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SHALE OIL RECOVERY SYSTEMS INCORPORATING ORE BENEFICIATION

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by
M.A. Weiss, I.V. Klumpar, C.R. Peterson & T.A. Ring

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Synthetic Fuels Center, Energy Laboratory
Massachusetts Institute of Technology
Cambridge, MA 02139

Abstract

This study analyzed the recovery of oil from oil shale by use of proposed systems which incorporate beneficiation of the shale ore (that is, concentration of the kerogen) before the oil-recovery step. The objective was to identify systems which could be more attractive than conventional surface retorting of ore. No experimental work was carried out. The systems analyzed consisted of beneficiation methods which could increase kerogen concentrations by at least four-fold. Potentially attractive low-enrichment methods such as density separation were not examined. The technical alternatives considered were bounded by the secondary crusher as input and raw shale oil as output.

A sequence of ball milling, froth flotation, and retorting concentrate is not attractive for Western shales compared to conventional ore retorting; transporting the concentrate to another location for retorting reduces air emissions in the ore region but cost reduction is questionable. 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 and such improvements, plus confirmation of other assumptions, could make high-enrichment beneficiation competitive with conventional processing.

Table of Contents

Table of Contents (continued)

Table of Contents (continued)

Table of Contents (continued)

List of Figures

List of Figures (continued)

List of Tables

1. SUMMARY

The purpose of this study has been to assess broadly the potential attractiveness of new large-scale systems for recovering oil from oil shale, systems which incorporate a beneficiation step to increase kerogen concentration substantially in the feed to a retort or other step for converting kerogen to oil.

The enormous magnitude of oil shale deposits in the United States is well known. More oil is contained in the rich thick deposits of Colorado's Piceance Basin oil shales alone than in all the proved petroleum reserves in the Middle East. However, a deposit is not a reserve. A deposit becomes a reserve only when exploitation of that deposit becomes technically and economically feasible. Recovery of oil from shale is not currently economically feasible by normal commercial criteria even though it is technically feasible. Therefore, there has been no commercial production of shale oil in the United States for over a century. At the date of this writing, no commercial-size shale oil plant is under construction or definitely committed for construction in the U.S. with the single exception of Union Oil Company's 10,000 barrel/day module in Colorado.

Oil Shale: A Lean Ore

The central economic problem in recovering oil from shale is the fact that even rich shales are lean ores. That is, only about 10-15% of the mass of rich shales is recoverable as marketable energy. The remaining 85-90% is worse than worthless; a cost is incurred in disposing of the residue acceptably. By contrast with other fossil fuel "ores," essentially 100% of petroleum (ex water) and natural gas, and perhaps 70-90% of most coals consist of energy, i.e., burnable material. Therefore, large amounts of shale rock must be mined, handled, processed, and disposed of in order to recover a relatively small amount of shale oil by traditional methods of surface retorting; that is expensive.

In some locations, a second key problem exists: heavy burdens imposed by the particular location. Colorado's Piceance Basin illustrates the problem. The terrain is difficult, making construction expensive. Water supplies are limited; their use for energy purposes provokes serious social and institutional debate even if the conventional economics are clear. Population is sparse; the infrastructure does not exist to provide and support the people needed to build and operate an industry. Environmental strictures may limit the size of the industry supportable, or at least require more extensive (and expensive) emission controls.

One technical approach to the lean-ore problem is in situ recovery. By leaving all or most of the rock in the ground and processing it there, materials handling problems are reduced. Several methods of in situ recovery have been proposed and researched. Although technical feasibility has been demonstrated--oil can be produced--economic feasibility has not and the future for in situ recovery is not clear.

Beneficiation of Lean Ores

A different technical approach was examined in this study: beneficiation. Beneficiation is the process for converting a lean ore into a rich feed, thus reducing some of the downstream materials processing and handling required. (Although, of course, adding the beneficiation step itself.) Beneficiation also offers the possibility of transporting the enriched feed at reasonable cost to another location for further processing--thus avoiding some of the problems of constructing and operating the recovery step (i.e. converting the kerogen to oil) at the site of the ore body.

Although beneficiation of oil shale has not even been thoroughly researched, beneficiation of metal-containing ores is standard commercial practice in the United States. Beneficiation of another energy ore, coal, is practiced widely on a commercial scale for the removal of inert rock and some pyrite. More sophisticated methods for coal beneficiation can be expected in the future.

The power of beneficiation can be illustrated by the fact that it makes possible the commercial exploitation of ores that are even leaner (in an economic sense) than oil shale is. For example, copper ores containing less than 0.5% copper are processed routinely, yielding less than $8 worth of refined metal per ton of ore. Analogous numbers for molybdenum are about $15/ton ore. By contrast, 30 gallon/ton oil shale yields oil worth $25/ton ore assuming that the oil is valued at $35/barrel. Therefore, the notion that oil shale can be beneficiated commercially cannot be rejected out of hand as having no parallel in other experience.

Steps in the Recovery System

We picture a system incorporating beneficiation for oil shale recovery to have the following major steps conceptually (although not necessarily literally since some steps may be combined); this process sequence is shown graphically in Figure 2-1, p. 2-5.

1. Comminution of crushed shale (from surface or underground mining) to particles of "liberation size", i.e. fine enough so that individual particles are primarily kerogen or primarily mineral.
2. Separation of the mixture of particles into two parts, a kerogen-rich concentrate and a mineral-rich tailings with disposal of the tailings; this step and the preceding one will occur at or adjacent to the mine site.
3. Optionally, transport of the concentrate (for example, by slurry pipeline) to another location for further processing; the other location may offer more favorable construction, socioeconomic, or environmental conditions, or may have existing facilities (e.g., reactor capacity) which can be used.
4. Recovery of crude shale oil (and gas) from the concentrate by retorting or other methods and disposal of the spent mineral matter, perhaps after recovering energy from the residual carbon by combustion or gasification.

Beneficiation's Advantages and Disadvantages

4Compared to conventional shale oil systems consisting of surface retorting of whole ore, systems incorporating beneficiation have potential advantages and disadvantages. The most important advantages are as follows:

- Enabling kerogen and shale oil recovery from lean shales as well as rich ones; surface mining would be more attractive if lean shales in the overburden could be processed.
- Reducing the mass of material processed (and thus the size of the retort or other recovery equipment) in the recovery step to yield a given amount of oil.
- Reducing the mass of spent shale (the rock subjected to pyrolysis) generated and the environmental problems associated with its disposition.

In addition, a beneficiation process of the type we considered--one that results in very high enrichment (say, four-fold) of finely divided ore--has the potential for:

- Transporting the enriched material in a slurry pipeline out of the oil shale region for remote conversion of kerogen to oil, as noted above.
- Using technologies other than, and perhaps superior to, retorting for recovery of oil from the kerogen.

The most important corresponding disadvantages are as follows:

- Cost of the beneficiation process (capital, energy, water, ore loss, other operating costs).
- Large tailings stream for disposal.

-- Possible problems in handling and separating fine particles during oil recovery.

The essential trade-off is obvious: can the cost ("cost" in all senses) of adding the high-enrichment beneficiation process be offset by the benefits (again, in all senses) of easier kerogen conversion? Answering that question was the major objective of this study.

We did not examine low-enrichment (say 30% to 50%) beneficiation, achievable by density separations for example. Such separations are under commercial investigation and are potentially attractive since they are relatively inexpensive and can capture retort credits directly.

Base Case Comparisons

In order to assess the potential benefits of incorporating ore beneficiation into a shale oil recovery system, the most direct approach simply compares systems with and without beneficiation. Such a comparison was made for "Base Case" systems, that is, systems based on technology that seems likely to perform reasonably as assumed even though the technology has not been demonstrated on a commercial scale or even on a pilot plant scale in some respects. The Base Case systems were selected and defined for mining through recovery of raw shale oil; the mining step is included because increased mining costs are incurred if kerogen recovery in the beneficiation step is less than 100%, although that increase may be offset if oil yield from kerogen concentrate in the recovery step exceeds the yield from conventional retorting. Upgrading and refining were not included because the impact of beneficiation on those steps should be minor.

The choice of conventional Base Case, without beneficiation, was the technology proposed for the Colony project, i.e. Tosco II retorting. Colony was chosen because more information has been published on the Tosco II technology than on any other technology and because Tosco II retorts, unlike most or all other developed U.S. retorts, seem able to handle a kerogen concentrate (although some modifications are required).

The Base Case including beneficiation assumed that the output of the Colony secondary crusher was ground to $<$ 20 microns in a multi-stage recycling ball mill circuit and froth flotated to separate kerogen particles from mineral particles. Oil was recovered from the dried and pelletized kerogen concentrate in Tosco II retorts.

For both Base Cases detailed flow sheets were prepared and estimates made of design criteria (Table 3-2, p. 3-7), equipment specifications (Table 3-3, p. 3-8), and materials, utilities, and labor requirements (Table 3-4, p. 3-9). Both Base Cases used the same original ore, 35 gallons/ton shale, and produced the same 50,000 barrels/stream day of raw shale oil. The beneficiation system assumed that the kerogen concentrate contained 88% of the kerogen in the ore and was enriched four-fold, i.e. to 140 gallons/ton.

Using the information described in the previous paragraph, and published estimates on the cost of the Colony project, capital and operating costs were calculated for the two Base Cases. The history of cost estimates of synthetic fuels plants does not inspire high confidence in the state of the art for making such estimates. The more speculative the technology, the greater the probable error, and some of the technologies considered in this study are extremely speculative. It is not facetious to say that, in many cases, total costs of proposed synthetic fuels plants have not been known confidently to <u>one</u> significant figure, e.g. Colony, although numbers are often tabulated with two to three significant figures (as here) or even more. Therefore, the absolute levels of costs cited in this study should be regarded with at least the same skepticism deserved by other estimates. However, we tried to estimate costs in a consistent way so that relative comparisons should have reasonable validity even if the absolute numbers are crude.

The table below summarizes the results where capital costs are expressed in 1981 dollars for an instantly built plant, and total annual costs are the sums of annual operating costs plus a 25% annual capital charge (details in Tables 3-7, 3-8, and 3-9, pp. 3-25, 3-26, and 3-30). The costs include only the mining, beneficiation, and pyrolysis sections of the plant. All numbers exclude other plant sections such as upgrading, land, and offsites (which make up 64% of the total capital cost in the conventional Base Case) because they are relatively unaffected by incorporating beneficiation.

Table 1-1

Base Case Costs for Mining Through Pyrolysis Sections

(1981 Dollars, Millions, for Instant Plants)

	Conventional	With Beneficiation
Capital Costs		
Mining	290	320
Beneficiation	-	550
Pyrolysis	770	220
Grand Total	1060*	1090
Annual Costs		
Operating	107	186
Capital Charge @ 25%	265	273
Grand Total (Rounded)	370	460

*Total plant capital cost including upgrading, offsites, and land is about $2910 million.

The capital costs are the same, well within the accuracy of the estimates, but the operating costs for beneficiation are $90 million/year higher primarily due to power costs for grinding of $70 million/year.

Since the assumptions for the less-proved beneficiation case are probably optimistic relative to the conventional case, our conclusion is that the beneficiation technology assumed for the Base Case is not an attractive prospect for a development program. Therefore we examined some alternative technologies to attempt to identify more promising prospects.

Comminution, Separation, and Recovery Alternatives

Improvement of comminution technology is hampered by ignorance about the minimum amount of energy required to reduce shale to liberation sizes (averaging about 5 microns). Ball mills and other commercial grinding equipment probably consume 100 to 1000 times as much energy as is theoretically necessary. Three new equipment designs were proposed: a stationary spiral ball mill (SSM), an autogenous shear mill (ASM), and a pneumatic impact mill (PIM) (Figures 4-5 to 4-8, pp. 4-29, 4-30, and 4-33). Both the SSM and the ASM might use about half the energy consumed in ball milling. The SSM might also cut the capital cost in half, but the ASM's capital costs would be much higher (Table 4-1, p. 4-43). The PIM was too speculative for cost and energy estimation.

Separation of comminuted particles is measured by the separation efficiency, a percentage defined as R_1 minus R_2, where R_1 and R_2 are the percentages of kerogen and mineral respectively recovered in the kerogen concentrate. High separation efficiencies are required to get a concentrate with both high kerogen recoveries from ore and high enrichment ratios. For example, recovering 90% of the kerogen from a 35 gallon/ton ore with an enrichment ratio of 4 corresponds to a separation efficiency of 84%. Of the many separation techniques considered, only the techniques which depend on wettability (like froth flotation) seem both practical and capable of high separation efficiencies. Two identified wettability alternatives to froth flotation are selective shear aggregation (SSA) and direct pelletization (DP). The cost estimates for these alternatives (Table 5-5, p. 5-33) show that DP might be cheaper than froth flotation, reducing capital and operating costs by one-quarter to one-third. Therefore, although DP is highly speculative it deserves more scrutiny if beneficiation systems are to be studied.

Alternatives to retorting for recovering oil from kerogen concentrate can be justified if the alternative is cheaper, gives improved yields or product quality, or is superior on environmental or other grounds, other considerations equal. A closer examination of existing data showed that alternative processes are generally more complex and thus more costly than retorting, that only modest yield improvements over optimized retorting are possible even theoretically, and that quality improvements are not likely to materially affect the total cost of refining to marketable transportation fuels since deep hydrogenation or other severe processing is needed in any case. One possibly attractive alternative to retorting is supercritical extraction (SCE), a process being developed for coal and other materials. Some speculative cost estimates on SCE (Table 6-5, p. 6-25) show system capital and operating costs similar to those of the Base Case systems--largely because of the assumption of improved yield (120% of Fischer Assay) which reduces the throughputs and costs of the mining and beneficiation sections.

Ultimate Process Possibilities

Since the obvious ball-milling-froth flotation Base Case sequence looked unattractive under the conditions assumed, we considered other conditions and other combinations of beneficiation technology to see if process sequences could be identified that were competitive with conventional ore retorting (details in Table 7-1, p. 7-4).

The same Base Case technology for Western shales was about cost-equivalent to conventional retorting if grinding costs were halved. The same cost-equivalence resulted for Eastern shales but at an absolute cost level about three times as high as for Western shales.

Several additional and more extreme assumptions were made in various combinations, namely: enrichment ratios were increased close to the theoretical limits; kerogen recovery during separation was increased to 97%; conversion of kerogen to oil was increased to 120% of Fischer Assay; further 50% cuts in grinding costs were made; and, finally, the recovery section was eliminated thus feeding the kerogen directly to a cracking/fractionation column. Obviously each of these technical targets would be difficult to reach, and reaching them all would be virtually impossible.

The various combinations resulted in total annual costs up to 50 or more percent lower than the conventional Base Case for both Eastern and Western shales (with the former remaining about three times as costly as the latter). However, because the assumptions made were so extreme, that potential cost reduction should be taken to mean only that there is no evident theoretical barrier to a competitive beneficiation process; the issue, rather, is how close we can come to the theoretical limits in each step. In addition, the total costs of a shale-derived transportation fuel must also include costs arising from other plant sections which will usually show little or no reduction from introducing beneficiation.

Remote Recovery

As noted previously, kerogen concentrate may be transported away from the ore site for remote conversion of kerogen to oil, e.g. by retorting. This option arises from the fact that the kerogen concentrate is a reasonably rich energy source, similar to some coals which can be and are economically transported, rather than a lean rock which cannot be moved any significant distance at reasonable cost. In assessing the costs and benefits of remote recovery, it is important to remember that beneficiation per se enables remote location of only the conversion section; remote upgrading and refining of raw shale oil can be done in a conventional system. Our analysis was confined to Western locations because, on balance, plants processing Eastern shales are likely to have net incentives to remain at or adjacent to the ore sites. Unlike the Western situation, the socioeconomic, environmental, and construction problems should be no more severe at Eastern ore sites than at any reasonable remote location.

Our estimate of slurry pipeline costs is equivalent to about 1 cent per mile per contained barrel of oil for distances greater than 100 miles or so. Other ancillary costs can be expected. The direct benefits due to relocating the conversion step appear to be small (considering construction costs and regional labor and socioeconomic costs). The major effect is reduction of perhaps 90+% of all regional air pollutants emitted except particulates. Such pollutants arise primarily from high temperature gas reactions associated with retorting (and downstream steps) and those plant sections are elsewhere.

On balance, the option to recover oil from kerogen remotely may be helpful in some circumstances, but it is not likely to provide a major incentive to switch to a beneficiation-based system.

Conclusions

Recovering oil from Western shales by surface retorting of ore is an expensive but essentially straightforward technology. Alternatives to retorting are generally more complex and more expensive without sufficiently compensating advantages. The currently feasible technology options for high-enrichment beneficiation seem to be no exception to that generalization even though low-enrichment separations may be attractive supplements to conventional retorting systems. Although beneficiation is relatively more advantageous for Eastern shales (because retorting costs start higher, and savings in retorts must pay for beneficiation), the total oil costs are still much higher than those for Western shales.

High-enrichment beneficiation systems can have some benefits which are difficult to assess in traditional engineering-economic terms, namely, enabling exploitation of lean shales as well as rich ones (and perhaps encouraging surface mining and labor saving) and major reduction of air emissions (ex particulates) in the ore region through remote location of the kerogen conversion plant. However, these benefits are not likely to be decisive for beneficiation without improvement in the traditional economics relative to conventional ore retorting.

Such improvement is certainly possible in principle. We cannot justify a development program now on high-enrichment beneficiation but we do recommend a modest program of basic and applied research. That research ought to be aimed initially at the fundamentals of shale comminution, characterization, and separation, and at exploring the feasibility of supercritical extraction for oil recovery.

2. INTRODUCTION

The purpose of this study is to assess broadly the potential attractiveness of new large-scale systems for recovering oil from oil shale, systems which incorporate a beneficiation step to increase kerogen concentration substantially in the feed to a retort or other step for converting kerogen to oil.

The enormous magnitude of oil shale deposits in the United States is well known. More oil is contained in the rich thick deposits of Colorado's Piceance Basin oil shales alone than in all the proved petroleum reserves in the Middle East. However, a deposit is not a reserve. A deposit becomes a reserve only when exploitation of that deposit becomes technically and economically feasible.

Recovery of oil from shale is not currently economically feasible by normal commercial criteria even though it is technically feasible. Therefore, there has been no commercial production of shale oil in the United States for over a century. (Shale oil was a commercial product in the U.S. before the first successful petroleum well was drilled in 1859.) As of the date of this writing, no U.S. commercial-size shale oil plant is under construction or definitely committed for construction with the single exception of Union Oil Company's 10,000 barrel/day module in Colorado. And that module will receive assistance from the U.S. Synthetic Fuels Corporation in the form of price guarantees for its product with an ultimate maximum government liability of $400M; current estimates of the total construction cost of the module are about $570M.

The central economic problem in recovering oil from shale is the fact that even rich shales are lean ores. That is, only about 10-15% of the mass of rich shales is recoverable as marketable energy. The remaining 85-90% is worse than worthless; a cost is incurred in disposing of the residue acceptably. By contrast with other fossil fuel "ores," essentially 100% of petroleum (ex water) and natural gas, and perhaps 70-90% of most coals consist of energy, i.e., burnable material. Therefore, large amounts of shale rock must be mined, handled, processed, and disposed of in order to recover a relatively small amount of shale oil by traditional methods of surface retorting; that is expensive.

In some locations, a second key problem exists: heavy burdens imposed by the particular location. Colorado's Piceance Basin illustrates the problem. The terrain is difficult, making construction expensive. Water supplies are limited; their use for energy purposes provokes serious social and institutional debate even if the conventional economics are clear. Population is sparse; the infrastructure does not exist to provide and support the people needed to build and operate an industry. Environmental strictures may limit the size of the industry supportable, or at least require more extensive (and expensive) emission controls.

One technical approach to the lean-ore problem is in situ recovery. By leaving all or most of the rock in the ground and processing it there, materials handling problems are reduced. Several methods of in situ recovery have been proposed and researched, most conspicuously by Occidental and its subsidiaries. Although technical feasibility has been demonstrated--oil can be produced--economic feasibility has not and the future for in situ recovery is not clear.

A different technical approach is proposed in this study: beneficiation. Benefication is the process for converting a lean ore into a rich feed, thus reducing some of the downstream materials processing and handling required. (Although, of course, adding the beneficiation step itself.) Beneficiation also offers the possibility of transporting the enriched feed at reasonable cost to another location for further processing--thus avoiding some of the problems of construction and operating the total recovery system at the site of the ore body.

2.1 The Logic of Beneficiation

Although beneficiation of oil shale has not even been thoroughly researched, beneficiation of metal-containing ores is standard commercial practice in the United States. Beneficiation of another energy ore, coal, is practiced widely on a commercial scale for the removal of inert rock and some pyrite. More sophisticated methods for coal beneficiation can be expected in the future.

The power of beneficiation can be illustrated by the fact that it makes possible the commercial exploitation of ores that are even leaner (in an economic sense) than oil shale is. For example, copper ores containing less than 0.5% copper are processed routinely, yielding less than $8 worth of refined metal per ton of ore. Analogous numbers for molybdenum are about $15/ton ore. By contrast, 30 gallon/ton oil shale yields oil worth $25/ton ore assuming that the oil is valued at $35/barrel. Therefore, the notion that oil shale can be beneficiated commercially cannot be rejected out of hand as having no parallel in other experience.

We picture a system incorporating beneficiation for oil shale recovery to have the following major steps conceptually (although not necessarily literally since some steps may be combined); they are also shown in Figure 2-1:

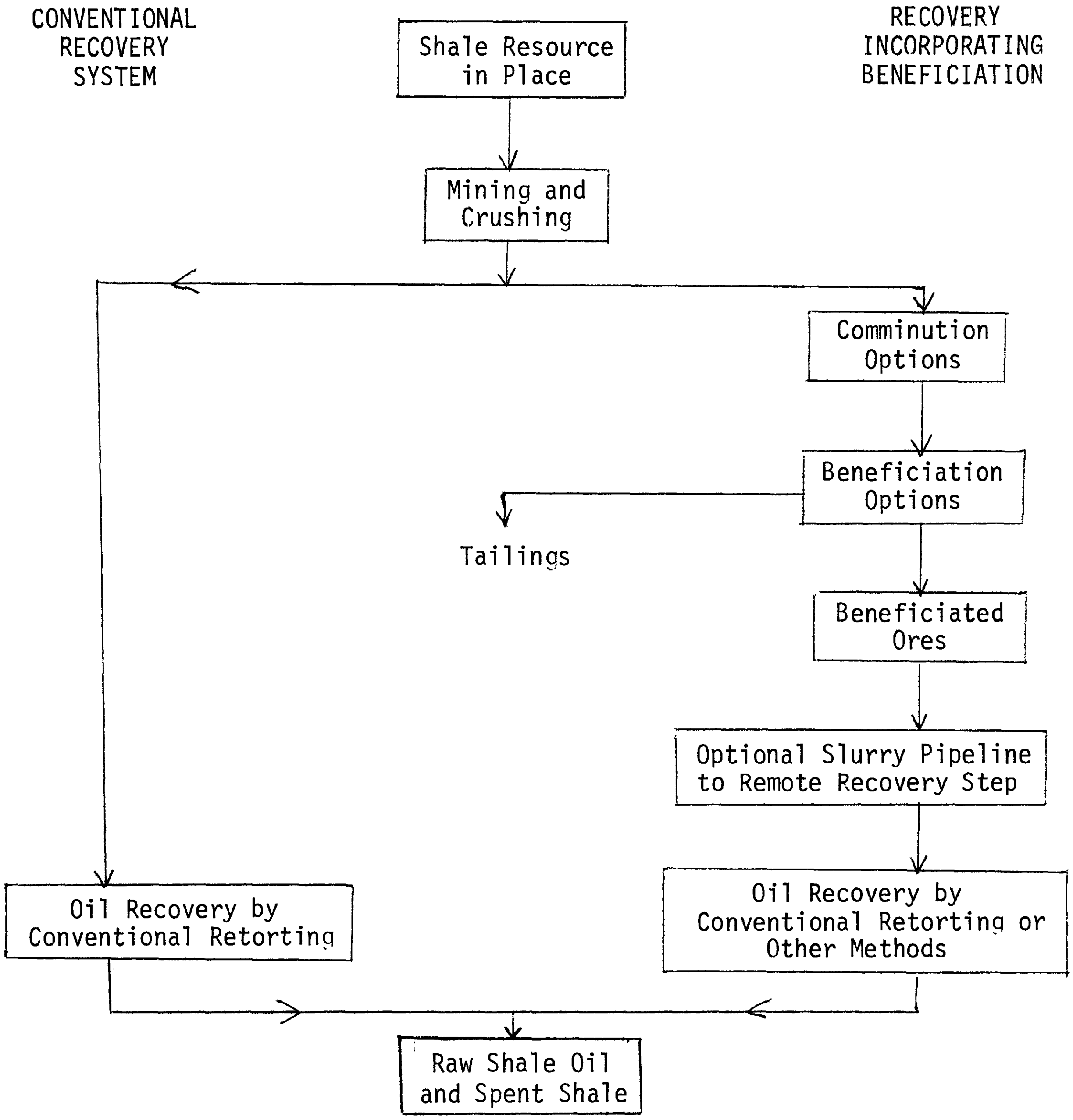
1. Comminution of crushed shale (from surface or underground mining) to particles of "liberation size", i.e. fine enough so that individual particles are primarily kerogen or primarily mineral.
2. Separation of the mixture of particles into two parts, a kerogen-rich concentrate and a mineral-rich tailings with disposal of the tailings; this step and the preceding one will occur at or adjacent to the mine site.
3. Optionally, transport of the concentrate (for example, by slurry pipeline) to another location for further processing; the other location may offer more favorable construction, socioeconomic, or environmental conditions, or may have existing facilities (e.g., reactor capacity) which can be used.
4. Recovery of crude shale oil (and gas) from the concentrate by retorting or other methods and disposal of the spent mineral matter, perhaps after recovering energy from the residual carbon by combustion or gasification.

Compared to conventional shale oil systems consisting of surface retorting of whole ore, systems incorporating beneficiation have potential advantages and disadvantages. The most important advantages are as follows:

-- Enabling kerogen and shale oil recovery from lean shales as well as rich ones; surface mining would be more attractive if lean shales in the overburden could be processed.

Figure 2-1

PROCESS BLOCKS FOR CONVENTIONAL AND BENEFICIATED-BASED SHALE OIL RECOVERY



-- Reducing the mass of material processed (and thus the size of the retort or other recovery equipment) in the recovery step to yield a given amount of oil.

-- Reducing the mass of spent shale (the rock subjected to pyrolysis) generated and the environmental problems associated with its disposition.

In addition, a beneficiation process of the type we are considering--one that results in very high enrichment of finely divided ore--has the potential for:

-- Transporting the enriched material in a slurry pipeline out of the oil shale region for remote processing, as noted above.

-- Using technologies other than, and perhaps superior to, retorting for recovery of oil from the kerogen.

The most important corresponding disadvantages are as follows:

-- Cost of the beneficiation process (capital, energy, water, ore loss, other operating costs).

-- Large tailings stream for disposal.

-- Possible problems in handling and separating fine particles during oil recovery.

The essential trade-off is obvious: can the cost ("cost" in all senses) of adding the high-enrichment beneficiation process be offset by the benefits (again, in all senses) of easier oil recovery? Answering that question is the major objective of this study.

We did not examine low-enrichment (say 30% to 50%) beneficiation, achievable by density separations for example. Such separations are under commercial investigation and are potentially attractive since they are relatively inexpensive and can capture retort credits directly.

2.2 Current Status

There is a modest degree of current interest in oil shale beneficiation methods; two recent reviews (Refs. 1,2) describe much of the activity. Most of the published private-sector work is devoted to processes which use differences in density to separate particles of crushed ore which are substantially larger than the particle liberation size. In effect such methods tend to separate naturally rich ore strata from naturally lean strata rather than separating (in principle) discrete mineral particles which have been detached from discrete kerogen particles. (Without grinding to liberation sizes, it is impossible to achieve high enrichment ratios and high recoveries simultaneously by density methods alone.)

Density forms an obvious basis for separation since the density of the kerogen averages about 1.07 and the density of the minerals averages about 2.7 in Western shale. The mixtures which constitute the natural ores thus cover a range in density as illustrated below:

Gal/Ton Assay	Average Density
15	2.42
30	2.15
45	1.94
60	1.80

In a recent publication, Larson (Ref. 3) of Gulf used heavy media of different densities to make sink-float separations of shale samples; however there is no report about the research proceeding to development or further stages. An important observation, which illustrates the limitations of density separation for simultaneous high enrichment and high recovery, was that there was no difference in results with ore ground to 45 microns (still well above the liberation size) compared to crushed ore in the 1/4- to 3-inch range.

Roberts and Schaefer Resource Service (Ref. 4) has also recently reported separation of lean (12-13 gal/ton) ores into concentrates and tailings in a heavy-medium cyclone (densities of about 2 to 2.4). Kerogen recovery fell off sharply as enrichment ratio increased. Plans for further work on the cyclone separation are not clear.

Natural or induced optical properties have been used (Refs. 1,5) to separate ores into lean and rich cuts. But again, the potential for enrichment is limited (as with density separations) and no development work is reported under way.

High-enrichment beneficiation has been researched abroad on European shales for some years (e.g., Refs. 6,7). In fact, current interest in the U.S. was stimulated by the 1979 paper of Fahlstrom of Boliden AB in Sweden (Ref. 8). Fahlstrom stated that Boliden's proprietary methods of grinding and froth flotation would give concentrates with about four-fold kerogen enrichment at 90% kerogen recovery from Colorado shales; such concentrates could provide the basis for a superior shale oil recovery system. However, Boliden has declined to provide further information and has been reported as doing no further work (Refs. 1,9).

Past reviews allude to the technical feasibility of high-enrichment beneficiation (e.g., Ref. 10), citing research both here and abroad. However, authors like Williamson (Ref. 10) and more recently Reisberg (Ref. 11) have been generally pessimistic about commercial applicability--usually on the grounds of high grinding costs.

The level of current U.S. activity on high-enrichment processes is low, judging by published material. The only private-sector work published in recent years has been that of Reisberg of Shell (Ref. 11). However, Reference 1 reports that TRW Energy Systems is working on a proprietary process using a single unspecified liquid which can recover up to 98% of the kerogen in 10-mesh shale ore with up to 10-fold enrichment ratios! SRI has U.S. Department of Energy support for research on froth flotation; no major report has been published yet but some initial data, e.g., as reported by Ref. 1, show results less promising than those claimed by Fahlstrom (Ref. 8).

To sum up, there seems to be a slow arousal of interest in oil shale beneficiation in recent years but the level is still quite low as judged by the amount of work publicly funded or privately funded and published. There are some indications that cancellations and delays of commercial shale oil ventures are encouraging people to rethink the total technological systems, and that may result in more attention to beneficiation.

2.3 The Scope of This Report

The studies included in this report are limited in scope to the following extent:

- No experimental work was undertaken.
- Mining alternatives were not examined; the feedstock was always assumed to be the product of the secondary crusher at the mine site.

- -- No separate upgrading or refining step was examined; the product was always assumed to be raw shale oil (plus associated gas and carbon) although some modest degree of upgrading may occur incidentally in some oil recovery processes considered.
- -- Only beneficiation processes capable of high enrichment at high recoveries were considered; we did not examine low-enrichment beneficiation (e.g., by density separation of crushed rock) which may be attractive but could not change the essential technology or location of the oil recovery step.

The following sections constitute the remainder of the report:

Base Case Systems

A comparison of Tosco II surface retorting of ore and of a kerogen concentrate prepared by ball milling and froth flotation. Comparisons include both capital and operating costs.

Comminution Alternatives

Limitations of ball milling; proposed new methods of comminuting oil shale to particle-liberation sizes including discussions of energy consumption and cost considerations in the several alternatives.

Separation Alternatives

Properties of kerogen and mineral matter which can serve as the basis for a separation process, and some separation efficiencies for froth flotation and for alternative processes. Discussion of engineering and cost aspects.

Recovery Alternatives

Processes for recovering oil from kerogen concentrate which may be superior to Tosco II retorting in terms of oil yield or quality; some estimates of performance and cost.

Alternative Systems

Examination of Base Case technology for other operating assumptions, and evaluation of other system technologies and parameters needed to make beneficiation more attractive.

Remote Recovery

Advantages and disadvantages of transporting the kerogen concentrate away from the ore site to more desirable locations at various distances from the site.

R&D Opportunities

Technical questions on which research seems justified on the grounds that there is potential for an attractive beneficiation-based system.

Dr. I. V. Klumpar of the Energy Laboratory prepared all the detailed flowsheets and cost estimates reported and wrote most of Sections 3, 6, and 7 and the appendices. Prof. C. R. Peterson of the Department of Mechanical Engineering wrote most of Section 4, "Comminution Alternatives". Prof T. A. Ring of the Department of Materials Science and Engineering wrote most of Section 5, "Separation Alternatives". Dr. M. A. Weiss of the Energy Laboratory was Principal Investigator and wrote most of the remainder of the report.

3. BASE CASE SYSTEMS WITH AND WITHOUT BENEFICIATION

In order to assess the potential benefits of incorporating ore beneficiation into a shale oil recovery system, the most direct approach simply compares systems with and without beneficiation. Such a comparison is made in this section for "Base Case" systems, that is, systems based on technology that seems likely to perform reasonably as assumed even though the technology has not been demonstrated on a commercial scale or even on a pilot plant scale in some respects. More speculative systems are considered in subsequent sections.

3.1 Scope of the Base Case

As one of the objectives of this study is to analyze several oil shale beneficiation and extraction alternatives, a common point of departure is needed for their comparison. Accordingly, Base Case systems were selected and defined for mining through recovery of raw shale oil; the mining step is included because increased mining costs are incurred if kerogen recovery in the beneficiation step is less than 100%, although that increase may be offset if oil yield from kerogen concentrate in the recovery step exceeds the yield from conventional retorting. Upgrading and refining were not included because the impact of beneficiation on those steps should be minor.

For the conventional alternative, i.e. without beneficiation, the TOSCO II process as applied in the Colony Project (Ref. 12) is used as the Base Case. Colony was the proposed commercial plant on which most technical and economic information were available when this study started. In addition, TOSCO II retorting seems able to handle kerogen concentrates, though some modifications to the conventional retort train are needed, unlike most or all of the other surface retorting systems in advanced development in the U.S. TOSCO II will be referred to as "shale pyrolysis" in this report. The raw shale input and oil production for Colony are 66 k tons and 50 k bbls per stream day, respectively. This is the only non-beneficiation alternative considered because it is not a purpose of this study to compare processes for the extraction of oil from non-beneficiated shale.

For the Base Case with beneficiation, ball milling and froth flotation followed by concentrate pyrolysis by a modified TOSCO process were selected. Froth flotation is the most widely used technology in the concentration of natural resources containing less than 20% of the constituent sought. The application of froth flotation to oil shale has been investigated by several other researchers, e.g. Ref. 8, 13. A block diagram of the benefication process is shown in Figure 3-1.

3.2 Ore Specification and Mining Technology

To bracket a reasonable range of potential oil shale types, two ores were selected, a rich Western and a lean Eastern shale. Their specifications are listed in Table 3-1. The Western ore is the ore planned for the Colony project (Ref. 12). It is used in the Base Case. (Results for the Eastern ore, whose properties were obtained from Reference 15, are discussed in Section 7.) A fourfold enrichment by beneficiation is postulated for the Base Case with the other

Figure 3-1

OIL SHALE BENEFICIATION BY FLOTATION

Material Balance in k tons/day

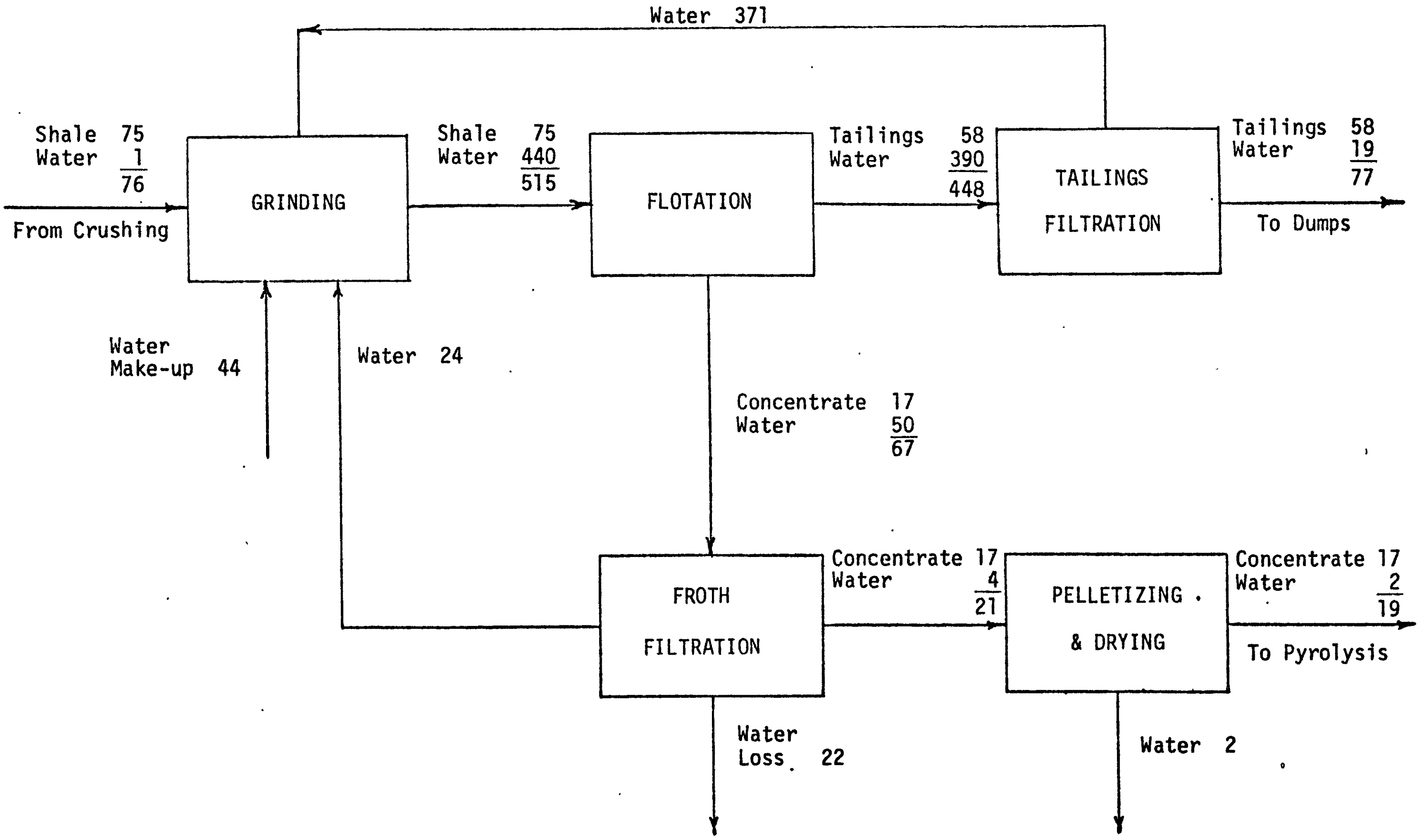


Table 3-1

Typical Oil Shale Specifications

Western Shale

Oil Content (Fischer assay)	35 gal/ton

Average Mineral Composition:

Mineral	Composition wt. %
Dolomite	32
Calcite	16
Quartz	15
Illite	19
Albite	10
K feldspar	6
Pyrite	1
Analcime	1
Total	100

Probable Composition of Organic Matter:

Component	Average wt. % organic matter
Carbon	80.52
Hydrogen	10.30
Nitrogen	2.39
Sulfur	1.04
Oxygen	5.75
H/C atomic ratio	1.54

Liberation Particle Size:	90% less than 20 microns; mass median, 5 microns
Moisture	1%

Eastern Shale

Oil content (Fischer assay)	10 gal/ton

Average Mineral Composition:

Mineral	Composition wt. %
Quartz	22
Feldspar	9
Illite & minor kaolinite	31
and muscovite	31
Carbon	13.6
Total Organic Matter	16-22
Pyrite and marcasite	11
Chlorite	2
Iron oxides	2
Tourmaline, zircon, and apatite	1

Probable Composition of Organic Matter:

Component	Average wt. % organic matter
Carbon	82.0
Hydrogen	7.4
Nitrogen	2.3
Sulfur	2.0
Oxygen	6.3
H/C atomic ratio	1.08

Liberation Particle Size:	90% less than 20 microns mass median, 5 microns
Moisture	5%

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*Sources: References 12, 15, 47

characteristics derived from the flotation process design to be discussed later. The enrichment ratio of four is with the range claimed by Fahlstrom (Ref. 8).

The ore is assumed to be mined by the room and pillar method as described in Colony's environmental documentation (Ref. 16). The mining section includes primary crushing, coarse raw shale storage, secondary crushing (to 0.5 inch), enclosed storage of crushed shale, the linking belt conveyor system, spent shale disposal, and the required truck fleet. Cone and impact crushers are considered for primary and secondary crushing, respectively. The spent shale is disposed of by mixing with other solids, spreading in a nearby gulch, and compacting.

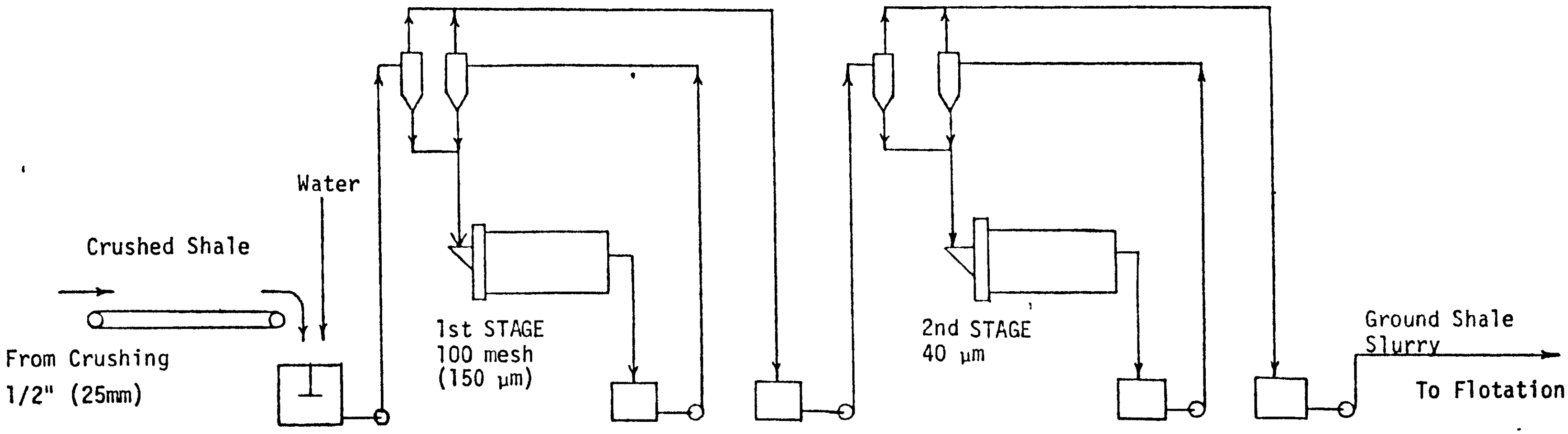
3.3 Comminution by Ball Milling

As the primary and secondary crushing stages are not affected by the various process alternatives we are examining, the term "comminution" will be used in this report only for the tertiary and any subsequent stages which are investigated separately with each alternative. There is no comminution in the Base Case without beneficiation. For flotation, the crushed ore is ground to less than 40 microns in a two-stage ball mill system shown in Figure 3-2.

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. Design criteria, major equipment specifications and requirements for materials, utilities and labor are listed in Tables 3-2, 3-3, and 3-4, respectively.

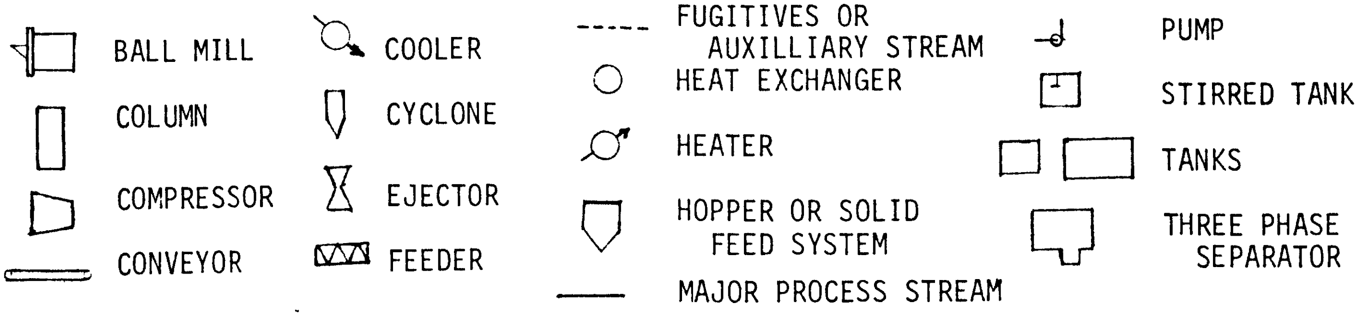
Figure 3-2

OIL SHALE GRINDING



Note: Parallel equipment not shown.

LEGEND*



*This legend also applies to all succeeding Figures.

Table 3-2

Design Criteria

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, lbs/lb shale	0.182
Fractionation products, wt. %	
gas	25.1
naphtha	10.4
gas oil	45.6
bottoms oil	18.9
Balls, lbs/lb 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, k tons/day	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/ft^2/hr, tailings	50
concentrate	25

Table 3-3

Major Beneficiation Equipment

First stage ball mills:	
24' dia x 36'	12
Motors: 4400 HP	24
Second stage ball mills, same specs	12
Motors: 4400 HP	24
Rougher and middling cells, 1000 ft^3	56
Regrind ball mills:	
24' dia x 36'	3
Motors: 4400 HP	6
Rougher cleaner cells, 300 ft^3	25
Middling cleaner cells, 300 ft^3	52
Tailings cleaner cells, 300 ft^3	18
Tailings filters, 3000 ft^2	22 + 4*
Tailings filters, 3000 ft^2	10 + 2*
Rotary dryers, 15' dia x 60'; 450 HP	8
Pelletizers, 14' dia; 54 HP	14 + 2*

*Standby equipment.

Table 3-4

Requirements for Materials, Utilities, and Labor

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 Pyrolysis	Concentrate Pyrolysis	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 lb/hr	120	40	-
Air*, k scfm	1	0.3	5

3. **Manning**

Grinding	37
Flotation	42
Shale pyrolysis	94
Concentrate pyrolysis	54

*Power included above.

**Includes heat required to evaporate water in concentrate.

Based on Swedish and SRI experimental work (Ref. 8,13), the comminution section was designed in more detail using copper industry practices. The key parameter is the work index that controls equipment size and power requirements. The ball mill work index range of 16 to 78 reported for oil shale (Ref. 17) is one of the highest among minerals and fossil fuels.

3.4 Separation by Froth Flotation

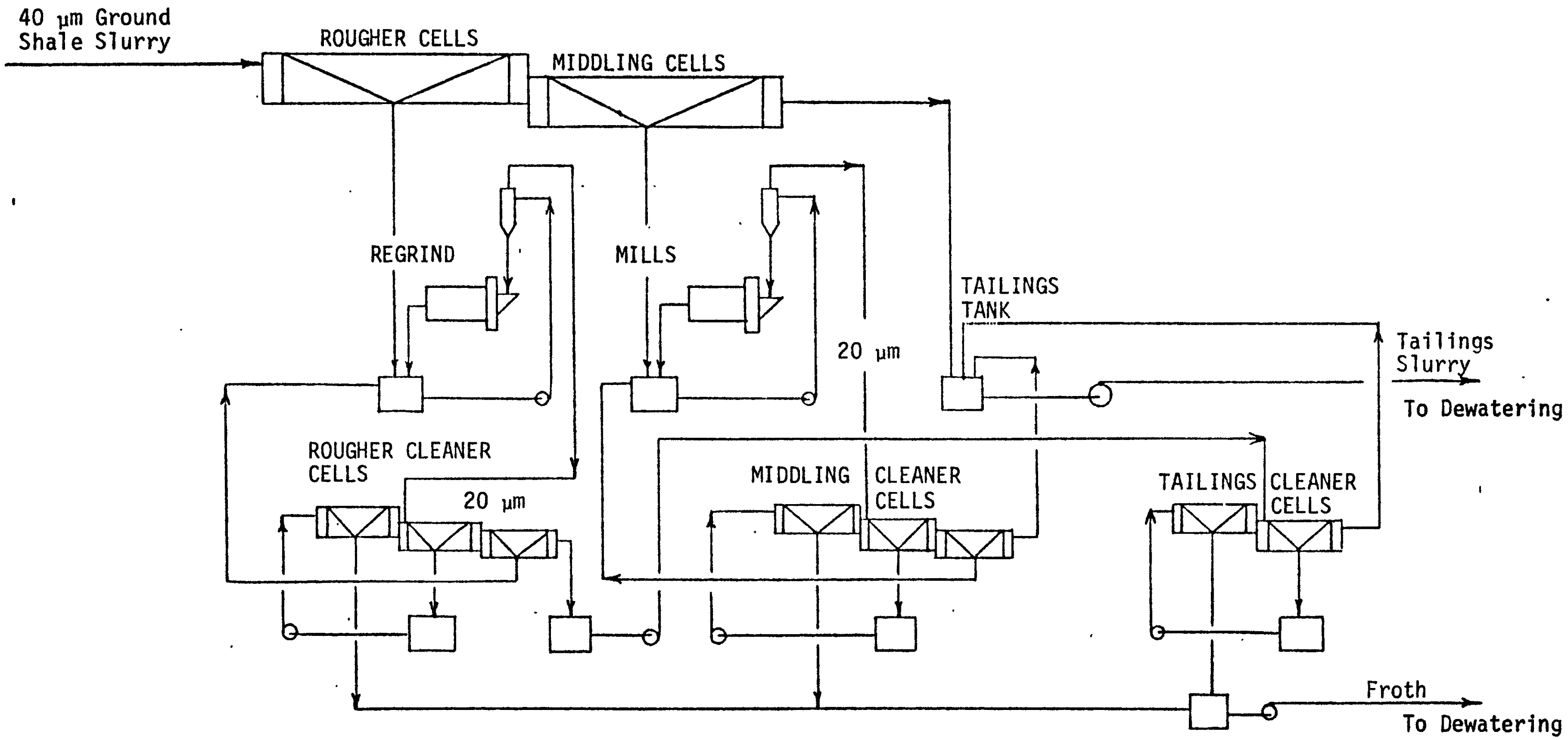
In froth flotation, finely ground oil shale in a water slurry is stirred with froth forming agents in a set of flotation cells. The kerogen-rich particles concentrate in the froth while the tailings particles depleted of kerogen stay in the bulk of liquid. The concentrate and tailings are subsequently 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.

Light alcohol is used as the frother. The adhesion of kerogen-rich particles to the air bubbles is enhanced by collectors such as pine oil or medium boiling shale oil. The pH of the slurry is maintained in the slightly alkaline region by adding lime as a conditioner.

The flotation process is shown in Figure 3-3. Design criteria, major equipment specifications, and requirements for materials, utilities and labor are listed in Tables 3-2, 3-3, and 3-4, respectively. The material balance for the entire beneficiation plant is included in Figure 3-1. The shale-water slurry from the second grinding stage together with the froth forming agents is pumped to the rougher cells. The underflow

Figure 3-3

OIL SHALE FLOTATION



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 middling cleaner cells, respectively, each comprising three stages.

The complex recycling system among these stages and the regrind mills is apparent in Figure 3-3. 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 Dewatering Sections of the benefication plant are shown in Figure 3-4. The concentrate and tailings slurries are both first filtered. The tailings filter cake is trucked to the spent shale disposal area (see Section 3.2) while the concentrate filter cake is pelletized, dried, and conveyed to Pyrolysis. The filtrates are recycled to Grinding.

3.5 Pyrolysis

The process design of Pyrolysis was based chiefly on the environmental documentation of the Colony Project (Ref. 16). The flowsheet for the Base Case without beneficiation is shown in Figure 3-5 while the design criteria are listed in Table 3-2. The plant has six parallel trains. The raw shale from the second stage crusher 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

Figure 3-4

OIL SHALE CONCENTRATE AND TAILINGS DEWATERING

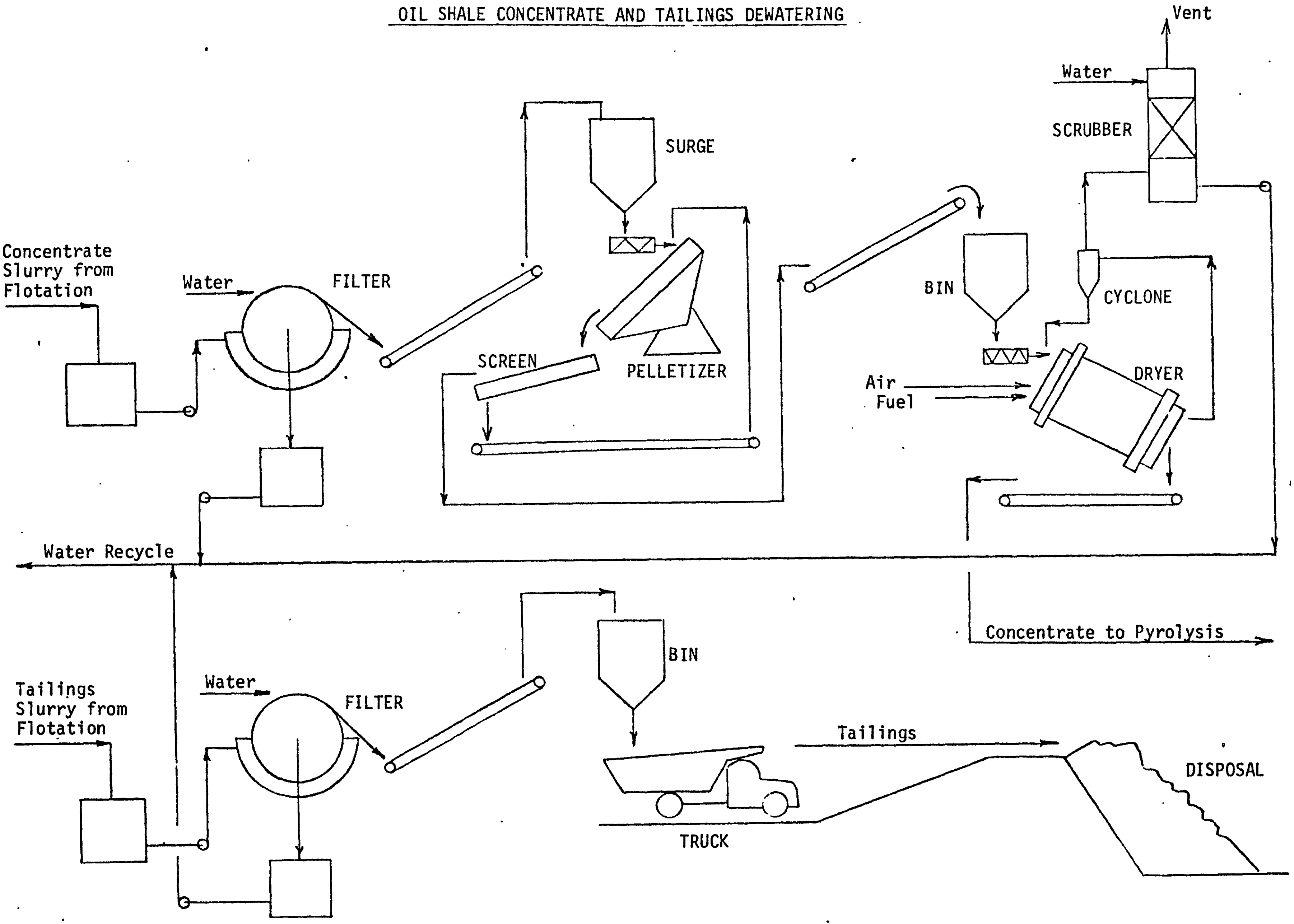
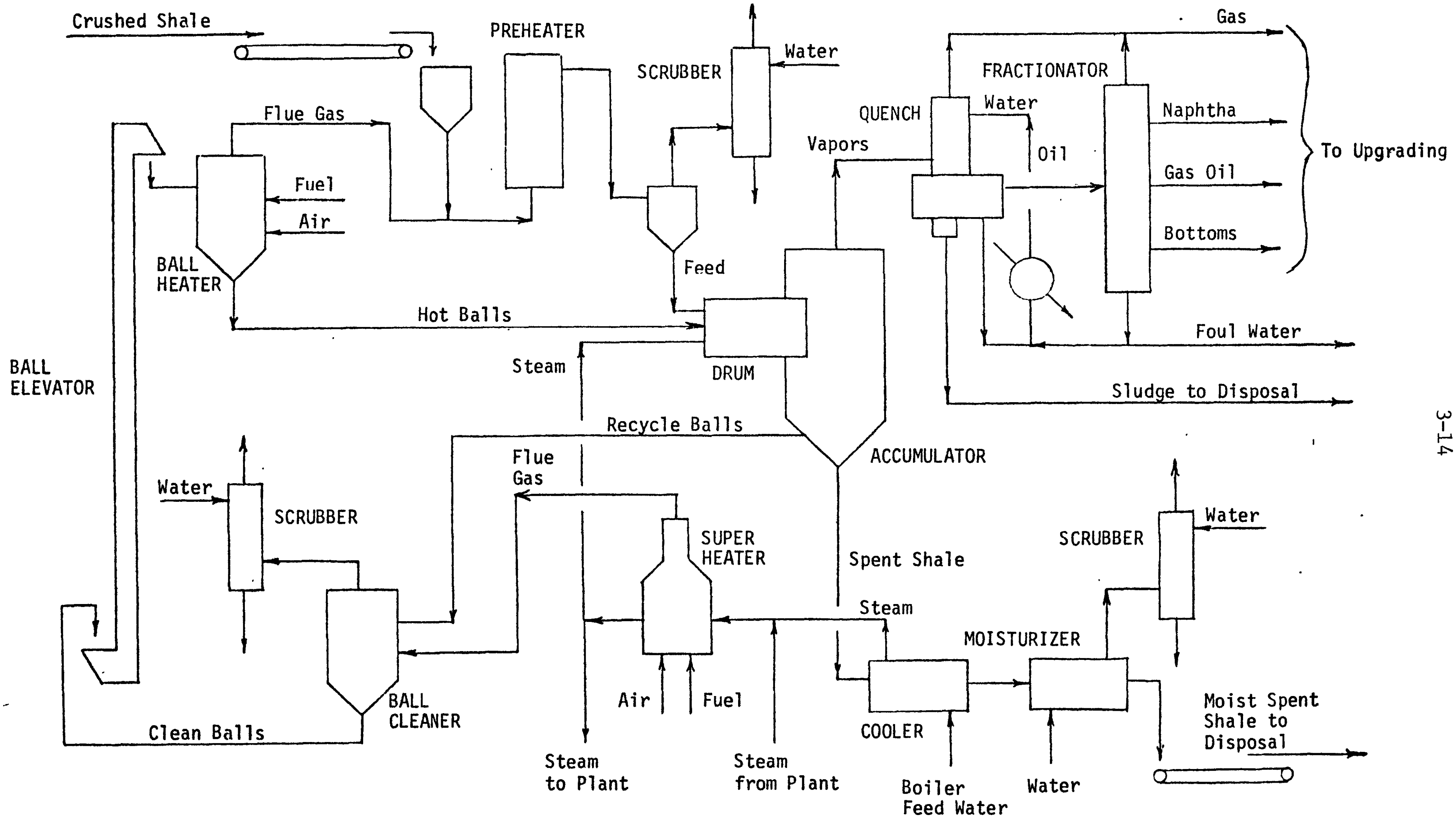


Figure 3-5

SHALE PYROLYSIS



the shale and balls are intimately mixed before they pass into the accumulator.

Overhead vapors include hydrocarbons, carbon monoxide and dioxide, ammonia, hydrogen sulfide, water, and hydrogen. They are quenched with water and separated into gas, sponge oil, gas oil, bottoms oil, and foul water in a fractionator. The flowsheet in Figure 3-5 shows the net naphtha outflow instead of sponge oil with naphtha leaving, and a sponge oil recycle entering the fractionator. The naphtha separation together with the processing of the other streams are part of the upgrading section.

The spent shale is separated from the balls in a rotating trommel screen at the bottom of the accumulator and discharged through a cooler (waste heat) boiler to a moisturizer. The moist spent shale 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. The steam facilities are not fully shown in Figure 3-5 because they are integrated with the steam generator for the entire plant.

The Base Case with beneficiation requires adjustments in the pyrolysis section. While the flowsheet remains the same, the feed rate in terms of dry concentrate decreases by a factor of four. Two parallel trains at a 75% capacity are needed. As the concentrate has a higher moisture content (see Figure 3-1) than the normal ore, the preheater and ball circuit capacities per ton of dry feed together with fuel requirements increase accordingly. No direct steam injection in the retort is required and the moisture content in the overhead vapor is higher than without beneficiation.

In the Base Case without beneficiation, spent shale contains all the rock of the original raw shale with a few percent of unrecoverable kerogen or its non-volatile organic derivatives. In the case with beneficiation, 93% of the rock goes to tailings. Accordingly, the residue separated from the balls at the bottom of the retort has a relatively higher content of organics, approximately four times as much as in the case without beneficiation. Because of its higher organic content and heating value, it may be attractive to burn the residue for heat but that option was not examined. No moisturizer is needed because the residue is mixed with the large amount of tailings before being disposed of. The amounts of waste streams for both Base Cases are stated in Section 3.6 below.

The changes in the pyrolysis section due to beneficiation discussed so far are relatively easy to account for in the preliminary process design that is required for the present study. However, 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 exacerbated 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. A quantitative assessment of the entrainment problem is difficult because of a lack of appropriate data on concentrate particles and pellets. Therefore, a contingency was added to the pyrolysis cost for antientrainment devices, see Section 3.7 below. In addition, the analysis of alternatives to retorting for oil recovery provides an opportunity to identify processes more suitable for handling the fine concentrate.

Design criteria, and requirements for materials, utilities and labor are listed in Tables 3-2 and 3-4, respectively.

3.6 Environmental Aspects

The waste effluents for the Base Cases and corresponding pollution control equipment are summarized in Table 3-5. For the case without beneficiation, the data are based on the environmental documentation (Ref. 16).

For the beneficiation case, the effluent flowrates have been increased or decreased in proportion to the throughput of the corresponding process stream. The capacities of pollution control equipment in the Mine Crushing and Waste Disposal Section have to be larger because more shale is needed due to kerogen losses in the Beneficiation section. On the other hand, the effluent flowrates are lower in Pyrolysis because less mass of concentrate than ore is processed. However, the capacity of the feed preheat system including the ball cleaner is larger than would correspond to the enrichment ratio of 1 to 4 because of the higher moisture content of the concentrate. The spent shale moisturizer is absent because the residue is mixed with a larger volume of wet tailings.

The largest effluent stream in the Beneficiation Section is the dryer offgas which was estimated based on the dryer system design (see Figure 3-4). Tailings are accounted for in the waste disposal area.

Table 3-5

Pollution Control

Dashes (--) indicate flowrates that vary widely or do not significantly affect control equipment selection and sizing. NA stands for Not Applicable.

Section and Equipment or Facility	Material Controlled	Type of Control	Flow Rate Unit	Base Case Flowrate Without Beneficiation	Base Case Flowrate With Beneficiation
Mining					
Mine stockpile	Shale dust	Water sprays	--	--	--
Primary crusher	Air+shale dust	Fabric filters	k acfm	62	72
Secondary crusher	Air+shale dust	Fabric filters	k acfm	70	83
Storage building	Air+shale dust	Fabric filters	k acfm	36	42
Waste disposal	Spent shale or tailings, etc.	Landfill	k tons/day	55	65
Conveyors	Shale dust	Foam sprays	--	--	--
Beneficiation					
Pelletizers	Concentrate	Fabric filters	--	NA	--
Dryers	Air+concentrate dust	Cyclone, scrubbers	k acfm	NA	200
Conveyors	Concentrate dust	Foam sprays	--	NA	--
Pyrolysis					
Preheat system	Air+feed dust, Hydrocarbons	Scrubbers,	k acfm	210	83
		Thermal oxidizers	k acfm	350	140
Ball cleaners	Flue gas+feed dust	Scrubbers	k acfm	44	17
Moisturizers	Air+spent shale	Scrubbers	k acfm	44	NA
Conveyors	Spent shale or residue dust	Foam sprays	--	--	--

Dusts from the pelletizers and conveyors are a relatively small pollution source.

Gas and liquid effluents from the pyrolysis step originate primarily with the kerogen and thus (except for water) are not greatly changed by switching from ore to concentrate. But gas effluents from the preheat and ball systems are reduced almost three-fold while solid wastes (spent shale) from pyrolysis are reduced more than four-fold. These environmental credits for beneficiation are offset to an unknown degree by the problems of tailings (and associated water) disposal including but not limited to the liberation of soluble salts, trace elements, and residual organic material, e.g. References 31 and 71.

3.7 Cost Estimation

The history of cost estimates of synthetic fuels plants does not inspire high confidence in the state of the art for making such estimates. The more speculative the technology, the greater the probable error, and some of the technologies considered in this study are, of course, very speculative. It is not facetious to say that, in many cases, total costs of proposed synthetic fuels plants have not been known confidently to one significant figure, e.g. Colony (Ref. 56), although numbers are often tabulated with two to three significant figures (as here) or even more.

Therefore, the absolute levels of costs cited in this study should be regarded with at least the same skepticism deserved by other estimates. However, we have tried to estimate costs in a consistent way so that relative comparisons should have reasonable validity even if the absolute numbers are crude.

3.7.1 Capital Cost Methodology

Preliminary capital cost estimation differs from the definitive cost estimates that precede actual construction. The latter are based on firm quotations from equipment vendors, and commodity material and labor "take-offs" from detailed drawings. The most accurate preliminary estimates start with costing of individual equipment, shown on a process flowsheet, using telephone quotations, historic data or literature. All other cost items such as installation, buildings, offsites and indirects, are estimated through factors based on equipment cost as explained in Appendix A (Ref. 78). If process flowsheets and/or sizes of individual equipment are not available, or the time-consuming accurate method is not warranted, various short-cut methods are used that can be classified in the following two categories:

- Sectional methods scale up or down cost of plant sections that are identical or similar, using empirical exponents, e,

$$\text{cost} = (\text{reference cost})(\text{capacity}/\text{reference capacity})^{e}$$

  A typical exponent for processing plants is 0.6. Within a narrow capacity range, mine operations can be scaled with an exponent of 1.0, keeping a small portion of the cost constant. In multiple train plants, the number of trains is first roughly adjusted to the desired capacity, and the above equation then applied to the individual trains.

- Modular methods estimate capital of chemical plants as an exponential function of plant capacity, number of modules, temperature, pressure, material of construction, and possibly an additional parameter. The disadvantage of this method is the ambiguity in the module definition that ranges from a major equipment unit to a functional unit (unit operation) to a chemical step. A modification of the modular method, discussed in Appendix B, was developed for this study (Ref. 79). It precisely defines twelve types of "process modules" and extends the correlation beyond the chemical industry.

Using these methods, cost comparisons for the Base Case were made. They were limited to the Mining through Pyrolysis sections; upgrading was not considered. For the conventional Base Case without beneficiation, cost estimates were based on the last published itemized data for the Colony project (Ref. 18). Those data, dated September 1977 and reproduced in Table 3-6, resulted in a total capital cost of about one billion 1977 dollars. A more realistic total capital cost in 1981 for an "instant plant" was about three billion dollars; escalation and interest during construction are excluded. Therefore, we escalated the 1977 items to a new total of about three billion dollars when we began this study. Recent events place the Colony total at five to six billion in as-spent 1981-1986 dollars, e.g. Ref. 56. However we did not further escalate our total and the relative comparisons of shale processes are unaffected by the level of the total.

To arrive at updated numbers useful for our purposes, the data of Table 3-6 were treated in the following way:

Table 3-6

Breakdown of Investment for the Colony Project

(September 1977 Dollars)

DESCRIPTION	Capital Investment $ Million
Contractor Estimates	
Mining, Crushing and Spent Shale Disposal	113
Pyrolysis and Oil Recovery	266
Oil Upgrading and Hydrogen Plant	102
By-Product Recovery	59
Utilities and General Facilities	177
Subtotal*	717
Reserve Costs	131
Other Owner Costs	
Mine and Spent Shale Disposal Mobile Eqipment	28
Catalysts and Chemicals	16
Spare Parts	4
Project Management and Plant Staffing	30
Taxes and Insurance During Construction	7
Community Assistance Costs	30
Precommitment Costs	10
Mine Predevelopment Costs	11
Prepaid Process Licenses	2
Miscellaneous Other Costs	10
Working Capital	22
Plant Fixit and Start-up Allowance	32
Subtotal	202
GRAND TOTAL INVESTMENT	1,050

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*The subtotal of $717 million is an update estimated by an engineering contractor. Allocation of costs to the individual units is by TOSCO after analysis of the detailed contractor estimates.

Source: Reference 18

- Individual operations of the Colony Project, such as mining, crushing, and pyrolysis, were compared with similar other facilities.
- Other information on the Colony Project was obtained informally to assist us in estimating changes in scope and a rough cost distribution among pyrolysis, offsites, and the rest of facilities.
- As there are not enough engineering data available for a budget estimate, the modular capital estimation method described above was applied to Shale Pyrolysis which, as new technology, is most difficult to evaluate by comparing to other known processes.

The following conclusions were drawn from the cost comparisons, specific Colony Project information (Ref. 75), and the modular estimate.

(a) The major cost increases since 1977 are due to vastly expanded offsites including additional roads, dams, a shale transportation tunnel, and a grass roots township with all buildings and services constructed from scratch.

(b) Substantial scope changes occurred in the upgrading section (e.g., arsenic removal) resulting in cost changes.

(c) Pyrolysis was underestimated by about 25%.

(d) There is essentially no change in the mining, crushing, and waste disposal estimate. The comparatively low costs are realistic because the Colony shale is easy to mine.

(e) The last estimate increase in 1982 were caused primarily by site-specific offsites which would have little effect on the Mining through Pyrolysis sections covered by this study.

3.7.2 Capital Estimate: Conventional Retorting

The capital estimate for the conventional Base Case, without beneficiation, totals $2910 million and is broken down in the first column of Table 3-7. The first item in each section is the escalated cost from Table 3-6. The next items are the "other owner cost" components specific for each section. The remaining "other owner costs" were prorated. "Reserve costs" are denoted as Land Acquisition in Table 3-7.

Each Table 3-7 item was escalated from September 1977 to mid 1981 using the Chemical Engineering magazine plant cost index. The cost of Pyrolysis and Oil Processing (the latter being the sum of oil upgrading, hydrogen plant, and by-product recovery) were increased to reflect both the underestimate and scope changes. The item "additional roads and dams, tunnel, and township" represents the offsites expansion. As project contingency was included in the 1977 estimate, only process contingencies were added in Table 3-7 at various percentages according to the estimate uncertainties.

The pyrolysis cost was independently confirmed using the modular method discussed in Section 3.7.1. The two estimates differed by six percent, i.e. they were identical well within the precision of estimation.

3.7.3 Capital Estimate: With Beneficiation

Capital costs for the Base Case with beneficiation are summarized in the second column of Table 3-7 and broken down in Table 3-8. The total is $2950 million, essentially no different from the $2910 million

Table 3-7

Total Capital Summary ($M)

(Mid-1981 Dollars)

	Base Case			
	Without Beneficiation		With Beneficiation	
Mine				
Mining, crushing, and solids disposal	160		180	
Mobile equipment	40		45	
Predevelopment	15		17	
Other costs (prorated)	25		28	
Subtotal	240		270	
Process contingency (20%)	50		50	
Total		290		320
Beneficiation				
Grinding			216	
Flotation and dewatering			126	
Other costs (prorated)			35	
Subtotal (Rounded)			380	
Contingency: Project (15%)			60	
Process (30%)			110	
Total				550
Pyrolysis				
Feed preparation, retorting, and oil recovery	490		144	
Other costs (prorated)	60		16	
Subtotal	550		160	
Process contingency (40%)	220		60	
Total		770		220
Oil Processing				
Upgrading, H_2 plant, and byproduct recovery	360		360	
Catalysts and chemicals	20		20	
Other costs (prorated)	40		40	
Subtotal	420		420	
Process contingency (30%)	130		130	
Total		550		550
Offsites				
Utilities and general facilities	250		250	
Community assistance	40		40	
Additional dams and roads, tunnel, township	460		460	
Other costs (prorated)	50		50	
Subtotal	800		800	
Process contingency (40%)	320		320	
Total		1120		1120
Land acquisition	190		190	
GRAND TOTAL		2910		2950

Table 3-8

Beneficiation Capital Cost

Item	$M
A. Oil Shale Grinding	
Mills	72
Other equipment	5
Subtotal - purchased equipment	77
Installation	83
Subtotal - direct cost	160
Indirects (35%)	56
Subtotal	216
Working capital, start-up, and other non-depreciable cost (10%)	22
Subtotal	238
Contingency: project (15%)	36
process (30%)	71
Total Capital (Rounded)	350
B. Flotation and Dewatering	
Flotation	
Regrind mills	9.0
Cells and other equipment	4.3
Subtotal - purchased equipment	13.3
Installation	21.7
Subtotal - direct cost	35
Concentrate Filtration	
Purchased equipment	3.4
Installation	6.7
Subtotal - direct cost	10
Pelletizing	
Purchased equipment	2.3
Installation	3.0
Subtotal - direct cost	5
Drying	
Purchased equipment	9.3
Installation	12.1
Subtotal - direct cost	21
Tailings Filtration	
Purchased equipment	7.3
Installation	14.6
Subtotal - direct cost	22
Total direct cost	93
Indirects (35%)	33
Subtotal -depreciable capital	126
Working capital, start-up, and other non-depreciable cost (10%)	13
Subtotal	139
Contingency: project (15%)	21
process (30%)	42
Total capital (Rounded)	200
GRAND TOTAL	550

estimated for the conventional Base Case. The purchased equipment items are sums of individual equipment costs. The installation, indirects, working capital, start-up and other non-depreciable items were estimated using percentage factors, some of which are indicated in Table 3-8. Installation factors are not shown because they differ for various equipment types, and commodity material and labor sub-items such as concrete, steel, piping, electricals, instrumentation, insulation, and paint.

Individual equipment cost and installation factors used in this study are based on copper industry experience. They were compiled from numerous engineering reports. The compilation of equipment cost is part of the ASPEN documentation (Ref. 54) while the factors are defined and tabulated in Appendix A. Pyrolysis and beneficiation capital were independently confirmed using the modular estimate method discussed in Section 3.7.1. The two estimates differed seven percent; again, the two estimates were identical for all practical purposes.

The other sections of the Base Case without beneficiation were scaled up or down from the alternative without beneficiation and are shown in the last column of Table 3-7. The Mine costs are higher for the beneficiation alternative because more shale has to be mined to compensate for the losses in Beneficiation. Concentrate pyrolysis cost is lower because only two trains are required, each at 75% capacity of those for Shale Pyrolysis but with a larger preheat circuit. Oil Processing and Offsite costs were assumed the same for both alternatives. That assumption is an approximation, consistent with the scope of this study.

3.7.4 Annual Cost

Annual costs for the Base Cases with and without beneficiation were calculated as the sums of two elements: (1) annual operating costs and (2) annual capital charges. Annual operating costs include items like wages and salaries, purchased chemicals, purchased utilities, and other materials and services consumed as a result of operating the plant. Annual capital charges are calculated most conveniently as a percentage of total "instant" capital and they account for depreciation, return on investment, interest during construction, startup difficulties, and other costs such as property taxes and insurance. The flat annual capital charge is a crude but customary way to account for capital costs and the level can be easily adjusted to account for different types of financing, rates of return, or other particular circumstances.

In the computations of annual cost in this section:

- Both annual operating costs and capital charges were calculated for only the battery limits of Mining through Pyrolysis sections since other sections should not be affected significantly by the presence or absence of a beneficiation section; thus less than half of the total capital investment of the plant is included in the sections considered. Battery limits exclude offsites.
- The annual capital charge was assumed to be 25%. For all-equity financing, and reasonable construction, startup, and other assumptions, a 25% capital charge should result in about a 12 to 15% discounted cash flow rate of return on investment in the sections considered.

The total annual cost for the Mining through Pyrolysis sections amounts to $370 million for the conventional Base Case and $460 million for the beneficiation Base Case. Details are shown in Table 3-9.

In both Base Cases, most of the total annual costs arise from capital charges rather than annual operating costs, emphasizing the usual importance of investment to the economics of shale oil plants. But since the capital costs for the two cases were about the same (see Table 3-7), the differences in these two particular Base Cases result from a difference in operating costs, specifically in the large cost of electric power for grinding in the beneficiation step.

Based on these results, our obvious conclusion is that beneficiation using the technology selected is not an attractive alternative to the conventional process sequence for extracting oil from Western oil shale. Making it attractive will require:

- major advances in the technology and economics of the beneficiation step, specifically by reduction of the investment and power consumption in grinding, and/or
- capturing major credits elsewhere in the total system by virtue of remote recovery or by recovery by means superior to conventional retorting.

The following sections consider some of the potentially attractive options.

Table 3-9

Annual Operating Costs: Mining Through Pyrolysis Sections

Item	Unit Cost or Percent	Base Case Without Beneficiation Cost $M/yr	Base Case With Beneficiation Cost $M/yr
1. Mine Operating Costs			
Mining and haulage		16	18
Crushing		3	4
Waste disposal		2	2
Subtotal		21	24
2. Pyrolysis Operating Costs			
Fuel	$3/M Btu	30.7	11.8
Power	5¢/kWh	0.2	0.1
Water	40¢/k gal	0.6	0.1
Steam cost/credit	$4/k lbs	3.8	-1.3
Chemicals	-	1.0	1.0
Oper. labor (OL)	$13/hr	2.5	1.5
Superv. and services	40% OL	1.0	0.6
Overhead	40% OL & ML	7.2	2.4
Maint. labor (ML)	2% capital	15.4	4.4
Oper. & maint. supplies	3% capital	23.1	6.6
Subtotal (Rounded)		86	27
3. Beneficiation Operating Costs			
Fuel	$3/M Btu		5.9
Power	5¢/kWh		70.8
Water	40¢/k gal		1.3
Chemicals	-		3.7
Grinding balls	avg. $300/ton		14.8
Oper. labor (OL)	$13/hr		2.1
Superv. and services	40% OL		0.8
Overhead	40% OL+ML		7.4
Maint. labor (ML)	3% capital		16.5
Oper. & maint. supplies	2% capital		11.0
Subtotal (Rounded)		--	134
Total Operating Costs (Rounded)		107	186
Capital Charges for Plant Sections 1, 2, and 3 Above	25%	265	273
Total Annual Cost for Plant Sections 1, 2, and 3 Above (Rounded)		370	460

4. COMMINUTION ALTERNATIVES

The preceding section, describing "Base Case" beneficiation systems using ball milling, confirms the conventional wisdom (e.g., Ref. 10) that the costs of comminution are a heavy burden for a beneficiation system to bear. Therefore, methods of comminution which might be significantly better than ball milling are of interest. In this section, the limitations of ball milling are described and the potential advantages of other new comminution methods and devices are discussed.

4.1 Conventional Milling

4.1.1 Ball Milling Background

Conventionally, fine grinding of material is accomplished in ball mills or, increasingly, in autogenous or semi-autogenous mills. As the names imply, ball mills utilize a tumbling mass of balls (typically steel); autogenous mills use a tumbling mass of the ore to be ground; and semi-autogenous mills use a mixture of the two. While there are significant differences in their operating economics, they are for our purposes essentially similar in physical operation. That is, they all accomplish energy input via mechanical lifting of the grinding media, and some of this energy is subsequently delivered to the material being ground as the media tumble back down within a rotating drum. We shall see that this indirect and random delivery of the useful energy component is at the heart of the poor efficiency of such devices.

It is widely accepted that ball mills are extremely inefficient devices (autogenous and semi-autogenous mills are typically slightly less efficient). Some (Ref. 19,20) put the efficiency at "no more than a fraction of one percent," while others (Ref. 21) state that, "Up to 99% of the energy consumed in grinding these ores may be expended in the movement of machinery, with noise and heat the undesirable by-products, leaving only one percent of the applied energy for size reduction." Efficiency is here defined as the ratio of the minimum energy input required for the observed fragmentation to the actual mechanical energy input (i.e., to drive the rotating drum).

Minimum required energy is generally taken to be the surface energy of the new surfaces created by fragmentation, in which case the efficiency is indeed very low. Some would argue that this is an unduly pessimistic assessment in that fragmentation by <u>any</u> process actually creates more new surface, in the form of multiple micro cracks, than that of the surfaces of separation.* Perhaps a "fracture energy" which includes the energy of these micro cracks as well as the separation surfaces should be used. In any case, and by any reasonable measure, the efficiency of conventional milling processes is <u>very</u> low, leaving plenty of room for improvement.

Milling is a highly empirical "science," perhaps more so than many of its practitioners realize. A major review of the state of the art (Ref. 21) concluded that, "...there is evidence that many of today's crushing and grinding techniques not only are inefficient and antiquated

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*But this micro crack energy input is not necessarily wasted: it weakens the material for subsequent fracture and the micro crack surfaces may ultimately become separation surfaces.

but have little or no scientific base of understanding." Being "antiquated" is of course no reason to discard a technology so long as it functions economically. Over the years conventional crushing and grinding technology has done just that, providing reliable low-cost comminution with relatively simple, albeit "antiquated", devices. Now, with rapidly increasing energy costs (and declining ore grades), the very low efficiencies of conventional practices render them uneconomic in many situations. This is quite apparent in oil shale beneficiation schemes, but it is also of concern in other areas. This broad concern prompted the review of Reference 21 which notes that crushing and grinding consume roughly 2% of the entire electric power generated in this country!

Despite the existence of several theories of comminution, the science is largely, almost exclusively, an empirical one, aimed at deriving relationships to permit the design, selection, and optimization of conventional components and grinding circuits. An extensive and continuing industry program has been carried out in the effort to meet the needs of the mineral industry. While this work has been necessary and is to be commended, it is unfortunate that a proportionate basic study program has not been possible as well, especially in a field where the gap between actual performance, no matter how carefully optimized, and reasonable physical potential is so very great.

4.1.2 Theoretical Considerations

At least three "theories" have been advanced to describe comminution, as summarized by Bond (Ref. 22) in presenting his "third theory of comminution."

Rittinger states that the energy necessary to break a particle of diameter D is proportional to D^2, which obviously is in keeping with the preceding surface energy concept (or the surface plus micro crack areas if the latter are proportional to the separation surface area). Kick states that the energy is proportional to the volume of the particle or to D^3. Physically this dependence might be explained in terms of a volume energy storage, as will be discussed shortly. Bond, in his third theory of comminution, suggests the energy should be proportional to $D^{2.5}$, largely, it seems, because this exponent is halfway between the preceding two. "Proof" of his theory is offered in the form of ball (and rod) mill data rather than direct measurement of material behavior. While material properties no doubt have some influence on ball mill performance, tests using a device believed to be about 99% inefficient are predominantly tests of the device rather than of the material. Such data may determine empirical means for the design and application of the device but they should not be interpreted as in any way defining comminution behavior in general. As a case in point, MacPherson (Ref. 23), in setting out a "procedure to arrive at an energy-efficient autogenous grinding plant," suggests tests in a standard laboratory ball mill, "to determine 100% power efficiency." Ball mill performance may well be a legitimate standard by which to judge autogenous mill performance, since the latter are typically slightly less efficient, and his procedure an effective way to approach that standard, but the phrase "100% power efficiency" sounds a good deal more impressive than it really is. Empirical procedures, though useful and necessary within their proper place, have a tendency to become "laws" that inadvertently mask basic phenomena and inhibit fresh insights.

To return to basics, we can suggest a "fourth theory of comminution" that, while certainly not proved or even strongly promoted here, will shed some light on basic issues and may help to illustrate the potential beyond that of conventional practices.

As for the first three theories of comminution, let us for the moment assume that the material to be fragmented is homogeneous in the sense that its behavior does not vary with particle size over the range of sizes of interest.* There can be little doubt that the minimum energy for fragmentation is related to the creation of new surfaces. Whether we consider an absolute minimum surface energy or a somewhat greater fracture energy as previously discussed, the conclusion is the same: minimum energy input to fracture a particle of "diameter" D is proportional to D^2, suggesting the Rittinger theory.**

Next consider how the energy for fragmentation is actually applied. In all practical devices the energy is applied by external forces, usually compressive, which distort the particle, storing energy within the <u>volume</u> of the particle, and creating stresses therein. For example, consider the simple case of a uniaxially loaded cube of dimension D. For elastic deformation the energy-stress relationship is

$$\text{Energy} = \frac{1}{2}\frac{\sigma^2}{E} D^3$$

where E is the elastic modulus of the material and @ is the (uniform) stress. Thus the energy necessary to reach a particular failure stress

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*There certainly <u>are</u> size effects, particularly when considering grinds fine enough to liberate individual minerals, but we shall mention these complications later.

**Our assumption of uniformity means that the ratio of micro crack surface area to separation surface area is also uniform.

(an assumed constant material property) is proportional to D^3, suggesting the Kick theory. For other shapes and for other than elastic deformation the conclusion is the same: energy to reach a fixed failure stress at some point within the particle is proportional to D^3, although the stress may not be uniform and the proportionality constant may be a good deal more complex.

When failure occurs stored energy (at least the elastic portion thereof) flows to the fracture to cause its propagation, but there is nothing that says the stored energy must be just sufficient to drive a crack through the particle. For "large" particles, having a large volume-to-surface ratio, there will be excess stored energy which will appear as kinetic energy of the fragments, stress waves moving within the fragments, noise, and so on. For "small" particles, having low volume-to-surface ratios, there will be insufficient stored energy and the particle will not fragment until further energy is put in from external forces. The distinctions "large" and "small" will depend upon material properties (strength, modulus of elasticity) and particle shape. Although there are insufficient data at present, this change of behavior may well be significant when grinding to liberation sizes.

The preceding arguments suggest the "fourth theory of comminution," shown graphically in Figure 4-1:

o For very small particles, energy stored during particle loading is insufficient to drive a crack through the particle: hence actual energy input for fracture will be proportional to D^2.*

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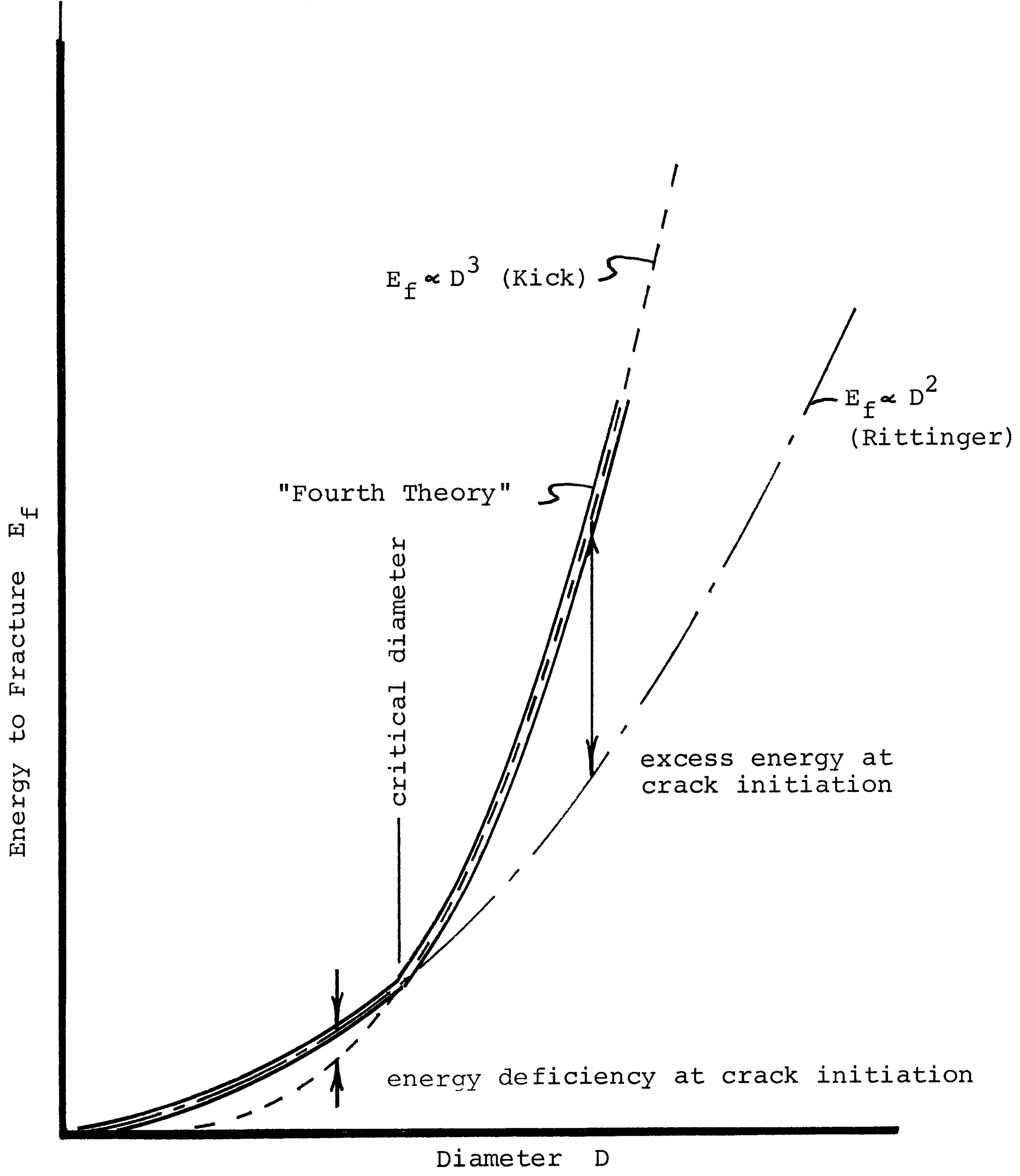
*Strictly speaking, at separation the fragments even in this case will carry with them some stored energy, but once a fracture has started the general stress levels, and consequently the stored energy, should be small.

Figure 4-1

"FOURTH THEORY OF COMMINUTION"

(for uniform material)



- For large particles, energy stored during particle loading is more than enough to drive a crack through the particle: hence actual energy input in a practical device will be proportional to D^3.
- A critical particle size, dependent upon material strength, elastic modulus, and particle shape, may be defined which divides these two regions and for which the stored energy is just sufficient to drive a crack through the particle.

Like Bond's third theory, this "fourth theory" is in a sense halfway between those of Rittinger and Kick, but it does seem to have some basis in physical behavior. Indeed, if one were to try to match the discontinuous fourth theory curve with a single exponent, $D^{2.5}$ would be better than either D^2 or D^3, but it is extremely unlikely that this detailed material behavior would appear as a 2.5 power law governing ball mill performance when something like 99% of the energy consumption is not material-related.

In reality, of course, many other material properties might mask this uniform material behavior. Particle failure is influenced by flaws within the particle, and the population of such flaws decreases as particle size decreases (i.e., the material becomes stronger). At liberation, particles are individual minerals and grain size is clearly not negligible. It may well be that the "fourth theory" size effect is masked by such other phenomena. On the other hand, material flaws may be essentially absent in very small particles.

The concept seems worthy of further study, particularly for very fine grinding as is necessary for oil shale. In any case, this presentation should serve to illustrate that there are material and particle behavior issues of great importance that are not to be found in empirical correlations of ball mill performance.

4.2 Some Identifiable Losses in Conventional Ball Milling

For a process consisting of roughly 99% losses, loss mechanisms should not be hard to identify. Indeed, they are not, but, at present anyway, they are difficult if not impossible to quantify. One can separate the loss mechanisms into two major categories: one mechanical, having to do with the transfer of useful energy from the grinding media to the particles being ground; and the other best described as fluid, having to do with the transport of particles through the grinding zone. Other loss mechanisms are present of course, such as mechanical drive losses, but these are either minor or easily remedied if they are not. This distinction of loss mechanisms, though the two are interrelated, will be useful in Section 4.3 where we attempt to define new design paths for improved processes.

Mechanical losses within a conventional ball mill stem largely from the fact that energy input is to the grinding media rather than to the material being ground. Although the full range of grinding action is no doubt very complex, some straightforward calculations for simple interactions, presented in following sections, will illustrate the enormous potential for losses in trying to deliver this energy for useful purposes.

Fluid losses are harder to quantify, but it is generally believed that failure to promptly remove ground material from the grinding zone results in large losses in "regrinding." It is possible, though this is pure speculation at this time, that "regrinding" losses are not large, as we shall discuss briefly in Section 4.3. Be that as it may, the effective use of fluid action to move material to the proper grinding locations and to remove it when finished is no doubt an important aspect of efficient mill design (unless, of course, some other means of material transport is used).

The following four subsections discuss simple interactions between grinding balls and relatively small particles.

4.2.1 Limited Effective Impact Area

The effective impact area between two spheres is a function of sphere diameter and the size of particle to be trapped between spheres. "Impact" here refers to ball-to-ball contact, whether it occurs at relatively high velocity between balls tumbling on the surface of the charge or more slowly elsewhere within the charge. Figure 4-2a illustrates the limiting radius R_c at which a particle of radius R_p can be captured between two spheres of radius R_b. Thus capture radius is given by

$$R_c^2 = (R_b + R_p)^2 - R_b^2$$

or

$$R_c^2 = 2R_bR_p + R_p^2 \simeq 2R_bR_p \tag{4-1}$$

for $R_b >> R_p$. The capture area A_c is then

$$A_c = \pi R_c^2 \simeq 2\pi\ R_bR_p \tag{4-2}$$

The capture volume which might contain particles of radius R_p is simple the area times $2R_p$ or

$$V_c = 4\pi R_p^2 R_b \tag{4-3}$$

Note that this is a somewhat optimistic estimate in that, if the volume were full of such particles, the two balls could not approach to contact as assumed in these derivations. Note also that in comparison to ball volume or any other characteristic volume, this is a very small volume. For example, for 50 μ particles and 100 mm balls,

$$\frac{V_c}{V_b} = \frac{4\pi R_p^2 R_b}{\frac{4}{3}\pi R_b^3} = 3(\frac{R_p}{R_b})^2 = .75 \times 10^{-6}$$

Though small volume per se has no direct influence on efficiency, this suggests that the active portion of a ball mill is a very small fraction of the total machine volume. Section 4.2.5 will indicate that this small volume can indirectly and adversely affect efficiency.

Figure 4-2b illustrates the interaction between a sphere and a flat plate for particles of radius R_p. This might characterize the interaction against the shell liner or another grinder geometry, as described in Section 4.4. In this case the capture terms are given by

$$R_c^2 = (R_b + R_p)^2 - (R_b - R_p)^2 = 4R_b R_p \tag{4-4}$$

$$A_c = 4\pi R_b R_p \tag{4-5}$$

$$V_c = 8\pi R_p^2 R_b \tag{4-6}$$

Each is twice its corresponding ball-to-ball counterpart.

Figure 4-2a

CAPTURE GEOMETRY FOR EQUAL BALLS

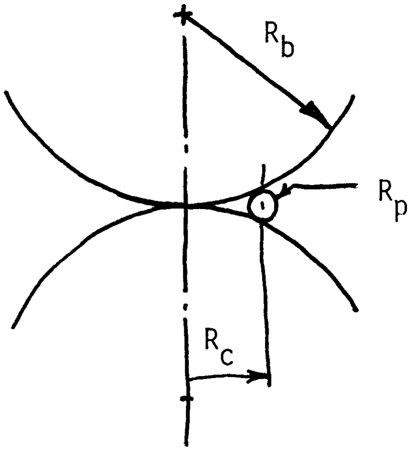
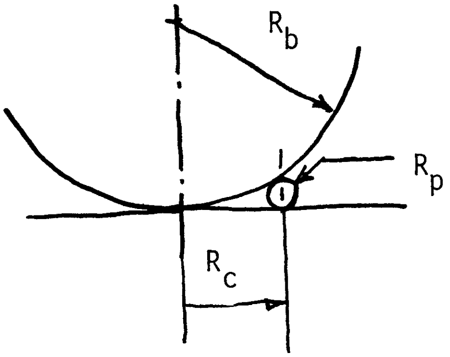


Figure 4-2b

CAPTURE GEOMETRY FOR BALL AND FLAT PLATE



4.2.2 Insufficient Charge per Impact

The preceding capture volume is of no use whatever if there is no material within the volume when the balls approach each other. Thus one obvious loss mechanism is insufficient material within the mill, or a distribution system that does not keep all of the interactions supplied with sufficient material. Indicative of the indirect energy input and this loss mechanism, the power input to a ball mill loaded with balls but with no material to be ground is essentially the same as that to a properly loaded mill.

4.2.3 Non-Uniformly Loaded Impact

It has already been pointed out that the capture volume calculated is an optimistic estimate since, with material present, the balls do not approach to contact. For the same reason, the presence of a few larger particles would shield smaller particles from action even though the latter are within the computed capture volume. It appears then that the ideal distribution system should provide uniformly sized material for each such interaction.

4.2.4 Insufficient Energy per Impact

Clearly, if an impact does not occur with sufficient energy to fracture material within the capture volume, the energy of that impact is largely wasted. Some damage may occur that contributes to later fracture and there may be some useful rebound of the balls, but the overall effect is not highly efficient. This loss limits the upper size of feed material particles but, as will be seen in the following section, excessive energy is a far more likely loss when grinding small particles.

4.2.5 Excessive Energy per Impact

The very small capture volume described in Section 4.2.1 means that, for small particles, only a very small energy can be usefully absorbed per impact. Excessive energy, besides adding to power consumption, adds to wear of the balls.

Consider the highly optimistic case in which the entire capture volume is full of uniformly sized particles, and ask, how far must a single ball drop to deliver just the required energy?

The volume of material to be acted upon is the capture volume, V_c, times $1 - \varepsilon$, where ε is the void fraction. For simplicity let us assume that the energy necessary to cause fracture is proportional to the volume of material, in keeping with the Kick theory and the "fourth theory" for particles above the critical size.

The energy required is then

$$W\Delta h = V_c(1 - \varepsilon)E_c$$

where

W = ball weight = $4/3\ \pi w R_b^3$

w = ball weight per unit volume (or net weight if immersed)

Δh = necessary height of fall

E_c = fragmentation energy per unit volume.

Then, with equation (4-3) for V_c,

$$\frac{4}{3}\pi w R_b^3\ \Delta h = 4\pi w R_p^2 R_b (1 - \varepsilon)E_c$$

Solving for Δh,

$$\Delta h = \frac{3E_c(1 - \varepsilon)}{w} \left(\frac{R_p}{R_b}\right)^2 \tag{4-7}$$

E_c is of course the source of much of the confusion in this science, and the simple format chosen here, with E_c independent of size, may not be correct, but this expression can be evaluated to provide an estimate of Δh. Surely it must be easier to fragment a collection of small particles than to fragment the same material in solid form under an indentor. For an indentor penetrating a massive solid sample, the energy per unit volume fragmented, called the specific energy, is typically about half the compressive strength of the material. Thus for oil shale having a compressive strength of 12,000 psi, E_c must be less than about 6,000 in lb/in^3. For a densely packed monolayer of uniform particles, requiring the maximum energy input, the void fraction might be about 0.3. For steel balls then

$$\Delta h = \frac{3 \times 6000(1 - .3)}{.285} \left(\frac{R_p}{R_b}\right)^2$$

$$= 4.42 \times 10^4 \left(\frac{R_p}{R_b}\right)^2 = 4.42 \times 10^4 \left(\frac{D_p}{D_b}\right)^2$$

For 50 μ particles and 100 mm balls,

$$\Delta h = 4.42 \times 10^4 \left(\frac{50 \times 10^{-6}}{100 \times 10^{-3}}\right)^2 = .011 \text{ inch.}$$

The typical ball drop, of the order of one ball radius per impact, is of course much greater than this, about 180 times greater in this case.

The numbers are approximate, and even the format may be questioned, but all assumptions are believed to be in the direction of maximizing the required drop. Hence the message is clear: near the small particle end of the grinding process, excessive energy per impact can cause very low efficiency even if all other features are ideal.

The fact that Δh is proportional to the square of the particle-to-ball radius ratio (in this constant E_c format) also illustrates that for coarse particles insufficient energy can be a problem. For example, if one considers 10 mm feed particles and 100 mm balls, Δh is found to be 442 inches by this formula. Of course, this simple format (or any other for that matter) is not likely to be valid over so broad a size range, particularly with such rough estimates of material properties.

4.3 New Design Paths for Improved Milling Processes

The preceding sections, while certainly not rigorous or precise, are sufficient to point out some of the shortcomings of present practices in fairly basic terms. These in turn suggest new design paths by which departure from continued empirical polishing of the present very inefficient process may prove beneficial. The purpose in this section is to reduce the foregoing observations to concise yet general statements that point out these new paths and, perhaps, suggest new concepts and approaches. They have suggested the concepts to be presented in Section 4.4, but their utility is believed to go beyond those few concepts.

Basically, conventional ball milling is seen to suffer from an indirect energy transfer process which, particularly for small particles, is demonstrably highly inefficient for a number of reasons. Further, it appears that much greater attention must be paid to the correct distribution of the material as it is ground, and to the prompt removal of finished product.*

It is also important to stress in this section, while still generalizing, the great need for a better understanding of actual comminution processes. So long as one is willing to blast particles with orders of magnitude greater energy than required, and to suffer something like 99% losses in the process, then the details of particle fragmentation are unimportant. But if we wish to operate with considerably less excess energy then we need to know with some accuracy just what energy is required in order that we can avoid the losses of either insufficient or excessive energy input.

4.3.1 New Design Paths for Improved Energy Efficiency

With the clear technical design goal of substantially improved energy efficiency, the preceding observations can be reduced to a few rather specific paths to be explored by new design concepts. Following the distinctions of Section 4.2, these thoughts can be divided into "mechanical" and "fluid" mechanisms.

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*Curiously, the preceding sample analyses suggest that regrinding of finished material may not be as important as generally believed. If impact involves only very small volumes of small particles, and the smallest particles are likely to be shielded by larger particles in the capture volume, it is possible that very little regrinding of finished particles ever takes place.

To reduce internal mechanical losses, the designer should strive to:

- Provide direct energy input to the particles being fragmented.
- If energy input must be indirect, as for example input through a grinding medium, then arrange for a non-random transfer of that energy to the material being ground and increase the "capture volume" per interaction.
- Match the energy, however imparted, to the size and properties of particle(s) involved (with some safety factor of course).

To improve the "fluid" behavior the designer should strive to:

- Direct fluids to effectively and promptly remove particles of the desired product size as they are formed.
- Use fluids, in keeping with the third comment above, to match particle sizes to the local grinding action and to minimize size variations at any one location.
- For other than fluid transport means, these same features should be sought.

4.3.2 Other Design Goals

Up to this point this discussion has been concerned only with the energy efficiency of milling processes, as well it should for a process of such low efficiency. It is recognized that a practical concept must fulfill a number of other requirements that are presently satisfied, in varying degrees, by conventional practices and equipment. These include rugged, reliable operation, generally associated with simple design; acceptable size and capital cost for a given capacity; long and economic wear life of component parts and grinding media (if appropriate);

controllability and capacity to accommodate varying feed properties, as discussed for example by Bassarear (Ref. 24) in describing the performance of a range of autogenous mill circuits; and perhaps other criteria.

In Section 4.4, three distinct concepts are presented which attempt to follow some of these design paths. These examples, together with the preceding general statements, may stimulate others to generate additional new, and perhaps better, concepts.

4.4 The Stationary Spiral Ball Mill

4.4.1 Basic Design Aims

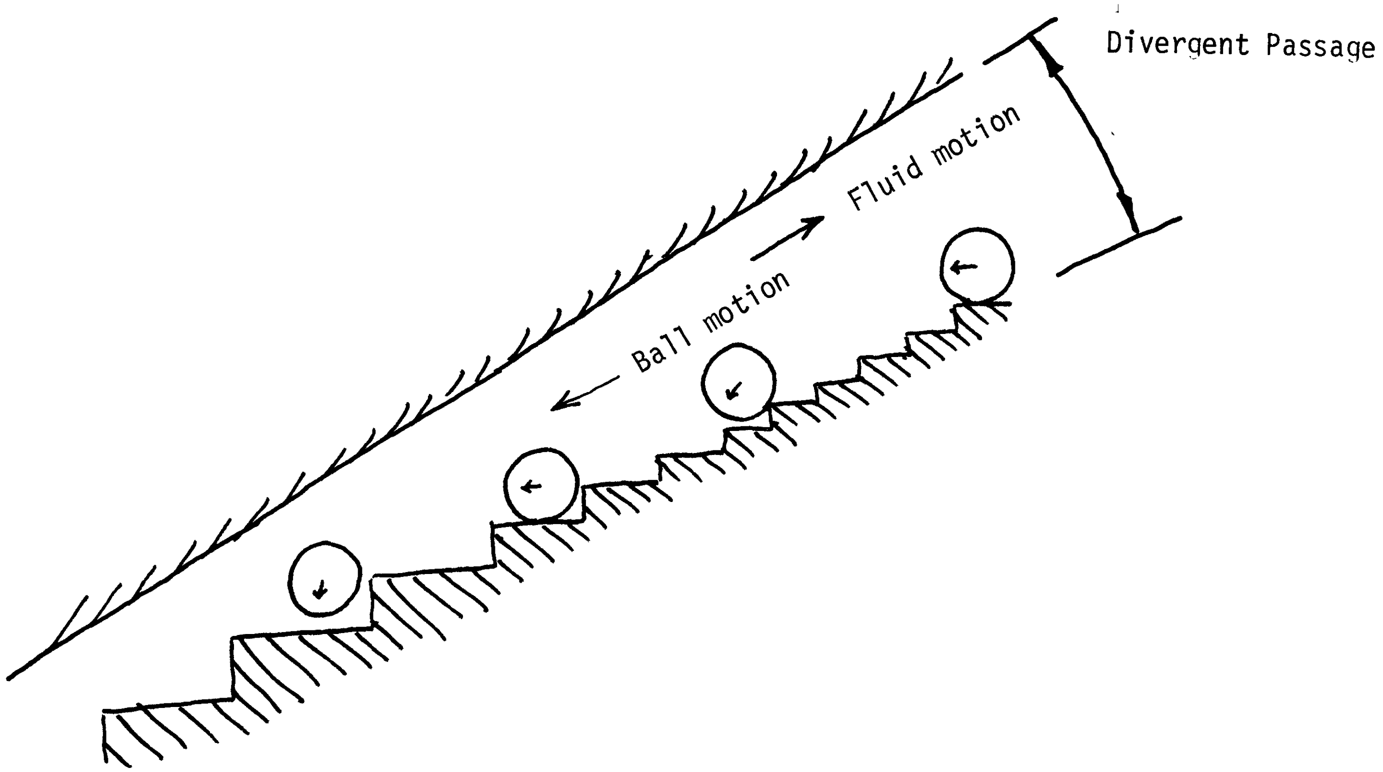
The stationary spiral ball mill utilizes an indirect energy transfer process consisting of conventional balls moving downward through the material being ground, but in all other respects it follows the preceding suggestions. Its design goals include:

- Full and non-random use of the balls.
- Enlarged capture volume.
- Matching of impacts to local energy requirements.
- Prompt removal of finished product.
- Size segregation of material to match local grinding action.

The concept is best illustrated in simple linear form as sketched in Figure 4-3. 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

Figure 4-3

SCHEMATIC LINEAR VERSION OF STATIONARY SPIRAL BALL MILL



requirements and, further, by impacting 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.

4.4.2 Some Design Variables

Design variables of interest in terms of the mechanism details include step sizing, step surface area, and fluid velocity distribution.

Vertical step dimension is what determines the impact energy of the ball and, together with the fluid segregation of particles according to size, offers the opportunity to reduce the losses associated with excessive energy per impact. Preceding calculations based upon conservative (i.e., large) estimates of small particle energy requirements indicate that the steps near the top of the ramp should be very small. For the model selected (constant fragmentation energy per unit particle volume) this calculation indicates the steps should grow in size with the square of the particle diameter as they approach the bottom of the ramp. However, the actual energy requirements and their variation with particle size are precisely the basic information that cannot be learned from existing empirical comminution information. Thus design of these details must await a better understanding of the basic material behavior, including specific tests for the oil shale in question.

High energy efficiency will require that the material layer on the surface of the steps be kept thin, perhaps approaching a monolayer. Thus, as the material is ground to smaller diameter, it must be spread over a larger area. In terms of the impact area concept discussed in Section 4.2.1, the capture volume decreases with the square of particle diameter (Equation 4-6). Thus the number of impacts for a given quantity of material must increase inversely with the square of the particle diameter. If one thinks in terms of a stepped ramp of generally constant slope, this requirement is at least partially provided for. That is, with required step height decreasing with particle size (also with the square of diameter according to the previous model), a great many more steps can be provided per length of ramp near the top in comparison with the bottom. For the spiral (rather than linear) arrangement suggested in Section 4.4.3, additional ramps can also be provided near the top of the mill.

Fluid velocity must continously decrease in order to provide the desired classification of material. Mean fluid velocity, for the linear device of Figure 4-3, can be set simply by proper sizing of the divergent passage. Of course, local perturbations will occur, as where the flow travels over the lip of a step, and the downward motion of the balls constitutes a major perturbation. The latter should be generally helpful in that the resultant turbulence near the step surfaces should prevent stagnation of material outside the impact areas.

For low solids density flow the velocity-transport relationship might be similar to the transport of sediment in river beds, but for high density flow, as anticipated here, transport phenomena will be similar to those in fluid classifiers already in use in the minerals processing industry.

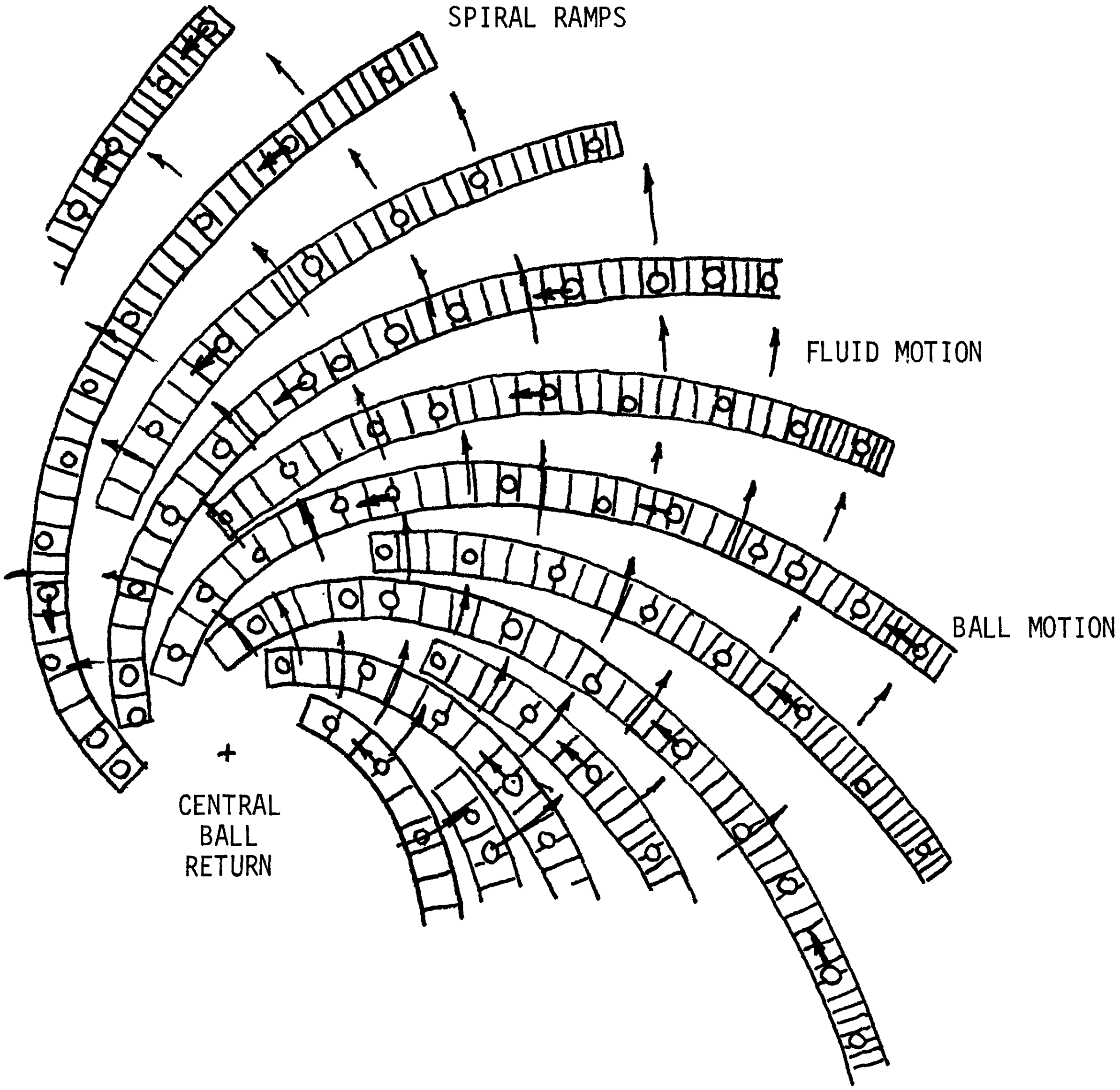
4.4.3 A Suggested Overall Arrangement

In addition to the descent ramp, the mill also needs a ball elevation system, a coarse material feed system, and a fluid circulating system. The linear arrangement of Figure 4-3 is certainly functional, and radial arrangements of such devices utilizing a single central ball elevator would be advantageous. Such an arrangement, for a large number of radial ramps, might appear as a cone, although there would be wasted space between ramps near the periphery.

A spiral arrangement, sketched in Figure 4-4, offers further advantages and is suggested here for further study. In plan view, a series of spiral (rather than radial) ramps is suggested. Spiraling inward at a constant angle to the local radial direction, these ramps would describe log spiral paths. To utilize added space near the periphery, additional ramps could be added, providing additional impact action where needed. These intermediate ramps would end short of the center and balls discharged from them would be fed to the ball elevator without descending to the level of the center.

Figure 4-4

PLAN VIEW - STATIONARY SPIRAL BALL MILL



In cross section the arrangement would appear as a cone, and a conical cover or shroud is provided to confine the outward flowing fluid. Because of the increasing radial flow area, a constant step-to-shroud gap would provide a velocity which decreases linearly with increasing radius, but a contoured shroud could provide any other variation.

A spiraling outward flow at right angles to the inwardly spiraling ramps as shown in the flow pattern is likely to provide a better distribution of material over the impact areas. Incoming fluid having a tangential velocity component will, with the absence of guide vanes, naturally follow a log spiral path--hence the suggested log spiral ramps since they would make a constant angle with the local fluid velocity.

Ridges along the edges of the ramps, to constrain balls within the ramp, might also serve to generate local low velocity zones (like a snow fence) to concentrate material in the paths of the impacting balls. However, these and other local flow disturbances may also cause undesirable stagnation and excessive buildup of material. Careful design and testing would be required if such behavior is troublesome.

4.4.4 Power and Performance Estimates

Power is consumed primarily in elevating balls from their discharge point to the top of the machine, just as it is in the outermost layer of balls in a ball mill. This power is expended, hopefully in useful fashion, as the balls descend the ramps, again just as in a conventional mill, but in this case the impact energies, particle sizes, and particle quantities are much more closely controlled.

Power consumption is equal to the rate of lifting balls which, in turn, is equal to and limited by the rate of ball descent. Ball descent rate down a series of steps can be calculated if one knows or assumes the residual velocity after each impact, but the answer (particularly for small steps) is essentially equal to the assumed residual velocity. As a first estimate for rough sizing purposes, one can simply assume that the descent rate is roughly equal to that experienced in conventional ball mills. From continuity, the descent rate of surface balls in a ball mill must be about equal to the surface speed of the mill, which, for a given mill diameter is well known (typically about 75% of critical speed). Put another way, the power density, say in power per unit surface area, will be about the same for the spiral mill as for balls along the surface of a conventional mill. Thus, if the outer balls in a conventional mill absorb about half the power, then a spiral mill will consume about half the power of a conventional mill of the same size (same plan area, roughly same volume).

The throughput per kilowatt-hour of the spiral mill should be much better than that of a conventional mill because of matched impact energies, better and classified solids distribution, prompt removal of finished product, and larger capture volume per impact. Of these features only the latter can be quantified at this time: because of the flat surface impact, the spiral mill has twice the capture volume per impact.* So long as the impact energy is sufficient (and it is

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*By providing a concave trough-like step surface the multiple can be further increased, up to a theoretical multiple of twice the square root of R_2/R_p for a close fitting 180° trough engagement (about 89 for 50 μ particles and 100 mm balls). Such a full engagement trough would be impractical and would present material distribution problems, but another doubling or so of capture volume in a shallow trough seems entirely reasonable.

excessive), twice the capture volume per impact means simply twice the output per energy expenditure. Hence the spiral mill should be at least twice as effective as the conventional mill.

In summary, for mills of about the same size, the spiral mill should absorb about half the power of a conventional mill while processing at least the same amount of material.

4.5 Autogenous Shear Mill

The autogenous shear mill represents quite a different pursuit of the new design paths of Section 4.3.1, and yet it appears to satisfy most of them as well as some of the "other design goals" of Section 4.3.2.

4.5.1 Basic Design Aims

The design aims include direct energy input, effective and graded removal of small particles as formed, rugged simple design, low wear of component parts, and high power density. These design aims can best be illustrated in terms of the suggested overall design.

4.5.2 Suggested Overall Design

The autogenous shear mill is designed so that virtually all of the input energy is consumed directly by shale fragments shearing against one another. It is "autogenous" in the sense that high forces and most highly loaded relative motion occur between shale fragments, rather than between shale and metal, in an attempt to minimize machine wear. The relative motion is, on a gross scale, the shearing of one mass of granular shale against another similar mass. This may improve efficiency since rock shear strength is typically much less than compressive strength, but on a local scale it is likely that most particle-to-particle interactions are compressive.

The basic concept is shown in cross-section in Figure 4-5. A vaned rotor rotates 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 particles 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. Among the major design features are the following:

1) Above the tops of the vanes the two feed columns are separated by cylindrical shells so that each column can enter its respective vaned zone without rotation relative to the vanes and consequent wear of their upper edges.

2) The rotor and stator cross sections shown in Figure 4-6 provide relatively deep (in the radial direction) pockets to establish solid body rotation (or non-rotation) of each column. Motion relative to the vane surfaces is just the very low velocity downward feed motion of the material.

Figure 4-5

VERTICAL SECTION OF AUTOGENOUS SHEAR MILL

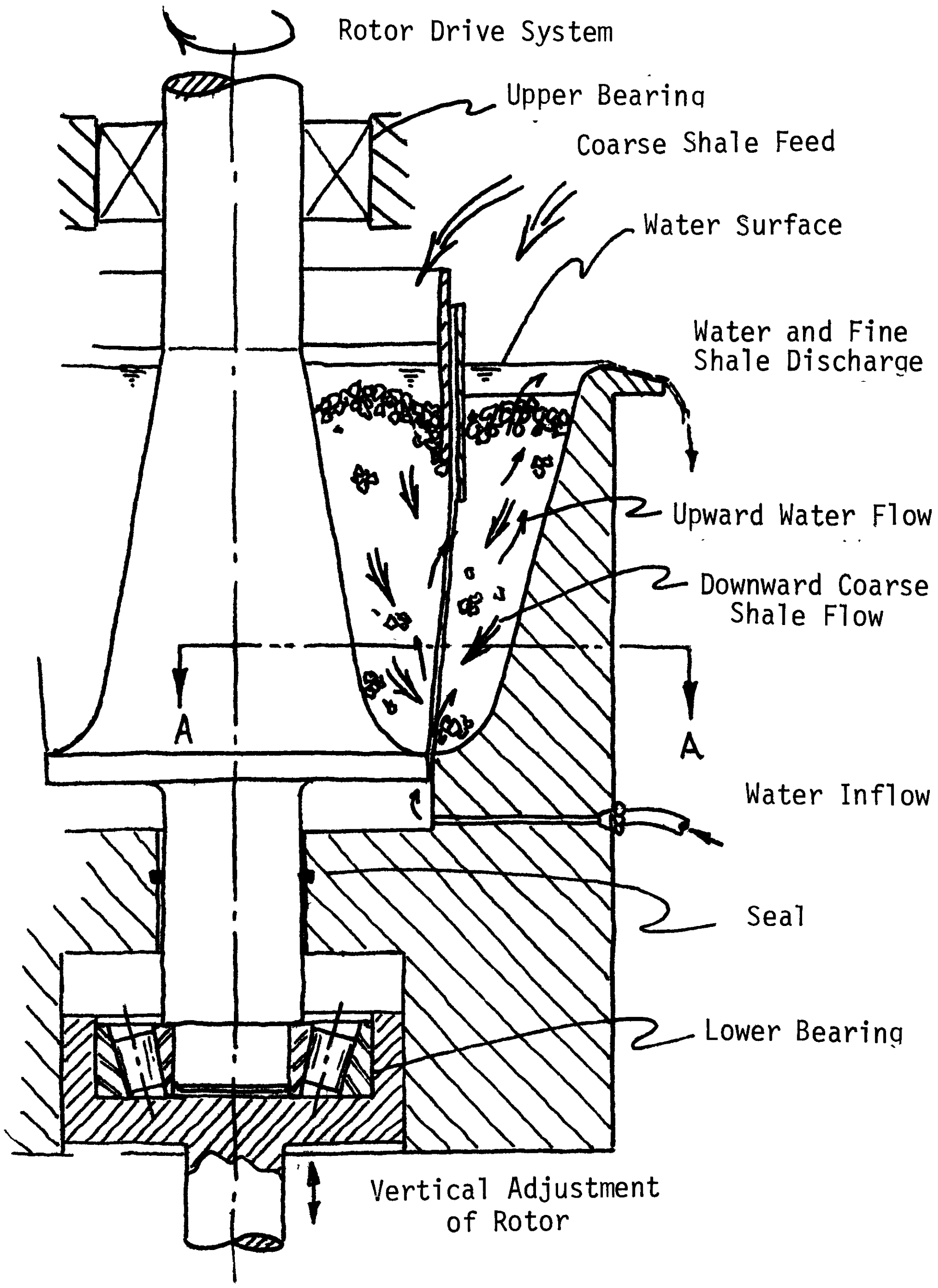
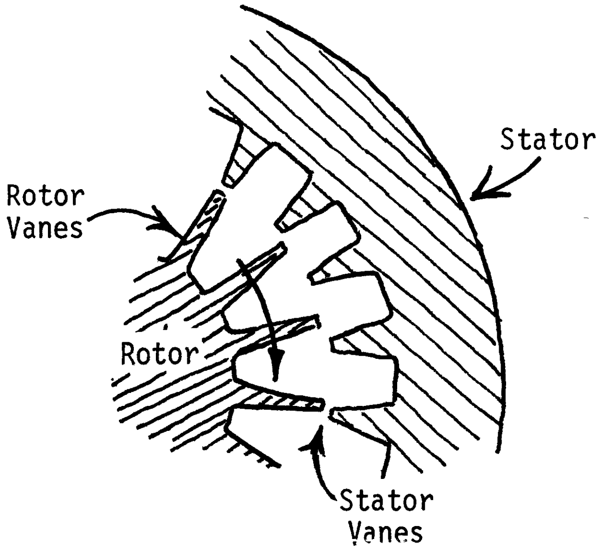


Figure 4-6

ROTOR CROSS SECTION FOR AUTOGENOUS SHEAR MILL



3) Feed material flows downward by gravity acting both on the material within the pockets and the head of material above the vaned portions. It is urged inward to the interface as it progresses downward by the tapered pocket shape.

4) To minimize wear of the vane tips, and to avoid very high local loading, the gap between rotor and stator is equal to or slightly larger than maximum feed size (perhaps about one inch). Wear will inevitably occur, and the vane tips should be hard faced, protected by tungsten carbide inserts, or designed for periodic simple replacement.

5) Crushed material is carried upward from the shear zone and out at the top of the stator by an upward fluid flow. By regulating the flow velocity, and perhaps the concentration of the resultant slurry, the size of material carried from the machine can be controlled. Intermediate sized material will be dropped out within voids in the coarse feed material and carried back to the shear zone by the motion of the latter.

In addition to these basic features, numerous design details can be suggested at this time. It is likely that, whatever the feed size and rotor-to-stator vane gap, bridging will occur frequently, creating local high forces. Obviously the spacing of rotor and stator vanes must not be the same in order to avoid periodic very high torques. Much smoother operation will be assured if one or both sets of vanes are spiraled about the rotation axis to eliminate simultaneous convergence along the length of a vane. Such spiraling can also be used in an auger-like fashion to force feed the material, minimizing or eliminating the need for a head of material above the machine (although variation of such a head may make a convenient means of machine control).

Water should be fed upward at the bottom of the machine in such a way as to flush fragments from the lower bearing and seal area. Recycled relatively dirty water could be used to keep solids out of the general area while clean make-up water could be introduced through a gland to protect the lower seal from all dirt.*

To provide adjustment for the rotor-to-stator vane gap, either for differing feed sizes or to compensate for wear, the two elements can be slightly conical as shown in Figure 4-7. Vertical adjustment of either element thus varies the gap (in a manner now used to vary the throat gap in gyratory and cone crushers).

4.5.3 Power Considerations

Power consumption of the autogenous shear mill may be estimated in terms of the shear strength exhibited by a granular material. When confined by a normal stress N, such material develops a shear strength S. Plotting S against N for a range of crushed rocks yields a linear curve of slope ϕ such that

$$S = N \tan\phi \tag{4-8}$$

and, for the typical case ϕ is about 30° so

$$S = .577N \tag{4-9}$$

Referring to Figure 4-8, normal stress at the cylindrical shear zone is simply the hydrostatic pressure of the column of crushed shale if the vanes are not spiraled. With considerable material vibration it is

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*The lower bearing could be eliminated if the rotor were cantilevered from above, but bearing protection is not that difficult and the lower bearing position does provide a much more rugged design.

Figure 4-7

CONICAL GEOMETRY FOR AUTOGENOUS SHEAR MILL

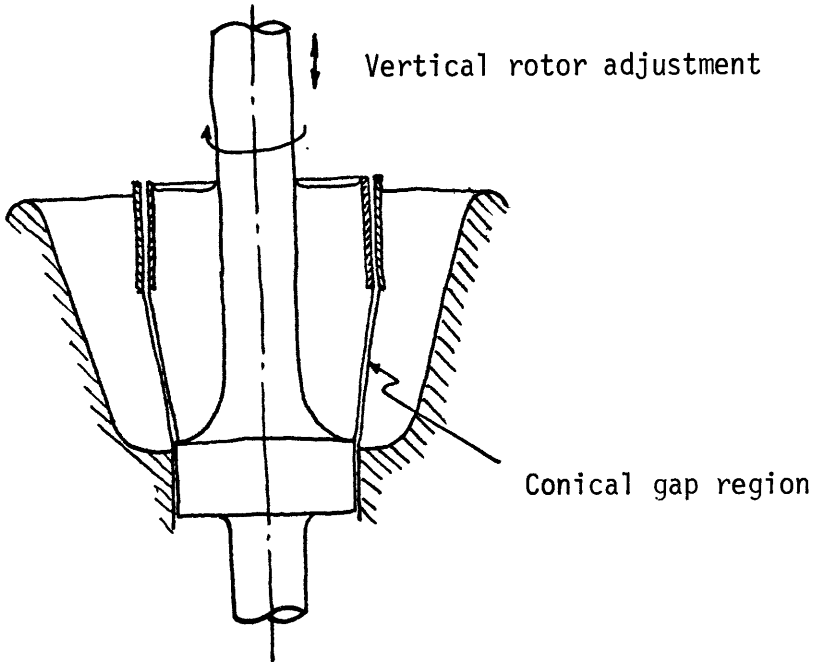
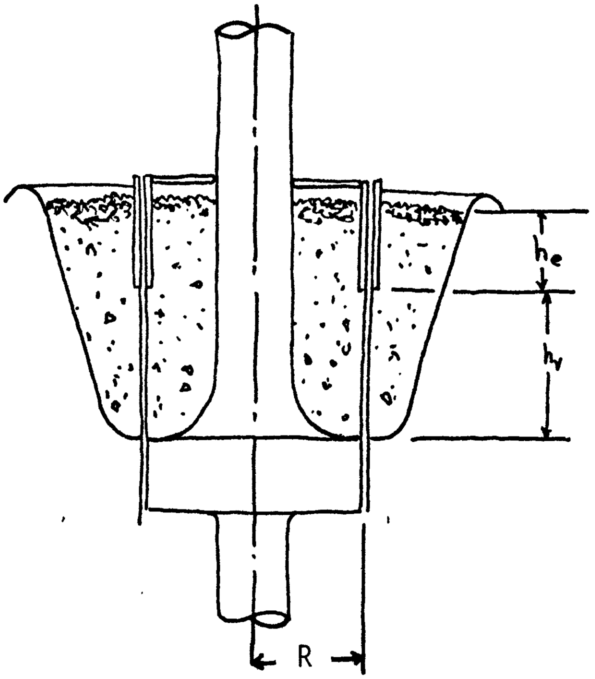


Figure 4-8

AUTOGENOUS SHEAR MILL NOMENCLATURE



reasonable to assume that the crushed material will behave essentially as a fluid in this regard. Then the mean hydrostatic pressure is

$$N = w(\frac{h_v}{2} + h_e) \qquad (4\text{-}10)$$

where N = mean hydrostatic stress at mid-vane height

w = bulk weight per unit volume of feed

h_v = vane height

h_e = external head

The total torque on the rotor then is:

$$\text{Torque} = 2\pi R^2\, h_v\, w(\frac{h_v}{2} + h_e)\tan \phi \qquad (4\text{-}11)$$

Power is given by

$$\text{H.P.} = \frac{\text{Torque (ft-lb) x rpm}}{5252} \qquad (4\text{-}12)$$

At this point it is necessary to estimate what the speed of such a machine might be. Lacking a more specific guideline, it is assumed that the speed might be similar to that of a tunnel borer or large boring tool of the same diameter. As a rough rule of thumb, such machines are limited to 120/D rpm where D is diameter in feet. In effect this rule establishes a maximum tip speed for such boring devices. For our grinder then,

$$\text{rpm} = 60/R \qquad (4\text{-}13)$$

Combining equations (4-9), (4-11), (4-12), and (4-13), and for a bulk weight of 90 lb/ft^3,

$$\text{Power} = 4.15\ Rh_v\ w(\frac{h_v}{2} + h_e) \qquad (4\text{-}14)$$

For a "reasonable" example, consider a mill 10 feet in diameter and 10 feet high. According to the above estimates it would rotate at just 12 rpm and, with no external head, absorb just over 1000 horsepower. With 10 feet of external head it would absorb just over 3000 horsepower. Spiral vanes, which would be used in any case, would increase these powers (or decrease the required external head). Without tests, of course, the rather low speed is only a very rough estimate based upon "similar" rugged service of a not-so-similar device.

Estimation of the throughput of such a device is more difficult, there being no known data on the grinding effectiveness of a shearing interface between two masses of granular material. Qualitatively the autogenous shear mill would seem to be more effective than a conventional ball mill in several categories discussed in Section 4.2. The energy is imparted directly to the shale, avoiding the random loss mechanisms of a ball mill. It can be argued that there can be neither excessive nor insufficient energy losses at each interaction. Insufficient energy losses are avoided because, compared to individual fragment interactions, the mill is a massive and irresistible drive. At the same time, excessive energy cannot be delivered because the energy delivered is only that necessary to overcome resistance to motion. In effect, the machine is a "displacement" mechanism (as contrasted to a "load" mechanism) that delivers just the force needed to produce a fixed displacement and no more. This behavior may be particularly helpful with a somewhat resilient material like oil shale. Similarly, limited target area and insufficient charge per interaction are not loss mechanisms because essentially all of the power input is due only to direct interactions and if there are too few there will be little power input. One loss which can occur despite these arguments is insufficient stress. That is, the coarse fragments may slide and roll over one another in the shear zone without generating sufficient stress to cause fractures. However, this behavior, which would also lead to low power input, can be simply overcome by increasing the external head on the machine, thereby increasing the loading in the shear zone. With all these arguments relative to the selected losses within a ball mill it seems reasonable to assume that the autogenous shear mill will be at least twice as efficient as a ball mill.

The autogenous shear mill is not unlike a gyratory crusher in general configuration. It is a somewhat more complex shape but not subject to the enormous loads of the latter. Furthermore, it uses about the same power: a gyratory crusher having 5-foot feed openings (i.e., over 10 foot inlet diameter) and an 8-foot mantle diameter is listed at 1000 horsepower. However the autogenous shear mill has relatively low speed power, thus requiring a stronger drive train. It is estimated, therefore, that the autogenous shear mill will be about twice as costly per horsepower as a gyratory crusher. In summary then, the autogenous shear mill is estimated to be about twice as productive per horsepower as a ball mill, and twice as costly per horsepower as a gyratory mill.

4.6 Pneumatic Impact Mill

4.6.1 Basic Design Aims

The pneumatic impact mill employs a gas stream to accelerate solid particles and impact them against a solid target, offering the ability to closely match particle energy to fracture requirements. The basic design aims are then direct energy input to the particles; energy matched to particle requirements; selective and prompt discharge of final particles; and low wear rate of components. Pneumatic conveying of material permits relatively simple separation according to size and, hence, the opportunity to treat rather narrow size ranges in individual stages of a multi-stage process. On the other hand, if the simple constant energy per unit volume model used in Section 4.2.5 is correct, then the required impact velocity is independent of size and the only sizing necessary will be to remove finished particles while recycling all oversize. That is

$$\frac{4}{3}\pi\rho R_p^3 \frac{V_{imp}^2}{2} = \frac{4}{3}\pi R_p^3 E_c \qquad (4\text{-}15)$$

where V_{imp} = impact velocity to fracture

ρ = particle density.

In terms of the loss mechanisms discussed in Section 4.2, this approach offers great potential, but it may introduce a new loss mechanism of its own. It is possible that an appreciable fraction of solid particles will not impact effectively against a solid target. Particularly for very small particles, individual impacts may be cushioned by the gas, and some particles may simply follow fluid streamlines and miss the impact target. Since small particles are of major concern, this may be a serious limitation of this approach.

Energy input to the gas (air or perhaps steam) can be efficiently accomplished and this can be converted directly into particle kinetic energy in a simple nozzle. Energy will be wasted in discharge kinetic energy of the gas, but this should be an acceptable loss since the mass flow of gas can be low relative to the solids mass flow.

As an added advantage in some situations, milling will be a dry process so that dewatering will not be necessary if a dry separation process is to be used. Also, for oil shale, fracturing of individual and separate particles should avoid smearing of kerogen onto mineral particles.

4.6.2 Some Suggested Design Details

The overall arrangement consists of nozzles directing jets at solid targets, particle pickup systems such as vacuum collectors, and pneumatic separation means such as cyclone separators. All are relatively simple and well known devices. To assure reasonable component life it will be necessary to avoid wear of machine components. Nozzles can be designed using peripheral clean gas cushions to minimize particle-to-wall contact in high velocity areas. The target itself, an unavoidably high wear area, can be simply coarse oil shale fragments, in which case target wear particles become useful product. For example, a moving conveyor loaded with shale can provide a continuously renewed target surface. Like the preceding concept, the device is an "autogenous" mill in this respect.

4.6.3 Power Considerations

More than either of the other concepts, the pneumatic impact mill can match the energy requirements of any particle size. Consequently it offers the potential of quite high efficiency, but, by the same reasoning, estimation of its performance demands knowledge of the actual energy requirement. Therefore, because any estimate at this time would require an assumption of this unknown requirement, and also because there is serious doubt that the concept will function for the very small particles of ultimate interest, this approach will not be examined further at this time.

4.7 Basic Research Requirements

The foregoing discussions all point to the necessity of basic research to establish the actual minimum energy requirements for comminution of oil shale and for many other materials as well. In simple terms, if we wish to avoid expenditure of excessive energy in comminution then we must know, with some accuracy, how much energy must be expended. The need for basic understanding of the comminution process goes beyond merely setting goals for purposes of performance evaluation: it includes basic understanding of the physics of the material behavior that can be expected to suggest new and substantially improved means of comminution. Such an understanding would be of value for all conceivable comminution devices in the sense that the information needed is material behavior, not device behavior.

The autogenous shear mill concept of shearing a coarse granular material against itself should be investigated. The configuration proposed seems reasonable, and in fact it would seem to be a convenient geometry for experimental studies. The apparent advantages of the basic concept justify a basic study of the phenomenon whether or not this particular design concept is the best.

Comminution research should proceed in close cooperation with separation studies for the products of comminution. It is quite likely that comminution processes can substantially affect downstream separation behavior. Of particular interest for oil shale beneficiation, smearing of the kerogen on mineral particles can obviously confuse the separation issues. Thus a careful characterization of the various comminution processes with respect to the liberation of minerals and with respect to subsequent separation processes should proceed in parallel with both basic comminution studies and separation studies.

4.8 Process Design and Cost

The engineering and economic analysis of novel comminution equipment will be limited to the stationary spiral and autogenous shear mills because the pneumatic impact mill is too conceptual at this point to allow even preliminary equipment design and costing.

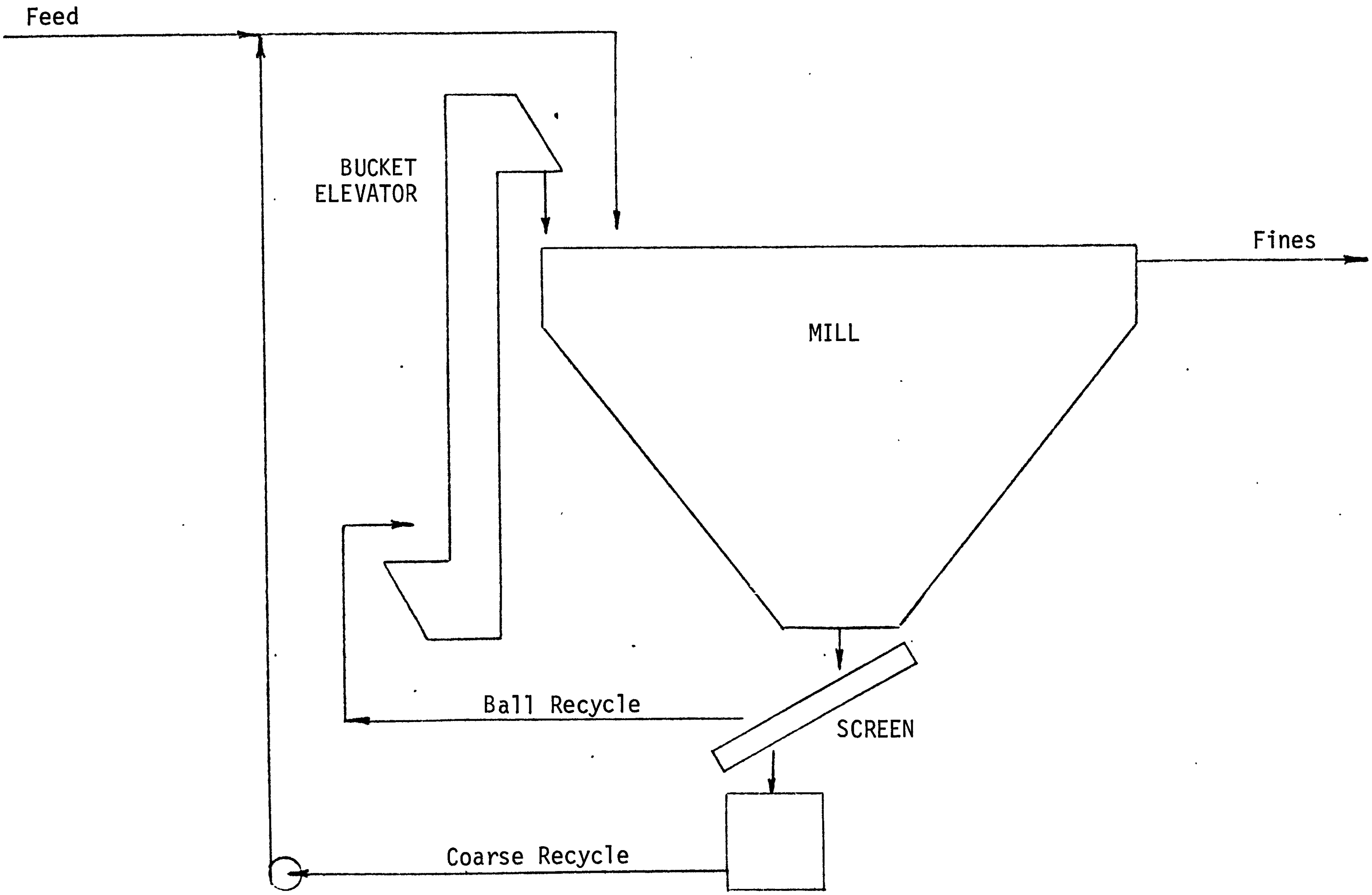
4.8.1 Stationary Spiral Mill

A conceptual equipment design indicated that it is possible to build Stationary Spiral Mills (SSM) at a size that would allow the replacement of ball mills at least on a one-for-one basis. The diameter and height of the unit would be within the 24' to 28' and 28' to 36' ranges, respectively. The equipment would have to be field assembled because of the large diameter of the housing.

Ancillary equipment would include a screen to separate the balls from the coarse shale fraction, a bucket elevator for the balls, and a slurry tank and pump to recycle the coarse fraction (see Figure 4-9). The screen might be an integral part of the mill. The other process

FIGURE 4-9

STATIONARY SPIRAL MILL



equipment of the SSM grinding section would be the same as that for ball mill grinding (Figure 3-2).

The SSM cost should be substantially lower than that of a ball mill because the former is essentially an immobile vessel with complex internals. The cost of the ancillary equipment including drives should be only a fraction of the SSM proper. On the other hand, a ball mill is a rotating body that, while simple, has to have heavy walls to withstand the constant impact of tumbling balls. The electric motors of the SSM ancillaries should be at least one half the size of those for a ball mill.

An analysis of the approximate range of SSM costs was made based on costs of similar equipment such as spiral gravity concentrators, jigs, and various solid blenders with relatively light internal moving parts. The costing of the ancillaries was straightforward. The ASPEN Cost Data Bank (Ref. 54) and Mullar (Ref. 74) were used for the estimates. The analysis indicated that the SSM cost including ancillaries should be at least one half that of the ball mill with the same capacity of 6250 tons/day dry shale.

4.8.2 Autogenous Shear Mill

The conceptual equipment design of the Autogenous Shear Mill (ASM) was done based on a large-size (10' diameter x 10' long) gyratory crusher made by Allis-Chalmers (Ref. 55). An ASM of this size would require a 1,000 kW motor. To take advantage of the economy of scale, the 1,000 HP ASM was scaled-up to a volume not exceeding the ball mills used in the Base Case with beneficiation.

The Base Case requires 26,400 HP per train per grinding stage for a 4-train plant (see Table 3-3). As the ASM should use at the most 50% of the power of a ball mill, the scale-up factor is 0.5 x 26,400/1,000 = 13.2. Assuming that power is proportional to the rotated volume, the diameter and height of one scaled-up ASM should be 20' to 24' and 24' to 33', respectively, driven by two 6,600 kW motors. Four units of this size would be used per stage and would not require any ancillary equipment. The other process equipment can be assumed unaffected by the change from three ball mills to one ASM per train.

4.8.3 Results and Conclusions

Capital costs and power consumption for three comminution options--the Base Case ball milling, the SSM, and the ASM--are summarized in the following table:

Table 4-1

Comminution Options

Option	Capital Cost, $M	Power Consumption, MW
Base Case: Ball Milling	350	180
Stationary Spiral Mill	170	90
Autogenous Shear Mill	680	90

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.

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.

5. SEPARATION ALTERNATIVES

The "Base Case" beneficiation systems described in Section 3 assume that froth flotation--of the type described in Refs. 8 and 13--would be used to separate kerogen and mineral particles ground to liberation size by ball milling. However, froth flotation is only one of several processes that are technically possible to bring about a high-enrichment high-recovery separation. This section describes the processes that we have been able to identify.

5.1 The Basis for Oil Shale Separations

In principle kerogen can be separated from oil shale using any physical property for which kerogen and the mineral have different values. Since there are many oil shale minerals, each with different physical properties, it is necessary to consider only the major mineral components. For Western U.S. oil shales, the major mineral components are dolomite, calcite, and clay minerals; for Eastern U.S. oil shales, the major mineral components are clays and quartz.

Most physical separation methods work best when the oil shale has been comminuted to liberation sizes. That is, when the powder consists of individual mineral particles and individual kerogen particles. For Western and Eastern oil shales the liberation size is on the order of 10 μm (Ref. 25). Due to the plasticity of kerogen, true liberation is probably not attainable with present grinding methods because kerogen will undoubtedly smear on mineral surfaces. For this reason as well as the cost of grinding to ~10 μm, most separation experiments have been performed on larger-sized oil shale powders. As a result, a particular separation is limited by the degree of liberation of the feed oil shale.

Each separation method can be categorized by the physical property upon which the separation is based. The following categories are described: density, wettability, solubility, electrostatic, magnetic, optical, friability, bioleach, and stickiness. This list reflects minerals processing technology and is by no means complete. Further research on the physical properties of kerogen and oil shale minerals would be useful to identify potential new separation techniques.

5.1.1 Density Separations

In principle, individual particles of kerogen (specific gravity 1.07) can be separated from individual particles of mineral (specific gravity 2.2-2.9) by placing the ground oil shale in a heavy medium, a fluid with a specific gravity between 1.07 and 2.2. In the heavy medium, the kerogen particles will float and the mineral particles will sink, facilitating a separation. At the small liberation size required for oil shale, the terminal settling velocity is very low. To improve the rate of disengagement, gravity can be replaced by a larger centrifugal force. The various density separation methods are reviewed below. These methods are not capable of the high-enrichment separations sought in this study but they might be useful pre-grinding steps under some circumstances.

5.1.1.1 Gravity Settling

Knowles (Ref. 26) performed gravity separations by allowing a 30.7 gallons per ton (gpt) oil shale ground to -37 μm to settle for 3 days in liquids with various specific gravities ranging from 1.18 to 2.14. The highest enrichment obtained was with specific gravity 2.14 recovering 77% of the kerogen in a 43 gpt concentrate.

Other settling experiments were performed by Larson et al. (Ref. 3). These authors ground a 25 gpt oil shale sample to either - 3 + 1/4 inch or -3/4 + 1/4 inch. Each grind was subjected to a five-stage sink-float series consisting of specific gravity stages of 1.80, 1.95, 2.10, 2.25, and 2.40. The sink material from the previous (lower specific gravity stage) was sent to the next (higher specific gravity) stage. For the 3-inch grind, 65% of the oil shale averaging 18 gpt was collected in the float from two consecutive stages (i.e., 2.10 and 2.25). Grinding to -3/4 inch gave only 50% of the oil shale averaging 17.5 gpt in the float from those two consecutive stages. A typical result is tabulated below for the -3/4 + 1/4 inch grind:

Float	Recovery (%)	GPT
1.80	2.74	70.4
1.95	13.90	36.0
2.10	31.38	29.0
2.25	45.33	17.4
2.40	6.25	10.7
Sink 2.40	0.39	0.4
	Average:	25

A high enrichment and a low recovery are observed at 1.80. The enrichment decreases as the recovery increases for higher specific gravities. Other oil shale samples with grades of 23, 30, and 35 gpt were examined with generally similar results.

Kaczynski (Ref. 27) performed similar experiments on a 13.8% kerogen oil shale with nine size fractions covering a range of -1.5 inch to

+200 mesh. The smaller size fraction gave better separations below a specific gravity of 2.3, indicating that improved liberation improves separation. Recovery of 90% of the kerogen with an ore rejection of 50% was achieved at a sizes smaller than 28 mesh.

5.1.1.2 Centrifugation

Knowles (Ref. 26) centrifuged two oil shale samples: one 52 gpt and the other 12.5 gpt. Each of these shales was ground to +840-590 μm and +250 - 180 μm and centrifuged with liquids of specific gravity 1.8, 2.1, or 2.35. For the 52 gpt shale, the large grind gave very little enrichment for all liquids except specific gravity 1.8. For this liquid 45 percent of the kerogen was recovered in a 71 gpt concentrate. The smaller grind gave similar results. For the 12.5 gpt shale, the large grind gave very little enrichment for all liquids except specific gravity 2.35. For this liquid 56% of the kerogen was recovered in a 15 gpt concentrate. The smaller grind gave similar results for specific gravities 1.8 and 2.1. For specific gravity 2.35, 52% of the kerogen was recovered in a 19 gpt concentrate.

Thomas and Lorenz (Ref. 28) used a centrifuge to investigate how kerogen is bound to the oil shale minerals. They pretreated oil shales with acid to dissolve carbonates, and sodium hydroxide to dissolve the clay minerals. Their results show that only iron oxides are closely associated with kerogen.

5.1.1.3 Hydrocyclone

Roberts and Schaefer Resource Services, Inc. (Ref. 4) performed heavy medium cyclone separations on a 13 gpt oil shale using various specific gravities from 1.98 to 2.39. Their results are tabulated below. The highest recovery was 82% giving a low-grade 18.5 gpt concentrate. The highest-grade concentrate was 37.6 gpt at a low recovery of 19.2%.

Specific Gravity	Recovery (%)	GPT
1.98	19.2	37.6
2.10	42.9	31.9
2.22	57.7	27.1
2.25	64.8	24.9
2.39	82.0	18.5

Lopachenok et al. (Ref. 29) performed experiments on "Shantsy" region shale using an 0.35-meter hydrocyclone. The authors evaluated the efficiency of size classification for various operating conditions. No data were reported for kerogen enrichment or recovery.

5.1.2 Wettability Separations

Kerogen has very different surface properties from the oil shale minerals. Kerogen is oil wetting while the minerals are water wetting. Wettability difference forms the basis for three separations methods: (1) froth flotation, (2) selective shear aggregation, and (3) direct pelletization. The fundamentals of these separation methods are that kerogen particles will be collected at either a water/gas or water/oil interface, while the mineral particles will remain in the water. The small particle size required for kerogen liberation contributes to gangue entrainment which decreases the grade of the concentrate.

5.1.2.1 Froth Flotation

Froth flotation of oil shale has a long history dating from a patent by S. Dolbear in 1924 (Ref. 30). This patent does not give quantitative data on kerogen recovery and concentrate grade, but does provide a description of the process which is substantially unchanged. In this separation method oil shale is pulverized and made into a water slurry. The slurry is aerated forming a froth. The froth is collected and retorted to produce shale oil. The minerals remain in the water slurry and are discarded. Two flotation additives are generally used. One called a frother is used to produce a stable froth. The other called a collector is used to increase the hydrophobicity of the kerogen particles. Typical frothers are pine oil and alcohols while typical kerogen collectors are alkanes and medium boiling point shale oil fractions.

Kaczynski (Ref. 27) and Knowles (Ref. 26) performed flotation experiments on oil shale slurries. Each used an ionic surfactant as a flotation collector and observed non-selective separations. The ionic surfactant adsorbs on the kerogen surface and renders it hydrophilic. Hydrophilic kerogen particles are not collected at the water/gas interface and remain in the slurry like the mineral particles. For this reason using the proper flotation collector is very important.

A description of the froth flotation process for oil shale is given by Fahlstrom (Ref. 8). The author suggests that flotation should be carried out at 5-15% solids by weight ground to 80% passing 15-20 μm for a time of 3-10 minutes depending on the oil shale. A medium boiling shale oil or amyl alcohol should be used as a flotation collector in conjunction with a "light" frother. The paper does not give any specific results but suggests that a 85-95% rejection of ash-forming minerals can be expected. For a Western oil shale with 16.3% kerogen, a 90% kerogen recovery can be expected giving a 56% kerogen concentrate corresponding to an enrichment ratio of 3.4. From mass balance considerations Fahlstrom asserts that the kerogen content of the feed influences the percent rejection of ash-forming minerals. Thus a higher-grade shale fed to flotation will give a lower percent rejection of ash at a constant kerogen recovery.

Hanna and Rampacek (Ref. 31) give results of froth flotation tests performed on finely ground Eastern and Western oil shales. A Wyoming oil shale with 60 gpt was concentrated to 80 gpt with a 93% recovery of kerogen. An Alabama oil shale with 10 gpt was concentrated to 18 gpt with a 96.6% recovery of kerogen.

Rosar et al. (Ref. 32) in a United States patent describe a froth flotation separation of oil shale rich in inorganic sodium compounds, principally nahcolite, dawsonite, and trona. In this disclosure the ore is ground to liberation and slurried in a sodium carbonate/bicarbonate brine. After aeration the organic-rich fraction of the ore is recovered in the float portion and the sodium compounds are recovered in the non-float portion. Details of a flotation series are disclosed where rougher, cleaning, and scavenger flotations are performed. Three examples are given with sufficient data to be useful. These examples are tabulated below.

Feed (gpt)	Recovery (%)	Concentrate (gpt)
7.6	78.8	15.9
13.6	66.3	26.8
18.4	59.2	43.1

Each example shows a high kerogen recovery with enrichment ratios on the order of two.

A current DOE contract with SRI International entitled, "Concentrating Oil Shale by Froth Flotation" is being performed by G. Krishnan; only preliminary results are yet available (Ref. 13). To date some froth flotation experiments have been carried out on both Eastern and Western oil shale. For a 13.3% kerogen Western oil, shale fine grinding to 10-15 μm was necessary to liberate the kerogen. A small amount of pine oil used as a frother-collector agent increased recovery from 70 to 90 %. The concentrates obtained from these flotation experiments had a grade of 20% kerogen. For an 8% kerogen Eastern oil shale fine grinding to 98% less than 44 μm was required for liberation. Pine oil used as a frother-collector agent increased the kerogen recovery from 78 to 92%. But this increase was associated with a decrease in concentrate grade from 19 to 16.4% kerogen. Regrinding the 19% kerogen concentrate for 3-1/2 hours gave a 30% kerogen concentrate upon flotation with a 90% recovery of the kerogen. Further regrinding and reflotation yielded a 37% kerogen concentrate with an overall recovery of 62%. This corresponds to an enrichment ratio of 4.5.

This completes the review of the flotation literature available in English. Chemical Abstracts provide several references in Russian concerning the flotation of oil shales. In the Soviet Union flotation concentrates are used to fire furnaces and boilers. English translations of two papers by Lopachenok (Ref. 33,34) have been obtained. Unfortunately, these papers do not give any information on grade and recovery. Other papers by Proskuryakov, et al. (Ref. 35 to 39) are not available in English. Only their abstracts have been translated. These abstracts are reviewed below.

Grind	Recovery	Grade	Source
-75-200 mesh	80%	62-75%	Proskuryakov, et al. (Ref. 37)
-200-325 mesh	70%	72-82% (rougher)	Proskuryakov, et al. (Ref. 38)
		85-90% (cleaner)	

A grind of -70 mesh was adequate to liberate the majority of organic particles while grinds of 200 to 325 mesh were necessary to produce a high-grade concentrate. The relationship between grade and recovery was typical (i.e., grade increased as recovery dropped off). A variety of collectors were used in the flotation process. The most popular were either a crude shale oil with a boiling point between 200 and 300°C, or a pine oil. Sodium silicate was used to depress silica flotation. Flotation was performed in slightly basic pH. Pulp densities between 15-35% solids were used. Concentrate grade increased as the pulp density decreased.

5.1.2.2 Selective Shear Aggregation

Selective shear aggregation is performed by subjecting a water slurry of finely ground oil shale to a low rate of shear. The low shear rate forces hydrophobic kerogen particles together with sufficient energy to squeeze the water out from between the particles causing aggregation. The water wetting mineral particles require a higher collision force to aggregate since water has an affinity for the particles. Once aggregated, the kerogen particles can be collected by various means.

Hanna and Rampacek (Ref. 31) suggest that this technique, also called selective flocculation, can be used for oil shale separations. No data, however, were provided.

Ring (Ref. 40) described a similar process where the selective shear aggregation was performed in a high specific gravity salt solution instead of water. As the kerogen aggregates formed in the shear field, they floated to the top of the vessel. Due to the 2-50 mm size of the aggregates, their terminal velocity was larger than the terminal velocity of individual kerogen particles. Experiments with a Western oil shale ground to 90% under 5 μm suspended in a 1.4 specific gravity salt solution showed an enrichment ratio of three.

5.1.2.3 Direct Pelletization

Direct pelletization consists of ball mill grinding oil shale in a mixture of water and oil. The results of this process are paste-like organic pellets which contain kerogen, as well as a water slurry of the minerals.

Quass (Ref. 41) ground a South African torbanite with water in a porcelain ball mill. Oil (unspecified) was added in sufficient quantity to form a paste with the kerogen and grinding was continued. Mineral matter became suspended in the aqueous phase and was discarded. In this process the ash content of the oil shale was reduced from 40% to 10%. Down and Himus (Ref. 42) used a similar technique to study the chemical composition of kerogen.

Himus and Basak (Ref. 43) ground a New Brunswick oil shale in a heavy gas oil. Water was added and grinding continued for 16 hours. The ash content of the oil shale was reduced from 58 to 34%.

Reisberg (Ref. 11) added 400-800 ml of water, 10 lbs of grinding media, 10-200 gm of -100 mesh oil shale and 50-100 ml of heptane to a 5.5 gal ball mill for 1 hr. After 1 hr the aqueous slurry was removed and replaced with fresh water. A small sample of the organic phase was taken and the milling operation repeated as many times as necessary. Too little heptane made the organic phase difficult to separate. Too much heptane formed voluminous amounts of organic phase which entrained gangue. The optimum conditions gave pellets ~1 cm in diameter. The resulting grade after each cycle for a 15% kerogen oil shale is tabulated below.

Cycle	Grade (% kerogen)
1	64
2	69
3	75
4	81

The grade improves most drastically for the first cycle, less for subsequent cycles.

Smith and Higby (Ref. 44) treated a Western oil shale with 5% acetic acid to remove the carbonate minerals prior to grinding in a water octane mixture. The aqueous mineral slurry was removed and replaced repeatedly with fresh water until no further mineral matter was observed. In this process the mineral content of the oil shale was reduced from 75 to 16%.

5.1.3 Solubility Separations

The two constituents of oil shale are solubilized by different reagents. The minerals are typically soluble in acid solutions while the kerogen is soluble to small varying degrees in organic solvents. From these physical properties, two types of separation are possible: (1) solution of kerogen and (2) solution of minerals.

5.1.3.1 Solution of Minerals

Down (Ref. 45) describes an analytical technique to obtain nearly pure kerogen samples from oil shales. In this work five oil shales were treated with a three-step acid leach which included:

1) 5N HCl at its boiling point for 2 hrs
2) HNO_3 (specific gravity 1.12) at 25° C for 100 hrs
3) 5N HCl + HF at 25°C for 1 hr.

The results of this treatment gave ash rejections greater than 92% for all five oil shale samples. The details of a similar analytical technique are described by Guthrie (Ref. 46).

5.1.3.2 Solution of Kerogen

Guthrie (Ref. 46) digested two Western oil shales for 24 hours in the following solvents at their boiling points: ethanol, methanol, ethyl ether, chloroform, carbon disulfide, carbon tetrachloride, benzol, gasoline, turpentine, and pyridine. All of the solvents were nearly inert except pyridine which dissolved 30% of the kerogen. Results of similar experiments are summarized by Baughman (Ref. 47) and Williamson (Ref. 10). No solvent listed in these references dissolved more than 30% of the kerogen in any oil shale at sub-decomposition temperatures.

5.1.4 Electrostatic Separations

Ground oil shale is spread on a grounded metal rotor. In one area of the rotor the oil shale is subjected to a corona discharge. The corona charges all of the particles. Relatively non-conducting kerogen particles discharge slowly and stick to the rotor. At another location on the rotor an A.C. corona discharges the kerogen particles and they fall into a hopper. Mineral particles, which are better conductors, discharge quickly and follow a free fall trajectory into a separate hopper.

Ring (Ref. 48) evaluated a Karpco electrostatic separator for the separation of oil shale. Various size fractions of Western oil shale were separated giving nonselective separations. When the dust particles were removed from the ground shale enrichment ratios of ~1.4 were observed.

5.1.5 Magnetic Separations

Ground shale flows down a chute through a magnetic field. Particles with sufficiently high magnetic susceptibilities move into the magnetic field and are concentrated. Preliminary experiments at M.I.T.'s Magnet Laboratory by Kelland (Ref. 49) using a +355-606 μm fraction Western oil shale gave an enrichment ratio of ~1.3.

5.1.6 Optical Separations

A single layer of crushed shale is dropped in front of an array of photoelectric detectors. When a kerogen-rich particle is "seen" by a detector a jet is engaged and the particle is directed into a separate concentrate hopperate.

Occidental Research Corporation has patented the "Oxylore" process (Ref. 53) based on optical sorting of shale particles labelled with a surface-active fluorescent dye. The unit has a 20-inch wide rock curtain and an array of 40 ultraviolet light sources, photoelectric detectors, and water jets. Each jet may be activated as many as 50 times/second. In tests with a -2 + 1 inch 14 gpt oil shale, a recovery of 58% of the kerogen was obtained in a 21 gpt concentrate. This process has been successfully piloted at rates up to 150 tons/hr for limestone sorting.

5.1.7 Stickiness Separations

Brison and Tangle (Ref. 50) describe a separation procedure where different materials absorb different amounts of heat from a radiant energy source. Impingement on a heat-sensitive surface results in only the hot particles sticking to the surface. This process has not yet been evaluated for oil shale separations. The most likely reason is that the liberation size for oil shale is too small.

5.1.8 Friability Separations

It is well known that a small degree of oil shale beneficiation can be obtained by selective sizing. When oil shale is ground, the small-size fractions are leaner than the large-size fractions. This behavior is due to the preferential crushing of the more friable lean material. The Bureau of Mines suggests that a 1 to 4 gpt enrichment can be obtained with proper design of crushing and screening circuits.

Fishback and Petticrew (Ref. 51) describe a conceptually similar separation process in a patent assigned to the Superior Oil Company. This process subjects oil shale to an agitated aqueous medium where a portion of the clay mineral is disintegrated and flushed away. In this process a 30.6 gpt oil shale ground to -3 inch was concentrated to 38 gpt with a recovery of 92.7% of the kerogen.

5.1.9 Bioleach Separations

Bioleaching of oil shale has been investigated by Meyer and Yen (Ref. 57). The bacterium "Thiobacillus spp" was used to oxidize sulfur to produce sulfuric acid which dissolved up to 98% of the dolomite and calcite in the shale. Overall weight losses up to 40% are achieved when oil shale is leached for 14 days. Since the bacteria oxidize sulfur, the possibility exists for kerogen desulfurization.

5.1.10 Other Types of Separation

TRW Energy Systems is developing a proprietary oil shale separation process (Ref. 1). The process is described as one that is of the chemical/physical type which uses a single liquid to liberate the kerogen from a 10 mesh feed. The process is not solvent extraction, acid leaching, or froth flotation. The resulting product is described as a semi-solid of nearly pure kerogen. The first scale-up from laboratory-scale operations is to be initiated.

5.2 Comparison of the Methods of Separation

The literature reviewed in the previous section presents a very complex picture of oil shale separations. There are two reasons for this. First, each worker uses a different grade of oil shale in his experiments. Generally, these grades are usually from two categories--Eastern oil shale at ~10 gpt and Western oil shale at ~35 gpt. The other reason for the complex picture is that three terminologies are used to describe the separation: (1) recovery and grade, (2) enrichment ratio, and (3) ash rejection.

Part of the problem can be eliminated if a common terminology is used to describe the separation process. However, this does not allow comparisons between different grades of oil shale fed to the various separation processes. When "separation efficiency" is used to describe the separation process a consistent comparison can be made for all grades of oil shale feed.

5.2.1 Separation Efficiency

The separation efficiency (η) is defined as (Ref. 52)

$$\eta = R_1 - R_2$$

where R_1 and R_2 are the percentages of kerogen and minerals recovered in the beneficiated product, respectively. A conversion of the enrichment ratios used in the literature review and separation efficiency is tabulated below:

	Separation Efficiency (%)*	
Enrichment ratio	35 gpt	10 gpt
4	84	70
3	74	61
2	53	43

*Assumes 90% recovery of kerogen.

High enrichment ratios correspond to high separation efficiencies for the 35 gpt feed. For a particular enrichment, a lower separation efficiency is required for the 10 gpt feed.

5.2.2 Separation Efficiency Comparison

The separation efficiencies of the separations described in the literature review have been calculated where possible. The highest and lowest values reported are listed in Table 5-1. From this table it can be seen that two classes of separation exist: (1) high enrichment separation with $\eta > 66\%$ and (2) low enrichment separation with $\eta < 50\%$. In the high enrichment category are: froth flotation, selective shear aggregation, direct pelletization, and acid solubility.

Table 5-1

Oil Shale Separation Comparison

	Separation Efficiency (%)		References
	Low	High	
Density Separations			
Gravity Settling	5	29	3,26
Centrifuge	0	10	26
Hydrocyclone	13	33	4
Wettability Separations			
Froth Flotation	44	80	8,31
Selective Shear Aggregation	--	66	40
Direct Pelletization	69	96	11,44
Solubility Separations			
Mineral Solubility	96	99	45,46
Kerogen Solubility	0	28	46
Electrostatic Separation	0	29	48
Magnetic Separation	--	25	49
Optical Separation	0	21	53
Friability Separation	5	20	51
Bioleach Separation	20	44	57

In the low enrichment category are all of the other separation methods. The stickiness separation and the TRW process are not listed since sufficient data are not available.

At this juncture it is useful to point out that three of the four high enrichment separation processes are based on wettability differences. This suggests that surface forces are most effective in selectively recovering kerogen. Body force separations like density, magnetic, and electrostatic separations are not as effective as surface forces in the recovery of kerogen.

5.2.3 Concentrate Pelletization

Depending on the method of separation, either a dry or wet kerogen concentrate may need to be aggregated to a larger particle size for retorting or other means of oil recovery. If the kerogen concentrate is wet, dewatering will be required before pelletization. Dewatering will probably be performed by filtration (or centrifugation) followed by drying. Similar processing is performed on ceramic clays, which are of a similar particle size. Pelletization of the dried kerogen concentrate could probably be performed with a pan nodulizer typical of those used to pelletize iron ore or alumina.

5.2.4 Tailings Disposal

The beneficiation of oil shale produces a huge volume of finely divided mineral gangue requiring disposal. Due to its void fraction, the gangue volume will be at least 1.3 times volume mined. The methods of tailings disposal will differ depending on the type of separation used. If the separation was performed dry, cementation will be required to prevent tailings dump dust storms. Retort spent shale can probably be used to help cement since it is often cementitious and is another waste stream of the process. For very high enrichment separations, sufficient retort spent shale may not be produced and additional cement will be required.

If the separation was performed wet, dewatering will be required. Dewatering a slurry of finely divided mineral particles of similar particle size is presently performed by the phosphate industry in Florida. In this industry, dewatering is performed by a thickener followed by a settling pond. This process is plagued with low dewatering rates. In some cases, more than 10 years are required to remove enough water for agricultural use of the tailings. The volume of tailings ponds required for the oil shale industry could be approximately 4.3 times the mined volume. The volume of water tied up in the ponds could be 3 times the mined volume. (This analysis assumes 30% solids by volume in the pond.) As a result of this excessive water loss, an alternate process sequence consisting of filtering (or centrifugation) would probably be required for the mineral gangue. Similar processing is performed on ceramic clays. Such processing could possibly decrease the water volume losses to approximately 50% of the mined volume, albeit at an increased cost.

5.3 Technical Feasibility of High Enrichment Separations

The technical feasibility for the high enrichment separations will be considered next in the context of a large-scale production facility.

5.3.1 Technical Feasibility of Froth Flotation

Froth flotation is used on a large scale for the beneficiation of copper ore. A major difference between copper ore and oil shale beneficiation is that the liberation size for copper is much larger (~100 μm). Therefore, more grinding will be required for oil shale beneficiation. Research on the froth flotation of oil shale has suggested high separation efficiencies under conditions similar to copper ore flotation. For this reason scale-up problems are not expected to be different from those encountered with copper. One disadvantage of this process is that both the concentrate and the tailings will contain water. The water attached to the concentrate will require a heat load to evaporate it. The tailings slurry will have to be concentrated before disposal. Water lost with tailings disposal may be significant.

5.3.2 Technical Feasibility of Selective Shear Aggregation

Shear aggregation has been demonstrated only at the lab scale. There are many uncertainties of scale-up in this relatively simple process. This process suffers the same ills as froth flotation, producing a wet concentrate and a tailings slurry. Again, water losses may be significant.

5.3.3 Technical Feasibility of Direct Pelletization

Direct pelletization has been demonstrated only at the lab scale. There are many uncertainties of scale-up in this process. Further uncertainties exist with respect to heptane losses and water losses.

5.3.4 Technical Feasibility of Acid Solution

Acid solution is an analytical technique to obtain pure kerogen for analysis. The technique requires huge volumes of different acid solutions for small amounts of oil shale. As such, this process is not likely to have any commercial importance.

5.3.5 Technical Feasibility Conclusions

In summary, the technical feasibility of froth flotation as a large-scale, high-enrichment separation process for oil shale has a high probability of success. The other high-enrichment wettability separations (i.e., selective shear aggregation and direct pelletization) are much more uncertain. It is improbable that acid solubility will be used on a commercial scale for the separation of kerogen from oil shale. Lack of information about the TRW process and stickiness separation make the determination of their technical feasibility impossible. Further laboratory research in these areas is warranted.

5.4 Process Design of Selective Shear Aggregation

The simplified flowsheet of Selective Shear Aggregation (SSA) including grinding is shown in Figure 5-1. Only major equipment units are indicated while surge bins, hold tanks, pumps, short conveyors, and other minor units are omitted. This arrangement is similar to that in Figures 3-2 and 3-4. The material balance is reported in Figure 5-2 and process data are listed in Table 5-2.

Figure 5-1

SELECTIVE SHEAR AGGREGATION

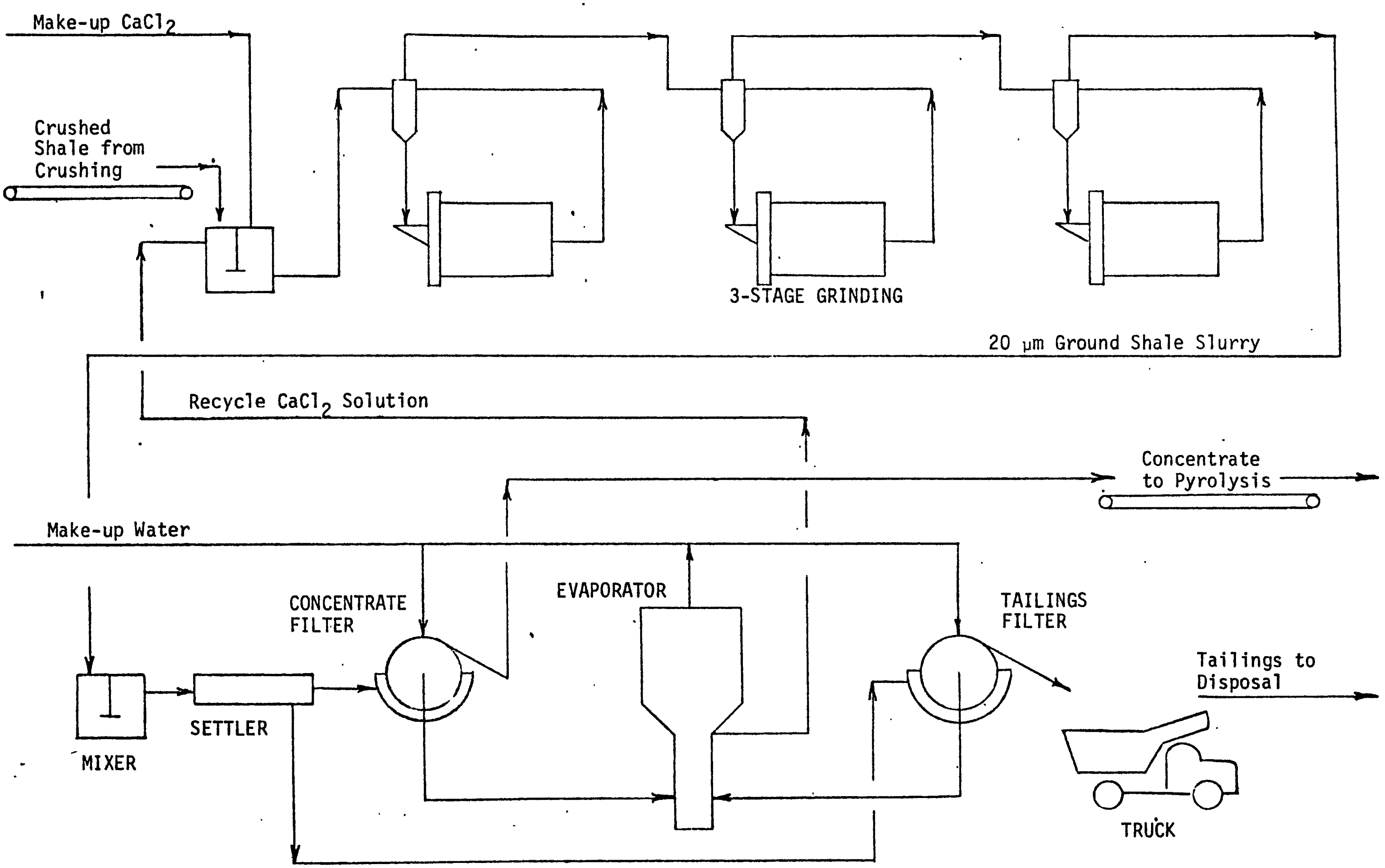


Figure 5-2

OIL SHALE BENEFICIATION BY SELECTIVE SHEAR AGGREGATION

Material balance in k tons/day

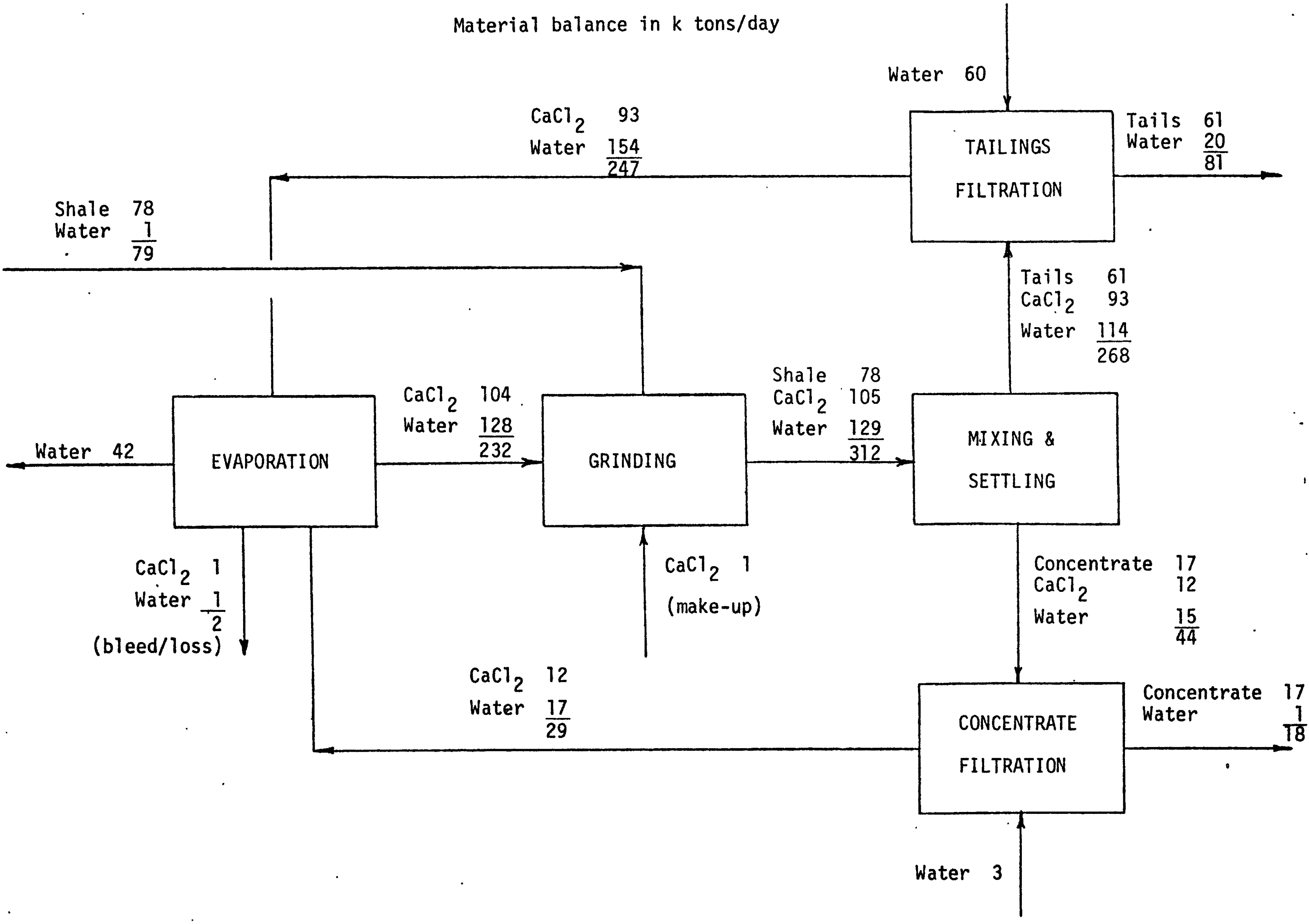


Table 5-2

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 x 36'
Motor size	4.4 MW
Mixers: Number	14
Size	17' diam x 15'
Settlers: Number	14
Size	108' x 40' x 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

The shale is mixed with a recycled concentrated calcium chloride solution, ground to minus 20 microns 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. (Only one effect is shown in the flowsheet as a symbol for the entire unit.) 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 grinding while only the rougher and middling cell froth are reground in the flotation alternative.

5.5 Process Design of Direct Pelletization

A flowsheet of Direct Pelletization (DP) including grinding and waste disposal in a tailings pond is shown in Figure 5-3. The material balance including heptane recovery is reported in Figure 5-4 and process data are listed in Table 5-3.

Figure 5-3

DIRECT PELLETIZATION

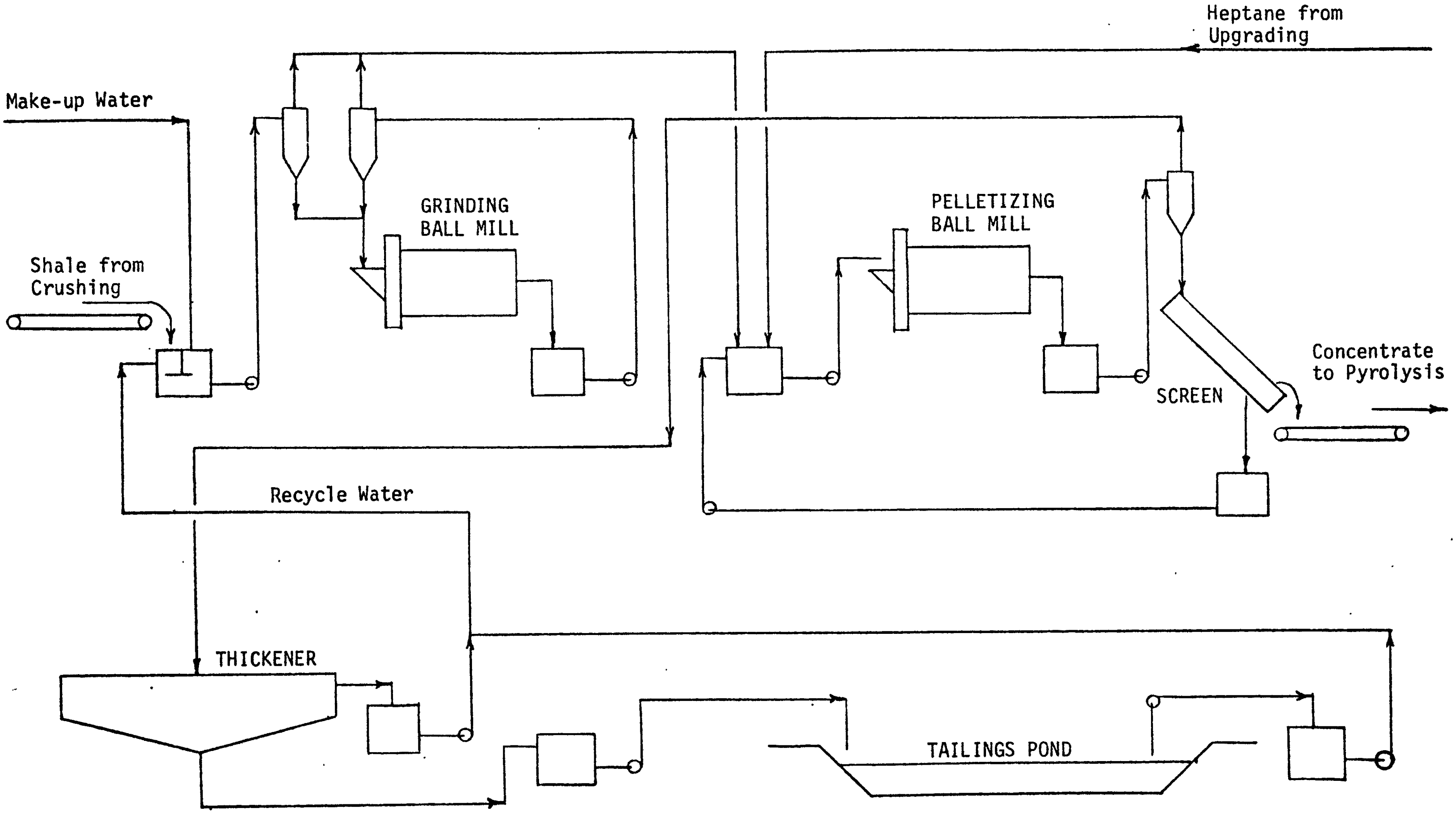


Figure 5-4

OIL SHALE BENEFICIATION BY DIRECT PELLETIZATION

Material balance in k tons/day unless indicated otherwise

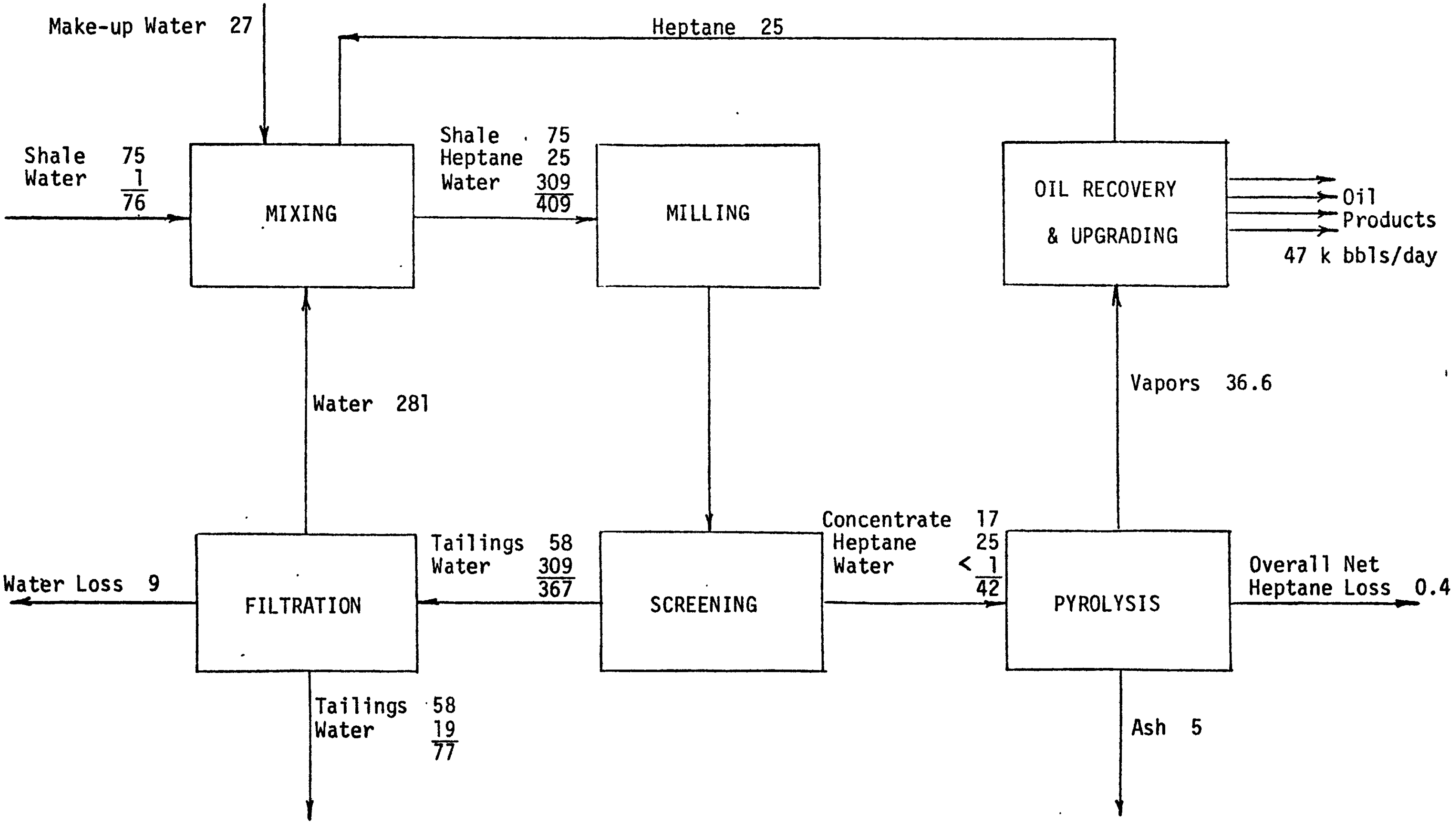


Table 5-3

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
Motors: Pelletizing: No. per mill	1
Size	2.9 MW
Screens: Number	12
Size	10' x 16'
Thickeners: Number	4
Diameter	700'

The shale is mixed with recycle water, ground to minus 100 mesh 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 residence time and power requirements for the pelletization operation. Thickeners and a tailings pond are proposed because slurry handling is the least expensive system for 100 mesh particles. The small volume pyrolysis residue would be also dumped into the pond. The thickener settling rates were selected based on minerals industry experience and have to be confirmed. The tailings pond differs from the Base Case solid spent shale disposal and the cost difference is difficult to assess since it is very site-dependent.

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 would be easily removed from the recycle water if necessary. However, the clean-up of the tailings slurry would be extremely expensive because it would require such techniques as steam stripping or solvent extraction of large slurry volumes. Lacking design data on the pelletizing ball mills, it was assumed that they are the same size as the grinding equipment but require only one-third the power.

5.6 Economics of Alternative Separation Processes

Capital costs of Selective Shear Aggregation (SSA) and Direct Pelletization (DP) were estimated using the factor method described in Section 3.7.1 based on ASPEN equipment cost compilation (Ref. 54) and Appendix A. A 40% process contingency was used for SSA and DP compared to 30% for froth flotation (in the Base Case) because of the additional uncertainties.

Annual costs were also calculated for each alternative using the methods described in Section 3.7.4. That is, annual operating costs were added to a 25% annual capital charge to estimate total annual costs.

The results (details are shown in Tables 5-4 and 5-5) show that compared to froth flotation, SSA has significantly higher capital and annual costs. But DP has 24% lower annual cost because of both lower (17%) capital cost and lower (41%) power consumption. Thus, of the beneficiation options considered, direct pelletization is speculative but the most interesting looking technology.

Table 5-4

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	6
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
Contingency:			
Project (15%)	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 5-5

Annual Cost of Alternative Separation Processes

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 Beneficiation Sections				
Annual Operating Costs		158	274	114
Annual Capital Charge at 25%		218	273	180
Grand Total Annual Costs (Rounded)		380	550	290

6. RECOVERY ALTERNATIVES

This section examines some possibilities other than conventional surface retorting for recovering oil and associated gas from the product of the separation step, kerogen concentrate.

6.1 Rationale for Examining Alternative Methods

Pyrolysis, i.e. heating to the temperature of decomposition in a retort (a surface vessel or underground chamber), has been the almost universal means of converting the kerogen in oil shale to shale oil. Pyrolysis times, temperatures, geometries, heat transfer arrangements, spent shale treatment, and other details have varied widely but the basic oil recovery mechanism has remained the thermal decomposition of kerogen in a more or less inert atmosphere.

A comparatively small effort has been devoted to investigating other recovery methods, notably solvent extraction and retorting in the presence of hydrogen. But these alternatives to straightforward pyrolysis have shown potential only under unusual circumstances and have not progressed to any large-scale development or demonstration programs that we know of, much less commercialization.

However, the availability of a concentrate rich in kerogen rather than a natural ore lean in kerogen suggests that the alternatives to pyrolysis be reexamined. The optimum method for converting a concentrate to oil may resemble the method that is optimum for a petroleum residuum or a high quality coal rather than for a lean shale ore. For example, the kerogen concentrate that results from four-fold enrichment of Colony ore contains about 80% organic matter (140 gpt)--roughly equivalent to the percentage in dry subbituminous coal.

In considering alternatives to retorting, the objective is to find an alternative that:

- Gives the same results but is cheaper
- Gives yields superior to retorting without offsetting costs
- Gives product quality superior to retorting without offsetting costs
- Or, has some other desirable characteristic (e.g. more benign environmentally, lower labor requirements) without offsetting costs.

Some general comments on these possible advantages are appropriate before looking at the specific cases examined in the following sections. We emphasize that the advantages are not likely to be realized unless the alternative (to retorting) process is able to exploit the different nature of the concentrate compared to natural ore; otherwise, the alternative would be useful on natural ore too.

An extraction process with competitive yields seems to require either extraordinary solvency by the solvent, or "extraction" at incipient pyrolysis conditions where the distinction is unclear between what is directly dissolved in solvent and what is dissolved only after thermal decomposition. In the latter case, the "solvent" serves primarily as a heat transfer medium for pyrolysis. In the former case, extraordinary solvency may be achievable by using conventional solvents at supercritical temperatures, i.e. at temperatures above the critical temperature where the solvent cannot be liquefied regardless of the pressure. It seemed conceivable that supercritical extraction might proceed by either or both of the mechanisms described above to recover oil competitively with retorting if the feedstock contained less inert material--as the kerogen concentrate, in fact, does. Thus, supercritical extraction was evaluated.

Retorting in the presence of hydrogen has been demonstrated to recover shale oil from ore with both yields and quality equal to or superior to those from conventional retorting. The real issue here is whether the likely increases in yield and/or quality are sufficient to offset the obvious added costs of generating hydrogen and retorting under pressure. The availability of the kerogen concentrate also makes it possible to consider hydrogen-retorting processes suitable for coal and residua that would be unsuitable for natural shale ores.

The increase in yield potentially available from non-retorting processes is limited. Developers of current conventional retorts claim high yields now from Western shales relative to Fischer Assay, e.g. 95% for Paraho direct, 100% for Union B, 105% for Lurgi (Ref. 5). Other laboratory data suggest that time and temperature optimization may be able to increase those retorting yields to about 110%, e.g. Ref. 58. How much more _can_ be recovered in principle?

Stanfield (Ref. 59) reports that about 75-88% (depending on ore richness) of the heating value of the kerogen of Western shales is contained by the oil from a Fischer Assay. Therefore, if the alternative recovery process produced only "oil" from kerogen, _no_ gas and _no_ char, its heating value would be equivalent to 111 to 133% of Fischer Assay--compared to the 110% which may be achievable from optimized retorting. Thus, the maximum theoretical increase in yield is modest.

Adding hydrogen can result in increased yield and energy output, but at the cost of energy input to produce and react the hydrogen. The energy and economic balances must be calculated for specific cases, especially for Eastern hydrogen-poor shales where large gross yield increases have been reported.

Potential changes in product quality are likely to be modest improvements for hydrogen retorting and significant debits for extraction. If an extract has an atomic ultimate analysis similar to kerogen, that extract would have less hydrogen, more nitrogen, more sulfur, and more oxygen than raw shale oil from retorting; all those differences are unfavorable.

Hydrogen retorting produces oils which, compared to conventional retorting, may be lighter but which have negligibly different nitrogen contents. Nitrogen content controls the cost of upgrading required to ultimately produce marketable transportation fuels.

6.2 Retorting in the Presence of Hydrogen Under Pressure

A hydrogenation process specific for oil shale was developed by the Institute of Gas Technology (IGT) under the name Hytort (Ref. 60). Other studies of hydrogen retorting of oil shale have been reported by Texaco (Ref. 61) and Esso (Ref. 62). But the Hytort process has been developed farthest and is used as the model here even though it is intended primarily for Eastern shales (where hydrogen retorting significantly increases yield) and for natural ore (since it is a vertical shaft process depending on physical integrity of ore fragments).

The flowsheet of the Hytort process adjusted to the treatment of oil shale concentrate is shown in Figure 6-1. Process data are listed in Table 6-1. 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 and preheated, to the lower part of the reactor. The solid residue is withdrawn from the bottom and disposed of in the same way as in the Base Case with beneficiation.

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 Upgrading. The water from these three separation steps is sent to the foul water treatment plant that is outside of the oil recovery system. The sludge from the quench and scrubber separators is added to the other solid wastes.

The overhead gas from the scrubber is divided into three streams. One portion is directly recycled to the reactor while the second 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 shown in Figure 6-2. It includes a furnace, shift reactors, and a hydrogen cleaning system.

The Hytort process has been developed through the pilot plant stage as a more efficient shale oil extraction process. However, it has the

Figure 6-1

DIRECT HYDROGENATION

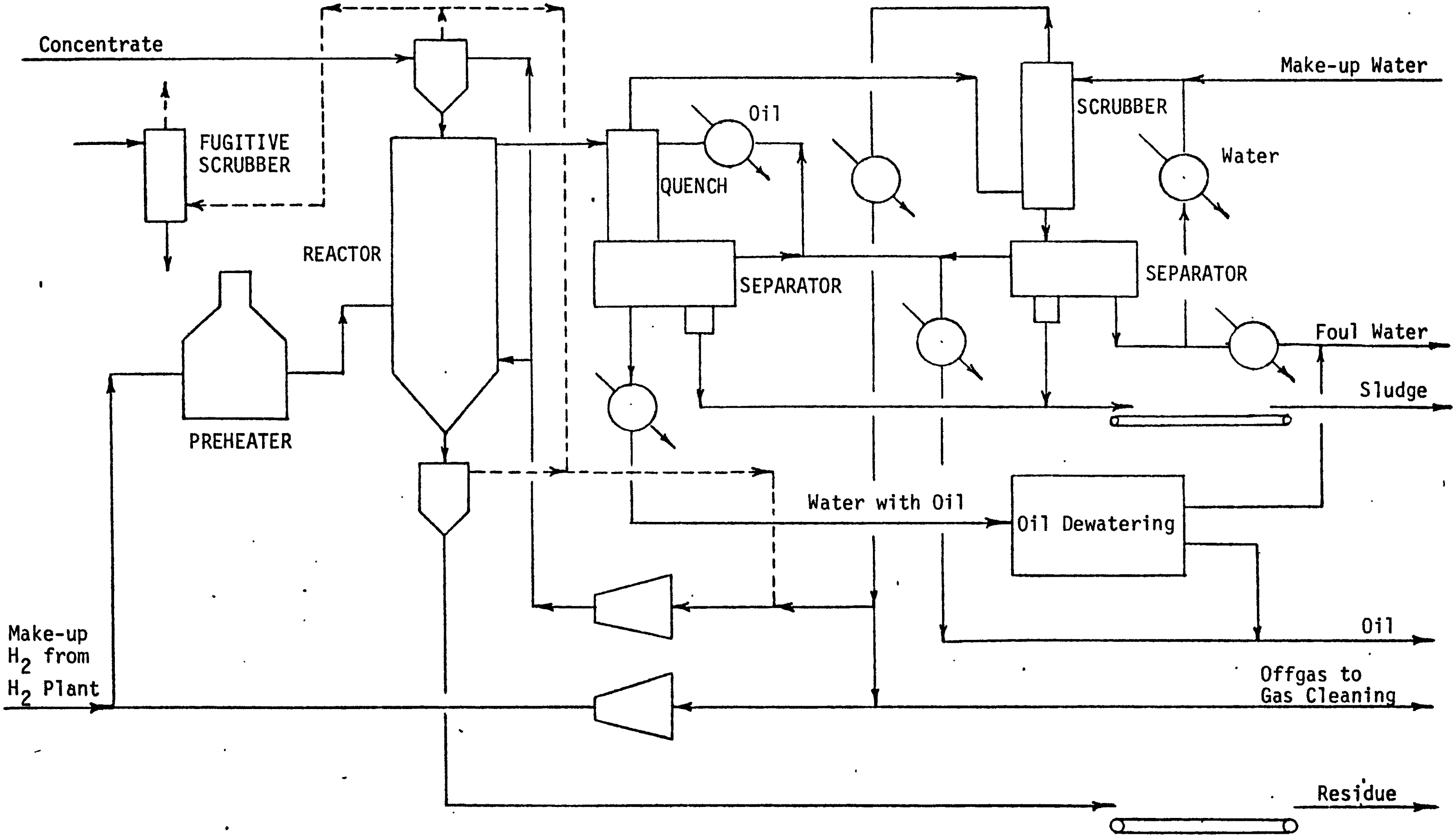


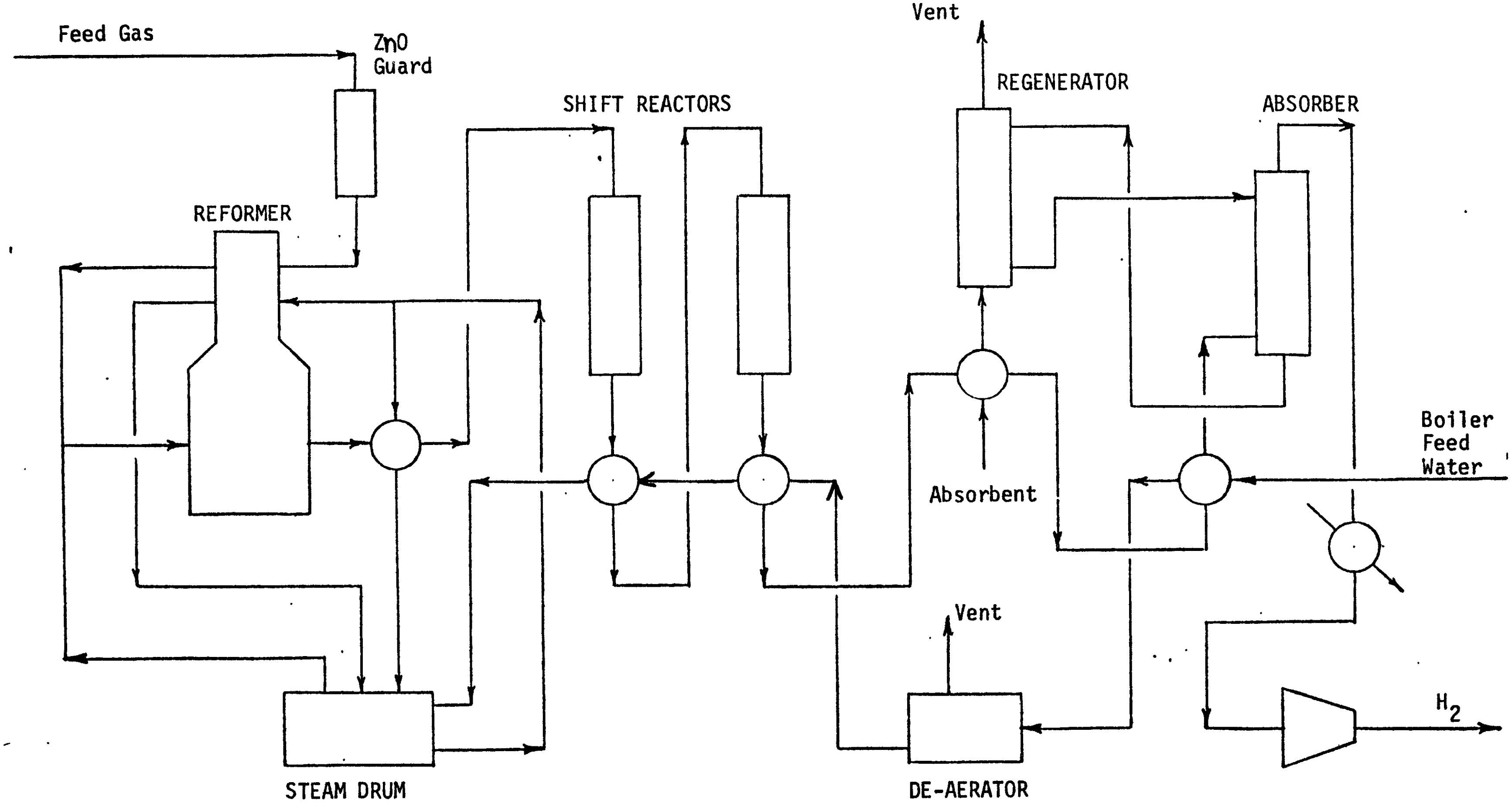
Table 6-1

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

Figure 6-2

HYDROGEN PLANT



disadvantage of heating the reactor with diluted hydrogen which requires the circulation of large amounts of gas at elevated pressures. An additional drawback of the Hytort application to shale concentrate is fine particle entrainment.

Even if the feed is pelletized, fines would undoubtedly be generated by attrition and require an efficient solid-gas separation system to prevent clogging of the feed equipment, damage to compressors, deposits on heat exchange surfaces, and other problems. The process could also be made more energy-efficient by including, ahead of the quench, a waste heat boiler (WHB) preceded by an electrostatic precipitator to protect the WHB cooling surfaces.

IGT claims that Hytort produces a higher grade oil than other retorting systems, but, as discussed in Section 6.1, the value of the improvement is small. In addition, there is a price to be paid for the front-end hydrogenation as will be shown below in Section 6.6. To decide whether it is better to hydrogenate in the retort or during upgrading would require a thorough analysis of the entire plant which is beyond the scope of this study.

6.3 Recovery in the Presence of a Hydrogen Donor Solvent

This section considers the hydrogenation of kerogen concentrate using a hydrogen donor solvent, or technology analogous to the Exxon Donor Solvent (EDS) process for liquefaction of coal. Our analysis was based on a report on the pilot plant developed by Exxon (Ref. 63). The flowsheets are shown in Figures 6-3 and 6-4, and the process data listed in Table 6-2.

Figure 6-3

DONOR SOLVENT HYDROGENATION

Extraction Section

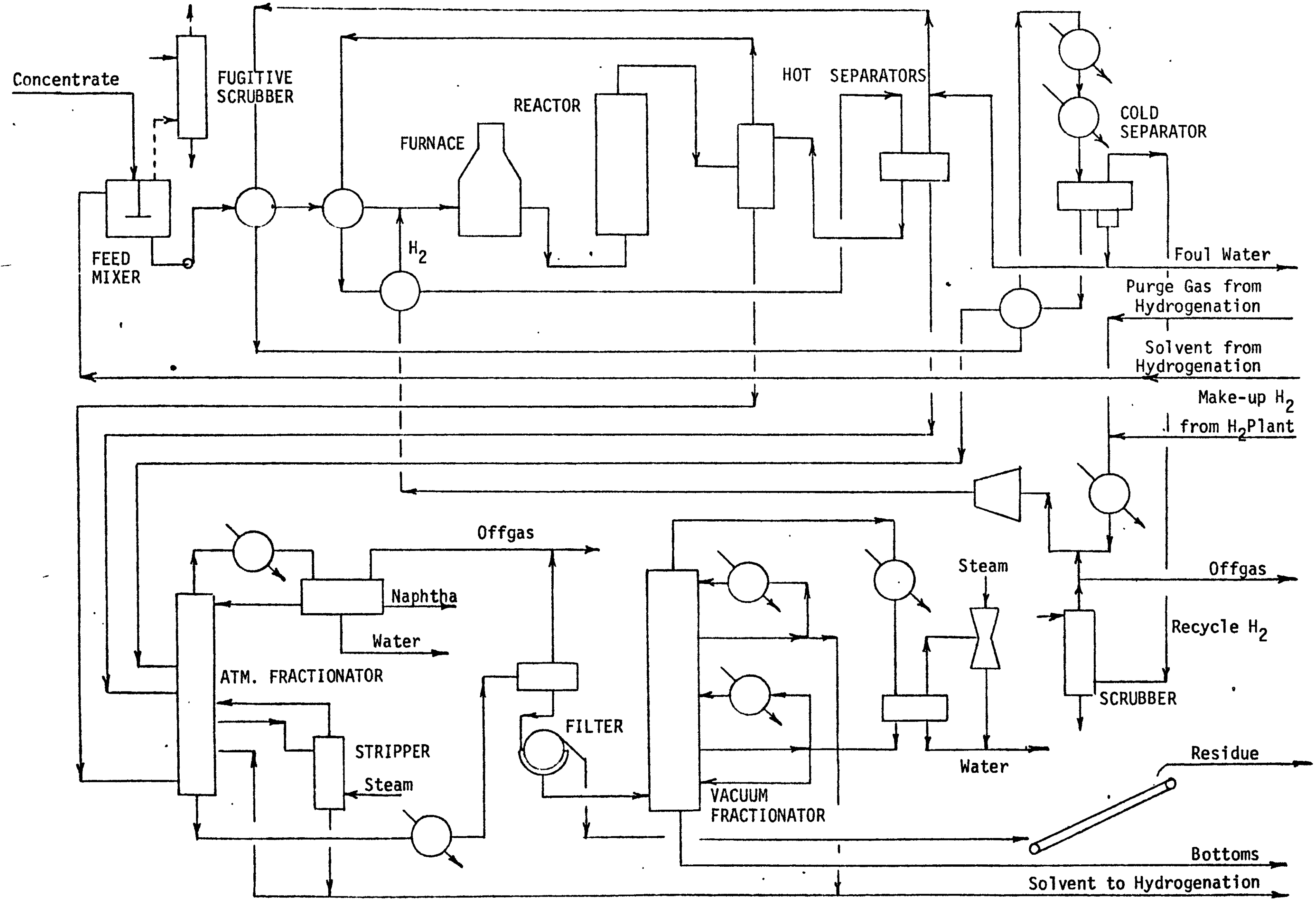


Figure 6-4.

DONOR SOLVENT HYDROGENATION

Hydrogenation Section

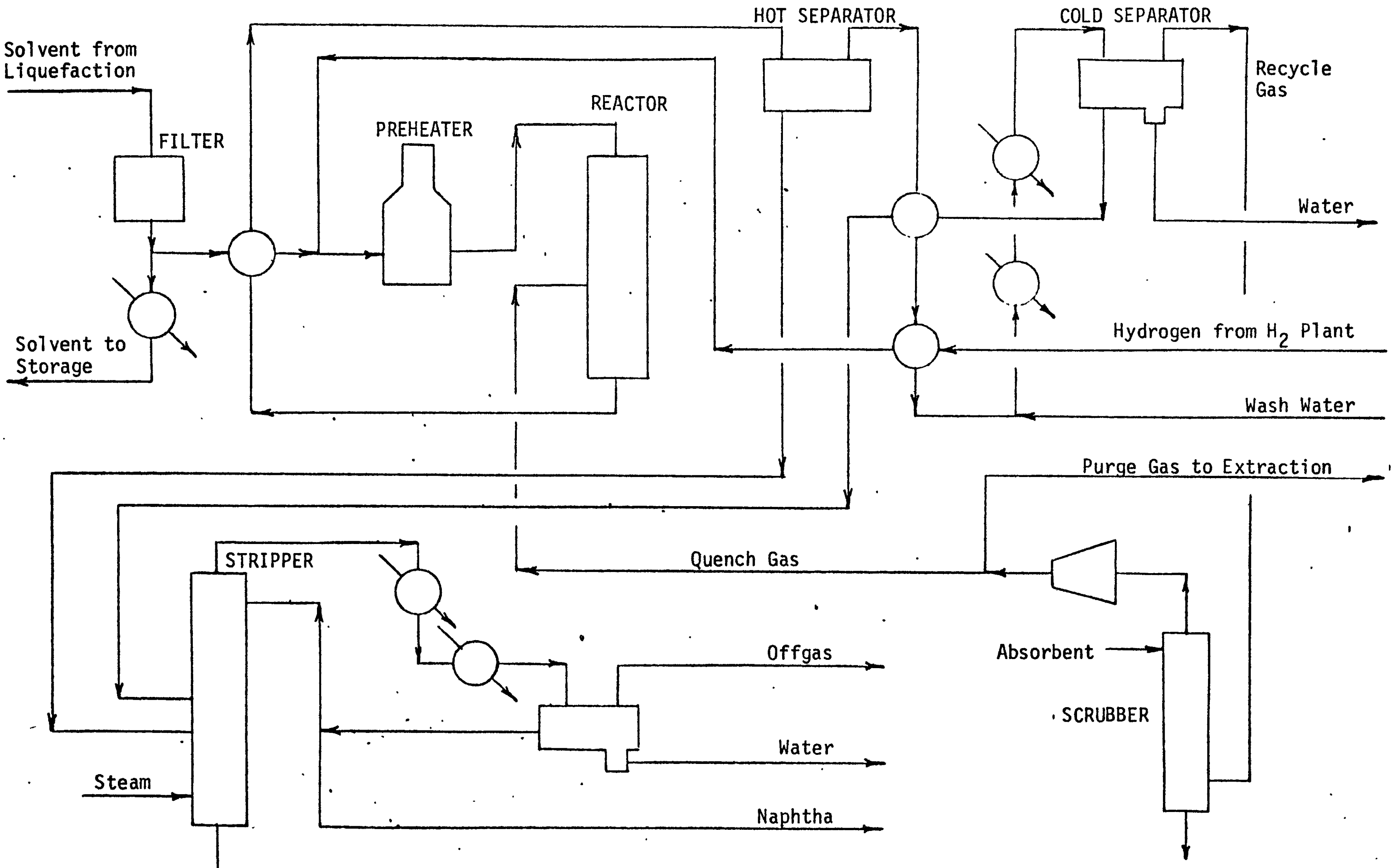


Table 6-2

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, MBtu/hr	1090
Power, kW	25
Water, k gal/min	0.85
Steam, k lbs/hr	67

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 the Upgrading section. 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 Upgrading.

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 reformer plant, and a hydrogen-rich purge gas from Hydrogenation. The resulting hydrogen recycle gas is pumped to the preheating furnace. Foul water from the cold separator is treated outside of the Extraction section.

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 (see Figure 6-2) 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 exchangers, direct water injection, and a hot and cold separator. The liquid fractions from both stages are fed to a steam stripper.

Offgases, naphtha, and water are separated in an overhead condensate tank and sent to the Upgrading section. 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. The cold separator water is treated outside the Extraction section.

The major advantage of the EDS system is replacement of the gas-solid system with a liquid-solid alternative which facilitates mass and heat transfer in the reactor. Another positive feature of the original EDS process is the complete liquefaction of coal that avoids most of the cumbersome downstream solid-fuel handling. The disadvantage of the original EDS is the high pressure required to liquefy coal together with the 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 the positive EDS features.

EDS conditions are probably more severe (and more expensive) than necessary to convert kerogen, and incomplete conversion or another liquid phase may cause solid-liquid separation problems after conversion. On the other hand, the EDS slurry reactor avoids the problems of feeding rock fragments into a high pressure reactor and of entraining fines in a gas stream. Pelletization of the EDS feed is probably not necessary and is not assumed in this study.

The design of the downstream liquid-solid separation system, and the assessment of whether it is preferable to the front-end gas-solid handling would require experimental data. A filtration step following the atmospheric fractionator was included in this study only to allow an approximate cost estimate.

EDS will probably yield a higher grade oil than the Base Case. The comments on this point in Section 6.1 apply also to EDS.

6.4 Recovery by Supercritical Extraction

Supercritical Extraction (SCE) of various solids utilizes the order-of-magnitude increase in dissolution power of some light organic liquids which are compressed and heated above the critical temperature. The potential application to oil shale concentrate is based on the SCE process for coal developed by the British Coal Board and Catalytic, Inc. (Ref. 64). The flowsheet is shown in Figure 6-5 and process data are listed in Table 6-3.

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. (Toluene is the solvent used by the British Coal Board. Some other solvent may be optimum for kerogen without significant effect on the flowsheet or economics.) 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 Upgrading section. 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

Figure 6-5

SUPERCRITICAL EXTRACTION

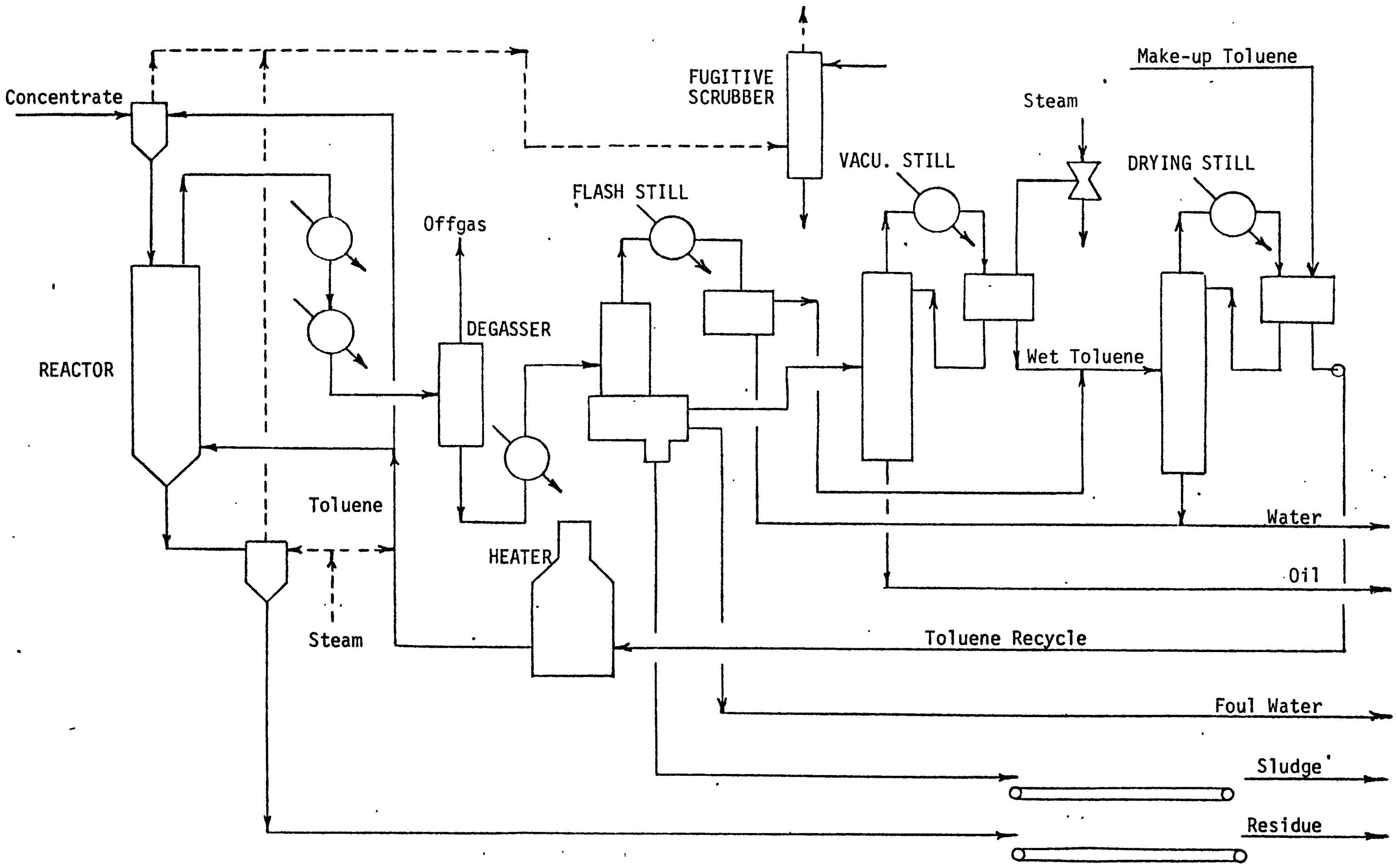


Table 6-3

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, MBtu/hr	410
Power, kW	67
Water, k gal/min	1.08
Steam, k lbs/hr	10

bottom of the flash still and are periodically withdrawn. The sludge is combined with the other solid waste while the foul water is treated outside the Extraction section. The organic phase from the flash still is separated into oil and wet toluene in a vacuum still. The oil is further processed in Upgrading while the combined toluene streams are dewatered in the drying still and pumped back to the reactor.

SCE is a flexible method used in the processing of various materials ranging from soybeans to fossil fuels. Air Products and Chemicals, the parent corporation of Catalytic, Inc., conducted its bench scale coal extraction experiments and is currently working on the application of SCE to tar sands but no technical information is available to the public.

While batch reactors were considered for coal extraction, the tar sand process design is reportedly based on continuous units which permit a substantial decrease in the number of reactors. There were eight units in the 10,000 tpd coal project (Ref. 64). Our study assumes only one reactor. Another change in the tar sand process is the replacement of lock hoppers with a slurry pumping and preheating system apparently similar to that of the EDS process (Figure 6-3).

SCE appears to be an elegant technology. It 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 as discussed in conjunction with the EDS process in Section 6.3.

6.5 Flash Pyrolysis

Flash pyrolysis depends on thermal decomposition in an essentially inert atmosphere as conventional retorting does. But the geometries and flows are so different, and the application to comminuted shale is so obvious, that we examined flash pyrolysis along with the other recovery alternatives.

Flash pyrolysis of coal was proposed (Ref. 65) based on bench scale experiments. The process adapted to kerogen concentrate is represented by the flowsheet in Figure 6-6 and process data are listed in Table 6-4.

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 then 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

Figure 6-6

FLASH PYROLYSIS

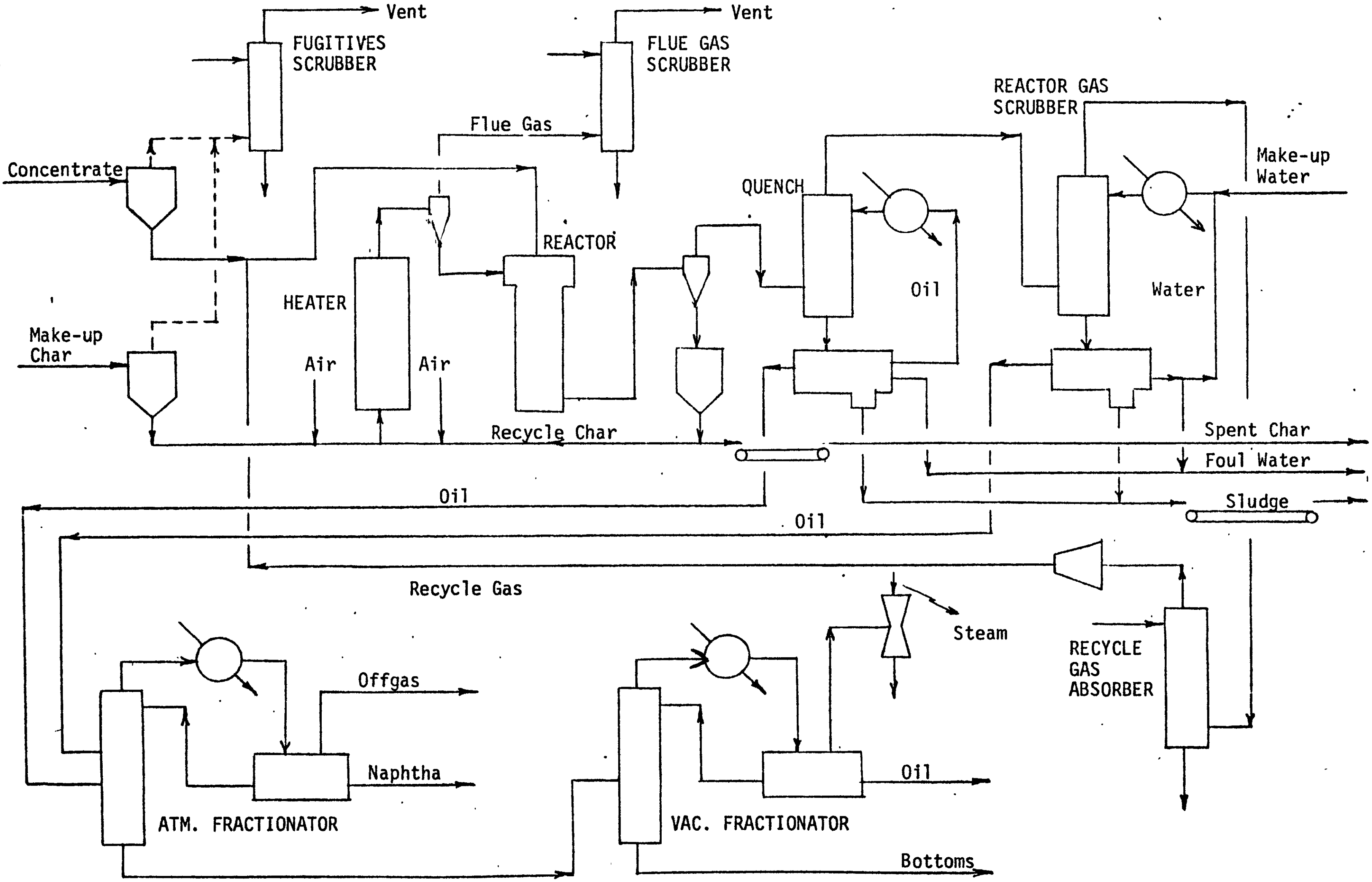


Table 6-4

Flash Pyrolysis Process Data

Reactor: Temperature, °F	1200
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, MBtu/hr	500
Power, kW	110
Water, k gal/min	1.12
Steam k, lbs/hr	1.0

separates into offgas and naphtha. The products of the vacuum fractionation, oil and bottoms, are further processed in the Upgrading section.

Foul water separated in the quench and scrubbing steps is treated outside the Pyrolysis section. Sludge accumulated in the respective separators is periodically withdrawn and combined with the other solid waste. Gas from the scrubber is purified in an absorber and recycled to the concentrate pneumatic feed system.

The major advantage of the Flash Pyrolysis process is that it 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 bed system. Also, of the alternatives compared here, process data on Flash Pyrolysis are most speculative.

Problem areas of Flash Pyrolysis are similar to those of the Hytort Process. The major issue is again particle carryover to the downstream parts of the process. Another issue is the provision of heat by the combustion of carbon on the spent shale.

6.6 Process Design and Economics

There are too few published data on individual equipment units for the Hytort, Exxon Donor Solvent (EDS), Supercritical Extraction (SCE), and Flash Pyrolysis processes to allow a rigorous process design of individual equipment and the use of the factored estimation method discussed in Section 3.7. However, there are published cost data prepared by various organizations at various times on all the processes except Flash Pyrolysis.

An analysis of the Hytort data and comparison with cost of similar processes indicated that the capital estimate was too optimistic. Accordingly, it was increased assuming an underestimate of the same magnitude as in the 1977 Colony figures. The EDS and SCE estimates appeared to be reasonable. However, the SCE estimate is based on ten semibatch reactors while a fully continuous process would need only one reactor. The direct adjustment from ten to one reactor is not possible because reactor costs are not separated from the other items in the SCE publication.

The process and cost engineering work on the adjustment of the four processes to shale concentrate was done in the following way.

- Flowsheets for the extraction plant (excluding upgrading) were prepared. They are shown and discussed above in Sections 6.2 through 6.5.
- Approximate material balances were computed. As discussed in Section 6.1 above, one of the purposes of investigating alternative recovery methods was to identify a process with a potentially higher yield. There is no clear experimental evidence about increased yield from any of the four selected alternatives. Therefore, an optimistic yield equal to 120% of Fischer Assay (compared to 90% for the Base Case) was assumed for each alternative. The flowrates of the major streams in Tables 6-1 through 6-4 are based on the material balance that assumes that 120% yield.

Table 6-5

Comparison of Alternative Shale Oil Recovery Processes

Item	Ore Processing	Concentrate Processing				
	Retort Pyrolysis	Retort Pyrolysis	Direct Hydrogenation	Donor Solvent Hydrogenation	Supercritical 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						
1. Mining	290	320	250	250	250	250
2. Beneficiation	--	550	420	420	420	420
3. Recovery	770	220	520	610	260	640
Total	1060	1090	1190	1280	930	1310
Annual Cost, $M/yr						
1. Mining	21	24	18	18	18	18
2. Beneficiation	--	134	105	105	105	105
3. 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

- The published Hytort, EDS, and SCE capital costs were adjusted to recovery from shale concentrate as defined in the flowsheets. Plant sections beyond recovery, such as Upgrading, were excluded. Direct costs were scaled up or down by sections, e.g., EDS recovery versus hydrogen plant. Indirect and non-depreciable items were reestimated in a uniform way.
- Comparing the Base Case to adjusted costs based on poorly documented estimates of three different processes prepared by three different organizations might be unpersuasive. In addition, no cost data were available on Flash Pyrolysis and on single-reactor SCE. To provide a check, the capital costs of all four processes including beneficiation were independently estimated using the modular method discussed above in Section 3.7. Those independent estimates were consistent with the adjusted published estimates within the accuracy of either.
- The material balance and literature provided sufficient data on approximate requirements for process materials and utilities. The numbers of operators were estimated based on the flowsheets. The other annual cost items are proportional to operating labor or capital costs.

Economics of the four alternative recovery processes are compared with the two Base Cases in Table 6-5. 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. 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. Section 7 presents a further discussion of system comparisons.

7. ALTERNATIVE SYSTEMS

The three preceding sections were concerned with individual steps in the overall system to see if alternative steps preferable to those in the Base Case could be identified. The purpose of this section is to see whether those alternatives can be combined with achievable operating parameters to arrive at a total beneficiation system which would look attractive compared to the Base Case without beneficiation.

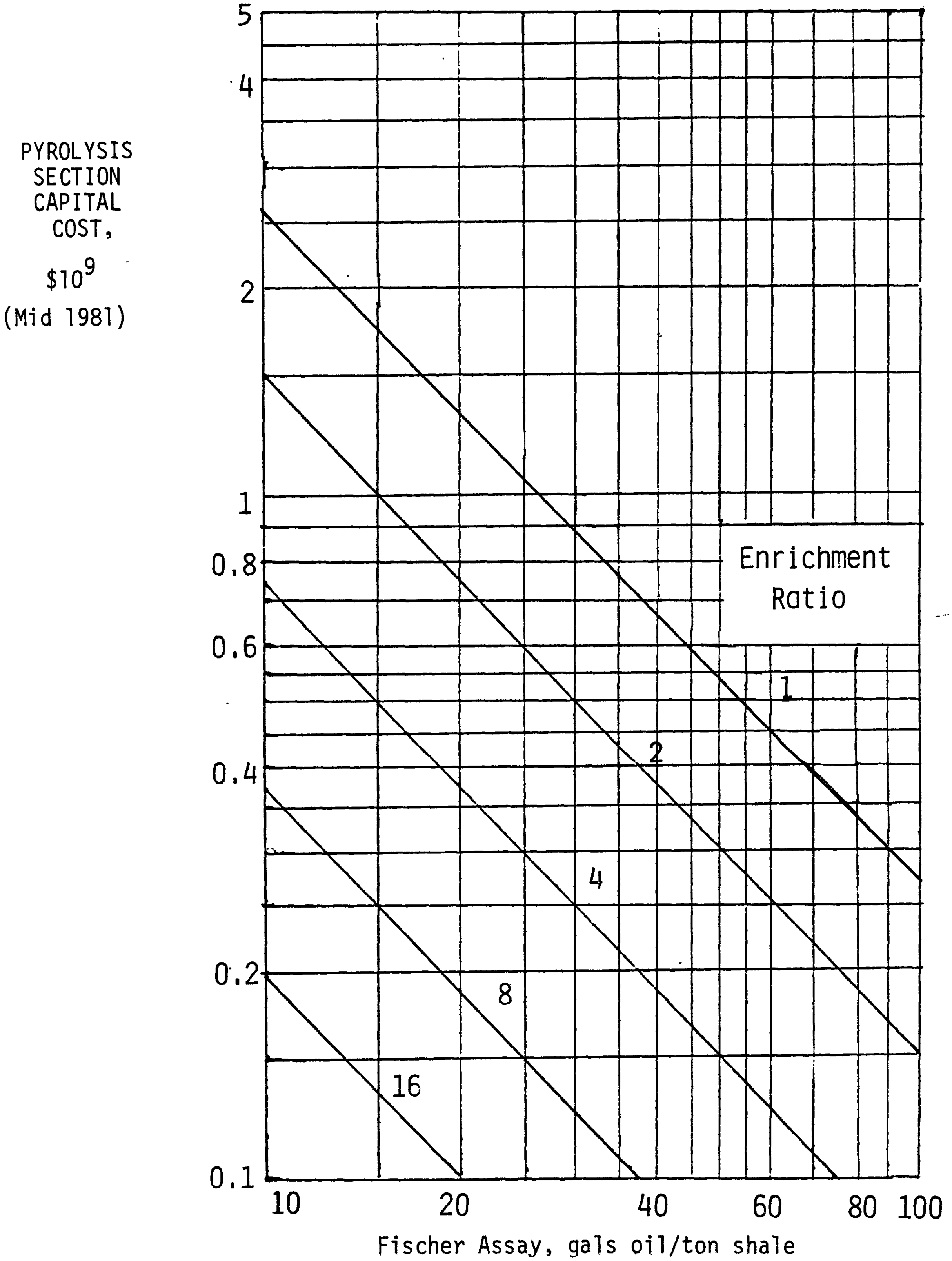
We start by assuming use of the same technology used in the beneficiation Base Case (i.e. ball mill grinding, froth flotation, and Tosco II pyrolysis). We can then perform a sensitivity analysis, an examination of the effects of changes in operating parameters on costs. The following parameters were examined:

- Ore assay
- Enrichment ratio
- Grinding energy
- Separation efficiency
- Kerogen and oil recoveries

Ore assay is related primarily to the difference between Eastern and Western shales and it has a very large effect on capital cost of the pyrolysis section. That capital cost is plotted versus ore assay and enrichment ratio in Figure 7-1. As annual costs are controlled by capital, the plot is a convenient simple indicator of the effect on total costs of the two major parameters. Figure 7-1 is a simplistic diagram which assumes that the capital cost of retorting is directly proportional to the total amount of mass that must be retorted to produce a barrel of shale oil; that assumption is a reasonable approximation for most retort

Figure 7-1

SIMPLISTIC SENSITIVITY ANALYSIS OF PYROLYSIS



designs and becomes more accurate over modest assay ranges with smaller throughputs in each module and with fewer common facilities capable of realizing economies of scale. The base starting point of the diagram is the Colony capital cost of $770 million at 35 gpt and an enrichment ratio of one.

A Western shale with a 35 gal/ton assay cannot be enriched even theoretically beyond about a 5 to 1 ratio. In general, no substantial savings can be realized by pushing the enrichment to the limit because the first savings are the largest ones. Figure 7-1 also shows the tremendous cost difference between direct retorting of Western and Eastern shales, the latter having a typical assay of 10 gal/ton.

Although Tosco II may not be the optimum technology for processing unbeneficiated Eastern shale, no other optimum has been convincingly identified. For example, the Davy McKee study on Eastern shale (Ref. 73) showed that a modified Paraho retort was cheaper than the Hytort process despite the fact that Paraho was originally developed for Western shales and Hytort for Eastern shales. A diagram like Figure 7-1, with a somewhat different base reference point, should be applicable for Paraho retorting although it is not clear that Paraho--a solids gravity-flow shaft retort with gas upflow--could handle a pelletized kerogen concentrate.

The economic sensitivity of the beneficiation process, from mining through recovery, to the major parameters is presented in concise form in Table 7-1 in terms of total annual costs broken down by plant sections. (Total annual costs are calculated, as in Section 3, as the sum of annual operating costs plus a 25% capital charge.) The two sections of the table represent typical Western and Eastern oil shales. The first

Table 7-1

Sensitivity Analysis of Shale Oil Mining Through Recovery

Annual cost in millions of mid-1981 dollars (unless indicated otherwise)

Case	Mine	Grinding	Separation	Recovery	Total Rounded	% Savings
Western Shale, 35 gal/ton						
No beneficiation (Base Case)	94	--	--	277	370	--
Four-fold enrichment (Base Case):						
High grinding cost	104	187	85	82	460	-24
Low grinding cost	104	95	80	82	360	3
Target Process:						
With recovery:						
High grinding cost	65	89	46	86	290	22
Low grinding cost	65	48	46	86	250	32
Without recovery:						
High grinding cost	65	89	46	10[a]	210	43
Low grinding cost	65	48	46	10[a]	170	54
Eastern Shale, 10 gal/ton						
No beneficiation	260	--	--	970	1230	--
Four-fold enrichment:						
High grinding cost	280	650	290	240	1460	-19
Low grinding cost	280	330	270	240	1120	9
Fourteen-fold enrichment:						
High grinding cost	280	650	270[b]	82	1280	-4
Low grinding cost	280	330	265[b]	82	960	22
Target Process:						
With recovery:						
High grinding cost	180	310	160	86	740	40
Low grinding cost	180	160	160	86	590	52
Without recovery:						
High grinding cost	180	310	160	10[a]	660	46
Low grinding cost	180	160	160	10[a]	510	59

(a) Assuming high kerogen concentrate is fed directly to cracking/fractionation column.

(b) Decrease in regrind cost is partially offset by higher flotation cost.

two numerical lines are the Base Cases without and with beneficiation. The cases denoted "low grinding cost" indicate the effect of arbitrarily decreasing both grinding capital cost and power requirements to one half of those in the "high grinding cost" cases. One potential way of accomplishing this is the replacement of ball mills with stationary spiral mills (see Section 4.8.1) but the analysis in Table 7-1 does not depend on that specific change. The Eastern shale section has two extra cases for the fourteen-fold enrichment.

The effects of the other major parameters are shown by means of the so-called Target Process which assumes achievement of the most favorable conditions, often at the near-theoretical limits. It is defined as follows:

- An increase in enrichment ratio to 5:1 and 17:1 for Western and Eastern shales, respectively.
- A one-tenth increase in kerogen recovery during separation, e.g. from 88 to 97% of the ore kerogen for Western shales.
- An increase in pyrolysis (or other "recovery") oil yield, i.e., from 90 to 120% of Fischer assay.
- A further 50% decrease in grinding capital and power costs for the high grinding cost cases, e.g. equivalent to the savings from an increase in allowable particle size from minus 40 microns to minus 100 mesh such as in Direct Pelletization. Accordingly, the low grinding cost cases of the Target Process represent a <u>total</u> reduction in grinding capital costs and power consumption of 75% with respect to the "high grinding cost" cases.

- Elimination of the pyrolysis or other recovery section by feeding the high kerogen concentrate directly to a cracking/fractionation column, thus skipping the retort or reactor with material handling, heating, cooling, and phase separation equipment (last two cases of each section). A nominal $10 million is included for fractionation and down-stream sludge separation instead of the complete recovery section.

The following comments should help to interpret Table 7-1. Mining and grinding of Eastern shale is more expensive than that of the Western ore primarily because of large volumes involved. However, as Eastern mines are open pits while the Western operations are underground, the cost ratio of the former to the latter is 2.2 to 2.7 on a per ton basis (Ref. 9). In the Target Process, the increase in separation efficiency, enrichment ratio, and oil yield reduce the requirements for oil shale from 66 yo 45 k tons/day with the corresponding cuts in mining and beneficiation cost.

In the four- and fourteen-fold enrichment alternatives, the low grinding cost separation is less expensive than the high grinding cost case because these alternatives are based on flotation which includes regrind circuits. However, there is no such difference for the Target Process because it assumes no regrind as, for example, in Direct Pelletization. The savings in separation cost going from a four- to a fourteen-fold enrichment are not significant because the reduction in regrind cost is partially offset by more expensive flotation to attain the higher kerogen concentration.

Recovery costs of four times enriched Western shales and fourteen times enriched Eastern ores are equal within the accuracy of this study because of the same volume processed. However, they are slightly lower than for the Target Process because a more efficient recovery process, such as Supercritical Extraction, costs more than retort pyrolysis. This increased cost is only partially offset by the higher enrichment ratio assumed for the Target Process.

The following conclusions can be drawn from the sensitivity analysis represented by Table 7-1. Decreased grinding cost barely compensates for the higher beneficiation cost in the Base Case. High enrichments and high grinding efficiency significantly reduce Eastern shale processing with respect to the case without beneficiation, but total costs are still more than twice those of the Western shale cases. Substantial savings can be attained only under the favorable conditions of the Target Process applied to Western ores if grinding costs can be reduced and/or the extraction step skipped. The Eastern shale case is not attractive at all. While substantial relative savings can be attained with fourteen-fold enrichment or the Target Process, the absolute costs are still too high to make the case fly.

8. REMOTE RECOVERY

In the preceding sections it has been assumed that the entire process system, up through the production of raw shale oil at least, was located at or adjacent to the site of the ore body. However, as noted in Section 2, one potential advantage of a system incorporating beneficiation is that the kerogen concentrate may be transported away from the ore site and processed elsewhere to recover oil, upgrade, and refine. This option arises from the fact that the kerogen concentrate is a reasonably rich energy source, similar to some coals which can be and are economically transported, rather than a lean rock which cannot be moved any significant distance at reasonable cost.

Compared to the ore site, recovery from concentrate at a remote site may provide one or more of the following incentives. Some of these incentives can be exploited in part by remote upgrading and refining of conventionally produced raw shale oil:

- A location more tolerant of environmental emissions and wastes.
- A location with better site conditions for construction and/or operation such as climate, geology, and terrain.
- Location in an industrial area where the required infrastructure is in place and where skilled workers are available.
- Integration of the processing plant with a new refinery.
- Utilization of idle refinery capacity by retrofitting some plant sections to serve as compatible shale oil recovery or upgrading sections.

- Proximity of markets for the products. A single continuous transportation system for the concentrate is likely to be less expensive than multiple and possibly intermittent shipments of a variety of products.

As is often the case in cost-benefit analysis of complex and speculative systems, it is easier to quantify the costs than the benefits, and the costs will be considered first.

8.1 Transportation Costs for Kerogen Concentrates

Transportation cost obviously depends on transportation mode and the preferred mode will, in turn, depend on the distance, existing transportation systems, and geographic conditions. Highway trucking makes sense for short distances, under-utilized roads, or in conjunction with a railroad that is too far to be reached by belt conveyors. Barge transportation doesn't apply to the Western deposits and is probably limited to a small number of Eastern locations. The two realistic transportation modes are rail and slurry pipeline.

Slurry transportation requires large amounts of water. The problem can be alleviated by building a smaller parallel water pipeline to recycle the concentrate filtrate and/or bring make-up water if it is more plentiful at the recovery site.

The available literature on slurry transportation is several years old and is limited to coal and mineral products with particle sizes above 100 microns. High inflation rates and changing financing conditions introduce errors when the late 1970's costs are updated. The rheological behavior of the minus 20 micron concentrate is not known but that behavior might differ significantly from coal, copper concentrates, ground limestone, and other mineral products.

Published coal slurry transportation costs include slurry preparation at the beginning of the pipeline and dewatering at the end. As these costs are constant, while the pipeline cost varies with length, the tariff in ¢/ton-mile is distance-dependent. Copper concentrate transportation tariffs are not affected by pipeline length because they exclude slurry preparation and dewatering but the costs are not as well documented as the coal data.

A DOE report (Ref. 66) that gives a cost breakdown of coal slurry pipelines was used in this study. The cost was escalated to 1981 and scaled down from a 38-inch to a 20-inch pipe diameter. The latter is a conservative estimate based on hydraulic calculations and assumes a solids concentration of 50 vol. %. Because of the uncertain non-Newtonian behavior of the shale concentrate, several equations published by Wasp (Ref. 67) and Perry (Ref. 68) were used to assist in these calculations.

The result for kerogen concentrate is an approximate tariff of 3¢/ton-mile; it assumes a 1,000 mile pipeline but should be independent of distance above 100 miles because pumping stations are located at 60- to 80-mile intervals. The tariff is in good agreement with Wasp's charts (Ref. 67) considering escalation, contingency, and the 25% annual capital charge used in this study versus the 15% used by Wasp. The parallel water pipeline would increase the tariff to about 4.5¢/ton-mile. A separate study based on industrial experience and References 67, 68, 76, 77 indicated that comparable railroad freight ranges from 2 to 6¢/ton-mile, depending on distance, local conditions, and whether unit trains are used and new track construction is required. A rate of 3¢/ton-mile is equivalent to 1¢/barrel-mile for 100% Fischer Assay recovery of a kerogen concentrate of 126 gallons/ton.

For comparison with transporting oils, Occidental (Ref. 70) estimated 1977 costs of piping shale oil to be 0.7 to 1.4¢/ton-mile at rates of 150,000 tons/day to 30,000 tons/day respectively.

8.2 Benefits of Remote Recovery

This section attempts some illustrative quantifications of the benefits for remote recovery listed in the introductory paragraphs of Section 8. The examples are confined to Western shales because, on balance, plants processing Eastern shales are likely to have net incentives to remain at or adjacent to the ore sites. Unlike the Western situation, the socioeconomic, environmental, and construction problems should be no more severe at Eastern ore sites than at any reasonable remote location.

Construction Site: Process plant construction in the Piceance Creek Basin has been estimated to cost perhaps 40% more than construction of the same facilities at a standard Gulf Coast location. The potential benefit that beneficiation brings to remote recovery is confined to the recovery section (using retorting, for example) because (a) the beneficiation section must remain at the ore site, and (b) in a conventional process sequence, raw shale oil from a retort could also be piped away for remote upgrading and refining if there were a net incentive to do so.

Considering the Base Case with beneficiation (Table 3-7), the capital cost of the recovery (pyrolysis) section plus a prorated share of offsites is $370 million. At a Gulf Coast location, that capital cost might be reduced to $260 million for a gross saving (ex transportation costs) of $110 million or 4% of the original total capital cost of $2950 million.

Integration into New or Existing Refineries: The benefits accruing to a beneficiation system per se are again limited to the recovery step because raw shale oil from retorting can also be transported (and more cheaply than kerogen concentrate) for remote upgrading and refining although reduction in oil pour point is necessary, e.g. by using pour point depressants, dilution, or visbreaking. In a new refinery, a new recovery reactor must be provided as a discrete unit, and no savings can result; there may be savings through integration of offsites. In an existing refinery, there is no way we can predict the availability of idle equipment suitable for the recovery step. An optimistic overall assumption is that no new offsite investment will be required, but that a recovery reactor must be constructed at Gulf Coast costs. That assumption results in a maximum Base Case gross cost reduction of $210 million or about 7% of the original total capital cost.

Labor Requirements and Socioeconomic Problems: In the Western shale areas, population is sparse and community facilities do not exist to serve the large new construction and operating forces required for shale oil plants. A "typical" 50,000 barrel/day plant will require a primary on-site operating labor force of about 1000 to 1500 workers and a total induced population in the region of about 5 to 10 times that size (Ref. 71,72). The provision and caring for large new labor forces (and their families and community infrastructure) results in high economic and social disruption costs. A technology which reduces the region's labor requirements is thus of real value.

A beneficiation-based system with remote recovery affects local labor requirements by:

- Increasing the mining labor force modestly to the extent that kerogen is lost in the recovery step, say 10%, if the mining technology is unchanged. (The mining labor force could be decreased substantially, say 50% or more, if the use of beneficiation to process lean ores makes it desirable to substitute surface mining, with lean strata in the overburden, for underground mining. But that substitution does not depend on remote processing.)
- Adding a new labor force to operate the comminution, separation, and tailings disposal systems.
- Subtracting the labor force required to operate the recovery (pyrolysis) section and spent shale disposal.
- Altering somewhat the labor force needed for offsites.

On balance, the introduction of both beneficiation and remote recovery should have only a small effect if the system includes underground mining. Mining accounts for about half the total work force and is increased, not decreased, by beneficiation. The beneficiation plusses should roughly balance the pyrolysis minuses.

Our conclusion is that remote recovery in a beneficiation-based system will not reduce significantly the on-site labor force or total induced local population. Beneficiation itself could encourage a shift from underground to surface mining, but that shift is likely to be dictated by environmental, site-specific, and traditional economic considerations rather than by socioeconomics.

Environmental Effects: Remote recovery of oil from concentrate can remove from the ore site all wastes and emissions arising from the recovery section (e.g. retorting), displacing them to some other location where environmental constraints and/or impacts may be less severe. There would be no effect on the mining and beneficiation sections attributable to remote recovery, and no effect on the upgrading and refining sections attributable to beneficiation.

For the Base Case including beneficiation, estimated air emissions from the recovery section are shown in the following table. The numbers were calculated from the Colony data shown in Reference 71 reduced by 60% to allow for the smaller retorting section required to handle concentrate rather than ore:

Table 8-1

Estimated Atmospheric Emissions from the Recovery Section in the Beneficiation Base Case (Tonnes/Day)

SO_2	Partic.	NO_x	HC	CO
0.6	0.7	6.2	1.2	0.2

Except for particulates, major atmospheric emissions in a shale oil facility are entirely due to high temperature gas reactions such as combustion which occur primarily in the recovery section (and secondarily in internal combustion equipment and explosives used in other upstream sections). Therefore, remote recovery results in elimination of 90+% of regulated air pollutants except particulates.

The impact of remote recovery on total particulates produced at the site is not clear because (a) we have limited understanding of the production of particulates that would result from introducing the beneficiation section, and (b) data like those of Table 8-1 refer only to controlled particulates and ignore fugitive dust (i.e. any dust that escapes from a source other than a stack or duct) which results from blasting, mining, and other solids handling activities in the open.

Remote recovery also results in eliminating site production of retort waste water and of spent shale. That elimination is not likely to be a major consideration since the facility can be designed for zero discharge of waste water, and spent shale can be mixed with tailings (as is assumed in Sections 3 and 5) for disposal.

8.3 Conclusions

There clearly are benefits associated with processing shale oil at a location remote from the rugged, dry, sparsely populated, and environmentally pristine Western ore sites. Many of the benefits accrue to the remote upgrading and refining of raw shale oil produced by conventional surface or in situ retorting. Whether or not the costs of remote transport offset those benefits goes beyond the scope of our study. The cost-benefit ratio could be very attractive in principle for an industry producing a million barrels a day pipable to large refining complexes with idle capacity. But shale oil plants will be built project-by-project, firm-by-firm. Realizing the big picture attractiveness is not obvious from the perspective of the individual project owner participating in a slowly evolving industry.

The introduction of beneficiation enables remote processing to move upstream one more step to include conversion of kerogen to oil as well as oil upgrading and refining. However, the addition of that step substantially increases the cost of remote transport (due to moving a water slurry rather than an oil) and yields modest new benefits: little or no labor saving and socioeconomic relief, and small potential construction savings, but a major reduction in the regional burden of air emissions other than particulates.

On balance, the option to recover oil from kerogen remotely may be helpful in some circumstances but it is not likely to provide a major incentive to switch to a beneficiation-based system.

9. RESEARCH AND DEVELOPMENT OPPORTUNITIES

Our conclusion from the preceding sections of this report is that our present state of knowledge about comminution, separation, and oil recovery of oil shale does not justify a major development effort aimed at a system incorporating high-enrichment beneficiation that would be obviously attractive compared to systems without beneficiation. But our present state of knowledge is primitive in many respects. We think that a modest program of basic and applied research _is_ justified and could lead to a more optimistic view of the prospects of beneficiation systems. There seems to be no theoretical barrier to major improvement. Therefore, this section briefly notes some initial technical questions which warrant answering through research, with development programs and other research (on tailings disposal, for example) to follow if those questions are answered favorably.

Comminution: Most capital and operating costs incurred by introducing high-enrichment beneficiation are due to the comminution section. As noted in Section 4 and Reference 21, comminution is a poorly understood and extremely inefficient unit operation. Major improvements in comminution could not only make shale beneficiation competitive but could lead to new opportunities for deep cleaning of coal and to new processing techniques for other ores. Recent reports of advances in commercial comminution equipment (e.g. Allis-Chalmers, Ref. 80) lead to optimism that major improvements are achievable and would go a long way toward making high-enrichment beneficiation competitive.

Comminution research should focus on determining the minimum amount of energy required to fracture different ores and on understanding how to apply forces to the ore to impart that energy with greatest efficiency. Understanding of that type is a prerequisite to designing comminution devices if we are to have any reasonable chance of major practical improvements in energy consumption, where "major" means a reduction of at least a factor of two or (we hope) much more. (Also, see Section 4.7, p. 4-39.)

Characterization: Research in the area of characterization should answer two fundamental questions.

First, where is the kerogen located in oil shale powders liberated by various methods of comminution? A perfect separation cannot be obtained without perfectly liberated kerogen. Smearing of kerogen on mineral surfaces, as the wettability separations suggest, decreases separation efficiency. Various comminution methods should show decreased kerogen smearing and, as such, improved separation efficiency.

Second, what other physical properties could be used to separate kerogen from oil shale? And at what temperature is the physical property difference the greatest? To answer these questions the physical properties of kerogen and minerals should be measured separately at various temperatures. Emphasis should be on the physical properties which could be used for separations (i.e., conductivity, magnetic susceptibility, absorptivity, etc.).

Separation: Given the conclusion that comminution is the most expensive part of benefication, it seems unlikely that research on high enrichment separations is called for until more economical comminution methods are available--especially because comminution and separation are not independent steps. However, if an efficient dry separation was available, the economic picture would be greatly improved. For this reason, further research is warranted on dry separations requiring little grinding. Examples of this kind of separation are hard to imagine, but the stickiness separation may be in this category.

Recovery: Bench-scale research is warranted on supercritical extraction of kerogen concentrate; the concentrate can be prepared by existing laboratory methods like those discussed in Section 5. Experiments on yield and quality as a function of solvent type and operating conditions (ratios, time, temperature, etc.) should result in data that will permit confident estimates of whether supercritical extraction really is an attractive alternative to retorting. The prospects of the other recovery methods discussed do not look promising enough to justify research at this time.

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Appendix A

UPDATING AND EXTENDING FACTORED CAPITAL ESTIMATION

Ivan K. Klumpar
Energy Laboratory, Massachusetts Institute of Technology
Cambridge, MA

Excerpts from paper presented at National Meeting of the American Institute of Chemical Engineers, Orlando, FL, March 1982, Microfiche No. 24b.

Capital cost of industrial facilities is estimated in various ways depending on the engineering information available, accuracy required, and effort justified. In estimating work, the plant is always broken down into items, the various methods differing in the structure and detail of this breakdown. In the process industry, most methods divide plants into process equipment, and bulk or commodity materials such as piping, brick or control instruments. Individual equipment and commodities have a material and labor component, the latter representing the man-hours to set the equipment unit or install the particular commodity material. In addition, earthwork, freight, indirects and other items are accounted for.

In preliminary estimating, the most widely used technique is the factor method. Process equipment cost is based on telephone quotations, similar units in previous jobs, or literature. Commodities are estimated on multiplying equipment costs by specific factors. Overall factors are used for buildings, utilities, and other non-process "systems" without breaking these items down into equipment and commodities. Additional factors apply to site development, freight, indirects, and other cost.

Capital estimation involves several steps. In each step a group of items is added. The total of the previous step is the basis for the factors used in the next step. Alternatively, some items may be estimated as functions of capacity, e.g., buildings costs as floor area times a factor, or power supply costs in proportion to kWh requirements. A typical sequence of steps is shown in Table I. The less common items are defined in Exhibit I. The items of the last step are listed only for completeness but won't be discussed in this paper.

There are many variations of the factor method. Many cost items shown in Table I are often divided into subitems listed in Exhibits I and II (first column). Some variations use different factor bases for items and subitems that are included in a single step in Table I. For example, it was proposed to estimate utility supply subitems based on the sum of installed equipment cost, process buildings and general facilities. It appears to be most logical to use the sum of purchased equipment cost as the base for capitalized spare parts but to include them in miscellaneous direct cost. However, some experts incorporate spares in working capital. The worst inconsistencies among the variations of the factor method are in the area of miscellaneous direct and field indirect costs.

The inconsistencies of assigning cost items to various computational steps may cause misinterpretation of cost data and make the use of many data sources difficult if not impossible. To remedy this problem, a standardization effort within the American Association of Cost Engineers has been initiated in conjunction with this study. In development of capital estimation models, it is of utmost importance to properly define all items. Moreover, the method should be made flexible enough to allow for alternative computational routes and extensive use of outside data.

Some literature data on capital estimation factors have not made the distinction between the material and labor component of individual cost itesm. This approach was adequate in the old days of low inflation and as long as the application of the factor method was limited to certain traditional petroleum and petrochemical geographic areas. However, high inflation rates in the early 1970's brought along significant differences among the escalation rates of various equipment, material and labor categories. Also, new government regulations have affected some cost items more than others. The result has been a distortion of the factors developed in the 1950's and 60's. In addition, as the method has been used for estimating plants located in different geographic areas, the deviations in labor productivity and wage rates have caused gross errors.

Most cost items can be separated into a material and labor component

$$C = K_M + K_L \qquad (2)$$

where

$$K - fB \qquad (3)$$

There are expections, such as capitalized spare parts which don't have a labor component. The material factor base may differ from the labor factor base. For example, Hackney bases commodity labor on commodity material rather than on purchased equipment. All costs have to refer to the same year. For instance, if purchased equipment costs from different sources are used as the factor bases, they have to be calculated or de-escalated to the same reference year,

$$B = Ey = (I_y/I)\ E \qquad (4)$$

where E and I refer to any year while the subscript y denotes the reference year.

The "f" in Equation 3 is a "dollar" factor because its dimension is \$/\$. For labor components, an "hour" factor, h, is preferable. The two factors are related by the following equation:

$$f = w(p/p_0)h \qquad (5)$$

where productivity, p, and wage rate, w, refer to the actual geographic area while p_0 refers to standard conditions on which h is based. Over a limited period of time, the material dollar factor should be independent of inflation and geographic location for all practical purposes. The labor dollar factor should be insensitive to inflation but affected by geography while the opposite is true about the labor hour factor.

It is proposed to revise the factors developed here before 1975 and keep updating them particularly during high inflation periods. As te first step in this direction, factors were compiled from projects of major engineering companies issued in the 1975 through 1980 period which had a relatively low inflation rate. Commodity material and installation labor hour factors for selected equipment types are reported in Tables II and III respectively. The term installation labor denotes the sum of setting and commodity crafts. Gulf Coast has been selected as the standard geographic area and mid 1980 as the reference year for the material cost base in Table III. The corresponding labor dollar factors are also shown in Table III using Southern California as the geographic area and 1975 as the reference year for comparison. Table IV lists the implied wage rates and cost indices. "Gulf Coast" is used to denote the geographic area around the mouth of the Sabine River in Texas and Louisiana. Table V presents factors for major miscellaneous direct and indirect items.

SYMBOLS

A	operational additional input (various dimensions)
B	factor base such as purchased equipment cost, commodity cost or the sum of previously established cost items ($)
C	computed cost item ($)
E	purchased equipment cost which is also the basic input ($)
I	cost index (no dimension)
K	material or labor component ($)
S	size, capacity or capacity parameter (various dimensions)
$a,b,c,$	equipment cost correlation coefficients (no dimension)
f	capital estimation factor ($/$)
p	average construction labor productivity in a geographic area (no dimension)
q	quality assurance transform coefficient (no dimension)
r	radiation protection transform coefficient (no dimension)
s	seismic protection transform coefficient (no dimension)
w	average wage rate in a geographic area ($/man-hour)

Subscripts

E	equipment
K	material or labor component
L	labor
M	material
y	reference year
0	standard geographic conditions

Table I

INVESTMENT COMPUTATION STEPS

Step	Cost Item
1.	Purchased equipment + Setting Labor + Commodities
2.	Total-Installed equipment + Process buildings + Utility supply + Utility distribution + General facilities
3.	Total-Plant direct + Site development + Miscellaneous direct
4.	Total - Direct + Field indirect
5.	Total - Field construction + Engineering and home office + Project management + Owner's cost
6.	Total-Depreciable capital excluding contingency + Contingencies
7.	Total-Depreciable Capital + Land and other non-depreciable capital + Working capital + Start-up (optional) + Investment expense
8.	Grand total - investment

Table II

COMMODITY MATERIAL FACTORS

Equipment Category and Type(if needed)	Factors As Percent of Purchased Equipment Cost							
	Concrete Foundation	Piping	Steel Supports	Instrumentation	Insulation	Electrical	Painting	Total Commodities
1	2	3	4	5	6	7	8	9
Agitator	-	-	-	4	-	24	1	29
Autoclave	6	43	4	6	5	5	1	70
Bin, hopper	6	-	15	5	4	2	1	33
Blower	7.5	17	-	3.5	-	34	1	63
Centrifuge	9.5	25	10	10	4.5	27	1	87
Compressor	6.5	22	8	7	3	29.5	1	77
Conveyor	12.5	-	25	7.5	-	20	1	66
Crane	10	-	30	-	-	20	2	62
Crystallizer- See Evaporator								
Cyclone	2	9	22	3	10	47	1	94
Dryer	10	10	10	6	-	20	1	57
Dust Collector	22	9	13	9	10	36	1	100
Ejector	-	60	5	-	5	-	1	71
Electrolytic Cells[1]	6	12	-	2	-	-	-	20

(Table continued on next page)

Table II (continued)

COMMODITY MATERIAL FACTORS

Equipment Category and Type(if needed)	Concrete Foundation	Piping	Steel Supports	Instrument-ation	Insulation	Electrical	Painting	Total Commod-ities
1	2	3	4	5	6	7	8	9
Evaporator (Concentrator or crystallizer)	6.5	40.5	6.5	12.5	5	5	1	77
(Long tube or forced circulation)	9	110	1	16	3	-	0	139
Fan	10	-	12.5	2.5	-	25	1	51
Feeder	5	-	20	13	-	25	3	66
Filter (Plate and Frame)	4	31	-	-	3	-	1	39
(Other)	14	60	15	5	-	10	5	109
Heat Exchanger	9	48	7	8	3	2	1	78
Pump	5	32	-	3	-	33	1	74
Screen	10	-	10	8	-	20	1	49
Scrubber (Peabody, Croll-Reynolds, etc.)	7	11	6	7	-	6	3	40
(High energy Venturi)	35	9	15	10	10	6	1	86

(Table continued on following page)

Table II (continued)

COMMODITY MATERIAL FACTORS

Equipment Category and Type(if needed)	Concrete Foundation	Piping	Steel Supports	Instrumentation	Insulation	Electrical	Painting	Total Commodities
1	2	3	4	5	6	7	8	9
Tank								
(Atmospheric)	9	47	7	8	5	4	1	81
(Pressure, horizontal)	7	42	-	5	5	4	1	64
(Pressure, vertical)	11	63	9	9	9	4	1	106
Thickener	15	29.5	9.5	2.5	-	6.5	1	64
Tower	12	54	9	11	9	5	2	102
Weighing equipment	4	-	4	-	-	4	2	14

[1] Electric commodity material for electrolytic cells is included in equipment.

TABLE III - Installation Labor Factors

Equipment		1980 Gulf Coast Labor Hour Factor For:			1975 Southern California Labor Dollar Factor For:		
Category	Type Specification[a]	Equip. Setting	Commodities	Commodities	Equip. Setting	Commodities	Commodities
		In Man-Hours/$100 of:			In % Based on $'s of:		
		Equipment	Equipment	Comm. Matl.	Equipment	Equipment	Comm. Matl.
1	2	3	4	5	6	7	8
Agitator	Turbine or propeller, SS or rubber-lined CS	0.7	1.4	4.9	12	24	83
Autoclave	--	0.4	2.8	4.0	8	48	69
Bin, hopper	--	0.3	1.4	4.2	6	22	67
Blower	Rotary	0.2	2.6	4.1	4	40	63
Centrifuge	--	0.6	3.5	4.0	12	57	66
Compressor	--	0.2	2.9	3.8	5	56	73
Conveyor	Belt or screw	0.9	4.4	6.7	17	70	106
Crane[b]	--	0	3.6	5.9	0	57	92
Crystallizer	SS	0.4	3.4	4.4	8	57	74

(more)

TABLE III - Installation Labor Factors (continued)

1	2	3	4	5	6	7	8
Cyclone	--	0.5	3.0	3.2	9	51	54
Dryer	Rotary	0.7	5.7	10.0	15	90	158
Dust Collector	Bag house	0.4	6.3	6.3	8	101	101
Ejector	SS	0.1	7.3	10.3	2	126	177
Electrolytic Cell[c]	Lined concrete	0.6	0.6	3.1	12	9	45
Evaporator	Concentrator, brick lined	0.3	2.9	3.8	6	50	65
	Long tube or forced circulation	0.5	10.3	7.5	10	178	128
Fan	--	0.6	2.3	4.6	12	37	73
Feeder	Rotary cup, SS wet parts	0.5	2.9	4.4	9	47	71
	Rotary vane, SS wet parts	1.1	3.8	5.7	20	61	92
	Vibrating pan, SS wet parts	0.4	3.0	4.6	7	49	74
Filter	Plate & frame	0.2	1.0	2.6	4	17	44
	Rotary drum	0.6	1.0	0.9	12	17	16
Heat Exchanger	Fixed tube sheet, CS/SS	0.1	3.1	4.0	1	53	68
	Plate, SS	0.6	2.9	3.8	13	50	64
	Shell & Tube, CS	0.1	4.9	6.2	12	79	100
	Shell & Tube, SS	0.1	2.9	3.7	1	49	63

(more)

Table III - Installation Labor Factors (continued)

1	2	3	4	5	6	7	8
Pump	Centrifugal	0.6	3.0	4.1	12	58	78
	Gear	0.5	3.0	4.0	10	57	77
	Metering	1.1	3.7	5.0	21	71	96
	Positive displacement	0.6	2.7	3.7	12	52	70
	Rotary	0.6	3.0	4.0	13	57	77
	Vertical sump	1.1	3.7	5.0	21	71	96
Screen	Stationary	1.0	3.9	7.9	17	61	124
	Vibrating	0.6	2.2	4.6	11	35	71
Scrubber	Peabody, Croll-Reynolds, etc.	0.5	2.2	5.5	10	34	85
Tank	Agitated	0.2	4.1	5.0	5	69	85
	Process[d], SS, vertical	0.1	4.2	4.0	2	71	67
	Process, lined concrete	0.5	4.3	5.3	9	73	90
	Storage, CS, horizontal	0.3	3.1	3.9	7	52	64
	Storage, CS, large vertical[e]	0.2	3.1	3.9	5	52	64
Thickener	--	6.9	3.3	5.1	132	52	81
Tower	Tray or packed	0.5	4.6	4.6	11	78	76
Weighing Equipment	Scale	0.6	4.1	29.0	10	61	436

(more)

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[a] If not indicated otherwise, material of construction is carbon steel (CS).

[b] Setting cost is included in equipment cost.

[c] Electric labor is included in equipment setting.

[d] For CS, use horizontal storage tank data with increased instrumentation.

[e] For tanks below 10,000 gal, use horizontal tank data.

Table IV - Wage Rates and Cost Indices

Construction labor cost index is 166 and 200 for May 1975 and June 1980, respectively.

	Wage Rate,*	Material Cost Index**	
Equipment and Material Type	$/man-hr	May '75	June '80
Fabricated Equipment		191	291
Autoclave, bin, crystallizer, cyclone, dust collector, evaporator, tank, tower	10.30		
Ejector, scrubber, heat exchanger	10.60		
Process Machinery		183	272
Agitator, feeder, screen, weighing equipment	9.10		
Centrifuge, dryer, filter	10.30		
Conveyor, thickener	10.10		
Pumping Equipment		206	332
Blower, fan	10.30		
Compressor, pump	9.10		
Electrical Equipment		142	205
Electrolytic cell	10.40		
Commodities			
Concrete foundations	7.10	176	237
Piping, valves, and fittings	9.00	217	329
Steel supports	8.10	196	298
Instrumentation	9.00	179	247
Electricals	8.90	142	205

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* Southern California, May 1975

** Chemical Engineering Magazine

Table V

INDIRECT AND MISCELLANEOUS DIRECT FACTORS

Cost Item	Base	Percent	
		Range	Average
Field personnel*	Total Direct	4-21	14
	Direct Labor	30-270	73
Other Field Indirect*	Total Direct	7-44	21
	Direct Labor	8-330	112
Engineering and Home Office	Field Construction	3-31	13
Project Management	Field Construction	2-20	6
Owner's Cost	Field Construction	0-5	4
Freight**	Plant Direct	4-10	7
Other Miscellaneous Direct**	Plant Direct	1-11	6

*Subitem of field indirect cost, see Table I.

**Subitem of miscellaneous direct cost, see Table I.

EXHIBIT I -- DEFINITIONS OF SOME COST ITEMS

PURCHASED EQUIPMENT COST

FOB manufacturer, i.e., excluding freight.

COMMODITIES

Concrete foundations, steel supports, piping including valves and fittings, electricals including panels, instrumentation including panels, insulation, painting, and miscellaneous (e.g., individual equipment fire protection).

UTILITIES

Electric power, fuels, process and cooling water, steam, plant air, heating and cooling fluids (such as brine or dowtherm), etc.

UTILITY SUPPLY

Facility for utility generation, regeneration or receiving "across the fence." E.g., power plant and/or main transformer; water treatment plant and cooling tower; air compressor station.

UTILITY DISTRIBUTION

Power lines and pipelines between utility supply facility and consumption areas including substations, pressure reduction stations, etc., but excluding

EXHIBIT I (Continued)

connections to equipment units.

GENERAL FACILITIES

Maintenance shops; warehouses; receiving storage and shipping facilities [1]; central control rooms; garages; administrative buildings; cafeterias and change houses; first-aid stations; waste disposal facilities [2]; general fire protection systems [3]; laboratories; communication systems [3]; computer facilities [3]; etc.

SITE DEVELOPMENT

Site cleaning; earthwork dredging and piling; landscaping; fences; roads, trades and walkways; sewers; parking and other paved areas; yard lighting; etc.

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[1] Alternatively considered part of process sections.

[2] Alternatively considered process sections (e.g., SO_2 scrubbing), utilities (e.g., waste water treatment for possible recirculation), site development (e.g., tailings ponds, wharfs) or service systems[3] (blow down lines and stacks.

[3] Alternatively combined with utility distribution into "service systems."

EXHIBIT I (Continued)

CONTINGENCIES

It is convenient to distinguish process contingency from definition contingency. The former expresses uncertainties in technology while the latter refers to potential problems of translating technology into hardware and unexpected events during construction.

OTHER TERMS

For miscellaneous direct, field indirect, and engineering and home office see Exhibit II.

EXHIBIT II -- Sheet 1

INDIRECT AND MISCELLANEOUS DIRECT COST DISTRIBUTION

ITEM	Process Industry Samples									Nuclear Ind. Samples				
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
INDIRECTS														
TEMPRОARY CONSTRUCTION*														
BUILDINGS & UTILITIES										X	X	X		
ROADS														
FENCES										X		X		X
RENTALS (TRAILERS, TOILETS, ETC)														
FIRST AID & SAFETY														
POWER & UTILITY BILLS														
WEATHER PROTECTION														
UNALLOCABLE EQUIPMENT & TOOLS														
RENTAL CONSTRUCTION EQUIP.*		X									X	X		
PURCHASED CONSTRUCTION EQUIP.* SUBT. - CONSTRUCTION EQUIP.				X						X				
SMALL TOOLS		X		X							X	X		X
CONSTRUCTION CONSUMABLES				X						X				
TRANSPORTATION & HANDLING													X	
UNALLOCABLE LABOR														

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* ERECTION (OR RENTAL), AND MAINTENANCE COST (IF APPLICABLE)

EXHIBIT II -- Sheet 2

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
FIELD NON-MANUAL PAYROLL														
CLERICAL														
ENGINEERING										X				
SUBT-FIELD STAFF											X			
SUPERVISION														
FRINGE BENEFITS														
SOCIAL SECURITY														
UNEMPLOYMENT INSURANCE														
WORKMEN'S COMPENSATION														
HEALTH INSURANCE														
OTHER														
SUBT.-PAYROLL INS. & TAXES		X		C*										
PREMIUMS (OVERTIME, ETC.)														
HOLIDAYS														
VACATION														
SICK TIME														
OTHER NON-PRODUCTIVE														
SUBT.-NON-PROD. PAYROLL		X		C*										

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* FOR CRAFT LABOR ONLY.

EXHIBIT II -- Sheet 3

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
<u>MISCEL. FIELD EXPENSE</u>														
NON-PAYROLL TAXES		X			X								X	
NON-PAYROLL INSURANCE				X							X		X	
SUBT.-TAXES & INSURANCE														
TRAVEL & SUBSISTENCE		X		C*							X	X		
MOVING & RELOCATION												X		
PERSONNEL TRANSPORTATION				C*										
SUBT.-PERSONNEL RELATED														
TESTING		X			X					X			X	
VENDOR TECHNICIANS					X									
OUTSIDE ENGINEERING					X									
OFFSITE STORAGE				X										
SUBT.-FIELD SERVICES														
TELECOMMUNICATIONS												X		
POSTAGE												X		
COMPUTER USE												X		
OTHER OFFICE EXPENSE												X		
SUBT.-OFFICE EXPENSES													X	
GUARDS & SECURITY														
COMPANY CHARGES				X										

EXHIBIT II -- Sheet 4

FIELD INDIRECTS SUMMARY

		1	2	3	4	5	6	7	8	9	10	11	12	13	14
GUARDS & SECURITY	A														
OFFICE EXPENSE	B													X	
BURDEN	C=A+B														
FIELD SERVICES	D											X			
BURDEN & SERVICES	E=C+D	X													
PERSONNEL RELATED	F														
NON-PAYROLL TAXES & INSUR.	G													X	
COMPANY CHARGES	H														
MISC. FIELD EXPSE.	I=E+F+G+H														
FRINGE BENEFITS	J					X								X	
OVERHEAD	K=F+G+J	X			X							C*			
FRINGES & MISCEL.	L=I+J			X								X	X		
FIELD STAFF (NON-SUPERV.)	M														
SUPERVISION	N		X	X										X	
NON-MANUAL	O=M+N	X											X		
FIELD ADMIN.	P=E+O				X							X			
UNALLOCABLE LABOR	Q	X	X		X										
FIELD PERSONNEL	R=M+Q			X											
UNALLOCABLE EQUIP. & TOOLS	S	X		X								X	X	X	
TEMPORARY CONSTRUCTION	T	X	X	X	X							X	X	X	
FIELD EXPSE.	U=C+G+H+O+Q+S+T					X									
FIELD INDIRECTS	V=U+D+F+J										X				

EXHIBIT II -- Sheet 5

		1	2	3	4	5	6	7	8	9	10	11	12	13	14
ENGINEERING											X			X	X
PAYROLL & FRINGES*	A			X											
MATLS. & OUTSIDE SERVICES	B														
PERSON. RELATED & OFFICE EXPSE.*	C	X													
SUBT.-"DIRECT"	D=A+B	X													
SUBT.-MISCEL.	E=B+C			X											
PROCUREMENT	F	X													
OTHER HOME OFFICE DEPTS															
PAYROLL & FRINGES*	G			X								X	X		
PERSON. RELATED & OFFICE EXPSE*	H			X								X	X		
OTHER CONTRACT COST	I	X		X								X	X		
SUBT.-"DIRECT"	J=G+I														
SUBT.-MISCEL,	K=H+I														
SUBT.-ADMIN.	L=G+I	X													
ENGINEERING & HOME OFFICE SUMMARY															
"DIRECT"	M=D+F+J				X										
OVERHEAD	N=C+H				X										
TOTAL	O=M+N														

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* FOR BREAKDOWN OF FRINGES, PERSONNEL RELATED AND OFFICE EXPENSE, SEE FRINGE BENEFITS AND MISCELLANEOUS FIELD EXPENSE ABOVE.

EXHIBIT II -- Sheet 6

	1	2	3	4	5	6	7	8	9	10	11	12	13	14
SUMMARY OF INDIRECTS														
FIELD INDIRECTS							X		X	X				
EXCLUDING SUPERVISION						X								
INCLUDING FREIGHT														
SUPERVISION														
PROCUREMENT							X							
SUPERV. & PROCUREMENT						X								
FREIGHT								X						
PROJECT MGMT. INCLUDING FEE	X	X	X	X						X	X			
BOND		X												
START-UP			X							X			X	
OWNER'COST														
ENGINEERING & HOME OFFICE		X												
EXCL. PROCUREMENT, INCL. FEE						X	X							
INDLUDING FEE					X				X					
TOTAL INDIRECTS														
EXCLUDING FREIGHT								X						

EXHIBIT II -- Sheet 7

ITEM	PROCESS INDUSTRY SAMPLES									NUCLEAR IND. SAMPLES				
	1	2	3	4	5	6	7	8	9	10	11	12	13	14
<u>**MISCELLANEOUS DIRECT**</u>														
<u>CRAFT RELATED (SEE ALSO INDIR.)</u>														
PAYROLL INSURANCE & TAXES*		X		X									X	
NON-PRODUCTIVE PAYROLL*				X										
SUBT.-FRINGE BENEFITS	X													
TRAVEL & SUBSISTENCE	X													
COMMOD. & S/C LABOR ADJUST.					X									
<u>FREIGHT</u>				X										
<u>TESTING</u>														
<u>QUALITY ASSURANCE</u>														
<u>NON-PAYROLL TAXES</u>						X								
<u>START-UP</u>				X						X			X	
<u>OTHER DIRECT</u>														
SPARE PARTS				X										
INITIAL OPER. SUPPLIES				X										X
DEMOLITION & DISMANTLING				X							X			
SOIL & WATER INVESTMENT				X										

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* FOR BREAKDOWN, SEE INDIRECTS

Appendix B

RAPID CAPITAL ESTIMATION BASED ON PROCESS MODULES

Ivan V. Klumpar, Richard F. Brown, and Jörg-W. Fromme
Massachusetts Institute of Technology, Cambridge, MA

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Review and Improvement of Existing Methods

There is a great need for approximate capital cost estimates of processing plants based on block diagrams and descriptions that define the unit operations but do not specify all equipment. A case in point is cost and profitability evaluations at early stages of an R&D project when not enough process data are available. Another example is economic analyses that include a process or several processes as an element not requiring a detailed flowsheet and process design because the time available or accuracy of other elements do not warrant the effort. As a matter of fact, quick low-accuracy estimates of ± 30% and up play an important role in cost engineering, although it is not readily admitted.

Several methods have been proposed for these types of approximate estimates [1, 2, 3, 4, 5, 6, 7]. In essence, they determine investment, I , as a function of plant capacity, temperature, pressure, material of construction, and the number of process modules that constitute the process. The latter variable is a specifically defined group of equipment. Additional parameters were also introduced. In mathematical form,

$$I = f(K, t, p, M, N, A) \tag{1}$$

Symbols are listed at the end.

Table I summarizes the main features of the published methods. While all of them were proposed in the form of algebraic equations, certain variables are defined by graphs or tables by some authors. It can be seen that, for capacity, temperature, pressure, and material of construction, either a value specific for the process or a weighted average is used. The specific value may be product flowrate , maximum temperature or pressure, and an overall rating of material construction. Averaging is based on stream flowrates, number of process modules, or a combination of variables. Flowrates are expressed in mass, volume, or moles.

What makes this type of approximate capital cost estimating unique is the concept of characterizing the process by the number of normalized modules. Most authors use Zevnick's and Buchanan's [1] "functional units" that roughly correspond to the traditional unit operations of Chemical Engineering. Other authors use process "steps" based on process chemistry or "basic items" that represent major equipment such as columns and storage tanks while neglecting minor units such as pumps and head tanks.

The "module concept" makes the methods apparently general and easy to use but introduces a fundamental uncertainty. Zevnick and Buchanan define their functional units only by way of examples. The breakdown of the process into modules requires experience and makes the technique rather subjective. Taylor's [6] chemical steps may considerably vary in the number and type of equipment that they include. Allen's and Page's [4] basic items are far less controversial but, to a certain degree, defeat the purpose of the method because they require developing a flowsheet and selecting equipment.

Another problem of the module concept is that simple counting of modules does not account for the varying complexities of processes and the differences in the equipment used. Most of the authors try to resolve the complexity issue by introducing additional parameters listed in the last column of Table I, but only Allen and Page have considered the effect of different equipment types. The problem with the more sophisticated improvements of the simple Zevnick-Buchanan method is that they require more information on the process and are more time consuming.

All published methods have a common disadvantage in that they were tested for a relatively narrow range of capacities and limited segment of the process industries. Most methods apply to the petroleum and chemical industries. Bridgwater's [5] specialization in metal extraction processes is unique. The limitation to a narrow capacity range and a specific class of processes explains the surprisingly high accuracies claimed by the authors, in one case as favorable as ± 15%.

The present study had the following objectives:

- Define the process modules precisely.
- Develop a quick and simple method that requires minimum process data.
- Make the method applicable to a wide range of process industries and plant capacities, but particularly to the insufficiently covered area of solids processing such as ore and coal beneficiation, coal and oil shale conversion, and extractive metallurgy.

The study is discussed in the subsequent sections. Details can be found elsewhere [8].

Process Module Definitions

Eleven types of process modules are defined in this study based on the change of a major parameter. An additional module type is assigned to storage. The twelve process module types are characterized in Table II. As a rule, a module includes a main piece of equipment such as absorption column and auxiliary units such as pumps, tanks and heat exchangers. However, the parameter change alone is not sufficient for a precise module definition. While it may exclude some auxiliary equipment units, it needs additional conventions to distinguish between single and multiple units as well as small and large equipment such as tanks or conveyors. An additional problem is overlapping in operations and equipment. For example, phase change operations such as evaporation are almost always accompanied by heat exchange, and columns and stirred tanks are included in Type 9, 11, and 12 modules.

The following rules have been developed to avoid ambiguity in module definitions with Table II used to establish a hierarchy of modules:

- A piece or group of equipment is assigned a module type number according to the highest parameter being changed. For example, leaching due to a chemical reaction is Type 12, while leaching caused by selective physical dissolution of a mineral component is Type 11. The dissolution of a pure substance is classified as Type 9.
- A heat exchanger is considered part of a higher ranking module if one of its streams is fully associated with that module, and the other stream is a utility stream or a process stream associated with the same module at least partially.

A stream is fully associated with a module if it flows from and to it, while it is partially associated with a module if only its origin or destination is that module. Examples are a water cooled absorbent cooler of an absorption tower or a distillation column reflux condenser that exchanges heat with cold column feed.

- A heat exchanger is considered a Type 1 module if it is between other independent modules, i.e., if each of its streams is fully or partially associated with another higher ranking module. An example is a water cooler of a reactor product to be separated in a distillation column.
- The material handling modules of Type 3 may include several pieces of equipment in a straight path, such as a bin, feeder, and several conveyors constituting a transport system. However, a branched system has to be broken down into two or more straight series. Also, fixed transport modules of the bin-feeder-conveyor variety have to be classified separately from mobile systems such as truck fleets.
- Systems of tanks and pumps with more than one inlet and more than one outlet streams are considered independent Type 4 modules. Other equipment of that type is included with an adjacent module. Details are discussed elsewhere [8].
- Tanks, silos, bins, and hoppers for the storage of a raw material, product, or a group of auxiliary process materials constitute separate Type 8 modules. They have to be distinguished from surge vessels such as hold-up or head tanks, and day bins or feed hoppers, which are combined with the adjacent module that they serve. Surge vessels for raw materials, products,

and auxiliary process materials, if used, obviously have smaller volumes than their storage counterparts. All intermediate product tanks, bins, and hoppers are, as a rule, classified as surge vessels rather than Type 8 storage modules. An exception might be storage between plant sections with a hold-up of more than a day.

- Parallel units performing the same operation listed in Table II are considered a single module. For example, two trains each including a stirred tank reactor, filter, and liquid-liquid extraction column with all the required pumps, surge tanks, and a single filter cake conveyor constitute three modules (Types 12, 9, and 11, respectively).
- A multistage system performing one Table II operation, such as a cascade of leach tanks, is considered a single module. On the other hand, if two or more operations are involved in each stage, every operation in each stage is a module. For example, a four-stage countercurrent decantation system that includes a stirred tank and thickener in each stage constitutes eight modules, total.

Data Sources and Correlations

Twenty plants listed in Table III were selected for this study. It was attempted to cover processes involving solids, liquids, and gases, and the combination of these phases with an emphasis on the extraction of natural resources and related technologies. For each plant a complete engineering documentation was available consisting of detailed flowsheets, a process description, equipment specifications, and costs based on quotations, and an itemized plant cost estimate. The latter included installation material and labor, buildings, and other direct battery limits cost. The twenty documentation packages were prepared by various reputable engineering companies. A wide range of capacities and other parameters were covered; see Table IV. It is believed that the capacity and cost ranges are wider than those in the literature.

Published studies indicate that, in addition to the number of process modules, plant cost is most sensitive to capacity. Accordingly, special consideration was given to the measuring of plant capacity in this study. Both volumetric and mass flowrates were used for this purpose. Molar flowrate was ruled out because of the difficulty of establishing average molecular weights of such natural materials as coal or ore. As mass flowrate has given better correlations, volumetric flowrate won't be discussed here. Plant capacity is expressed in terms of average throughput, i.e., the sum of all module throughputs divided by the number of modules. The module throughput, in turn, is the sum of the individual flowrates of all module inlet and outlet streams divided by two. Utility streams are not counted, except for Type 1 modules. The advantage of this averaging method is its simplicity and applicability to processes with great differences in raw material and product flowrates, with large recycles or with a product that cannot be measured in terms of flowrate. The latter case is represented by nuclear power plants. An example of a process with a large raw material to product ratio is the mining and extraction of copper. Another advantage of the throughput averaging method is that it can be used even if a few minor module inlet and outlet flowrates are not known. The error caused by the omission of such streams is insignificant in most cases.

In some cases, approximate estimates of minor stream flowrates suffice.

Battery limits direct costs, C , of the twenty plants were first correlated by the following modified DeCicco equation:

$$C = kFNG^{e} \tag{2}$$

The correlation is plotted in Figure 1. This method postulates two constants, the proportionality factor, k , and the exponent, e . The complexity factor is a variable defined as

$$F = 2(10^{T+P+M}) \tag{3}$$

where factors T and P are based on the extreme temperature and pressure, respectively, while M corresponds to the average corrosivity of the process: where

$T = 1.8 \times 10^{-4}(t-27)$ for $t \geqslant 27°$ C (hot process) (4)

$T = 2.0 \times 10^{-3}(27-t)$ for $t < 27°$ C (cold process) (5)

$P = 0.1 \log p$ for $p \geqslant 1$ atm (pressure process) (6)

$P = 0.1 \log (1/p)$ for $p < 1$ atm (vacuum process) (7)

M ranges from 0 to 0.4; see Table V.

The straight line in Figure 1 was established using the GLM regression procedure [9]. The constant factor, k , and exponent, e , together with the statistical data are summarized in Table VI. The 0.57 exponent is close to the 0.6 proposed by DeCicco for capacities falling within the range of the present study.

The following four attempts were made to improve the Equation 2 correlation:

A. The constant exponent, e, was replaced by a variable. This technique gave best results and will be discussed in detail in the next section.

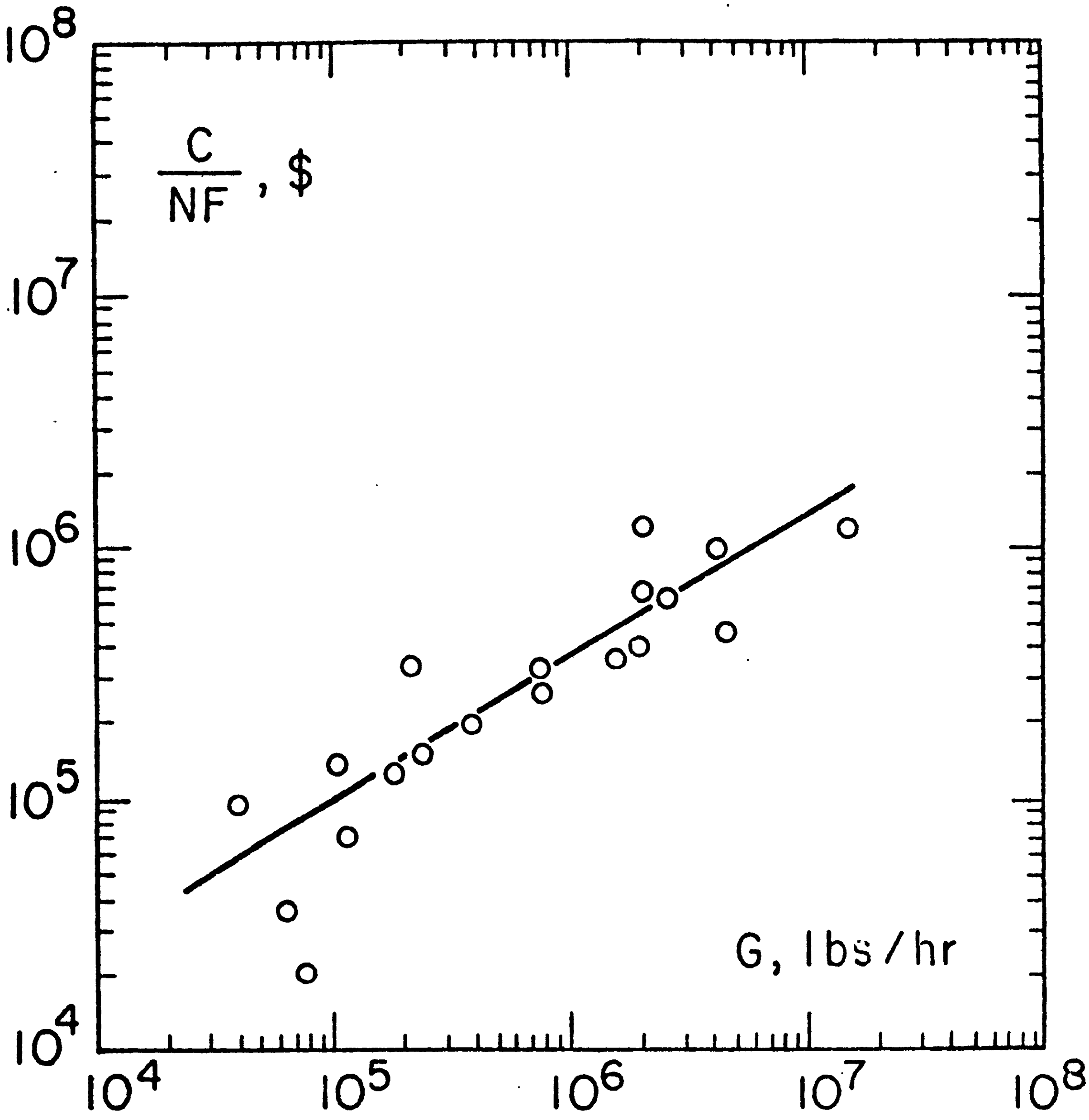

Figure 1 - Constant Exponent Correlation

B. The constant exponent, e, was replaced by a weighted average exponent computed for each plant from individual module exponents assigned according to equipment type based on literature [4,10,11]. The average exponent was weighted by the number of modules.

C. Instead of the constant proportionality factor, k, a variable was introduced that was defined as a function of module distribution in the plant.

D. A combination of the techniques B and C above was used.

The alternative techniques B, C, and D did not result in a better correlation than Equation 2.

Selected Correlation

The best correlation of battery limits direct cost with process modules, plant capacity, temperature, pressure and material of construction was obtained by postulating a constant proportionality factor, k , and a variable exponent, v ,

$$C = kFNG^{v} \tag{8}$$

The exponent was assumed to vary as a function of module distribution in the following form:

$$v = \sum_{i=1}^{n} c_i x_i \tag{9}$$

The coefficients, c_i , were established by a statistical method [9]. Three of them are zero for all practical purposes, which indicates that Equation 9 is statistically insensitive to module types 6, 7, and 10. The interpretation is provided elsewhere [8]. The numerical value of the constant, $k = 1.1(10^2)$ in 1981 dollars, does not differ from that of Equation 2 within the accuracy of the analysis.

The exponent is computed with the following equation:

$$v = 0.83x_1 + 1.05x_2 + 0.59x_3 = 0.47x_4 + 0.59x_5 + \\ + \ 1.07x_8 + 0.60x_9 + 0.83x_{11} + 0.40x_{12} \qquad (10)$$

The constant and exponent of Equations 2 and 8 and the statistical data are compared in Table VI. As the correlations are based on 20 plants, the 40% deviation line means that four and two estimates based on Equations 2 and 8, respectively, are outside the ± 40% range. However, the accuracy of the recommended Equation 8 is actually higher than this routine interpretation of the statistics indicates. The 10% points that have a deviation exceeding ± 40% represent Processes 5 and 7 in Table III. The former is 53% underestimated because it has only a few modules, one of them being an extremely expensive reactor. Process 7 is 61% overestimated because it is a reconstructed plant with some unidentified retrofitted old equipment. If these two odd processes are left out, the accuracy is ± 30% in 94% cases. The two plants were included to demonstrate that the method applies with lower accuracy even to extreme cases.

The recommended correlation was tested [8] using a coal gasification project [12]. The predicted direct cost was within 9% of the detailed estimate, reinforcing the confidence in Equations 8 and 10. While most of the plants underlying these equations were deliberately chosen from the neglected mining and solids processing technologies, the inclusion of four typical fluid processes (Nos. 11, 12, 16, and 17 in Table III) and the good agreement with DeCicco's formula indicate that the correlation applies to a wide range of process industries.

It can be concluded that the new correlation of direct plant cost with process modules, capacity, and process conditions is an improvement over published methods in that it removes the ambiguity of defining the modules and covers a wider range of capital costs and technologies.

Symbols

A	additional parameter, various dimensions
C	battery limits direct cost, $
F	complexity factor, no dimension
G	average throughput, lbs/hr
I	investment cost, defined in various ways, $
K	plant capacity, various dimensions
M	material of construction factor, no dimension
N	number of process modules, no dimension
P	pressure factor, no dimension
R	statistical coefficient of determination, no dimension
T	temperature factor, no dimension
c	coefficient, no dimension
e	constant scale-up exponent, no dimension
f	function of
i	process module type, no dimension
k	proportionality factor, ($)(hrs)/(lbs) per module
n	number of module types, no dimension
p	pressure, atm
t	temperature, ° C
v	variable scale-up exponent, no dimension
x	fraction of the number of modules that are of type i, no dimension

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TABLE I - Type of Variables Used in the General Investment Estimation Equation 1

Author	Capacity K	Temperature t	Pressure p	Material Factor M	No. of Modules N	Notes
Zevnick & Buchanan (1)	Mass flow of product	Maximum	Maximum	Overall	Functional units defined by examples	Graphical correlation converted to equations by DeCicco (2)
Gore (5)	Average volumetric flow	Maximum	Not used	Overall	Zevnick's & Buchanan's functional units	Additional parameter: empirical recycle factor
Stallworthy (3)	Mass flow of products	Average weighted by stream flows	Average weighted by stream flows	Average weighted by stream flows	Avg. No. of functional units weighted by stream flows	-
Bridgwater (7)	Avg. mass flow based on input and output	Avg. weighted by No. of functional units	Avg. weighted by No. of functional units	Not used	Zevnick's & Buchanan's functional units	-
Taylor (6)	Mass flow of product	Not directly used	Not directly used	Not used	Steps based on process chemistry	Additional parameter: tabulated function of T, P, N for each stream
Allen & Page (4)	Molar flow of product	Graphical function based on maximum	Graphical function based on maximum	Overall	Basic items based on major equipment	Additional parameters: (a) Stream configuration, (b) Tabulated exponents and reference costs for equipment types

TABLE II - Process Module Type and Characterization

Type	Parameter Changed	Example of Operations	Example of Main Equipment
1	Temperature	Heat exchange	Heater, cooler, exchanger
2	Pressure (requiring energy input)	Compression, evacuation	Compressor, blower, fan vacuum pump
3	Location of solids and molten solids	Receiving and unloading, transport, storing reclaiming, packaging, loading, and shipping	Car dumper, conveyor systems, reclaimer, crane and ladle, strapping station
4	Location of liquids	Liquid collection and distribution (with at least 2 inlet and 2 outlet streams)	Systems of tanks, sumps, and pumps
5	Particle size	Comminution; agglomeration	Crusher, ball mill; balling disk
6	Particle size distribution	Screening, classification	Screen, mechanical classifier, cyclone
7	Solid body shape	Casting, extrusion, briquetting	Casting wheel, extruder, press
8	-	Storage*	Tank farm, silo system, stockpile, pond
9	Number of streams	Liquid or solid mixing, dissolution, multiphase stream generation; phase separation	Stirred tank, kneader, emulsifier; filter, thickener, cyclone
10	Phase	Melting, evaporation, sublimation, condensation, solidification	Furnace, evaporator, splash condenser, scraped wall condenser
11	Phase distribution of components	Absorption, adsorption, desorption, drying, extraction (liquid-liquid or leaching), washing, distillation	Column, dryer, stirred tank or leach system
12	Composition or isotope	Chemical or nuclear reaction	Reactor, reverberatory furnace, electrolytic cell system

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* Cannot be defined in terms of parameter change

TABLE III - Investigated Processes

1. Metal ore mining and hydrometallurgical extraction
2. Solution mining
3. Waste metal reclamation
4. Mineral concentration by grinding and froth flotation
5. Combined smelting and converting
6. Metal ore mining and pyrometallurgical extraction
7. Grinding, flotation, and electric smelting
8. Grinding, flotation, and combined smelting and converting
9. Recovery of a non-ferrous metal from solution
10. Counter-current leaching and washing
11. Carbon monoxide production from natural gas
12. Metal reduction in solution
13. Electrowinning
14. Transportable solid-liquid-gas process plant
15. Ore storage, reclamation, and grinding
16. Ammonia manufacture
17. SO_2 scrubbing and sulfur production
18. Flash smelting
19. Material handling and electric furnace processing
20. Coal cleaning

TABLE IV - Range of Parameters Covered

Parameter	Unit	Low	High
Number of modules	-	6	97
Average capacity	k lbs/hr	40	15,000
Battery limits direct cost	\$ k (1981)	740	152,000
Battery limits installed cost	\$ k (1981)	430	77,000
Maximum temperature	°F	40	3,100
Maximum pressure	Psia	15	2,600
Overall corrosivity	Zevnick-Buchanan material factor	0	0.3*

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* The maximum material factor is 0.4.

TABLE V - Material of Construction Factors

M	Material of Construction
0	Carbon steel, cast iron, wood
0.1	Al, Cu, brass, 400 series stainless steel, lined carbon steel
0.2	Ni, monel, inconel, 300 series stainless steel
0.3	More expensive alloys such as Hastelloy
0.4	Precious metals

TABLE VI - Correlation Parameters

Item	Constant Exponent Correlation (Equation 2)	Recommended Correlation (Equations 8 and 10)
Exponent	0.57	0.4 to 0.9
Constant, k, expressed in (\$)(hrs)/(lb) per module	1.1×10^2	1.1×10^2
Percentage of points that have a deviation of less than:		
± 30%	70	85
± 40%	80	90
Coefficient of determination, R^2	0.75	0.85