PREDICTED COSTS OF ENVIRONMENTAL CONTROLS FOR A
COMMERCIAL OIL SHALE INDUSTRY

Volume 1 — An Engineering Analysis

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Denver Research Institute
Denver, Colorado

and

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U. S. DEPARTMENT OF ENERGY

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PREFACE

The U.S. Department of Energy funded this project as part of an attempt to assess realistically the possibility of an oil shale industry contribution to domestic petroleum production. This two volume report assesses the costs of environmental controls for a commercial (not pilot or prototype) oil shale industry using currently available technology. The task was performed by an assemblage of engineers, scientists, and economists with experience in natural resource evaluation, oil shale technology, and environmental engineering. Early drafts of this report were reviewed by groups within DOE, by members of the National Laboratory System, and by private consultants. Their comments were weighed and incorporated into this final draft.

The authors are aware of the current sociopolitical milieu. This work can have impact. As the editors put the final touches on these volumes the President of the United States called for massive public support of a synfuels industry. Of course, there are a range of opinions. Like the authors of the recently published report of the Energy Project at the Harvard Business School, we hope our work helps "the United States to come to terms with the realities of the energy problem, not with romanticisms, but with pragmatism and reason."

Finally, sadly, we dedicate this volume to the memory of the late Dr. Charles Prien, gentleman, scientist, and a co-author of this report.

C. Gale and M. Shaffron
DRI, August 1, 1979

ABSTRACT

The pollution control costs for a commercial oil shale industry were determined in a joint effort by Denver Research Institute, Water Purification Associates of Cambridge, and Stone and Webster Engineering of Boston and Denver. Four commercial oil shale processes were considered. The results in terms of cost per barrel of syncrude oil are predicted to be as follows: Paraho Process, \$0.67-\$1.01; TOSCO II Process, \$1.43-\$1.91; MIS Process, \$2.02-\$3.03; and MIS/Lurgi-Ruhrgas Process, \$1.68-\$2.43. Alternative pollution control equipment and integrated pollution control strategies were considered and optimal systems selected for each full-scale plant. A detailed inventory of equipment (along with the rationale for selection), a detailed description of control strategies, itemized costs and predicted emission levels are presented for each process. Capital and operating cost data are converted to a cost per barrel basis using detailed economic evaluation procedures. Ranges of cost are determined using a subjective self-assessment of uncertainty approach. An accepted methodology for probability encoding was used, and cost ranges are presented as subjective probability distributions. Volume I presents the detailed engineering results. Volume II presents the detailed analysis of uncertainty in the predicted costs.

ACKNOWLEDGMENTS

Seldom, if ever, in the memory of those who participated in this research, has a more difficult project been encountered. When the engineering results were complete we discovered we had created an information explosion. The results might never have been communicated but for Herculean efforts by the DRI Word Processing Center, particularly Carolyn Bauer, Don Weiss, Bev Slater, and Lorna Schnase, and the efforts of the DRI Chemical Division secretaries Martha Lee, Carole Taylor and Kathy Allmer.

Also we acknowledge the support and encouragement of the Division of Environmental Impacts, Office of Technology Impacts, Office of Environment, U.S. Department of Energy.

Our editors, Mike Shaffron and Chuck Gale, of the DRI Chemical Division, struggled mightily in order to make this volume readable. In addition to integrating the work of nine authors, they found and corrected literally hundreds of inconsistencies in the working drafts of this report. All remaining errors and failures to communicate are the responsibility of the authors.

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The right to search for truth implies also a duty; one must not conceal any part of what one has recognized to be true.

-Albert Einstein

1.0 EXECUTIVE SUMMARY

1.1 INTRODUCTION

The Federal Nonnuclear Energy Research and Development Act (PL 93-577), passed in December of 1974, began a program of national dedication to research, development and demonstration of nonnuclear energy technologies. This act also mandates that attention be given to environmental concerns at every stage of the RD&D programs for any nonnuclear energy technology.

In response to this mandate, the U.S. Department of Energy has sought to integrate the necessary research to assure protection of the environment into all projects as a technology moves through the research and development phases.[1] The responsibility for this mission rests with the Office of Environment which sponsored the research presented in this report.

Oil shale represents a vast alternate energy source which can be readily converted to a liquid synfuel. Known U.S. resources which could be tapped using existing technology are estimated to be equivalent to 600 billion barrels of oil. Total U.S. resources, including leaner shales (up to 10 gallons of oil per ton) are estimated to be approximately 28 trillion barrels.[2] Yet, shale oil has never been produced commercially in the U.S., even though the technology exists and the reserves of known high grade ore alone are inconceivably large.

Production of oil from oil shale has been on the verge of becoming a commercial reality a number of times in the last 30 years, but each time the barrier of economic uncertainty, above all others, has arisen, dimming the hopes for full development of this alternate energy source. This uncertainty has become a larger and more complex problem in the seventies with the recognition of a number of critical environmental concerns.

This report addresses these critical environmental concerns from a technical standpoint and assesses the economic uncertainties which accompany them. The outputs of this research program are a large data base (over 190 tables) relevant to the consideration of environmental issues surrounding oil shale processing, a critical evaluation of appropriate environmental control technologies, integrated pollution control strategies, and more detailed estimates of environmental control costs than have been available in the past.

1.2 OBJECTIVES

The estimates of cost to produce a barrel of oil from oil shale are infamous for having tripled between 1973 and 1976.[3, 4] Estimates since then have understandably lacked high credibility. The U.S. Department of Energy has investigated the causes of this underestimation error[5] and found that cost estimates increased as the level of detail in plant design information increased. Other factors, such as inflation, were found to be much less important.

Preliminary estimates of the cost of environmental control for a commercial oil shale industry were developed by the Office of Environment, DOE, using a generalizing approach. This produced broad ranges with substantial uncertainty.[6] In an effort to avoid a dramatic underestimation error, and to reduce the uncertainty hampering decision-making processes,[7] DOE contracted for this research effort to look at cost in detail on a specific process by process basis.

The scope of the effort also limited consideration to only the direct environmental control costs. Only costs associated directly with equipment and procedures for pollution control of wastes and effluent streams within development sites were considered. Such problems as increases or decreases in Colorado River salinity due to withdrawal of water for processing or addition of low salt, treated water were not addressed.

This study was also limited to consideration of commercial oil shale processes (and not the first commercial plant) in the Piceance Creek Basin in Western Colorado, containing the richest oil shale deposits in the nation and an environmentally sensitive area. Other parts of the Green River formation and Eastern shales were not considered. Only commercially available

environmental control technologies were considered. This allowed more exact estimates to be made since these technologies are well understood and current cost data are available, but overlooks the possibility of higher levels of control at the same or less cost using emerging technology.

DOE also asked for extensive backup data for all cost estimates so that experts in the field of pollution control can judge the soundness of the approach and determine the impact of a broad spectrum of alternate processing and pollution control assumptions. Finally, DOE asked for a thorough analysis of the residual uncertainties in the results of this study.

1.3 THE APPROACH

1.3.1 Environmental Regulation

An essential ingredient in any detailed analysis of environmental control costs is specification of the emission levels which are to be achieved. Two levels of emissions were considered, where appropriate, for this study. One level, termed the "less strict regulatory scenario," was generated by extrapolation from the most strict existing regulations as applied to other industries, for example, power generation. A "more strict regulatory scenario" was developed to reflect the maximum extent of regulation possible under current and anticipated legislated authority and regulation.

Use of a two regulatory scenario approach was not appropriate in some areas of pollution control. First, a two scenario approach is impossible when a zero discharge pollution control strategy is selected. Such a strategy was adopted for management of process waters. Second, in some instances the control technology selected produced a level of control greater than specified for the most strict scenario. In such a case, both scenarios become trivial; only one piece of equipment is specified for both scenarios. The actual regulation followed what can be termed "Best Available Technology." The regulatory scenarios used are summarized in Table 1-1.

1.3.2 Selection of Oil Shale Processes

Four oil shale processes were selected for detailed study. Two factors were judged most important in deciding upon which processes should be considered. It was felt that emphasis should be placed on actual processes

Table 1-1
SUMMARY OF REGULATORY SCENARIOS

Pollutant	Proposed Regulatory Scenarios		Scenarios Used
	Less Strict	More Strict	
Particulate Matter	90% Control of Vented Emissions	99% Control of Vented Emissions	99 + % Control
Hydrocarbons	Good Management	Controls on all Sources	Both
Nitrogen Oxides	No Controls	Adjustment of Combustion; available controls	More Strict Only
Sulfur Dioxide	90% removal of SO ₂	93.5% removal of SO ₂	BAT*, 99 + % Control
Hazardous Materials	Specific controls not found to be necessary		
<u>Waters</u>			
Process Waters	Zero Discharge		Zero Discharge
Mine Water			
Total Suspended Solids	30 mg/l	30 mg/l	nil
Total Dissolved Solids	723 mg/l	500 mg/l	<200 mg/l; <100 mg/l
Boron	1.0 mg/l	0.75 mg/l	1.0 mg/l; 0.06 mg/l
Fluoride	2.0 mg/l	2.0 mg/l	1.5 mg/l; 0.15 mg/l
Ammonia	-	0.2 mg/l	0.1 mg/l
Phenol	-	1.0µg/l	0.2µg/l
Solid Wastes	Sold Waste Disposal Regulations	Hazardous Waste Disposal Regulations	Both
Aresnic Wastes	Offsite Disposal		Offsite Disposal
Spent Catalysts	Recycle		Recycle
Spent MIS Retorts	No Treatment	Slurry Backfilling	Both

*Best Available Technology

with a high probability of commercialization in the near future. Since environmental control technology was to be considered in detail rather than in a generalized manner, the availability of extensive, detailed engineering data was important.

After selection of the processes to be considered, it was judged important to include a broad range of processing alternatives such as upgrading to syncrude and power generation within the four processes.

Table 1-2 lists the four processes and the processing alternatives associated with each. Consideration of each process is also limited to a specific development site, so the results of this study are both process and site specific. Development plans were not scaled to a common production level (for example, 50,000 bpd) to avoid introduction of any inaccuracies in this way. Actual anticipated commercial scale development plans were followed exactly.

A strict regulatory scenario for the Modified In Situ (MIS) process requires some treatment to prevent leaching and subsidence of the spent MIS retorts. The method selected for this study is slurry backfilling with spent shale provided by the Lurgi-Ruhrgas aboveground retorting process. Resource recovery is enhanced in the MIS/Lurgi process and the cost of environmental control is actually less per barrel of syncrude for the combination of MIS and Lurgi processes than for the MIS process alone.

1.3.3 Selection of Pollution Control Technologies

Detailed material, energy, and water balances were calculated around each retorting process before the selection of pollution control technologies. After selection of all controls, these balances were adjusted to reflect the addition of the controls. The preliminary material, energy, and water balances were first used to develop overall integrated pollution control strategies for each process, and to characterize and quantify all emission streams.*

Numerous environmental control technologies were considered for control of each emission stream. Final selection of the appropriate control

* Emission stream can be a process water stream which is actually recycled or reused but must be treated before recycling.

Table 1-2
SUMMARY OF OIL SHALE PROCESSES AND PROCESSING ALTERNATIVES SELECTED FOR STUDY

<u>Oil Shale Process</u>	<u>Production Rate</u>	<u>Location</u>	<u>Location Features</u>	<u>Processing Alternatives†</u>
Paraho	99,170 bpsd* Crude Shale Oil 155 MW Electricity	Anvil Points, Colorado	29 Gal/Ton Shale Dry	Open Cycle Power Generation
TOSCO II	47,000 bpsd Syncrude 4,300 bpsd LPG	Colony Site	35 Gal/Ton Shale Dry	Crude Oil Upgrading Power Purchased
Modified In Situ	57,000 bpsd Crude Shale Oil 97 MW Electricity	Tract C-a or C-b	25 Gal/Ton Shale Excess Mine Water	Open Cycle Power Generation No treatment for Abandoned Retorts
Modified In Situ + Lurgi-Ruhrgas	81,000 bpsd Crude Shale Oil 140 MW Electricity	Tract C-a or C-b	25 Gal/Ton Shale Excess Mine Water	Open Cycle Power Generation In Situ/Aboveground Combination Slurry backfilling of Spent Retorts

* Barrels Per Stream Day

† Ammonia and sulfur are recovered in all processes. All power gas is consumed on-site.

technologies was made after careful consideration of all parameters affecting the performance of a piece of equipment. Data are presented in this report to support the selections made.

A detailed list of the environmental controls for each process would be too extensive to present in this section of the report. For example, 113 separate pieces of air pollution control equipment are listed in Section 5.0 for the Paraho process.

1.3.4 Engineering Cost Estimates

Engineering cost estimates were not based on commonly available literature data or approximation procedures. Detailed specifications for each piece of control equipment were developed, including the level of control which must be achieved. (In many instances substantial engineering was required to develop the specification.) In almost all cases (except for very common small off-the-shelf items) actual price quotes for specific equipment were obtained from actual equipment vendors. Care was taken to insure a complete understanding of all auxiliary items included in the price. Installation costs for each piece of equipment were then itemized. Factors or vendor estimates were not used to determine installation costs. All required auxiliary items were included in the final installed capital cost estimate for each piece of equipment. Operating costs were also itemized, often with the assistance of the vendors.

Environmental control of solid waste primarily involves activities and procedures rather than actual equipment. An exhaustive list of these activities was compiled, and the cost for each was then determined individually. A complete, itemized inventory of these costs by year of operation for each oil shale process is presented in Section 7.0.

Environmental control costs were considered in detail in order to avoid the underestimation errors previously mentioned (which have characterized the estimates of the total cost of a barrel of shale oil). Although substantial uncertainty still surrounds the total oil shale production process (since no commercial size plant has ever operated in the U.S.), the environmental controls selected are well-established technologies which have been used on a commercial scale in other industries. Also the individuals directly involved in this project in developing the cost estimates have extensive,

current, hands-on experience in the design, installation, and operation of such equipment. The full cost of all pieces of equipment even marginally involved in pollution control were included even though such equipment may serve mainly a processing function.

1.3.5 Emission Estimates

For all emission streams, estimates have been made of emission levels for the full scale commercial processes. The control strategies developed and the control technologies selected are not necessarily the same as those proposed by the developers of each process. Hence, the emission estimates presented are not necessarily in agreement with the estimates which have been made by the developers.

This report can be viewed as an independent, in depth analysis of pollution control requirements for oil shale. Also the actual developers of the processes considered are currently involved in making accurate estimates of emission levels for pilot scale or single commercial module plants, not full scale developments. Consideration in detail of integrated environmental control systems for commercial scale plants may not have occurred.

The emission data contained in this report are too extensive to present in this section, but have been summarized in Sections 5.0, 6.0, 7.0 and 8.0.

1.3.6 Economic Analysis

Total capital and annual operating costs were converted to total annual costs and costs per barrel of oil, using a detailed economic evaluation. DOE asked that these costs be determined as an oil shale developer would determine them when evaluating the attractiveness of the investment opportunities.

Total capital was converted to annual capital charges using annual capital recovery factors. Annual capital recovery factors were derived considering all economic impacts on actual after-tax return on investment, such as the period of investment in construction, applicable tax laws, depletion allowances, severance taxes, allowable depreciation, etc., and derived to provide the required return on investment as a component of cost. Eight different capital recovery factors were necessary to convert all capital costs for all processes to annual capital charges.

The basic assumptions follow:

- o 13% discounted cash flow rate of return on investment
- o 100% equity
- o Maximum allowable depreciation
- o 25-year project life
- o 1 to 4 year investment period for construction
- o 3 to 4 year start-up profile to reach full capacity
- o 330 stream days/year

The basis upon which these assumptions were made is discussed in detail in Section 10.0.

1.3.7 Analysis of Uncertainty

Upon completion of the engineering portion of this research program, all results were analyzed in depth for uncertainty. An established procedure[8] for subjective probability encoding combined with a structure developed during the program for applying this procedure was used to quantify the uncertainty in the cost estimates. Subjective probability distributions were produced for all items which contribute to pollution control cost and which were found to contain any significant amount of uncertainty. These distributions were combined for each oil shale process, following the rules of probability, to produce overall probability distributions on cost for each process as a measure of uncertainty.

The entire analysis of uncertainty is presented as Volume II of this report. Subjective probability was used because all available experimental and engineering information was already considered in the engineering portion of the study. The result of the analysis of uncertainty is a systematic and complete examination of uncertainty and a quantification of the opinions of the individuals who participated in the research program.

1.4 THE RESULTS

The overall results of Volume I and Volume II are combined and summarized in Figure 1-1 and Table 1-3. Table 1-4 shows the results in terms of total capital cost and annual operating cost for each process.

FIGURE 1-1. SUMMARY OF ENVIRONMENTAL CONTROL COSTS
FOUR OIL SHALE PROCESSES. JAN. 1979 DOLLARS.

Source: Table 1-1, Volume II.

Table 1-3
SUMMARY OF ENVIRONMENTAL CONTROL COSTS OF FOUR OIL SHALE PROCESSES

<u>Oil Shale Process</u>	<u>Maximum Possible Cost*/Barrel Syncrude Equivalent</u>	
	<u>Less Strict Regulatory</u>	<u>More Strict Regulatory</u>
	<u>Scenario</u> (Dollars)	<u>Scenario</u> (Dollars)
TOSCO II	1.73	1.91
Paraho	0.99	1.01
Modified In Situ	2.68	3.03
Modified In Situ + Lurgi-Ruhrgas	2.28	2.43

* The full data presented in Figure 1-1 can be found in Table 1-1, Volume II.

Table 1-4
CAPITAL AND OPERATING COSTS* FOR POLLUTION CONTROL OF FOUR OIL SHALE PROCESSES

<u>Oil Shale Process</u>	<u>Regulatory Scenarios</u>			
	<u>Less Strict</u>		<u>More Strict</u>	
	<u>Capital</u> (10 ³ \$)	<u>Operating†</u> (10 ³ \$)	<u>Capital</u> (10 ³ \$)	<u>Operating</u> (10 ³ \$)
TOSCO II	67,183	14,610	88,007	16,802
Paraho	74,792	15,334	75,342	18,783
Modified In Situ	101,343	19,125	134,992	26,859
Modified In Situ + Lurgi-Ruhrgas	131,128	25,985	152,007	30,328

* Volume I values. Does not reflect analysis of uncertainty.

† Includes byproduct credits.

To make the per barrel results comparable between processes, a "syncrude equivalent" concept was developed. Table 1-1 shows the total per stream day output from each plant. The TOSCO II process produces an oil shale syncrude and LPG. These are high quality energy products and are worth substantially more per barrel than crude shale oil. The other processes produce electricity for export. This is also a high value form of energy. Simply dividing the total annual cost of pollution control by the number of barrels of liquid product output to obtain a per barrel cost does not truly reflect the cost per real unit output of energy value, and comparisons between processes are difficult. To overcome this, the total output of each plant has been converted to its equivalent in terms of shale oil syncrude, including electricity (in the TOSCO case, electricity purchased is subtracted from output). Total annual costs are then divided by like measures of output to produce a cost per barrel of syncrude equivalent.

Figure 1-1 shows that the cost of pollution control is expected to be more for the in situ processes than the aboveground processes. The difference can be attributed to the need to treat excess mine water in the in situ processes, and to a lesser extent to greater production of NH_3 and H_2S in the retort gas. This means that most of the higher cost for the in situ processes is due to factors of location rather than processing factors. Also, the results presented in Figure 1-1, and in the tables summarize calculations based on the maximum quantity of excess mine water requiring treatment. It is possible that much less water will require treatment. If the minimum possible quantity of mine water were assumed, the costs would be lower by 44¢ to 74¢ per barrel for the MIS process and 21¢ to 34¢ per barrel for MIS/Lurgi, depending upon the regulatory scenario. Such reductions would bring the total cost for pollution control for the in situ processes more in line with the costs for the aboveground processes.

The MIS/Lurgi process also appears to cost less than the MIS process for pollution control. However, Table 1-4 shows that the MIS/Lurgi costs are substantially higher. The output of the MIS/Lurgi process is also substantially more than the MIS process due to retorting of the mined shale aboveground. Hence, the cost per unit output for pollution control is less.

The cost difference between the TOSCO II and the Paraho processes also can be attributed partly to economies of scale. The proposed Paraho full

scale plant produces approximately twice as much product as the TOSCO II plant.

No information is presented on the total cost of the oil shale plants, so no comparison of total plant cost per unit output can be made. Hence, one commercial plant cannot be judged to be more economically attractive than another only on the basis of environmental control cost.

Figure 1-1 also shows approximate probability ranges for cost. The probability is judged to be zero that the actual costs will exceed the upper bound or be lower than the lower bound of these regions. It should be remembered that these conclusions are based on hard data but also reflect expert opinion.

1.5 CONCLUSIONS

All pollution control problems which were discovered could be managed well with existing technology, in order to meet the regulatory requirements assumed for the study. It appears that a high level of control can be achieved at reasonable cost. No consideration was given, however, to the impacts of the estimated emission levels. This was entirely outside the scope of this effort. The next step is to carefully analyze the impact of a commercial industry on a regional basis, assuming multiple plants in the Piceance Creek Basin area with emission level estimates which reflect good, integrated pollution control systems. Careful attention also should be paid to secondary environmental and socioeconomic impacts.

Uncertainty analysis identified three important areas of needed research in environmental control technology. These areas are:

- o H₂S removal from retort gas streams on a pilot scale.
- o Organic removal from retort process waters.
- o Slurry backfilling of spent MIS retorts.

Research programs are now underway in these areas.

Even though the pollution control costs for a commercial oil shale industry were studied in great depth, and the uncertainty in the results was also examined in depth, the true answers to the critical environmental questions addressed are still not completely understood. The true answers will only be known after commercial plants have been built and successfully operated.

Along the way, problems will be discovered which have not yet been imagined, and some problems will disappear. This is inherent in the process of developing any new technology.

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2.0 INTRODUCTION

2.1 SCOPE OF THE EFFORT

In order to understand the results of this study, great care must be taken to understand the approach taken and the assumptions made. The authors have in no way attempted to answer all questions concerning pollution control costs for oil shale production. Rather, the results are specific for the limited number of processes and controls considered. The results also depend on the assumptions which inevitably must be made to determine such costs. A number of specialists in oil shale processing and pollution control were assembled for this study, and the authors feel the assumptions made are reasonable.

Every effort has been made to provide flow rates, compositions, temperatures, quantities, etc. for all waste streams which require control, and to provide extensive information of the same kind on many process streams. These data are sufficient to allow the assessment of this approach and the consideration of alternate approaches in detail. The authors urge those who feel alternate approaches are more appropriate to calculate their results in the same detail as has been done in this study, thereby stimulating productive discussion of any differences.

2.1.1 The Process Models

Four specific process models were selected for this study. In most instances costs determined for one model cannot be generalized to any of the other models. Environmental control costs are different for each site and each process. Processing costs (excluding environmental control costs) are also different for each site and each process. A cost effective system can be determined through consideration of both environmental control costs and processing costs. This study deals with environmental control costs for the process models.

The selection criteria for the process models involved three considerations:

- o The processes selected should represent a broad cross section of processing alternatives.
- o Extensive data must be available on the processes selected.
- o The processes selected should have a reasonably high probability of reaching full commercialization.

The four process models and their products are shown in Figure 2-1. The major processing alternatives covered by the four process models are shown in Figure 2-2. Note that the process models are meant to represent full scale commercial operation of a mature industry. As mentioned before, the costs are process and site specific, so actual costs for specific processing alternatives cannot be separated or generalized. However, careful consideration of the details presented later in this report does give a meaningful quantitative feel for the impact of processing alternatives.

Within specific areas of each process, a number of processing alternatives are also available. For this study it was necessary to select specific processing alternatives. In some instances, an alternative selected was obviously the preferred choice. In other instances, a number of attractive approaches are available, but only one could be selected for in depth consideration. In the latter cases, an effort has been made to discuss the alternative choices.

As can be seen in Figure 2-1, the oil product output of the four processes has not been normalized to a single level for all processes, e.g., 50,000 barrels per day. Consideration was given to doing this at the beginning of the study, but the idea was rejected for several reasons. First, the processes generally contain multiple units of processing equipment (e.g., Paraho commercial scale operation will have 24 retorts). Hence, scaling the output to a new size becomes a complex problem. Also, the size and/or number of controls must change. The validity of cost data determined for such a situation would be uncertain. An example can be found in the MIS process.

Figure 2-1. THE FOUR OIL SHALE PROCESS MODELS AND THEIR OUTPUTS

<u>Process</u>	<u>Products/Outputs</u>
TOSCO II + Refinery Colony Development Site	<ul style="list-style-type: none"> o 47,000 bpsd* Syncrude o 4,300 bpsd LPG o 800 tpsd* Coke o 134 tpsd NH₃ o 192 tpsd Sulfur o 54,168 tpsd Spent Shale
Paraho Anvil Points	<ul style="list-style-type: none"> o 99,170 bpsd Crude Shale Oil o 146 tpsd NH₃ o 132 tpsd Sulfur o 155 MW Electricity o 126,100 tpsd Spent Shale
Modified <u>In Situ</u> Tract C-a or C-b	<ul style="list-style-type: none"> o 57,000 bpsd Crude Shale Oil o 281 tpsd NH₃ o 144 tpsd Sulfur o 97 MW Electricity o 1,500 to 8,500 gpm Water (12,400 acre ft/yr. max.) o 41,134 tpsd Raw Shale
Modified <u>In Situ</u> + Lurgi-Ruhrgas	<ul style="list-style-type: none"> o 81,000 bpsd Crude Shale Oil o 281 tpsd NH₃ o 172 tpsd Sulfur o 140 MW Electricity o 0 to 4,590 gpm Water (6,695 acre ft/yr. max.) o 5,250 tpsd Spent Shale

* bpsd = Barrels per stream day; tpsd = Tons per stream day.

Figure 2-2. MAJOR PROCESSING ALTERNATIVES COVERED BY PROCESS MODELS

Retorting Methods

- o Aboveground retorting with combustion
- o Aboveground retorting without combustion
- o Modified in situ retorting
- o Modified in situ retorting plus aboveground retorting

Refining

- o No refining of crude shale oil on-site
- o Refining of crude shale oil to syncrude on-site

Power Generation

- o No power generation on-site
- o Open cycle power generation on-site

Mining

- o Room-and-pillar mining
- o MIS mining plus rubblizing

Water Sources

- o River water
- o Aquifier water

Shale Grades

- o 25 to 35 gallons/ton

Water at Location

- o No significant mine water
- o Moderate mine water
- o Excess mine water

Solids Disposal

- o Aboveground raw shale disposal
- o Aboveground spent shale disposal
- o Spent in situ retort untreated
- o Spent in situ retort cemented

As designed, the process produces 57,000 barrels of oil per stream day. Assuming the expected yield in the retorting operation, and the grade of shale found at tracts C-a and C-b, and a certain size for the retorts, 40 retorts will be operating at one time delivering retort gas to 5 ammonia absorption systems. Reducing output to 50,000 barrels cuts the number of retorts to 35.1 and the number of absorbers to 4.4. Obviously, to again obtain whole numbers for the model, a series of complex adjustments must be made. Among these adjustments, the size of the ammonia absorbers would need to be scaled to a size different from that expected in the actual commercial plant before the cost of ammonia absorption could be determined. The cost for the artificially sized units would then only relate to the expected "real life" situation if costs were to vary linearly with size--an assumption which is seldom true.

The second reason for retaining the oil production level specified by the developers involves byproducts. The processes produce more than crude shale oil. Even if the oil yields are normalized, the outputs from the four plants are still not matched or comparable. A different approach has been used to allow the reader a rough means of comparing costs between processes, and will be discussed later in this section.

For a more detailed discussion of the process models, refer to Section 3.0.

2.1.2 The Regulatory Scenarios

To determine the cost of environmental controls, the process must be combined with regulatory scenarios--assumed environmental regulations for the industry which must be met. Two scenarios, "less strict" and "more strict," were applied to each pollutant or emission. The "less strict" scenarios were set assuming the regulations would not be unusually or unreasonably restrictive. Levels were often set by comparison with existing and/or proposed regulations for other industries. The "more strict" scenarios are meant to be as restrictive as current and anticipated regulation will allow, taken to their limits, and hence are meant to represent "worst case" situations. Costs determined for the "more strict" scenario cases are meant to be realistic "worst case" costs. In some instances even more restrictive regulation could be assumed, but to do so would represent such unreasonable and unfairly restrictive regulation as to be unrealistic.

The two scenario system encounters some important conceptual problems. A most obvious problem arises when a zero discharge control strategy is adopted, making a two scenario system inapplicable. In some instances, comparison of two different approaches to controlling a specific pollutant showed the less effective approach also to be the more expensive. In such a situation, logic dictates that only the more effective control be used. In some instances, only one effective means of controlling a particular pollutant is available, making it impossible to apply two scenarios.

As mentioned before, the results of this study are highly site and process dependent, and each process model has unique characteristics. Two effective means of controlling a pollutant may be available for one process model, but only one means available for another. Hence, although both process models may achieve the same level of control for that particular pollutant, one model may show results for two scenarios, while the other model has only one scenario. The environmental controls have been considered to be an integral part of the overall process. Switching from the "less strict" to "more strict" scenario for control of one pollutant may effect the control strategy for another pollutant as well. A major process waste stream found in one process model may not even exist in another process model. For a more detailed discussion of the regulatory scenarios refer to Section 4.0.

2.1.3 The Pollution Controls

Several basic assumptions must be understood which affect the selection of controls and the evaluation of the appropriateness of the controls selected. First, it is assumed that scaleup to commercial size is successful and the plants are producing oil in a routine manner with a minimum of serious interruptions. Provisions have been made for interrupted situations, but controls were not selected on a basis that such situations would represent the normal composition of effluent streams.

This first assumption is best understood by way of example. The question is often asked: How will the controls on the MIS process work if there is extensive channeling in the retorts? Extensive channeling would mean low oil yields. If one hypothesizes that such a problem is pervasive, one in essence has hypothesized that the MIS process is a failure and

will never be commercialized. Instead one must assume that the process is commercially feasible, therefore, channeling is not a big problem. The reader is cautioned not to make alternate assumptions which represent a failure to commercialize the retorting processes. To do so will lead to unrealistic answers.

The second important assumption which must be made is that existing commercially available control technologies will successfully transfer to oil shale production. In every case, the assumption of transferability was made after careful study, and the controls selected represent those most likely to transfer successfully. However, for some items some doubt inevitably still exists. Where serious question exists, conservative designs have been used which result in high costs. In general, such costs are high enough to accommodate switching to an alternate technology without altering the conclusions.

Data carefully considered in many cases indicate that a technology will successfully transfer. For example, spent shale fly ash resistivities have been determined by DRI for the temperatures and moisture contents encountered where electrostatic precipitators are proposed. Fly ash resistivity is a key characteristic affecting precipitator performance. Preliminary data from a Water Purification Associates treatability study in progress and EPA/Cincinnati experience in performing BOD analyses lead to the assumption that bio-oxidation may be a suitable control technology for organic removal in water treatment. A number of years of research experience in spent shale cementation and specific proprietary research on Lurgi spent shale cementation for slurry backfilling has lead to selection of that method for grouting MIS retorts.

The reader is cautioned not to make the alternate assumption that a particular control technology will not transfer without hard data showing this to be true. It is felt that data from field tests of at least pilot units operating on emission streams characteristic of commercial operation will be necessary before more certain assumptions can be made.

The cost estimates are also based upon a mature industry. It is expected that a first generation commercial plant will be more expensive than plants which are designed after gaining experience by building and operating a full-scale plant.

2.2 OVERALL RESULTS

2.2.1 Pollution Control Strategies

When the option is available to design any process from the beginning, pollution control is designed as an integral part of the process rather than as simply add-on equipment. This approach makes a tremendous difference in terms of both the degree of control achieved and the cost of control. Pollution control is accomplished, not only with equipment, but also with strategy. A process stream which might be viewed as a waste may be useful input in another part of the process, eliminating the need for a control, in an efficient and effective way.

Although they cannot be meaningfully summarized in this section, the pollution control strategies developed are an important result of this study. The authors sought optimal, but not the optimum, strategies. The optimum strategies will only develop as the processes reach commercial operation. Every effort is also made to point out attractive alternatives not adopted as part of the models used in this study.

2.2.2 Summary of Environmental Control Costs

All costs are given in January, 1979 dollars.

Processing vs. Environmental Control Costs: Although the initial goal of this study was to determine only the cost of environmental controls, at times it is impossible to separate processing costs from environmental control costs. Hence, the costs presented represent more than environmental control, and the reader is left to judge how any division should be made. The costs determined are the costs for the entire piece of equipment and/or operation even if only a portion of the function can be attributed to environmental control.

The problem is best understood by using an example. The costs presented for water pollution control is actually the total cost for installation and operation of integrated water management systems. Costs are included for such items as source water treatment for boiler feed and cooling tower feed. The pollution controls require both steam and cooling, but so do many processing activities. No accurate way was found to charge for only that portion

serving pollution control, and to try to do so would only have made the results confusing. The entire cost of slurry backfilling of the MIS retorts in the MIS/Lurgi process models is included, although this is primarily a production function with a coincidental environmental side-benefit. The reader, however, can compare this cost with cost estimates by others for grouting techniques which are primarily environmental costs.

Costs and Return on Investment: The costs of environmental controls determined and presented by this study are costs which will be reflected in the eventual sale price of the shale oil produced, i.e., the cost to the consumer. The costs presented include a reasonable return on invested capital. Investment in oil shale development must compete with other opportunities available to the investing companies, so it is essential to view the cost of environmental controls by including the reasonable rate of return necessary to attract capital. Economic analysis is presented in detail in Section 10.0 of this report.

Cost Comparisons Between Process Models: The usual procedure for presenting environmental costs involves dividing the total annual cost by the number of barrels of oil produced for the process being analyzed. The use of such a procedure, however, can result in an inequitable comparison of several processes. As discussed earlier, the process models used in this study produce export products in addition to oil, most notably electricity. Also, the TOSCO II process produces a high quality low nitrogen and sulfur synthetic crude oil, which is of higher value than the crude shale oil produced by the other processes. Hence even the crude oil products are not comparable. Also, certain environmental control costs are incurred in producing electricity and syncrude, and should be credited against the higher value of these products when such processing activities are present.

In order to make the costs per barrel of oil for the different process models comparable even in a rough fashion, the total annual environmental control costs are converted to a cost per barrel of equivalent synthetic crude oil by dividing by the anticipated production in a year of normal operations. Anticipated production is measured in terms of equivalent barrels of syncrude. The factors used to convert all products to equivalent syncrude are as follows:

- o 1 Barrel Crude Shale Oil = 0.9 Barrel Syncrude
- o 1 Barrel Syncrude = 6×10^6 Btu
- o 1 kW electricity = 1×10^4 Btu

The thermal efficiency for conversion of crude shale oil to low nitrogen and sulfur syncrude has been found to be between 87% and 93%, depending on the processes used. Hence a value of 90% has been chosen to convert crude shale oil to its equivalent of syncrude. Electricity is converted to Btu and then to syncrude equivalent using the above factors.

Table 2-1 shows the exportable production outputs of the four process models and their conversion to equivalent barrels of synthetic crude oil. These outputs are the factors used to convert total costs to cost per barrel of product.

Table 2-2 summarizes the pollution control costs for the four process models in terms of total annual cost, cost per actual barrel of oil product, cost per barrel of syncrude equivalent, and cost per million Btu equivalents. Four cost figures are given for the MIS and MIS/Lurgi processes. At this time, considerable uncertainty still exists in estimating the quantity of excess mine water which must be treated by these processes. To handle this uncertainty, estimates have been made of the maximum and minimum quantities of excess mine water which may be encountered. Calculations have then been made for treating both quantities under both scenarios, resulting in four cost figures rather than two.

The cost per barrel of syncrude equivalent or Btu equivalent figures should be used for comparing the processes studied in this program. Remember that even this comparison is somewhat rough. It may be necessary to use the cost per barrel of actual oil product to compare the results of this study to the results of other studies, hence they are given as well in Table 2-2.

2.3 CONCLUSIONS

Although conclusions can be drawn in detail from studying the following sections of this report, some overall observations can be made at this point which may be useful in understanding what follows.

Table 2-1. EXPORTABLE PRODUCTION OUTPUTS OF THE FOUR PROCESS MODELS

<u>Process</u>	<u>Barrels Shale Oil</u>	<u>Barrels Syncrude</u>	<u>Electricity Exported</u>	<u>Barrels Equivalent Syncrude</u>	<u>10⁶ Btu Equivalents</u>
TOSCO II	-	47,000 B/d syncrude +4,300 B/d LPG = 51,300 B/d Product = 1.6929×10^7 B/y	-95 MW* -3,800 B/d - 1.254×10^6 B/y	1.5675×10^7 B/y	9.405×10^7 /y
Paraho	99,170 B/d 3.2726×10^7 B/y	89,253 B/d 2.9453×10^7 B/y	155.5 MW 6,220 B/d 2.0526×10^6 B/y	3.1506×10^7 B/y	1.8904×10^8 /y
Modified In Situ	57,000 B/d 1.881×10^7 B/y	51,300 B/d 1.6929×10^7 B/y	97 MW 3,880 B/d 1.2804×10^6 B/y	1.8209×10^7 B/y	1.0925×10^8 /y
MIS/Lurgi	81,000 B/d 2.673×10^7 B/y	72,900 B/d 2.4057×10^7 B/y	140 MW 5,600 B/d 1.8481×10^6 B/y	2.5905×10^7 B/y	1.5543×10^8 /y

* Imported electricity charged against barrels of equivalent syncrude.

Table 2-2. SUMMARY OF POLLUTION CONTROL COSTS FOR OIL SHALE PLANTS

<u>Plant</u>	<u>Scenario</u>	<u>Total Annual Cost Millions of Dollars</u>	<u>Total Cost Dollars per Barrel Actual Oil Product</u>	<u>Total Cost Dollars per Barrel Syncrude Equivalent</u>	<u>Total Cost Dollars per 10⁶ Btu Equivalents</u>
TOSCO II	Less Strict	25.8	1.53	1.65	0.27
	More Strict	31.4	1.85	2.00	0.33
Paraho	Less Strict	27.8	0.94	0.88	0.15
	More Strict	31.4	1.06	1.00	0.17
MIS	Less Strict				
	Low mine water	38.1	2.24	2.09	0.35
	High mine water	46.2	2.72	2.54	0.42
	More Strict				
	Low mine water	41.0	2.42	2.25	0.38
	High mine water	54.5	3.21	2.99	0.50
MIS/Lurgi	Less Strict				
	Low mine water	49.8	2.07	1.92	0.32
	High mine water	55.3	2.30	2.13	0.35
	More Strict				
	Low mine water	50.8	2.11	1.96	0.33
	High mine water	59.6	2.47	2.30	0.38

Wherever possible, conservative designs or approaches were used for pollution control. In many instances, the equipment was also overdesigned. Hence, it is felt that the costs presented for the more strict scenarios should represent realistic "worst case" costs.

Although the costs of pollution control for the four process models considered in detail by this study do not appear to be prohibitive, such controls will represent investment of substantial quantities of capital. Considering the huge quantities of capital which will be necessary to develop the oil shale resource as a significant national source of energy, any increase in capital cost is significant even if a reasonable rate of return is available.

One can conclude from the emission levels predicted by this study that oil can be produced from oil shale with surprisingly low emissions. For example, the reader can compare air emissions from the four plants with emission levels from coal-fired power generation. The exportable electric power produced by the four process models is produced more cleanly than it would be by using a coal-fired conventional power plant. Hence, generation of exportable power represents an energy resource available in the region in an environmentally acceptable fashion as a byproduct of oil shale processing.

Both the MIS and MIS/Lurgi processes may produce substantial quantities of water in excess of process needs. After treatment, this water will be of high quality and should be considered a new resource. In essence, processing at tracts C-a and C-b will tap a vast aquifer (25,000,000 acre-feet) and develop this regional resource as a byproduct of oil shale processing. Water is a scarce commodity in this semiarid region and should prove valuable for a number of beneficial uses. The process models considered by this study define the method of final disposition of this treated water as aboveground disposal. The quality and quantity of this water would suggest, however, that a more useful disposition will be found.

The MIS/Lurgi process may be considered the more strict scenario for the MIS process, since it results in grouting of the spent MIS retorts, preventing leaching and subsidence. It is interesting to note that the pollution control costs per barrel of syncrude equivalent for the MIS/Lurgi process is less than for the MIS process, since more oil is also produced. The MIS/Lurgi process approach becomes even more attractive when one considers that a higher

utilization of the shale resource may be possible for MIS retorting if the spent retorts are cemented as proposed by project Rio Blanco and studied herein. A technology sharing agreement has recently been secured between Occidental Oil Shale and Rio Blanco, so the slurry backfilling procedure may eventually be useful to both.

Surprisingly, the MIS/Lurgi type approach, using slurry backfilling of the spent MIS retorts, may be appropriate technology only in the wet areas of the Piceance Creek Basin, since slurry backfilling requires large quantities of water not available elsewhere. Such a conclusion represents a reversal in some recent traditional thinking which says that in situ oil shale production should only be allowed in dry areas.

Lack of sufficient water has often been viewed as a significant barrier to oil shale production in the Piceance Creek Basin. The actual consumption of water by the four processes is given in Section 6.0 of this report and includes water cooling which is a highly consumptive use. Conversion to aerial cooling and adoption of several other conservation measures would be more expensive, but should significantly lower consumption. Also, as previously discussed, considerable quantities of excess water are produced by some processes and could be used to satisfy the water needs of others.

2.4 ASSESSMENT OF UNCERTAINTY

Upon completion of the engineering portion of this study, the authors have attempted a subjective self assessment of the uncertainty of the results. A report on the methodology employed and the results of this assessment of uncertainty will be presented as Volume II and will accompany this report. The reader is directed to Volume II for further interpretation of the results contained herein.

3.0 DESCRIPTION OF SELECTED OIL SHALE PROCESSING MODELS

Four model oil shale retorting plants analyzed here are (1) the Paraho process without refinery, (2) the TOSCO II process including refinery to produce low sulfur and low nitrogen syncrude, (3) the Modified In Situ (MIS) process without a refinery or auxiliary aboveground retorting, and (4) the MIS process without refinery but with Lurgi-Ruhrgas aboveground retorting of shale mined out to allow voids generation and with subsequent complete cementing by slurry backfilling of the rubble left after MIS retorting.

Aboveground Retorting Plants: Either Paraho retorts or TOSCO II retorts are normally located aboveground and shale is brought to them from underground or open pit mines. Room-and-pillar mining has been chosen in this study for both of these models.

Open pit mining had at one time been considered by Rio Blanco for development of Tract C-a using a combination of Paraho and TOSCO II retorts. TOSCO II retorts were used to retort the raw shale fines necessarily screened out of the feed to the Paraho retorts thus preventing choking of gas flow through the shale charge.

Fines in the Paraho retort are a problem. They result from raw shale crushing whether open pit or room-and-pillar mining is used. They may be screened out and dumped, fed to TOSCO II retorts or Lurgi retorts capable of handling fines, or briquetted with shale oil heavy ends as a binder and sent to the Paraho retorts. The latter is assumed in the Paraho plant variation selected for study.

In aboveground retorts such as TOSCO or Paraho types, good oil yield per ton of shale retorted is allowed by superior retort charge uniformity and good separation of the retorting and combustion zones. The disposal of spent shale poses a large problem. Resource utilization is low for room-and-pillar mining. Pillars must be left for roof support and only the richer shale horizons may be mined economically.

Resource utilization is also low with MIS retorting. The maximum attainable oil yield is less than theoretical (Fischer assay) by some 30 to 40% due to encroachment of the combustion zone into the retorting zone. Pillars of sizes larger than those of typical room-and-pillar mining must remain intact, again lowering resource utilization. On the other hand, large cavern heights are being proposed that may exploit considerable amounts of low grade shale not otherwise economically mineable. Slurry backfilling or complete grouting of the rubble in the spent in situ retorts may also provide strength sufficient for reduction of "pillar" proportions, disposal of much spent shale underground, prevention of groundwater migration through the spent rubble, and alleviation of ground subsidence.

Aboveground retorts are stationary thereby allowing for relatively compact piping systems for the handling of gas and oil. The large expanse of surface area associated with transient MIS retorting dictates that long horizontal underground passages and long vertical underground passages for gas flow be used as well as other long lines for oil product.

Modified In Situ Plants: In contrast to the aboveground oil shale retorting methods, all or much of the oil shale retorted by in situ methods is retorted at or near its original location. In "true" in situ retorting, little or no shale is extracted before retorting, although some sort of fracturing is usually proposed before hand. With "modified" in situ retorting some 20 to 25% of the shale is mined out and the rest is rubblized to provide relatively small shale chunks for better heat and mass transfer. This mining and rubblizing also provides uniform air and steam flow resistance thereby allowing uniform development of retorting and combustion fronts in the rubble. Occidental Burn 5 incurred bad channeling apparently due to poor voids distribution. Burn 6 is said to be progressing better due to a better void distribution method during rubblizing.

In one development of the Modified In Situ method, the interstitial spaces between the rubble chunks after retorting are backfilled with grout composed primarily of a slurry of finely divided spent shale produced by aboveground retorting of the initially mined out shale. Both the Lurgi-Ruhrgas retorting system and the TOSCO II retorting system can burn the finely divided coke bearing spent shale thereby burning off the residual coke and producing a more finely divided nearly organic-free shale ash.

During combustion of the carbonaceous spent shale, subtle to profound chemical changes of the matrix minerals occur. In general, at an optimum combination of temperature and residence time, considerable cementing potential may be developed in the shale ash as it is mixed with water. This reduces the water permeability of an intact grout to a very low value and gives moderate "cohesion" based compressive and shear strengths. Permeability may be further lowered and strength increased by addition of various additives to the grout. Rio Blanco project is currently studying the grout backfill concept.

Fenix & Scission, Inc. [1] have reviewed five basic systems of mining for Modified In Situ retorting. The room-and-pillar mining method is of interest. Room-and-pillar mining involves the removal of shale, forming rooms with a portion left for supporting pillars. The shale is then blasted to rubble for Modified In Situ retorting. Restabilization of burned-out retorts with a mixture of oil shale ash, expanding cement, and water would allow the burned-out retort to be used as a pillar and permit new retorts to be rubblized and burned immediately adjacent to the restabilized retorts. This latter concept improves resource recovery considerably. Thus, assuming mined shale retorted aboveground, resource "extracted" would increase from around 60% to 85%.

3.1 PARAHO DIRECT MODE PLANT

3.1.1 Basic Parameters Assumed for Paraho Plant

The Paraho direct mode retort, as contrasted with the Paraho indirect mode retort has been taken for this present analysis. Following Paraho Case One, (discussed below) a pipeline of unrefined crude would be produced. The location would be near Anvil Points, Colorado, the site of the current Paraho operations. Room-and-pillar mining techniques, as in the TOSCO process, will remove 29 gal/ton assay raw shale from the Mahogany zone. Retorting will yield 99,170 barrels per stream day of crude shale oil for 92% of Fischer assay recovery after deducting 830 bpsd for the fuel pool. No upgrading will be done on-site, but electric power can be exported using gas turbine power generation from the low Btu gas produced by the retorting process. Ammonia and sulfur recovery are indicated for pollution control.

Byproduct ammonia sales help pay for NH_3 and H_2S removal. Water for the process will be provided by expanding the current supply line from the Colorado River. Semicarbonaceous partially burned spent shale will be cemented with wetting and compaction to form an initial impermeable barrier or "bathtub" in the disposal area. The bulk of the spent shale will then be compacted above this barrier without wetting. There will be zero discharge of aqueous effluent.

3.1.2 Paraho Direct Mode Retort Plant Description

The Paraho Development Corporation and its contractors have designed full-scale commercial complexes for shale oil production, based upon the Paraho direct mode and indirect mode retorting processes, as operated in both the pilot plant and semiworks plant at Anvil Points, Colorado from 1974 to date.

One hundred thousand bpd of oil at the retorts has been examined by Paraho for three different processing sequences, as follows:

- Case One: A pour point depressant is added to produce a low pour point pipelineable crude shale oil product. The large quantities of low Btu gas also produced are converted to exportable electrical power (24 direct mode Paraho retorts are used).
- Case Two: The product shale oil is the same as in Case One, but a high Btu byproduct gas is also produced, for use as retort fuel. (24 indirect mode Paraho retorts are used.)
- Case Three: The product shale oil is the same as in Case One, together with a quantity of exportable SNG. (18 direct mode and 6 indirect mode Paraho retorts are used.)

Two additional cases were examined by Paraho, each involving substantial upgrading of the retorted raw shale oil to 88,500 barrels per stream day of a high grade syncrude product. There would also be produced each stream day some 190 tons of sulfur and 415 tons of ammonia, mostly from the attached refinery. There would be no excess (exportable) power. A supply of 8,300 acre-feet of water would be required annually.

The two processing sequences for syncrude production include the following:

Case Four: Coking of the raw crude, and hydrotreating of the resulting liquids to produce a low sulfur and nitrogen, low pour point syncrude. (A combination of 12 direct mode and 12 indirect mode Paraho retorts was assumed.)

Case Five: Direct hydrogenation of the raw crude to a low pour point, low sulfur and nitrogen syncrude. (A combination of 18 direct mode and 6 indirect mode Paraho retorts was selected.)

A total of 24 direct heated Paraho retorts of 42-ft. outside diameter and 72-ft. high would be required to produce the 99,170 bbl per stream day of oil. The huge complex would include, in addition to the mine and underground primary crushing plant, a surface secondary crushing and screening plant, the retorts, each with blowers and an electrostatic precipitator for oil aerosol collection, as well as several prime movers. Separate gas turbines may be used for the burning of the retort gas from small groups of retorts. Separate NH_3 water scrubbers, H_2S scrubbers, gasoline plants, etc. would be needed for each group. Three ancillary units are required when grouping 8 retorts for each gas turbine, compressor, NH_3 scrubber, H_2S scrubber, gasoline plant, etc.

We have chosen for the present study to use atmospheric pressure NH_3 and H_2S scrubbers rather than high pressure scrubbers and not to use a gasoline recovery plant. Thus eight retorts can be manifolded to any number of NH_3 scrubber and H_2S scrubber units in parallel at atmospheric pressure. The three gas turbine units each servicing eight retorts, may be linked parallel thereby allowing for gas combustion with power generation as shown in Figure 3-1. Three water scrubbers for NH_3 removal and three Stretford H_2S scrubbers are assumed per 24 retorts to be connected as in Figure 3-1. Figure 3-3 shows a simplified overall flow diagram of the entire plant variation selected.

3.1.3 Plant and Mine Sites

For purposes of evaluation of the Paraho commercial plant design, a site was chosen on Naval Oil Shale Reserve No. 1 in Garfield County, Colorado, approximately three miles north of the present Paraho Oil Shale Demonstration plant at Anvil Points, Colorado. The topography of the area is hilly, with

FIGURE 3-1. ASSUMED ARRANGEMENT OF ONE PARAFFIN RETORTING AND GAS PURIFICATION SECTION

FIGURE 3-2. PARAHO DIRECT MODE RETORT FLOW DIAGRAM

FIGURE 3-3 FLOW SCHEME OF PARAHO PLANT VARIATION SELECTED

elevations varying from 8000 to 9160 feet. Access would be via a connecting road to Colorado Highway 13.

The plant site and proposed mining area are such that the processing area is immediately adjacent to the mine shafts, which extend 1055 feet downward to the mine floor. Two large retorted shale disposal areas sufficient for a project life of 20 years (at a disposal rate of up to 150,000 tons per day of spent shale plus fines), are located approximately 0.5 mile both north and south of the plant processing area.

3.1.4 Mining and Crushing

A shale layer (the Mahogany Ledge) 76-ft. thick and averaging 29 gal/ton has been selected for mining. It is estimated that this layer averages 2040 barrels per acre-foot mined. An overall extraction efficiency for room-and-pillar mining is 62%. The mineable reserves are 738 million barrels for the 7680 acres to be mined during the 20-year life of the commercial project.

Mining will be by a modified room-and-pillar system, on two levels simultaneously. Conventional room-and-pillar mining will be used on the upper 40 ft. of the shale layer, and a bench method will be used on the lower 36 ft. During the initial preproduction period of 3½ years, while the production shaft, three ventilation shafts, and the man-and-materials shaft are excavated, some 8 million tons of ore would be mined and stockpiled.

The mine, as planned, is composed of a series of mining panels 8 rooms wide and 16 crosscuts deep, separated by 70 ft. barrier pillars. Both the rooms and crosscuts will be 55 ft. wide, with 60 ft. × 88 ft. pillars within the panels. ANFO will be used for primary blasting, at the rate of 0.51 lbs. of explosive per ton of broken shale.

Approximately 53,500 tons of oil shale will be mined per shift, and hauled to eight primary crushers located underground below mine floor level, at the bottom of the main production shaft. Each primary crusher consists of a 48-inch × 72-inch toothed single roll capable of reducing the mined shale to minus 12-inch. The resulting material is hoisted to the secondary crushing plant at the surface.

Water sprays at the underground truck dumping points and in the primary crushing area will be used for dust suppression, together with a

wetting agent for the less strict scenario. Five baghouses will be added here for the more strict scenario. Fugitive dust from the mining shaft will be dispersed into two 40-ft. diameter exhaust ventilation shafts with a total capacity of 10 million. The main production shaft and a 40-ft. diameter intake ventilation shaft will serve as air intakes. The intake air will be heated during colder months by waste retort gases. In addition to the above four shafts, a 30-ft. diameter man-and-materials shaft is also planned.

When the minus 12-inch primary crusher product reaches the surface it is sent to either a 7000 ton plant surge pile, or to a 625,000 ton stockpile, and then to the secondary crushing plant. There, double roll crushers reduce the plus 6-inch oversize to minus 6-inch, which is then combined with the primary crusher undersize and sent to the tertiary crushing and screening plant.

In the tertiary plant eight double roll crushers reduce the plus 3-inch to minus 3-inch, after which the minus 3-inch plus 3/8-inch fraction is separated for retort feed. The minus 3/8-inch fines, which amount to 8,300 tons per stream day (5% of the mined shale) could be sent to the retorted shale disposal pile but are assumed to be briquetted and sent to the retorts.

Particulate emissions in the crushing areas are to be controlled by bag collectors, as discussed later in this report. All conveyor systems are to be covered.

3.1.5 Retorting

The Paraho retort will be discussed in further detail later in this chapter. Briefly, it is a vertical refractory-lined vessel some 42-ft. outside diameter, 39-ft. inside diameter, and 72-ft. high from its lower discharge mechanism to the top feed bin. It is of gas-combustion-type design, with gravity flow of shale downward through four successive zones: mist formation and shale preheating, retorting, combustion (or heating), residue cooling and gas preheating.

Operating Modes: The retort may be operated in either a direct-heated or indirect-heated mode. In the direct mode, Figure 3-2, the retort offgas, at approximately 150°F, is passed through a coalescer (mist separator) and electrostatic precipitator, where product shale oil is removed. Approximately

two-thirds of the denuded product gas, with a heating value of approximately 100 Btu/scf, is then recycled with admixed air to both the combustion zone and the retorting zone of the retort. The recycle gas and a portion of the residual carbon on the spent shale, burn in the combustion zone. This burn furnishes the heat required for retorting. The remaining one-third of the product gas can be used as in-plant fuel or for power generation in gas turbines.

In the indirect mode, not used in this study, there is no combustion zone in the retort. The retort off gases, at approximately 300°F are therefore of high heating value, 800 Btu/scf. After removing product oil, a portion of the retort gases are reheated in an external recycle gas heater. The hot recycle gas is returned to the retort to pyrolyze the shale. The fuel for the recycle gas heater can be a portion of the high Btu retort offgases, product shale oil or shale fuel oil, or other external-source fuel.

Retort Design Parameters: The design parameters for a single commercial size direct mode retort, including the operating conditions, yields, compositions of product gas, and product oil properties are shown in Table 3-1. The shale mass throughput rate is 455 lbs./hr. sq. ft.

In the direct heating mode the retort, offgas leaves the retort at 150°F. After removal of the product oil, approximately two-thirds (14,790 scf/ton of shale) of the total direct mode retort gases (21,990 scf/ton) are recycled to the retort and burned. The remaining 7,200 scf/ton of product gas are treated to remove NH_3 and H_2S , and sent to the plant fuel pool. This product gas is low in heating value, approximately 100 Btu/scf, because of dilution with retort combustion products.

Heat and Material Balances: Heat and material balances for a single Paraho commercial retort operating in a direct heating mode are shown in Table 3-2. The raw shale throughput rate is 546,600 lbs/hr per retort, which is 1/24 of the total 157,420 tons/day for the bank of 24 retorts in the total plant. Although we have grouped the retorts into three batteries of eight, they may also be grouped into four batteries of six retorts each, with each battery fed by a single raw shale conveyor in a Paraho plant.

Table 3-1

Design Parameters of Single Paraho Direct Heated Mode Retort
(From Reference [2])

Operating Conditions

Shale Mass Rate, lbs/hr/ft ²	455
Air to Top Distributor, scf/ton	3,810
Air to Middle Distributor, scf/ton	850
Total Air, scf/ton	4,660
Recycle to Top Distributor, scf/ton	1,360
Recycle to Middle Distributor, scf/ton	1,320
Recycle to Bottom Distributor, scf/ton	12,110
Total Recycle, scf/ton	14,790
Distributor Cooling Water, gal/ton	430
Retort Offgas, °F	150
Retorted Shale Exit Temperature, °F	380
Distributor Cooling Water Rise, °F	18
Retort Pressure Drop, Inches H ₂ O/ft. bed	1.0
Retort Bed Depth, feet	25.5

Yields

Raw Shale Grade, gpt	29
Oil Collected, Volume % Fischer Assay	92
Product Gas, scf/ton	7,200
Retorted Shale, weight % Raw Shale	81

Product Gas Composition (Wet basis)

	<u>Wet Vol. %</u>
H ₂	3.539
O ₂	0.743
N ₂ + Ar	51.342
CO	1.600
CO ₂	17.572
CH ₄	1.858
C ₂ H ₄	0.743
C ₂ H ₆	0.768
C ₃ H ₆	0.364
C ₃ H ₈	0.380
C ₄ s (MW = 57.0)	0.323
C ₅ s (MW = 71.5)	0.129
C ₆ + (MW = 96.2)	0.622
H ₂ S	0.242
NH ₃	0.566
H ₂ O	19.210
Molecular weight	28.43
	100.000

Table 3-1
(continued)

Product Oil Properties

Gravity, °API	21.2
Specific Gravity @ 60°F	0.9267
Pour point, °F	< 80
Viscosity, SUS @ 100°F	162.9
Total Sulfur, Weight %	0.63
Total Nitrogen, Weight %	1.90
Ramsbottom Carbon, Weight %	1.59
BS & W, Volume %	0.14
Carbon, Weight %	84.20
Hydrogen, Weight %	11.69
Oxygen, Weight %	1.24

Table 3-2
Heat and Material Balances
Single Paraho Direct Mode Commercial Retort
(From Reference [2])

<u>Inputs</u>		
	<u>Lbs/hr.</u>	<u>MMBtu/hr.*</u>
Raw Shale Retort Feed (50°F)	546,600	-3.06
Recycle Gas (225°F)	303,780	12.94
Air (175°F)	95,880	2.24
Combustion Heat	---	128.7
	<u>946,260</u>	<u>140.82</u>
<u>Outputs</u>		
	<u>Lbs/hr.</u>	<u>MMBtu/hr.*</u>
Offgas with oil (150°F)	508,380	11.31
Retorted Shale (380°F)	442,740	29.84
Water Vaporization	---	11.94
Carbonate Decomposition	---	33.69
Pyrolysis	---	22.48
Distributor Cooling	---	17.58
Skin Heat Loss & Unaccounted	<u>-4,860</u>	<u>13.98</u>
	<u>946,260</u>	<u>140.82</u>

* Enthalpies shown are based on a 77°F datum temperature.

3.1.6 Shale Oil Recovery Processing Sequences

The variation in the processing sequence downstream from the bank of 24 commercial retorts assumed for the commercial plant in this study is summarized in the generalized flowsheet shown in Figure 3-3. This flowsheet also shows the crushing and retorting systems.

Twenty-four Paraho direct heated retorts are used. The low Btu product gas obtained is treated to remove ammonia and sulfur, and used to generate electrical power for plant use and for export. The crude shale oil produced is given no further treatment other than the addition of a pour point depressant, before being sent to the shale oil product pipeline.

3.1.7 Product Gas Upgrading

The product gas is upgraded or treated for ammonia and sulfur removal.

In the upgrading sequence, the product retort gases could be, in one variation, cooled, compressed, and then further cooled. The resulting condensate (sour water plus hydrocarbons) could be sent to a Phosam Wastewater Treatment unit. The denuded, cooled gas would continue to the ammonia absorption unit where it would be washed to remove ammonia. This ammonia-containing water would also be sent to the Phosam Wastewater Treatment unit, and the water-free gas would continue on to a Stretford sulfur recovery unit for H_2S removal.

The Phosam Wastewater Treatment unit includes both H_2S and NH_3 strippers. The sour water from compression would enter the H_2S stripper, and the ammonia-containing water would enter the NH_3 stripper. The outputs from the Treatment unit are anhydrous ammonia, which is sent to market, stripped water, which is returned to the product gas ammonia removal unit, and H_2S , which joins the product gas entering the Stretford unit.

The compression step, although favored by Paraho, has been omitted from the presently selected variation for study for the sake of simplicity. Also, we have replaced the Paraho favored Chevron Wastewater Treatment plant by an equivalent Phosam Wastewater Treatment plant. The resulting scheme is shown in Figure 3-4.

FIGURE 3-4
FLOWS AND AUXILIARY EQUIPMENT
OF BUREAU OF MINES GAS COMBUSTION RETORT

The Stretford Unit oxidizes H_2S to elemental sulfur of 99.5+% purity, which is removed for sale. The resulting purified fuel gas, containing approximately only 10 ppm residual H_2S , is sent to the fuel pool for use as an in-plant fuel, or as a power production fuel.

3.1.8 Disposal of Retorted Shale

The direct mode Paraho retorting of 157,000 ton per stream day of raw shale results in the production of 126,130 tpsd of spent shale. In addition to this, there will be produced 5% (by weight) of the mined shale as minus 3/8-inch rejected shale fines from the crushing operations. These fines, 8,280 tons per stream day, could also be discarded with the crushed shale, but we have elected to briquette these and send them to the retorts as suggested by the Bureau of Mines.

The retorted shale leaves the retorts at 380°F and is sent to a moisturizer where approximately 5% water is added for cooling and dust control. The resulting material is held in surge bins prior to being transported to the disposal site.

The retorted shale will be conveyed from its surge bins to the disposal area where it will be spread using 100 ton bottom-dump trucks. This will also provide primary compaction. Compactors at the site will increase the in-place density to approximately 90-100 lbs./cu. ft. for the lining and valley dam.

The Paraho valley disposal site plan calls for the construction of a leachate-impervious lining of compacted shale in the valley disposal area, and a cross-valley earthen dam, also of retorted shale, to be built at the mouth of the valley. A catchment basin would contain any water runoff. The retorted shale would be dumped and compacted in the compacted-shale-lined "bathtub" behind the dam, and revegetated. Coherez, a dust-controlling agent, would be sprayed on the retorted shale, as it is deposited. The dumps will be developed with a 4:1 slope and 40-ft. lifts. Each lift will have a 40ft. wide berm for access and maintenance. Detailed discussion of spent shale disposal is contained in Section 7.0.

3.1.9 Material Balance Around Paraho Retorts

Table 3-3 shows the material balance around 24 direct heated mode Paraho retorts in a plant producing 99,170 bpsd crude shale oil from 29 gal/ton oil shale. A further detailed balance is to be found in Section 6.0.

3.1.10 Supporting Facilities

The major utility requirements for the Paraho plant selected are shown in Table 3-4. Included are power input requirements (or exportable power), steam needed, water required (and available), and plant fuel and diesel fuel demands.

Power: Electrical power is to be furnished by an adjoining externally-owned generating plant. The excess low Btu gas is available to generate some 155,500 kW of exportable power to the grid.

Water: The estimated makeup water requirements (water needed minus water available), are discussed in detail in Section 6.0. This water will be furnished, as raw makeup water, from the Colorado River, and suitably pretreated.

Waste Treatment: A wastewater treatment plant and sewage plant will be operated for processing all plant effluents prior to further use. This is also discussed in Section 6.0.

Product Pipeline: A shale oil product pipeline could provide a connection with existing crude terminal facilities 464 km (290 miles) northward at Casper, Wyoming. The estimated cost is \$75 million (in 1976 dollars).

3.1.11 Energy Balance Around Paraho Retorts

Table 3-5 shows the energy balance around the 24 Paraho direct heated mode retorts selected for the present study.

3.1.12 Paraho Direct Mode Retort Detailed Description

A detailed description of the evolution and operation of the Paraho direct mode oil shale retort follows:

Table 3-3
Producing 99,170 BPSD Shale Oil
Producing 99,170 bpsd Shale Oil
from 29 gal/ton Oil Shale

(Adapted from Reference [3])

(24 Direct Heated Retorts)

Raw Shale Mined, tpsd	157,421
Raw Shale Retorting Rate, tpsd	157,421
Retorted Shale Produced, tpsd	126,137
Retorted Shale to Disposal, tpsd	126,137

Retort Production

Crude Shale Oil, bpsd ("Dry" oil containing 1.5% H ₂ O)	100,000
Net Gas, after cleaning, lb/sd, mw = 29.93, 7.59 Vol.% H ₂ O	85,704,000

Net Production

Crude Shale Oil, bpsd	99,170
Ammonia, tpsd (2,000 lb. tons)	144
Sulfur, tpsd (2,000 lb. tons)	132

tpsd = Tons per stream day

bpsd = Barrels per stream day

mmsfsd = Millions of standard cubic feet per stream day

foe = Fuel oil equivalent (total barrels of fuel oil
of equivalent heating value)

Table 3-4

Major Utility Requirements for Paraho Plant Selected

	Power kW	Steam lbs/hr	Water Required gpm	Circulating Cooling Tower Water gpm, $\Delta T=30^{\circ}\text{F}$	Plant Fuel foe bpsd	Diesel Fuel bpsd
Mining, Crushing, and Shale Disposal	77,500	-	2,129	-	-	830
Retorting and Product Gas Upgrading	90,000	225	1,829	69,185	-	-
Offsites, Auxiliaries	10,000	100	478		6,320	-
Subtotal Required	177,500	325	4,436	69,185	6,320	830
Power Plant: Required, or (Produced)	(333,000)	-	-0-	-	18,530	-
Net Totals Required	(155,500)	325	4,436	69,185	24,850	830

Table 3-5
ENERGY BALANCE AROUND 24 PARAHO RETORTS
PRODUCING 99,170 bpsd CRUDE SHALE OIL
FROM 29 GAL/TON OIL SHALE
77°F DATUM TEMPERATURE

Stream	Flow 10 ³ lb/hr	T °F	P psia	Sp. Ht. Btu/lb °F	HHV Btu/lb	Energy Flow 10 ⁹ Btu/hr
<u>IN</u>						
Raw shale and moisture ^(a)	13,118	50		0.243	2,668	34.92
Process air ^(b)	2,339	50	12.5	0.240	--	-0.02
Compression energy ^(c)	(90 MW)					<u>0.31</u>
TOTAL						<u>35.21</u>
<u>OUT</u>						
Spent shale ^(a)	10,511	380		0.21	180	2.56
Product gas ^(d)	3,594	225	12.5	0.26	1,686	6.06
Product oil ^(a)	1,352	120	60	0.62	18,400	24.91
Cooling water evaporated ^(e)	302				(1,400)	0.42
Carbonate conversion ^(f)	1,392				(500)	0.70
Losses						<u>0.56</u>
TOTAL OUT						35.21

NOTES:

- (a) Raw shale. Higher heating value (HHV) and specific heat estimated from data in Cameron Engineers Synthetic Fuels Data Handbook. [4]
- (b) Process air. This is the uncompressed air.
- (c) Compressing energy. $2,339 \times 10^3$ lb/hr air from 12.5 to 15 psia, and $3,571 \times 10^3$ lb/hr product gas and $7,294 \times 10^3$ lb/hr recycle gas from 10 psia to 15 psia. After allowing for machine efficiencies.
- (d) Product gas. HHV calculated from composition, Table 3.
 HHV = 125 Btu/scf $\equiv$ 1,686 Btu/lb (wet basis)
 HHV = 156 Btu/scf $\equiv$ 1,919 Btu/lb (dry basis)
- (e) Cooling water evaporated. Reference [3] gives 430 gal. cooling water per ton shale retorted for cooling gas distributors, $\Delta T = 18^\circ\text{F}$. At 1,400 Btu/lb evaporated, CW evaporated = 302×10^3 lb/hr, or 604 gpm.
- (f) Carbonate conversion. $1,392 \times 10^3$ lb/hr $\text{CaCO}_3 \rightarrow \text{CaO}$, based on heat consumption of 500 Btu/lb CaCO_3 .

Similarity of Paraho Direct Mode Retort and Bureau of Mines Gas Combustion

Retort: The Paraho direct mode retort is essentially the Bureau of Mines "gas combustion" retort. They are identical in general principle and similar in apparatus and operating conditions. No particular attempt to generate and save ammonia is made. Even though ammonia may be evolved in the high temperature gas zone and/or spent shale combustion zone, much of it is destroyed in the recycling process.

Both retorts have high thermal efficiencies. The cool inflowing recycle gas retrieves much heat from the outflowing partially-burned spent shale. Both retorts allow for rapid nucleation coupled with slow condensation rates for the oil, thereby favoring oil aerosol or mist formation. The mechanism involves the cooling of upward flowing oil vapors (bearing a large percentage of inert gases) by the downflowing cool raw shale feed. The oil from both retorts is removed as an aerosol. In practice, some capture of aerosol on the packed bed of raw shale occurs and a condition of refluxing and coking occurs. This produces a lower yield of oil, more carbon on the shale at the retorting zone, and oil of altered properties such as lowered Conradson carbon assay and somewhat lower viscosity and pour point. The spent shale exit grate and raw shale spreader of the Paraho retort was patented by Cameron and Jones [5], formerly with the Bureau of Mines Oil Shale Project.

Certain features of a Bureau of Mines gas combustion retorting plant were distinctive.[6,7] These included the proposal to use a gas turbine in burning the low Btu make gas, and the use of charcoal beds for removing H_2S (and a "smaller" concentration of NH_3) from this gas before the gas turbine. Included was a proposal to use slurry pipelines[6,7] to carry much of the spent shale to underground mine rooms for disposal. Paraho has also proposed use of a gas turbine[8] in a proposed full size single module demonstration plant.

Operation of Bureau Gas Combustion Retort: Bureau of Mines Bulletin 635[9] describes the development of the gas combustion oil shale retorting process. Figure 3-4 shows the flows and auxiliary equipment for the retort[10]. The apparatus is essentially an internal combustion shaft kiln. Within the retort, the shale moves downward as a flowing bed. It is preheated by upward counterflowing retort gases which are composed of new retort offgas and

recycled gas from which the oil has been removed. Shale in a central region of the bed becomes heated to approximately 900°F. This temperature is sufficient to cause pyrolysis of the bitumen and kerogen in the shale. The retort gases containing oil, etc., flow upward in the gas stream. The shale, now retorted or "spent," moves downward through a combustion zone where a mixture of air and low Btu recycle gas is being introduced into the shaft through tuyeres. This recycle gas contains several combustibles such as CO, CH₄, and oil light ends, but not enough retort gas is present in the air/gas mixture to allow combustion to occur in the duct leading to the tuyeres. The low Btu recycle gas along with similar gas moving upward through the shaft at this elevation burns with the air (hence the term gas combustion retort) and supplies much of the heat required for retorting. In the combustion zone, part of the carbon or coke left on the spent shale after pyrolysis is burned. Part of the unretorted shale escaping the upper retorting zone is also burned here. A shale residue containing around 4% carbon remains after complete retorting (without combustion) of shale (for approximately 25 gallon per ton Fischer assay shale). Half or more of this carbon is burned in the direct mode retort.

Green River shale contains considerable dolomite, calcite, and other mineral carbonates, and these thermally dissociate to some extent during retorting. They require considerable heat for dissociation upon which they liberate CO₂, and change to CaO, MgO, etc. It is desirable to limit the extent of this dissociation in order to conserve heat and minimize the concentration of CO₂ in the retort offgas. High concentrations of inerts (CO₂, N₂) in the gas tend to lower its heating value and increase the size requirements for the equipment used for oil-gas separations.

Much of the cool recycle gas burned in the retort is introduced at the bottom of the retort shaft where it flows countercurrent to the outflowing burned shale being withdrawn. In this way, the shale can be cooled to as low as 380°F and it is thus made easy to handle and much heat otherwise lost is saved. This simple heat-saving scheme is one of the attractive features of the gas combustion retort.

The retort offgases consist of the large volume of low Btu recycle gas and the oil aerosol from the pyrolysis zone. The oil aerosol is removed

from these gases. Sometimes they are led first through "rotoclones" (centrifugal blower separators) and then through electrostatic separators. Due to the low degree of agitation of the shale bed and its action as a dust removal device itself, there is less dust in the retort offtake gases than in those of some other retorts. This is a second attractive feature of the gas combustion retort. In some other retorting systems (such as TOSCO II, Lurgi, and a number of fluidized bed systems using admixture of raw shale with heat transfer balls and attrition, with hot spent shale and mixing, or with hot spent shale and fluidization) dust in the retort offgases is troublesome.

An interesting feature of Figure 3-4 is that a gas temperature of approximately 150°F may be selected instead of 130°F. This higher temperature will prevent aqueous condensation with the oil product.

Scale-Up of Bureau of Mines Gas Combustion Retort: For a period of years, a 20-inch diameter 6 ton per day retort was tested under various operating conditions. Then a 25 ton per day 6 ft. x 10 ft. rectangular retort was tested. Upon discontinuing its program at Anvil Points near Rifle, Colorado in 1956, the Bureau of Mines had attained mass flow rates of 300 lbs/hr/ft² with its Gas Combustion Retort.

Development Engineering, Inc. (DEI), a six company consortium of Mobil, Humble, Pan American, Sinclair, Phillips, and Continental leased Anvil Points and continued development of the gas combustion retort between 1964 and 1967. They further studied the 6 ton and 25 ton per day Bureau of Mines retorts. They then rebuilt the 150 ton per day unit. They were able to further increase the mass flow rate in this work and obtain oil yields as high as 87% of Fischer assay. Difficulties with fines in the charge at higher rates of gas and shale throughput, and difficulties of bridging with rich shales were encountered. Apparently the Paraho retort version of the gas combustion retort has overcome these difficulties to a considerable degree through the use of their 10.5-ft. diameter 450 tons per day nominal throughput unit.

In 1972, DEI obtained a lease on the Bureau of Mines oil shale plant and mine. DEI hoped to prove that the plant and mine technologies were indeed based on patents of John B. Jones of DEI. This technology included improvements to the gas combustion retort and an indirect heated retort such as the

Petrosix retort. The Paraho Development Corporation was formed with DEI as a subsidiary and in 1973, with 17 participants, the Paraho Oil Shale project began with Arthur G. McKee and Co. as engineering contractor. The 17 Paraho participants are Atlantic Richfield, Carter Oil (Exxon), Chevron Research (Std. of California), Cleveland-Cliffs Iron Co., Gulf Oil, Kerr-McGee, Marathon Oil, Arthur G. McKee, Mobil Research, Phillips Petroleum, Shell Development, Sohio Petroleum, Southern California Edison, Standard Oil Co. (Indiana), Sun Oil, Texaco, Webb-Chambers-Gary-McLoraine Group.

Thus the Paraho retort evolved from the gas combustion retort selected by the U.S. Bureau of Mines for extensive development. It is appropriate but not perfect for retorting calcareous Colorado-Utah-Wyoming Green River Formation oil shales. Much NH_3 , however, is lost in the direct heating mode system while the indirect mode produces very little NH_3 . The two modes of operation of the basic retort shaft and of the associated refining plants have their counterparts in much Bureau experimental work and conjecture.

The oil yields and properties of both modes are rather similar to those from the simple laboratory Fischer assay retort. The Conradson carbon assay of the oil is, however, lower due to oil refluxing in the direct mode retort. (This also lowers the oil's pour point). The high pour point of the shale oil from the Green River Formation has long been of concern, for there is a possibility that during winter, the oil in an unheated long pipeline of crude shale oil would become too viscous to pump. The Paraho direct mode oil has a pour point slightly lower than that of Fischer assay and other retorts.

Operation of Paraho Direct Mode Retort: Figure 3-5 is one flow scheme shown for the Paraho direct mode retort.[11] An alternate scheme omits the condenser. Near the bottom of the retort, cooling of the downflowing charge is accomplished before the gas mixture is autoignited. There is some question as to the extent of various combustion reactions in the various zones above the middle and top distributors but a rather uniform temperature is obtained throughout the entire combustion zone (see Figure 3-6) thus minimizing carbonate decomposition, shale ash sintering, and clinker formation. The coke or carbon dispersed in the retorted shale is said to be the principle fuel to heat the process.[11]

FIGURE 3-5. ONE FLOW SCHEME SHOWN OF PARAHO DIRECT MODE RETORT

FIGURE 3-6
TEMPERATURE PROFILE IN PARAHO DIRECT
MODE RETORT*

*from U.S. Patent 3,736,247

Tests performed in 1974, and 1975 showed higher oil recovery, lower coke or carbon in the retorted shale, and lower carbonate decomposition than expected. It was also possible to feed greater amounts of raw shale fines smaller than 1/2-inch. After a 56-day run, 10,000 bbls of shale oil were shipped to the Gary Western Refinery at Grand Junction, Colorado. Seven different military fuels were produced in cooperation with Sohio Petroleum Company[8].

The Paraho direct mode retort and the Bureau of Mines gas combustion retort are operationally similar. One distinction between them is the addition of air to the lower recycle gas of the Paraho retort. The Paraho retort is thereby said to derive its heat from carbon combustion rather than simply gas combustion.

Sufficient bed porosity must be maintained in the Paraho retorts. Decreased bed porosity increases resistance to countercurrent gas flow. For this reason, the smaller shale fines are screened out before the shale is fed to the retorts. The smallest fines would tend to be blown up into the product oil-gas stream and increase the dust content of the crude oil. Rat holing, channeling, and blocking of gas add further complications when too large a fraction of fines is present in the feed.

The captured oil contains dust from the retort. This generally necessitates the distilling of all product crude oil so that the dust, along with some heavy ends, may be separated from the lighter fractions. The dust and heavy ends are then sent to a coker. The coker pyrolyzes the heavy ends producing gas, lighter fractions of oil, and coke. The dust becomes incorporated in the coke thus lowering its quality and limits its uses, but the dust is neatly disposed of. Recently the coker has been omitted from a White River Shale project plan.[12] The coker and entire oil refinery have been omitted in the Paraho model selected in this present study.

Scale-Up and Efficiency of Paraho Direct Mode Retort: Figure 3-7 originates in a paper by McKee and Kunchal,[13] and shows a block diagram of a projected shale oil plant using 16 Paraho direct mode retorts. These retorts require a total of 160,000 tons per stream day of mined 30 gal/ton shale. A prerefining delayed coking step produces 2,600 tons per stream day of byproduct coke. Also fine shale screenings to the extent of 5% of the mined shale must be

FIGURE 3-7
BLOCK DIAGRAM OF A PROJECTED PARAKO OIL SHALE PLANT
OVERALL ENERGY BALANCE FOR A 100,000 B.P.D. SHALE OIL PLANT
BASIS: PARAKO INDIRECT MODE RETORTING
30 B.P.T. OIL SHALE (Billions of BTU's)

discarded.[13] This is a smaller fraction than the earlier estimated 8% rejected amount reported in the earlier "Paraho Prospectus for Full Size Module." [8] It is based on further operation of the Paraho semiworks unit.

Discarding the shale screenings, of course, would be a resource waste, and due to its poor marketability, the coke from a refinery coking step would also be a probable source of energy waste. A White River Project [12] scheme has avoided coking by hydrotreating uncoked raw retort oil. A Rio Blanco [14] scheme burned half of the delayed coker coke with oxygen from an oxygen plant to produce much of the hydrogen needed for shale oil hydrotreating. This method was chosen rather than the otherwise more economical steam reforming of high calorie hydrocarbon gases from cooperating TOSCO II retorts, delayed cokers, and other operations.

Paraho has proposed that each retort have an outside diameter of 42 feet. Operating pressure is to be slightly above atmospheric. Early criteria for the design of the retort are compared with the experimental values later obtained in the Paraho semiworks retort as recently reported. [13, 11] (See Table 3-6.) Also shown for comparison are some selected figures from the Bureau of Mines pilot plant "gas combustion" retort [9] and Bureau of Mines projections for a 56-foot diameter and 60-foot diameter retort. [6] The closely related Union Oil Company Retort [15] is also compared in its semiworks form.

The major ancillary equipment required for each retort module in the proposed Paraho Direct Mode plant includes a recycle gas blower, air blower, inert purge gas compressor, and two electrostatic precipitators. The air and recycle gas blowers are run by a gas turbine according to designs in the recent prospectus. [40] All other prime movers would be electric motors. An inert gas is introduced into the shale feed chutes just above their discharge level. This inert gas will be under pressure slightly higher than retort gas pressure. The top of the retort will therefore be sealed against escaping retort gases. Retorted shale is removed from the bottom of the vessel through a battery of rotary seals. Dust collector systems are provided for both the entry and discharge ends of the retort.

Many versions of ancillary equipment exist for this process. For example, for the White River Project [12] proposes steam power for driving electric alternators whose electric output would be used for driving the

Table 3-6
Comparison of Gas Combustion and Union Oil Retorts

		Early Paraho Design Criteria ⁴⁰	Paraho Experimental Retort ⁴¹	USBM Exptl. Gas Combustion Retort 150 tpd ⁴²	USBM Pro- jected Gas Combustion Retort 56' Dia ¹⁹	Union Oil Company Retort ¹⁵	Paraho Design This Study
Maximum Throughput Rate	lb/hr ft ²	700		299	500		455
Liquid Recovery	% of Fischer Assay	95	97	82.8	94.4	92 ?	92
Oil Shale	Inches	$\frac{1}{4}$ to $3\frac{1}{2}$	$\frac{1}{4}$ to $3\frac{1}{2}$	3/8 to 3	3/16 to 3		3/8 to 3
Gas Produc- tion Heating Value	Btu/Ton Shale	800,000	632,000	785,000	441,000	1,124,000	923,800
Fischer Assay of Feed	gal/ton	30	28	28	30	30	29
Air Rate	scf/ton	4,800		3,940		10,150	4,660
Product Gas Rate	scf/ton	--	--	--	--	--	7,200
Recycle Gas Rate	scf/ton	16,000		13,240	82%	-0-	14,790
Seal Gas	scf/ton	840		--		-0-	

retort blowers, mining, crushing and conveying, offsite power, and other plant power. Since an elaborate hydrotreating scheme was proposed by White River, a deficiency of electric power of 32.5% was counteracted by burning some syn-crude product. Alternately, an early Rio Blanco project plan[14] obtained plant and retort blower power from purchased electricity. TOSCO and Colony have similarly planned for purchased electricity usage in their TOSCO II, process. The Bureau of Mines gas combustion retort, in recently proposed form [6] has also used gas turbines burning low Btu retort gas to generate electric power. The retort blowers and other equipment are then driven by electric motors.

Ammonia Evolved During Oil Shale Retorting: In the shale oil industry of Scotland, ammonia production was of greater importance than oil production. The shale was heated to a much higher temperature than that possible in the low temperature TOSCO II or Lurgi processes, and steam was injected to secure a maximum ammonia yield. In Scotland, the combustion gases were generally kept separate from the retort gases and heat was transferred through the retort walls although it was possible to secure partial internal combustion by limited air injection. Air injection tripled the shale throughput rate with no loss of oil yield per ton and an ammonia yield reduction of only about 27%.[16] High temperatures (1400°F) were attained by the shale at the bottom of the moving charge in the Scottish Pumpherston retorts.

Ammonia has also been produced in the experimental NTU (Nevada-Texas-Utah) internal combustion batch retort with Green River shale. Here again, high maximum shale temperatures were attained by internal combustion.

Karrick[17] has stated that the nitrogen compounds in oil shale do not yield much ammonia until 750°C (1382°F) has been attained and steam is necessary. In the Pumpherston retort, 65% of the nitrogen in the shale was converted to ammonia and Karrick predicted that this recovery could be expected from American shales if sufficient steam were used and 1500°F were attained. Figure 3-8 shows nitrogen content as a function of Fischer assay in Colorado Green River shale as found by Bureau of Mines work.[18] Earlier Bureau of Mines data for Green River shale reported by Gavin[19] gave a somewhat different correlation summarized by the straight line in Figure 3-8. Thus, roughly, the nitrogen content of Green River shale varies directly with

FIGURE 3-8. NITROGEN CONTENT OF GREEN RIVER OIL SHALE vs FISCHER ASSAY.

oil yield. The Devonian Shales of the eastern United States contain more nitrogen for a given oil yield than Green River shales but again there is a direct linear relation.[19]

Experiments by the Bureau of Mines[19] indicated that conditions of producing the oil from shale largely determine the amount of nitrogen that remains in the spent shale to be acted on by the steam at subsequent high temperature. "Rapid production from the shale--rapid heating and rapid removal of vapors--yields an oil containing a relatively high percentage of nitrogen and a spent shale with a relatively low percentage of nitrogen." "Slow production of oil reverses this effect." [These results suggest that coking of the shale oil occurs to a greater extent with greater oil vapor residence time and "deeper" coking. It is known that shale oil nitrogen tends to be concentrated in shale oil coke during coking.]

Gavin[19] suggested that treating spent shale from preliminary oil retorting in a separate gas producer independent of the retort should increase ammonia yield almost to the total theoretical amount possible from the nitrogen in the spent shale. But he also stated that in view of the more recent, less expensive ammonia production methods [Haber Process], it may be more profitable to neglect ammonia production if increased shale throughput per retort were thereby possible.

Thus in the direct mode gas combustion retort, an opportunity for ammonia production exists that is not present in the low temperature TOSCO II, Lurgi, and Paraho indirect heated retorts where shale combustion at high temperatures with steam does not occur. "Gas combustion retorts" here includes the Paraho direct mode, Bureau of Mines, Union Oil, Fushun, and other retorts where a high shale temperature is obtained, generally through internal combustion, and steam from recycle gases or from steam injection is present. MIS retorts should also be included, particularly when steam injection is used along with the air feed.

Table 3-7 summarizes data found on ammonia or ammonium sulfate production from various retorts and shales.

The Paraho and Bureau of Mines gas combustion retorts burn approximately 67-75% of the ammonia they produce. This burning occurs through recycling of this fraction, along with recycle gas, through the gas combustion zone of the retort.

Table 3-7. Ammonia From Various Shales and Retorts

1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	
														Lb NH ₃ Equiv. Produced Per 1000 lb Oil Theor. Produced (Fischer Assay)	Lb NH ₃ Equiv. Per 1000 lb oil Produced In Total Retort Gas (Assuming No NH ₃ in Recycle Gas Burned)			
Shale Source	Shale Fischer Assay, GPT	Retort	Recycled Gas Make Gas Ratio	Steam used	Oil Yield, % of Assay	Nitrogen in Shale Feed, %	Temperature of Combustion or Hottest Zone	% of N ₂ in Shale Converted to NH ₃	% of N ₂ in Shale Converted to NH ₃ and N	% of N ₂ in Shale Deposited in Spent Shale	NH ₃ in Retort Gas % Dry Basis	(NH ₄) ₂ SO ₄ Produced, lb/Ton Shale	Lb NH ₃ or NH ₃ Equiv. Per 1000 lb Oil Produced	1000 lb Oil Produced (Fischer Assay)	1000 lb oil Produced In Total Retort Gas (Assuming No NH ₃ in Recycle Gas Burned)	Reference	Date of Reference	
Green River	Variable	Fischer	0	no	100	Variable	900°F	9%	12%	38%		-					USBR, RI 4825 ⁵⁶	
Green River	-40	Lab. Tube Furnace	0	no	100	0.50	932°F	6%	-								Maier & Drapeau ^{106, 107}	
Green River	28	Paraho Direct	3.0	no	-97	-0.43	-1300°F				0.249	-	3.34	3.24	13.4	Jones ⁴¹	1976	
Green River	28	Paraho Indirect	0	no	-97	-0.43	1000°F				1.2	-	1.309	1.270	1.27	Jones ⁴¹	1976	
Green River	29	Paraho Direct	2.04	no	-92	-0.45	1300°F				0.1839	-	2.72	2.50	7.60	DRI ¹⁰²	1977	
Green River	26.5	Paraho Direct	2.04	no	-97	-0.41	1300°F				0.700	-	10.9	9.52	33.2	NH ₃ from Heistand ¹⁰³	1978	
Green River	-45	Indirect (George)	0	no		0.6 ¹⁸	788°F				Very Small	Small		7.18	7.18	George ⁶⁷		
Green River	-45	Indirect & Steam Injection (George)	-	yes		0.6 ¹⁸	>1112°F		0			19.5 to 25.2		to 9.29	to 9.29	George ⁶⁷		
Green River	-30?	Union Oil A	High	no	92?		Very Hot											
Fushun	15	Fushun	High	no	-88.4		Very Hot					25.0	31.2	27.6	31.2	48, 45, 52, 54		
Green River	23	Occidental MIS	0	yes	70	-0.36					0.6**	-	22.8	15.96	22.8	NH ₃ from Skogen ¹⁰⁴ for 15 GPT Shale*		
Green River	22.5	Laramie 10 Ton NTU	0	yes		-0.35						22.9 Equiv.		16.86	16.86	Martel & Harak ¹⁰¹		
Green River		Livermore 2 ft. Dia.	yes								0.005-0.065							
Scotch	19.4-57.8	Pumpherson	0	yes			1500°F	65%				36-98		28.8	28.8			
Nova Scotia	18		0	yes								35		32.2		McKee ⁵⁵ p. 46		
Antigonish County	24.5											30.2		20.4		McKee ⁵⁵ p. 48		
Quebec	24.0											22		15.2		McKee ⁵⁵ p. 48		

*Perhaps should be higher for 23 gpt shale.

The ammonia produced in both the Paraho direct mode and MIS retorts is at a concentration level too low to be very profitable to recover, yet too high to be ignored as it would then cause air and/or water pollution. Moreover, there is evidence that the burning of the ammonia may produce some NO_x in the retort gas.

The Manchurian Fushun oil shale retort[5,16,20,21] does not recycle ammonia to the combustion (gas producer) section but scrubs it out of the entire retort gas stream before part of the stream "circulates." In this way, from 1,360,000 tons of 15 gal/ton oil shale, 17,000 tons per year of $(\text{NH}_4)_2\text{SO}_4$ and 70,000 tons of shale oil was produced or the equivalent of 2,187 tons per year of NH_3 which is 0.03 lb NH_3 /lb shale oil. A sulfuric acid scrubber was used to capture the NH_3 as sulfate. At an assumed oil specific gravity of 0.93 this is equivalent to 11.31 lb NH_3 /bbl oil. Oil yield was apparently $\frac{11.47}{15} 100 = 76.4\%$ of Fischer assay.

In Figure 3-5 a condenser is shown in the entire retort gas stream of a Paraho direct mode retort. Present trends show a preference to maintain the gases at a temperature sufficient to avoid any water condensation. Depending on the extent to which water does condense some ammonia will be collected in the condensed retort water primarily as ammonium carbonate and secondarily as ammonium sulfide. The amount collected will depend, in part, on the amount of water condensed. A countercurrent water scrubber can be used to very effectively collect the ammonia especially in the presence of large amounts of CO_2 . This is generally not considered for the entire retort gas stream, apparently because of the much larger volume of gas that must be handled. Usually an ammonia scrubber is applied only to the product gas purged out before the recycle gas stream. For the Paraho retort, the scrubber cross section can be reduced to about 1/3- to 1/4-size of an alternatively positioned scrubber. This results in lesser amount of NH_3 collected; the rest being recycled to the retort and burned.

With large amounts of CO_2 present in the retort gas of a gas combustion retort, H_2S is not appreciably captured along with the NH_3 . This is the case for a Paraho direct mode retort or MIS retort using carbonate rich Green River shale. The Fushun shale is very low in carbonate hence the CO_2 concentration in the Fushun retort would be relatively low as little mineral carbonate dissociation would be expected. This may explain the use of sulfuric

acid scrubbing at Fushun rather than water scrubbing, for much ammonium sulfide might otherwise be collected presenting a difficult separation problem before NH_3 could be produced. In the primitive Fushun works, shipment of ammonia might have been more difficult than shipment of ammonium sulfate.

As will be discussed under Alternate Processing Approaches, Section 3.5, a choice exists in water scrubbing ammonia out of the entire retort gas stream or out of only the smaller product gas stream. Much of the decision depends on the value of the byproduct ammonia and the quantity that can be collected. Recent gas analyses of the Paraho direct mode gas seem to indicate higher ammonia contents than were previously reported. Previous lower figures may have been due to partial condensation of retort water in places where NH_3 could be captured but not fully accounted for. Small pilot plant scale apparatus, for example, could have cold spots. The NH_3 concentration recently reported in MIS retort gas is also high and tends to confirm the higher Paraho value when allowance is made for burning of that part of the NH_3 in the Paraho recycle gas.

Ammonia absorption represents an important and expensive part of the overall pollution control strategy. The specifics of handling ammonia are discussed in Sections 5.0 and 6.0.

3.2 TOSCO II PROCESS

3.2.1 Basic Parameters Assumed for TOSCO II Plant

The TOSCO II process, for the purpose of this project, would be located in the upper Parachute Creek area on the Colony property. Room-and-pillar mining in from the Piceance Creek Basin edge face will produce 35 gal/ton raw shale from the Mahogany zone of the Green River Formation. The water source for the process will be the upper Colorado River at Grand Valley. Crude shale oil (55,000 bpsd) will be upgraded to 47,000 barrels per stream day of high quality syncrude and 4,300 bpsd of LPG using the proposed TOSCO refining process. Crude shale oil yield will be 100% of Fischer assay. The high Btu gas produced by the retorting process will be consumed on-site by the retorting and refining operations. Ammonia and sulfur will be recovered as byproducts. Electric power will be purchased from off-site. Water will be considered a scarce resource and will be recycled or reused. Overall there

will be zero discharge of aqueous effluents. Carbonaceous spent shale will be disposed of in Davis Gulch with wetting and compaction.

3.2.2 TOSCO II Plant Description

The TOSCO II Process has been developed over the past 20 years, from bench-scale through a 24 ton per day pilot plant, to a 1,000 ton per day semiworks unit. Both the semiworks unit and an associated room-and-pillar mine in western Colorado were operated from 1964 to 1972. During this period over one million tons of shale were mined, of which 290,000 tons were later retorted in the semiworks plant.[28, 29]

A full-scale 66,000 tons per stream day commercial plant has subsequently been designed for location on the Dow West property, Parachute Creek, Garfield County, Colorado. This plant would produce, on a daily basis: 47,000 barrels of low sulfur fuel oil, 135 tons of ammonia, 193 tons of sulfur, and up to 4,300 barrels of LPG for sale. There would also be obtained 800 tons of green coke daily.

The commercial process, as designed, includes an underground mine, retorting and oil recovery plant, and upgrading plant. All in-plant fuels will be furnished by the process itself, including the diesel fuel for the mine and the fuel gases (or fuel oils) for the retorting and upgrading operations. Electrical power will be purchased outside.

The air emissions, effluents and solid wastes from the mine, retorting plant, and upgrading units have been computed for the commercial plant complex. The total hourly air emissions for our less strict control scenario are estimated to be as follows: 283 lbs. SO_2 ; 1,934 lbs. NO_x ; 62.3 lbs. CO; 494 lbs. particulates; 345 lbs. total hydrocarbons. These emissions are within the current specifications of the Colorado Air Quality Control Regulations.

No discharge of aqueous effluents are expected from the plant complex to the environment. Treated plant waters and other aqueous discharges will be used to moisturize the processed shale for disposal. Any leaching of the disposal pile will be captured in a catchment reservoir and returned to the pile.

Approximately 20,220,000 tons of solid plant wastes will be disposed of annually, or an average of 55,400 tons daily. Some 97% of this waste, or

54,200 tons per day, will be processed shale (and its dust). An additional 1,285 tons per day will be raw shale dust. The remaining solids will be comprised of spent catalyst materials, sludges, arsenic-laden solids, and processed plant sanitary wastes.

Figure 3-9 is a block diagram of the TOSCO retorting and upgrading units.

Figure 3-10 is a flow diagram of the initial operations of the Colony adaptation of the TOSCO II retort in the variation chosen in this study to satisfy the "less strict" pollution control scenario. A "more strict" scenario involves inclusion of electrostatic precipitators and baghouses in place of some wet scrubbers for some particulate controls. It also involves classification of spent shale as a "hazardous waste" and construction of a shale-cement liner for containing the dumped spent shale.

3.2.3 Colony Underground Mine

The proposed Colony commercial plant includes a conventional underground room-and-pillar mine designed to recover about 60% of the in-place 35 gal/ton shale from a 60 ft. seam of the Mahogany Ledge located approximately 900 ft. below the surface. Access to the mine will be from a 4.5 acre portal bench through eight 30 ft. x 30 ft. adits, four on the west and four on the east of Middle Fork Canyon. The mine will produce 66,000 tons per stream day of oil shale from a 60 ft. seam of the upper Mahogany Zone. Some six 50 ft. x 30 ft. ventilation openings, without scrubbers, will be provided for air circulation and for removal of fumes. The diesel equipment used underground will utilize mounted catalytic scrubbers, with a total NO_x emission discharge rate of 250 lbs./hr.

The mining cycle will include the conventional sequence of drilling, charging, blasting (3 times daily, between shift changes), rock pile wetting, loading, hauling, scaling, and roof bolting. The run of the mine shale will be conveyed to a primary crusher located outside the mine, on the canyon floor below the portal bench.

3.2.4 Crushing, Screening, Conveying Operations

In order to prepare the feed for the proposed Colony retorting and upgrading plant on the Middle Fork of Parachute Creek, the minus 48-inch

FIGURE 3-9

BLOCK FLOW DIAGRAM TOSCO II RETORTING AND UPGRADING UNITS

FIGURE 3-10 FLOW DIAGRAM OF TOSCO II PLANT VARIATION SELECTED

plus 8-inch run of the mine shale will be trucked to the top of the primary crusher, which is to be located on the valley floor at the mine portal bench. There, a single tooth roll with a maximum capacity of 4,500 tons/hour will reduce the shale to a minus 8-inches coarse feed. This feed is then conveyed to a 15 acre, 200 ft. high, coarse ore stockpile with a capacity of 1.5 million tons. From the pile the ore is conveyed to the final crushing units at the main plant site on top of the plateau, 900 ft. above the mine portal bench. The final crushers will reduce the coarse feed to minus 1/2-inch, for charging to the retort. An alternative crushing sequence might be primary crushing to minus 2 inch in a double-tooth roll, followed by a final crushing to minus 1/2-inch in a short-head cone gyratory crusher, or an impact crusher.

Dust control in the enclosed primary crusher area will be accomplished by baghouse filters. Particulate emissions in the 62,200 cubic feet per minute of stack gas emissions from the primary crusher area are estimated to be 8 lbs. per hour.

An enclosed inclined conveyor system will transport the coarse ore from the primary crusher to a storage pile at the final (secondary) crushing plant on top of the plateau. At the final crushing plant a bank of 10 impact crushers with a total capacity of 2,700 to 3,000 tons per hour will reduce the coarse ore to minus 1/2-inch fine ore feed for the retorts. Dust control, once again, will be accomplished through a baghouse filter system. The fine ore feed will be conveyed to a three silo storage area with a total capacity of approximately 15,000 tons. A baghouse filter will control dust in the storage area.

3.2.5 Retorting and Oil Recovery Unit

The heart of the Colony processing sequence is the TOSCO II pyrolysis (retorting) unit and associated oil recovery equipment. The commercial plant will be composed of six retorting/oil recovery units or "trains," each with a design capacity of up to 11,000 tons/day of raw shale.

The minus 0.5-inch raw shale from the final crusher is first fed to a dilute phase fluidized bed, where it is preheated to about 400°F with flue gases from the ball heater. The cooled flue gases are separated from the preheated shale and venturi wet-scrubbed in the least strict control scenario to remove particulates, which are disposed of as a sludge (A, Figure 3-10).

Some total 137 tons per day of particulates will be recovered and discarded on the processed shale pile from six trains. The resulting preheat system gases (B, Figure 3-10) will be vented to the atmosphere at about 130°F. Fabric filters are substituted for the wet scrubbers in the more strict control scenario.

The preheated shale is fed to a horizontal retort (pyrolysis drum), together with approximately 1.5 times its weight in hot ceramic balls from a ball-heater, in order to raise the shale to a pyrolysis temperature of 900°F and convert its contained organic matter to shale oil vapor. The shale vapors are withdrawn and fed to a fractionator, for hydrocarbon recovery. The mixture of balls and spent shale are discharged through a trommel, in order to separate the now only warm balls from the spent shale.

The warm balls are purged of dust with flue gases from a steam preheater, and the dust removed from the flue gases in the least strict control scenario by a high energy venturi wet-scrubber. The total of six 11,000 tons/day "trains" of the commercial plant will produce sludge dust (C, Figure 3-10) for disposal from this source. The resulting scrubbed flue gases are discharged to the atmosphere. In the most strict control scenario this stream and the shale wetter stream is combined and sent to a hot electrostatic precipitator.

The dust-free warm balls are returned to the ball heater via the ball elevator. In the ball-heater they are reheated to about 1300°F using in-plant fuel, and recirculated to the pyrolysis drum.

The hot spent shale (denuded of oil) is cooled to about 150°C (300°F) in a rotating drum cooler, and moisturized to about 14% water content with 1,500 gallons/minute per train of NH₃-free, H₂S-free, and CO₂-free water from the plant's foul water stripper unit. The wetted shale, at below 200°F, is transported to the disposal site. The steam-dust mixture produced in the moisturizer is, in the least strict control scenario, venturi wet-scrubbed to remove dust, and then vented to the atmosphere (E, Figure 3-10). (The total of six trains discharge 44 lbs./hr. of particulate matter to the atmosphere from the moisturizer scrubber stacks, and dispose of 6.5 tons/day of dust in wet sludge (F, Figure 3-10).

The shale oil hydrocarbon vapors from the pyrolysis drum are separated into water, gas, naphtha, gas oil, and bottom oil in a fractionator.

The condensed water is sent to the foul water stripper to remove NH_3 and H_2S , and then used to moisturize processed shale. The gas and naphtha is piped to the gas recovery and treating unit, the gas oil to a hydrogenation unit, and the bottoms oil to the delayed coking unit.

The sludges from the three wet venturi scrubbers per train, described above, are all discarded on the spent shale pile.

The process heat for the entire six trains of the retorting and oil recovery plant is intended to be fuel gases, liquid butanes, and/or fuel oil from the retorting and subsequent upgrading units.

3.2.6 Upgrading Units

As shown in Figure 3-9, the upgrading section of the commercial plant consists of the following units: Gas recovery and treating, hydrogen production, gas oil hydrogenation, naphtha hydrogenation, ammonia separation, sulfur recovery, delayed coking and foul water stripping. These upgrading units process the individual fractionator product streams from all six pyrolysis and oil recovery trains.

Gas Recovery and Treating Unit: The gas and raw naphtha from the six shale oil fractionators, delayed coker, and the naphtha and gas oil hydrotreaters are all fed to the Gas Recovery and Treating Unit, where they are separated into stabilized naphtha, LPG, butanes, butanes/butylenes, and fuel gas. The stabilized naphtha is sent to the naphtha hydrotreater. The butanes/butylenes are used as in-plant fuel, as is also part of the treated fuel gas. The remainder of the fuel gas is fed to the hydrogen plant. The LPG is sent to storage, for sale. Acid gas from the amine treaters is fed to the sulfur unit. Since all streams exiting from the Gas Recovery and Treating Unit are sent to other units, there are no emissions to the atmosphere, or byproduct liquid effluents.

Hydrogen Unit: Hydrogen is needed for the naphtha and gas oil hydrotreaters, to remove nitrogen and sulfur and saturate olefins. This hydrogen is produced from a portion of the fuel gas from the Gas Recovery and Treating Unit, using a conventional steam reforming process at (1400-1600°F), with the usual nickel catalyst plus follow-on iron/chromium oxide and copper/zinc shift catalysts.

As noted in the flowsheet the various processing steps include desulfurization, reforming, CO conversion to CO₂, CO₂ removal, and methanation.

A three-step desulfurization is necessary to prevent downstream catalyst poisoning. The sequence involves hydrodesulfurization to H₂S, solvent absorption of the H₂S, and trace removal of any residual sulfur compounds over a ZnO catalyst.

In the reforming furnaces and follow-on CO shift converters, there are produced a high concentration H₂ gas with low residual CO content, plus CO₂ and H₂O. A solvent absorption unit removes the CO₂, which is then desorbed and discharged to atmosphere. The resulting gas stream is passed over a nickel-based methanation catalyst, where any residual hydrogen-rich product gas is sent to the naphtha and gas oil hydrotreaters.

The gaseous emissions from the Hydrogen Unit are (a) the flue gases from the reforming furnaces (two stacks per furnace), each of which burns 15,500 lbs./hr. of in-plant treated fuel gas; and (2) the CO₂ discharged to atmosphere from the single 3-ft. diameter, 210-ft. high stack of the CO₂ absorption unit.

In addition to these gaseous emissions there will be discarded annually on the processed shale pile, or sent offsite for reclamation some 67 tons of spent catalysts from the Hydrogen Unit, as itemized below:

	<u>Tons</u>
Spent HDS catalyst	34
Spent ZnS catalyst	7
Spent Fe-Cr catalyst	10
Spent Cu-Zn catalyst	16

and 876 tons of spent aqueous caustic.

Gas Oil Hydrogenation Unit: The feeds to this Unit are the gas oil streams from the oil recovery fractionators and the delayed coker unit, and hydrogen from the Hydrogen Unit. The feedstreams are first heated to reaction temperature in a furnace and passed over a proprietary catalyst to remove arsenic, which would foul the subsequent hydrogenation catalysts. Some 530 lbs. of arsenic will be removed daily by this method, and shipped off-site for disposal. The purified feed is then hydrogenated over an HDN catalyst to reduce nitrogen and sulfur, and to hydrogenate the unsaturated hydrocarbons present.

The hydrogenated stream is washed with 240 gpm of water from the ammonia separation unit, in order to absorb NH_3 and H_2S , and the sour water separated and returned to the ammonia separation unit. The washed "hydro-treated gas oil" stream is then fractionated to produce overhead gas and naphtha (which are returned to the Gas Recovery and Treating Unit), 380 bbl per day of diesel fuel for mine and in-plant use, and a low sulfur (0.8%) and nitrogen hydrotreated gas oil for sale.

The gaseous emissions from the Gas Oil Hydrogenation Unit are from two 2.5-ft. diameter by 75-ft. high stacks of the 2 reactor furnaces; and from the 5.7-ft. diameter by 150-ft. high single stack of the reboiler furnace. All three furnaces burn treated in-plant fuel gas or fuel oil.

Solid wastes include up to 130 tons per year of spent HDN catalyst, and the aforementioned spent arsenic catalyst.

Naphtha Hydrogenation Unit: The stabilized naphtha stream from the Gas Recovery and Treating Unit is catalytically hydrogenated to remove sulfur and nitrogen and to saturate the olefins present, in the Naphtha Hydrogenation Unit. Prior to hydrogenation, arsenic is removed over a proprietary catalyst. Some 59 lb/day of arsenic are produced, for discard on the spent shale pile.*

The arsenic-free feed is hydrogenated in the presence of an HDN catalyst and the reactor effluent is washed with stripped water from the ammonia separation unit, in order to remove H_2S and NH_3 . The resulting sour water is sent to the Ammonia Separation Unit for purification.

The resulting H_2S and NH_3 -free mixed hydrocarbon stream is fed to an absorber to remove heavier hydrocarbons from the overhead gases. The remaining overhead gases are returned to the Gas Recovery and Treating Unit. The treated naphtha product is blended with the previously mentioned treated gas oil product to form the low sulfur fuel oil for sale.

The normal gaseous emissions from the single Naphtha Hydrogenation Unit are from the reactor feed heater stack. This stack is 2.5-ft. in

* Actual discard weight is 300 lbs. of a solid waste containing 20% As (ARCO, private communication).

diameter and 75-ft. high. The solid wastes from the Unit, in addition to the arsenic mentioned above, include as much as 75 tons per year of spent HDN catalyst.

Delayed Coker Unit: In the delayed coker unit the heavy bottom oil from each of the oil recovery fractionators is converted into lighter fractions and byproduct coke. The naphtha and gas produced are returned to the Gas Recovery Unit. The gas oil fraction is fed to the Gas Oil Hydrogenation Unit, and the condensed water is sent to the Foul Water Stripper Unit. Approximately 800 tons of coke (one-third of the bottoms oil) is expected to be produced daily. Pending establishment of a market for this coke it will be stored.

The bottoms oil is charged to the lower section of the coker unit fractionator, where it is contacted with hot vapors from the coking drums. The lighter ends (gas, water, naphtha, gas oil) are fractionated off overhead and distributed to other upgrading units (see above). The heavier oil remaining is raised to cracking temperature in a furnace heated by treated fuel gas or fuel oil, and fed to the coking drums, where coke and hot oil overhead vapors are produced. The latter are recycled to the coker fractionator.

The gaseous emissions from this Unit are those from the single 7.5-ft. diameter, 175-ft. high stack of the coker furnace. The only solid waste is the green coke previously mentioned.

Ammonia Separation Unit: The sour waters from the gas oil and naphtha hydrogenation units are sent to the Ammonia Separation Unit. After degassing to remove light hydrocarbons, H_2S is stripped off and sent to the Sulfur Recovery Unit. Ammonia is then removed in an ammonia stripper, compressed, and cooled to liquid ammonia for sale. Some 135 tons of anhydrous ammonia are estimated to be produced daily. The stripped water is returned to the hydrotreaters, or used to moisturize processed shale. There are no expected atmospheric emissions or other effluents from this Unit.

Foul Water Stripper Unit: The fouled waters from elsewhere in the upgrading plant are all sent to a Foul Water Stripper Unit distillation column, where NH_3 and H_2S are taken off overhead and transferred to the Sulfur Recovery Unit. Skimmed oil from the condenser is returned to the Retorting and Oil Recovery

Unit, as is also the stripped water itself. There are no atmosphere emissions or effluent wastes from this Unit.

Sulfur Recovery Units: As generally proposed by Colony, the acid gases from the Ammonia Separation, Foul Water Stripper, and Gas Recovery and Treating Units were fed to two conventional Claus-type Sulfur Recovery trains, to convert H_2S to liquid elemental sulfur. Some 194 tons of sulfur were expected to be produced daily. The tail gases from the Claus trains were fed to a standard Wellman-Lord sulfur recovery process, to further reduce their residual sulfur content to approximately 250 ppm by volume. The SO_2 produced (by oxidation) was recycled to the Claus trains.

There were no air emissions from the two Claus trains. It was expected, however, that some SO_2 would be emitted to the surrounding atmosphere with the 129,800 cubic ft. per minute of 125°F "clean" offgases discharged from the 8-ft. diameter, 210-ft. high tail gas stack of the Wellman-Lord unit.

Other wastes from the sulfur recovery units would have included approximately 75 tons per year of spent alumina catalyst from the Claus trains and neutralized acidic water from the Wellman-Lord Unit which would have been be discharged to the oily water sewer.

The more recently introduced Stretford process for H_2S removal has been costed instead of the Claus-Wellman-Lord system in the present study. Two Stretford absorbers are used, one just after the LPG-Gasoline recovery plant sponge absorber and the other just after the raw naptha stabilizer in the C_3/C overhead stream in the Gas Recovery and Treating Unit and hence must operate at almost 150 psi. These positions were chosen so the sponge oil absorber could finish scrubbing heavy oil, tar, and dust from the gas to be lead to the Stretford units. In this way a pure yellow sulfur can be produced rather than possibly a black oily product. Figure 3-10 shows the location of these high pressure Stretford absorbers while Figure 3-11 shows the arrangement of the amine absorbers they replace. A low pressure third Stretford absorber is placed instead of the Claus unit itself to remove H_2S from the gas from the ammonia separation unit.

By using the Stretford Process even lower sulfur emissions (as H_2S) are possible than are possible for the replaced system (as SO_2) using simpler

FIGURE 3-II. DEVELOPER'S GAS RECOVERY AND TREATING UNIT FOR
TOSCO II / COLONY PLANT.

equipment. Moreover there is no spent alumina catalyst or neutralized acid water to be handled.

3.2.7 Disposal of Solid Wastes

Some 20,220,000 tons* of solid plant wastes must be disposed of annually, or an average of 55,400 tons daily. Some 96.2% of this waste, 53,300 tons per day, is processed shale (and its dust). An additional 1,285 tons per day, is raw shale dust. The remaining solids are spent catalyst materials, sludges, arsenic-laden solids, and processed, plant sanitary wastes.

The processed shale from the Retorting and Oil Recovery Unit is moisturized with 14% of plant process water, and transported by closed conveyor to the disposal site. At the disposal site the waste shale will be transferred to trucks, and a processed shale embankment will be created in the form of a compacted landfill with a typical average density of 85 to 95 lbs/cu ft. A drainage system will be provided, together with a catchment basin for runoff. The drainage flumes will allow for a maximum rainfall from a 100-year 24-hour storm.

After final contours are established, contained salts in the top of the pile will be leached down into the pile, a 2-foot layer of topsoil added, and a revegetation program initiated. The latter will include the requisite chemical fertilization and maintenance over a period of years to produce a stable, self-sufficient soil cover of about 45% grasses, 40% shrubs, and 15% forbs.

3.2.8 Supporting Facilities

Water Supply and Treatment: The proposed commercial plant will require diversion of 12.5 cubic feet per second from the Colorado River at Grand Valley, Colorado. (This is equivalent to 9,000 acre-feet/yr., of which some 7,200 acre-feet/yr. may be obtained from Green Mountain Reservoir, through a Bureau

* Figures are given on an equivalent, moisture-free basis, exclusive of the water content of moisture processed shale and shale dust sludges.

of Reclamation contract. Groundwater wells may also provide a supplementary source of water.

At Grand Valley the river water will be settled to remove sediment, and treated with alum flocculant, plus a proprietary coagulant. The lime sludge so produced will be used in the pyrolysis unit for moisturizing processed shale.

The cooling water constantly circulating in the plant system will be reduced in temperature in an induced-draft multiple-cell cooling tower. All water used at the plant site will either be consumed or evaporated. There will, therefore, be no effluent discharge into Parachute Creek or any other surface stream or groundwater source.

Reservoirs: Two reservoirs will be constructed, one each in the Davis Gulch and Middle Ford (of Parachute Creek) drainages. The Davis Gulch Dam will provide a catchment basin below the processed shale pile, with a capacity of 310 acre-feet. The Middle Fork Dam will be a flood control and water storage reservoir located upstream from the mine bench.

Service Corridor and Product Pipeline: A 15 mile-long service corridor will be developed in Parachute Creek Valley, from the plant site to associated facilities located north of the town of Grand Valley, Colorado. The corridor will include three pipelines (water, ammonia, LPG), and a paved two-lane highway. Terminal storage facilities will be located at the Grand Valley end of the service corridor, and a railroad spur line constructed to these facilities.

A 16-inch diameter, 194 mile-long product fuel oil pipeline is proposed to be constructed from the plant site to a 120,000 barrel floating-roof storage tank to be located at Lisbon Valley Station in Utah. Subsequent connections to a major interstate pipeline at Aneth, Utah will be made.

Powerlines: It will be necessary to provide at least two 230,000 volt transmission lines to the mine and plant, each capable of providing 100% of the plant's ultimate electrical requirements. The lines may be built by Public Service of Colorado. These two lines will either be in completely separate corridors, or at least sufficiently separated in the same corridor so

that any reasonably credible emergency would not cause both to fail simultaneously. Final routes have not been selected as yet. Two powerlines are required because the shale oil refining process utilized by Colony will require a power supply with a very high degree of reliability. A sustained outage would have serious implications since the processing units contain high temperature and high pressure services utilizing catalysts and equipment which could be damaged by repeated loss of electric power. A one to three day time period would be required to restart these units in the event of a catastrophic power failure.

The projected electric load for construction and operation is as follows:

Construction, plant site and Grand Valley, temporary	3 megawatts
Preproduction Mining	10 megawatts
Water Pumping	10 megawatts
Grand Valley Facilities	5 megawatts
Plant Full-Scale Operation (Including Mine)	100 megawatts

Steam Generation: There will be four utility boilers at the plant site for steam generation. These will burn in-plant fuel gas. There are 52,000 cubic feet of total stack gases from four 55-ft. high boiler stacks, with an exit temperature of 400°F. The total fuel consumption of the boilers is 200 million Btu's per hour.

3.2.9 Plant Fuels

The plant and mine are expected to be self-sufficient in their fuel requirements once full production is attained, burning treated fuel gas, liquid butanes/butenes, and distillate fuel oils (usually gas oil) produced in the various plant units described earlier. The diesel fuel for the mine and processed shale disposal area equipment, 300-400 bpd, will come from the Gas Oil Hydrogenation Unit, as previously noted.

A possible fuel schedule for the entire plant complex at full-scale operation is given below, based upon the following average heating values: fuel gas 22,200 Btu/lb; fuel oil 19,500 Btu/lb; C₄ liquids 21,200 Btu/lb.

Table 3-8. TOSCO/Colony Plant Fuel

<u>Units</u>	<u>Firing Duty, 10⁶ Btu/hr.</u>		
	<u>Fuel Gas</u>	<u>Fuel Oil</u>	<u>C₄ Liquids</u>
Pyrolysis and Oil Recovery	708	755	528
Hydrogen	632	-	-
Gas Oil Hydrogenation	-	-	87
Naphtha Hydrogenation	10	-	-
Sulfur Recovery	10	-	-
Delayed Coker	88	-	-
Utility Boilers	-	93	-
Totals	1,448	848	615

It is necessary to emphasize that the individual allocation of fuels given above is subject to considerable alteration, in response to changing fuel prices, market demands, and variations in-plant operating conditions. It can be expected, however, that the total fuel heating requirements for the entire plant will not vary substantially from that shown.

3.2.10 Material Balances

Table 3-9A is the material balance around the TOSCO II/Colony plant as shown in Figures 3-9 and 3-10 with the variation of Stretford instead of amine H₂S scrubbers in the less strict pollution control variation. Table 3-9B is the material balance for the more strict variation in which venturi scrubbers have been replaced by bag filters. Details to these tables are given in Section 6.0.

3.2.11 Energy Balance

Table 3-10 is the the energy balance around the TOSCO II/ Colony plant.

3.3 MODIFIED IN SITU PLANT

3.3.1 Basic Parameters Assumed for Modified In Situ Plant

For the most part, the Modified In Situ process proposed by Occidental is used for this case. Occidental's pilot plant experience at Logan Wash has provided data which is useful in analyzing this process, and

Table 3-9A
MATERIAL BALANCE AROUND TOSCO II PLANT PRODUCING 47,000 BPSD
UPGRADED SHALE OIL FROM 35 GAL/TON OIL SHALE. FLOWS IN 10³ LB/HR

<u>Stream</u>	<u>Total</u>	<u>H₂</u>	<u>Moisture</u>	<u>H₂O equiv</u>	<u>C</u>
<u>IN</u>					
Raw shale, 35 gal/ton	5465	116	146	1189	902
Water to pyrolysis	1190	-	1190	1190	-
Steam to pyrolysis	195	-	195	195	-
Steam to H ₂ production	246	-	246	246	-
Oxygen for H ₂ combustion, pyrolysis	125	-	5	5	-
Water to coker	<u>50</u>	<u>-</u>	<u>50</u>	<u>50</u>	<u>-</u>
TOTAL IN	7271	116	1832	2875	902
<u>OUT</u>					
Spent shale & sludge (dry)	4514	6	-	57	158
Water evaporated on shale	250	-	250	250	-
Water leaving with spent shale and sludges	632	-	632	632	-
Water in scrubber stack gases	540	-	540	540	-
Foul water, pyrolysis	213	-	213	213	-
Foul water, coker, gas treating	36	-	36	36	-
Compressor condensate	23	-	23	23	-
Fuel burned	138	22	-	201	110
Coke (product)	67	2	5	27	61
NH ₃ + S	27	2	-	18	-
LPG	32	6	-	52	26
Product oil	552	76	-	686	476
Diesel fuel to mine	6	1	-	7	5
CO ₂ released	<u>241</u>	<u>-</u>	<u>133</u>	<u>133</u>	<u>66</u>
TOTAL OUT	7271	115	1832	2875	902

Table 3-9B
MATERIAL BALANCE FOR TOSCO II PLANT IN WHICH VENTURI SCRUBBERS
HAVE BEEN REPLACED BY BAG FILTERS. OIL PRODUCTION AS FOR TABLE 3-9A
FLOWS IN 10^3 LB/HR^(a)

Stream	Total	H	Moisture	H ₂ O	C
<u>IN</u>					
Raw shale, 35 gal/ton	5465	116	146	1189	902
Water to pyrolysis ^(b)	853	-	853	853	-
Steam to pyrolysis	195	-	195	195	-
Steam to H ₂ production	246	-	246	246	-
Oxygen for H ₂ combustion, pyrolysis	125	-	5	5	-
Water to coker	50	-	50	50	-
TOTAL IN	6934	116	1495	2538	902
<u>OUT</u>					
Spent shale & sludge (dry) ^(c)	4514	6	-	57	158
Water evaporated on shale ^(c)	235	-	235	235	-
Water leaving with spent shale	632	-	632	632	-
Water in scrubber stack gases ^(d)	218	-	218	218	-
Foul water, pyrolysis	213	-	213	213	-
Foul water, coker, gas treating	36	-	36	36	-
Foul water, compressor condensate	23	-	23	23	-
Fuel burned	138	22	-	201	110
Coke (product)	67	2	5	27	61
NH ₃ + S	27	2	-	18	-
LPG	32	6	-	52	26
Product oil	552	76	-	686	476
Diesel fuel to mine	6	1	-	7	5
CO ₂ released	241	-	133	133	66
TOTAL OUT	6934	115	1495	2538	902

(a) Flow rates as in Table 6-17 notes, unless otherwise noted below.

(b) Water to pyrolysis. See Tables 6-20 and 6-21.

(c) Water evaporated on shale. The 500 gpm used in Table 6-17 less the 30 gpm used in the venturi scrubber, gives 470 gpm or 235×10^3 lb/hr.

(d) Water in scrubber stack gases. (Shale preheat and ball elutriator stacks). See Table 6-20. In Figure 6-6 the ball elutriator stack gas is shown condensed with the moisture stack gas.

Table 3-10
ENERGY BALANCE FOR TOSCO II PLANT PRODUCING
47,000 BPSD UPGRADED SHALE OIL
77°F DATUM TEMPERATURE

Stream	Flow 10 ³ lb/hr	T °F	P psia	Sp. Ht. Btu/lb. °F	HHV Btu/lb	Heat Content 10 ⁹ Btu/hr
<u>IN</u>						
Raw shale and moisture ^(a)	5465	50		0.235	3208	17.77
Electric energy	68 MW					<u>0.23</u>
TOTAL IN ^(c)						<u>18.00</u>
<u>OUT</u>						
Spent shale and moisture ^(d)	5146	200		0.32	344	1.95*
Coke ^(e)	67	120			13850	0.93
Oil ^(f)	552	120		0.62	19500	10.78
LPG ^(e)	32				21560	0.69
Diesel fuel to mine ^(g)	6				19500	0.12*
Fuel burned ^(h)	149					
Cooling water ⁽ⁱ⁾	761				1400	1.06*
Water evaporated on shale	250				1000	0.25*
NH ₃ ^(f)	11				9668	0.11
S ^(j)	16				3963	0.06
Losses ^(k)						<u>2.05*</u>
TOTAL OUT						<u>18.00</u>

- (a) Raw shale to retort. Specific heat and HHV data for 35 gal/ton shale from Cameron Engineers Handbook.
- (b) Electric power. Total power to plant given in Reference 28 as 95 MW. About 9.8 kWh/ton shale required for mining and crushing, leaving: $(98 - 9.8 \times 66/24) = 68$ MW to process.
- (c) Steam supplied to retorting/upgrading is raised using internally generated energy. Other input streams have negligible thermal contribution.
- (d) Spent shale. Specific heat and HHV data from Cameron Handbook, adjusted to 14% moisture content.

Notes--Table 3-10 (Concluded)

- (e) HHV from Ref. 6.
- (f) Product oil; Ref. 1 gives HHV = 19,500 Btu/lb. Specific heat from Cameron Engineers' Synthetic Fuels Data Handbook.
- (g) Diesel fuel. Assumed to be similar to crude shale oil.
- (h) Fuel burned. The energy consumed or lost as a result of fuel burning is distributed as follows:
 - a) in the products dealt with separately.
 - b) removed by evaporating cooling water, see (i).
 - c) dissipated in the stack gases or by air cooling--included in losses.
- (i) Cooling water. Ref. 3 gives 1522 gpm or 761×10^3 lb/hr evaporated. Using 1400 Btu dissipated/lb water evaporated gives 1.06×10^9 Btu/hr. This is 19.5% of the unrecovered heat, which is in fair agreement with the 18% given in Ref. 6.
- (j) HHV from Ref. 5.
- (k) Losses. By difference.

recent data on site C-b helps to further define the process. However, useful process information has also been taken from Project Rio Blanco for Tract C-a. Overall, much less information is available on these Modified In Situ (MIS) processes than is available on the Paraho and TOSCO II aboveground processes. The location for the MIS process may be either Tract C-a or C-b on the top of the Piceance Creek Basin. Both tracts are sufficiently similar as to be interchangeable for the purposes of this study. The Occidental MIS mining technique with rubblizing of vertical retorts is assumed. The water source will be the upper aquifer in the Piceance Creek Basin collected as mine water by the processing activities. A large excess of mine water over process needs now appears certain and is considered.

Rubblized shale of 25 gal/ton average grade will be retorted in place with steam and air injection producing 57,000 bpd of crude shale oil for a yield of 57.2% of Fischer assay. No upgrading on site will occur. Low Btu gas produced by the retorting will be consumed to generate steam and to produce electricity by gas turbine power generation. Ammonia and sulfur will be recovered as byproducts. Some 20% of the raw shale will be mined to produce a void space before rubblizing, and this raw shale will be disposed of aboveground. The spent shale will be left in the burned out retorts without further treatment. There will be significant quantities of excess mine water for surface disposal, reinjection into the aquifer or evaporation by ponding.

The Occidental Modified In Situ system for Tract C-b would use rubblized "retorts" 200 ft. x 200 ft. x 310 ft. high. Eight retorts per cluster, 32 clusters per panel are planned. The C-b tract will bear approximately 15 panels equivalent with 200 ft. barrier pillars surrounding each panel and the property boundaries. At 57,000 bpd production rate, a panel would last around four years. Thus Tract C-b would last roughly 60 years.

3.3.2 Description of Occidental Modified In Situ Retorting Process [2]

The Occidental Petroleum Corporation's involvement in oil shale technology is relatively recent. In 1972, Garrett Research and Development Company (now Occidental Research and Development), a subsidiary of Occidental Petroleum Corporation, announced plans for the field testing of a Modified In Situ shale oil recovery scheme.

Work began in the summer of 1972 on the D.A. Shale property at the head of Logan Wash, outside of Debeque, Colorado. In the ensuing months, several research retorts, each 30-ft. on a side and 72-ft. high were prepared and ignited.

At the end of 1974, the project was transferred to Occidental Oil Shale, Inc., a subsidiary of the Occidental Oil & Gas Production Division. Development of a commercial size retort in the commercial mine, located in a canyon off the north side of Logan Wash was concurrently begun. The commercial mine is being developed at the new location because there is insufficient room at the head of Logan Wash (the research mine location) to permit large-scale mining, and also because the research mine is located just below the Mahogany Ledge and too high for the construction of commercial size retort columns.

The first commercial size retort (Retort No. 4), with 120 ft. by 120 ft. cross section and 250 ft. height containing 15 gpt rubblized shale, was ignited from the top on December 10, 1975. The ignition was successful, with oil being recovered from the sump at the bottom of the column and sustained combustion and temperature control achieved by recycle of a portion of the retort gas [2]. A total of 27,500 barrels of oil was produced from Retort No. 4. Retort No. 5 was unsuccessful in that voids distribution was poor and bad channeling of the fire front developed, causing poor oil yield. Retort No. 6, nearly through burning, is more successful.

Occidental has spent over \$45 million in the last five years in the development of the Modified In Situ process. Occidental hopes to attract sufficient support for the construction of a 5,000 bpd demonstration mine and retort.

The D.A. Shale property contains relatively low grade shale (15 gpt) and may be marginal for commercial operation. In November 1976, upon the withdrawal of the Shell Oil Company from the C-b Oil Shale Project, Occidental and Ashland announced that a "letter of agreement" arrangement had been made for Occidental to join Ashland in the C-b Oil Shale Project. The C-b Tract contains much higher grade shale and will lead to major changes in Occidental's development plans. Occidental was to be the operating group for the Tract C-b consortium. Modifications to the C-b Detailed Development Plan incorporating Modified In Situ retorting, have been submitted to the Area Oil Shale Supervisor. Since that time, Ashland has withdrawn from the project.

3.3.3 MIS Retort Preparation

The Occidental process involves three basic steps. The first step is the mining out of approximately 20 to 25% of the oil shale deposits (preferably lower grade shale or barren rock if aboveground retorting is not also used) at the upper and/or lower level of the shale layer, or in other geometry. This is followed by drilling longholes from the mined-out space into the shale, loading with an ammonium nitrate-fuel oil (ANFO) explosive, and detonating it with appropriate time delays so that the broken shale will fill both the volume of the mined out space and the volume of the shale column before blasting. Connections are finally made to both the top and bottom of the prepared retort and retorting is conducted. Figure 3-12 shows the finished retort and connections.

The results from the Occidental experiments indicated a retort burn rate of 0.54 in./hr., thus the production period of a 120 ft. x 120 ft. x 250 ft. high retort is 232 days, "and the production rate of crude shale oil from a commercial size retort is 218.5 bpd. . . 229 retorts would be required to operate simultaneously to produce 50,000 bpd of crude shale oil if the average Fischer assay of the shale zone is 15 gpt. For a shale zone with an average Fischer assay of 25 gpt., a minimum of 149 retorts would still be required if the production goal 50,000 bpd of crude shale oil were to be realized[2]. . . ."

The Modified In Situ process proposed for the C-b Tract will contain retorts with a 200 ft. x 200 ft. cross section and 310 ft. height. The retorting configuration will consist of 8 retorts in a cluster, and 32 clusters in a panel. At the proposed production rate 57,000 barrels per day, a panel will last for about 4 years. The C-b property encompasses about 11 complete panels, plus some additional groups of clusters (equal to about 4 additional panels) filling the remaining areas. The panels are bordered by 200-ft. barrier pillars along and within property boundaries.[2]

For the Modified In Situ process proposed for the C-b Tract, steam will be piped to the retorts along with combustion air to promote a water gas reaction. Steam addition may increase retorting efficiency to approximately 78%. Only 40 retorts would be required to operate simultaneously to

FIGURE 3-12. FLOW DIAGRAM OF MIS PLANT SELECTED

produce 57,000 bpd of crude shale oil, as the result of higher retorting efficiency, larger retort cross sectional area, and the higher grade of shale found in the C-b Tract. "In the construction of the commercial size retort, Occidental plans mining at two levels. The upper mining level will be a complete heading at or near the top of the retort, and will serve as a base from which vertical longholes will be drilled for the loading of explosives." [2]

3.3.4 MIS Retort Operation

Retorting is initiated by heating the top of the rubblized shale column with a flame from compressed air and oil, propane, or natural gas. After several hours the flame is extinguished, and the compressed air flow is maintained, utilizing the carbon in the retorted shale at the top to sustain combustion. The hot gases from the combustion zone move downwards to pyrolyze the shale just below that zone, producing gases, water vapor, and shale oil mist which collects in trenches at the bottom of the rubblized column. Oil production precedes the advancing combustion front by 30 to 40 ft. The shale oil and some byproduct water are collected in a sump and pumped to storage. The offgas consists primarily of gases from shale pyrolysis, carbon dioxide and water vapor from the combustion of carbonaceous residue, carbon dioxide from the decomposition of inorganic carbonate (primarily dolomite and calcite), and hydrogen from the water gas reaction as well as N_2 . Part of this offgas could be recirculated to control the oxygen level of the incoming air and thereby control retorting temperature. However, the design plans for Tract C-b call for steam injection into the retort to promote the water gas reaction which also cools the combustion zone. The offgas from retorts with steam injection has a heating value of approximately 100 Btu/scf. This offgas, after being discharged through gas blowers, is cooled and passed through a Stretford-type hydrogen sulfide removal system, and part through a thermal oxidizer and steam generation system, where steam needed for the retorting operation and perhaps other plant uses may be generated by the combustion of the treated offgas. In the Stretford process, hydrogen sulfide present in the retort offgas is absorbed in an alkaline liquid and directly oxidized to sulfur by dissolved vanadium compounds and an oxidation catalyst. The hydrogen sulfide level in the treated offgas is expected to be reduced to 15 ppmv. [2]

"On the surface, free water is separated from the product oil by gravity in a primary separator. The wet oil is then pumped to standard electrostatic oil-field type heater/treaters for secondary water separation. Essentially water-free oil obtained after this second step may be stabilized and sent to product storage." [2] With proper storage stabilization (C_3 removal) may not be required.

"The crude shale oil produced from the Occidental process at the D.A. Shale property reportedly has a specific gravity of 0.904 (API gravity of 25°), a pour point of 70°F , a sulfur content of 0.71 weight percent and a nitrogen content of 1.50 weight percent. The crude shale oil is also reportedly free of solids and may be usable directly as boiler fuel. Occidental has indicated that tests conducted with the crude shale oil show that its direct use as boiler fuel would meet the current NO_x standards. This implies that less than 20% of the fuel nitrogen contained in the crude shale oil is converted to NO_x during the combustion process. For most other fuels, the conversion of chemically bound nitrogen to NO_x under normal boiler operating conditions is significantly higher and amounts to around 50%. At the present time, the information available is insufficient to assess the validity of the NO_x tests with Occidental in situ crude shale oil" [2].

3.3.5 MIS Plant Flow Diagrams of Variation Analyzed

Figure 3-12 is a flow diagram of the MIS plant selected for study.

3.3.6 MIS Plant Material and Energy Balances

Table 3-11 is the material balance around the the Modified In Situ retort producing 57,000 bbl per day crude shale oil from an assumed 164,000 tons per day of shale of approximately 25 gal/ton assay oil shale. At this assay a yield of 57.2% of assay is calculated. At 60% yield the assay should be 23.8 gal/ton and so forth. The Fischer assay of the section of formation selected is dependent on a number of factors and will vary from tract to tract. These yields are conservative and yields above 65% may eventually be obtained. Further details of the retort material balance are to be found in Section 6.0.

Table 3-12 is the energy balance around the MIS plant. The low Btu gas produced is largely burned in gas turbines for electric power.

TABLE 3-11
MATERIAL BALANCE AROUND MIS PRODUCING 57,000 BBL/DAY
CRUDE SHALE OIL FROM ~ 25 GAL/TON OIL SHALE^(a)
FLOWS IN 10³ LB/HR

Stream	Total	H	Moisture	H ₂ O Equiv	C
<u>IN</u>					
Raw shale retorted ^(b)	13,720	200	274	2,077	1,577
Steam	848	-	848	848	-
Air, + 1% moisture	3,109	-	31	31	-
Carbonates decomposed (75%)		-	-	-	562
Seepage					
TOTAL IN	<u>17,677</u>	<u>200</u>	<u>1,153</u>	<u>2,956</u>	<u>2,139</u>
<u>OUT</u>					
Spent shale	10,091	47	-	423	728
Water leaving with oil	386	-	386	386	-
Water vapor leaving with gas	917	-	917	917	-
Retort gas (dry, NH ₃ & H ₂ S free)	5,464	43	-	388	745
Product oil (1.5% moisture)	754	88	11	805	636
Sulfur	12	-	-	-	-
Total NH ₃	23	4		37	
Coke	<u>30</u>	<u>-</u>	<u>-</u>	<u>-</u>	<u>30</u>
TOTAL OUT	<u>17,677</u>	<u>182</u>	<u>1,314</u>	<u>2,956</u>	<u>2,139</u>

NOTES:

- (a) Scheme based essentially on the Occidental Oil Shale scheme as presented in Figure III-J, Reference [30].
- (b) Shale retorted. Reference [1] shows 41,134 t/d shale mined. If this is 20% of total shale, shale retorted = $4 \times 41,134$ t/d or $13,720 \times 10^3$ lb/hr. For 25 gal/ton shale, C = 11.5%, H = 1.46%. Following Reference [31] free and combined water taken as 2%.

Table 3-12
ENERGY BALANCE AROUND MIS RETORTS PRODUCING 57,000 BBL/DAY
CRUDE SHALE OIL FROM 25 GAL/TON OIL SHALE
77°F DATUM TEMPERATURE

Stream	Flow 10 ³ lb/hr	T °F	P psia	Sp. Ht. Btu/lb. °F	HHV Btu/lb	Heat Content 10 ⁹ Btu/hr
<u>IN</u>						
Shale retorted ^(a)	13,720	50		0.21	2,241	30.67
Steam ^(b)	848	500	463		(1,190)	1.01
Air	3,109	50		0.24		-0.02
Electric power ^(c)	50 MW					<u>0.17</u>
TOTAL IN						<u>31.82</u>
<u>OUT</u>						
Spent shale	10,913	200	-	0.22	600	6.82
Retort gas (dry) ^(d)	5,414	140		0.26	841	4.82
Product oil ^(e)	754	150		0.62	18,500	14.00
Water vapor with gas ^(f)	917				(1,081)	0.99
Sulfur	12				3,963	0.05
NH ₃	23				9,668	0.21
Coke	30				13,850	0.42
Carbonate conversion ^(g)	4,683				(500)	2.34
Losses						<u>2.17</u>
TOTAL OUT						<u>31.82</u>

NOTES:

- (a) Shale retorted. Sp. Ht. and HHV from Cameron Engineers Handbook for 25 gal/ton shale.
- (b) Steam. 450 psig superheated steam. Enthalpy relative to water at 77°F = 1,190 Btu/lb.
- (c) Electric power. Process blowers (induced draft in retort) Reference [30] p. III-39 gives 50.1 MW, or, at 3,415 Btu/kWhr is equivalent to 0.17×10^9 Btu/hr.
- (d) Product gas. HHV based on Table 6-35 treated gas composition, corrected for NH₃ and H₂S removed.
- (e) Product oil. From Reference [2]. HHV 18,500 Btu/lb. Sp. Ht. = 0.62.
- (f) Water vapor with gas. Enthalpy above 77°F - 1,081 Btu/lb.
- (g) Carbonate conversion. 562 lb C released (from Table 6-33) is equivalent to 4,683 lb CaCO₃ decomposed. Heat consumed ~ 550 Btu/lb CaCO₃.

Table 3-13 is the composition of the retort gas assumed in the above and subsequent calculations.

Table 3-14 is an estimation of the electric power exported from the Modified In Situ plant.

Table 3-15 is the overall water balance for the selected plant.

3.4 MODIFIED IN SITU PLUS LURGI-RUHRGAS ABOVEGROUND RETORTING COMBINATION

This last case is the most recently proposed Rio Blanco plan. For the modified in situ portion of the processing, the basic parameters and data discussed in 3.3 will be followed. The retorting process to be used on the mined shale will be the Lurgi-Ruhrgas process. Finely divided noncarbonaceous spent shale is produced, a part of which can be slurried and pumped back into the spent in situ retort. The slurry then sets as a strong impermeable grout. Part of the spent shale will be disposed of aboveground as not enough void space in the spent in situ retort exists to accept the total volume.

3.4.1 Basic Parameters Assumed

Rio Blanco may use either the TOSCO II process or the Lurgi-Ruhrgas process for retorting the mined shale. We have chosen to study the latter because of greater accessibility of data and the known cementing activity of the Lurgi-Ruhrgas spent shale.

Production volume for the process will be 81,000 barrels per calendar day (57,000 Bbls from the MIS Process and 24,000 Bbls from the Lurgi-Ruhrgas Process) of crude shale oil from 25 gal/ton average grade shale. The water source will be the upper aquifer in the Piceance Creek Basin, collected as mine water. There will be moderate to no excess mine water over process needs. Location will be either tract C-a or C-b. All low Btu gas produced by retorting will be consumed on site. Electricity will be provided by gas turbine power generation. Ammonia and sulfur will be recovered as byproducts.

TABLE 3-13
COMPOSITION OF MIS RETORT GAS^(a)

Constit	Mwt	Raw Retort Gas		Cooled & Scrubbed Gas	
		Mass %	Volume %	Mass %	Volume %
N ₂	28	43.930	44.699	51.244	55.338
CO ₂	44	38.153	24.704	40.471	27.865 ^(d)
CO	28	0.759	0.772	0.885	0.958
H ₂	2	0.298	4.246	0.348	5.271
CH ₄	16	0.650	1.158	0.758	1.763
C ₂ H ₆	29	0.315	0.309	0.267	0.383
C ₃ H ₈	43	0.232	0.154	0.271	0.191
C ₄ H ₁₀	60	0.618	0.309	0.721	0.364
O ₂	32	0.086	0.077	0.100	0.095
H ₂ S	34	0.184	0.154	0.201	0.179
NH ₃	17	0.369	0.618	-	-
H ₂ O	18	14.405	22.800	4.633	7.594
TOTAL		100.	100.	100.	100.
H ^(c) (dry basis)		0.835		0.796	
C (dry basis)		14.245		13.76	
Mwt		28.49		30.29	
Gas rate, wet, 10 ³ lb/hr:		6,366		5,482 ^(d)	
dry, 10 ³ lb/hr:		5,449		5,229	

(a) Values presented here are as used in calculations and are not intended to represent the accuracy of the composition.

(b) Based on composition on dry basis provided by DRI.

(c) Not including hydrogen in the water vapor and H₂S.

(d) About 46×10^3 lb/hr CO₂ are removed in gas treatment.

TABLE 3-14
ESTIMATION OF ELECTRIC POWER EXPORTED FROM THE MIS PLANT

Raw retort gas rate (dry)	$5,414 \times 10^3$ lb/hr
Treated gas rate (dry) ^(a)	$5,218 \times 10^3$ lb/hr
Calculated HHV, treated gas, dry basis (65 Btu/scf)	791 Btu/lb
Gross heat available	4.13×10^9 Btu/hr
Heat required for steam raising	
848×10^3 lb/hr steam for retort, 75% efficiency ^(b)	1.36×10^9 Btu/hr
100×10^3 lb/hr steam at 85% efficiency ^(c)	0.12×10^9 Btu/hr
Total heat for steam raising	1.48×10^9 Btu/hr
Net heat available for power generation	2.65×10^9 Btu/hr
Electricity generated, open gas cycle ^(d)	203 MW
Electricity consumed ^(e) :	
Retorting, gas treating	71 MW
Mining	35 MW
Total consumed	106 MW
Net electricity available for export	97 MW

-
- (a) Table 3-13, scrubbed and cooled gas rate, less than H₂S removed in Stretford.
- (b) Raised in "thermal sludge" unit.
- (c) Steam for other process uses, Reference [30], Figure III-J.
- (d) Raised in gas turbine cycle at 13,000 Btu/kWh. The combined cycle scheme proposed by the developers was not followed in view of the low heating value of the gas, and to keep the analysis comparable to that for the Paraho plant.
- (e) As given on page 3-39, Reference [30]. It has been assumed that the gas is not compressed prior to the Stretford gas treatment unit.

TABLE 3-15
WATER BALANCE WITH MINE WATER INPUT FOR MODIFIED IN SITU RETORTING TO
PRODUCE 57,000 BBLS/DAY OF SHALE OIL AND 132 MW OF ELECTRIC POWER^(a)

<u>IN</u> (Makeup to plant, or produced in retorting)	<u>GPM</u>
MIS Retort Water	
Water condensed from gas	1,340
Water separated from oil	772
Mine Water	7,656 ^(c)
Plant and Mine Area Runoff	125 ^(d)
Disposal Area Runoff	-0- ^(e)
	<u>9,893</u>
 <u>OUT</u> (Consumed or discharged)	
Steam into Retort	1,696 ^(f)
Cooling Tower Evaporation and Drift	1,634 ^(g)
Shale Disposal, Dust Control and Revegetation	685 ^(h)
Potable/Sanitary and Service	31 ⁽ⁱ⁾
Mine Uses (including Dust Control)	210 ^(j)
Losses, Steam System and Condensate Treatment	83 ^(k)
Excess Mine Drainage Water to Disposal	5,554 ^(l)
	<u>9,893</u>

(a) Generated by gas turbine system.

(b) Gas cooled to 100°F prior to Stretford process; no compression.

(c) Quantity required for balance in scheme shown in Figure 2.

(d) Quantity estimated.

(e) Assumed negligible.

(f) Assumed to be the same as in Figure III-J, Reference [30].

(g) See Table 6-42.

(h) Estimated at 100 lb. water/10³ lb spent shale, which is equal to 10% by weight of 40,000 tons/day of disposed shale (cf. Table 7.11, Reference [32]).

(i) Consumption of 31 gpm for 1,600 people (see p. 297, Reference [32]).

(j) Mine use at 32 lb water/10³ lb mined out shale (see p. 195, Reference [32]).

(k) See Figure 3.

(l) Based on minimum and maximum values of mine drainage water given in references found. Mine water for disposal for disposal may vary from about 1,500 gpm to 8,500 gpm.

3.4.2 Crushing

The crushing operations for the Lurgi-Ruhrgas process will be similar to those for the other aboveground retorting processes. Both primary and secondary crushing of the shale will be performed aboveground. Primary crushing will reduce the size of the shale to minus 8-inches. Secondary and possibly tertiary crushing will further reduce the shale size to minus 0.25 inches. The production rate required for the proposed plant is 41,000 tons/day of shale.

3.4.3 Lurgi-Ruhrgas Plant Flow Diagram

In the Lurgi-Ruhrgas process[33, 34], shown in Figure 3-13 in a variation largely proposed by Lurgi-Ruhrgas, crushed oil shale of minus 0.25 inch size is fed through a feed hopper to a double screw mixer, where it is thoroughly mixed with a large quantity (6 to 8 times greater) of hot (1200°F) circulating shale residue. The fresh shale feed is heated to 950°F within a very short time, resulting in the evolution of gas, shale oil vapor and water vapor. The circulating heat carrier and the partially retorted fresh shale feed are then dropped from the screw mixer into the surge hopper, where residual oil components are distilled off.

The mixture of heat carrier and retorted shale residue is passed to the lower section of the lift pipe, where combustion air (500°F) is introduced, raising the mixture pneumatically to the collecting bin. The carbon contained in the retorted shale residue is burned during the transport process. The heat carrier is then separated from the flue gases in the collecting bin. The fines are carried through with the flue gas stream. The coarse grained shale residue accumulates in the lower section of the collecting bin and flows continuously to the mixer. Some provision may be required for partial removal of shale residue if the dust removal is not sufficient to prevent solids accumulation. Some shales may disintegrate to the extent that an additional coarse-grained heat carrier, such as sand or low-grade shale should be used. The combustion air supplied to the lift pipe is preheated by countercurrent heat exchange with the flue gas stream.

The volatile gas product stream from the retorting of the oil shale is virtually the same composition as TOSCO II retort gas. It is passed

FIGURE 3-13. LURGI-RUHRGAS OIL SHALE RETORTING PROCESS

through two series-connected cyclones. The dust separated in these cyclones is returned to the recycle system. The gas stream then enters a sequence of three scrubbing coolers. The first scrubbing cooler is designed to operate at a higher temperature to remove the residual dust from the gas stream by washing and circulating condensed heavy oil. In the next scrubbing cooler, major condensation of the oil takes place at a temperature above the dew point of water to produce a dust-free heavy oil. Final cooling of the condensate occurs in the last scrubbing cooler, after the condensate has been cooled in air and water coolers. The condensate is separated into middle oil and gas liquor in an oil/water separator. Finally, the gas is scrubbed with sponge oil for the recovery of C_3+ . For removal of H_2S in the gas a Stretford Process has been chosen for the present study although other methods are available. It will be a separate plant from the Stretford plant used on the MIS gas. The gas from the Lurgi process is a high Btu gas and it is not desirable to mix it with the low Btu MIS gas.

The flue gas stream evolved in the lift pipe is dedusted in a cyclone after leaving the collecting bin, and then routed through a heat exchanger (for the preheating of combustion air), a waste heat boiler, a feedwater preheater, another cyclone, and, in the Lurgi-Ruhrgas proposed variation, a humidifier and a "cold-side" electrostatic precipitator before discharge to the atmosphere. In the humidifier, the flue gas stream is cooled from approximately 600°F to 300°F by water injection, and a portion of the shale dust contained in the flue gas is separated and discharged to a chain conveyor. The residual dust is removed from the flue gas stream in the electrostatic precipitator and discharged onto another chain conveyor. The two chain conveyors then carry the fine dust residue to the residue moistening screw mixer, where flue gas dust from the cyclone, heavy oil dust from the heavy oil dust removal step, and moistening water (along with the gas liquor) are added. The spent shale residue, requiring surface disposal, has a moisture content of approximately 10% water and a temperature of 150°F.

Dust is removed from the heavy oil obtained in the first scrubbing cooler by sedimentation and centrifuging. This occurs after the addition of naphtha which reduces the viscosity of the heavy oil. The final products obtained include a dust-free heavy oil, a middle oil, a Lurgi naphtha, and a distillation gas.

The "optional" moisturizer has been omitted from the variation of the Lurgi-Ruhrgas process selected for this study. This moisturizer cools the flue gas and moisturizes the shale ash. The moist shale ash has a lower resistivity and favors a better electrostatic precipitation. Without this moisturizer the electrostatic precipitator must operate as a "hot-side" unit at approximately 600°F. At 600°F the ash resistivity is low enough to give good precipitation although at intermediate temperatures it is not. Less water is used in cooling the ash in the dust cooler from 600°F to 150°F when compared to cooling the flue gas from 600°F to 300°F plus the ash from 600°F to 150°F. An ash cooler that handles larger amounts of cooling water spray and generated water vapor is required and a sizeable venturi wet scrubber or baghouse is needed to capture the dust in the water vapor stream. An alternate approach involves directing portions of the 200°F vapor stream from the ash cooler to the electrostatic precipitator inlet. This stream will cool and moisturize the precipitator inlet gas. The combined effect may slightly increase the ash resistivity without affecting the precipitator performance.

At the temperature of the hot precipitator, hydration of MgO in the ash will occur. The nature of the ash product affects grout preparation. Hydration of MgO will also occur in the "cold-side" precipitator of the original Lurgi-Ruhrgas design although in a different manner.

3.4.4 Slurry Backfilling and Grouting

The spent shale ash from the Lurgi-Ruhrgas process will be slurried with water and suitable additives for control of the slurry flow characteristics and cement quality. It will be pumped back into the voids of the previously retorted and cooled in situ caverns. Such a procedure would have the advantages of (a) providing a convenient method for disposal of approximately 80% of the shale ash produced from the 20 to 25% of the shale mined for the in situ operation, (b) rendering the retorted in situ shale essentially impervious to groundwater flow, (c) minimizing subsidence and (d) maximizing the oil recovery by allowing minimum support pillars.

The slurry backfilling system includes a series of shale ash silos, a conveyor and shale ash feed system, additive silos and feeding systems, a slurry mixing tank, a pumping and pipeline system, a 10-day capacity water holding pond, and necessary water flushing systems to clean out the mixing

tank, pipeline, and pumps when the system is shut down. The silos, holdup tanks, etc., are designed for minimum holdup capacity on the assumption that the slurry backfilling system will operate only when the Lurgi-Ruhrgas process is operating. Retorted shale ash from the Lurgi-Ruhrgas process is difficult to store in open piles. It absorbs moisture from air and rain and forms cements. The raw shale feed may be stored indefinitely.

3.4.5 Material Balance Around MIS Retorting Plant with Lurgi-Ruhrgas Aboveground Retorting and Slurry Backfill

Table 3-16 shows this overall material balance. Table 3-17 shows the material balance for Lurgi-Ruhrgas retorting only. An MIS retorting operation producing 57,000 bpd of shale oil from 25 gal/ton shale requires mining of approximately 570,000 ft³/day of raw shale (assuming a 20% porosity in the MIS retort). Assuming the retorted shale in the MIS retort shrinks approximately 10% in volume, the total volume of voids available for slurry backfill will be approximately 855,000 ft³/day. The 570,000 ft³/day mined shale will weigh 41,000 tons/day and will generate approximately 30,750 tons/day of shale ash. The slurry backfill may use 25,500 tons/day, leaving 5,250 tons/day for disposal in surface fills.

3.4.6 Energy Balance Around MIS Retorting Plant with Lurgi-Ruhrgas Aboveground Retorting and Slurry Backfill

Table 3-18 shows the overall energy balance around the plant.

3.5 OVERALL RESULTS AND DISCUSSION

3.5.1 Introduction

A number of variations of each of the four plants being considered here are possible. Generally these variations are due to alternate means of retort gas treatment. (1) Does one remove most of the gasoline and LPGs they contain through gas compression, cooling and absorption in circulating oil in a rather conventional gasoline recovery plant? (2) Does one remove H₂S by a Stretford unit before utilization of the gas by a gas turbine or steam boiler but instead remove the resulting SO₂ by a stack gas scrubber? (3) Does one remove and sell all the NH₃ using water scrubbing and steam stripping or use a

Table 3-16. Overall MIS/Lurgi-Ruhrgas Material Balance
(Excluding possible LPG and gasoline plant on Lurgi-Ruhrgas
retort gas, mine water source, dust control water, cooling
tower water loss, and misc. water flows)

<u>Stream</u>	<u>MIS Retort Section</u>	<u>Lurgi-Ruhrgas Retort Section</u>	<u>Overall Plant</u>
<u>IN</u>			
Raw shale retorted, lb/hr	13,720,000	3,410,000	17,130,000
Air, +1% moisture	3,109,000	1,411,400	4,520,400
Steam into MIS retort, lb/hr	848,000	-	848,000
Seepage into MIS retort	nil	-	nil
Sand, lb/hr	0	16,400	16,400
Water to cool Lurgi ash 206°-302° lb/hr	-	202,800	202,800
Water to make Lurgi ash slurry, lb/hr	-	1,203,500	1,203,500
Portland cement additive to slurry, lb/hr	-	51,150	51,150
	17,677,000	6,295,250	23,972,250
<u>OUT</u>			
Portland cement in MIS and surface fill slurry, lb/hr	-	51,150	51,150
Spent shale, lb/hr	10,091,000	2,557,500	13,470,500
Water evaporation to cool Lurgi ash, lb/hr	-	202,800	202,800
Slurry water, lb/hr	-	1,203,500	1,203,500
Lurgi-Ruhrgas flue gas, lb/hr	-	1,883,300	1,883,300
Water separated from oil, lb/hr	386,000	45,300	431,300
Water vapor leaving with retort gas, lb/hr	917,000	2,480	919,480
Retort gas (dry, NH ₃ & H ₂ S free), lb/hr	5,464,000 ^(a)	68,300 ^(b)	5,482,300
Product crude shale oil, lb/hr	754,000 ^(c)	330,900 ^(d)	1,084,900
Sulfur, lb/hr	12,000	2,375	14,375
Ammonia, lb/hr	23,000	nil	23,000
Coke (left on MIS spent shale)	30,000	0	30,000
	17,677,000	6,347,605	24,796,605

NOTES:

(a) Low BTU gas

(b) High BTU gas

(c) (1.5% moisture)

(d) Dry basis

TABLE 3-17
MATERIAL BALANCE LURGI-RUHRGAS RETORTING SECTION
ONLY WITH SLURRY BACKFILLING OF MIS RETORTS

	<u>No.</u>	<u>Stream</u>	<u>Total Lb/Hr</u>	<u>H₂O GPM</u>
<u>IN</u>	1.	Raw shale, 25 gallons/ton	3,410,000	136
	2.	Sand	16,400	
	3.	Air	1,411,400	
	4.	H ₂ O to cool ash 206-302 °F	202,800	406
	5.	Slurry H ₂ O	1,203,500	2408
	6.	Portland Cement and Additives	<u>51,150</u>	<u> </u>
			6,295,250	2950
<u>OUT</u>	7.	Sulfur	2,375	
	8.	Spent shale	2,557,500	
	9.	Water evaporation to cool shale ash	202,800	406
	10.	Slurry H ₂ O	1,203,500	2408
	11.	Flue gas	1,883,300	
	12.	Heavy oil	84,900	
	13.	Middle oil	246,000	
	14.	Product gas dry	68,300	
	15.	H ₂ O in product gas	2,480	5
	16.	Gas liquor (H ₂ O)	45,300	91
	17.	Portland Cement and Additives	<u>51,150</u>	<u> </u>
			6,347,605	2910

TABLE 3-18
ESTIMATION OF ELECTRIC POWER EXPORTED
FROM THE MIS/LURGI-RUHRGAS PLANT

Treated gas rate, dry	: MIS ^(a)	$5,218 \times 10^3$ lb/hr
	: Lurgi-Ruhrgas	65×10^3 lb/hr
Calculated HHV, dry basis:	MIS (65 Btu/scf) ^(a)	791 Btu/lb
	: Lurgi-Ruhrgas (707 Btu/scf)	8,908 Btu/lb
Gross heat available		5.23×10^9 Btu/hr
Heat required for steam raising ^(a)		<u>1.48×10^9 Btu/hr</u>
Net heat available for power generation		3.75×10^9 Btu/hr
Electricity generated, open gas cycle ^(b)		288
Electricity consumed	: MIS ^(a)	106 MW
	: Lurgi-Ruhrgas ^(c)	<u>42 MW</u>
Total consumed		148 MW
Net electricity available for export		<u>140 MW</u>

(a) As given in Table 3-14.

(b) Generated in gas turbine at 13,000 Btu/kWh.

(c) Scaled from value used for TOSCO II plant.

less expensive incidental absorber and instead burn much of the NH_3 in a turbine or boiler. (4) If compression of the gas is done for gasoline recovery, does one remove NH_3 by water scrubbing and H_2S by a high pressure Stretford unit or remove NH_3 by water scrubbing and remove H_2S somewhat less completely by high pressure "selective" absorption by ammonia liquor or amine liquor followed by treatment of H_2S subsequently stripped from the liquor by the Claus process? (5) In the Paraho process should NH_3 also be collected from the retort recycle gas?

Except for TOSCO II other variations are possible depending on whether the low Btu gas is burned to raise steam for power or is burned in a gas turbine for power. The high Btu TOSCO II gas is more useful for ball heating and a refinery hydrogen source than for power. A variation of TOSCO II is to burn the the residual coke from the spent shale thus eliminating possible "toxic chemicals" on it and raising much steam for heat or power.

Table 3-19 includes most of the above variations and shows some other options.

We have chosen for detailed costing the removal of H_2S by the Stretford process from the low Btu Paraho and low Btu MIS gases at atmospheric pressure. The Paraho and Rio Blanco organizations, on the other hand, have preferred to compress this gas before H_2S removal although Occidental does not seem to prefer this. Neither Occidental nor Rio Blanco have reported plans to use a gasoline-LPG absorption plant on the compressed low Btu gas although both Colony and Lurgi-Ruhrgas have proposed an absorption plant on the high Btu gas from the TOSCO II and Lurgi-Ruhrgas retorts, respectively.

Compression of the low Btu gas may be used for two of the operations given in Table 3-19 without harm (Stretford or NH_3 absorption in water) and is required for amine scrubbing and gasoline plant. Once the gas is compressed to as high as 150 psi for any reason, gas turbine power is indicated rather than boiler power. If the gas is never compressed to 150 psi, boiler power or regenerative gas turbines at 50 psi must be used. If the gas is compressed, the options (1) gasoline and LPG absorption, (2) pressurized Stretford H_2S removal, (3) pressurized NH_3 absorption by water scrubbing, and (4) pressurized selective H_2S absorption are possible. (See Table 3-19). Table 3-19 shows the variations selected for costing. In some cases important alternate

TABLE 3-19
Retort Gas Purification, LPG and
Gasoline Plant, and Power Generation Options

Operation	High BTU low NH ₃ gas from TOSCO II or Lurgi-Ruhrgas				Low BTU high NH ₃ gas from MIS retorts					Low BTU high NH ₃ gas from Paraho direct mode				
	Selected Lurgi variation	Selected TOSCO variation	Colony TOSCO plan	Lurgi plan	Selected variation	Variation II	Variation III	Variation IV	Variation V	Selected variation	Variation II	Variation III	Variation IV	Variation V
Atmospheric pressure (1) or high pressure (2) aqueous NH ₃ scrubbing					(1)	(2)	(2)	(2)	(1)	(1)	(2)	(2)	(2)	(1)
Atmospheric pressure (1) or high pressure (2) Stretford H ₂ S scrubbing	X	(2)			(1)	(2)			(1)	(1)	(2)			(1)
Selective high pressure H ₂ S Scrubbing (amine or other) followed by Claus process			(2)	X			(2)	(2)				(2)	(2)	
LPG and gasoline recovery by sponge oil absorption at moderate to high pressure			X	X		X	X	X			X	X	X	
Boiler and steam turbine, process heat boiler, hydrogen generation from high BTU gas, and/or steam for MIS retort	X	X	X	X	X	X	X	X	X	X	X	X	X	X
No H ₂ S scrubbing with (1), no NH ₃ scrubbing (a), or (2) NH ₃ scrubbing (b), either with SO ₂ scrubbing of					X	X	X	X		X	X	X	X	
Regenerative gas turbine, 50 psi, low temperature exhaust									X					X

Notes:

- (a) Multistage gas turbine combustor development needed for NO_x control.
(b) Single stage gas turbine combustor probably adequate.

variations have also been costed, particularly the removal of SO_2 after the gas turbine rather than removal of H_2S before the gas turbine.

Two methods for H_2S stripping are Stretford H_2S removal and selective regenerable H_2S absorption. The regenerable H_2S absorption unit may be based on rate selective H_2S absorption (as opposed to CO_2 absorption) in ammonia liquor or may be based on selective H_2S absorption in amine liquor. In both cases the H_2S with some CO_2 is stripped from the liquor and converted to sulfur in a Claus unit. H_2 and SO_2 in the Claus "tail gas" can be reduced considerably by attachment of any of a number of tail gas cleanup units such as a Wellman-Lord unit.

NH_3 absorption by a water scrubber is not shown for the TOSCO II retort gas or Lurgi-Ruhrgas gas. The quantity of NH_3 produced by these low temperature retorts is small. A water scrubber is used at the top of the TOSCO II retort gas fractionator. This scrubber will absorb the NH_3 .

A similar circumstance occurs in the final spray cooler for the Lurgi-Ruhrgas retort where a water spray is used, although small amounts of NH_3 would be collected. The final spray cooler is used primarily for gas cooling and partial drying before the gasoline plant. The equipment for NH_3 absorption from the Paraho gas and MIS gas is of major concern. The large gas flow requires three (Paraho) or five (MIS) large diameter absorbers. Large quantities of steam are required for stripping out the large amount of absorbed NH_3 .

No gasoline plant is involved in the MIS and Paraho variations costed for pollution control in this study. If used, additional steam for stripping the gasoline and LPG out of the recirculating oil liquor would be needed and the retort gas would need to be compressed. Since it needs to be compressed to 150 psi for a gas turbine, the gasoline plant should also operate at 150 psi as more complete capture of gasoline and LPG is possible at high pressure for a given stripping steam load. As much as 10% additional "crude shale oil" output, composed of high value LPG and gasoline components, may be possible from these gasoline plants. These light ends could be dissolved in the shale oil thereby lowering its viscosity and pour point. For this reason, the cost of pollution control per barrel of crude oil seems somewhat conservatively presented for the MIS and Paraho process, for eventually the gasoline plants seem probable and then the value of the

gasoline and LPG could be partially credited against compression of the retort gas and its cooling.

As shown in Table 3-19, boilers are used in combusting at least part of the retort gas for all retort models. A possible exception is the TOSCO II gas which may be used in ball heating and in the hydrogen plant. In this case, oil, LPG, and gas are options for firing the ball heater and boiler. The remainder of the retort gas from the MIS or Paraho retort will be burned in gas turbine electric power plants. This is proposed by Oxy, Rio Blanco, and other developers. Power may be generated more efficiently and less costly from low Btu gas (such as from the MIS and Paraho retorts) by use of gas turbines rather than by use of special burners, boilers, and steam turbines. Low Btu gas may be easier to burn when compressed in the gas turbine combustor than at atmospheric pressure in a boiler.

3.5.2 Alternate Processing Schemes for Paraho Direct Mode Plant as Influenced by Regulatory Scenarios

Switching from "least strict" to the "more strict" air pollution regulatory scenario does not cause variation in the retort gas purification systems for Paraho and MIS plants.

There is some doubt in regard to the amount of NO_x in the Paraho retort gas (resulting from combustion of NH_3 in the Paraho retort recycle gas). This NO_x might escape the ammonia scrubber and be picked up by the Stretford scrubber and carried in the Stretford liquor to the oxidizer. Here nitrates might be generated which would require more frequent changing of the Stretford liquor and higher operating cost. Nevertheless little or none of this NO_x may find its way out the Stretford air oxidizer stack. That NO_x passing through the Stretford absorber into the gas turbine combustor would be minor compared to that produced in the combustor. A strong detrimental effect of NO_x on the Stretford liquor may necessitate a change in the processing scheme. Further investigation of NO_x levels in both Paraho and MIS retort gases is necessary.

Paraho has reported results found by various groups showing a range of less than 50 ppm NO_x to less than 200 ppm NO_x in the direct mode retort gas. A burn in the Livermore Lab 2-foot diameter $\times$ 20-foot tall MIS simulator retort showed less than 10 ppm NO_x . If the NO_x is a problem in the Stretford process, and it results from combustion of NH_3 in the recycle gas burned in

the retort, then this can be eliminated by scrubbing it out of the entire retort gas stream rather than out of only the make gas purge stream. This requires an ammonia scrubber 3 to 4 times larger in cross section, but, as developed in the following section, this course may be justified based on ammonia byproduct value. Inexpensive steam from waste heat will be available for NH_3 stripping.

Since zero water discharge is attained by the Paraho processing complex taken for study, no change caused by water pollution regulatory scenario change is possible.

No variation is anticipated with variation in solid waste regulatory scenario. Yet, variations will occur if spent shale is classified as a hazardous material. For the latter case, a spent shale cement liner or "bathtub" might be built to enclose the spent shale. Such an approach has been taken in Section 7.0 of this report, and costs determined as a worst case.

3.5.3 Maximizing Ammonia Byproduct Recovery from Paraho Direct Mode Retort Through Scrubbing Total Retort Gas

Higher temperature retorting produces more ammonia than retorting at temperatures of approximately 900°F, such as in the Fischer assay retort, TOSCO II retort, or Lurgi-Ruhrgas retort. (See Figure 3-14). NH_3 production from Soldier Summit Green River shale is minor at approximately 900°F, but a large increase towards various plateaus (depending on moisture[22]) is present at temperatures of approximately 800°C (1472°F) and higher. This large difference in ammonia evolution is a major distinction between "low temperature" and "high temperature" oil shale retorting.

Column 16 of Table 3-7 shows the expected NH_3 produced per unit of oil when the NH_3 is scrubbed from the total retort gas (prior to recycling in Paraho direct mode units). The data from Heistand, obtained late in 1978, is higher than previous NH_3 data. The data for 15 gpt shale retorted for an Oxy MIS retort is also contained in Table 3-7. These data show equally appreciable ammonia yields from either Paraho or MIS retorts and, moreover, yields nearing the expected maximum from the approximately 65% conversion of the nitrogen in the shale predicted by Gavin [19] or from the curves of Figure 3-15 of Maier and Drapeau.[22]

The moderate costs of water scrubbing and NH_3 stripping are presented in a later section. These costs suggest that an improvement in economics

FIGURE 3-14. EFFECT OF TEMPERATURE AND MOISTURE ON NH_3 EVOLUTION IN GREEN RIVER SHALE RETORTING

(O'Heshi and Fukuesan)

FIGURE 3-15
FUSHUN (MANCHURIA). RETORTS

would be expected if the greater quantity of NH_3 were removed from the entire retort gases of the Paraho retort. This, to be practical, must be done at near atmospheric pressure as the recycle gas must be injected into the retort at near atmospheric pressure. Fortunately NH_3 is rather completely and easily removed by water absorption when much CO_2 is also present in the gas. This is the situation for both Paraho and MIS gas.

The Fushun retort gas contained less carbon dioxide than the Paraho and MIS gases. The Fushun shale is noncalcareous, in contrast with Green River shale whose mineral carbonates produce much CO_2 in the Paraho retort and MIS retort. Ammonia was scrubbed from the total of the Fushun retort gases with sulfuric acid to produce ammonium sulfate. Figure 3-15 shows this retort scheme. Figure 3-16 shows the flow scheme. A description of the Fushun retort, deposits, and system follows below.*

* Fushun Retort: The Fushun oil shale retort (see Figure 3-15) [16] was a modified Pintsch (Estonia) or Inabe (Japan) retort and was used on a large scale by the Japanese on vast beds of low-grade oil shale at Fushun (Fu-Sung) Manchuria [5]. This was a gas combustion retort somewhat similar in counterflow gas-to-shale heat transfer principle to the Bureau of Mines or Paraho gas combustion retorts. However, the producer gas section and shale distillation section were separated physically and the retorting shale was not exposed to air until retorting was finished. Then, in the lower producer gas section, air was admitted, and carbon was burned off the spent shale. Low calorie gas, after oil and NH_3 removal, was burnt in an external furnace then mixed with the producer gas stream and the hot mixture passed counterflow through the shale in the upper retorting section. Blast pressure drop and shale throughput rate was limited by the capability of a water seal at the burned shale ash offtake at the bottom of the producer retort section and, possibly, by a low heat transfer coefficient between the little burning heat transfer gas and the shale in the retort section. Each of these Fushun retorts handled about 50 tons of shale per day and retorting zone was at between 350° to 550°C (662° to $1,022^\circ\text{F}$).

The Fushun oil shale deposit has been mined since 1923, is of tertiary age, 450-ft. thick, and averages 15 gallon per ton Fischer assay [20]. It is underlain by coal and part of its viability was due to the fact it was stripped off of the coal anyway. Inferred reserves were 2,100,000,000 bbls. [21] In 1943 the Fushun oil shale plants, controlled then by the Japanese, processed 4,700,000 tons of shale and produced 1,317,000 bbls of shale oil [35]. During the Korean War operations were greatly expanded by the Chinese and in 1968 may have been producing 40,000 bbl per day. [21]

The Fushun shale is mostly SiO_2 and Al_2O_3 in ash analysis with little CaO and MgO hence carbonate must be nil. This must have contributed to the success of the Fushun retort, particularly in operation of

its producer gas section. Green River shale under such conditions would have evolved considerable CO due to carbonate calcination and not so much useful heat and CO gas would have been generated. The Green River shale, however, would be richer (and presumably more carbon would be available for the producer gas section) than Fushun shale. The status of the Fushun retort today is not known.

The somewhat similar Pintsch-type retorts (Kohlta Jarve retorts) in Estonia were modified from vertical gas flow to horizontal-transverse flow in 1961 to 1963 to give better shale bed heat distribution. The Estonian oil shale contains high concentrations of CaO in its ash (and hence high carbonate) but relatively low concentrations of MgO. This may contribute to the success of these retorts in Estonia. Green River shale MgO is mostly as Dolomite which thermally dissociates at a lower temperature than calcite, CaCO_3 .

FIGURE 3-16. FLOW CHART RETORTING FUSHUN OIL SHALE

The retort gases exit the Paraho direct mode retort at a temperature of 150°F. These gases may enter the ammonia stripper at this temperature, or they may be cooled. Cooler gases in the ammonia stripper reduce steam load and increase the ease of NH_3 capture. An economic balance of stripper spray tower gas cooling and value of ammonia saved would be required to set the recycle gas ammonia absorbed temperature in the absence of other requirements.

Several reasons for use of lower NH_3 absorber temperatures are:

- (1) A Stretford process following the absorber should preferably operate at below 120°F, 100°F, or lower to reduce deterioration of the costly Stretford liquor. If a different H_2S removal system is used or none is used at all this does not apply.
- (2) If the retort gases are to be compressed to approximately 50 psi for a regenerative gas turbine, or 150 psi for a simple gas turbine, or compressed as is needed for a selective amine H_2S absorber or a LPG and gasoline absorber plant, cooling the inlet gas reduces the size and hp requirements of the compressor.
- (3) If an amine H_2S absorber or gasoline plant is used, the cooler the inlet gas the more effective and efficient they are.
- (4) Lower gas temperature reduces the blower size and hp for moving recycle gas.
- (5) Cooler gas will condense out more oil aerosol that may be captured in the electrostatic precipitator.

Three reasons for maintaining the retort gas temperature are:

- (1) The retort is cooled less by the incoming streams and becomes more efficient.
- (2) Less equipment is required if the gas is not cooled at all.
- (3) If the gas is cooled below 150°F condensation of retort water will occur which tends to emulsify with simultaneously collected oil causing separation problems and corrosion.

3.5.4 Omission of Most NH_3 Recovery and Use of SO_2 Scrubber for Gas Turbine Stack in Paraho and MIS Processes

Use of little or no NH_3 absorber at all is possible if:

- (1) Two stage gas turbine combustion is feasible to inhibit the formation of NO_x from the burning NH_3 content of the retort gas and

- (2) The SO_2 formed from the H_2S in the gas turbine combustor may be scrubbed from the turbine exhaust by, possibly a lime scrubber (lime from burned spent shale). Use of a stack gas scrubber on a turbine exhaust necessitates the use of a regenerative turbine heat exchange to lower the turbine exhaust temperature from approximately 900°F to approximately 350°F .

If ammonia byproduct sales are not profitable and items (1) and (2) above are practical, burning of the ammonia may be the preferred choice. Compression for a gas turbine does not need cooling and the heat of compression could help keep the gas above the dew point. Concentrations of absolute water and oil increase with compression. Knock out drums may be necessary.

3.5.5 Gasoline and LPG Recovery Plant

Cool retort gases are needed for an absorption type LPG and gasoline recovery plant. Retort gas compression (50-150 psi), and subsequent cooling, promote condensation of a high percentage of the water content of the gas. This compression-cooling operation necessitates the need for ammonia recovery for cleaning the foul water produced by this operation.

Figure 3-17 shows the amount of gasoline and LPG in various retort gases per 100 lbs of theoretical oil from the shale. The temperature of the condenser or recovery system greatly influences this quantity. The MIS gas seems to contain more C_5+ than the Paraho gas, discounting for the apparently too high C_6+ point. The volume of MIS gas per ton of shale retorted is larger than that of Paraho gas and would hold more gasoline at a given temperature. The "improved Fischer assay" data are derived from gases of lower temperatures. The gases pass through a 32°F temperature and are therefore cooler than the TOSCO II gases which pass through a warmer plant condenser system.

The gas analyses used for calculations in this study are shown in Table 3-20. The hydrocarbon content of these gases is important in determining the necessity of an LPG-gasoline recovery plant.

The TOSCO II and Lurgi-Ruhrgas plants both have been shown with associated LPG-gasoline recovery plants. The TOSCO II gasoline plant pressure is nearly 150 psi.

FIGURE 3-17. LPG AND GASOLINE IN RETORT GASES

TABLE 3-20
COMPARISON OF VARIOUS RETORT GAS ANALYSES

HIGH BTU RETORT GAS:

	<u>Paraho Indirect Modé, Jones (11)</u>	<u>Lurgi Schmalfeld (36)</u>	<u>TOSCO Plant Goodfellow & Atwood (37)</u>
H ₂	24.8%	18.5%	22.51%
O ₂	-0-		-0-
N ₂ + argon	0.7	7.7	
CH ₄	28.7	11.9	15.23
CO	2.6	2.7	3.57
CO ₂	15.1	25.0	21.41
H ₂ S	3.5	46.2 g/NM ³ *	4.32
NH ₃	1.2		-
C ₂ ⁼	9	12.0	5.36
C ₂	6.9		10.36
C ₃ ⁼	5.5	8.9	3.68
C ₃			4.03
C ₄	2	C ₄ ⁺ 13.3	4.23
C ₅			2.54
C ₆			1.49
C ₇			0.832
C ₈			0.346
C ₈ ⁺			0.086
SO ₂		0.06 g/NM ³	
H.H.V.	885 BTU/SCF		

*Reference 36 lists a value of 1.98 g/NM³; the authors feel that this value is unrealistic. The value shown in the table is approximately the same as for TOSCO II and was used in this study.

TABLE 3-20 (Cont.)

PARAHO DIRECT MODE GAS:

	Jones (11) <u>Dry Basis</u>	(Union Oil Retort, 1954-1958 <u>Dry Basis (15)</u>	Wise & Prien (3) <u>Wet Basis</u>	WPA Wet Basis Derived from Wise & Prien <u>NH₃, H₂S & Temp.</u>
H ₂	2.5%	2.2%	3.59%	3.539%
O ₂	-0-	2.2 air	.76	0.743
N ₂ + argon	65.7	57.6	52.89	51.342
CH ₄	2.2	0.9	1.90	1.858
CO	2.5	4.6	1.63	1.600
CO ₂	24.2	30.3	17.93	17.572
H ₂ S	0.266	0.1	0.18	0.242 (0.3% dry)
NH ₃	0.2490	?	0.17	0.566 (0.7% dry)
C ₂ ⁼	0.7	0.6	0.76	0.743
C ₂	0.6	0.4	0.78	0.768
C ₃ ⁼	} 0.7	0.4	0.37	0.364
C ₃		0.2	0.39	0.380
C ₄ ⁼	} 0.4	0.3	} 0.33 MW 57.0	0.323
C ₄		0.1		
C ₅		0.1	0.13 MW 71.5	0.129
C ₆ +			0.63 MW 96.2	0.622
H ₂ O	-	-	17.56	19.210
			100.00	100.00
Molecular Wt.				28.43 MW
H.H.V. Dry	102 BTU/SCF	80	150 BTU/SCF	
H.H.V. Wet			120 BTU/SCF	

TABLE 3-20 (Cont.)

MIS GAS:

$$\frac{\text{Gas Rate } 6,366,000 \text{ lb/hr wet}}{\text{Oil Rate}^* 754.000 \text{ lb/hr } (.985)} = 8.571$$

	<u>DRI to Domahidy Wet Basis</u>	<u>WPA Wet Basis, Mass %</u>	<u>Lb Component in Gas Per 100 lb Oil Prod.</u>	
H ₂	4.246%	0.298% Mass	2.554%	
O ₂	0.0772			
N ₂ + argon	44.78			
CH ₄	1.158	0.650	5.571	
CO	0.772			
CO ₂	24.704			
H ₂ S	0.1544			
NH ₃	0.6176			
C ₂ ⁼	} 0.3088	0.315	2.699	} 7.285
C ₂				
C ₃ ⁼	} 0.1544	0.232	1.988	
C ₃				
C ₄	} 0.3088	C ₄ ⁺ 0.618	5.297	
C ₅				
SO ₂	(0.3088 ppm)			
H ₂ O	22.8			
	<u>100.08</u>			

*Oil contains 1.5% H₂O.

TABLE 3-20 (Cont.)

PARAHO DIRECT MODE GAS:

Product gas 3,571,000 lb/hr, wet
 Product oil "dry" 1,352,000 lb/hr = 2.641

	WPA Wet Basis Mass %	Lb Component in Gas Per 100 lb Oil Prod.	
H ₂	0.249% Mass	0.6576%	
CH ₄	1.046	2.763	
C ₂ ⁼	} 1.542	4.072	} 11.10
C ₂			
C ₃ ⁼	} 1.126	2.974	
C ₃			
C ₄	} 0.648	1.711	
C ₄			
C ₅	0.324	0.8557	
C ₆ +	2.105	5.559	

TABLE 3-20 (Cont.)

MIS GAS:

Dry Basis					WPA Table	DRI to Domahidy
S 11 (air)	S 13 (air + steam) 1:1					
	Livermore Labs.	Livermore Labs. Experimental	Livermore Labs. Model or Calc. from J.W. Smith Fischer Assays*	Wet Basis	Dry Basis	
		(early) (late)				
H ₂	1.9%	4.7 - 6.9%	3.6%	4.246%	5.5%	
O ₂	0.6	0.1	0.0	0.077	0.1	
N ₂ + argon	59.7	54.4 - 52.4	51.8	44.699	58	
CH ₄	1.0	1.1	1.0	1.158	1.5	
CO	3.1	1.3 - 2.4	1.4	0.772	1	
CO ₂	32.4	37.1 - 35.8	40.9	24.704	32	
H ₂ S	?	?	?	0.154	0.2	
NH ₃	?	?	?	0.618	0.8	
C ₂ ⁼	0.4	0.4		0.309	0.4	
C ₂			1.3*			
C ₃ ⁼	0.2	0.2		0.154	0.2	
C ₃						
C ₄	0.7 C ₄ +	0.7 C ₄ +		0.309 C ₄ +	0.4	
C ₅						
SO ₂					0.5 ppm	
	100.0				100.1	
H ₂ O					22.8	
H.H.V.	69	71 - 81	55			
				CS ₂	0.1 ppm	
				COS	1.9 ppm	

3.5.6 Alternate Processing Approaches for TOSCO II Plant

For TOSCO II, few alternate methods exist for the handling of NH_3 and H_2S . TOSCO II is a low temperature retort with short residence time and little ammonia is produced in the retort. The ammonia production from the TOSCO II system originates at the refinery, largely from hydrotreating the oil to make low N low S syncrude. A number of major variations of a shale oil refinery exist. Examining this is beyond the scope of the present study.

Shale Preheater: Much of the air pollution from the TOSCO II plant is derived from the shale preheater stack. Preheating the shale with waste heat is important to the thermal efficiency of the plant. The hydrocarbon emissions cited are based on preheating the shale to 400°F . A lower shale preheat would reduce the HC emissions but would lower plant efficiency. The waste heat used is derived from the hot flue gas issuing from the ball heater at a temperature of approximately 1400°F . This flue gas contains much NO_x from (1) the combustion of crude shale oil and (2) from its high temperature generation from nitrogen of the combustion air. Currently a multistage stepwise counterflow entrained solids shale preheater is being considered by TOSCO. This would allow exposure of the shale to lower inlet gas temperatures. Shale fines are rapidly heated to a high temperature at the inlet section resulting in the release of hydrocarbons into the gas stream.

Reduction of NO_x in the shale preheater stack gas could be achieved by burning refinery gas, retort gas, and possibly LPG instead of crude shale oil in the ball heater. Unfortunately, there is not enough gas available for this heat and other fuels must be burned. Were the TOSCO II process used in proximity to another retorting system producing much gas such as Paraho or MIS, gas would be available. Burning such lower Btu gas may take a special burner, however, without fortification with higher Btu gas or oil.

Spent Shale Burner: Much heat is potentially available in the carbon or coke left in the spent shale. With the TOSCO II process, the carbonaceous spent shale is dumped to waste. It may be possible to design a spent shale burner which can produce steam for power or plant heat. It would be desirable to transfer some of this heat to the ball heater thereby reducing the amount of fuel needed to drive the heater. A problem arises in the burning of spent

shale. The nitrogen in the spent shale tends to form NO_x rather than NH_3 . The NO_x emissions may be significant.

Ball Heater Design: One method of lowering the ball heater fuel requirement involves the use of a gas which flows countercurrent to ball circulation pattern. This would also allow lower inlet gas temperature to the shale preheater with lower expected HC emissions. The shale preheater, as previously discussed, would allow more efficient heat transfer from inlet flue gas to shale, thereby compensating for the lower heat in the flue gas.

3.5.7 Alternate Processing Approaches for Modified In Situ Plant

All alternate processing approaches previously discussed for the Paraho direct mode plant are applicable to the MIS plant. One minor difference exists. With the MIS process, some retort water condenses and drips down the retort rubble during the early stage of a burn. In the Oxy burn (retorts 4, 5, and 6), the water was alkaline. This resulted in capture of only small amounts of ammonia.

3.5.8 Alternate Processing Approaches for MIS Retorting with Lurgi-Ruhrgas Retorting and Grout Backfill

The TOSCO II process may be substituted for the Lurgi-Ruhrgas process for production of finely divided burned spent shale. This spent shale would possibly have properties similar to Lurgi-Ruhrgas spent shale and could be used, with additives, in slurry backfilling the burned out MIS retorts. The spent shale from a TOSCO II plant with shale burner retrofit might also be suitable.

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4.0 THE REGULATORY SCENARIOS

4.1 INTRODUCTION

Estimates of the environmental control costs for oil shale were made by combining the four process models (described in Section 3.0) with specific estimates of the future environmental regulations which might be applied to the mature industry. Where possible, two scenarios were developed for each pollutant. The "less strict" scenarios represent current permits and regulatory thinking extrapolated to commercial operations. And in many cases even these scenarios are quite restrictive. The "more strict" scenarios are meant to represent best estimates of the "worst case" which could possibly occur if current and anticipated environmental regulations are applied in the most strict manner possible.

Clearly, the setting of the regulatory scenarios can dramatically affect the control costs estimated in this study. Efforts were made to establish reasonable scenarios which do not overstate or understate the problem and which allow estimation of realistic "worst case" costs. This section summarizes the actual scenarios applied for each pollutant. Appendix 4.0 summarizes all pertinent laws and regulations, and describes the use of these in determining the actual regulatory levels for an industry.

4.2 THE APPROACH, ASSUMPTIONS AND LIMITATIONS

Since, for the most part, regulations do not now exist specifically for oil shale, any regulations assumed are necessarily extrapolations containing uncertainty. The regulation of a commercial oil shale industry will involve federal, state and local authorities, and it was necessary to work with all these levels of government for this study. Even within a single government agency, the answers to questions regarding probable regulations

were found to be different from person to person. To arrive at regulatory scenarios in the face of such uncertainties, the following approach was taken.

At the start of the program, Denver Research Institute lead the effort through contact with all appropriate federal, state, and local authorities who will be involved in regulation of the industry. Differences of opinion within a particular agency were determined by extended discussion of the particular issue with the personnel involved. For each pollutant, summaries of the applicable laws and regulations were then prepared along with levels of regulation suggested by the government authorities contacted. The summaries and suggestions were forwarded to DOE for review and revision. The review process within DOE generally involved direct discussion of the issues between DOE and the appropriate regulators. Final scenarios were then specified by DOE and, for the most part, approved by the agency which will actually be involved in the regulation. The only exception to this process was made in the case of RCRA regulation. DOE established the scenarios for RCRA without any discussions between DOE and EPA.

The scenarios assumed for this study were set by DOE and approved by the appropriate regulatory agencies. Even the "less strict" scenarios are often quite strict. Hence, the scenarios applied in this study represent one of several ways in which conservative design parameters were used to estimate control costs. The scenarios should not be interpreted to be recommendations for future regulation of the industry.

4.3 USE OF THE SCENARIOS

The scenarios listed in the following sections can be viewed as goals set during the beginning of the project as guidance for use by the engineers in selecting and designing the controls. The final level of control achieved for a particular pollutant is, in a number of instances, different from the scenarios and higher. Wherever possible, two scenarios were assumed in the beginning, but the final results, in some instances reflect application of only a single scenario. Such a result is inevitable, and is a reflection of the regulatory procedure.

Regulations governing the emission level for a particular pollutant can be specified in a variety of ways. First, actual allowable emissions can

be expressed in terms of quantities, e.g., pounds per day, parts per million, percent of total pollutant, etc. For this project, whenever possible the scenarios were specified using some sort of quantity measure. Regulation can also be specified in terms of "best available technology." In other words, the specifications may be set as the emission level achievable by applying the best available control technology to a particular pollutant stream. The regulations may specify the required equipment, or a quantity measure of emissions which are expected from the equipment, or both. The term "best available technology" also considers cost-benefit trade-offs. Finally, the regulations may be set in terms of the quality of the environment some place different from the location of the production operation. Good examples of this type of specification are the Colorado River salinity agreement between the U.S. and Mexico at the Mexican border, and air quality requirements in neighboring Class I areas. The eventual regulated emission levels will be determined by measurement. The optimum performance of "best available technology" or the actual extent of environmental degradation at target locations will be measured during operation of actual commercial plants before final decisions are made. Presently, the final requirements can only be estimated using engineering calculations and/or mathematical modeling. Modeling was not used for this study, but the results shown will be modeled in an EPA sponsored program to begin shortly involving Denver Research Institute, Stone & Webster, Engineering and General Electric TEMPO.

The manner in which the approach to development of regulations efforts this project is complex. Also the cost and performance of all controls can be interrelated. The complexity is illustrated by the following examples.

The allowable levels for SO_2 emissions were set at 90% and 93.5% control for this project. Such control can be achieved either by removing the sulfur before burning the retort gas, or by removing the SO_2 from the flue gas after combustion. Both alternatives and several technologies were originally considered. The process selected for sulfur removal is the Stretford process. Scrubbers are considered for SO_2 removal from the flue gas. Calculations and data on actual pilot plant experience indicate that the Stretford process should achieve 99% control or better, a level much in excess of the two scenarios. Both the 90% and 93.5% control levels can be achieved with SO_2

scrubbers, but with difficulty. Finally, comparing the cost of the two approaches, the SO₂ scrubbers are also found to be substantially more expensive than the Stretford process. Hence, the final answer for the control of SO₂ emissions appears actually as a single scenario which specifies use of the Stretford process to achieve 99% control, since the other alternatives provides less control at substantially higher cost.

The selection of particulate controls in the TOSCO II process represents a much less clear cut decision, and illustrates the need to consider the controls as part of the overall processing system. Baghouses achieve only a slightly higher degree of control than scrubbers, but are much more expensive. Both provide high levels of control. Given this information, cost-benefit considerations would suggest that scrubbers are the best choice. However, scrubbers use large amounts of water. Use of scrubbers will require additional water treatment capacity, and greater consumption of this scarce resource. Hence, selection of the best available technology may involve considerations other than level of control and cost of the specific pieces of equipment. In this case, the capital and operating costs for two scenarios are shown. The level of control for both equipment selections, however, is near to, or better than the upper scenario set at the beginning of the project.

The interdependency of cost figures and regulatory scenarios is illustrated by comparing the MIS and MIS/Lurgi processes. The MIS/Lurgi process is actually the more strict scenario for the MIS process, representing one means, and the authors feel the best means, of grouting spent MIS retorts. To provide the grouting material, an aboveground retorting process is added along with additional costs for air, water and solid waste controls. Not only is there a cost for grouting, but costs are added in all categories of pollution control. However, the amount of oil produced is substantially increased, and a higher utilization of the resource is achieved. The true cost of grouting can only be judged by considering the total cost of pollution control for the two process on a per barrel of oil basis including a credit for increased recoverable oil reserves to the MIS/Lurgi process. No such credit was allowed in this project.

The above examples illustrate the complexity involved in applications of regulatory scenarios to process models for the determination of

pollution control costs. The starting scenarios are presented next. The results of their application constitute the remainder of this report.

4.4 SCENARIOS FOR AIR POLLUTION CONTROL

The actual emission levels for each process are presented in Section 5.0 along with detailed descriptions of the control measures used.

4.4.1 Particulate Matter

The regulatory scenarios originally suggested for particulate control are as follows:

Less Strict: 90% Control of Vented Emissions.
Good Management of Fugitive Dusts.

More Strict: 99% Control of Vented Emissions.
Good Management of Fugitive Dusts.

Two levels of control were considered for the TOSCO and Paraho processes. The first level of control essentially provides the equipment specified by the developers. The higher level of control for the TOSCO process substitutes baghouses for venturi scrubbers. For the Paraho process, additional baghouses are added. Both sets of controls for both processes should meet the standards of the more strict of the original scenarios.

For the MIS and MIS/Lurgi processes, only one set of controls is provided. The controls specified represent the most logical choices and the best available technology. The performance meets or exceeds the more strict scenario. A lower level of control would represent an artificial case where sources of the vented emissions remain uncontrolled.

Fugitive dusts are controlled for all processes using water spraying, chemical spraying, binders, and paving of roads. All crushing and screening operations are enclosed and all materials are transported by covered conveyors vented through baghouses.

4.4.2 Hydrocarbons

The suggested regulatory scenarios for hydrocarbon control are as follows:

Less Strict: Good Management Practices.

More Strict: Controls on All Major Sources of Hydrocarbon Emissions.

Only one level of control was considered for hydrocarbon emissions from all processes. The Paraho process and MIS process do not have significant sources of hydrocarbon emissions. To handle startup, shutdown, and interrupt situations, the boilers and turbines are used as flares. The TOSCO process and MIS/Lurgi process would have significant hydrocarbon emission levels without controls. A low flare and thermal oxidizer are provided for both processes. The thermal oxidizer is sized to handle the normal hydrocarbon load and the low flare provides extra capacity for start up, etc. The TOSCO II process has a second thermal oxidizer as part of the shale preheater. A second scenario or lower level of control could only be achieved artificially by undersizing the units, and hence was not used.

Floating roof storage tanks are provided for all product storage as required by the Federal New Source Performance Standards. Addition of double seals was found to make an insignificant cost difference.

4.4.3 Nitrogen Oxides

Less Strict: No Controls.

More Strict: Adjustment of Combustion Conditions Where Possible. Appropriate commercially Available Controls.

Adjustment of processing parameters to minimize NO_x production was considered to be a no cost item or an item for which no cost could be determined. Only the TOSCO and Lurgi processes have potential direct processing sources of NO_x . Lurgi claims to produce negligible NO_x in their shale burner, and adjustment of conditions in the TOSCO ball heater and elutriator would affect a number of other processing parameters. (See Section 5.0). Actual

NO_x levels in the TOSCO design were also found to be low compared to other industries such as the power industry.

Under the more strict scenario for all processes, special burners for boilers are considered, and water injection systems are provided for turbine power generation. Although in the future, a catalytic converter for NO_x in flue gas may be developed, such equipment was not considered for this study since it is not a commercially available item.

For both scenarios for the Paraho, MIS, and MIS/Lurgi processes, ammonia absorption and recovery processes have been added to remove ammonia from the retort gas stream before it is burned. Although the combustion temperature of these low Btu gases might be low enough to allow burning of ammonia without formation of NO_x, the high ammonia content is a potential source of NO_x. The TOSCO II process does not produce ammonia in the retort gas, but rather ammonia is released in the refining operation. An absorption and recovery system is provided for the refinery to prevent ammonia emissions.

4.4.4 Carbon Monoxide

Two levels of control were not considered. In general, the oil shale processes are not significant sources of carbon monoxide. All fuels are burned completely resulting in essentially nil CO. No special controls are required.

4.4.5 Sulfur Dioxide

Less Strict: Removal of 90% of SO₂ from Flue Gas Emissions

More Strict: Removal of 93.5% of SO₂ from Flue Gas Emissions

The existing and proposed regulations for SO₂ emissions are quite complex and the level of control required can be specified in a number of ways. (See Appendix 4.0 for a detailed discussion.) The two levels of control shown above were originally selected by DOE and exceed the current proposed Federal New Source Performance Standards for power generation.

Only one set of controls is considered for this study, and the actual level of control achieved far exceeds both scenarios as discussed

before. The fuel burned by oil shale processes is retort gas. As a means of controlling SO_2 emissions, sulfur can be removed from this gas stream before combustion, using, in all cases considered in this study, the Stretford Process. Assuming all sulfur in the retort gases are in the form of H_2S , 99% removal or better can be expected. However, it is possible to encounter sulfur in other forms such as COS , C_2S , mercaptans and even SO_2 itself. The Stretford Process performs poorly if these species are present, and the liquor may be degraded by SO_2 .

It is necessary to consider critically the composition of the retort gas streams, in part, to determine if sulfur compounds other than H_2S are present. Little agreement was found between different literature sources of information on retort gas composition, and little information has been generated on minor gas components such as COS . All sources of available information were assembled, and new data, never before released, was provided by several developers. Gas compositions are shown in Sections 3.0, 5.0 and 6.0 and discussed in more detail.

It is concluded that COS , C_2S , SO_2 and mercaptan levels appear to be, and should be, low. The basic assumption must be made that the retorting processes are operating properly and maximizing the yield of oil. The gas stream is generated under reducing conditions which would not favor the production or survival of such compounds when retorting is operating properly.

For comparison purposes only, the cost for SO_2 flue gas scrubbers is also estimated and is discussed in Section 5.0. Such scrubbers are a less effective method of control.

4.4.6 Air Emissions of Hazardous Materials

Scenarios: Specific Controls for Hazardous Materials
 are not Necessary

The above scenario was assumed at the beginning of the study as a hypothesis. The hypothesis was not disproven.

The principal hazardous materials of concern in oil shale processing are heavy metals and organic compounds. Trace elements are tracked and their disposition discussed in Section 8.0. In general, they are not expected to

escape the environmental controls for other pollutants, hence specific controls are not required.

In general, organic compounds in the gas streams are either burned or become part of the fossil fuel product. Organics are reduced to low levels in water streams before use in dust suppression or cooling. Hence, emission of organics from water streams is not anticipated. Fugitive emissions of organics are controlled by thermal oxidizers, double sealed floating top tanks, etc.

4.5 SCENARIOS FOR WATER POLLUTION CONTROL

Two categories of waters are encountered in oil shale processing, and these waters are treated in this study in separate systems. First, all processes produce process waters, often called "retort water." This term is actually a misnomer since two distinctly different streams of process water from the retort can be identified. In the past it has been common practice to combine these two streams into one for analysis. In commercial operation, the MIS processes will have both streams, but they will be handled separately, and only one of the streams will be found in the aboveground processes. Sections 3.0 and 6.0 discuss this issue in more detail. The second category of water encountered in oil shale processing is mine or aquifer water. This water may be present in excess of process needs, and hence will require management for disposal.

The laws and regulations which apply to discharge of wastewaters and their application are discussed in detail in Appendix 4.0.

4.5.1 Process Waters

Both Scenarios: Zero Discharge of Process Waters

In designing water management systems for a new process where controls can be completely integrated into the production system, a general rule is applied when water is not produced in excess of process needs. All water is upgraded as necessary for recycle and/or reuse. In oil shale production, process waters do not exceed process needs, hence zero discharge is the outcome of applying a reasonable water management strategy. Such strategy is

discussed in detail for all four processes in Section 6.0, and can be seen to involve upgrading and reuse of other process waters in addition to retort water. Consideration of two regulatory scenarios for these waters becomes a moot point, however, since there is zero discharge.

4.5.2 Excess Mine Water

Mine water in excess of process needs can be managed in two ways. It can either be treated for aboveground discharge, or reinjected into the aquifer. For this study, treatment and aboveground discharge were considered, as they would represent development of a significant and valuable water resource in an arid region as a spin-off of oil shale processing. It is also felt that large volume sustained reinjection might prove to be complex. Two scenarios were considered as specifications which must be met to discharge such water.

Less Strict: Table 4-1

More Strict: Table 4-2

Table 4-1 summarizes the levels for certain criteria pollutants which must be achieved for discharge under the less strict scenario. Section 6.0 summarizes the actual levels achieved under the less strict scenario and the levels assumed in the mine water before treatment. The dissolved solids content is reduced well below the levels specified in the less strict scenario (to <200 ppm) by the treatment necessary to achieve the boron and fluoride specifications.

Table 4-2 summarizes the specifications for the more strict scenario. Again, Section 6.0 summarizes the actual levels achieved, and the levels assumed in the mine water before treatment. The more strict scenario governs the composition of the discharge water by the requirements to achieve such low levels for ammonia and phenol. The TDS level is potentially reduced to as low as 30 mg/l in the process of meeting these specifications. Boron and fluoride are also reduced to well below the requirements in the process.

It is informative to compare the levels of phenol and ammonia found in various water sources in the Piceance Creek Basin area to the levels specified for the more strict scenario. These data are presented in Tables 4-3 and 4-4. In the less strict scenario, no treatment is required for these two

Table 4-1.

LESS STRICT SCENARIO FOR EXCESS MINE WATER TREATMENT

<u>Parameter</u>	<u>Level</u>	<u>Reason</u>
Total Suspended Solids (mg/l)	30	Current Permits
Total Dissolved Solids (mg/l)	723	Colorado River Salinity Standard at Hoover Dam
Total Fluoride (mg/l)	2.0	Drinking Water
Total Boron (mg/l)	1.0	Current Permits
pH	6-9	Current Permits

Table 4-2

MORE STRICT SCENARIO FOR EXCESS MINE WATER TREATMENT

<u>Parameter</u>	<u>Level</u>	<u>Reason</u>
Total Suspended Solids (mg/l)	30	Current Permits
Total Dissolved Solids (mg/l)	500	Public Health Service Drinking Water Standard
Total Fluoride (mg/l)	2.0	Drinking Water
Total Boron (mg/l)	0.75	Agricultural Use
Ammonium-N	0.2 mg/l or 0.2 g/m ³ of Stream Flow	Cold Water Fishery Standard
Phenolic compounds (µg/l) (ppb)	1.0	Threshold for Tainting Fish Flesh

Table 4-3
SUMMARY OF BACKGROUND AMMONIA CONCENTRATION
IN THE PICEANCE CREEK BASIN

Location	Range (mg/l)	Geometric Mean (mg/l)	Number of Samples	Source
Rinky Dink Gulch	.12 - .06	.08	2	(a)
Slake Springs	.14 - .14	.14	1	(a)
Box Elder Gulch	.32 - .01	.02	7	(a)
Corral Gulch-West	.12 - .01	.01	20	(a)
Corral Gulch-East	.35 - .01	.03	30	(a)
Yellow Creek	.17 - < .1	.10	24	(a)
Yellow Creek	.52 - < .01	.03	34	(a)
White River Near Piceance Creek	.14 - < .1	< .1	96	(a)
Alluvial Aquifer, C-a	6.4 - < .1	.36	49	(a)
Upper Aquifer, C-a	1.8 - < .1	.3	(2)	(a)
Various Springs and Seeps, C-b	< .1	< .1	9	(b)
Drilling Effluent, C-a	-	.7	-	(c)
Dewatering Wells, C-a	-	.3	-	(c)
White River at Meeker	0 - .3	< .1	20	(d)
White River at Rangely	0 - .8	.1	22	(d)
Colorado River at New Castle	0 - .5	< .1	34	(d)
Colorado River Near Cameo	0 - .5	< .1	19	(d)
White River Below Piceance Creek	0 - .6	< .1	22	(d)
Upper Aquifer Composite	.3 - 1.2	.8	-	(e)

(a) Final Environmental Baseline Report, Tract C-a.

(b) Detailed Development Plan, C-b Oil Shale Project.

(c) NPDES Permit Application CO-00345045

(d) STORET Data

(e) Average of upper aquifer data assumed for this project.

Table 4-4
SUMMARY OF BACKGROUND PHENOLIC CONCENTRATION
IN THE PICEANCE CREEK BASIN

Location	Range	Geometric Mean	Number of Samples	Source
	(mg/l)	(mg/l)		
Corral Gulch-East	3.0 - 0.0	.05	-	(a)
Box Elder Gulch	3.0 - 3.0	3.0	1	(a)
Corral Gulch-West	3.0 - 1.0	1.7	2	(a)
Yellow Creek	1.0 - 0.9	.03	2	(a)
Alluvial Aquifer, C-a	13 - < 1	2.1	11	(a)
Upper Aquifer, C-a	174 - < 1	2.5	39	(a)
Springs and Seeps, C-b	0 - 37	15	-	(a)
Upper Aquifer Composite	1 - 40	2.5	-	(b)

(a) Final Environmental Baseline Report, Tract C-a.

(b) Average of upper aquifer data assumed for this project.

criteria pollutants. The differential cost for excess mine water treatment between the scenarios represents conceptually the cost of reducing the natural background levels of ammonia by only a small amount. This information can be used to judge, in a cost/benefit sense, the wisdom of requiring such strict regulation.

4.6 SCENARIOS FOR SOLID WASTE MANAGEMENT

Two categories of solid waste are encountered in oil shale processing. First, large quantities of raw (MIS process) or spent shale (Paraho, TOSCO, MIS/Lurgi) must be managed. Second, there are significant quantities of nonshale wastes from various processing and pollution control activities. These categories are considered separately in Section 7.0 in the discussion of solid waste management practices. The scenarios used are summarized in this section and are discussed in much greater detail in Section 7.0. The solid waste management regulations are discussed in detail in Appendix 4.0.

4.6.1 Raw and Spent Shale Waste Disposal Aboveground

The regulatory scenarios used for raw and spent shale management are as follows:

Less Strict: Application of Solid Wastes Disposal Regulations.

More Strict: Application of Hazardous Waste Disposal Regulations.

For the less strict scenarios, a composite of plans proposed by the oil shale developers and the requirements of the federal, state and local solid waste disposal regulations was used to create a model which was applied to all four processes. The developer proposals adopted in the model often represent more strict management than required by regulation of nonhazardous wastes. Application of Hazardous Waste regulations was requested by DOE to be used in developing "worst case" cost estimates. Neither the authors nor EPA feel that there is any probability that raw or spent shale will be classified as hazardous wastes and hence subject to these regulations. In response to the DOE request, a standardized model of a satisfactory hazardous waste disposal area was developed and applied to all four processes. See Section 7.0 for a detailed discussion of both models.

4.6.2 Nonshale Solid Wastes

All significant nonshale hazardous wastes are listed in Section 7.0. The final disposition of such wastes was determined by considering each separately. Only one treatment was developed for each such waste. Treatments used range from recycle, to offsite disposal as a hazardous waste, to management on-site in a separate hazardous waste landfill area, to management as a nonhazardous waste. This is detailed in Section 7.0.

4.6.3 Spent MIS Retort Management

Less Strict: Spent MIS Retorts Abandoned without Treatment.

More Strict: Spent MIS Retorts Grouted to Render Impervious to Leaching.

The question of leaching from spent MIS retorts is viewed by many as perhaps the most critical issue in all of oil shale processing. The cost of grouting spent MIS retorts to prevent leaching is presented in this study. The grouting technique chosen involves slurry backfilling with spent shale, produced by the Lurgi retort, as discussed in Section 3.0. The entire MIS-Lurgi process model can be viewed as the response to the more strict scenario. The basic MIS process model can be viewed as presenting the less strict scenario.

4.7 WORKER HEALTH AND SAFETY

Discussion of Worker Health and Safety is contained in Section 9.0.

APPENDIX 4.0

ENVIRONMENTAL REGULATIONS

This appendix summarizes in tabular form only those acts and regulations which would most affect the design and operation of a full-scale oil shale plant. Requirements which affect the rate and cost of the original development, such as the National Environmental Policy Act, are not included. The development process also requires over 400 permits and approvals which are not discussed here.

A.1 AIR QUALITY REGULATIONS

A.1.1 Particulate Matter

Federal New Source Performance Standards

Petroleum refineries, catalytic cracker	1.0 lb/1000 lb 30% opacity (6 min. exception)
Fossil fuel-fired steam generators	0.1 lb/mBtu
Electric utility steam generating units (proposed)	0.03 lb/mBtu 20% opacity

Colorado Air Quality Regulation No. 1

(Sulfur dioxide, particles and smoke)

All stationary sources	20% opacity
All fuel burning sources	0.5 - 0.1 lb/mBtu, depending on fuel consumption
All manufacturing processes, $\geq 60,000$ lbs/hr	$17.31 (P)^{16}$ where P = production in tons/hr
Fugitive dusts from all sources except those listed in attachment	20% opacity--No dust can be visibly transported off property. Practices such as watering of roads, dust palliatives, and paving may be required.

Proposed Colorado Air Quality Regulation No. 1

(Standards of Performance for new stationary sources)

This revision expands on the controls of fugitive dusts and requires good management practices such as wetting, the addition of dust palliatives, paving, and speed limits for haul roads. Dust is not to be visibly transported off haul roads.

Colorado Air Quality Control Regulation No. 6

(Standards of performance for new stationary sources)

Fossil fuel-fired steam generator	0.1 lb/mBtu
Petroleum refinery, catalyst regeneration	30% opacity 1 lb/1000 lb of coke

Proposed Colorado Air Quality Regulation No. 6

(Standards of performance for new stationary sources)

Fossil fuel-fired steam generator	0.1 lb/mBtu
All nonspecified fuel burning sources	$E = 0.5(Q)^{-0.26}$ where Q = heat input, mBtu/hr E = emission rate, lb/mBtu
1-250 mBtu/hr	

Prevention of Significant Deterioration

(Federal Clean Air Act)

PSD permitted increments for suspended particles ($\mu\text{g}/\text{m}^3$)

	<u>Class I</u>	<u>Class II</u>	<u>Federal NAAQS</u>	<u>State AAQS</u>
Annual Average	5	19	75/60 (secondary)	45
24-Hour Average	10	37	260/150 (secondary)	150

Visibility Impact

Visibility regulations are presently being promulgated but are not yet available. Since suspended particles degrade visibility it is conceivable that visibility regulations may impact the oil shale industry. Models are not sufficiently developed to predict the effect of an oil shale industry on visibility in adjoining areas.

A.1.2 Hydrocarbons

Federal New Source Performance Standards

Petroleum refineries, catalytic cracker	30% opacity (6 min. exception)
Storage vessels for petroleum liquids >40,000 gallon capacity	floating roof for $P > 570$ mm Hg vapor recovery system for $P > 570$ mm

Proposed Federal New Source Performance Standards

Storage vessels for Petroleum liquids, >40,000 gal of capacity >40,000 gal, $P > 78$ mm Hg	Double seals on external floating roof tanks
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Ambient Air Standards (Federal only)

3-hour average (6-9 a.m.)	160 $\mu\text{g}/\text{m}^3$
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Clean Air Act and Amendments

The clean air act requires the best available control technology (BACT). Lowest achievable emission rates (LAER) may be required for nonattainment areas.

This act also requires the EPA to promulgate regulations by August 1979; to prevent the significant deterioration in air quality (PSD) due to hydrocarbons regulations not yet available.

Colorado Air Quality Control Regulation No. 7

(Emissions of Hydrocarbon Vapors)

Regulations for hydrocarbon emissions vary among air pollution control areas, and increase in severity with increasing levels of ambient ozone. Regulations include:

Petroleum distillate and crude oil storage	Pressurized tanks, seals on floating roof tanks, vapor gathering system
Petroleum distillate transfer	Vapor recovery and sealing
Disposal of solvents	No more than one quart of photochemically reactive solvent may be disposed of in a manner which permits its evaporation into the air

In general, the regulations require good management practices.

Proposed Revisions to Colorado Air Quality Regulation No. 7

(Emissions of Hydrocarbon Vapor)

Transfer of petroleum distillate	80 mg of evaporation per liter of distillate
Volatile organic compounds from any source except:	15 lbs/day of photochemically reactive compounds or 3,000 lbs/day of organic compounds 20% of which are volatile. If a chemical oxidizer is used at least 90% of the volatile organic compounds must be oxidized CO ₂
(a) those sources controlling 85% of their hydrocarbon emissions	
(b) methane or ethane	
(c) transfer of petroleum distillates	
Petroleum Refineries	Process unit blowdowns and turn-arounds, pressure relief valves, vacuum equipment shall be vented to a smokeless flare or other combustion source

A.1.3 Nitrogen Oxides

Federal New Source Performance Standards

Coal-fired steam generator, >250 mBtu/hr	0.7 lb/mBtu
Gas-fired steam generator, >250 mBtu/hr	0.2 lb/mBtu
Stationary gas turbine, fuel >.15% N (proposed)	75 ppm @ 15% O ₂
Utility steam boiler fired with bituminous coal, shale oil, or fuels derived from coal (proposed)	0.5 lb/mBtu
Utility steam boiler fired with bituminous coal	0.6 lb/mBtu

Ambient Air Standards (as NO₂)

Annual average 100 µg/m³

Existing State and Federal ambient air standards are identical for NO₂. Neither has yet promulgated significant deterioration standards.

Clean Air Act and Amendments

Prevention of Significant Deterioration. The EPA is required to promulgate PSD regulations for NO_x by August 1979. Regulations are not yet available.

Nonattainment Factors. The oil shale region is occasionally over the 0.12 ppm ambient air standard for ozone. This fact may require strict NO_x controls.

Technology Requirements. The best available control technology (BACT) will be required. Lowest achievable emission rates (LAER) may be required if the area is interpreted to be in noncompliance for ozone.

Colorado Air Quality Control Regulation No. 6

(New Source Performance Standards)

No standards are listed specifically for oil shale. However the following standards may be relevant.

Gas fuel-fired steam generator	0.2 lb/mBtu, expressed as NO_2
Liquid fuel-fired steam generator	0.3 lb/mBtu, expressed as NO_2
Solid fuel-fired steam generator	0.7 lb/mBtu, expressed as NO_2

Proposed Revisions to Colorado Air Quality Control Regulation No. 6

(New Source Performance Standards)

NO_x emissions from shale oil processes and oil refineries are not addressed.

A.1.4 Hazardous Materials

Regulations

Clean Air Act

The Clean Air Act requires that standards for hazardous emissions, such as mercury, be set irrespective of cost or technical feasibility.

Colorado Air Quality Control Regulation No. 8
(Chemical Substances and Physical Agents)

- a. This regulation limits the emissions of those chemicals listed by the American Conference of Governmental Industrial Hygienists. Permissible emission rates depend on stack height as well as the threshold limit value.
- b. Emission Standard for Mercury.
- c. Intentions to regulate carcinogens are stated.

Proposed Colorado Air Quality Control Regulations No. 8
(Hazardous Air Pollutants)

- a. Lead. No stationary source shall emit lead in amounts sufficient to produce a concentration of $1.5 \mu\text{g}/\text{m}^3$ at a receptor site.
- b. Hydrogen sulfide. No stationary source shall emit hydrogen sulfide in amounts sufficient to produce a concentration of $75 \mu\text{g}/\text{m}^3$ (0.5 ppm) at a receptor site.

A.1.5 Carbon Monoxide

Colorado Air Quality Control Regulation No. 6
(Standards of Performance for New Stationary Sources)

Petroleum refinery, fluid cracking unit	0.05%
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Standards of Performance for New Stationary Sources
(Federal)

Petroleum refinery, fluid catalytic cracking unit	0.05%
---	-------

Ambient Air Standards

8-hour average	10 mg/m^3
1-hour average	40 mg/m^3

Existing Colorado and Federal standards are identical, neither has yet promulgated significant deterioration standards.

A.1.6 Sulfur Dioxide

Colorado Air Quality Control Regulation No. 1

0.3 lbs/bbl of shale oil produced

0.3 lbs/bbl of shale oil refined

After 1985 all sources must emit less than 5 tons per day of sulfur dioxide, and all vented waste gases must contain less than 500 ppm sulfur dioxide.

Proposed Revisions to Colorado Air Quality Regulation No. 1

No standards specific to oil shale are proposed. However, the following regulations are proposed:

Coal-fired operations	<250 MBtu/hr	1.8 lbs/MBtu
	≥250 MBtu/hr	1.2 lbs/MBtu
Combustion turbines	<250 MBtu/hr	1.3 lbs/MBtu
	≥250 MBtu/hr	0.8 lbs/MBtu
Petroleum refining		0.7 lbs/bbl of oil processed
Natural gas desulfurization; SO ₂ emissions ≥3 tons/day		No standard
Natural gas desulfurization; SO ₂ emissions ≥ tons/day		2.0 lbs/1000 cubic feet of delivered gas
All sources emitting more than three tons of sulfur dioxide per day		Best practical control technology

Colorado Air Quality Control Regulation No. 6

(New source performance standards)

Fuel with more than 230 mg/dscm of sulfur dioxide may not be burned in an oil refinery unless it is followed by a device to remove the resulting sulfur dioxide.

Proposed Revisions to Colorado Air Quality Regulation No. 6

(New source performance standards)

The proposed changes incorporate the standards for shale oil production and refining which are presently listed under regulation No. 1.

Liquid fuel-fired steam generator	0.3 lbs/mBtu
Solid fuel-fired steam generator	0.4 lbs/mBtu
Petroleum refinery	0.3 lbs/bbl of oil processed
Claus recovery plant	0.025% SO ₂ or 0.030% reduced S and 0.001% H ₂ S

Federal New Source Performance Standards

Oil refinery, fuel gas combustion	230 mg/dscm fuel gas content, unless controlled
Coal-fired steam generator, >250 mBtu/hr	1.2 lb/mBtu, removal of 70-90% depending on S content of coal
Oil-fired steam generator, >250 mBtu/hr	0.8 lb/mBtu
Fluid catalytic cracking units catalyst regenerator	0.025% SO ₂
Fuel gas combustion devices	
Claus sulfur recovery plants	
Stationary gas turbines (proposed)	150 ppm exhaust, <.8% S in fuel
Electric utility steam generating units, >250 mBtu/hr (proposed)	1.2 lbs/mBtu for solid fuel 0.8 lbs/mBtu for gaseous and liquid fuel plus 85% reduction of uncontrolled emissions--exception: 0.2 lbs/mBtu

Ambient Air Quality Standards

	PSD Permitted Increments (µg/m ³)				Ambient Air Quality Standards	
	Federal		Colorado		Federal	Colorado
	I	II	I	II		
Annual arithmetic mean	2	20	2	10	80	
24-hour maximum	8	91*	5	50	365	
3-hour maximum	25	512*	25	300	1300(Δ)	700

Δ = secondary standard

Visibility Impacts

Visibility regulations are presently being promulgated but are not yet available. Since sulfate aerosols have been implicated in visibility degradation, visibility regulations may limit sulfur dioxide emissions.

Control Technology

The best available control technology (BACT) will be required as a minimum.

A.2 WATER POLLUTANTS

Regulations for Implementation of the Colorado River Salinity Standards (Colorado Department of Health)

Objective: No salt return wherever practicable.

Discharges of salt may be permitted from new sources only if it is not practical, as determined by the Public Health Department, to prevent such discharges. In order to discharge salt into the Colorado River Basin, the industry must show that other alternatives are not practical. In permitting a discharge, the Public Health Department must consider "the impact of the total proposed salt discharge . . . on the lower mainstream, in terms of both tons per year and concentration, and the economic impact of the increased salt loads."

Federal Clean Water Act of 1977 and the Colorado Water Quality Control Act

These Acts establish three categories of pollutants: toxic, conventional, and nonconventional. A list of those substances presently listed as toxic under the Act is attached (Table A-1). Conventional pollutants will require "Best Practical Technology," (BPT) and other pollutants will require Best Available Technology (BAT). However, unlike toxic pollutants, nonconventional pollutants can be granted variances based upon economic considerations. In addition, the regulations can also require "best management practices." Effluent limitations for toxic pollutants relevant to the oil shale industry have not yet been set.

In addition, the Act provides a mechanism for setting ambient stream standards. In Colorado, ambient stream standards are set by the State. (See

attached Table A-2). In the event that BAT will not preserve the stream standards, stricter controls may be required.

Safe Drinking Water Act of 1974

This Act is primarily aimed at assuring the quality of drinking water. However, under this act the EPA must develop regulations for underground injection.

Rules for Subsurface Disposal Systems

(Colorado Department of Health)

Subsurface disposal systems may be operated only if the Water Quality Control Commission finds "beyond a reasonable doubt" that "no waters of the State will be polluted thereby," or that the resulting pollution will be "limited to waters in a specified limited area from which there is no risk of significant migration."

Regulations and Existing Surface Water Quality. In considering possible regulatory scenarios it is important to include water quality criteria that have been adopted by the State of Colorado. Table A-2 thus includes such criteria for a warm water aquatic life, agricultural uses, and Class 2 water supplies. While other classifications are also applied as can be seen in the fourth column, it is assumed that these three classifications include those most relevant to the oil shale region. It is important however to realize that classifications of streams throughout the oil shale area in Colorado do vary. As an example, Occidental's discharge permit No. CO-0033961 is based on the assumption that the receiving waters of Piceance Creek are a cold water fishery. On the other hand, the requirements for Rio Blanco's permit to discharge into Corral Gulch and Yellow Creek is based primarily on agricultural use. It is also important to realize that the criteria shown in Table A-2 differ from fixed standards. In translating criteria to standards the natural quality of the undisturbed stream can also be considered. The criteria shown in Table A-2 can be considered as presumptive standards for the purposes of this study.

It should also be noted that many of the virgin streams in the oil shale areas already exceed the criteria shown in Table A-2. Fox (1977) has

investigated water quality in the mainstream of the White River and its tributaries. In the White River, which represents drainage of the Piceance Creek and Yellow Creek as well as numerous other tributaries, values exceeded the criteria for TDS, TSS, aluminum, iron, copper, manganese, zinc, and phosphorus. Yellow Creek exceeded criteria for dissolved oxygen, alkalinity, TDS, TSS, sodium, iron, molybdenum, selenium, boron, and fluoride. Table A-3 also shows water quality from Corral Gulch, Piceance Creek and Black Sulfur Creek. The geometric mean values of sulfide, sulfate, and phenols all exceed the water quality criteria. Table A-4 indicates the extent of variation and water composition during high and low flows. The values of ammonia, arsenic, chromium, fluoride, mercury, phenolics, sulfate and sulfide all exceed the criteria during low water periods in Corral Gulch.

As these data suggest, setting standards for effluents in this region is further complicated by widely varying stream flows. Weeks et al. (1976) for example, have shown that during the period of 1941 to 1943 the mean monthly runoff through the Piceance Creek varied from 30 to approximately 1,500 acre-feet per month. The flow rate of this creek is less than 1 cfs 1% of the time.

The other major regulatory aspect to be considered the regulations for the implementations of the Colorado River Salinity Standards. This set of standards is discussed above. However, it should be added that the goal of these standards is to assure that the salinity (TDS) as measured at Hoover Dam is less than 723 $\mu\text{g/l}$. Zero discharge of dissolved solids is preferred whenever possible.

Wastewater composition. The next factor to be considered is the composition of the water which could be discharged. Table A-3 summarizes water quality expected from dewatering wells in the Rio Blanco project. Values approaching or exceeding the criteria can be seen for aluminum, ammonia, arsenic, boron, cadmium, fluoride, iron, lead, manganese, mercury, silver, sulfate and sulfide. Table A-5 from the Occidental area shows concentrations in groundwaters which exceed water quality criteria. Table A-5 also shows constituents in tract C-b (Occidental project) which exceed the most restrictive water quality criteria. Most noticeable are the relatively high values of boron and fluoride in the lower aquifer. The wide variability and

the relatively cleaner nature of the upper aquifer should also be noted. Table A-6 shows the range of concentrations found in retort waters produced under various conditions. While most groundwaters and mine drainage waters are similar to surface waters in the same area, the retort waters are distinctly contaminated.

In view of these uncertainties it is difficult to ascertain what effluent limitations the regulations will eventually require. The nondegradation requirement of the clean air act creates a similar situation. The question "How should natural nonattainment areas be treated?" is yet to be answered.

Current Rationale. In this regard, it is instructive to consider the rationale used by the Colorado State Department of Public Health in issuing NPDES permits to the Rio Blanco Company and to Occidental Oil Company. A summary of these permit requirements are shown in Table A-7. The Rio Blanco oil shale company originally intended to re-inject most of their dewatering waters into the same aquifer from which they were withdrawn; the NPDES permit was requested as an interim device until the re-injection system was in place and also as insurance against unexpectedly large volumes of water. It is designed to allow Rio Blanco to begin the modular testing stage of their process and is not intended for full scale operation. Because of the low flows of the receiving water (Yellow Creek) no dilution was taken into account. Downstream water usage consisting of irrigation, livestock watering and domestic wells was considered. The total suspended solids limit was based on state effluent standards and the short duration of the project. The total dissolved solids limit is the level for livestock without physiological upset and is a background concentration for Yellow Creek at its mouth. The one ton per day limit is included to provide exception to the Colorado River salinity standards. The fluoride requirement is based on limits for livestock and for drinking water. It is also similar to the background concentration in Yellow Creek. The boron limit is based on the sensitivity of tolerant crops. Oil and grease and pH standards are based on state effluent standards.

A similar approach was used in the Occidental permit. Requirements for this permit are essentially the same except for the addition of ammonia and phenol limitations. In this case, Piceance Creek was considered a cold water fishery which required limitations of ammonia-nitrogen of 0.197 $\mu\text{g/l}$ (at

a pH of 8.6) and phenol limits of 0.001 µg/l. Alternately the developer can risk regulating his discharge rate in accordance with the flow of Piceance Creek at rates of 0.2 grams of ammonia-nitrogen per kilometer of water and 1.0 milligrams of phenol compounds per cubic meter of flow in Piceance Creek. Occidental's permit suggests that they must either remove phenolic compounds and ammonia or impound dewatering waste so that they can be released proportional to the flow of Piceance Creek.

Due to the types of pollutants found (e.g., Table A-1) in retort waters, it is likely that zero discharge of pollutants will be required. This may be simply achieved since the retorting process is typically a net consumer of water.

A.3 RECLAMATION AND SOLID WASTE DISPOSAL

Colorado Mined Land Reclamation Act

This Act creates the Colorado Mined Land Reclamation Board and provides it with standard making authority. It establishes permit requirements for mining and requires a surety until reclamation is complete.

Rules and Regulations of the Colorado Mined Land Reclamation Board, July 1978

- a. Reclamation is required on all affected land.
- b. The affected land shall be regraded and all toxic and acid-forming materials will be disposed of in a manner "that will control unsightliness and protect the drainage system."
- c. Disturbances to the groundwater system will be minimized.
- d. Wildlife will be protected.
- e. Topsoil will be segregated, stored, returned, and if necessary, fertilized.
- f. Where revegetation is part of the reclamation plan, land shall be revegetated with self regenerative cover at least equal in extent to the original cover. Noxious weeds shall be controlled.
- g. Fire lanes and access roads will be constructed whenever necessary.

- h. Surety bonds or an equivalent security will be posted by the operator on the amount required by the Board.

Rules for Subsurface Disposal Systems

According to a recent interpretation by the Attorney General of Colorado, this act applies to spent shale left by in situ retorts. Thus, the operators must show that:

- a. "No water of the State will be polluted thereby," or
- b. the "pollution resulting therefrom will be limited to waters in a specified limited area from which there is not risk of significant migration. . . ."

Federal and State Water Pollution Regulations

(See Water Section)

These require that surface waters be protected against toxic runoff from storage piles. Occidental's permit requires that dams and impoundments be adequate to retain rainfall from a 25-year, 24-hour precipitation event of 2.1 inches.

Guidelines and Criteria for Review of Solid Waste Disposal Facilities

(Colorado State Department of Public Health)

These regulations set minimal requirements for the operation of solid waste disposal sites. These include that:

- a. Noxious odors and windblown debris must be controlled
- b. Design and operation shall include adequate drainage and shall prevent the pollution of surface and groundwater.
- c. Solid wastes shall be placed in the most dense volume practical.
- d. Surface waters shall be directed from the disposal site.
- e. When dry arroyos are selected as disposal sites, the design must provide a positive means of preventing washing of the solid wastes downstream in periods of runoff based on the 50-year peak discharge.

Resource Conservation and Recovery Act

This act requires the EPA to promulgate regulations for the treatment, storage, and disposal of solid wastes. Solid wastes are to be classified as normal, hazardous or special (e.g., most mining wastes).

Solid Waste Disposal Facilities (FR, 43, 4942 (1978))

- a. Solid waste disposal sites cannot be located in a flood plain, in recharge zone of a sole-source aquifer, nor in the critical habitat so as to jeopardize an endangered species.
- b. Nonpoint source discharges must be minimized, and point sources must comply with NPDES regulations.
- c. leachate shall be collected and treated.
- d. Monitoring of groundwater may be required.
- e. Access to facility must be controlled.

Proposed Hazardous Waste Disposal Regulations Pursuant to RCRA (FR, 43, 59013 [1978])

- a. A generator must prepare an annual report of all hazardous wastes generated, treated, stored, and disposed.
- b. A hazardous waste dump must comply with regulations for nonhazardous solid wastes.
- c. Diversion structures must be present for a 24-hour 25-year storm.
- d. Detailed waste and leachate sampling and analysis is required.
- e. Emergency equipment must be provided.
- f. Daily inspection is required.
- g. Closure shall be secured after complete to prevent access by humans or animals and to prevent discharges.
- h. Post-closure care must be provided for 20 years, including monitoring maintenance, and restructure access.
- i. Financial assurances must be provided for closure, post-closure, and accidents.
- j. Landfill design is carefully specified.
- k. The leachate sump must be adequate to contain at least three months leachate.

Table A-1
Toxic Pollutants

1. Acenaphthene
2. Acrolein
3. Acrylonitrile
4. Aldrin/Dieldrin
5. Antimony and compounds¹
6. Arsenic and compounds
7. Asbestos
8. Benzene
9. Benzidine
10. Beryllium and compounds
11. Cadmium and compounds
12. Carbon tetrachloride
13. Chlordane
14. Chlorinated benzenes
15. Chlorinated ethanes
16. Chloralkyl ethers
17. Chlorinated naphthalene
18. Chlorinated phenols
19. Chloroform
20. 2-chlorophenol
21. Chromium and compounds
22. Copper and compounds
23. Cyanides
24. DDT and metabolites
25. Dichlorobenzenes
26. Dichlorobenzidine
27. Dichloroethylenes
28. 2,4-dichlorophenol
29. Dichloropropane and dichloropropene
30. 2,4-dimethylphenol
31. Dinitrotoluene
32. Diphenylhydrazine
33. Endosulfan and metabolites
34. Endrin and metabolites
35. Ethylbenzene
36. Fluoranthene
37. Haloethers
38. Halomethanes
39. Heptachlor and metabolites
40. Hexachlorobutadiene
41. Hexachlorocyclohexane
42. Hexachlorocyclopentadiene
43. Isophorone
44. Lead and compounds
45. Mercury compounds
46. Naphthalene
47. Nickel and compounds
48. Nitrobenzene

Table A-1 (continued)

49. Nitrophenols
50. Nitrosamines
51. Pentachlorophenol
52. Phenol
53. Phthalate esters
54. Polychlorinated byphenyls
55. Polynuclear aromatic hydrocarbons
56. Selenium and compounds
57. Silver and compounds
58. 2,3,7,8 Tetrachlorodibenzo p dioxin
59. Tetrachloroethylene
60. Thallium and compounds
61. Toluene
62. Toxaphene
63. Trichloroethylene

¹The term "compounds" shall include organic and inorganic compounds.

Table A-2
Colorado Water Quality Criteria

Parameter	Aquatic Life Warm Water Biota	Agriculture	Water Supply Class 2	Strictest Classification
pH	6.5-9.0	X	5.0-9.0	6.5-9.0
Suspended Solids (mg/l)	X	X	X	X
TDS (mg/l)	Y	Y	Y	Y
Al (mg/l)	0.1(c)	X	X	0.1
As (mg/l)	0.05	.1	0.05	0.05
Ba (mg/l)	X	X	1.0	1.0
Be (mg/l)	1.1(a)	0.1	X	0.01
Cd (mg/l)	0.01(a)	0.01	0.01	0.0004
Cr (mg/l)	0.1	0.1	0.05	0.05
Cu (mg/l)	0.04(a)	0.2	1.0	0.01
Fe (mg/l)	1.0	X	0.3(f)	0.3)
Pb (mg/l)	0.15(a)	0.1	0.05	0.004
Mn (mg/l)	1.0	0.2	0.05	0.05
Hg (mg/l)	0.00005	X	0.002	0.00005
Mo (mg/l)	X	Y(c)	Y	Y
Ni (mg/l)	0.4(a)	0.2	X	0.05
Se (mg/l)	0.05	0.02	0.01	0.01
Ag (mg/l)	0.00025(a)	X	0.05	0.0001
Tl (mg/l)	0.015	X	X	0.015
U(b) (mg/l)	1.4(a)	2.0	5.0	0.05
Zn (mg/l)	0.60(a)	2.0	5.0	0.05

(c) Assigned on a case by case basis depending on the affected agriculture

(b) Toxic properties to aquatic life

(a) For water hardness $\geq$ 400 mg/l. Criteria vary with water hardness

Table A-2 (continued)
Colorado Water Quality Criteria

<u>Parameter</u>	<u>Aquatic Life Warm Water Biota</u>	<u>Agriculture</u>	<u>Water Supply Class 2</u>	<u>Strictest Classification</u>
Ammonia (mg/l as N)	0.10 unionized	X	0.5	0.02
Cyanide (mg/l)	0.005	0.2	0.2	0.005
Fluoride (mg/l)	X	X	1.4-2.4(d)	(2)
Nitrate (mg/l as N)	X	100	10	10
Nitrite (mg/l as N)	.5	10	1.0	0.05
Sulfide (mg/l)	0.002 undissociated	X	0.05	undissociated
B (mg/l)	X	0.75	X	0.75
Chloride (mg/l)	X	X	250	250
Mg (mg/l)	X	X	125	125
Sulfate (mg/l)	X	X	250	250
Phosphorus (mg/l)	Bioassay (e)	X	Bioassay (e)	-
Phenol (mg/l)	0.001	Y	0.001	0.001
Total alpha (pc/l) (excluding U & Rn)	15	15	15	15
Total beta (pc/l) excluding Sr-90	50	50	50	50
Cs-134 (pc/l)	80	80	80	80
Ra-226 & 228	5	5	5	5
Sr-90	8	8	8	8
Th-230 & 232	60	60	60	60
H-3	20,000	20,000	20,000	20,000
U-total	-	5	5	5

(e) Algal bioassay

(d) Depends on temperature. See National Interim Primary Drinking Water Regulations

X Numerical limits are generally not needed but may be set by the Division. Basic Standards apply.

Y Numerical values may be required in certain cases, but there is insufficient data for setting general criteria. Basic Standards apply.

Basic Standards:

Substances shall not be discharged which

- (a) form detrimental deposits
- (b) form floating debris or scum
- (c) produce disagreeable color, odor or other conditions
- (d) are harmful to the classified uses

TABLE A-3

DISCHARGE AND STREAM WATER QUALITY SUMMARY (RAW WATER)
(All values listed are geometric mean)

Param. No.	Item	1 Dewatering Wells	2 Drilling Corral Effluent	3 Gulch E. of Tract	4 Piceance Creek	5 Black Sulphur
410	Alkalinity (mg/l)	371	263	404	532	482
1106	Aluminum (mg/l)		0.15	0.02		0.01
71846	Ammonia (NH ₃) (mg/l)	0.3	0.70	0.03		
1000	Arsenic (ug/l)	<10.1	14.2	5.3		0
1005	Barium (ug/l)	<1002		<74		0
1010	Beryllium (ug/l)			<3.3		
440	Bicarbonate (mg/l)	429	455	492	649	588
1015	Bismuth (ug/l)			<10		
310	BOD (mg/l)			2.5		
1020	Boron (ug/l)	258	470	117		
71870	Bromide (mg/l)			0.10		0.05
1025	Cadmium (ug/l)	<10	10.7	3.4		
915	Calcium (mg/l)	54	70	86	79	120
445	Carbonate (mg/l)	<1.8	4.6	0	0	0
940	Chloride (mg/l)	10.5	23.8	9.8	16.1	11
1030	Chromium (ug/l)	<11	31	2.8		
340	COD (mg/l)	32		13.6		
31615	Coliform, Fecal (Col/100 ml)			2.1		
31501	Coliform, Total (Col/100 ml)			323		
80	Color (PCU)			4.5		
95	Conductivity, Spec. (umhos)	1227	1148	1231		1680
1040	Copper (ug/l)	<63	<100	2.2		
720	Cyanide (mg/l)	<0.01		0.01		
300	Dissolved Oxygen (mg/l)			8.3		
950	Fluoride (mg/l)	0.69	1.15	0.49	0.78	0.6
1120	Gallium (ug/l)			<4.6		
1125	Germanium (ug/l)		10.7	<12.4		
900	Hardness (Ca, Mg) (mg/l)	371	393	476	538	
1046	Iron (ug/l)	1708	95	31	41	6900
625	Kjeldahl Nitrogen (mg/l)				0.4	
1049	Lead (ug/l)	141	55	1.9		<50

TABLE A-3 (Continued)

Param. No.	Item	1 Dewatering Wells	2 Drilling Effluent	3 Corral Gulch E. of Tract	4 Piceance Creek	5 Black Sulphur
1133	Lithium (ug/l)	118		11.6		10
925	Magnesium (mg/l)	51	53	63	82	95
1055	Manganese (ug/l)	146	57	17	38	290
38253	MSAS (mg/l)			0.55		
71893	Mercury (ug/l)	3.5	<10	0.02		
1060	Molybdenum (ug/l)			37		
1065	Nickel (ug/l)	18.5		8.6		
71951	Nitrate (NO ₃) (mg/l)	0.26	0.14	1.22		
613	Nitrogen - Nitrite as N (mg/l)	0.07		0.02		
631	Nitrite Plus Nitrate (N) (mg/l)			0.32	0.27	0.66
603	Nitrogen, Ammonia (mg/l)			0.03		
607	Nitrogen, Organic (mg/l)			0.29		
1330	Odor (severity)			0		
550	Oil and Grease (mg/l)			2.9		
	Organic Carbon (mg/l)					
681	Dissolved	10.1		8.1		
689	Suspended			0.4		
650	Total			11.4		
	Pesticides (ug/l)			0		
400	pH*	6.58	7.26	7.83	8.0	
32730	Phenols (ug/l)	<3.66		0.05		
	Phosphate (mg/l)					
600	Dissolved (PO ₄)			0.06		
70507	Ortho (P)	<0.10	0.32	0.02	0.03	0.04
665	Total (P)			0.03		
935	Potassium (mg/l)	2.2	3.9	1.6	3.3	2.8
	Radioactivity					
1503	Gross Alpha (pc/l)#	2.9	4.2	<6.6		
9503	Radium 226 (pc/l)			.07		
80030	Natural Uranium (ug/l)			<9.7		
3501	Gross Beta (pc/l)					
80050	SR ⁹⁰ (pc/l)			3.1		
3515	CE ¹³⁷ (pc/l)			3.0		
1145	Selenium (ug/l)		<10	3.0		

* Normal mean and standard deviation

Calculated from gross alpha as natural uranium

TABLE A-3 (Concluded)

Param. No.	Item	1 Dewatering Wells	2 Drilling Effluent	3 Corral Gulch E. of Tract	4 Piceance Creek	5 Black Sulphur
955	Silica (SiO ₂)	24.2	29.4	21.1	17.6	20
1075	Silver (ug/l)	14.4		<1.5	34	
930	Sodium (mg/l)	180.3	189.4	109.2	184	150
931	Sodium Adsorption Ratio			2.2		
540	Solids, Dissolved (mg/l)	874	994	823	1086	1210
1080	Strontium (ug/l)			1685		3800
945	Sulfate (mg/l)	300	326	289	379	520
746	Sulfide (mg/l)	1.0	1.0	0.06		
11	Temperature (°F)*	51.7	65.1	52.6	29.7	
1100	Tin (ug/l)			<9.7		
1150	Titanium (ug/l)			6.0		
70	Turbidity (JTU)			8.2		
1085	Vanadium (ug/l)			6.8		
1090	Zinc (ug/l)	298	408	12.5		
1160	Zirconium (ug/l)			<18.4		

* Normal mean and standard deviation.

TABLE A-4
 BASELINE SURFACE WATER QUALITY SUMMARY
 CORRAL GULCH EAST OF TRACT C-a

Param. No.	Item	USGS DATA					NUS DATA				
		High	Low	Geo. Mean	Geo. Dev.	No.of Analy.	High	Low	Geo. Mean	Geo. Dev.	No.of Analy.
410	Alkalinity (mg/l)	739.00	268.00	403.86	1.08	58					
1106	Aluminum (ug/l)	100.00	10.00	24.28	2.12	11					
71845	Ammonia (NH ₄) (mg/l)	0.35	0.01	0.03	3.07	30	<0.10	<0.01	<0.09	1.60	24
1000	Arsenic (ug/l)	14.00	1.00	5.33	1.29	50					
1005	Barium (ug/l)	<200.00	<50.00	<74.09	1.81	45					
1010	Beryllium (ug/l)	<7.00	<2.00	<3.32	1.67	7					
440	Bicarbonate (mg/l)	901.00	327.00	492.00	1.08	58					
1015	Bismuth (ug/l)	<25.00	<5.00	<10.38	1.92	7					
310	BOD (mg/l)	2.60	2.00	2.25	1.14	3					
1020	Boron (ug/l)	670.00	60.00	116.78	1.22	53					
71870	Bromide (mg/l)	0.10	0.10	0.10	1.0	3					
1025	Cadmium (ug/l)	<60.00	1.00	3.39	3.33	48					
915	Calcium (mg/l)	110.00	55.00	85.56	1.07	58	88.00	54.00	78.91	1.13	24
445	Carbonate (mg/l)	0	0	0	1.00	49					
940	Chloride (mg/l)	19.00	5.90	9.78	1.18	58	25.00	8.00	10.41	1.25	24
1030	Chromium (ug/l)	50.00	<1.00	2.79	3.51	27					
340	COD (mg/l)	94.00	5.00	13.61	2.60	8					
31616	Coliform, Fecal (Col/100 ml)	220.00	0	2.07	89.66	6					
74050	Coliform, Total (Col/100 ml)	580.00	180.00	323.11	2.29	2					
80	Color (PCU)	17.00	0	4.51	2.22	32	25.00	2.50	10.52	2.41	24
95	Conductivity, Spec. (umhos)	1725.00	797.00	1231.54	1.07	59					
1040	Copper (ug/l)	13.00	1.00	2.15	3.02	46					
720	Cyanide (mg/l)	0.02	<0.02	0.01	1.21	25					
300	Dissolved Oxygen (mg/l)	13.00	4.80	8.32	1.25	53					
950	Fluoride (mg/l)	10.00	0.20	0.49	2.02	58	0.40	0.40	0.40	1.00	4
1120	Gallium (ug/l)	<8.00	<2.00	<4.58	1.73	7					

TABLE A-4 (Continued)
 BASELINE SURFACE WATER QUALITY SUMMARY
 CORRAL GULCH EAST OF TRACT C-a

Param. No.	Item	USGS DATA					NUS DATA				
		High	Low	Geo. Mean	Geo. Dev.	No. of Analy.	High	Low	Geo. Mean	Geo. Dev.	No. of Analy.
1125	Germanium (ug/l)	<40.00	<6.00	<12.39	2.13	7					
900	Hardness (Ca, Mg) (mg/l)	520.00	360.00	476.06	1.08	58	508.00	358.00	445.00	1.10	24
1046	Iron (ug/l)	240.00	10.00	31.25	2.35	55	<20.00	<20.00	<20.00	1.00	4
625	Kjeldahl Nitrogen (mg/l)	2.40	0.14	0.42	2.09	21					
1049	Lead (ug/l)	18.00	<1.00	1.90	2.09	46					
1130	Lithium (ug/l)	40.00	<5.00	11.59	2.35	46					
925	Magnesium (mg/l)	71.00	45.00	62.04	1.10	58	80.00	43.00	59.90	1.17	24
1055	Manganese (ug/l)	70.00	5.00	16.83	1.99	52					
38260	MBAS (mg/l)	3.00	0.10	0.55	11.08	2					
71890	Mercury (ug/l)	1.30	<0.01	0.02	3.50	46					
1060	Molybdenum (ug/l)	56.00	23.00	37.55	1.40	8					
1065	Nickel (ug/l)	<20.00	<4.00	<8.59	1.92	7					
71851	Nitrate (NO ₃) (mg/l)	3.00	0.93	1.22	1.14	20	1.00	0.20	0.28	1.75	24
71856	Nitrite (NO ₂) (mg/l)	0.10	<0.01	0.02	2.86	21	0.02	<0.01	0.01	1.15	24
631	Nitrite Plus Nitrate (N) (mg/l)	2.00	0.02	0.32	1.59	58					
603	Nitrogen, Ammonia (mg/l)	0.27	0.01	0.03	2.76	31					
607	Nitrogen, Organic (mg/l)	0.89	0.07	0.29	1.99	9					
1330	Odor (Severity)	3.00	0	0	8.31	57	0	0	0	1.0	24
550	Oil & Grease (mg/l)	16.00	<0.10	2.90	3.04	24					
	Organic Carbon (mg/l)										
631	Dissolved	14.00	1.70	8.12	1.73	10	91.00	<1.00	12.41	2.52	48
689	Suspended	8.70	0.30	0.35	1.40	10					
680	Total	15.00	9.80	11.37	1.27	3					
	Pesticides (ug/l)	0	0	0	1.00	6					
400	pH ⁴	8.90	7.10	7.88	0.31	58					
32730	Phenols (mg/l)	3.00	0	0.05	287.50	2					
	Phosphate (mg/l)										
660	Dissolved (PO ₄)	0.40	<0.01	0.06	2.11	58	0.04	0.01	0.03	1.46	24
671	Ortho (P)	0.13	<0.01	0.02	2.06	58					
665	Total (P)	0.06	0.01	0.03	2.02	31	2.10	0.03	0.04	1.55	24
935	Potassium (mg/l)	3.00	0.70	1.59	1.22	58	4.40	1.70	2.18	1.19	24

TABLE A-4 (Concluded)
 BASELINE SURFACE WATER QUALITY SUMMARY
 CORRAL GULCH EAST OF TRACT C-a

Param. No.	Item	USGS DATA					NUS DATA				
		High	Low	Geo. Mean	Geo. Dev.	No.of Analy.	High	Low	Geo. Mean	Geo. Dev.	No.of Analy.
	Radioactivity										
	Gross Alpha (pc/l)#	<8.84	<4.56	<6.60	1.29	7					
9511	Radium 226 (pc/l)	0.10	0.05	0.07	1.34	4					
80030	Natural Uranium (ug/l)	<13.00	<6.70	<9.70	1.29	7					
	Gross Beta (pc/l)										
80050	SR ⁹⁰ (pc/l)	7.50	2.00	3.10	1.59	7					
3515	CE ¹³⁷ (pc/l)	9.40	2.40	3.83	1.61	7					
1145	Selenium (ug/l)	8.00	1.00	2.96	1.39	45					
955	Silica (SiO ₂) (mg/l)	24.00	13.00	21.10	1.07	58	21.83	9.20	18.28	1.30	24
1075	Silver (ug/l)	<2.00	<1.00	<1.49	1.45	7					
930	Sodium (mg/l)	300.00	51.00	109.15	1.10	58	350.00	91.00	108.98	1.15	24
931	Sodium Adsorption Ratio	6.80	1.20	2.15	1.10	58					
70301	Solids, Dissolved (mg/l)	1140.00	522.00	822.87	1.07	57	1200.00	691.00	810.23	1.06	24
1020	Strontium (ug/l)	2000.00	1300.00	1685.55	1.15	11					
945	Sulfate (mg/l)	360.00	160.00	288.77	1.10	58	299.44	179.66	259.62	1.10	24
746	Sulfide (mg/l)	0.80	<0.01	0.06	5.17	25					
10	Temperature (°C)*	22.00	3.00	11.47	4.12	65					
1100	Tin (ug/l)	<20.00	<5.00	<9.65	1.75	7					
1150	Titanium (ug/l)	<10.00	<3.00	<6.02	1.67	7					
70	Turbidity (JTU)	1000.00	1.0	8.23	8.94	21	1000.00	0.30	1.79	5.41	24
1085	Vanadium (ug/l)	<16.00	3.00	6.83	1.97	7					
1090	Zinc (ug/l)	100.00	4.00	12.47	2.05	46					
1160	Zirconium (ug/l)	<40.00	<9.00	<18.35	1.86	7					

* Normal mean and standard deviation

Calculated from Gross Alpha as Natural Uranium

TABLE A-5
CONCENTRATIONS IN DEEP WELLS
WHICH EXCEED WATER QUALITY STANDARDS (COLORADO)

<u>Constituent</u>	<u>Most Restrictive Standard</u>	<u>Mean, Lower Aquifer</u>	<u>Maximum, Lower Aquifer</u>	<u>Mean, Upper Aquifer</u>	<u>Maximum, Upper Aquifer</u>
Aluminum	0.3	0.3	2.0	0.3	4.0
Ammonia	0.5	17	200	1.2	7.9
Arsenic	0.01	0.02	0.02	0.01	0.06
Boron	0.75	36	400	1.4	18
Cadmium	0.01	--	0.1	--	0.02
Chloride	250	1200	9800	26	53
Chromium	0.05	--	0.02	--	0.3
Copper	0.04	0.06	0.9	0.09	3.0
Fluoride	1.4	21	48	10	190
Iron	0.3	0.8	8.0	0.5	7.0
Lead	0.05	0.03	0.4	0.01	0.07
Manganese	0.05	0.1	0.6	0.1	0.6
Mercury	0.00005	0.0004	0.0027	0.0004	0.0031
Nickel	0.1	0.01	0.06	0.02	0.2
Radiation, Alpha	15	28	460	6	21
Selenium	0.01	0.004	0.02	0.006	0.03
Silver	0.00025	--	0.02	--	0.05
Zinc	0.6	0.2	4.0	0.2	2.0

TABLE A-6

RANGE OF CONCENTRATIONS FOUND IN RETORT WATERS
RETORT WATER COMPOSITION
(MG/L)

	<u>Minimum</u>	<u>Maximum</u>
Conductivity (μ mhos/cm)	15,100	193,000
Alkalinity (as CaCO_3)	18,200	110,900
COD	8,500	43,000
Carbon, organic	2,200	19,000
TDS	1,750	24,500
Ammonia	1,700	13,200
BOD ₅	350	5,500
Volatile acids (as HAC)	807	1,481
Hardness (as CaCO_3)	20	1,500
Total sulfur (as S)	14	2,320
Phosphorus (as P)	0.23	19
Iron	0.07	68
Nitrate (as N)	1.4	8.7
pH	8.1	9.4

Source: Ref. 26

TABLE A-7
PERMIT REQUIREMENTS

I. Rio Blanco Oil Shale Company Permit No. CO-0 345045

<u>Parameter</u>	<u>30 Day Average</u>	<u>Daily Maximum</u>
Flow	10 MGD	15 MGD
Total Suspended Solids	30 mg/l	45 mg/l
Total Dissolved Solids	1 ton/day 1200 mg/l from finished well or mine shaft 3000 mg/l from new wells being drilled	1800 mg/l
pH	6.0-9.0	6.0-9.0
Oil and Grease	--	10 mg/l No visible sheen
Total Fluoride	2.0 mg/l	3.0 mg/l
Total Boron	1.0 mg/l	1.5 mg/l

II. Occidental Permit No. CO-0033961

Discharge limitations are the same as above with the following additions:

- a. Ammonia discharges are based on the fraction of unionized ammonia, and are .197 mg/l of total ammonia nitrogen at a pH of 8.6. Alternately, 0.2 g of ammonia-nitrogen may be discharged per cubic meter of flow in the Piceance Creek (receiving water).
- b. Phenols are limited to .001 mg/l of discharge, or alternately to 1.0 mg of total phenol per cubic meter of flow in the Piceance Creek.

5.0 AIR POLLUTION CONTROL

5.1 INTRODUCTION

The order of magnitude cost of air pollution control equipment for the four processes was estimated as follows. First, the material balances and information on the processes were used to identify all sources of vented emissions and to quantify the streams. This information, along with performance standards and design specifications, where appropriate, was used to request current prices from the equipment manufacturers. Stone & Webster Engineering Corporation then estimated the cost of installation including all supporting equipment, and the total annual operating cost including utilities, labor, maintenance, chemicals, etc. Often the total cost of the actual piece of control equipment is small compared to the total installed cost.

Point source emission levels were determined in two ways. Expected emissions were calculated, based on performance data from manufacturers, for each piece of pollution control equipment. Data on specific points of uncontrolled emissions generally were obtained from the literature and may be traced to engineering studies by the private developers or government agencies. Occasionally, literature data was updated by the developers.

Pollution control strategies were developed jointly by Denver Research Institute, Stone & Webster Engineering and Water Purification Associates, (WPA), considering not only technical and engineering information, but also the regulatory issues involved. Air pollution control interacts strongly with water management, so WPA was an essential participant in development of the strategies. Good examples of this interaction is NO_x control and particulate control with wet scrubbers. To avoid excessive generation of NO_x , considerable ammonia is removed from the retort gas stream before combustion. The ammonia absorbers shown are considered to be air pollution control equipment. They also serve to cool the retort gas for entry into the

Stretford, to remove particulates, and to condense water from the gas stream. The scrubber water is treated in the water management system to remove and recover the ammonia. Hence, the ammonia recovery system is considered to be part of the water management costs. However, the ammonia storage and shipping facilities costs are again considered to be NO_x air pollution control costs.

5.2 OVERALL RESULTS AND CONCLUSIONS

The total capital and operating costs for air pollution control are shown in Table 5-1 for all the processes. Two scenarios are shown for the TOSCO II and Paraho processes but not for the MIS processes, as discussed in Section 4.0.

TOSCO II particulate controls for the more strict scenario are much more expensive than the less strict case due to the use of high efficiency baghouses or precipitators instead of scrubbers. The high cost of particulate control for the MIS/Lurgi process is due to a \$25.5 million expense for hot precipitators on the retort discharge, a high temperature and dust load emission which requires this special treatment not found in the other processes. As previously discussed, two reasonable choices of equipment for particulate control are not available for the MIS cases, and only one scenario is shown.

In general, two control costs are encountered for NO_x control. The majority of the total cost is associated with ammonia absorption from the retort gas streams. Only one scenario is considered for this item since it is assumed that with the present state-of-the-art, ammonia must be removed to prevent unreasonable NO_x emissions. The absorption systems were designed for maximum ammonia removal and subsequent recovery to defray the cost of such equipment to the maximum extent. The ammonia scrubbers are also designed to lower the retort gas temperature for entry into the Stretford unit.

A minor cost for water injection systems to the gas turbine power generation systems was also included in the total cost for NO_x control. Water injection was judged to be advisable to meet the proposed NSPS standards of 75 ppm NO_x for gas turbines. Since gas turbines are not unique to oil shale processing, and proposed standards exist which are unrelated to the other processing activities which may surround the turbine, only one scenario was used.

Table 5-1
 SUMMARY OF AIR POLLUTION CONTROL CAPITAL AND OPERATING COSTS
 1979 Dollars
 (10³ \$'s)

Process	Particulate Matter		Nitrogen Oxides		Sulfur Dioxide		Hydrocarbons		Totals	
	Capital	Operating	Capital	Operating	Capital	Operating	Capital	Operating	Capital	Operating
TOSCO II										
Less Strict	10,010	1,901	-0-	-0-	9,910	5,919	17,785	420	37,705	8,240
More Strict	30,860	1,277	-0-	-0-	9,910	5,919	17,785	420	58,555	7,616
Paraho										
Less Strict	15,730	1,463	9,280 ¹	1,357	20,180	4,404	720	-0-	45,910	7,224
More Strict	16,280	1,488	9,280	1,357	20,180	4,404	720	-0-	46,460	7,249
Modified In Situ	7,252	615	12,866 ²	1,252	26,210	6,155	240	-0-	46,568	8,022
MIS/Lurgi	35,800	1,651	12,866 ²	1,252	29,800	7,167	878	-0-	79,344	10,070

¹ Includes \$2,170,000 for Ammonia Storage.

² Includes \$3,060,000 for Ammonia Storage.

There are several reasons for TOSCO II having no direct NO_x control costs. First, power is not generated on site in the process model used, and power generation is not planned by Colony. Second, TOSCO II retort gas contains very little ammonia, so an absorber is unnecessary. Ammonia is removed in the refinery, however, so the water management system still has to provide for ammonia recovery.

SO_2 control is accomplished indirectly with Stretford processes which remove hydrogen sulfide from the retort gas stream before combustion. As previously discussed (Section 4.0), SO_2 flue gas scrubbers were also considered, but provided a much lower level of control at a substantially higher cost. Hence, only one scenario, the Stretford process, is presented.

Except for the TOSCO II process, the cost for hydrocarbon control is very modest and mainly consists of the differential costs for floating roof storage tanks. Low flare and thermal oxidizer systems are added to the TOSCO II and MIS-Lurgi processes, but these costs are minor. TOSCO II also has a second, very expensive thermal oxidizer as part of the shale preheater system. This control is the only technology of choice, so again, only one scenario is possible.

Table 5-2 summarizes the expected air emission levels from all processes, including fugitive emissions but not including the impact of employee automobile traffic. The total emissions from all four processes are rather low for the total exportable production of almost 300,000 barrels per day of oil and 500 MW of electricity. The two scenarios for particulate matter in the TOSCO II and Paraho processes show small improvement in total emissions, but at substantially increased cost, particularly for TOSCO II. The MIS processes produce less particulate matter. This is partly due to smaller scale ore crushing operations. For TOSCO II, Paraho and MIS, fugitive dusts are significant, but could be greatly reduced if continuous revegetation is practiced. (See Section 7.0.)

Nitrogen oxides are produced by any process which involves combustion of fuels. The largest emitters of NO_x are the gas turbine power plants. The maximum allowable NO_x emission levels are shown. It is possible that actual emission from power generation using the low Btu retort gases could be less. Since the TOSCO II process does not include power generation, the

Table 5-2
SUMMARY OF ESTIMATED AIR EMISSIONS
(Pounds/hour)

Process	Particulate Matter		Nitrogen Oxides	Sulfur Dioxide		Hydrocarbons	
	Uncontrolled	Controlled		Uncontrolled	Controlled	Uncontrolled	Controlled
TOSCO II							
Less Strict	28,223.2	494.2	1,933.8	32,200	88.5	696	344.7
More Strict	28,223.2	150.7			366		
Paraho							
Less Strict	83,948	727	3,427.4	22,000	220	196.3	39.9
More Strict	83,948	683					
Modified In Situ	2,375	107.1	3,199.2	24,000	240	223.1	97
MIS/Lurgi	26,666	117.1	3,944.4	29,200	292	300.8	33
Total, All Processes	Less Strict	1,445.4	12,504.8		1,074		514.6
284,170 barrels per day oil product	More Strict	1,057.9					

* Based on Table 3-8, TOSCO/Colong plant fuel use assuming 99% Stretford efficiency, 0.8% S in fuel oil, 99% SO₂ removal efficiency for scrubbers (less strict) where fuel gas used and combined 43% SO₂ removal efficiency for baghouses and ESP in place of scrubbers (more strict).

largest NO_x emitter is the ball heater-shale preheater system. Diesel equipment are also substantial contributors in all processes.

SO_2 emissions are low for all processes, particularly considering the high total Btu value of the fuel burned in each process.

Hydrocarbon emissions are also low for all processes. The TOSCO II process has the highest HC emissions due to the shale preheater system.

The actual equipment selections and descriptions of the emission streams they must handle are shown for all processes and scenarios in Tables 5-3 through 5-8. The information presented is sufficient for cost determination of the specified controls using an approach similar to this study. The actual designs of the Stretford units and the ammonia absorber systems are shown in Figures 5-1 and 5-2. Section 3.0 contains overall process diagrams showing the incorporation of pollution controls into the overall processing system.

5.3 CONTROL OF SPECIFIC POLLUTANTS

5.3.1 Particulate Matter

The recovery of shale oil requires the mining, material handling, and processing of large quantities of dry raw and spent shales. Since these operations are all dry, the potential for entrained particulate emissions is high and dust control equipment is required where processes interface with the atmosphere.

The strategy for controlling particulate matter emissions from the oil shale developments includes installation of dust collection equipment at the process emission point sources, and using soil and spent shale chemical stabilizers and paving of roads to control fugitive dust emissions. Process point sources are controlled with equipment such as wet venturi scrubbers, electrostatic precipitators, and fabric filters. The raw and spent shale characteristics indicate that these types of equipment should be capable of removing these particulates from air and other gas streams at efficiencies of 99.0-99.9%. High energy wet venturi scrubbers are expected to be capable of collecting up to 99% and fabric filters are expected to collect up to 99.9% of these particulates. Electrostatic precipitator performance is expected to

[illegible]

Table 5-5
AIR POLLUTION CONTROL EQUIPMENT--PARAHO OIL SHALE PROCESS
Less Strict Scenario

No.	ITEM	ORIGIN	GAS	GAS VOL. (acfm)	GAS TEMP. (°F)	DUST	DUST LOAD (gr/acf)
1	Fabric Filter	Raw Shale Storage	Air	150,000	60	Raw Shale	0.6
2	Fabric Filter	Secondary Crushing	Air	40,000	60	Raw Shale	6.0
1	Fabric Filter	Surge Bins	Air	78,000	60	Raw Shale	3.0
24	Fabric Filter	Tertiary Crush	Air	15,000	60	Raw Shale	6.0
24	Fabric Filter	Retort Feed	Air	20,000	60	Raw Shale	4.5
24	Fabric Filter	Retort Discharge	Flue Gas	20,000	200	Spent Shale	6.1
24	Venturi Scrubbers	Moisturizers	Air & Steam	28,000	184	Spent Shale	1.7
3	Stretford Unit Towers, 929×10^6 scfd fuel gas,						
1	Sulfur Recovery Unit, 132 tons/day sulfur						
3	Ammonia Absorption System, 3,608,000 lbs/hr, wet fuel gas, 5,650 ppmv NH_3 at 150°F, 11.17 psia						
1	Ammonia Storage System						

3* Flue Gas Desulfurization, 3,424,000 acfm, 300°F

* Calculated for cost comparison only. Not included as equipment in final design.

Table 5-6

[illegible]

Table 5-7
AIR POLLUTION CONTROL EQUIPMENT--MODIFIED IN SITU PROCESS
Both Scenarios

No.	ITEM	ORIGIN	GAS	GAS VOL. (acfm)	GAS TEMP. (°F)	DUST	DUST LOAD (gr/acf)
1	Fabric Filter	Shale Crusher	Air	280,000	60	Raw Shale	1.0
6	Fabric Filter	Mine Ventilation	Air	150,000	60	Raw Shale	0.2
2	Fabric Filter	Mine Shaft Conveyor Transfer Point	Air	56,000	60	Raw Shale	0.2
5	Stretford Unit Towers, $1,708 \times 10^6$ scfd fuel gas,						
1	Sulfur Recovery Unit, 144 tons/day sulfur						
5	Ammonia Absorption System, 5,947,000 lbs/hr, wet fuel gas, 6,200 ppmv NH_3 , 11.17 psia						
1	Ammonia Storage System						
3	Flue Gas Desulfurization, 4,759,000 acfm 300°F*						

* Calculated for cost comparison only. Not included in design.

FIGURE 5-1. SCHEMATIC FLOW DIAGRAM OF THE AMMONIA ABSORPTION SYSTEM

FIGURE 5-2. SCHEMATIC FLOW DIAGRAM OF THE RETORT GAS CLEANING SYSTEM INCLUDING THE STRETORD UNIT.

range between 99.0% and 99.9%. The point sources and the equipment selected to control particulate emission are described for each process below.

TOSCO II: The particulate matter emission rates and the order of magnitude costs of controlling these emissions for the TOSCO II Oil Shale Process are shown in Tables 5-9 and 5-10. The emission rates and control equipment costs are shown for two alternative methods for collecting entrained dust. Case I, Table 5-9, the less strict scenario, shows the rates and the order of magnitude costs for the control equipment described in the "Colony Final Environmental Impact Statement." Case II, Table 5-10, the more strict scenario, represents the emission rates and expected costs for alternative dry particulate dust collectors. The dry collectors considered are more effective and do not evaporate water. In Case II, fabric filters or electrostatic precipitators were substituted for the venturi scrubbers following the preheater, elutriator and shale wetter systems.

The preheat system exit gas temperature is approximately 500°F and contains raw shale dust. Since the raw shale dust may be flammable at this temperature, the use of electrostatic precipitators for particulate control will pose unacceptable fire and explosion risks. Teflon fabric filter material that can withstand the 500°F flue gas temperatures is now available. Therefore, it is expected that the preheat venturi scrubber could be replaced with baghouse dust collectors. This baghouse collector is expected to remove 99.7% (as compared to 97.3%) of the entrained raw shale dust and eliminate consumptive water use.

The shale wetter and elutriator exit gas characteristics indicate that neither precipitator or baghouse dust collectors are suitable for this service. The shale wetter gas is saturated with moisture at approximately 180°F and the combination of this high humidity with the spent shale dust will foul dry dust collectors. The elutriator exit gas temperature of 800°F is too high for fabric filter materials and may result in excessive maintenance for electrostatic precipitators. However, the characteristics of the gas produced by combining shale wetter and the elutriator gases indicates that either baghouse or electrostatic precipitator should be suitable for removing particulates. The temperature of the combined gas should be in the range of 400-500°F, however temperature excursions to 800°F are possible when the shale

Table 5-9
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For The TOSCO II/Colony Oil Shale Process
Particulate Matter
Less Strict Scenario

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks*		Efficiency %	Control Equipment	
			No Control lbs/hr	Controlled lbs/hr		Capital Cost M\$	Operating Costs M\$/Year
Preheat System	Venturi Scrubbers	6	11,435	243.4 270.5	97.3	3,610	1,015
Elutriator	Venturi Scrubbers	6		42.1 46.7		1,430	238
Shale Heater	Venturi Scrubbers	6	600	54.0 60.0	90.0	1,450	234
Fine Ore Storage	Fabric Filters	4	1,545	4.2 4.6	99.7	220	11
Fine Ore Crusher	Fabric Filter	1	11,808	32.8 36.3	99.7	1,500	85
Coarse Ore Crusher	Fabric Filter	1	2,610	7.3	99.7	340	19
SUB TOTALS			27,998	425.4	98.5	8,550	1,601
Coker Heater		1	1.1	1.1	0		(b)
Gas Oil Reactor Feed Htr.		2	.4	.4	0		(b)
Gas Oil Reboiler Furnace		1	1.8	1.8	0		(b)
Naptha Reactor Feed Htr.		1	.2	.2	0		(b)
Hydrogen Reforming Furnace		4	12.8	12.8	0		(b)
Boilers		4	3.1	3.1	0		(b)
Sulfur Plant Tail Gas		1	0.0	0.0	0		(b)
Mine Ventilator		3	25.0	25.0	0		(b)
Transfer Points		3	11.0	11.0	0		(b)
SUB TOTALS			55.4	55.4	0		
TOTAL Point Sources			28,053	481	98.3	8,550	1,601
Fugitive Dust Paving & Spraying			170.2	13.2	92.2	1,460 (c)	230 (d) 69
TOTAL All Sources			28,223.2	494.2	98.3	10,010	1,901

(a) Emission rates on first line are average, rates on second line are maximum values.

(b) Taken from references 1 & 2.

(c) Taken directly from 2.

(d) Includes road maintenance and 12¢/1000 gal for dust suppression waters.

Table 5-10
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For The TOSCO II/Colony Oil Shale Process
Particulate Matter
More Strict Scenario

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks (a)		Efficiency %	Control Equipment	
			No Control lbs/hr	Controlled lbs/hr		Capital Cost M\$	Operating Costs M\$/Year
Preheat System	Fabric Filters	6	11,435	30.9	99.7	16,430	395
Elutriator	Hot Precipitator	6	600	3.0	99.5	10,910	468
Shale Wetter		6					
Fine Ore Storage	Fabric Filters	4	1,545	4.2 4.6	99.7	220	11
Fine Ore Crusher	Fabric Filter	1	11,808	32.8 36.3	99.7	1,500	85
Coarse Ore Crusher	Fabric Filter	1	2,610	7.3	99.7	340	19
SUB TOTALS			27,998	82.1	99.7	29,400	978
Coker Heater		1	1.1	1.1	0		
Gas Oil Reactor Feed Htr.		2	.4	.4	0		
Gas Oil Reboiler Furnace		1	1.8	1.8	0		
Naphtha Reactor Feed Htr.		1	.2	.2	0		
Hydrogen Reforming Furnace		4	12.8	12.8	0		
Boilers		4	3.1	3.1	0		
Sulfur Plant Tail Gas		1	0.0	0.0	0		
Mine Ventilator		3	25.0	25.0	0		
Transfer Points		3	11.0	11.0	0		
SUB TOTALS			55.4	55.4	0		
TOTAL Point Sources			28,053	137.5	99.5	29,400	978
Fugitive Dust Paving & Spraying			170.2	13.2	92.2	1,460	230 69
TOTAL All Sources			28,223.2	150.7	99.5	30,860	1,277

(a)*Emission Rates on first line are average, rates on second line are maximum values.

wetter is out of service. Since shale wetter elutriator gas temperatures could exceed 500°F, electrostatic precipitators were selected to replace the venturi scrubbers. The precipitator is expected to collect 99.5% of the particulate and eliminate this consumptive water loss.

Comparison of the values in Tables 5-9 and 5-10 indicate that the total expected input to control equipment TOSCO II point source process emissions (excluding fugitive dust) is 27,998 lbs/hr and a capital expenditure of \$8,550,000 (Case I) and annual operating cost of \$1,602,000 are required to reduce emission by 98.5%. Higher emission reduction to 99.7% will require a capital expenditure of \$29,400,000 and a total operating cost of \$978,000. The lower operating cost for the higher overall removal efficiency is a result of the lower maintenance costs for electrostatic preicpitators and fabric filters. However, the increase in capital expenditures by \$20,850,000 would result in an additional 343.3 lbs/hr of particulate removal.

One advantage of using dry dust collectors for particulate control when possible, as previously mentioned, is the lower water usage. The evaporation of a total of 674 gpm water would be eliminated.

Paraho: The Paraho Oil Shale Process is an aboveground retorting and oil recovery system. Since 24 individual retorts are necessary to produce the required barrels of oil, the Paraho process will need the largest number of dust collectors, over 100 separate units. Fortunately, the dust collector sizes will be small and the overall costs reasonable. Since capital investment for baghouse dust collectors are expected to be similar to venturi scrubbers in the Paraho case, fabric filters were selected for all systems except for the shale moisturizers. Venturi scrubbers were selected for the shale moisturizer exit gases, because of the high relative humidity of this gas. This high relative humidity could blind bag material or cake and short-out precipitators.

The particulate matter emission rates and order of magnitude costs for dust control equipment are shown in Tables 5-11 and 5-12. The two cases shown are nearly identical except in Case I, the less strict scenario, the primary crusher dust emissions are suppressed with water sprays, but in Case II it was assumed that fabric filters were used to collect particulate matter at each primary crusher.

Table 5-11
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For the Proposed Paraho Oil Shale Process
Particulate Matter
Less Strict Scenario

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks		Efficiency %	Control Equipment	
			No Control lbs/hr	Controlled lbs/hr		Capital Cost M\$	Operating Costs M\$/Year
Raw Shale Storage Bldg.	Fabric Filter	1	734	2.2	99.7	820	49
Secondary Crushers	Fabric Filter	2	4,114	12.3	99.7	460	25
Surge Bins	Fabric Filter	2	2,026	6.1	99.7	410	23
Tertiary Crushers & Screen	Fabric Filter	24	18,514	55.5	99.7	2,600	107
Retort Feed	Fabric Filter	24	18,510	55.5	99.7	3,030	159
Retort Discharge	Fabric Filter	24	23,500	70.5	99.7	3,210	191
Shale Moisturiser	Venturi Scrubber	24	9,050	90.5	90.0	3,740	596
SUB TOTALS			76,448	293	99.6	14,270	1,150
Mining			175.2	2.7	98.5	Wet Suppression	(a)
Blasting			365	5.5	98.5	Wet Suppression	(a)
Primary Crusher			3,651	54.8	98.5	Wet Suppression	(a)
Mining Equipment & Disposal Equipment			43.6	43.6			(a)
Transfer			365	5.5	98.5	Wet Suppression	(a)
Ammonia Recovery Plant	No Direct Discharge						
Sulfur Recovery Plant	Moist Air						
SUB TOTALS			4,600	112	97.6		
TOTAL Point Sources			81,048	405	99.5	14,270	1,150
Fugitive Dust Paving & Spraying			2,900 (b)	322 (b)	88.9	1,460	230 (c) 83
TOTAL All Sources			83,948	727	99.2	15,730	1,463

(a) Taken from reference 4.

(b) Scaled from reference 4 to agree with solid waste disposal plans in Section 7.0.

(c) Includes road maintenance & 12¢/1000 gal for dust suppression waters.

Table 5-12
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For the Proposed Paraho Oil Shale Process
Particulate Matter
More Strict Scenario

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks		Efficiency %	Control Equipment	
			No Control lbs/hr	Controlled lbs/hr		Capital Cost M\$	Operating Costs M\$/Year
Primary Crushers	Fabric Filter	5	3,651	11.0	99.7	560	25
Raw Shale Storage Bldg.	Fabric Filter	1	733.7	2.2	99.7	820	49
Secondary Crushers	Fabric Filter	2	4,114	12.3	99.7	470	25
Surge Bins	Fabric Filter	2	2,026	6.1	99.7	410	23
Tertiary Crushers	Fabric Filter	24	18,514	55.5	99.7	2,600	107
Retort Feed	Fabric Filter	24	18,510	55.5	99.7	3,030	159
Retort Discharge	Fabric Filter	24	23,500	70.5	99.7	3,200	191
Shale Moisturizer	Venturi Scrubber	24	9,050	90.5	90.0	3,730	596
SUB TOTALS			80,099	304	99.6	14,820	1,175
Mining			175.2	2.7	98.5	Wet Suppression	
Blasting			365	5.5	98.5	Wet Suppression	
Mining Equipment & Disposal Equipment			43.6	43.6			
Transfer			365	5.5	98.5	Wet Suppression	
Ammonia Recovery Plant			No Direct Discharge				
Sulfur Recovery Plant			Moist Air				
SUB TOTALS			949	57.3	94.0		
TOTAL Point Sources			81,048	361	99.6	14,820	1,175
Fugitive Dust Paving & Spraying			2,900	322	88.9	1,460	230 83
TOTAL All Sources			83,948	683	99.2	16,280	1,488

The addition of a baghouse collector to each primary crusher did not significantly increase the costs but it is expected to improve air quality and working conditions in the mine where the primary crushers are located.

MIS: The MIS Oil Shale Process is a belowground, in situ, retorting system that requires the mining of only approximately 20-25% of the shale rock. Since the quantity of the bulk material mined and handled is much less than aboveground retorting systems and less crushing effort is required, fewer particulate control units are necessary. Based on the report entitled "Oil Shale Tract C-b Supplemental Material to Detailed Development Plan Modifications" filed by Occidental Oil Shale, Inc. and Ashland Oil, Inc. only the mine ventilation and bulk handling systems are expected to be equipped with particulate control equipment.

The expected total particulate point source emission rates and also the order of magnitude dust control equipment costs for MIS Process are listed in Table 5-13. These emission rates and costs are lower than those costs for aboveground retorting systems. These control equipment cost data were based on using fabric filters for particulate control. These equipment are most commonly used for dust control with bulk material handling systems.

The total costs for controlling point source particulate matter emissions are expected to be approximately \$5.8 million capital investment and \$0.34 million annually for operation.

MIS/Lurgi: The MIS/Lurgi Oil Shale Process consists of the MIS process and an aboveground retort (Lurgi Process) which treats the shale necessarily mined for the construction of in situ retorts. Table 5-14 tabulates the expected point source emission rates and the order of magnitude costs of control equipment for the Lurgi/MIS Process.

The emission rates and equipment costs for the MIS portion of this process are identical to those in Table 5-13. Since the Lurgi retorting system generates fine spent shale particulates, which are entrained in combustion products, the uncontrolled emission rates are expected to be much higher than those for MIS process. The Lurgi retort exit gas characteristics indicate that a hot electrostatic precipitator is a practical dust collector for this emission. (See Appendix 5.0.) The exit gas temperature of 608°F is too

Table 5-13
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For The Proposed MIS Oil Shale Process
Particulate Matter
Both Scenarios

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks		Efficiency %	Control Equipment Capital Cost \$	Operating Costs \$/Year
			No Control lbs/hr	Controlled lbs/hr			
Shale Crushing (Site Prep Temp.)	Fabric Filter	1	195.8	.6	99.7	182	8.5
Mine Ventilation	Fabric Filter	1	1,600	4.8	99.7	4,980	291
Mine Shaft Conveyor and Transfer Point	Fabric Filter	1	170.8	.51	99.7	630	34
SUB TOTAL			1,967	5.9	99.7	5,792	334
Temporary Power Generation		1	1.5	1.9	0		(a)
In Situ Gas Treatment		1	7.4	7.4	0		(a)
Ammonia Recovery		No Direct Discharge					
Sulfur Recovery Plant		Moist Air					
Steam Generator		1	16.3	16.3	0		(a)
SUB TOTAL			25.2	25.2	0		
TOTAL Point Sources			1,992	31.1	98.4	5,792	334
Fugitive Dust Paving & Spraying			383 (b)	76 (b)	80.2	1,460	230 51 (c)
TOTAL All Sources			2,375	107.1	95.5	7,252	615

(a) Taken from reference 5.

(b) Scaled from reference 5.

(c) Includes road maintenance and 12¢/1000 gal for dust suppression waters.

Table 5-14
Expected Emission Rates
Control Equipment & Order of Magnitude Costs
For The Proposed Lurgi/MIS Oil Shale Process
Particulate Matter
Both Scenarios

Source	Control Equipment	Number of Stacks	Total Emission Rates For All Stacks		Efficiency %	Control Equipment	
			No Control (c) lbs/hr	Controlled lbs/hr		Capital Cost M\$	Operating Costs M\$/Year
Shale Crushing Site Prep.	Fabric Filter	1	195.8	0.6	99.7	182	9
Mine Ventilation	Fabric Filter	1	1,600	4.8	99.7	4,980	291
Mine Shaft Conveyor Transfer Point	Fabric Filter	1	170.8	0.5	99.7	630	34
Primary Crusher	Fabric Filter	1	1,754	5.3	99.7	290	13
Secondary Crusher	Fabric Filters	4	7,935	23.8	99.7	1,182	58
Ore Storage	Fabric Filter	1	1,039	3.1	99.7	200	8
Lurgi Retort Discharge	Hot Precipitators	5	13,543	13.5	99.9	25,543	830
Moisturizer	Venturi Scrubbers	5	403	40.3	90.0	1,333	161
SUB TOTALS			26,641	91.9	99.8	34,340	1,404
Power Generation		1	1.5	1.5			
In Situ Gas Treatment		1	7.4	7.4			
Ammonia Recovery Plant		1	No Direct Discharge				
Sulfur Recovery Plant			Moist Air Only				
Steam Generator		1	16.3	16.3			
SUB TOTAL			25.2	25.2			
TOTAL Point Sources			26,666	117.1	99.6	34,340	1,404
Fugitive Dust			- (a)	- (a)		1,460	247 (b)
TOTAL All Sources			26,666	117.1	99.6	35,800	1,651

(a) Fugitive dusts will be very low since all traffic areas will be paved and the disposal area will be essentially cement. Actual levels have never been calculated.

(b) Includes road maintenance and 12¢/1000 gal for dust suppression waters.

(c) Values taken from Section 3.0 and proprietary sources.

high for fabric filters. High energy wet scrubbers are not expected to be capable of satisfactory performance and would evaporate large quantities of scarce water.

In addition, the Lurgi retorts require shale crushing, as do other aboveground retorts. The shale crushers are expected to be equipped with baghouse collectors for dust control. Table 5-14 shows that the retort discharge hot precipitator comprises about 2/3 of the total cost for controlling particulate matter emission from the MIS/Lurgi process. The high cost of the precipitator is due to the high gas temperature and dust loading. These "hot" precipitators will collect 99.9% of the input dust and actually handle 50% of the total dust produced by the process.

Fugitive Dusts: Fugitive dust is expected to be controlled by spraying and paving of roads for all processes. Costs are included for paving of 10 miles of road in each process. Cost of water consumed in dust suppression was set at 12¢/1,000 gallons, the same cost used for cooling and scrubbing water. The cost of chemical spraying in the disposal areas is included in Section 7.0 as a cost of solid waste management. Fugitive dust emissions were calculated from literature values for the processes and scaled to fit the disposal models used in this study. Fugitive dusts from paved areas were assumed to be negligible.

5.3.2 Nitrogen Oxides

The retorting of raw shale, containing approximately 0.2-0.5% nitrogen, will volatilize nitrogen compounds. Because of the reducing atmosphere in the retorts, the nitrogen forms ammonia (NH_3) gas. Ammonia has a low boiling point and will not condense during the recovery of the liberated kerogens, but remains as a gas with the light hydrocarbons. The noncondensable hydrocarbon gases (retort gas or fuel gas) are usually burned in process equipment and in stream or power generating systems.

The combustion of these retort gases without NH_3 removal can generate nitrogen oxides (NO_x) concentrations up to about 5,000 ppm in the flue gas (assuming all the N in NH_3 is converted to NO_x) or about 16 lbs NO_x per 10^6 Btu of heat input (assuming no oxidation of atmospheric nitrogen, N_2). Therefore, NO_x control plans must include effective NH_3 removal from the retort gases. Water scrubbing can reduce ammonia concentration in the retort

gases to 10-20 ppm and is expected to be the most practical NO_x control system for the oil shale developments. The absorbed ammonia can be stripped from the water and recovered as a byproduct.

After NH_3 removal from the retort gas, no additional capital expenditures are anticipated for controlling NO_x emission from the steam boilers. The burners required for stable combustion of the low Btu retort gas inherently generate low concentrations of NO_x . However, gas turbines may require water injection (flame quenching) to control NO_x emissions and for compliance with the proposed NSPS of 75 ppm NO_x adjusted 15% oxygen in the combustion products.

The ammonia removal systems for the Paraho, MIS, and MIS/Lurgi processes are the same, only the mass rates differ slightly.

The gases from eight retorts, following crude shale oil recovery, enter the ammonia absorber at approximately 150°F. Since downstream equipment (the Stretford Units) require 95°F gas temperature and since the lower temperatures increase the solubility of NH_3 in water, the retort gas is first cooled by direct contact with cold water. The absorption tower selected for this study is a combination of a cooler and absorber. The retort gas is cooled in the lower section of the tower and NH_3 is absorbed in the upper section. Water flow is countercurrent for absorption and cooling. A schematic flow diagram of the NH_3 absorption system is shown in Figure 5-1.

The control strategy for minimizing NO_x emissions from the four oil shale projects is discussed below.

TOSCO II: The TOSCO II Oil Shale process does not include a separate scrubber to remove NH_3 from the noncondensable retort gases that fuel the process equipment. Very little ammonia is produced by the combustion-free low temperature TOSCO II retorts, and the small amount generated is removed in the fractionator. The ammonia concentrations in the fuel gas are expected to be negligible.

In addition to the retorts, the TOSCO II oil shale development plan includes partial refining of the recovered shale oil and therefore will have numerous process heaters and boilers which emit nitrogen oxides. Table 5-15 lists all these possible sources of NO_x and the emission rate.

Table 5-15
 EXPECTED NO₂ EMISSION RATES FOR THE COLONY SHALE OIL PLANT
 Both Scenarios

Source	No. of Stacks	Heat Input 10 ⁶ Btu/hr	Total NO ₂ Emission Rates from all Stacks Lbs/hr (a)	Lbs No/10 ⁶ Btu
Preheater System	6	1,830.6	1,314.8	0.72
Elutriator System	6	94.3	113.4	1.2
Coke Heater	1	47.6*	8.4	0.18
Gas-Oil Reactor Feed Ht.	2	18	2.7	0.15
Gas-Oil Reboiler Fur.	1	75	11.2	0.15
Naphtha Reactor Feed Ht.	1	8.4	1.3	0.16
Hydrogen Reforming Fur.	4	548	82.2	0.15
Boilers Auxiliary	2	144	21.6	0.15
Sulfur Plant TGU	1	-0-	-0-	-0-
Shale Wetter	6	-0-	-0-	-0-
Fine Ore Storage	1	-0-	-0-	-0-
Fine Ore Crusher	4	-0-	-0-	-0-
Mine Ventilator	3	N/A	249.5 (b)	-0-
Disposal Traffic	—	—	129.9 (c)	—
Total for Plant Site			1,935.0	
Utility Boilers at Grand Valley	2	5*	0.8	0.18

(a) Values from Reference 2 unless otherwise specified.

(b) Private Communication.

(c) Scaled from Reference 4.

* Firing duty estimate.

As can be seen from Table 5-15, most of the fuel burning systems have very low NO_x emission rates with the exception of the preheat and the elutriator systems. These systems are expected to have higher NO_x emissions, due to their process characteristics and the use of liquid fuels containing organic nitrogen.

High flame and combustion gas temperatures are required in the ball heater (source of hot gases to preheat system) to obtain maximum oil recovery and maintain process efficiency. Unfortunately, these high flame temperatures also generate higher concentrations of NO_x through the fixation of atmospheric nitrogen. We anticipate that NO_x emission rates from the preheat system cannot be reduced, with combustion modifications, without reducing process efficiency substantially. Therefore, no attempt was made to estimate the cost of reducing NO_x emissions by redesigning the ball heater/preheat system. The reduction of NO_x by wet absorption or catalytic reduction with NH_3 have not been tested on this type of facility, and the applicability of these systems to the TOSCO II preheat system is questionable. The effect of the raw shale dust and hydrocarbons on the NO_x conversion catalyst and NO_x absorption liquids must be evaluated to determine the feasibility of using these systems, when they become commercially available.

The elutriator system also exhibits a high NO_x emission rate of 1.2 lbs. $\text{NO}_2/10^3$ Btu due to the fuel characteristics. Raw crude shale oil containing up to 2% nitrogen is expected to be burned in the steam superheater, the source of hot combustion gases for the elutriator. This high NO_x emission rate could be reduced to 0.2-0.3 lbs. $\text{NO}_x/10^6$ Btu by burning a lower nitrogen gaseous or liquid fuel, but the partial refining process will incur a performance penalty. Since the elutriator heat input and the emission of 113 lbs NO_2/hr is low, and since the total plant NO_x emissions are also relatively low, the burning of a higher grade fuel in the elutriator is not expected to be cost effective. The other (excluding preheat and elutriator) process heaters and steam boilers which burn clean fuel gas, produce less than 0.2 lbs $\text{NO}_2/10^6$ Btu each.

Because of the low NH_3 concentrations in the fuel gas (retort gas) and the use of this clean gas to provide the process energy, as well as the lack of power generation, no additional capital outlay is anticipated for controlling NO_x emission from TOSCO II.

Paraho: Light noncondensable hydrocarbon retort gases are recycled and burned directly in the Paraho retorts to provide the energy necessary to liberate the kerogens from the shale. During this combustion process, oxides of nitrogen may form; however the subsequent reducing atmosphere in the retort will reduce most of the NO_x to either NH_3 or N_2 . The Paraho retort gas is expected to contain higher concentrations, approximately 0.6% by volume, of ammonia than TOSCO II, because of higher retorting temperatures. The ammonia concentration in the retort gas is not expected to be reduced during the shale oil recovery. Therefore, to avoid the discharge of combustion products containing high concentrations of NO_x when the excess retort gases are burned in steam boilers or gas turbines, an ammonia absorption system is anticipated.

The ammonia absorbers (three absorbers one for each eight retorts) will cool and treat the retort gases following the shale oil recovery system; the cooled ammonia free retort gas is next treated for H_2S removal. Due to the large gas volume, three 30-foot diameter and 52-foot high absorbers are anticipated to reduce ammonia concentration to 10-20 ppm range. The total estimated order of magnitude costs of the absorption system are shown in Table 5-16.

The retort gases from the direct mode Paraho oil shale process have a very low Btu value, in the range of 100-250 Btu/SCF. When burned, these gases cannot produce the high flame temperatures necessary for the fixation of atmospheric nitrogen. Also, the burner designs necessary to provide a stable flame with this low Btu gas inherently generate low NO_x concentrations. Therefore, no differential costs were included for the steam boiler burner modifications for NO_x control. The steam boilers are expected to comply with the proposed New Source Performance Standard of 0.2 lbs $\text{NO}_2/10^6$ Btu of heat input, when the ammonia absorbers are removing 99.0% of the NH_3 from the retort gas.

The gas turbines may require water injection to comply with the proposed New Source Performance Standard, of 75 ppm of NO_x in the turbine exhaust at 15% oxygen. The order of magnitude costs of a water injection system has been estimated and is included in Table 5-16.

Modified In Situ: As the Paraho Oil Shale Process, the Modified In Situ Oil Shale Process generates a large volume of retort gases that have a low Btu

Table 5-16
 EXPECTED NO₂ EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE COSTS FOR THE PARAHO SHALE PROCESS
 Both Scenarios

Source	No. of Stacks	Heat Input 10 ⁶ Btu/hr	Total NO ₂ Emission Lbs/hr	Rates from all Stacks Lbs NO ₂ /10 ⁶ Btu	Control	Equipment
					Capital Costs \$	Operating Costs \$/year
Blasting	10		149			
Mining Equipment			565.5			
Disposal Traffic			297.9			
Energy Utilization						
Steam Boilers	2	1,480×10 ⁶	296	0.2		
Gas Turbines	3	4,330×10 ⁶	2,119 (75 ppm)	0.49	176,000	7,500
Sub Totals			3,427			
Ammonia Scrubber			-0-		6,394,000	1,350,000
Totals			3,427.4		6,570,000	1,357,000

Source: Reference 5.

value and contain high concentrations of ammonia (NH_3), about 0.6% by volume. If these retort gases are burned, without NH_3 removal, high concentrations of NO_x can be emitted with the combustion products.

The ammonia absorption system for the MIS process is expected to include five absorption towers approximately 30 feet diameter and 68 feet high. One tower is expected to handle the gases from eight retorts.

Steam boilers and/or gas turbines are expected to be capable of complying with the NSPS of $0.2 \text{ lb NO}_2/10^6 \text{ Btu}$ and 75 ppm adjusted to 15% oxygen respectively. The cost associated with a water injection system have been included in Table 5-17 for control NO_x emissions from the gas turbines.

MIS/Lurgi: The Modified In Situ/Lurgi Oil Shale Process will have the same NO_x emission rates and control equipment costs as the MIS process, Table 5-17, in addition to the NO_x emission rates and costs associated with the Lurgi retorts. These emission rates and order of magnitude costs are shown in Table 5-18.

The emission rates and control systems for the MIS portion of this process were described in the previous section of this report. Retort gas ammonia scrubbing and gas turbine water injection (flame quenching) are anticipated to be necessary for controlling NO_x emissions from this portion of the process.

The Lurgi retort gas is not expected to contain NH_3 and consequently may be burned in combustion systems without NH_3 scrubbing. The Lurgi retort gas is expected to be burned in the steam boilers or gas turbines along with the MIS retort gases. The MIS/Lurgi hourly emission rates are expected to be slightly higher than MIS, because more fuel will be burned in this combined process. However, the emission rates are expected to be in compliance with the proposed standard of $0.2 \text{ lbs NO}_2/10^6 \text{ Btu}$ and 75 ppm @ 15% O_2 for steam boilers and gas turbines respectively.

The Lurgi retorting system includes a fluidized bed spent shale oxidizer, which burns the residual organic carbon remaining in the spent shale. Since, during this oxidation the bed temperatures exceed the decomposition temperatures of the nitrogen compounds remaining in the spent shale, the formation of NO_x is expected. However, it is expected that the formation and NO_x emissions may be minimized by operating the fluidized bed at low

Table 5-17
 EXPECTED NO₂ EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE COSTS FOR THE MODIFIED IN SITU OIL SHALE PROCESS
 Both Scenarios

<u>Source</u>	<u>No. of Stacks</u>	<u>Heat Input 10⁶ Btu/hr</u>	<u>Total NO₂ Emission Lbs/hr</u>	<u>Rates from all Stacks Lbs NO₂/10⁶ Btu</u>	<u>Control</u>	<u>Equipment</u>
					<u>Capital Costs \$</u>	<u>Operating Costs \$/year</u>
Mining (blasting, handling, hauling, crushing)	1		300 (a)	--		
Disposal Traffic			97.2 (b)	--		
Energy Utilization						
Steam Boilers	2	1,480×10 ⁶	292	0.2		
Gas Turbines	5	2,650×10 ⁶	2,510 (75 ppm)	0.95	116,000	5,000
Sub Totals			3,199.2		116,000	5,000
Ammonia Scrubber			-0-		9,690,000	1,247,000
Totals			3,199.2		9,806,000	1,252,000

(a) Source is References 5 and 6 unless otherwise specified.

(b) Scaled from Reference 4.

Table 5-18
 EXPECTED NO₂ EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE COSTS FOR THE LURGI/MODIFIED IN SITU OIL SHALE PROCESS
 Both Scenarios

Source	No. of Stacks	Heat Input 10 ⁶ Btu/hr	Total NO ₂ Emission Lbs/hr	Rates from all Stacks Lbs NO ₂ /10 ⁶ Btu	Control	Equipment
					Capital Costs \$	Operating Costs \$/year
Mining (blasting, handling, hauling, crushing)	1		300	--		
Disposal Traffic			97.2 (a)	--		
Energy Utilization						
Steam Boilers	2	1,480×10 ⁶	292	0.2		
Gas Turbines	5	3,230×10 ⁶	3,060	0.95 (75 ppm)	116,000	5,000
Fluidized Bed	5		380 (b)			
Sub Totals			3,944.4			
Ammonia Scrubber			-0-		9,690,000	1,247,000
Totals			3,944.4		9,806,000	1,252,000

(a) Scaled from Reference 4.

(b) Proprietary source; see discussion previous page.

excess air levels. It was assumed that the fluidized bed combustion products would contain 100 ppm by volume of NO_x , since precise data was not made available.

5.3.3 Sulfur Dioxide

Oil shale developments have the potential for emitting large quantities sulfur oxides without control systems. The uncontrolled sulfur emission rates for each of the four developments are in the range of 100-200 tons/day. These emission rates approach those of a 100MW steam-electric power plant burning a low sulfur coal (without SO_2 scrubbers). As a result of ambient air quality requirements sulfur removal is expected to be required for all the developments considered.

The potential to emit these large quantities of sulfur oxides is a result of the sulfur in the retort gases which are used to supply process heating requirements and to generate steam and electric power. The retort gases contain in the range of 0.1-0.3% by volume of hydrogen sulfide (H_2S). The H_2S and organic sulfur compounds are contained in the raw shale and are liberated during retorting. If these retort gases were burned, without removing H_2S , sulfur dioxide concentration up to 2,000 ppm in the combustion products could be emitted.

The oil shale developers have two options for controlling sulfur emissions; these are H_2S removal prior to burning the retort gases or SO_2 scrubbing after combustion. Based on published reports, the four developers favor H_2S removal prior to combustion, although SO_2 scrubbing also has several advantages.

Sulfur dioxide removal from flue gas after combustion of the retort gases (H_2S is converted to SO_2) may be attractive due to the inherent alkalinity of the spent shale. Spent shale contains up to 30% lime and 10% soda ash by weight and is reactive with SO_2 in wet scrubbers. Chemical costs for sulfur emission control could be eliminated by using the alkalinity in the spent shale to remove SO_2 from steam or power generation equipment stack gases. When in situ retorts are used, spent shale remains in the ground and is not available for SO_2 scrubbing. However, these processes could use excess spent shale from nearby aboveground retorts for SO_2 scrubbing. Using the spent shale for SO_2 removal in wet scrubbers may also improve the disposal characteristics of the spent shale.

Because the flue gas volumes are greater than the fuel gas volumes prior to combustion, the SO_2 scrubber sizes will be larger than the H_2S absorbers. Consequently, the capital expenditure for SO_2 removal is expected to be much higher than H_2S removal. The order of magnitude capital costs for SO_2 removal for the Paraho and MIS processes were estimated for comparison with H_2S removal before burning of retort gases. These cost data confirmed that it is more economical to remove H_2S from the fuel gas and to produce elemental sulfur for sale.

Hydrogen sulfide can be removed from hydrocarbon gas streams by several methods. These methods were developed by the petroleum industry for the desulfurization of fuel oils and natural gases and are directly applicable for removing H_2S from retort gases. The H_2S removal process selection depends on gas characteristics and performance requirements. H_2S removal with a Stretford unit was selected as a basis for the study. The Stretford H_2S removal and sulfur recovery process offers the highest system performance with H_2S removal efficiencies up to 99.9%.

Table 5-19 lists the order of magnitude capital and operating costs for Stretford units for the four oil shale developments. Figure 5-2 shows a schematic of the Stretford as a part of the gas cleanup system.

Both the capital and operating costs are significantly affected by the fuel gas volumes requiring treatment and the sulfur production rates. The Paraho and MIS processes produce less sulfur than the TOSCO II process. However, due to the much larger volumes of fuel gas treated, the capital costs for these systems are proportionately higher. The TOSCO II Stretford process recovers the most sulfur, because during partial refining organic sulfur contained in the crude shale oil is converted to H_2S for removal in the Stretford system.

The operating costs for the Stretford units are in range of 4.4-7.2 million dollars. These high operating costs are due to the high chemical makeup costs for sodium vanadate and anthraquinone disulfonic acid. The annual chemical makeup costs are in the range of 2.1-3.3 million. Chemical makeup is necessary because a portion of the process reagents are lost due to entrainment in the elemental sulfur produced. Degraded chemicals must also be replaced. Based on information obtained from one of the Stretford system vendors, chemical makeup cost may be reduced 50% by

Table 5-19
 EXPECTED SULFUR EMISSION RATES EXPRESSED AS SO₂ AND STRETFORD SYSTEMS ORDER OF MAGNITUDE COSTS
 Both Scenarios

Process	Potential SO ₂ Emissions	Controlled SO ₂ Emissions	Removal Efficiency %	Stretford Systems Costs M\$	
	lbs/hr	lbs/hr		Capital \$	Operating \$/yr
TOSCO II	8,620*	88.5** 366	99.0 95.8	9,910	5,919
Paraho	22,000	220	99.0	20,180	4,404
Modified In Situ	24,000	240	99.0	26,210	6,155
MIS/Lurgi	29,200	292	99.0	29,800	7,167

* Based on total sulfur in fuel burned (see Table 3-8), not on sulfur removed by Stretford Process.

** See footnote to Table 5-2.

recovering the purged chemicals. The capital costs for the purge stream recovery system were not available and could not be included in this study. The recovery of the purged chemical could reduce operating expenses, but would increase the total plant investment costs.

The costs of the Stretford systems compared to the costs of SO₂ scrubbing for Paraho and MIS processes indicate that SO₂ removal capital and operating costs are higher than H₂S removal as shown below.

TABLE 5-20
COST COMPARISON OF H₂S TO SO₂ REMOVAL (10⁶ \$)
1979 Dollars

	<u>Capital Costs</u>		<u>Operating Cost/Yr</u>	
	<u>SO₂</u>	<u>H₂S</u>	<u>SO₂</u>	<u>H₂S</u>
Paraho	47	20	4.9	4.4
MIS	59	26	6.5	6.2

Besides the economic advantages, Stretford systems also have capability of higher performance, use less water, do not scale up, and do not generate large volumes of waste sludge for disposal. Stretford units produce elemental sulfur, a salable byproduct. For this evaluation, credit for sulfur sales was included, but the value of produced sulfur was set at a minimum value of \$5/ton considering the shipping requirements.

5.3.4 Hydrocarbons

During normal operation, the emission of gaseous hydrocarbons (HC) are expected to be low from all the oil shale processes considered, without control equipment, if good maintenance and operating practices are followed. Good maintenance and operating practices include prompt repair of leaky equipment and cleanup of spilled hydrocarbons, control of operating variables to avoid venting and excess leakage, control of air or oxygen to combustion equipment to assure complete combustion, etc. To reduce and to maintain these low emission rates during emergencies or process upsets, thermal oxidizers and flare stacks may be required.

Generally a flare stack is necessary for venting the total hydrocarbon gases contained in plant equipment during emergencies. These stacks can be designed to effectively burn all hydrocarbon gases. However, for

complete combustion of small volumes of hydrocarbon gases released from pressure relief valves, etc. during process upsets, a separate thermal oxidizer is anticipated. Also, to reduce HC emission, the storage tanks will be equipped with floating roofs.

The HC emission rates, the order of magnitude costs for thermal oxidizers, and differential costs for floating roof storage tanks are discussed separately below for each oil shale process.

TOSCO II: Expected HC emission rates for the TOSCO II Oil Shale Process are shown in Table 5-21. These expected emission rates are low during normal operation. The HC emission rates during emergencies and process upsets are not readily predictable and are not included. However, to burn vented HC gas from the TOSCO II process during process upsets, a thermal oxidizer in addition to the normal process flare is anticipated.

The Colony Oil Shale Development Plan includes the TOSCO II shale retorting system and partial refining of the recovered crude oil. (See Section 3.0 for more detailed description). Since the partial refining system is operated at elevated pressures, intermittent venting of hydrocarbon gases during process upsets may be necessary. Therefore, it is anticipated that a thermal oxidizer capable of burning up to 20% of the process gases at 99% efficiency may be required in addition to the normal process flare stack. The hydrocarbon emissions from the retort system, during start-up, shutdown, and process upsets may be also burned in the thermal oxidizer. When these gas volumes are too large for the thermal oxidizer, they are expected to be flared. This double flare system (thermal oxidizer and flare stack) is expected to burn 99% of the vented HC emissions; consequently, HC emissions are expected to be low.

The preheat system is the only process equipment that emits hydrocarbons during normal operation. The system is expected to be equipped with an effective thermal oxidizer between preheat stages.

However, volatile HC gases are stripped and emitted from the final preheat stages. Due to the high volumetric gas flow rates and the high temperature necessary for oxidation of the entrained HC gases, thermal oxidation is not expected to be practical following the last preheat stage. The order of magnitude costs for the preheat system thermal oxidizer are presented in Table 5-21.

Table 5-21

EXPECTED HYDROCARBON EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE
COSTS FOR THE PROPOSED TOSCO II COLONY SHALE OIL PLANT

Both Scenarios

Source and (Control Equipment)	No. of Stacks	Total Emission Rates for All Stacks		Control Equipment		
		No Control	Controlled	Collection	Capital	Operating
		<u>Lbs/hr</u>	<u>Lbs/hr¹</u>	<u>Efficiency</u> <u>%</u>	<u>Investment</u> <u>\$</u>	<u>Costs</u> <u>\$/yr</u>
Preheat system (Thermal Oxidizer)	6	481.5	270	55%	16,330,000**	\$420,000
Mine Ventilator	3	50	50			
Other Source	29	6.4	6.4			
Shale Oil Storage Tanks (Floating Roof)	(11 Tanks)	150	17.5 ²	88.3	830,000* ³	
Disposal Traffic		8.1	0.8 ⁴	90.1		
Emergencies and Process Upsets (Thermal Oxidizers)					625,000	
TOTALS		696	344.7	50.5	17,785,000	\$420,000

* Cost difference between double sealed floating roof and conventional cone roof tanks.

** Private communication with TOSCO.

NOTES TO TABLE 5-21

1. Source is References 1 and 2 unless otherwise indicated.
2. Following Tankage assumed:

Type	Diameter	No.	Contents	Capacity	60°F	M
Floating Roof	150'	2	#6	111,000	0.00004	190
Floating Roof	200'	2	Gas Oil	482,000	6.35	70
Floating Roof	200'	1	Naphtha	260,000	1.3	80
Floating Roof	150'	1	Fuel Oil & Diesel	155,000	0.0085	130
Floating Roof	150'	2	Fuel Oil	230,000	0.0085	130
Cone Roof	30'	1	Diesel	3,200	0.0074	130
Cone Roof	20'	1	Diesel	1,700	0.0074	130

Calculation based on References 8 and 9. Assume average wind velocity equals 6.52 mph [7] (Actual lower at TOSCO and Paraho, see [4]); average yearround temperature of stored oil assumed equals 60°F (Actual area average temperature equals 44°F [10]); floating floor, double seal, good condition, white paint; $H = 15$, $\Delta T = 20^\circ\text{F}$. Emissions with no controls determined by calculations for cone roof tanks.

3. Costs determine using References 8 and 11, and adjusting to current dollars using 7% annual inflation.
4. Scaled from Reference 4.

Table 5-22
 EXPECTED HYDROCARBON EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE
 COSTS FOR THE PARAHO OIL SHALE PROCESS
 1979 Dollars

Source and (Control Equipment)	No. of Stacks	Both Scenarios		Control Equipment		
		Total Emission Rates for All Stacks		Collection Efficiency %	Capital Investment \$	Operating Costs \$/yr
		No Control lbs/hr	Controlled lbs/hr ¹			
Mine Ventilator Equipment	10	35.7	3.6			
Disposal Traffic	--	18.8	1.9			
Energy Utilization	3	18.8	18.8			
Shale Oil Storage Tanks (Floating Roof)	(3 Tanks)	123	15.6 ²		720,000	
TOTALS		196.3	39.9	79.7	720,000	

¹ Source is Reference 4 unless otherwise specified.

² Calculations based on References 8 and 9. Assumes p = 4.50 psia, wind velocity 6.52 mph, T = 60°F, floating roof, double seal, white paint, H = 15', ΔT = 20°F. No controls = cone roof. 3 × 200' diameter tanks, 700,000 barrels crude shale oil.

Paraho: The expected emission rates for the Paraho Direct Mode Oil Shale Process are shown in Table 5-22. These very low emission rates represent expected normal operating conditions. Most of the total hydrocarbons are emitted by the diesel shale mining and transport equipment, and the storage tanks.

Hydrocarbon emissions during emergencies and upset conditions are expected to be vented to the process steam boilers or energy production system. The Paraho Direct Mode Oil Shale Process produces large volumes of low Btu hydrocarbon gas which must be burned in steam boilers or other energy producing systems. This equipment is expected to be capable of burning minor volumes of vented gases at efficiency greater than 99.0%. The costs associated with burning vented HC gases in steam boiler or other such combustion equipment is expected to be minimal.

The crude shale oil storage tanks should be equipped with floating roofs to minimize the HC emissions. The differential cost for floating roof tanks as opposed to cone roof tanks is shown in Table 5-22.

MIS: The Modified In Situ (MIS) Oil Shale Process, like Paraho, is also expected to have low hydrocarbon emission during normal operating conditions. The expected emission rates are shown in Table 5-23.

The control strategy for reducing HC emission from the MIS process is the same as the one discussed previously for Paraho. Process vent gases are expected to be burned in the steam or power generating equipment and floating roof shale oil storage tanks are expected to be used. The differential costs for floating roof storage tanks are shown in Table 5-23. The burning of process vent gases in the steam/power generating systems is not expected to have significant associated costs.

As in the Paraho process the diesel mining and transport equipment and storage tanks are the major HC producers during normal operating conditions.

MIS/Lurgi: The hydrocarbon emissions from the MIS-Lurgi Oil Shale process expected during normal operating conditions, are shown in Table 5-24.

The control strategy for the MIS/Lurgi process is expected to include a thermal oxidizer to burn the vented gases from the Lurgi process during emergencies.

Table 5-23
 EXPECTED HYDROCARBON EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE
 COSTS FOR THE MODIFIED IN SITU OIL SHALE PROCESS
 1979 Dollars

Both Scenarios

Source and (Control Equipment)	No. of Stacks	Total Emission Rates for All Stacks		Control Equipment		
		No Control lbs/hr	Controlled lbs/hr ¹	Collection Efficiency %	Capital Investment \$	Operating Costs \$/yr
Mine Vent Diesel Equipment	1	120	12			
Disposal Traffic	--	6.1	0.6 ²			
In situ Gas Treatment and Energy Utilization	5	15	15			
Shale Oil Storage Tank (Floating Roof)	(2 Tanks)	82	10.4		240,000	
TOTALS		223.1	97	56.5	240,000	

¹ Source is References 5 and 6 unless otherwise specified.

² Scaled from Reference 4.

³ Same assumptions as Paraho; 2 × 200' diameter tanks; 400,000 barrels crude shale oil.

Table 5-24
 EXPECTED HYDROCARBON EMISSION RATES CONTROL EQUIPMENT ORDER OF MAGNITUDE
 COSTS FOR THE LURGI/MODIFIED IN SITU OIL SHALE PROCESS
 1979 Dollars

Source and (Control Equipment)	No. of Stacks	Both Scenarios		Control Equipment		
		Total Emission Rates for All Stacks		Collection Efficiency %	Capital Investment \$	Operating Costs \$/yr
		No Control lbs/hr	Controlled lbs/hr			
Mine Vent						
Diesel Equipment	1	120	12			
In situ Gas Treatment and Energy Utilization	5	15	15			
Shale Oil Storage Tank (Floating Roof)	(3 Tanks)	165	5.9 ¹		360,000	
Disposal Traffic	--	0.8	0.1			
Emergencies and Process Upsets (Thermal Oxidizer)	1				518,000	
TOTALS		300.8	33.0	89.0%	\$878,000	

¹ Same assumptions as Paraho except: $P = 5.05$, $3 \times 200'$ diameter tanks, 570 barrels crude shale oil.

² Scaled from Reference 4.

The costs associated with the thermal oxidizer and floating roof storage tanks are shown in Table 5-24.

5.4 CAPITAL AND OPERATING COST BREAKDOWN FOR AIR POLLUTION CONTROL

A more complete breakdown of operating costs for the controls discussed in this section is presented in Appendix 5.0.

REFERENCES

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Appendix 5.0

ELECTROSTATIC PRECIPITATION OF OIL SHALE ASH

Electrostatic precipitators of conventional design operate most efficiently when the resistivity of the ash to be trapped is between 10^{10} ohm-cm. Resistivity of a sample of Lurgi ash was measured at DRI; the graph (Figure 5-A1) shows that the resistivity between 608°F and 700°F will allow for efficient trapping in a high temperature precipitator.

FIGURE 5-A1. LURGI ASH RESISTIVITY

Table 5-A1
 Air Pollution Control Equipment
 Operating Costs 1979 Dollars
 TOSCO II/Colony Oil Shale Process
 Less Strict Scenario
 \$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemicals</u>	<u>O & M</u>	<u>Totals</u>
Venturi Scrubber	Shale Wetter	182,000		15,000		36,570	233,570
Fabric Filter	Coarse Ore Crusher	15,300				3,400	18,700
Fabric Filter	Fine Ore Crusher	69,800				15,100	84,900
Fabric Filter	Storage	9,040				1,890	10,930
Venturi Scrubber	Preheat System	916,200		6,930		91,370	1,014,500
Venturi Scrubber	Elutriator	188,290		14,760		36,100	239,150
Stratford System	Retort Gas	<u>2,037,225</u>	507,162	6,693	3,113,710	<u>254,000</u>	<u>5,918,790</u>
		3,417,855				438,430	7,520,540
Thermal Oxidizer	Preheat System	<u>280,000 (fuel gas)</u>				<u>140,000</u>	<u>420,000</u>
	Totals	3,697,855	507,162	43,383	3,113,710	578,430	7,940,540

Table 5-A2
 Air Pollution Control Equipment
 Operating Costs 1979 Dollars
 TOSCO II/Colony Oil Shale Process
 More Strict Scenario
 \$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemicals</u>	<u>O & M</u>	<u>Total</u>
Fabric Filter	Preheat System	324,300				70,500	394,800
Fabric Filter	Fine Ore Storage	9,040				1,890	10,930
Fabric Filter	Fine Ore Crusher	69,820				15,100	84,920
Fabric Filter	Coarse Ore Crusher	15,300				3,400	18,700
Electrostatic Precipitator	Elutriator/Shale Wetter	375,200				92,700	467,900
Stretford System	Retort Gas	<u>2,037,225</u>	507,162	6,693	3,113,710	<u>254,000</u>	<u>5,918,790</u>
		2,830,885				437,590	6,896,040
Thermal Oxidizer	Preheat System	<u>280,000 (fuel gas)</u>				<u>140,000</u>	<u>420,000</u>
	Totals	3,119,885	507,162	6,693	3,113,710	577,590	7,316,040

Table 5-A3
 Air Pollution Control Equipment
 Operating Costs 1979 Dollars
 Paraho Oil Shale Process
 Less Strict Scenario
 \$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemicals</u>	<u>O & M</u>	<u>Total</u>
Fabric Filter	Raw Shale Storage Bldg.	40,200				8,300	48,500
Fabric Filter	Secondary Crusher	20,600				4,160	24,760
Fabric Filter	Surge Bins	19,020				4,160	23,180
Fabric Filter	Tertiary Crusher & Screens	83,760				18,480	107,240
Fabric Filter	Retort Feed	131,260				27,300	158,560
Fabric Filter	Retort Discharge	164,130				27,300	191,430
Venturi Scrubber	Shale Moisturizer	462,660		38,160		95,050	595,870
Stretford System	Retort Gas	1,376,546	349,672	4,538	2,141,640	529,600	4,401,996
Ammonia Absorption System	Retort Gas	<u>365,323</u>	<u>-</u>	<u>929,978</u>	<u>-</u>	<u>54,610</u>	<u>1,349,916</u>
	Total	2,668,504	349,672	972,676	2,141,640	768,960	6,901,452
Flue Gas Desulfurization	Energy Production System Flue Gas	3,072,000		50,360		1,747,000	4,869,360

Table 5-A4
Air Pollution Control Equipment
Operating Costs 1979 Dollars
Paraho Oil Shale Process
More Strict Scenario
\$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemical</u>	<u>O & M</u>	<u>Total</u>
Fabric Filter	Primary Crushers	21,520				3,850	25,370
Fabric Filter	Raw Shale Storage Bldg.	40,200				8,300	48,500
Fabric Filter	Secondary Crushers	20,600				4,160	24,760
Fabric Filter	Surge Bins	19,020				4,160	23,180
Fabric Filter	Tertiary Crushers	88,760				18,480	107,240
Fabric Filter	Retort Feed	131,260				27,300	158,560
Fabric Filter	Retort Discharge	164,130				27,300	191,430
Venturi Scrubber	Shale Moisturizer	462,660		38,160		95,050	595,870
Stretford System	Retort Gas	1,376,546	349,672	4,538	2,141,640	529,600	4,401,996
Ammonia Absorption System	Retort Gas	<u>365,328</u>	<u> </u>	<u>929,978</u>	<u> </u>	<u>54,610</u>	<u>1,349,916</u>
	Total	2,690,024	349,672	972,676	2,141,640	772,810	6,926,822
Flue Gas Desulfurization	Energy Production System Flue Gas	3,072,000		50,360		1,747,000	4,869,360

Table 5-A5
 Air Pollution Control Equipment
 Operating Costs 1979 Dollars
 Modified In Situ Oil Shale Process
 \$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemical</u>	<u>O & M</u>	<u>Total</u>
Fabric Filter	Shale Crushing (Site Prep. Temp.)	7,140				1,400	8,540
Fabric Filter	Mine Ventilation	241,260				49,845	291,105
Fabric Filter	Mine Shaft Conveyor & Transfer Point	27,690				6,090	33,780
Stratford System	Retort Gas	3,002,910	381,400	4,951	2,335,839	429,600	6,154,700
Ammonia Absorption System	Retort Gas	<u>452,523</u>	<u> </u>	<u>709,560</u>	<u> </u>	<u>84,840</u>	<u>1,246,923</u>
	Total	3,731,523	381,400	714,511	2,335,839	571,775	7,735,048
Flue Gas Desulfurization	Energy Production System Flue Gas	4,368,508	-	71,590	-	2,012,076	6,452,174

Table 5-A6
Air Pollution Control Equipment
Operating Costs 1979 Dollars
Modified In Situ/Lurgi Oil Shale Process
\$/yr

<u>Equipment</u>	<u>Location</u>	<u>Power</u>	<u>Steam</u>	<u>Water</u>	<u>Chemical</u>	<u>O & M</u>	<u>Total</u>
Fabric Filter	Shale Crushing (Site Prep. Temp.)	7,140				1,400	8,540
Fabric Filter	Mine Ventilation	241,260				49,845	291,105
Fabric Filter	Mine Shaft Conveyor & Transfer Point	27,690				6,090	33,780
Fabric Filter	Primary Crusher	10,720				2,270	12,990
Fabric Filter	Secondary Crusher	47,390				10,580	57,970
Fabric Filter	Ore Storage	6,710				1,320	8,030
Hot Precipitator	Lurgi Retort Discharge	608,730				221,500	830,230
Venturi Scrubber	Moisturizer	122,290		10,080		28,550	160,920
Stretford System	MIS Retort Gas	3,325,577	459,452	5,905	2,799,025	577,200	7,167,159
Ammonia Absorption System	MIS Retort Gas	<u>452,523</u>	<u> </u>	<u>709,560</u>	<u> </u>	<u>84,840</u>	<u>1,246,923</u>
	Total	4,850,030	459,452	725,545	2,799,025	983,595	9,817,647

6.0 WATER POLLUTION CONTROL

6.1 OVERALL RESULTS AND CONCLUSIONS

6.1.1 Results

Oil shale retorting produces water, partly by combustion of hydrogen and oxygen, and partly due to the release of combined and free moisture in the shale. Most of this water leaves the retort in the vapor phase with the raw gas, and is recovered as a foul condensate when the gas is cooled prior to purification. Ammonia, along with carbon dioxide and some H_2S dissolves in the condensate, and some of the volatile organics in the gas will condense as well. Inorganic salts are normally not important in the gas condensate.

Some of the water produced during retorting may condense out in the retort or in the oil/gas separators. This water must be separated from the oil, an operation that may prove difficult due to emulsification and properties peculiar to shale oil. For this reason it would be useful, for example, to operate the retorting section of the Paraho process at a temperature such that the water remains in the vapor phase until the gas treatment section. We assume such operation here. However, in in situ retorting, condensation within the "retort" cannot be avoided as the vapors pass through the cold unretorted rubblized shale before being brought to the surface. This retort condensate will leach out inorganic salts, particularly bicarbonates, as it percolates to the base of the retort, and will be of a quality significantly different from that of the gas condensate.

Because of the mine location, a further water stream associated with the Modified In Situ (MIS) operation considered in this study is mine drainage water. These waters are released from aquifers at various levels (alluvial, upper and lower) in the oil shale deposits, and must be either consumed, discharged to a surface stream, or re-injected into the aquifers. As with the retort condensate, mine drainage waters contain considerable amounts of

leached inorganic salts, but are otherwise clean in the sense that they are not associated with the retorting operation and do not contain significant concentrations of dissolved gases and organics. Excess mine drainage water that is surface discharged generally will require some treatment, depending on discharge permits, but should nevertheless be regarded as a water resource and not a wastewater. Re-injection, which is not considered here, is another possible means of mine water disposal and would probably require less treatment than for surface discharge; however, in this case the water can no longer be regarded as a resource.

The production rate of the above water streams, normalized to the oil production rate, are summarized in Table 6-1 for the four plant schemes studied. Cooling tower blowdown and blowdown associated with the plant steam system can also be considered to be produced in the process, and are included in the table. Also shown are the major water requirements. In developing the water management schemes, the water produced is used, where possible, to satisfy these water requirements. Treatment is provided to achieve the quality desirable for each reuse application.

As seen from Table 6-1, the water requirements of both the Paraho and TOSCO plants exceed the quantity of water produced, and makeup water from the Colorado river must be provided. In the case of the MIS plant, excess mine drainage water is available and in the present scheme is treated and discharged. The quantity of excess water may range from 3 to 11 mgd and should prove to be a valuable commodity in this water short region. In the case where in situ retorting is combined with surface retorting, backfill of the in situ retorts with the surface retorted shale was investigated as a possible environmental control. Water requirements for slurring the spent shale and moistening the in situ retort amount to nearly 8 mgd, and may, in fact, exceed the amount of water available.

In the TOSCO II scheme the spent shale is moistened to about 14% water by mass resulting in a requirement of some 2.4 mgd of low grade water. The use of venturi scrubbers in the TOSCO II retorting section results in a further relatively high water use rate for dust control of 0.9 bbl/bbl oil as shown in Table 6-1. Replacing the venturi scrubbers with bag filters reduces this water requirement to about 0.4 bbl/bbl oil as at the Paraho plant.

Table 6-1
SUMMARY OF MAJOR WATER STREAMS AT THE FOUR OIL SHALE PLANTS
FLOWS GIVEN IN BARRELS WATER/BARREL PRODUCT OIL^(a)

<u>Stream</u>	<u>Paraho</u>	<u>TOSCO II</u>	<u>MIS</u>	<u>MIS + Lurgi</u>
<u>Water required</u>				
Cooling tower makeup	0.9	1.5	1.3	1.0
Boiler feedwater makeup ^(b)	~	0.9	1.0	0.7
Dust control	0.1	0.4	0.1	0.1
(Spent) shale disposal	0.4	1.3 ^(c)	0.1	2.3 ^(d)
Revegetation	0.2	0.2	0.3	-0-
TOTAL	<u>1.6</u>	<u>4.3</u>	<u>2.8</u>	<u>4.1</u>
<u>Water produced</u>				
Cooling tower blowdown ^(e)	0.4	0.5	0.3	0.2
Boiler blowdown	~	0.1	~	~
Retort Condensate	~	0.3 ^(f)	0.5 ^(g)	0.3 ^(g)
Gas condensate	0.2	~	0.8	0.6
Mine drainage water ^(h)	~	~	4.8	3.2
TOTAL	<u>0.6</u>	<u>0.9</u>	<u>6.4</u>	<u>4.3</u>
Net Water Produced (consumed)	(1.0)	(3.4)	3.6	0.2

- (a) Actual flow rates of all the water streams for the four plant schemes considered are shown on Figs. 6-1, 6-7, 6-14 and 6-18.
- (b) Includes steam consumed in retort.
- (c) Spent shale moistened 14% according to TOSCO scheme.
- (d) Surface retorted burned shale slurried for backfill.
- (e) Between 2-4 cycles of concentration, depending on the low quality water needs of the plant.
- (f) From oil fractionator; does not contact shale.
- (g) Condensed in retort; leaches out considerable inorganic salts.
- (h) The calculations are based upon mine water flow as listed on the Figures 6-14 and 6-18. It is realized that mine water flow could range from 4,000-10,000 gpm.

Water used for dust control and shale disposal need not necessarily be of high quality, and suitably treated process condensate or blowdown is generally satisfactory. The treatment train selected for the process condensate consists essentially of ammonia and acid gas removal units followed by organic removal. This treated water is in most designs then used to provide some of the cooling tower makeup. With the exception of the MIS/Lurgi process in which the spent shale is slurried, the largest water consumption is for cooling. The use of treated process condensate waters as makeup provides several benefits. First, the amount of fresh water makeup required is reduced, second, the treated gas condensate is expected to be low in inorganic salts so reducing scaling problems, and third, there is considerable evidence that biological reduction of the organic loading will occur in the cooling tower circuit [1-3]. Normal cooling tower operating procedures should not be followed; correct operation of organically loaded cooling circuits is discussed in References 1-6.

Blowdown from the cooling tower is blended with other blowdown streams and provides the makeup for low quality water users. The organic concentration has at this stage been reduced by a specific organics removal treatment, degradation in the cooling tower, and by dilution on blending with other blowdown streams. As mentioned, the inorganic content of the gas condensate is expected to be low, and so should not contribute significantly to the inorganic character of the low quality makeup stream. In addition, due to the relatively high demand for low quality water, cooling tower cycles have been kept low, and the total dissolved solids content of the blended low quality water stream is always less than double that in the source water (Colorado river water or mine drainage water). Even with low cooling tower cycles, the blowdown streams are generally insufficient for supplying all the low quality water needs. Consequently source water is needed as well, and may be selectively directed to those applications where continued contact with plant personnel is possible, e.g., dust control at the mine.

Toxic metals do occur in oil shale and may volatilize during the retorting process and end up in the oil, process condensate or gas stream. In addition, soluble toxic metal salts may be leached into the water in processes in which condensation occurs within the retort. In a study of trace elements in simulated in situ retorting it was found, however, that most of these

elements remained in the shale [44]. Elements that do volatilize in the combustion region were found to condense out on cooler shale lower down, only to revolatilize on approach of the flame front. If such a succession of volatilization/condensation were to occur in practice, it was conjectured that those toxic elements that do leave the retort would do so in a wave near the end of the burn, and could be more easily controlled.

On the basis of these findings and the limited available published data, it was assumed that toxic metals would probably not be present in process condensate in important concentrations. For this reason, and because of the lack of adequate published information, and in view of the fact that no condensate is discharged from the plant, a detailed study of toxic metals in process condensate was not made. Any present in the gas condensate could be expected to follow the route of the other inorganic constituents described above. In the case of the one retort condensate occurring in this study (MIS) in which inorganics and consequently toxic metals might be significant, the problem is avoided by recycling the condensate to the retort via a "thermal sludge" steam raising system. Inorganics would be removed from the cycle in the blowdown sludge streams. However, toxic metals constitute an important environmental issue and one which requires further investigation. See Section 8.0 for a more detailed discussion.

The total source water makeup for all plant needs is given in Table 6-2. The makeup quantities are calculated on the assumption that all effluent streams are reused and that, apart from excess mine drainage water, no wastewater is discharged. The water treatment costs are summarized in Section 10.0. The costs shown include makeup treatment for the cooling tower and steam boilers, source water clarification, as well as the necessary treatment of all blowdown and effluent streams. In the sections that follow, a breakdown of the costs is given with the description of the water management scheme at each plant.

6.1.2 Conclusions

Water consumed in producing shale oil or a synthetic crude is somewhat different among the various processes examined but in no case is it excessive. The values at the low end of consumption are significantly lower than that required to convert coal to liquid fuels [10] and at the high end

Table 6-2
SOURCE WATER MAKEUP REQUIRED AT THE FOUR OIL SHALE PLANTS
ALL WATER STREAMS ASSUMED TO BE REUSED FOLLOWING APPROPRIATE TREATMENT

<u>Process</u>	<u>Barrels per Barrel Actual Oil Product</u>	<u>Barrels per Barrel Syncrude</u>	<u>Barrels per Barrel* Syncrude Equivalent</u>	<u>Gallons per 10⁶ Btu* Equivalent</u>
Paraho	1.13		1.18	8.2
TOSCO II	-	3.44-3.89	3.72-4.20	26.0-29.4
MIS	1.28		1.32	9.3
MIS/Lurgi	2.91		3.01	21.0

* See Table 2-2, Section 2.0 for quantities.

are comparable. The low values of about 1.5 barrels of water per barrel of equivalent syncrude are found for the Paraho and MIS processes. Water consumptions about three times higher are found for the TOSCO and MIS/Lurgi processes because of the methods assumed to be used for the disposal of the spent shale.

Water consumption figures are based on complete recycle and reuse. It is shown that methods are, in principle, available and water management schemes can be devised to treat the water to the extent necessary for this purpose. It must be emphasized, however, that no overall treatment scheme has yet been tried out on oil shale process waters on a demonstration plant scale or in some instances even on a pilot plant scale and this should be done. In the MIS case, where there is excess mine water, methods such as reverse osmosis exist to treat the water to a condition acceptable for surface discharge, though again they have not been tested either on pilot scale or on the scale of water volumes expected for commercial production.

The treatment schemes suggested in the study are not necessarily optimal ones and further research and development is required to define the optimal procedures, particularly for the treatment of the retort water. Even with the conservative treatment methods assumed, the cost of water treatment for internal reuse and recycle is not large, with estimated costs ranging between 23¢ and 76¢ per barrel of equivalent syncrude. Only when excess mine water must be treated for surface discharge, as for the MIS case, does the cost rise to as high as \$1.74 per barrel. Even here, however, this figure may be considered an upper limit since no credit has been taken for the high quality water produced, which may prove to be a valuable commodity in the arid regions of the area.

The question of toxic elements in the produced waters arises principally in connection with the low quality waters used for surface disposal of spent shale, and to a lesser extent with the low quality waters used for dust control. This problem has not been addressed in detail, but does not appear from this study to present a problem, although further investigation would be desirable. We emphasize, however, that toxic trace elements are important environmental issues and further study is recommended.

6.2 PROCESS FLOW DIAGRAMS AND WATER BALANCES

Where available, developers' designs were used to determine overall material balances for the oil shale plants. Known or estimated composition data were then used to complete carbon and hydrogen elemental balances. This provided a check on the consistency of the data used, and more important from the water management viewpoint, enabled the water streams associated with retorting, in particular the water formed by combustion, to be estimated. Three hydrogen related values were calculated for each process stream:

- (a) hydrogen not associated with water,
- (b) moisture (H_2O),
- (c) water equivalent,

where the water equivalent is given by $9 \times (\text{hydrogen not associated with water}) + (\text{moisture})$. The water equivalent column in the material balances must always balance, as hydrogen is not created or destroyed in the process. However, there is generally a difference in the moisture column which is due to water formation by combustion of hydrogen (either in the shale or in the fuel) with oxygen. In the Paraho and MIS processes the water formed by combustion amounts to about half the net retort water produced, or about 0.15 bbl/bbl oil produced. In the TOSCO process the water produced by combustion is either lost in the flue gases, or is offset by hydrogen production in the upgrading sections.

The water oriented material balances, the resulting water management scheme, and the size and costs of the required water treatment systems, are presented for each plant in the sections which follow. All necessary pre-treatments, e.g., oil water separation, filtration, chemical addition, and pH adjustment are assumed to be provided but not necessarily detailed in the text. Background information on the two major condensate treatment processes, ammonia recovery and biological oxidation, is given in Section 6.3. Details not directly related to water management of the oil shale plants have been given in Section 3 and are not repeated here.

6.2.1 Paraho Plant

The material balance for the Paraho plant is given in Table 6-3 and the overall water balance, showing a breakdown of water supply and

Table 6-3
MATERIAL BALANCE AROUND PARAHO RETORTS, DIRECT HEAT MODE, PRODUCING
99,170 BPSD CRUDE SHALE OIL FROM 29 GAL/TON OIL SHALE.
FLOWS IN 10^3 LB/HR

<u>Stream</u>	<u>Total</u>	<u>H</u>	<u>Moisture</u>	<u>H₂O equiv</u>	<u>C</u>
<u>IN</u>					
Raw shale ^(a)	13,118	243	250	2,434	1,677
Carbonates converted ^(b)	-	-	-	-	167
Process air, wet ^(c)	2,339	-	23	23	-
TOTAL IN	15,457	243	273 ^(h)	2,457	1,844
<u>OUT</u>					
Spent shale, prior to moistening ^(d)	10,511	13	-	116	225
Product oil, "dry" ^(e)	1,352	158	20	1,442	1,138
Product gas, wet ^(f)	3,571	49	440	880	481
NH ₃ ^(g)	12	2	-	19	-
Sulfur ^(g)	11	-	-	-	-
TOTAL OUT	15,457	222	460 ^(h)	2,457	1,844

NOTES:

- (a) Raw shale value from Reference 7. According to data in Reference 8 combined water content $\approx$ free water content $\approx$ 0.95%. The dry (combined water only) shale was taken to have an organic content of 16.0%, and H₂ = 1.85%, C_{org} = 12.91% by mass.
- (b) Carbonates converted. According to Reference 8, some 24% of the carbonates are decomposed to release CO₂. The equivalent amount of C released must be added to the organic carbon.
- (c) Process air. Dry rate from Ref. 7. Moisture content assumed to be 1% by mass.
- (d) Spent shale. Value from Reference 7. Hydrogen content by balancing H₂O equivalent column gives 0.12% H₂ by mass. Carbon content by balancing C column gives C = 2.14%.
- (e) Product oil rate. Based on 99,170 BPSD product plus 830 BPSD to fuel pool; SG = 0.9267. Composition from Reference 7, C = 84.20%, H = 11.69%. According to Reference 9, product oil contains 1.5% H₂O. This oil taken to be the "dry" product. It was assumed that temperatures are adjusted such that no water condensate is removed from the oil phase.
- (f) Product gas rate. Value from Reference 7. NH₃ and S given separately. Gas composition give in Table 6-9.
- (g) Calculated from gas composition, Table 6-9.
- (h) Material balance shows that $(460-273) \times 10^3$ lb/hr or 187×10^3 lb/hr water is formed by combustion. Reference 8 data suggest 232×10^3 lb/hr water formed by combustion.

consumption, is given in Table 6-4. Tables 6-3 and 6-4 were used to construct Figure 6-1, the water management scheme for the Paraho plant. The figure shows the net water streams requiring treatment and should not be viewed as a detailed water balance. For instance the 586 gpm shown coming from the retorting and gas treatment block is made up of the water produced by combustion together with the moisture in with the shale and the combustion air, less the water leaving with the oil and product gas. These values are given in Table 6-4, together with an explanation or reference as to their origin.

Source water in the amount of 3,276 gpm is required as makeup for plant needs, most of which is used for cooling. All the source water is shown to be clarified in Figure 6-1. This is probably useful to prevent nozzle clogging, etc., even for the low quality revegetation and dust control use. The composition of the source water, clarified water, and cooling tower makeup water is shown in Table 6-5, and the cost of clarification is estimated according to Table 6-6.

Most of the plant cooling load is for cooling the gas distributors in the retort, and for cooling the retort gas prior to the Stretford purification unit. It has been assumed that the gas is not compressed upstream of the purification step, as this would substantially increase the cooling requirements. For reasons of water conservation, and possibly energy conservation, compression without cooling just upstream of the gas turbines seems preferable. See Section 3.0 for a discussion of alternatives. The cooling load is summarized in Table 6-7, together with an indication of scale control measures and treatment costs for the cooling tower makeup.

The plant steam load is shown in Table 6-8. Live steam is not used so the makeup to the boiler feedwater is only 9 gpm, which is sufficient to compensate for blowdown and losses. Treatment of the blowdown consists of strong acid cation exchange, decarbonation and strong base anion exchange. The alkali and acid regenerant wastes are mixed and taken to the equalization pond. Some reduction in the amount of regenerant waste can be had at the expense of increased system complexity by use of weak acid/strong acid cation beds in series and countercurrent regeneration. The estimated cost of boiler feedwater treatment is included in Table 6-8. These costs as well as those for cooling tower makeup treatment and source water clarification are mostly process and not environmental costs, but have nevertheless been included for

Table 6-4
OVERALL WATER BALANCE FOR PARAHO PROCESS PRODUCING 99,170 BPSD CRUDE SHALE OIL
WATER MANAGEMENT SCHEME IS SHOWN IN FIGURE 6-1

<u>IN</u>	<u>10³ lb/hr</u>	<u>gpm</u>
Source (a)	1,638	3,276
Net water produced by combustion (b)	187	374
Moisture in with shale	250	500
Moisture in combustion air	23	46
Runoff (c)	120	240
Mine water		~0
Total	<u>2,218</u>	<u>4,436</u>
 <u>OUT</u>		
Water in product oil	20	40
Spent shale and sludge moistening (5%)	525	1,050
Spent shale disposal, dust control, revegetation (d)	223	446
Cooling water evaporation and drift (e)	740	1,479
Dust control: At mine	317	633
: At plant (crushing and fugitive dust)	192	384
Potable, service; water consumed (assuming 2,600 employees (f))	26	54
Losses, condensate treatment and steam cycle (g)	28	56
Losses, product gas (h)	147	294
Total	<u>2,218</u>	<u>4,436</u>

(a) By difference.

(b) See note (g), Table 6-3.

(c) Value from Reference 7. Corresponds to ~15 inches rain/yr falling on 300 acres.

(d) Value from Reference 7. Total moisture on spent shale amounts to 71 lb H₂O/10³/lb spent shale as given in Reference 10.

(e) See Table 6-7.

(f) Based on data given in Reference 10.

(g) See Figure 6-1.

(h) See Figure 6-2 and Table 6-9.

Table 6-5
COMPOSITION OF MAKEUP WATER FOR PARAHO PLANT

Constituent	Source Water ^(a)	Clarified Source Water ^(b)	Cooling Tower Makeup ^(c)	
	mg/l	mg/l as CaCO ₃		
Na ⁺	145.0 ^(d)	334.1 ^(d)	334.1	334.1
K ⁺	5.3	6.8	6.8	6.8
Ca ⁺⁺	72.0	179.3	179.3	179.3
Mg ⁺⁺	19.0	77.9	77.9	77.9
HCO ₃ ⁻	168	137.8	124.3	37.3
Cl ⁻	205	289.0	289.0	289.0
NO ₃ ⁼	8.6	7.0	7.0	7.0
SO ₄	158	164.3	177.8	264.7
SiO ₂	7.0			
TDS	734			
pH	7.4			5.8 ^(e)

- (a) Colorado river near Cameo, Colorado. From "Quality of Surface Water of the U.S.," Parts 9 & 10, p. 33, USGS 1970.
 (b) Following 30 ppm alum dosage and clarification. See Table 6-6.
 (c) After neutralization of 80% of alkalinity with H₂SO₄. See Table 6-7.
 (d) Adjusted by 18.0 mg/l as CaCO₃ for ionic balance.
 (e) pH $\approx$ 7 in circulating cooling water.

Table 6-6
COST OF SOURCE WATER CLARIFICATION FOR PARAHO PLANT

Clarification of 3,276 gpm Colorado river water by addition of 30 ppm alum and two hours retention in a 74 ft diameter clarifier.

Installed clarifier cost:	\$208,000
<u>Operating Cost</u>	<u>\$/yr</u>
Maintenance. 4% of capital investment	8,300
Chemicals	<u>28,400</u>
TOTAL OPERATING COST	36,700

If capital is amortized @ 15%/yr, total cost = 4.4¢/10³ gal source water treated.

Table 6-7
COOLING TOWER MAKEUP TREATMENT FOR PARAHO PLANT

<u>Plant Cooling Load</u>	<u>Circulating Water</u> ($\Delta t = 30^{\circ}\text{F}$)	<u>Water Evaporated</u> (1400 Btu/lb evap)
	<u>gpm</u>	<u>gpm</u>
Retort gas distributors ^(a)	28,200	604
Retort gas cooling prior to NH_3 scrub ^(b)	24,640	528
NH_3 scrubber wash water cooler ^(b)	3,550	76
Stripped condensate cooler ^(c)	2,875	51
NH_3 recovery, Phosam W ^(d)	<u>9,920</u>	<u>212</u>
TOTAL	69,185	1,471
Cooling tower blowdown ^(e)		1,096 gpm
Makeup, allowing for ~0.5% drift loss		2,575
Cycles of concentration		2.3
Acid addition to prevent CaCO_3 scaling		= 70% of alkalinity ^(f) = 482 tons H_2SO_4 /yr = \$24,100/yr = $24/10^3$ gal
Cost of acid at \$50/ton = \$24,100/yr		
Estimated total cost of cooling water treatment		= $204/10^3$ gal ₃ /yr = $\$224.7 \times 10^3$ /yr

(a) From Reference 7.

(b) See Figure 6-2.

(c) Cooling prior to biological oxidation may be effected in part by spray cooling in the equalization pond, but would result in greater evaporation losses and be environmentally less acceptable due to windage losses of the organically contaminated water.

(d) See Figure 6-3.

(e) Blowdown sufficient for low quality plant uses. See Figure 6-1

(f) Assuming all makeup water is of composition shown under "clarified source water" in Table 6-5. As shown in Figure 6-1, the source water is diluted with some treated process condensate. CaSO_4 scaling of the acidified source water will occur only after about 8 cycles of concentration.

Table 6-8
BOILER FEEDWATER TREATMENT FOR PARAHO PLANT

<u>Plant Steam Load</u>	<u>10³ lb/hr</u>	<u>gpm</u>
NH ₃ recovery, Phosam-W ^(a) 600 psi	169	338
60 psi	36	72
Stretford gas treatment ^(b)	20	40
Miscellaneous uses ^(c)	100	200
	325	650

Boiler feedwater makeup, for 1% blowdown, 1% losses = 9 gpm

Cost of Boiler Feedwater Demineralization

Capital investment	\$234,000
Operating costs	\$ 17,600/yr

With amortization of capital @ 15% (\$35,100/yr) this would amount to about \$9/10³ gal treated. (d)

- (a) See Figure 6-3.
- (b) Value given in Reference 7.
- (c) Estimated from difference between total and specified uses, Reference 7.
- (d) Including regenerant blowdown of 3.6 gpm. High cost due to relatively small volumes treated--70% of cost for capital amortization and labor.

Figure 6-1. Major water streams for Paraho direct heated process producing 99,170 BPSD crude shale oil from 29 gal/ton shale. Recirculating streams are not shown.

completeness. If desired, the reader may back out the process portion of the total costs to obtain an environmental control cost.

The condensate stream from the retort gas cooler contains dissolved ammonia, carbon dioxide and hydrogen sulfide along with condensed organics and possibly some inorganic salts. Several analyses of the Paraho retort water have been published [12-14]. The published data have a wide range in concentration probably reflecting differences in retort operating conditions, shale, and the relative quantities of retort and gas condensate. In the scheme presented here it has been assumed that temperatures in the gas/oil separators are controlled to virtually eliminate the retort condensate leaving with the oil. The gas condensate from the gas coolers is then the only process condensate requiring treatment. Rather than use published data for the dissolved gas content, these concentrations have been calculated from the gas composition shown in Table 6-9. Liquid composition was determined using van Krevelen [11] vapor-liquid equilibrium data for the system $\text{NH}_3\text{-CO}_2\text{-H}_2\text{S-H}_2\text{O}$, and based on the flow scheme shown in Figure 6-2. Note that the figure is included to show the general flow scheme used for the equilibrium calculations; the detailed gas cooling/scrubbing scheme is discussed in Section 5.0.

As seen in Table 6-4, ammonia is produced at a rate of about 12,000 lb/hr or 146 t/d and, if recovered, can be sold to partly offset the cost of its removal from the condensate. Several processes may be used, including the Chevron [15] and Phosam-W [16] processes. In the Chevron process, which has been suggested by the Paraho developers for ammonia recovery, the H_2S is preferentially stripped from the water prior to ammonia stripping. Due to the high CO_2 concentration in the retort gas, however, H_2S concentrations in the liquid phase are low (~ 200 mg/l), so the Phosam-W process has been presented in this study. The selection of either process is not expected to significantly affect the overall water treatment costs.

The flow scheme for the Phosam-W process, including influent and effluent concentrations of the major streams, is shown in Figure 6-3. A detailed description of the process and design procedures for the water stripper and ammonia fractionator are available elsewhere [17, 18]. Design procedures for the ammonium phosphate absorber and stripper are proprietary to USS Engineers and Consultants (UEC). Sizes of these units were estimated on the basis of designs supplied by UEC (see e.g. Ref. 17), and used for the cost

Table 6-9
COMPOSITION OF PARAHO DIRECT HEATED MODE RETORT GAS

Values taken from Reference 7, but adapted (a) to allow for higher water content resulting from higher temperatures used to prevent water from condensing with oil, and (b) to allow for reportedly (DRI private communication, December 1978) higher NH_3 and H_2S concentrations. Composition on wet basis. Values given are as used in calculations and are not intended to reflect the accuracy of the analyses.

Constituent	Mwt	Raw Retort Gas		Cooled & Scrubbed Gas	
		Mass %	Vol. %	Mass %	Vol. %
H_2	2	0.249	3.539	0.274	4.107
O_2	32	0.836	0.743	0.922	0.862
N_2 + Argon	28	50.564	51.342	56.390	60.027
CH_4	16	1.046	1.858	1.152	2.156
CO	28	1.576	1.600	1.737	1.857
CO_2	44	27.195	17.572	28.326	19.270
H_2S	34	0.289	0.242	0.304	0.268
NH_3	17	0.338	(0.3% dry) 0.566	0.0005	9 ppm
$\text{C}_2^=$	28	0.732	(0.7% dry) 0.743	0.806	0.862
C_2	30	0.810	0.768	0.893	0.891
C_3	42	0.538	0.364	0.592	0.422
C_3	44	0.588	0.380	0.648	0.441
C_4	57	0.648	0.323	0.714	0.375
C_5	71.5	0.324	0.129	0.358	0.150
C_6^+	96.2	2.105	0.622	0.320	0.722
H_2O	18	12.162	19.201	4.564	7.590
TOTAL		100.00	100.00	100.00	100.00
TOTAL C		13.49		14.42	
TOTAL H_2^*		1.47		1.56	
Molecular wt.		28.43		29.93	

* Excluding H_2 in H_2O

	IN			OUT	
	10 ³ lb/hr	ppm (Vol)		10 ³ lb/hr	ppm (Vol)
NH ₃	12.2	5,650	NH ₃	0.016	9
CO ₂	984.0	175,000	CO ₂	960.0	132,600
H ₂ S	10.5	2,420	H ₂ S	10.3	2,680
H ₂ O	440.00		H ₂ O	141.0	

FIGURE 6-2. PRODUCT GAS COOLING AND AMMONIA SCRUBBING FOR PARAHO PLANT PRODUCING 99,170 BBL/DAYS CRUDE SHALE OIL.

Figure 6-3. Ammonia separation by Phosam-W for Paraho plant producing 99,170 bbl/day crude shale oil.

estimate presented in Tables 6-10 and 6-11, but have not been specified in Table 6-11.

Volatile organics stripped in the water stripper of the Phosam-W plant may be carried over to the ammonia fractionator. Caustic soda is added to the fractionator column to hold down these organics which are then returned with the bottoms to the water stripper, and exit in the treated water along with organics that are not stripped. The organics must be removed from the water prior to its reuse.

According to Cook [19], organic acids in process water from Green River oil shale are mainly highly biodegradable straight chain carboxylic acids, and the phenols content is low. Further, some preliminary data do suggest that biological oxidation is a feasible option for organics removal [20]. Poor performance could result in practice if the condensates treated contain a high inorganic salt content (see Table 6-33), or high $\text{NH}_3/\text{H}_2\text{S}$ levels resulting from inadequate stripping. However, as in the design considered here, the process waters leave the retort in the vapor phase, they do not leach inorganic salts. Further, adequate stripping is provided and it has consequently been assumed that biological oxidation is a feasible treatment procedure. A suitable process scheme is shown in Figure 6-4. Although at least one investigation is in progress in which design parameters are being determined [21], results were not available at the time of writing. The costs and equipment sizes presented in Tables 6-13 and 6-14 for the system shown in Figure 6-4 were based on design data for condensate from the H-Coal process [22].

As seen from Table 6-12, the estimated cost for biological oxidation is $\$6.7/10^3$ gallons. This is excessive and is a consequence of the conservative biokinetic constants used for plant design. While it is anticipated that actual costs will be less, if, in fact, costs are high, alternative processes will be used. Extraction, resin or carbon adsorption and wet air oxidation, either alone or in combination, may be used. Costs for these processes should not exceed that shown for biological oxidation, which may be regarded as an upper limit for organics removal.

Following organics removal, the condensate stream is blended with Colorado River source water and used as cooling tower makeup for organics polishing. Cooling tower blowdown along with other blowdown streams as shown

Table 6-10
AMMONIA RECOVERY FROM PARAHO RETORT GASES

The treatment scheme is shown in Figures 6-2 and 6-3.

	<u>gpm</u>
Volume of condensate collected on cooling gas ^(a)	586
Volume of wash water from NH ₃ scrub ^(a)	<u>704</u>
Total feed rate to NH ₃ recovery system ^(b)	1,290
Capital investment, as per Table 6-11	$\$6.3 \times 10^6$
<u>Operating costs</u>	<u>10³ \$/yr</u>
Maintenance and labor	387
Steam @ \$3/10 ⁶ Btu	5,105
Cooling water at 12¢/10 ³ gal circulated	485
Chemicals, electricity	<u>145</u>
TOTAL OPERATING COST	6,122
Credit, NH ₃ sales; 146 t/d @ \$100/ton	4,818
NET OPERATING COST	<u>1,304^(c)</u>

If capital amortization @ 15% is included (\$946,000/yr), total net cost per 1,000 gal water treated: $\$3.6/10^3$ gal^(c)

- (a) See Figure 6.2 for gas cooling and scrubbing scheme. Concentrations calculated using van Krevelen [11] equilibrium data.
- (b) Feed composition and NH₃ recovery scheme shown on Figure 6-3.
- (c) The cost of the Phosam-W process was scaled from data available for similar, but not identical, feed conditions. USS Engineers and Consultants, Inc. can provide cost optimized plant designs for specific feed conditions that would reduce or might eliminate the net cost shown. The high cost obtained here results from the high steam usage in combination with an energy cost of \$3/10⁶ Btu, or about \$20/bbl crude shale oil.

Table 6-11
MAJOR EQUIPMENT AND ESTIMATED COSTS FOR
PHOSAM-W AMMONIA RECOVERY AT THE PARAHO PLANT

<u>Vessels and Columns</u>	<u>Size</u>	<u>Material</u>	<u>Cost 10³\$</u>
Water stripper	9' diameter, 80' shell + 15' skirt	CS	90
Phosam absorber	--	CS/SS	392
Phosam stripper			
NH ₃ fractionator	4' diameter 70' shell + 15' skirt	CS/SS	190
<u>Heat Exchangers</u>	<u>Area ft²</u>		
Lean solution cooler	10,400	CS/SS	149
Solution exchanger	2,000	SS	66
Stripper condenser	7,500	SS	170
Fractionator condenser	4,500	CS	49
Water stripper reboiler	6,000	CS/SS	63
Absorber cooler	6,000	CS/SS	62
Phosam stripper reboiler	4,800	CS/SS	78
Fractionator reboiler	1,600	CS/SS	21
Miscellaneous equipment (flash drums, storage tanks, etc.)			86
Total equipment cost			1,416
Installation cost (including pumps, instrumentation)			3,890
Total installed cost			5,306
Royalty			1,000
Total Capital Investment			6,306

Table 6-12
ORGANICS REMOVAL FROM STRIPPED CONDENSATE AT PARAHO PLANT

The biological oxidation^(a) scheme is shown in Figure 6-4. It was sized on the basis of 10,000 mg/l BOD in the influent water^(b), and biokinetics constants established for H-Coal process condensate. An effluent concentration of 50 mg/l BOD (99.5% removal of BOD) was selected for the design. Feed rate = 540 gpm.

Capital investment, as per Table 6-13 \$13.85 × 10⁶

<u>Operating Cost</u>	<u>10³ \$/yr</u>
Labor, Maintenance	300
Electricity 1 MW @ 3¢/kWhr	238
Nutrients	<u>70</u>
TOTAL OPERATING COST	<u>608</u>

If capital amortization @ 8% is included (\$1,100,000), the total cost per 1,000 gal water treated = \$6.7/10³ gal.

- (a) Biological oxidation has yet to be demonstrated as a feasible process for organics removal from oil shale process condensates. It is possible that inhibiting agents may be present and would require removal prior to the biological step. Laboratory studies are being performed by Water Purification Associates under a grant from the Laramie Energy Research Center, DOE [21]. A very conservative design based on reported H-Coal process condensate biokinetic coefficients has been presented here, and the cost should be regarded as an upper limit of the same order as other normally more expensive treatments, e.g., wet air oxidation.
- (b) Based on several published water compositions. See e.g., References 12-14.

Table 6-13
MAJOR EQUIPMENT AND ESTIMATED COSTS FOR BIOLOGICAL
OXIDATION AT THE PARAHO PLANT

	<u>Size</u>	<u>Cost 10³\$</u>
Aeration basin	1.77 × 10 ⁶ ft ³	11,500
Clarifier	44 ft diameter	100
Vacuum filter	420 ft ²	400
Equalization basin	100,000 ft ³	100
Aerators		500
Pumps & miscellaneous		<u>1,250</u>
TOTAL COST		13,850

Table 6-14
MISCELLANEOUS WATER MANAGEMENT COST, PARAHO PLANT

<u>Treatment</u>	<u>Flow gpm</u>	<u>Capital Investment 10³\$</u>	<u>Operating Cost 10³ \$/yr</u>
Oil water separation ^(a)			
Process condensate	586	60	-
Runoff	272	30	-
"Domestic" waste ^(b)	28	47	0.8
Equalization basin, plant blowdown streams (3 day retention)	1,378	5,600	-
TOTAL:		<u>5,737</u>	<u>0.8</u>

- (a) It is assumed that the oil in the gas condensate and runoff can be separated in conventional API separators, and that emulsification is not a problem.
- (b) Package biological unit. Draws approximately 3 kW.

Figure 6-4. Organics removal by biological oxidation for Paraho plant producing 99,170 BPSD crude shale oil.

in Figure 6-1, is taken to an equalization pond from where it is distributed as needed for dust control, shale disposal and revegetation. Cost of the equalization pond is included in Table 6-14 along with other miscellaneous treatment costs. These include the domestic waste treatment (package biological unit) and oil water separators.

A summary of the costs for the treatment schemes considered for the Paraho plant is given in Table 6-15.

6.2.2 TOSCO II Plant

Two retorting schemes were considered. In the first, shown in Figure 6-5, the developer's design [23, 24] using venturi scrubbers for dust control was followed. In Section 5 it was shown that improved air pollution control could be achieved by replacing the venturi scrubbers with a bag filter and electrostatic precipitator. This system significantly reduces the amount of water vapor lost up the retort stacks, as seen in Figure 6-6.

The TOSCO II process includes partial upgrading of the shale oil to a synthetic crude, and the material balances shown in Tables 6-16 and 6-17 for the two schemes considered, include the hydrotreating operation. The material balance data was used to determine the retort water streams shown in Tables 6-18 and 6-19, and summarized in balance form in Table 6-20. As seen, the water balance is reduced by 674 gpm by eliminating the venturi scrubbers and this is reflected in the drop in source water requirement from 5,825 gpm to 5,151 gpm shown in Figures 6-7 and 6-8, respectively. Sources of other water streams shown in these figures are given in Table 6-21, the overall water balance for the TOSCO II process.

As for the Paraho plant, source water is from the Colorado River. Its composition is repeated in Table 6-22 along with composition changes on clarification and pretreatment for the cooling circuit. Clarification costs are shown in Table 6-23 for the two source water flow rates, and cooling water treatment/costs in Table 6-24. The plant steam load, estimated from data in Reference 25, is presented in Table 6-25 along with boiler feedwater demineralization costs.

The process condensate from the retorting section contains only about 5,000 mg/l NH_3 [28], compared to the calculated 18,000 mg/l in the

Table 6-15
SUMMARY OF WATER TREATMENT COSTS AT THE PARAHO PLANT

	<u>Capital</u> <u>10³ \$</u>	<u>Operating</u> <u>10³ \$/yr</u>
Source water clarification	208	36.7
Cooling water treatment		224.7
Boiler feedwater demineralization	234	17.6
Process condensate: NH ₃ removal	6,306	1,304 ^(a)
: Organics removal	13,750	608
Miscellaneous (Table 6-15)	<u>5,737</u>	<u>0.8</u>
TOTALS	<u>26,235</u>	<u>2,191.80</u>

(a) Includes byproduct credit of \$4,818,000.

Table 6-16
MATERIAL BALANCE AROUND TOSCO II PROCESS PRODUCING 47,000 BPSD
UPGRADED SHALE OIL FROM 35 GAL/TON OIL SHALE. FLOWS IN 10^3 LB/HR

Stream	Total	H ₂	Moisture	H ₂ O equiv	C
IN					
Raw shale, 35 gal/ton ^(a)	5,465	116	146	1,189	902
Water to pyrolysis ^(b)	1,190	-	1,190	1,190	-
Steam to pyrolysis ^(c)	195	-	195	195	-
Steam to H ₂ production ^(d)	246	-	246	246	-
Oxygen for H ₂ combustion, pyrolysis ^(e)	125	-	5	5	-
Water to coker ^(f)	50	-	50	50	-
TOTAL IN	7,271	116^(w)	1,832^(w)	2,875	902
OUT					
Spent shale & sludge (dry) ^(g)	4,514	6	-	57	158
Water evaporated on shale ^(h)	250	-	250	250	-
Water leaving with spent shale and sludges ⁽ⁱ⁾	632	-	632	632	-
Water in scrubber stack gases ^(j)	540	-	540	540	-
Foul water, pyrolysis ^(k)	213	-	213	213	-
Foul water, coker, gas treating ^(m)	36	-	36	36	-
Compressor condensate ^(p)	23	-	23	23	-
Fuel burned ^(q)	138	22	-	201	110
Coke (product) ^(q)	67	2	5	27	61
NH ₃ ^(s)	27	2	-	18	-
LPG ^(s)	32	6	-	52	26
Product oil ^(t)	552	76	-	686	476
Diesel fuel to mine ^(u)	6	1	-	7	5
CO ₂ released ^(v)	241	-	133	133	66
TOTAL OUT	7,271	115^(w)	1,832^(w)	2,875	902

NOTES:

- (a) Raw shale. Reference 23 gives 61,000 t/cd or 66,000 t/sd. Shale dust lost = 425 t/sd. Rate to retort = 65,575 t/sd or $5,465 \times 10^3$ lb/hr. Reference 24 gives C_{org} = 16.5%, H₂ = 2.12%, and combined water = 1.15%. Reference 25 gives surface moisture as 1.52%. Total moisture = 2.67% or 146×10^3 lb/hr.
- (b) Water to pyrolysis. See Tables 6-18 and 6-20.
- (c) Steam to pyrolysis. References 23 and 25 give 195×10^3 lb/hr. Fed to pyrolysis drum.
- (d) Steam to H₂ production. Reference 23 gives 246×10^3 lb/hr. About 224×10^3 lb/hr is consumed (see Table 6-25), the remainder appears as condensate.
- (e) Oxygen required for H₂ combustion, pyrolysis. This oxygen enters with air at ball heater, and leaves system as water. See under 7 in Table 6-18.

- (f) Water to coker unit. Reference 24 gives 50×10^3 lb/hr.
- (g) Spent shale and sludge. Reference 23 gives $4,514 \times 10^3$ lb/hr on a dry basis. Balancing C and H_2O columns gives $C_{org} = 3.5\%$, $H_2 = 0.13\%$. Reference 23 suggests $C_{org} = 4.49\%$ and $H_2 = 0.44\%$.
- (h) Water evaporated on shale. This is the water vaporized in moisturizing and dust control of the spent shale. Reference 23 gives 500 gpm or 250×10^3 lb/hr.
- (i) Water on spent shale. Taken to be 14% of dry shale and sludge rate.
- (j) Water in scrubber stack gases. (Shale preheat and ball elutriator stacks). See Table 6-18.
- (k) Foul water, pyrolysis. Reference 23 gives 425 gpm or 213×10^3 lb/hr.
- (m) Foul water, coker and gas treater. Data not found in literature. Value given here obtained by balancing 'total' column of material balance.
- (n) Compressor condensate, hydrogen production section. Reference 23 gives 45 gpm or $\sim 23 \times 10^3$ lb/hr.
- (p) Fuel burned. Combustion rates from Reference 23.
 - (1) Fuel gas. $1,448 \times 10^6$ Btu/hr at 22,000 Btu/lb or 65.8×10^3 lb/hr gas burned. C and H_2 content of fuel gas estimated from composition in Reference 24. $C = 73\%$, $H_2 = 17.5\%$ by mass.
 - (2) Fuel oil. 848×10^6 Btu/hr at 19,500 Btu/lb, or 43.5×10^3 lb/hr fuel oil burned. Composition from Reference 27, $C = 86.1\%$, $H_2 = 13.8\%$.
 - (3) C_4 liquids. 615×10^6 Btu/hr at 21,200 Btu/lb, or 29×10^3 lb/hr C_4 liquids burned. If mean composition is $C_4H_{9.5}$, $C = 83.5\%$, $H_2 = 16.5\%$.
- (q) Coke. Reference 23 gives 800 t/d or 67×10^3 lb/hr. Reference 26 gives composition as $C = 91\%$, $H = 3.6\%$ (dry) and moisture = 7%.
- (r) Ammonia and sulfur. Reference 23 gives 11.3×10^3 lb/hr NH_3 and 16.1×10^3 lb/hr S.
- (s) LPG. Reference 23 gives 4,300 bbl/day. Data for propane (Reference 27); $C = 81.7\%$, $H_2 = 18.3\%$, S.G. = 0.508.
- (t) Product oil. Reference 23 gives 47,000 bbl/day, $API^\circ = 44$ (SG = .806). Composition for upgrade shale oil from Reference 27, $C = 86.1\%$, $H_2 = 13.8\%$.
- (u) Diesel fuel to mine. 0.3 gal fuel/ton shale mined. S.G. = 0.9.
- (v) CO_2 released, hydrogen production section. Reference 24 gives $CO_2 = 1,300$ t/d or 108×10^3 lb/hr and accompanying water vapor as 1,600 t/d or 133×10^3 lb/hr.
- (w) The H and moisture columns need not necessarily balance as some steam is converted to hydrogen in steam reforming and some H converted to water in combustion. The fact that they do balance indicates that, within the accuracy of the material balance, the hydrogen formed by steam reforming equals the hydrogen combusted in the ball heater.

Table 6-17
MATERIAL BALANCE FOR TOSCO II PROCESS IN WHICH VENTURI SCRUBBERS
HAVE BEEN REPLACED BY BAG FILTERS. OIL PRODUCTION AS FOR TABLE 6-16.
FLOWS IN 10^3 LB/HR^(a)

Stream	Total	H ₂	Moisture	H ₂ O	C
IN					
Raw shale, 35 gal/ton	5,465	116	146	1,189	902
Water to pyrolysis ^(b)	853	-	853	853	-
Steam to pyrolysis	195	-	195	195	-
Steam to H ₂ production	246	-	246	246	-
Oxygen for H ₂ combustion, pyrolysis	125	-	5	5	-
Water to coker	50	-	50	50	-
TOTAL IN	6,934	116	1,495	2,538	902
OUT					
Spent shale & sludge (dry) ^(c)	4,514	6	-	57	158
Water evaporated on shale ^(c)	235	-	235	235	-
Water leaving with spent shale	632	-	632	632	-
Water in scrubber stack gases ^(d)	218	-	281	218	-
Foul water, pyrolysis	213	-	213	213	-
Foul water, coker, gas treating	36	-	36	36	-
Foul water, compressor condensate	23	-	23	23	-
Fuel burned	138	22	-	201	110
Coke (product)	67	2	5	27	61
NH ₃ + S	27	2	-	18	-
LPG	32	6	-	52	26
Product oil	552	76	-	686	476
Diesel fuel to mine	6	1	-	7	5
CO ₂ released	241	-	133	133	66
TOTAL OUT	6,934	115	1,495	2,538	902

NOTES:

- (a) Flow rates determined as in Table 6-16 notes, unless otherwise noted below.
- (b) Water to pyrolysis. See Tables 6-19 and 6-20.
- (c) Water evaporated on shale. The 500 gpm used in Table 6-16 less the 30 gpm used in the venturi scrubber, Figure 6-5, gives 470 gpm or 235×10^3 lb/hr.
- (d) Water in scrubber stack gases. (Shale preheat and ball elutriator stacks). See Table 6-19. In Figure 6-6 the ball elutriator stack gas is shown combined with the moisture stack gas.

Table 6-18
WATER STREAMS AROUND TOSCO II RETORT FOR PROCESS
PRODUCING 47,000 BPSD OF UPGRADED SHALE OIL

	gpm
1. Water out in preheat stack. Reference 23 gives gas rate as 1,272,000 acfm at 127°F. If gas is 90% saturated, then water rate in gas is:	810*
2. Water out in ball circulation stack. Reference 23 gives gas rate as 265,800 ACFM at 146°F. If 90% saturated, water rate in gas is:	270*
3. Water out in shale moisturizer stack. Reference 23 gives gas rate as 266,400 ACFM at 184°F. Referemces 23-25 give 500 gpm in the gas stream, which is equivalent to 70% saturated:	500*
4. Moisture from fractionator.	
(a) With retort gas. Using 70 lb gas/2000 lb shale (Reference 25, p. 10) gives 191×10^3 lb/hr gas. If saturated at 120°F, water rate is:	40*
(b) Foul water. Reference 23 and 24 give:	425
(c) With naphtha. Estimated:	32
5. Water with sludges. Assuming wet sludge is 38% H ₂ O. Dry sludge rates from Reference 23:	
(a) Preheat scrubber. 860 t/d dry sludge	88
(b) Ball circulation scrubber, 65 t/d dry sludge	7
(c) Shale moisturizer. 43 t/d dry sludge	5
6. Process steam (References 23 and 24)	390
7. Water of combustion. Reference 23, p. 40 gives fuel rate to pyrolysis. From the hydrogen content of the respective fuels, the equivalent water is calculated to be 135×10^3 lb/hr or 270 gpm. Allowing for moisture in combustion air, and in accordance with Reference 23:	280
NOTE: The equivalent oxygen in the moisture formed by combustion must be included in the mass balance. This amounts to $135 \times 10^3 \times \text{lb/hr H}_2\text{O}$ or $120 \times 10^3 \text{ lb/hr O}_2$. The extra moisture allowed for with the combustion air is $5 \times 10^3 \text{ lb/hr}$.	
8. Makeup water to the pyrolysis unit. By balance (see Figure 6-5), makeup water is 740 gpm. This water is distributed to the venturi scrubbers in proportion to their dry sludge rates:	
(a) Water to shale heater scrubber	660
(b) Water to ball circulation scrubber	50
(c) Water to moisturizer scrubber	30
9. Water to shale moisturizer. If total shale and sludge contain 14% moisture: water out with shale = $0.14 \times 4,514 \times 10^3 = 632 \times 10^3 \text{ lb/hr}$ or 1,264 gpm. Water to moisturizer = moisture out with shale and moisture in moisturizer stack - moisture in sludges - moisture to moisturizer venturi scrubber = $(1,264 + 500 - 88 - 7 - 30)$:	1,639

(a) Data essentially as given in References 23-25; see also Reference 26.

*Total evaporated water is 1,620 gpm as given in Reference 24.

Table 6-19
WATER STREAMS AROUND MODIFIED TOSCO II RETORT IN WHICH
VENTURI SCRUBBERS HAVE BEEN REPLACED BY BAG FILTERS

	<u>gpm</u>
1. Water out in preheat stack. Assume the surface moisture in the shale (138 gpm) and water formed by combustion (270 gpm) are distributed to the preheat stack and ball circulation stack in the same proportion as in Table 6-18; 75% in preheat stack, 25% in ball elutriator stack,	
$0.75 \times (138 + 270) =$	327
2. Water in ball elutriator stack gases,	
$0.25 \times (138 + 270) =$	109
3. Moisturizer scrubber stack. The 500 gpm from Table 6-18 less the 30 gpm previously fed to the venturi scrubber	470
4. Water to shale moisturizer. This is now the only makeup water stream to the pyrolysis unit. By balance (Fig. 6-6) makeup water =	1,705

Other streams as per Table 6-18.

Table 6-20
WATER BALANCE FOR TOSCO II RETORT FOR PROCESS
PRODUCING 47,000 BPSD UPGRADED SHALE OIL

	Developer's Design ^(a) gpm	Modified Design ^(b) gpm
<u>IN</u>		
Combined and surface water with shale	292	292
Water formed by combustion, ball heater	280	280
Process steam	390	390
Venturi scrubbers: Preheater	660	-
Ball circulation	50	-
Moisturizer	30	-
Spent shale moisturizer	<u>1,639</u>	<u>1,705</u>
TOTAL IN	3,341	2,667
<u>OUT</u>		
On spent shale and sludge	1,264	1,264
In stack gases: Preheater	810	327
Ball circulation	270	109
Moisturizer	500	470
In raw retort gas	40	40
Foul water separated	425	425
In naphtha	<u>32</u>	<u>32</u>
TOTAL OUT	3,341	2,667

- (a) Using venturi scrubbers for dust control as in Figure 6-5. Data as in Tables 6-16 and 6-18.
- (b) Dust control by bag filters as in Figure 6-6. Data as in Tables 6-17 and 6-19.

Table 6-21
OVERALL WATER BALANCE FOR TOSCO II PROCESS
PRODUCING 47,000 BPSD UPGRADED SHALE OIL

	Developer's Design ^(a)		Modified Design ^(b)	
	10 ³ lb/hr	gpm	10 ³ lb/hr	gpm
<u>IN</u>				
Source ^(c)	2,912	5,825	2,576	5,151
Moisture in with shale ^(d)	146	292	146	292
Net water produced by combustion ^(e)	-	-	-	-
Runoff ^(f)	55	110	55	110
Mine water	-	-	-	-
TOTAL IN	3,113	6,227	2,777	5,553
<u>OUT</u>				
Moistened spent shale ^(d)	632	1,264	632	1,264
Moisture out with coke ^(d)	5	10	5	10
Spent shale ^(g) Disposal dust control	125	250	125	250
Revegetation	178	355	178	355
Dust control ^(g) At mine	175	350	175	350
At plant	130	260	130	260
Cooling water, evaporation and drift ^(h)	765	1,530	765	1,530
Potable, service; water consumed (1,300 employees) ^(g)	12	24	12	24
Losses, condensate treatment ⁽ⁱ⁾	26	52	26	52
Losses, consumption, steam cycle ^(j)	142	286	143	286
Losses, process stacks ^(k)	923	1,846	586	1,172
TOTAL OUT	3,113	6,227	2,777	5,553

- (a) Retort water shown in Figure 6-5; overall water management in Figure 6-7.
(b) Retort water shown in Figure 6-6; overall water management in Figure 6-8.
(c) By balance on Figures 6-7 and 6-8.
(d) See Table 6-16.
(e) Note (w), Table 6-16 shows that net water formed by combustion is zero.
(f) Based on 15" rain/yr falling on 175 acres, 80% collection.
(g) Values based on data in Reference 10.
(h) See Table 6-24.
(i) See Figures 6-7, 6-9 and 6-10.
(j) Table 6-25 shows unspecified steam losses as 282 gpm. Values used here (286 gpm) to close balance.
(k) See Tables 6-16 and 6-17; sum of scrubber stack gases, water evaporated on shale, and water released with the CO₂ purge in the hydrogen production section.

Table 6-22
COMPOSITION OF MAKEUP WATER FOR TOSCO II PLANT

Constituent	Source Water ^(a)		Clarified Source Water ^(b)	Cooling Tower Makeup ^(c)
	mg/l	mg/l as CaCO ₃	mg/l as CaCO ₃	
Na ⁺	145.0 ^(d)	334.1 ^(d)	334.1	334.1
K ⁺	5.3	6.8	6.8	6.8
Ca ⁺⁺	72.0	179.3	179.3	179.3
Mg ⁺⁺	19.0	77.9	77.9	77.9
HCO ₃ ⁻	168	137.8	124.3	24.9
Cl ⁻	205	289.0	289.0	289.0
NO ₂ ⁻	8.6	7.0	7.0	7.0
SO ₄ ⁻	158	164.3	177.8	290.1
SiO ₂	7.0			
TDS	734			
pH	7.4			5.4 ^(e)

- (a) Colorado river near Cameo, Colorado. From "Quality of Surface Water of the U.S.", Parts. 9 & 10, p. 33, USGS 1970.
- (b) Following 30 ppm alum dosage and clarification. See Table 6-24.
- (c) After neutralization of 80% of alkalinity with H₂SO₄. See Table 6-25.
- (d) Adjusted by 19.0 mg/l as CaCO₃ for ionic balance.
- (e) pH $\approx$ 7 in circulating cooling water.

Table 6-23
COST OF SOURCE WATER CLARIFICATION FOR TOSCO II PLANT

Clarification by addition of 30 ppm alum and two hours retention in a clarifier.

	Developer's Design (a)	Modified Design (b)
Source water flow rate, gpm	5,725	5,151
Clarifier, diameter, ft	69	65
<u>Cost of installed clarifier, 10³\$</u>	384	358
<u>Operating Cost</u>	<u>10³\$/yr</u>	<u>10³\$/yr</u>
Maintenance; 4% of capital investment	15	14
Chemicals	<u>44</u>	<u>44</u>
TOTAL OPERATING COST	59	58
If capital is amortized @ 15%/yr, total cost, per thousand gal source water treated	4.3¢	4.6¢

- (a) See Figures 6-5 and 6-7.
(b) See Figures 6-6 and 6-8.

Table 6-24
COOLING TOWER MAKEUP TREATMENT FOR TOSCO II PLANT

Plant cooling load ^(a) :	1,520 gpm evaporated
Cooling tower blowdown ^(b)	724 gpm
Makeup, allowing for ~ 0.5% drift loss	2,254 gpm
Cycles of concentration	3.1
Acid addition to prevent CaCO ₃ scaling	= 80% of alkalinity
	= 482 ton H ₂ SO ₄ /yr
Cost of acid @ \$50/ton = \$24,100/yr. or 2.3¢/10 ³ gal makeup.	
Estimated cost of cooling water treatment	= 20¢/10 ³ gal
	= \$214 × 10 ³ /yr

(a) Cooling load according to Reference 25.

(b) Sufficient water for shale moistening--see Figure 6-7.

Table 6-25
BOILER FEEDWATER TREATMENT FOR TOSCO II PLANT

<u>Plant Steam Load^(a)</u>	<u>10³ lb/hr</u>	<u>gpm</u>
Total steam raised	1,373	2,745
<u>Consumptive steam users</u>		
Pyrolysis	195	390
Hydrogen production	224	448
Unspecified, and losses	<u>141</u>	<u>282</u>
Total consumption	560	1,120
Blowdown	108	215
Makeup to boilers	<u>668</u>	<u>1,335</u>
Demineralizer regenerant	<u>55</u>	<u>110</u>
TOTAL MAKEUP TO DEMINERALIZERS	<u>723</u>	<u>1,445</u>

Cost of boiler feedwater demineralization

<u>Capital Investment</u>	$\$2.05 \times 10^6$
<u>Operating Costs</u>	
Labor, maintenance	$\$85 \times 10^3/\text{yr}$
Chemicals	<u>$\\$750 \times 10^3/\text{yr}$</u>
TOTAL OPERATING COST	$\$835 \times 10^3/\text{yr}$

If capital is amortized @ 15%/yr, total cost is $\$1.66/10^3$ gal treated.

(a) Values taken from or estimated from Reference 25 data.

Figure 6-5. The TOSCO II retorting system showing water streams for the production of 47,000 BPSD of upgraded shale oil from 35 gal/ton oil shale. Water streams determined in Table 6-19.(Figure adapted from Ref.23).

Figure 6-6. Water streams for the modified TOSCO II process in which venturi scrubbers have been replaced by bag filters. Oil production as for Figure 6-5; water streams determined in Table 6-20.

Figure 6-7. Major water streams for TOSCO II process producing 47,000 BPSD of upgraded shale oil, and 4,300 BPSD LPG from 66,000 ton/day 35 gal/ton shale.

Figure 6-8. Major water streams for modified TOSCO II process in which venturi scrubbers have been replaced by bag filters. Other conditions as for Figure 6-7.

Paraho gas condensate. Following the developer's water treatment scheme [24, 25], the TOSCO retort condensate is stripped in a foul water stripper and ammonia recovery is not attempted. The stripper, an 85 ft tall 5 ft diameter unit fitted with a 2,700 ft² reboiler, is shown in Figure 6-9 along with the compositions of the influent and effluent streams. Costs and steam requirements are summarized in Table 6-26.

About 255 gpm wash water/condensate from the gas-oil and the naphtha hydrogenation units is treated in the ammonia recovery section. Based on a total ammonia production of 134 ton/day, the NH₃ content of the wash water is about 41,000 mg/l. The H₂S content, calculated from the difference between the total sulfur production of 192 ton/day and the sulfur recovered from the foul water stripper, is 30,000 mg/l. The high H₂S/NH₃ equivalence ratio in the hydrogenation wash water is not obtained in the retort/gas condensates due to the high concentrations of CO₂ in the gas phase and the association of dissolved NH₄⁺ ions in the liquid phase with HCO₃⁻ ions in preference to HS⁻ ions [11]. The Chevron [15] process is a candidate treatment scheme for the hydrogenation unit wash water, but we have selected the Phosam-W process in line with the systems used in this study for the other oil shale plants. A possible advantage of the Phosam-W is that the stripped gases from the foul water stripper may be routed to the Phosam adsorber for recovery of the ammonia in the retort condensate. This may prove desirable if ammonia concentrations in this condensate turn out to be higher than measured so far [28], or if destruction of the ammonia in the stripped gases is a problem. The Phosam-W process is shown in Figure 6-10, with costs and equipment summarized in Tables 6-27 and 6-28 respectively.

Most of the treated water from the Phosam-W process is recycled to the scrubbers at the hydrogenation units with only a small 24 gpm stream, equivalent to the water added to the wash streams by condensation, being blown down. A build-up of organics in the wash water circuit would require a large bleed to the organic removal section. Organics removal is by biological oxidation following the scheme used at the Paraho plant. Costs have been scaled directly from those determined for the Paraho process condensate, and are shown in Table 3-29.

Miscellaneous treatment costs are presented in Table 6-30 and overall costs are summarized in Table 6-31.

Table 6-26
FOUL WATER STRIPPING, TOSCO II PLANT

The stripper and flow rates are shown in Figure 5^(a).

Stripping costs

Capital Investment	$\$776 \times 10^3$
<u>Operating costs</u>	<u>10^3 \$/yr</u>
Maintenance and labor ^(b)	89
Steam @ $\$3/10^6$ Btu	<u>551</u>
TOTAL OPERATING COST	<u>640</u>

With capital amortized @15%/yr, total cost per 1,000 gal water treated = $\$3.45/10^3$ gal.

- (a) Composition of foul water based on data in Reference 28, p. A1-6 to A1-8.
- (b) Maintenance @ 4% of capital investment, labor at \$20/hr, 1/3 man per shift.

Table 6-27
AMMONIA RECOVERY FROM HYDROTREATING CONDENSATES, TOSCO II PLANT

Ammonia recovery is by the PHOSAM-W scheme shown in Figure 6-10. (a)
Wash water rates from Reference 23: Naphtha hydrogenator 255 gpm
Oil hydrogenation 255 gpm
Total feed to NH₃ recovery 510 gpm

Cost of NH₃ recovery (b)

Capital Investment, as per Table 6-28	\$6.04 × 10 ⁶
<u>Operating costs</u>	<u>10³ \$/yr</u>
Maintenance and labor	377
Steam @ \$3/10 ⁶ Btu	4,075
Cooling water at 12¢/10 ³ gal circulated	519
Chemicals, electricity	<u>145</u>
TOTAL OPERATING COST	5,116
Credit for NH ₃ sales, 134 t/d @ \$100/ton	<u>4,422</u>
NET OPERATING COST	694

With capital amortization @ 15%/yr, total net cost per 1,000 gal water treated: \$6.6/10³ gal (a).

- (a) NH₃ composition based on total ammonia production, Table 6-16.
H₂S composition based on difference between total S production, Table 6-16, and the H₂S in the retort gas. Reference 24 gives 4.32 volume % or 4.61 mass % H₂S in the product gas, and 70 lb gas/ton shale or 191 × 10³ lb gas/hr.
(b) See note (c), Table 6-10.

Table 6-28
MAJOR EQUIPMENT AND ESTIMATED COSTS FOR PHOSAM-W
AMMONIA RECOVERY AT THE TOSCO II PLANT

<u>Vessels and Columns</u>	<u>Size</u>	<u>Material</u>	<u>Cost 10³\$</u>
Water stripper	6'9" diameter 52 ft shell + 15' skirt	CS	65
Phosam adsorber			
Phosam stripper	--	CS/SS	407
NH ₃ fractionator	4' diameter 70' shell + 15' skirt	CS/SS	190
<u>Heat Exchangers</u>	<u>Area ft²</u>		
Lean solution cooler	9,500	CS/SS	130
Solution exchanger	2,000	SS	66
Stripper condenser	7,500	SS	170
Fractionator condenser	4,100	CS	45
Water stripper reboiler	4,200	CS	40
Absorber cooler	5,500	CS/SS	58
Phosam stripper reboiler	4,400	CS/SS	70
Fractionator reboiler	1,500	CS/SS	19
Miscellaneous equipment (flash drums, storage tanks, etc.)			84
Total equipment cost			1,344
Installation cost (including pumps, instrumentation)			3,696
Total installed cost			5,040
Royalty			1,000
Total Capital Investment			6,040

Table 6-29
ORGANICS REMOVAL FROM STRIPPED CONDENSATE AT TOSCO II PLANT

In view of the uncertainty surrounding the design of a biological unit, a specific design was not made for the TOSCO plant. Instead, the very conservative design for the Paraho plant, Table 6-12, was scaled directly:

Total cost per 1,000 gal water treated =	\$6.7/10 ³ gal
Feedwater rate	439 gpm
<u>Capital Investment</u>	\$11.2 × 10 ⁶
<u>Operating cost</u>	<u>\$494,000/yr.</u>
Estimated effluent concentration	< 50 mg/l BOD

Table 6-30
MISCELLANEOUS WATER MANAGEMENT COSTS, TOSCO II PLANT

	<u>Flows gpm</u>	<u>Capital Investment 10³\$</u>	<u>Operating Cost 10³ \$/yr</u>
Oil water separation: (a)			
Process condensate	461	48	--
Runoff	126	15	--
"Domestic" waste (b)	15	31	0.6
Equalization basin, blowdown streams (3 day retention)	1,650	<u>6,480</u>	<u>--</u>
TOTAL		6,574	0.6

- (a) Assuming conventional oil water separation is adequate.
(b) Package biological unit. Draws approximately 2.2 kW.

TABLE 6-31
SUMMARY OF WATER TREATMENT COSTS AT THE TOSCO II
PLANT PRODUCING 47,000 BPSD UPGRADED SHALE OIL

	Capital 10 ³ \$	Operating 10 ³ \$/yr
Source water clarification	384 ^(a)	59 ^(a)
Cooling water treatment	--	214
Boiler feedwater demineralization	2,050	835
Process condensate: Foul water stripping	776	640
NH ₃ recovery	6,040	694 ^(b)
Organics removal	11,200	494
Miscellaneous (Table 6-30)	<u>6,574</u>	<u>0.6</u>
TOTAL	<u>26,998</u>	<u>2,936</u>

(a) Costs for modified design slightly lower. See Table 6-17.

(b) Includes byproduct credit.

Figure 6-9. Foul water stripper for TOSCO II plant producing 47,000 bbl/day upgraded shale oil.

Figure 6-10. Ammonia recovery by Phosam-W for TOSCO II plant producing 47,000 bbl/day upgraded shale oil.

6.2.3 Modified In Situ Process

The water balance for a Modified In Situ retort burning shale with a mean grade of 25 gal/ton and with a nominal production of 57,000 bpsd is shown in Figure 6-11. The water rates shown were determined as indicated in Table 6-32, the material balance for the retort. As seen, there are two process condensate streams produced. The retort condensate is derived from water vapor condensing in the retort; this water percolates through the unretorted shale and leaches out inorganic salts, mainly as bicarbonates. A probable composition is given in Table 6-33, which suggests that the ammonia concentration is low, and H_2S probably negligible. The gas condensate on the other hand leaves the retort in the vapor phase and is condensed from the gas in a scheme similar to that shown in Figure 6-2 for the Paraho plant. It is not expected to contain significant concentrations of inorganic salts, but the dissolved gases will, if in equilibrium with the retort gas composition of Table 6-34, be about 21,000 mg NH_3/l , 42,000 mg CO_2/l and 100 mg H_2S/l . Organics will be present in both condensate streams.

A further, and probably major water stream associated with MIS retorting, is mine drainage water. Expected flow rates range from 4,000 to 10,000 gpm [30, 35]. Based on the compositions shown in Table 6-35, the mine drainage water is of fair quality and may be used directly for many plant requirements, including cooling. However, discharge permits, depending on their stringency, will require lowering of the TDS and removal of boron, fluoride, and possibly ammonia and phenol, before discharge to a surface stream. Although re-injection might require little or no treatment, it is concluded in Reference 31 (Vol. 2, p. 203) that it does not appear to be an attractive method for disposing of large volumes of water for a sustained period. The reasons given are that a relatively large number of wells would be required in locations far from the mine, miles of buried pipeline would be needed to deliver the water, maintenance would be difficult, and leakage of highly saline water might result. For these reasons re-injection was not considered.

Two levels of treatment corresponding to the less strict and more strict regulations discussed in Section 4.0 were considered for surface discharge of the excess mine water. The constituents covered by each regulation and the corresponding concentration limits are given in Tables 6-36 and 6-38.

Table 6-32
MATERIAL BALANCE AROUND MIS RETORT PRODUCING 57,874 BPSD
CRUDE SHALE OIL FROM ~ 25 GAL/TON OIL SHALE^(a)
FLOWS IN 10³ LB/HR

IN	Stream	Total	H	Moisture	H ₂ O equiv.	C
	Raw shale retorted ^(b)	13,720	200	274	2,077	1,577
	Steam ^(c)	848	-	848	848	-
	Air ^(c) , + 1% moisture	3,109	-	31	31	-
	Carbonates decomposed (75%) ^(d)	-	-	-	-	562
	Seepage ^(e)	-	-	-	-	-
	TOTAL IN	17,677	200	1,153	2,956	2,139
OUT						
	Spent shale ^(f)	10,913	47	-	423	728
	Water leaving with oil ^(c)	386	-	386	386	-
	Water vapor leaving with gas ^(g)	917	-	917	917	-
	Retort gas ^(c,g) (dry, NH ₃ and H ₂ S free)	5,414	43	-	388	745
	Product oil (1.5% moisture) ^(h)	754	88	11	805	636
	Sulfur ^(g)	12	-	-	-	-
	Total NH ₃ ⁽ⁱ⁾	23	4	-	37	-
	Coke ^(j)	30	-	-	-	30
	TOTAL OUT	17,677	182	1,313	2,956	2,139

NOTES:

- (a) Scheme based essentially on the Occidental Oil Shale scheme as presented in Figure III-J, Reference 30.
- (b) Shale retorted. Reference 30 shows 41,134 t/d shale mined. If this is 20% of total shale, shale retorted = $4 \times 41,134$ t/d or $13,720 \times 10^3$ lb/hr. For 25 gal/ton shale, C = 11.5%, H = 1.46%. Following Reference 35, free and combined water taken as 2%.
- (c) Value from Reference 30.
- (d) Carbonates decomposed. A carbon balance on Reference 35 data suggest most of carbonates are decomposed. Using inorganic analysis of Reference 35, carbon produced by 100% carbonate conversion is 5.47% of shale retorted. Assume 75% carbonate decomposition, i.e., 4.1% of shale retorted.
- (e) Seepage - assumed negligible.
- (f) Spent shale. The total mass spent shale, as well as H and C concentrations, were determined by closing the material balance, to get:
Spent shale = 79% of raw shale
Carbon content = 6.7%
Hydrogen content = 0.43%
- (g) See Table 6-34 for gas composition.
- (h) Product oil. 56,874 bbl/d, SG = 0.91, (Reference 31), C = 84.2%, H = 11.69 by mass. Assumed to contain 1.5% H₂O by mass.
- (i) 99% of total NH₃ assumed to be in retort gas. See Table 6-34.
- (j) Coke. Reference 35 suggest about 4% of theoretical oil yield, or 4.6% of actual oil yield, forms coke.

Table 6-33
COMPOSITION OF MIS RETORT DRIPPINGS^(a)

<u>Constituent</u>	<u>Concentration mg/l</u>
Ca ⁺⁺	20
Mg ⁺⁺	17
Na ⁺	3,600
K ⁺	100
CO ₃ ⁼	660
HCO ₃ ⁻	11,960
SO ₄ ⁼	1,200
Cl ⁻	280
NO ₃ ⁻	1.6
F ⁻	39
B	25
SiO ₂	40
NH ₃	720
TDS ^(b)	5,980
pH	8.5

(a) Composition supplied by DRI. It is assumed that this water can be fed directly to the thermal sludge unit for steam raising.

(b) Solids remaining in thermal sludge unit, assuming all HCO₃ is stripped. 5,980 mg/l $\approx$ 50 lb sludge/10³ gal.

Table 6-34
COMPOSITION OF MIS RETORT GAS^(a)

Constituents	MWt	Raw Retort Gas ^(b)		Cooled & Scrubbed Gas	
		Mass %	Vol. %	Mass %	Vol. %
N ₂	28	43.930	44.699	49.656	54.265
CO ₂	44	38.153	24.704	42.306	29.411 ^(d)
CO	28	0.759	0.772	0.858	0.937
H ₂	2	0.298	4.246	0.337	5.156
CH ₄	16	0.650	1.158	0.735	1.405
C ₂	29	0.315	0.309	0.356	0.376
C ₃	43	0.232	0.154	0.262	0.186
C ₄ ⁺	60	0.618	0.309	0.678	0.356
O ₂	32	0.086	0.077	0.097	0.093
H ₂ S	34	0.184	0.154	0.206	0.185
NH ₃	17	0.369	0.618	--	--
H ₂ O	18	14.405	22.800	4.490	7.630
Total		100	100	100	100
H ^(c) (dry basis)		0.835		0.796	
C (dry basis)		14.245		13.76	
mw		28.49		30.59	
Gas rate, wet, 10 ³ lb/hr:		6,366		5,632 ^(d)	
dry, 10 ³ lb/hr:		5,449		5,379	

(a) Values presented here are as used in calculations and are not intended to represent the accuracy of the composition.

(b) Based on composition on dry basis provided by DRI.

(c) Not including hydrogen in the water vapor and H₂S.

(d) About 46 × 10³ lb/hr CO₂ are removed in gas treatment.

Table 6-35
WATER QUALITY DATA FOR OIL SHALE MINE AQUIFERS IN PICEANCE CREEK BASIN

Constituent	Concentration (mg/l unless noted)			
	Upper Mine Aquifer (a)	(b)	Lower Mine Aquifer (a)	Values Adopted for this Study
Bicarbonate	350-2,100	482	265-4,300	750
Boron	<0.16-11	0.3	.05-12	2-4
Calcium	5.4-52	35	4-28	50
Carbonate	0-53	0.9	0-360	50
Chloride	4.1-151	12	1-700	20
Fluoride	3.1-19	0.4	6.5-45	15
Magnesium	4.2-54	52	1.9-29	60
Silica	10-19	26	2-19	15
Sodium	200-780	212	143-2,320	300
Sulfate	<5-370	325	<4-350	350
TDS	750-1,800	905	356-5,747	1,350
pH (units)	8.3-8.9	6.8	8.3-9.3	~ 8.5
Phenol ^(c) , µg/l	<1-174	2.5		2.5
Ammonia (c)	<1-174	0.3		
(d)		1.2	17	1.2

- (a) Data from range published in Reference 36, unless otherwise noted.
 (b) Data from mean values published in Reference 37, unless otherwise noted.
 (c) Rio Blanco data, as provided by DRI.
 (d) Occidental data, as provided by DRI.

Table 6-36
EXCESS MINE WATER TREATMENT, LESS STRICT REGULATIONS^(a)

Constituent	Concentration in mg/l			% Removal
	Mine water ^(b)	R.O. Discharge	Proposed Regulation ^(e)	
Ammonia	1.2	0.18	-	85 ^(d)
Bicarbonate	750	175	-	77 ^(d)
Boron	2-4	0.6-1.2	1.0	70 ^(e)
Calcium	50	0.5	-	99 ^(f)
Carbonate	50	5.0	-	90 ^(f)
Chloride	20	1.2	-	94 ^(f)
Fluoride	15	1.5	2.0	90 ^(f)
Magnesium	60	0.6	-	99 ^(f)
Phenol	0.0025	0.001	-	55 ^(c)
Silica	15	2.6	-	83 ^(d)
Sodium	300	18	-	94 ^(f)
Sulfate	350	10.5	-	97 ^(d)
TDS	1,350	<200	723	
pH	~ 0.85		6-9	
Relative flow	1	0.85		

Costs^(g)

Capital (incl. dam): \$5.342 to 23.738 million
Operating: \$591,000 to 3,351,000/yr.

(a) Treatment according to Figure 6-12.

(b) As given in Table 6-35.

(c) As given in Section 4.

(d) Based on data in Reference 38 for B-9 Permeators.

(e) Based on data in Reference 39 for B-9 Permeators.

(f) Based on data in Reference 41.

(g) Based on R.O. capital cost of (\$1.8/gal/day) and operating cost of \$0.75/10³ gal and flow of 1,500 gpm to 8,500 gpm through the R.O. system. Actual flow to R.O. will depend on the mine water production rate and its contaminant concentration. The retention dam, sized for 10,000 gpm flow, 1 week retention, is a 28 acre, 10 ft deep pond. \$50,000/acre used for costing. Cost of concentrate treatment handled separately; see Figure 6-15.

Table 6-37
EXCESS MINE WATER TREATMENT, MORE STRICT REGULATIONS^(a)

Constituent	Mine water ^(b)	Concentration in mg/l			Proposed ^(e) Regulation	Overall Removal
		Stage 1 Feed	Stage 2 Feed	Stage 2 Discharge		
NH ₃	1.2	1.15	0.17	0.10	0.2	91.7
HCO ₃ ⁻	750	805	185	43	-	94.3
B	2-4	2.4-4.8	0.7-1.4	.04-.08	0.75	98.2
Ca ⁺⁺	50	456	0.5	0.01	-	99.9
CO ₃ ⁼	50	53	10.5	2.1	-	95.8
Cl ⁻	20	19.0	1.1	0.07	-	99.6
F ⁻	15	14.8	1.5	0.15	2.0	99.0
Mg ⁺⁺	60	54.7	0.5	0.01	-	99.9
C ₆ H ₅ OH	0.0025	0.0035	0.0016	0.0002	0.001	93.7
SiO ₂	15	15.4	2.6	0.5	-	96.7
Na ⁺	300	285	17	1	-	99.7
SO ₄ ⁼	350	324	9.7	0.3	-	99.9
TDS	1,350			<100	500	
pH	~ 8.5		10		6-9	
Relative Flow	1	1,106	0.956	0.860		

Costs^(e) Capital (incl. dam): \$7.967 to 38.615 million
Operating: \$986,000 to 5,585,000/yr.

(a) Treatment according to Figure 6-13(A).

(b) As given in Table 6-35.

(c) As given in Section 4.

(d) Stage efficiencies as in Figure 6-13(A), and Table 6-36.

(e) Based on capital costs of \$2.75/gal/day and operating costs of \$1.25/10³ gal for two stage R.O. scheme and excess mine water flow rate from 1,500 to 8,500 gpm. Cost of concentrate handling costed separately; see Figure 6-15.

Table 6-38
ALTERNATIVE SCHEME FOR EXCESS MINE WATER TREATMENT,
MORE STRICT REGULATIONS^(a)

Constituent	Mine water ^(b)	Concentration in mg/l		Proposed ^(c) Regulation	Overall Removal %
		R.O. Feed	R.O. Discharge		
NH ₃	1.2	1.0	0.5(d)	0.2	58
HCO ₃ ⁻	750	265	61	-	92
B ⁺⁺	2-4	2	0.2-0.8	0.75	80
Ca ⁺⁺	50	0.5	-	-	~100
CO ₃ ⁼	50	50	5	-	90
Cl ⁻	20	20	1.2	-	94
F ⁺⁺	15	15	1.5	2.0	90
Mg ⁺⁺	60	0.6	-	-	~100
C ₆ H ₅ OH	0.0025	0.0025	0.001	0.001	55
SiO ₂	15	15	2.6	-	83
Na ⁼	300	270	16	-	94
SO ₄ ⁼	350	350	11	-	97
TDS	1,350		<100	500	
pH	~ 8.5	~ 9(e)		6-9	

Cost^(f): Capital (incl. dam): \$5.5 to 24.6 million.
Operating: \$725,000 to 4,000,000/yr.

- (a) Treatment according to Figure 6-13(B). Ion exchange train may contain specific ion removal resins if necessary.
- (b) As given in Table 6-35.
- (c) As given in Section 4.
- (d) Further reduction of ammonia effected in aerated pond and cascade.
- (e) pH adjustment to level required for obtaining desired boron and phenol removal.
- (f) Weak acid exchange capital at 7¢ to 5¢ per gallon per day and operating cost at 17¢ to 13¢ per 1,000 gallons in flow range 1,500 to 8,500 gpm. R.O. costed at as in Table 6-37. Addition of specific ion exchange resins will increase costs. Cost of concentrate treatment not included, see Figure 6-15.

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Figure 6-11. Water streams around MIS retort burning 164,640 t/d rubblized shale and producing 56,874 bbl/day crude shale oil.

Two processes were considered for obtaining a high purity product: electrodialysis and reverse osmosis [40, 41]. However, electrodialysis will only separate those molecules which are in ionic form in solution. Boron, for example, requires that the solution pH be about 8.5 to 9 to become ionic. In general electrodialysis also does not have a capability for separating soluble organic molecules. Reverse osmosis, on the other hand, has a moderate capability for separating soluble organic molecules, but a very poor capability for separating boron from acidic waters. It is only at a quite high pH ($\sim 9.5-10$) that a 75% rejection of boron is attained. Fluoride rejection is typically about 90%. Both of the processes require a moderate to good level of prefiltration to remove suspended solids which will be contained in the mine water. The mine water is alkaline and electrodialysis would require pretreatment with sulfuric acid to prevent scaling on the membranes. Acid addition or chelating agents would also be required for the reverse osmosis system to prevent precipitation of salts. We emphasize that both of the systems would provide a product with a lower total dissolved solids than required for discharge. As a rule, reverse osmosis would give the lowest TDS product with a typical value for the mine water considered of from 100-200 mg/l, while the electrodialysis product might range from 200-400 mg/l. These values are illustrative only and are controlled by the system economics. If total dissolved solids content were the only discharge requirement to be met, then a significant reduction in treatment cost could be achieved by treating only part of the water and blending it with untreated water from which only the suspended solids have been removed. However, for the contaminant levels shown in Table 6-35, it is not likely that the discharge limits specified in Table 6-37 could be met with blending. Of course, specific pollutant removal might still be practiced which would enable blending. This decision is also an economic one.

For this study reverse osmosis (RO) was selected as the basic treatment process for the excess mine drainage water. The single stage scheme shown in Figure 6-12 is considered adequate for the less strict regulations and should result in the effluent composition listed in Table 6-36. The dissolved solids content is reduced well below the level imposed by the regulation, but the overall salt removal effectiveness is governed by the boron and fluoride specifications. For the more strict regulations a second RO

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Figure 6-12. Excess minewater treatment scheme for less restrictive regulations. Rejection efficiencies are exemplary and will depend on membrane and water characteristics.

stage operated at pH $\approx$ 10 is added primarily to meet the boron and phenol limits. The treated water (Table 6-37) is now of a quality far in excess of what would normally be regarded as a discharge stream and in practice alternative schemes related to the actual mine water composition would present themselves. For example, a significant fraction of the total flow could bypass one or both the RO stages. This would be particularly true if some phenol and ammonia were stripped (biological removal would be insignificant in view of the low nutrient level) in the pond and cascade. The rate of stripping would depend on the aeration efficiency, pH, and ambient temperature. If in fact substantial stripping occurred, the RO units could be replaced altogether or in part by an ion exchange system which might include specific ion removal resins. A combination weak acid resin, high pH RO scheme is presented in Figure 6-13 (B) and Table 6-38 by way of example.

Obviously several treatment schemes will adequately handle the excess mine drainage water. Final selection will depend on the influent concentration, flow rate, discharge permit and economics. The single stage RO treatment cost may be considered a lower bound, with the two stage scheme presenting an upper limit. Note that disposal of the concentrate stream, which will amount to about 15% of the total flow, has not been included in the cost estimates. As shown in the water management scheme of Figure 6-14, the RO concentrate can be conveniently combined with the retort condensate and used for steam raising in a "thermal sludge" type unit. This is discussed below.

Derivation of the plant water streams shown in Figure 6-14 are referenced in Table 6-39, the plant water balance. All the source (mine drainage) water is clarified and costs are shown in Table 6-40. The plant cooling load is detailed in Table 6-41 and the steam requirements in Table 6-42. It has been assumed that steam will be used as a source of heat for the thermal sludge system. The system normally shown with the commercially available thermal sludge units incorporates a heat transfer medium heated in a furnace. A direct fired reboiler may not be used as scaling will occur at hot spots on the boiler tubes. The choice between a heat transfer medium and steam is an economic one, and will depend on the amount of steam being raised at the plant.

Table 6-39
WATER BALANCE WITH MINE WATER INPUT FOR MODIFIED IN SITU RETORTING TO
PRODUCE 57,000 BBLS/DAY OF SHALE OIL AND 97 MW OF ELECTRIC POWER^(a)

<u>IN</u> (Makeup to plant, or produced in retorting)	<u>GPM</u>
MIS Retort Water	
Water condensed from gas	1,340 ^(b)
Water separated from oil	772 ^(c)
Mine Water	7,656 ^(c)
Plant and Mine Area Runoff	125 ^(d)
Disposal Area Runoff	-0- ^(e)
	<u>9,893</u>
 <u>OUT</u> (Consumed or discharged)	
Steam into Retort	1,696 ^(f)
Cooling Tower Evaporation and Drift	1,636 ^(g)
Shale Disposal, Dust Control, & Revegetation	685 ^(h)
Potable/Sanitary & Service	31 ⁽ⁱ⁾
Mine Uses (including Dust Control)	210 ^(j)
Losses, Steam System and Condensate Treatment	83 ^(k)
Excess Mine Drainage Water to Disposal	5,554 ^(l)
	<u>9,893</u>

- (a) Generated by gas turbine system.
(b) Gas cooled to 100°F prior to Stretford process; no compression.
(c) Quantity required for balance in scheme shown in Figure 6-14.
(d) Quantity estimated.
(e) Assumed negligible.
(f) Assumed to be the same as in Figure III-J, Reference 30.
(g) See Table 6-41.
(h) Estimated at 100 lb water/10³ lb spent shale, which is equal to 10% by weight of 40,000 ton/day of disposed shale (cf. Table 7-11, Reference 10).
(i) Consumption of 31 gpm for 1,600 people (see p. 207, Reference 10).
(j) Mine use at 32 lb water/10³ lb mined out shale (see p. 195, Reference 10).
(k) See Figure 6-14.
(l) Based on minimum and maximum values of mine drainage water given in References 30-34, mine water for disposal may vary from about 1,500 gpm to 8,500 gpm.

Table 6-40
COST OF MINE WATER CLARIFICATION FOR THE MIS PLANT^(a)

Clarification of up to 11,000 gpm for highest quantity of mine water requires three 80-ft. diameter clarifiers. Lower bound of mine water clarification based on two 2,000 gpm and two 59-ft. diameter clarifiers. Both cases require addition of 30 ppm alum and two hours retention.

<u>Installed Clarifier Costs</u>	<u>High Mine Water</u> \$696,000	<u>Low Mine Water</u> \$320,000
<u>Operating cost</u>	<u>\$/yr</u>	<u>\$/yr</u>
Maintenance @ 4% of capital	27,800	12,800
Chemicals	96,400	39,400
TOTAL OPERATING COST	124,200	52,200

If capital is amortized @ 15%/yr, total operating cost = 4.4¢/10³ gal for high mine water treated, and = 11¢/10³ gal for low mine water treated.

- (a) All the mine drainage water is taken to the clarifiers. The clarified water is then distributed according to the various plant requirements as shown in Figure 6-14. Excess mine water not required for plant needs is treated for river discharge as shown in Figures 6-12 and 6-13.

Table 6-41
COOLING TOWER MAKEUP TREATMENT FOR THE MIS PLANT

<u>Plant Cooling Load</u>	<u>Circulating Water ($\Delta t = 30^\circ\text{F}$) gpm</u>	<u>Water Evaporated (1,400 BTU/lb evap.) gpm</u>
Retort gas cooling prior to NH_3 scrub(a)		
NH_3 scrubber wash water cooler(a)	47,660	1,020
Stripped condensate cooler(b)	3,672	78
NH_3 recovery, Phosam-W	5,500	118
	<u>19,050</u>	<u>410</u>
TOTAL	<u>75,882</u>	<u>1,626</u>
Cooling tower blowdown(c)	470 gpm	
Makeup, allowing for ~ 1% drift	2,104 gpm	
Cycles of concentration	4.5	
Acid addition to prevent CaCO_3 scaling:	= 95% of alkalinity(d)	
	= 2,178 ton H_2SO_4 /yr	
Cost of acid @ \$50/ton = \$108,000/yr or \$11.8¢/10 ³ gal makeup.		
Estimated cost of cooling water treatment	= 30¢/10 ³ gal	
	= \$300 × 10 ³ /yr	

- (a) See Figure 6-2 for cooling scheme.
 (b) Cooling stripped condensate to 120°F prior to equalization and biological oxidation.
 (c) Blowdown sufficient for low quality plant uses; see Figure 6-14.
 (d) Assuming total makeup stream has Ca and alkalinity of mine water, Table 6-35.

Table 6-42
BOILER FEEDWATER TREATMENT FOR THE MIS PLANT

Plant Steam Load	10^3 lb/hr	gpm
NH ₃ recovery, Phosam-W ^(a) 600 psi	294	588
60 psi	54	108
Stretford gas treatment ^(b)	35	70
Thermal sludge unit ^(c)	933	1,866
Miscellaneous uses ^(d)	50	100
	1,366	2,732

Boiler feedwater makeup, for 1% blowdown, $\frac{1}{2}\%$ losses = 41 gpm

Costs of boiler feedwater demineralization

Capital Investment \$360,000

Operating Costs \$ 64,400/yr

If capital is amortized @ 15%/yr, total cost is \$5.50/10³ gal treated.

- (a) See Figure 6-16.
 (b) Scaled from value given for Paraho plant.
 (c) Assuming thermal sludge requires 10% more steam than it produces.
 (d) Estimated.

% REJECTION

	STAGE 1	STAGE 2
NH ₃	85	40
S	70	95
F	90	90
Phenols	55	90

(A)

% REMOVAL

NH ₃	15
Ca	99
Mg	99
HCO ₃	65

% REJECTION

NH ₃	50
S	80
F	90
Phenol	60

(B)

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Figure 6-13. Two possible excess minewater treatment schemes for more restrictive regulations. Treatment efficiencies shown are exemplary. Final scheme selected will depend on flow and concentrations of feed minewater.

Figure 6-14. Major water streams for modified in situ shale oil plant producing 57,000 bbl/day crude shale oil and 97 MW electricity. No surface retorting. Alternative schemes for handling more or less mine water than the 7656 gpm shown here are indicated in Figure 6-15.

The management of the condensate and RO concentrate streams is shown in more detail on Figure 6-15. All the RO concentrate together with as much retort condensate and, if necessary, gas condensate, to provide the retort steam requirements, go to the thermal sludge unit. The retort condensate is not expected to have a high NH_3 concentration (Table 6-33) and preliminary stripping to prevent an NH_3 build-up, is probably not required. If the quantity of excess mine drainage water is at the lower end of the predicted range (Table 6-39, note 1), significant quantities of gas condensate may be required as makeup to the thermal sludge. Prestripping may then be required, but will not significantly alter costs as the gas condensate is in any case stripped in the ammonia recovery section.

Vendor costs for the thermal sludge unit were not obtained. Instead costs were estimated as shown in Table 6-43. Capital charges are insignificant relative to energy requirements, here based on an assumed 10% excess steam. Costs can obviously be considerably reduced by using units with good heat utilization.

As at the Paraho and TOSCO II plants, ammonia is recovered by the Phosam-W process as shown in Figure 6-16, and costed in Table 6-44. The biological oxidation system for organics removal was again scaled from the Paraho system, and the resulting costs shown in Table 6-45. Additional water treatment costs at the MIS plant are given in Table 6-46, while overall costs are summarized in Table 6-47.

6.2.4 Modified In Situ with Lurgi-Ruhrgas Retorting of Mined Shale

The water streams associated with the surface retorting are shown on Figure 6-17. Condensate streams are small relative to those from the in situ retort. The 90 gpm shown leaving with the oil stream is, following separation, taken to the thermal sludge unit, while the portion of the 5 gpm vapor leaving with the gas that is condensed is added to the 1,320 gpm of MIS gas condensate. Neither stream affects the condensate treatment costs presented in the previous section.

The major change, as far as water management is concerned, is that in the surface retorting scheme the burned spent shale is slurried and returned to the mine as backfill. Slurrying requires 3,220 gpm, while a further 2,180 gpm are required for moistening the retorted shale in the mine. Of the

Table 6-43
THERMAL SLUDGE UNIT, MIS PLANT

This unit is used to raise low quality steam for the MIS retort. Makeup water is the concentrate effluent from the excess mine water treatment system (Figures 6-12 and 6-13) and from the stripped retort water condensate. Long term variations in the quantity of mine water concentrate may be balanced against the quantity of water sent to ammonia recovery as shown in Figure 6-15.

Thermal sludge unit costs^(a)

Capital cost	\$1,000,000
Installed cost	\$3,000,000
<u>Operating Costs</u>	<u>10³ \$/yr</u>
Maintenance @ 4% of capital	120
Labor $\frac{1}{2}$ man/shift, \$20/hr	88
Chemicals	40
<u>Steam.</u> Assuming 10% excess steam is required, steam rate = $1.1 \times 1,696 = 1,866$ gpm. As this unit is considered to be an environmental control system, steam charge is only for the additional 10%.	<u>2,020</u>
TOTAL OPERATING COST	2,268

If amortized @ 15%/yr, total cost per 1,000 gal water treated = $\$3.34/10^3$ gal.

(a) Costs approximate. Estimated by WPA on basis of 10 stage stripping column and reboiler.

Table 6-44
AMMONIA RECOVERY AT THE MIS PLANT

The treatment scheme is shown in Figures 6-16.

Volume of condensate collected on cooling gas ^(a)	1,320
Volume of wash water from NH ₃ scrub ^(a)	740
Total feed rate to water stripper, NH ₃ recovery plant ^(b)	<u>2,060</u>
<u>Capital Investment</u>	$\$8.6 \times 10^6$
<u>Operating Costs</u>	<u>10³ \$/yr</u>
Maintenance and labor	478
Steam @ \$3/10 ⁶ Btu	8,650
Cooling water @ 12¢/10 ³ gal circulated	1,087
Chemicals and electricity	<u>230</u>
TOTAL OPERATING COST	10,445
Credit, NH ₃ sales: 281 t/d @ \$100/t	<u>9,273</u>
NET OPERATING COST	<u>1,172</u>
With capital amorized @ 15%/yr, total net cost per 1,000 gal water treated: \$2.27/10 ³ gal ^(c)	

- (a) Concentrations in water calculated using van Krevelen's [11] equilibrium data.
 (b) See Figure 6-16 for feed composition.
 (c) See note (c), Table 61-10.

Table 6-45
ORGANICS REMOVAL FROM STRIPPED CONDENSATE AT MIS PLANT

As discussed in Table 6-12 reliable data on biological oxidation of oil shale process condensates is not available. In view of the uncertainty surrounding the design of a biological unit, a specific design was not made for the MIS plant. Instead, the very conservative design for the Paraho plant was scaled directly:

Cost per 1,000 gal water treated	=	\$6.7/10 ³ gal
Feed water rate		1,224 gpm
<u>Capital Investment</u>		\$31.2 × 10 ⁶
<u>Operating Cost</u>		\$1.4 × 10 ⁶ /yr
Estimated effluent concentration		< 50 mg/l BOD

Table 6-46
MISCELLANEOUS WATER MANAGEMENT COSTS, MIS PLANT

<u>Treatment</u>	<u>Flow</u> <u>gpm</u>	<u>Capital</u> <u>Investment</u> <u>10³ \$</u>	<u>Operating Cost</u> <u>10³ \$/yr</u>
Oil water separation ^(a)	2,060	180	-
Runoff	141	17	-
"Domestic" waste ^(b)	19	26	0.9
Equalization basin, plant	882	<u>3,718</u>	<u>-</u>
blowdown streams			
(3 day retention)		3,941	0.9

(a) It is assumed that oil can be separated in conventional API separators and that emulsification is not a problem.

(b) Package biological unit. Draws approximately 3 kW.

- (a) Not including excess mine water treatment costs.
- (b) Range for 1,500 gpm, less strict regulations, to 8,500 gpm, more strict regulations. No credit taken for water by-product.
- (c) Probably not required.
- (d) Including byproduct credit of \$9,273,000
- (e) Assuming high level of mine water.

Figure 6-15. Schematic showing generation of steam for retorting from minewater concentrate and retort condensate. In the possible event that RO treatment of the excess minewater is not required, or insufficient minewater concentrate is available for the steam requirements, additional makeup is available from the gas cooling condensate. Alternatively, retort water can be taken to ammonia recovery if minewater flows are excessive.

Figure 6-16. Ammonia separation by Phosam-W for the MIS plant producing 57,000 bbl/day crude shale oil.

Figure 6-17. Schematic showing water streams for the Lurgi-Ruhrgas retort producing 24,400 BPSD shale oil from 41,000 t/d shale mined from MIS retorts.

slurry water, about 406 gpm are evaporated on cooling the ash, with the remainder forming a slurry of 35% H₂O by mass. The water required for moistening the retorted in situ shale was calculated on the basis of 3.75 gallons absorbed per cubic foot of retorted shale.

The total water required for backfill is 5,400 gpm and will come from the mine drainage water. According to the water balance shown in Table 6-48, 6,884 gpm of mine drainage water are required by the MIS/Lurgi-Ruhrgas retorting scheme. Provided the quantity of mine water is near the upper end of the 4,000-10,000 gpm range, water supply will not be a problem. However, if the mine water rate drops much below 7,000 gpm, additional makeup water will have to be obtained, or an alternative spent shale disposal scheme will have to be adopted.

As in the previous section, it is assumed that 7,650 gpm of mine water are available, leaving 874 gpm excess water for treatment and discharge. The overall water management scheme for this quantity of mine water is shown in Figure 6-18. Mine water flows different from the 7,656 gpm (but greater than 6,884 gpm) can be accommodated by adjusting the amount of retort and gas condensate, and untreated mine water, taken to the thermal sludge system.

Due to the large quantity of mine water used for slurring, the costs for mine water clarification and excess mine water treatment are lower than in the previous section. Costs for the scheme considered here are summarized in Tables 6-49 and 6-50. The surface retort imposes a small additional cooling load on the plant as shown on Table 6-51, and some of the miscellaneous treatment costs, shown in Table 6-52, are increased due to the larger number of personnel and plant surface area. The remaining water treatment costs remain the same as in the previous section, but have been included in the cost summary, Table 6-53, for convenience. Water treatment costs on a dollar/year basis are not significantly changed by the introduction of surface retorting; however, the amount of oil produced is increased resulting in a decrease in the cost per barrel of oil produced. In addition, the amount of excess mine water is considerably reduced, resulting in a corresponding reduction in costs for its disposal.

Table 6-48
OVERALL WATER BALANCE FOR MIS + LURGI-RUHRGAS RETORTING SCHEME
PRODUCING 81,000 BPSD SHALE OIL AND 140 MW ELECTRICITY

IN (Makeup to plant or produced in retorting)	<u>10³ lb/hr</u>	<u>gpm</u>
Source (Mine water) ^(a)	3,442	6,884
Moisture in with shale, Lurgi-Ruhr gas retort ^(b)	68	136
MIS retort water; gas and oil condensate ^(c)	1,056	2,112
Runoff ^(d)	80	160
	<u>4,646</u>	<u>9,292</u>
OUT (Consumed or discharged)		
Steam to MIS retort ^(e)	848	1,696
Slurry water, Lurgi-Ruhr gas ^(b)	1,610	3,220
Water to saturate residue shale in retorted cavern	1,090	2,180
Dust control	148	295
Cooling water, evaporation and drift ^(f)	867	1,735
Potable, service; water consumed	19	37
Losses, water treatment and steam cycle ^(g)	41	83
Losses, Lurgi-Ruhr gas retort and product gas ^(b)	23	46
	<u>4,646</u>	<u>9,292</u>

(a) Water required for plant needs; not including excess mine drainage water.

(b) As shown in Figure 6-17.

(c) Taken from Table 6-39, MIS section.

(d) Estimated surface area ~ 200 acres.

(e) See Table 6-51.

(f) 400 plant personnel in addition to the 1,600 mine/plant personnel for the MIS case.

(g) As shown in Figure 6-18.

Table 6-49
COST OF MINE WATER CLARIFICATION FOR THE MIS/LURGI PLANT^(a)

Clarification of up to 6,000 gpm for highest quantity of mine water requires two 71.5-ft. diameter clarifiers. Lower bound of mine water clarification based on one 2,000 gpm and 59-ft. diameter clarifier. Both cases require addition of 30 ppm alum and two hours retention.

	<u>High Mine Water</u>	<u>Low Mine Water</u>
<u>Installed Clarifier Costs</u>	\$393,000	\$160,000
<u>Operating Cost</u>	<u>\$/yr</u>	<u>\$/yr</u>
Maintenance @ 4% of Capital	15,800	6,400
Chemicals	49,600	19,700
TOTAL OPERATING COSTS	65,400	26,100

If capital is amortized @ 15%/yr, total operating cost = 4.40/10³ gal for high mine water treated, and \$5.3/10³ gal for low mine water treated.

(a) Mine water used for spent shale reinjection is not clarified.

Table 6-50
EXCESS MINE WATER TREATMENT, MIS + LURGI-RUHRGAS PLANT

Based on a maximum mine drainage flow rate of 11,250 gpm, and plant requirements of 6,880 gpm, a maximum of 4,370 gpm excess water will require treatment prior to stream discharge. The costs given below are for this projected maximum flow.

1. Treatment according to less strict regulations^(a):
Capital: \$12.884 million; operating cost: \$1,723,000/yr.
2. Treatment according to more strict regulations^(b):
Capital: \$20.533 million; operating cost: \$2,871,000/yr.
3. Alternative treatment, more strict regulations^(c):
Capital \$15.7 million; operating cost: \$2,112,000/yr.

- (a) See Figure 6-12 and Table 6-36.
(b) See Figure 6-13(A) and Table 6-37.
(c) See Figure 6-13(B) and Table 6-38.

Table 6-51
COOLING TOWER MAKEUP TREATMENT FOR THE MIS + LURGI-RUHRGAS PLANT

Plant cooling load	Circulating Water ($\Delta t = 30^{\circ}\text{F}$) gpm	Water Evaporated (1400 BTU/lb) gpm
MIS Section ^(a)	75,882	1,626
Lurgi-Ruhr gas ^(b)	<u>4,670</u>	<u>100</u>
	80,552	1,726
Cooling tower blowdown	578 gpm	
Makeup, allowing for $\frac{1}{2}\%$ drift	2,313 gpm	
Cycles of concentration	4	
Acid addition to prevent CaCO ₃ scaling	95% of alkalinity ^(c) = 2,297 ton H ₂ SO ₄ /yr	
Cost of acid at \$50/ton = \$113,900/yr or 1.8¢/10 ³ gal makeup		
Estimated cost of cooling water treatment = 30¢/10 ³ gal = \$328,000/yr		

(a) See Table 6-41.

(b) Estimated; mainly for oil condenser, retort section.

(c) Assuming total makeup stream as Ca and alkalinity concentrations of mine water, Table 6-35.

Table 6-52
MISCELLANEOUS WATER MANAGEMENT COSTS, MIS + LURGI-RUHRGAS PLANT

Treatment	Flow gpm	Capital Investement 10 ³ \$	Operating Cost 10 ³ \$/yr
Oil water separation ^(a)			
Gas treatment condensate	2,060	180	-
Runoff	184	22	-
"Domestic" waste ^(b)	25	42	0.6
Equalization basin, plant blowdown streams (3 day retention)	1,000	<u>4,215</u>	<u>-</u>
		4,459	0.6

(a) It is assumed that oil can be separated in conventional API separators and that emulsification is not a problem.

(b) Package biological unit. Draws approximately 3 kW.

Table 6-53
SUMMARY OF WATER TREATMENT COSTS, MIS + LURGI-RUHRGAS PLANT

	<u>Capital</u> <u>10³ \$</u>	<u>Operating</u> <u>10³\$/yr</u>
Mine water clarification (d)	160	26.1
Cooling water treatment		328
Boiler feedwater demineralization ^(a)	360	64.4
Retort condensate: Foul water stripping ^(a)	-	-
Thermal sludge ^(a)	3,000	2,268
Gas condensate: NH ₃ removal ^(a)	8,600	1,172 ^(b)
Organics removal ^(a)	31,200	1,400
Miscellaneous (Table 6-52)	<u>4,459</u>	<u>0.6</u>
TOTAL ^(b)	<u>47,779</u>	<u>5,259</u>

Excess mine water treatment costs (c): Capital \$0 to 20.533 million
Operating cost \$0 to 2,871,000/yr

- (a) Treatment and costs identical to those for MIS plant with no surface retorting.
- (b) Including byproduct credit of \$9,273,000.
- (c) Upper value based on maximum anticipated mine water flow and more restrictive regulations.
- (d) Assuming low level of mine water

Figure 6-18. Major water streams for modified in situ Lurgi Ruhrgas shale oil plant producing 81,000 bbl/day crude shale oil. Minewater rate of 7,656 gpm from MIS scheme with no surface retorting.

6.3 PROCESS CONDENSATE TREATMENT

Major contaminants in the process condensate are dissolved ammonia and the acid gases CO_2 and H_2S , along with organics. Inorganics are not expected to be a problem in the gas condensate, (see e.g., Reference 28,) but will be present in waters condensed on spent or unretorted shale. Some information is available on the distribution of toxic metals among the spent shale, oil, water and gas streams [44], and it appears that while most of these elements remain in the spent shale, some could be present in the retort condensate. See Section 8.0.

Of the retorting schemes considered in this study, only the in situ system produces retort condensate and the concomitant inorganic and toxic metal problems. However, low grade steam is required in the MIS retort, and can be conveniently raised from the retort condensate in a thermal sludge [42] type system. Nonvolatile salts will be recovered as a sludge and can be disposed of in a hazardous waste landfill (see Section 7). Volatiles will be stripped off and returned to the retort with the steam. Here they might either be adsorbed on the spent shale (see e.g., Reference 45), destroyed in the combustion region, or leave with the oil, water or gas streams. In view of the partitioning of the elements [44], and the low concentration of NH_3 in the retort condensate (Table 6-33), a build-up of pollutants in the condensate/steam circuit is not expected, but could, if needed, be controlled by blowdown with the sludge. Detailed information on the application of thermal sludge to MIS retort condensate has not been published; it is understood, however, that the system is being used with success [46].

Removal of dissolved gases and organics from the gas condensate are discussed below.

6.3.1 Dissolved Gas Removal

Thermal stripping is the only process considered economical for removal of NH_3 , CO_2 and H_2S from the condensate. Recovery of the stripped ammonia for sale is feasible and may offset or completely cover the treatment costs, depending on the ammonia throughput rate. The ammonia recovery process used in this study is the USS Engineers and Consultants' (UEC) Phosam-W process [16], in which the ammonia in the stripped gas mixture is absorbed into,

and then recovered as aqua ammonia from an ammonium phosphate solution. Saleable anhydrous ammonia can then be produced by fractionation of the aqua ammonia.

The degree to which the gases are to be removed from the condensate determines the size of the stripping column and the steam rate. CO_2 strips off easily and is not a problem. The NH_3 and H_2S normally come off together at rates determined by, inter alia, their relative concentrations. Sufficient ammonia should be left in the treated water to provide a nutrient for biological oxidation. In fact the ammonia remaining after stripping can be adjusted such that the concentration in the cooling tower makeup is as high as about 200 mg/l. Ammonia does not concentrate in the cooling tower, and should not exceed the odor threshold in the cooling tower plume for makeup concentrations up to 500 mg/l. However, copper should not be used in the cooling system in the presence of ammonia. The presence of low concentrations of ammonia (< 200 mg/l) in the blowdown will not cause difficulties in the low quality water uses, and may in fact be beneficial when used for revegetation.

However, H_2S should be removed down to about 2 mg/l and it turns out that to strip down to this H_2S level, the NH_3 required stripping down to less than 100 mg/l. This provided a good wash water for scrubbing the retort gases, but additional nutrient is required for those waters taken to the biological unit. Another possibility is to reduce the degree of stripping and follow it up with a specific H_2S removal step. Oxidation with peroxide, chlorine or ferric chloride may prove economical and be used in practice. This route was not costed, but is not expected to significantly alter the costs in this study.

6.3.2 Organics Removal

Resin or carbon adsorption [43], and wet air or biological oxidation are among the possible methods for controlling the dissolved organics. Carbon adsorption is normally cost effective only as a polishing step and could find application as a backup to resin adsorption or biological oxidation. Resin adsorption has been investigated as a possible means of organics control [43, 45] and may prove feasible, particularly if mixed resin beds are used to cover the range of organics present. Published analyses [19, 28] indicate that the bulk of the organics present are carboxylic acids and neutral compounds. The polymeric adsorbents should be selected accordingly.

It has been stated that the majority of the organics present are biodegradable [19], and preliminary studies with actual retort waters have to some extent supported that conclusion [20, 47]. Difficulties that have arisen may be attributed in part to inadequate pretreatment or a high inorganic loading which may include specific toxic substances. In the water management schemes presented here, only gas condensate, (containing few inorganics) required treatment for reuse, and in all cases proper pretreatment is provided. The two basic types of biological treatment are aerobic processes in which oxygen (air) addition is required, and anaerobic processes which take place in the absence of dissolved oxygen. The most familiar example of aerobic treatment is the conventional activated sludge process, and the most familiar example of anaerobic treatment is the septic tank. Biological treatment is very widely assumed to be the state-of-the-art procedure for purifying the largest and dirtiest effluent water stream from a coal gasification plant, namely the process condensate stream (also called quench water stream) which is characterized by a very high level of organic contamination measured as a high chemical and biological oxygen demand.

Anaerobic biological treatment is mostly used for the digestion and disposal of sludge from an aerobic treatment and is not used widely on industrial waste. Since oxygen is not required, the energy consumption is cut to about 1/10th of that for aerobic treatment and the sludge production is less than for aerobic treatment which makes the sludge disposal problem an easier one. A recent study [47] of the anaerobic treatment of oil shale retort water reported the following conclusions:

1. The retort water studied had to be pretreated to remove toxic constituents (ammonia and sulfide) and to add deficient nutrients (calcium, magnesium and phosphorus) before it could be successfully treated with the anaerobic fermentation process. Pretreatment included pH adjustment, air stripping and skimming, and nutrients addition.
2. A digested sludge from a conventional municipal sewage treatment plant was successfully acclimated to the retort water studied.
3. A major fraction of the organics in the retort water studied was stabilized by conversion to CH_4 and CO_2 using the anaerobic

fermentation process. BOD_5 and COD removal efficiencies were 76-80%. Within the limits of experimental error, the same removal rate was obtained for both BOD_5 and COD.

4. The effluent from anaerobic fermentation of the retort water studied (BOD_5 : 530--580 mg/l) might be suitable for treatment by conventional aerobic processes.
5. The growth of the methane formers, which stabilize the organics, is nutrient limited in the retort water studied.
6. The pretreatment of the retort water studied removed 49% of the BOD_5 . This was probably due to the reduction in solubility of high molecular weight fatty acids. At neutral pH they precipitate out of solution and do not exert a BOD.

The long hydraulic residence times required for anaerobic treatment (about 50 days in the study, with a reduction to 2-3 days possible in practice by cell recycle techniques) requires excessively large holding ponds and high capital costs. Anaerobic treatment consequently does not appear to be economical.

The most widely assumed aerobic procedure is the activated sludge process. The design of this type of biological treatment plant is fundamentally empirical. The rate of removal of BOD is measured under various conditions of biomass concentration, contaminant concentration, and oxygen concentration, within the range of interest and using the specific wastewater. From the experimental results, rate constants are calculated and a plant can then be sized to operate under the chosen, optimum conditions. All biological treatment designs must be based on experience with the wastewater in question.

There is no operating experience with wastewater from an oil shale plant. There is, however, published experience with related wastewater from coke ovens and from coal conversion condensates. We selected the biokinetic constants published for the H-Coal wastewater [22] for this study in the belief that they will provide a conservative, safe design. While preliminary published data on aerobic treatment suggests that aerobic treatment is feasible, the H-Coal biokinetic constants produced designs that are only marginally economical. If future studies show that biological oxidation rates are of the

same magnitude as those used here, alternative schemes should prove to be more economical. The costs estimated in this study are, therefore, regarded as providing an upper limit for organics removal costs.

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7.0 MANAGEMENT OF SOLID WASTES

7.1 OVERALL RESULTS AND CONCLUSIONS

Each of the four process models considered in this project generates large quantities of solid wastes which require substantial effort for proper disposal. The calculations undertaken herein are an attempt to estimate the magnitude of this required effort. Above ground disposal of solid wastes using a landfill type design is used.

While considerable discussion of this method of disposal can be found both in development plans [1-4, 21], and in other publications on oil shale pollution control [5-20] such discussions were found, without exception, to be generalizations lacking the detail necessary for the calculation of costs. Only in a few instances were actual costs found. In the first report of such figures [10], the derivation of the cost figures was not given, and the other studies appear to have referenced the first [9, 17, 18]. Hence, the cost figures derived in this study represent a new and independent derivation.

Final selection of solid waste disposal sites, and even the method of disposal itself, have not been made by the developers for any of the processes with complete certainty. Also, the potential aboveground disposal areas which may be used, vary greatly in geometry. Selection of specific sites for this project would have required extensive calculation to determine disposal parameters without improving the accuracy of the predicted costs. Hence a generalized disposal area model was used which the authors feel allows good estimates to be made of the magnitude of the problem without making the calculations unreasonably complex, and yet represents in a justifiable way all activities which would be expected for proper solid waste management.

The overall costs for management of solid wastes for each of the models are summarized in Table 7-1. As mentioned in Section 4.0, Regulatory Scenarios, two scenarios have been applied in each case. In the first

scenario, a hybrid of the plans proposed by the developers, incorporating the best features from all plans, was used to handle the spent shale and some other wastes as nonhazardous solid wastes, complying with all applicable regulations for this category of waste.²² In the more strict scenario, regulations for disposal of hazardous wastes were applied.²³ In only a few instances, which will be discussed in detail later, did the application of hazardous waste regulations deviate from strict compliance with the regulations as proposed by EPA. In these instances, these deviations will probably be allowed under the "note" system adopted by the EPA.

A prime example of such a deviation is placement of a 6-inch soil cover on the spent shale added to the disposal area each day. Such a soil cover is intended to minimize the possibility of "fire, explosion and/or harboring, feeding and breeding of land burrowing animals." Such events would not be expected to occur under the disposal conditions used, hence the soil cover was deemed unnecessary. Furthermore, such soil cover would cause instability in the disposal pile and would require an inordinate amount of soil. Assuming a 1×10^6 barrel per year industry in Colorado composed of the same processes used for this study, such a soil cover would result in removal of soil to a depth of one foot from 1,240 square miles of Colorado in 25 years. The "note" system adopted by EPA in the proposed regulations allows EPA to waive requirements which are, in a particular situation such as this, unnecessary and unreasonable.

7.2 SOLID WASTE MANAGEMENT MODELS

7.2.1 The Regulatory Scenarios

Less Strict Scenario: In order to develop solid waste management models, it is first necessary to define in some detail the regulatory parameters which must be met. For the less strict scenario it was assumed that solid waste disposal regulations [22] must be met. All actions taken to meet these regulations as summarized in Appendix 4.0 are included as costs in both scenarios since these actions are common to both. Matters concerning location are assumed to involve no cost. For any particular location these restrictions result in a go, no-go decision rather than corrective action. The disposal

TABLE 7-1. SUMMARY OF SOLID WASTE MANAGEMENT
COSTS FOR THE FOUR OIL SHALE PLANTS

Process	Less Strict Scenario		More Strict Scenario	
	Capital Cost 10 ³ \$	Operating Costs 10 ³ \$/yr.	Capital Costs 10 ³ \$	Operating Costs 10 ³ \$/yr.
Paraho	2,647 ^(b)	4,787	2,647 ^(b)	6,447
TOSCO	2,454	2,569	2,454	4,818
MIS	2,012 ^(c)	3,375	2,012 ^(c)	5,254
MIS/Lurgi Pollution Control and site closure Sludge Disposal	1,005	920	1,118	1,873
Slurry Back- filling	<u>3,000</u>	<u>7,350</u>	<u>3,000</u>	<u>7,350</u>
TOTAL	4,005	8,270	4,118	9,223

- a. The Uniform Annual Operating Cost of the direct operating cost (including the effective cost of establishing trust funds as required under the more strict scenarios) expressed as a uniform sum for each of the 25 years of operation. See Section 10.4.2 for more details.
- b. Plus an additional capital expenditure of \$2,647,000 in Year 12.
- c. Plus an additional capital expenditure of \$1,152,000 in Year 12.

plans proposed by the developers which have been synthesized into the solid waste management models in many ways exceed the minimum requirements for correct nonhazardous solid waste disposal.*

* Cannot be located in a flood plain; cannot be located in a critical habitat area; cannot be located at the recharge zone of a sole source aquifer which is the principal source of human drinking water, etc.

More Strict Scenario: Use of EPA hazardous waste proposed regulations [23] for the more strict scenario should not be construed to mean that raw or spent shale will ever be classified as hazardous wastes. Nor should it be construed that the authors or the EPA have suggested these materials would ever be classified as hazardous wastes. Rather, the Department of Energy has asked that this scenario be applied as a "worst case" example to determine the upper limit of costs for solid waste management. Since no commercial shale oil production facilities exist at present, no regulations for such an industry exist. A more likely "worst case" scenario for future regulation of raw and spent shale under RCRA would be classification as "other mining waste."

The new proposed regulations for hazardous wastes under RCRA specify certain characteristics which are considered indicative of a hazard. These characteristics are:

- Ignitability
- Corrosivity
- Reactivity
- Toxicity
- Infectiousness
- Radioactivity
- Phytotoxicity
- Teratogenicity and Mutagenicity

Current proposed rules will rely upon the first four characteristics for determining whether or not a substance is a hazardous waste because reliable tests can be run for these characteristics. Detailed testing protocols have been given for these four characteristics. Raw and spent shale would be expected to test positive only in the toxicity category, if at all.

Rather than speculating as to the results of the toxicity test, it was actually run for this analysis on a sample of Paraho spent shale, and this sample was found not to meet the criteria for a hazardous waste. The results are summarized in Table 7-2, along with maximum allowable levels of the 8 trace elements. The results shown in Table 7-2 should be viewed as tentative and require some qualification.

First, the extraction procedure was run only once. For the results to be viewed with some certainty, the test should be run several more times. Second, other types of spent shale should be run rather than generalizing the

TABLE 7-2. COMPARISON OF TOXIC WASTE EXTRACT RESULTS
AND MAXIMUM ALLOWABLE TRACE ELEMENT LEVELS

<u>Contaminants</u>	<u>Level Found (as noted)</u>	<u>Maximum Allowed (PPM)</u>
Arsenic	< 0.1 ppm	0.50
Barium	< 10 ppm	10.0
Cadmium	< 0.05 ppm	0.10
Chromium	100 ppb	0.50
Lead	220 ppb	0.50
Mercury	14 ppb	0.02
Selenium	23 ppb	0.10
Silver	30 ppb	0.50

results to all spent shales, although similar results would be expected. However, the test conditions used were somewhat more severe than required. Paraho spent shale is normally gravel and sand size. The sample used for the test had been finely powdered for use in other experiments. Also the agitation rate used during extraction was much greater than required (the reader is referred to Reference 23 for the exact protocol). Hence, higher levels of these trace elements may have been extracted than would be found if the procedure were followed exactly. Given the above uncertainties, it is probably safe to assume Paraho spent shale will not be classified as hazardous waste under the current proposed regulations, and it is unlikely that other spent shale would give significantly different results. For a much more detailed discussion of trace elements in raw and spent shale see Section 8.0.

The suggestion has been made within DOE that carbonaceous spent shale could be classified as hazardous due to the organic residue found on it. The authors believe this possibility is remote. Even if such classification occurred, such spent shale would not be handled as a hazardous waste. The proposed regulations specify the following preference:

"Where practical, disposal of hazardous waste shall be avoided and alternatives such as destruction, treatment to render the waste nonhazardous, or treatment for purposes of resource recovery and reuse shall be employed."

The underlined above option is both available and attractive.

TOSCO is now designing a spent shale burner [24] which can utilize the carbon residue as an energy source, producing a burnt shale with essentially nil organic carbon. The system involves two fluidized beds. In the first, the carbon residue is burned and the heat used to generate steam. In the second fluidized bed, the hot burned shale is cooled, generating more steam and preheating the combustion air for the burner. Conservative calculations of the heat liberated and recovered by burning the 54,168 tons per stream day of spent shale indicates recovery of 4.0626×10^{10} Btu of heat as live steam. Using the value assumed throughout this study of \$3/mBtu, the value of such heat recovery would be \$121,000 per day or approximately \$40 million per year--an amount sufficient to amortize a \$180,000,000 capital investment (see Section 10.0 for the details of such a calculation). Use of a shale burner would provide energy which, without such a burner, is supplied by consuming a portion of the fuel produced in the TOSCO refinery and by purchasing electricity from offsite. Assuming the steam produced can be used for process steam and power generation, all steam and electricity needs could be satisfied and a small excess of electric power for export produced. Because fuel from the refinery is not burned, the output of the plant would be increased by the equivalent of around 5,500 barrels per stream day of liquid fuels, and would switch from being an importer of electric power to being an exporter.

Whether or not this savings, with an opportunity cost value of approximately \$40 million per year, could actually be realized was not considered in detail. However, it certainly seems reasonable to say that the cost of installing and operating a shale burner and small power plant should at least be a breakeven proposition, and would solve the environmental problem which has been hypothesized. It should also be noted that a conventional boiler would no longer be required.

Proper control of the burning temperature and residence time in both the TOSCO burner and Lurgi burner can produce a noncarbonaceous spent shale

with good cementation properties [25-28, 43].* Such spent shale could be disposed of as a cement with very low permeability ($\leq 1 \times 10^{-7}$ cm/sec.), high strength and good durability. Such a treatment would represent complete conversion to a nonhazardous form, a treatment preferable to disposal of a hazardous waste.

Deviations from Hazardous Waste Regulations: In the instances discussed below, the procedures applied to disposal of spent and raw shale deviate from those specified for "on-site" disposal of hazardous waste. The proposed regulations provide a system of "notes" after certain standards which serve the following purpose [23]:

"Generally the notes authorize the Regional Administrator to allow deviation from a specific requirement when the applicant for a permit demonstrates that an alternate requirement or an existing natural condition at the site will achieve at least an equivalent degree of containment, destruction, or environmental protection as Subpart D design or operating requirements."

The following deviations were chosen as reasonable by the authors, and should be viewed as assumptions made for the analysis:

Daily Soil Cover: Deemed unnecessary and undesirable in that it could lower disposal pile stability and interfere with proper compaction (See above discussion).

Hazardous Waste Management Instruction Course: Deemed unnecessary since only one type of hazardous material is handled.

Fire Protection System: Since no fire hazard exists, no fire protection system is necessary.

Natural Impermeable Soil Barrier (10 feet): Since valley disposal will be used with a slope, a clay liner was deemed dangerous in that it could cause disposal pile slippage. Alternate appropriate impermeable barriers were substituted.

* The exact conditions required are the subject of Current DRI research and research elsewhere and are proprietary.

7.2.2 Disposal Site Model

For all four processes, valley disposal of solid wastes is assumed. Rather than using an irregular shape such as might be found in nature, a simplified rectangular disposal area model is used allowing easy calculation of the total area, depth, length of dams, drainage ditches, runoff ditches, etc.

The valley is assumed to be shallow with a 1:4 rise up the valley. On level ground at the head of the valley, a dam with an impervious barrier and an impervious face is constructed. The dam has a 1:4 rise on the front face, a 1:2 rise on the back face and a 50' wide road on the crest. Spent shale is filled behind the dam up the valley to a maximum height of the top of the valley. The dam serves to stabilize the pile and trap leachate. Figure 7-1 and 7-2 shows several views of the dam and fill area model.

All developers propose to construct a dam at the downhill end of their solid waste disposal areas [1-10, 12, 13, 19-21]. The TOSCO plan presents the most detail concerning such a dam [8, 21]. Some developers have proposed using considerably steeper valleys [2, 13].

Figure 7-2 shows a top view of the completely filled disposal area. As shown, the entire area is surrounded with ditches of sufficient capacity to divert runoff from a 100-year, 24-hour storm around the disposal area. The face of the spent shale, as shown, contains shallow unlined drainage ditches to catch runoff from the spent shale and carry it to main lined drainage ditches running down the face of the disposal area to join the diversion ditches at the dam face. All ditches and the leachate sump drain into an imperviously lined catchment pond. In general, leachate and storm runoff collected in the catchment pond is used, when available, for dust suppression in the disposal area.

The soil cover in the valley is removed as the disposal area is filled, until enough soil is obtained for revegetation use. It was assumed that the soil will be stored above the valley in an open disposal area. It is advantageous to leave the stored soil completely exposed to the elements, as this improves soil microorganism population, improving the viability of the soil as a growth medium for later revegetation [29, 30, 31, 41, 42].

The disposal area behind the dam is lined with a "soil cement" type impermeable liner of 6 inch or 1 foot depth. Specific construction of the

Valley Fill Disposal

$$h = \text{Dam Height} = 6.8718 \times 10$$

$$b = 4.1231d + 50$$

$$a = 4.1231d \times 2 \div 6$$

$$c = 4.1231d \times 4 \div 6$$

SIDE VIEW

FRONT VIEW UP VALLEY

TOP VIEW OF DAM

FIGURE 7-1. GENERAL DISPOSAL AREA DESCRIPTION

FIGURE 7-2.TOP VIEW OF FINISHED FILL

liner will be discussed later under each process. The liner is placed moving ahead of the spent shale, rather than lining the entire valley at the start. The lining is considered to be an annual operating cost.

The spent shale is placed in the disposal area in 2 foot layers using bottom dump trucks loaded from covered conveyers which transport the shale to the dump area. The trucks and staggered wheel compactors are used to compact the spent shale sufficiently. As the shale is placed and compacted, the surface is sprayed with a chemical wetter/binder. A product named Coherez distributed by Witco Chemical Company is recommended, mixed at 1:8 with water and applied at a rate of 680 gallons/acre [13]. The face of the spent shale pile is smooth rolled to prevent fugitive dusts [10]. Larger rock, uncrushed low grade shale, etc. from the mine and retorting is placed next to the impervious liner to facilitate leachate drainage under the spent shale to the leachate sump.

The catchment pond allows management of leachate and storm runoff to achieve zero discharge from the disposal area. In general, the water collected can be used for dust control in the area, but especially heavy runoff can be drained to the equalization basin shown in the water management schemes for the processes (Section 6.0) and incorporated into the water use and treatment systems for the overall processes.

Considerable research has been conducted and is still underway on revegetation. Revegetation--successfully establishing a self-sufficient native plant population--can be accomplished readily in a number of ways, and has been extensively reported in the recent literature [29-34]. For the purposes of this study, an optimum scheme was suggested by Dr. Wayne Cook of Colorado State University [35] one of the principal researchers in the area of revegetation of disturbed land in the Piceance Creek Basin.

Revegetation will proceed in the following steps. A top dressing of uncompacted spent or raw shale is spread on the disposal area to a depth of 3 feet. This layer serves to anchor long rooted trees and shrubs. The entire area is then covered with 2 feet of soil which has been stored since the construction of the disposal area. To this soil is added a phosphate fertilizer such as triple superphosphate at the rate of 60 lbs. P_2O_5 per acre. The area will then be seeded with 8 lbs/acre of mixed seed containing all plant varieties needed for revegetation. Such seeds are now collected by hand

at a cost of approximately \$4 per pound, but greenhouses are now being established to produce these seeds in quantity. The eventual price is expected to be lower, but the \$4 level was used for this study. At the end of the first and second year, nitrogen fertilizer such as ammonium phosphate is applied at a rate of 60 lbs. N per acre. No irrigation is required for successful revegetation, although water has been provided for this purpose in the water management schemes of this study. Three years are required for establishment of a full vegetative cover, and five years are necessary for trees and shrubs to reach maturity. Throughout the entire operation of the disposal area, access is restricted by a fence surrounding the area. During revegetation, the fence is maintained to prevent animals from damaging the new growth.

Throughout the operation of the disposal area, and during the post-closure maintenance period, monitoring wells are used to measure local groundwater quality. Three wells are placed near the fill down gradient from it, and one well is placed up gradient. A fifth is placed under the facility within the zone of aeration, between the bottom of the landfill and the top of the underlying aquifer. The wells and the leachate sump are sampled and detailed analyses performed quarterly throughout the operational and post-closure periods. The analyses are those specified in the proposed regulations [23] plus additional trace element monitoring. Extensive background data on groundwater quality and movement is already available [7, 36, 40].

7.2.3 Special Actions for Hazardous Waste Management

Many of the activities already proposed by the developers fulfill requirements of RCRA for hazardous waste management as specified in the proposed regulations. Hence, many cost items which apply to hazardous wastes are found in both scenarios. The following items, however, are unique to the more strict scenario:

- Disposal Area Guards
- Daily Visual Inspection
- Annual Waste Sampling
- Daily Operating Log
- Monitoring Report
- Annual Report
- Closure Trust
- Post-Closure Trust
- Special Liability Insurance

Any disposal area for hazardous wastes must have controlled entry. For the more strict scenario, extra personnel are provided as guards to secure the area against trespassing by unauthorized personnel. Any hazardous waste disposal area must also be inspected in detail daily. Extra personnel with the sole job of providing such inspection are provided.

RCRA provisions for hazardous wastes also require recordkeeping and reporting activities. Wastes are to be sampled and analyzed annually, and the results reported annually, along with the groundwater monitoring data, to the EPA Regional Administrator. A daily operating log must be maintained and available for inspection by EPA at all reasonable times. This log should contain the results of the daily inspections and a daily record of the quantity and nature of all wastes placed in the facility. An annual report on all disposal activities must also be prepared and submitted to the EPA Regional Administrator. Additional personnel have been provided for these recordkeeping and reporting activities.

Finally, certain measures must be taken to assure the financial responsibility of the operator of the disposal area. A major part of this responsibility relates to closure of the disposal area. Revegetation of the spent shale can be planned to proceed stepwise as the valley is filled, or can occur after operations have ceased. For the purposes of this study, it is assumed that revegetation will start in the last year of activity for the disposal area for both scenarios. Such an assumption allows ready application of the RCRA regulations. To comply with the proposed RCRA regulations, a trust fund is established before disposal begins to cover the entire cost of revegetation. A second trust is established at the same time to provide funds for 20 years of post-closure monitoring and maintenance of the area. The monitoring will be the same as described before. Finally, an insurance policy is provided in the amount of \$5,000,000 per occurrence and \$10,000,000 annual aggregate for claims arising out of injury to persons or property from the release or escape of the hazardous raw or spent shale into the environment. Such coverage is provided for both sudden and accidental release, and for gradual or steady state release or escape of hazardous raw or spent shale to the environment. The insurance is maintained continuously throughout the operation of the disposal area.

7.2.4 Management of Wastes Other Than Raw and Spent Shale

Solids: Oil Shale processing will produce a variety of solid wastes other than raw or spent shale. A majority of these wastes are the result of pollution control activity. These wastes are listed in Table 7-3 and their eventual disposition shown. For the most part the disposition of these wastes is self-explanatory, but a few features deserve discussion.

First many of these wastes can be handled without significant increase in pollution control costs. Raw shales are generally returned to the retort except in the MIS case. High volume processing wastes such as sulfur and coke are sold as byproducts, although this may not be done profitably. No credit has been given for sulfur recovery in Section 5.0. Bio-oxidation sludges are placed in the spent shale disposal area, preferably near the outer surface, to provide some nutritive value for deep rooted plants established in the later revegetation. These sludges from the treatment of process condensate are expected to be low in inorganic salts since the water is condensed from the retort gas stream. This feature of the process and wastewater management models is discussed more fully in Sections 3.0 and 6.0. Spent shale fines are sent to the disposal area along with the rest of the spent shale. Spent catalysts are shipped to the manufacturers or nearby suitably equipped refineries for recovery.

Two wastes shown in Table 7-3 deserve special mention. The thermal sludge boiler produces a major, nonshale waste which contains the salts and heavy organics from the MIS process retort waters. To manage this waste, a smaller landfill disposal area meeting the requirements of RCRA for hazardous waste disposal is used. The landfill is separate from the raw and spent shale disposal area proposed for the MIS and MIS/Lurgi models. Arsenic wastes are also produced by the TOSCO refinery. Rather than handling this small volume waste on-site, it is shipped to an offsite disposal area equipped to handle such wastes. The cost of such handling is estimated.

Liquids: Water treatment, boiler operation, cooling tower operation, and other processing activities produce liquid wastes which must also be managed. Disposition of these wastes has not been discussed elsewhere in this report, so it will be presented here. Tables 7-4 through 7-7 present the results for

TABLE 7-3. DISPOSITION OF SOLID WASTES OTHER THAN SPENT SHALE

<u>Waste Description</u>	<u>Source</u>	<u>Process</u>	<u>Quantity Per Stream Day</u>	<u>Disposition</u>
Bio-oxidation Sludge	Process Condensate	Paraho	18,900 lbs	Spent Shale
		TOSCO II	15,300 lbs	" "
		MIS	42,800 lbs	" "
		MIS-Lurgi	42,800 lbs	" "
	Domestic Waste	Paraho	56 lbs	" "
		TOSCO II	30 lbs	" "
		MIS	28 lbs	" "
		MIS-Lurgi	36 lbs	" "
Raw Shale Fines	Baghouses	Paraho	1,137,752 lbs	Briquette Plant
		TOSCO II	655,579 lbs	Retort
		MIS	47,057 lbs	Shale Disposal
		MIS-Lurgi	303,756 lbs	Retort
	Scrubber or Pptr.	TOSCO	27,400 lbs (wet)	Spent Shale Disposal
Spent Shale Fines	Scrubbers	Paraho	195,480 lbs (dry)	" " "
		TOSCO II	12,950 lbs (dry)	" " "
		MIS-Lurgi	8,704 lbs (dry)	" " "
	Baghouses	Paraho	562,308 lbs	" " "
	Precipitator	TOSCO II	14,328 lbs	" " "
		MIS-Lurgi	324,707 lbs	" " "
Sulfur	Sulfur Recovery	Paraho	261,360 lbs	Sell as By-product
		TOSCO II	382,540 lbs	" " "
		MIS	285,120 lbs	" " "
		MIS-Lurgi	355,000 lbs	" " "

Table 7-3. (Concluded)

<u>Waste Description</u>	<u>Source</u>	<u>Process</u>	<u>Quantity Per Stream Day</u>	<u>Disposition</u>
API Separator Sludge	Oil/Water Separator	Paraho	-	Return to Retort
		TOSCO II	-	" " "
		MIS	-	Incinerate
		MIS-Lurgi	-	Return to Retort
Snake Oil Coke	Refinery	TOSCO II	1,600,000 lbs ^(a)	Sell as By-Product
Spent Catalysts	"	"	873 lbs ^(a)	Ship to Manufacturer for Recovery
Arsenic	"	"	590 lbs ^(a)	Ship to Hazardous Waste Disposal Off- Site
Thermal Sludge	Thermal Sludge Boiler	MIS	160,000 lbs	On-Site Hazardous Waste Landfill
		MIS-Lurgi	81,260 lbs	

^(a)Reference 8, average values.

Table 7-4. DISPOSITION OF LIQUID EFFLUENT STREAMS FROM THE PARAHO PLANT

<u>Stream</u>	<u>Flow (gpm)</u>	<u>Disposition</u>
Underflow, source water clarifier	30	Spent Shale Disposal
BFW/Cooling Tower Blowdown	1,106	Dust Control/Shale Disposal
Process Condensate ^(a)	586	Cooling Tower Makeup
Potable/Sanitary Wastes ^(a)	28	Spent Shale Disposal
Runoff	272	Dust Control/Shale Disposal

(a) After Treatment

Table 7-5. DISPOSITION OF LIQUID EFFLUENT STREAMS
FROM THE TOSCO II PLANT

<u>Stream</u>	<u>Flow (gpm)</u>	<u>Disposition</u>
Underflow, Source water Clarifier	60	Spent Shale Moistening
BFW/Cooling Tower Blowdown	1,049	Spent Shale Moistening
Process Condensate ^(a)	434	Spent Shale Moistening
Potable/Sanitary Wastes ^(a)	14	Dust Control/Spent Shale Disposal
Runoff	126	Dust Control/Spent Shale Disposal

(a) After Treatment

Table 7-6. DISPOSITION OF LIQUID EFFLUENT STREAMS FROM THE MIS PLANT

<u>Stream</u>	<u>Flow (gpm)</u>	<u>Disposition</u>
Underflow, Minewater Clarifier	380	Dust Control/Shale Disposal
Concentrate, Minewater Treatment	924	Thermal Sludge/Retort Steam
BFW/Cooling Tower Blowdown	483	Dust Control/Shale Disposal
Retort Condensate ^(a)	772	Thermal Sludge/Retort Steam
Gas Condensate ^(a)	1,272	Cooling Tower Makeup
Potable/Sanitary Waste ^(a)	13	Shale Disposal
Runoff	141	Cooling Tower Makeup

(a) After Treatment

Table 7-7. DISPOSITION OF LIQUID EFFLUENT STREAMS FROM THE MIS/LURGI PLANT

<u>Stream</u>	<u>Flow (gpm)</u>	<u>Disposition</u>
Underflow and bypass, Minewater Clarifier	5,780	Low Quality Water Uses ^(a)
Concentrate, Minewater Treatment	122	Thermal Sludge/Retort Steam
BFW/Cooling Tower Blowdown	610	Low Quality Water Uses ^(a)
Retort Condensate ^(b)	862	Thermal Sludge/Retort Steam
Gas Condensate ^(b)	1,272	Cooling Tower Makeup
Potable/Sanitary Wastes	17	Low Quality Water Uses ^(a)
Runoff	160	Cooling Tower Makeup

(a) Includes dust control, spent shale slurring and makeup to thermal sludge unit.

(b) After Treatment

the four processes. The nature and disposition of these wastes is self-explanatory and does not need further discussion. Section 6.0 contains additional background information and flow diagrams for water management, and Section 3.0 contains process descriptions.

7.3 SPECIFIC PARAMETERS AND COSTS FOR SOLID WASTE MANAGEMENT

7.3.1 Disposal Area Dimensions

The disposal area dimensions used for cost calculations for all four processes are summarized in Tables 7-8 through 7-11. To the best of our knowledge, this represents the first time the magnitude of the shale disposal activities has ever been presented in such detail. The exercise truly gives a feel for the vast amount of material which will be handled, and yet the problem does not appear unmanageable. The verticle rise of the landfill for the Paraho and TOSCO process is much larger than that for the other two processes. This is because much deeper valleys are available to Paraho and TOSCO than are available to the others.

A large number of alternate designs are possible for the disposal areas, and no attempt was made to suggest an optimal design. Rather, an attempt was made to present a reasonable and conservative design, which admittedly is an entirely synthetic model since the dimensions of the valleys are taken as regular in shape, while in actuality such areas will be of irregular shape. However, changes in the overall design, which the reader might suggest, should have small impact on the conclusions, since the important feature of any design will be the quantity of material handled. Such quantities are accurately and fairly presented by the models used herein.

7.3.2 Summary of Disposal Costs

Itemized breakdowns of capital and operating costs for the four processes are presented in Tables 7-12 throught 7-15, along with the assumptions behind the figures.

The overall assumptions behind the deviations for each process are as follows.

Table 7-8. PARAHO SPENT SHALE DISPOSAL AREA

Item	Dimensions
Spent Shale Amount	4.162×10^7 tons/year or 3.426×10^7 yd ³ /year
Spent Shale Average Density	90 lbs./ft ³
Verticle Rise Landfill	2,500 feet
Total Operating Time	12.5 years each
Spent Shale Bed Depth	530 feet
Retention Dam Height	364 feet
Disposal Area Surface Area	1,080 acres
Catchment Pond Capacity	340 acre-feet
Area Width	4,240 feet
Area Maximum Length	12,185 feet
Lined Diversion Ditches	23,174 feet
Unlined Diversion Ditches	7,228 feet
Drainage Ditches	~ 20,000 feet
Leachate sump drain	2,635 feet
Fencing, 3 gates	30,403 feet

Table 7-9. TOSCO II SPENT SHALE DISPOSAL AREA

Item	Dimensions
Spent Shale Amount	1.7875×10^7 tons/year or 1.4712×10^7 yd ³ /year
Spent Shale Average Density	90 lbs./ft ³
Verticle Rise Landfill	2,500 feet
Total Operating Time	25 years
Spent Shale Bed Depth	490 feet
Retention Dam Height	337 feet
Disposal Area Surface Area	991 acres
Catchment Pond Capacity	310 acre-feet
Area Width	3,920 feet
Area Maximum Length	12,020 feet
Lined Diversion Ditches	6,776 feet
Unlined Diversion Ditches	22,854 feet
Drainage Ditches	~ 20,000 feet
Leachate sump drain	2,470 feet
Fencing, 3 gates	29,630 feet

Table 7-10. MIS RAW SHALE STORAGE AREA

Item	Dimensions
Raw Shale Amount	1.3574×10^7 tons/year or 1.0474×10^7 yd ³ /year
Raw Shale Average Density,* 30% Voids	96 lbs./ft ³
Verticle Rise Landfill	1,000 feet
Total Operating Time	12.5 years
Raw Shale Bed Depth	462 feet
Retention Dam Height	317 feet
Disposal Area Surface Area	420 acres
Catchment Pond Capacity	131 acre-feet
Area Width	3,696 feet
Area Maximum Length	5,905 feet
Lined Diversion Ditches	6,422 feet
Unlined Diversion Ditches	10,668 feet
Drainage Ditches	~ 8,000 feet
Leachate sump drain	2,355 feet
Fencing, 3 gates	17,090 feet

* Determined from Reference 43.

Table 7-11. MIS/LURGI SPENT SHALE DISPOSAL AREA

Item	Dimensions
Spent Shale Amount	1.9058 × 10 ⁷ tons/year or 1.6608 × 10 ⁷ yd ³ /year
Spent Shale Average Density	85 lbs./ft ³
Verticle Rise Landfill	1,000 feet
Total Operating Time	25 years
Spent Shale Bed Depth	261 feet
Retention Dam Height	179 feet
Disposal Area Surface Area	217 acres
Catchment Pond Capacity	68 acre-feet
Area Width	2,088 feet
Area Maximum Length	5,074 feet
Lined Diversion Ditches	3,668 feet
Unlined Diversion Ditches	8,716 feet
Drainage Ditches	~ 8,000 feet
Leachate sump drain	1,524 feet
Fencing, 3 gates	13,232 feet

Table 7-12.

ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE PARAHO PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Retaining Dam				
Dam Fill (a)	1st & 13th yr	960	1st & 13th yr	960
" "	2nd-9th and 14th-21st yr	425	2nd-9th and 14th-21st yr	425
Soil Cement Face	1st & 13th yr	354	1st & 13th yr	708
" " "	2nd-9th and 14th-21st yr	156	2nd-9th and 14th-21st yr	312
Leachate Sump Drain	1st, 2nd, 13th & 14th yr	84	1st, 2nd, 13th & 14th yr	84
Disposal Area Construction				
Soil Removal	1st & 13th yr	2770	1st & 13th yr	2770
" "	2nd-11th and 14th-23rd yr	920	2nd-11th and 14th-23rd yr	920
Impermeable Lining	1st & 13th yr	1853	1st & 13th yr	3706
" "	2nd-11th and 14th-23rd yr	926	2nd-11th and 14th-23rd yr	1853
Ditch Construction				
Unlined Storm-Runoff Diversión (j) [C]	0 & 12th yr	30	0 & 12th yr	30
" " "	13th and 25th yr	5	13th and 25th yr	5
Lined Storm-Runoff Diversión (b) [C]	0 & 12th yr	36	0 & 12th yr	36
" " "	12th, 13th, 24th & 25th yr	6	12th, 13th 24th & 25th yr	6
Drainage on Shale	6th and 18th yr	15	6th and 18th yr	15
" " "	7th-11th and 19th-23rd yr	9	7th-11th and 19th-23rd yr	9
" " "	12th, 13th 24th & 25th yr	5	12th, 13th 24th & 25th yr	5

Note: [C] indicates that the item was treated as a capital cost.

Table 7-12.(Cont.) ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE PARAHO PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Dust Control				
Chemical Spraying	1st-25th yr	1343	1st-25th yr	1343
Smooth Rolling	1st-25th yr	93	1st-25th yr	93
Catchment Pond ^(c) [C]	0 & 12th yr	1422	0 & 12th yr	1422
Fencing [C]	0 & 12th yr	164	0 & 12th yr	164
"	6th & 18th yr	164	6th & 18th yr	164
"	13th & 25th yr	82	13th & 25th yr	82
Monitoring Wells, 5 @ 500' Average (d) [C]	0 & 12th yr	113	0 & 12th yr	113
Guards, 3 Men, 24 hrs/d, 365 d/yr (e)	---		1st-25th yr	
Annual Report, 2 Men, 2 wks/yr	---		1st-25th yr	3.2
Operating Log, 1 Man, 10 hrs/wk, 52 wk/yr	---		1st-25th yr	10.4
Daily Visual Inspection, 1 Man 365 d/yr	---		1st-25th yr	58
Monitoring Sampling and Analysis, 2 Men, 3 wks, 4x/yr	1st-25th yr	19	1st-25 yr	19
Spent Shale Sampling Anal- ysis, 2 Men, 3 wks/yr	---		1st-25th yr	5
Liability Insurance ^(f)	---		1st-25th yr	80
Revegetation			Trust: 0 and 12th yr (k)	4522
Soil Placement	13th and 25th yr	5670	---	
Triple Superphosphate	13th and 25th yr	13.4	---	
Nitrogen	14th, 15th, 26th and 27th yr	23	---	

Table 7-12.(Cont.) ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE PARAHO PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Seeds	13th and 25th yr	35	---	
Planting	13th and 25th yr	26	---	
Post Closure Monitoring and Maintenance				
Less Strict: 2 Men, 52 wks/yr, 5 yrs (g)	13th-17th and 25th-29th yrs	83		
More Strict: Maintenance and Monitoring			Trust: 1st-25th yr	79.3
Hauling and Spreading Bio-oxidation Sludges ^(h)	1st-25th yr	7.9	1st-25th yr	7.9
Four-Wheel Drive Truck ⁽¹⁾	1st, 6th, 11th, 16th & 21st yrs	10.0	1st, 6th, 11th, 16th & 21st yrs	10.0
Indirect and Distributable Costs	1st-25th yrs plus	10	1st-25th yrs plus	10
" " " [C]	0 yr & 12th yr	882	0 yr & 12th yr	882
" " "	1st yr	1738	1st yr	1738
" " "	2nd & 14th yr	100	2nd & 14th yr	
" " "	13th yr	1789	13th yr	1789
" " "	25th yr	41	25th yr	41

Notes: Table 7-12.

- (a) Approximately 1.4 yrs of retorting at full capacity required to generate enough spent shale for dam.
- (b) Lining is cement.
- (c) Lined with hypolon to make impervious.
- (d) Cased and backfilled as per regulations.
- (e) One full-time guard per gate.
- (f) Estimated upper limit for annual insurance premium. Actual premium determined after engineering evaluation of disposal area by insurance underwriter. Source: John Buckley, Howden Agencies, 6 Commerce Drive, Cranford, N.J. 07016 (201) 272-2500.
- (g) Maintenance suggested by developers, to be provided until all plant species reach maturity.
- (h) Average hauling distance assumed.
- (i) Used for maintenance and inspections in addition to routine operations.
- (j) Two disposal areas are required. Hence, capital items occur twice.
- (k) Factor of 0.781 used for present value of future payment in 12.5 years at a compounded interest of 2%. See proposed regulations for method of calculation.
- (l) Average annual expense of \$71,200. Factor of 14.680 used for future value of the sum of an annuity at 2% compounded interest with annual payments for 13 years. Actual contributions for 25 years since 2 areas required. Annual contribution determined as follows: $\$71,200 \times 16.35 + 14.680 = 79,300$ (see regulations).

Table 7-13.

ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE TOSCO II PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Retaining Dam				
Dam Fill ^(a)	1st yr	501	1st yr	501
" "	2nd-20th yr	170	2nd-20th yr	170
Soil Cement Face	1st yr	159	1st yr	318
" " "	2nd-20th yr	71	2nd-20th yr	142
Leachate Sump & Drain	1st yr	78	1st yr	78
" " " "	2nd yr	79	2nd yr	79
Disposal Area Construction				
Soil Removal	1st yr	948	1st yr	948
" "	2nd-22nd yr	472	2nd-22nd yr	472
Impermeable Lining	1st yr	881	1st yr	1762
" "	2nd-22nd yr	457	2nd-22nd yr	913
Ditch Construction				
Unlined Storm-Runoff Diversion [C]	0 yr	30	0 yr	30
" " "	25th yr	6	25th yr	6
Lined Storm-Runoff Diversion (b) [C]	0 yr	36	0 yr	36
" " "	24th & 25th yr	6	24th & 25th yr	6
Drainage on Shale	17th-21st yr	10	17th-21st yr	10
" " "	22nd-25th yr	5	22nd-25th yr	5
Dust Control				
Chemical Spraying	1st-25th yr	897	1st-25th yr	897
Smooth Rolling	1st-25th yr	38	1st-25th yr	38
Catchment Pond ^(c) [C]	0 yr	1297	0 yr	1297
Fencing [C]	0 yr	160	0 yr	160
"	12th yr	160	12th yr	160
"	25th yr	83	25th yr	83

Note: [C] indicates that the item was treated as a capital cost.

Table 7-13. (Cont.) ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE PARAHO PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Monitoring Wells, 5 @ 500' Average (d) [C]	0 yr	113	0 yr	113
Guards, 3 Men, 24 m/d, 365 d/yr (e)	---		1st-25th yr	526
Annual Report, 2 Men, 2 wks/yr	---		1st-25th yr	3.2
Operating Log, 1 Man, 10 wks/yr	---		1st-25th yr	10.4
Daily Visual Inspection, 1 Man 365 d/yr	---		1st-25th yr	58
Monitoring Sampling and Analysis, 2 Men 3 wks/yr	1st-25th yr	19	1st-25th yr	19
Spent Shale Sampling and Analysis, 2 Men, 3 wks/yr	---		1st-25th yr	5
Liability Insurance ^(f)	---		1st-25th yr	80
Revegetation			Trust: 0 yr ^(k)	3206
Soil Placement	25th yr	5145	---	
Triple Superphosphate, 60 lbs P ₂ O ₅ /acre, \$140/ ton, \$50/ton Shipping	25th yr	12.3	---	
Nitrogen, 60 lbs N/acre as (NH ₄) ₂ O ₄ , \$110/ton, \$50/ton Shipping	26th & 27th yr	21	---	
Seeds, \$4/lb, 8 lbs/acre	25th & 27th yr	32	---	
Planting, \$3/lb of seed	25th yr	24	---	
Post Closure Monitoring & Maintenance				
Less Strict: 2 Men, 52 wks/yr 5 yrs (g)	25th-29th yr	83	---	
More Strict: 5 yrs, 1 Man, 52 wks/yr, 15 yrs, Monitoring: 2 Men 3 wks, 4 x/yr, 20 yrs			Trust: 1st-25th yr (e)	36

**ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE PARAHO PLANT**

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Handling and Spreading Bio-oxidation Sludge (h)	1st-25th yr	6.3	1st-25th yr	6.3
Four Wheel Drive Truck (i)	1st, 6th, 11th, 16th & 21st yr	10	1st, 6th, 11th, 16th & 21st yr	10
Arsenic Waste Disposal (j)	1st-25th yr	9	1st-25th yr	9
Indirect and Distributable Costs	1st-25th yr	10	1st-25th yr	10
plus				
" " " [C]	0 yr	818	0 yr	818
	1st yr	1802	1st yr	1802
	2nd yr	100	2nd yr	100
	25th yr	50	25th yr	50

Notes: (Table 7-13)

- (a) Approximately 2-8 years of retorting at full capacity required to generate enough spent shale for complete dam.
- (b) Lining is cement.
- (c) Lined with Hypalon to make impervious.
- (d) Cased and backfilled as per regulations.
- (e) One full-time guard per gate.
- (f) Estimated upper limit for annual insurance premium for a hazardous waste disposal area. Source: John Buckley, Howden Agencies, 6 Commerce Drive, Carnford, N.J. 07016 (201) 272-2500
- (g) Maintenance provided until all plant species reach maturity.
- (h) Average hauling distance assumed to be 2 miles.
- (i) Used for maintenance and inspection.
- (j) Waste is 20% as by weight; very low solubility; estimated annual leaching from total annual production volume is 28 lbs. Cost determined for shipping and disposal at hazardous waste landfill in Denver at Lowry Bombing Range. Estimated 1 hr/ton at \$40/hr required for special handling at dump site.
- (k) Factor of 0.61 used for present value of future payment in 25 years at a compounded interest of 2%.
- (l) Average annual expense of \$71,200. Factor of 32.671 used for future value of the sum of an annuity at 2% compounded interest with annual payments for 25 years. Annual contribution determined as follows:
 $\$71,200 \times 16.35 + 32,671$ (see regulations).

Table 7-14.

ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE MIS PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Retaining Dam				
Dam Fill	1st & 13th yr	1500	1st & 13th yr	1500
" "	2nd-9th and 14th-21st yr	750	2nd-9th and 14th-21st yr	750
Soil Cement Face	1st & 13th yr	250	1st & 13th yr	500
" " "	2nd-9th and 14th-21st yr	127	2nd-9th and 14th-21st yr	254
Leachate Sump Drain	1st & 13th yr	73	1st & 13th yr	73
" " "	2nd & 14th yr	70	2nd & 14th yr	70
Disposal Area Construction				
Soil Removal	1st & 13th yr	853	1st & 13th yr	853
" "	2nd-11th and 14th-23rd yr	375	2nd-11th and 14th-23rd yr	375
Impermiabie Lining	1st & 13th yr	428	1st & 13th yr	856
" "	2nd-11th and 14th-23rd yr	222	2nd-11th and 14th-23rd yr	444
Ditch Construction				
Unlined Storm -Runoff Diversion [C]	0 & 12th yr	5	0 & 12th yr	5
" " " "	13th & 25th yr	2	13th & 25th yr	2
Lined Storm-Runoff Diversion(a) [C]	0 & 12th yr	10	0 & 12th yr	10
" " " "	12th, 13th, 24th & 25th yr	2	12th, 13th, 24th, & 25th yr	2
Drainage on Shale	9th & 21st yr	3	9th & 21st yr	3
" " "	10th-13th and 22nd-25th yr	2	10th-13th and 22nd-25th yr	2
Dust Control				
Chemical Spraying	1st-25th yr	896	1st-25th yr	896
Catchment Pond (b) [C]	0 and 12th yr	550	0 and 12th yr	550
Fencing [C]	0 and 12th yr	90	0 and 12th yr	90

Note: [C] indicates that the item was treated as a capital cost.

Table 7-14. (cont.)

ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE MIS PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Fencing	6th & 19th yr	90	6th & 19th yr	90
"	13th & 25th yr	52	13th & 25th yr	52
Monitoring Wells	0 & 12th yr	113	0 & 12th yr	113
Guards	-		1st-25th yr	526
Annual Report	-			3.2
Operating Log	-		1st-25th yr	10.4
Daily Visual Inspection	-		1st-25th yr	58
Monitoring Sampling & Analysis	1st-25th yr	19	1st-25th yr	19
Spent Shale Sampling & Analysis	-		1st-25th yr	5
Liability Insurance (c)	-		1st-25th yr	80
Revegetation	-		Trust: 0 & 12th yr (d)	1861
Soil Placement	13th & 25th yr	2336	-	
Triple Superphosphate	13th & 25th yr	5.2	-	
Nitrogen	14th, 15th, 26th & 27th yr	9	-	
Seeds	13th & 25th yr	13.4	-	
Planting	13th & 25th yr	10	-	
Post Closure Monitoring & Maintenance				
Less Strict: See Table 7-12	13th-17th, and 25th-29th yr	83.2		
More Strice: See Table 7-12.			Trust: 1st-25th yr (e)	79.3

Table 7-14. (Cont.)

ITEMIZED COST FOR SOLID WASTE
MANAGEMENT FOR THE MIS PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Hauling & Spreading Bio-Oxidation Sludge	1st-25th yr	17.1	1st-25th yr	17.1
4-Wheel Drive Truck	1st, 6th, 11th, 16th & 21st yr	10	1st, 6th, 11th 16th & 21st	10
Thermal Sludge Disposal ^(f) [C]	0 yr	573	0 yr	573
" " "	1st-25th yr	437	1st-25th yr	437
Indirect & Distributable Costs	1st-25th yr	10	1st-25th yr	10
	<u>plus</u>		<u>plus</u>	
" " " " [C]	0 yr	671	0 yr	671
" " " "	1st yr	1949	1st yr	1949
" " " "	2nd & 14th yr	100	2nd & 14th yr	100
" " " " [C]	12th yr	384	12th yr	384
" " " "	13th yr	2286	13th yr	2286
" " " "	25th yr	40	25th yr	40

Notes: (Table 7-14)

- (a) Lining is cement.
- (b) Lining is hypolon.
- (c) Estimated upper limit for annual insurance premium. Actual premium determined after engineering evaluation of disposal area by insurance underwriter. Source: John Buckley, Howden Agencies, 6 Commerce Drive, Crawford, NJ 07016 (201) 272-2500.
- (d) Factor of 0.781 used for present value of future payment in 12.5 years at a compounded interest of 2%.
- (e) Average annual expense of \$71,200. Factor of 14.680 used for future value of the sum of an annuity at 2% compounded interest with annual payments for 13 years. Calculation same as Paraho case.
- (f) Costs for separate hazardous waste disposal area to manage sludge from thermal sludge boiler.

Table 7-15. ITEMIZED COST FOR ABOVE GROUND SOLID WASTE
MANAGEMENT FOR THE MIS-LURGI PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Disposal Area Construction				
Soil Removal	1st yr	187	1st yr	187
" "	2nd-20th yr	101	2nd-20th yr	101
Ditch Construction				
Unlined Storm-Runoff Diversión [C]	0 yr	4.8	0 yr	4.8
Line Storm-Runoff Diversión [C]	0 yr	13	0 yr	13
Dust Control				
Surface Spraying	1st-25th yr	326	1st-25th yr	326
Catchment Pond [C]	0 yr	298	0 yr	298
Fencing [C]	0 yr	75	0 yr	75
"	12th yr	75	12th yr	75
"	25th yr	35	25th yr	35
Monitoring Wells [C]	-		0 yr	113
Guards	-		1st-25th yr	526
Annual Report	-		1st-25th yr	3.2
Operating Log	-		1st-25th yr	10.4
Daily Visual Inspection	-		1st-25th yr	58
Monitoring Sampling and Analysis	-		1st-25th yr	19
Spent Shale Sampling and Analysis	-		1st-25th yr	5
Liability Insurance	-		1st-25th yr	80
Revegetation			Trust: 0 yr ^(a)	561
Soil Placement	25th yr	878	-	
Phosphate	25th yr	4.7	-	
Nitrogen	26th & 27th yr	8.1	-	
Seeds	25th yr	12.1	-	

Note: [C] indicates that the item was treated as a capital cost.

Table 7-15. (Cont.) ITEMIZED COST FOR ABOVE GROUND SOLID WASTE
MANAGEMENT FOR THE MIS-LURGI PLANT

Cost Item	Less Strict Scenario		More Strict Scenario	
	Timing of Cost	Cost (10 ³ \$)	Timing of Cost	Cost (10 ³ \$)
Planting	25th yr	4.7	-	
Post Closure Monitoring & Maintenance				
Less Strict	25th-29th yr	83.2		
More Strict			Trust: 1st-25th yr (b)	36
Hauling & Spreading Bio-Oxidation Sludge	1st-25th yr	17.1	1st-25th yr	17.1
4-Wheel Drive Truck	1st, 6th, 11th, 16th & 21st yr	10	1st, 6th, 11th, 16th & 21st yr	10
Thermal Sludge Disposal [c]	0 yr	409	0 yr	409
" " "	1st-25th yr	420	1st-25th yr	420
Indirect & Distributable Costs	1st-25th yr	10	1st-25th yr	10
	<u>plus</u>		<u>plus</u>	
" " " " [c]	0 yr	205	0 yr	205
" " " "	1st yr	175	1st yr	175
" " " "	2nd & 25th yr	50	2nd & 25th yr	50

Notes: Table 7-15.

- (a) Factor of 0.610 used for present value of future payment in 25 years at a compound interest of 2%.
- (b) Average annual expense of \$71,200. Factor of 32.781 used for future value of the sum of an annuity at 2% compounded interest with annual payments for 25 years. Calculation same as TOSCO case.
- (c) Cost for separate hazardous waste disposal area to manage sludge from thermal sludge boiler.

Paraho Process: The dam is constructed of spent shale compacted with 56,250 ft/lbs per square foot effort. Extensive studies of the physical characteristics of Paraho spent shale have shown such compaction to produce a material with a permeability of 9.7×10^{-8} cm/sec., [5, 12] which is equivalent to the impermeable clays specified in the RCRA hazardous waste proposed regulations.²³ The entire volume of spent shale used to construct the dam is compacted to the same extent.

The dam outer face is covered with a 6-inch or 1-foot thick "soil-cement" type of cover, where spent shale is used in place of soil in the cement. The permeability of the face is further lowered by this treatment and protected against leaching and erosion. The entire disposal area surface behind the dam is also covered to a depth of 6 inch or 1 foot with the same type of soil cement after removal of the native soil cover, forming an impermeable barrier under the spent shale which is tough and will not cause slippage of the pile. The native soil is removed and the soil cement put in place stepwise as needed as the pile moves up the valley, rather than being completely installed at the start. The drainage ditches on the shale pile itself are lined to prevent erosion. The dam is built up slowly as the spent shale disposal pile depth increases avoiding creation of an area behind the dam which would retain runoff water.

Two such disposal areas will be required for a 25-year operating schedule which has been assumed for the processes. See Section 10.0 for the economic treatment of this staggered investment.

TOSCO Process: For the TOSCO process, spent shale is also used for dam construction is compacted with 56,250 ft/lbs/ft² effort. As in the Paraho case, the disposal of TOSCO spent shale has been studied extensively [14]. These studies have shown that a permeability of approximately 1×10^{-6} cm/sec can be obtained with such compaction. Again, the dam face is lined to prevent leaching and erosion, and to provide an impermeable barrier. The valley floor is also lined with impermeable soil cement made from spent shale as in the Paraho case.

One disposal area will be sufficient for 25 years of operation.

MIS Process: Spent shale with cementation properties is not available in the MIS process for dam construction and valley lining. Hence, for the MIS case a cement faced earthen dam with internal impermeable barrier will be constructed using the raw shale as fill. The valley is lined with three feet of soil cement formed from available soil in the area. No special compaction effort is used in placement of the shale. The finished surface of the shale pile is sprayed with organic binder to control fugitive dusts. In all other respects, the management of solid wastes is handled as in the Paraho and TOSCO cases.

It should be mentioned that application of hazardous waste disposal regulations to raw shale is perhaps illogical and inappropriate. First, raw shale, although freshly mined, is still native rock. Only the act of moving it from one place to another qualifies it for application of this more strict scenario, since no processing has occurred which alters the hazard it poses, other than exposing fresh surfaces. Second, Occidental Oil Shale views this raw shale as a material which is stored for later aboveground processing. Until the idea of aboveground processing is completely abandoned, this raw shale is not classified as a waste and does not fall under the jurisdiction of RCRA. The proposed regulations²³ define a waste as "any material which is: (1) not reused (that is, the material is abandoned or committed to final disposition..." The storage area for such a material "in-process", however, must comply with all applicable air and water emission standards, and many of the provisions of the disposal model used here do serve this purpose. Completing the fill area with revegetation goes beyond storage and suggests final disposition. Two areas are needed for 25 years of operation.

MIS/Lurgi: The aboveground disposal area for the MIS/Lurgi process is small because most of the spent shale is placed back into the burned out retorts by slurry backfilling. The material remaining for disposal provides special advantages for environmentally safe disposal. The spent shale is noncarbonaceous and has the ability to cement readily, forming strong, impermeable ($<1 \times 10^{-7}$ cm/sec) and durable material [5, 25-28]. Hence no dam or valley lining is required since the entire pile is essentially cement. All other features of the disposal model such as diversion ditches and catchment ponds are retained. Also, both regulatory scenarios are applied, although the

authors find it inconceivable that this spent shale would ever be found to be a hazardous waste.

7.4 MANAGEMENT OF SPENT MIS RETORTS

7.4.1 The Regulatory Scenarios

The preparation and burning of an MIS retort to produce oil is discussed in detail in Section 3.0. After retorting is completed, a large cavern remains which contains considerable void space. Both tracts C-a and C-b are in an areas of the Piceance Creek Basin where substantial amounts of groundwater are found, and the potential exists for leaching from such spent retorts. Many people believe that contamination of the aquifer and even surface groundwater could occur from such leaching, but no direct research is yet available to define the problem, if any, in a precise way. The potential also exists for subsidence around abandoned retorts.

Two scenarios were considered by this study for the final disposition of the spent MIS retorts. In the less strict scenario, the spent retorts are abandoned without any further treatment. In the more strict scenario, the spent retorts are grouted to seal them against leaching and subsidence. A discussion of the possible impact of leaving spent retorts without further treatment is not germane to this study, although some theoretical projections are available [44], and other extensive studies are underway [45]. Rather, this study attempts to determine the cost of effectively grouting to seal spent retorts, as an incremented cost which would be incurred if grouting is required by regulation.

In discussing the possibility of future regulation with the EPA, a surprising discovery was made. Strictly speaking, spent in situ retorts may not be defined as a waste, and may not be subject to existing regulations.

Many grouting techniques can be proposed for sealing spent retorts, and recent cost projections of \$5 to \$278 per barrel of oil for grouting each retort have been made by others [46]. Clearly such a cost would be prohibitive. Unfortunately the authors of the above projections did not present the details used in calculating such costs, so their results can not be compared with the techniques used in this study. The results of this study are quite different.

7.4.2 Slurry Backfilling Costs for Spent MIS Retorts

The procedures for slurry backfilling of spent retorts is described in detail in Section 3.0. Basically, the raw shale which is mined to provide void space for the MIS retort, is retorted above ground using the Lurgi-Ruhrgas process to produce a noncarbonaceous spent shale. A portion of this spent shale is slurried with water and pumped back into the spent MIS Retorts, filling the void space and setting as a impermeable cement. The details of the slurry backfilling system and the formulation of the slurry are proprietary and cannot be discussed in greater detail than is presented in this report.

The costs shown in Table 7-16 relate to the production and placement of a cement which will prevent teaching and subsidence of spent MIS retorts. These costs can be viewed as pollution control costs. From a production standpoint, it is desirable to do more. Cementation of spent MIS retorts with a stronger cement would allow retorting of shale support pillars between retorts, and enhance resource utilization. To produce such a cement with current technology would be more costly. However, research is underway to find ways to produce economically, a very strong cement from spent shale. Eventually, slurry backfilling may be used both to protect the enviornment and to increase oil recovery appreciably from a given plot of land.

7.5 ALTERNATE APPROACHES TO SOLID WASTE MANAGEMENT

7.5.1 Landfill Disposal

Working within the framework of the EPA proposed regulations for hazardous wastes, as is the case with the disposal models presented in this section, several attractive, common sense procedures could not be incorporated into solid waste management. One of the most important changes which is suggested as an alternate approach is continuous revegetation. As can be seen from Tables 7-12 through 7-15, a huge expense is incurred in removing the initial soil cover in the valley and later replacing the soil for revegetation. A majority of this cost can be attributed to hauling to and from the soil storage area. A much more sensible approach would be to remove the soil in front of the advancing disposal pile, and place the soil on finished sections of the pile, revegetating continuously as the disposal area is filled.

Table 7-16. ITEMIZED COSTS FOR SLURRY
BACKFILLING OF SPENT MIS RETORTS

<u>Equipment, Installed</u>	<u>Cost, 10³\$</u>
10,000 Ton Storage Silo	1,000
Conveyer and Weighing System	30
Additive Storage Silos, 2 @ 500 tons	100
Mixing Tank and Mixer, 30'x30', 200 hp	100
Pumps, 4 @ 150 hp	80
Installation and Controls	80
Piping, 7,500', 18", 250 psi., Carbon Steel	225
Welding and Installation	50
Holding Tanks, 2 @ 100,000 gallons	100
Holding Pond, Butyl Lined, 23.6 Acre-feet	800
Miscellaneous Equipment and Contingencies	<u>435</u>
Total Capital	3,000
<u>Operating Expenses, Annual</u>	<u>Cost, 10³\$/year</u>
Power, 7,008 mw	200
Labor	700
Chemicals	<u>6,450</u>
Total Annual Operating Expenses	7,350

Not only would a large dollar savings result, but this approach would be preferable from an environmental and aesthetic standpoint. Continuous revegetation would allow exposure of a minimum disposal area and hence reduce fugitive dusts significantly. The procedures followed in the models discussed in this section result in maximum exposed spent shale.

However, the financial accountability provisions of the proposed regulations cannot accomodate the suggested alternative. Establishment of a revegetation trust is meaningless if this activity is viewed as a continuous operating procedure. Notification of intent to close also cannot be given because closure is an ongoing activity almost from the start. When the disposal area reaches maximum capacity, closure is completed almost simultaneously and has been completed for most of the area much earlier. The requirement then for post closure monitoring and maintenance also becomes definitely difficult. Maintenance will have already occurred on most parts of the disposal area for a varying number of years, up to, in some cases, the required 20-year period. It is suggested that provisions be made in the new proposed RCRA regulations for continuous revegetation when it is feasible and environmentally desirable.

Some concern should also be expressed for lining the valleys used for spent shale disposal with impermeable barriers. Such linings should prevent leachate from reaching ground waters, but ground waters may also be trapped under the impermeable linings causing potential disposal area instability. Of critical concern, in addition to leaching, should be disposal pile stability. From the engineering standpoint, the most common type of slope failure, and one which is most amenable to quantitative investigation, is the rotational shear slip which occurs in cohesive soils [25]. Similar problems may be anticipated in building up deep dumps.

From the compressive and triaxial strength tests of spent oil shale ash burned in the range of 1300°F with 10% water added [25], it is noted that the angle of internal friction is approximately 8° and the cohesive strength is approximately ten times that of a stiff clay. Thus, the physical characteristics of the spent shale itself would allow higher angles of slope and/or higher critical heights than for stiff clays. However, impermeable barriers which may trap ground water may cause foundation failure, allowing rotational shear slip from the bottom, and hence failure of the disposal area which would not otherwise occur without an impermeable barrier.

7.5.2 Disposal in the Mine

The developers of the four processes considered in this study have mentioned that disposal of spent shale (raw shale in the MIS case) back in the mine may be investigated in the future. Such a procedure may prove to be attractive from aesthetic, environmental and resource utilization standpoints for some of the processes. Backfilling of mines with tailings [49, 50, 52, 53] or other materials [51] is used in other mining industries. All the spent shale produced by retorting may not be placed back in the mine, but certainly a much smaller above ground disposal area would result. Assuming the spent shale could be adequately compacted floor to ceiling, this disposal approach might also provide roof support sufficient to allow mining of the pillars which otherwise represent an unused shale resource [53].

7.5.3 Commercial Uses of Spent Shale

Although commercial use of the vast quantities of spent shale which will be produced is unlikely, a portion of this material which this study has considered as waste may find its way into use rather than disposal. A number of uses have been investigated [26]. Among the attractive alternatives are: Asphaltic concrete; highway use lightweight aggregate; structural concrete and concrete blocks, rubber filler; road base and sub-base materials; building brick; and portland cement. The possibility that such markets may exist for even a small portion of the spent shale is worth further investigation as the industry moves toward commercialization.

7.6 INTERPRETATION OF RESULTS

The authors wish to re-emphasize that the above ground disposal models used to determine costs do not represent the only approach possible. The models chosen represent stringent solid waste management procedures, and, as such, should give reasonable upper limits for cost.

The authors suggest that the reader who finds fault with the models should determine the cost for his own model in equal detail. Doing so should facilitate and stimulate productive discussion of the differences.

8.0 TRACE ELEMENTS

8.1 SUMMARY AND CONCLUSIONS

This chapter considers the source, partitioning, and fate of the trace elements Ag, As, B, Be, Cd, Cr, Cu, F, Mn, Hg, Mo, Ni, Pb, Se, U and Zn. Laboratory and pilot-scale studies of the retorting process serve as a basis to speculate on the route of trace elements in full-scale oil shale plants equipped with complete pollution controls.

Water Discharges. In the less strict scenario all trace elements with the possible exception of B and F are below the strictest Colorado water quality criteria. In the more strict scenario concentrations are typically reduced below measurable levels.

Solid Wastes. The trace element content of spent and raw shale is similar to soils in the Piceance Creek Basin; trace elements from windblown dust originating from these sources should not present a problem. In waste from the thermal sludge unit the elements As, B, F, Hg, Mo, and Se should be elevated over local soils, although not necessarily to hazardous levels. Trace elements in leachate from the spent shale pile are recycled into the process and are therefore not in a waste stream; however, concentrations in leachate should be typical of surface waters with the possible exceptions of As, Mo and Se, which could be elevated.

Air Emissions. Laboratory studies indicate that the elements B, Be, Cd, Co, Cr, Cu, F, Mn, Mo, Ni, Pb, V and Zn do not partition measurably into the retort gas, but that 0-90% of the Hg may do so. Results for As and Se are less clear. Once in the retort gas stream, Hg is likely to be removed at numerous sites in the plant, including the Stretford unit (all processes),

catalyst and catalyst protectors (TOSCO II), product oil (Paraho), scrubbing liquors, and electrostatic precipitators (Paraho). Under completely uncontrolled conditions, Hg emissions could vary from negligible amounts up to approximately 27,000 g/d (~10 times permitted levels), although the latter would require a set of unlikely circumstances.

8.2 INTRODUCTION

It is well established that volatile trace elements in coal such as As, Hg and Se are partitioned to the smallest particles during the combustion process. For example, Wallace [1] and Davison et al. [2], in their studies of a coal-burning power plants, have shown that Pb, Tl, Sb, Cd, Se, As, Zn, Ni, Cr and S are at least ten times more concentrated in the smallest fly ash particles than in the bulk coal ash. Radian Corporation [3] in their study of three different coal burning power plants has shown that the elements Sb, As, B, Cd, Cl, Cr, Co, F, Pb, Mg, Hg, Mo, Ni, Se, Ag, S, U, V and Zn are at least three times more concentrated in the entrained fly ash particles leaving the stack than in the bulk coal ash. In the Radian study 86-96% of the Hg in the coal traveled through the particulate control equipment and left the stack as entrained fly ash particles. It has been proposed that this partitioning of trace elements is due to volatilization during the combustion process followed by condensation on available fly ash surfaces. Linton et al. [4] measured elemental surface concentrations on fly ash particles, and found that the elements C, K, Na, P, S, Be, Cr, Cl, Mn, Pb, Tl, V, Zn and occasionally Ca are concentrated preferentially on the surface.

Because the volatilization process should apply generally to combustion or roasting processes, it is necessary to investigate similar partitioning of trace elements during oil shale retorting. The interest in oil shale arises because of the large mass of ore which must be retorted, with the potential to release large masses of trace elements per unit of energy produced. (The heat content of 30 gal/ton oil shale is 2,800 Btu/lb. compared to approximately 10,000 Btu/lb. for coal.)

However, since major differences exist between coal combustion and oil shale retorting in regard to the combustion process, temperature, mineral content, and organic matrix, the results discussed above cannot be merely

extrapolated to assess oil shale retorting. In addition, the emission control equipment differs significantly for the two processes as described in Chapters 5 through 7.

Interest in the mobility of trace elements during oil shale processing is also due to two additional considerations. First, excess mine drainage waters can contribute toxic trace metals to surface waters, rendering them less desirable for agricultural or drinking purposes. Second, the oxidizing, fracturing, and grinding of the oil shale may change (increase or decrease) the mobility of trace elements in spent shale.

This chapter considers the source, partitioning, and fate of trace elements during the oil shale retorting processes.

8.3 MINE DRAINAGE WATER

8.3.1 Composition of Untreated Mine Drainage Water

As discussed in Chapter 3, excess groundwater may be discharged during mining or dewatering operations. This water will then be used in the retorting process or will be treated and discharged to a surface stream.

As an aid in estimating the levels of trace elements in such discharges, Table 8-1 summarizes mean concentrations of groundwaters in tracts C-a and C-b, along with concentration in Corral Gulch (east of tract C-a) and miscellaneous springs and seeps.

At first glance wide variations are apparent; these deserve closer attention. Notice that concentrations in the lower aquifer are elevated for B and F compared to the upper aquifer. The variation is reasonable because the two aquifers are isolated and vary widely in gross parameters such as salinity. For the purpose of this study it is assumed that the lower aquifer will not contribute to mine drainage water.

The variation in data can also be explained in part by comparing the concentration levels to the lower working concentration limits (LWL) for a local Denver laboratory, typical of water quality laboratories which are using "Standard Methods." As can be seen, concentrations of Ag, As, Be, Cd, Cr, Cu, Mo, Ni, Pb and Se are close to or below the LWL, suggesting that analytical error could be a major source of variation for these species. Alternative explanations are that samples were preconcentrated prior to analysis or that more sensitive (but probably less precise) analytical methods were employed.

In addition, the methods which were used to calculate the averages in Table 8-1 must be considered. When concentrations are presented as upper limits (e.g., $\leq 10 \mu\text{g/l}$), they may be included or not, thereby affecting the calculated average.

Such explanations do not account for the variation seen between the upper aquifer in tract C-a and tract C-b for F (30X) and Pb (13X). For the purpose of this study the estimated average concentrations of F and Pb must be considered uncertain.

The last column in Table 8-1 represents the estimated average trace element content in mine drainage waters extrapolated from concentrations in the upper aquifer. Ranges are also presented for F and Pb. This extrapolation is based on the assumptions that mine drainage waters are typical of the aquifer which is contacted, and that the lower aquifer will remain isolated (see Chapters 3 and 6).

8.3.2 Discharge Composition

Although the treatment methods for mine drainage waters are designed to reduce salinity, suspended solids, fluorine, boron, ammonia and phenol, significant amounts of additional trace elements would be removed simultaneously. Table 8-2 summarizes the estimated removal efficiencies for some of the more important trace elements for a single-stage reverse osmosis (RO) unit, as is included in the "less strict" treatment scheme. Removal efficiencies are based on measurements reported in the literature for more concentrated solutions, or were estimated based on chemical similarity [9, 10].

Considering only the effect of the reverse osmosis units shown in Figures 6-12 and 6-13, Table 8-2 summarizes the expected concentrations of trace elements in discharge waters, as well as the total mass of trace elements reporting to the thermal sludge unit. The latter was calculated by assuming that 100% of trace elements report to the thermal sludge unit, and therefore represents a worst case situation.

The untreated mine water should be clean enough to meet the strictest Colorado water quality criteria for As, Be, Cr, Ni, Se, and U, and clean enough to meet agricultural criteria for all remaining elements with the possible exception of B. (B and F are discussed more thoroughly in Chapter 6). The less strict treatment scheme should reduce all trace elements,

Table S-1 Trace Constituents in Groundwaters

	Tract C-a				Tract C-b			Lower Working Concentration Limit, Denver Laboratory	Values Adapted For This Study
	Dewatering Wells (b)	Corral Gulch East of Tract C-a (b)	Upper Aquifer (e)	Upper Aquifer (c)	Springs and Seeps (c)	Upper Aquifer (d)	Lower Aquifer (d)		
Ag ($\mu\text{g/l}$)	14.4	<1.5	12	--	<7		20	.05 (a), 10	10
As ($\mu\text{g/l}$)	<10	5.3	10	30	3	14	30	2 (a), 10	10
B (mg/l)	.26	.12	.33	1.6	.7	1.8	60	.05	1.0, .4-2
Be ($\mu\text{g/l}$)	--	<3.3	<10	1	2	--	--	5	2
Cd ($\mu\text{g/l}$)	<10	3.4	9.6	2	1.0	--	--	2	4
Cr ($\mu\text{g/l}$)	<11	2.8	12	<10	<10	--	--	10	<10
Cu ($\mu\text{g/l}$)	<63	2.2	74	30	80	--	--	10	50
F (mg/l)	.69	.49	.4	16	1.2	12	23	.1	7, .5-20
Mn ($\mu\text{g/l}$)	150	17	130	70	200	82	86	5	100
Hg ($\mu\text{g/l}$)	3.5	.02	3.6	<1	.6	.6	.6	.02	2
Mo ($\mu\text{g/l}$)	--	37	130	--	15	27	57	5 (a), 200	80
Ni ($\mu\text{g/l}$)	19	8.6	19	30	12	19	12	10	20
Pb ($\mu\text{g/l}$)	141	1.9	173	50	14	13	34	1 (a), 50	100, 10-200
Se ($\mu\text{g/l}$)	--	3.0	<9.8	5	<5	--	--	5	<5
Zn ($\mu\text{g/l}$)	298	12.5	263	60	130	310	290	5	200
U ($\mu\text{g/l}$)	--	<9.7	--	--	--	30	.02	--	20

(b) NPDES Permit Application No. CO-00345045 (reference 5)

(2) Detailed Development Plan and Related Material, Ashland Oil Inc., Shell Oil Co., 1976 (reference 6)

(3) First Year Environmental Baseline Program, Ashland Oil, Inc., Shell Oil Co., 1976 (reference 7)

(4) Final Environmental Baseline Report for Tract C-a and Vicinity, Gulf Oil Corporation, Standard Oil Co. (Indiana), 1977 (reference 8)

(a) The lower limit is for flameless atomic absorption, a technique which has only recently become common in routine analytical services laboratory. The second number is for alternate techniques which commonly pre-dated the flameless technique.

Table 8-2 FATE OF TRACE CONSTITUENTS IN MINE WATER

Constituent	Mine Water Concentration	Strictest Water Quality Specification	Estimated Removal Efficiency, One R.O. Pass	Discharge Concentration, Less Strict Regulation	Discharge Concentration, More Strict Regulation	Maximum Quantity Reporting to Thermal Sludge Per Day	
						874 gpm	6478 gpm
Ag (ug/l)	10	.1	98%	.2	<.01	48 g	360 g
As (ug/l)	10	50	50%	.6-1.2	<3	48 g	360 g
B (mg/l)	0.4-4	0.75	80%	.04	.04-.08	19 kg	140 kg
Be (ug/l)	2	10	98%	<.5	<.01	10 g	74 g
Cd (ug/l)	4	.4	99%	0.04	<.01	19 g	.74 g
Cr (ug/l)	<10	50	95%	1.5	<.01	48 g	360 g
Cu (ug/l)	50	10	99.5%	0.05	<.01	240 g	1.8 kg
F (mg/l)	0.5-20	1.4	90%	0.05-1.5	0.15	96 kg	710 kg
Mn (ug/l)	100	50	99.99%	0.01	<.01	478 g	3.5 kg
Hg (ug/l)	2	.05	99%	0.02	0.0002	10 g	74 g
Mo (ug/l)	80	Undetermined	95%	4	0.02	380 g	2.9 kg
Ni (ug/l)	20	50	98%	.4	<.01	96 g	710 g
Pb (ug/l)	10-200	4	95%	0.2-4	0.004-0.08	960 g	7.1 kg
Se (ug/l)	<5	10	50%	3	<2	24 g	180 g
U (ug/l)	20	50	90%	2	.2	96 g	710 g
Zn (ug/l)	200	50	50%	40	8	960 g	7.1 kg

with the possible exception of B and Ag, to levels well below the strictest criteria, while the most strict scenario typically reduces concentrations of trace elements below measureable amounts.

The reject water carrying the concentrated minor and major elements is sent to the thermal sludge unit. Using the mine water source figures from Chapter 6, the maximum quantity of trace elements reporting to the thermal sludge from the excess mine water treatment units has been calculated on a per day basis. Those figures appear in the last two columns of Table 8-1 under their respective feed rates. The total trace element collection in the thermal sludge is calculated in Section 8.6.

8.4 TRACE ELEMENTS AND ELEMENTAL BALANCES IN THE RETORTING PROCESS

8.4.1 Typical Concentrations in Oil Shale, Shale Oil, Spent Shale, and Product Gas

Table 8-3 lists the concentrations of trace elements which have been measured in raw oil shale. The TRW/DRI study shown in column 1 is actually a review of seven previous studies; the numbers listed are typical of Colorado oil shales. The concentrations listed in column 2 are measurements made by Shendrika and Faudel on oil shale from the Mahogany Zone of the Colony site in Colorado. Three samples were analyzed in this study. The values from Fox et al. are also for Colorado oil shale. As can be seen, the concentrations of all elements except mercury and selenium from the various studies agree within a factor of three.

The variance in the mercury concentrations may be due to contamination from drilling mud which is often used in obtaining core samples, since it has been reported that mercury is a common contaminant of drilling muds. Shaw [11] reported that mercury levels measured in oil shale cores were typically 0.4 $\mu\text{g/g}$ but that these levels typically dropped to 0.2 $\mu\text{g/g}$ when drilling mud was removed more thoroughly. Therefore, a Hg concentration of 0.2 $\mu\text{g/g}$ was adopted for this study.

The variance apparent in the selenium measurements could be due to differences in the analytical procedures used. The determination of selenium even in relatively pure waters has been difficult and the method employed by Shendrikar and Faudel is likely to be subject to interferences. For these

Table 8-3. Concentrations of Selected
Trace Elements in Raw Oil Shale ($\mu\text{g/g}$)

	TRW/DRI Review (12)	Shendrikar and Faudel (13)	Fox et al, (14)	Values Adapted for This Study
	Green River Formation	Colony Site Mahogany Zone	Colorado	
As	35	60	28-51	40
Be	1.3	1.0	--	1
B	65	63.3	--	60
Cd	1	1.25	.59-1.3	1
Cr	34	41.7	--	35
Co	10	6.5	--	10
Cu	37	47.5	29-41	40
F	1000	1162	--	1000
Hg	.4	--	.06-.08	.2
Mn	250	230	--	250
Mo	10	30	--	10
Ni	25	23.9	--	25
Pb	20	29.3	17-24	20
Se	1.5	14.6	--	1.5
Sb	1	--	--	1
V	100	57.1	--	100
Zn	70	65	53-67	70

reasons the lower value for selenium represented by TRW/DRI review is adopted for this study.

Column 4 in Table 8-3 summarizes the concentrations of trace elements in raw oil shale which have been postulated for purposes of this study.

Table 8-4 lists typical concentrations for these same trace elements in spent shale. As can be seen, the different studies typically agree within a factor of 2 except for selenium. For the purpose of this study the lower level of 0.3 $\mu\text{g/g}$ will be accepted for the same reasons discussed in the previous paragraph.

A comparison of Tables 8-3 and 8-4 indicates that the concentrations of trace elements are similar in raw shale and spent shale with the possible exceptions of mercury and selenium, both of which are depleted approximately five times in the spent shale. This has led to the hypothesis that these elements volatilize from the raw shale during the retorting operations and ultimately deposit in the product or effluent streams. However, it must be emphasized that the concentrations in Table 8-3 and 8-4 were not necessarily measured on the same oil shale, so that such effects may be due to natural variability.

The data in Table 8-4 also indicates that the trace element content of spent shale is not significantly different from typical soils in the Piceance Creek Basin. Had the trace element content of spent shale been particularly high, one might argue that windblown spent shale would generate undue exposure to trace elements. This route of exposure does not appear to be a problem and will not be considered further in this report.

Table 8-5 tabulates the concentrations of trace elements in crude shale oil. This table indicates a somewhat larger variation in the concentrations measured by the various investigators, possibly due to the difficulty in determining the trace elements at the levels shown in Table 8-5.

Table 8-5 also indicates that trace elements in shale oil with the exception of arsenic, are typically present at levels less than or in the range of typical crude petroleum oils. As an example the mercury content of shale oils is typically eighty times less than that found in California crude. The pathway and fate of arsenic will therefore be considered in more detail in the following sections, but the remaining trace elements in shale oil do not appear to be reasons for concern.

Table 8-4 Concentrations of Selected
Trace Elements in Spent Shale (µg/g)

	TRW/DRI Review (12)	Shendrikar and Faude1 (13)	Fox, et al., (14)	TRW/DRI Review (12)
	Misc. Shales	Colony Site Fischer Assay	Colo.Oil Shales Simulated In Situ Retort	Soils in Piceance Creek Basin
As	25	60	26-76	6
Be	1	1.2	--	2.4
B	100	54	--	61
Cd	.5	1.5	.63-.91	-
Cr	100	53.8	--	60
Co	16	8.0	--	8
Cu	26	55	39-56	30
F	900	1352	--	500
Hg	.04	-	<.01-.42	.04
Mn	250	275	--	490
Mo	10	37.7	--	5
Ni	38	27.5	--	21
Pb	34	34.5	25-40	26
Se	.3	16.7	--	.3
Sb	.5	-	--	.9
V	80	70.6	--	56
Zn	35	76	69-111	80

Table 8-5 Concentrations of Selected Trace Elements in
Shale Oil ($\mu\text{g/g}$)

	TRW/DRI (12)	Shendrikar & Faudel (13)	Fox <u>et al.</u> (14)	TRW/DRI (12)
	<u>Miscellaneous</u>	<u>Fischer Assay</u>	<u>Simulated In Situ Retort</u>	<u>Miscellaneous Crude Petroleum</u>
As	20	21	26-76	.002-.7
Be	.02	0.0	--	.0005-.5*
B	5	.48	--	.002-.2*
Cd	.1	0.0	.59-1.3	.003-1
Cr	.2	.8	--	.002-.7
Co	1	.26	--	.003-13
Cu	.5	.25	.042-12	.2-.9
F	1	<1.0	--	.004
Hg	.3	--	.029-.35	.03-23
Mn	1	.05	--	.05-1.2
Mo	2	.50	--	8
Ni	5	1.4	--	.6-100
Pb	.5	.14	<.6-2.1	1-4*
Se	.4	0.0	--	.4-1.1
Sb	.02	--	--	.06-.30
V	3	.48	--	.7-1100
Zn	2	1.4	.41-3.6	.7-10

*Residual Fuel Oil #6

Table 8-6 presents the range of concentrations measured in whole retort waters. Although each of the authors indicates that concentration ranges may vary an order of magnitude, the ranges presented by the various authors typically overlap. Although values higher than those shown in Table 8-6 for copper were reported by Fox *et al.* [14], these were not included due to likely contamination problems in the retort water collection system. Of the various processes examined, the TOSCO II process produced the lowest levels of trace elements in the retort waters, similar to the waters produced by the Fischer Assay. Except for the TOSCO II process, there is not sufficient data to differentiate between waters produced by the various retorting processes. Although not shown here, the trace element content of retort waters is in the range encountered in typical municipal effluents with the possible exception of arsenic, boron, fluoride and mercury [12].

Leachates from spent shale piles are described in Table 8-7. Three types of experiments are summarized. In the "blender experiments" spent shale and water are vigorously mixed. These experiments indicate that Se, Mo, B and F--those elements forming anionic species under basic conditions--are most readily extracted from spent shale. The remaining elements shown in column 2 are not appreciably extracted.

In the column experiments shown in Table 8-7 a column of spent shale is eluted with water and the various fractions of the eluant are collected.

The only data from actual field plots is shown in column 4. The wide range of concentrations shown can be explained by the different types of samples that were collected: The low concentrations are typically from surface runoff during natural rainfall, while the high concentrations represent leachate that was obtained by forcing water through the spent shale pile. It should be realized that rainfall was insufficient to produce measurable amounts of leachate.

With the possible exception of Se, Mo, and As, concentrations in leachate do not differ greatly from typical groundwaters in tracts C-a and C-b. (See Appendix 4.0, Section A-3 through A-5.) It should be realized, however, that leachate and runoff from spent shale piles will be collected and will serve as a source of water for the retorting process. No discharge of runoff or leachate is anticipated.

Table 8-6 Concentrations of Selected Trace
Elements in Whole Retort Waters (µg/ml)

	TRW/DRI (12)		Shendrikar and Faude1 (13)	Fox <u>et al</u> , (14)
	<u>Simulated In Situ</u>	TOSCO & Paraho	Fischer Assay	<u>Simulated In Situ</u> retort
As	.3-6	1	0.0	.23-6.5
Be		-	0.0	--
B	.3-6	.4-5	.55	--
Cd	.001-.003	-	0.0	.001-.005
Cr	<.012-.02	.007-.3	.004	--
Co	.07-.7	.005	.005	--
Cu	.003-5	.2	.16	.03-.46
F	25	.3-7	2.6	--
Hg	<.01-.4	-	--	.01-.34
Mn	.02-.1	.02-.3	.02	--
Mo	.1-.47	.006-.1	.006	--
Ni	.26-1	.03-.2	.034	--
Pb	.01-.1	<.002-.2	0.0	<.02-.22
Se	.005-1.0	-	.1	--
Sb	.007-.016	-	--	--
V	.07-1.2	.002-.03	0	--
Zn	.04-5	.04-.4	.045	.2-6.3

Table 8-7 Concentration of Trace Elements
Leached from Spent Oil Shale (12)

	<u>Blender Experiments</u>		Column Experiments	Leachates from TOSCO II Field Plots ($\mu\text{g/ml}$)
	($\mu\text{g/ml}$)	% of element in shale	($\mu\text{g/ml}$)	
Cd	.004	.8	--	.003-.006
Se	.05	17	.005-.05	.005-2
Mo	<.05-8	2-80	.2-10	.04-74
Pb	<.05	.01-.4	--	.003-.009
As	.1	.4	.002-.04	.005-.2
Cr	<.05-1.3	.05-1	--	.004-.07
Cu	.01	.04	--	.02-.2
B	.4-12	1-12	5-10	.02-.9
Zn	.02-.12	.05-3	--	.01-3
F	3-80	2-9	10	.02-17
Be	--	--	--	.006
Hg	--	--	--	.00002-.0005
Sb	--	--	--	.001-.003
Co	--	--	--	.001-.04
Ni	--	--	--	.05-.6
V	--	--	--	.003-.1
Mn	--	--	--	.004-.5

Few measurements have been made of trace elements in the product gas, either as entrained particulate matter or as gaseous compounds. The available data is summarized in Table 8-8. The studies by Fruchter et al. (1977), include both gaseous compounds and particulate matter less than 0.5 mm in diameter. The measurements by Fox et al. [14], include both particulate and gaseous mercury. The measurements by Shendrikar and Faudel [13] were performed on the retort gas from a Fischer Assay and should therefore be predominantly gaseous or very fine particulate matter. The measurements by both Fruchter and Fox were performed on a simulated in situ retort.

8.4.2 Elemental Balances

Most evidence for emissions in the retort gas have come indirectly from mass balance studies. Here one attempts to account for the entire mass of each trace element in the raw shale by analyzing the product oil, water, and spent shale. Ideally the trace element content of the gas stream should be measured as well, although the gas measurements are difficult and consequently have seldom been performed. In addition to providing estimates of gaseous emissions, mass balance studies also provide a quality assurance function in the sense that mass closure implies that the analyses have been completed with sufficient precision and that the total mass of each element is accounted for. Mass balance studies also provide information on the volatilization mechanism discussed earlier.

The results of five mass balance studies are shown in Table 8-9. In this table the fraction of total recovered element which is found in each product is presented. The percent of unrecovered mass represents that fraction of the original element which cannot be found in the spent shale, shale oil, or water. Hence a positive percent unrecovered suggests the escape of a gaseous product. The data from Shendrikar and Faudel [13] and Donnell and Shaw [16] are from Fischer Assays of Colorado oil shale. Retorting was carried out at 500°F in an inert atmosphere. The remaining studies were carried out in the Laramie Energy Research Center's controlled state simulated in situ retort and Lawrence Livermore Laboratory's 125 kg oil shale in situ retort. The controlled state retort contains 20 kg of oil shale and the Laramie retort contains 125 kg. The simulated in situ retorts were operated with both inert and oxidizing atmospheres in the temperature range of

Table 8-8 Measured Concentrations of Gases and Fine
Particulate Matter in Product Gas ($\mu\text{g}/\text{SCM}$)

	Fruchter et al. (15) <u>($\mu\text{g}/\text{SCM}$)</u>	Fox et al. (14) <u>($\mu\text{g}/\text{SCM}$)</u>	Shandrikar & Faude1 (13) <u>($\mu\text{g}/\text{SCM}$)</u>
As	15		.5
Cr	92		
Fe	126		
Hg	2.4	0-3	
Zn	40		

Table 8.9 Trace element partitioning during laboratory retorting (%).
The percentages represent the fraction of recovered element
found in oil, water, and spent shale.

	Snedorikar & Faudel (13)				Wildeman & Meglen (17)				Fox et al. (14)				Fox et al. (12)				Donnell & Shaw (16)			
	Fischer Assay				Simulated <i>In Situ</i> Retort				Simulated <i>In Situ</i> Retort				Simulated <i>In Situ</i> Retort				Fischer Assay			
	Spent Shale	Oil	Water	% Unrecovered	Spent Shale	Oil	Water	% Unrecovered	Spent Shale	Oil	Water	% Unrecovered	Spent Shale	Oil	Water	% Unrecovered	Spent Shale	Oil	Water	% Unrecovered
As	94.8 - 95.2	4.8 - 5.2	0.0	10.4 - +12.6	94.7	5.2	.1	+1.6	98.6 - 99.2	.6 - 1.4	0.0	-43 - +21								
Be	100.0	0.0		-5.4 - +2.8																
B	99.8 - 99.9	.1		+2.4 - +29.1																
Cd	100.0	0.0		-6.2 - +.4					99.5 - 100.0	<.1 - .5	0.0	-20 - +26								
Cr	99.8	.2		-7.3 - +4.3																
Co	99.5 - 99.6	.4 - .5		-11.7 - -.3																
Cu	99.9	0 - .1		+4.3 - +3.3	-	-	-	-	99.9 - 100.0	0.0 - .1	0.0	-12 - -.8								
F	99.9	0.0		-14.0 - +3.4																
Hg	-	-		-96 - +42	-	-	-	-	23 - 100.0	<16 - 98	<.7 - 1.3	+51 - +86	57 - 100	<10 - 43	<.1 - 20	<0 - 88	1.9	24.9	15.5	57.7
Mn	99.9	0.0		-7.2 - +.8																
Mb	98.7 - 99.8	.2 - .3		-4.5 - +.4																
Ni	98.9 - 99.2	.8 - 1.1		-3.5 - +.8	99.5	.4	.1	+2.3												
Pb	99.9	.1		-4.1 - +2.2	-	-	-	-	100.0	0.0	0.0	-29 - -9.1								
Se	99.9	0.0		+9.1 - +5.1	93.5	4.8	1.7	+14.1												
Sb	-	-		-																
V	99.9	.1		-8.0 - +.8																
Zn	99.5 - 99.7	.3 - .4		-6.1 - +.7	99.9	.1	0.0	+17.2	99.8 - 100.0	0 - .2	0.1	-1.1 - -2.8								

540-1,030°C. The data in Table 8-9 do not include measurements from the Livermore retort for Cu, Pb and Zn, since copper and brass fittings apparently contaminated the products [14].

In Table 8-9 the percent of unrecovered mass for Be, Cd, Cr, Cu, F, Mn, Mo, Ni, Pb, V, and Zn varies both positively and negatively and is well within experimental error. These elements therefore show no evidence of vaporization or exit with the product gas stream. Co indicates a slight excess of recovered material over the original mass in the raw shale. This is probably due to measurement errors or slight contamination from a metal surface. Co therefore appears to be retained entirely by the spent shale and shale oil. A cursory examination of the data for B suggests the loss of up to 29 percent of B with the gas. However, examination of the original reference indicates that at least 94 percent of B is recovered in two out of three samples; the data for B is therefore insufficient to imply escape with the gas phase. The remaining elements--Hg, As, and Se--deserve further comment.

Of these three, Hg has been by far the most extensively studied. Fox et al. [18], have measure the mercury directly in the emitted gaseous stream from Laramie's controlled state in situ retort. These measurements accounted for essentially all of the unrecovered mercury. That is, the mercury not accounted for in the spent shale, shale oil, or water was in fact found in the gas stream in either the particulate or gaseous phase. Under the same circumstances, Fox et al. also show that the mercury migrates in a band ahead of the flame front, leaving the oil shale at approximately 300°C and recondensing between 75°C and 300°C. Under the experimental conditions employed, the mercury tends to exit in the gas phase during the final third of the retorting cycle.

This behavior helps to explain the wide variations in the recovery of mercury which have been reported. Although Hg may leave the oil shale as a fine aerosol or as a gas, it may condense at unexpected points in the retort or condensation train; the condensed mercury may become re-entrained in subsequent runs or may be entirely missed by chemical analyses. As an example, Donnell and Shaw [16], found that a large fraction of Hg was deposited on the glassware during the Fischer assay, thereby normally escaping chemical analysis. Another explanation for the erratic results shown in Table 8-9 is that channelling of gas flows allows pockets of Hg to be retained

by the shale, or that cooler parts of the simulated retort retain part of the mercury. The volatilization-recondensation process could also create a non-homogeneous distribution of Hg in the spent shale, thereby making accurate analyses more difficult. It is interesting that in the range of 500-1,030°C, Hg imbalances do not appear to be a function of temperature. Lacking further data, it is assumed for the purpose of this report that Hg partitioning occurs equally in each of the retorting processes. In summary it is clear that in laboratory studies up to 88% or as little as zero percent of the Hg in raw oil shale partitions into the retort gas where it can be present in either a gaseous or particulate state. Although the volatilization of Hg in laboratory studies appears well established, the extrapolation of this mechanism to a full scale retort must involve a good deal of speculation.

The escape of As and Se in retort gas is not nearly as well established. The concentration of As listed in Table 8-8 is not sufficient to account for the missing As shown in Table 8-9, column 4. This situation is further complicated by the difficulty in obtaining accurate and precise measurement for As and Se using the normal chemical methods. However the consistency of positive values for unrecovered selenium does suggest that up to ten percent of the selenium is lost with the retort gas.

The negative values for unrecovered As reported by Fox et al. [14], shown in Table 8-9 need further discussion. All imbalances for As reported by this study are negative for the Livermore 125 kg retort and positive for the Laramie 20 kg controlled state retort. This fact suggests a source of contamination in the Livermore retort. Eliminating the data from the Livermore retort leaves only positive values for unrecovered arsenic in Table 8-9, suggesting that under laboratory conditions small amounts of arsenic may escape with the gas stream. This data suggests that up to 10% of the arsenic in the raw shale escapes with the gas stream. As for Hg, extrapolating the data in Table 8-9 to a full scale retort would also involve a good deal of speculation: Unlike Hg, concentrations of As and Se in the retort gas have not been established by direct gas analysis.

The mass balance studies discussed in the previous paragraphs have also established other measurements which will be useful in determining the fate of trace elements in the oil shale retorting process:

- o In the water phase 75% of the Pb, 50-85% of the Zn, 75% of the As, and 67% of the Hg in retort water occurs as particulate matter.
- o Upon storage, Hg in retort water gradually precipitates, presumably due to bacterial action.
- o The concentration of Hg in product oil gradually increases with increasing retort temperature from approximately 0.1 ppm at 510°C to 0.35 ppm at 1,030°C.

8.5 MERCURY BALANCES

In this section an attempt is made to extrapolate the laboratory data described above to a full-scale oil shale retorting plant. As can be seen by examining each process in detail, substantial differences other than the scale of operation exist between the laboratory models and the anticipated retorting plants, so that emission rates will be presented as uncertain variables. Emphasis is placed especially on the atmospheric emissions of Hg.

8.5.1 Paraho Process

Examination of Figure 3-4 indicates some of the major differences between the Paraho process and the laboratory model systems. The Fischer assay co-condenses water and shale oil so that intimate contact is achieved between the two phases; the Paraho process condenses the water and oil phase separately. The simulated in situ laboratory studies blend the condensed water and oil from each of a series of traps in order to assure homogeneity; the Paraho process keeps each stream separate in order to minimize contamination and treatment problems. The laboratory studies employ a series of condensers to remove the oil; the Paraho process employs a spray coalescer and an electrostatic precipitator to remove suspended particles as well as product oil. Unlike the laboratory model systems, the Paraho process also returns approximately 67 percent of the product gas to the retort, thereby providing a possible means of re-depositing suspended or gaseous Hg onto the spent shale.

In the overall Paraho process shown in Table 3-4, the laboratory studies are most relevant to the hot (300-700°C) middle section of the Paraho retort. The studies by Fox et al. [14, 18] and Donnell and Shaw [16] indicate that at these temperatures between 0-90% of the Hg is partitioned to the gas

phase with the remainder going to the spent shale and product oil, present as a mist or in the gaseous state at these temperatures. As the gas proceeds upward through the cooler (65°C) sections of the retort, Hg should re-deposit, forming an enriched band (Fox et al., [18]). During sufficiently long operation the re-deposition of Hg in the top of the retort should equilibrate, until a point at which the gas leaving the top of the retort should contain 0-90% of the Hg entering with the raw shale. While one could argue for either the high or low extreme, the authors believe additional refinement is not justified.

At this point let us estimate for the purpose of this study the concentration of Hg in the product oil, retort gas, and spent shale as it leaves the hot section of the retort. Assume that 0-90% of the total Hg in the raw shale exits in the retort gas stream. Since the concentration of Hg in the shale oil is lower than many petroleum crude oils, and since the shale oil is a rather minor sink for Hg, its exact concentration in the oil is not critical. Assume therefore, consistent with Table 8-5, a concentration of 0.1 µg/g in the shale oil, with the remainder of the Hg staying with the spent shale as it leaves the hot zone of the retort and proceeds to the cooler zone.

Based on these assumptions, the Hg balance for the hot section of the Paraho retort is shown in Table 8-10. Here the recycled product gas is not included as input. Gas volumes were calculated based on the ideal gas law with $T = 66^{\circ}\text{C}$ (150°F), $P = 0.67$ atm., and the mole weight = 28.4 g/mole. As can be seen, the upper bound for Hg in retort gas exceeds the maximum permissible emissions of 2300 g/d by about ten fold.

What then are possible removal mechanisms for Hg in the plant complex? Examination of Figure 3-4 suggests the following possibilities:

- o Redeposition on the cooled, spent shale via recirculated gas
- o Removal of particulate forms in the mist separator and electrostatic precipitator
- o Condensation in the cooler and ammonia absorber
- o Scrubbing in the H_2S removal system

The remainder of this chapter will evaluate the feasibility of these mechanisms in view of chemical principals.

Table 8-10 Mercury Partitioning Adapted for
this Study for the Hot Section of
the Paraho Direct-Heated Retort

<u>IN</u>	Total Mass (10 ³ lb/hr)	Mercury		
		Concentration	Mass g/d	%
Raw Shale	13,118	200 ng/g	28,587	100.0
Process Air	<u>2,339</u>	~0	<u>~0</u>	<u>~0</u>
Total	15,457		28,587	100.0
<u>OUT</u>				
Spent Shale, leaving the hot section of the retort	10,511	25.0-249 ng/g	27,114- 1,386	94.8- 4.8
Oil Mist from the retort	1,352	.1 µg/g	1,473	5.2
Excess				
Product Gas, Wet, Including NH ₃ and S	<u>3,594</u> <u>5.73x10⁷ m³/d</u>	0-450 µg/m ³	<u>0-25,728</u>	<u>0-90.0</u>
Total	15,457		28,587	100.0%

Any Hg in the form of suspended particles would be removed with almost 100% efficiency (>99%) by the combination of the coalescer and electrostatic precipitator. Although the Hg most likely leaves the oil shale as a vapor, condensation into particulate form may occur rapidly upon cooling. (Unfortunately, the laboratory studies did not differentiate particulate and gaseous Hg.) Possible mechanisms of particle formation include the reactions

where s and g denote solids and gases. For reaction (1) thermodynamic equilibrium is expressed as [19]

$$\log \rho_{\text{Hg}} = 11.46 - \frac{3177}{T} - \log T \quad (4)$$

where ρ = concentration Hg vapor (g/m^3), and T = temperature ($^{\circ}\text{K}$). For the temperature of the electrostatic precipitator, $\rho_{\text{Hg}} = 0.36 \text{ g/m}^3$, well above the expected concentration in the gas stream. Hence, Hg would not condense to elemental, liquid droplets.

For reaction (2) the vapor concentration of Hg is approximated [19] by

$$\log \rho_{\text{Hg}} = 15.345 + \log \frac{(H_2)}{(H_2S)} - \frac{5163}{T} - \log T \quad (5)$$

where (H_2) and (H_2S) are the partial pressures of H_2 and H_2S as given in Table 6-3. For 66°C , $\rho_{\text{Hg}} = 0.056 \text{ g/m}^3$. Except for the possibility of adsorption by active adsorbents, Hg is likely in the vapor phase at this stage in the Paraho process.

However, for the sake of completeness, it is noteworthy that if all of the Hg in the retort gas is removed by the precipitator or coalescer, and is thereby introduced into the product oil, the final concentration in oil would be about $1.7 \text{ } \mu\text{g/g}$, at least 90% as suspended solid material. This value is well with the range of normal crude petroleum oils (TRW/DRI, 1977) and should be amenable to well-established refining procedures.

Since approximately 67% of the retort gas is recycled into the retort, it is conceivable that this same fraction of Hg in the retort gas is redeposited on the cooled, spent shale. This possibility is suggested by the findings of Fox et al. [18], who observed Hg condensing in a band ahead of the flame front in the temperature range 75-300°C, well within the range of the lower section of the retort. Other than expecting increased Hg in the spent shale, further speculation is unwarranted.

After the precipitor, the gases are cooled and scrubbed with water at 35°C (95°F) in order to remove excess water from the gas stream. Previous laboratory studies have exposed retort water intimately to shale, retort gases, and shale oil, and have observed little partitioning to the water phase. Concentrations of Hg in the range 0.01-0.3µg/g of retort water are common. Assuming a net production rate for condensate water of 327×10^3 lb/hr (Figure 6.3) and the above concentration range, 4-1000 g/d of Hg would be partitioned into the condensate water to eventually end up in the solid waste disposal pile. Compared to the total mass of Hg expected in the disposal site, 1000 g/d is a minor perturbation and will not be considered further.

As the gases are cooled in the scrubber system, the possibility of enhancing reactions (1)-(3) arises. However, the concentration of Hg vapor at equilibrium according to reactions (1) and (2) is 45,000 and 1900 µg/m³ respectively compared to a predicted maximum of 450 µg/g in Table 8-10, so that condensation appear unlikely. However, more reactive species or cold spots in the cooling system would begin to cause the formation of HgS.

Of particular interest is the Stretford unit designed to remove H₂S. In this device the H₂S is absorbed in an alkaline solution and is oxidized to elemental S. Although the majority of the oxidation supposedly occurs after the scrubbing section, and although S is removed from the recirculating absorbing solution, the scrubbing solution must nevertheless contain a limited amount of freshly formed S. The reaction

is a well known mechanism for removing mercury vapor in a laboratory, and would likely provide an efficient method for removing Hg vapor. For reaction (6) the equilibrium vapor concentration is given [19] by:

$$\log \rho_{\text{Hg}} = 13.064 - \frac{6215}{T} - \log T \quad (7)$$

For $T = 35^{\circ}\text{C}$ (95°F), $\rho_{\text{Hg}} = 2.5 \times 10^{-10} \text{ g/m}^3$ well below the expected concentration and efficient removal of Hg vapor is favored.

Assuming that all of the Hg in the gas stream (25,728 g/d) is removed by the Stretford unit and exits with the elemental S ($\sim 3.57 \times 10^6 \text{ lb/hr}$) the resulting Hg concentration in the S would be $240 \text{ }\mu\text{g/g}$.

In summary, the route of Hg in the Paraho plant is highly speculative and should be examined experimentally. However, likely sinks for Hg are sulfur, the spent shale, and the product oil. The escape of Hg to the atmosphere would require that Hg remain in the gaseous state past the electrostatic precipitator, followed by rapid conversion to a fine aerosol which could escape the Stretford. Under this unlikely set of circumstances, Hg would be routed to the gas combustor and would escape with the flue gas.

8.5.2 Modified In Situ Process

Partitioning of Hg in the MIS plant can be discussed best by reference to Figure 8-1, which shows a simplified schematic of the most relevant processes.

This process differs significantly from the laboratory studies. In particular, in the MIS process two water streams are produced and kept separate, while in the laboratory simulations all water streams are combined. The particle size and the gas composition is certainly less uniform in the MIS process than in simulation studies. In addition, the MIS process cools the gas stream to 35°C , while the simulated in situ laboratory studies cool the gases and oils to as low as -78°C [18].

In spite of these differences, the subsystem in the MIS process which most closely resembles the laboratory procedures is shown inside the dashed line in Figure 8-1. Table 8-11 presents total mass and Hg balances for this system based on the data in Table 6-33 and Figure 6-16. Although the steam from the thermal sludge unit could contain traces of Hg originating in the retort water, such Hg input would be negligible; it is therefore assumed here that all Hg input to the enclosed system in Figure 8-1 is contained in the raw shale. (The fate of Hg in the retort water is discussed in Section 8.6 in more detail.)

FIGURE 8-14 MIS PROCESS SCHEME

Table 8-11. Mercury Partitioning Adapted for the Modified In Situ Process. Partitioning applies to the system inside the dashed line in Figure 8.1 and does not apply to the entire plant.

IN	Total Mass (10 ³ lb/hr)	Mercury		
		Concentration	Mass g/d	%
Raw Shale	13,720	200 ng/g	29,899	100
Air	3,109	~0.0	~0	0
Steam	848	~0.0	~0	0
Total	17,677		29,899	
<u>OUT</u>				
Spent Shale Left in Retort	10,141	250-9 ng/g	27,866-957	93.2-3.2
Total Oil (a)	784	~100 ng/g	807	2.7
Retort Water	386	~100 ng/g	419	1.4
Condensate Water (b)	740	~100 ng/g	807	2.7
		(10-300)	(90-2,422)	(.3-8.1)
Gas (c)	5,626 (7.83x10 ⁷ m ³ /d)	0-340 µg/m ³	0-26,909	0-90%
Total	17,677		29,899	100

(a) Includes 1.5% H₂O and 30x10³ lb/hr of coke. (See Table 6-33.)

(b) From Figure 6-16.

(c) Dry gas +S+NH₃+H₂O vapor leaving retort with gas-condensate solution.

In Table 8-11 typical concentrations of Hg were assigned to oil, retort water, and condensate based on the data in Tables 8-5 and 8-6. 0-90% of the total Hg was assigned to the retort gas, and the balance to the spent shale. As before, in the worst case Hg in the untreated gas exceeds the permitted emission rate by ten fold and could conceivably present an environmental problem. The volume of gas was calculated for 35°C (95°F) and atmospheric pressure.

Consider first the impact of Hg in the condensate water. To what extent is Hg enriched in the disposal site due to the addition of condensate water? Since about 20% of the total shale is removed and added to the disposal site, and since ~2.7% of the total Hg is in the condensate water, Hg in the disposed shale will be enriched by about $2.7(1/0.25) \approx 10\%$, an insignificant amount.

As in the Paraho process, the likely removal mechanisms for Hg depend on whether it is present as entrained particulate matter or as a gas. Again, considering only the Hg(g), H₂S, H₂ system, the equilibrium concentration for gaseous Hg is given by equation (5). For the gas concentration shown in Table 6-35, $\rho_{\text{Hg}} = 3400 \text{ } \mu\text{g}/\text{m}^3$ at 35°C, well above the expected concentrations shown in Table 8-10. It therefore appears likely that the Hg would enter the Stretford unit as a gas, where it would be removed efficiently and would exit the MIS system with the product S. For an S production rate of $12 \times 10^3 \text{ lb/hr.}$, and assuming all of the gas-borne Hg is trapped by the S, the maximum resulting concentration of Hg in the S would be 200 $\mu\text{g}/\text{g}$.

The MIS system shown in Figure 8-1 includes little capability of removing submicron particles. A large fraction of Hg present in particulate form could therefore be emitted to the atmosphere unless particulate control equipment is included in the retort gas stream.

8.5.3 TOSCO II Process

Partitioning of Hg in the TOSCO II process can best be discussed by reference to Figure 8-2, which illustrates those sections of the plant complex which are most relevant to this discussion.

Table 8-12 gives mass and Hg flows for the subsystem inside the dotted line in Figure 8-2. As in the previous sections, 0-90% of the Hg was assigned to the gas stream. The oil and water streams were assumed to contain

FIGURE 8-2 TOSCO II PROCESS SCHEME

Table 8-12. Mercury Partitioning Adapted for the TOSCO II Process. Partitioning applies to the system inside the dashed line in Figure 8-2.

<u>IN</u>	Total Mass (10^3 lb/hr)	Hg Balance		
		Concentration	Mass (g/d)	%
Oil Shale, dry (a)	5,465	200 ng/g	11,909	100
Steam (b)	172	~0	0	0
Total	5,637		11,909	100
<u>OUT</u>				
Spent Shale, dry (a)	4,514	200-3 ng/g	10,890-167	91.4-1.4
Oil, total (c)	728	~100 ng/g	~793	~6.7%
Condensate water (a)	213	~100 ng/g	~226	~1.9%
Gas, dry (c)	182	0-4,700 $\mu\text{g}/\text{m}^3$	0-10,723	0-90%
	(2.28×10^6 m^3/d)			
Total	5,637		11,909	100

(a) From Table 6-18.

(b) See Figure 6-6 steam in = 390 gpm process steam + 280 gpm
combustion steam - 327 gpm from
preheater = 343 gpm = 172×10^3 lb/hr

(c) Total gas + oil = 910×10^3 lb/hr by difference
Product is apportioned 20%
to gas stream and 80% to unrefined
oil, typical of results in a Fischer
Assay. Gas volume was calculated on
the assumptions 35°C, .73 atm, Mwt = 30.

100 $\mu\text{g/g}$ of Hg, and the variability of these streams was considered because of their minor significance. The remainder of the Hg was then assigned to the spent shale. For the purposes of this table, the total oil includes all oil fractions shown in Figure 8-2. A simplified schematic of the TOSCO II process showing only those sections most relevant to a hypothetical Hg balance can be seen, the Hg in the untreated gas stream contains about four times the permissible daily emissions.

At 35°C (95°F), $\text{H}_2 = 22.5$ Volume %, and $\text{H}_2\text{S} = 1.34$ Volume % (Goodfellow and Atwood, 1974), the vapor phase equilibrium Hg pressure is given by equation (5) as $p = 2,000 \mu\text{g/m}^3$. In comparison, Table 8-12 indicates that the maximum expected Hg concentration in the gas is $4,700 \mu\text{g/m}^3$ in the section at atmospheric pressure, or $\sim 50,000 \mu\text{g/m}^3$ in the high pressure (150 psia) section of the gas treatment system. The tendency, then, would be for at least part of the Hg to form solid HgS , which could either deposit or remain entrained as a dust.

Discussion of the TOSCO II process is complicated by the refinery system which actually uses the retort gas as a refinery feedstock as well as fuel. (In fact, the fraction of gas burned as fuel is variable and could, in theory, be adjusted to minimize Hg emissions.) In the refinery, retort gases are exposed to elemental S (in the Stretford unit), to numerous catalysts and catalyst protectors (such as Co, ZnO , Ni, Fe, and Cr, and molybdate), and to numerous scrubbing solutions. Cooled surfaces also provide sites for condensation and deposition.

Although it is unlikely that a large fraction of Hg in the original retort gas could pass through the numerous treatment processes and escape with the combusted gas, the complexity of the plant facility precludes meaningful prediction of the Hg sinks. Possible sinks include the synthetic crude oil, the S product, spent catalyst or catalyst protectors, and the fuel gas. Table 8-13 summarizes the highest possible rate of Hg removal in each of these sources based on the assumption that the maximum Hg in the retort gas all reports to a single end product.

Mercury contained in gas which is combusted would most likely escape to the atmosphere. Hg in the spent catalysts and catalyst protectors would terminate in the solid waste disposal site, discussed in Section 7 of this report. The maximum Hg concentration in the synthetic crude oil product is

Table 8-13. Highest Possible Removal Rates for Hg in
Possible Links in the TOSCO II Process

<u>Product</u>	<u>Total Mass (10³ lb/hr)</u>	<u>Mass Hg (g/d)</u>	<u>Concentration Hg (μg/g)</u>
Sulfur (a)	16.1	0-10,723	0-60
Synthetic Crude Oil (b)	552	0-10,723	0-1.8
Spent Catalyst		0-10,723	
Fuel Gas	138	0-10,723	0-7

(a) From Table 6-17, footnote r

within the range of petroleum crude oils; its ultimate deposition is therefore generic to oil refinery processes and not unique to shale oil, and is outside the scope of this study. Hg in the S product would be in the stable form of HgS, the naturally occurring form of the ore.

8.5.4 MIS/Lurgi Plant

Although the combined MIS-Lurgi process provides advantages in terms of solid waste and water management, similar advantage regarding Hg emissions are not apparent. The atmospheric Hg emissions from the MIS section of this combined process would be unchanged, and are discussed in Section 8.5.2. The Lurgi process will process an additional 20% of oil shale, and therefore increase the maximum potential atmospheric emissions of Hg 20% over the MIS process alone to a total of $(26,909)(1.20) = 32,290$ g/d.

The Lurgi process resembles the TOSCO II process but does no partial refining. Hence catalyst and catalyst protector are not possible Hg sinks. As in the other processes, S from the Stretford unit is a likely sink for Hg from the Lurgi process.

8.6 TRACE ELEMENTS IN SOLID WASTES FROM THE THERMAL SLUDGE UNIT

Wastes from the thermal sludge units will contain trace elements which must be disposed of in the solid waste disposal site. This section estimates the quantities of trace elements expected from the thermal sludge boilers.

8.6.1 MIS Plant

The thermal sludge units will receive trace elements from two sources, the reject water from the excess mine water treatment and the condensed water from the MIS retorts. The maximum quantities of trace elements from the mine water treatment were calculated in Section 8.3. The data used to estimate the maximum trace element concentrations of in situ retort water are from the TRW/DRI studies [12] and J. P. Fox [18]. Both studies rely upon simulated in situ retort experiments at LLL and LERC to produce typical retort waters. Table 8-14 lists the key trace elements, their range of concentrations observed in retort waters, and the maximum amount which could be received in the thermal sludge unit.

Table 8-14. Trace Elements in the Thermal Sludge Unit for the MIS Plant.
Concentrations of trace elements in retort water are estimated
from studies of simulated in situ retorts.

Constituent	Trace Elements from Retort Water		Trace Elements from Mine Water (c) (g/d)	Total Sludge to Waste	
	Estimated Concentration ($\mu\text{g/l}$)	Maximum Mass (d) (g/d)		Estimated Concentration (e) ($\mu\text{g/g}$)	Maximum Mass (g/d)
Ag	13 a	54	360	6	410
As	230-6,500 a	27,000	360	400	27,000
B	300-6,000 b	25,000	140,000	2,000	165,000
Be	ND a, b	-	74	>1	>74
Cd	1-5 a	21	.74	.3	22
Cr	12-20 b	84	360	6	440
Cu	30-460 a	1,900	1,800	50	3,700
F	25,000 b	170,000	710,000	10,000	810,000
Mn	20-100 b	420	3,500	50	3,900
Hg	10-340 a	1,400	74	20	1,500
Mo	100-470 b	1,900	2,900	70	4,800
Ki	260-1,900 b	4,200	710	70	4,900
Pb	20-220 a	930	7,100	100	8,000
Se	5-1,000 b	4,200	180	60	4,400
U	ND	-	710	10	7,710
Zn	200-6,300 a	26,000	7,100	500	33,000

a Reference - Fox, 1977

b Reference - TRW, 1977

c Table 8-2

d At 772 gpm (4.21×10^6 l/d)

e Total sludge production rate = 7.29×10^4 kg/d.

The calculations employ the following assumptions:

- o The experimental conditions of the simulated in situ retorts provide a good approximation of the trace element leaching that would occur during production. Possible deviations to this assumption include differences in particle size, compaction, and residence time of the water. In addition, retort water is collected at a much higher temperature in the production scale retort than in the simulated retort
- o The highest observed concentrations are used in the calculations in order to represent worst case levels
- o The consumption rate of retort water by the thermal sludge unit is 772 gpm in accordance with Figure 6-14. This assumption is not as straightforward as it may seem, since the production retort recycles water while the simulated in situ retorts typically do not.
- o Total solids reporting to the thermal sludge unit from the retort water was estimated by assuming a TDS of 5.98 g/l. (See Table 6-34.) The thermal sludge unit was also assumed to receive all of the dissolved solids (1,350 g/l) from the treated mine water (6,478 gpm). Total sludge production was therefore 7.29×10^4 kg/d.

8.6.2 MIS/Lurgi Plant

Trace element balances for the thermal sludge unit in the MIS-Lurgi plant were calculated in the same manner as above, and are summarized in Table 8-15. Calculations were performed as follows and are best understood with reference to Figure 6-18.

- o Concentrations of trace elements in retort and mine water are assumed to be identical to the MIS process. Whenever a range of concentrations is given, the highest value was employed.
- o The thermal sludge unit receives 772 gpm from the MIS retort and 90 gpm from the Lurgi retort. A total input of 862 gpm of retort water was therefore assumed.
- o The thermal sludge unit receives all of the concentrate from the treatment of 874 gpm of mine water. It also receives 712 gpm of water consisting mainly of partially treated mine water. For the purposes of Table 8-15, the thermal sludge unit was assumed to receive all of the trace elements in $874 + 712 = 1,586$ gpm of mine water.
- o As for the MIS plant, the total sludge production rate was estimated by assuming a TDS of 5.98 g/l in the retort water and

1.35 g/l in the mine water, and by assuming that all of the solids in the treated mine water report to the thermal sludge unit.

Table 8-15. Trace Element Balances for the Thermal Sludge Boiler in the MIS/LURGI Plant. Concentrations of trace elements in retort water are estimated from studies of simulated in situ retorts.

Constituent	Trace Elements from Retort Water		Trace Elements from Mine Water		Total Sludge to Waste	
	Estimated Concentration ($\mu\text{g/l}$)	Maximum Mass (c) (g/d)	Estimated Concentration (d) ($\mu\text{g/l}$)	Maximum Mass per Day (e) (g/d)	Estimated Concentration (f) ($\mu\text{g/g}$)	Maximum Mass (g/d)
Ag	13 a	61 g	10	86	4	150
As	230-6,500 a	30,000	10	86	700	31,000
B	300-6,000 b	28,000	1,000	8,600	900	37,000
Be	ND a, b	-	2	-	-	-
Cd	1-5 a	23	4	35	1	58
Cr	12-20 b	94	<10	<86	<2	90-180
Cu	30-460 a	2,200	50	432	70	2,600
F	25,000 b	120,000	7,000	61,000	5,000	180,000
Mn	20-100 b	470	100	860	30	11,300
Hg	10-340 a	1,600	2	17	40	1,600
Mo	100-470 b	2,200	80	700	70	2,900
Ni	260-1,000 b	4,700	20	170	100	4,900
Pb	20-220 a	1,000	100	860	50	1,900
Se	5-1,000 b	4,700	<5	43	100	4,700
U	ND	-	20	170	>4	>170
Zn	200-6,300 a	30,000	200	1,700	800	31,000

ND - No data or not detected

a Reference

b Reference

c @ 862 gpm (4.70×10^5 l/d)

d Table 8-1

e @ 1,586 gpm (8.64×10^5 l/d)

f Total sludge production rate = 3.98×10^7 g/d

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9.0 WORKER HEALTH AND SAFETY

9.1 RESULTS AND CONCLUSIONS

Any mining operation must comply with Mining Safety and Health Administration regulations. The cost of such compliance will depend upon the configuration of the mine and the type of mining. Since many safety features are now built into the equipment as a standard, the cost of these safety features cannot be determined as separate items.

In order to determine the impact of MSHA regulations on oil shale production, significant features unique to this type of operation were sought. Requirements common to all types of mining operations are considered to be normal production activities.

The only unique impact that health and safety activities have on oil shale development costs appears in mines employing in situ retorting, which requires the maintenance of proper air quality underground. Regulations have not yet been promulgated for this activity. Special regulations allowing exceptions to existing regulations will be required. Work has recently begun in this area. The expected unique features of MSHA regulation on in situ oil shale production will not impose a significant cost on operations so the exact costs were not determined.

9.2 MSHA COMPLIANCE REQUIREMENTS

All oil shale mining is now regulated by the metals and nonmetals requirements of CFR 303; part 57 is the most important. The existing regulations forbid the use of underground fires. Therefore, during pilot operations it has been necessary to negotiate a special modification agreement before each burn. The following measures are typical details of a modification agreement necessary for in situ burns:

1. Mine ventilation is twice normal.
2. CO, Methane, H₂S, and O₂ are continuously monitored at a number of locations within the mine.
3. Retort chambers are maintained during the burn at below ambient pressure.
4. Compressed air life-support refuges are located throughout the mine.
5. "Instant on" alternate power supplies are provided to back up the utility supply. Backup power is considered necessary in order to comply with items 1-4 above.
6. Retort gases are monitored for oxygen so that the oxygen levels can be maintained below the lower explosive limit.

The current modification agreement approach provides a close working relationship between the developers and MSHA, but there is a need for more routine inspection and approval processes in the near future. With the assistance of MSHA, the developers are now helping to draft new regulations for in situ retorting. Considerable manpower has been committed to this task by MSHA, the developers, and the mining union.

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10.0 ECONOMIC ANALYSIS

10.1 RESULTS OF ECONOMIC EVALUATION

The total annual cost for each oil shale pollution control or group of controls has been determined by calculating the annual operating costs, plus a capital charge based on a year of normal output. The capital charge provides for recovery of capital expenditures and the provision of an acceptable return on investment to the developer after payment of taxes, and is levied uniformly on each barrel of shale oil (or equivalent products) produced. Finally, these annual costs are converted to a cost per barrel of equivalent synthetic crude oil by dividing by the anticipated production in a year of normal operation.

The results are summarized in Table 10-1, while Tables 10-2 through 10-9 provide a breakdown of costs by type of control for each of the eight major cases considered. In the latter tables, the cost category "Miscellaneous Water Management (A)" includes all water-related equipment not separately identified and installed in Year 0 (i.e., one year before production). Depending on the process this may involve:

- Cooling tower makeup
- Oil/Water separation
- Boiler feedwater demineralization
- Foul water stripping
- Thermal sludge unit

Similarly, "Miscellaneous Water Management (B)" includes all equipment that is installed in Year -3, (i.e., 4 years before production). These items are as follows:

- Source water clarification
- Domestic waste treatment
- Equalization basin
- Excess mine water treatment (MIS and MIS/Lurgi process only)

The costs presented in Tables 10-1 through 10-9 have, as far as possible, been calculated on the basis that an oil shale developer might use,

Table 10-1

SUMMARY OF POLLUTION CONTROL COSTS FOR OIL SHALE PLANTS

Plant	Scenario	Total Annual Cost Millions of dollars	Total Cost Dollars per barrel of syncrude equivalent
Paraho	Less Strict	27.8	0.88
	More Strict	31.4	1.00
TOSCO II	Less Strict	25.9	1.65
	More Strict	31.4	2.00
MIS	Less Strict		
	Low mine water	38.0	2.09
	High mine water	46.2	2.54
	More Strict		
	Low mine water	41.0	2.25
	High mine water	54.5	2.99
MIS-Lurgi	Less Strict		
	Low mine water	49.8	1.92
	High mine water	55.3	2.13
	More Strict		
	Low mine water	50.8	1.96
	High mine water	59.6	2.30

Table 10-2
POLLUTION CONTROL COSTS FOR PARAHO PLANT, LESS STRICT SCENARIO

POLLUTION CONTROL COSTS FOR PARADO PLANT, LESS STRICT SCENARIO							TOTAL COST
		CAPITAL COST	ANNUAL COSTS (thousands of dollars)				dollars per barrel equivalent syncrude
			CAPITAL RECOVERY ^(a)	OPERATING COSTS		Total	
				DIRECT ^(b)	INDIRECT		
<u>WATER MANAGEMENT</u>							
(i)	Ammonia Recovery	6,306	1,013	6,122	(4,884) ^(d)	2,251	0.07
(ii)	Organics Removal	13,750	2,210	608	366 ^(c)	3,184	0.10
(iii)	Misc. Water Mgmt. (A)	324	52	242	0 ^(c)	294	0.01
(iv)	Misc. Water Mgmt. (B)	5,855	1,358	37	165 ^(c)	1,560	0.05
	Subtotal	26,235	4,633	7,009	(4,353)	7,289	0.23
<u>AIR POLLUTION CONTROL</u>							
(v)	Oxides of Nitrogen	9,280	1,491	1,350	210 ^(c)	3,051	0.10
(vi)	Sulphur Dioxide	20,180	3,243	4,402	179 ^(e)	7,824	0.25
(vii)	Hydrocarbons	720	116	--	20 ^(c)	136	0.004
(viii)	Particulates	14,270	2,293	1,150	359 ^(c)	3,802	0.12
(ix)	Fugitive dust	1,460	235	313	29 ^(c)	577	0.02
	Subtotal	45,910	7,378	7,215		15,390	0.49
<u>SOLID WASTE MANAGEMENT</u>							
(x)	Pollution Control, Year 0:	2,647					
	Sludge Disposal & Year 12:	2,647	467	4,787 ^(g)	(121) ^(f)	5,133	0.16
	Site Closure						
TOTALS		74,792+	12,478	19,011	(3,677)	27,812	0.88

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

+Capital in Year 0 only.

Table 10-3
POLLUTION CONTROL COSTS FOR PARAHO PLANT, MORE STRICT SCENARIO

		ANNUAL COSTS (thousands of dollars)				Total	TOTAL COST dollars per barrel equivalent syncrude
		CAPITAL COST	CAPITAL RECOVERY ^(a)	OPERATING COSTS			
				DIRECT ^(b)	INDIRECT		
<u>WATER MANAGEMENT</u>							
(i)	Ammonia Recovery	6,306	1,313	6,122	(4,884) ^(d)	2,251	0.07
(ii)	Organics Removal	13,750	2,210	608	366 ^(c)	3,184	0.10
(iii)	Misc. Water Mgmt. (A)	324	52	242	0 ^(c)	294	0.01
(iv)	Misc. Water Mgmt. (B)	5,855	1,358	37	165 ^(c)	1,560	0.05
	Subtotal	26,235	4,633	7,009	(4,353)	7,289	0.23
<u>AIR POLLUTION CONTROL</u>							
(v)	Oxides of Nitrogen	9,280	1,491	1,350	210 ^(c)	3,051	0.10
(vi)	Sulphur Dioxide	20,180	3,243	4,402	179 ^(e)	7,824	0.25
(vii)	Hydrocarbons	720	116	--	20 ^(c)	136	0.004
(viii)	Particulates	14,820	2,382	1,175	374 ^(c)	3,931	0.12
(ix)	Fugitive dust	1,460	235	313	29 ^(c)	577	0.02
	Subtotal	46,460	7,467	7,240	812	15,519	0.49
<u>SOLID WASTE MANAGEMENT</u>							
(x)	Pollution Control, Year 0:	2,647					
	Sludge Disposal & Year 12:	2,647	467	8,262 ^(h)	(187) ^(v)	8,542	0.27
	Site Closure						
TOTALS		75,342+	12,567	22,511	(3,728)	31,350	1.00*

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

*Column items do not add to totals due to rounding.

⁺Capital in Year 0 only.

Table 10-4
POLLUTION CONTROL COSTS FOR TOSCO II PLANT, LESS STRICT SCENARIO

POLLUTION CONTROL COSTS FOR TOSCO II PLANT, LESS STRICT SCENARIO							TOTAL COST
		CAPITAL COST	ANNUAL COSTS (thousands of dollars)				dollars per barrel equivalent syncrude
			CAPITAL RECOVERY ^(a)	OPERATING COSTS		Total	
				DIRECT ^(b)	INDIRECT		
WATER MANAGEMENT							
(i)	Ammonia Recovery	6,040	971	5,116	(4,455) ^(d)	1,632	0.10
(ii)	Organics Removal	11,200	1,800	494	298 ^(c)	2,592	0.17
(iii)	Misc. Water Mgmt. (A) ^(p)	2,889	464	1,689	14 ^(c)	2,167	0.14
(iv)	Misc. Water Mgmt. (B)	6,895	1,599	60	193 ^(c)	1,852	0.12
	Subtotal	27,024	4,834	7,359	(3,950)	8,243	0.53
AIR POLLUTION CONTROL							
(v)	Oxides of Nitrogen	--	--	--	--	--	--
(vi)	Sulphur Dioxide	9,910	1,593	5,919	(272) ^(e)	7,240	0.46
(vii)	Hydrocarbons	17,785	2,858	420	488 ^(c)	3,766	0.24
(viii)	Particulates	8,550	1,374	1,602	179 ^(c)	3,155	0.20
(ix)	Fugitive dust	1,460	235	299	30 ^(c)	564	0.04
	Subtotal	37,705	6,060	8,240	425	14,725	0.94
SOLID WASTE MANAGEMENT							
(x)	Pollution Control, Sludge Disposal & Site Closure	2,454	351	2,569 ^(g)	(33) ^(c)	2,887	0.18
		=====	=====	=====	=====	=====	=====
TOTALS		67,183	11,245	18,168	(3,558)	25,855	1.65

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

Table 10-5
POLLUTION CONTROL COSTS FOR TOSCO II PLANT, MORE STRICT SCENARIO

		ANNUAL COSTS (thousands of dollars)				TOTAL COST	
		CAPITAL COST	CAPITAL RECOVERY ^(a)	OPERATING COSTS		dollars per barrel	
				DIRECT ^(b)	INDIRECT	equivalent	
						syncrude	
<u>WATER MANAGEMENT</u>							
(i)	Ammonia Recovery	6,040	971	5,116	(4,455) ^(d)	1,632	0.10
(ii)	Organics Removal	11,200	1,800	494	298 ^(c)	2,592	0.17
(iii)	Misc. Water Mgmt. (A)	2,889	464	1,689	14 ^(c)	2,167	0.14
(iv)	Misc. Water Mgmt. (B) ^(q)	6,869	1,593	59	193 ^(c)	1,845	0.12
	Subtotal	26,998	4,828	7,358	(3,950)	8,236	0.53
<u>AIR POLLUTION CONTROL</u>							
(v)	Oxides of Nitrogen	--	--	--	--	--	--
(vi)	Sulphur Dioxide	9,910	1,593	5,919	(272) ^(e)	7,240	0.46
(vii)	Hydrocarbons	17,785	2,858	420	438 ^(c)	3,766	0.24
(viii)	Particulates	29,400	4,725	978	796 ^(c)	6,499	0.41
(ix)	Fugitive dust	1,460	235	299	30 ^(c)	564	0.04
	Subtotal	58,555	9,411	7,616	1,042	18,069	1.15
<u>SOLID WASTE MANAGEMENT</u>							
(x)	Pollution Control, Sludge Disposal & Site Closure	2,454	351	4,818 ^(h)	(82) ^(w)	5,087	0.32
TOTALS		88,007	14,590	19,792	(2,990)	31,392	2.00

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

Table 10-6
POLLUTION CONTROL COSTS FOR M15 PLANT, LESS STRICT SCENARIO

POLLUTION CONTROL COSTS FOR HIS PLANT, LESS STRICT SCENARIO							TOTAL COST
		CAPITAL COST	ANNUAL COSTS (thousands of dollars)			Total	dollars per barrel equivalent syncrude
			CAPITAL RECOVERY ^(a)	OPERATING COSTS			
				DIRECT ^(b)	INDIRECT		
WATER MANAGEMENT							
(i)	Ammonia Recovery	8,600	1,542	10,445	(9,339) ^(k)	2,648	0.15
(ii)	Organics Removal	31,200	5,594	1,400	857 ⁽ⁱ⁾	7,851	0.43
(iii)	Misc. Water Mgmt. (A)	3,557	638	2,632	23 ⁽ⁱ⁾	3,293	0.18
(iv)	Misc. Water Mgmt. (B)						
	Low mine water ^(s)	9,406	2,433	644	252 ⁽ⁱ⁾	3,329	0.18
	[High mine water ^(r)	28,178	7,290	3,476	707 ⁽ⁱ⁾	11,473	0.63]
	Subtotal (Low mine water)	52,763	10,207	15,121	(8,207)	17,121	0.94
AIR POLLUTION CONTROL							
(v)	Oxides of Nitrogen	12,866	2,307	1,247	333 ⁽ⁱ⁾	3,887	0.21
(vi)	Sulphur Dioxide	26,210	4,699	6,155	333 ⁽ⁱ⁾	11,187	0.61
(vii)	Hydrocarbons	240	43	--	7 ⁽ⁱ⁾	50	0.003
(viii)	Particulates	5,792	1,039	334	157 ⁽ⁱ⁾	1,530	0.08
(ix)	Fugitive dust	1,460	262	281	34 ⁽ⁱ⁾	577	0.03
	Subtotal	46,568	8,350	8,017	864	17,231	0.94*
SOLID WASTE MANAGEMENT							
(x)	Pollution Control,	Year 0: 2,012					
	Sludge Disposal &	Year 12: 1,152	364	3,375 ^(g)	(45) ^(j)	3,694	0.20
	Site Closure						
TOTALS (Low mine water)		101,343 ⁺	18,921	26,513	(7,388)	38,046	2.09*
				[High Mine Water		46,190	2.54*]

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

*Column items do not add to total or subtotal due to rounding.

⁺Capital in Year 0 only.

Table 10-7
POLLUTION CONTROL COSTS FOR MIS PLANT, MORE STRICT SCENARIO

ANNUAL COSTS (thousands of dollars)							TOTAL COST
CAPITAL COST	CAPITAL RECOVERY ^(a)	OPERATING COSTS		Total	dollars per barrel equivalent syncrude		
		DIRECT ^(b)	INDIRECT				
<u>WATER MANAGEMENT</u>							
(i) Ammonia Recovery	8,600	1,542	10,445	(9,339) ^(k)	2,648	0.15	
(ii) Organics Removal	31,200	5,594	1,400	657 ⁽ⁱ⁾	7,851	0.43	
(iii) Misc. Water Mgmt. (A)	3,557	638	2,632	23 ⁽ⁱ⁾	3,293	0.18	
(iv) Misc. Water Mgmt. (B)							
High mine water ^(t)	43,055	11,138	5,710	1,096 ⁽ⁱ⁾	17,917	0.98	
[Low mine water ^(u)	12,031	3,112	1,039	315 ⁽ⁱ⁾	4,466	0.25]	
Subtotal (High mine water)	86,412	18,912	20,187	(7,390)	31,709	1.74	
<u>AIR POLLUTION CONTROL</u>							
(v) Oxides of Nitrogen	12,865	2,307	1,247	333 ⁽ⁱ⁾	3,882	0.21	
(vi) Sulphur Dioxide	26,210	4,699	6,155	333 ⁽ⁱ⁾	11,187	0.61	
(vii) Hydrocarbons	240	43	--	7 ⁽ⁱ⁾	50	0.003	
(viii) Particulates	5,792	1,039	334	157 ⁽ⁱ⁾	1,530	0.08	
(ix) Fugitive dust	1,460	262	281	34 ⁽ⁱ⁾	577	0.03	
Subtotal	46,568	8,350	8,017	864	17,231	0.94*	
<u>SOLID WASTE MANAGEMENT</u>							
(x) Pollution Control, Year 0:	2,012						
Sludge Disposal & Year 12:	1,152	364	5,254 ^(h)	(73) ^(x)	5,545	0.30	
Site Closure							
<u>TOTALS (High mine water)</u>							
	134,992+	27,262	33,458	(6,599)	54,485	2.99*	
			[Low Mine Water		40,034	2.25*]	

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

* Column items do not add to total and subtotal due to rounding.

+ Capital in Year 0 only.

Table 10-8
POLLUTION CONTROL COSTS FOR MIS-LURGI PLANT, LESS STRICT SCENARIO

POLLUTION CONTROL COSTS FOR MIS-LURGI PLANT, LESS STRICT SCENARIO						TOTAL COST
	CAPITAL COST	ANNUAL COSTS (thousands of dollars)			Total	dollars per barrel equivalent syncrude
		CAPITAL RECOVERY ^(a)	OPERATING COSTS			
			DIRECT ^(b)	INDIRECT		
WATER MANAGEMENT						
(i) Ammonia Recovery	8,600	1,542	10,445	(9,371) ⁽ⁿ⁾	2,616	0.10
(ii) Organics Removal	31,200	5,594	1,400	849 ^(m)	7,843	0.30
(iii) Misc. Water Mgmt. (A)	3,562	639	2,660	14 ^(m)	3,313	0.13
(iv) Misc. Water Mgmt. ^(B)						
Low mine water ^(u)	4,417	1,143	27	126 ^(m)	1,296	0.05
[High mine water ^(t)	<u>17,534</u>	<u>4,536</u>	<u>1,789</u>	<u>444^(m)</u>	<u>6,769</u>	<u>0.26]</u>
Subtotal (Low mine water)	47,779	8,918	14,532	(8,382)	15,068	0.58
AIR POLLUTION CONTROL						
(v) Oxides of Nitrogen	12,866	2,307	1,247	328 ^(m)	3,882	0.15
(vi) Sulphur Dioxide	29,800	5,343	7,167	334 ^(o)	12,844	0.50
(vii) Hydrocarbons	878	157	--	25 ^(m)	182	0.01
(viii) Particulates	34,340	6,157	1,404	938 ^(m)	8,499	0.33
(ix) Fugitive dust	<u>1,460</u>	<u>262</u>	<u>247</u>	<u>34^(m)</u>	<u>543</u>	<u>0.02</u>
Subtotal	79,344	14,226	10,065	1,659	25,950	1.00*
SOLID WASTE MANAGEMENT						
(x) Pollution Control, Sludge Disposal & Site Closure	1,005	161	920 ^(g)	(2) ^(m)	1,079	0.04
(xi) Slurry Backfilling	<u>3,000</u>	<u>538</u>	<u>7,350</u>	<u>(157)^(m)</u>	<u>7,731</u>	<u>0.30</u>
Subtotal	4,005	699	8,270	(159)	8,810	0.34
<hr/>						
TOTALS (Low mine water)	131,128	23,843	32,867	(6,882)	49,828	1.92*
			[High Mine Water		55,301	2.13*]

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

* Column items do not add to total or subtotal due to rounding.

Table 10-9
POLLUTION CONTROL COSTS FOR MIS-LURGI PLANT, MORE STRICT SCENARIO

	CAPITAL COST	ANNUAL COSTS (thousands of dollars)				Total	TOTAL COST dollars per barrel equivalent syncrude
		CAPITAL RECOVERY ^(a)	OPERATING COSTS				
			DIRECT ^(b)	INDIRECT			
<u>WATER MANAGEMENT</u>							
(i) Ammonia Recovery	8,600	1,542	10,445	(9,371) ⁽ⁿ⁾	2,616	0.10	
(ii) Organics Removal	31,200	5,594	1,400	849 ^(m)	7,843	0.30	
(iii) Misc. Water Mgmt. (A)	3,562	639	2,660	14 ^(m)	3,313	0.13	
(iv) Misc. Water Mgmt. (B)							
High mine water ^(t)	25,183	6,515	2,937	625 ^(m)	10,077	0.39	
[Low mine water ^(u)	4,417	1,143	27	126 ^(m)	1,296	0.05]	
Subtotal (High mine water)	68,545	14,290	17,442	(7,883)	23,849	0.92*	
<u>AIR POLLUTION CONTROL</u>							
(v) Oxides of Nitrogen	12,866	2,307	1,247	328 ^(m)	3,832	0.15	
(vi) Sulphur Dioxide	29,800	5,343	7,167	334 ^(o)	12,844	0.50	
(vii) Hydrocarbons	878	157	--	25 ^(m)	182	0.01	
(viii) Particulates	34,340	6,157	1,404	938 ^(m)	8,499	0.33	
(ix) Fugitive dust	1,460	262	247	34 ^(m)	543	0.02	
Subtotal	79,344	14,226	10,065	1,659	25,950	1.00	
<u>SOLID WASTE MANAGEMENT</u>							
(x) Pollution Control, Sludge Disposal & Site Closure	1,118	179	1,873 ^(h)	(21) ^(y)	2,031	0.08	
(xi) Slurry Backfilling	3,000	538	7,350	(157) ^(m)	7,731	0.30	
Subtotal	4,118	717	9,223	(178)	9,762	0.38	
TOTALS (High mine water)	152,007	29,233	36,730	(6,402)	59,561	2.30*	
			[Low Mine Water		50,780	1.96*]	

Note: Lettered footnotes are explained in Appendix 10.0, Section B.

* Column items do not add to total and subtotal due to rounding.

that is, they provide for an acceptable return on investment. This return was chosen as the equivalent of a 13% internal rate of return on a discounted cash flow analysis.

It should be pointed out, however, that there is another way of viewing these results. If a developer did not incur the pollution control costs, he would make additional profit. This profit would be taxed at the developer's combined (state and federal) marginal income tax rate of 50.6%,* and hence it can be argued that the effective cost to a developer is approximately half of the cost calculated in this chapter.

The costs presented in this chapter do not include an allowance for the cost of working capital. Since the addition of pollution controls does not affect the quantity of work in progress, the most significant components of working capital would be maintenance supplies and chemicals, and inventories and accounts receivable associated with byproducts. None of these items would be particularly large and hence the cost of financing the working capital associated with pollution controls has been neglected.†

In Section 10.2, the basis of the treatment of capital costs is explained, and the effect on the resulting capital recovery factor of various different assumptions is explored. Section 10.3 discusses operating costs, while Section 10.4 explains the special treatment of solid waste management costs.

Section 10.5 discusses the sensitivity of the overall results to the various economic assumptions. It is shown that the most critical assumptions are those concerned with capital charges, the capital recovery factor being particularly sensitive to assumptions about the timing of capital expenditures, and to the discount rate used to calculate capital charges. However, if the total pollution control cost is recalculated using conservative as

*This argument disregards a small change in severance tax levied.

†For example if the working capital was 60 days' operating costs, and this were required to return 13% per year on equity (this cannot be expensed for tax purposes, approximately doubling the effective cost--see Section 10.4.3) the cost of working capital would be about 2 to 3% of the cost of pollution control. However, if this capital were a loan at 10% interest (which can be expensed for tax purposes), the cost of working capital would drop to about 1% of the total cost of pollution control.

opposed to "middle of the road" assumptions with respect to these two factors, the cost increases by 23 to 30%.

Appendix 10.0, Section A illustrates the calculation of a capital recovery factor. Appendix 10.0, Section B provides specific details of the method of calculation used for some items in Tables 10-2 through 10-9.

10.2 TREATMENT OF CAPITAL COSTS

10.2.1 Introduction

In conducting the economic analysis, the overall objective was to simulate industry's evaluation procedures and associated assumptions as closely as possible. Industry normally evaluates capital investment projects by the discounted cash flow (DCF) method, and requires that the DCF rate of return (ROR) be greater than some specified value for the project to constitute an acceptable investment.*

In practice, industry is unlikely to make an economic analysis of the separate cost of pollution controls for two major reasons: first, that pollution controls are not "optional" investments that require economic justification for installation; and second, that in some cases it is very difficult to separate equipment required for pollution control from that required for the process itself. Nevertheless, if industry were to analyze the cost of pollution control separately from the total investment in a project, it would almost undoubtedly use a method that provided an acceptable return on the capital invested in the control facilities.

10.2.2 Use of Capital Recovery Factors

Because industry generally makes its economic evaluations using the DCF method, simple amortization or the equivalent is not an adequate way of treating the capital component of total cost. The most accurate approach

*Other criteria may be used in addition to the minimum acceptable DCF ROR, or to choose between alternative projects. However, provision of an acceptable DCF ROR is almost invariably the preeminent economic criterion for the evaluation of major capital investment projects, such as mines and chemical plants.

would be to use a DCF analysis in which cash flows with and without the pollution controls were developed. Each cash flow would determine the selling price that provided the required DCF ROR. The cost attributable to pollution controls would be the difference between these two prices. This approach would accurately account for the timing of pollution control expenditures and for the effects of taxation on costs.*

There are two reasons why the approach outlined above was not used in this study. In the first place, accurate data on the total costs of producing shale oil were not available. Secondly, the huge effort involved in developing even approximate incremental cash flow calculations, with their attendant depreciation and taxation schedules, for all the alternatives under consideration would have been far beyond the time and resources available for this study.†

The concept of a "capital recovery factor" is useful for situations such as this where large numbers of alternatives need to be evaluated. A capital recovery factor can be calculated by simulating a cash flow evaluation on the basis of a standard set of assumptions concerned with the phasing of project income and capital expenditures, taxation details and required DCF ROR. The basis of the calculation is to determine the necessary income (in addition to operating costs) that would provide the specified DCF ROR on capital, after payment of income taxes.

The capital recovery factor, r , is defined as the proportion of total capital expenditure that must be added to operating costs in each year of normal output in order to provide the specified DCF return on investment. Since income during start-up years is prorated on the basis of output, the capital charge may readily be attributed to each unit of production, e.g., per barrel of shale oil. The calculation of a typical capital recovery factor is explained in Appendix 10.0, Section A.

*Note that an incremental approach, in which calculations are based on the difference between the two cash flows, would not be accurate since calculation of the depletion allowance in order to determine tax liability requires a full cash flow (see Section 10.2.3).

†In this study 10 different pollution controls for two scenarios for each of four processes were evaluated, making 80 basic combinations.

In this study, the capital recovery factor has, of necessity, been applied only to the cost of the pollution controls, i.e., the analysis is an incremental analysis. Note, however, that the use of a capital recovery factor is just as accurate as a full cash flow evaluation, provided that both are based on identical assumptions.*

Table 10-10 lists capital recovery factors, based on a 25-year project life and a 13% DCF ROR, for a number of cash flow profiles that are relevant to oil shale pollution control. Table 10-11 provides an analysis of the sensitivity of the value of "r" to different assumptions about the required DCF ROR, taxation details, project life and capital expenditure profile.

10.2.3 Assumptions that Influence Capital Charges

This section discusses the rationale behind the choice of assumptions used in calculating capital recovery factors, and indicates some of the changes that would occur if alternative assumptions were made.

Required DCF Rate of Return: Table 10-11 shows that the value of "r" is highly sensitive to the required DCF ROR and hence this choice is critical. For example, increasing the required ROR from 13 to 15%, increases the capital recovery factor for the case considered from 16.07% to 19.13%.

The minimum acceptable DCF ROR is normally not divulged by companies, and in any event is influenced by alternative investment opportunities and other factors. However, there is broad confirmation that a rate between 12 and 15% is appropriate for evaluating oil shale investments[1]. In addition, recent published evaluations of oil shale economics have used 12% [2]; 10, 13 and 15% [3]; and 15% [4,5]. One source has referred to 12 or 13% as "modest" [6], while another implies that 15% is attractive [7].

*In practice, a full DCF evaluation will usually introduce additional data, and adjust the assumptions to match the details of each situation being considered. "Fine tuning" of this type is not feasible if a capital recovery factor is used; however, as is the case in this study, more than one factor can be employed to account for major differences in assumptions between the alternatives.

Table 10-10
CAPITAL RECOVERY FACTORS (PERCENT)

<u>Control Category</u>	<u>Year(s) in which investment is made</u>	<u>Capital Recovery Factor (r)</u>	
		<u>TOSCO II and Paraho Plants</u>	<u>MIS and MIS + Lurgi Plants</u>
Air and most water	0	16.07%	17.93%
Misc. Water Management(B)	-3	23.19	25.87
Solid waste	0	14.32	15.98
Solid waste (double investment)†	0,12	17.63	19.67

Assumptions:

- (i) 13% DCF ROR
- (ii) 100% equity
- (iii) 20% investment credit
- (iv) 50.6% effective income tax rate
- (v) 13 year DDB* depreciation for air and water, 8 year DDB depreciation for solid waste
- (vi) 25 year project life
- (vii) Start-up profiles are as follows:

<u>Year</u>	<u>TOSCO II and Paraho Plants</u>	<u>MIS and MIS + Lurgi Plants</u>
1	0.56	0.25
2	0.83	0.50
3	1.00	0.75
4		1.00

Each factor represents the proportion of a normal year's output (1.00) produced in the specified year.

*DDB = Double declining balance.

†In the solid waste management plans for the Paraho and MIS processes, capital investment is required in Year 12 as well as Year 0, as two disposal sites are necessary.

Table 10-11
CAPITAL RECOVERY FACTOR SENSITIVITY ANALYSIS

The following "base case" (for air and most water pollution controls, investment in Year 0, TOSCO II or Paraho Plant) yields a capital recovery factor of 16.07%.

Base case assumptions:

- (i) 13% DCF ROR
- (ii) 100% equity
- (iii) 20% investment credit
- (iv) 50.6% effective income tax rate
- (v) 13 year DDB depreciation
- (vi) 25 year plant life
- (vii) Start-up profile is as follows:

Year 1	0.56
Year 2	0.83
Year 3	1.00

The following capital recovery factors were calculated when certain of the base case assumptions were changed. In every case only the item listed below was changed, all other assumptions remaining constant.

	Capital Recovery Factor (Percent)
(a) 12% DCF ROR	14.62
(b) 15% DCF ROR	19.13
(c) 20-year plant life	16.82
(d) 30-year plant life	15.69
(e) 10% investment credit	18.70
(f) Investment in Year -3, credit/depreciation in Year -2	23.19
(g) Investment in Year -3, credit/depreciation in Year -1	27.46
(h) Typical complete plant investment profile of 15% in Year -3, 20% in Year -2, 40% in Year -1, and 25% in Year 0, credit/ depreciation in Year 1	21.24

The figure of 13% was chosen to lie towards the low side of the 12 to 15% range mentioned above. There are three reasons for selecting a required DCF ROR that is less than the upper rate of 15%. First, one function of the required rate of return is to cover sunk costs, such as exploration and lease acquisition costs incurred prior to project evaluation. Pollution controls should not significantly affect these costs, and hence this portion of the required ROR is not applicable.

Secondly, companies normally vary their ROR criteria to reflect a perceived degree of risk, raising the minimum acceptable expected return for risky projects.* However, the risk associated with oil shale pollution controls is probably low compared to the overall technical and economic risks involved in a complete oil shale project, and hence it can be argued pollution controls should not be costed using a "risky" DCF ROR.†

Finally, it should be borne in mind that costs developed in this report are not intended to apply to the first commercial plants, but are expected to be representative of a more mature oil shale industry, in which many of the risks (especially technological risks) should have been reduced below their present levels. Most of the DCF ROR criteria quoted earlier were applied to evaluations of plans for prototype commercial plants, and hence it is possible that in the future acceptable DCF ROR's will move towards the lower end of the 12 to 15% range.

All evaluations were made on the basis of 100% equity. This is a common practice in industry evaluations, and is consistent with the DCF ROR values discussed above.

*Variations in the level of risk and sunk costs account for variations in the required DCF ROR between different commodities. For example, a uranium investment project would probably require a significantly higher ROR than 15%, which is the highest normally suggested for oil shale.

†A countervailing argument is that the entire project, including pollution controls must provide an acceptable return on investment, and that reducing the DCF ROR on pollution controls would require an increase of the ROR on the remaining investment. However, the effect of adding low-risk pollution controls would be to reduce the overall riskiness of the project (at least for some types of risk) thereby justifying a slightly lower required DCF ROR.

Effective Income Tax Rate: Colorado income tax is levied at 5%, and is deductible for the purpose of calculating federal income tax. Hence the effective overall income tax rate is:

$$5 + (1 - 0.05)48 = 50.6\%$$

The Investment Credit: Under the Energy Tax Act of 1978 (PL 95-618), an additional 10% investment credit (i.e., 20% in total) is allowable for equipment used producing or extracting shale oil from oil shale. It is not allowable on equipment used for hydrogenation or refining, but water management facilities are specifically included.

The 20% investment credit has been used in calculating all capital recovery factors.* In the case of the TOSCO II process there are pollution controls for hydrogen sulphide and hydrocarbons on the upgrading facility. To a large extent these controls could be attributed to the basic retorting process, as some control would be needed even if the upgrading facility were not built. Nevertheless, if a conservative viewpoint is adopted and only 10% investment credit is allowed ($r = 18.70\%$; see Table 10-11) on the \$10.5 million capital involved in this equipment, the pollution control cost would be increased by about 1.9 cents per barrel, a very small discrepancy.

The Depletion Allowance: The depletion allowance has not been incorporated into the calculation of taxes. The justification for this is as follows. The percentage depletion allowance is 15% on the "gross income" from an oil shale property. (Gross income would probably be calculated on the value of the shale oil before any upgrading, by analogy with tax decisions on other minerals.) In this case, since the sales or transfer price of shale oil (and hence, gross income) is independent of pollution control costs, the depletion allowance will not affect those costs. However, there is a limitation that the percentage depletion allowance cannot exceed 50% of the taxpayer's taxable income from the property, computed without allowance for depletion. Since pollution control costs reduce the taxable income they could affect the

*This implies that the Energy Tax Act will be extended to be in force when commercial plants are constructed.

depletion allowance if it was limited under the above rule, and this would then be a cost attributable to pollution control. While this might well be the case in a start-up year, it appears that this limit is most unlikely to apply during a normal year's operation since capital-related costs are a high proportion of total costs (based on data for the TOSCO II process [3]). Hence, for the purpose of developing capital recovery factors the effect of the federal percentage depletion allowance on pollution control costs has been disregarded. This may introduce minor errors during start-up years, but complete project cost data are not available to permit the effect to be calculated (see Section 10.2.2).

Depreciation: The lower limit of the IRS' Asset Depreciation Range (Section 167 of the Internal Revenue Code) for assets used in mining is 8 years, while that for petroleum refining is 13 years. Oil shale surface retorting equipment is largely analogous to oil refining equipment, and hence 13 years was selected as the depreciation life for air and water pollution control equipment. It was also selected for the equipment used in the slurry backfilling in the MIS/Lurgi plant as this equipment is similar to surface retorting equipment. On the other hand, solid waste management is very largely related to mining, and an 8 year life was assumed for all major equipment used for solid waste management. (Also see Section 10.4.1) Double declining balance (DDB) with switch-over to straight line was used for all depreciation calculations.

Under Section 169 of the Internal Revenue Code, certain qualifying equipment such as pollution control equipment may be amortized over 5 years. However, only straight line amortization may be used and the investment credit is forfeited. The effect of these restrictions is to make this provision for special amortization less attractive than regular depreciation for oil shale projects [9], and hence the provision was not incorporated into the calculations.

Project Timing Factors

Project Life: The expected project life (measured from the commencement of production) will be determined by exhaustion of the oil shale reserves, or

technological obsolescence.* Sources suggest that planned project lives used for evaluations of oil shale developments range from 18 to 30 years, with 20 and 25 years as popular nominal figures [8]. Twenty-five years was chosen for this evaluation, but as Table 10-11 shows, the capital recovery factor is comparatively insensitive to this assumption, decreasing by only 1.13% as the life is changed from 20 to 30 years.

Start-up Profile: Different start-up profiles were used for the plants using surface retorting, and those that involve in situ retorting. These assumptions are given in Table 10-10. The profile used for the surface retorting operations (TOSCO II and Paraho) was based on a published schedule for the TOSCO II process [3], and appears to be in line, or conservative with respect to industry expectations [10]. That for the MIS and MIS/Lurgi plants was selected as a compromise between the expectations of two oil shale developers [10]. The slower build-up of output in comparison with surface retorting reflects the sequential development and firing of a series of in situ retorts.

Note that operation for 330 days per year implies a normal production capacity factor of 90%. Again, this is reasonably in line with industry expectations: Paraho considers that this factor would be appropriate for their process [11]. TOSCO II economics have been evaluated on this basis [3], and an in situ developer has used the same factor for evaluations [12]. However, another developer is assuming 83% of the theoretical 365 days per year capacity [12].

Expenditure of Capital: The majority of oil shale development plans have envisaged $3\frac{1}{2}$ to 4 years of construction before the commencement of retorting, although this could be extended by the need to sink shafts [12]. However, the nature of all the air pollution control equipment, the solid waste management facilities, and part of the water management equipment is such that it would be installed in the year preceeding production (i.e., in Year 0). Some of the water management equipment would be required at or near to the beginning of

*In this report, the first year of production is designated Year 1. Note that with this nomenclature, investment occurs in negative years, i.e., Year -3 through Year 0.

construction,* i.e., in Year -3 [13]. In calculating capital recovery factors for the latter equipment, it has been assumed that the investment credit and depreciation can be taken in the year following installation (Year -2), since the assets will be actively placed in service in that year. If, however, it was necessary to wait until Year 1 before claiming these allowances, Table 10-11 shows that the capital recovery factor would be raised by about 4½ percentage points.

A typical time profile for investment in a complete oil shale development might be as follows: Year -3, 15%; Year -2, 20%; Year -1, 40%; Year 0, 25% [14]. For aboveground retorting this leads to a capital recovery factor of 21.24% (if the investment credit and depreciation is claimed in Year 1) compared to 16.07% for investment in Year 0, but 23.19% for investment in Year -3 (with investment credit and depreciation claimed in Year -2. (See Table 10-11.) This shows that the assumption that most of the pollution control equipment is installed in the year prior to production results in a significant cost reduction with respect to the alternative assumption that the equipment is installed with the same time profile as the complete plant.

It may be tempting to simply assume that all capital expenditures follow the time profile for a complete oil shale development. However, the authors' consider that this is not as accurate as the adopted assumption of some water-related capital expenditures in Year -3 and air, solid waste, and the remaining water-related expenditures in Year 0. As noted in Section 10.2.2, pollution control costs would ideally be determined by evaluating two cash flows, one with and one without the pollution controls. This would focus attention on the most appropriate timing for the additional expenditures, allowing for any constraints on the sequence of the installations, or the total volume of site-work in any period. While the authors' assumptions may be oversimple, they should be closer to engineering reality than the more arbitrary allocation of pollution control expenditures to the same phasing as the much larger total mining and retorting investment.

*i.e., Source water clarification, domestic waste treatment, equilization basin and excess mine water treatment. These processes have been grouped together under the category "Miscellaneous Water Management (B)."

10.3 ANNUAL OPERATING COSTS

The term "operating cost" excludes all capital-related items and considerations of income tax. Operating costs have been divided into two subcategories: direct and indirect costs.

The category of direct operating costs includes the following:

- o labor
- o maintenance
- o chemicals, etc.
- o utilities

Indirect operating costs includes:

- o byproduct credits
- o property tax and insurance on equipment
- o severance tax credit

Direct operating costs for air and water controls are developed in Sections 5 and 6. These sections also specify the byproduct credits. Other components of indirect operating costs are calculated separately. Costs for solid waste management require special treatment and are discussed in Section 10.4.

10.3.1 Direct Operating Costs

The bases of the cost items are outlined below.

Labor: All categories of labor were costed at \$20.00 per hour, which can be taken as including a payroll overhead. As well as fringe benefits, this covers general and administrative costs.

Maintenance: Maintenance costs include both labor and materials.

Utilities: In making an overall assessment of the economics of an oil shale project, self-supplied utilities (such as electrical power and steam) would not be charged as project costs and would be accounted for by reducing the net energy output. However, it is necessary to include charges for these items when calculating pollution control costs. Both electrical power and process steam are costed at \$3.00 per million Btu.

Cooling water for pollution controls is costed at 12¢ per 1,000 gallons circulated, which is essentially a charge for the use of the cooling

tower. This excludes the cost of treating the makeup water which is costed separately.

10.3.2 Indirect Operating Costs

Byproduct Credits: Credits for the value of byproducts from the pollution control were allowed at the following rates:

Ammonia	\$100 per ton
Sulphur	\$ 5 per ton

The quantities produced are given in Table 2-1.

Property Tax and Insurance: In making economic evaluations, industry practice varies between allowing 1½% [10] to over 4% [2] of the capital invested for the annual cost of plant insurance and local property taxes. Three percent of the capital cost of the pollution control equipment was chosen as a representative value.

For 1979, a credit of 30% of the property tax paid (due in 1980) can be deducted from the following year's (i.e., 1981) Colorado income tax. This has not been taken into account in the calculations, as opinion is divided about the likelihood of this credit being extended to other years [15,16].

Severance Tax Credit: Under Colorado HB 1076, enacted in 1977, severance tax is levied on the production of a commercial oil shale facility at the rate of 4% of the "gross proceeds" for surface retorted oil, and 3% for shale oil produced by in situ methods. "Gross proceeds" is defined as the value of the oil shale at the point of severance, and is calculated by subtracting costs (e.g., retorting and mining) from the gross sales income. Hence, since pollution controls add to costs, they reduce the gross proceeds by a corresponding amount. Hence a 4% (3% for in situ) credit for severance tax not paid should be deducted from the pollution control costs.

However, while operating costs are clearly allowable in calculating gross proceeds, return on capital does not appear to be (the statute refers to allowing "costs, including direct and indirect expenditures for: (a) equipment and machinery..."). Hence, when this credit is calculated, the capital charge must be replaced by some form of depreciation. The severance tax credit calculations are based on direct operating costs, plus 4% of the capital

invested (except for solid waste with investments in Year 0 and Year 12, where 8% of the Year 0 capital investment was used). These figures allow for uniform capital amortization over the project lives.*

For the MIS/Lurgi plant, the severance tax credit was prorated between the 3% in situ and 4% surface retorting rates on the basis of heat output. This lead to a combined rate of 3.30%.

It could certainly be claimed that some or all of the other items included in the indirect costs should be included in the cost base for the calculation of the severance tax credit, or that the capital amortization could be treated some other way. However, since this credit cannot exceed 4% of the total cost per barrel, and is typically 3% or less, this is not an important issue.

10.4 COSTS FOR SOLID WASTE MANAGEMENT

10.4.1 Introduction

The data for solid waste management were developed in the form of a cash flow from Year 0 to beyond Year 25. (See Tables 7-12 through 7-15.) Even after the initial 2 or 3 years, the series is not uniform, especially in the cases (Paraho and MIS plants) where two disposal sites must be used. Closure costs extend the series beyond Year 25.

All costs incurred in Year 0 were taken as capital costs. Where two sites are used, any cost in Year 12 that corresponded to a cost in Year 0 was also taken as a capital cost. All other costs were treated as operating costs. This represents a simplification, since there are a few items (such as fencing and truck purchase) that should probably not be expensed in the year of purchase. However, these items are only a very small proportion of the total costs. A more serious question might arise over some of the abnormally high costs incurred in Year 1 (and Year 13 for the two site cases). However, since these costs are always followed by a sequence of several years' costs relating to the same item, expensing appears to be justified. For example,

*There is also an exemption to the tax of the first 10,000 barrels per day of shale oil (or 15,000 tons per day of oil shale, whichever is greater), but this was disregarded as it would be small for a mature industry.

there are extra costs for spent shale disposal and lining in Year 1, due to the need to establish a dam and line part of the disposal area, but in subsequent years similar activities continue, albeit at a lower level of expenditure. Thus these costs can be reasonably regarded as continuing expenses, rather than capital investment.

The capital portion of the solid waste management costs was treated in the same way as other pollution controls, using the capital recovery factors given in Table 10-10. The method of dealing with operating costs is outlined below, while that for the trusts is explained in Section 10.4.3.

10.4.2 Direct Operating Costs for Solid Waste Management

Because of the irregular nature of the solid waste management cash flow (after subtraction of capital items) the approach used was to discount the entire series at 13% (i.e., the time value of money used elsewhere in this evaluation) and to equate this to a uniform series of expenditures, also discounted at 13%. This provides an equivalent annual cost (or "levelized cost") which is somewhat analogous to a capital charge. A uniform series (rather than one with a start-up profile) was used because the costs developed in Chapter 7 reflect operation at the normal level of production from Year 1 onwards, i.e., they do not reflect the reduced production in Years 1 and 2, and Year 3 for the in situ cases.

Table 10-12 gives the annual "levelized cost" for each plant and scenario, and compares this with actual anticipated cost in Year 4, which is the first year in which all extra early year expenses and start-up effects have been eliminated. The "levelized cost" is some 5-20% higher than the cost in Year 4 largely because of the higher expenses (relative to the quantity handled) in the first two years of operation.

10.4.3 Treatment of Trust Contributions

The more strict scenarios involve the creation of two trusts associated with solid waste management. One is an annual payment in Years 1 through 25, while the other is a lump sum paid in Year 0 (and in Year 12 where two disposal sites are required).

The lump sum trust payment was converted to an equivalent annual cost, using a similar technique to that described above. However, in this

Table 10-12

PARTIAL OPERATING COSTS FOR SOLID WASTE MANAGEMENT

Plant	Scenario	"Levelized" Annual Cost (Thousands of dollars)	Cost in Year 4
Paraho	Less Strict ^a	4787	3900
	More Strict ^a	6447	5666
TOSCO II	Less Strict ^a	2569	2149
	More Strict ^a	3792	3359
MIS	Less Strict ^a	3375	2854
	More Strict ^a	4334	3885
MIS/Lurgi	Less Strict ^b	920	874
	More Strict ^{a,b}	1614	1576

^a All costs exclude the cost of the trusts (see Section 10.4.3)

^b Excludes the cost of slurry backfilling.

case the usual start-up profiles for the first two or three years were incorporated, in order to obtain a uniform distribution of trust-related costs over each unit of output.* (See Appendix 10.0, Section B for further detail.)

The treatment of these trusts for tax purposes might attract a special ruling, but the situation appears to be analogous to that of reclamation expenses, which cannot be deducted until the expenditures have actually occurred. (In contrast, payments to a trust for black lung disease have been ruled to be deductible [17].) If the contributions are not deductible when incurred, then their effective cost to the developer is substantially increased, because this money is committed, but cannot be written off against taxes. In attributing the effect of these trust payments to operating costs, the payments were multiplied by 2.024 ($=100/[100 - 50.6]$), since the effective tax rate is 50.6%. This treatment does not allow for the fact that the trust payments can be deducted after the expenditures have been incurred. However, the latter takes place no earlier than Year 26, and could be several years later, so the extent of the overestimate of this cost is very small in terms of present value, since the major payments occur in Year 0 (and Year 12 for some cases).

Table 10-13 presents the equivalent annual costs for the sum of the two trusts.

TABLE 10-13
EQUIVALENT ANNUAL COSTS FOR BOTH TRUSTS

Plant	Equivalent Annual Costs (thousands of dollars)	
	Direct Annual Cost	Annual Cost With Tax Effect
Paraho	897	1,815
TOSCO II	507	1,026
MIS	455	920
MIS/Lurgi	128	259

*In the case of the regular operating costs discussed in Section 10.4.2, the costs for the first 2 or 3 years were higher than would actually be incurred because they did not reflect the lower than normal output in these years. Hence, these costs were equated to a uniform series since the data were based on uniform output. However, the trust contributions are independent of the start-up profile, so this profile was incorporated into the calculation of the equivalent annual cost to correctly attribute this cost to each unit of output.

The sum of the "levelized" annual cost (Table 10-12) and the equivalent annual cost for the trusts, with tax effect (Table 10-13) constitute the direct operating costs for solid waste management. This cost is termed the "uniform annual operating cost" in Table 7-1.

10.4.4 Indirect Operating Costs for Solid Waste Management

Indirect operating costs for solid waste management were calculated using the same algorithm as for other indirect operating costs (see Section 10.3.2.), except that the trusts were not included in the direct cost base (see Appendix 10.0, Section B). As noted in Section 10.3.2., 8% amortization was used for calculating the severance tax credit where there is capital investment in Years 0 and 12.

10.5 SENSITIVITY ANALYSIS

A feature of economic evaluations is that they require a number of assumptions which can reflect not only future uncertainty (e.g., government taxation practices) but also differing company policies (e.g., required DCF ROR). Unlike technological uncertainties, which can generally be reduced with further research (e.g., process efficiency, groundwater yield to mines), economic uncertainties often cannot be resolved at the time of a feasibility study (e.g., future tax rates, or rulings on the tax treatment of items such as trust contributions). Hence, economists usually conduct sensitivity analysis to answer a series of "what if?" questions, concerned with the sensitivity of the results of the economic analysis to both variables that are beyond the firm's control, and to contingencies such as construction delays.*

Most sensitivity analyses are concerned with capital-related assumptions. In Section 10.2.3. the effect on the capital recovery factor, r , of alternative assumptions is discussed, and is illustrated in Table 10-11. It is the intention that from these data, the reader can gauge the approximate sensitivity of the results to different economic assumptions and DCF ROR

*Assumptions about markets are frequently an important part of a sensitivity analysis, but are not relevant to this study.

criteria, and perform such sensitivity analyses as he desires. Unlike the technical assumptions, which tend to be conservative (see Volume II of this report), the assumptions used in the economic analysis are intended to reflect "middle of the road" views. Since the effect of cumulating a number of conservative assumptions (or a number of liberal ones) tends to be multiplicative, it is important not to place too much emphasis on scenarios containing many conservative or liberal assumptions.

To illustrate the sensitivity of the results to alternative capital-related assumptions Table 10-14 shows the effect of adopting the conservative assumptions of increasing the DCF ROR to 15% and assuming that the phasing of all pollution-related capital expenditures is the same as that for a complete oil shale development, with depreciation over 13 years,* starting with Year 1. (As noted in Section 10.2.3, changes in the DCF ROR and project timing assumptions have the greatest effect on the value of "r".) Although the capital recovery factor for these conservative assumptions is 26.27%† for an above-ground retorting plant (cf. 16.07% for air and most water equipment for the same plant), Table 10-14 shows that the overall effect on pollution control costs is not large, with increases of 22.8 to 30.3% or 22 to 69 cents per barrel of equivalent syncrude.

The assumptions about operating costs that might be subject to scrutiny are those relating to labor productivity (in constant dollar terms) and any overhead charges (e.g., charges to support head office costs. Labor is a minor item in the operating costs, and while there are many ways in which overhead can be levied, in the authors' experience a labor overhead is most common. The average labor cost of \$20 per hour to include both fringe benefits (e.g., social security) and labor overhead is probably adequate.

*This is clearly inappropriate to the two site solid waste management scenarios, but the capital involved is comparatively small.

†The capital recovery factors used in Table 10-14 were specifically calculated for the given assumptions. However, a rough estimate of the capital recovery factor for a case involving multiple variations in assumptions may be obtained by summing the differences from the "base case" (e.g., $19.13 + 24.3 - 16.07 = 24.30$, cf. 26.27 for this example). The smaller the change in "r," the less inaccurate is this procedure.

Table 10-14
SENSITIVITY ANALYSIS FOR CHANGES IN
CAPITAL-RELATED ASSUMPTIONS

Plant	Scenario	Total Cost, Dollars per Barrel Syncrude Equivalent		Percentage Difference
		Standard* Assumptions	Conservative† Assumptions	
Paraho	Less Strict	0.88	1.11	+25.8%
	More Strict	1.00	1.22	+23.1
TOSCO II	Less Strict	1.65	2.06	+24.8
	More Strict	2.00	2.55	+27.2
MIS	Less Strict			
	Low mine water	2.09	2.70	+29.3
	High mine water	2.54	3.19	+25.7
	More Strict			
	Low mine water	2.25	2.87	+27.4
	High mine water	2.99	3.68	+22.8
MIS/Lurgi	Less Strict			
	Low mine water	1.92	2.51	+30.3
	High mine water	2.13	2.74	+28.2
	More Strict			
	Low mine water	1.96	2.54	+29.7
	High mine water	2.30	2.91	+26.7

*Standard assumptions are specified elsewhere in this report, and include 13% DCF ROR and investment in Year 0 or Year -3. (See Table 10-10.)

†Conservative assumptions increase DCF ROR to 13% and invest 15% in Year -3, 20% in Year -2, 40% in Year -1, 25% in Year 0 with depreciation over 13 years commencing in Year 1. All other assumptions are unchanged. (r = 26.27% for Paraho and TOSCO II plants, r = 29.68% for MIS and MIS-Lurgi plants.)

In view of the small magnitude of labor costs in pollution control activities no sensitivity analysis has been attempted. Nevertheless, it is possible that some companies would impose overhead charges on pollution controls using some other basis, and not covered by the required DCF ROR. This would raise operating costs.

REFERENCES

1. Based on private communications from sources at a number of mining and oil companies.
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3. J. Nutter and C. Waitman, Oil Shale Economics Update, Tosco Corporation, Los Angeles, California, 1978. (Presented at AIChE, S. California Section Annual Technical Meeting, Anaheim, CA, 18 April 1978.)
4. N. Ericsson and P. Morgan, "The Economic Feasibility of Shale Oil: An Activity Analysis," The Bell Journal of Economics, 9(2): 457-487, 1978.
5. H. Pforzheimer, "Paraho--Outlook for Commercial Development," Quarterly of the Colorado School of Mines, 71(4): 65-84, 1976.
6. J. Schanz, Jr. and H. Perry, "Oil Shale--A New Set of Uncertainties," Natural Resources Journal, 18(Oct): 775-785, 1978.
7. A. Grossman, "Economic Evaluation of Combined In Situ and Surface Retorting of Oil Shale." In: Tenth Oil Shale Symposium Proceedings, (J. Reubens, ed.) Colorado School of Mines, Golden, Colorado, 1977, pp. 9-15.
8. Based on published Environmental Impact Statements, private communications from oil shale developers and references 2 and 3.
9. Telephone interview, L. Spencer, Gulf Minerals, Denver, Colorado, 16 February 1979.
10. Private communications from more than one oil shale developer.*
11. Telephone interview, R. Heistand, Paraho Development Corporation, Rifle, Colorado, 28 February 1979.
12. Private communication from an oil shale developer.*
13. These decisions were made by the authors of this report, but are supported by published Environmental Impact Statements and by private communications from oil shale developers.
14. Authors' assumption, based on private communications from two oil shale developers.

*References 10 and 12 appear more than once in this chapter. This does not imply that the same developers are involved in each case.

15. Personal interview, R. Venn, Colorado Division of Property Taxation, Denver, Colorado, 2 March 1979.
16. Telephone interview, J. O'Donnell, Colorado Division of Property Taxation, Denver, Colorado, 22 March 1979.
17. Telephone interview, P. Korth, Gulf Mineral Resources Co., Denver, Colorado, 14 March 1979.

APPENDIX 10.0

SECTION A, CALCULATION OF CAPITAL RECOVERY FACTORS

A sample calculation of a capital recovery factor ("r") is shown in Table 10.A.1. This particular factor is based on the following assumptions:

- (i) capital is 100 percent equity
- (ii) all capital is expended in Year 0 (i.e., the year before production commences)
- (iii) production as follows
 - Year 1 : 56% of full output
 - Year 2 : 83% of full output
 - Years 3-25 : full output
- (iv) 20% investment credit
- (v) 13 year double declining balance depreciation (with switch-over to straight line)
- (vi) 50.6% effective income tax rate
- (vii) Required DCF ROR = 13% (after tax)

The details of the calculation are as follows. Column [2] shows a capital expenditure of \$1,000 in Year 0. (Expenditure is shown as a negative quantity, while income and credits are positive.) In Year 1 (the first year of production) the investment credit of \$200 (20% of \$1000) is allowable, together with the first year's depreciation. In Years 2 through 14 the remaining depreciation is allowable, as shown in column [3]. Column [3] sums to \$1200 as the investment credit is additional to the regular depreciation. Column [4] shows the value of these allowances to the company, assuming that the company or its parent(s) have profits from other operations against which to claim the allowances.* The \$200 investment credit is a direct charge

*If the company were independent, and had only the one project, it would have to wait to claim the credit and depreciation allowances until it had sufficient income to cover them. Since this is equivalent to deferring income it would increase the capital charge due to the time value of money.

Table 10-A-1 EXAMPLE OF CAPITAL RECOVERY FACTOR CALCULATION

YEAR	GROSS CAPITAL	ALLOWANCES		OPERATING INCOME		DCF FACTORS @ 13%	NET PRESENT VALUES		
		CREDIT/ DEPRECIATION	VALUE AT 50.6% TAX EFFECT	GROSS	NET AFTER 50.6% TAX		INCOME	ALLOWANCES	CAPITAL
[1]	[2]	[3]	[4]	[5]	[6]	[7]	[8]	[9]	[10]
0	(1,000)					1.0000			(1,000.00)
1		200.00	200.00	.56x	.277x	.8850	.2448x	177.00	
2		153.85	77.85					68.90	
3		130.18	65.87	.93x	.410x	.7831	.3211x	51.58	
4		110.15	55.74	1.00x	.494x	.6931	.3424x	38.63	
5		93.21	47.16	1.00x	.494x	.6133	.3030x	28.92	
6		78.87	39.91	1.00x	.494x	.5428	.2681x	21.65	
7		66.73	33.77	1.00x	.494x	.4803	.2373x	16.22	
8		56.46	28.57	1.00x	.494x	.4251	.2100x	12.15	
9		51.76	26.19	1.00x	.494x	.3762	.1858x	9.85	
10		51.76	26.19	1.00x	.494x	.3329	.1645x	8.72	
11		51.76	26.19	1.00x	.494x	.2945	.1455x	7.72	
12		51.76	26.19	1.00x	.494x	.2607	.1288x	6.83	
13		51.76	26.19	1.00x	.494x	.2307	.1140x	6.04	
14				1.00x	.494x	.2042	.1009x	5.35	
15				1.00x	.494x	.1807	.0893x		
16				1.00x	.494x	.1599	.0790x		
17				1.00x	.494x	.1415	.0699x		
18				1.00x	.494x	.1252	.0618x		
19				1.00x	.494x	.1108	.0547x		
20				1.00x	.494x	.0981	.0485x		
21				1.00x	.494x	.0868	.0429x		
22				1.00x	.494x	.0768	.0379x		
23				1.00x	.494x	.0680	.0336x		
24				1.00x	.494x	.0601	.0297x		
25				1.00x	.494x	.0532	.0263x		
25				1.00x	.494x	.0471	.0233x		
TOTALS		<u>1,200.00</u>	<u>706.01</u>				<u>3.3631x</u>	<u>459.57</u>	<u>(1,000.00)</u>

$$3.3631x = 1,000 - 459.57; \quad x = \frac{540.43}{3.3631} = 160.69; \quad r = \frac{160.69}{1,000} = 16.07\%$$

against tax payable, and hence is transferred directly to column [4], but the depreciation only reduces the project's profits, so its value to the company is the reduction in the tax liability, i.e., 50.6% of each year's depreciation allowance.

Column [5] shows a stream of operating income (i.e., income after deducting regular operating costs) that results from the \$1000 investment. In a full output year (i.e., Years 3 through 25) the income is designated as "x", but in the start-up years it is less, i.e., 0.56x in Year 1 and 0.83x in Year 2. The benefit to the company of this stream of operating income is only 49.4% of the gross operating income, since the company must pay 50.6% tax on its profits. This net value is shown in column [6].

Column [8] is the present value (referred to Year 0) of the net operating income (column [6]) after multiplying by the 13% discount factors shown in column [7]. Columns [9] and [10] are the discounted values of the effective value of the allowances (column [4]), and the capital (column [2]), respectively. The totals of columns [8] and [9] are the net present values of the after-tax income and the tax-reducing effect of the allowances. By adding these totals to total of column [10],^{*} and equating to zero, we can determine the value of x at which the net present value of the entire cash flow is zero when discounted at 13%.

$$\text{i.e., } 3.3631x + 459.57 - 1000 = 0$$

$$\begin{array}{ccc} [8] & [9] & [10] \end{array}$$

This value of x, \$160.69, is therefore the magnitude of the gross operating income in a full production year that is required to provide a 13% DCF ROR on \$1000 capital. Thus the capital recovery factor, r, is 160.69/1000, i.e., 16.07%. This means that to provide the specified 13% on investment, a capital charge of 16.07% of the total capital must be added to normal year annual operating costs. Note, however, that these operating costs should not include any depreciation or other capital related costs, such as interest.

^{*}In this case column [10] only comprises one entry, but in general it may contain the discounted values of capital expenditures spread over several years.

Because the operating income was made proportional to production in the initial years, this implies that the capital charge is applied uniformly to each unit of output, facilitating the calculation of a cost per barrel, or other measure of production.

SECTION B, DETAILS OF CERTAIN CALCULATIONS

Each note below refers to a superscript letter on Tables 10-2 through 10-9.

(a) Capital recovery factors were taken from Table 10-10. Note that each case involves three different values of r : one for Miscellaneous Water Management (B), one for air and all other water-related capital, and one for solid waste disposal.

(b) Direct operating costs do not include byproduct credits.

(c) Indirect operating costs were calculated as follows:

Tax and insurance = 3% of capital cost

Severance tax credit = 4% [Direct operating cost + 4% of capital cost (25 year amortization)].

Hence, indirect operating cost = 0.0284 [Capital cost] - 0.04 [Direct operating cost].

(d) Indirect operating costs were calculated as in (c), less NH_3 credits as follows:

TOSCO II: 134 tpsd @ \$100/ton = \$4,422,000

Paraho: 146 tpsd @ \$100/ton = \$4,818,000

(e) Indirect operating costs were calculated as in (c) less sulphur credits as follows:

TOSCO II: 192 tpsd @ \$5/ton = \$316,800

Paraho: 132 tpsd @ \$5/ton = \$217,800

(f) Indirect operating costs were calculated as follows:

Tax and insurance = 3% of capital cost

Severance tax credit = 4% [Direct operating cost + 8% of Year 0 capital cost (12½-year amortization)].

Hence, indirect operating cost = 0.0268 [Capital cost (Year 0)]
-0.04 [Direct operating costs].

(g) Direct operating costs for solid waste management were calculated by discounting the annual cash flow derived from Tables 7-12 through 7-15 at 13% per year with respect to Year 0, and dividing by 7.330. (The factor 7.330 is the net present value of a uniform cash flow of \$1 for Years 1 through 25, discounted at 13% per year to Year 0).

The results, termed the "levelized" annual costs are presented in Table 10-12.

(h) Direct operating costs for solid waste management were calculated as in (g), but also include the trusts.

The lump sum trust payments were converted to an equivalent annual series as described below. This equivalent annual payment was added to the regular annual trust payment, to provide the direct annual cost as shown in Table 10-13. However, this direct cost was divided by 0.494 to allow for the effect of tax (see Section 10.4.3). This "with tax effect" is also shown in Table 10-13.

The direct annual operating cost for solid waste management is the sum of the "levelized" annual cost from Table 10-12 and the equivalent annual cost of both trusts, with tax effect, from Table 10-13.

Calculations of Equivalent Annual Series for Trusts.

(i) TOSCO II. The Year 0 payment was divided by 6.8075. This factor represents the net present value at Year 0 of \$0.56 in Year 1, \$0.83 in Year 2 and \$1 in Year 3 through 25, discounted at 13% per year.

(ii) Paraho. The Year 0 payment was divided by 5.5314. This factor is 6.8075 calculated as in (i) above, divided by 1.2307 which reflects the net present value of \$1 in

Year 0, plus \$1 in Year 12 (i.e., it provides for the two equal trust payments).

(iii) MIS/Lurgi. The Year 0 payment was divided by 6.1015. This factor represents the net present value of \$0.25 in Year 1, \$0.50 in Year 2, \$0.75 in Year 3 and \$1 in Years 4 through 25.

(iv) MIS. The Year 0 payment was divided by 4.9577. This factor is 6.1015 divided by 1.2307 (see (ii) and (iii) above).

(i) Indirect operating cost was calculated as follows: Tax and insurance = 3% of capital cost. Severance tax credit = 3% [Direct operating cost + 4% of capital cost (25-year amortization)]. Hence, indirect operating cost = 0.0288 [Capital cost] - 0.03 [Direct operating cost].

(j) Indirect operating costs were calculated as follows: Tax and insurance = 3% of capital cost. Severance tax credit = 3% [Direct operating cost + 8% of capital cost (12½-year amortization) + 4% of capital cost (25-year amortization)]. Hence indirect operating cost = 0.0288 [Capital cost (Year 0-Year 12)] + 0.0276 [Capital cost (Year 12)] - 0.03 [Direct operating cost].

(k) Indirect operating cost was calculated as in (i), less NH₃ credit as follows:

$$281 \text{ tpsd} @ \$100/\text{ton} = \$9,273,000.$$

(l) Indirect operating cost was calculated as in (i), less sulphur credit as follows:

$$144 \text{ tpsd} @ \$5/\text{ton} = \$237,600.$$

(m) Indirect operating costs were calculated as follows: Tax and insurance = 3% of capital cost. Severance tax credit = 3.30% [Direct operating cost + 4% of capital cost (25-year amortization)]. Hence, indirect operating cost = 0.02868 [Capital cost] - 0.033 [Direct operating cost].

- (n) Indirect operating cost was calculated as in (m), less NH_3 credit as in (k).
- (o) Indirect operating cost was calculated as in (m), less sulphur credit as follows:
172 tpsd @ \$5/ton = \$283,800
- (p) Developer's design for clarifier.
- (q) Modified design for clarifier.
- (r)* Clarifier sized for 11,000 gpm, excess mine water treatment for 8,500 gpm.
- (s)* Clarifier sized for 4,000 gpm, excess mine water treatment for 1,500 gpm.
- (t)* Clarifier sized for 6,000 gpm, excess mine water treatment for 4,370 gpm.
- (u) Clarifier sized for 2,000 gpm, no excess mine water treatment.
- (v) Indirect cost calculated as in (f), but cost of trust excluded from direct operating cost.
- (w) Indirect cost calculated as in (c), but cost of trust excluded from direct operating cost.
- (x) Indirect cost calculated as in (j), but cost of trust excluded from direct operating cost.
- (y) Indirect cost calculated as in (m), but cost of trust excluded from direct operating cost.

*In more strict scenarios, costs were based on treatment using two RO stages, as opposed to the "alternative treatment" of a weak acid IX and a RO stage.