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SHALE BENEFICIATION AND OIL RECOVERY FROM THE CONCENTRATE

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ABSTRACT

A flow sheet and economic feasibility study of oil shale beneficiation and recovery showed that conceivable improvements in grinding, separation, and recovery could make a beneficiation-based system more attractive than conventional ore retorting although current technology (based on ball milling, froth flotation, and retorting) is less attractive. Experiments are needed.

1. INTRODUCTION

The purpose of this study was to assess broadly the potential attractiveness of new large-scale systems for recovering oil from oil shale, systems which incorporate a beneficiation step to substantially increase kerogen concentration in the feed to a retort or to other equipment for converting kerogen to oil. The work was sponsored by the U.S. Department of Energy and is reported in detail elsewhere [1].

The enormous magnitude of oil shale deposits in the United States is well known. More oil is contained in the rich tick deposits of Colorado's Piceance Basin oil shales alone than in all the proved petroleum reserves in the Middle East. However, oil recovery from shale, while technically feasible, is uneconomical because only about 10–15% of the mass of rich shales can be converted to marketable energy.

A potential solution to this problem is beneficiation. While its application to oil shale has not been thoroughly researched, benefication of metal-containing ores and coal is standard commercial practice. However, concentration of metal values or combustible substances in lean ores by benefication is costly. Obviously, benefication of oil shale has to be coupled with a recovery process that would compensate for the increased upstream costs by taking full advantage of the processing characteristics of the concentrate.

2. METHODOLOGY

The study is based on Western oil shale with a Fischer assay of 35 gallons oil/ton as a basis, exemplified by the Tosco Colony deposit [2]. The equipment and process alternatives investigated are summarized in the second, third, and fourth rows in Fig. 1 For each alternative, a flowsheet, material balances, and design criteria were developed, and capital and operating costs were estimated. Process design for the mining and crushing operation was not done but its costs were roughly estimated and incorporated into the final evaluation. Out of the eleven alternatives examined, four were based on actual test data, three assumed that experi-

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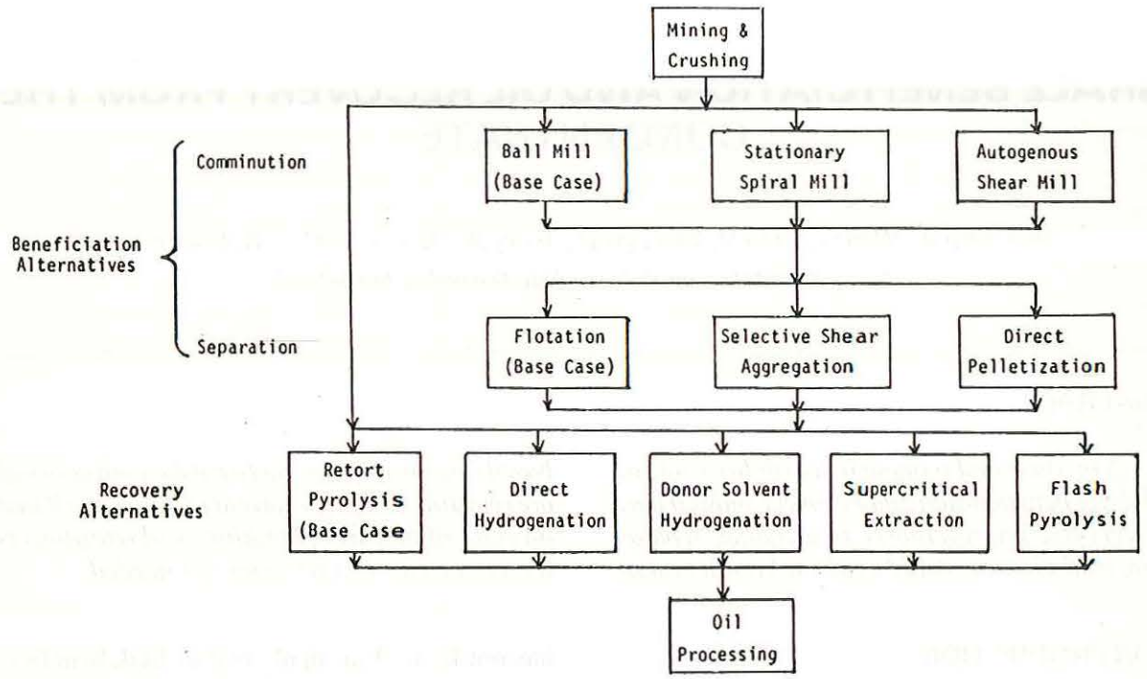


Fig. 1. Oil shale processing options.

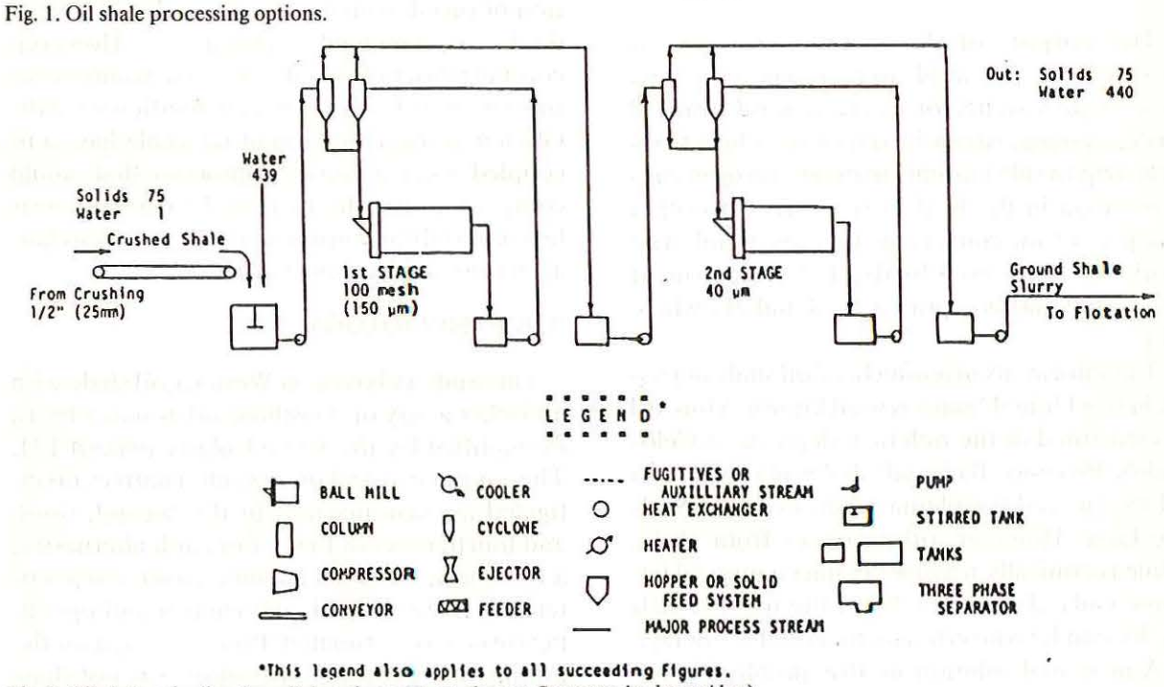


Fig. 2. Oil shale grinding (parallel equipment not shown; flowrates in *k* tons/day).

TABLE 1

Design criteria (Base case)

A. Pyrolysis and fractionation	
1. Material balance	
Raw shale (*k* tons/day)	66
Shale oil recovery (%)	90
Moisture (%)	
raw shale	1.4
spent shale	14
pyrolysis vapor	1.2
Pyrolysis vapors (1bs/1b shale)	0.182
Fractionation products (wt.%)	
gas	25.1
naptha	10.4
gas oil	45.6
bottoms oil	18.9
Balls (1bs/1b shale)	1.5
2. Temperatures (°F)	
Shale feed after preheater	500
Shale feed to retort	900
Balls to retort	1300
Flue gas after preheater	130
Spent shale after cooler	300
Spent shale after moisturizer	200
B. Beneficiation	
1. Material balance	
Raw shale	75
Enrichment factor	4.
Beneficiation efficiency (%)	88
Overall water losses (%)	5.
Specific gravities	
shale	2.2
kerogen	1.07
Pulp densities (%)	
flotation feed	15
concentrate slurry	25
tailings slurry	13
Moisture (%)	
tailings filter cake	25
concentrate filter cake	20
concentrate pellets	10
pellets to dryer	20
pellets after dryer	10
2. Grinding power requirements (kWh/ton)	
Work index	38
First stage (0.5" to 100 mesh)	25
Second stage (100 mesh to 40 microns)	26
Regrind (40 to 20 microns)	22
3. Equipment parameters	
Flotation residence time (based on feed), min.	8.
Filtration capacity (gal/ft²/hr)	
tailings	50
concentrate	25

TABLE 2

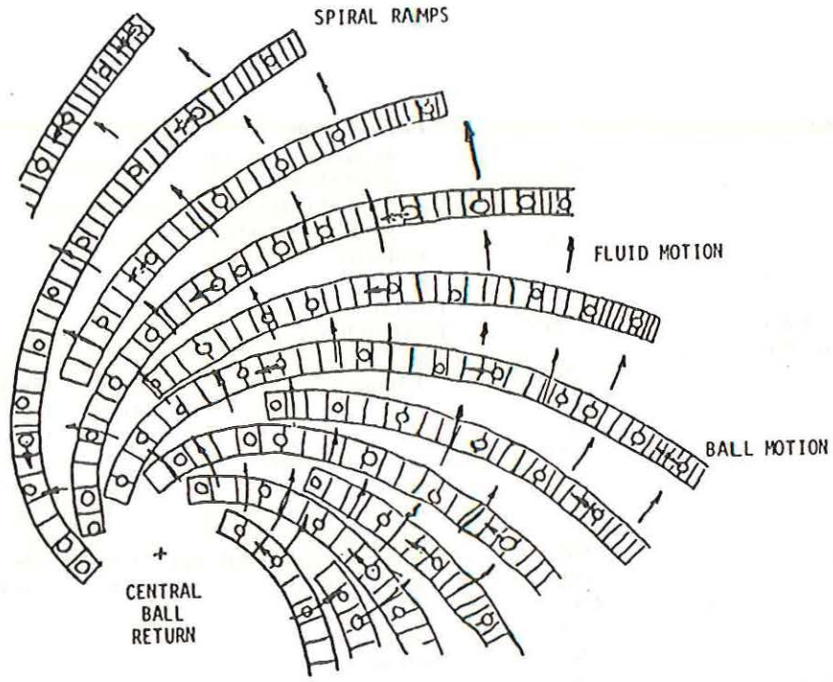
Requirements for materials, utilities and labor (base case)

1. Process materials (lbs/ton dry feed)			
Grinding balls			
forged steel (stage 1)			0.8
cast steel (stage 1)			0.7
cast steel (stage 2)			1.1
cast steel (regrind)			1.1
Collector			0.1
Conditioner			7.0
Frother			0.2
Pyrolysis balls			2.0
2. Utilities	Shale pyrolisis	Concentrate pyrolisis	Beneficiation
Fuel (M Btu/hr)	1300	500**	250
Power (MW)	0.44	0.11	180
Make-up water (*k* gal/min)	3.2	0.7	7
Steam, generated (*k* 1b/hr)	120	40	-
Air* (*k* scfm)	1	0.3	5
Manning			
Grinding		37	
Flotation		42	
Shale pyrolysis		94	
Concentrate pyrolysis		54	

*Power include above.
**Includes heat required to evaporate water in concentrate.

mental results using coal would also apply to using oil shale, and four are entirely speculative. Therefore there is high uncertainty in the absolute evaluations although the relative results should be reasonable. The oil upgrading plant and other facilities, such as offsites, were not included in this study.

The three communication alternatives were compared with each other. Three benefication alternatives were defined as flotation, selective shear aggregation, and direct pelletization, each including ball mills as the base case comminution equipment. Similarly, five recovery alternatives were defined by combining the five processes in the fourth row in Fig. 1 with the base case ball mill–flotation option. Retort pyrolisis of the crushed, unbeneficiated shale constituted the sixth recovery option. The three benefication alternatives were then compared with each other; so were the six recovery alternatives.



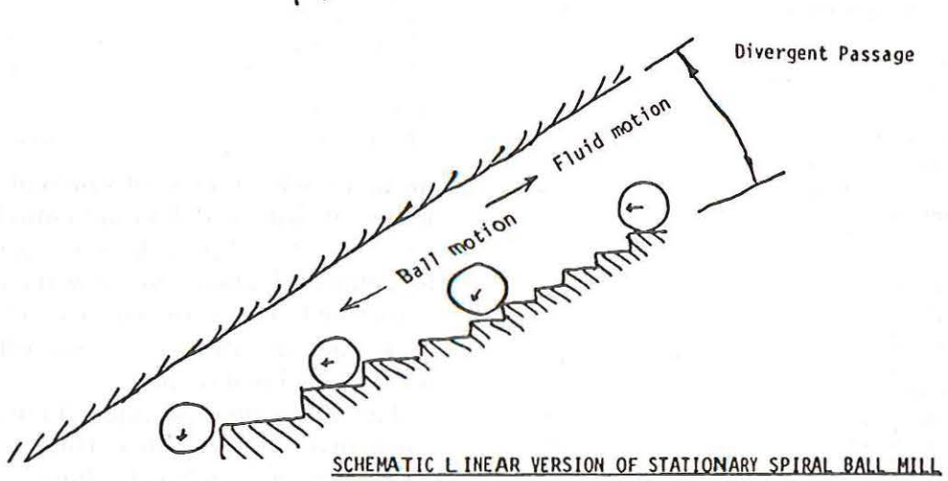


Fig. 3. Stationary spiral ball mill (plan view above).

In the subsequent sections, the individual alternative processes will be discussed first. The comparisons will be presented and conclusions drawn in the final section.

3. BALL MILLING

Process design (Fig. 2; Tables 1,2) is based on the industrial practice of ore concentrating and a specific oil shale study supported by lab experiments [3]. The crushed shale is mixed with water and fed to feed cyclones of the first grinding stage. The coarse fraction passes to the ball mills together with material from the recycle cyclones. The fines from both sets of cyclones are pumped to the second grinding stage which consists of the same type of equipment as the first stage.

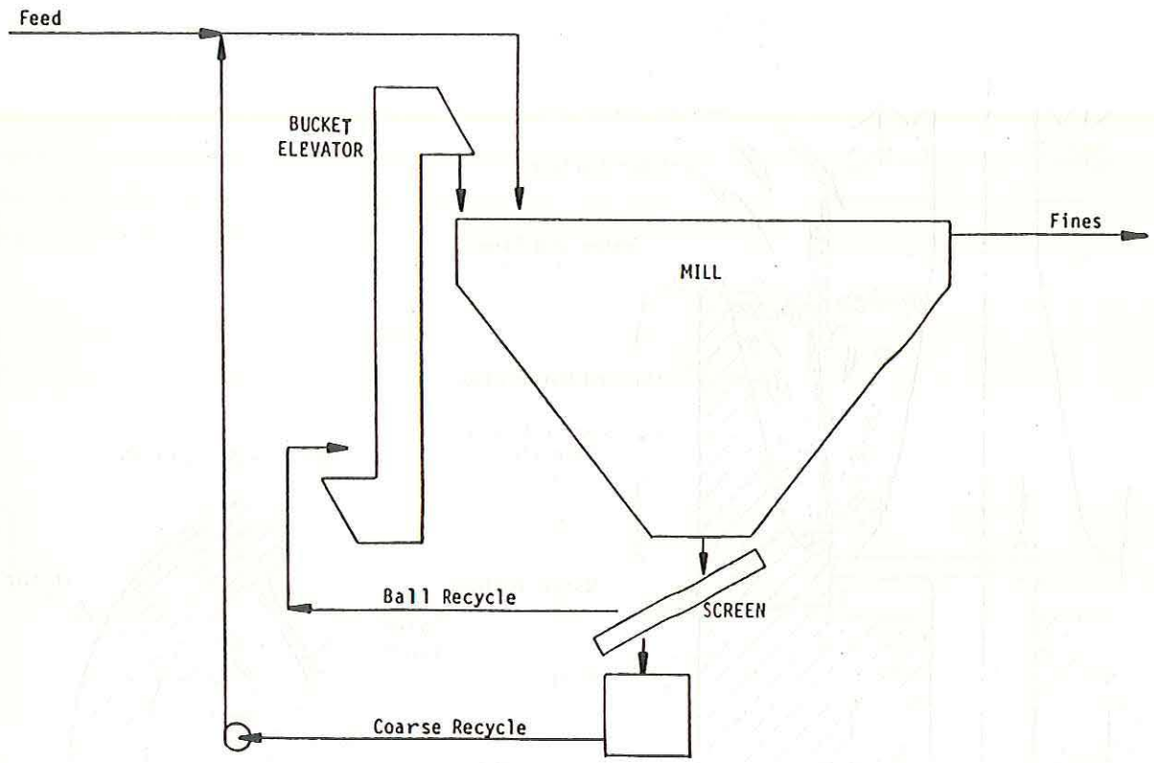


Fig. 4. Stationary spiral mill.

4. STATIONARY SPIRAL MILL

This machine was proposed by C.R. Peterson based on a theoretical study [1]. The stationary spiral ball mill utilizes an indirect energy transfer process consisting of conventional balls moving downward through the material being ground. The concept is best illustrated in simple linear form as sketched in Fig. 3. A stream of balls rolls down a stepped ramp to crush material deposited on each of the step surfaces. In this non-random descent, each step can be sized to match the local particle energy requirements and, further, by impact against a flat surface, the capture volume per impact is double that of a conventional ball mill.

While the stream of balls moves down the steps an upward fluid flow moves the material to be ground up the steps. The fluid passage is designed with a divergence so that fluid velocity decreases as it progresses upward. Coarse material is fed into the mill at the bottom and carried upward by the fluid, but only so far as the local velocity can carry each particle. The fluid acts not only as a transport mechanism, but also as a classifier, dropping out particles at different positions in accordance with their size. Once a particle drops out, it remains stationary until it is crushed, whereupon the smaller fragments are again transported uphill and dropped out at new size-dependent positions, and so it continues until particles are small enough to be carried out of the top of the device. This classification of material according to size is what permits matching the step sizes to the local energy requirements.

Ancillary equipment would include a screen to separate the balls from the coarse shale fraction, a bucket elevator for the balls, and a slurry

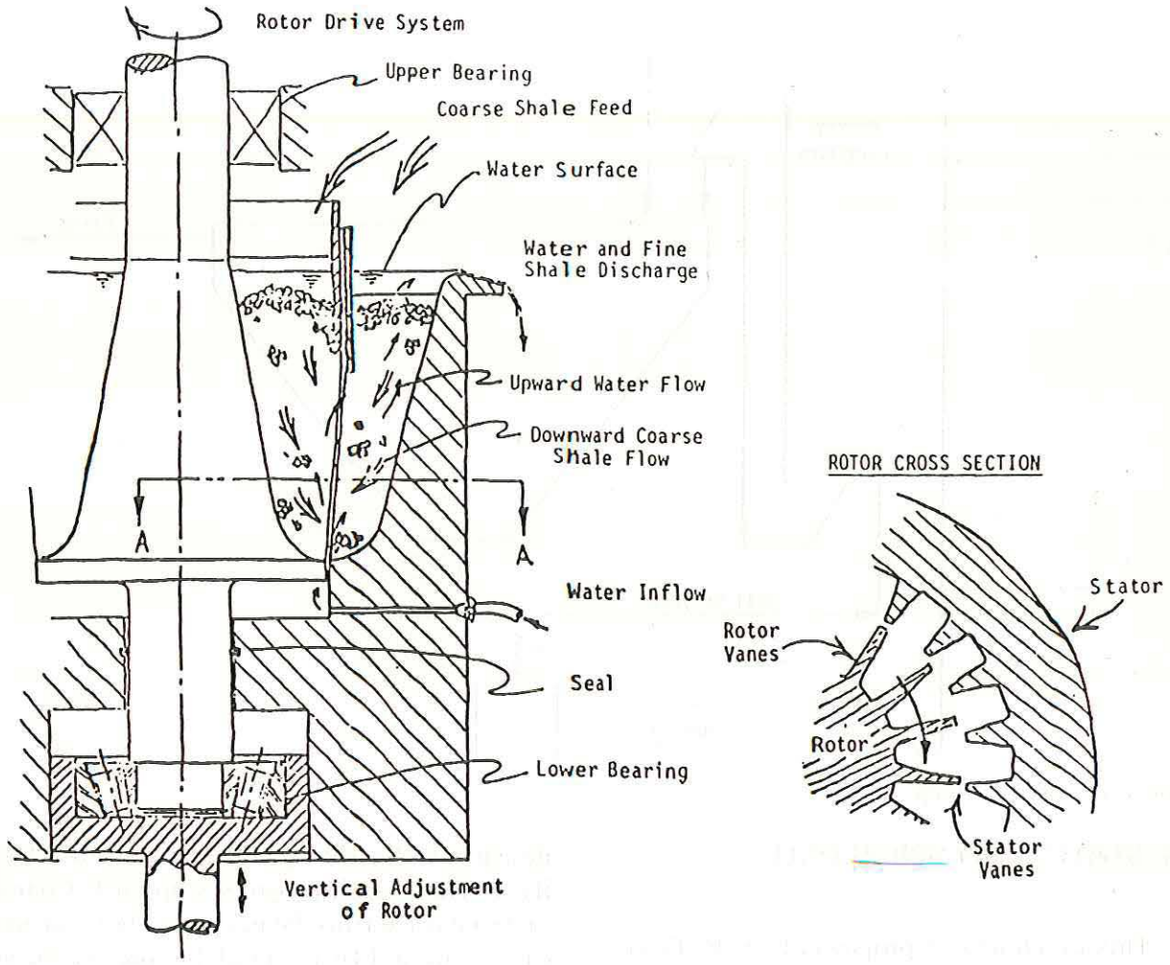


Fig. 5. Vertical section of autogenous shear mill.

tank and pump to recycle the coarse fraction (Fig. 4). The screen might be an integral part of the mill. The other process equipment of the grinding plant section would be the same as that for ball mill grinding (Fig. 1).

5. AUTOGENOUS SHEAR MILL

This machine was proposed by C.R. Peterson on a theoretical study [1]. The design aims include direct energy input, effective and graded removal of small particle as formed, rugged simple construction, low wear of component parts and high power density. The autogenous shear mill is designed so that virtually all of the input energy is consumed directly by shale fragments shearing against one another (Fig. 5).

A vaned rotor turns about a vertical axis within a vaned stator. The two vaned elements could counter-rotate, but there seems little advantage to this greater complexity. A separate downward flowing column of shale fragments is top-fed into each vaned element. The basic object of the design is to establish relative rotation between these columns and to concentrate their interactions in a relatively thin shear zone where comminution will take place, probably as much by attrition of coarse parti-

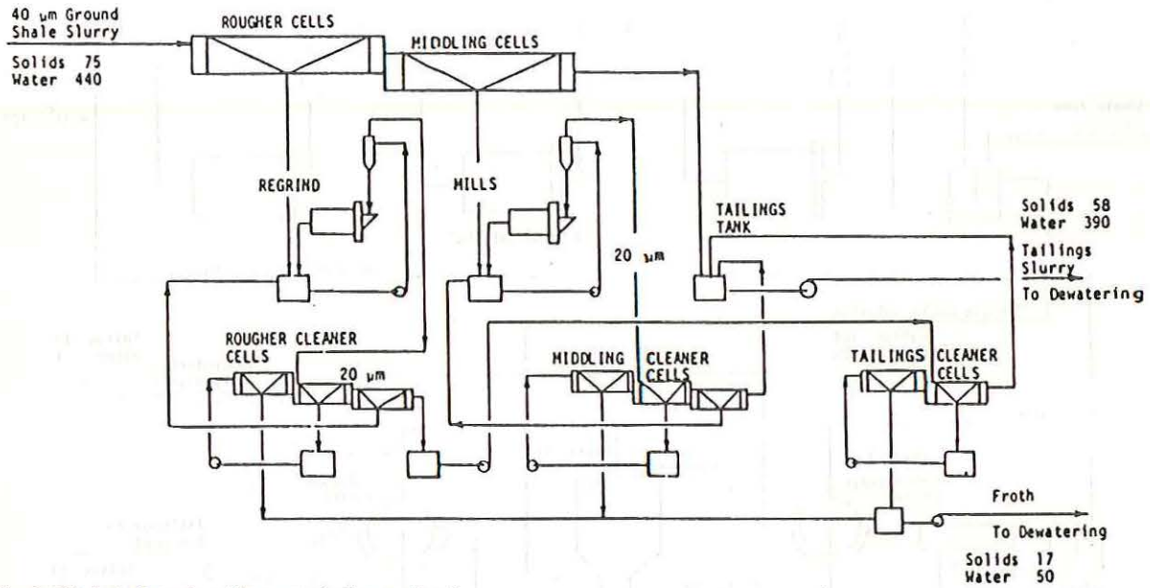


Fig. 6. Oil shale flotation (flowrates in k tons/day).

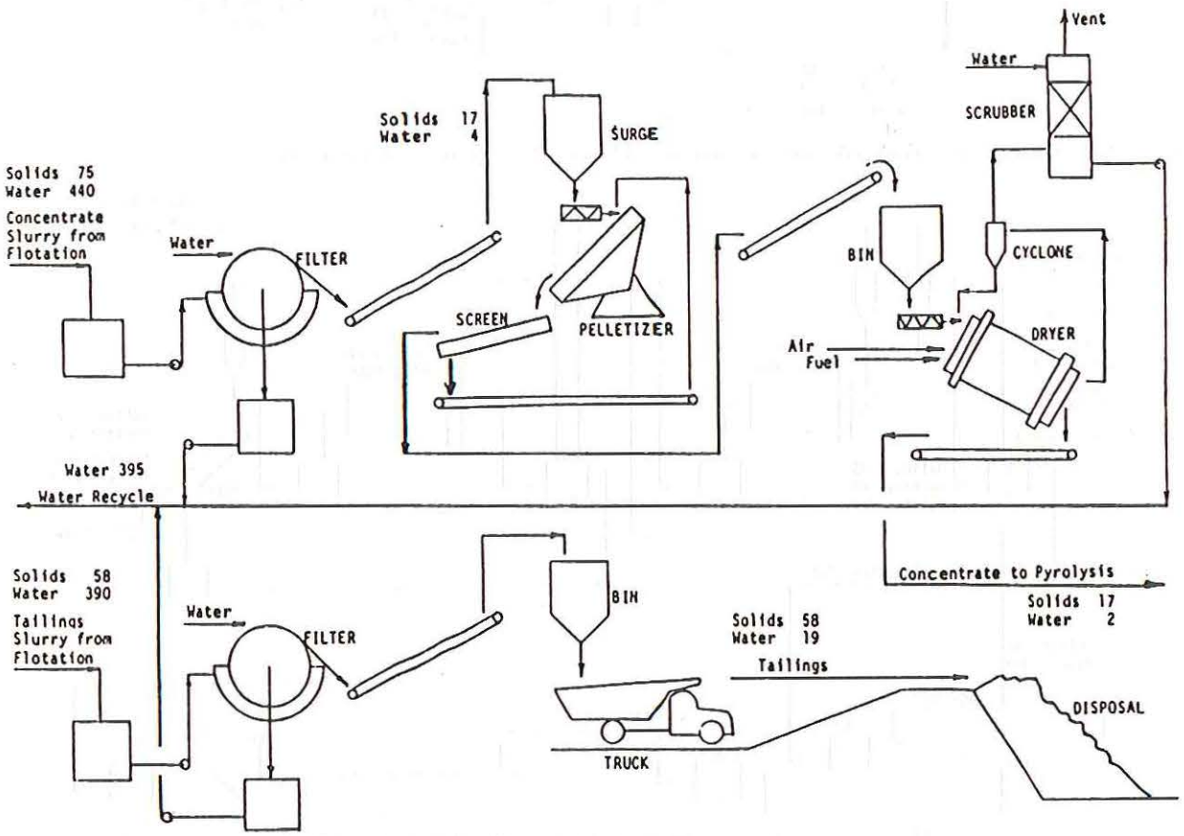


Fig. 7. Oil shale concentrate and tailings dewatering (flowrates in k tons/day; water losses = 24).

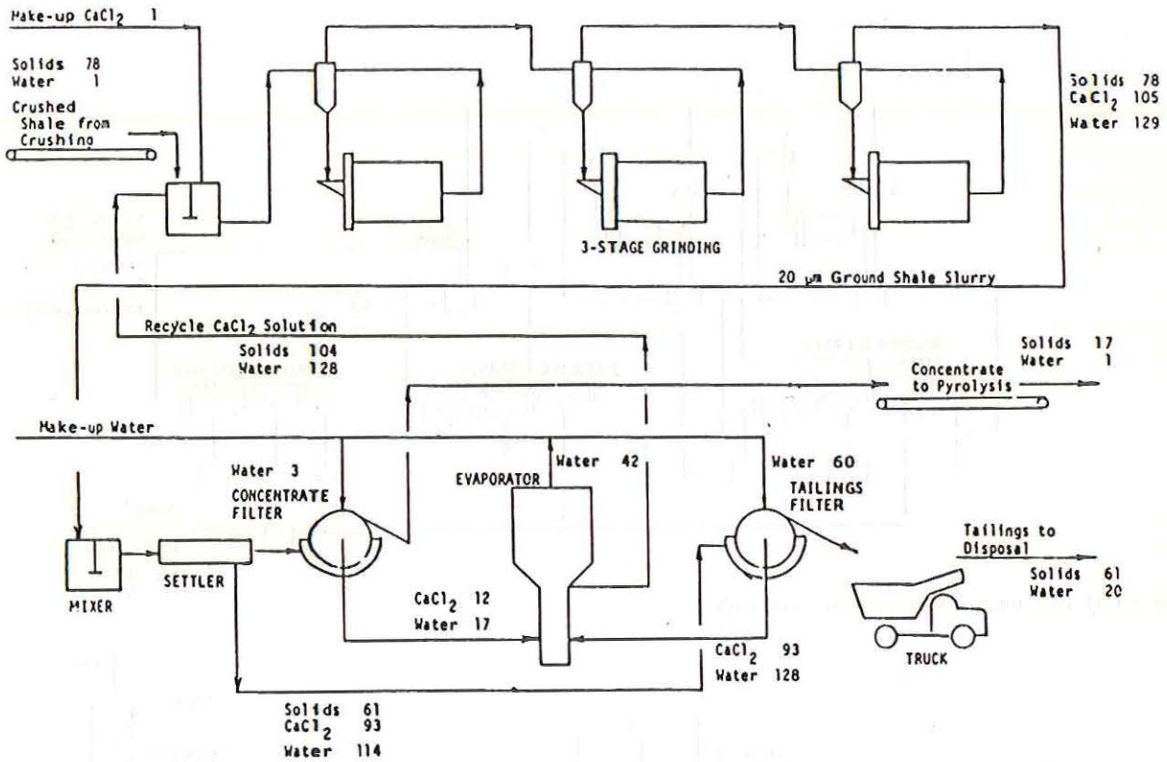


Fig. 8. Selective shear aggregation (flowrates in k tons/day; bleed and losses: $CaCl_2=1$, water$=1$).

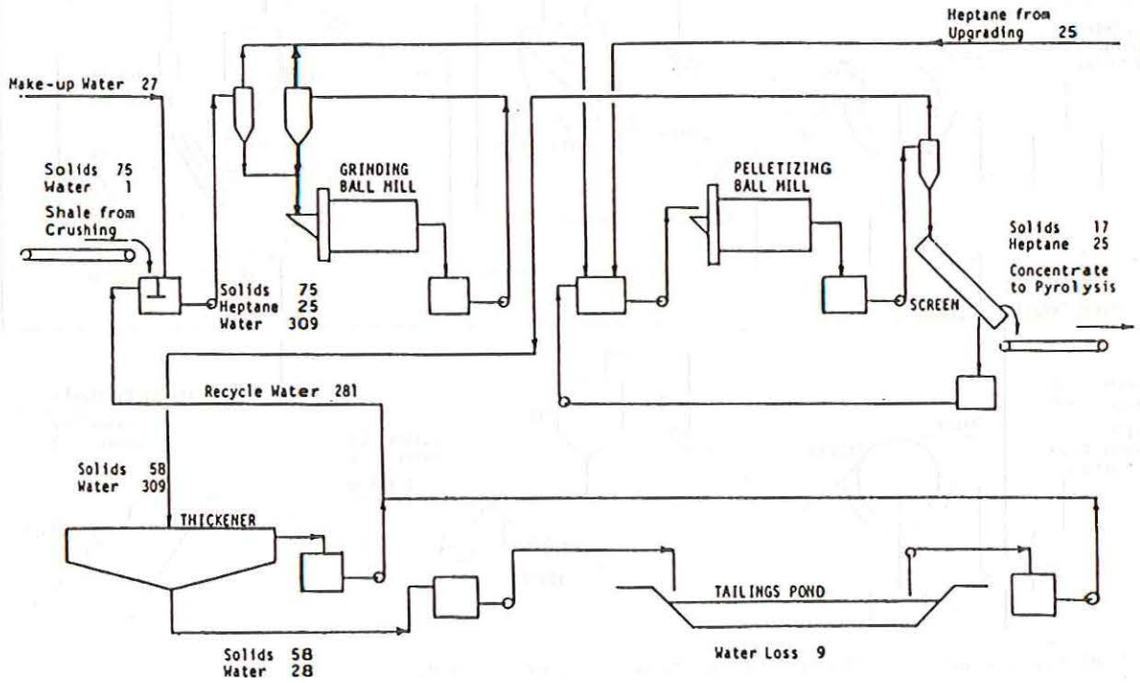


Fig. 9. Direct pelletization (flowrates in k tons/day).

TABLE 3

Selective shear aggregation process data

Recovery	85%
Feed particle size, 80% less than	20 microns
Grinding stages	3
Shale slurry pulp density	25 wt.%
Concentrate slurry pulp density	40 wt.%
Calcium chloride concentration	45 wt.%
Calcium chloride loss	1%
Mixer residence time	10 min
Settling rate	0.5 gal/sq.ft/min
Filtration rate	25 gal/sq. ft/hr
Wash water displacement	3 fold
Moisture content	
Concentrate	6 wt.%
Tailings	25 wt.%
Electric power	240 MW
Make-up water	12 *k* gal/min
Steam	1670 lbs/hr
No. of operators	55
Ball mills	
No. of trains per stage	4
No. of mills per train	3
No. of motors per mill	2
Mill size	24′dia×36′
Motor size	4.4 MW
Mixers	
Number	14
Size	17′diam×15′
Settlers	
Number	14
Size	108′×40′×3′6″
Filters	
Size	3000 sq ft
Number for concentrate	9+2 spare
Number for tailings	4+1 spare
Evaporators: Size	50,000 sq ft
No. of effects	4
No. of trains	4

cles as by crushing. It is a further objective to minimize heavily loaded relative motion between shale fragments and metal surfaces in an attempt to minimize wear of the latter. Water is fed upward at the bottom of the machine in such a way as to flush fragments from the lower bearing and seal area.

6. FROTH FLOTATION

Process design (Figs. 6,7; Tables 1,2) for this step is again based on general industrial practice of ore concentrating as well as a specific oil

TABLE 4

Direct pelletization process data

Recovery	88%
Feed particle size (80% less than)	100 mesh
Grinding stages	1
Shale slurry pulp density	18 wt.%
Heptane content in concentrate	60 wt.%
Moisture content	
Concentrate	less than 1%
Tailings	25 wt.%
Heptane loss	1.5%
Water loss (excluding tailings pond)	3%
Thickener settling rate	3 lbs/(hr) (sq ft)
Electric power	106 MW
Make-up water	4 *k* gal/min
No. of operators	35
Ball mills	
No. of trains per stage	4
No. of mills per train	3
Motors	
Grinding	
No. per mill	2
Size	4.4 MW
Pelletizing	
No. per mill	1
Size	2.9 MW
Screens	
Number	12
Size	10′×16′
Thickeners	
Number	4
Diameter	700′

shale study supported by lab experiments [3]. The finely ground oil shale in a water slurry is stirred with froth forming agents in a set of flotation cells. The kerogen-rich particles concentrate in the froth while the tailings particles depleted of kerogen stay in the bulk of liquid. The concentrate and tailings are subsequently separated as overflow and underflow, respectively. To increase the process efficiency, the process is staged. The concentrates from the first two stages are reground and refloated.

The shale–water slurry from the second grinding stage plus the froth forming agents are pumped to the rougher cells. The underflow goes to the middling cells and from there, in turn, to the tailings tank. The overflow from the rougher and middling cells is reground separately in ball mills and fed to the rougher and

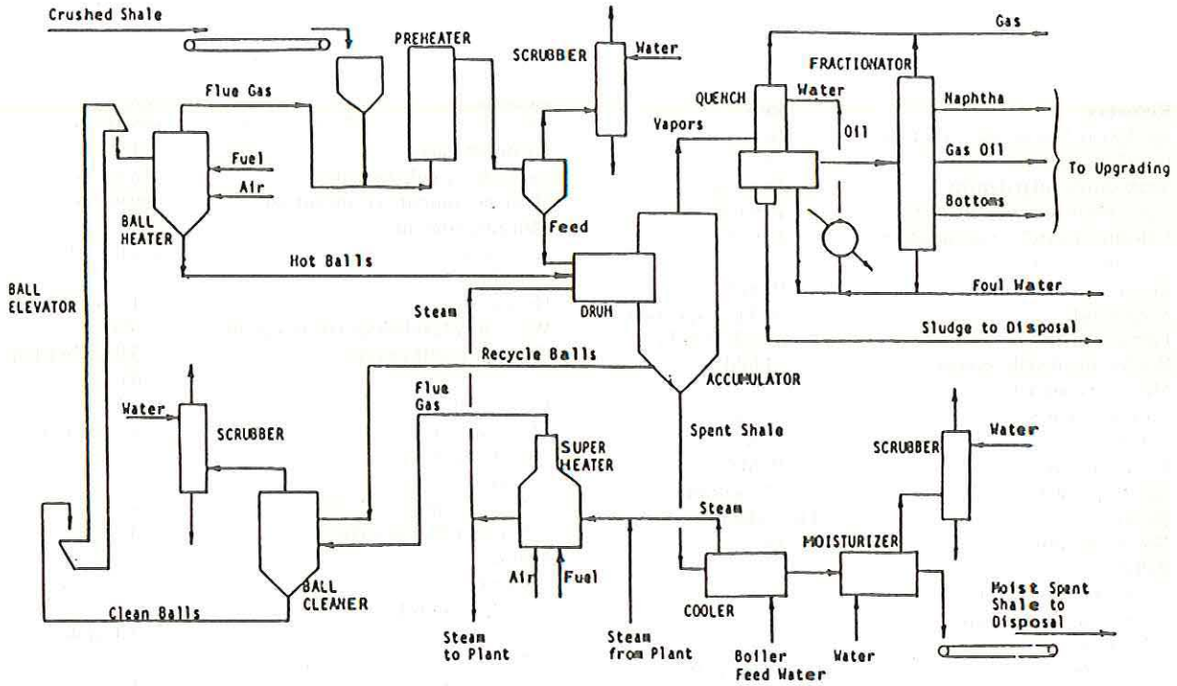


Fig. 10. Retort pyrolysis ("shale" denotes either crushed shale or shale concentrate, "upgrading" is part of oil processing).

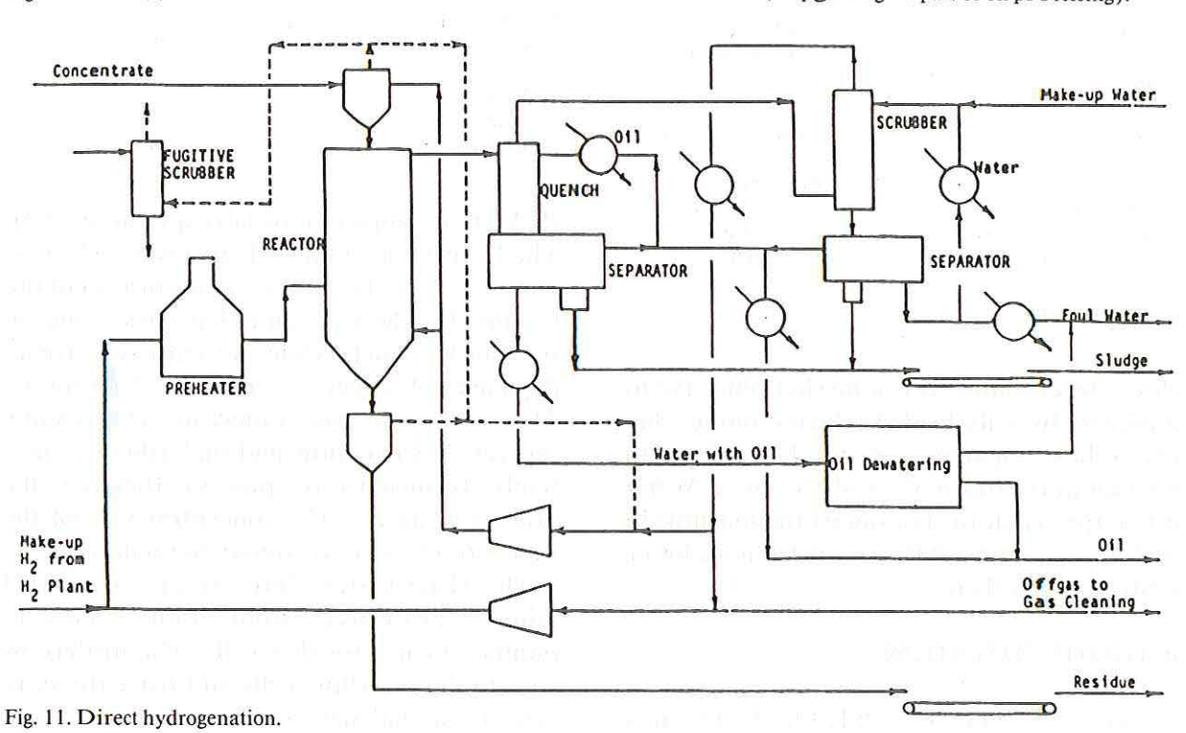


Fig. 11. Direct hydrogenation.

TABLE 5

Direct hydrogenation process data

Reactor	
Temperature (°F)	1290
Pressure (psig)	425
No. of process modules	65
Avg. flowrate (*k* tpd)	114
Material balance (*k* lbs/hr)	
Concentrate	1030
Preheated gas to reactor	940
Cool gas to reactor	2580
Residue	40
Reactor vapors	5510
Scrubber overhead gas	3810
Raw oil	790
Make-up hydrogen	80
Hydrogen plant feed gas	240
No. of operators	56
Utilities	
Fuel (MBtu/hr)	1720
Power (kW)	45
Water (*k* gal/min)	2.8
Generated steam (*k* lbs/hr)	486

middling cleaner cells, respectively, each comprising three stages. The underflow of the last rougher cleaner stage is refloated in the two-stage tailings cleaner cells with an interstage recycle. The overflow from the first stages of all cleaner cells is the final concentrate while the underflow from the last middling and tailing cleaner stage passes to the tailings tank.

The concentrate and tailings slurries are both first filtered. The tailings filter cake is trucked to the spent shale disposal area while the concentrate filter cake is pelletized, dried, and conveyed to Pyrolysis. The filtrates are recycled to grinding.

7. SELECTIVE SHEAR AGGREGATION

Process design (Fig. 8, Table 3) uses a method that was proposed by T.A. Ring based on limited lab tests [4]. The shale is mixed with a recycled concentrated calcium chloride solution, ground in a three-stage system of ball mills, and fed to a set of mixers. The shear forces of slow stirring agglomerate a kerogen-rich concentrate that is separated as overflow from the tailings in settlers. The dense calcium chloride acts as a heavy medium enhancing separation. Both concentrate and tailings are filtered and washed. The calcium chloride concentration in the combined filtrate is increased in a multiple-effect evaporator with forced circulation and vapor recompression. The tailings are transported to the waste disposal site while the concentrate is conveyed to pyrolysis.

Major problem areas are the confirmation of agitation speed and residence time in the mixers, as well as the determination of settling and filtration rates. The process avoids the complex system of froth flotation cells but pays a price in that the energy-intensive evaporation step is needed to maintain the calcium chloride concentration. It should also be noted that the total amount of shale has to be subjected to the third stage regrinding while only the rougher and middling cell froth are reground in the flotation alternative.

8. DIRECT PELLETIZATION

Process design (Fig. 9, Table 4) uses a method that was proposed by J. Reisberg [5] and T.A. Ring [6] based on several experimental studies. The shale is mixed with recycled water, ground in a one-stage system of conventional ball mills, and fed together with recycled heptane to a set of pelletizing ball mills. Heptane combines with kerogen, thus enhancing the extraction of a concentrate from the shale and its agglomeration to pellets that substantially exceed the particle size of the ground shale. The tailings slurry is separated in cyclones, thickened, and pumped to a waste disposal pond. A slurry of undersized particles is separated by means of a screen and recycled to the pelletizing mill. The full-size pellets are conveyed to Pyrolysis.

Major problem areas are tailings handling, heptane recovery, heptane contamination of aqueous process streams, and confirmation of

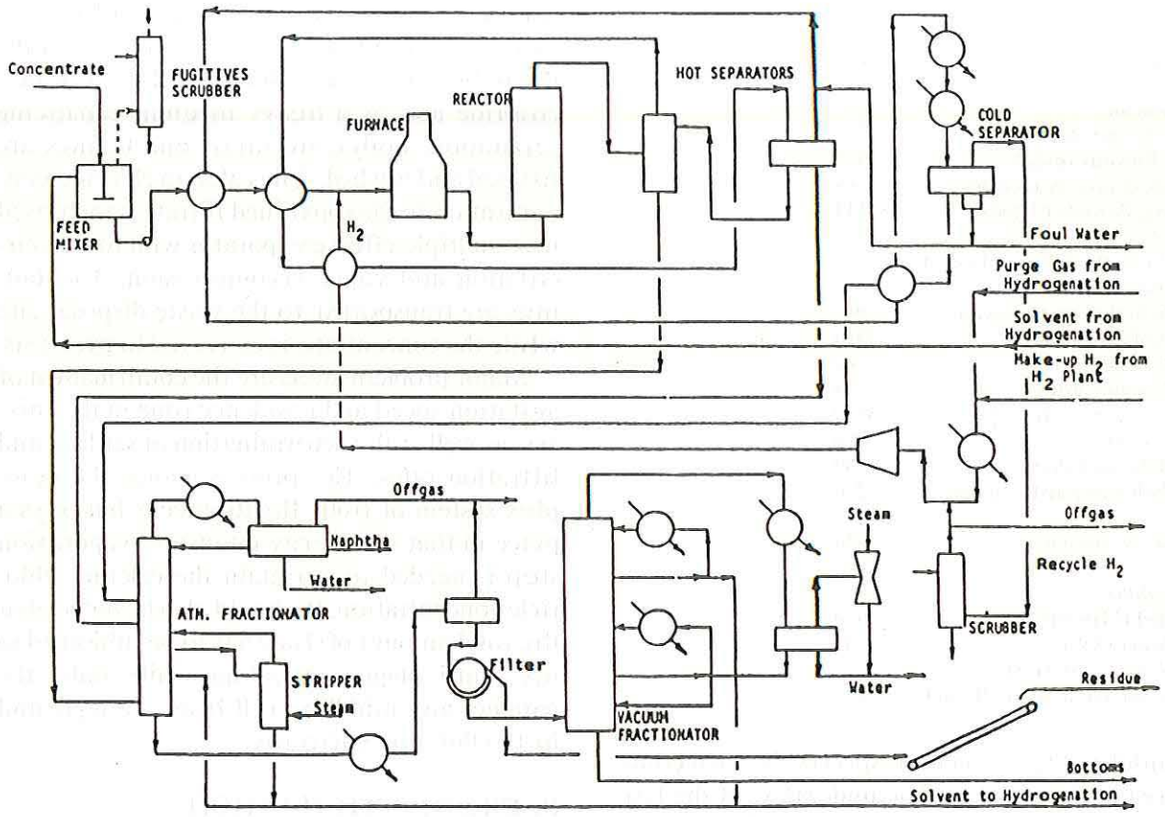


Fig. 12. Donor solvent hydrogenation – extraction section.

residence time and power requirements for the pelletization operation. The flowsheet assumes that heptane would become a part of the concentrate pellets and any traces in the aqueous streams would not violate environmental rules in the grinding, thickening, or tailings pond areas. Heptane could be removed from the recycle water if necessary. However, the clean-up of the tailings slurry might be extremely expensive.

9. RETORT PYROLYSIS

The process design (Fig. 10; Tables 1,2) is based cheifly on the environmental documentation of the Tosco Colony Project [7]. The plant has six or two parallel trains for the shale or concentrate, respectively. The raw material is preheated with flue gases from the ball heater and fed into the retort together with steam and hot ceramic balls that act as a heat transfer medium. The retort includes a rotating inclined drum in which the feed and balls are intimately mixed before they pass into the accumulator.

Overhead vapors are quenched with water and separated into gas, sponge oil, gas oil, bottoms oil, and foul water in a fractionator.

The spent shale or concentrate is separated from the balls in a rotating trommel screen at the bottom of the accumulator and discharged through a coller (waste heat boiler) to a moisturizer. The moist spent material is then taken by conveyor to the waste disposal area. The balls are recycled to the retort drum via a cleaner and heater. In the cleaner, dust is removed from the balls using flue gases from a steam superheater.

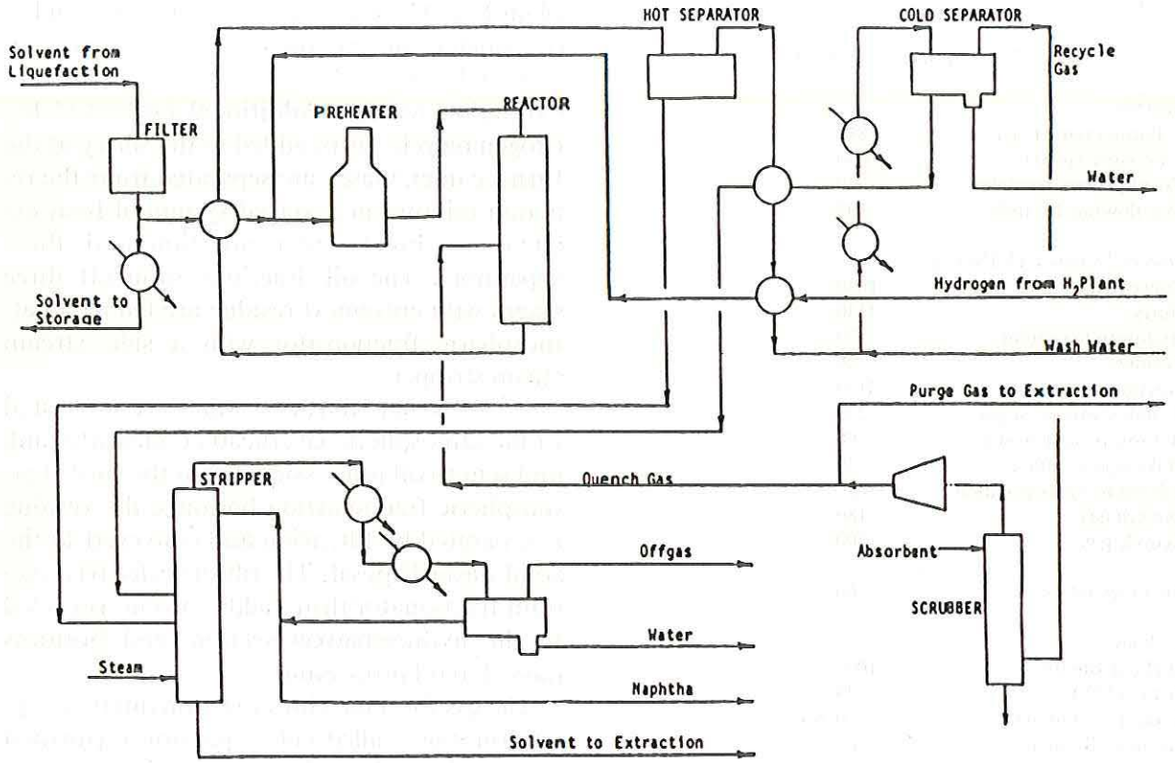


Fig. 13. Donor solvent hydrogenation – hydrogenation section.

The original concentrate particle size constitutes a major uncertainty and potential problem. While it can be assumed that the high kerogen content provides a strong initial bond for the concentrate pellets, the pellets are likely to disintegrate under the impact of the hot ceramic balls. The resulting entrainment of fines will be exacerabated by the higher vapor velocities per ton feed unless the accumulator diameter is substantially increased. Dust collection and sludge separation equipment downstream, with a recycle to the retort, might be required.

10. DIRECT HYDROGENATION

Among the published direct hydrogenation processes, Hytort was selected for process design (Fig. 11, Table 5) because pilot plant data were available [8]. The concentrate is fed to a multi-stage counter-current moving bed reactor to be contacted with recycle gas and hydrogen. Portions of the recycle are injected into the feed system, bottom of the reactor, and after being mixed with make-up hydrogen, to the lower part of the reactor. The solid residue is withdrawn from the bottom and disposed of. The overhead reactor vapors are first quenched with oil and then scrubbed with water. The quench and scrubber towers recycle the cooling liquid via water coolers. The oil–water emulsion that settles in the lower part of the quench separator is dewatered in another cleaning step. The oil phases from the quench separator, scrubber separator, and dewatering equipment are combined and pumped to oil processing.

The overhead gas from the scrubber is divided into three streams. One portion is directly recycled to the reactor while the second

TABLE 6

Donor solvent hydrogenation process data

Reactor	
Temperature (°F)	840
Pressure (psig)	2000
No. of process modules	79
Avg. flowrate (*k* tpd)	102
Material balance (*k* lbs/hr)	
Concentrate	1030
Slurry	4130
Hydrogen to reactor	170
Residue	40
Fractionator feed	4100
Scrubber overhead gas	200
Hydrogen plant feed gas	90
Make-up hydrogen	30
Solvent to hydrogenation	3080
Quench gas	180
Recycle gas	200
No. of operators	66
Utilities	
Fuel (M Btu/hr)	1090
Power (kW)	25
Water (*k* gal/min)	0.85
Steam (*k* lbs/hr)	67

is first mixed with make-up hydrogen and preheated. The third gas stream is cleaned outside the oil recovery system and combined with other oil processing offgases. A part of these is used to make hydrogen in a standard reforming plant.

The Hytort process has the disadvantages of fine particle entrainment and the need for heating the reactor with diluted hydrogen which requires the circulation of large amounts of gas at elevated pressures. The process is claimed to produce a higher grade oil than other retorting systems, but the value of the improvement is small and there is a price to be paid for the front-end hydrogenation.

11. INDIRECT HYDROGENATION USING A DONOR SOLVENT

Process design (Figs. 12,13; Table 6) adapts the Exxon Donor Solvent (EDS) process for liquefaction of coal that was tested in a pilot plant [9]. The concentrate is slurried with hydrogenated solvent, preheated in heat exchangers and a furnace, and fed to a co-current extraction reactor. Additional preheated hydrogen recycle gas is added to the slurry at the furnace inlet. Gases are separated from the reaction mixture in a staged system of heat exchangers, direct water injection and three separators. The oil fractions from all three stages with entrained residue are fed to an atmospheric fractionator with a side stream steam stripper.

Offgases, naphtha, and water are separated in the atmospheric overhead condensate tank and sent to oil processing. From the cooled atmospheric fractionation bottoms, the residue is separated by filtration and conveyed to the solid waste disposal. The filtrate is fed to a vacuum fractionator that yields solvent, recycled to the hydrogenation section, and bottoms passed to oil processing.

The gas from the third reaction mixture separation stage, called cold separator, is purified in a scrubber. It is then mixed with make-up hydrogen from a reforming H_2 plant, and a hydrogen-rich purge gas from hydrogenation. The resulting hydrogen recycle gas is pumped to the preheating furnace.

The solvent from the fractionator is preheated in a heat exchanger and furnace, and fed to the hydrogenation reactor. Preheated hydrogen from a reforming plant is added at the furnace inlet. Reactor temperature is controlled by injecting recycled cold quench gas. Gases are separated from the reaction mixture in a staged system of heat exchanges, direct water injection, and a hot and cold separator. The liquid fractions from both stages are fed to a stream stripper.

Offgases, naphta, and water are separated in an overhead condensate tank and sent to oil processing. The hydrogenated solvent from the bottom is returned to extraction. The gas from the cold separator is purified in a scrubber, and recycled as quench gas to the reactor after a portion is purged to extraction.

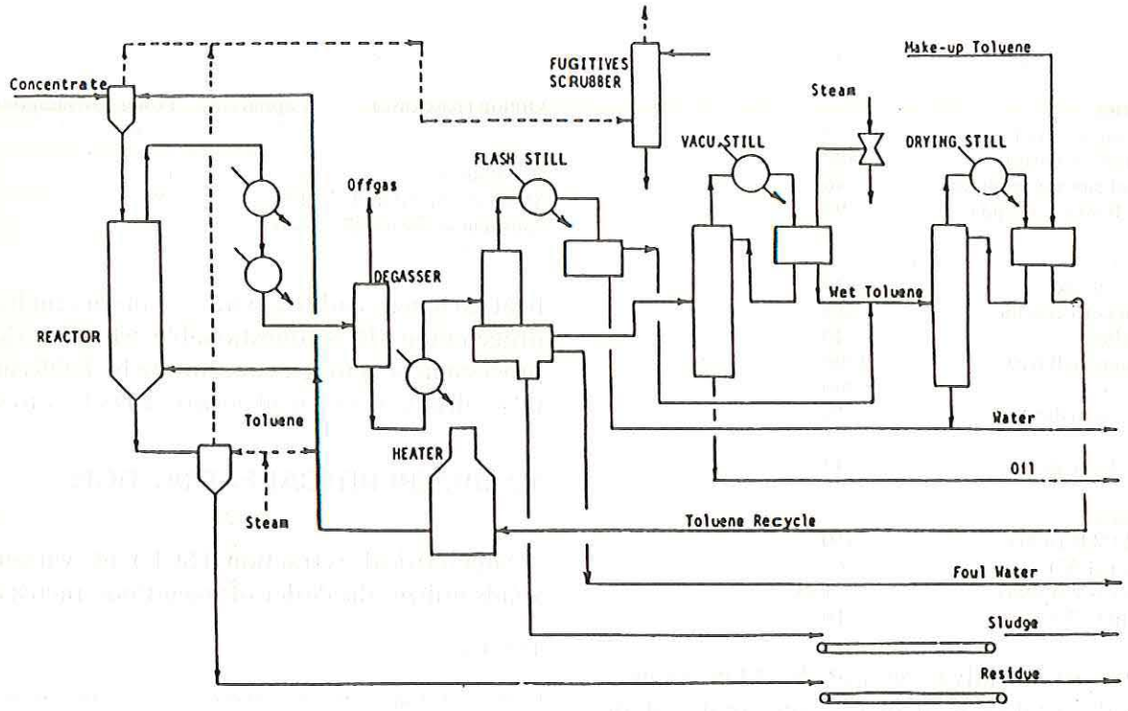


Fig. 14. Supercritical extraction.

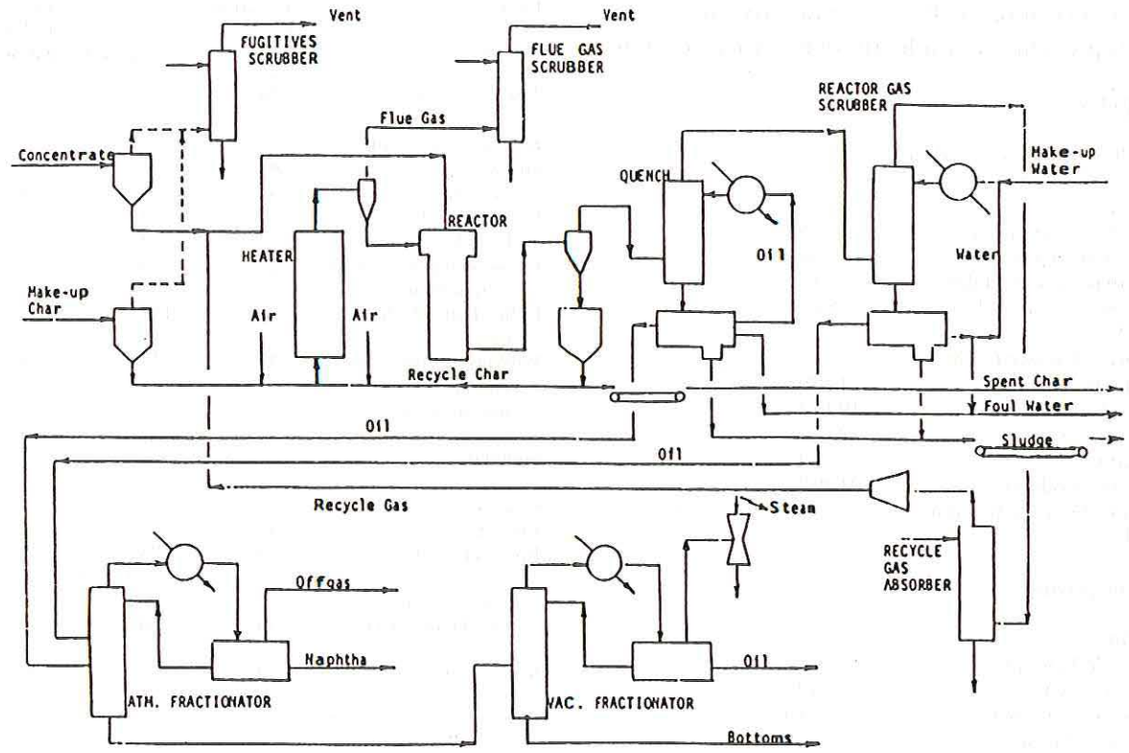


Fig. 15. Flash pyrolysis.

TABLE 7

Supercritical extraction process data

Reactor	
Temperature (°F)	600
Pressure (psig)	580
No. of process modules	46
Avg. flowrate (*k* tpd)	95
Material balance (*k* lbs/hr)	
Concentrate	1030
Toluene to reactor	880
Residue	40
Vacuum still feed	1790
Raw oil	990
Make-up toluene	10
No. of operators	42
Utilities	
Fuel (M B tu/hr)	410
Power (kW)	67
Water (*k* gal/min)	1.08
Steam (*k* lbs/hr)	10

The major advantages of the EDS systems are the replacement of the gas–solid system with a liquid–solid alternative and complete liquefraction of the feed. The disadvantages are high pressure, complexity caused by efficient heat exchange, and the extra step of solvent hydrogenation. It is questionable whether the processing of kerogen concentrate by EDS can take full advantage of all positive EDS features.

TABLE 8

Flash pyrolysis process data

Reactor	
Temperature (°F)	1,200
Pressure (psig)	30
No. of process modules	68
Avg. flowrate (*k* tpd)	200
Material balance (*k* lbs/hr)	
Concentrate	1,030
Recycle gas	10,000
Char to reactor	10,000
Residue	40
Reactor products	21,030
Atmospheric fractionator feed	770
No. of operators	40
Utilities	
Char (M B u/hr)	500
Power (kW)	110
Water (*k* gal/min)	1.12
Steam (*k* lbs/hr)	1.0

TABLE 9

Comminution data

Option (base case)	Capital cost (SM)	Power consumption (MW)
Ball milling	350	180
Stationary spiral mill	170	90
Autogenous shear mill	680	90

12. SUPERCRITICAL EXTRACTION

Supercritical extraction (SCE) of various solids utilizes the order-of-magnitude increase

TABLE 10

Capital cost of alternative separation processes (millions of mid-1981 dollars)

Item	Flotation (base case)	Selective shear aggregation	Direct pelletization
Mining section	320	340	320
Beneficiation section			
Grinding	160	240	80
Separation	35	23	75
Concentrate dewatering	10	4	5
Tailings dewatering	22	9	13
Drying and pelletizing or evaporation	26	50	
Indirects at 35% direct cost	89	114	61
Working capital, startup, etc., at 10% direct plus indirect	35	44	23
Subtotal	377	484	258
Contigency			
Project	57	73	39
Process (40%)*	113	193	103
Beneficiation total (rounded)	550	750	400
Mining plus beneficiation total	870	1090	720

*30% for base case.

TABLE 11

Annual cost of alternative separation process (millions of mid-1981 dollars per year)

Item	Unit cost or percent	Flotation (base case)	Selective shear aggregation	Direct pelletization
Mine operating cost		24	25	24
Beneficiation operating cost				
Fuel	$3/MBtu	5.9	–	–
Power	5 cents/kWh	70.8	94.5	41.7
Water	40 cents/ *k* gal	1.3	2.3	0.8
Steam	$4/*k* lb	–	52.6	–
Grinding Balls	—	14.8	17.2	11.4
Chemicals	—	3.7	32.8	9.2
Oper. Labor (OL)	$13/hr	2.1	1.5	0.9
Supervision and Services	40% OL	0.8	0.6	0.4
Overhead	40% OL+ML	7.4	9.6	5.2
Maint. labor (ML)	3% capital	16.5	22.5	12.0
Op. and maint. supplies	2% capital	11.0	15.0	8.0
Subtotal (rounded)		134	249	90
Totals for mining and benefication sections				
Annual operating costs		158	274	114
Annual capital charge at 25%		218	273	180
Grand total annual costs (rounded)		380	550	290

in dissolution power of some light organic liquids which are compressed and heated above the critical temperature. Process design (Fig. 14, Table 7) adapts the SCE process for coal developed by the British Coal Board and Catalytic, Inc. based on bench scale experiments [10].

Concentrate is fed to the reactor at elevated pressure via a system of alternating lock hoppers pressurized by a portion of the preheated toluene while the bulk of the solvent flows counter-currently through the reactor. The residue is separated at the reactor bottom, stripped with steam in outlet lock hoppers, and conveyed to the waste disposal area. The reaction mixture is cooled and its pressure released in two stages. First, offgases are separated in a degasser and sent to the oil processing. Second, a portion of the toluene is evaporated in a flash still, condensed to remove some of the water, and passed to the toluene drying column. Sludge and another portion of water accumulate at the bottom of the flash still and are periodically withdrawn. The organic phase from the flash still is separated into oil and water toluene in a vacuum still. The oil is further processed in oil processing while the combined toluene streams are dewatered in the drying still and pumped back to the reactor.

SCE combines the simplicity of retorting with the advantages of slurry feeding over gas–solid handling without using the high pressure of the EDS system. However, even if the reactor is designed as a counter-current slurry-fed bed of concentrate pellets, attrition and entrainment might still cause a solid-liquid separation problem downstream.

TABLE 12

Comparison alternative shale oil recovery processes

Item		Ore processing	Concentrate processing				
		Retort pyrolysis	Retort pyrolysis	Direct hydrogenation	Donor solvent hydrogenation	Super-critical extraction	Flash pyrolysis
Recovery conditions							
Recovery rate (%)		90	90	120	120	120	120
Temperature (°F)		1300	1300	1290	840	600	1200
Pressure (psig)		15	15	425	2000	580	30
Capital ($M)							
Mining		290	320	250	250	250	250
Beneficiation		--	550	420	420	420	420
Recovery		770	220	520	610	260	640
Total		1060	1090	1190	1280	930	1310
Annual cost ($M/yr)		21	24	18	18	18	18
Mining							
Beneficiation		--	134	105	105	105	105
Recovery							
Fuel	$3/MBtu	30.7	11.8	40.6	25.7	9.7	7.8*
Power	5¢/kWh	0.2	0.1	0.0	0.0	0.0	0.0
Water	40¢/kgal	0.6	0.1	0.5	0.2	0.2	0.2
Steam	$4/klb	3.8	−1.3	−15.3	2.1	0.3	0.0
Chemicals		1.0	1.0	9.7	3.4	5.7	1.0
Oper.labor (OL)	$13/man-hr	2.5	1.5	1.5	1.8	1.1	1.1
Supervision & services	40% OL	1.0	0.6	0.6	0.7	0.5	0.4
Overhead	40% OL+ML	7.2	2.4	4.8	5.6	2.5	5.6
Maint. labor (ML)	2% capital	15.4	4.4	10.4	12.2	5.2	12.8
Op. & maint. supplies	3% capital	23.1	6.6	15.6	18.3	7.8	19.2
Subtotal–recovery (rounded)		86	27	68	70	33	48
Capital charges (25% capital)		265	273	298	320	232	328
Total		370	460	490	510	390	500

*Char at $2/MBtu

13. FLASH PYROLYSIS

Process design (Fig. 15, Table 8) adapts a bench-scale process for coal [11]. Flash pyrolysis depends on thermal decomposition in an essentially inert atmosphere as conventional retorting does. While the geometries and flows are different, the application to cumminuted shale appears to be excellent.

Concentrate is fed to a mixing chamber on top of the reactor via a pneumatic transport system that uses recycle gas. The feed is brought instantaneously to reaction temperature through intimate contact with hot char, and the

mixture subsequently flows through the main body of a co-current entrained-bed reactor. Char and residue are separated in hot cyclones and recycled to an entrained-bed heater where the solids are heated by burning additional make-up char in air. The build-up of residue in the char circuit is controlled by purging a stream of spent char to the solid waste disposal. The bulk of preheater flue gases is separated in cyclones before the char is fed to the reactor mixing chamber.

The vapors from the reactor outlet cyclones are first quenched with oil and than scrubbed with water. The quench and scrubbing towers recycle the cooling liquid via water coolers. Oil and water are separated in tanks at the bottom of the quench and scrubbing towers. Both oil streams are fed to an atmospheric fractionator, the bottoms of which are passed to a vacuum fractionator while the overhead condensate separates into offgas and naphtha. The products of the vacuum fractionation go to oil processing. Gas from the scrubber is purified in an absorber and recycled to the concentrate pneumatic feed system.

The major fuel advantage of the flash pyrolisis process is that is can presumably handle fine particles and that heat is supplied to the reactor by a solid medium. Both the feed and the heat transfer medium are transported pneumatically rather than mechanically. The disadvantage is the large circulation volume required by the entrained system. Also, of the alternatives compared here, process data on flash pyrolisis may be most speculative.

Problem areas of flash pyrolysis are again particle carryover to the downstream parts of the process, and provision of heat by the combustion of carbon on the spent shale.

14. ECONOMIC EVALUATION AND CONCLUSIONS

Because of the wide variations in availability and accuracy of engineering and cost data on the alternatives outlined in Table 1, a special cost estimating method was devised for the economic evaluation of the selected options. The method and detailed estimates are published elsewhere [12]. From the results of the cost comparisons summarized in Tables 9 through 12, the following conclusions can be drawn.

14.1 Comminution alternatives

Although the autogenous shear-mill promises a significant reduction in power consumption from the base case, that reduction is more than offset by a doubled capital cost and a drastic change in design would be required to arouse interest (Table 9). The stationary spiral mill, on the other hand, cuts the base case costs in half and further examination is warranted to see if this advantage can be realized or perhaps increased.

14.2 Beneficiation alternatives

Compared to froth flotation, selective shear aggregation has significantly higher capital and annual costs while direct pelletization has 24% lower annual cost because of both lower (17%) capital cost and lower (41%) power consumption (Tables 10 and 11). Thus, of the beneficiation options considered, direct pelletization is speculative but the most interesting looking technology.

14.3 Recovery alternatives

For each alternative, capital and operating costs of the preceding steps in the system (mining and beneficiation) are lower by 25% because of the optimistic assumption that yield will be 33% higher than from Tosco II retorting (Table 12). Even so, only the system incorporating supercritical extraction has lower total capital or operating costs than Tosco II retorting of concentrate, and total annual costs are

roughly breakeven with the Base Case and no beneficiation.

15. SUMMARY

The sequence of ball milling, froth flotation, and retorting concentrate is not attractive for Western shales compared to conventional ore retorting. The high capital and energy costs result largely from the ball milling step which is very inefficient. Major improvements in comminution seem achievable through research. Such improvements, plus confirmation of other assumptions, could make high-enrichment beneficiation competitive with conventional processing. Among potential alternative processes for the recovery of oil from beneficiated shale, only supercritical extraction appears to be competitive with ore retorting, but experimental verification is necessary.

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