/Yr) from SRC yield via filtration.

TABLE 26 - cont'd

C. MANUFACTURING COSTS, 10^6 \$/YR

	Antisolvent Deashing	Filtration Deashing Base Case	Filtration Deashing Alt. Base Case
1. Operating Costs	(3.7)	(4.1)	(0.9)
2. Capital Charges	15.5	27.3	27.3
TOTAL MANUFACTURING COSTS	11.8	23.2	26.4

D. UNIT MANUFACTURING COSTS, ¢/10^6 BTU (Based on 56 x 10^{12} BTU/Yr. Product SRC, HHV of 37,216 kJ/kg, or 35,274 BTU/kg

	Base Case	Alt. Base Case
1. Filtration Deashing	41.4	47.1
2. Antisolvent Deashing	21.1	21.1
DIFFERENCE	20.3	26
	(\$1.37/Bbl/SRC)	(\$1.78/Bbl SRC)

The design criteria selected to determine the economics for the base filtration case shown in Table 26 reflect a rather high degree of filter reliability (95%), which has not yet been achieved on a continuous basis in actual practice. We have evaluated the economics of the filtration process for a wide range of operating parameters including various filtration rates, filteraid consumptions and sparing factors.

Table 27 illustrates the effect on the economics of rotary precoat filtration of varying these criteria for the same capacity plant. The table indicates that for less optimistic design criteria, the manufacturing costs associated with filtration would increase significantly. For example, if the onstream filtration rate was 488 kg/hr/m^2 and filter reliability was equal to 80% of calendar time, the deashing cost for the filtration process would increase by 22 ¢/10^6 BTU SRC product for a total cost of 63 ¢/10^6 BTU. Comparing the two processes on this basis would result in savings of 42 ¢/10^6 BTU for antisolvent deashing which is about double the advantage shown for the base case.

TABLE 27

SENSITIVITY EFFECTS(1) ON ROTARY PRECOAT FILTRATION ECONOMICS

VARIATIONS(2)	CHANGE IN MANUFACTURING COST,(3) Δ ¢/10^6 BTU (HHV)		
1. Filtration Rate			
488 kg/hr/m^2	+ 13		
977 kg/hr/m^2	- 6		
2. Filter Aid Consumption			
0.05 kg/kg ash	- 9		
0.2 kg/kg ash	+ 12		
3. Filter Sparing Factor	Filtration Rate - kg/hr/m^2		
	488	732	977
20% (80% availability)	+ 9	+ 7	+ 6
40% (60% availability)	+ 27	+ 22	+ 18

NOTES:

(1) Sensitivities are relative to base case (41 ¢/10^6 BTU SRC product) presented in Table 26.

(2) Base case assumes:

- filtration rate = 732 kg/hr/m^2
- filter aid consumption = 0.10 kg/kg ash
- filter sparing factor = 5%
- overall SRC yield = 70.7 x 10^{12} BTU/Yr.

(3) In order to obtain individual differences relative to Antisolvent Deashing, add 20¢/10^6 BTU.

APPENDIX A

ANALYTICAL METHODOLOGY

TABLE A-1

ANALYTICAL METHODOLOGY

Determination	Analytical Method Employed
Specific Gravity	ASTM D-287
Ash (Solid samples)	ASTM D-271
Ash (Liquid samples)	ASTM D-482
Benzene Insolubles	ASTM D-367-67 Modified
Sulfur Content	ASTM D-1552
Carbon, Hydrogen, Nitrogen Contents	Via Perkin Elmer Model 240 Elemental Analyzer
Vacuum Distillation Analysis	Improved ASTM D-1160-61
Softening Point (R&B)	ASTM D-36-70
Proximate Analysis (Coke)	ASTM D-3172-73
Quinoline Insolubles	ASTM D-2318-66 Modified

APPENDIX B

MISCELLANEOUS ANALYTICAL DATA

TABLE B-1 (1)

ELEMENTAL ANALYSIS OF VACUUM DISTILLATE (800°F- or 700°K-) OF ANTISOLVENT FREE OVERFLOW PRODUCT

Carbon Content, Wt.%	89.44
Hydrogen Content, Wt.%	6.95
Sulfur Content, Wt. %	0.48
Nitrogen Content, Wt. %	0.69
Ash Content, Wt. %	Nil
Oxygen Content (Via Dif.), Wt.%	2.44

(1) Refer to Table 4

TABLE B-2 (1)

ELEMENTAL ANALYSIS OF VACUUM DISTILLATE (550°F- or 560°K-) OF ANTISOLVENT FREE UNDERFLOW PRODUCT

Carbon Content, Wt.%	87.68
Hydrogen Content, Wt. %	8.23
Sulfur Content, Wt. %	0.19
Nitrogen Content, Wt. %	0.40
Ash Content, Wt. %	Nil
Oxygen Content (via Dif), Wt.%	2.44

(1) Refer to Table 5

TABLE B-3 (1)

APPROXIMATE ELEMENTAL ANALYSIS OF COKE SAMPLES

Lab Coking Run No.	LC-131	LC-133-A
Source of Coker Feedstock	Wilsonville Dry Mineral Residue	DA-53-2 Underflow
Elemental Analysis		
Carbon Content, Wt. %	25.06	64.62
Hydrogen Content, Wt. %	0.78	1.96
Sulfur Content, Wt. %	3.32	2.06
Nitrogen Content, Wt. %	0.20	1.46
Ash Content, Wt. %	69.41	29.81
Organic Oxygen Content (Via Dif.), Wt.%	1.23	0.1

(1) Due to the difficulty in sampling this type of material and the low levels of the hydrogen content, the hydrogen and carbon contents must be considered approximate.

TABLE B-4

UNDERFLOW DERIVED ASH COMPOSITIONAL ANALYSIS FOR PILOT PLANT DEASHING, RUN DA-53-2

	Ignited
Phos. Pentoxide (P_2O_5), Wt. %	0.09
Silica (S_iO_2), Wt. %	50.00
Ferric Oxi[illegible]Fe_2O_3), Wt.%	19.92
Alumina (Al_2O_3), Wt.%	20.92
Titania (TiO_2), Wt. %	0.72
Lime (CaO), Wt. %	3.00
Magnesia (MgO), Wt.%	1.01
Sulfur Trioxide (SO_3), Wt.%	1.72
Potassium Oxide (K_2O), Wt.%	0.52
Sodium Oxide (Na_2O), Wt.%	0.52
Undetermined, Wt.%	0.10
	100.00

Silica Value = 67.63

T_{250} Poises = 1615°K

TABLE B-5

MOISTURE FREE GAS CHROMATOGRAPHY ANALYSIS OF A SAMPLE WITHDRAWN FROM THE BENCH COKING GAS HOLDER AT THE END OF RUN LC-133-A

Component	Wt. %
Hydrogen	6.86
Carbon Monoxide	8.40
Carbon Dioxide	4.59
Methane	52.30
C_2	8.64
C_3	5.31
C_4	3.05
C_5	0.98
C_6 and C_6+	9.87
	100.00