EPA 600/7-88-003

NTIS PB88 196027

AIR POLLUTION CONTROL ALTERNATIVES FOR SHALE OIL PRODUCTION OPERATIONS

By:

H. J. Taback
R. J. Goldstick

KVB, INC.
Engineering and Research Division
Irvine, CA 92714

EPA Contract No. 68-03-3166

Project Officer

Edward R. Bates
Air and Energy Engineering Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

Prepared For:

U.S. Environmental Protection Agency
Office of Research and Development
Washington, D.C. 20460

December 1987

NOTICE

The information in this document has been funded wholly or in part by the United States Environmental Protection Agency under Contract Number EPA No. 68-03-3166 to Metcalf & Eddy, Wakefield, MA. It has been subject to the Agency's peer and administrative review, and it has been approved for publication as an EPA document. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted and used, the related pollution impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The EPA's Air and Energy Engineering Research Laboratory (Research Triangle Park, NC) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report is a compilation of air pollution control technology specifically applicable to the various processes employed in the production of shale oil, along with a forecast of air pollutant emission rates achievable using that technology. The results should assist developers and permit writers in selecting controls and predicting emissions for any future oil shale projects.

Frank Princiotta, Director
Air and Energy Engineering Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, NC 27711

ABSTRACT

The available air emission data and air pollution control technology relevant to the production of shale oil are consolidated, evaluated and presented in a manner to be useful to project developers in preparing environmental impact statements and permit applications under the Clean Air Act as well as to their respective regulatory approval agencies. The processes covered include subsurface and surface mining; raw shale sizing and handling; all of the various retorting schemes, both in-situ and above-ground; spent shale combustion; spent shale disposal and product upgrading. The air pollution control technology covered includes most of the traditional processes for nitrogen oxide (NO_x), sulfur compounds, particulate, volatile organic compounds (VOC) and carbon monoxide (CO) control. However, some recently-developed processes and processes not discussed elsewhere in the oil shale literature are included; such as catalytic mufflers for NO_x, VOC and CO control on surface and subsurface vehicles; a new wet scrubbing process (caustic/charcoal/sodium hypochlorite solution) for organic sulfur compounds, staged combustion for NO_x control of high nitrogen fuels; spent shale absorption for sulfur oxides; improved filter bag materials, moving bed granular filters and dry venturis for fine particulate control; and dry sorbent injection for sulfur oxide control.

Data extracted from PSD permit applications from seven different shale oil projects are analyzed and compared. Data consistencies and inconsistencies are noted for future reference. Finally, five shale oil recovery processes, considered to be the best of the state of the art, are analyzed at three levels of emission control. The baseline is comparable to that used in the PSD permit applications. The first improvement level represents the use of existing proven technology used commonly in other industries. The second and greater improvement level represents the additional use of technology that may still require some development before committing it to a full scale shale oil production application.

The interesting conclusion is made that if the second level of improved control technology is applied, the overall emission levels in terms of weight of emissions per unit of oil produced will be essentially the same whether the oil is produced in either in-situ or above-ground retorts and whether the retorts are either directly or indirectly heated.

Based on this highest degree of control, the emissions that might be expected from a shale oil production plant under the best conditions are approximately:

	$Kg/1000m^3$ Oil Produced
CO	200
VOC	100
NO_x	700
SO_x	200
TSP	200

TABLE OF CONTENTS

Section			Page
Foreword			iii
Abstract			iv
Figures			viii
Tables			xii
Acronyms and Abbreviations			xv
Conversion Factors			xvii
1.0	EXECUTIVE SUMMARY		1-1
	1.1	Introduction	1-1
		1.1.1 Background	1-1
		1.1.2 Objective	1-2
		1.1.3 Approach	1-3
	1.2	Overview	1-3
	1.3	Retort Process Review	1-5
		1.3.1 Retort Gas	1-6
		1.3.2 Sulfur Gases	1-8
		1.3.3 Nitrogen Gases	1-9
		1.3.4 Specific Process Review	1-10
	1.4	Air Pollution Control Technology Review	1-12
		1.4.1 Dry Venturi	1-13
		1.4.2 Fugitive Dust	1-13
		1.4.3 Staged Combustion	1-13
		1.4.4 Activated Carbon and Hypochlorite	1-14
		1.4.5 SO_x Absorption on Spent Shale	1-14
		1.4.6 Catalytic Mufflers	1-14
	1.5	Prevention of Significant Deterioration (PSD) Permit Applications	1-14
		1.5.1 Total Facility Emissions	1-15
		1.5.2 Individual Unit Operations	1-17
	1.6	Retort Process Evaluation	1-21
		1.6.1 Shale Oil Recovery Plant	1-21
		1.6.2 Methodology	1-22
		1.6.3 Mining, Solids Handling and Upgrading Emissions	1-24
		1.6.4 Retort Gas Emissions	1-25
		1.6.5 Retort Gas Combustion and End-of-Pipe	1-26
		1.6.6 Emissions from Retort Gas Combustion	1-29

Section		Page
2.0	RETORT PROCESS REVIEW	2-1
2.1	Background	2-4
2.1.1	Chemistry	2-5
2.2	Formation and Fate of Pollutants	2-14
2.2.1	Criteria Pollutants (SO_x, NO_x)	2-14
2.2.2	Trace Elements	2-25
2.3	Retort Processes	2-27
2.3.1	Above-Ground Retorts	2-31
2.3.2	In-situ Retorts	2-33
2.4	Process Descriptions	2-35
2.4.1	Paraho Retort Process	2-35
2.4.2	Hytort Oil Shale Retorting	2-37
2.4.3	T^3 Oil Shale Retorting	2-39
2.4.4	Lurgi Retorting Process	2-40
2.4.5	Tosco II Oil Shale Retorting	2-42
2.4.6	Union A, B, C and SGR	2-46
2.4.7	Circular Grate Retort	2-49
2.4.8	Fluidized-Bed (Chevron STB Oil Shale Retorting)	2-51
2.4.9	Cascading-Bed Retort and Combustor	2-55
2.4.10	Occidental Vertical Modified In-situ (VMIS)	2-55
2.4.11	Horizontal In-situ Retorting (Geokinetics)	2-59
3.0	AIR POLLUTION CONTROL TECHNOLOGY REVIEW	3-1
3.1	Particulate Emission Controls	3-6
3.1.1	Point Source Technologies	3-6
3.1.2	Fugitive Dust Technologies	3-50
3.2	Nitrogen Oxides Control	3-96
3.2.1	Staged Combustion	3-98
3.2.2	Exhaust Gas Treatment	3-115
3.3	Sulfur Compound Control	3-134
3.3.1	Stretford Process (H_2S)	3-135
3.3.2	Lo-Cat Process (Reduced Sulfur Compounds)	3-143
3.3.3	Unisulf Process (H_2S)	3-151
3.3.4	Alkaline Scrubbing	3-157
3.3.5	Activated Carbon and Hypochlorite Process	3-160
3.3.6	Claus Process (H_2S)	3-179
3.3.7	Operating Requirements for Claus Process	3-188
3.3.8	Flue Gas Desulfurization (Wet and Dry)	3-199
3.3.9	SO_x Absorption in Spent Shale (ASSP)	3-217
3.4	Engine Emissions of CO and HC	3-223

Section		Page
4.0	PSD PERMIT APPLICATION ANALYSIS	4-1
4.1	Particulate Emissions	4-22
4.1.1	Mining - Below-Ground (Primary Category 1)	4-23
4.1.2	Above-Ground Mining Emissions (Primary Category 2)	4-24
4.1.3	Retort Operations and Retort Gas Combustion	4-25
4.1.4	Upgrading and Miscellaneous	4-26
4.2	Total Facility Gaseous Emissions	4-27
4.3	Emission Source Breakdown	4-31
4.3.1	Carbon Monoxide	4-31
4.3.2	Hydrocarbons	4-34
4.3.3	Nitrogen Oxides	4-38
4.3.4	Sulfur Oxides	4-42
4.4	PSD Application Data Inconsistencies	4-45
4.4.1	Carbon Monoxide	4-45
4.4.2	Sulfur Oxides	4-47
4.5	Actual Permitted Emission Rates	4-48
5.0	PROCESS ANALYSIS	5-1
5.1	Methodology	5-3
5.1.1	Mining	5-4
5.1.2	Retort	5-5
5.1.3	Gas Utilization and End-of-Pipe	5-11
5.2	Results	5-14
5.2.1	Direct Combustion - Case #1	5-14
5.2.2	Direct Combustion - Circular Grate Case #2	5-20
5.2.3	Indirect Combustion - Gas Recycle Case #3	5-23
5.2.4	Indirect Combustion - Solids Recycle Fluidized Bed Spent Shale Combustor Case #4	5-26
5.2.5	Modified In-situ with Indirect Combustion Gas Recycle Above-Ground Process Case #5	5-30
5.2.6	Summary	5-33
6.0	REFERENCES	6-1

APPENDICES

A.	Trace elements	A-1
B.	Codisposal emissions	B-1
C.	Summary of PSD permit application emissions	C-1

FIGURES

Number		Page
1-1	Total Facility Emissions Summary	1-16
1-2	Carbon Monoxide Emissions	1-18
1-3	Hydrocarbon Emissions	1-18
1-4	Nitrogen Oxides Emissions	1-19
1-5	Sulfur Oxides Emissions	1-19
1-6	Particulate Emissions	1-20
1-7	Particulate Emissions from Retort Gas Combustion for Five Cases	1-30
1-8	Nitrogen Oxide Emissions Summary for Five Cases	1-30
1-9	Sulfur Oxide Emissions Summary for Five Cases	1-31
1-10	Nitrogen Oxide Emissions for Five Cases	1-33
1-11	Particulate Emissions for Five Cases	1-34
1-12	Sulfur Oxide Emissions for Five Cases	1-34
2-1	In-situ Retort Processes - General	2-2
2-2	Above-Ground Retort Processes - General	2-3
2-3	Carbonate Reaction Scheme	2-9
2-4	Sulfur Reactions in Recycled Shale Process	2-11
2-5	Nitrogen Reactions During Char Combustion	2-13
2-6	Partitioning of Sulfur During Retorting	2-18
2-7	Nitrogen Conversion Mechanisms	2-20
2-8	Nitrogen Distribution During Retorting	2-22
2-9	Ammonia Evolved from Fischer Assay	2-23
2-10	Schematics of Above-Ground Retorting Processes	2-29
2-11	Schematics of Established In-Situ Retorting Processes	2-30
2-12	Paraho Retort	2-36
2-13	Simplified Schematic of the Hytort Process	2-38
2-14	Schematic of the T^3 Process	2-41
2-15	Lurgi-Ruhrgas Retorting Process	2-43
2-16	Fines Type Tosco II Retort	2-44
2-17	Union Processes	2-47
2-18	Superior Direct-Heated Process	2-50
2-19	Schematic of a Fluidized-Bed Retort System	2-52
2-20	Two-Stage Fluidized-Bed Retort	2-53
2-21	Concept for Cascading-Bed Retorting System	2-56
2-22	Cascading-Bed Oil Shale Combustor	2-57
2-23	Soak Tank Portion of Cascading-Bed Retort	2-58
2-24	Occidental Vertical Modified In-situ Process	2-60
2-25	An Example of a Geokinetics Horizontal In-situ Retort	2-61
3-1	Test Sheet	3-2
3-2	Isometric View of a Two-Compartment Pulse-Jet Fabric Filter	3-8
3-3	Predicted and Observed Result for Bench Scale Fabric Filter Tests	3-11
3-4	Fabric Filter Field Test Results for Different Bag Materials	3-12
3-5	Filter Pressure Drop for Reverse Air-Jet Operation	3-13

Number		Page
3-6	Costs for Pressure Flow, Intermittent Duty, Mechanical Shaker Standard Construction Baghouses	3-16
3-7	Costs for Continuous Duty, Pulse Jet Standard Construction Baghouses	3-17
3-8	Costs for Continuous Duty, Pressure Flow, Mechanical Shaker Standard Construction Baghouses	3-18
3-9	Costs for Continuous Duty, Pressure Flow, Reverse Air Standard Construction Baghouses	3-19
3-10	Costs for Continuous Duty, Reverse Air Custom Construction Baghouses	3-20
3-11	Standard Venturi Scrubber	3-24
3-12	Efficiency Performance of Venturi Scrubbers	3-25
3-13	Cut Diameter for Gas-Atomized Spray Scrubbers	3-29
3-14	Venturi Cut Diameter vs Liquid-to-Gas Ratio	3-30
3-15	Schematic Arrangement of Wire/Plate and Wire/Tube Precipitators	3-33
3-16	Drop in Precipitation Rate W_e with Increasing Fly Ash Resistivity for a Representative Group of Precipitators	3-38
3-17	Relation of W_e to Coal Sulfur Content for Flue Gas Temperatures in the Neighborhood of 420K as Determined by Several Investigators	3-38
3-18	Theoretical Precipitator Collection Efficiencies at Different Migration Velocities	3-39
3-19	Capital Investment for Collectors on 500 MW (Net) Power Plants	3-45
3-20	Purchase Price of Dry-Type Electrostatic Precipitators	3-46
3-21	Dry Venturi	3-48
3-22	Sources of NO_x Emissions Using Western Kentucky Coal and a Divergent Injector with 600°F Preheat	3-106
3-23	Low NO_x Combustor for Sewage Sludge	3-108
3-24	Emissions from Isothermal Combustion of Retorted Shale	3-111
3-25	Arrhenius Plot of Rate Data for Nitric Oxide Reduction by Retorted Oil Shale	3-112
3-26	NO Reduction Variation with Temperature	3-113
3-27	Exit NO and NH_3 Concentrations for Laboratory Test with Simulated Flue Gas	3-116
3-28	Effect of Temperature on NO Reductions with Ammonia Injection	3-117
3-29	Effect of Initial NO Level with NH_3 Injection	3-119
3-30	Coal Firing with Ammonia Flue Gas Injection, 80 hp Boiler	3-121
3-31	Ammonium Bisulfate/sulfate Equilibrium with Field Operating Conditions Overplotted	3-124
3-32	Simplified Flow Diagram of the Stretford Pilot Plant	3-137
3-33	Stretford Capital Investment Cost	3-142
3-34	Flow Scheme for Conventional Lo-Cat Unit	3-145
3-35	Unisulf Process	3-153
3-36	Centrifuge Capacity for Separating Sulfur from Aqueous Feed Slurry	3-155

Number		Page
3-37	Alkaline Scrubbing Process Schematic for H_2S Removal	3-158
3-38	Removal Efficiency vs Selectivity for Alkaline Scrubber	3-161
3-39	Two-Stage Alkaline Scrubber	3-162
3-40	Tray Tower Scrubber with Isolated Liquid Inlets	3-163
3-41	Teller System Flow Sheet	3-164
3-42	Equilibrium H_2S Concentration in Flue Gas	3-167
3-43A	Schematic Diagram of One Version of PPRIC/BCRC Scrubbing Process Using a TESI Cross-Flow Scrubber	3-171
3-43B	Schematic Diagram of the PPRIC/BCRC Scrubbing Process in a Packed Counter-Current Scrubber	3-171
3-44	Kraft Recovery Boilers TRS Removal Data	3-173
3-45	TRS Emissions Inlet vs Outlet Condition 2	3-174
3-46	TRS Emissions Inlet vs outlet Condition 1	3-175
3-47	Kraft Boiler with Direct Contact Evaporation	3-177
3-48	Straight-Through Claus Process	3-182
3-49	Claus Plant Installed Costs	3-189
3-50	Estimated Claus Plant Capital Investment Cost	3-190
3-51	Shell Claus offgas Treating (SCOT) Process Schematic Flow Diagram	3-192
3-52	Capital Investment Cost for Scot Units	3-198
3-53	Typical Process Flow Diagram for Lime/Limestone Scrubbing	3-201
3-54	Schematic of a Coal-Fired Power Plant with Dry-Sodium Injection for SO_2 Control	3-209
3-55	SO_2 Removal as a Function of Normalized Stoichiometric Ratio	3-212
3-56	Sodium Utilization as a Function of NSR and Particle Size	3-214
3-57	ASSP Process Flow Diagram	3-218
3-58	Effect of Flue Gas Oxygen on SO_2 and NO_x Emissions	3-222
3-59	Effect of Flue Gas Oxygen on CO and Trace Hydrocarbon Emissions	3-222
3-60	Catalyst Efficiency for Oxidation of CO vs Exhaust Temperature, Engine 1-Various Speeds and Loads	3-227
3-61	Catalyst Efficiency for Oxidation of CO vs Exhaust Temperature, Engine-2 Various Speeds and Loads	3-227
3-62	Effect of Load and Catalyst on Exhaust Odor Intensity	3-230
5-1	Possible Processing Combinations	5-2
5-2	Case #1 - Direct Heat	5-15
5-3	Case #2 - Circular Grate	5-22
5-4	Case #3 - Indirect Heat, Gas Recycle	5-25
5-5	Case #3 - Solids Recycle	5-27
5-6	Case #3 - Modified In-situ (Indirect Heat Above-Ground)	5-31
5-7	Nitrogen Oxide Emissions from Retort Gas Combustion	5-35
5-8	Sulfur Oxide Emissions from Retort Gas Combustion	5-36
5-9	Particulate Emissions from Retort Gas Combustion	5-37

Number		Page
5-10	Particulate Emission for Total Facility	5-39
5-11	Nitrogen Oxide Emission for Total Facility	5-40
5-12	Sulfur Oxide Emissions for Total Facility	5-41

TABLES

Number		Page
1-1	Principal Air Emission Sources	1-4
1-2	Retort Gas Production Rates	1-7
1-3	Partitioning of Nitrogen in Retort	1-9
1-4	Retort Processes Reviewed	1-11
1-5	Air Pollution Control Technologies	1-12
1-6	PSD Permit Applications in Data Base	1-15
1-7	Five Processes for Analysis	1-22
1-8	Design Parameters Used in Analysis	1-24
2-1	Retort Gas Compositions (Solids Recycle)	2-7
2-2	Chemical Analysis for Various Oil Shales	2-14
2-3	Sulfur Gases from Three Raw Shales	2-16
2-4	Distribution of Raw Shale Sulfur in Products	2-17
2-5	Nitrogen Concentration in Oil Shale Components and other Fuels	2-21
2-6	Nitrogen Partitioning from PSD Permit Applications	2-24
2-7	Nitrogen-Containing Species in Retort Off-gas	2-25
2-8	Untreated Retort Gas Characteristics	2-31
3-1	Characteristics of Selected Filter Fabrics	3-9
3-2	Approximate Guide to Estimate Gross Cloth Area	3-21
3-3	Bag Prices	3-22
3-4	Typical Pressure Drops for Venturi Scrubbing Systems	3-27
3-5	Typical Value of Design Variables for Commercial ESPs	3-35
3-6	Typical Average Migration Velocities Encountered in Commercial ESP Systems	3-41
3-7	Typical Application Data for Electrostatic Precipitators	3-42
3-8	Commonly Encountered Precipitator Problems	3-43
3-9	Performance Characteristics of the Dry Venturi-Baghouse System	3-51
3-10	Soil Movement Dust Suppressants	3-55
3-11	Summary of Results	3-58
3-12	Exposed Area and Storage Pile Dust Suppressants	3-62
3-13	Test Plots Application Data	3-71
3-14	Weed Growth Quantification	3-73
3-15	Proper Size Gradation for Unpaved Road Surface	3-78
3-16	Improper Size Gradation	3-78
3-17	Unpaved Road Dust Suppressants	3-81
3-18	Assumptions for Cost-Effectiveness Analysis	3-88
3-19	Preliminary Cost-Effectiveness Comparison to Achieve 50 Percent Control	3-89
3-20	Comparison of Measured Control Efficiencies	3-92
3-21	Total Capital Investment and Annual Revenue Requirements for Improved Thermal $DeNO_x$ Applied to Coal-Fired Boilers	3-126

Number		Page
3-22	Cost Effectiveness for Improved Thermal $DeNO_x$ Applied to Coal-Fired Boilers	3-127
3-23	Total Capital Investment for Thermal $DeNO_x$ and Selective Catalytic Reduction for a Coal-Fired 500 MW Boiler	3-129
3-24	Cost Effectiveness Comparison of Improved Thermal $DeNO_x$ vs Kawasaki Industries SCR for a Coal-Fired 500 MW Boiler	3-129
3-25	SCR Costs	3-133
3-26	Design Data for a Stretford Unit Treating Rectisol Offgas at El Paso Natural Gas Burnham Coal Gasification Facility	3-140
3-27	Typical Compositions of Stretford Absorber Offgas in Beavon Sulfur Removal Plants	3-140
3-28	Operating Requirements for the Stretford Process	3-143
3-29	Chemical Costs for Lo-Cat	3-148
3-30	Capital Costs for Lo-Cat Plant	3-150
3-31	Cost Basis for Lo-Cat	3-151
3-32	Commercial Nucleation Process Applications	3-178
3-33	Teller Recovery Boiler System Performance	3-180
3-34	Sulfur Recovery Variation Relative to Acid Gas Feed Composition and Number of Catalytic Stages	3-185
3-35	Reported Claus Plant Operating Data	3-187
3-36	Operating Requirements for the Claus Process	3-188
3-37	Typical Compositions of Gas Streams in Claus and Scot Units	3-195
3-38	Sulfur Concentration in Scot Offgas	3-196
3-39	Sodium Reagent Materials	3-210
3-40	Costs of SO_2 Control for 90 Percent Removal	3-216
3-41	Cost Comparison for ASSP	3-220
3-42	Catalytic Converters	3-224
3-43	Engines Tested	3-225
3-44	Fuel Properties	3-225
3-45	Emission Rates of CO and HC	3-228
3-46	Acceptable Maximum Emission Rates	3-229
3-47	Exhaust Odor Intensity	3-231
3-48	Aldehyde Emissions	3-233
3-49	Sulfate Emissions	3-234
4-1	Sources and Controls for PSD Applications	4-2
4-2	Controlled Particulate Emissions	4-23
4-3	Controlled Below-Ground Particulate Emissions	4-23
4-4	Controlled Above-Ground Mining and Material Handling Particulate Emissions	4-25
4-5	Controlled Particulate Emissions from Retort Operation	4-25
4-6	Controlled Particulate Emissions from Upgrade and Miscellaneous Above-Ground Operations	4-27
4-7	Total Facility Controlled Gaseous Emissions Summary	4-27
4-8	Increase in NO_x Emissions from Fuel Nitrogen	4-29
4-9	Effect of Organic Sulfur in Retort Gas on Sulfur Oxide Emission	4-30

Continued

Number		Page
4-10	Controlled Carbon Monoxide Emissions	4-31
4-11	Controlled Carbon Monoxide Emissions for Below-Ground Mining	4-32
4-12	Controlled Carbon Monoxide Emissions from Retort and and Gas Combustion	4-33
4-13	Carbon Monoxide Emissions from Upgrading Operations	4-34
4-14	Controlled Hydrocarbon Emissions	4-35
4-15	Controlled Hydrocarbon Emission from Below-Ground Mining	4-36
4-16	Controlled Hydrocarbon Emission from Retort and Gas Combustion	4-36
4-17	Controlled Hydrocarbon Emissions from Upgrading Operations	4-37
4-18	Controlled Nitrogen Oxide Emissions	4-38
4-19	Controlled Nitrogen Oxide Emissions from Below-Ground Mining and Material Handling	4-39
4-20	Controlled Nitrogen Oxide Emissions from Retort and Other Gas Combustion	4-40
4-21	Controlled Nitrogen Oxide Emissions from Upgrading Operations	4-41
4-22	Controlled Sulfur Oxide Emissions	4-42
4-23	Controlled Sulfur Oxide Emissions from Below-Ground Mining	4-43
4-24	Controlled Sulfur Oxide Emissions from Retort and Other Gas Combustion	4-44
4-25	Controlled CO Emissions from White River and Syntana	4-45
4-26	Comparison of Permit Application and Actual Permitted Emission Rates	4-49
5-1	Design Parameters	5-4
5-2	Controlled Emission Rates for Mining	5-5
5-3	Retort Gas Production Rates	5-5
5-4	Effect of Organic Sulfur on Sulfur Emission	5-7
5-5	Summary of Estimated SOx Emission Rates	5-8
5-6	Estimated Emission of Nitrogen Oxide from Combustion of Retort Gas	5-10
5-7	Controlled Particulate Emissions	5-16
5-8	Controlled Sulfur Oxide Emissions from Retort Gas Combustion	5-17
5-9	Controlled Nitrogen Oxide Emission	5-18
5-10	Total Facility Controlled Emissions Case 1 - Direct Combustion	5-19
5-11	Total Facility Controlled Emissions Case 2 - Direct Combustion	5-21
5-12	Total Facility Controlled Emissions Case 3 - Indirect Combustion	5-24
5-13	Total Facility Controlled Emissions Case 4 - Indirect Combustion	5-28
5-14	Total Facility Emissions Case 5 - Modified In-situ	5-32

LIST OF ABBREVIATIONS AND ACRONYMS

ACGIH	American Conference of Government Industrial Hygienists.
APC	Air Pollution Control.
ASSP	Adsorption on Spent Shale Process - a SO_x control.
BACT	Best Available Control Technology.
BAF SET	Baffled Settling - a dust suppression technique.
BCRC	British Columbia Research Council.
BERC	Bartlesville Energy Research Center.
COS	Carbonyl Sulfide.
$DeNO_x$	Exxon's NO_x control involving ammonia injection.
EDTA	Ethylene diamine tetra acetic acid.
EPA	Environmental Protection Agency (U.S.)
EPRI	Electric Power Research Institute.
ESP	Electrostatic Precipitator
FBN	Fuel Bound Nitrogen
FEL	Front end loader.
FGD	Flue Gas Desulfurization.
FGR	Flue Gas Recirculation.
HC	Hydrocarbon.
KVB	Authors of this report.
LBL	Lawrence Berkeley Laboratories.
LLNL	Lawrence Livermore National Laboratory
MIS	Modified In-situ
MW	Megawatts
MeSH	Methyl mercaptan.
NAAQS	National Ambient Air Quality Standards

NO_x	Nitrogen oxides - air pollutants.
NSPS	New Source Performance Standards - EPA's allowable new source emissions.
NSR(1)	New Source Review - a Clean Air Act designation for the permitting requirements for an area not in compliance with the NAAQS.
NSR(2)	Normalized Stoichiometric Ratio.
NTU	Number of Transfer Units - a mass transfer concept.
PCTM	Pollution Control Technical Manual.
PM	Particulate matter - an air pollutant.
ppmV	Parts per million on a volumetric basis.
PPRC	Pulp & Paper Research Institute of Canada.
PSD	Prevention of Significant Deterioration - a Clean Air Act designation for the permitting requirements for an area in compliance with the NAAQS.
SCA	Specific collection area.
SCOT	Shell Claus Offgas Treating
SCR	Selected Catalytic Reduction - a NO_x reduction technique using ammonia and a catalyst.
SGR	Steam - Gas Recirculation - a retorting process.
SNG	Synthetic Natural Gas.
SO_x	Sulfur oxides - air pollutants.
STB	Stage Turbulent Bed - a type of fluid bed combustor.
TESI	Teller Environmental Services, Inc.
TLV	Threshold Limit Value.
TRS	Total Reduced Sulfur
TRW	TRW, Incorporated
UEF	Uncontrolled Emission Factor.
VMIS	Vertical modified in-situ - a retorting process.
VOC	Volatile Organic Compound.

LIST OF
CONVERSION FACTORS
SI Units to Metric or English Units

To Obtain	From	Multiply By
g/Mcal	ng/J	0.004186
10^6 Btu	GJ	0.948
Btu	gm cal	3.9685×10^{-3}
lb/10^6 Btu	ng/J	0.00233
ft	m	3.281
in.	cm	0.3937
ft^2	m^2	10.764
ft^3	m^3	35.314
lb	kg	2.205
Fahrenheit	Celsius	$t_F = 9/5(t_c) + 32$
Fahrenheit	Kelvin	$t_F = 1.8t_K - 460$
psig	Pa	$P_{psig} = (P_{pa})(1.450 \times 10^{-4}) - 14.7$
psia	Pa	$P_{psia} = (P_{pa})(1.450 \times 10^{-4})$
iwg (39.2°F)	Pa	$P_{iwg} = (P_{pa})(4.014 \times 10^{-3})$
10^6 Btu/hr	MW	3.413
GJ/hr	MW	3.60
ft^3/min	m^3/s	2119

To obtain ppm at 3% O_2 of	Multiply* Concentration in ng/J by
Natural Gas Fuel	
CO	3.23
HC	5.65
NO or NO_x	1.96
SO_2 or SO_x	1.41
Oil Fuel	
CO	2.93
HC	5.13
NO or NO_x	1.78
SO_2 or SO_x	1.28
Coal Fuel	
CO	2.69
HC	4.69
NO or NO_x	1.64
SO_2 or SO_x	1.18

*These conversions depend on fuel composition. The values given are for typical fuels.

LIST OF
CONVERSION FACTORS
English or Metric Units to SI Units

Continued

To Obtain	From	Multiply By
ng/J	$lb/10^6$ Btu	430
ng/J	g/Mcal	239
GJ	10^6 Btu	1.055
m	ft	0.3048
cm	in.	2.54
m^2	ft^2	0.0929
m^3	ft^3	0.02832
kg	lb	0.4536
Celsius	Fahrenheit	$t_c = 5/9\ (t_F - 32)$
Kelvin	Fahrenheit	$t_K = 5/9\ (t_F - 32) + 273$
Pa	psig	$P_{pa} = (P_{psig} + 14.7)(6.895 \times 10^3)$
Pa	psia	$P_{pa} = (P_{psia})(6.895 \times 10^3)$
Pa	iwg (39.2°F)	$P_{pa} = (P_{iwg})(249.1)$
MW	10^6 Btu/hr	0.293
MW	GJ/hr	0.278

To obtain ng/J of	Multiply* Concentration in ppm @ 3% O_2 by
Natural Gas Fuel	
CO	0.310
HC	0.177
NO or NO_x (as NO_2)	0.510
SO_2 or SO_x	0.709
Oil Fuel	
CO	0.341
HC	0.195
NO or NO_x (as NO_2)	0.561
SO_2 or SO_x	0.780
Coal Fuel	
CO	0.372
HC	0.213
NO or NO_x (as NO_2)	0.611
SO_2 or SO_x	0.850

*These conversions depend on fuel composition. The values given are for typical fuels.

SECTION 1.0

EXECUTIVE SUMMARY

1.1 INTRODUCTION

For over ten years, the EPA has supported research on the control of air pollutants from shale oil production. This report is a culmination of that effort.

1.1.1 Background

Production of oil from oil shale has been on the verge of becoming a commercial reality a number of times in the last thirty years. However, the barrier of economic uncertainty, related primarily to the availability and price of conventional crude oil, has resulted in limited commercial development efforts. The recent (1986) collapse in the price of crude oil (from $28/bbl to $12/bbl) illustrates the volatile nature of the crude oil market and further enhances the uncertainty and risk associated with the major capital investment required to construct and operate a shale oil recovery plant.

The economics of shale oil recovery are complicated further by the environmental concern associated with oil shale recovery. The mining and processing of oil shale to produce a refined shale oil product can produce a variety of air pollutants, many of which may have adverse environmental impacts if not properly managed.

Under the Clean Air Act (PL 95-95) shale oil project developers must: (a) employ Best Available Control Technology (BACT), (b) insure that National Ambient Air Quality Standards (NAAQS) are not violated, (c) avoid violating the prevention of significant deterioration (PSD) ambient air quality increments, (d) refrain from significant degradation of visibility in mandatory Class I areas and (e) obtain up to one year of baseline data before applying for a PSD permit to construct and operate. These requirements can add substantial costs and complexity to the design and construction of an oil shale recovery plant.

Currently, the most significant aspect of the Clean Air Act to oil shale development is the PSD program. The energy rich oil shale region of Colorado, Utah and Wyoming enjoys some of the cleanest air in the United States. The maximum size of an oil shale project in Colorado may be limited by the emissions of sulfur dioxide, nitrogen dioxide and particulates resulting from that specific project and possibly also by visibility impacts. The Air and Energy Engineering Research Laboratory, which is part of EPA's Office of Research and Development, has conducted an on-going research program to provide technical support to EPA's Program Offices, EPA Regional Offices, other federal and state agencies, and industry. The major efforts in this research program have been directed towards quantifying the air emissions, waste water discharge and solid wastes associated with the various types of recovery facilities.

This program has consisted of a combination of laboratory and field tests to determine actual waste stream quantities and compositions from retorting processes and engineering studies to evaluate various pollution control techniques.

The EPA's engineering study series titled Pollution Control Technical Manuals (PCTM) in particular provides a comprehensive analysis of the air, water and solid emissions from specific shale oil recovery plants. (These PCTMs are cited later in the report.)

1.1.2 Objective

The objective of this report is to consolidate available air emissions data and air pollution control (APC) technology relevant to oil shale processing operations for use by proponents (i.e., PSD, or NSR applicants) and their respective approval agency. Five basic questions are answered in this report:

- What shale oil production processes are available and how do they function?
- What are the specific sources of air pollutants from those processes?
- What APC technology options are applicable to each source; how do they function; what removal efficiency can be

expected; what do they cost; and what rationale should be used to select the most effective one?

- What mass emissions per unit of throughput (e.g., kg/1000 m^3 of oil) will be released by the various processes?

- What answers to the above questions have been proposed in actual PSD permit applications?

1.1.3 Approach

In preparing this report, the authors have reviewed and summarized the pertinent literature. A comprehensive bibliography is included as Section 6 and key design manuals (such as the EPA's PCTMs) are extensively referenced and summarized rather than completely reproduced. Interviews with key industry and government agency personnel were held to gain their latest experiences, impressions of process performance, and their intentions with regard to future developments. An evaluation was then made of the emission factors for mining, retorting, and upgrading processes. A matrix is presented summarizing APC options for each unit process of certain selected shale oil production schemes. Each of the standard APC techniques identified in the matrix is synopsized while the newer and more innovative APC techniques are discussed more extensively. Performance and cost data are presented along with all known limitations.

Finally, as an example of specific case studies, the APC alternative for five shale oil production processes are presented along with their associated mass emission rates: (1) direct heated (e.g. Paraho); (2) circular grate (e.g. Superior, Allis-Chalmers, Dravo); (3) indirect heated (e.g. Union B); (4) recycled solids (e.g. Chevron, Lurgi); and (5) modified in-situ (e.g. Occidental).

1.2 OVERVIEW

A shale oil recovery facility consists of many varied unit operations. The raw shale is mined and then retorted to produce gas, oil and spent shale. The spent shale is conveyed to a solid waste disposal site. The retort gas may be treated to remove nitrogen and sulfur gases and then burned to produce steam and electricity. The recovered oil is treated to remove

sulfur and nitrogen and then stored. The primary sources for air emissions from these processes are shown in Table 1-1.

TABLE 1-1. PRINCIPAL AIR EMISSION SOURCES

Process	Principal Emissions*	Source
Mining	PM	Drilling, blasting, materials handling, crushing, vehicles
	CO	blasting, vehicles
	NO_x	Vehicles
Retort Gas Combustion	PM, NO_x, SO_x	Retort gas
Spent Shale Disposal	PM	Materials handling, erosion
Upgrade	HC	Storage

*PM = particulate matter
CO = carbon monoxide
NO_x = nitrogen oxides
SO_x = sulfur oxides
HC = hydrocarbon (volatile organic compounds)

The potential (i.e., uncontrolled) emissions from retort gas combustion are determined by the composition of the raw shale, the type of retort used to separate the oil and gas from the raw shale, and the retort gas combustion process. The retorting process determines the quantity of retort gas produced, its heating value, the concentration of sulfur and nitrogen gases. These factors (retort gas quantity, heating value, and concentration of sulfur and nitrogen gases) determine the applicability and effectiveness of the various air pollution control (APC) schemes that might be used.

The chemistry of the retort process, source and type of sulfur and nitrogen gases, a generic description of the types of retort processes developed and a discussion of eleven specific processes being considered for full scale application are described in Section 2.0.

Emissions from mining, spent shale disposal and upgrading operations are similar to other mining and oil refining operations. A technology review of all air pollution controls that are applicable to an oil shale recovery facility is presented in Section 3.0. Included are process descriptions, application experience or development status, and economics.

The developers who have proposed full-scale, oil shale recovery facilities have submitted Prevention of Significant Deterioration (PSD) Permit Applications. These PSD applications represent a significant data base of emission sources, controls and controlled air emissions. Information from these sources was computerized and sorted to determine typical emissions from each specific unit operation in a complete shale oil recovery facility. The results from this analysis are presented in Section 4.0.

In Section 5.0 the retort process information from Section 2.0, the air pollution control technology information in Section 3.0 and the unit operation emission data from Section 4.0 are used to evaluate five types of retort processes. For each process three scenarios are considered. The first scenario (base case) uses air pollution controls proposed in PSD permit applications. The reader should note that the PSD permit finally issued often called for more stringent controls than proposed in the original permit application. The second scenario (Alternate 1) uses either additional, improved or more appropriate control technology. The third scenario (Alternate 2) uses a further improvement in control technology where available. The major conclusion from these five analyses is that, although the base case emissions can vary for the individual process configurations by as much as one order of magnitude, the application of appropriate controls essentially results in air emissions being similar for each of the five processes considered.

1.3 RETORT PROCESS REVIEW

The type of retorting process determines the subsequent emissions of sulfur (SO_x) and nitrogen (NO_x) from the combustion of the retort gas. The following is a summary of the findings of this study.

Oil shale contains organic material, (kerogen and bitumen) associated with various minerals including sulfur [(organic and FeS_2) and nitrogen [organic and inorganic (in the form of buddingtonite -- an ammonium alumina silcate)]. To recover the organic content in the form of oil and gas, the raw shale is heated to 500 C to 800 C, decomposing the high molecular weight organics to oil and gas. This may leave a carbonaceous residue (char) on the retorted (spent) shale.

Two types of processes have been developed -- above-ground and in-situ. The above-ground process requires mining, crushing, screening and transfer of the raw shale to the retort and then removal and disposal of the spent shale. The true in-situ process provides heat to the oil shale directly by feeding oxygen to burn part of the organic content (the carbonaceous char). The hot combustion gases heat the remaining shale bed to decompose the kerogen and bitumens. The modified in-situ process (MIS) is a hybrid of the above-ground and in-situ processes. Approximately 25 percent of the shale is mined and retorted above-ground. The remaining shale is rubblized and retorted in place.

In the above-ground process, heat can be provided to the retort either directly or indirectly. The direct-heated retort is similar to an in-situ retort in that the heat is provided by feeding air (oxygen) to combust the carbonaceous char. The resulting hot combustion gases flow through the raw shale, heating it to the retorting temperature.

In the indirectly heated process, heat is provided to the retort by heating either a gas or a solid which is recycled to the retort. Since no combustion takes place within the retort, the retort gas is not diluted by combustion air and, consequently, lower volumes of high Btu content gas are produced.

1.3.1 Retort Gas

The emissions from the combustion of the retort gas are determined by the:

- volume and heating value of the retort gas
- presence of sulfur in the gas

- presence of nitrogen as ammonia or organic nitrogen compounds

The retort gas flow rates for the three basic types of retorts are shown in Table 1-2.

TABLE 1-2. RETORT GAS PRODUCTION RATES
(Taback, H. J., et al., 1986)

Type Retort	Gas Produced m^3 of gas/m^3 of Oil	Heating Value gram-calories/liter
In-situ	7000	9000
Direct combustion	1800	9000
Indirect combustion	180	90,000

The in-situ process produces the highest retort gas flow rate due to the combined effect of the higher retort temperatures converting more of the kerogen to gas and the higher dilution gas flow required to provide adequate oxygen to burn the shale. The direct combustion retort has similar conditions (i.e. high temperatures and requirement for adequate oxygen for combustion) but to a lesser degree than the in-situ retort and, consequently, has lower retort gas flow rates. The indirect combustion process has the lowest retort gas flow rate due to the lower retort temperatures and low gas flow rate with no reduction from combustion air.

The heating value of the retort gas is determined by the amount of combustion air. In-situ and direct combustion retorts produce low heating value gas at 9000 gram-calories/liter (100 Btu/scf) and the indirect combustion retort produces high heating value gas at 90,000 gram-calories/liter (1000 Btu/scf).

The retort gas flow and heating value determine the net exhaust gas flow after combustion. Since air pollution control performance is often determined by a pollutant exit concentration (ppm or g/m^3), high gas flow rates can result in comparatively higher pollutant emissions.

1.3.2 Sulfur Gases

During retorting, sulfur in the raw shale is partitioned to the spent shale (60 percent), oil (10 percent) and retort gas (30 percent). The significant variations in raw shale sulfur content, percentage of sulfur partitioned to the gas phase and the chemical structure of these sulfur gases result in the sulfur gas clean-up strategy being quite difficult.

Sulfur emissions can be reduced either by removing the sulfur prior to combustion or by adding a flue gas desulfurization process after combustion. Because the combustion process reduces the pollutant concentrations and increases the gas flow rate, the more cost-effective technique often is sulfur removal prior to combustion.

The form of the sulfur in the gas is extremely important when considering sulfur removal processes. Sulfur recovery processes that have been considered for cleaning the retort gas prior to combustion are not effective in removing the organic sulfur compounds, which can amount to as much as 16 percent of the total sulfur in the retort gas. Consequently, the effectiveness of these clean-up processes is dependent on the amount of organic sulfur relative to H_2S. Even high efficiencies of H_2S removal (99%) may not be sufficient to meet the Colorado standard on sulfur emissions below 850 kg/1000 m^3 of oil (0.3 lb/bbl).

To avoid the costly alternative of adding an "end of pipe" flue gas desulfurization, two alternatives can be considered. The first, the activated carbon and hypochlorite H_2S removal process, is an improvement on the H_2S scrubbing process, because it also removes organic sulfur species. Therefore, this process is effective for removal of the sulfur gases prior to combustion, and eliminates the need for more expensive post-combustion control. The activated-carbon and hypochlorite process can remove 99+% of H_2S and 90-98+% of the organic sulfur gases. This results in a net sulfur removal efficiency of 99% and a sulfur emissions (SO_x) of 500 kg/1000 m^3 oil (0.17 lb/bbl) even when the organic sulfur gases are 15% of the total sulfur.

The second alternative to the use of post-combustion SO_x control is the indirect combustion-solids recycle retort process, which limits sulfur gas

emissions by the chemistry of the retort and combustion process. The sulfur contained in the retort gases from the recycle solids process represent only one percent of the total sulfur content in the feed. [The remaining sulfur is partitioned to the oil (10%) and the spent shale (89%)]. Therefore, the amount of organic sulfur is minimal, and the H_2S removal processes alone are sufficient to reduce the sulfur emissions below the regulatory limit.

1.3.3 Nitrogen Gases

The removal of nitrogen gases also is difficult to predict due to the degree of variability in nitrogen content in shale, partitioning between gas, oil and spent shale and chemical form of the gaseous nitrogen species.

Using the data reported in the PSD permit applications and the EPA's Pollution Control Technical Manuals, the partitioning of nitrogen was estimated as shown in Table 1-3.

TABLE 1-3. PARTITIONING OF NITROGEN IN RETORT
(Taback, H.J., et al., 1986)

Process		% of Raw Shale Nitrogen in Product Spent Shale	Oil	Retort Gas
In-situ (MIS)		21	25	54
Solids Recycle	(Lurgi) (Chevron)	5	55	10*
Direct Combus'n	(Paraho)	33	37	30

* remaining nitrogen content in the spent shale after retorting is burned in the lift pipe or fluidized combustor and exits with the flue gas.

The nitrogen content of the retort gas consists primarily of ammonia with smaller amounts of other nitrogen compounds. In an investigation of nitrogen-containing species from an in-situ and above-ground retort process, hydrogen cyanide, various nitriles, pyridine, methyl and diethyl aniline and other nitrogen gas species were identified. The organic nitrogen content of the retort gas was found to be as much as 1 - 2% of the ammonia content.

The presence of organic nitrogen species presents the same problem for limiting fuel related NO_x emissions as that described above for the SO_x emissions; namely that the removal processes generally considered are not effective in reducing the organic nitrogen content of the retort gas.

The primary nitrogen removal technique considered is removal of ammonia from the retort gas by a water wash absorption tower followed by an ammonia recovery stripper. The outlet ammonia concentration is determined by the effectiveness of the ammonia absorber. At atmospheric pressure, the equilibrium exit partial pressure for ammonia at 50 C is 0.5 mm Hg.

The nitrogen content of the treated retort gas and the subsequent NO_x emissions from combustion of the retort gas is determined by:

- the exit gas ammonia concentration
- the amount of retort gas produced by the retort
- the amount of nitrogen partitioned to the retort gas
- the percentage of nitrogen present as organic nitrogen compounds

The processes that use the combustion of high-nitrogen-content spent shale for energy recovery can produce high NO_x emissions if proper staging of the combustion is not used. The PSD permit applications NO_x emissions were based on a high estimate of 15% for the conversion of the nitrogen in the spent shale to NO_x in the combustor. This level of conversion also was determined by Lawrence Berkeley Laboratory investigators who did not attempt to stage the combustion.

The principle of NO reduction in a staged combustor can be applied to reduce these excessive NO_x emissions to approximately 3% nitrogen conversion to NO_x.

1.3.4 Specific Process Review

A description of the retort processes listed in Table 1-4 also is provided in Section 2.0.

TABLE 1-4. RETORT PROCESSES REVIEWED

Developer/ Process Name	Flow Pattern	Heat	Fines	Spent Shale Comb.	Comment
Paraho	Counter-current, gas up, solids down	Direct	No	Partial	
Hytort	Counter-current, gas up, solids down	Indirect	No	No	Hydrogenation
T^3	Batch	Direct	No	Yes	Two vessels
Lurgi	Solids recycle mixer	Indirect	Yes	Yes	
Tosco II	Solids mixer	Indirect	Yes	No	Ceramic balls provide heat
Union A	Counter-current gas down, solids up	Direct/	No	Yes	
Union B		Indirect	No	No	
Union SGR		Indirect	Yes	Yes	
Union C		Indirect	Yes	Yes	
Superior, Dravo, Allis-Chalmers	Travelling grate, cross-flow	Direct/ indirect	No	Yes	
Chevron	Fluidized bed, recycle solids	Indirect	Yes	Yes	
Lawrence Livermore	Cascading bed,	Indirect	No	Yes	
Modified In-Situ	Below-ground, semi-batch cross-flow above-ground any of above	Direct	NA	Yes	
In-situ	Semi-batch	Direct	NA	Yes	

1.4 AIR POLLUTION CONTROL TECHNOLOGY REVIEW

Descriptions of the air pollution control technologies that can be applied to oil shale facilities are provided in Section 3.0. The technologies presented are listed in Table 1-5. The technologies indicated by an asterisk have been recently developed and have potential application for oil shale processing. For these technologies, the descriptions are extensive, providing as much design information as available. The remaining technologies have been previously presented extensively in the oil shale literature, and the descriptions are meant more as an overview than a detailed analysis.

TABLE 1-5. AIR POLLUTION CONTROL TECHNOLOGIES

Pollutant		Control Technology
Particulate (Point Sources)		Baghouse
		Venturi scrubber
		Electrostatic precipitator
	*	Dry Venturi
(Fugitive Sources)	*	Surfactants
	*	Liners
	*	Wind screens
	*	Chemicals
	*	Water spray
Nitrogen Oxides	*	Staged combustion
	*	Ammonia injection
		Selective catalytic reduction
Sulfur Oxides		Stretford
		Lo-Cat
		Unisulf
		Alkaline scrubber
	*	Activated carbon and hypochlorite
		Claus
		Scot (Shell-Claus Offgas Treating)
		Flue Gas desulfurization (wet & dry)
	*	SO_x Absorption on spent shale
CO and HC	*	Catalytic mufflers

*Indicates systems given greater emphasis in this report because they are not covered in other shale oil documents.

1.4.1 Dry Venturi

The dry venturi scrubber is similar to a wet venturi scrubber except that the dry venturi is based on the principle of providing the targets necessary for sub-micron particle removal as a dry solid particle rather than as a liquid droplet. In a standard wet venturi, particles are captured by inertial impact; the liquid targets are created within the venturi by transferring velocity energy into shearing action.

The capture of the particles in the dry venturi is only the first step in the removal processes. The particles still have to be removed from the gas stream. In a completely dry process, i.e., no downstream processing after the venturi, a solids capture device (baghouse or ESP) is required to remove the solid particles [captured particulate and targets] from the gas stream.

1.4.2 Fugitive Dust

Oil shale recovery that uses above-ground retorting requires extensive mining and spent shale disposal. These mining and disposal operations require storage piles for raw shale feed. They also include soil removal and storage, spent shale hauling, and disposal sites for spent shale and raw shale fines. These processes are standard mining operations and the fugitive dust emissions and control techniques have been defined in detail. To provide the reader with access to this information, it is presented in Section 3.1.2.

1.4.3 Staged Combustion

The high nitrogen content of the retort gas and the spent shale result in high NO_x emissions when either is burned. Staged combustion is the control technology that converts the fuel nitrogen to N_2 rather than NO and maintains a flame temperature low enough to prevent the reaction of the oxygen and nitrogen in combustion air. Staging the combustion (fuel-rich zones followed by fuel-lean zones) can eliminate the need for end-of-pipe controls and essentially requires a comprehensive design and prototype testing program to determine optimum operating conditions. This control technology is discussed in Section 3.2.1.

1.4.4 Activated Carbon and Hypochlorite

As discussed above, the presence of organic sulfur compounds decreases the effectiveness of sulfur removal processes to the point where additional end-of-pipe controls (flue gas desulfurization) is required. In processes like Stretford, Lo-Cat, etc., the H_2S is scrubbed but the reduced organic sulfur compounds are not. The activated-carbon and hypochlorite process removes the organic sulfur species as well as the H_2S, which eliminates the need for further controls. These systems have been installed in the pulp and paper industry on black liquor recovery boilers, but there is no operating experience on oil shale retort gas. A description including process performance is presented in Section 3.3.5.

1.4.5 SO_x Absorption on Spent Shale

The energy content of the carbonaceous residue on the spent shale (char) represents approximately four percent of the total recoverable energy, and most developers are considering recovery of this energy by combustion. In addition, the presence of carbonates in the spent shale results in the absorption of any SO_2 present on the spent shale. Consequently, the retort gas and the spent shale can be burned together in a fluidized or cascading bed combustor with minimal SO_x emissions (10-20 ppm exit concentration). This process is discussed in Section 3.3.9.

1.4.6 Catalytic Mufflers

A significant source of carbon monoxide (CO) and hydrocarbon (HC) emissions from a complete facility are the vehicle and engine emissions associated with below-ground mining and above-ground transportation, grading, hauling, etc. Catalytic converters have been developed to reduce these CO and HC emissions by as much as 90 percent. This catalytic conversion process is discussed in Section 3.4.

1.5 PREVENTION OF SIGNIFICANT DETERIORIATION (PSD) PERMIT APPLICATIONS

Seven developers have submitted PSD Permit Applications to state agencies in Utah and Colorado with extensive information on emission sources, controls and permitted emission rates. To evaluate controlled emission rates,

particularly from operations other than retort gas combustion, this PSD information was computerized and sorted to determine average emissions and relative percentages of emissions for each unit process. The seven PSD permits analyzed are listed in Table 1-6.

TABLE 1-6. PSD PERMIT APPLICATIONS IN DATA BASE

Project	Location	bbl/ day	Oil Product. m^3/Day	Retort Process
Cathedral Bluffs	Rio Blanco County, CO	12,000	1,900	Modified in-situ with Union above-ground retort
Clear Creek	Grand Valley, CO	100,000	15,900	Chevron - fluidized bed with solids recycle
Utah Cottonwood Wash	Green River Basin Basin, UT	31,500	5,000	T^3 retort with fluidized bed combustion of retort gas & fines
Paraho - Ute	Uinta Basin, UT	42,000	6,700	Paraho/direct heated
Syntana	Uinta County, UT	57,200	9,100	Superior - retort indirect heat with Tosco II retort for fines
Union Facility Phase I-Union B	Parachute Creek, CO	10,000	1,590	Indirect combustion gas recycle
White River Project	Vernal, UT	106,000	16,900	Superior-direct heated Union B-indirect heated Tosco II-fines retort

1.5.1 Total Facility Emissions

The total facility emission for the five criteria pollutants are shown in Figure 1-1. There is wide variation in these emission rates directly dependent on the type of retort process used.

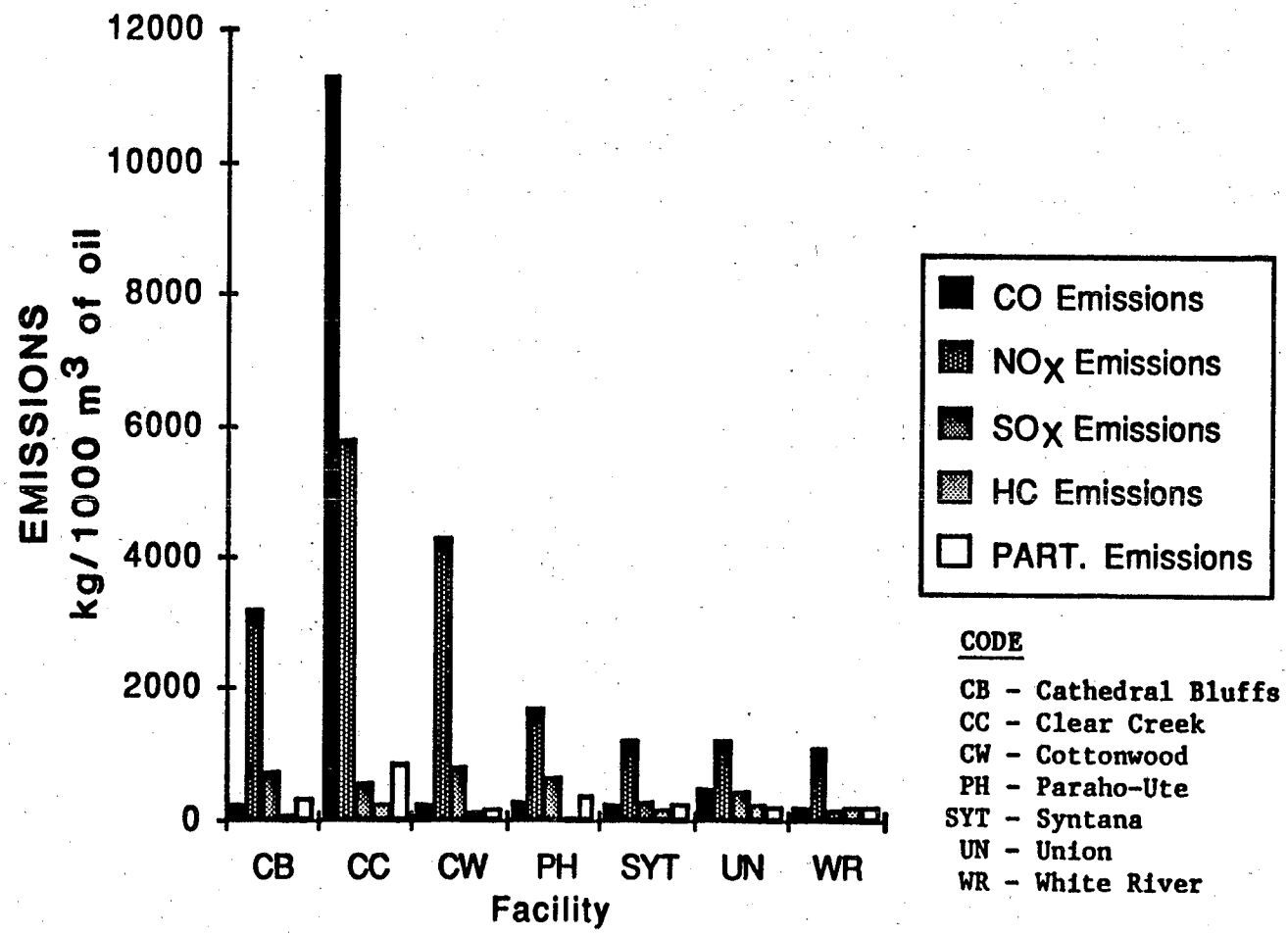


Figure 1-1. Total facility emissions summary (Taback, H.J., et al., 1986).

1.5.2 Individual Unit Operations

The data provided in the PSD permit applications were evaluated to determine typical emission rates for the various unit operations. The emission rates from individual sources in the various facilities for the criteria pollutants are shown in Figures 1-2 through 1-6. For this analysis, all emissions associated with combustion of the retort gas (i.e. upgrade heater, retort heater) are considered as part of the emissions from the retort operation. In actual practice, the retort gas may not be burned on site depending on the price and availability of alternate fuels (oil or coal) or the use of a high Btu retort gas as a source of hydrogen for the upgrading process.

The carbon monoxide emission sources (Figure 1-2) are blasting, below ground vehicles, above-ground vehicles, and the combustion of retort gas in the retort and upgrading process. The hydrocarbon emission sources (Figure 1-3) are primarily mining vehicles, oil storage and fugitive emissions in the upgrading and retort gas combustion. Nitrogen oxide emission sources (Figure 1-4) are primarily from retort gas combustion and mining vehicles. The only significant sulfur oxide emissions source (Figure 1-5) is combustion of the retort gas, with mining and upgrading adding a relatively small amount.

The reader should note that nitrogen and sulfur oxide emission estimates provided by the developers in their PSD permit applications were based on assumptions with regard to the percentage of organic sulfur and nitrogen gases. Consequently, the actual emission rates could be much higher than the estimates presented in the PSD applications.

The particulate emission sources (Figure 1-6) are those associated with below-ground mining (drilling, blasting, conveying, crushing, engines), above-ground mining (surface soils removal, second and third degree crushing, conveying, storage and spent shale disposal), retort gas combustion (steam generator, retort heater and upgrade heaters) and above ground vehicles and fugitive emissions from truck traffic.

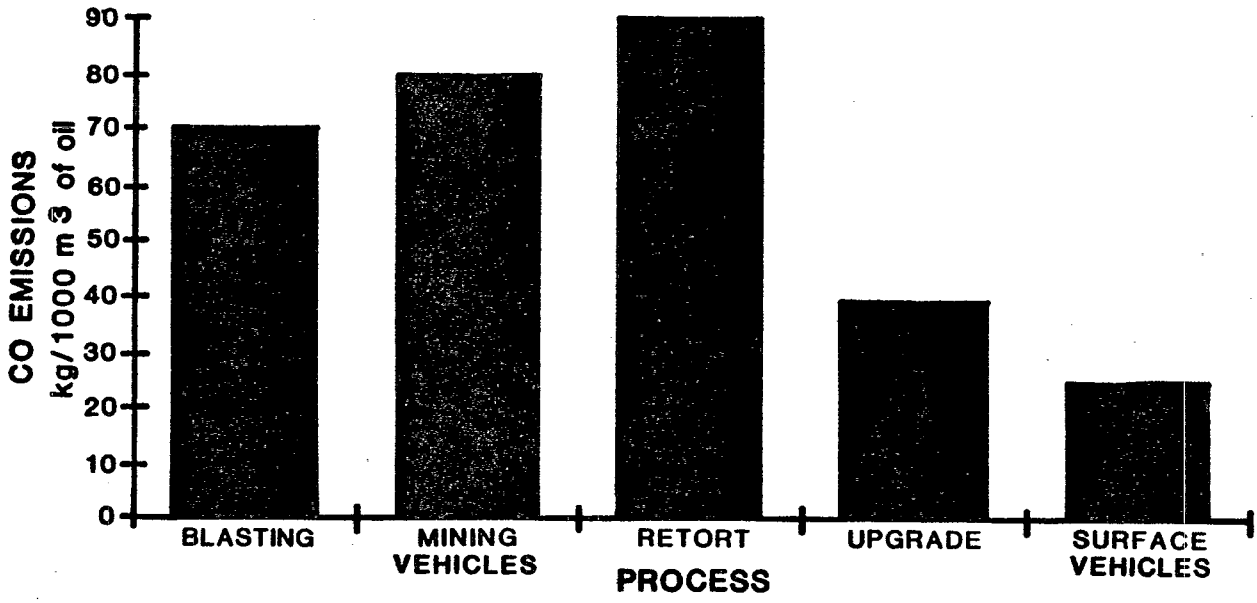


Figure 1-2. Carbon monoxide emissions (Taback, H.J., et al., 1986).

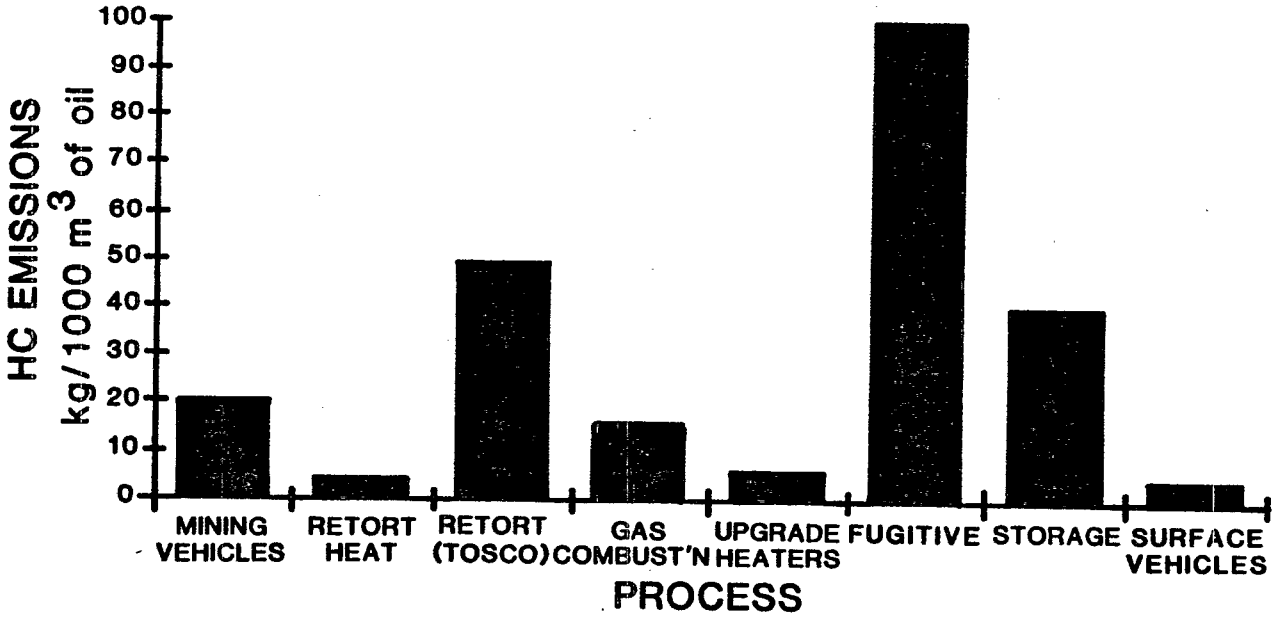


Figure 1-3. Hydrocarbon emissions (Taback, H.J., et al., 1986).

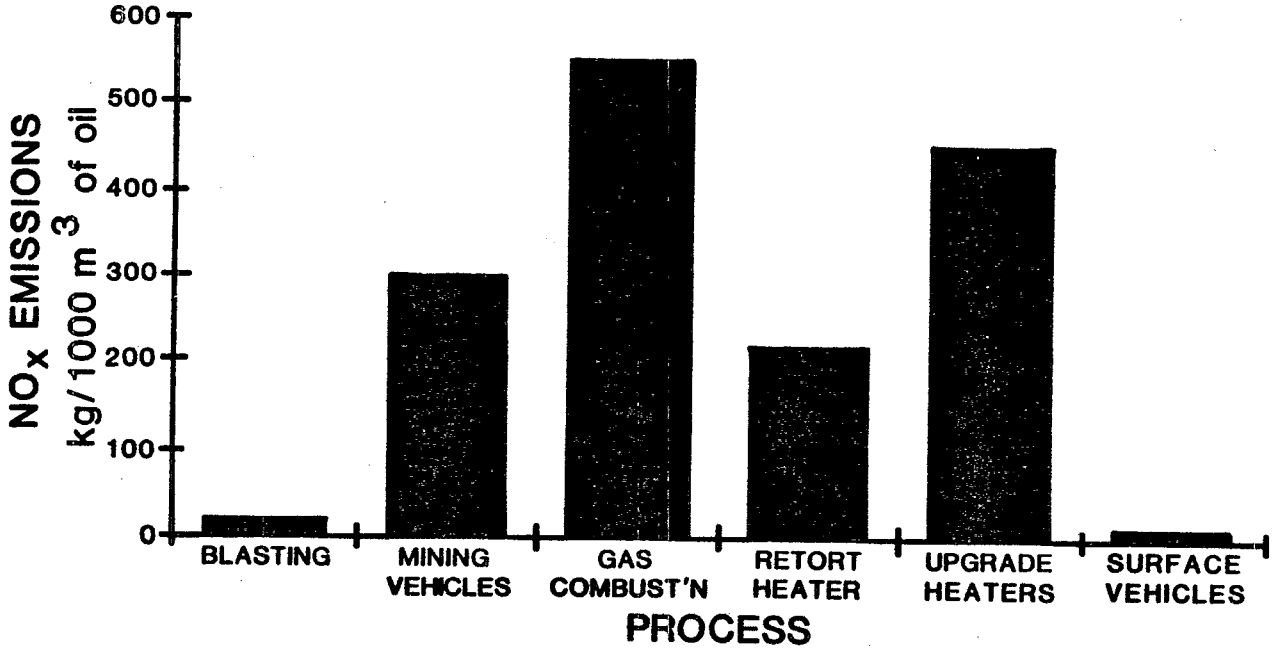


Figure 1-4. Nitrogen oxides emissions (Taback, H.J., et al., 1986).

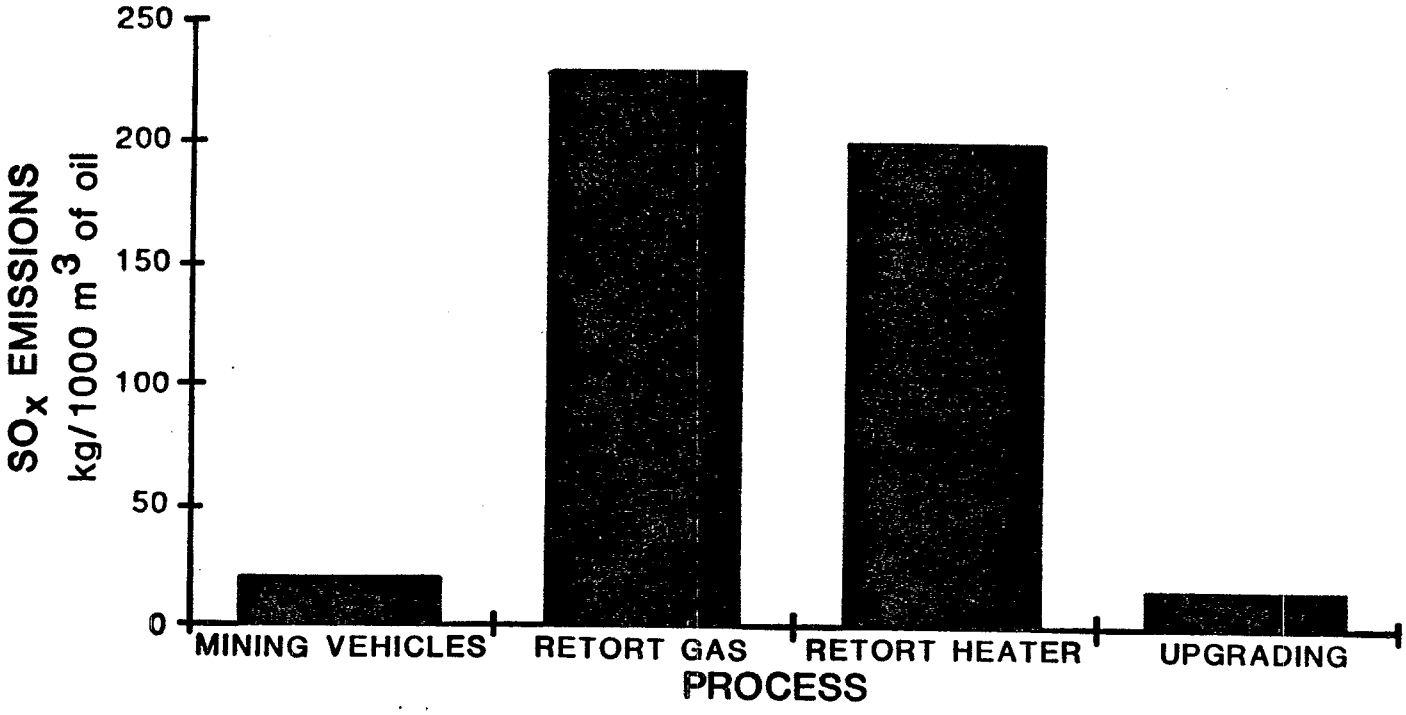


Figure 1-5. Sulfur oxides emissions (Taback, H.J., et al., 1986).

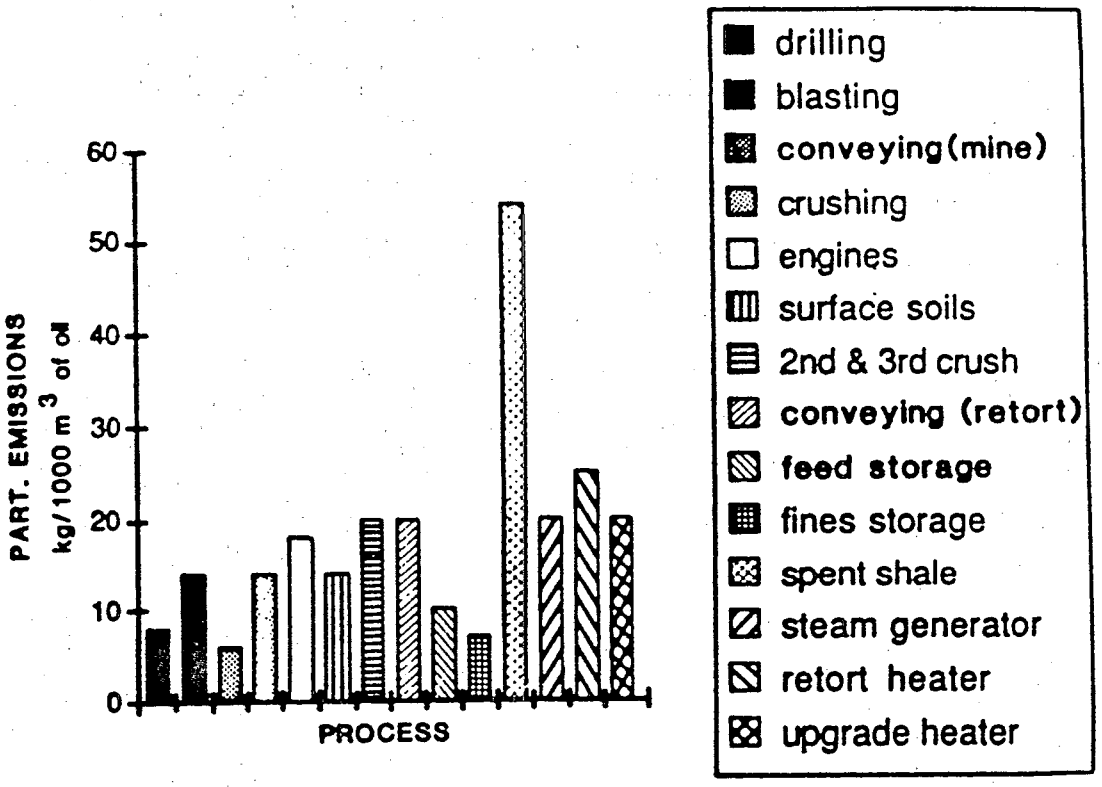


Figure 1-6. Particulate emissions (Taback, H.J., et al., 1986).

1.6 RETORT PROCESS EVALUATION

Development of oil shale production processes have led to a variety of retort designs. The total facility emissions as reported in the PSD permit applications for seven potential shale oil recovery plants were shown in Figure 1-1. The degree of variation in total facility emissions is considerable with complicating tradeoffs. For example, the Clear Creek facility with a Chevron solids recycle retort has very high NO_x and CO emissions. Conversely, the Union facility with a gas recycle retort has much lower CO and NO_x emissions but higher SO_x emissions.

The purpose of the analysis in Section 5.0 is to evaluate these various processing schemes and determine the effect of improved air pollution controls. The information developed indicates that, with the proper selection of air pollution control techniques, the air emissions for each of these processes can be held to essentially equivalent values.

1.6.1 Shale Oil Rcovery Plant

A shale oil recovery plant is quite complex, involving many varied operations. The unit operations required to recover the oil from the shale include:

- mining
  - below-ground
  - above-ground
- retorting
- product recovery
- removal of nitrogen (ammonia and organic nitrogen gases) from the retort gas
- removal of sulfur (hydrogen sulfide and organic sulfur gases) from the retort gas
- gas utilization (retort gas combustion)
- end of pipe controls
- upgrading
- spent shale disposal

Within each of these unit operations, there can be a number of process alternatives. The shale can be mined in an open pit mine, a room and pillar mine or the oil recovered without mining with an in-situ process. The retort heat can be provided by combustion of the spent char within the retort or with a recycled stream, which can be either gas or solid. Each of these variations can affect the emission rates. Consequently, consideration of all the potential processing schemes can be extensive.

The specific process combinations, analyzed in Section 5.0, are summarized below in Table 1-7.

TABLE 1-7. FIVE PROCESSES FOR ANALYSIS (Taback, H.J., et al., 1986)

Case #	Mining	Retort
1	open pit	direct combustion heat (e.g. Paraho)
2	room & pillar	direct combustion heat - circular grate (e.g. Superior, Dravo, Allis Chalmers)
3	room & pillar	indirect combustion gas recycle (e.g. Union)
4	room & pillar	indirect combustion solids recycle (e.g. Lurgi, Chevron)
5	modified in-situ	in-situ and indirect heat gas recycle above ground

For each analysis, a base case scenario representative of the process configuration proposed by the developers is presented and the criteria pollutant emissions are determined. Then two alternative processing schemes to reduce these emissions to their lowest levels are considered.

1.6.2 Methodology

For this analysis, the oil recovery facility was divided into three basic categories: mine, retort and upgrade. The emissions associated with mining and solids handling (primarily particulates, carbon monoxide and

nitrogen oxides) are similar to other mining operations. The data provided in the PSD permit applications was used to develop emission rates for each type of unit operations (i.e. blasting, drilling, vehicles, conveying, crushing, etc.). The emissions associated with the upgrading process (hydrocarbons from storage and fugitive sources), other than those from the combustion of the fuel gas, are similar to other oil refining operations and, again, the PSD permit applications were used to develop estimated emission levels.

The combustion of the retort gas can be the principal source of emissions from a facility and the source most affected by the particular retort process and gas clean-up scheme used.

The primary concerns of this analysis are emissions of nitrogen and sulfur oxides and particulates. Emissions of carbon monoxide and hydrocarbons generally are consistent for all processes with a few exceptions. The combustion of the spent shale can produce very high carbon monoxide emissions, and this will be discussed below for that particular process. High hydrocarbon emission rates can result from certain types of retort processes (i.e., Tosco II) that involve direct contact heating of raw shale or a heat carrier with flue gas. However, this process was not included in this analysis.

To evaluate the many process variations and develop the data necessary to estimate emission levels, it was first necessary to determine what process combinations are feasible. The basic design parameters that affect the pollutant emissions for each process were defined. These design parameters are shown in Table 1-8.

For each of the unit operations, the design parameters were applied as indicated by either the retort process conditions, performance of the pollution control equipment or the reported emissions from the PSD applications. The following discussion presents the rationale for choosing the specific design parameters used in the analysis.

TABLE 1-8. DESIGN PARAMETERS USED IN ANALYSIS
(Taback, H.J., et al., 1986)

Unit Operation	Design Parameter
Mining	Type of Mining open pit room and pillar in-situ
Retort	• Retort Gas Produced, m^3/m^3 of oil • Heating Value of Retort Gas, kJ/m^3 • Partitioning of sulfur and nitrogen
Product Recovery	• None- The product recovery process has no significant effect on emissions
NH_3 Removal	• NH_3 exit concentration, ppm • Organic nitrogen content of retort gas
H_2S Removal	• H_2S exit concentration, ppm • Organic sulfur content of retort gas • Organic sulfur gas removal efficiency %
Gas Utilization	• Boiler-dilution ratio (dry gas/fuel) • Spent shale combustion exit concentrations for NO_x, SO_x & CO
End of Pipe Controls	Particulate - Baghouse - exit loading, g/m^3 • Sulfur - FGD - exit SO_x, ppm • Nitrogen Ammonia injection - exit NO_x Staged Combustion - exit NO_x & CO

1.6.3 Mining, Solids Handling and Upgrading Emissions

The values from the PSD permit applications presented for mining and upgrading emissions were used in the overall facility emission estimates. The emissions from retort gas combustion were calculated based on retort and controls used.

The choice of mining technique determines the emission rates. The values used were developed from the PSD analysis for room and pillar mining and from the literature for open pit mining.

1.6.4 Retort Gas Emissions

The emissions from the combustion of the retort gas are determined by the:

- volume and heating value of the retort gas
- presence of sulfur in the gas
- presence of nitrogen as ammonia or organic nitrogen compounds

The retort gas flow rates for three types of retorts were shown in Table 1-2.

The design conditions used for the H_2S removal process determines the residual H_2S and organic sulfur in the retort gas that eventually is emitted as sulfur oxides. The processes considered are:

1. Direct or indirect conversion of the sulfur (e.g. Stretford, Lo-Cat, Unisulf, alkaline or amine scrubbing)
   - H_2S exit concentration = 50 ppm
   - organic sulfur assumed at 5% of total sulfur in retort gas - no removal
2. activated carbon-hypochlorite process
   - H_2S exit concentration = 10 ppm
   - organic sulfur - 90% removal

Two process operations result in the direct emission of sulfur oxides from the retort; the circular grate direct heated retort and the fluidized bed combustion of the spent shale. The design conditions for these two processes were taken from the literature.

Circular Grate Retort - SO_x = 175 ppm in retort gas (White River Shale Project, 1981)

Fluidized Bed Spent Shale Combustor Flue Gas - SO_x = 20 ppm (Chevron Shale Oil Co., 1982)

The primary nitrogen removal technique considered is removal of ammonia from the retort gas by a water wash absorption tower followed by an ammonia recovery stripper. The outlet ammonia concentration is determined by the effectiveness of the ammonia absorber. At atmospheric pressure the equilibrium exit partial pressure for ammonia at 50 C is 0.5 mm Hg.

The nitrogen content of the treated retort gas and the subsequent NO_x emissions from combustion of the retort gas is determined by:

- the exit gas ammonia concentration (660 ppm)
- the amount of retort gas produced by the retort
- the amount of nitrogen partitioned to the retort gas
- the percentage of nitrogen present as organic nitrogen compounds

The processes that utilize the combustion of high-nitrogen-content spent shale for energy recovery can produce high NO_x emissions if proper staging of the combustion is not used. The PSD permit applications NO_x emissions were based on a high estimate of 15% for the conversion of the nitrogen in the spent shale to NO_x in the combustor. This level of conversion also was determined by Lawrence Livermore Laboratory investigators (Taylor, R. W., et al., 1985) who did not attempt to stage the combustion.

The principle of NO reduction in a staged combustor can be applied to reduce these excessive NO_x emissions to approximately 3% nitrogen conversion to NO_x.

1.6.5 Retort Gas Combustion and End-of-Pipe

The End-Of-Pipe controls are those either added after combustion of the retort gas to remove particulate, NO_x and SO_x or incorporated as part of the combustion process as in staged combustion for NO_x control.

A. Particulate--

For particulate control, two alternatives were considered. The first is the base case using a standard baghouse. The second control technique is the combined dry venturi-baghouse.

The dry venturi-baghouse combination provides for particulate control that is somewhat independent of type of particulate. The applicants for PSD permits all considered a minimum particulate exit loading of 0.07 g/m^3 (0.03 gr/scf) which was based on standard technology within the limits of the unknowns associated with oil shale particulate. By capturing the small particles on larger target particles of specified physical properties, the dry venturi eliminates the major uncertainties in designing baghouses with respect to particulate type and size.

B. Sulfur Oxides--

If the H_2S (and organic sulfur) removal is not sufficient to reduce the sulfur emissions to an acceptable level, a post combustion flue gas desulfurization system must be added. This could be either a wet or dry scrubber.

The second sulfur control technique is the use of a spent shale combustor. The combustion of the spent shale has two important advantages:

1. recovery of the energy value of the char, and
2. reduction of the sulfur oxide emissions due to the scrubbing nature of the spent shale.

However, the spent shale combustion also has two distinct disadvantages:

1. high emissions of NO_x from the nitrogen in the spent shale, and
2. high emissions of CO due to incomplete combustion.

These emissions (NO_x and CO) will be discussed in the following section.

The design conditions for SO_x emissions used in the analysis are:

1. Flue Gas Desulfurization - 50 ppm SO_x exit concentration
2. Combustion of spent shale with retort gas
   - 10 ppm SO_x
   - 300 ppm NO_x
   - 1000 ppm CO

C. Nitrogen Oxides--

Two controls were considered for reducing post-combustion NO_x emissions. The first is ammonia injection; this technique has been applied successfully in utility boilers and could be used when the retort gas is burned in a conventional boiler for steam and electrical generation.

The second NO_x control considered is staged combustion which has particular advantages for spent shale combustion due to the ability to adequately control the fuel related NO_x. The staged combustion could be applied to either the conventional boiler or the spent shale combustor.

The use of selective catalytic reduction (SCR) was not considered due to the potential for poisoning the catalyst. The retort gas particulates contain a wide variety of heavy metals which have a deleterious effect on catalyst life. In addition, there are still a number of unknown factors which can effect the long term catalyst performance that have not been completely identified. For example, it had been assumed that the mercury associated with the retort gas was in the form of elemental mercury and would be substantially removed prior to combustion during the standard gas cleaning (ammonia and hydrogen sulfide removal) processes. However, it has been shown that the mercury may be present primarily as methyl mercury which is volatile and is present in the retort gas during combustion. Consequently, any post-combustion control processes must be capable of handling these emissions of elemental and oxides of mercury. This is only one instance where unknown factors could have a negative affect on catalyst performance. Consequently, due to the inherently variable nature of the retort gas from an oil shale retort and the known presence of many catalyst poisons, the use of SCR was not considered a viable alternative.

There is a tradeoff between the NO_x and CO emissions in the spent shale combustor. As indicated above, the NO_x and CO emissions are approximately 300 and 1000 ppm, respectively. Tests by J&A Associates (Van Zanten and Haas, 1986) indicate that the excess O_2 has a dramatic effect on NO_x and CO. As the O_2 level increases from one percent, the NO_x rises and the CO drops. Because shale combustors operate at low temperatures, approximately 700 C, the effects of temperature were found to be negligible.

In addition, the staged combustion technique for control of the fuel related NO_x, depends on low excess oxygen (actually sub-stoichiometric combustion) which further increases the CO emissions.

The fluidized-bed combustor has limitations in process control which result in the inability to provide conditions that result in adequate staging for NO_x control. However, the cascading-bed combustor is designed as a staged device. Consequently, combustion conditions can be controlled at each stage of the process, alternating between fuel rich and fuel lean zones to reduce the NO formed to N_2 and complete the combustion of the CO formed to CO_2.

There are no specific test results of the staged combustion with spent shale to support this hypothesis; however, the reaction kinetics of the reduction of NO to N_2 with spent shale have been investigated (Taylor, R. W., et al., 1985) and the engineering design of the cascading-bed combustor is ideal for a staged system with easy means for controlling the process conditions. Consequently, it is the opinion of the authors that proper design and development would result in minimal NO_x and CO emissions.

The design conditions for post-combustion NO_x control are:

1. NH_3 injection - 20 ppm NO_x exit concentration
2. cascading-bed combustor - 50 ppm NOx
   - 50 ppm CO

1.6.6 Emissions from Retort Gas Combustion

The emissions from combustion of the retort gas for particulates, nitrogen and sulfur oxides also are shown in Figures 1-7, 1-8, and 1-9.

These figures indicate that there is wide variation in emission levels for the five processes based on the technology proposed in the PSD permit applications (Base Case conditions). For particulates, the emission levels vary from 200 to 800 kg/1000 m^3 of oil; for nitrogen oxides the emission levels vary from 1000 to 8000 kg/1000 m^3 of oil; for sulfur oxides the emission levels vary from 350 to 3000 kg/1000 m^3 of oil.

The first alternative considered was the use of the activated carbon enhanced H_2S removal process, an acid wash for improved ammonia removal and

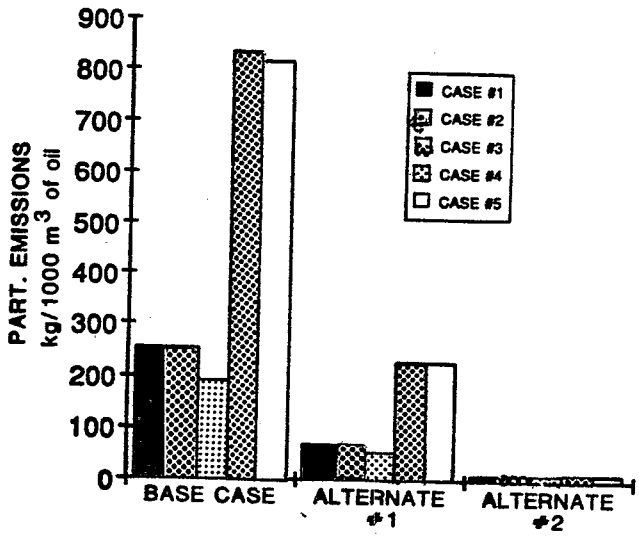


Figure 1-7. Particulate emissions from retort gas combustion for five cases (Taback, H.J., et al., 1986).

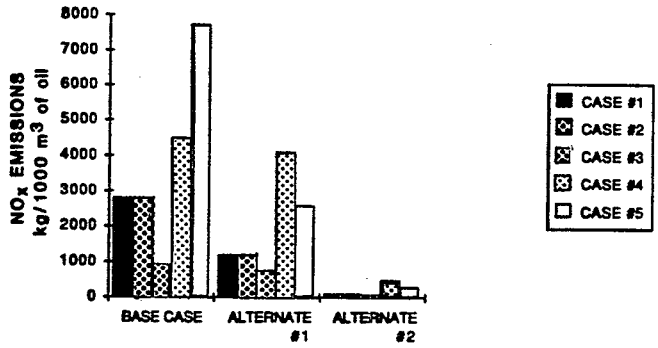


Figure 1-8. Nitrogen oxides emissions summary for five cases (Taback, H.J., et al., 1986).

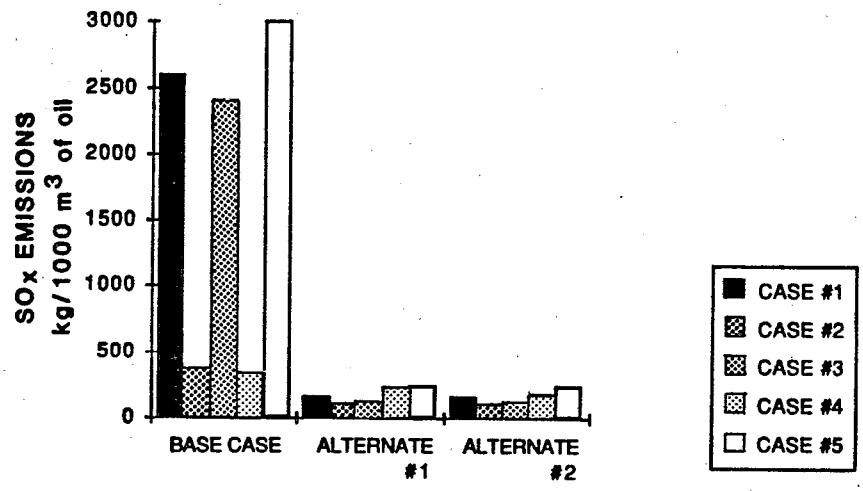


Figure 1-9. Sulfur oxides emissions summary for five cases (Taback, H.J., et al., 1986).

the addition of a dry venturi - baghouse for post combustion particulate control. Referring to Figures 1-7, 1-8, and 1-9, the emission levels for alternate #1 show considerably less variation, particularly for sulfur oxides (range from 100 to 250 kg/1000 m^3 of oil) and particulates (range from 50 to 200 kg/1000 m^3 of oil). The variation of nitrogen oxide emissions is still considerable, ranging from 1000 to 4000 kg/1000 m^3 of oil. Essentially, the acid wash only removes the residual ammonia without reducing the organic nitrogen content and has no effect on the thermal NO_x; therefore, there is relatively little improvement in the NO_x emission rate.

The second alternative considered was the use of ammonia injection for NO_x control from boiler and/or furnace combustion, the use of staged combustion for control of NO_x emissions from the spent shale combustor and the dry venturi-baghouse with an increased space velocity which improves collection performance at the expense of increased pressure drop. Again, referring to Figures 1-7, 1-8, and 1-9, it is apparent that the addition of these controls essentially levels the performance of all five processes.

The particulate, nitrogen oxide and sulfur oxide emission levels for alternate #2 conditions along with the total facility emissions are shown in Figures 1-10, 1-11, and 1-12. The particulate emissions (Figure 1-10) still show variation from 4 to 12 kg/1000 m^3 of oil. However, the absolute value is considerably less than the particulate emissions from the mining and solids handling operations and the total facility is essentially equivalent for all five cases ranging from 180 to 200 kg/1000 m^3 of oil.

The nitrogen oxide emissions, Figure 1-11, range from 75 to 500 kg/1000 m^3 of oil. While this is still a significant variation, again the absolute magnitude of the values is such that the net variation in the total NO_x emissions for the five facilities is less than 2 to 1 ranging from 400 to 800 kg/1000 m^3 of oil.

The sulfur oxide emissions (Figure 1-12) range from 100 to 250 kg/1000 m^3 of oil and are essentially the same for the total facility as there are no other significant sources of sulfur emissions.

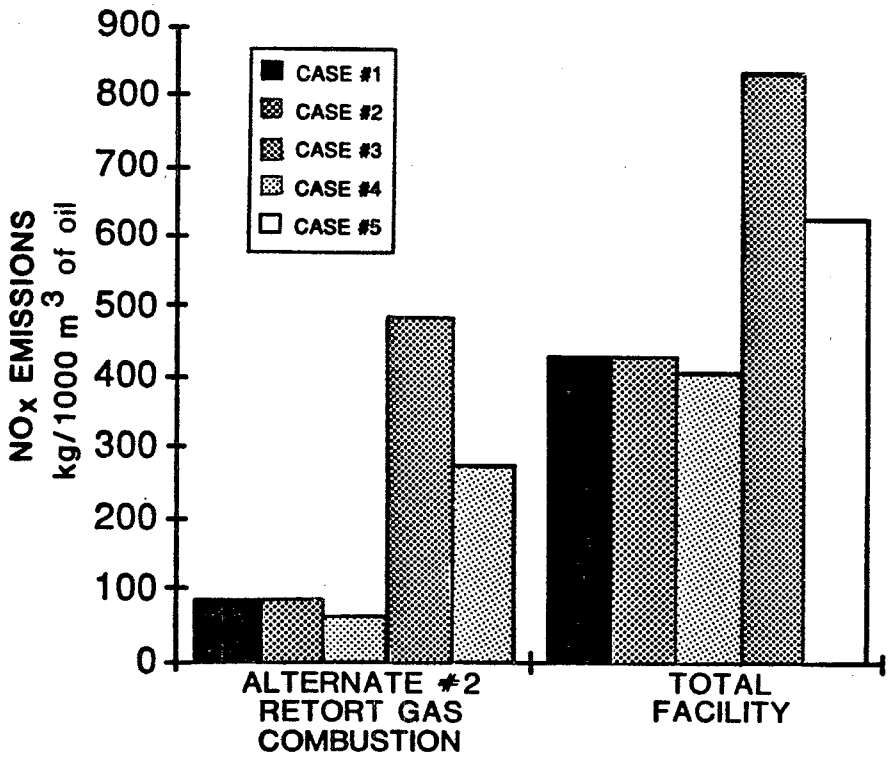


Figure 1-10. Nitrogen oxides emissions for five cases (Taback, H.J., et al., 1986).

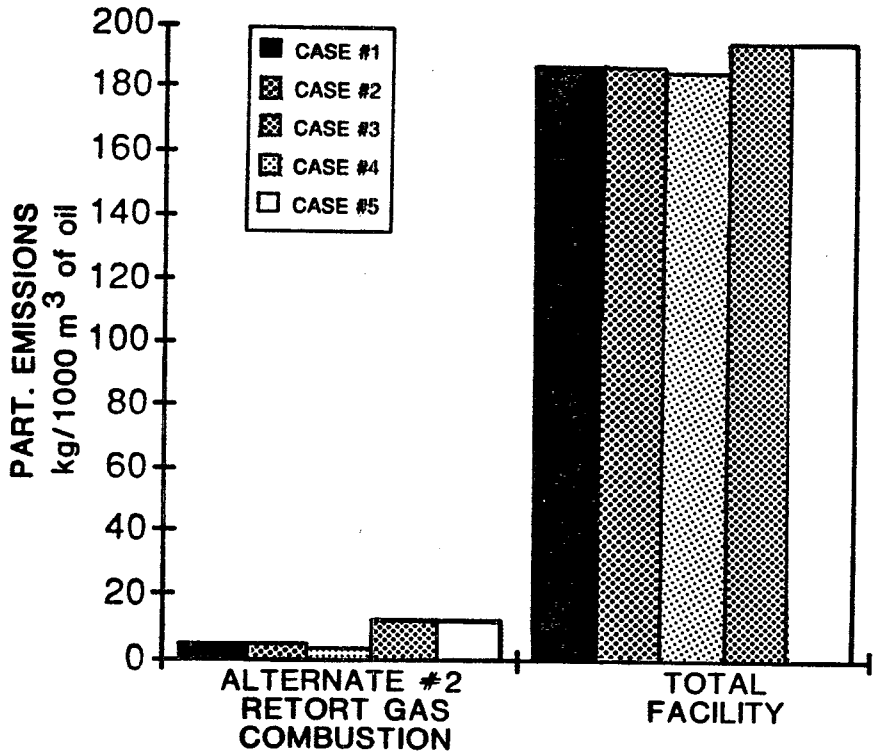


Figure 1-11. Particulate emissions for five cases (Taback, H.J., et al., 1986).

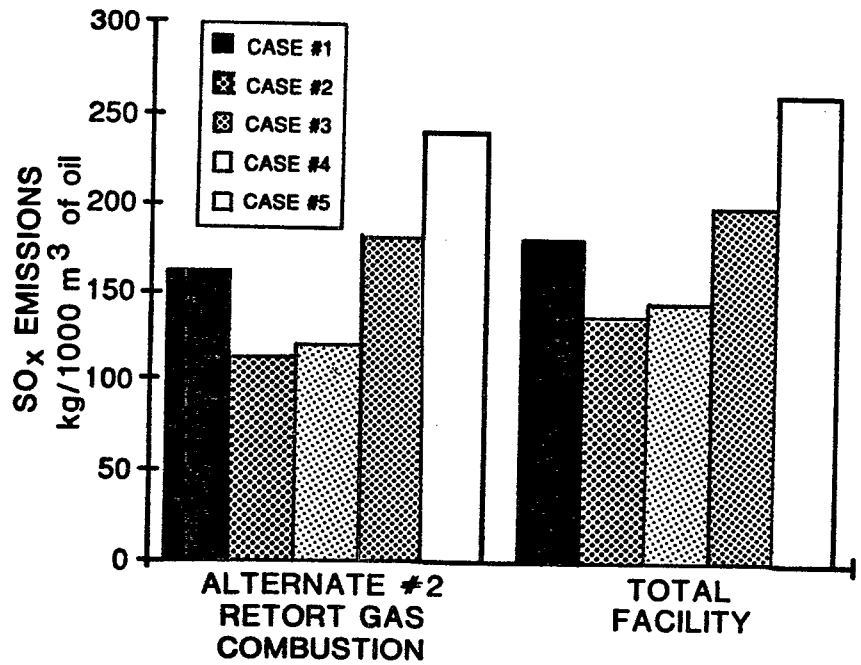


Figure 1-12. Sulfur oxides emissions for five cases (Taback, H.J., et al., 1986).

The basic conclusion derived from the above analysis is that, although the air emission levels for the different retort processes with controls proposed in PSD permit applications can vary considerably, sometimes by as much as two orders of magnitude, the application of control techniques that are either improvements over proposed technology or more suitable for a specific application, result in similar emission levels for all five processes considered. This statement needs to be qualified by the fact that some of the control techniques considered have not been applied specifically to the shale oil recovery process, and therefore might not be considered as BACT. However, these techniques have been proven at the full scale level in various other difficult control applications.

SECTION 2.0

RETORT PROCESS REVIEW

A complete shale oil processing facility consists of the following eight operations:

- mining,
- retorting,
- spent shale disposal,
- oil recovery,
- retort gas treatment,
- retort gas utilization,
- raw shale oil upgrading, and
- oil storage.

Although this report addresses the air pollution control techniques for all aspects of shale oil production, this section deals primarily with the retorting process, retort gas treatment, and retort gas utilization, since these are the areas where the type of retorting process can have a significant effect on the air emissions. The mining and spent shale disposal emissions are similar to those from other solids handling operations; the upgrading and storage facilities are similar to other oil refining operations.

The shale retort converts the raw shale solids to crude shale oil, retort gas, and spent shale. There are many different retorting processes ranging in size and complexity from an in-situ retort where the retort operation takes place within the shale formation with subsequent limitations on process control and operation, to the above-ground cascading-bed retort-combustor, which has considerable advantages in both process control and reduced emission rates, but requires mining of large amounts of raw shale. Schematics for the two general types of retort processes, underground (in-situ) and above-ground are shown in Figures 2-1 and 2-2, respectively.

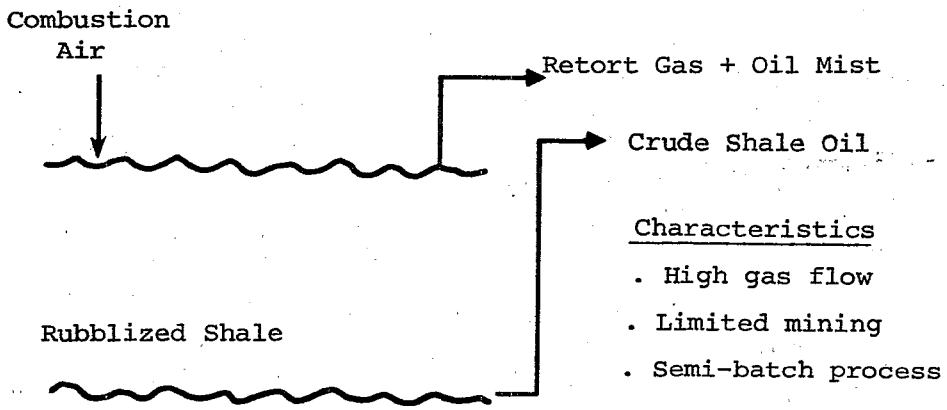
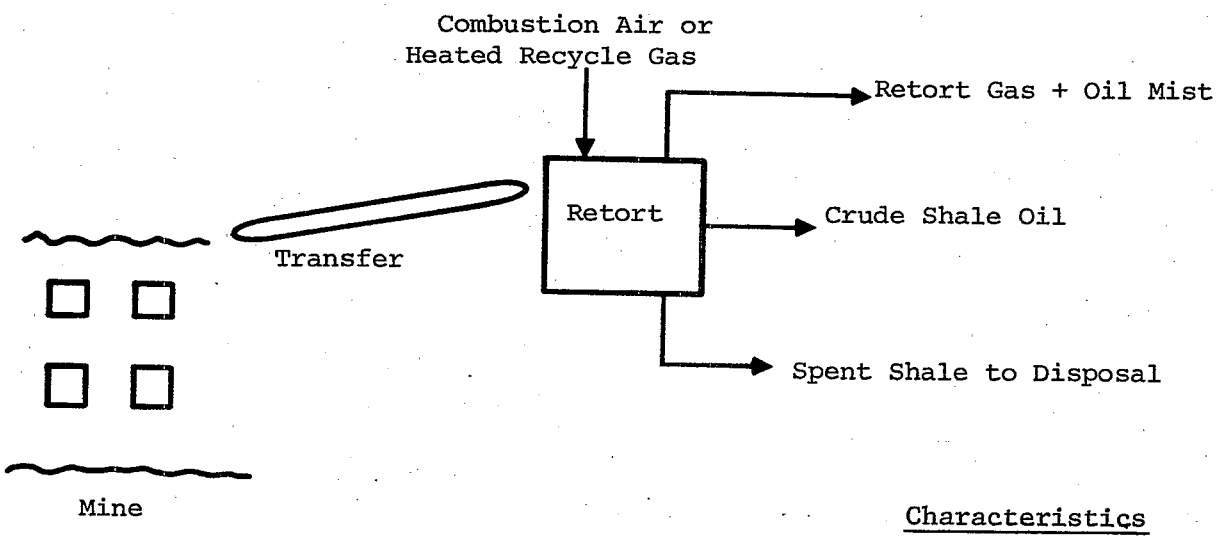


Figure 2-1. In-situ retort processes - general.



Characteristics

- Requires mining
- Lower gas flows
- Higher product recovery
- Continuous process

Figure 2-2. Above-ground retort processes - general.

The type of retort process can have a considerable effect on air emissions. These process differences can limit the sulfur emissions due to direct contact of the retort gas with oxidized spent shale; they can affect the relative amounts of organic sulfur to H_2S by varying steam concentration, or they can affect the quantities of gas and the concentrations of pollutants in that gas. Many of these different processes and their respective effects on the air pollution control strategy are described in this section.

2.1 BACKGROUND

Recovery of oil from oil shale involves the process of heating the raw shale to approximately 500 C to break down the organic content into its oil and gas components. This process is referred to as "retorting". Differences in processes involve variations in the techniques of transferring of heat to the raw shale. This background section presents information on various process configurations with particular emphasis on their respective pollution sources and emissions of criteria pollutants.

In direct-heated retorts, heat is provided to the shale by combustion within the retort to produce shale oil, retort gas, and spent shale. Then, the produced gases are scrubbed to remove ammonia and sulfur compounds. The gaseous emissions of SO_x and NO_x are determined primarily by the performance of the clean-up processes. The chemistry of the individual process and the method of heating the shale affect the generation of H_2S, organic sulfur, NH_3, and organic nitrogen in the retort and the resulting concentrations of these pollutants in the gas stream. These conditions significantly affect the gas clean-up process selected. The discussion in Section 4.0, PSD Permit Application Analysis, and Section 5.0, Process Analysis, reviews and evaluates these emission control design considerations.

In indirect heated retorts, heat is applied to a recycle stream of either gas or solids outside the retort. When gas is used as the recycle heating medium, the main effect is to eliminate the presence of combustion air within the retort which results in a lower retort gas volume of high Btu content gas.

When recycled solids are the heating medium, the solids are heated by combustion of the char which, in addition to recovering the energy value of the char and minimizing the SO_x emissions (Section 2.1.1.B), also oxidizes FeS to Fe_2O_3. The recycle of Fe_2O_3 to the retort has the effect of reducing the formation of H_2S within the retort, thereby reducing the necessary controls for limiting SO_x emissions.

The following section describes the major chemical reactions occurring in the retort and spent shale combustor.

2.1.1 Chemistry

The following discussion presents information on the specific chemical reactions that pertain to the generation and transformation of criteria pollutant emissions.

A. Retort--

Oil shale consists primarily of varying amounts of organic material (kerogen and bitumen), dolomite, calcite, quartz, analcime, feldspars, illite, Mg-siderite, pyrite (inorganic sulfur), and nitrogen (organic and inorganic) as well as trace amounts of many elements. During the retorting process, the organic material decomposes to oil, hydrocarbon (HC) gases (including hydrogen and CO), water, carbon dioxide, hydrogen sulfide, ammonia, a carbonaceous residue (char) and numerous minor components. Some of the major reactions occurring in the retort at the typical retort temperature of 500 C are:

1. Kerogen and Bitumen ---> HC(gas) + oil(liquid) + H_2S(gas) + organic S (gas) + CO_2 (gas) + H_2O (gas) + char (solid)
2. $FeS_2 + H_2$ ---> FeS + H_2S (gas) + oil S (liquid)
3. Organic S ---> H_2S (gas) + organic S (gas) + oil S + char S (solid)
4. $2H_2S + Fe_2O_3 + H_2$ ---> $2FeS + 3H_2O$
5. Buddingtonite ($NH_4AlSI_3O_8$) ---> NH_3 (gas) + organic N (gas)
6. Organic N ---> NH_3 (gas) + organic N (gas)

Reaction 1 is the primary reaction occurring in the retort and is the pyrolysis of kerogen and bitumen to form C_1-C_7 hydrocarbons which comprise the oil and gas. The thermal efficiency of the processes (i.e., the relative amount of oil and gas produced per unit of energy input) can vary considerably. However, most of the candidate processes for commercial application have process efficiencies of approximately 78 percent in the form of oil and gas with an additional 4 percent contained in char remaining with the spent shale.

In reactions 2 and 3 sulfur gases (H_2S and organic sulfur gases) are formed from inorganic sulfur (FeS_2) and organic sulfur material in the raw shale. Since FeS is stable under retort conditions, as much as 50-60 percent of the inorganic sulfur remains with the retorted shale.

In the direct heated and indirect heated with gas recycle processes, i.e., without external combustion and recycle of hot solids, organic sulfur gases are produced with the H_2S significantly complicating gas clean-up. In indirect heated solids recycle processes, recycled shale is present, which contains iron oxides and, as Reaction 4 indicates, some H_2S is removed from the gas phase by reaction with iron oxides to form the stable FeS. A major concern of air pollution control process design is kinetic and thermodynamic requirements to maximize the H_2S removal.

In reactions 5 and 6 nitrogen-containing gases, primarily NH_3 with other organic nitrogen gases, are produced from inorganic nitrogen (buddingtonite) and organic nitrogen in the shale.

The gas formed in the retort is a mixture of the gaseous components listed in Reaction 1 to 6 plus N_2, CO, and H_2O. Typical retort gas compositions are shown in Table 2-1. A breakdown of the sulfur gases (H_2S, thiophenes, carbonyl sulfide, etc.) and the nitrogen gases (NH_3, etc.) are presented in Section 2.2.

TABLE 2-1. RETORT GAS COMPOSITIONS (SOLIDS RECYCLE)
(Mole %)
(Adpated from Dept. of Energy, 1979, and Ondich, 1983)

	In-Situ	Union B Indirect Heat	Lurgi	Paraho Direct Heat
H_2	6.5	16.2	8.2	3.5
CO	1.2	3.9	0.8	1.6
CO_2	17.3	11.9	7.4	17.6
N_2	42.0	1.9	0.8	51.3
CH_4	1.1	15.5	3.9	1.9
C_2H_2	--	0.1	--	--
C_2H_4	0.1	1.2	1.4	0.7
C_2H_6	0.2	5.3	1.4	0.8
C_3H_6	0.1	2.6	1.2	0.4
C_3H_8	0.1	2.3	0.6	0.4
H_2S	0.12	2.01	0.12	0.24
NH_3	0.40	0.13	0.37	0.57
Water	30.8	31.52	55.03	19.2

B. Spent Shale Combustion--

In some processes, spent shale is burned to recover energy associated with with residual char. The resulting energy can be used to produce energy for either the retort or cogeneration of electricity. The reactions occurring in spent shale combustion are numerous and have significant effects on both the process performance and emission rates, as presented below.

1. Process efficiency -- The thermal efficiency of the overall retorting process is improved by spent shale combustion, primarily due to the energy value of the char, but also from any unretorted kerogen present after pyrolysis. At the typical combustion temperature range of 500-800 C, the major reactions are:

7. Kerogen + O_2 ---> CO_2 + H_2O + SO_2 + NO_x
8. Char + O_2 ---> CO_2 + H_2O + SO_2 + NO_x
9. Char + NO ---> N_2 + CO_2 + H_2O

The reaction of char with NO can be used to limit NO_x emissions by using staged combustion. Limiting the combustion temperature also limits the thermal NO_x formation. See Section 3.2.1.B.

Combustion of the spent shale can have a negative effect on the energy recovery if the combustion conditions are not adequately controlled. At elevated temperatures, carbonate decomposition can occur resulting in endothermic reactions that consume available energy. These reactions are:

	Heat of Reaction
10. $CaCO_3 \longrightarrow CaO + CO_2$	-420 cal/g
11. $CaMg(CO_3)_2 \longrightarrow CaO + MgO + 2CO_2$	-400 cal/g
12. $2CaCO_3 + SiO_2 \longrightarrow Ca_2SiO_4 + CO_2$	-270 cal/g
13. $CaMg(CO_3)_2 + 2SiO_2 \longrightarrow CaMgSi_2O_6 + 2CO_2$	-200 cal/g

If allowed to occur, these reactions result in a loss of thermal efficiency. Figure 2-3a shows the detailed reaction mechanism for these reactions and Figure 2-3b shows the CO_2 evolution rate as a function of temperature. At temperatures below 700 C, the CO_2 evolution and the energy loss from carbonate decomposition is minimal.

2. Sulfur reactions -- The use of a spent shale combustor limits the SO_x emissions due to reactive elements in the spent shale. The principal sulfur reactions in the combustor are:

- spent shale combustion

14. $4FeS + 7O_2 \longrightarrow 2Fe_2O_3 + 4SO_2$

15. $2SO_2 + O_2 + 2MCO_3 \longrightarrow 2MSO_4 + 2CO_2$,
    where M = Ca or Mg

- recycled spent shale combustion

16. $MS + 2O_2 \longrightarrow MSO_4$

17. $4Fe_3O_4 + 1/2O_2 \longrightarrow 6Fe_2O_3$

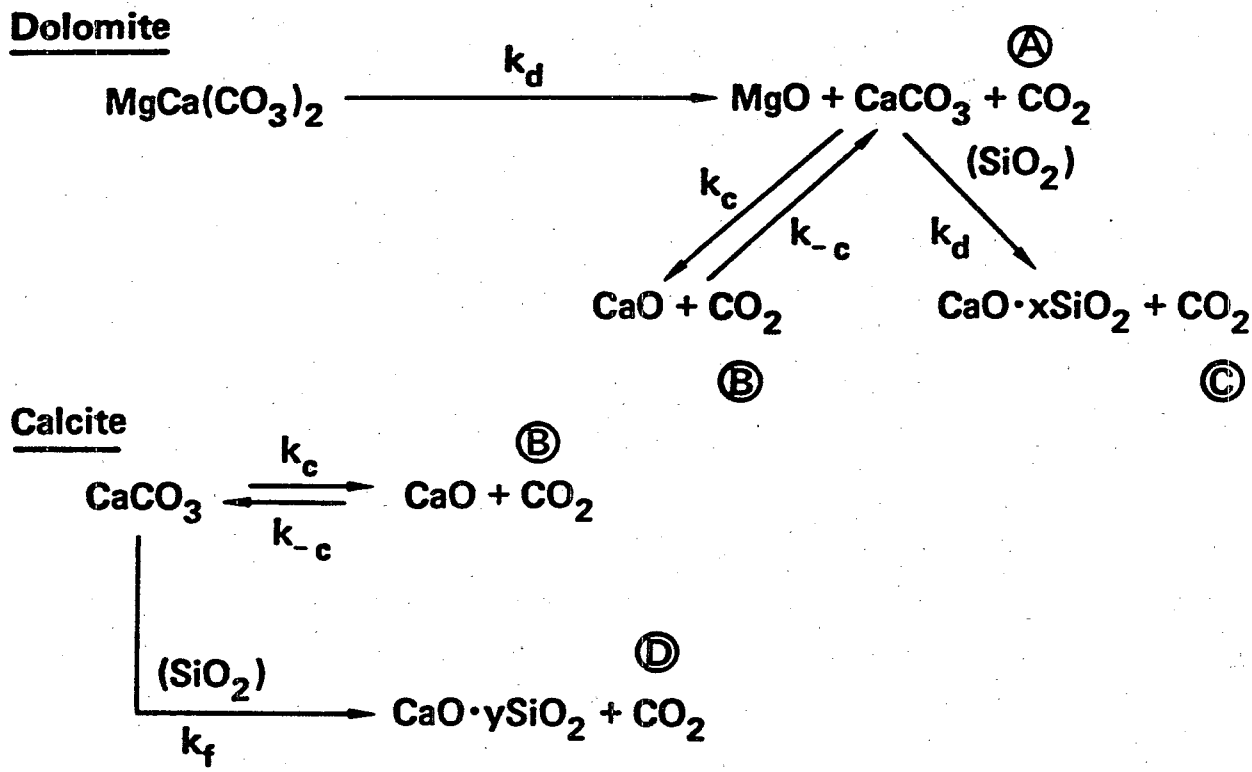


Figure 2-3a.

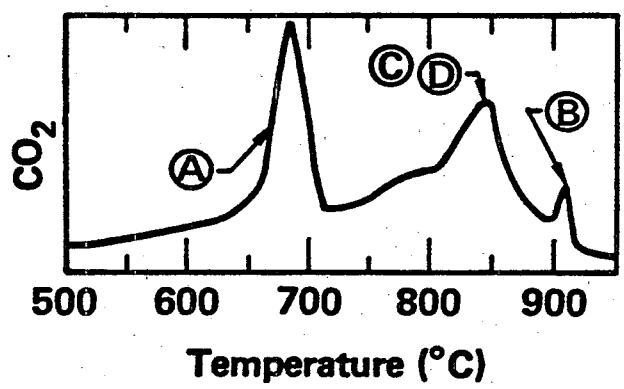


Figure 2-3b.

KEY

K_d = Decomposition rate constant

K_c = Calcination rate constant

K_{-c} = Reverse calcination rate constant

(A)
(B) Indiate temp. at which corres-
(C) ponding reactions take place.
(D)

Figure 2-3. Carbonate reaction scheme (Burnham, 1985).

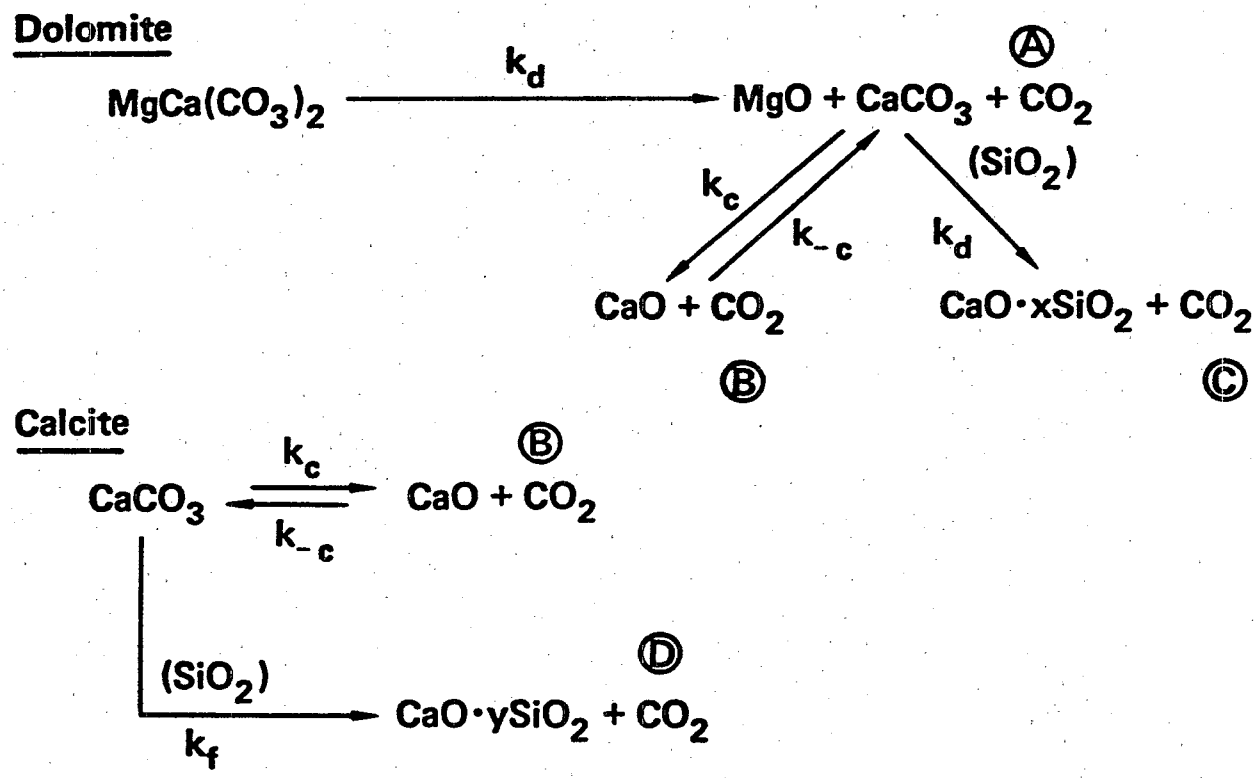


Figure 2-3a.

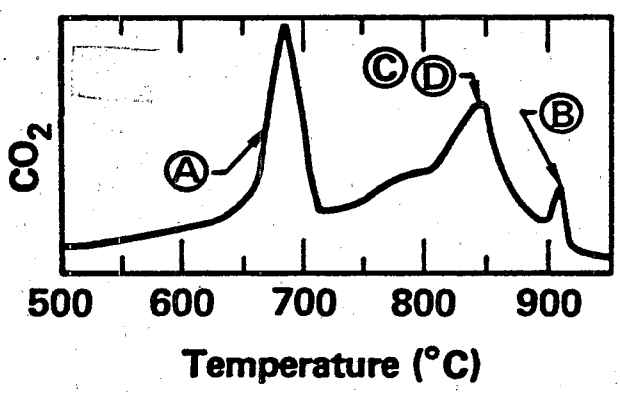


Figure 2-3b.

KEY

K_d = Decomposition rate constant

K_c = Calcination rate constant

K_{-c} = Reverse calcination rate constant

(A)

(B) Indiate temp. at which corres-

(C) ponding reactions take place.

(D)

Figure 2-3. Carbonate reaction scheme (Burnham, 1985).

Calcium and magnesium carbonate are present in both the raw and spent shale and react with the sulfur dioxide according to Reaction 15 thereby removing the sulfur from the flue gas.

Reaction 14 shows the oxidation of FeS to Fe_2O_3 and SO_2, the primary source for the sulfur in the spent shale. It is this reaction combined with Reaction 4 (the reaction of Fe_2O_3 with H_2S) that produces the synergistic effect of the recycled solids processes. Figure 2-4 illustrates this effect. Combusting the spent shale converts the FeS sulfur to SO_2 and Fe_2O_3. The SO_2 is removed from the gas phase (Reaction 15) and the Fe_2O_3 is recycled with the solids to the retort. In the retort the Fe_2O_3 removes H_2S from the gas phase (Reaction 4). The sulfur is essentially contained in a closed loop within the retort and combustor, exiting only as FeS on the combusted shale and sulfur in the oil.

The combustion of the spent shale therefore not only increases the thermal efficiency of the process, but also removes the gaseous sulfur from the flue gas eliminating the need for additional flue gas controls for the portion of the energy generated in the combustor. Since the concentration of calcium and magnesium carbonates is considerably greater than the amount of sulfur in the spent shale, the spent shale has additional capacity to remove SO_2 from retort offgas and from combustion of raw shale (which also contains carbonates). Because spent shale combustion has the potential for also burning both raw shale fines and retort gas, it also can be used for limiting SO_x emissions from these energy conversion processes.

3. <u>Nitrogen reactions</u> -- The combustion of the spent shale results in the following reactions for nitrogen containing species:

18. Char + O_2 ---> NO + NH_3 + CO_2 + H_2O + CO

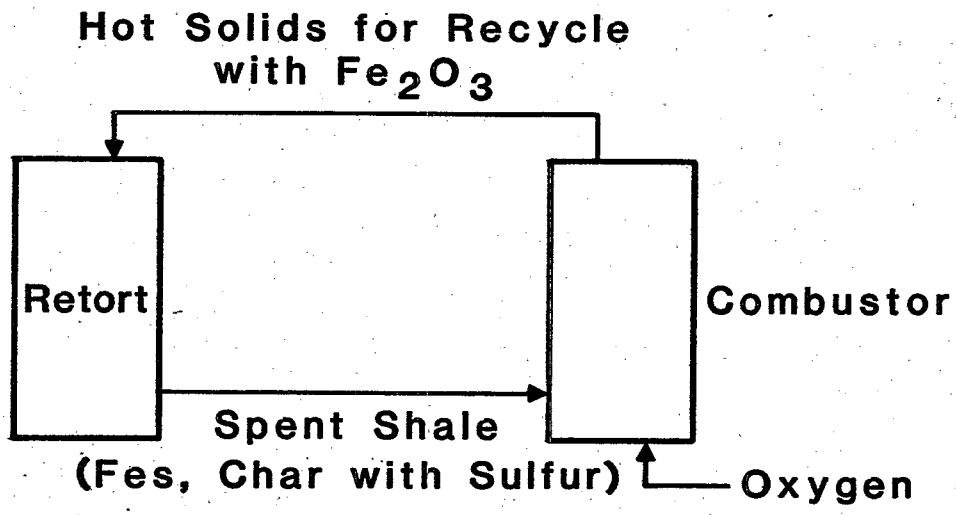
19. Char + NO ---> N_2 + CO_2 + H_2O

20. NO + NH_3 ---> N_2 + H_2O + H_2 (with & without catalyst)

21. NO ---> N_2 + O_2 spontaneous decomposition

22. NH_3 ---> N_2 + H_2 (catalyzed)

23. NH_3 + O_2 ---> N_2 + H_2O + NO



Sulfur Production

$FeS_2 + H_2 \rightarrow FeS + H_2S$

Sulfur Consumption

$1/2\ H_2 + 1/2\ Fe_2O_3 + H_2S \rightarrow$
$FeS + 3/2\ H_2O$

Combustion Reactions

$FeS + O_2 \rightarrow Fe_2O3$

$MS + O_2 \rightarrow MSO_4$

$Char + O_2 \rightarrow CO_2 + H_2O + SO_2 + NO$
$M = Ca, Mg$

$SO_2 + O_2 + MCO_3 \rightarrow MSO_4 + CO_2$

Figure 2-4. Sulfur reactions in recycled shale process (KVB, Inc.).

24. Buddingtonite ($NH_4AlSi_3O_8$) ---> NO + NH_3

Reactions 18, 23 and 24, the combustion of the char, ammonia and mineral Buddingtonite, respectively, form NO. The relative importance of the contribution from inorganic Buddingtonite at normal combustion temperatures has been shown to be minimal (Burnham, 1985).

The NO then can react with char in accordance with a carbon reduction reaction:

25. C + 2NO ---> N_2 + CO_2

and with any ammonia present according to Reaction 20. These reactions are shown in Figure 2-5. In kinetic studies to determine the mechanism for NO reduction (Taylor, 1983), found that the CO_2 evolution rate was not consistent with Reaction 25 and proposed that hydrogen in the char was the actual reducing agent.

The application of these reactions to reduce NO is actually the application of staged combustion, the principal of Reaction 25, and ammonia injection, Reaction 20. Since there are a number of sources of each component, (e.g., carbon can come from retorted shale as char or raw shale fines as kerogen, ammonia from reactions 18 and 23 or externally from the retort offgas), there are many variables that can affect the net NO formation. Consequently, the potential exists to limit NO emissions by proper process design. The cascading combustor described below is suitable for staged control of operating conditions (excess oxygen percent, temperature) as combustion proceeds and should therefore be suitable for designing to minimal NO levels.

As an example of how effective combustion staging can lower NO_x emission rates when burning high N_2 fuel, consider the experiments conducted on sewage sludge by Haug, et al (1981). The digested sewage sludge fuel was dried thoroughly and was in powder form. It had a heating value of approximately 2.8 kcal/g, an ash content of approximately 50 percent and a nitrogen content of approximately three percent. This was burned in a conventional fluidized-bed reactor first in an incineration mode with 30 percent excess air then in a pyrolysis mode (i.e., starved air) with 30

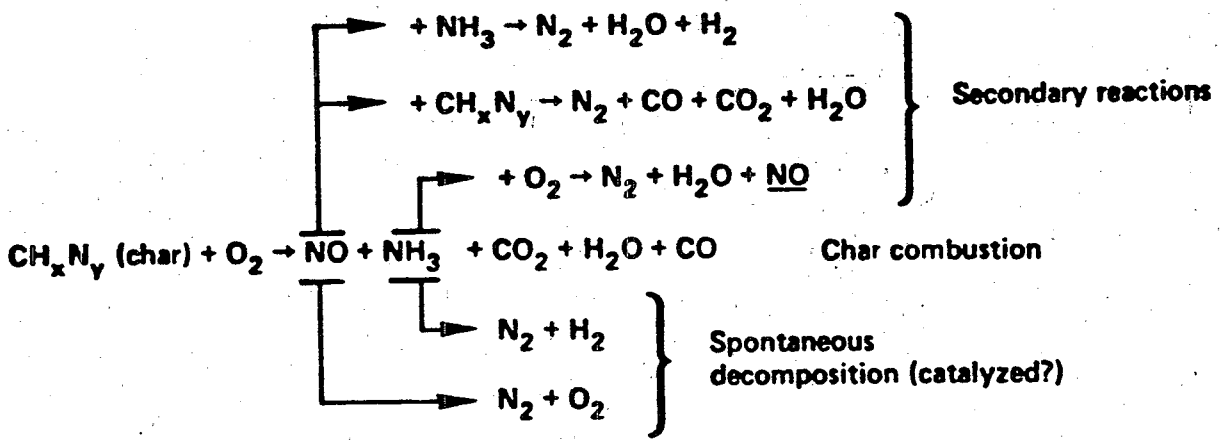


Figure 2-5. Nitrogen reactions during char combustion (Taylor, et al., 1985b).

percent excess air added to the gases in stages downstream of the bed. In the incineration mode the NO_x ranged from 500 to 1200 ppmV (at 5.5 percent O_2). In the optimum staging configuration, the NO_x was lowered to the range of 50-75 ppmV (at 5.5 percent O_2), a reduction of over 90 percent.

2.2 FORMATION AND FATE OF POLLUTANTS

2.2.1 Criteria Pollutants (SO_x, NO_x)

The pollutants emitted from an oil shale recovery process (no solids recycle) are directly related to the composition of the oil shale. The analyses of various oil shales is shown in Table 2-2.

The sulfur and nitrogen weight percent varies considerably, from 0.12 percent to 1.2 percent for nitrogen and 0.23 percent to 1.4 percent for sulfur. The wide variation in the concentration of the source material requires that process and emission controls be designed to limit subsequent pollutant emissions.

TABLE 2-2. CHEMICAL ANALYSIS FOR VARIOUS OIL SHALES
(Dept. of Energy, 1979)

Location	Oil Yield (gal/ton)	Chemical Composition, wt% C	CO_2	C acid	C org.	H	S	% Inorg. N	N
Tract Ca	34.3	21.9	16.0		17.5	2.4	NA	0.61	21
Tract Ca	24.8	16.6	19.7		11.2	1.7	1.1	0.54	43
Anvil Points	61.0	30.4	12.3		27.0	3.6	0.84	0.95	22
Anvil Points	30.6	19.6	20.6		14.0	1.8	NA	0.43	10
Anvil Points	22.4	15.6	17.5		10.8	1.6	0.63	0.41	27
Anvil Points	8.1	10.0	21.4		4.1	0.6	0.23	0.12	13
Co Core #3	22.0	12.6	12.3		9.2	1.7	1.4	1.2	78
Naval Resrv.1	1.9	4.7	11.4		1.5	0.5	0.87	0.54	92
Anvil Points	25.0	15.6	17.1		11.0	1.6	0.69	0.47	35
Geokinetics	4.7	6.5	15.2		2.4	0.5	NA	0.20	67
Horse Draw Co	25.0	19.2	29.5		11.1	2.0	0.60	0.45	32

A. Sulfur--

Raw oil shale contains a significant amount of sulfur. During retorting, the sulfur is converted to various organic and inorganic compounds and partitioned to the retort gas, oil and spent shale. The use of the retort gas (combustion or shale) and oil dictates that these sulfur compounds be recovered and/or controlled to prevent excessive emissions of reduced and oxidized sulfur species.

1. Formation -- Oil shale from the Mahagony basin contains up to 2 percent sulfur (by weight), with the average sulfur concentration being nearly one percent. Approximately 73 percent of the sulfur in raw shale is found as pyrite (FeS_2) (Williamson and Mechior, 1980) and approximately 23 percent as organic sulfur with the remaining 4 percent in the form of sulfates.

The disposition of sulfur during processing depends on both the original form of the sulfur and retort operating conditions. The primary concern when evaluating retort processes with respect to removal of sulfur gases is the composition and chemical structure of these sulfur compounds. The results from determining the composition of the sulfur gases from retorting three raw shales are shown in Table 2-3.

TABLE 2-3. SULFUR GASES FROM THREE RAW SHALES
(Wong, et al., 1985)

Sulfur Gas	Tract Ca 18 gal/ton mole %	Anvil Points 24 gal/ton mole %	Geokinetics 56-57'(4.7) 5 gal/ton mole %
Hydrogen sulfide	90.3	86.2	84.2
Methanethiol	0.38	0.55	1.5
Ethanethiol	0.26	0.47	0.67
Propanethiol	0.01	0.02	0.08
Dimethyl disulfide	0.005	0.01	0.03
Carbonyl sulfide	1.71	2.66	4.57
Carbonyl disulfide	6.92	6.39	0.24
Thiophene	0.21	0.39	1.37
Methyl thiophene	1.11	3.27	7.40
Total raw shale S, %	1.7	0.7	0.1
Conversion of S to S gases, %	20.2	7.9	1.8

The dominant inorganic sulfur reaction has been shown to be:

$FeS_2 + H_2 \longrightarrow FeS + H_2S$ (gas) + organic S (gas) + oil S

Since the FeS is not easily reduced to Fe at temperatures below 900 C, this reaction is consistent with the observation that over 60 percent of the inorganic sulfur remains in the retorted shale.

Reactions of organic sulfur compounds in raw shale have not been determined. However, the evolution temperatures and source of sulfur for individual gaseous species has been investigated (Wong et al, 1985). Raw shale and acid treated raw shale were pyrolized and evolved gas species were measured. Treating the raw shale with acid removes FeS_2, leaving only organic sulfur. Consequently, by comparing the results from the raw shale and the acid-treated shale, the specific source of the sulfur gases can be determined.

The H_2S evolves from the organic sulfur at temperatures of 250 to 400 C and from the FeS_2 at 500 C. The CS_2 evolves from the FeS_2 at

temperatures greater than 500 C. Thiophenes come from the organic sulfur at temperatures of 300 to 450 C. The methanethiol, ethanethiol and propanethiol are generated from both the organic sulfur compounds at 250 to 400 C and from the pyrite (FeS_2) at temperatures greater than 500 C.

The reactions can be summarized as:

26. $FeS_2 + 2H_2 \xrightarrow{\Delta} FeS$ (solid) + H_2S (gas) + CS_2 (gas) + methanethiol + ethanethiol + propanethiol

27. organic S $\xrightarrow{\Delta}$ H_2S (gas) + thiophenes + methanethiol + ethanethiol + propanethiol

2. Partitioning of sulfur during pyrolysis -- In the pyrolysis of oil shale, the sulfur is distributed between the gas, oil and spent shale. The results of the distribution analysis for various samples are shown in Table 2-4.

TABLE 2-4. DISTRIBUTION OF RAW SHALE SULFUR IN PRODUCTS (Dept. of Energy, 1979)

Sample	Avg. Sulfur Content, wt.%	% of Raw Shale Sulfur in Product		
Site		Spent Shale	Oil	Gas
Anvil Points	0.6	69	11	20
Anvil Points	1.29	78	8	14
Anvil Points	0.6	74	15	11
Geokinetics	0.5	86	7	7
Tract C-a	1.4	60	8	32
Tract C-a	1.4	66	5	28
Colony	0.8	69	15	19

The amount of sulfur partitioned to the gas phase varies from 7 to 32 percent of that contained in the raw shale. Figure 2-6 shows the partition data from Table 2-4.

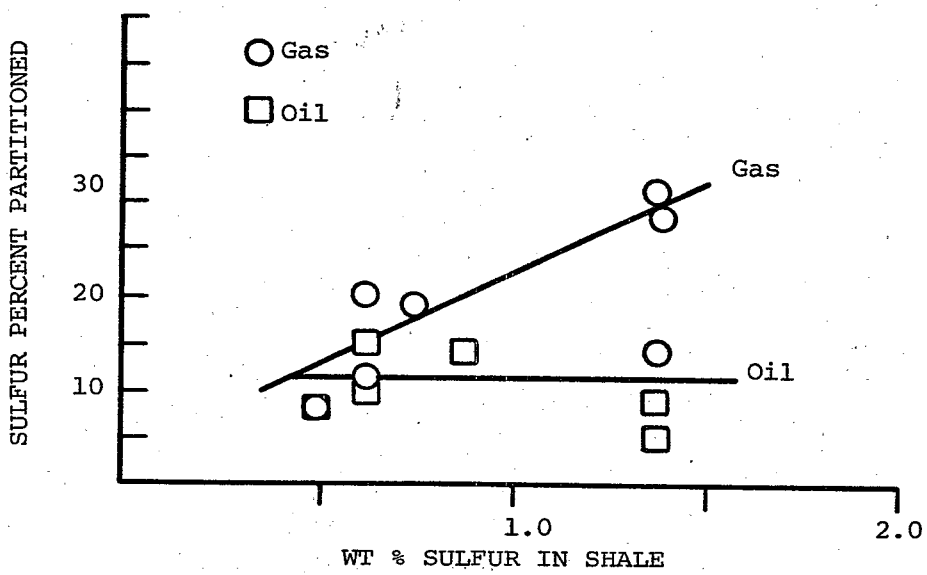


Figure 2-6. Partitioning of sulfur during retorting (Table 2-4 data).

B. Nitrogen Oxides--

Nitrogen oxides emissions (NO_x) from oil shale retorting are due to both fuel nitrogen (i.e., nitrogen compounds originating in the raw shale and released during the retorting process) and thermal NO_x from combustion of either the retort gas or the carbonaceous spent shale. In this section, only fuel-related nitrogen emissions are discussed. Thermal NO_x emissions are discussed in Section 3, Air Pollution Control Technology, and Section 5, Process Analysis.

1. Formation -- The fuel-related nitrogen emissions are determined by the nitrogen content of the raw shale, the chemical form of the nitrogen, the partitioning of the nitrogen to the gas, oil and spent shale and the chemical form of the gaseous nitrogen species. The composition of a number of raw shales from the Mahagony zone western states were presented previously in Table 2-2. The average nitrogen composition is 0.5 weight percent with values ranging from 0.1 to 1.2 weight percent. The nitrogen is present as both inorganic nitrogen compounds, primarily buddingtonite, and organic nitrogen compounds associated with the organic kerogen and bitumen.

The nitrogen conversion mechanisms are shown in Figure 2-7. Organic nitrogen is either deposited in the oil or forms HCN and NH_3 in the gas phase. Inorganic nitrogen decomposes to NH_3. The NH_3 formed can be oxidized to NO or reduced to N_2. The retorting process redistributes the nitrogen from the kerogen and shale rock to the shale oil and char. Typical nitrogen concentrations for shale and other fuels are shown in Table 2-5.

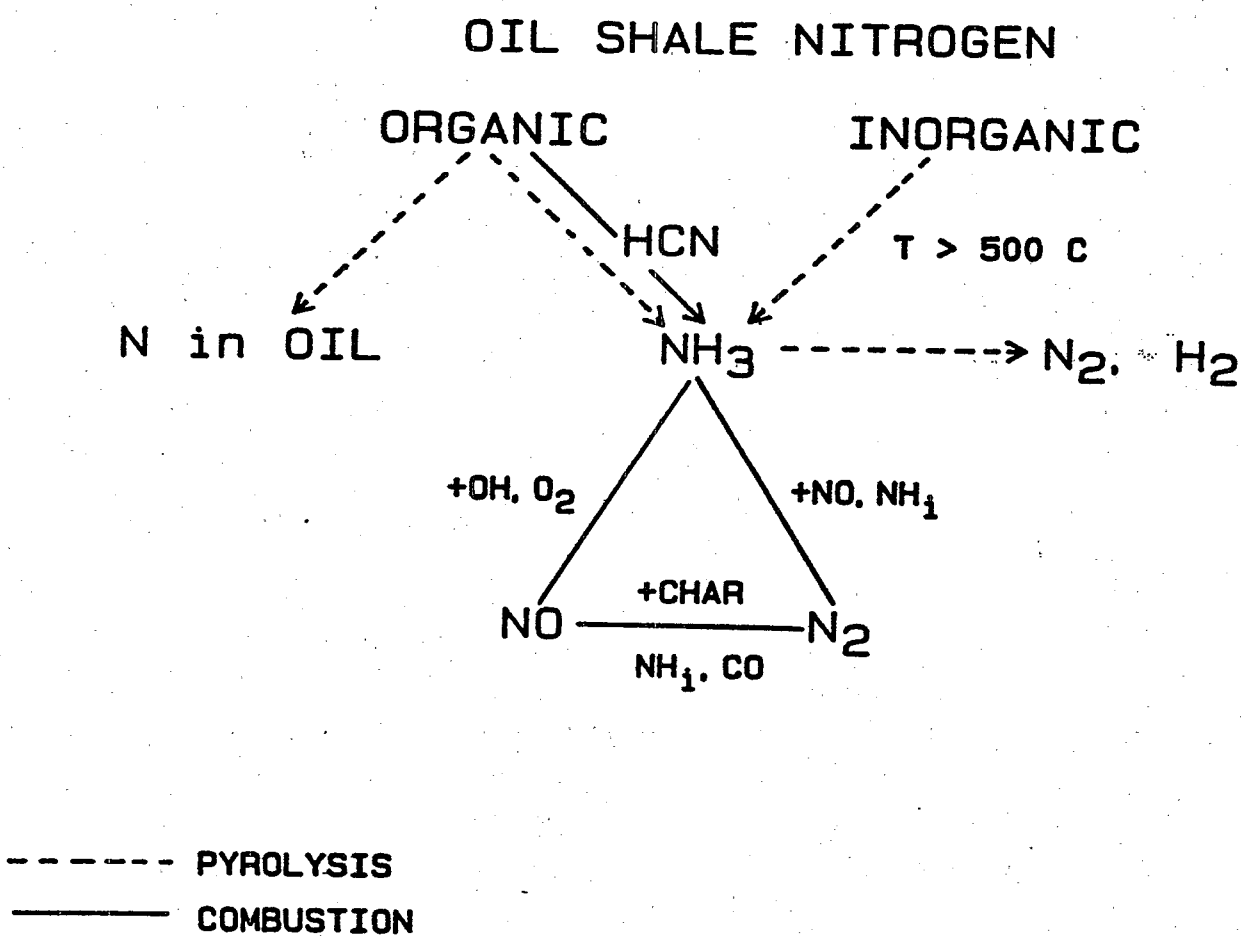


Figure 2-7. Nitrogen conversion mechanisms (Oh, et al., 1985).

TABLE 2-5. NITROGEN CONCENTRATION IN OIL SHALE COMPONENTS AND OTHER FUELS (Oh, et al., 1985)

Component	wt % N
Raw shale	0.5
Kerogen	3.0
Oil shale char	7.0
Shale oil	2.0
Other fuels:	
Crude oil	0.1
Coke	1.0
Coal	1.5

2. Partitioning -- Nitrogen is distributed into four components: oil, gas, spent shale and condensed water. Figure 2-8 shows the relative distribution. At retort temperatures of 500 C and 700 C, the nitrogen distribution is:

Component	% of nitrogen (data from Fig. 2-8) at 500 C	700 C
Spent shale	60	40
Oil	33	30
Water	4	15
Gas	3	10

The rapid increase in nitrogen (primarily ammonia) in the gas phase with temperature is also indicated by the ammonia evolution rate which varies with temperature shown in Figure 2-9. At 500 C, ammonia evolution is minimal with rapid increases between 500 and 700 C (Oh, 1985).

Using the data reported in the PSD permit applications presented in Section 4, the partitioning of nitrogen for the modified in-situ (MIS), Lurgi solids transfer and Paraho direct combustion was calculated and the results are shown in Table 2-6.

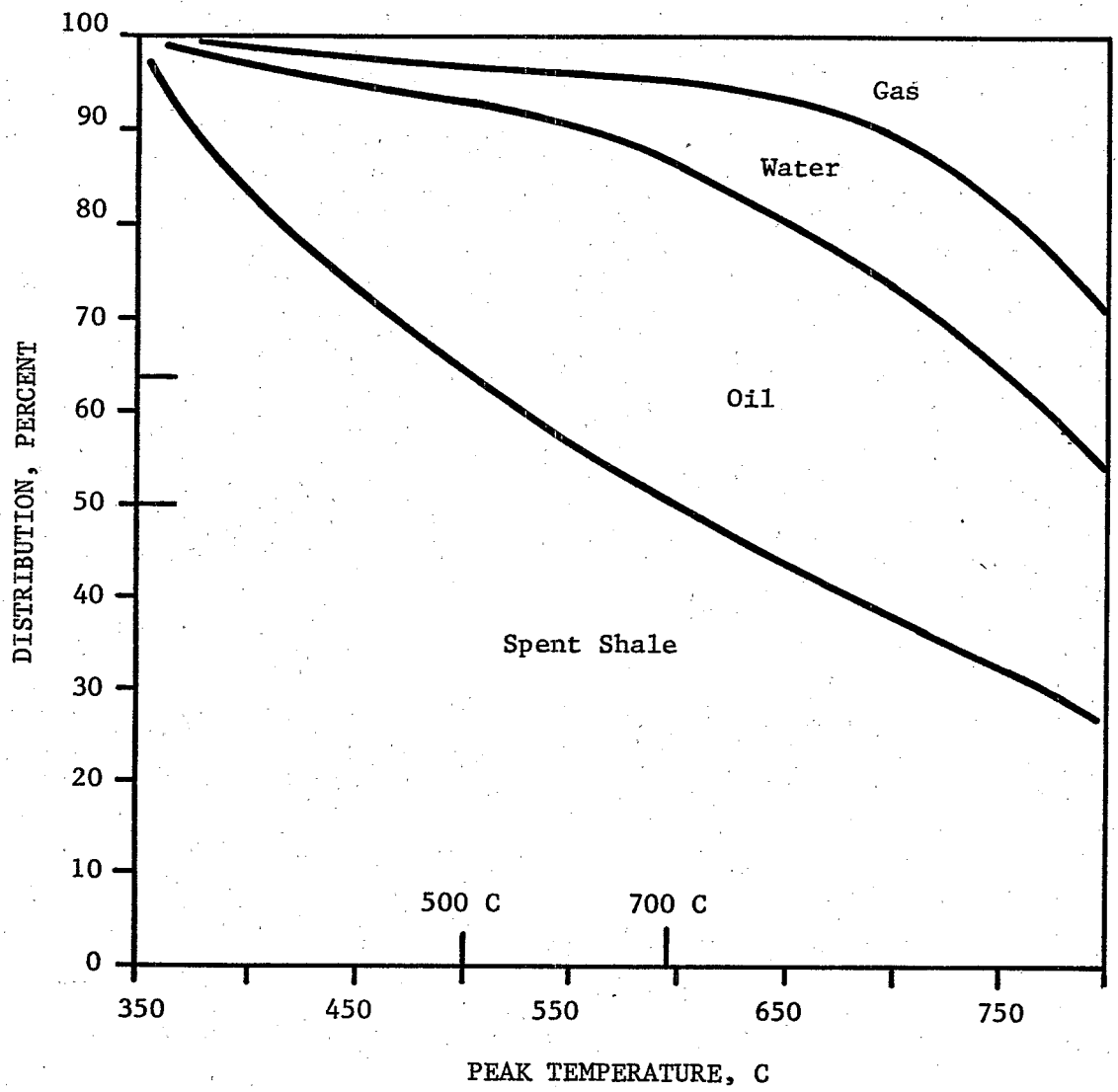


Figure 2-8. Nitrogen distribution during retorting (Oh, et al., 1985).

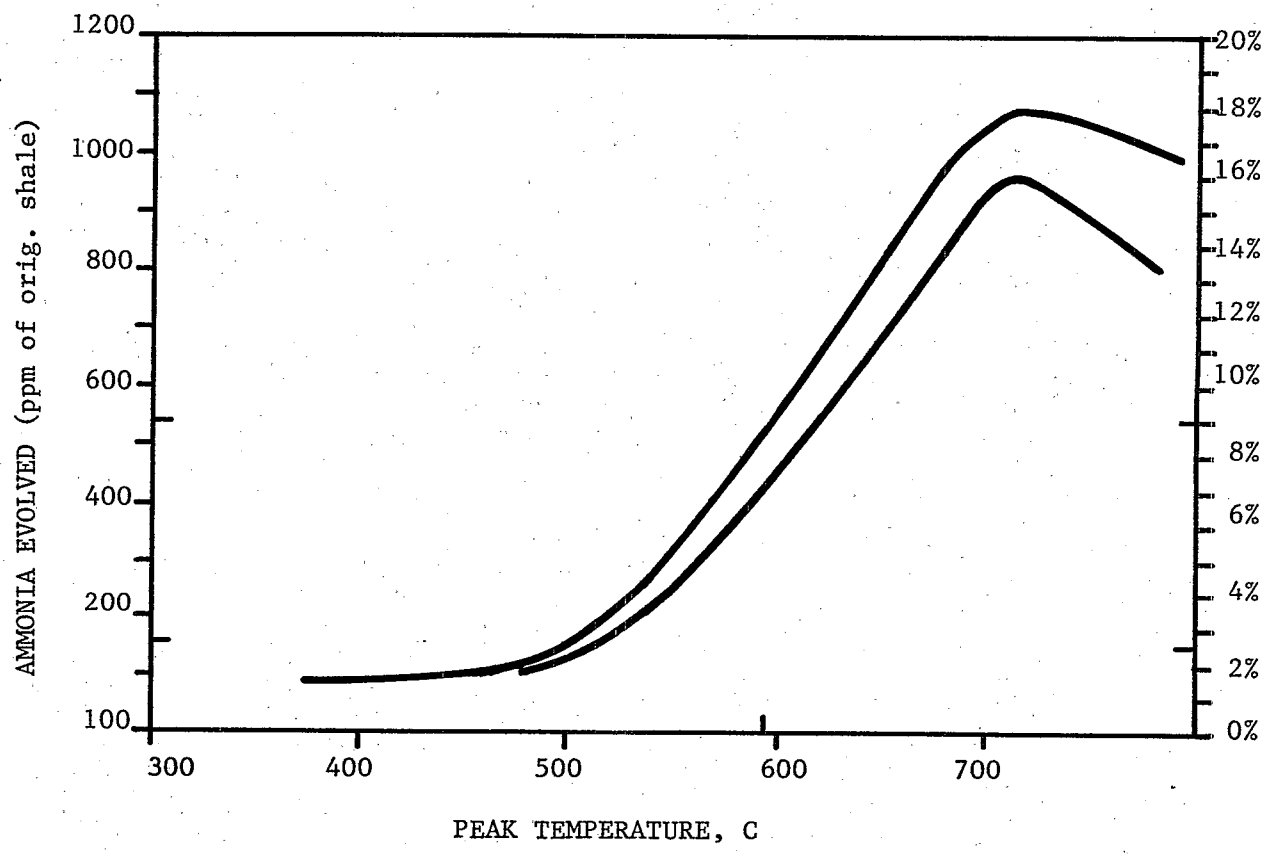


Figure 2-9. Ammonia evolved from Fischer Assay (Oh, et al., 1985).

TABLE 2-6. NITROGEN PARTITIONING FROM PSD PERMIT APPLICATIONS

Site	% of Raw Shale Nitrogen in Product			
	Spent shale	Oil	Water/Gas	Retort Temp. C
Tract C-b (modified in-situ)	21	25	54	1000 High
Tract C-b (Lurgi solids recycle)	5(35*)	55	10	500 Low
Paraho (direct combustion)	33	37	30	500-700 Medium

*Much of the nitrogen content in the spent shale, after Lurgi retorting, (as shown in parentheses) is burned in the lift pipe combustion and exits with the flue gas.

These data are consistent with the trend of the curves in Figure 2-8. The higher temperature retort, modified in situ, has the highest percentage of nitrogen in the gas phase, while the lower temperature direct combustion retort has lower nitrogen and the solids transfer Lurgi, with the lowest retort temperature, has the lowest amount of nitrogen in the gas phase.

3. Nitrogen species -- The trace nitrogen species in retort offgas are primarily ammonia with small amounts of other nitrogen compounds. In an investigation of nitrogen-containing species from an in-situ and an above-ground retort process, Sklarew (1984) identified the compounds in Table 2-7.

TABLE 2-7. NITROGEN-CONTAINING SPECIES IN RETORT OFFGAS (Sklarew, 1984)

Compound	ppm
HCN	6 - 39
Acetonitrile	17 - 58
propionitrile	16 - 26
isobutyronitrile	7 - 16
pyrrole or pyridine	6 - 45
butanenitrile	*
pentanenitrile	*
pyrrolidine	*
benzylamine	*
N-methylaniline	*
o-toluidine	*
N,N-diethylaniline	*
ammonia	1 - 3%

*Detected but below limit of quantification (approx. 1 ppm)

2.2.2 Trace Elements

The principal source of gaseous trace element emissions at an oil shale facility is the retort itself. The elevated temperatures, from 500 to 750 C in surface retorts and up to 1200 C in in-situ retorts, mobilize trace elements associated with kerogen and mineral phases and redistribute them among product gases, oil, process water, and retorted shale.

The quantity of trace elements volatilized during retorting has been studied by making direct, gas-phase measurements and by mass balances in which gas phase emissions are calculated by difference. These studies have revealed that trace element emissions during retorting are controlled primarily by retorting temperature and configuration and operation of the bed of raw shale and secondarily, by shale source, heating rate, and retorting atmosphere.

The type of process determines the effective retorting temperature. Thus, higher temperature in-situ retorts will revolatilize greater amounts of

trace elements than will the lower temperature processes. However, the relative trace element concentrations using the in-situ processes will be lower due to the high amount of dilution gas.

The flow configuration of the retort gas also effects the trace element concentration. Trace elements volatilized within the retort may be adsorbed onto raw shale or retorted shale, thus reducing and/or retarding their release from the retort. Thus, whether the bed is stationary or mobile and whether the gas flow is co-current or counter-current with the bed will affect the release of trace elements.

In-situ retorts are stationary beds of shale with a moving reaction zone. Some trace elements volatilized in the reaction zone may be deposited ahead of the reaction zone on cool raw shale and are not released from the retort until late in a run when the bottom of the bed is sufficiently hot. These volatile trace elements are released to the offgases in pulses towards the end of a run.

In surface retorts, which have moving shale beds with stationary reaction zones, volatilized trace elements are continuously released into the reaction zone with organic vapors and can be partially deposited either on the retorted shale or raw shale, depending on whether co-current or counter-current gas flow is used. Trace element emissions from above-ground continuous retorts are relatively uniform throughout a run, with variations due primarily to changes in raw shale feed composition.

At pyrolysis temperatures of 500 C, typical of above-ground, indirectly heated retorts, the only trace elements that are volatilized in significant quantitites are mercury and cadmium. At higher temperatures typical of above-ground, direct-combustion retorts (up to 750 C) and all in-situ processes (up to 1200 C), other trace elements, notably arsenic and selenium, also may be volatilized. Certain other trace elements (e.g., the halides and volatile organometallic compounds, such as nickel carbonyl) may also be released. However, there are no definitive data currently available.

The reader is referred to Appendix B, Trace Elements from Oil Shale Retorts, for a more thorough discussion of the individual trace elements and

their distribution during the retorting process. It should be noted that the majority of this work has been performed using small laboratory retorts or at pilot-scale, demonstration facilities without control technology for the criteria pollutants. Therefore, the applicability of this information to a commercial oil shale plant with control technology in place is uncertain. Retort gases at a commercial plant will be variously compressed, heavier hydrocarbons removed, sulfur and perhaps NO_x may be reduced, and the gases may be mixed with natural gas and burned. Thus, the information presented in the Appendix should be considered to be uncontrolled emission data.

2.3 RETORT PROCESSES

Retort processes can be categorized with respect to both location, above-ground or in-situ (underground), and heating method, direct or indirect. The above-ground retorting steps include:

- mining and handling shale -- removal of the raw shale, crushing and screening, conveying to retort, feeding retort, discharge from retort
- retorting - heating the raw shale to produce oil and gas
- product recovery - separation of oil from gas stream
- product upgrading - removing sulfur, nitrogen and trace elements from the oil product, improve oil quality
- offgas clean-up - removal of H_2S (and organic sulfur) and NH_3 (and organic nitrogen) from retort gas prior to combustion or sale
- spent shale disposal

Emissions associated with the raw shale mining, crushing, handling and spent shale disposal are primarily particulates from solids handling and CO and NO_x from vehicles and other combustion engines. As these emissions are similar to those from other ore mining processes the reader is referred to Section 3.0, Control Technology Review, and the EPA's publication AP-42 (1985) of emission factors for additional background on these techniques and their emissions. Also, the discussions in Section 4.0, PSD Permit Application

Analysis, and Section 5.0, Process Analysis, provide detailed information on the various emission sources and control techniques.

Emissions associated with the upgrading process are NO_x, CO and SO_x from boilers, heaters and other combustion devices plus hydrocarbons from product handling and storage. These processes and emission sources are similar to those in the petroleum refining industry, and the reader is again referred to Section 2 and the EPA's publication AP-42 for a detailed discussion. The discussions in Section 4.0, PSD Permit Application Analysis, and Section 5.0, Process Analysis, also provide detailed information on the various emission sources and control techniques.

The following discussion will focus primarily on the retorting processes, their variations and the effect of the type of process on the retort gas volume, pollutant concentration and subsequent ease of control.

Essentially, the following four techniques have evolved to provide efficient retorting processes: (These process are shown in Figures 2-10 and 2-11.)

- direct heating - above ground with combustion in a countercurrent retort
- direct heating - above ground with combustion in a crossflow moving grate retort
- indirect heating - above ground with recycled solids or gas
- in-situ - below ground - direct heating with combustion

The approximate retort gas flow rates and pollutant concentrations for each of these process types is shown in Table 2-8.

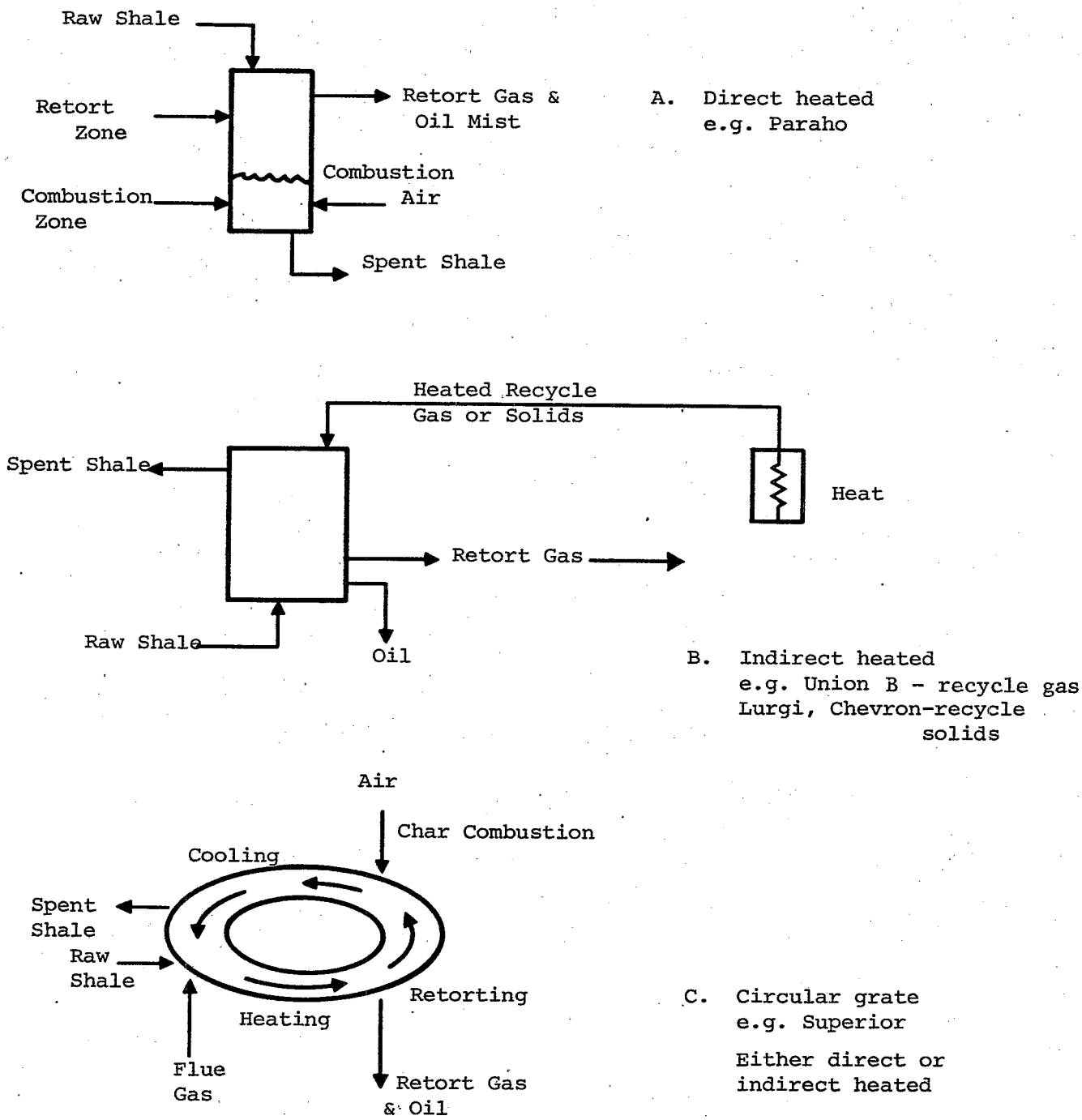


Figure 2-10. Schematics of above-ground retorting processes (KVB, Inc.).

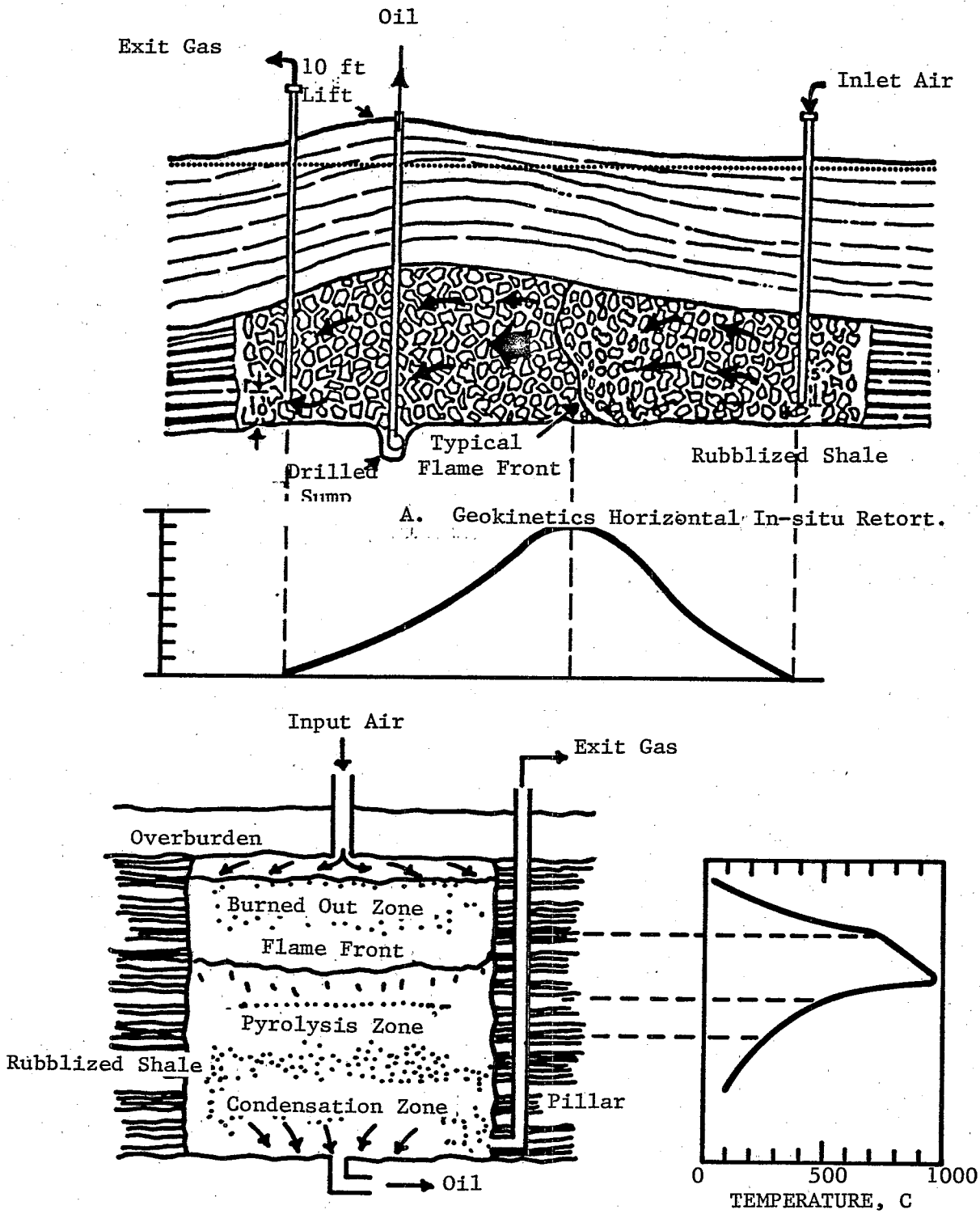


Figure 2-11. Schematics of established in-situ retorting processes (Agarwal, 1986).

TABLE 2-8. UNTREATED RETORT GAS CHARACTERISTICS

Type of Heat	Location	m^3 Retort Gas/ m^3 Oil Produced	H_2S ppm	SO_2 ppm
direct-combustion	in-situ	7000	1200	0
direct-combustion	above ground (circular grate)*	2000	0	80
direct-combustion	above ground (Paraho)	2000	4000	0
indirect-no comb.	above ground (Union, Chevron, etc.)	200	40000	0

*Circular grate operates with excess air -- others are starved air.

The gas production rate varies from a minimum of 200 m^3 retort gas/m^3 of oil for the indirect heating mode to a maximum of 7000 m^3 retort gas/m^3 of oil in-situ direct combustion mode, or a 35:1 increase. This is the same 35:1 ratio which appears in the pollutant concentration.

2.3.1 Above-Ground Retorts

Surface retorts are moving-bed, continuous reactors. Various types involve cocurrent or countercurrent flows, and examples in which the gas sweep is up, down, or horizontal may be found.

A. Direct Heating - Countercurrent Flow --

In the direct heating mode (Figure 2-10a) combustion air is introduced into the retort to burn the residual carbon and provide the heat necessary for the retort process. Raw shale and hot combustion gas flow countercurrently, the shale is heated and retorted producing oil and retort gas. However, the internal combustion results in the dilution of the retort offgas producing a large quantity (2,000 in-situ, m^3/m^3 of oil) of low heating value [9 kJ/kmole (100 Btu/scf)] gas with a relatively low pollutant concentration (1500 to 4000 ppm H_2S). Consequently, it is difficult to condition and use the retort offgas.

B. Indirect Heating--

In the indirect-heating mode (Figure 2-10b), the retort heat is provided by either recycled retort gas or recycled solids. Due to the absence of combustion in the retort, this process produces a low quantity of retort gas (200 m^3/m^3 of oil) of high heating value [90 kJ/kmole (1000 Btu/scf)] with relatively high pollutant concentrations (e.g., 35,000 to 40,000 ppm H_2S). However, the heat for retorting must be supplied by burning some source of fuel, either the offgas, part of the produced oil, natural gas or the carbon residue in the spent shale.

Since the spent shale from an indirect heated retort contains about four percent carbon, it is feasible to burn it to recover the additional energy content. The spent shale is conveyed to a combustion reactor where oxygen is introduced and the char is burned. A spent shale combustor also can be used to burn unprocessed raw shale fines and low Btu gas, again with minimal sulfur emissions due to the chemical nature of the spent shale. Combining the spent shale combustion with recycle solids heating results in the low H_2S and SO_x emissions discussed above in Section 2.1.1.4.

C. Direct Heat - Cross Flow Moving Grate Retort--

The moving grate is a continuous above-ground retorting method that uses a cross flow between the shale and the heating gas. The cross flow pattern provides improved control of the relative gas flow rates and various zone temperatures compared with the countercurrent direct heating mode. The moving grate approach also provides for easy transfer of the spent shale solids from the retorting zone to the spent shale combustion zone.

The process utilizes a moving bed retort (Figure 2-10c). Raw shale is fed onto a moving grate where it is first preheated with exhaust gases. The shale then moves into the retorting zone where it is heated to retorting temperature and oil is evolved and carried away with the circulating gas heating media. Variations of the process involve multiple retorting zones with individual temperature control to produce specific oil products (heavy, light, etc.). The retorted shale is then passed through the carbon recovery zone where residual carbon is partially oxidized, a portion of the low heating

value retort gas is burned and oil evolution is completed. The spent shale then continues into the shale cooling zone. Recycle gas is heated in the shale cooling zone by recovering sensible heat from the hot retorted shale. The shale is further cooled by ambient air which is heated for use in the direct fired burners or alternately in the indirect heater.

The retort offgas and combustion flue gas are combined prior to oil recovery. The oil and water are removed from the circulating gas which is then divided into a recycle stream for spent shale combustion and a low heating value gas for steam generation.

This system can be operated in either the direct or indirect heating mode. However, the development work and proposed plants are all based on the direct heating mode that produces a relatively large quantity of retort gas with a low pollutant concentration. As this process includes the combustion of the spent shale with the low Btu gas, the reactions described above for spent shale combustion are applicable. Therefore, the sulfur present in the retort gas is removed and the exit gas SO_x concentration is low (50 to 100 ppm SO_2). However, since the combusted solids are not recycled to the retort, the reactions limiting H_2S formation do not apply.

2.3.2 In-Situ Retorts

Above-ground processing of oil shale requires mining large amounts of solids and the subsequent disposal of approximately 80 percent of the material mined. This large amount of solid handling and disposal is costly. An alternative to above-ground processing is in-situ retorting which recovers the energy content of the oil shale without requiring solids removal, handling and disposal. In-situ processing also allows consideration of low quality shale (10 gal/ton) which would be uneconomical by above-ground techniques. The solids handling emissions are decreased, however, the increased gas flow rates required for in-situ retorts can result in increased pollutant emissions from the gas conditioning and combustion processes.

The various direct combustion processes for in-situ oil shale retorting are basically similar with regard to the reactions that take place, although they differ in geometry and flow configuration. Each may be

characterized as a semi-batch, fixed-bed reactor in which reaction zones, commonly called fronts, are swept through the bed by a stream of injected gas. The sweep contains oxygen to sustain the combustion and inert gases, steam or recycled produced-gas to control it. Differences exist principally in the flow patterns that prevail.

Two processes that have been developed for commercialization are the horizontal in-situ retort and the modified in-situ retort (Figure 2-11). In a vertical modified in-situ (VMIS) retort, the major dimension of the bed is vertical, the fluids flow downward, and gravity and pressure forces are parallel. This may be considered most likely to produce uniform, one-dimensional flows with nominally flat, horizontal fronts. An upward-flow, vertical retort might exhibit reduced yields owing to the tendency of the produced fluids to drain back into the reaction zones. A horizontal retort, with both its major dimension and flow oriented horizontal is characterized by nominally one-dimensional flow, however, gravitational forces act perpendicularly to the imposed flow. Generally, vertical plane fronts would be preferred, but because of the gravitational effects, the fronts may tend to tilt or override and be more difficult to control.

The behavior of an in-situ retort is most easily illustrated by considering the reactions occurring in various sections in an in-situ retort. After an initial startup period, the combustion retorting process reaches a quasi-steady state characterized by a slowly moving temperature wave which moves progressively through the retort. The temperature distribution defines the reaction zones depicted schematically in Figure 2-11.

Near the exit, there is relatively cool, raw shale, and the dominant process is water condensation. Except for a period near the end of a run, the gas will therefore exit at close to the dew point temperature. Somewhat higher temperatures exist in the next higher levels of the retort where oil may condense and the water originally in place will vaporize. As the fronts move through the retort, the oil is condensed ahead and revaporized repeatedly. Kerogen decomposition occurs in the next higher levels of the retort. Conversion rates become appreciable as the shale approaches 300 C (570°F), and decomposition is essentially complete at 500 C (900°F).

The highest temperatures in the retort occur in the combustion zone, the region next upstream from the retorting zone, where the injected oxygen is consumed in burning the carbon residue within the particles and the coke deposited on their surfaces. This normally supplies all the heat necessary to sustain retorting. Produced gas and oil will also burn if present in the oxygen-containing region. Optimum control will therefore aim at separating the combustion and retorting zones as much as possible.

It is this region of high temperature combustion plus the variation in bed temperature in the retorting zone due to localized heating that results in higher sulfur, nitrogen and trace element gas emissions from in-situ retorts.

The region near the inlet of the retort is the gas heating zone where the spent shale is cooled as heat is transferred from the solids to the incoming gas. The combustibles in any produced gas that is recycled will burn in this region when they reach their ignition temperature.

2.4 PROCESS DESCRIPTIONS

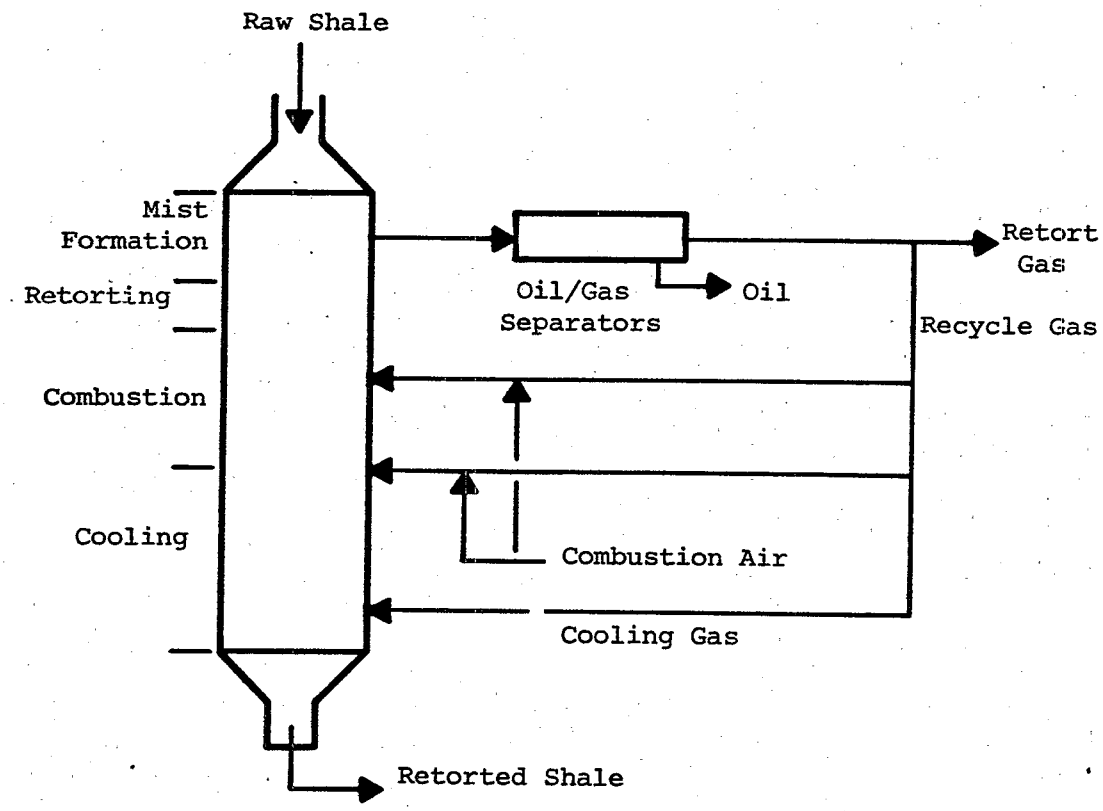
The following discussion covers brief descriptions of eleven specific processes being considered for full scale commercial retorting. This information was obtained primarily from previous reviews of oil shale retort processes. The reader is referred to references Agarwal, A. (1985), Shih, C. C. (1979), TRW (1979), and LLNL (1985), for more detailed description of each process as well as other processes that have been considered.

2.4.1 Paraho Retort Process

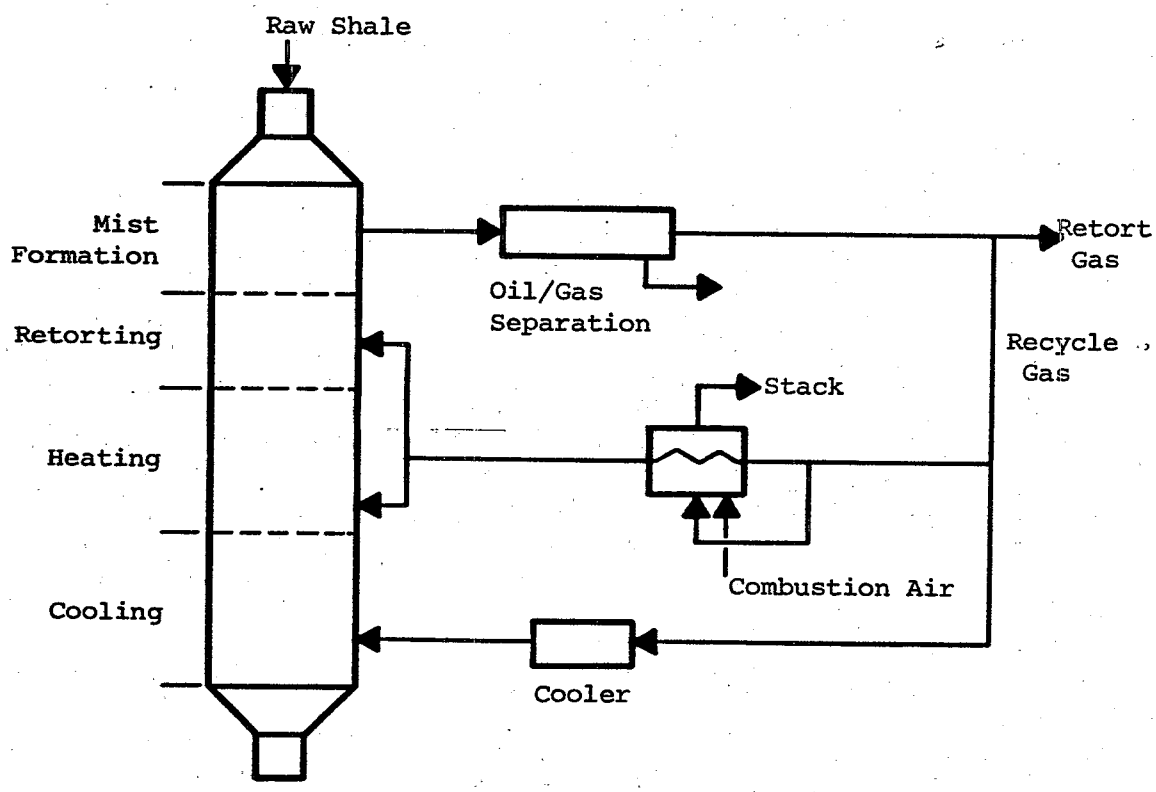
The Paraho retort (Figure 2-12) is a continuous vertical kiln for retorting oil shale. Crushed shale flows downward and is contacted with a countercurrent stream of hot gases which provide sufficient heat content to pyrolyze the kerogen in the shale. The oil is carried out of the top of the retort as a stable mist with the gas stream, and the retorted shale is removed from the bottom.

A. Directly Heated Mode--(Figure 2-12a)

Raw shale is distributed evenly across the top of the bed. Shale flows downward and is warmed by ascending hot gas in the mist formation zone. Next,



A. direct combustion



B. indirect heated

Figure 2-12. Paraho retort (adapted from Agarwal, 1985).

the preheated shale passes through the retorting zone where the kerogen is decomposed into oil, gas, and char. The char remaining on the retorted shale serves as fuel for the process in the combustion zone. Air for combustion is distributed evenly across the bed in an air-gas mixture by at least two levels of distributors. In the lower section of the retort, the shale is cooled by bottom recycle gas and this gas, in turn, is preheated before entering the combustion zone. The cooled, retorted shale passes through a rotary seal and is conveyed to a shale disposal area. Oil, as a stable mist, is swept out the top of the retort through the offgas collector and is separated from the gas in a coalescer-electrostatic precipitator system.

B. Indirectly Heated Mode--(Figure 2-12b)

In the indirect mode the retort heat is provided by heating the recycle gases in an external heater. The offgas product is not diluted by combustion gases and has a high energy content [78 kJ/kmole (885 Btu/scf)]. However, the spent shale contains two to four percent carbon residue representing an energy loss and decreased efficiency.

The principal zones within the retort are essentially the same as the direct mode, except that in the indirect mode, the combustion zone becomes the heating zone.

Because of changes in the nature and flows of the recycle gas, the temperatures of both the offgas and retorted shale are somewhat hotter for the indirect mode. Both the retorted shale and the offgas temperatures average approximately 150 C (300°F).

2.4.2 Hytort Oil Shale Retorting

Eastern shales produce a much lower yield of oil plus gas than western shales despite the fact that the organic carbon content is similar. This is due to the lower hydrogen-carbon ratio present in Eastern shale as well as structural differences. The Institute of Gas Technology has found that retorting in the presence of high partial pressure of hydrogen greatly increases the yield of oil and gas and has developed the HYTORT process.

The HYTORT Retort (Figure 2-13) consists of an indirect heated moving-bed-type retort with countercurrent contacting of hydrogen and shale. Crushed

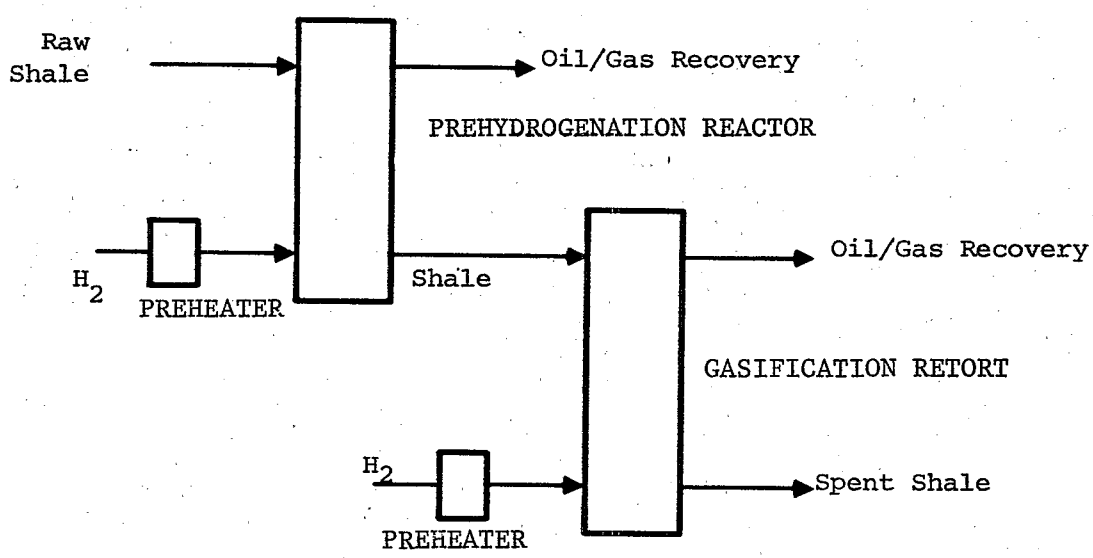


Figure 2-13. Simplified schematic of the Hytort process (KVB, Inc.).

shale is first heated with recycle hydrogen to about 425 C (800°F) for shale prehydrogenation. This prehydrogenation stage allows for greater recovery of the carbon content of the shale. The shale then is contacted concurrently with hot reactant hydrogen increasing its temperature to 650 C (1200°F) resulting in shale gasification. The pressures used have been from (1400 to 3400 kPa (200 to 500 psi) and average temperatures from 650 to 800 C (1200 to 1475°F). In the bottom of the retort, the spent shale is cooled by countercurrent contact with warm (preheated) hydrogen which carries that heat upward to effect the retorting. For Western shales, the retort would operate at 1700 kPa (250 psig).

Spent shale is removed from the reactor through liquid sealed lockhoppers. Product gas is scrubbed to recover water and shale oil and then treated to remove acid gases. A bleed stream of the clean gas is sent to the steam reformer where hydrocarbons are converted into hydrogen. All the make up hydrogen requirements of hydroretorting and catalytic hydrotreating are to be produced in this fashion and it is expected that no hydrogen will need to be purchased.

HYTORT shale oils from both eastern and western shales have relatively low viscosities and pour points. They can be pumped through pipelines in the raw state, in contrast to most shale oils from thermal retorting processes. HYTORT shale oil from Colorado shale is lighter in color and contains 30 percent less sulfur than oils from conventional retorting, because of the high pressure hydrogen treatment. However, like all shale oils, HYTORT oil requires catalytic hydrotreating to reduce its sulfur and nitrogen to acceptable levels.

2.4.3 T^3 Oil Shale Retorting

The NTU/T^3 consists of the NTU batch retorting process with significant improvements based on recent technology development including the vertical modified in-situ oil shale process.

In the T^3 process a mixture of air and steam instead of air and recycled gas is used for retorting. The use of steam results in an increase in

the retorting rate, control of maximum retort temperature, and the water gas shift reaction,

$$CO + H_2O \longrightarrow CO_2 + H_2$$

The net result of this reaction is to increase the combustion rate of residual carbon in spent shale, increase the hydrogen production per ton of oil shale, and reduce the oil loss due to thermal cracking.

The T^3 retort process is a semi-batch process where the two reaction vessels alternate between retorting and retorted shale combustion and cooling (Figure 2-14). The process operates in the following manner. At the end of a cycle, the Vessel A that had been functioning as the retorted shale combustion and cooling vessel has been filled with raw shale and will now function as the retort vessel. No shale is added during the retorting process and the retort heat is provided by the air-steam mixture from Vessel B. As the raw shale is retorted, the retorted shale front moves down the vessel, much in the same way that the front moves down the vertical in-situ retort. The retorting is complete when the vessel is filled with hot, retorted shale.

While Vessel A is functioning as the retort, Vessel B is functioning as the retorted shale combustion and cooling vessel. Air and water are introduced to the bottom of the vessel, which contains hot, retorted shale from the previous cycle, and the water is vaporized to steam while cooling the hot shale. The air-steam mixture flows up the vessel and burns the residual carbon char on the retorted shale. The heat released is contained in the air-steam mixture which then flows to Vessel A where the gas flow provides the heat for retorting the raw shale in Vessel A.

As the lower section of the hot shale is cooled, it is discharged and an equal volume of raw shale is introduced in the top of Vessel B. At the completion of the cycle, Vessel B is filled with preheated raw shale ready to be retorted and Vessel A is filled with hot, retorted shale ready to be burned for its char value and then cooled and discharged.

2.4.4 Lurgi Ruhrgas Retorting Process

The Lurgi Ruhrgas Retorting Process is an indirectly heated continuous process in which a fine-grained heat carrier is circulated from a combustion

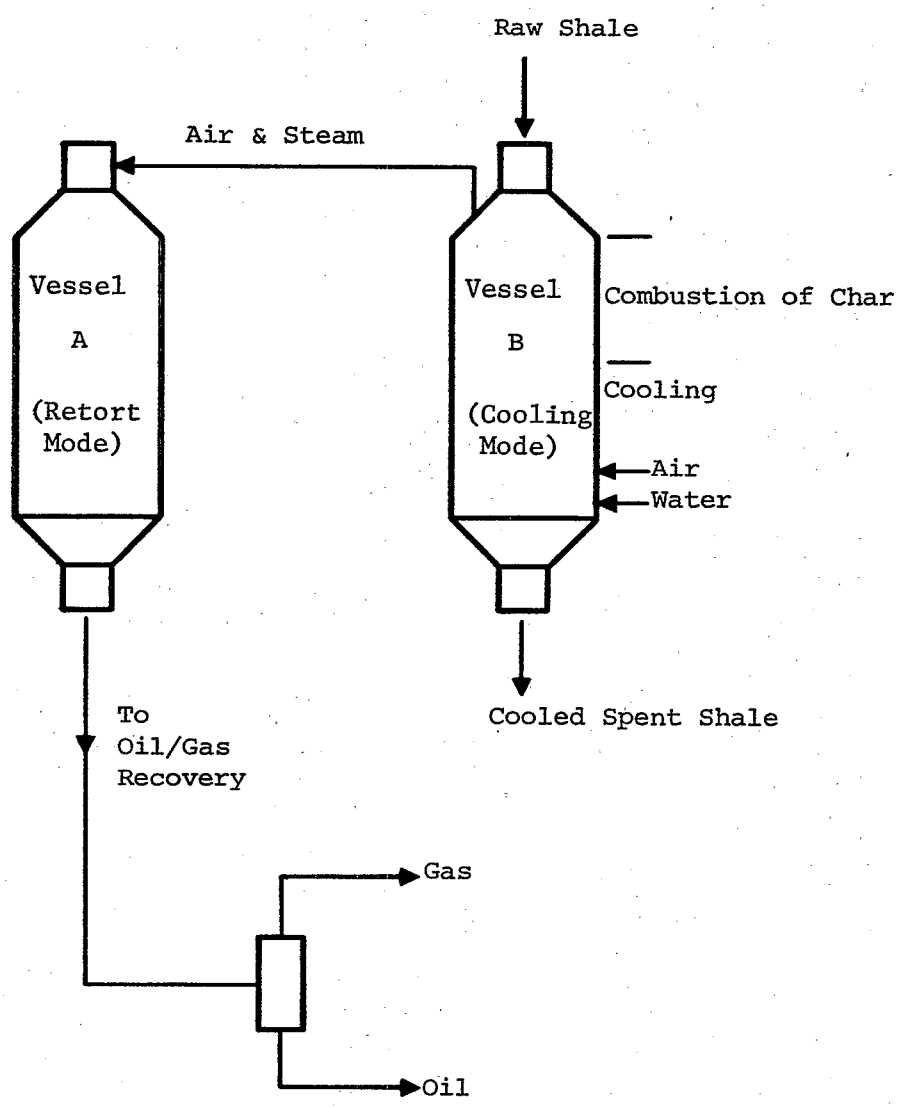


Figure 2-14. Schematic of the T^3 process (KVB, Inc.).

zone to a retorting zone to supply the retort heat (solids recycle heating). Heat is transferred from the hot recirculated solid to shale feed in a horizontal screw mixed retort (Figure 2-15).

Raw oil shale (including fines) is mixed with six to eight times as much hot spent shale. The raw shale feed is rapidly heated to about 450 to 600 C (840 to 1100°F) where retorting occurs. The retorted shale mixture passes to the lift pipe where preheated combustion air is introduced to burn residual carbon from the spent shale. The combustion gas and heated shale residue are separated at about 600 to 700 C (1100 to 1300°F). The heated spent shale is recycled to provide the heat for the retorting process.

The volatile gas stream contains oil product, retort offgas containing H_2S and NH_3 and entrained dust. Cyclones are used to remove the bulk of the entrained dust. The gas is then scrubbed with oil in a series of absorption units for product recovery. The retort gas is then treated to remove the NH_3 and H_2S and burned to produce steam and/or electricity.

Combustion gas leaving the collection bin enters the primary cyclone where most of the dust is removed. The gases are passed through the heat recovery train, secondary cyclone, humidifier, and finally an electrostatic precipitator before being discharged to the atmosphere. The spent shale combustion results in the reaction of the sulfur gases reducing the SO_x to a concentration of 70 to 100 ppm.

2.4.5 Tosco II Oil Shale Retorting

The process is indirectly heated and employs a solid-to-solid heat exchange (between hot ceramic balls and raw shale) as a means for providing the heat of retorting. While this process could be termed a solids recycle heater, the solid material being recycled is not spent shale so the reduction of H_2S evolution is not obtained. The integral parts of the TOSCO II retorting system are: retort and accumulator; product fractionator; processed shale removal system; ceramic ball system; and raw shale preheat system (Figure 2-16).

The raw shale is fed into a slightly inclined, rotating drum where it is preheated to approximately 500°F. Ceramic balls, at 1.5 times the shale

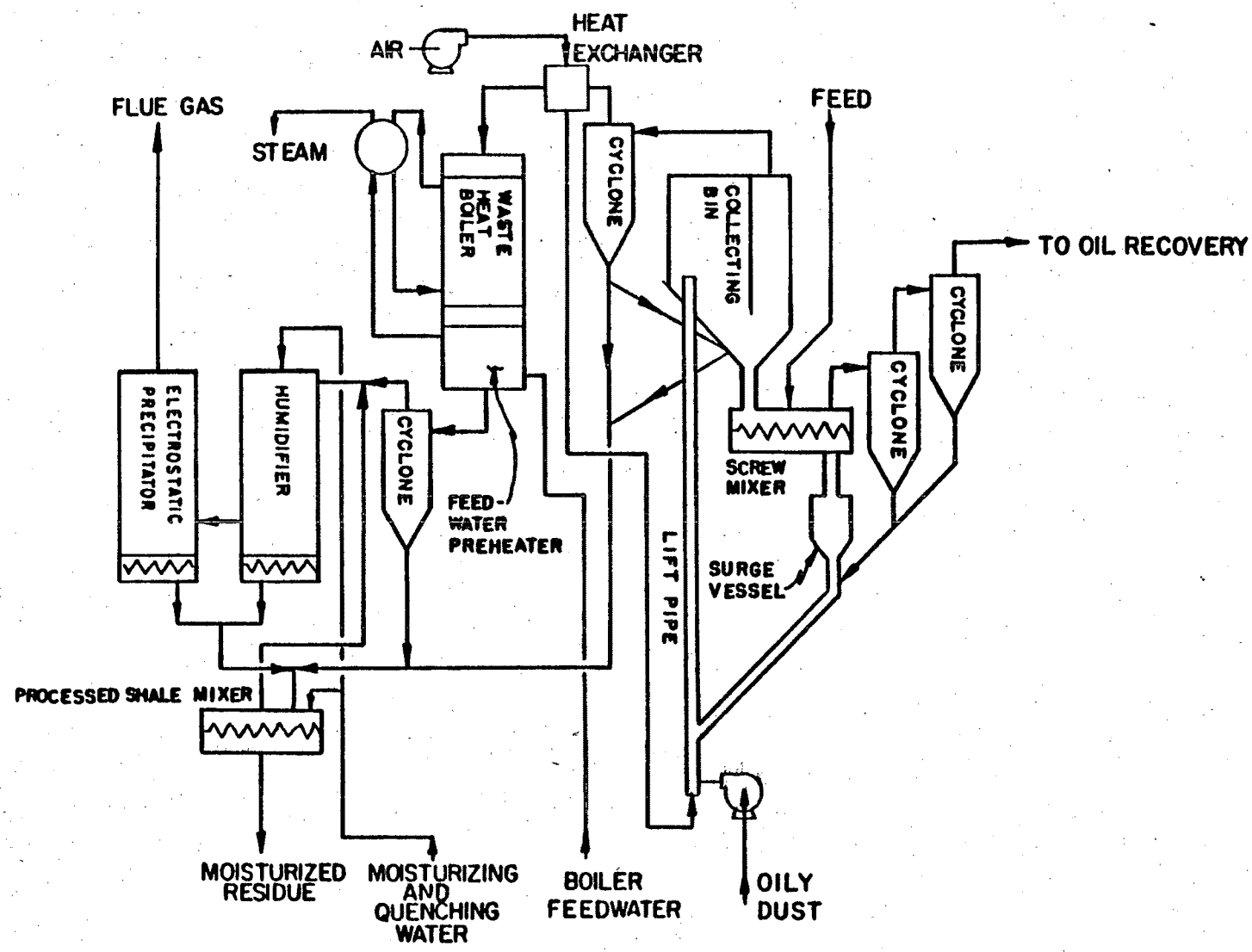


Figure 2-15. Lurgi-Ruhrgas retorting process (Agarwal, 1986).

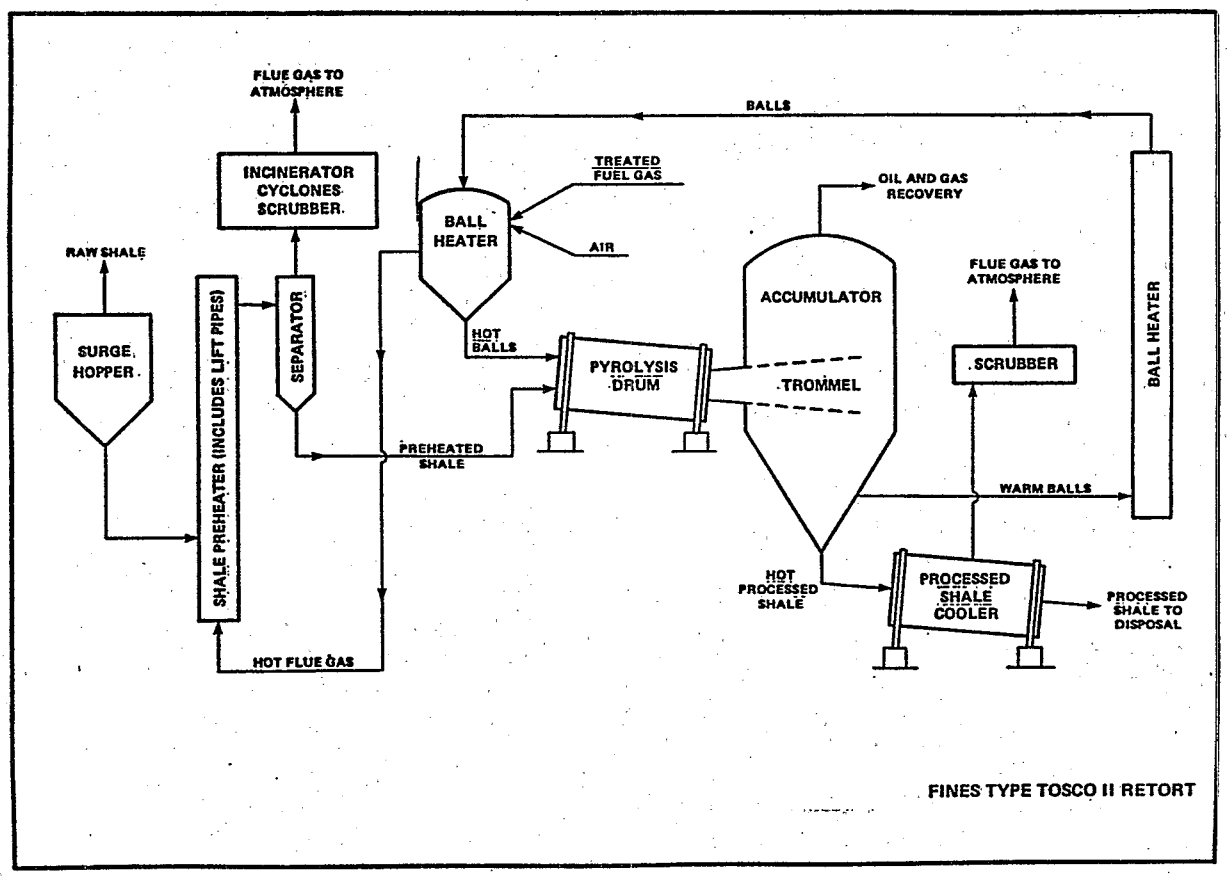


Figure 2-16. Fines type Tosco II retort (Quintana Minerals Corp.-PSD Permit Application, 1982).

mass flow rate, and previously heated to about 800 C (1300°F), are also added to the retort. The rotating, mixing action results in pulverization of the raw shale. Heat transfer from the ceramic balls raises the shale temperature to approximately 500 C (900°F), and pyrolysis, or retorting, of the kerogen in the shale occurs. The pyrolysis vapors and the mixture of balls and pyrolyzed shale are then taken to an accumulator vessel. This accumulator consists of a rotating, perforated screen or trommel which retains the balls but allows the pulverized shale to pass through, thus affording a separation of the two. The pyrolysis vapors are removed from the vapor dome at the top of the accumulator and sent to a fractionator for oil recovery, while the ceramic balls are sent for recycling and the processed shale is eventually sent for disposal.

In the oil recovery section the pyrolysis vapors are separated by the fractionator into gas, naphtha oil, gas oil, bottom oil, and gas condensate, or foul water. Each stream is sent to its respective processing unit for appropriate treatment.

Processed shale dust, contained with the ceramic balls as they emerge from the accumulator, is removed by hot flue gas from the steam superheater. The particulate matter is subsequently converted to a sludge in the venturi wet scrubber and sent to the disposal area. The clean flue gas is emitted to the atmosphere through the scrubber stack. The clean ceramic balls are then transported by a bucket elevator to the ball heater for heating and recycling back to the retort. In the ball heater, treated fuel gas and shale oil are burned by atomizing the fuels with air in a vertical combustion chamber at the top of the vessel. Hot flue gas thus generated passes downward, concurrently with the balls, thereby heating them. The flue gas is separated from the balls in the gas disengagers and the hot balls are returned to the retort.

The hot spent shale from the accumulator is taken to rotating drum steam generator/cooler where it is cooled to about 150 C (300°F) by indirect heat transfer to the feedwater to generate some process steam. The cooled spent shale is then taken to another rotating drum and processed shale moisturizing water is added. Steam incidentally produced during the moisturizing operation entrains some processed shale dust which is removed in the venturi wet scrubber; the steam, along with a little particulate matter,

is released to the atmosphere through the scrubber stack. The spent shale, cooled to below 100 C (200°F) is moisturized to approximately 14 percent water by weight (to aid compaction and dust control) and then transported to the disposal area.

The disengaged hot flue gas is used to preheat the raw shale feed, thereby recovering most of the waste heat. The preheating system consists of a series of three lift pipes (or preheat zones), a thermal oxidizer (incinerator), cyclones and wetscrubbers. The flue gas is introduced at the bottom of the last lift pipe (first preheat zone), where the raw shale stream from the second lift pipe (second preheat zone) is also received. The solids are lifted pneumatically and heat transfer from the gas to the shale occurs. The preheated shale is accumulated in a collecting bin at the top of the lift pipe and sent to the retort. Residual dust in the flue gas is separated by a cyclone and added to the feed going to the retort.

Since the flue gas temperature is at its highest when introduced to the last lift pipe (first preheat zone), it partially retorts the very fine shale, which results in hydrocarbon vapor release into the flue gas. Therefore, the flue gas is introduced to a thermal oxidizer (located between the first and second preheat zones) to burn the hydrocarbons so that the hydrocarbon emission to the atmosphere is not excessive. Some shale oil, liquids, and air are also added to the oxidizer to sustain combustion. The resulting flue gas is cooled, and then introduced to the bottom of the other two lift pipes. At this point, the temperature of the flue gas is low enough so that the extent of retorting of the fine shale is less than in the first preheat zone. Hydrocarbons released in these two lift pipes are emitted with the flue gas, without incineration.

2.4.6 Union A, B, C and SGR

All four of the processes developed by Union Oil Company are variations of an upflow vertical kiln retort using a rock-pump shale feeding mechanism (Figure 2-17). The retorts are known as Retort A, Retort B, Retort C and SGR-3. In all three processes, the oil shale is pushed upward through the retort while the gas flow is downward. The Union Retort A is a direct-heated retort where the heat for pyrolysis is generated by internal combustion

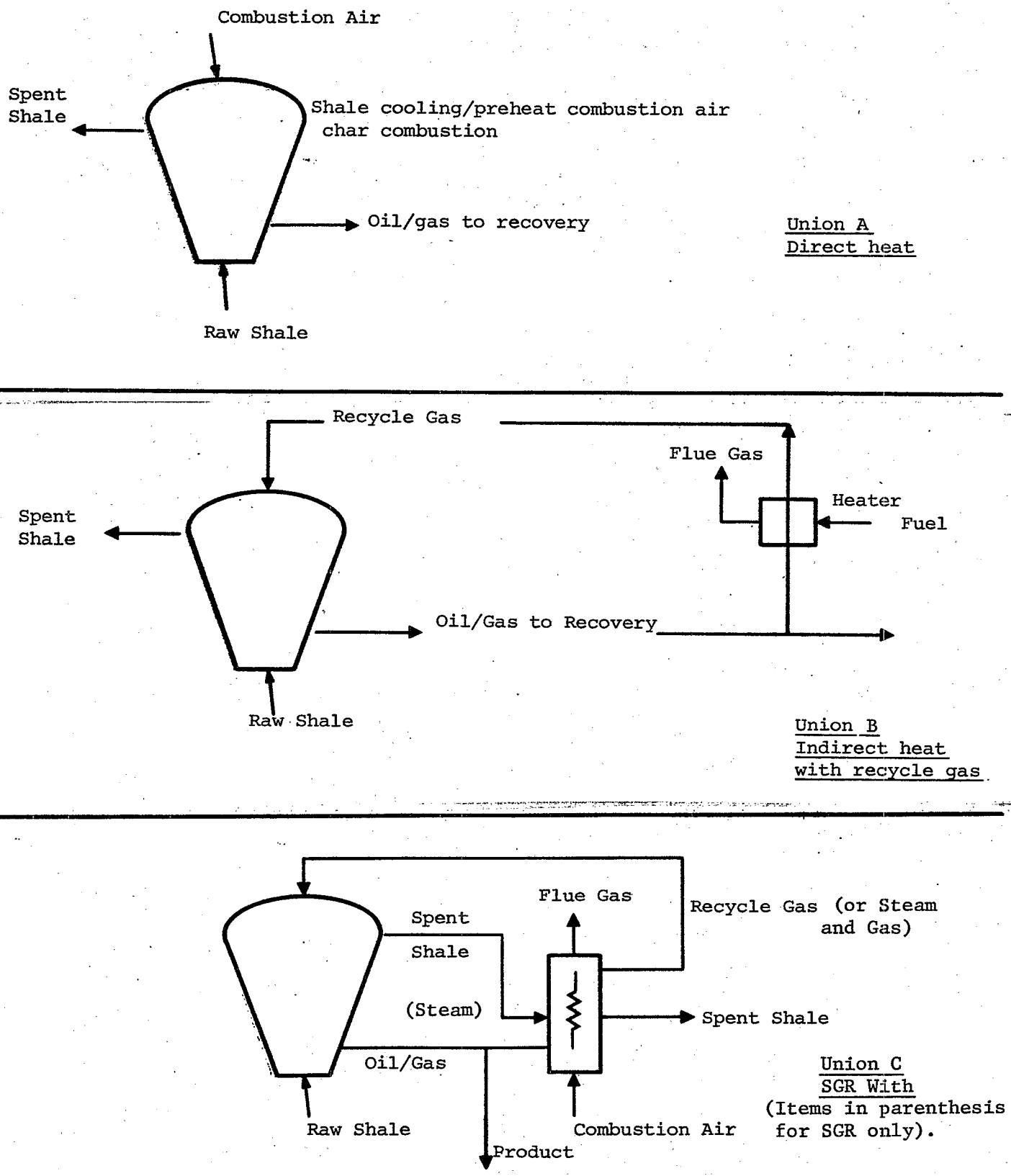


Figure 2-17. Union processes (KVB, Inc.).

of residual carbon on the spent shale. In Retorts B and C and SGR, the heat is supplied indirectly by externally heated hot gases.

A. Union A--

In the Union A process, air is introduced at the top of the retort to support combustion of residual carbon on retorted shale fragments in the upper portion of the retort. This combustion provides the heat necessary to retort the upward-moving shale.

As the shale enters the retort, it is contacted by a countercurrent flow of hot gases leaving the combustion and/or retorting zones above. In this lower zone, the gas-cooling zone, heat is transferred from the hot gases to the cool incoming shale. The feed is preheated, and the oil vapors are condensed while the gases are cooled. Passing up through the retort, the preheated shale enters the retorting or pyrolysis zone. The gases passing downward in this area heat the shale to a temperature such that organic material is pyrolyzed, producing shale oil vapor, product gas, and residual carbon on the surface of the shale. As the gaseous products of pyrolysis are evolved, they are swept downward by the flow of combustion gases. The downward sweep of pyrolysis to the cooler region of the retort helps limit product degradation (cracking and coking).

The retorted shale fragments then enter the combustion zone where they encounter preheated combustion air passing down through the retort. This air sustains the combustion of the residual carbon that remains on the surface of the shale after retoring. Excessive temperatures must be avoided to limit fusion of the shale and formation of clinkers.

The hot spent shale leaves the combustion zone and enters the top most zone, the air preheat zone. In this area, heat from the spent shale is transferred to the cool incoming combustion air, thus preheating the air before entry into the combustion zone. Upon leaving this zone, the shale is pushed up over the top of the retort and falls into the ash disposal chute.

B. Union B--

The Union B is an indirect retorting process based on the upflow of shale. The process is similar to the directly heated retort described above

with the exception that the retort heat is supplied by externally heated gas resulting in lower retort gas volumes with high pollutant concentrations.

C. Union C and SGR--

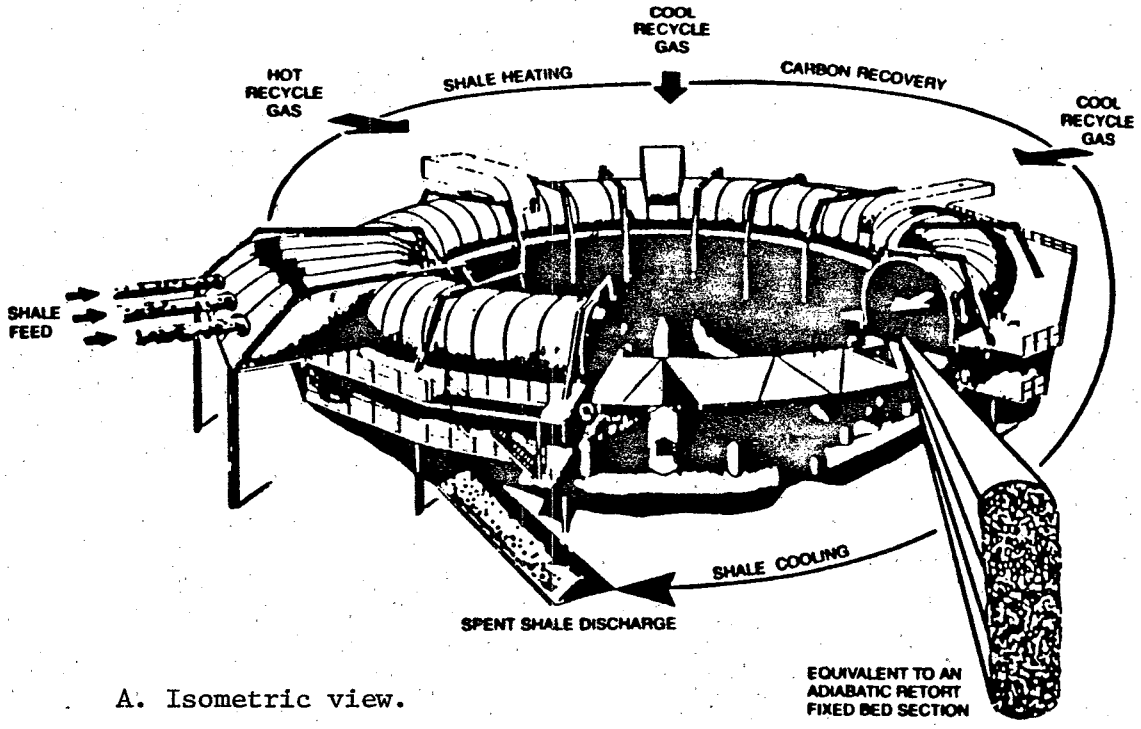
The Union C and Steam-Gas Recirculation Retort (SGR-3) are similar to the Union B retort (downward gas flow with indirect heat) but also provide for the combustion of the residual carbon in a separate vessel thus producing enough hot flue gas to supply all of the retort heat requirements. The spent shale combustion results in energy recovery and reduction of sulfur emissions. In the SGR system the addition of steam increases the retorting rate and energy recovery due to the water/gas shift reaction as used in the T_3 process.

2.4.7 Circular Grate Retort

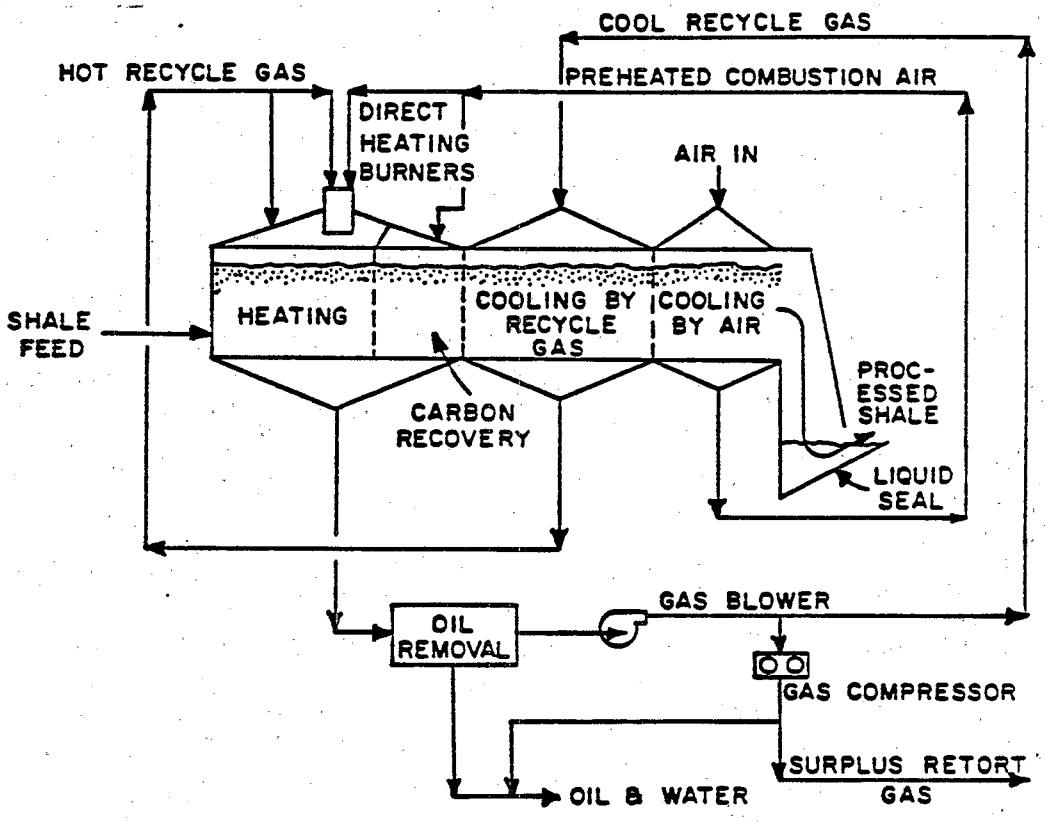
The circular grate is a continuous moving bed above-ground retort. The doughnut-shaped retort is divided into five separately enclosed sections: a loading zone, a retorting zone, a residual carbon recovery zone, a cooling zone, and finally an unloading zone (Figure 2-18). The ability to control the process conditions as the solids move through the retort allows for variations in retort and combustion conditions to provide specific results.

The raw shale is fed to the traveling grate so that the finest material is at the bottom and coarsest on top. The prepared bed of shale passes first into the retorting zone where it contacts a stream of hot gases, and retorting occurs as the hot gases pass down through the bed of shale. The oil and gas mixture passes to a separator/condenser system where shale oil is recovered, and the cooled recycle gases then pass through the retorted shale bed in the cooling zone of the retort to cool the processed shale before dumping and provide heat for the recycle gas.

The shale is cooled further by ambient air which is heated for use in the direct fired burners or alternately in the indirect heater. This energy recovery improves thermal efficiency. The rotating grate is sealed between stationary hoods on top of the bed and the windboxes underneath by water troughs on both sides of the shale bed.



A. Isometric view.



B. Straight Path Representation
(The circular path of the solids bed is pictured as a straight path for clarity)

Figure 2-18. Superior direct-heated process (Agarwal, 1986).

Preheated recycled gas from the shale coolng zone is further heated by mixing with hot combustion gases of 1370 to 1650 C (2500 to 3000°F). The resulting gas mixture, at 680 to 815 C (1250 to 1500°F) is introduced into the shale heating zone completing its cycle.

Product gas is taken as a slip stream from the recycling heating media gas. A high Btu gas of 14,900 to 22,400 kJ/m^3 (400 to 600 Btu/scf), depending on grade of feed shale) is produced by using the indirect heating mode. The gas produced from direct heating is diluted by combustion gases (CO_2 and N_2) and has a heating value of 3000 to 4200 kJ/m^3 (80 to 130 Btu/scf) depending on shale grade.

This process takes advantage of the chemistry of spent shale combustion to limit the SO_x emissions from this portion of energy generation. The moving grate process also has the capability to control the combustion conditions for staging and NO_x emission control.

2.4.8 Fluidized-Bed (Chevron STB Oil Shale Retorting)

The Staged Turbulent Bed (STB) retort is a small-particle indirectly heated retorting process in which retorted shale is burned in a separate combustor and recycled to the retort to supply the process heat requirement. The flow scheme for the core of the process is shown schematically in Figure 2-19.

The environment in the retort section, Figure 2-20, can best be described as a staged, moving bed of particles, in which a portion of the particles are "fluidized." Solids first move downward through the retort. The internal design creates a staging effect approaching a plug flow pattern with countercurrent gas flow thereby minimizing the residence time required to assure complete retorting. The solids then flow upward in the second stage with co-current gas flow. The two-stage approach provides for equal residence times for large and small particles. During the downflow, the heavier particles drop quickly, however, during the upflow, the larger particles rise more slowly.

Within each stage, conditions have all the appearances of fluidized state, although the superficial gas velocity is far below the minimum

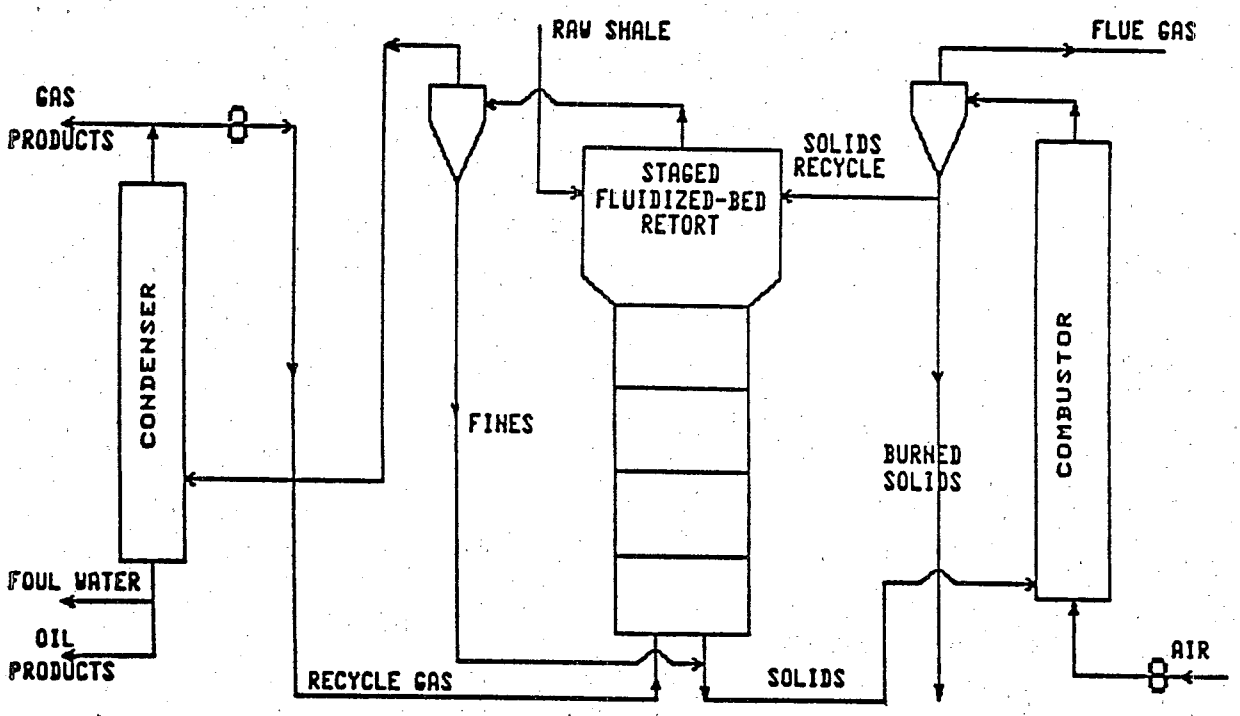


Figure 2-19. Schematic of a fluidized-bed retort system (Diaz & Braun, 1985).

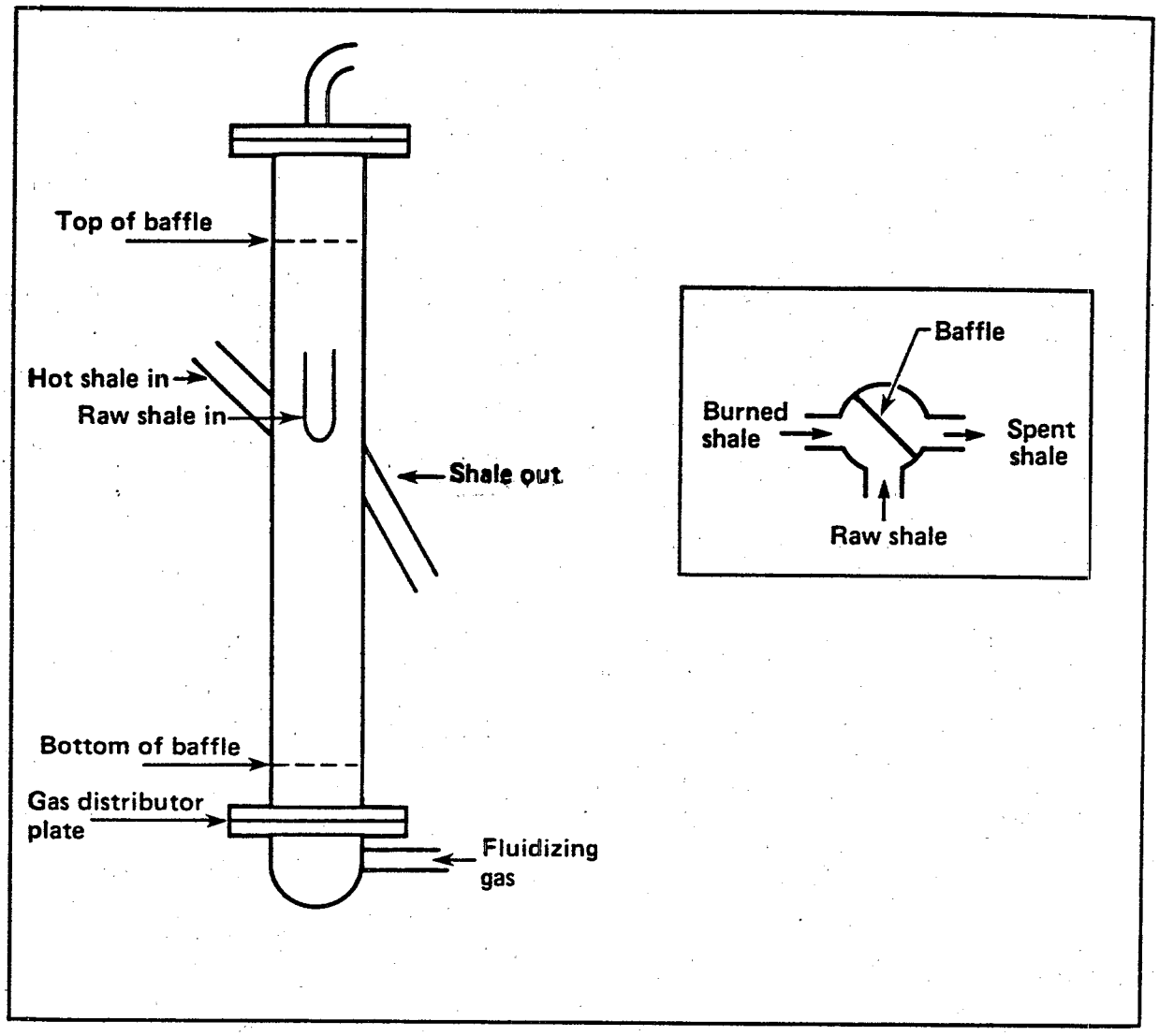


Figure 2-20. Two-stage fluidized-bed retort (Mallon, 1985).

fluidization velocity of many of the particles. Rapid local mixing and a high rate of heat transfer keep the temperature profile in the retort very nearly isothermal.

The retort can be operated with a stripping gas consisting of either recycled product gas, steam, or an inert gas.

The fresh shale throughput for this process is 5-10 times that of most other retorting processes. The use of small shale particles, the excellent heat transfer conditions in the retort, and the "staging" effect in the retort make the residence time required for complete retorting quite low. The extremely low residence time required in the combustor is a consequence of the high reactivity of the carbonaceous residue.

The combination of thermal shock experienced by the fresh shale on entering the retort, removal of the binding kerogen, and turbulent conditions within the retort causes some breakup of the shale. The finest particles (<200 mesh) are swept overhead from the retort with the product vapors. Most of the elutriated particles are recovered from the vapor stream before condensation of the oil. The recovered particles, collected by primary and secondary devices, contain carbonaceous residue; and they are, therefore, sent to the combustor section for recovery of their fuel value.

In the product recovery section, the product vapors are condensed in stages producing several oil fractions and foul water. The initial condensation stage has to be capable of handling some solids, as the fines removal equipment is never 100 percent efficient. The non-condensible gases are either produced directly or a portion is recycled to the retort for stripping. With very lean shales, some or all of the product gas can be used as a supplemental fuel in the combustor.

The combustion section receives coarse retorted shale withdrawn from the bottom of the retort and retorted shale fines removed from the product vapors. Air injected at the bottom of the combustor both transports the shale pneumatically and combusts the carbonaceous residue. The temperature of the solids rises rapidly. At the outlet of the combustor, a solid separator splits the stream once again into fine and coarse fractions. Most of the hot,

coarse fraction is recycled to the retort to provide the necessary process heat. The excess coarse shale is sent on to the heat recovery section, from which it eventually discharges as wet coarse spent shale. The fine fraction exits the separator with the flue gas. This stream is also sent to the heat recovery section where it is cooled and then separated into a wet fine spent shale stream and a flue gas stream. The SO_x content of the flue gas is claimed to be unusually low (<20 ppm) because of the scrubbing action of the shale itself.

2.4.9 Cascading-Bed Retort and Combustor

The Cascade-Bed Reactor has been developed by Lawrence Livermore National Laboratory (LLNL) to take advantage of the significant performance improvement inherent in retorts employing the solids recycle heating and spent shale combustion process while making significant improvements in various portions of the process.

The Cascade-Bed system is shown in Figure 2-21. Raw shale and hot burned shale are mixed by gravity flow in a series of inclined planes (Figure 2-22). The mixed shale is then retorted on a moving packed-bed pyrolyzer (Figure 2-23). The retort gas is removed at several locations to prevent fluidization of the solids.

A lift pipe, sized for optimum pneumatic transport (a significant advantage of this system over the combined lift pipe-combustor in the fluidized bed system) conveys the retorted shale to the top of a cascading-bed combustor (Figure 2-22). Gravity-delayed fall of the shale through this combustor yields sufficient residence time for burning the char to reach the burned-shale temperature required for retorting. Low pressure drop and low elutriation of fines are major advantages of using a cross-flow of air.

An additional advantage is that cross-flow combustion allows for easy control of the operating conditions (excess oxygen and temperature) on each stage which can be used to reduce NO_x emissions.

2.4.10 Occidental Vertical Modified In-situ (VMIS)

The Occidental Vertical Modified In-situ (VMIS) process is a combination of above-ground and in-situ processing. True in-situ processing,

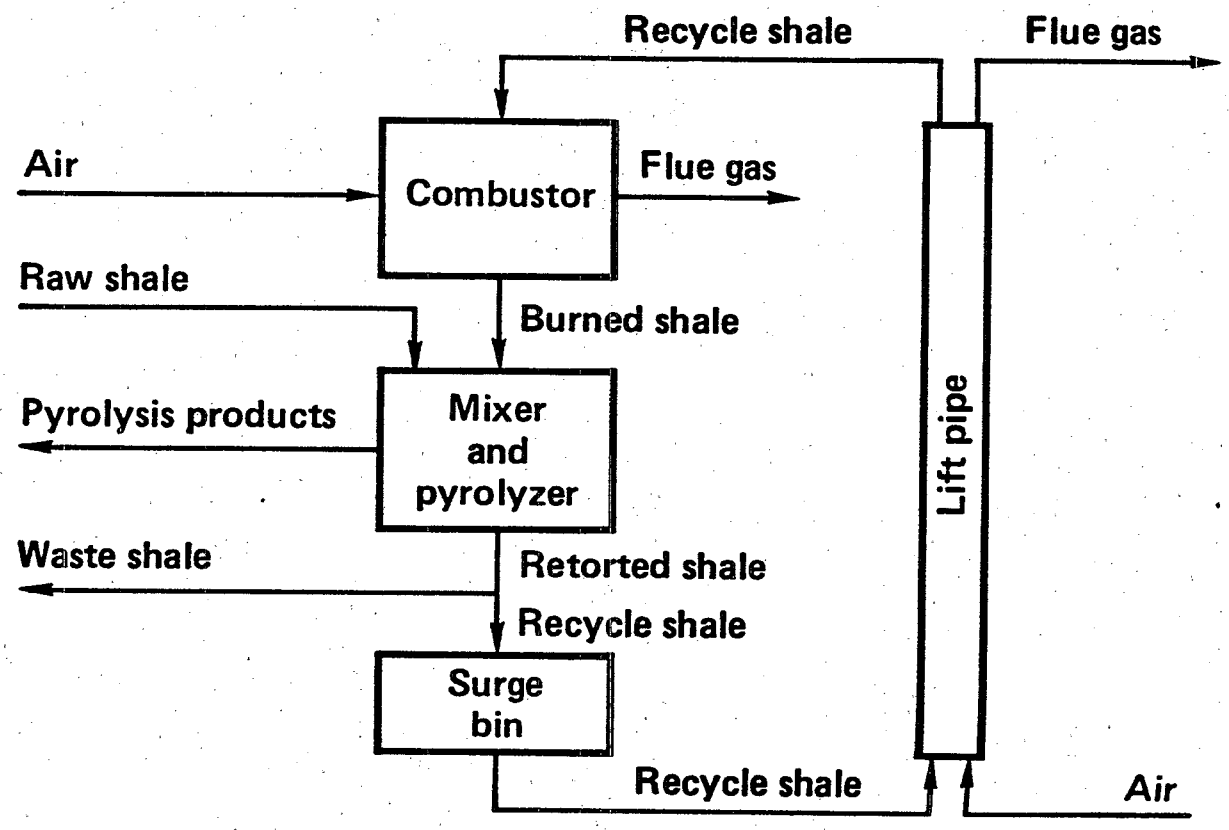


Figure 2-21. Concept for cascading-bed retorting system (Braun, et al, 1985).

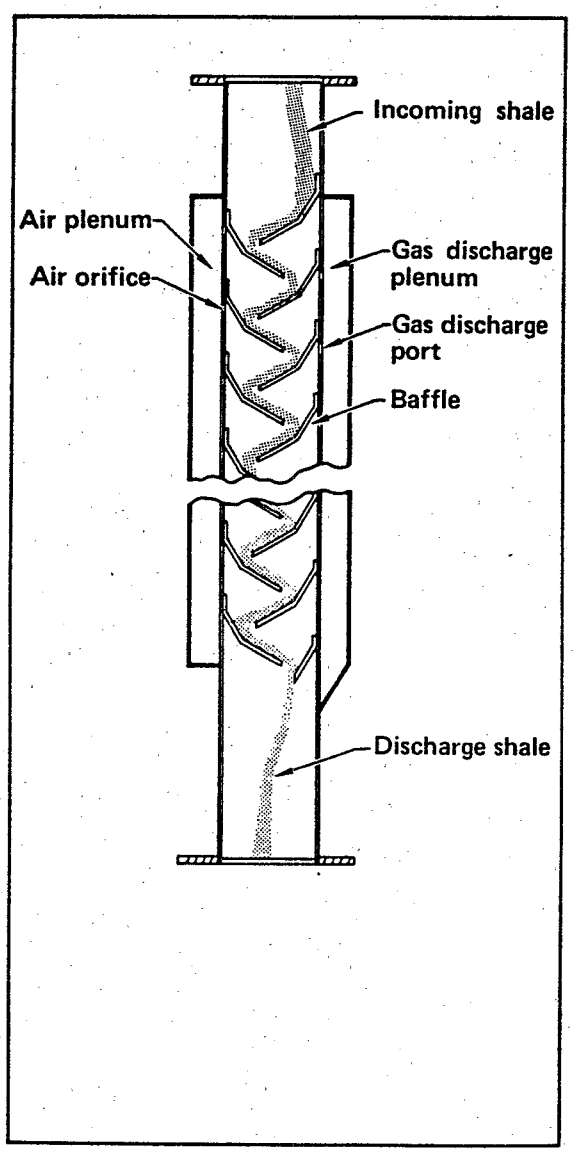


Figure 2-22. Cascading-bed oil shale combustor (Cena, 1985).

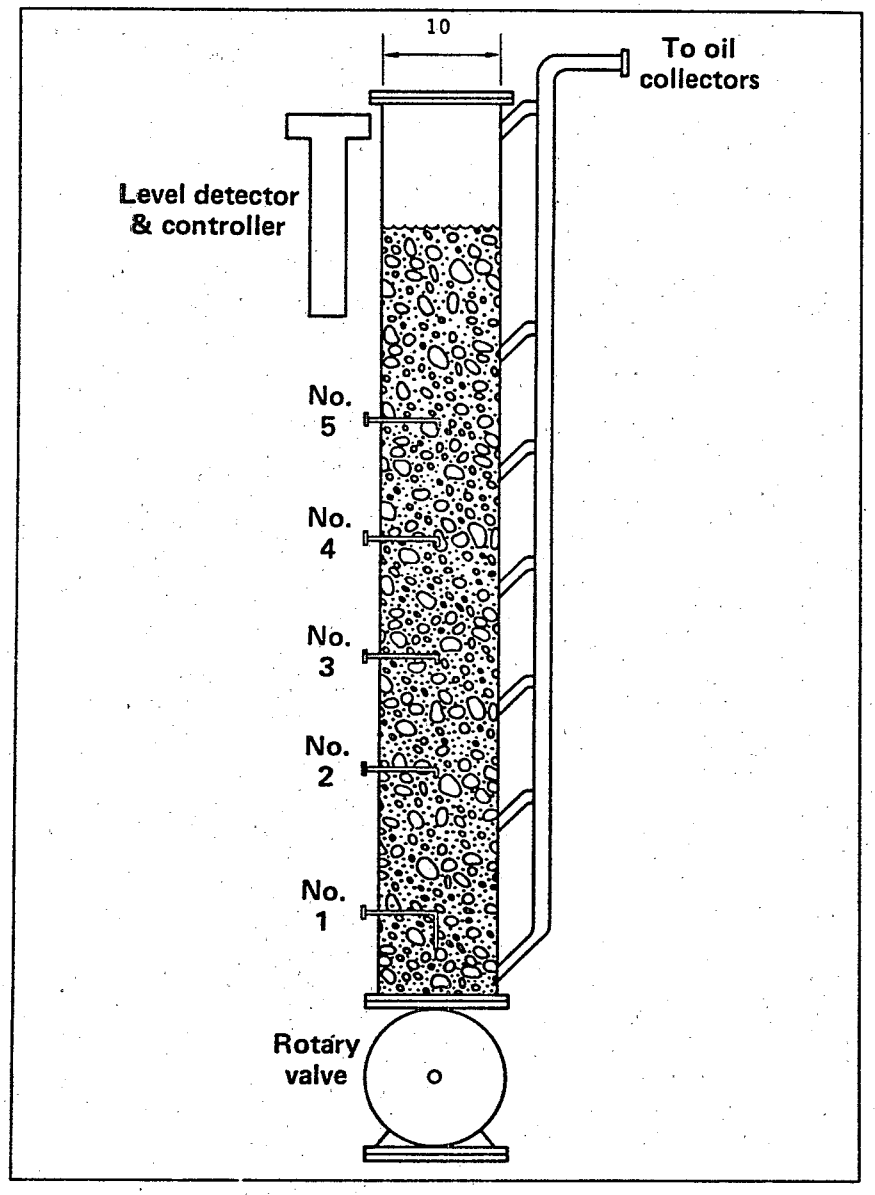


Figure 2-23. Soak tank portion of cascading-bed retort (Cena, 1985).

i.e., processing all of the shale without mining, is difficult to control and consequently presents serious limitations on the recovery of the total potential energy source. [The primary considerations in maximizing yield and minimizing operating costs in in-situ processing are the gas flow rate, temperature profile and gas pressure drop.] The Occidental process is based on the fact that a greater degree of predictability and hence control of both the initial retort bed conditions and the operating parameters can be obtained when the degree of rubblization to prepare the shale for retoring is closely controlled. The method of achieving this control is to first remove 20-30 percent of the shale rock by "room and pillar" mining and then rubblize the remaining shale into the complete volume. The concept is illustrated in Figure 2-24. This technique results in more uniform gas flow rates and easier control over retort conditions to improve yield.

A complete Occidental MIS retort facility consists of both an above-ground retort to recover the oil and gas from the 20-30 percent of the raw shale mined to prepare the in-situ retort. Occidental Shale Oil Corporation has considered various above-ground processes for their Cathedral Bluffs facility, however, they have not made a final decision as to which process they will use.

2.4.11 Horizontal In-Situ Retorting (Geokinetics)

In the GKI horizontal in-situ retorting process, a specific pattern of blast holes is drilled from the cleared surface through any overburden and into the oil shale bed. Explosives are placed in these holes and detonated by use of a carefully timed and planned blast system. The blast yields a well-fragmented mass of shale with high permeability and also produces a slightly sloping (approximately four deg) bottom surface that allows the produced oil to drain into a sump for collection. The fragmented zone constitutes the in-situ retort. The void space in the fragmented zone comes from lifting the overburden, producing a small uplift of the surface as shown in Figure 2-25. Submerged-type oil well pumps are used to lift the recovered oil to surface storage tanks.

Burning charcoal is introduced into drilled holes at the upper end of the rubblized zone to ignite the retort. Air inlet piping is also installed

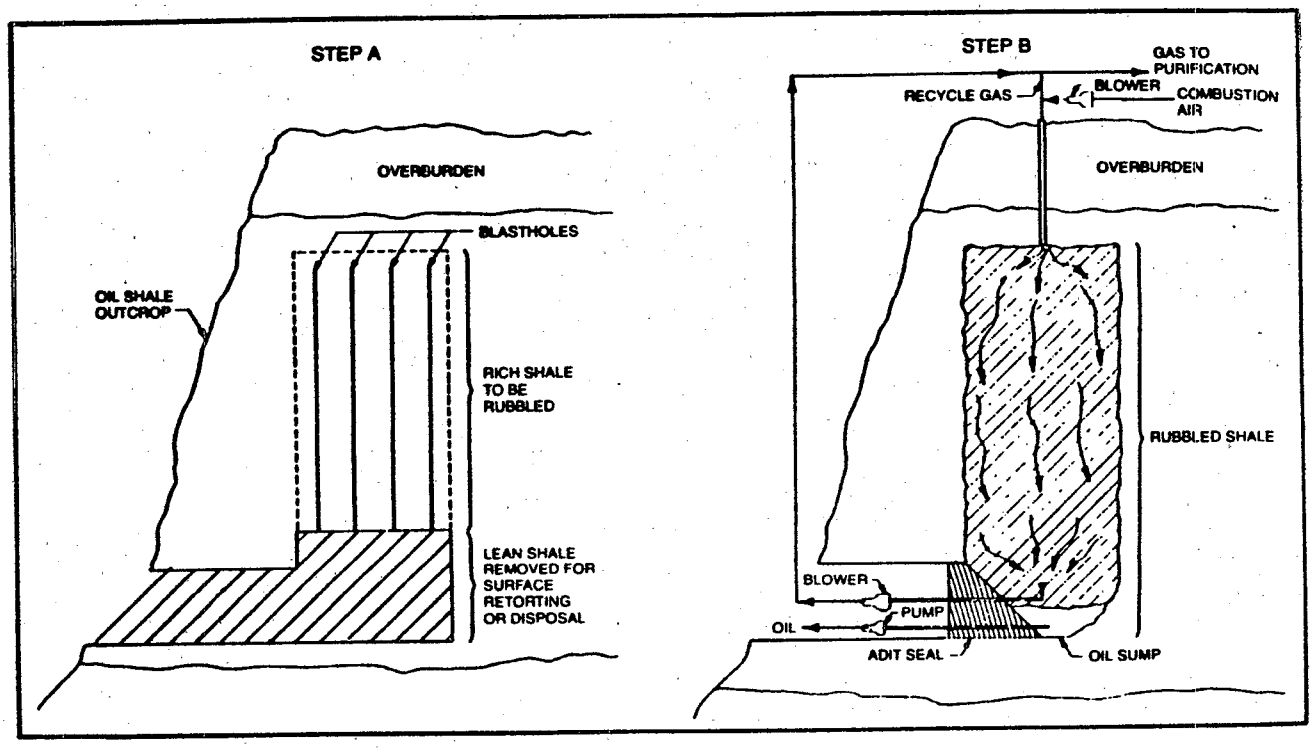


Figure 2-24. Occidental vertical modified in-situ process. (Sladek, 1975. Reprinted by permission of Colorado School of Mines, Golden, CO).

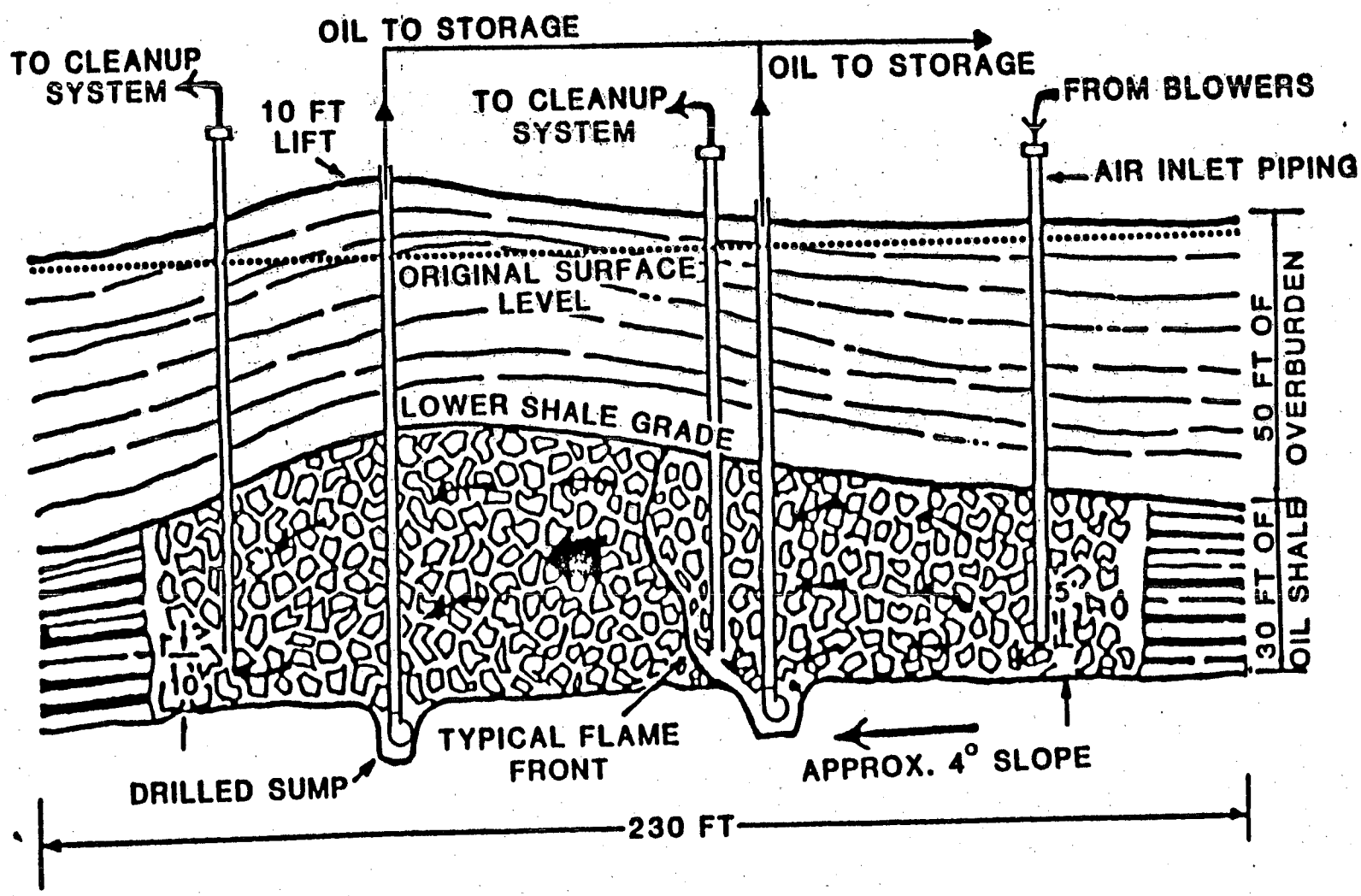


Figure 2-25. An example of a Geokinetics Horizontal In-situ Retort (Taback 1985a).

at this end of the retort. The burn front, consisting of a vertical wall approximately 30-ft high, travels toward the deep or low end of the retort. The objective is to retort the shale from one end to the other in a plug-flow fashion by maintaining a burn front that occupies the entire cross section of the bed. Typically the front travels at speed of one foot per day. At normal production with two retorts operating, the GKI plant produced approximately 400 barrels/day of shale oil.

SECTION 3.0

AIR POLLUTION CONTROL TECHNOLOGY REVIEW

This section is concerned with those air pollution control (APC) methods identified as applicable to the various shale oil production processes and those aspects of its upgrading and refining performed in the vicinity of the production sites. Extensive detailed information is available in the literature on most of these methods. All of those relevant documents and the test references that are especially appropriate for a given method are listed in the bibliography. The objective of this section is to provide the reader with information regarding the alternatives which are available, the expected emission reduction efficiency, the associated cost, descriptions of the method and other decision-making information, such as how extensively the method has been used in the same application and what, if any, problems have been experienced. The reader is also referred to Section 4.0 where the PSD permit applications for a number of shale oil projects are reviewed and the APC methods are identified.

The level of detail for this section is such that a process designer could, on the basis of the information presented, select candidate APC measures (and alternative) to be confirmed in the detailed design. Where the EPA or other agency has a comprehensive report on the subject, this will be noted and the level of detail in this report will be less than for a more recently developed process that may not be as well documented in the literature. Certain shale oil specific problems (such as scrubbing organic sulfur from the retort offgas) are specifically addressed in this report.

To help focus on the various APC processes as applied to oil shale, a matrix (Figure 3-1) was prepared matching each step in the shale oil recovery process with the various APC technologies considered. The matrix serves as an index indicating the section of this report where the technology is presented and also indicating an approximate control efficiency possible with that measure.

	Point Sources																		
	Particulate							NO_x			SO_x								
	Cyclone	Bag-house	Wet Venturi	ESP	Dry Venturi	Enclosures Chutes etc.	Water & Chem. Spray.	Staged Comb.	Amm. Inj.	SCR	Stret-ford	Low-Cat	Uni-sulf	Alk. Scrub.	Act. Carb.	FGD Wet	FGD Dry	Spent Shale Absorb.	Other
Section No.	None	3.1.1A	3.1.1B	3.1.1C	3.1.1D	*	3.1.2	3.2.1	3.2.2A	3.2.2B	3.3.1	3.3.2	3.3.3	3.3.4	3.3.5	3.3.8	3.3.8	3.3.9	
Below-Ground Activities																			
Drilling	●					●	●												
Blasting							●												
Removal	●																		
Conveying		●		●		●	●												
Transferring		●		●			●												
Primary Crush & Screen		●	●	●		●													
Sec. Crush & Screen		●		●		●													
In-mine Vehicles	See Mobile and Fugitive section of this table																		
Mine Vents	●	●	●	●		●			●										
Surface Handling																			
Surface Soils - Removal							●												
- Load							●												
- Haul							●												
- Wind							●												
Raw Shale - Crush & Screen		●																	
- Convey & Transfer		●				●	●												
- Haul						●	●												
- Dump						●	●												
Storage - Deposit						●	●												
- Maintenance							●												
- Removal							●												
- Wind							●												
On-Site Vehicles	See Mobile and Fugitive section of this table																		
Retort Feed Processes																			
Silo - Conveyor		●				●	●												
- Discharge						●	●												

Figure 3-1. Test Sheet (1 of 4).

	Point Sources																		
	Particulate							NO_x			SO_x								
	Cyclone	Bag-house	Wet Venturi	ESP	Dry Venturi	Enclosures Chutes etc.	Water & Chem. Spray.	Staged Comb.	Amm. Inj.	SCR	Stret-ford	Low-Cat	Uni-sulf	Alk. Scrub.	Act. Carb.	FGD Wet	FGD Dry	Spent Shale Absorb.	Other
Section No.	*	3.1.1A	3.1.1B	3.1.1B	3.1.1D	*	3.1.2	3.2.1	3.2.2A	3.2.2B	3.3.1	3.3.2	3.3.3	3.3.4	3.3.5	3.3.8	3.3.8	3.3.9	
Fines - Conveyor & Transfer	●	●	●	●		●													
- Storage	●	●					●												
- Wind							●												
- Maintenance							●												
Retorted Shale		●	●	●		●	●												
- Conveyor						●	●												
- Transfer						●													
- Haul						●													
- Dump							●												
- Storage						●	●												
- Wind							●												
- Maintenance																			
Combustion Sources																			
Retort Gas Treatment	●	●	●	●	●						●	●	●	●	●	●	●	●	
Heaters & Lift Pipes	●	●	●	●	●														
Elutriators (Tosco)	●	●	●	●												●	●		
Moisturers (Tosco)			●																
Upgrading Heaters		●		●	●			●	●	●							●	●	
Power Generation	●	●		●	●			●	●	●							●	●	
Miscellaneous																			
Valves, Flanges, Pumps, etc.																			VOC - Maintenance
API Separators																			VOC - Enclosure
Storage - Crude Shale Oil																			VOC - Vapor Recovery
- Hydrotreated Shale Oil																			VOC - Vapor Recovery
- Fuel Oil																			VOC - Vapor Recovery
- Diesel Fuel																			VOC - Floating Roof Tank
- Gasoline																			VOC - Floating Roof Tank
- Naphtha																			VOC - Floating Roof Tank
Cooling Towers																			Water Treatment

Figure 3-1. Continued (Sheet 2 of 4).

	Mobile and Fugitive Sources										
	Particulate			CO	HC						
	Water Spray.	Chemic. Spray.	Enclosure	Veh. Cat. Mfflr.	Veh. Cat. Mfflr.	Incin.	Floating Roof Tanks	Vapor Rcvry	Maint-ence	Enclosure	Other
Section No.	3.1.2	3.1.2	3.1.2	3.4	3.4	*	3.1.2*	*	*	*	
Below-Ground Activities											
Drilling	●	●	●								
Blasting											Explosive Selection
Removal	●										
Conveying	●		●								Fines to Baghouse or ESP
Transferring		●									
Primary Crush & Screen			●								Fines to Baghouse or ESP
Sec. Crush & Screen			●								Fines to Baghouse or ESP
In-mine Vehicles				●	●						
Surface Handling											
Surface Soils - Removal	●	●		●	●						
- Load	●	●	●	●	●						
- Haul	●	●		●	●						
- Wind	●	●									
Raw Shale-Crush & Screen			●								Fines to Baghouse or ESP
Sec. & Ter. Crush & Screen			●								Fines to Baghouse or ESP
Convey & Transfer	●		●								Fines to Baghouse or ESP
Haul	●	●	●								
Dump	●	●	●								
Storage - Deposit	●		●								
- Maintenance		●		●	●						
- Removal	●			●	●						
- Wind	●	●	●								
On-Site Vehicles				●	●						
Retort Feed Processes											
Silo - Conveyor	●		●								Fines to Baghouse or ESP
- Discharge			●								

Figure 3-1. Continued (Sheet 3 of 4).

	Mobile and Fugitive Sources										
	Particulate			CO	HC						
	Water Spray.	Chemic. Spray.	Enclosure	Veh. Cat. Mfflr.	Veh. Cat. Mfflr.	Incin.	Floating Roof Tanks	Vapor Rcvry	Maint-ence	Enclosure	Other
Section No.	3.1.2	3.1.2	3.1.2	3.4	3.4	*	3.1.2*	*	*	*	
Fines - Conveyor & Transfer	●	●	●								
- Storage	●	●									Fines to Baghouse or ESP
- Wind	●	●	●								
- Maintenance	●	●									
Retorted Shale - Conveyor	●		●								
- Transfer	●		●								Fines to Baghouse or ESP
- Haul	●	●	●	●	●						
- Dump	●	●	●								
- Storage	●	●									
- Wind	●	●	●								
- Maintenance	●	●		●	●						
Miscellaneous											
Valves, Flanges, Pumps, etc.									●		
API Separators								●	●	●	
Storage - Crude Shale Oil							●	●	●		
- Hydrotreated Shale Oil							●	●	●		
- Fuel Oil							●	●	●		
- Diesel Fuel							●	●	●		
- Gasoline							●		●		
- Naphtha							●		●		
Cooling Towers											Water Treatment

*Indicates that technology is not specifically presented in this report because it was considered an incidental aspect of the overall shale oil production process. For example, the control of fugitive HC emissions are standard refinery practice and only incidental to the upgrading of shale oil.

Figure 3-1. Continued (Sheet 4 of 4).

3.1 PARTICULATE EMISSION CONTROLS

3.1.1 Point Source Technologies

A. Baghouse--

The broad terminology for this technology is fabric filtration. However, in most practical applications particulate collection is performed using an array of fabric bags with a suitable positioning and holding device inside of an enclosure called a baghouse. While the general principles of operation are similar to a household vacuum cleaner, the industrial version requires bag materials which will collect submicron particles and withstand the process gas temperature and the wear and tear associated with collection of the trapped particles. Another critical matter is the design of the bag attachment. It must clamp the bag to prevent leakage of the particles under the joint (called sneakage) while avoiding adverse stress on the bags which cause early failure. A properly designed baghouse provides for continuous air flow through the unit while the bags are cleaned sequentially. The bags often are compartmentalized and cleaned a compartment at a time. Often an extra compartment or two are provided to allow for bag replacement or other maintenance while the unit remains in operation.

Baghouse technology has improved rapidly. Except for gases with high temperature, condensed moisture, or certain corrosive constituents it is the preferred point-source control technology. This is especially true since a fine particulate emission standard is in existence. Baghouses achieve a high removal efficiency. Their actual efficiency is a matter of the dust loading; i.e., the higher the dust loading, the higher the removal efficiency. Practically, the baghouse can be characterized better by its outlet dust loading in grains per dry standard (g/dscf) cubic feet or grams per dry cubic meter. Typical performance efficiencies exceed 99.99 percent for particle sizes down to 0.5 micrometers and 99.9 percent below 0.5 micrometer. Dust loadings of less than 0.002 g/m^3 (approx. 0.001 gr/dscf) can be achieved.

1. Process Description--Fabric filter systems typically consist of cloth bags or envelopes, suspended or mounted in such a way that the collected particles fall into a hopper/bin for disposal when dislodged from the fabric. Bags usually have the conveying gas flow from inside the bag to the outside;

envelopes, from the outside in. The accumulated material is dislodged by such devices as bag shakers, a pulse-jet reverse gas-flow (which reverses total air flow in a bag), and a reverse-jet, ring-slit device (which moves along the bag). The sieving action of the dust layer accumulating on the fabric surface predominates shortly after the cleaning operation. However, the slight passage of material through and around the bag attachment during the cleaning cycle is felt to contribute most of the exhaust dust loading.

With shakers or the pulse-jet equipment, removal occurs on a scheduled basis when the flow in a bag or a series of bags is reversed momentarily. Only those bags being cleaned at any moment are out of use; the rest of the unit keeps operating. An isometric view of a two-compartment pulse-jet fabric filter is provided in Figure 3-2. The reverse-jet ring-slit device works continuously to remove the collected particulates as it travels along the bag during filtration. In this technique, bags remain in service while being cleaned. Because it removes most of the accumulation, tightly felted fabric works best, and the initial particulate layer becomes less significant.

Fabric filters can be categorized as to the type of service and frequency of bag cleaning. Intermittent fabric filters are cleaned after filtering is completed (i.e., after the process stream is secured or shutdown) usually at the end of each day. These fabric filters operate with low dust loadings since they cannot be cleaned while on stream. In continuous duty fabric filters, cleaning of a portion of the filters occurs at periodic intervals while the remainder continue to process gases. These filters are more expensive than the intermittent types due to the accessories required in the cleaning process and the additional filter area required for continuous operations.

A variety of fabrics are available for use in filtration systems. A selected listing of materials and their properties is presented in Table 3-1. Tabulated air permeabilities reflect fresh fabric performance and will decrease upon use. Other fabrics which are available include graphitized fiber, and polyethylene. Cotton and nylon, although the least expensive fabrics listed, have poor resistance to acid attack. Orlon, Dacron, and polypropylene resist attack by acids and may be suitable at low temperatures. Fiberglass and Teflon offer higher temperature acid resistance

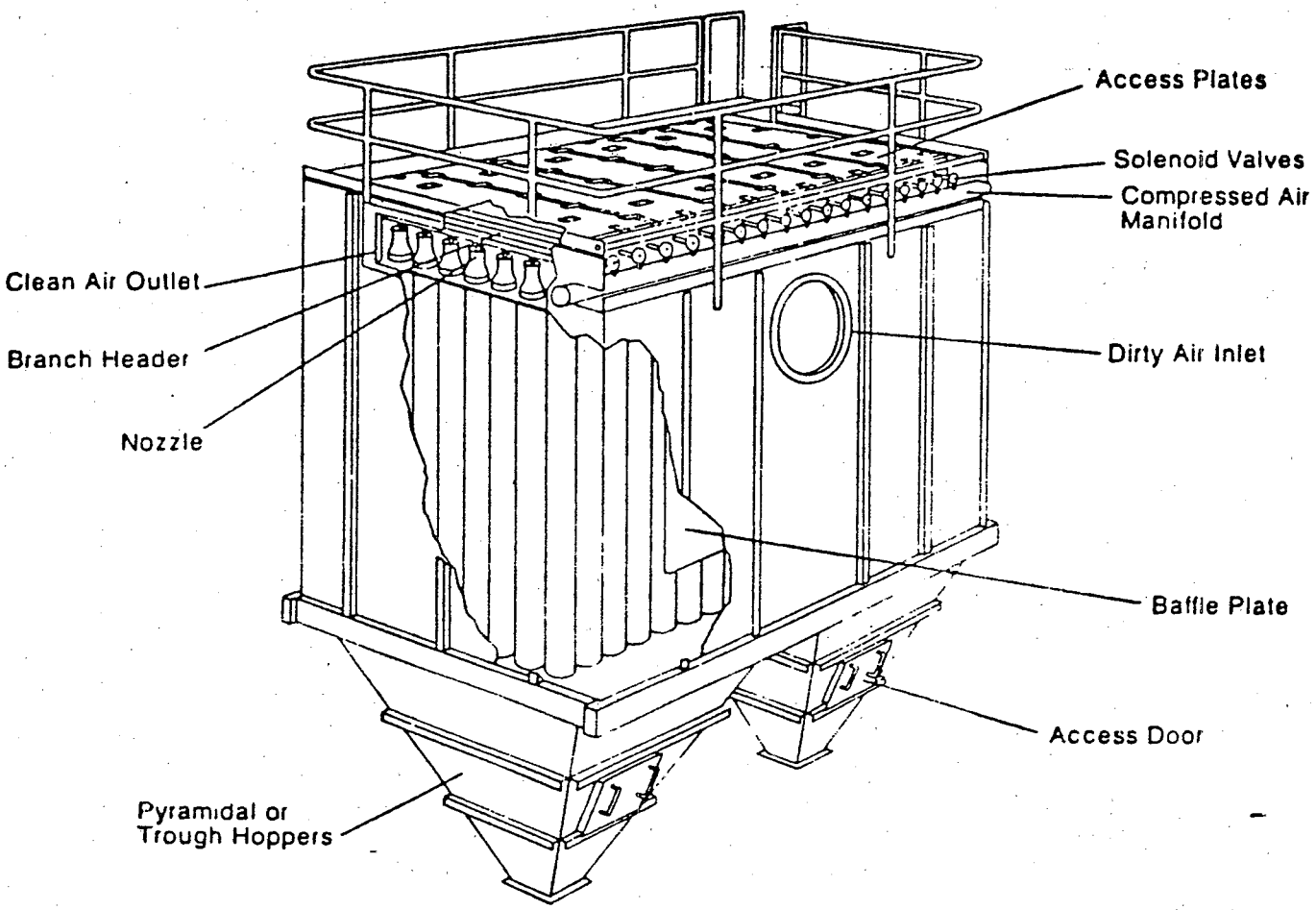


Figure 3-2. Isometric view of a two-compartment pulse-jet fabric filter (Ondich, 1983).

TABLE 3-1. CHARACTERISTICS OF SELECTED FILTER FABRICS (Ondich, 1983)

Fiber	Operating Temperature Exposure, K		Supports Combustion	Air Permeability (m/s)[a]	Composition	Resistance to[b]				Cost Rank[c]
	Long	Short				Abrasion	Mineral Acids	Organic Acids	Alkali	
Cotton	350	380	Yes	0.05-0.10	Cellulose	G	P	G	G	1
Wool	370	390	No	0.10-0.30	Protein	G	F	F	P	7
Nylon**	370	390	Yes	0.07-0.15	Polyamide	E	P	F	G	2
Orlon	390	410	Yes	0.10-0.23	Polyacrylonitrile	G	G	G	F	3
Dacron**	410	440	Yes	0.05-0.30	Polyester	E	G	G	G	4
Polypropylene	370	390	Yes	0.03-0.15	Olefin	E	E	E	E	6
Nomex**	490	530	No	0.12-0.27	Polyamide	E	F	E	G	8
Fiberglass	560	590	Yes	0.05-0.35	Glass	P-F	E	E	P	5
Teflon**	500	530	No	0.07-0.33	Polyfluoro-ethylene	F	E	E	E	9
GORE-TEX® Membrane	500	530	No		Teflon	E	E	E	E	10

[a] m/s at a pressure drop of 125 Pa
[b] P = Poor, F = Fair, G= Good, E = Excellent
[c] Cost rank, 1 = lowest cost, 10 = highest cost

** DuPont registered trademark

-- although Teflon is more costly. A Teflon or a Teflon-Orlon mixture called MTI may be used when significant concentrations of fluorides are present. With the exception of wool and fiberglass, most listed materials provide good resistance to attack by alkaline substances. A recently-developed bag material, GORE-TEX membrane, is a micro-porous membrane of expanded polytetrafluoroethylene (PTFE) laminated to the surface of various fabric substrates. It has been found to exhibit excellent filtration properties. The special property of this bag material is that it has high membrane surface filtration and does not depend on the buildup of a filter cake. Thus, the removal efficiency remains high immediately after the cleaning cycle and before a new filter cake has been formed. This material is a proprietary product of W.L. Gore & Associates, Inc., Elkton, MD.

The applicability of fabric filters depends upon the emission limitations, temperature, moisture content, the bag replacement schedule of the fabric filter, and other aspects. Temperature of the gas imposed limitations upon usable materials of construction for bags, as well as affecting system size with regard to volumetric flow. Moisture or other condensibles can affect performance of the fabric filter and may render it inoperative.

Laboratory studies with fabric filters have demonstrated a strong correlation between outlet concentration and face velocity (air-to-cloth ratio) for a given loading and type of fabric. Predicted and observed outlet concentrations for bench scale tests indicate that outlet concentration increases to a certain extent and then tapers off as air-to-cloth ratio (face velocity or fabric loading) is increased. The results of these tests are presented in Figure 3-3. This effect has also been substantiated by field pilot studies (refer to Figure 3-4) although there are some inconsistencies which could be due to control problems in field experimentation.

The pressure drop across the baghouse generally varies as the square of the face velocity (air-to-cloth ratio) at constant dust loadings. However, it is difficult to relate pressure drop directly to the collection efficiency of the unit. The filter pressure drop is influenced by the frequency and completeness of bag cleaning. The effects of cleaning frequency on the filter pressure drop for a reverse air-jet type of cleaning mechanism are shown in Figure 3-5. At the lower dust loading value of 1 g/am^3, this curve indicates

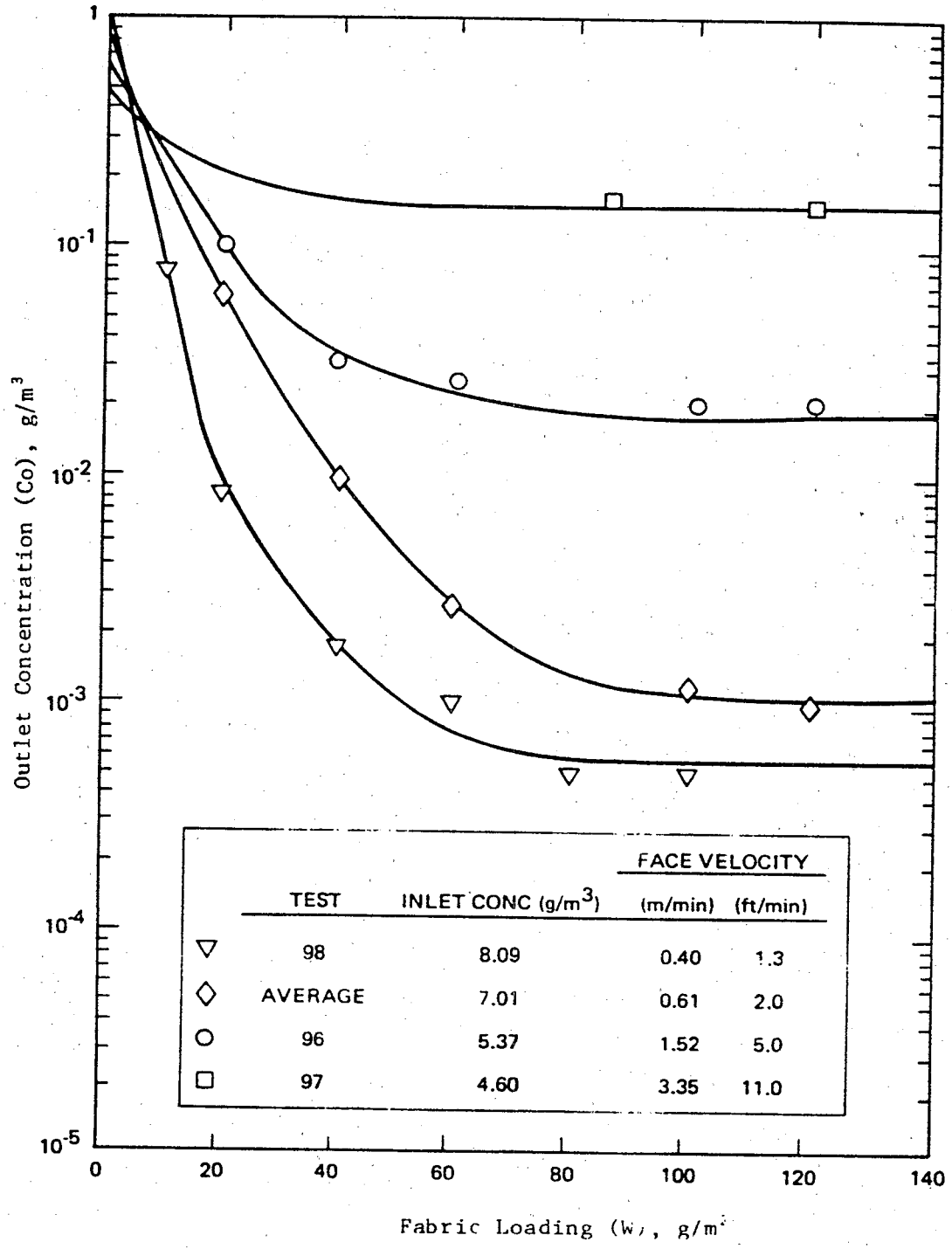
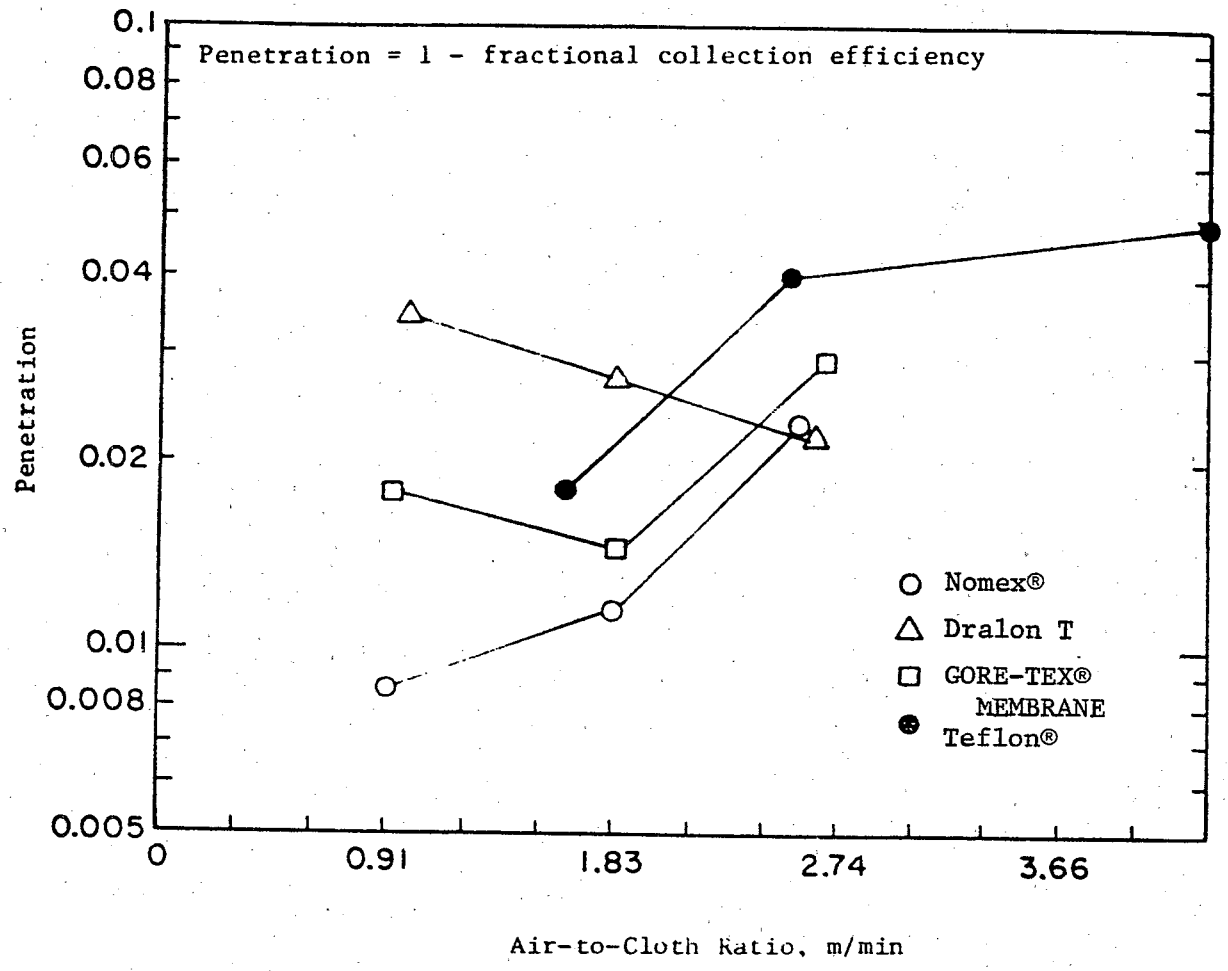


Figure 3-3. Predicted and observed results for bench scale fabric filter tests. GCA fly ash and Sunbury fabric (Ondich, 1983).



Nomex® and Teflon® are registered trademarks of E. I. duPont de Nemours & Co., Inc.

GORE-TEX® MEMBRANE is a registered trademark of W.L. Gore & Associates,Inc.

Figure 3-4. Fabric filter field test results for different bag materials (Ondich, 1983).

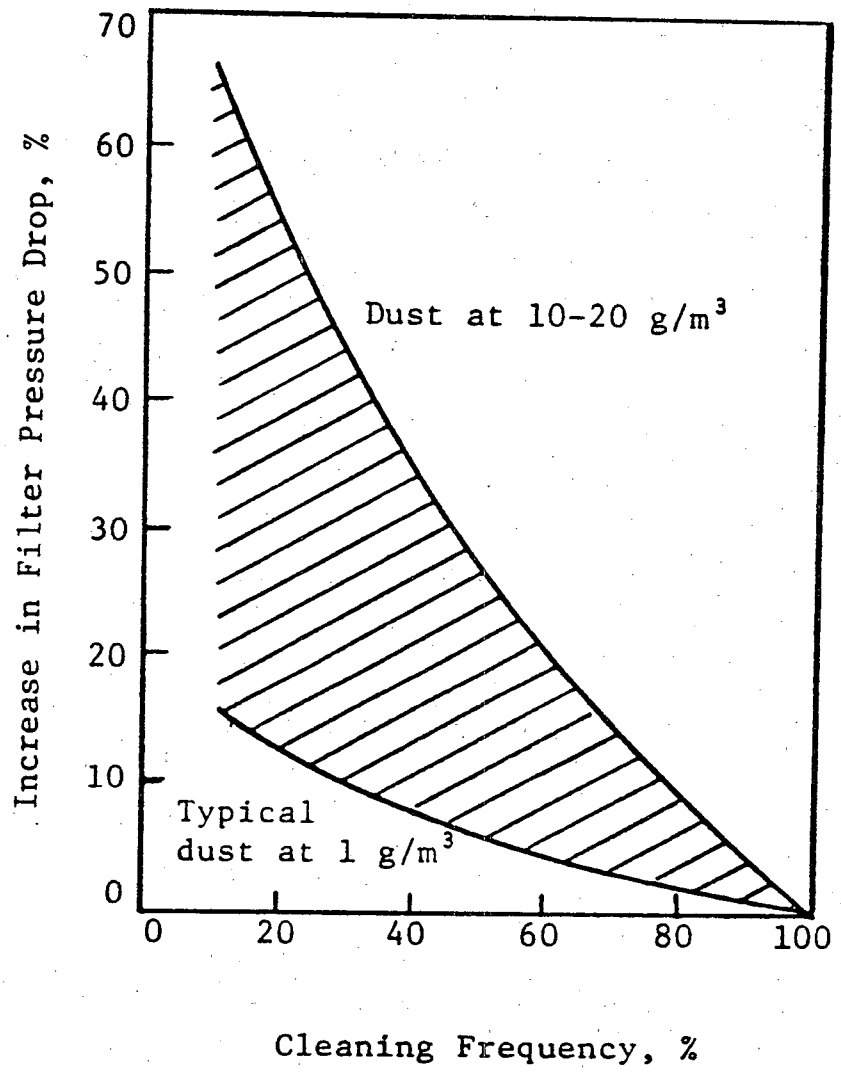


Figure 3-5. Filter pressure drop for reverse air-jet operation (Ondich, 1983).

that operating the cleaning mechanism only 30 percent of the time causes an increase in pressure drop of only 10 percent. Because of the wear on most fabrics imposed by cleaning, these values indicate the desirability of accepting minimal pressure drop increase while reducing bag maintenance. The need for more frequent cleaning at higher dust loadings also is indicated by the data shown in Figure 3-5.

The design of a baghouse requires careful consideration and analysis of the above factors along with the information and advise of reliable equipment suppliers who will ultimately warrant the unit's performance. A smaller air-to-cloth ratio results in a large and more expensive unit. However, this is offset by lower energy and bag replacement cost because of the lower pressure drop and bag cleaning frequency. Cleaning frequency involves a tradeoff of bag wear versus pressure drop. This is somewhat subjective and depends on the equipment supplier's recent experience with his attachment design and selected cleaning method. Acid and moisture conditions in the gas must be considered in selecting case material.

Since most users prefer to purchase a unit on a competitive basis, it is advisable to allow as much design freedom as possible consistent with obtaining the desired performance guarantees backed up by demonstrated achievements. Failure to achieve a performance guarantee may be as much of a hardship on the user as the supplier if the entire process is down due to a malfunctioning baghouse.

The time to isolate a broken bag can be reduced drastically if detectors are placed on the outlet of each compartment. These detectors actually indicate the opacity of the gas stream from the compartment and allow operators to isolate quickly the faulty compartment for repair.

Leaks through bypass dampers also affect performance since leakage can approach five percent. Leakage can be reduced essentially to zero by including two louvered dampers with a continuous purge of clean reverse air to block uncleaned gas.

2. Process Economics -- Costs of fabric filter baghouses depend upon: (1) type of fabric, (2) air-to-cloth ratio, (3) intermittent or continuous duty, (4) pressure or suction type construction, (5) standard or custom design, (6) type of cleaning mechanism, and (7) materials of construction.

Costs for mechanical shaker, pulse-jet, reverse-air, and custom baghouses are presented in Figures 3-6 though 3-10. The prices are based on net cloth area which is determined from the specified air-to-cloth ratio. The net cloth area is the average filter area seen by the air flow. This may differ from the gross cloth area (which is the total cloth area in the baghouse) due to the need to isolate cloth area when continuous cleaning is used. For an intermittent system, the net and gross cloth areas are the same, since the baghouse is cleaned after a filtering cycle. For continuous filters, factors to calculate gross cloth area are provided in Table 3-2. Using this calculated gross area and type of filter desired, bag costs can be determined from Table 3-3.

Baghouse prices are flange-to-flange, including the basic baghouse without bags, three meter support clearance, and inlet and exhaust manifold. Pressure baghouses are designed for three kPa gauge pressure and suction fabric filters are designed for a negative pressure of five kPa gauge. Custom baghouse prices are more a function of specific requirements than of pressure or suction construction, so prices do not differentiate between pressure or suction. Custom baghouses are designed for continuous operation and normally use reverse air cleaning. All baghouses except custom baghouses are generally factory-assembled. The operating requirements for fabric filters primarily consist of labor, replacement bags, and power. Labor requirements include two to four man-hours per shift for operations and one to two man-hours per shift for maintenance. Bag life is estimated at 0.3 to 5 years with an average of 1.5 years. Power requirements for shaker and blower motors are given as approximately 4 kW/1000 m^2 of cloth area. Power usage will depend on dust loading and cleaning cycle.

B. Venturi Scrubber--

1. Process Description--A Venturi scrubber is one of several types of wet scrubbers and is the most commonly used for large scale control of particulates. It uses the venturi effect to create water droplets in the high

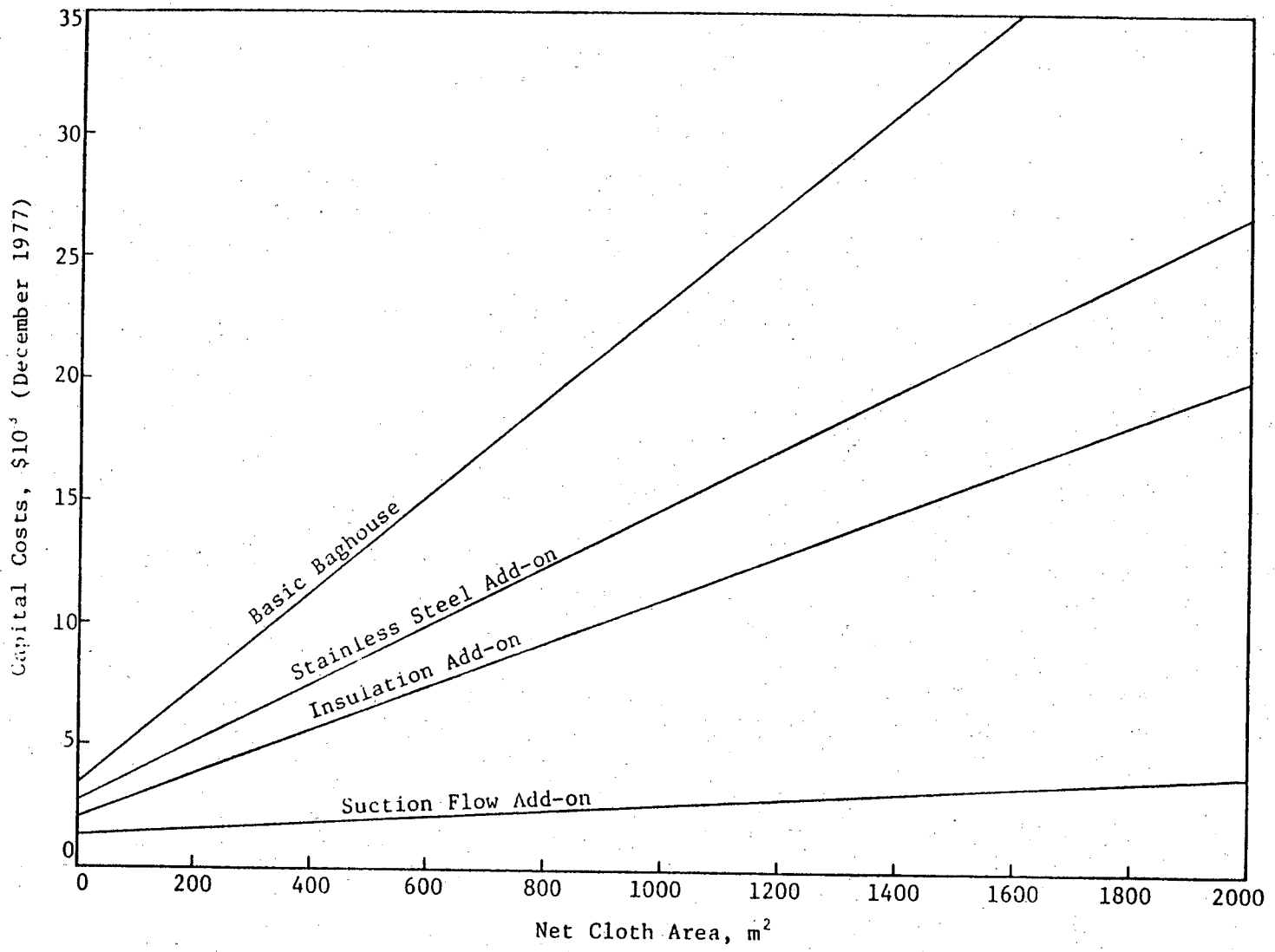


Figure 3-6. Costs for pressure flow, intermittent duty, mechanical shaker standard construction baghouses, exclusive of bags (Ondich, 1983).

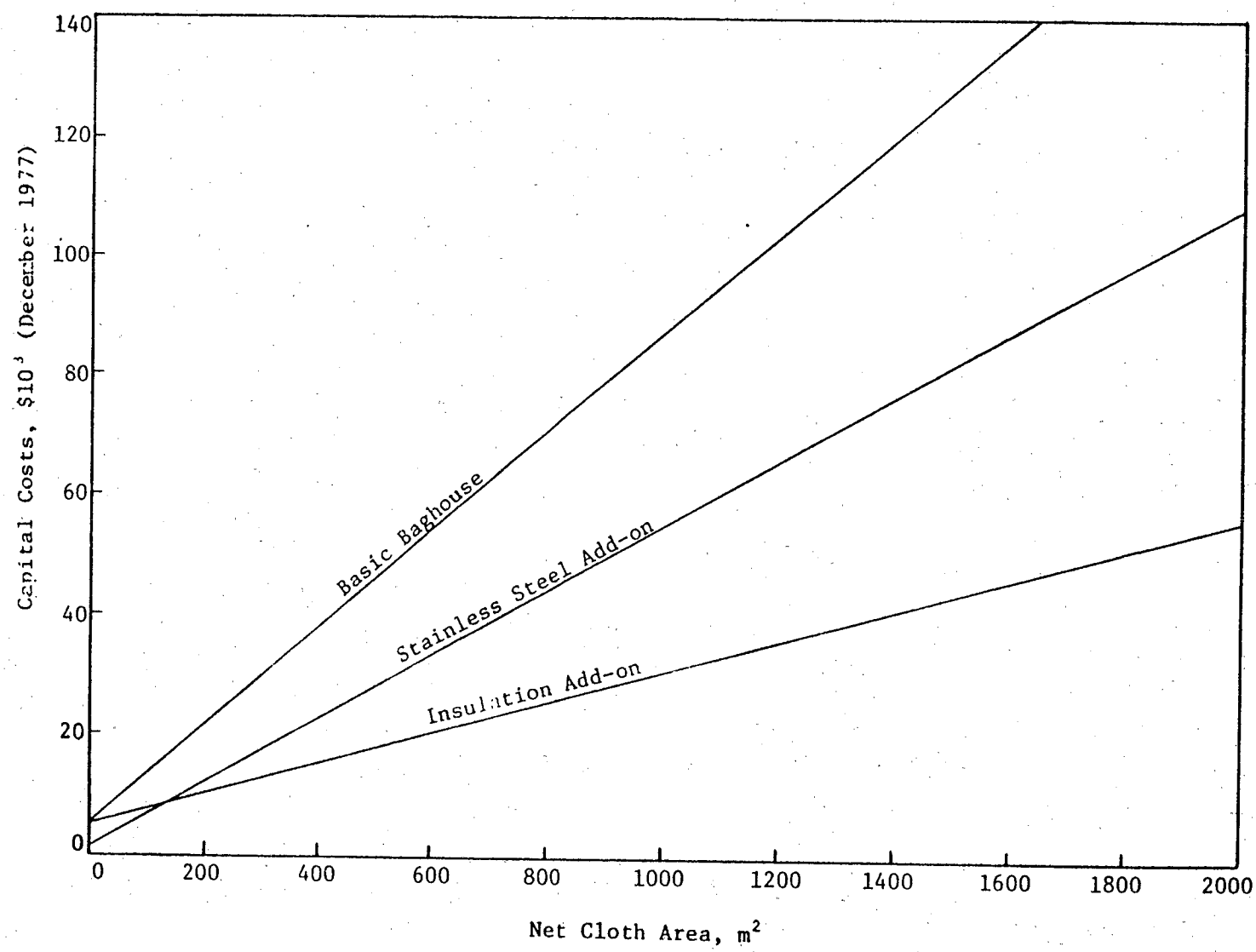


Figure 3-7. Costs for continuous duty, pulse jet standard construction baghouses, exclusive of bags (Ondich, 1983)..

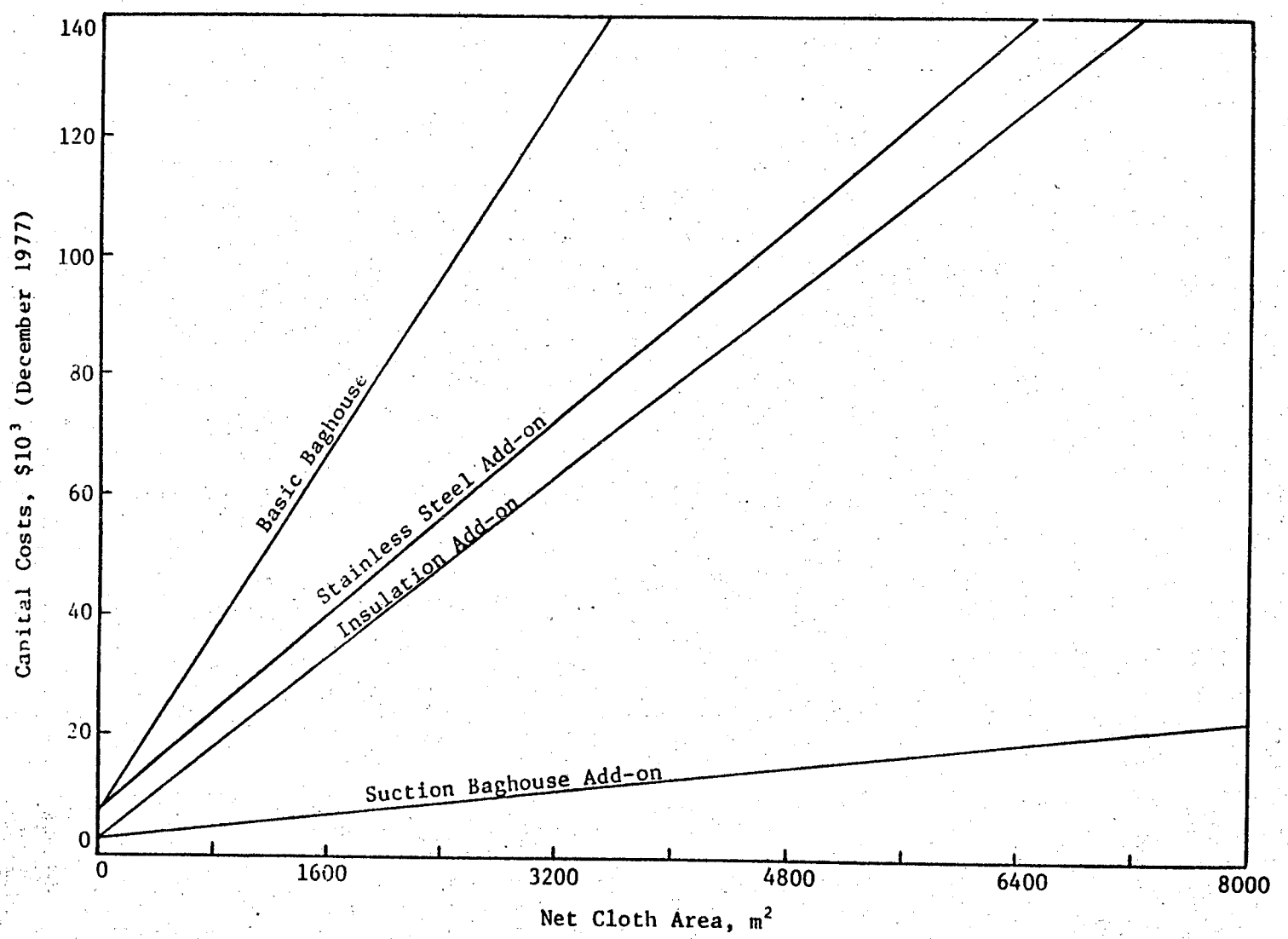


Figure 3-8. Costs for continuous duty, pressure flow, mechanical shaker standard construction baghouses, exclusive of bags (Ondich, 1983),

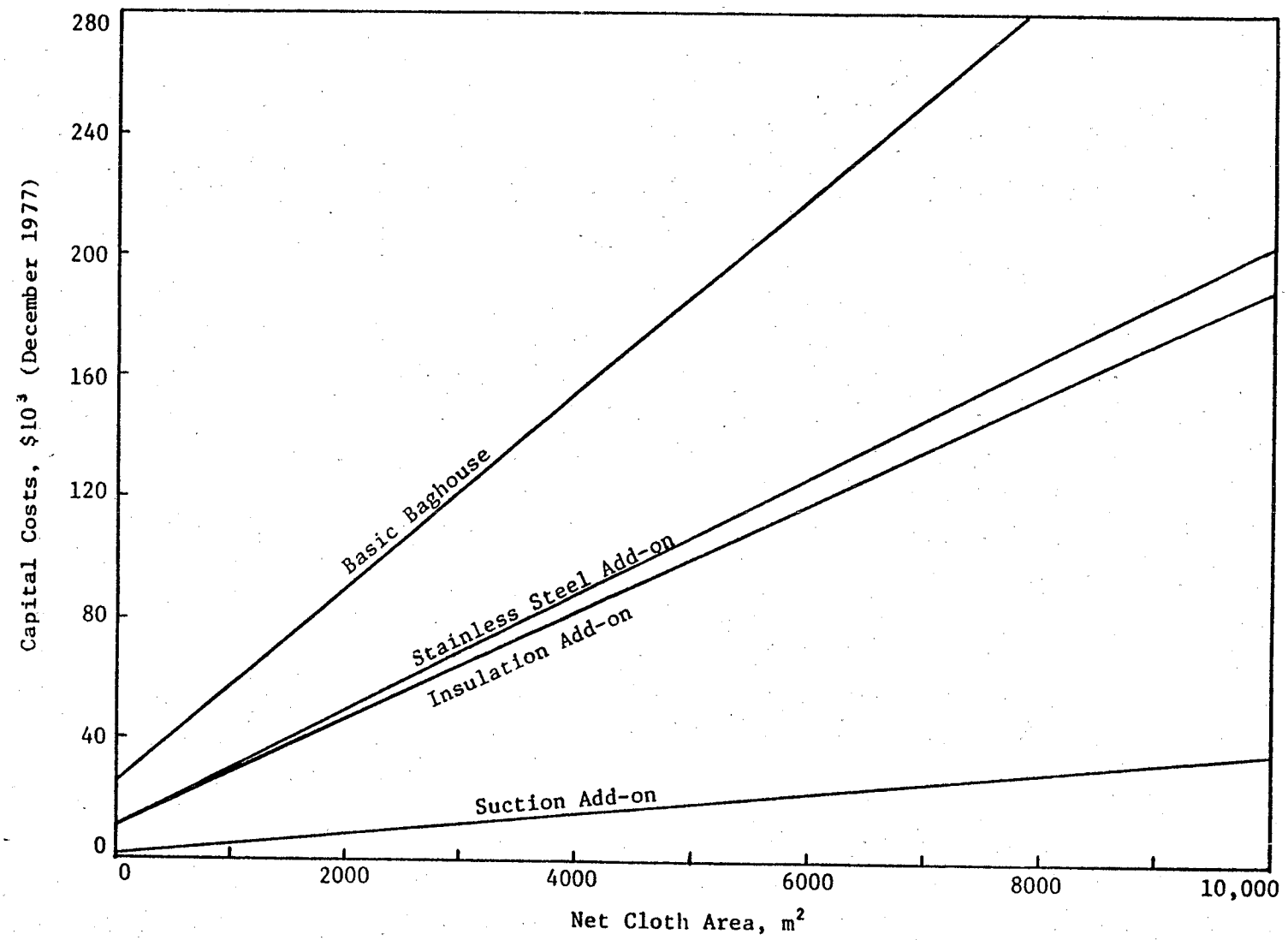


Figure 3-9. Costs for continuous duty, pressure flow, reverse air standard construction baghouses, exclusive of bags (Ondich, 1983).

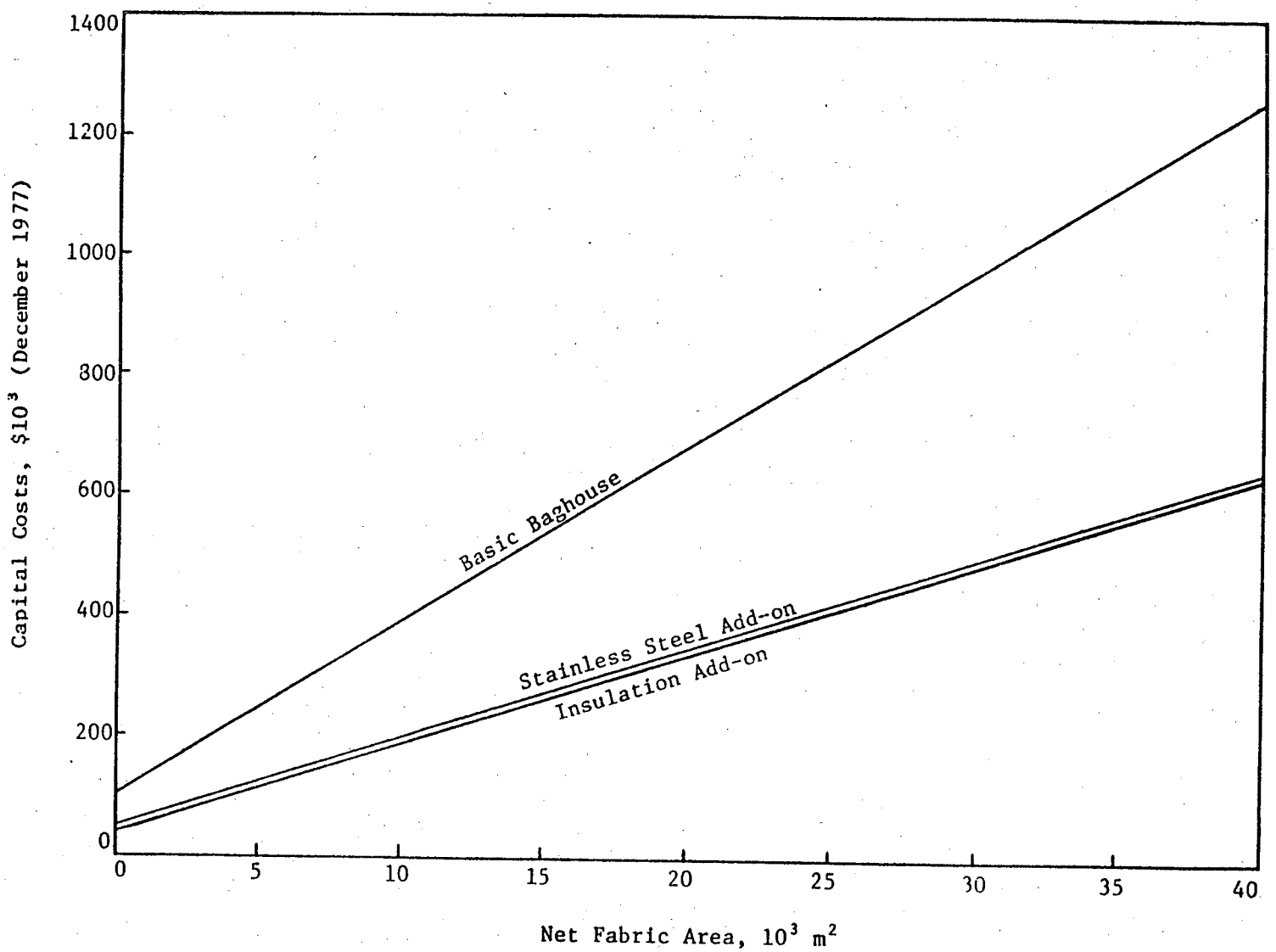


Figure 3-10. Costs for continuous duty, reverse air custom construction baghouses, exclusive of bags (Ondich, 1983).

TABLE 3-2. APPROXIMATE GUIDE TO ESTIMATE GROSS CLOTH AREA
(Ondich, 1983)

Net Cloth Area, m^3	Gross Cloth Area m^3
1 - 370	Multiply by 2
370 - 1100	Multiply by 1.5
1100 - 2200	Multiply by 1.25
2200 - 3300	Multiply by 1.17
3300 - 4400	Multiply by 1.125
4400 - 5600	Multiply by 1.11
5600 - 6700	Multiply by 1.10
6700 - 7800	Multiply by 1.09
7800 - 8900	Multiply by 1.08
8900 - 10,000	Multiply by 1.07
10,000 - 12,300	Multiply by 1.06
12,300 - 16,700	Multiply by 1.05
16,700 On Up	Multiply by 1.04

TABLE 3-3. BAG PRICES ($\$/m^2$)[a]

Class	Type	Dacron	Orlon	Nylon	Nomex®	Glass	Poly propylene	Cotton	GORE-TEX® Membrane
Standard	Mechanical shaker, <1900 m^2	7.30	11.90	13.80	21.00	9.20	11.90	8.10	c
	Mechanical shaker, >1900 m^2	6.50	9.20	12.70	19.20	8.10	10.00	7.30	c
	Pulse jet [b]	11.00	17.30		24.00		12.70		c
	Reverse air	6.50	11.00	12.70	19.20	8.10	10.00	7.30	c
Custom	Mechanical shaker	4.60	6.50	8.10	11.90	5.40	6.50	7.30	c
	Reverse air	4.60	6.50	8.10	11.90	5.40	6.50	7.30	c

[a] Based on prices in Ondich (1983) adjusted for inflationary factors to 1986.

[b] For heavy felt, multiply price by 1.5

[c] GORE-TEX® Membrane is applied as a laminate over any of the listed fabrics. The GORE-TEX® membrane increases the bag cost by a factor of 3 to 4.5. A custom glass bag with GORE-TEX® membrane would cost approximately $\$20/m^2$ (J. R. Darrow, W. L. Gore & Associates, Inc., 1986).

GORE-TEX® Membrane is a registered trademark of W. L. Gore & Associates, Inc.

Nomex® is a registered trademark of E. I. du Pont de Nemours & Co., Inc.

velocity gas stream. These droplets coalesce and collide with and scrub the particulates. Use of baffles and mist eliminators in conjunction with the velocity decrease returns the droplets with entrained particulates to the liquid phase.

In venturi type scrubbers (Figure 3-11) mixing is achieved by using fans to accelerate the incoming gas stream to velocities in the range of 46 to 120 m/second in the converging venturi section. As high velocity gas enters the throat of the venturi, it encounters a stream of scrubbing liquor flowing down the walls of the chamber. The liquor is introduced by flooding, sprays, or weirs, and a spray of atomized liquid is directed into the gas stream ahead of the venturi throat. As the thin sheet of liquid reaches the venturi throat, it is sheared off by the gas stream to become a mass of droplets entrained in the atomized sprays. Typically these droplets are 25 to 100 μm in size. In the turbulent zone just beyond the entrance to the throat, these relatively massive droplets move much more slowly than dry particulates in the gas stream. As a result, collisions between droplets are frequent, so most particles are captured by the scrubbing liquor.

As droplets with their burden of particulates move through the venturi, the gas/droplet mixture decelerates, and the droplets collide with each other to form even larger and heavier drops. When they enter the next stage, these are separated from the gas in a cyclonic (or other) separation device.

Inertial impaction is the most important mechanism for collection of fine particulates. As particle size decreases, theory indicates that a higher pressure drop is required to maintain a given particle removal efficiency. For a given dust, as the pressure drop is increased, finer water droplets are atomized to interact with the dust particles through impingement and agglomeration with a consequent increase in removal efficiency. Increasing the pressure drop can be accomplished by either increasing the gas stream throat velocity, increasing the scrubber liquor flow rate, or both. The relationship between particle size and pressure drop is shown in Figure 3-12. The relationship between pressure drop and collection efficiency is the same for all types of venturi scrubbers irrespective of the size, shape, or general configuration of the scrubber. Venturi scrubbers normally are operated at

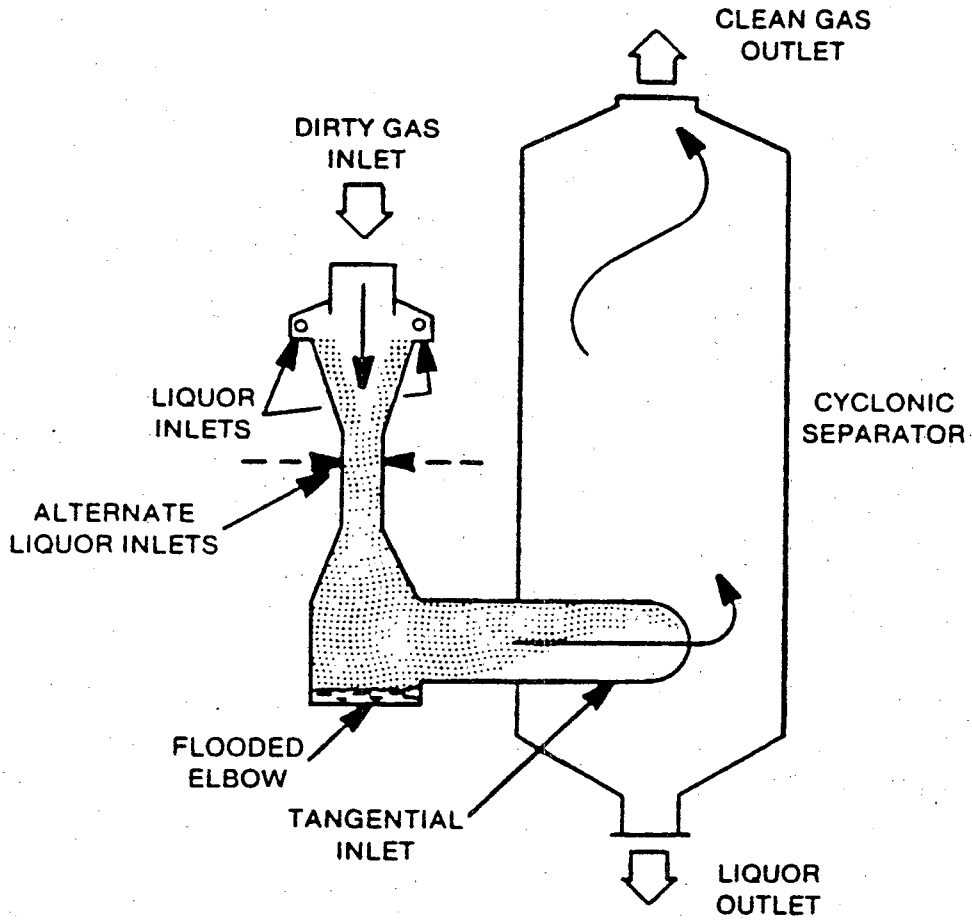


Figure 3-11. Standard Venturi Scrubber (Ondich, 1983).

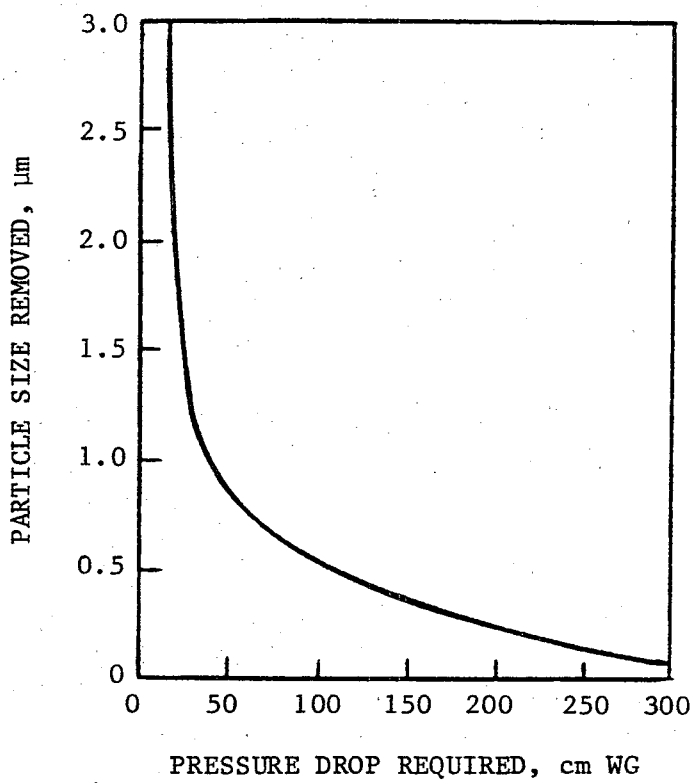


Figure 3-12. Efficiency performance of venturi scrubbers. (Ondich, 1983).

pressure drops of between 1.5 and 20 kPa, depending on the characteristics of the dust, and at liquor flow rates of 0.4 to 2.7 L/min per gas flow of 1 m^3/min. The collection efficiencies range from 99+ percent for one µm (micron) or larger sized particles to 90 to 99 percent for particles below one µm size.

A separator for removal of the agglomerates from the gas stream is provided downstream of the scrubber. These separators are usually of the cyclone type in which the gas stream and agglomerates are given a cyclonic motion which forces the liquid and particles to impinge on the walls of the separator by centrifugal force. The separator normally consists of a cylindrical tank with a tangential inlet located at the lower side of the tank and an exhaust outlet located at the top of the tank on the centerline axis. A cone bottom with outlet is provided to collect the liquid slurry. The collected particles settle to the bottom of the cone and are sent to the water treatment facility while the cleaner liquid above the sediment is removed and recycled to the scrubber.

For hot processes, a considerable amount of water is vaporized in the scrubber and upstream equipment (e.g., quencher) which must be handled by the fan. Although the gas volume is reduced, a large portion remains as water vapor which results in higher horsepower requirements and in higher operating costs. To alleviate this condition, a gas cooler can be incorporated into the separator to cool and dehumidify the gas stream. Several types of gas coolers are used for this purpose; one type employs spray banks of cooling water followed by impingement baffles while a second type utilizes flooded plates or trays with either perforated holes or bubble caps to permit passage of the gas stream through the cooling water bath. Several plates or trays can be used in sequential stages to provide the necessary cooling and contact time.

Scrubber selection for particulate emission control depends upon the particulate loading, particle size distribution, and required removal efficiency. Generally, venturi scrubbers are used in those process areas where wet collection is necessary or appears desirable due to the characteristics of the gas stream such as the presence of tar, gums and other materials that tend to blind or plug bag filters. Table 3-4 gives typical values of pressure drops for selected processes and equipment where venturi scrubbers are utilized for particulate control.

TABLE 3-4. TYPICAL PRESSURE DROPS FOR VENTURI SCRUBBING SYSTEMS (Ondich, 1983)

Process	Pressure Drop, kPa
Basic oxygen furnaces	10.0 - 15.0
Brick manufacturing	0.7 - 8.7
Clay refractories	2.7
Coal-fired boilers	3.7
Detergent manufacturers	2.5 - 10.0
Ferroalloy plants	10.0 - 19.9
Glass manufacturing	16.2
Gray iron foundaries	6.2 - 14.9
Kraft recovery furnaces	3.7 - 7.47
Lime kilns	3.0 - 10.0
Petroleum catalytic cracking	10.0
Phosphate fertilizer	3.7 - 7.47
Phosphate rock crushing	2.5 - 5.0
Secondary aluminium	7.5

Venturi scrubbers produce a significant quantity of liquid waste which may be discharged to a settling pond or to a water treatment plant after solids removal. The quantities of wastewater discharged are difficult to predict since systems use different L/G ratios as well as different degrees of recirculation. A venturi scrubber on a pulverized coal boiler operating at an L/G ratio of 0.9 L/m^3 with no recirculation will discharge about 2000 L/min. Usually this discharge is pumped to a settling pond where the fly ash settles to the bottom and the liquid is either discharged, evaporated, or recycled. Pond liners may be used to prevent leaching of any metals or chemicals into the soil and surrounding water table. Although intrusion upon a local water body or supply is always possible, good operating procedures can minimize this potential impact. Since the properties of pond discharge waters are dependent upon the process particulate and gas composition, it is not possible to specify average values.

2. Process Performance--Collection performance of venturi scrubbers is affected by the following parameters: (1) pressure drop, (2) liquid-to-gas ratio (L/G), (3) gas velocity, and (4) particle size distribution.

Particle cut diameter is the diameter of the particle which will be collected by the scrubber at 50 percent efficiency. Particle-cut diameter is a frequently used parameter for describing the particle collection efficiency of venturi scrubbers. One reason for this is that plots of collection efficiency versus particle diameter tend to be rather steep in the region where inertial impaction is the predominant collection mechanism. A plot of cut diameter versus pressure drop for a gas-atomized spray scrubber is provided in Figure 3-13. This plot is based on experimental data from large venturis, orifices, and rod-type units, plus mathematical model. A plot of particle size vs pressure drop has been illustrated earlier in Figure 3-12.

The relationship between liquid-to-gas ratio and cut diameter for venturis has been plotted (Ondich, 1983) with a liquid density of water, 1000 kg/m^3, and a gas viscosity of 1.8×10^{-5} Pa·s (equal to the viscosity of air at one atmosphere and 298 K). Plots of liquid-to-gas-ratio versus cut diameter with gas velocity and pressure drop as parameters are presented in Figure 3-14. The factor f, the ratio of relative velocity between gas and liquid droplet to gas velocity, is approximately 0.25 for fly ash and

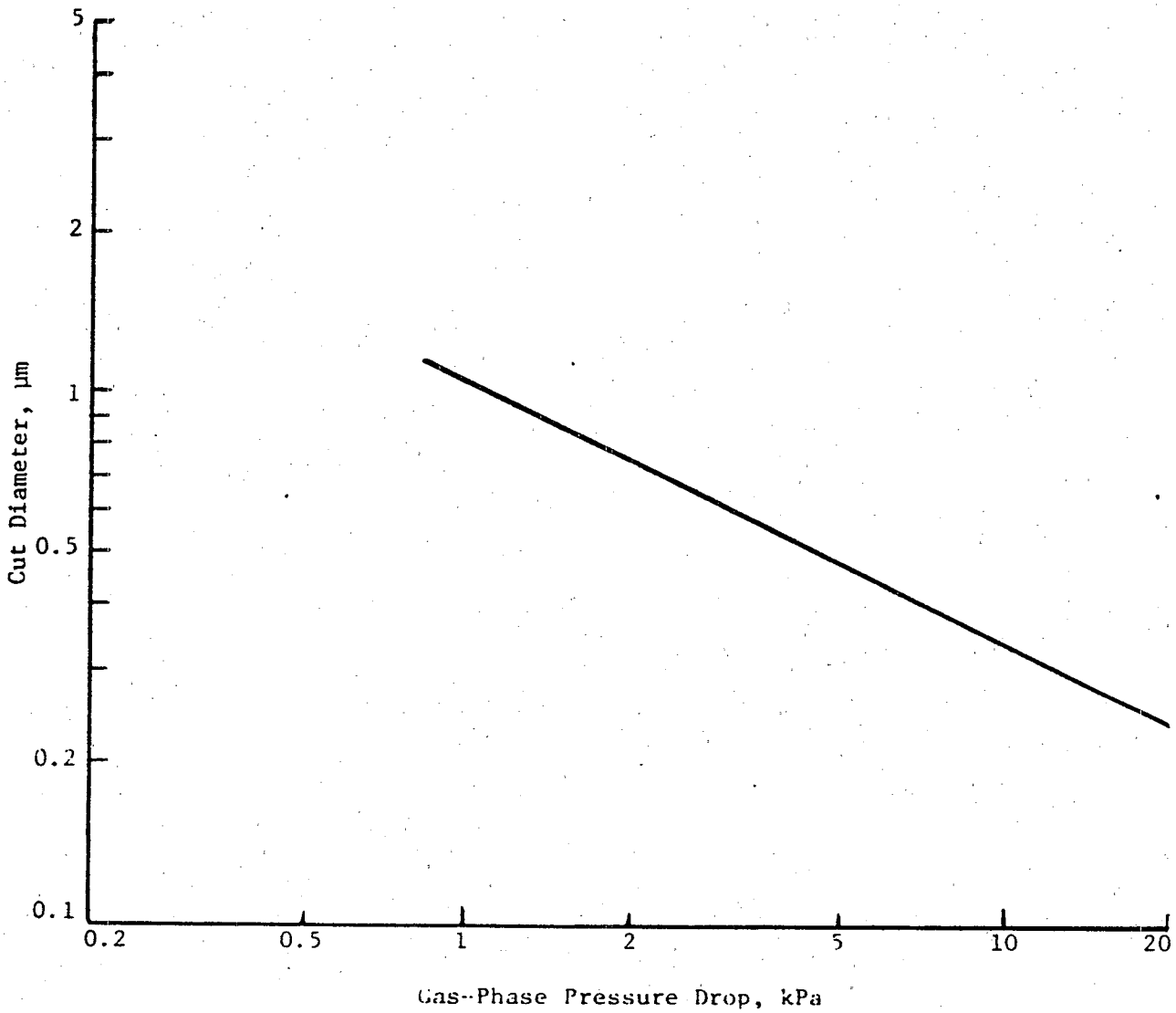


Figure 3-13. Cut diameter for gas-atomized spray scrubbers (Ondich, 1983).

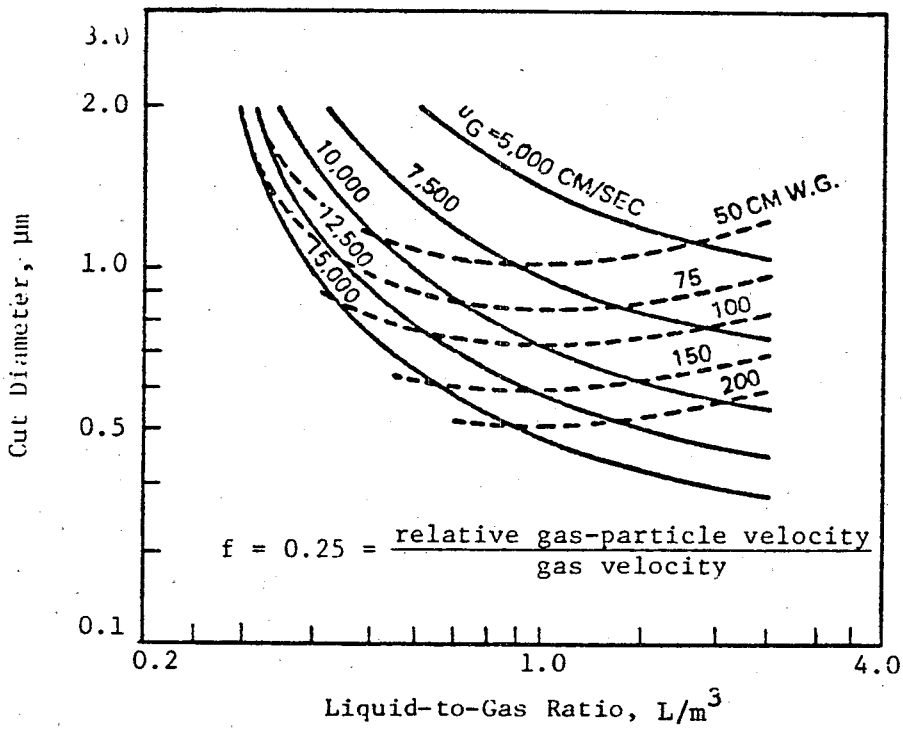


Figure 3-14. Venturi cut diameter vs liquid-to-gas ratio (Ondich, 1983).

hydrophobic aerosols and significantly higher for hydrophilic (water attracted) aerosols. This figure may be used to determine approximate operating conditions once the required cut diameter has been estimated. Care must be used to ensure that adequate liquid is supplied to provide good gas sweeping; a minimum liquid-to-gas ratio of approximately 1 liter/m^3 recommended.

It has been shown that the pressure drop across a venturi is proportional to the square of gas velocity and directly proportional to the liquid-to-gas ratio. Therefore, within limits, increasing gas velocity will result in increasing pressure drop and decreasing the device cut diameter, other things being equal. Velocity data are not available for survey sites equipped with venturis, although typical gas velocities employed commercially are 30 to 120 m/s. The low end of this range, 30 to 45 m/s, is typical of power plant applications while the upper end of the range has been applied to lime kilns and blast furnaces.

3. Process Reliability--Maintenance is critical in venturi scrubbing systems when corrosive gases (such as sulfur oxides) are present. In addition, handling of large amounts of water containing potentially corrosive or abrasive materials causes more problems than those found in dry systems. Hence, frequent and thorough inspection of equipment is essential for reliable operation.

4. Process Economics--Costs of venturi scrubbers depend upon: (1) volumetric flow, (2) operating pressure, and (3) materials of construction. For venturi scrubbers constructed of 3.2mm carbon steel and gas capacities up to about 5700 actual m^3/min, ventrui scrubber costs can be estimated with the following equation (Vatavik & Neveril, 1981):

$$C = 7117 + 14.4\ V - 0.0011\ V^2 \qquad (3\text{-}1)$$

where:

C is the venturi scrubber cost, $\$10^3$ and

V is the inlet gas flow rate, actual m^3/min.

These costs are flange-to-flange costs (December 1977 dollars) and include the venturi, elbow, separator, pumps, and controls. Correlations for adjusting these costs for different vessel thicknesses, materials of construction, and internal coolers are presented by Vatavik & Neveril, 1981.

C. Electrostatic Precipitator--

1. Process Description--Electrostatic precipitators use electrical rather than mechanical forces for the removal of suspended particulates in a gas stream. The process encompasses three basic functions: the charging of particles, the collection of the charged particles on an electrode of opposite polarity, and the removal of the collected particles. Particles suspended in the gas stream are charged by passing through a high voltage direct-current corona established between a discharge electrode, usually a small diameter wire which is maintained at high voltage, and a grounded collecting surface (collecting electrode). As the particles pass through the corona, they are bombarded by negative ions emanating from the discharge electrode and charged within a fraction of a second. The charged particles, influenced by electric field forces, migrate toward the grounded collecting surface where they are deposited and held by electrical, mechanical, and molecular forces. Particulates adhering to the collecting surface are periodically dislodged by mechanical rappers, acoustical horns or by flushing with water and fall to a hopper from which they are subsequently removed.

A majority of the industrial electrostatic precipitators (ESPs) used today are the single-stage, wire and plate type. Charging and collection take place in the same section of the ESP. The collecting surface consists of flat parallel plates spaced from 15 to 30 cm apart with wire or rod discharge electrodes located between the plates. The plates usually range from 4 to 12 m in height and 4 to 7 m in length. Plate-type precipitators are typically used for dry particulate collection. In tube-type ESPs, the collecting surface consists of a cylinder with the discharge electrode centered along its longitudinal axis. They are generally used for wet gas cleaning. With either type, a complete precipitator consists of many units in parallel and in series. A schematic arrangement of wire/plate and wire/tube ESPs is shown in Figure 3-15.

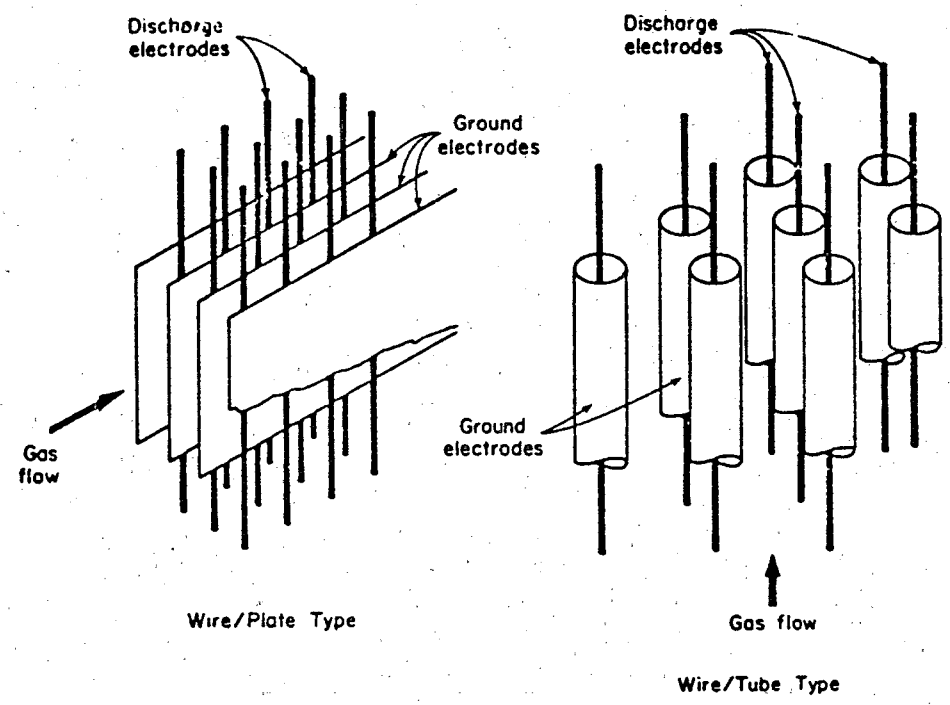


Figure 3-15. Schematic arrangement of wire/plate and wire/tube precipitators (Ondich, 1983).

In both ESP types, there are four main components: electrode system, precipitator casing, collected material removal system, and power supply.

In the single-stage wire/plate ESP the discharge electrodes may be round wire, square twisted rods, ribbons, etc. The choice of construction material is a function of the corrosive service. The most common collection electrode in the wire/plate design is a smooth plate with vertical interlocking baffles. These baffles provide strength and also produce near-zero velocity conditions as the gas flows in a normal direction to them. Other special plate electrode configurations are rod curtains, zigzag plates, and various hollow electrodes with pockets on the outside surfaces for discharging the collected dust to the hopper from quiescent gas zones. As in the case of the discharge electrodes, the plates are fabricated from a wide choice of construction materials, depending upon the degree of corrosive service. The high-voltage discharge electrodes are suspended vertically between each pair of collection electrodes. They are carefully centered between the collection electrodes to ensure proper corona gaps. Typical design parameters for commerical wire/plate precipitators are shown in Table 3-5 (Ondich, 1983).

The precipitator casing may be fabricated from a wide selection of materials including mild steel, lead-lined steel, brick, and concrete. The ESP may include several fields in the direction of the gas flow. Each field section can be treated as a separate unit module so that additional fields can be tied easily into the system as the need for increased collection efficiency arises. The entire casing is supported on a steel base that rests on support steel. Main support columns, attached to the steel base, support structural members near the top of the precipitator from which the grounded electrodes are suspended. The casing roof also is supported from these structural members. The discharge electrodes are hung from high-voltage insulators located on and supported by the roof. These insulators, together with gas seals and rapper mechanisms, are enclosed in an attic space above the casing roof. The storage hoppers for the wire/plate ESP are located under the collection electrodes and are a structural continuation of the casing.

Collected dust from the ESP is removed by mechanical or acoustical rapping and washing. Mechanical rapping methods usually consist of mechanical actuated "hammers", which are driven electrically or pneumatically. The

TABLE 3-5. TYPICAL VALUE OF DESIGN VARIABLES FOR COMMERCIAL ESPs (Ondich, 1983)

Design Variable	Normal Range of Values
Plate spacing	0.20 to 0.28 m
Gas velocity	0.61 to 2.4 m/s
Vertical height of plates	3.7 to 7.3 m
Horizontal length of plates	0.5 to 1 x height
Applied voltage	30 to 75 kV
Gas temperature	Up to 643 K; standard 810 K high temperature 977 K; special
Treatment time	2 to 10 s
Draft loss	25 to 125 Pa
Efficiency	Up to 99.9%, although usually 90 to 98%
Corona current	0.033 to 3.3 mA/m of wire
Field Strength	276 to 590 kV/m

impact blow from the hammer is transmitted vertically to the freely suspended collection electrode so that the dust is released in the direction of gravity. Rappers are provided for the discharge electrodes as well. In certain installations particles have collected in peripheral areas and built up to impede gas flow. Conventional rapping is ineffective. If the material collected is friable (ie. dry and powdery), it can be dislodged by directing acoustical frequency sound at the mass. In the case of some viscid dusts the removal of collected material can be accomplished by washing down the plate. Periodic water sprays are used, and by proper cycling of these sprays to maintain a wetted electrode, reentrainment can be completely eliminated.

The electrical energy requirement for an electrostatic precipitator is that necessary to produce an effective corona. The power supply must deliver a unidirectional negative current to the discharge electrodes at a potential very close to that which will produce arcing across the electrodes. The value of the potential difference used in the single-stage precipitator is usually in the range of 20,000 to 100,000 volts. The current delivered may vary from 20 to 500 mA. Power requirements are relatively low since only the dust is treated rather than the total gas flow. Solid state rectification is used in most new ESPs. Selenium and silicon rectifiers provide reliable service with long life. The ESP potential is maintained at the optimum value by a spark counter or current-sensing feedback circuit.

The single-stage wet wire/tube precipitator design is better suited for wet collection applications. This type of precipitator is built in a cylindrical shell. The collection electrodes consist of nested pipes which are connected and sealed to header sheets attached to the shell. The discharge electrodes are supported above the header sheet and are suspended axially in the collection electrode pipes. Water is introduced above the header sheet and flows over carefully leveled weirs at the tops of the pipes to form a water film on the inner walls of the pipes. The charged particles are collected in the water film, where their charge is neutralized, and drained off from the bottom of the ESP with the water. Because wet collection is involved, materials of construction are usually corrosion resistant.

The power supply is similar to that for the plate type EPS. However, it is simpler in concept because the vertical gas flow pattern eliminates the need for sectionalization.

Electrostatic precipitators are used extensively on large volume applications where the fine dust and particulate matter is less than 10 to 20 um in size with a predominant portion in the sub um range. The precipitators can achieve high efficiencies (in excess of 99 percent) depending on the resistivity of the particulate matter and the characteristics of the gas stream. Wet or dry particulate can be collected including highly corrosive materials if the units are suitably constructed. Precipitators can be used at temperatures up to 810 K but are normally operated at temperatures below 650 K. The static pressure drop through the units is low, usually up to 12 kPa for units operating at gas velocities of 0.6 to 2.4 m/sec. Safety precautions are always required since the operating voltages are as high as 100,000 volts. The overall size of electrostatic precipitators is comparable to fabric filters (baghouses). Space requirements are an important factor in the layout and design of the facilities.

2. Process Performance--The three most important design criteria for ESPs are the precipitation rate (W_e), the specific collection area (SCA), and the gas velocity (Q_G). Because precipitation rate can vary with resistivity, particle size distribution, gas velocity distribution, rapping, and electrical factors, an effective rate parameter or migration velocity usually is adopted. Variation of this parameter with fly ash resistivity and coal sulfur content is shown in Figures 3-16 and 3-17. Particulate resistivities of over 10^9 to 10^{10} ohm-cm were found to decrease collection efficiency. This loss results since some high resistivity particles are not able to dissipate their charge as they attach to the collector electrode. Eventually, an electric potential can build up at the surface of the collected particulates giving rise to tuft-like discharges of polarity opposite to that of the collector. Subsequently, arriving particles are repelled rather than collected at these sites (Ondich, 1983).

The relationship of gas flows to theoretical collection efficiency as a function of migration velocity is represented in Figure 3-18. Migration

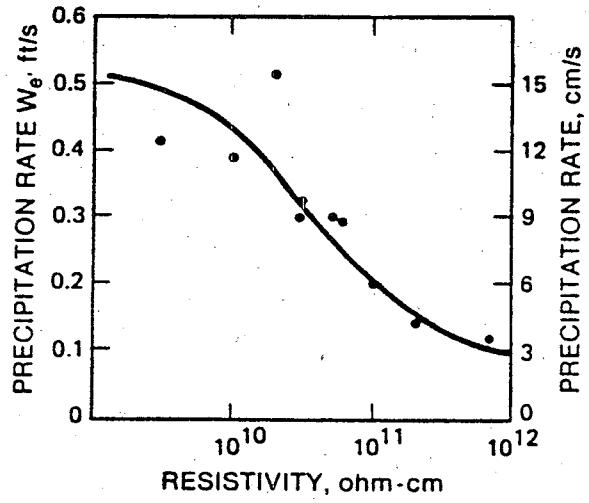


Figure 3-16. Drop in precipitation rate W_e with increasing fly ash resistivity for a representative group of precipitators. (Ondich, 1983).

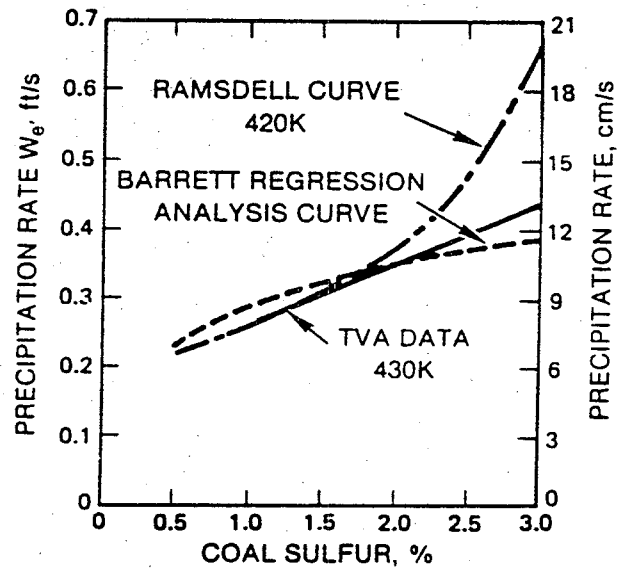


Figure 3-17. Relation of W_e to coal sulfur content for flue gas temperatures in the neighborhood of 420K as determined by several investigators. (Ondich, 1983).

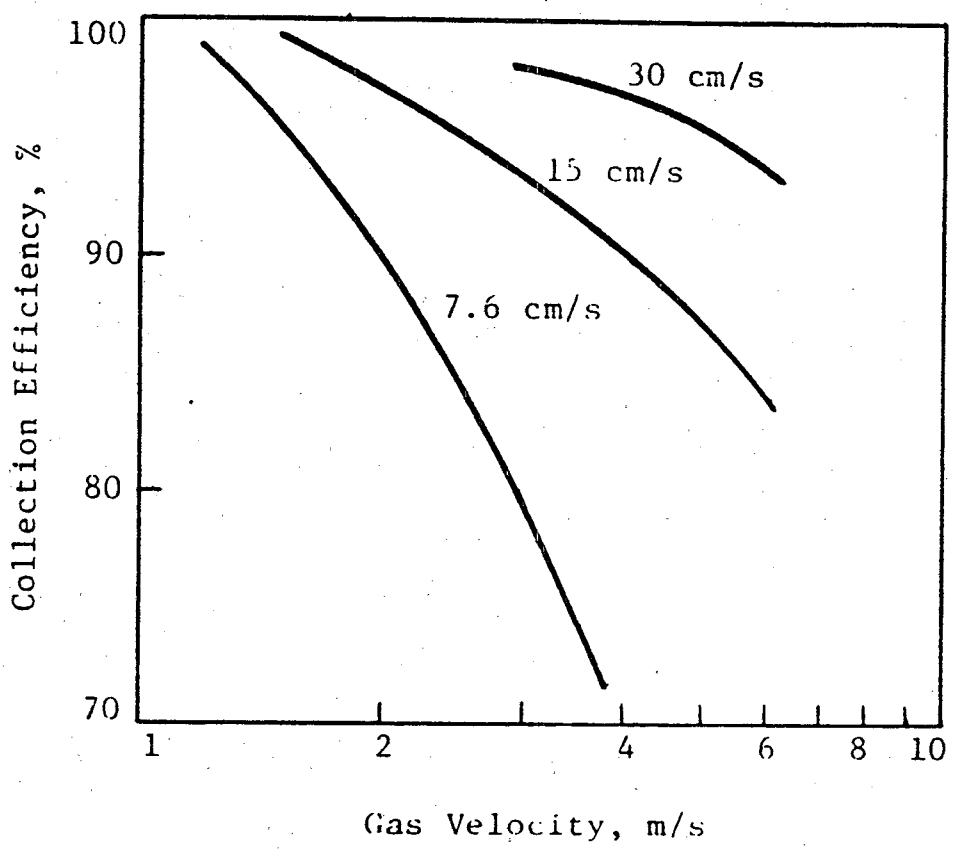


Figure 3-18. Theoretical precipitator collection efficiencies at different migration velocities. (Ondich, 1983).

velocity is an empirical constant used for sizing ESPs in certain applications. Suggested values of migration velocities for some of these applications are tabulated in Table 3-6.

It should be noted that migration velocity is directly proportional to particle size. Therefore, the larger the particle size, the larger the migration velocity and the higher the particulate collection efficiency. Gas velocity in the ESP is extremely important since collection is highly sensitive to velocity variations. The gas velocity at which maximum efficiency can be attained depends on such factors as plate configuration, precipitator size, and the judicious use of flow distributors required to minimize velocity gradients. The design velocity limit for high efficiency fly ash precipitators is about 1.5 to 1.8 m/s.

In general, the performance of a given ESP unit is a function of "the size of the box" (plate area and depth), the particle resistivity and size distribution, the electrical parameters defining particle charge and field strength, and proper operation and maintenance of equipment. Typical collection efficiencies that can be achieved by ESPs in various applications are listed in Table 3-7.

There are two types of reliability problems encountered with ESPs: mechanical and operational. Mechanical problems include misalignment of electrodes, breakage of corona wires by fatigue or by electrical burning, and air in-leakage into the hoppers. Operational problems commonly encountered include poor electrical set adjustments, shorted corona sections (e.g., caused by broken wires), overloading the precipitator by excessive gas flow, and failure to empty hoppers of collected dust. A more complete list of the problems most frequently encountered is given in Table 3-8.

ESP reliability can be improved by increased sectionalization; precipitators can have as many as 100 independent electrical sections. To minimize the effect of ash-valve failures or plugged hoppers, the trend is toward locating a hopper under each section. A well-designed ESP may have 5 to 25 percent of the bus sections out of service during the operational year. The loss in efficiency must be offset by additional collection areas beyond the plate area added for performance efficiency contingencies. Commonly, this

TABLE 3-6. TYPICAL AVERAGE MIGRATION VELOCITIES ENCOUNTERED IN COMMERCIAL ESP SYSTEMS (Ondich, 1983)

Application	Migration Velocity (w), m/s
Pulverized coal (fly ash)	0.10 to 0.13
Paper mills	0.076
Open-hearth furnace	0.057
Secondary blast furnace (80% foundry iron)	0.12
Gypsum	0.16 to 0.19
Hot phosphorous	0.027
Acid mist (H_2SO_4)	0.057 to 0.076
Acid mist (TiO_2)	0.057 to 0.076
Flash roaster	0.076
Multiple hearth roaster	0.079
Portland cement manufacturing (wet process)	0.10 to 0.11
Portland cement manufacturing (dry process)	0.057 to 0.070
Catalyst dust	0.076
Grey iron cupola (iron to coke ratio - 10)	0.030 to 0.036

TABLE 3-7. TYPICAL APPLICATION DATA FOR ELECTROSTATIC PRECIPITATORS (Ondich, 1983)

Industry	Application	Gas Flow m^3/s	Temperature K	Dust Concentration g/m^3	Dust Size % <10 μm	Collection Efficiency %
Electric Power	Flyash-Pulverized Coal Boiler	24-380	400-590	09.-11	25-75	95-98
Portland Cement	Dust from Kilns	24-470	420-670	1.1-34	35-75	85-99+
	Dust from Dryers	14-47	320-450	2.3-34	10-60	95-99
	Mill Ventilation	0.9-4.7	310-340	0.05-1.1	100	95-99
Steel	Blast Furnace Gas Cleaning	9.4-47	310-340	0.05-1.1	100	95-99
	Tar Collection From Coke Oven Gases	24-94	310-340	0.23-2.3	100	95-99
	Fume Collection From Open Hearth and Electric Furnace	14-35	420-640	0.11-6.9	95	90-99
Nonferrous Metals	Fume From Kilns, Roasters, and Sinter Machine	2.4-470	340-870	0.11-114	10-100	90-98
Pulp and Paper	Kraft Soda Fume	24-94	410-450	1.1-4.6	99	90-95
Chemical	Acid Mist	1.2-9.4	310-370	0.02-2.3	100	90-99
	Gas Cleaning, SO_2 CO_2, etc.	2.4-9.4	290-370	0.02-2.3	100	90-99
Petroleum	Powdered Catalyst	24-70	450-560	0.23-57	50-75	99-99.9
Gas	Tar Removal From Gas	0.9-24	280-340	0.02-0.5	100	90-98

TABLE 3-8. COMMONLY ENCOUNTERED PRECIPITATOR PROBLEMS (Ondich, 1983)

Fundamental Problems

1. High resistivity particles
2. Reentrainment of collected particles
3. Poor gas flow
4. Insufficient or unstable rectifier equipment
5. Insufficient number of corona sections
6. Improper or incompatible rapping
7. Gas velocity too high
8. Aspect ratio too small
9. Precipitator size too small

Mechanical Problems

1. Poor electrode alignment
2. Distorted or skewed collecting plates
3. Vibrating or swinging corona wires
4. Excess dust deposits on corona electrodes and/or collecting plates (sometimes cemented on)
5. Formation of dust mountains in precipitator inlet and outlet ducts
6. Gas turning vanes and/or gas distribution screens plugged with dust
7. Air inleakage into hoppers, shells, or gas ducts
8. Gas leakage around precipitation zones and/or through hoppers

Operational Problems

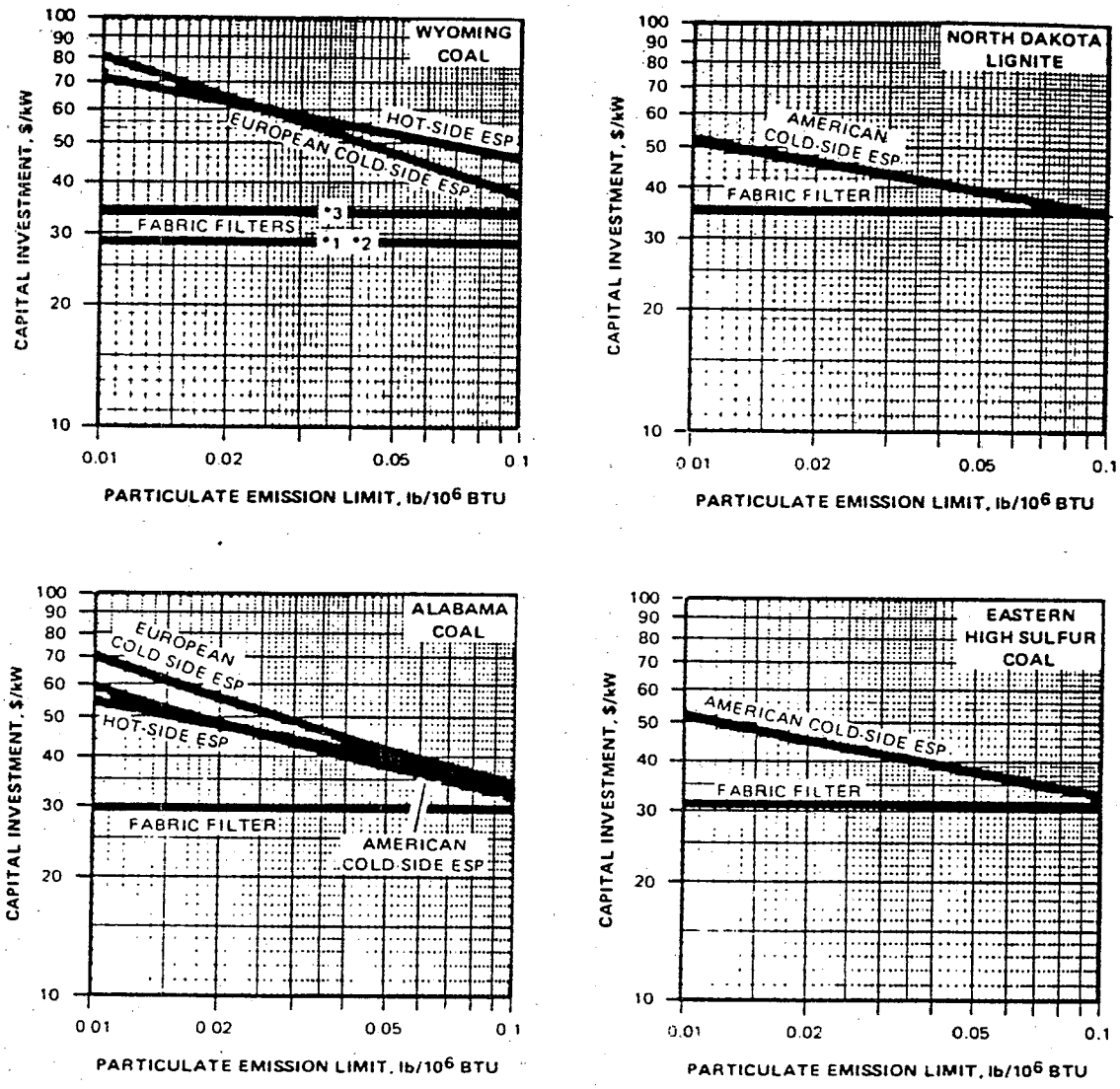
1. Full or overflowing hoppers
2. Shorted corona sections (e.g., broken wires)
3. Rectifier sets or controls poorly adjusted
4. Precipitator overloaded by excessive gas flow
5. Precipitator overloaded by excessive dust concentration
6. Process upsets (e.g., poor combustion, steam leaks)

extra area is expressed in terms of extra fields. For instance, a precipitator that could meet guarantees with four fields may have a fifth field installed to account for normal deterioration and contingency.

3. Process Economics--Costs for ESPs are given in Figure 3-19 (Ondich, 1983). This figure illustrates capital investment costs for precipitators used on 500 MW_e pulverized coal fired power plants. These costs are based on 20 different designs and estimates. Included in the estimates are material and labor costs for the installation of the collectors and associated equipment. Added to these costs are differential and indirect field costs, engineering costs, fee at 3 percent, and contingency and miscellaneous costs at 10 percent.

Costs for ESPs increase significantly as emission limits become more restrictive. Note that capital investment increases about 20 percent when the outlet emission is halved. Costs also vary depending upon the type of coal and properties of the fly ash. To meet NSPS standards for electric utilities of 13 ng/J requires a capital investment from $42/kW to $57/kW (1978 dollars) depending upon the type of coal being fired. Costs also increase with increase in size as shown in Figure 3-20 for a dry ESP. Although these costs (no data given) may be substantially lower than current market costs, they provide a relative cost vs. size factor. Many items influence costs, and for wet precipitators, the cost may be 2 to 10 times higher than for dry ESPs (Ondich, 1983).

Annual operating and maintenance costs for ESPs are generally composed of maintenance labor and materials, maintenance supervision, plant general and administration, and electrical power costs. Maintenance labor and material costs are approximately 2 percent of the total capital investment. Maintenance supervision and plant general and administrative costs are approximately 5 percent and 15 percent of maintenance labor and material costs, respectively. Electrical power requirements average 0.012 kW/m^3 for high efficiency ESPs and 0.009 kW/m^3 for other ESPs. High efficiency ESPs are those units which have collection efficiencies of over 99 percent.



*1 20-COMPARTMENT FABRIC FILTER WITH 2 YEAR BAG-REPLACEMENT
*2 20-COMPARTMENT FABRIC FILTER WITH 4 YEAR BAG-REPLACEMENT
*3 40-COMPARTMENT FABRIC FILTER WITH 2-YEAR BAG-REPLACEMENT

Notes: 1 lb/10^6 Btu = 430 ng/J. Costs are in 1978 dollars.

Figure 3-19. Capital investment for collectors on 500 MW (net) power plants (Ondich, 1983).

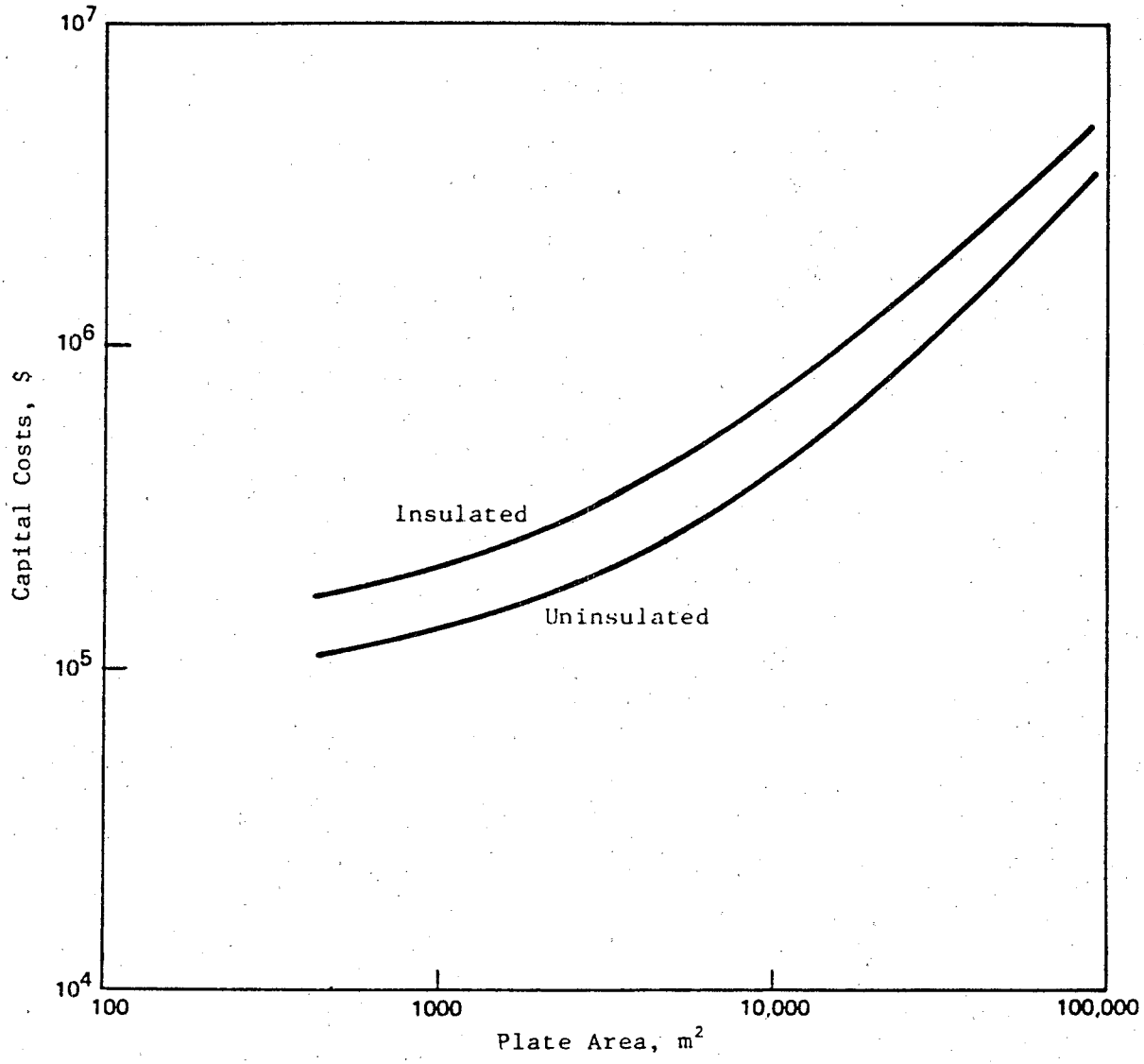


Figure 3-20. Purchase price of dry-type electrostatic precipitators. (Ondich, 1983).

D. Dry Venturi Scrubber--

The dry venturi scrubber is based on the principal of providing the targets necessary for sub-micron particle removal as dry solid particles rather than as liquid droplets.

In a standard wet venturi, particles are captured by inertial impact; the liquid targets are created within the venturi by transfering velocity energy into shearing action.

The capture efficiency for a venturi can be represented by the equation:

$$\text{Efficiency of capture} = \frac{f\,(V_r\,Dp)}{D_t} = f\,(N_s)$$

where V_r = relative velocity between particles and targets
D_p = pollutant particle diameter
D_t = target droplet (or solid) diameter
N_s = separation number

The diameter of the target droplets in a wet venturi is inversely proportional to the gas velocity. Therefore, high velocities, and the resulting high pressure drop, are required to achieve the small diameter targets necessary for a high capture efficiency.

In the dry venturi system developed by Teller Environmental Systems shown in Figure 3-21, the targets are supplied to the system as a dry powder. The required gas velocity is reduced because the targets already have been created, both gas velocity and the energy input are reduced. Small solid targets in the range of 5 to 25 microns permit high capture efficiencies with only a few percent of the power required for a high energy wet venturi.

The dry venturi introduces the target material in a dispersed state at a high velocity relative to the gas stream. The target size, population density and velocity can be varied for specific applications to improve capture efficiency. Existing systems use a separation function number (N_s) of 1.2 to capture 0.5 μm size particles with a 20 μm size target.

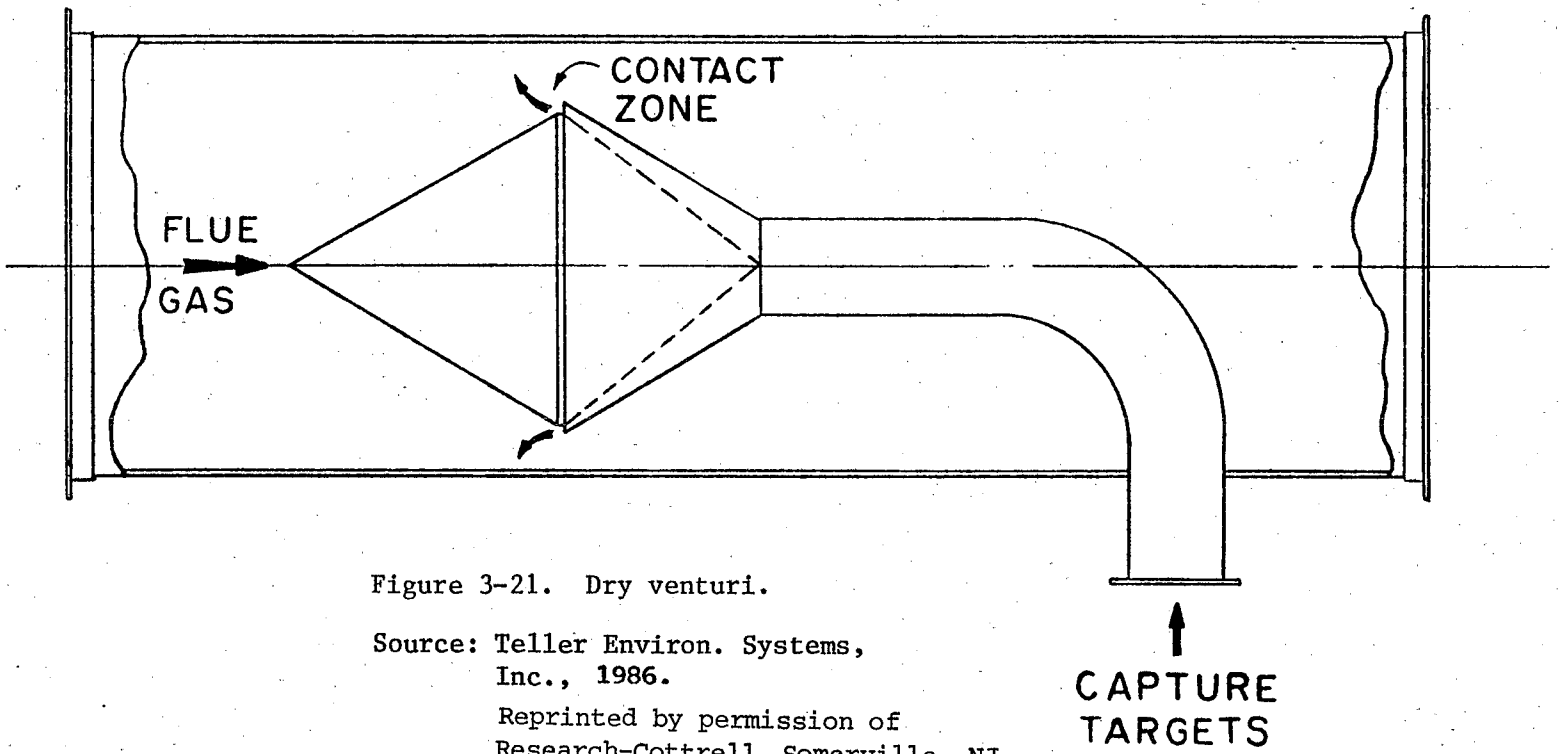


Figure 3-21. Dry venturi.

Source: Teller Environ. Systems, Inc., 1986.

Reprinted by permission of Research-Cottrell, Somerville, NJ.

The capture of the particles in the dry venturi is only the first step in the removal processes. The dry venturi is used primarily as an augmentation device to improve the particulate collection efficiency, to facilitate the removal of the collected particulate matter from the collector surface and possibly to provide a degree of acid gas control.

Commercial systems have demonstrated a total power consumption for the dry venturi and baghouse/ESP of 0.35 to 0.5 kW/1000 m^3/hr of gas (Teller, 1985). If additional processing is required, e.g. acid gas removal with a wet scrubber, the wet scrubber can also serve as a particulate removal system since the particulate diameter of the particulates has been increased from the original particulate diameter (i.e. 0.5 μm) to the target diameter (i.e. 20 μm).

1. Dry Venturi - Baghouse/ESP -- The use of the dry venturi followed by a baghouse permits collection of deliquescent, cohesive, adhesive submicron, and combustible particulates where the use of a baghouse was previously restricted or where ESP collector plates could not be continuously cleaned. These devices now can function with submicron particles and hydrophilic or cohesive particulates without blinding or high pressure drops. It also functions safely with combustible particulates. In systems where SO_x or other acid gases as well as particulate control is required, the use of SO_2-reactive powder (like Nahcolite or Trona) as the target material can serve a dual purpose. The use of dry sorbent for SO_x control is discussed further in Section 3.3.8D.

The use of crystalline targets results in exceedingly low pressure drops in the baghouse even when capturing submicron particulates. As a result, the bag cleaning cycle can be extended from the normally anticipated 5 to 30 minutes to 4 to 36 hours. This results in extended bag life from decreased shaking requirements and from the elimination of the permeation of the submicron particles into the interstices of the cloth.

Residence time of the cake on the cloth is increased by a factor of approximately 30. As a result, where a system is providing the dual service of particulate and acid gas removal, residual unreacted reagent is accumulated in the cake which increases the residence time for the acid gas reaction.

Performance characteristics - Performance data for the dry venturi - baghouse combinations is shown in Table 3-9.

2. Dry Venturi - Wet Scrubber -- The use of the dry venturi followed by a wet scrubber results in the scrubber functioning as both a particulate removal device and gas absorption unit. While wet scrubbing will remove large diameter particulate from a gas stream, it is generally not effective in removing particles less than 5 μm. However, by increasing the particle size from 0.5 μm to 20 μm, the use of dry venturi upstream of the wet scrubber results in effective particulate removal.

This combination can be particularly effective for oil shale processing in that a wet scrubber for gas clean-up is required and its dual use for particulate control can eliminate the need for an end-of-pipe baghouse or ESP. The reader is referred to Section 3.3.5 for additional discussion of this combination.

3.1.2 Fugitive Dust Technologies

Dust is generated from the detonation of explosives; the loading, handling, crushing, screening, and stockpiling of raw shale; the handling and disposing of spent shale; and the moving of transport vehicles over unpaved and paved roads. In an underground mine, the generation of dust is of concern for workers' health, safety and comfort. But the dust that escapes the mine must traverse the mine vents where it can be readily collected in conventional filters (electrostatic precipitators, baghouses or granular filters). Above ground, the normal handling, stockpiling and disposal are the subject of special procedures to be discussed below. Where open pit mining is employed, other special controls are employed since the blasting is unconfined.

The following discussion of fugitive dust emission controls is presented in relationship to the operations in which dust is produced: (1) material handling, (2) wind erosion, (3) vehicle traffic, and (4) miscellaneous processes.

A. Shale and Overburden Handling--

The movement of raw shale, spent shale and occasionally overburden material is performed by bulldozers, front-end loaders and conveyors.

TABLE 3-9. PERFORMANCE CHARACTERISTICS OF THE DRY VENTURI-BAGHOUSE SYSTEM (PEI, 1985)

Industrial Application	Typical Emissions: Particulate: Composition	Typical Emissions: Particulate: Size, Micron	Typical Emissions: Particulate: Loading, gm/ncm	Typical Emissions: Gases: Compos.	Typical Emissions: Gases: Conc. ppmV	Performance: ΔP, mm H_2O: Dry Venturi	Performance: ΔP, mm H_2O: Bag-house	Performance: Shake Cycle, Hours	Performance: Part. Emiss. gm/ncm	Performance: Acid Gases Collctd.
Fiberglass	Na_2,NaF B_2O_3,H_3BO_3	0.05-2	0.2-0.6	HF SO_2,SO_3	200-400 200-400	25	80	24	.005-.009	HF,SO_x
Secondary Aluminum	C,Oil, Al_2O_3, AlF_3,$AlCl_3$	0.05-30	0.02-0.4	HCl,Cl_2 HF,SO_2, SO_3	5-100	20	150	48	.003-.007	HCl,HF
Municipal Incinerator	S,O_2, PbO,ZnO, C,Oil	0.1-20	0.8-4	HCl HF	0-600 0-50	20	90	8	.007-.014	HCl,HF SO_x
Coke Calinciner						25	125	8	.009-.014	SO_x
Glass						30	100	16	.009-.019	SO_x

Particulate emissions from bulldozers come from two points, the bulldozer tracks and the bulldozer blade. The bulldozer tracks reentrain dirt much the same as wheels, except probably with additional grinding action. The bulldozer blade top and sides are emissions points as dirt slides off the blade. This is particularly true of the top of the blade where thin layers of dirt can easily be carried off by the wind.

Emissions from front-end loaders come off the tracks or wheels as well as the loader bucket. With the loader bucket, the usual source of emissions is spillage from the bucket as the bucket is being raised.

Both the bulldozer and loader drop their loads of shale into a truck bed or into a conveyor. Dust is emitted by two mechanisms as the material is dropped. First, the wind picks up shale particles from the edges of the mass of material being dropped. Second, the air turbulence caused by the displacement of air up out of the truck or conveyor bed caused by the mass of dirt moving downward, brings up soil already in the bed and soil from the edge of the dirt mass being dropped.

Emissions from conveyors loaded are caused primarily by wind and can be controlled by proper screening or enclosing.

1. Quantification of Emissions--An emission factor was developed for bulldozing activity on overburden in coal mines where silt values ranged from 3.8 to 15.1 percent, and moisture ranged from 2.2 to 16.8 percent. The emission factor includes emissions from both the tracks and the blade and is (PEDCo 1981):

$$TSP = \frac{5.7\ s^{1.2}}{M^{1.3}} \qquad \text{(Eq. 3-2)}$$

where TSP = emissions of total suspended particulate in lb/hour
s = silt, percent
M = Moisture, percent

The emission factor for front-end loader operations appearing in EPA's Compilation of Air Pollutant Emission Factors (1982) was developed based on

material handling operations within a steel mill. All sources (track, tires, bucket, dump) are represented by the factor which is given as:

$$E = (0.0013) \frac{\frac{s}{5} \frac{U}{5} \frac{H}{5}}{\left(\frac{M}{2}\right)^2 \left(\frac{Y}{6}\right)^{0.33}} \qquad \text{(Eq. 3-3)}$$

where E = TSP emission factor, lb/ton

s = material silt content, (%)

U = mean wind speed, m/s (mph)

H = drop height, m (ft)

M = material moisture content, (%)

Y = dumping device capacity, m^3 (yd^3)

The silt and moisture terms describe the general dustiness of the material being moved. Three of the variables deal with the material dump cycle. Emissions increase with increased wind speed (blowing of dirt from the dirt mass edges), increased drop height (greater turbulence caused by material drop), and decreased bucket size (more dirt mass edge per unit of volume).

2. Principles of Control--Because the movement of shale continually exposes new material, the control measure also must be continuous. For emissions from bulldozers or front-end loaders, the only method to control dust emissions is to spray the area being worked at frequent intervals of 30 minutes to 2-3 hours. The spray can be water or surfactant (to minimize the amount of water used), and can be sprayed from a mobile tower. Spraying moistens the shale particles on the surface but not all the shale being moved. However, the material below the surface is frequently more moist than that on the surface. The surface spray reduces emissions from the track or wheels, as well as emissions from the bucket and material being dropped.

There have been limited experiments on attaching sprays directly to bulldozers or front-end loaders (PEI, 1985). Several operational problems develop. The machine either must be outfitted with a large tank or with an umbilical cord to a tank -- neither is desirable. The spray nozzles must be attached to the blade/bucket or on arms reaching over the blade/bucket. Maintenance is difficult with either approach. Lastly, machine operators don't like to work in the misty conditions.

Area spraying does seem to reduce the dust emissions from material being dropped despite the fact that most of the material is never treated with the spray, and significant emissions are still present. The basic methods of control are (1) to induce moisture into the drop cycle (increasing the moisture term in Equation 3-3) and (2) to decrease windspeed around the drop receptacle (decreasing the windspeed in Equation 3-2). Neither of these practices is used widely by the mining industry for truck loading, although they are commonly used in the aggregate industry when dumping material into bins. The most efficient method of spraying moisture on material being dumped is to construct a mobile frame under which a truck can be driven. The truck bed is thus positioned under a series of nozzles which produce a curtain of water across the horizontal surface of the bed. The spray is operated only during the dumping. The system either can be activated by the truck driver, or remotely by the front-end load operator, but is not operated continuously. Water, surfactant, or foam may be used. As the material is dumped through the spray curtain, the edges of the soil mass are moistened. More importantly, as the upward turbulence of air brings dirt upward out of the bed, the dust is caught in the spray curtain and falls back into the bed. Portable windscreens also may be used to control emissions from the dump cycle. The windscreen can be positioned to shield the dump area as well as the loading operation. The screen height should be at least one foot above the height of the front-end loader bucket drop height. The screen should be two screen heights wider than the width of the area being worked. Screen porosity should be 50 percent. The screen will shelter a down-wind distance of about 7 to 10 screen heights with as much as a 50 percent reduction in windspeed at the surface.

Wind screens do not reduce the formation of dust by the material handling operation. However, the diminished wind velocity causes the dust to drop back to the ground sooner. If the plume from the material drop rises above the height of the screen, there is no control on that portion of the plume. In fact, wind eddies from the windscreen may carry the dust farther.

3. Products--Products available are mainly surfactants, which can be used for the area spray and for the spray curtain, foams which can be used for the spray curtain, and windscreens. Data on these products are shown in Table 3-10.

TABLE 3-10. SOIL MOVEMENT DUST SUPPRESSANTS (PEI, 1985)

Product Name	Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Liquid Chemicals (Surfactant)								
Compound MR	Johnson-March Corp. 3018 Market Street Philadelphia, PA 19104 (215) 222-1411		X			1000:1	na	4.00/gal
Compound MR 20/40			X			2000-4000:1	na	5.25/gal
DCF-20	Calgon Corp. P.O. Box 1346 Pittsburgh, PA 15230 (412) 777-8000		X			50-200:1	20-50 lbs product to 1000 tons of material	4.76/gal
DCL-163			X				10-45 lbs product to 1000 tons of material	9.26/gal
DCL-1870			X			1000-5000:1	2-10 lbs product to 1000 tons of material	14.58/gal
Dustalloy	Wen-Don Corp. P.O. Box 13905 Roanoke, VA 24034 (703) 982-0561		X			1000-2000:1	not available	not available
GCP 200	Betz Laboratories, Inc. Somerton Rd. Trevose, PA 19047 (215) 355-3300		X			a		
GCP 201			X			a		
GCP 202			X			a		

(Continued)

[a] Application and price information is confidential

TABLE 3-10. Continued

Product Name	Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
			Liquid Chemicals (Surfactant continued)					
M070E	Mona Industries, Inc. P.O. Box 425 76 E. 24th Street Paterson, NJ 07544 (210) 345-8220		X			not available	not available	6.30/gal
Sterox DF	Monsanto Company 800 N. Lindbergh Blvd. St. Louis, MO 63166 (314) 694-1000		X			not available	not available	6.35/gal
Sterox ND			X			not available	not available	6.35/gal
Sterox NJ			X			not available	not available	6.35/gal
			Foams					
Aquadyne Dust Suppression System	Motmoco, Inc. P.O. Box 300 Paterson, NJ 07543 (201) 345-6202		X			na	na	b
			Windscreen					
Dusttamer	Julius Koch, Inc. P.O. Box A-995 New Bedford, MA 02741 (617) 995-9565				Wind Screen	na	na	2.07-2.95/linear foot, 3 ft. width

[b] Products are sold as turn-key systems. Price varies with application and size of system

Suppliers of these products often sell spray nozzles, masts and spray curtains.

4. Effectiveness--The only testing the effectiveness of dust control measures effectiveness was performed by PEDCo Environmental, Inc. (1984). Four control measures were evaluated. Control Measure 1 consisted of spraying the active working area of the Front End Loader (FEL) and dump truck with water at 4 L/m^2 (0.9 gal/yd^2). Control Measure 2 was identical to Measure 1 but a surfactant was added to the water to form a 1:1000 dilution of surfactant to water. It was found that somewhat less water was needed for these tests, only 3.4 L/m^2 (0.75 gal/yd^2).

Control Measure 3 consisted of an array of 12 spray nozzles on the sides of the dump truck emitting a spray curtain of the water/surfactant mixture at a rate of 6.8 L/m^2 (1.5 gal/yd^2). This method was used to control emissions only during dumping. Control Measure 4 consisted of four spray nozzles at the corners of the truck bed to disperse a foam spray curtain at a rate of 1.8 L/m^2 (0.4 gal/yd^2). As in Method 3, the foam spray was actuated only during dumping.

The results are summarized in Table 3-11. Water spraying over the area being worked by the FEL and truck controlled efficiencies for the fine particles (<2.5 micrometers), more effectively (64 percent) than it controlled on the larger particles (<30 micrometers) (42 percent). Adding surfactant to the water increased control efficiencies while allowing the quantity of water used to be reduced.

Both the spray curtain control measures were less effective than the area spray with the water/surfactant mixture. The water curtain was somewhat better for control of dust from the dump cycle than the foam curtain. It is believed that higher efficiencies could be achieved with the water spray by redesigning the spray system. If this dumping control was used in conjunction with the water/surfactant area spray, the resulting overall control efficiency would probably be significantly greater than for either one alone.

B. Wind Erosion--

1. Dust Producing Mechanisms--Wind erosion from exposed areas or piles includes shale particle transport by surface creep, saltation, and

TABLE 3-11. SUMMARY OF RESULTS (PEI, 1985)

Operation	Control Measure	Control Efficiency FP	Control Efficiency TSP
FEL travel and scraping	Area Spray-Water 4 L/m^3 (0.9 gal/yd^2)	65	42
	Area Spray-Water/Surfactant 3 L/m^2 (0.75 gal/yd^2)	70	63
FEL dump	Water Curtain 7 L/m^2 (1.5 gal/yd^3)	56	50
	Foam Curtain 2 L/m^2 (0.4 gal/yd^3)	41	46

FP = Fine particles

TSP = Total Suspended Particles

FEL = Front End Loader

suspension. Surface creep describes the rolling and sliding movement of particles across a surface. These particles generally have a diameter in excess of 1,000 μm. Saltation is a term used to describe the hopping and bouncing movement of a particle. These particles are lifted by the wind but are too heavy to remain airborne. Particles moved by saltation have diameters ranging from 80 to 1,000 μm. Particles smaller than 80 μm are generally suspended. PEI (1985) reported that from 3 to 40 percent by weight of the total soil particle loss from exposed areas may be attributed to suspension, from 50 to 75 percent of particle loss to saltation, and 5 to 25 percent of the loss to surface creep.

Wind erosion is largely an intermittent activity that occurs above a threshold wind velocity. Estimates of this threshold velocity have varied, and range from about 10 to 20 mph across different soil types, aggregates and meteorological conditions.

2. Quantification of Emissions--Various researchers have attempted to quantify emissions from exposed areas and from piles. None are totally satisfying.

The emission factor most commonly used to estimate emissions from exposed areas is (EPA 1974):

$$E_s = AIKCL'V' \qquad \text{(Eq. 3-4)}$$

where:
- E_s = suspended particulate fraction of wind erosion losses of tilled fields, tons/acre/year
- A = portion of total wind erosion losses that would be measured as suspended particulate, estimated to be 0.025
- I = soil erodibility, tons/acre/year
- K = surface roughness factor, dimensionless
- C = climatic factor, dimensionless
- L' = unsheltered field width factor, dimensionless
- V' = vegetative cover factor, dimensionless

Values from this equation can range from .001 to 8.25 tons/acre-year, but more generally range from .05 to .75 tons/acre-year. The equation is based on the premise that wind erosion varies with the soil particle size (A), soil characteristics (I and K), moisture and windspeed (C), field width (L') and vegetative cover (V').

The most commonly used emission factor equation for wind erosion from storage piles is (EPA 1979):

$$E = 1.7 \left(\frac{s}{1.5}\right) \left(\frac{365-p}{235}\right) \left(\frac{f}{15}\right) \text{ (lb/day/acre)} \qquad \text{(Eq. 3-5)}$$

where:
- E = total suspended particulate emission factor
- s = silt content of aggregate (%)
- p = number of days with $\geq$ 0.01 in. of precipitation per year
- f = percentage of time that the unobstructed wind speed exceeds 12 mph at the mean pile height

The premise of the equation is that wind erosion emissions vary with soil particle size, mositure and windspeed.

3. Principles of Control--Control systems work in one of two ways: (1) By reducing windspeed on the soil surface; or (2) By forming a new, less erodible soil surface.

Methods to reduce windspeed at the soil surface are:

1. Cover pile with wind impervious fabric or vinyl
2. Erect windscreen
3. Pile orientation and pile shape

These control methods have the effect of reducing the "C" term in Equation 3-4 (measures 1 and 2), or reducing the area exposed to maximum windspeeds (measure 3) which influences the "C" and "L'" terms in Equation 3-4.

Methods to form a new, less erodible surface, are:

1. Water spraying which compacts and weights soil particles
2. Chemical dust suppressant application which forms a crust over the existing soil, or binds the top soil particles together.
3. Establish vegetation. The roots bind the soil together, while the stems reduce windspeed at the surface.

These methods change the I, K, C, and V' factors in Equation 3-4.

4. Product-Types, Application Rates, Availability and Costs--Products for dust control of exposed areas and undisturbed storage piles are the same. Product categories are:

1. Liners and fabrics that are impermeable to the wind. Some are impermeable to liquids
2. Windscreens which decrease windspeed on the downwind side.
3. Spray systems and foams which are sprayed every few hours to cover or moisten the soil
4. Liquid chemicals which form a soil admixture. These products are sprayed on every few weeks. Liquids can be bitumens, adhesives, salts, or binders with grass seed.

Product name, address, phone number, application method and cost are shown in Table 3-12.

Liners will not allow water or many chemicals to pass. Fabrics will allow liquids to pass, and may not be tolerant of certain chemicals. Because fabrics are more commonly used for prevention of soil erosion, chemical compatibility testing has not been performed. The spray systems and foams are intended for small areas as a temporary dust control measure. The control would last only for a period of hours.

The greatest diversity of products occurs under the category of liquid chemicals. Listed first are oil-based products, many of which are primarily marketed for haul road control. Next listed are adhesives which encompass a wide range of products. For example, Bio Cat 300-1 is marketed as a soil enzyme. Some are lignons such as Flambinder, Lignosite, and Norlig A. Others are polymers of various sorts such as AMSCO-RES 4281 (Carboxylated Styrene-Butadiene Copolymer), Curasol (Synthetic Resin), Genaqua (Vinyl Acetate Resin) and Soil Seal (Latex Acrylic Copolymer). The polymers are applied as a water soluble liquid, but supposedly cure from a non-water soluble material.

Terra Tack I, Terra Tack III, Terra Tack AR and Sherman Mulch can be impregnated with grass seed. The products contain a binder to hold the soil while the grass grows. Other similar products are available and are routinely used to vegetate highway excavations after construction. A detailed Handbook for use of materials to quickly revegetate soils of low productivity is available (EPA 1975).

5. Effectiveness--

(a) Exposed Areas--Wind erosion control from the standpoint of stabilizing mineral wastes and soil stabilization in connection with construction projects

TABLE 3-12. EXPOSED AREA AND STORAGE PILE DUST SUPPRESSANTS (PEI, 1985)

Product Name	Manf. Address, Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Mulch, Liners, Fabrics								
Enviromat	International Minerals & Chemical Corp. 421 East Hawley St. Mundelein, IL 60060 (312) 566-2500				Liner	na(1)	na	4.50/yd^2
Gagle Liner	Duane W. Gagle Co. P.O. Box 441 Bartlesville, OK 74003 (918) 337-0129				Liner	na	na	4.00-5.00/yd^2 installed
Mirafi Fabrics	Director of Mirafi Celanese Fibers Marketing 1211 Ave. of the Americas New York, NY 10036 (212) 719-8000				Fabric	na	na	0.90-1.20/yd^2
Sherman Process (mulch)	KPN International, Inc. 19 Pebble Road Newtown, CT 06470 (203) 426-3639				Mulch	na	na	800.00/acre
Spray Systems, Foams								
Omega Foam Dust Suppressant Sys.	Valerin Technologies, Inc. Technical Center 87 Great Valley Pkwy. Great Valley Corp. Cntr. Malvern, PA 19355 (215) 296-7322		X			not available	not available	~~(b)~~

(Continued)

TABLE 3-12. Continued

Product Name	Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Sani Blanket	Sani Foam, Inc. 1370 Logan Ave. Suite D Costa Mesa, CA 92626 (714) 557-5070	X				none	1-2" layer	0.11/ft^2
				Liquid Chemicals (Bitumens)				
AMS 2200	ARCO Mine Sciences 1500 Market St. P. O. Box 7258 Philadelphia, PA 19101 (215) 557-2000	X				4:1	0.5	1.75/gal
AMS 2300		X				1:1	0.75	1.85/gal
Coherex CRF	Witco Chemical Golden Bear Div. P. O. Box 378 Bakersfield, CA 93302 (805) 393-7110	X				4:1	0.5	1.50/gal
Docal 1002	Douglas Oil Co. 3160 Airway Ave. Costa Mesa, CA 92626 (714) 540-1111	X				2:1	0.1	0.67/gal
Peneprime	Utah Emulsions Co. P. O. Box 248 North Salt Lake, UT 84054 (801) 292-1434	X				None	0.5	1.25/gal

(Continued)

TABLE 3-12. CONTINUED

Product Name	Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Petro Tac P	Syntech Products Corp. 520 E. Woodruff Ave. Toledo, OH 43624 (419) 241-1215	X				5:1	0.25-0.75	1.55
Resinex	Neyra Industries, Inc. c/o Petroleum Products,Inc. P. O. Box 493 Valparaiso, IN 46383 (219) 465-1300	X				10:1	1.25-5.0	1.48/gal
Retain	Dubois Chemical Co. 3630 E. Kemper Road Sharonville, OH 45241 (513) 769-4200	X				10:1	0.4	5.55/gal
			Liquid Chemicals (Adhesives)					
Bio Cat 300-1	Applied Natural Syst.Inc. 35 E. Lake Mead Dr. Henderson, NV 89015 (702) 451-6010	X				66:1	2.0	19.95/gal
CPB-12	Wen-Don Corp. P. O. Box 13905 Roanoke, VA 24038 (703) 982-0561		X			10:1	1.0	7.50/gal

(Continued)

TABLE 3-12. CONTINUED

Product Name	Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Curasol AK	American Hoechst Corp. Industrial Chemicals Route 202-206 North Somerville, NJ 08876 (201) 231-2000	X				22:1	0.2	6.26/gal
DCL-40A	Calgon Corp. P. O. Box 1346 Pittsburgh, PA 15230 (412) 777-8000	X				2-10:1	0.27-1.1	4.14/gal
DCL-1801		X	X			66-200:1	0.5-0.8	9.20/gal
DCL-1803		X	X			100-200:1	0.5-0.8	20.64/gal
DG-859	Betz Laboratories, Inc. Somerton Road Trevose, PA 19047 (215) 355-3300	b						
DG-873		b						
Dust Ban 6500	Nalco Chemicals Co. 2901 Butterfield Rd. Oak Brook, IL 60521 (312) 887-7500	X	X			20-100:1	0.25-1.0	0.67/gal
Dust Ban 7991		X				20-100:1	0.25-1.0	9.20/gal
Dust Ban 8820			X			10:1	0.25-1.0	5.37/gal
Dust Bond 100	Research Products, Inc. 4222 No. 39th Ave. Phoenix AZ 85019 (602) 269-7891	X				None	0.17	0.40
Dust-Set	Mateson Chemical Corp. 1025 E. Montgomery Ave. Philadelphia, PA 19125 (215) 423-3200	X				500:1	0.17	8.00

(Continued)

TABLE 3-12. CONTINUED

Product Name	Telephone Number	Application Method				Application Rate		FOB Price Before Dilution, Dollars
		Topically By Truck	Topically By Fixed Mast	Mixed In Place	Other	Dilution	Applied gal/yd^2	
Dustbinder 124	Union Carbide Corp. Mining Chemicals 270 Park Ave. New York, NY 10017 (203) 794-2000	X				10-15:1	1.0	4.50
Flambinder	Flambeau Paper Co. P. O. Box 340 Park Falls, WI 54552 (715) 762-3231	X				5.5:1	0.5	0.15
GCP 203	Betz Laboratories, Inc. Somerton Road Trevose, PA 19047 (215) 355-3300	b						
Lignosite	Georgia-Pacific Corp. P. O. Box 1236 Bellingham, WA 98227 (206) 733-4410	X				4:1	0.5	152.00/ton
M166	Dowell Div. of Dow Chemical U.S.A.	X	X			16-20:1	0.4-0.6	5.10
M167	P. O. Box 4378 Houston, TX 77210 (800) 645-9355	X	X			16-20:1	0.4-0.6	5.55

(Continued)

TABLE 3-12. CONTINUED

Product Name	Manf. Address, Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Norlig A	Reed Lignin, Inc. 120 E. Odgen Ave.	X				1.1	2.8	0.77
Norlig 12	Suite 106 Hinsdale, IL 60521 (312) 887-9640	X				2.4 lb/gal	2.8	0.23
Orzan AL-50	Crown Zellerbach Corp.	X				10:1	1.2-6.3	0.20
Orzan DSL	Chemical Prod. Div.	X				10:1	1.2-6.3	0.20
Orzan GL-50	Camas, WA 98607 (206) 834-4444	X				10:1	1.2-6.3	0.20
Res 661	Union Chemicals Div. Union Oil Co. of Calif.	X				8:1	0.2	3.10/gal
Res 3078	1445 Alondra Blvd. La Mirada, CA 90638	X				8:1	0.2	4.04/gal
Res 4281	(714) 523-5120							
Rezosol 5411-B	E.F. Houghton & Co. Madison & Van Buren Aves. Valley Forge, PA 19482 (215) 666-4105	X				30:1	1.0	6.48/gal
SP-301	Johnson-March Corp. 3018 Market St.	X				None	0.25	2.15/gal
SP-400	Philadelphia, PA 19104 (215) 243-1700	X				None	0.25	3.95/gal
Soil Gard	Walsh Chemical Corp. 207 Telegraph Dr. Gastonia, NC 28052 (704) 865-7451	X				5-15:1	0.25-0.8	9.09/gal

(Continued)

TABLE 3-12. CONTINUED

Product Name	Manf. Address, Telephone Number	Application Method: Topically By Truck	Application Method: Topically By Fixed Mast	Application Method: Mixed In Place	Application Method: Other	Application Rate: Dilution	Application Rate: Applied gal/yd^2	FOB Price Before Dilution, Dollars
Soil-Sement	Midwest Ind. Supply, Inc. P. O. Box 8431 Canton, OH 44711 (216) 499-7888	X				5:1	0.25	2.32.gal
	Sulfur Products 575 Market St. San Francisco, CA 94105 (415) 894-6723							
Terra Tack I	Grass Growers	X				0.5 lb/gal	0.16	3.36/lb
Terra Tack III	425 Cottage Place	X				0.25 lb/gal	0.33	3.95/lb
Terra Tack AR	Plainfield, NJ 07060 (201) 755-0923	X				0.25 lb/gal	0.10	3.06/lb
WESLIG 120	WESCO Technologies, Ltd. P. O. Box 3880 San Clemente, CA 92672-1680 (714) 661-1142	X	X			6.7-10:1	0.25	0.42/gal
Woodchem LS	Woodchem, Inc. P. O. Box A Oconto Falls, WI (414) 846-2839	X 54154				None	1.5	0.17
55-03 Terraset	Celtite, Inc. 150 Carley Court Georgetown KY 40324 (502) 863-6800			X		1.4-10:1	not available	not available
81-03 Polybind DLR		X				10:1	0.5	not available
81-85 Polytack		X				0.04 lb/gal	0.26	not available

(Continued)

TABLE 3-12. CONTINUED

Product Name	Manf. Address, Telephone Number	Application Method: Topically By Truck	Topically By Fixed Mast	Mixed In Place	Other	Application Rate: Dilution	Applied gal/yd^2	FOB Price Before Dilution, Dollars
		Liquid Chemicals (Salts)						
Calcium Chloride, Flake	Allied Chemical Corp. Ind. Chem. Div. P. O. Box 6 Solvay, NY 13209 (315) 487-4000	X					1.0-1.5 lb	0.07/lb
Calcium Chloride, Liquid		X				None	0.4	0.225/lb
DP-10	Wen-Don Corp. P. O. Box 13905 Roanoke, VA 24038 (703) 982-0561	X				None	0.5	1.95/gal
Dust Ban 8806	Nalco Chemical Co. 2901 Butterfield Rd. Oak Brook, IL 60521 (312) 887-7500	X				None	0.25-0.5	0.22/gal
Dustgard ($MgCl_2$)	Great Salt Lake Minerals & Chemicals Corp. P. O. Box 1190 Ogden, UT 84402 (801) 731-3100	X				None	0.5	0.24/gal
Sodium Silicate(N)	The PQ Corp. P.O.Box 840 Valley Forge, PA 19482 (215) 293-7200	X				4:1	not available	0.69/gal
Sodium Silicate(O)		X				4:1	not available	0.71/gal

(a) Not applicable

(b) Application and price information is confidential

have been examined in several studies. None has been performed in conjunction with improvement of air quality, or control of dust emissions from hazardous waste sites. Testing has consisted of compressive strength testing, resistance to water erosion and weatherability. Weathering tests consists of placing a weighted amount of soil of known moisture content in a sheet pan, spraying the soil with a dust-suppressant, exposing the sample to weather, and reweighing the pan with moisture correction. The soil loss is the loss in weight through the period. These tests give qualitative results, but are several abstractions away from a large exposed area for the following reasons.

1. The soil is not naturally compacted in the baking pan.

2. The soil is not of the same thickness as would be found in place. The sample soil is less then two inches thick. Moisture probably behaves differently in this circumstance.

3. Using a hand spray bottle for suppressant application may not simulate the use of a high-powered sprayer on a large exposed area.

4. The before and after weights are compromised by dust and organic matter falling onto the test sheet.

None of the tests have involved ambient air sampling. References are Bureau of Reclamation (1977); Bureau of Reclamation (1982); U.S. Army Engineer Waterways Experiment Station (1977).

The only testing with ambient air measurements was performed by PEDCo Environmental, Inc. (1984a). A chemical tracer (zinc oxide) was applied to 50 x 50 foot test plots. Dust suppressants then were applied. Sampling was performed for several weeks with passive air samplers. The dust collected from the ambient air was analyzed with atomic absorption spectroscopy to determine the presence of zinc. Zinc concentrations above the natural background level occurring in the soil (75 ppm) indicated failure of the dust suppressant formed crust.

Materials tested, dilution, and application rates are shown in Table 3-13. Selection of the products shown for testing did not mean they were worse or better than other products available. These same products are listed in Table 3-12.

TABLE 3-13. TEST PLOTS APPLICATION DATA (PEI, 1985)

Dust Suppressant Name	Formulation	Appli. Conc.	Application Rate
Soil seal	Latex acrylic copolymer	3%	4 L/m^2 (1.0 gal/yd^2)
AMSCO-RES 4281	Carboxylated styrene-butadiene copolymer	20%	3 L/m^2 (0.6 gal/yd^2
Fiber mat	Non-woven geotextile	8 oz/yd^2	3-12 ft rolls
Flambinder	Lignon sulfanate	17%	2 L/m^2 (0.5 gal/yd^2)
Genaqua	Vinyl acetate resin	10%	0.9 L/m^2 (0.2 gal/yd^2)
Curasol	Synthetic resin	3%	1.3 L/m^2 (0.3 gal/yd^2)
M166/M167	Latex	7% (M166)+ 0.2% (M167)	2 L/m^2 (0.5 gal/yd^2)
CRF	Petroleum resin	25%	2 L/m^2 (0.5 gal/yd^2)
Sherman process (no grass seed)	Straw mulch bound with emulsified asphalt	--	--
Sherman process (with grass seed)	Straw mulch bound with emulsified asphalt	--	--
Terra Tack I	Vegetable gum	0.3%	6 L/m^2 (1.4 gal/yd^2)

After 16 to 30 days from application, all crusts remained intact. After 30 to 44 days from application, only the M166/M167 crust was intact. The zinc tracer values increased with time, representing the progressive failure of the crust over time.

Visual examination of the plots during the course of the tests revealed almost immediate plant growth on the initially bare plots. The vegetation eventually overran all of the test plots, totally destroying the dust controlling crusts and rendering the test plots indistinguishable from the surrounding study area. Even the fiber mat covering Plot 3 was overtaken by vegetation growing through the mat. This necessitated the use of a preemergent herbicide on most of the subsequent test plots. The use of the preemergent herbicide marked by decreased the amount of vegetation although a few plants still appeared on each plot.

Weed growth is quantified in Table 3-14. The presence of zinc in various segments of the test plot on July 20 is shown in this table. The first column indicates the saltation (ambient air) catch sample. Values range from 55 to 121 ppm. Of course the crust was very rich in zinc, since that is where the tracer had been added. (Values ranged from 163 to 544 ppm.) Below the crust, values were at or near background. The soil around the weed stems, however, apparently was composed of destroyed crust, because zinc levels ranged to 546 ppm — very nearly the same as in the crust levels. The loose soil around the weed stems was crumbly, had a very erodible texture, and would be highly subject to wind erosion.

An alternate procedure to eliminating vegetative growth would be to encourage it. Products are available that are temporary soil binders impregnated with grass seed. While grass was beginning to grow, the same problem described above would occur. Assuming a thick stand of grass did grow, control probably would not be 100 percent since there could always be some loose dirt between grass stems. Chemical dust suppressants sprayed on thick grass stands would not be effective because the suppressant would stick to the stems and little would penetrate the soil.

It is apparent that 100 percent effective control of wind eroded particulates will require higher dust suppressant concentrations and/or

TABLE 3-14. WEED GROWTH QUANTIFICATION (PEI, 1985).

Initial 8 Plots, July 20
(ppm)

Product	Saltation Sample	Crystal Sample	Sub-crust Sample	Soil Around Plant Stems
1	85	544	47	499
2	121	413	79	546
4	55	291	50	263
5	90	433	46	546
6	67	366	69	239
7	67	190	71	193
8	72	163	91	108

multiple applications beyond the measures tested in the field study. Considerations must also be made for the effects of weather and vegetation. Precipitation is detrimental to those suppressants that are water soluble (i.e., lignon sulfonate). Control of plant growth is essential if the crust formed by a dust suppressing product is to remain intact.

(b) Storage Piles--Various studies were located that evaluated the effectiveness of dust suppressants or windscreens in controlling fugitive dust from storage piles.

The use of chemical dust suppressants have been evaluated in two studies. The first study used a wind tunnel placed over a coal pile for the evaluation (Midwest Research Institute 1983). Because coal is a very different material than contaminated soil, results are only partially applicable. Control techniques studied were: (1) a 17 percent solution of Coherex in water applied at an intensity of 3.4 L/m^2 (0.74 gal/yd^2), and (2) a 2.8 percent solution of Dow Chemical M-167 Latex Binder in water applied at an average intensity of 6.8 L/m^2 (1.5 gal/yd^2). The control efficiency of Coherex applied at the above intensity to an undisturbed steam coal surface approximately 60 days before the test, under a wind of 15.0 m/s (33.8 mph) at 15.2 cm (6 in.) above the ground, was 89.6 percent for total particulate (TP) and approximately 62 percent for <15 (IP) and <2.5 (FP) micrometer size particles. The control efficiency of the latex binder on a low volatility coking coal two days after application, under a 14.3 m/s (32.0 mph) wind speed at 15.2 cm (6 in.) above the ground, was 37 percent for TP and near zero for IP and FP. However, when the wind speed was increased to 17.2 m/s (38.5 mph), the control efficiency increased to 90 percent for TP, 68.8 percent for IP, and 14.7 percent for FP. The efficiency under the same wind speed, 17.2 m/s, decayed four days after application to 43.2 percent for TP, 48.1 percent for IP, and 30.4 percent for FP (PEI, 1985).

Another study evaluated the use of chemical dust suppressants on a topsoil pile (PEDCo 1984c). Measurements were made with the RAM-1, a light-scattering instrument. Control efficiencies of more than 50 percent were estimated. Plots of emission rates indicated a lower rate of wind erosion than for an untreated pile, and wind erosion was not initiated until a higher threshold windspeed. The report concluded that the use of a chemical dust

suppressant was superior to a windscreen in controlling dust, in terms of effectiveness, cost, and mobility around the pile.

In conclusion, on chemical dust suppressants applied to inactive piles, control efficiencies of at least 50 percent are achieved. However, data indicate that dust suppressants cannot approach 100 percent effectiveness. Data reported are for an undisturbed pile and material being added or removed would necessitate retreatment. However, material cost is quite low, and only the distrubed area would need to be retreated. For piles with vehicle travel on them, control measures suitable to controlling vehicle reentrainment would need to be used as opposed to the materials listed in this section.

PEDCo (1984c) studied windscreens using RAM-1 aerosol monitors and windspeed sensors interfaced with a portable computer to give real-time data results. The analysis indicated that the windscreen did not produce significant reductions in concentrations in the less than 10 mmicrometer respirable size range. The screen did reduce windspeeds by the amount anticipated, but this did not result in commensurate reductions in particulate concentrations coming from the pile.

An explanation for the windscreen's performance is that wind erosion emission rates in the less then 10 micrometer size range are fairly constant at windspeeds above the threshold of about seven mph (hourly average). The additional emissions associated with high wind erosion losses at high windspeeds are larger particles that are not detected by the RAM-1's. The windscreen may be effective in stopping or reducing the movement of these large particles, but many of them do not stay airborne because of their relatively large size, so they present less of a threat of off-site exposure.

In summary, the results of all studies of reductions in windspeed caused by windscreens are in fair agreement. Reductions in total dust concentrations are achieved but reductions in concentrations of the smaller respirable sizes are questionable. A PEDCo study of particles in the less than 10 micrometer size respirable range shows no consistent benefit from the windscreen, but acknowledge that positive control efficiencies of larger size particles is

likely. However, control of the smaller size particles is more important because they are in the respirable range and because their small size will allow the wind to transport them far off-site.

C. Vehicle Emissions--

Dust from vehicles is produced by the action of the tire on the road which grinds the road surface and forces particles backwards and up. It also is produced by wind currents created by the moving vehicle which lift the dust from the roadway and the shoulder. Thus, both the road and the road shoulder are involved in this problem. If the road is unpaved, it should be as compacted (no loose particles) as possible. If the road is paved, it should be clear of windblown dust and spills. Road shoulders should be as compacted as possible to make it difficult for wind currents to lift particles.

1. Quantification of Emissions--For unpaved roads, the emission factor for an uncontrolled road (PEI Assoc. 1985) is:

$$E = k(5.9)\left(\frac{s}{12}\right)\left(\frac{S}{30}\right)\left(\frac{W}{3}\right)^{0.7}\left(\frac{w}{4}\right)^{0.5} \qquad \text{(Eq. 3-6)}$$

where E = Emissions, lb of $\leq$30 micrometer particulate
k = particle size multiplier (dimensionless) = 0.80
s = silt content of road surface material, percent
S = mean vehicle speed, mph
W = mean vehicle weight, tons
w = mean number of wheels

For paved roads the emission factor (PEI Assoc. 1985) is:

$$E = k(0.90)\ I\ \left(\frac{4}{n}\right)\left(\frac{s}{10}\right)\left(\frac{L}{1{,}000}\right)\left(\frac{W}{3}\right)^{0.7} \qquad \text{(Eq. 3-7)}$$

where E = Emissions, lb of $\leq$30 micrometer particulate
I = industrial augmentation factor (dimensionless), ranging from 1.0 to 7.0, usually 3.5
n = number of traffic lanes
s = surface material silt content, percent
L = surface dust loading, lb/mile
W = average vehicle weight, tons

Particles $\leq$30 micrometers are likely to stay in the air at distances greated than several hundred yards from the source. Particles greater than 100 micrometers usually settle out within 20 to 30 feet of the edge of the

road. Particles 30 to 100 micrometers probably will settle out within a few hundred feet of the road.

By examining the variables in the equation, the factors influencing dust emissions can be analyzed. For unpaved roads, emissions increase with increases in silt content in the road surface material, vehicle speed, vehicle weight and number of wheels. For paved roads, emission increase with an increase in silt content of the surface material, the quantity of material on the road, and vehicle weight. (It would seem that speed should be a factor for paved roads, as it was for unpaved, but apparently speed did not meet the statistical requirements to be entered into the equation.)

2. Principles of Control--The effect of dust suppressants on unpaved roads is to aid in compacting the dust particles. The ability of a road material to be compacted is highly dependent on the size gradation of that material. A proper size gradation for a roadway surface is shown in Table 3-15. The results of an improper size gradation are shown in Table 3-16. With too much gravel, there will be relatively little dust (until tires grind the gravel creating silt size particles), but the aggregate will be pushed to the side of the road. Any dust suppressant will pass through the top roadway surface and provide little control. If there is too much sand, bearing capacity will be poor, and any dust suppressant that attempts to from a crust will not work due to rutting. The most common of these conditions is excess silt and clay. For this condition, the dusting rate is high, since dust suppressants do not effectively penetrate the surface. In the rain, the road becomes muddy, rutted, and slippery. The dust suppressant will aggrevate these conditions.

For each unpaved road where dust control is required, roadway samples should be taken and a size gradation obtained. If the roadway aggregate does not meet the specifications on Table 3-15, additional material of the missing sizes should be added to the roadway. Without the proper size gradation of particles no chemical dust suppressant or watering, will be satisfactory.

The method of control on paved roads is to sweep, vacuum or flush the dust from the paved road surface (i.e., reduce the "L" term in Equa. 3-7). These methods tend to remove coarse particles more effectively than fine

TABLE 3-15. PROPER SIZE GRADATION FOR UNPAVED ROAD SURFACE (PEI, 1985)

Sieve Size	Percent Passing	Soil Type
1"	100	Gravel
3/4"	85-100	"
3/8"	65-100	"
#4	55-85	"
#10	40-70	Sand
#40	25-45	"
#200	10-25	Clay, Silt

TABLE 3-16. IMPROPER SIZE GRADATION (PEI, 1985)

Material in Excess	Bearing Capacity	Amount of Dust	When Wet	Action of Dust Suppressant
Gravel	Good	Little	OK	Drain through top level of soil. Little control.
Sand	Poor	Some	Soft	Drain through top level of soil. Little control.
Silt/Clay	Very Poor	Great	Mud/Ruts/ Slippery	May not penetrate. Will aggravate mud, ruts and slippery conditions.

particles, i.e., the silt term in Equation 3-3. Any paved road dust control program must concentrate on removing the fine material from the street.

Water is the most commonly used dust suppressant for unpaved roads. The quantity and frequency of application depend on the road material, the traffic level and the weather conditions. Chemical dust suppressants can be more effective than water alone and are sometimes more cost-effective than water where usable water is not readily available at the site, which is commonly the case in oil shale projects.

A comprehensive questionnaire survey indicated that about 40 manufacturers marketed various products for unpaved road dust suppression. Available products were divided into the following four categories:

Salts--Hygroscopic compounds that extract moisture from the atmosphere and dampen the road surface, e.g., calcium chloride, magnesium chloride, hydrated lime, sodium silicates, etc.

Surfactants--Compounds capable of reducing liquid surface tension thereby allowing available moisture to wet more dust particles per unit volume, e.g., soaps, detergents, Dust-set, Monawet, etc.

Adhesives--Compounds that are mixed with native soils to form a crust-like new surface, e.g., calcium lignon sulfonate, sodium lignon sulfonate, ammonium lignon sulfonate, etc.

Bitumens--Compounds derived from petroleum that are mixed with native soils to form a cohesive surface, e.g., Coherex, asphalt, oils, etc.

Salts, adhesives, and bitumens can be applied topically (sprayed on the road surface) or mixed in place (blade mixed with the top four to six inches of the roadbed) at intervals of weeks or months. Surfactants are routinely added to the water in water wagons and applied at intervals of hours. In selecting a chemical dust suppressant type, the type of roadway aggregate should be considered as shown in Table 3-15.

3. Product Availability and Cost--Names of commercial products, their application method, dilution, rate and cost are shown in Table 3-15. Also shown is a telephone number for the main office where each of the products is manufactured.

The products shown in Table 3-17 represent suppressants available in 1984. Some products may be no longer available while some new products may not be listed. Products listed are not endorsed over products not listed. The results of control efficiency evaluations are described in the next section.

The cost elements involved in dust suppression include labor and material associated with road surface preparation, the dust suppressant used, application, and road maintenance (grading, watering, and aggregate). A recent study (PEI Assoc. 1985) cited total costs of applying specific types of dust suppressants at a rate and frequency to achieve a 50 percent control level in a coal mine. These data are directly applicable to an oil shale mine. Assumptions used for the analysis are shown in Table 3-18. These assumptions are based on the following.

- Product costs were obtained from each vendor and represent the least expensive per gallon cost available shipped to both a Wyoming and an Illinois destination.
- Labor and machinery values are industry averages obtained from mine personnel by PEI. Rates vary by mine depending on local labor contracts, machinery type and age.
- The water was assumed to be free. The reader should adjust these figures for actual water costs in his area.
- Activity parameters (miles graded per hour, etc.) are industry averages and vary by mine. Actual time to apply the suppressants for the tests was not used since the relatively short test segments did not facilitate optimum application procedures. Identical parameters were used for all chemicals mixed in place, with a second set of activity parameters for all topical applications.

These assumptions were used to calculate costs associated with the use of chemicals and water for dust suppression. The analysis of chemical dust suppressants was limited to calcium chloride (mixed and topical applications) and lignon (mixed in place only).

The results of the comparison of cost effectiveness to achieve a minimum 50 percent control level are shown in Table 3-19. The limited results show that the topically applied salt or the mixed in place adhesive is more

TABLE 3-17. UNPAVED ROAD DUST SUPPRESSANTS[a] (PEI, 1985)

Product Name	Manf. Address, Telephone Number	Application Method		Application Rate		FOB Price Before Dilution Dollars/ gallon
		Topical	Mixed in Place	Dilution[b]	Applied gal/yd^2	
			Salts			
Dowflake (dry product)	Dow Chemical Larkin Laboratory Midland, MI 48640 (517) 636-0949	X	X	NA	1.55 lbs.	0.0725
DP-10	Wen-Don Corp. P.O. Box 13905 Roanoke, VA 24038 (703) 982-0561	X		None	0.5 (2 L/m^2)	1.95
Dust Ban 8806	Nalco Chemical Co. 2901 Butterfield Rd. Oak Brook, IL 60521 (321) 887-7500	X		None	0.25-0.5 (1-2 L/m^2)	0.22
Dustgard ($MgCl_2$)	Great Salt Lake Minerals and Chemical Corp. P.O. Box 1190 Ogden, UT 84402 (801) 731-3100	X	X	None	0.5 (2 L/m^2)	0.24

[a] Products listed are not endorsed over products not listed.
[b] Water: product.
(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method: Topical	Application Method: Mixed in Place	Application Rate: Dilution[b]	Application Rate: Applied gal/yd^2	FOB Price Before Dilution Dollars/gallon
			Salts (continued)			
Liquidow	Dow Chemical Larkin Laboratory Midland, MI 48640 (517) 636-0949	X	X	None	0.27-0.6 (1.2-3 L/m^2)	0.20
Sodium Silicate (N)	The PQ Corporation P.O. Box 840 Valley Forge, PA 19482 (215) 293-7200		X	4:1	NA	0.69
Sodium Silicate (O)			X	4:1	NA	0.71
			Surfactants			
M070E	Mona Industries, Inc. P.O. Box 425 76E 24th Street Paterson, NJ 07544 (210) 345-8220	X		NA	NA	6.30
Sterox DF/ND/NJ	Monsanto Company 800 N. Lindbergh Blvd. St. Louis, MO 63166 (314) 694-1000	X		NA	NA	6.35

(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method		Application Rate		FOB Price Before Dilution Dollars/ gallon
		Topical	Mixed in Place	Dilution[b]	Applied gal/yd^2	
			Adhesives			
Bio Cat 300-1	Applied Natural Systems 35 E. Lake Mead Drive Henderson, NV 89015 (702) 451-6010	X	X	66:1	2.0 (9 L/m^2)	19.95
DCL-1801	Calgon Corp. P.O. Box 1346 Pittsburgh, PA 15230 (412) 777-8000	X	X	66-200:1	0.5-0.8 (2-3 L/m^2)	9.20
DCL-1803		X	X	100-200:1	0.5-0.8 (2-3 L/m^2)	30.64
Dust Bond 100	Research Products, Inc. 4222 North 39th Ave. Phoenix, AZ 85019 (602) 269-7891	X	X	None	0.17 (0.8 L/m^2)	0.40
Dust-Set	Mateson Chemical Corp. 1025 East Montgomery Ave. Philadelphia, PA 19125 (215) 423-3200	X	X	500:1	0.17 (0.8 L/m^2)	8.00
Dustbinder 124	Union Carbide Corp. Mining Chemicals 270 Park Ave. New York, NY 10017 (203) 794-2000	X		10-15:1	1.0 (5 L/m^2)	4.50

(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method		Application Rate		FOB Price Before Dilution Dollars/ gallon
		Topical	Mixed in Place	Dilution[b]	Applied gal/yd^2	
		Adhesives (continued)				
Flambinder	Flambeau Paper Company P.O. Box 340 Park Falls, WI 54552 (715) 762-3231	X	X	5.5:1	0.5 (2 L/m^2)	0.15
Haul Road Dust Control	Midwest Industrial Supply, Inc. P.O. Box 8431 Canton, OH 44711 (216) 499-7888	X		33:1	NA	3.75
Lignosite	Georgia-Pacific Corp. P.O. Box 1236 Bellingham, WA 98227 (206) 733-4410	X	X	4:1	0.5 (2 L/m^2)	152.00/ton
Norlig A	Reed Lignin, Inc. 120 East Ogden Ave. Suite 106 Hinsdale, IL 60521 (312) 887-9640	X	X	1.1	2.8 (13 L/m^2)	0.765
Norlig 12		X	X	2.4 lbs/ gal	2.8 (13 L/m^2)	0.228/lb
Orzan AL-50/Orzan DSL/ Orzan GL-50	Crown Zellerbach Corp. Chemical Products Division Camas, WA 98607 (206) 834-4444	X	X	10:1	1.2-6.3 (5-30 L/m^2)	0.20

(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method: Topical	Application Method: Mixed in Place	Application Rate: Dilution[b]	Application Rate: Applied gal/yd^2	FOB Price Before Dilution Dollars/gallon
		Adhesives (continued)				
Soil-Sement	Midwest Industrial Supply, Inc. P.O. Box 8431 Canton, OH 44711 (216) 499-7888	X		5:1	0.25 (1 L/m^2)	2.32
Soiltex	Protex Industries, Inc. 1331 West Evans Ave. Denver, CO 80223 (303) 935-3566		X	4:1	4.8 (20 L/m^2)	0.33
Suferm	Chevron Chemical Co. Sulfur Products 575 Market Street San Francisco, CA 94105 (415) 894-6723	X	X	None	0.2 (0.9 L/m^2)	1.88
WESLIG 120	WESCO Technologies, Ltd. P.O. Box 3880 San Clemente, CA 92672-1680 (714) 661-1142	X		6-10:1	0.25 (1 L/m^2)	0.42
Woodchem LS	Woodchem, Inc. P.O. Box A Oconto Falls, WI 54154 (414) 846-2839	X	X	None	1.5 (7 L/m^2)	0.17

(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method: Topical	Application Method: Mixed in Place	Application Rate: Dilution[b]	Application Rate: Applied gal/yd^2	FOB Price Before Dilution Dollars/ gallon
			Bitumens			
AMS 2200	ARCO Mine Sciences 1500 Market Street P.O. Box 7258 Philadelphia, PA 19101 (215) 557-2000	X	X	4:1	0.5 (2 L/m^2)	0.5
AMS 2300		X	X	1:1	0.75 (3 L/m^2)	0.75
Coherex	Witco Chemical Golden Bear Division P.O. Box 378 Bakersfield, CA 93302 (805) 393-7110	X	X	10:1	0.5 (2 L/m^2)	1.25
Docal 1002	Douglas Oil Co. 3160 Airway Ave. Costa Mesa, CA 92626 (714) 540-1111	X	X	2:1	0.1 (0.4 L/m^2)	0.67
Peneprime	Utah Emulsions Co. P.O. Box 248 North Salt Lake, UT 84054 (801) 292-1434	X		None	0.5 (2 L/m^2)	1.23
Petro Tac P	Syntech Products Corp. 520 E. Woodruff Ave. Toledo, OH 43624 (419) 241-1215	X	X	1:5	0.24-0.75 (1-3 L/m^2)	1.55

(continued)

TABLE 3-17. (Continued)

Product Name	Manf. Address, Telephone Number	Application Method		Application Rate		FOB Price Before Dilution Dollars/ gallon
		Topical	Mixed in Place	Dilution[b]	Applied gal/yd^2	
			Bitumens			
Resinex	Neyra Industries, Inc. c/o Petroleum Products, Inc. P.O. Box 493 Valparaiso, IN 46383 (219) 465-1300	X	X	10:1	1.25-5.00 (5-20 L/m^2)	1.48
Retain	Dubois Chemical Co. 3630 East Kemper Road Sharonville, OH 45241 (513) 769-4200	X	X	10:1	0.5 (2 L/m^2)	5.55

TABLE 3-18. ASSUMPTIONS FOR COST-EFFECTIVENESS ANALYSIS (PEI, 1985)

Cost Item	$/h	Hours/Mile Chemicals Mixed	Hours/Mile Chemicals Topical	Hours/Mile Water	Frequency
Application					
Surface Prep	75	16	8	0	1/application
Application	45	8	2	0.15	1/application
Grading Cost	75	0	0	4	Water, 1/wk Chemicals, 1/ application
Remaining Watering	95	0.15	0.15	0.15	Water, 1.5/h Chemicals, 1/ shift

TABLE 3-19. PRELIMINARY COST-EFFECTIVENESS COMPARISON TO ACHIEVE 50 PERCENT CONTROL (PEI, 1985)

Control[2]	Cost of Chemical Application/ Mile, $[1] East	West	Cost of Grading Watering/Week, $ Grading	Water	Applications Required to Average 50% Control	Cost Per Week East	West
Salt							
Mixed	7240	11,300	0	143	1/4 weeks	1950	2970
Topical	3260	5,100	0	143	1/4 weeks	960	1420
Adhesive							
Mixed	4800	7,600	0	143	1/4 weeks	1340	2040
Water			375	1710	1/week	2100	2100

(1) Includes surface preparation, material cost, and application. Material cost is delivered cost in East (southern Illinois) and West (Rock Springs, WY). Material cost is: Liquidow, $0.36/gallon East, $0.47/gallon West; Flambinder $0.33/gallon East, $0.47/gallon West. Assumes 50-ft and 60-ft wide road in East and West.

(2) The application intervals required could not be estimated for adhesive--topical, surfactant, or bitumens. Comparative costs could not be calculated.

cost-effective than watering. However, the choice of dust suppressant strategies also should include other considerations related to road construction and spillage as explained in the next section.

Reapplications of the chemicals probably would result in higher control efficiencies than the initial application because of residual traces of the material. Therefore, this analysis based on initial applications may overestimate the cost of a long-term chemical program. Conversely, watering displays no such cumulative control effects. The analysis was performed for the mine and for the haul road. The heavy vehicles and high speeds make dust suppressions more difficult on the haul road where more frequent application would be required.

Equation 3-6 has a factor of S/30 to describe the effect of speed on emissions. For example, a change in speed from 30 to 20 mph would result in a 33 percent reduction in emissions. While this factor may overestimate emission reductions with reduced speed, the principle holds. The cost of speed control is increased labor and equipment time to haul material.

The base emission factor for an unpaved road versus a paved road given in Equation 3-2 and 3-3 is 5.9 lbs/mile traveled versus 0.09 lbs/mile traveled, a reduction of 98.5 percent. This control is far more efficient than water, chemicals speed control or housekeeping. However, to maintain that control efficiency, the paved street must be continually cleaned.

Costs vary between area of the country and the thickness of pavement required to support truck weight. The average cost to blacktop a road suitable for over-the-road trucks is about $140,000 per mile of 2-lane road, plus street cleaning costs.

Paved roads, which become dirt-laden due to spills, track-on and windblown dust, must be cleaned to remove all loose dirt, particularly the fine particles. Manual cleaning may be adequate for short sections of road. However, it is very labor intensive. Street sweeping is most common. However, street sweeping is relatively ineffective at removing fine particles. In one series of tests, material 74 to 177 microns in size was applied to a paved street at a loading of 600 grains per square foot. Removal efficiency was 46 to 63 percent. Silt size particles, the particle size most

likely to be entrained, are less than 74 micron in size. Removal efficiency of this size particle is probably less than 46 percent. In addition, street cleaning itself makes dust due to tires rolling on the road, brushing dry pavement, and wind turbulence caused by exhaust and vehicle movement. As will be discussed in the next section, the net cleaning impact on improving ambient air will be small or negative unless the street is very dirty.

Vacuum sweeping is more efficient than sweeping. In the above experiment, collection efficiencies of 90 to 92 percent were noted. Again, collection of silt particles would probably be less, and again, some dust emissions are caused by the sweeper itself.

Street flushers hydraulically move street debris from the street surface to the gutter. Often flushing is used as an aid to sweeping, and not the sole method of cleaning. Flushing before sweeping washes street dirt to the curb for collection by motorized sweepers. This type of flushing ordinarily employs smaller quantities of water and lower nozzle pressures. The benefits of flushing after sweeping are that the entire pavement is made cleaner and that only small quantities of dirt are washed into inlets and catch basins. Again, however, flushing is more effective at removing larger particles.

4. Effectiveness--Watering test results are shown in Table 3-20. In general, watering once per hour will have a control effectiveness of 50 percent. Watering twice per hour, or once every two hours, will have an effectiveness of about 75 and 30 percent respectively. Greater effectiveness may be realized during evening hours and during humid periods.

No surfactant tests were conducted. Efficiencies at the same level of water use should exceed plain watering. Although the objective of using a surfactant is to reduce water use, the effectiveness of decreased watering with a surfactant has not been tested.

To obtain the nearly 100 percent control needed at some sites, it is recommended that watering be initially applied at 0.125 gallons/yd^2 every 20 minutes. If muddy conditions develop, reduce to 30 minutes or more as possible to achieve near total dust control.

TABLE 3-20. COMPARISON OF MEASURED CONTROL EFFICIENCIES (PEI Assoc. 1985)

Control	No. of Samples	Location of Test	Vehicle Type[a]	Time After Application	Control Efficiency	Particle Size[b]
Water	3	I&S[c]	HD	0.5-4.5h[d]	96-55	TP
					98-50	IP
					98-61	FP
	9	Coal Mine	HD	0-1h	69-59	TSP
					73-61	IP
					58-54	FP
	3	Coal Mine	HD	0-0.5h	88	TSP
				0-0.25h	97	TSP
	26	Coal Mine	HD	0.3-1.0h	75-25	TSP
	3	I&S	HD	1.0-4.8	98-61	TP
					98-78	IP
					98-79	PM_{10}
					96-67	FP
	24	Coal Mine	HD	0.5-2.0h	77-12	TSP
	12			1.0-2.0h	66-31	IP
	24			0.5-2.0h	60-15	FP
Bitumens	2	I&S	LD	< 1 day	97-84	TSP
					97-83	FP
	4	I&S	HD	0-48h	98-92	TP
					96-91	IP
					97-90	FP
	5	I&S	LD	25-51h	100-94	TP
					99-91	IP
					97-94	FP
	2	I&S	NR	28-29 days[e]	95-90	TP
	24	Other	LD	0-3.5 mo	80-0	TSP
	24	Other	LD	0-3.5 mo	86-36	TSP

(Continued)

TABLE 3-20. CONTINUED

Control	No. of Samples	Location of Test	Vehicle Type[a]	Time After Application	Control Efficiency	Particle Size[b]
	4	Coal Mine	HD	0-3 days	89-59	TSP
	4	Coal Mine	HD	3.5 weeks	35	TSP
	8	I&S	HD	2-116 days	100-21	TP
					100-0	IP
					100-0	PM_{10}
					99-0	FP
	8	I&S	HD	7-77 days	87-16	TP
					86-22	IP
					97-36	PM_{10}
					100-25	FP
	4	I&S	HD	4-35 days[f]	98-88	TP
					98-91	IP
					100-90	PM_{10}
					100-25	FP
	30	Coal Mine	HD	1-4 weeks	64-0	TSP
	18				85-0	IP
	30				88-0	FP
Adhesives	3	Other	HD	0.5-1 day	83-91	TSP
	36	Coal Mine	HD	1-4 weeks	80-0	TSP
	25				91-0	IP
	36				75-0	FP

(Continued)

TABLE 3-20. CONTINUED

Control	No. of Samples	Location of Test	Vehicle Type[a]	Time After Application	Control Efficiency	Particle Size[b]
Salts	1	Coal Mine	HD	3 mo[g]	95	TSP
					95	IP
					88	FP
	34	Coal Mine	HD	1-7 weeks	83-0	TSP
	26				74-0	IP
	34				80-0	FP
Surfactants	27	Coal Mine	HD	1-6 weeks	87-0	TSP
	20				68-0	IP
	27				85-0	FP

[a] HD Heavy duty
LD Light duty
NR Not reported

[b] TP Total particulate
TSP Total suspended particulate (<30 m)
IP Inhalable particulate (<15 m)
PM_{10} Particulate matter <10 m
FP Fine particulate (<2.5 m)

[c] Iron and Steel
[d] Testing began at 3:00 p.m. and contmnued past dusk
[e] Time since third application
[f] Time since second application
[g] Road was watered prior to test

Data for the effectiveness of chemical dust suppressants also is shown in Table 3-20. Data vary widely depending on days since application, application rates, traffic volumes, vehicle size, the receiving surface, and testing methodologies. In the first week after application, efficiencies of 80 percent or greater can be achieved. After one month, values of 40 to 60 percent are common under heavy duty vehicle use. These data are for initial applications. Almost no data are available for chemical reapplications. Higher efficiency values could reasonably be expected.

From an air quality perspective, the relative merits of topical application versus mixed in place application are unclear. It appears that salt and bitumen generally performs better when topically applied. Conversely, lignon was better when mixed in place.

In light of the greatly higher cost involved with mixed-in-place application as opposed to topical application, these test results suggest that the $CaCl_2$ and bitumens be applied topically. At any rate, mixed-in-place application is usually only recommended at the time of initial application.

To achieve the total control of dust when necessary, it is recommended that a second chemical dust suppressant application be made 4 to 10 days after the initial application. The time should be based on observation. Time between applications can be gradually lengthened to near 30 days assuming spillage and track-on are being controlled.

When hourly watering is compared with application of chemical dust suppressants (and once/shift watering), every four weeks costs and control efficiencies are similar (see Section 3.2), depending on the chemical used. Other considerations can impact the selection as indicated in Table 3-20. Chemicals often may be the material of choice, since containment of contaminated water runoff from the road may be problematical.

However, chemical dust suppressants are not feasible at sites where road construction is so poor that the road must be regraded or rebuilt with new aggregate after all major storms. Regrading or rebuilding almost totally destroys chemical effectiveness. Thousands of dollars of chemicals could conceivably could be wasted within days of application.

Control effectiveness testing of paved roads has dealt primarily with the effectiveness of sweeping on city streets in lowering ambient air quality. City streets are relatively clean to begin with as compared to paved industrial roads which would be more similar to streets at an oil shale site used for material hauling.

Results on the effect of cleaning non-industrial streets are virtually all in agreement. Based on results from several cities, it appears that street cleaning of unpaved roads makes no significant difference in ambient air quality, and may make air quality slightly worse. A frequent explanation is that more dust is generated during the sweeping, than fines are removed from the road.

Street sweeping of industrial roads is much more limited. Based on four exposure profiler tests per control (MRC 1983) in a steel mill, the highest measured values for the control efficiency of vacuum sweeping, occurring 2.8 hr (midpoint of test) after vacuuming, were 69.8 percent for TP. Another test 4.1 hours after vacuuming showed a control efficiency of 16.1 percent. The control efficiency for water flushing at 2.2 L/m^2 (0.48 gal/yd^2), approximately 40 minutes after application, was 54.1 percent for total particulates. A subsequent test after 3.6 hours showed a value of 44.1 percent. The control efficiency for flushing and broom sweeping approximately 40 minutes after application with water applied at 2.2 L/m^2 (0.48 gal/yd^2), was 69.3 percent for TP. After 2.8 hours, the control efficiency was 34.6 percent. Control efficiency decay on a paved road is more a function of how much material is being deposited on the road from spilling and windblown dust, than actual decay in control efficiency (assuming control measurements are made under dry road conditions). However, because some of the steel plant tests were performed immediately after flushing, portions of the levels of control being measured are probably the effect of moisture.

3.2 NITROGEN OXIDES (NO_x) CONTROL

In a shale oil plant, NO_x is produced by the various combustion processes associated with retorting and the generation of heat and power. NO_x is formed by the oxidation of nitrogen (N_2) in the combustion air or in the fuel. "Thermal NO_x" is the term used for the NO_x produced from the N_2 in the air and

"fuel NO_x" is the term used for that produced from the nitrogen bound into the fuel.

The control of thermal NO_x is accomplished best by combustion design which prevents the formation of NO_x rather than destroying it. If combustion designs (or modifications in the case of existing equipment) techniques are not effective, there are two exhaust gas cleanup processes which can be employed: thermal DeNOx and selective catalytic reductions. Combustion design is always the most cost effective NO_x control on new equipment; but, sometimes, is subject to certain process limitations such as requirements to produce a high gas temperature or a high heat release rate.

One of the combustion design concepts which produces low NO_x emissions is called "staged combustion." Stage combustion involves a sequencing of the addition of combustion air to the fuel system. This lengthens the flame (or combustion reaction zone) and provides time for certain combustion chemical reactions to take place in a sequence that will avoid NO_x formation. It also provides an opportunity to control the temperature of the reaction zone which is important in NO_x prevention. Certain processes cannot accommodate the staged combustion characteristics; and, therefore, exhaust gas cleanup concepts must be employed on these processes.

Both of the exhaust gas NO_x removal processes available involve the use of ammonia or urea to reduce the NO_x to N_2 and H_2O. One process (Thermal $DeNO_x$*) involves the injection of ammonia in the exhaust gas stream at a precise temperature of 950° ± 30°C. Typically, 50 percent control efficiency can be achieved with this medium cost process. The use of urea to reduce the NO_x is in the development and evaluation stages. A California utility currently is experimenting with the use of urea in this application. The other NO_x removal technique involves the use of a catalyst which lowers and widens the temperature range over which the ammonia is effective. This process, referred to generically as selective catalytic reduction (SCR), is available from several manufacturers who feature, among other things, different structural forms of the catalyst. The catalyst is expensive. In a dirty gas stream, especially one containing trace metals, the catalyst life

*Patented by Exxon Corp., New Jersey

could be prohibitively short. SCR systems can achieve approximately 85 percent removal when the catalyst is uncontaminated.

NO_x removal by wet scrubbing is not effective for combustion gases because the NO_x is predominately NO which is relatively insoluble in water. NO_2 is relatively soluble in water and can be successfully scrubbed. Therefore, if the exhaust has high NO_2, a wet scrubber could be considered for NO_x reduction. But no process to first oxidize NO to NO_2 then scrub it has been applied on a significant commercial basis.

This presentation on NO_x controls is concerned primarily with combustion staging and provides only summary and reference material on Thermal $DeNO_x$ and SCR.

3.2.1 Staged Combustion

The mechanisms involved in the formation and emission abatement of NO_x are extremely complicated and only beginning to become understood by the researchers in the field. Most of the recent NO_x research has dealt with coal, primarily with pulverized coal burners. Many experiments have been conducted to quantify physical and chemical reaction rates in order to develop a consistent model. One thing has been apparent from this research, i.e., the NO_x that is produced is dependent on the broad parameters of time, temperature and stoichiometry. The time phasing of volatilization, mixing, and chemical reaction are critical. These rates can change by orders of magnitude with only a 100 C temperature change. The local fuel/air ratio is a critical factor in the formation of NO_x. It is difficult to design a low NO_x combustion system that will work every time for every fuel. Scaling up from pilot sizes to full scale can change results significantly.

In the coal-burning utility industry, manufacturing firms build boilers to meet a size specification and must guarantee the NO_x emission level. Their basic approach has been to use discrete burner modules in which the combustion and NO_x level have been established before the hot gas is discharged to the heat exchange section. Whether the boiler has 10 or 20 burners, each of them is the same and the NO_x which is emitted can be predicted by burning the specified coal in a subscale boiler in the manufacturer's plant.

The modular technique is not so readily adaptable to an oil shale retort because shale oil retorting itself is still in a state of evolution. Nevertheless, it is possible to apply the staged combustion principles early in the retort design and then test and adjust them throughout the pilot plant and semi-works stages. Because the staged combustion principles still are evolving, it is important that any design team include a staged combustion researcher who keeps up with the current state-of-the-art.

The principles discussed below are simplified and philosophical rather than rigorous. They are intended to inform the retort designer of the complexities, not to provide a design procedure. For standard combustion equipment (boilers, heaters, incinerators, etc.) that may be acquired for a shale oil project, the discussion which follows should provide some advice on what features to expect and what NO_x emission levels to anticipate.

A. Thermal NO_x--

In a conventional combustor burning gas, oil or coal with excess air in the range of 30 percent, the flame temperature is 1600 C (3000°F) or higher. At these temperatures, a fraction of the oxygen and nitrogen present in the combustion air will form NO_x in the thermal fixation reaction:

$$N_2 + O_2 \rightleftarrows 2NO.$$

This phenomenon is well known. In the early 1970s the first NO_x reduction techniques addressed this phenomenon.

It was reasoned that the rate determining mechanism for NO_x formation was the reaction:

$$N_2 + O \rightleftarrows NO + N$$

which is highly temperature dependent and occurs to an appreciable extent only at temperatures above 1600 C (Hayhurst and Vince, 1980). The principal source of oxygen atoms for this reaction is the dissociation of molecular O_2, although other reactions probably contribute oxygen atoms.

The free nitrogen further reacts with molecular oxygen:

$$N + O_2 \rightleftarrows NO + O$$

Even in fuel rich flames where hydroxal radicals, OH, are present, NO can be formed from the reaction:

$$N + OH \rightleftarrows NO + H$$

Thus, early NO_x control involved schemes to produce lower flame temperatures. This approach is still essentially valid for nitrogenless fuel. The research in this area continues. The early theory failed to account for experimentally derived NO formation rates in the flame region. It is not the purpose here to explain the many theories now found in the literature. However, to illustrate the complexity of the combustion chemistry in thermal NO formation and to help indicate why an emperical approach to NO_x control is taken by combustion hardware designers, the following equation proposed by Bartok, et. al. (1972) is given:

$$\frac{d[NO]}{dt} = \frac{2\,k_o\,[O]^2}{k_1\,[O_2] + k_4\,[NO] + k_4\,[OH]_{eq}}\; k_3 k_5\,[N_2]\,[O_2]$$

$$+ k_3\,k_7\,[N_2]\,[OH]_{eq} - k_4\,k_6\,[NO]^2 - \frac{k_4\,k_8\,[NO]^2\,[H]_{eq}}{k_o\,[O_2]}$$

where the various k_i are equalibrium constants for the N_2, O_2, N, O dissociation and recombination reactions and the symbol, $[\ ]_{eq.}$, indicates equilibrium concentrations of the species inside the brackets.

Based on this general understanding of the thermal NO_x formation mechanisms, various control concepts involving modifications to both hardware and operating procedures have been identified and evaluated to reduce NO_x emissions from boilers, heaters, etc.

In principle, thermal-NO_x can be minimized using the following techniques:

- Reduce the peak temperature
- Lower the local N_2 concentrations at peak temperature
- Reduce the residence time at peak temperature, and
- Decrease the local O_2 concentrations at peak temperature

Lowering the N_2 levels requires the use of oxygen-enriched air, which is not considered cost-effective. Therefore, efforts in the field have focused on the decreasing of O_2 levels, peak temperature, and residence time in the NO_x-producing regions of a furnace. On a macroscopic scale, techniques such as low-excess-air (LEA) combustion and off-stoichiometric combustion (OSC) have been used to lower local O_2 concentrations in utility and industrial boilers. In addition, flue-gas recirculation (FGR), water injection, load reduction, and reduction of air preheat have been used in boilers to control thermal NO_x by lowering the peak flame temperatures. FGR also decreases combustion-gas residence time and local O_2 levels, but its primary effect is to lower peak temperatures.

Thermal NO_x formation is largely a burner phenomenon, since the temperature and availability of oxygen in the bulk gas downstream is usually too low to support NO formation. To a large measure, the amount of NO formed is dependent upon the design of the burner. A burner that is efficient in promoting complete mixing and combustion close to the burner will have combustion products that will experience a subsequent high temperature time history. Hence, the rate of NO_x formation is high. In contrast, a burner that promotes a delayed mixing of the primary and secondary air streams will result in a long, low-turbulent flame with reduced peak flame temperatures. Similarly, a burner that can operate at reduced excess air levels will have lower emissions than a burner that requires a higher levels of excess air.

It has been demonstrated through numerous studies on existing on-line units that modifications to the combustion process can be implemented using existing equipment. This is accomplished in several ways.

A relatively simple technique is to use the flexibilities in combustion air damper position and degree of swirl which exist on nearly every burner. The local fuel and air distribution at each individual burner can thereby be affected to various degrees and operation of the unit optimized with regard to NO_x emissions and minimum excess air level.

An OSC approach involves staging the combustion process by operating the burners at fuel-rich conditions and introducing the remaining air required to complete combustion in the bulk gas region. This can be accomplished either by using existing "NO_x ports", where present, or by operating with less than a full complement of burners in service. In this manner, the fuel is distributed to fewer burners whereas the air supply, which is maintained through the out-of-service burners, remains fixed. The local fuel/air ratio at the in-service burners is therefore higher. Using this technique, the formation rate of NO is lowered by reducing the availability of oxygen in the primary combustion zone. It also maintains a lower flame temperature because the final air addition takes place in a portion of the combustion chamber where heat removal occurs. This technique requires a field test program to optimize the burner out-of-service pattern to be used. Since each unit (even units of identical design and construction) operate differently, this generally requires that a program be conducted on each unit to minimize the formation of NO_x.

In an FGR system, the recirculated flue gases are inert materials which absorb a part of the energy released during the combustion process and thereby reduce the peak temperatures achieved. This technique, however, requires the existence of a windbox flue gas recirculation system which is not always available on industrial boilers of the size used for shale oil projects.

Another relatively simple approach involves the adjustment of combustion air preheat temperature. However, this technique, however, causes several practical operational problems and reduces the fuel use efficiency.

Although NO_x formation is a burner phenomenon, the furnace design can have a substantial influence on the amount of NO_x produced. This influence is manifested in several ways. The temperature of the bulk gas, which controls

the rate of temperature decay of the combustion products and thus the rate of decrease in NO_x formation, is fixed by the heat release rate per unit volume and is effected locally by circulation patterns and the promixity of cold surfaces. Bulk gas recirculation into this primary combustion zone following cooling by the furnace walls can affect NO_x formation by reducing peak flame temperature. The existence of "NO_x ports" allows greater flexibility in the implementation of staged combustion operation.

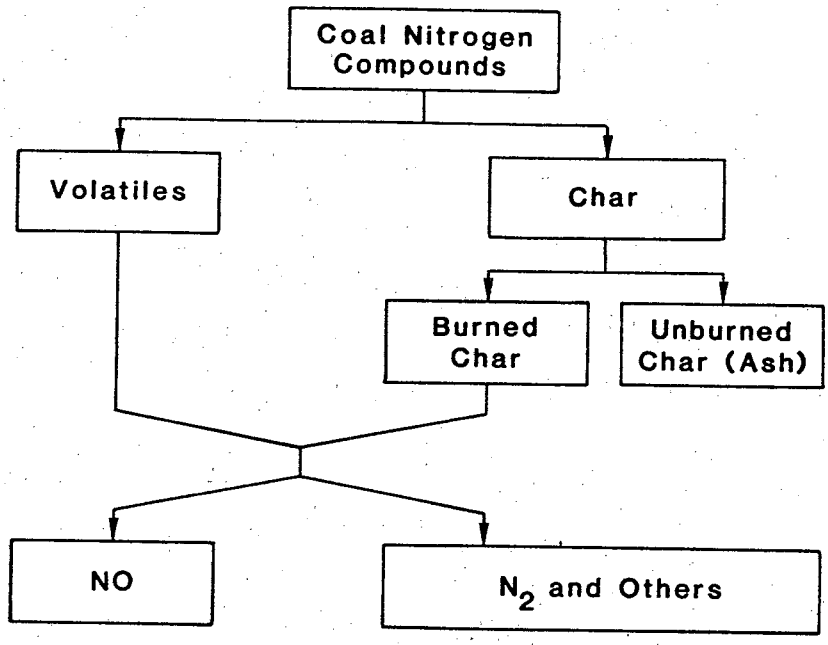
B. Fuel NO_x--

Fuel NO_x is the principal form of NO_x to be controlled in a shale oil project. While there may be conventional boilers firing natural gas or light oils where thermal NO_x is a concern, the major concern is with the retoring process or any combustor using the shale, the shale oil, other nitrogen containing oil or coal as fuel.

There are several extensive documents treating fuel NO_x controls as associated with coal firing primarily in utility boilers [Thompson, et al, 1985; Huang, 1981; and KVB, 1977]. This discussion will acquaint the reader with some of the technical factors and suggest certain approaches that might be incorporated into any new system involving the burning of shale, shale oil, etc. These principles may also suggest modifications to R&D burners as test data are obtained.

Two mechanisms for the formation of NO_x from fuel-nitrogen have been proposed. The combustion of the fuel yields volatile matter released from the fuel particles and char. Fuel-NO_x can be created in combusting both forms of fuel as either volatile-NO_x or char-NO_x. These two reactions are interrelated (as are thermal-NO_x and fuel-NO_x), because any fuel-nitrogen not emitted as volatiles must remain as part of the char.

The fate of fuel nitrogen is shown schematically in the diagram which indicates the complex manner in which combustion takes place.



The first step in the process, the pyrolysis of the fuel, determines the fractions of fuel-nitrogen to be associated with volatiles and char. The manner in which the volatiles evolve depends on the time-and-temperature history of the fuel particles. Tests by Pohl and Sarofim (1977) have demonstrated that about 10 percent (by weight) of coal is volatilized before any fuel-nitrogen is released. It is believed that this reflects the time required to break down the structures of the nitrogen-containing compounds. After the initial delay, fuel-nitrogen is released at a rate higher than, but proportional to, the rate of incremental changes in weight of the coal particles.

The precise form in which fuel-nitrogen is released is not known. Similarly, no complete reaction mechanisms for the combustion of the volatile fuel-nitrogen components have been developed, although it has been postulated that ammonia and hydrogen cyanide are the major intermediates for NO_x formation. (In burning the retorted shale the fuel is principally char. Thus the nitrogen may not be in the form of hydrogen compounds.) Laboratory studies

have demonstrated that the conversion of fuel nitrogen to NO_x is strongly affected by the local stoichiometry. However, contrary to the trends for thermal NO, the effects of temperature on the formation of volatile NO_x are small, because of two opposing phenomena. An increase in the pyrolysis temperature will generally increase the rate of release of volatile hydrocarbons as well as of nitrogen compounds. Therefore, the local fuel fraction will increase, resulting in more fuel-rich combustion and a decrease in the conversion rate of volatile nitrogen to NO_x. The combined effect produces the relative independence of volatile NO_x on temperature.

As the fuel particles leave the flame zone, the combustion of the volatiles essentially is complete; the remainder of the combustibles, as well as fuel nitrogen, can be oxidized via surface reactions on the char material. The rate of conversion of the char's fuel nitrogen is determined by the stoichiometry at the surface and in the pores of the char. Low conversion rates of char nitrogen compounds are attributable to local oxygen depletion (and, possibly to the chemical reduction of the fuel NO_x on the char's surface).

Based on Pershing and Wendt's (1976 a,b and 1977) work with coal, there probably is a weak dependence of char NO formation on temperature. Because both volatile NO and char NO are considered to be relatively temperature-independent, fuel NO formation is considered to be relatively temperature-independent as well. This relatively independence is in contrast to the strong temperature dependence of thermal NO formation.

The magnitude of the fuel-NO_x can be expressed by both the quantity of NO_x emitted and the fraction of fuel-nitrogen converted to fuel-NO_x. It has been experimentally determined that approximately 80 percent of the total NO formed from coal combustion can be attributed to fuel-nitrogen, depending upon combustion conditions. Pershing and Wendt's (1976, b) method of determination is to compare the quantity of NO emitted for a coal/air flame as opposed to a coal/oxygen (plus non-nitrogen inert gas) flame. In this procedure, the source of nitrogen for thermal NO is eliminated, and fuel NO alone is determined. As illustrated in Figure 3-22, the bulk of the fuel NO is generated from volatile fuel components (Pershing and Wendt, 1977, b).

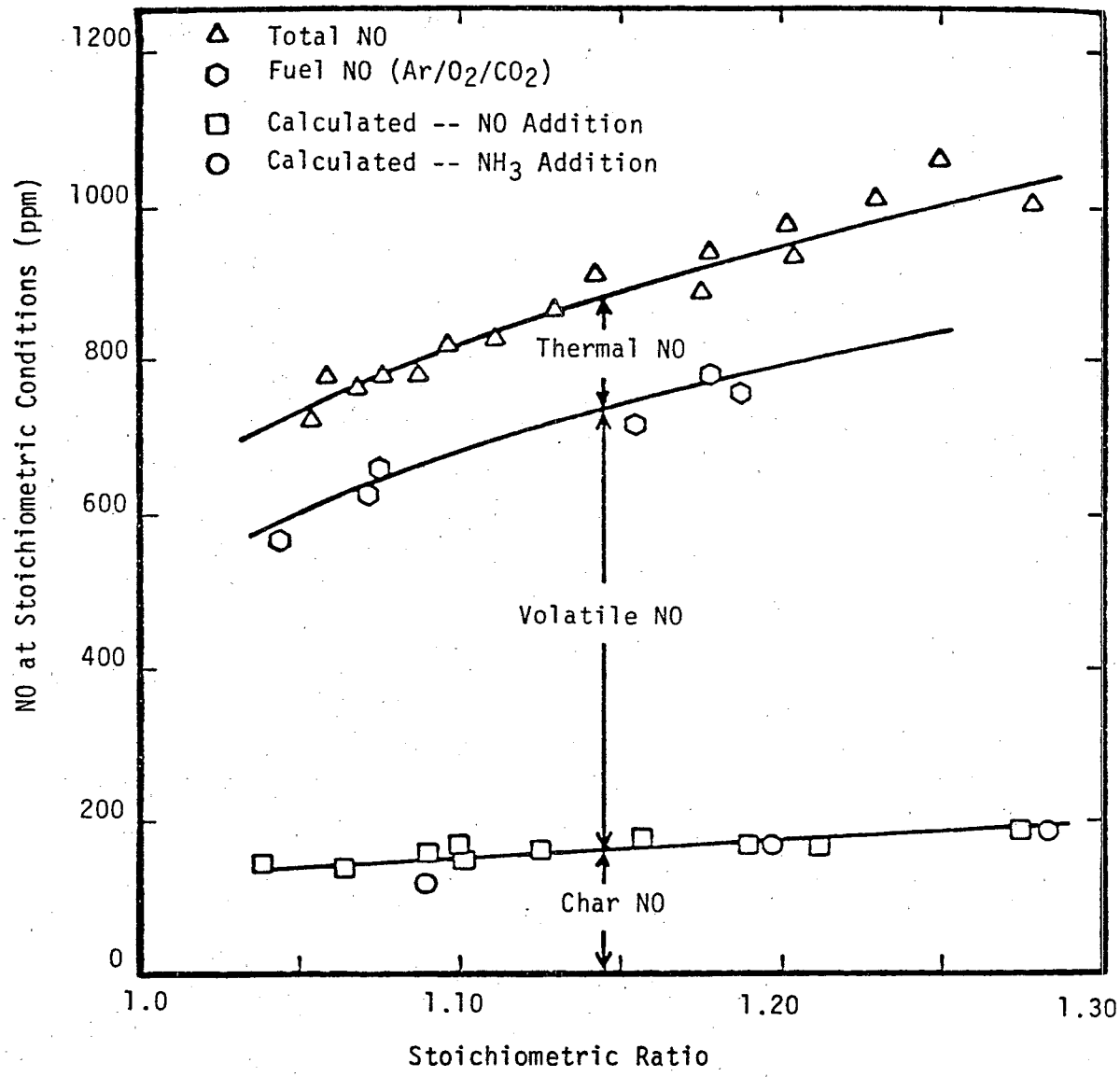


Figure 3-22. Sources of NO_x emissions (baseline conditions) using western Kentucky coal and a divergent injector with 600°F preheat (Wendt, 1980). Reprinted by permission of J.O.L. Wendt, Dept. of Chemical Eng., Univ. of Arizona, Tucson, AZ.

Pershing and Wendt (1976, a) have recorded coal-nitrogen conversions of 15 to 30 percent in tests using a variety of coals and burners. Although the conversion rate is low, the amount of fuel NO produced is high because of the large quantity of fuel nitrogen contained in coal. These experiemental findings, obtained using small-scale laboratory burners are higher than obtained in some coal-burning utility and industrial boilers.

While most NO_x reduction research has been performed on coal, the City of Los Angeles conducted some interesting experiments with dried sewage sludge [Haug, 1981]. Los Angeles plans to initiates operation of the Hyperion Energy Recovery Systems (HERS) in 1986. The plant uses a multiple effect evaporation process to produce a bone dry fuel in powder form yield approximately 3000 Cal/gm (5500 Btu/lb). This fuel has approximately 3 percent bound nitrogen. It will be burned in a substoichiometric fluidized bed combustor followed by a two stage air addition in an afterburner. In designing their combustion system, an extensive NO_x reduction program was conducted on a subscale system. Beginning with an unstaged incineration mode in the fluidized-bed, the NO_x emissions ranged from 500 to 1200 ppmv at an excess O_2 level of approximately 5 percent. After testing several staging configurations, a NO_x emission level of 50 ppmv at 5.5 percent O_2 (60 to 90 ppmv at 3 percent O_2) was achieved on a continuous basis. This is a reduction of over 90 percent compared with the incineration mode.

The low NO_x configuration consisted of substoichiometric combustion in the fluidized bed (gasification mode) followed by injection of secondary air into the freeboard of the fluidized-bed reactor and then injection of additional air at two locations in an after-burner section. A schematic of the essential elements of the system is shown in Figure 3-23.

In the evaluation of this configuration measurements were made of the NO_x and ammonia contents of the pyrolosis gas emitted from the fluidized bed. NO_x values measured by TECO chemiluminescence analyzer indicated levels of 2 to 15 ppmv. Some concern was expressed regarding accuracy of this instrument in a reducing atmosphere. Nevertheless, these data indicate that the NO_x concentration at that point was comparatively low. The ammonia measurements (by wet chemistry) at a point where the air supplied was only 60 to 63 percent of stoichiometric ranged from 1700 to 2700 ppmv. When the air supplied was 83 percent of stoichiometric, the ammonia concentration was 230 ppm.

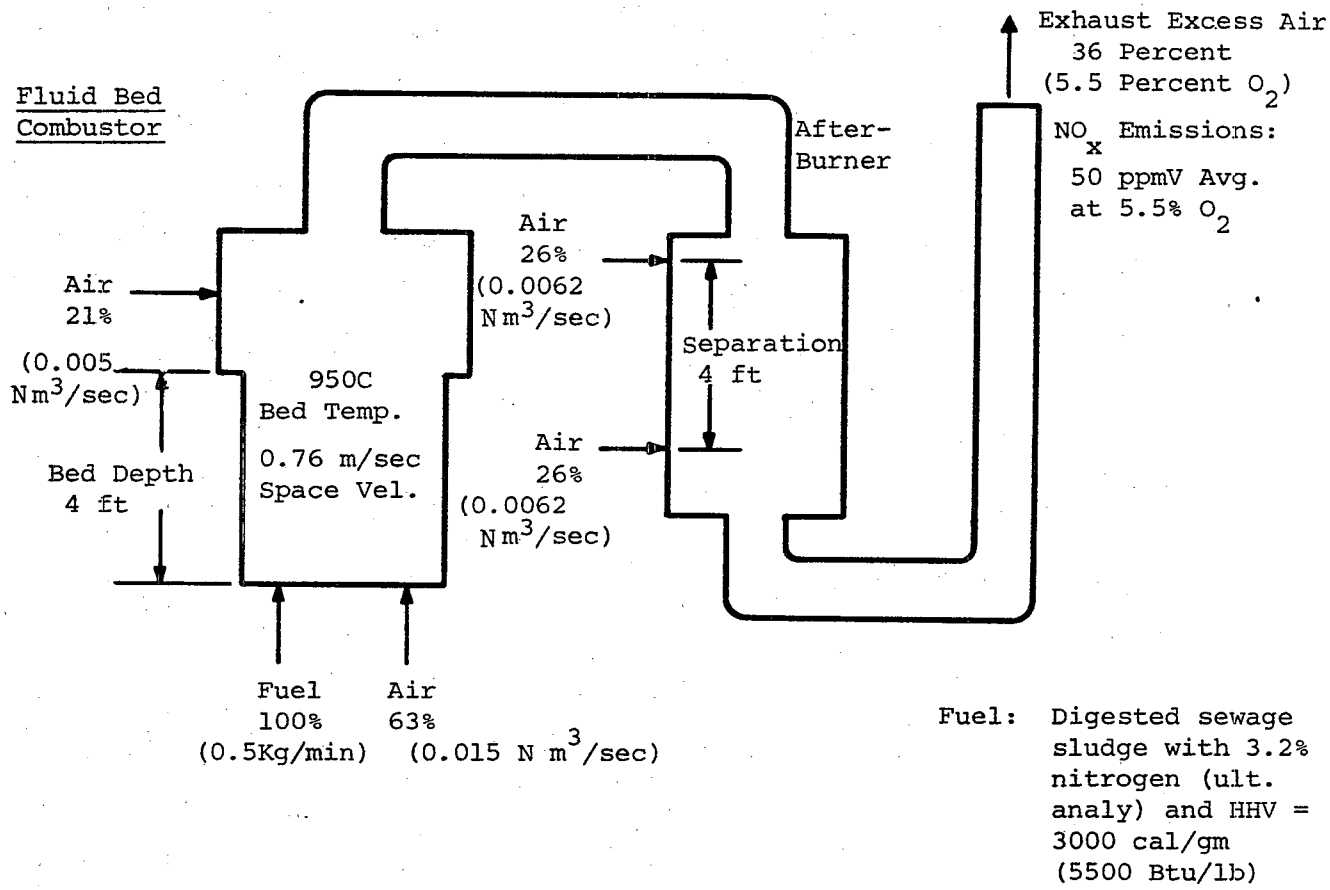


Figure 3-23. Low NO_x combustor for sewage sludge (Haug, 1981). Reprinted by permission of T. Haug, City of Los Angeles, CA.

The presence of NH_3 in the pyrolysis gas is consistent with the fact that starved air conditions in the fluidized bed should limit the availability of oxygen to oxidize fuel-bound nitrogen to nitrogen gas or NO_x. Thus, a fraction of the fuel bound nitrogen (FBN) would be expected to leave the reactor in the reduced NH_3 form. Furthermore, the concentration of NH_3 is inversely proportional to the fraction of stoichiometric air supplied. If excess air were supplied to the FBR, very little NH_3 would be expected in the exhaust gas. The concentration of NH_3 increased to several hundred ppm at 83 percent of stoichiometric air and several thousand ppm at 60 to 63 percent of stoichiometric air. It is interesting to note that if all fuel nitrogen were converted to ammonia, the resulting concentration in the pyrolysis gas would be on the order of 20,000 ppmv [Haug, 1981]. Thus, a relatively small percentage of fuel nitrogen leaves in the NH_3 form when the fluidized bed air supply is 60 percent of stoichiometric or above.

An estimated nitrogen balance about the fluidized bed when operated in a gasification mode at about 60 to 65 percent stoichiometric was made by Haug (1981). Ammonia and NO_x concentrations were assumed to be 2300 and 10 ppmv, respectively. Organic nitrogen in the pyrolysis gas was estimated from the measurement of the quantity and nitrogen content of tars and oils. This analysis indicated that approximately 10 percent of the fuel nitrogen will appear as ammonia, 0.5 percent as organic nitrogen in tar and oil, and only 0.05 percent as NO_x. The remaining 90 percent will be released as N_2. As indicated above, the increase in total air supply to 80 percent of stoichiometric reduces the ammonia to 1 percent.

An interesting result was that the expected "thermal $DeNO_x$ reaction did not occur. As discussed in Section 3.2.2., the injection of ammonia into NO_x bearing combustion gas will lower the NO_x emissions. This process is characterized by a narrow temperature window of approximately 900 to 1000 C. This would suggest that the afterburner temperature could have a critical effect on NO_x emissions since there are ammonia-like species in the pyrolosis gas. Tests were run between 760 and 920 C with an approximately constant level of 100 to 120 ppmV. At the higher temperatures, the CO and hydrocarbon burnout was improved. Raising the temperature to 1000 C caused some NO_x increase to 150 ppmV. However, no distinct window was observed.

In a series of laboratory tests where retorted Colorado oil shale was burned in both the cascading bed and fluidized bed combustor, it was found (Taylor, et al. 1985 a) that the fraction of total nitrogen which is released as NO was lower than expected for this high-nitrogen fuel. In investigating the kinetics of the NO formation, it was found that the NO is released well after the CO_2 from the oxidation of carbon. This is shown in Figure 3-24. This phenomenon indicates that it is quite possible to burn the char to CO_2 while limiting the NO emissions from the combustion of the fuel-bound nitrogen. The temperature dependence of the nitric oxide reduction by retorted oil shale is shown as an Arrhenius plot in Figure 3-25 (Taylor, et al 1985, b). The reaction for the nitric oxide reduction is:

$$\text{char} + NO \longrightarrow CO + N_2.$$

The effects of this temperature dependence are such that at 400-500 C this reaction is quite rapid and results in an effective reduction of NO in the combustor. These results are shown in Figure 3-26 (Taylor, et.al 1985 b), where it is indicated that NO concentration greater than 150 ppm would result at low combustion temperatures, but that the NO decreases to below 50 ppm at temperatures above 400 C. This test indicates that the potential for limiting the NO emissions from the retorted shale combustor exists and appears to be primarily dependent on proper design of the combustor to provide adequate staging (fuel-rich zones followed by oxygen rich zones).

The following statements summarize the staged-combustion strategy for controlling fuel NO_x from shale oil processing:

- For standard items of combustion equipment (boiler, heaters, etc.) where nitrogen bearing fuels are planned:
  - select units equipped with proven low-NO_x burner systems
  - insist on experimental evidence to substantiate manufacturer's warranted NO_x level
  - provide manufacturers with detailed fuel composition data for their analyses
  - if possible, provide sufficient quantity of the proposed fuel to allow the manufacturer(s) to test their burner systems with the proposed fuel. NO_x emissions are

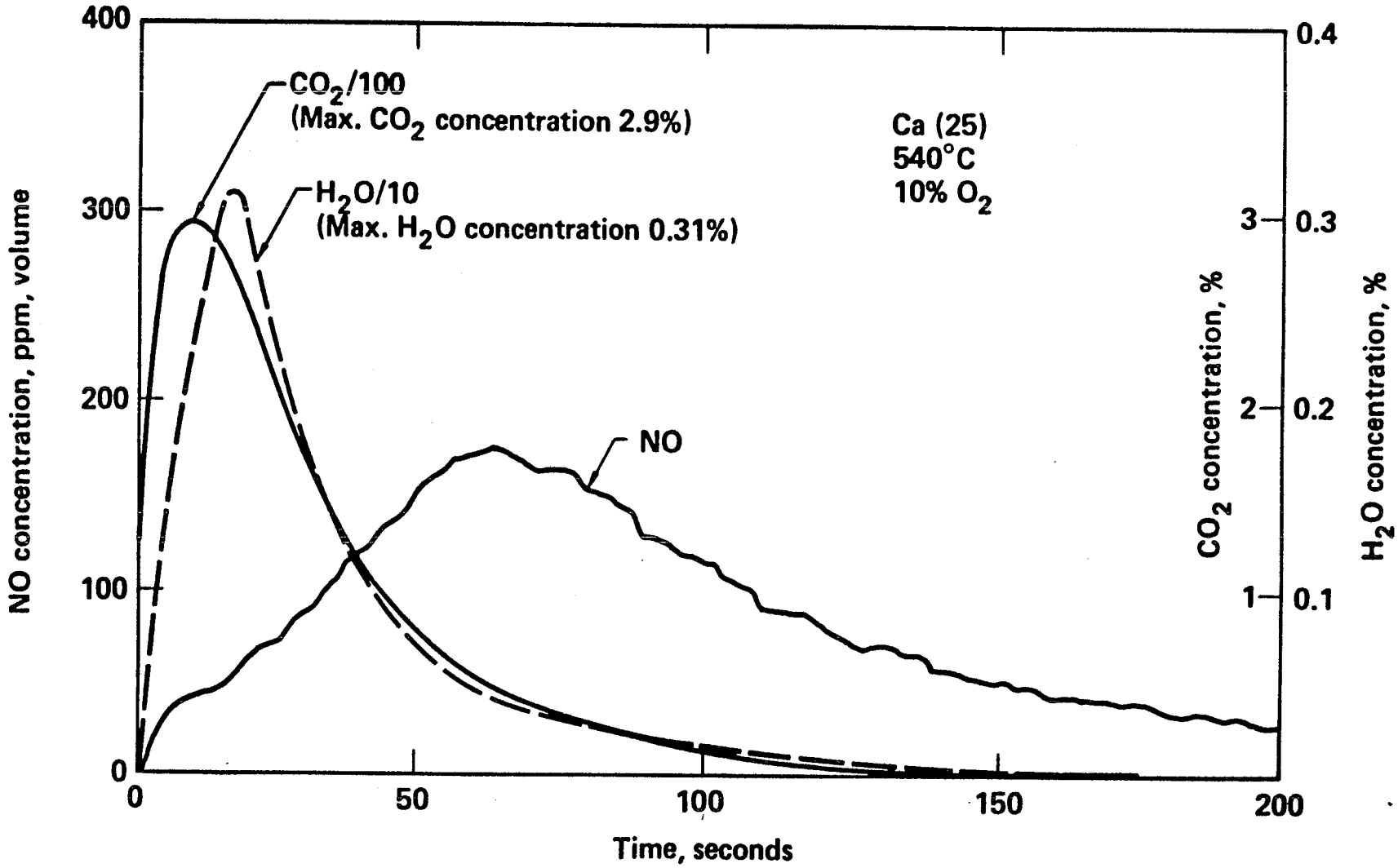


Figure 3-24. Emissions from isothermal combustion of retorted shale. (Taylor, et al., 1985b).

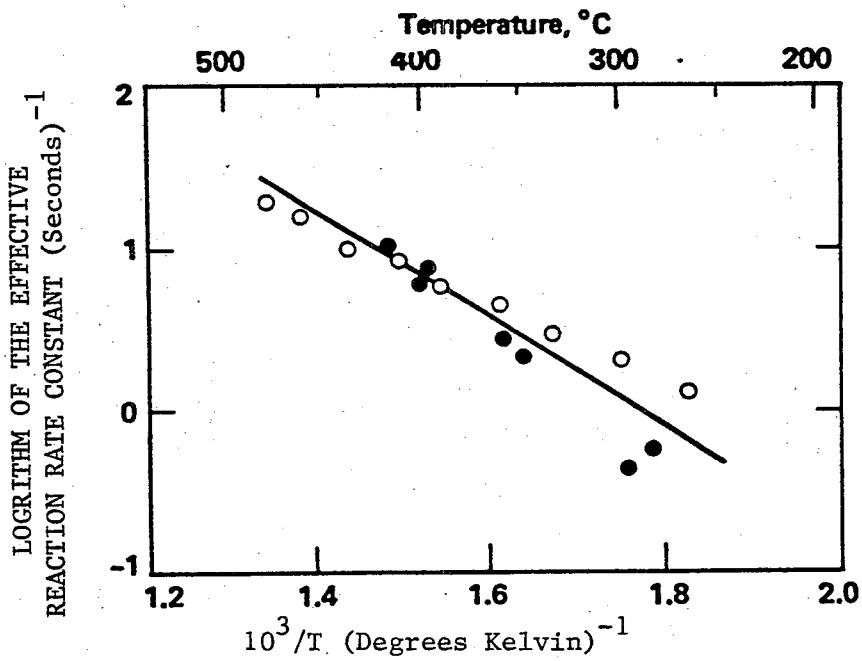


Figure 3-25. Arrhenius plot of rate data for nitric oxide reduction by retorted oil shale (Taylor, et al., 1985b).

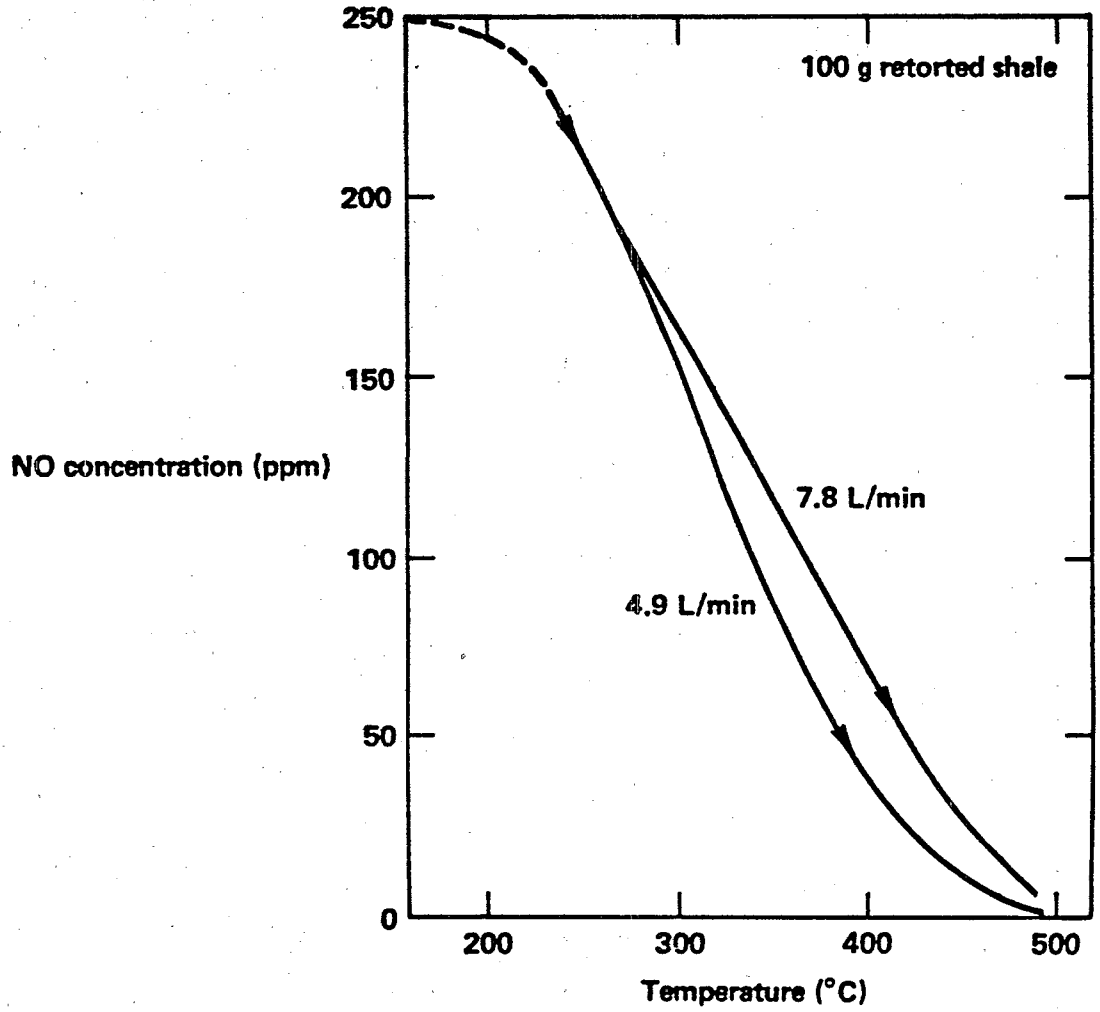


Figure 3-26. NO reduction variation with temperature (Taylor, et al., 1985b).

primarily a function of the fuel/burner system provided that the bulk gas temperature in the remainder of the system does not exceed 1400 C.

- For customized systems (shale and shale oil burner, etc), where the system is designed and developed for a specific process:

  - First, design the system to provide a substoichiometric combustion (or pyrolysis) process to release the volatile nitrogen in the fuel in an atmosphere where the released nitrogen will take the N_2 form rather than oxidize to NO_x.

  - Provide sufficient residence time for this process to take place before injecting secondary air to complete combustion.

  - When introducing the secondary air, let the bulk gas temperature increase only as high as necessary to achieve CO and hydrocarbon burnout or to meet any other process requirements. (Limit the temperature to 1400°C if possible).

  - Bulk temperature control can be achieved by heat removal or by recirculating cooled exhaust. (In systems with gases containing abrasive particles that could erode steam tubes or refractory, recirculated exhaust gas is a proven technique).

  - Subscale and pilot plant test are essential to determine the achievable levels of NO_x control. Many of the NO_x producing processes take place in localized zones. It is important to appreciate these phenomena in scaling the combustion process up or down.

  - At every level of design including the full scale unit, it is essential to provide as much leeway as possible for combustion modifications during startup. Equally important is to provide gas sampling ports throughout the unit for diagnostic purposes.

  - The achievable NO_x emissions from a procedure of this type can only be predicted in broad terms. There is no reason to expect any correlation between the level of fuel nitrogen and the NO_x emissions. The level will depend on the skill of the designer and the patience (i.e., funding available) of the owner. With this in mind, the authors feel the range of emissions that can be expected is as follows:

NO_x Emission ppmV @ 3% O_2	Probability of Achieving This Level or Lower
200	95
100	50
75	25

- The cost impact of the above procedure will be difficult to separate from the overall development costs. Since all process development goes through subscale, pilot plant, semi-works and full scale development, these NO_x control procedures are just an added concern that would be difficult to price. Undoubtedly the best investment in this regard is to secure services of combustion scientists and engineers who are most current on NO_x abatement technology especially those involved in coal combustion research.

3.2.2 Exhaust Gas Treatment

A. Ammonia Injection (Thermal $DeNO_x$)--

1. Process Description -- Ammonia has been the most widely investigated NO_x reducing agent and the process patent for ammonia injection is held by Exxon Research and Engineering as Thermal $DeNO_x$ (Patent No. 3,900,554). It involves a chain reaction mechanism between NO_x and the radicals produced by the decomposition of ammonia. The overall reaction is [KVB, Inc. 1977]:

$$NH_3 + NO + 1/4\ O_2 = N_2 + 3/2\ H_2O$$

Laboratory tests of ammonia injection using simulated flue gas and a gas/oil-fired combustion tunnel experiment are shown in Figures 3-27 and 3-28, respectively [KVB, 1977]. Dramatic NO reductions occur in the temperature range of 870 to 1040 C for NH_3 to NO molar ratios near 1.0 or greater. At temperatures below 870 C, NH_3 and NO concentrations remain near their original levels which indicate that the reaction activity level is low (Figure 3-28). After a peak reduction of NO, in the temperate range of 930 to 980 C, the

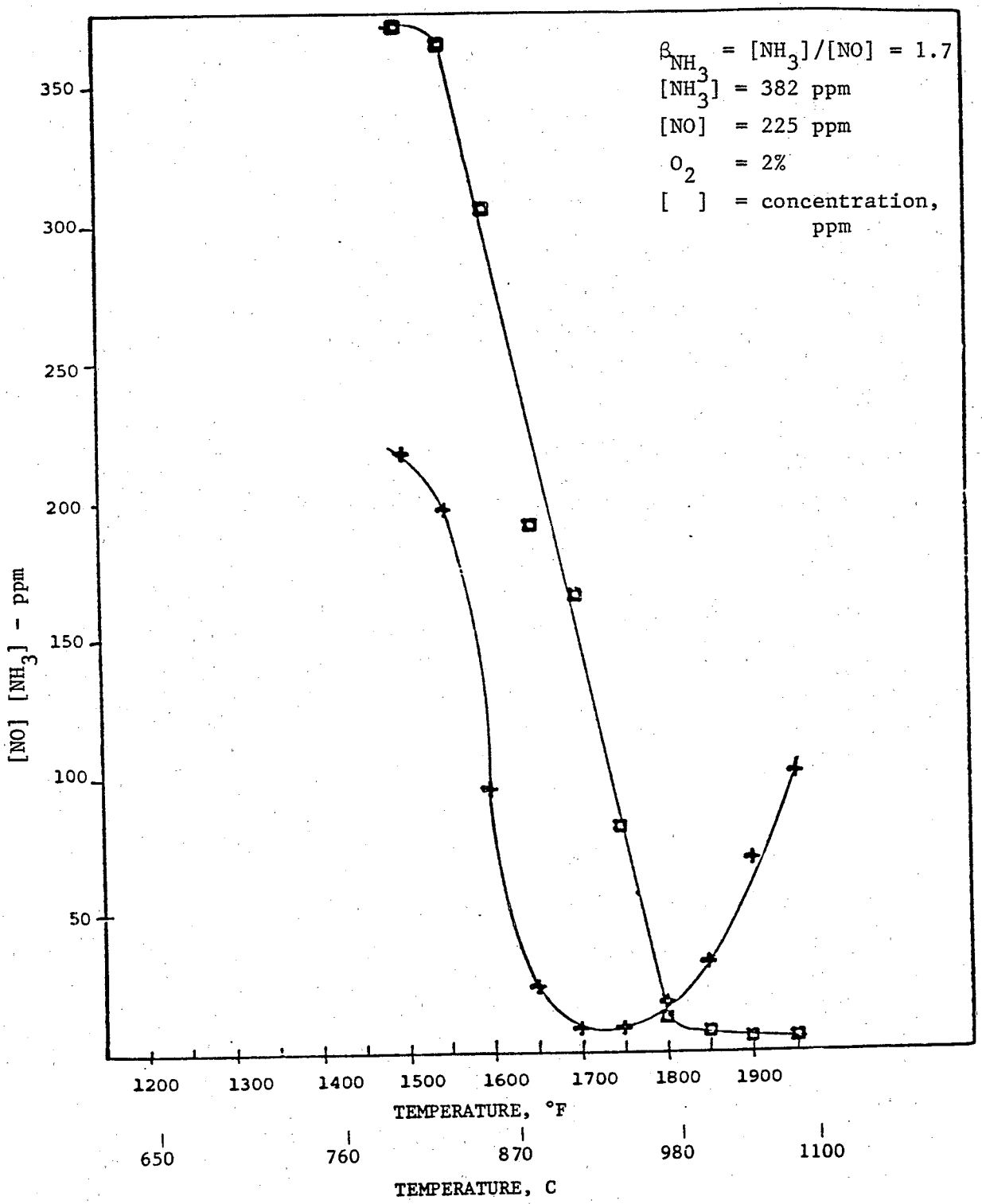


Figure 3-27. Exit NO and NH_3 concentrations for laboratory test with simulated flue gas (KVB, 1977).

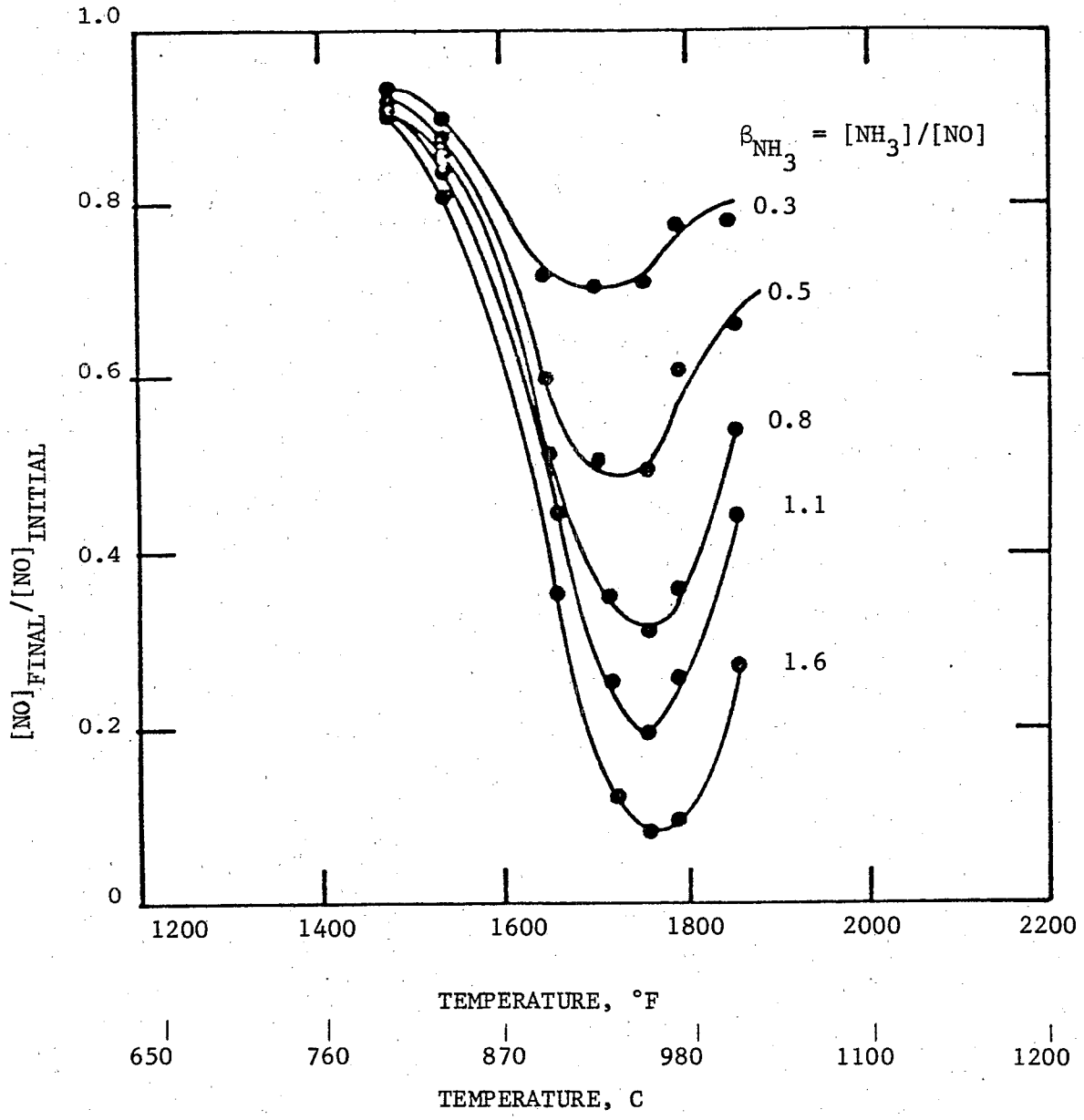


Figure 3-28. Effect of temperature on NO reductions with ammonia injection (excess oxygen 4%, initial [NO] 300 ppm) (KVB, 1977).

emission of NO begins to rise. At the same time, the ammonia carryover remains at the minimum level. A reaction of ammonia and oxygen occurs at these higher temperatures, competes with the NO and NH_3 reaction for the ammonia, and prevents effective NO reduction. The overall reaction of ammonia and oxygen is:

$$NH_3 + 5/4\ O_2 \rightarrow NO + 3/2\ H_2O$$

In extreme cases, the final NO level can be higher than the initial, due to the production of NO from the reaction of NH_3 with O_2. Because of the two competing reactions, the selective reduction of NO with ammonia is a very temperature-sensitive process.

The effect of a variation of the NH_3/NO ratio is shown in Figure 3-28. For a constant reaction temperature, the NO emission decreases as the concentration of NH_3 which is available to react with NO increases. At the optimum temperature and a NH_3/NO ratio of approximately 2, there is very little room for further improvement. For the laboratory scale data presented, the maximum reduction efficiency achieved is approximately 95 percent of the original NO concentration (Figure 3-28).

The variation in reduction with the initial NO concentration is shown in Figure 3-29 for the injection temperature of 960 C. The dependence of NO reduction efficiency upon both NO and NH_3 concentration points to the need of a well designed injection system. A well designed system would provide thorough mixing, efficient use of ammonia, and minimum ammonia carryover.

Oxygen levels between two percent and six percent were investigated by KVB (1977) and were found to have only a small effect on NO reduction. Very little effect was noted at peak reduction temperatures, but increases in O_2 level increased NO removal at lower temperatures (approximately 870 C). Increases in oxygen levels also decreased the ammonia carryover for all temperatures.

Although most of the laboratory tests have been performed with gas and oil-fired systems, a brief series of tests was performed on an 80 hp firetube boiler burning bituminous coal to evaluate the applicability of NH_3 injection on coal-fired systems. These preliminary tests allow comparison with the

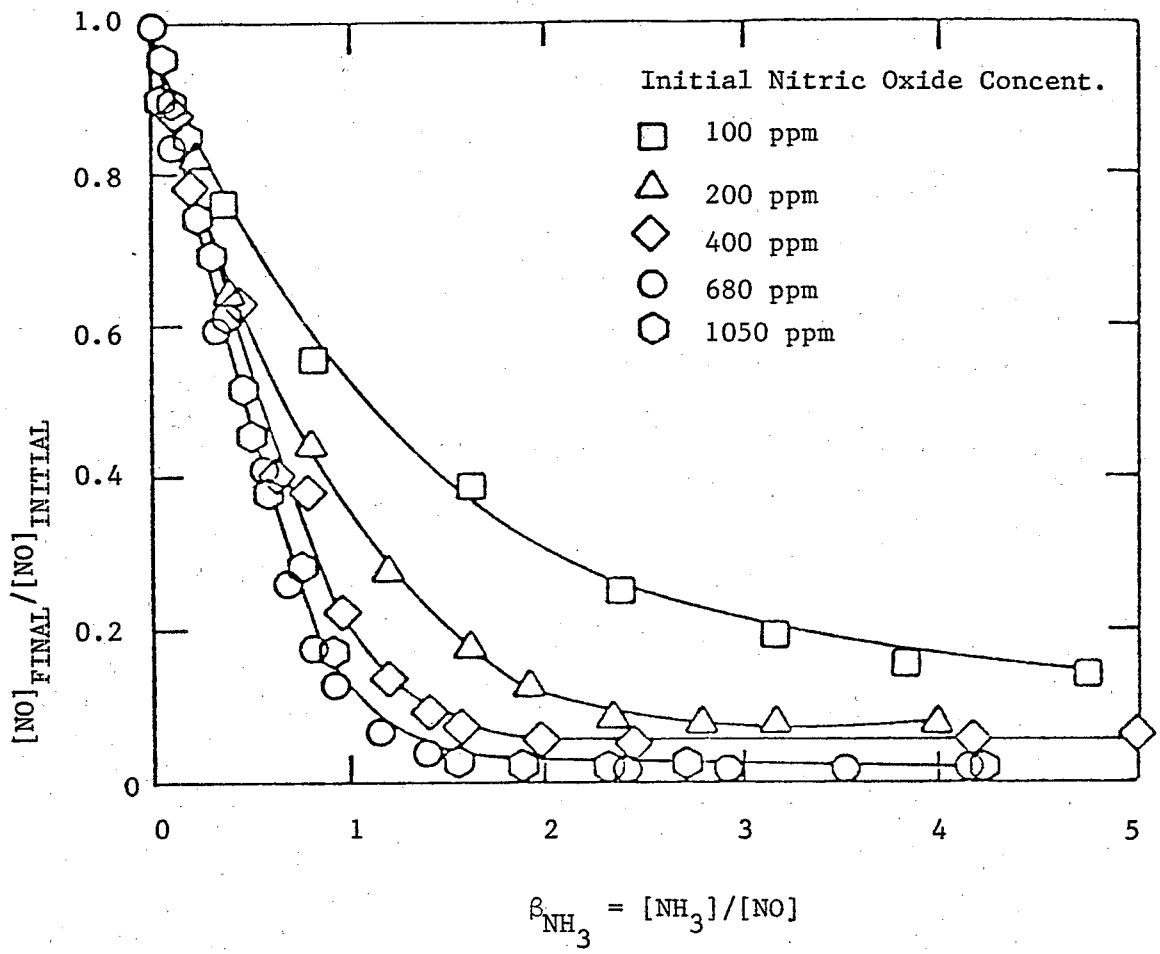


Figure 3-29. Effect of initial NO level with NH_3 injection (960 C, 2% excess oxygen) (KVB, 1977).

natural gas combustion tunnel experiments, and show a good similarity in NH_3 injection effectivenesss (Figure 3-30).

2. Application Experience

Thermal $DeNO_x$ can be a viable control measure for certain applications with removal efficiencies in the range of 50 to 70 percent. The system is readily applied to a low-sulfur low-particulate gas stream at a constant and uniform gas temperature of 950 C. Since there are few of these applications, problems arise in using the system in retrofit shifting load and sulfur bearing fuel applications. However, in the twelve years since it was patented, the process has evolved to a point where it is a highly-developed and pratical means of achieving NO_x reduction in all types of stationary firing equipment (Hurst, 1985).

The problems in the past have involved the design of an injection system to distribute the ammonia to the gas where the temperature was the critical 950 C. In constant-load systems, this was achieved by careful temperature probing and the use of an injection grid. A full scale system was installed on a 235 megawatt oil and gas fired utility boiler by the Los Angeles (California) Department of Water and Power (Dziegiel et al, 1983). The system startup was in May 1982. The initial NO_x reduction observed was 40 to 50 percent with ammonia breakthrough less than 65 ppm.

Ammonia breakthrough is a problem because of odor and the formation of ammonia bisulfate and other ammonia salts which condense in the exhaust stream causing fouling plus visibility and fine particulate emission problems.

The more recent technology involves the use of novel injector designs and the development of a kinetic model which together may substantially improve the system performance and lower costs. The kinetic model has been successful in assessing the process capability through a more precise calculation of the interaction of flue gas constituents and physical conditions including temperature, time, initial NO_x, NH_3/NO_x ratio, O_2, H_2O and free radicals. This technology has enabled extension of the temperature window to over 1200°C under certain conditions of the other kinetic parameters. The model can be used to determine the optimum amount of NH_3 which should be injected in a given location. Also the model permits accurate prediction of

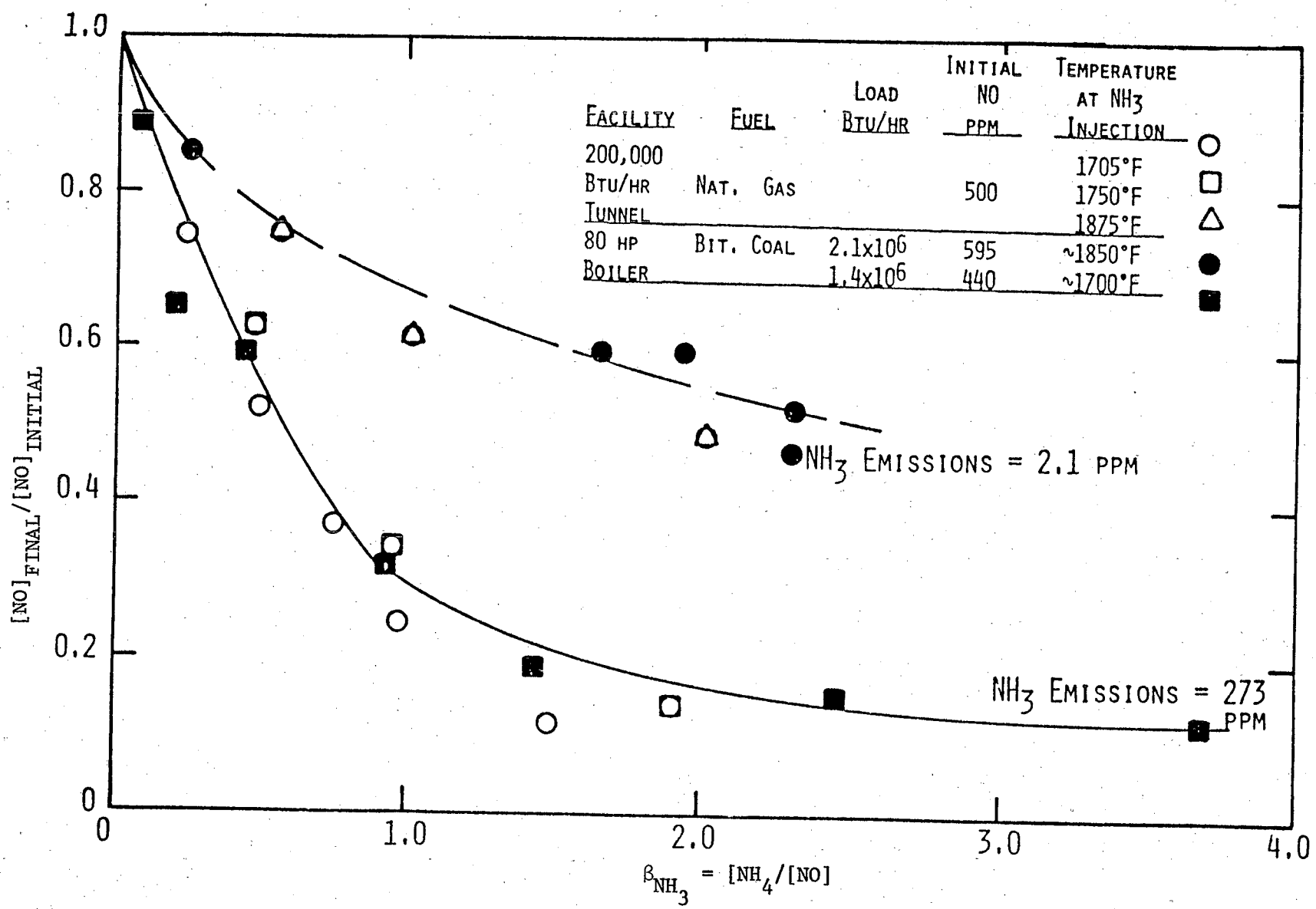


Figure 3-30. Coal firing with ammonia flue gas injection, 80 hp boiler (KVB, 1977).

the quantity of unreacted ammonia. The kinetic model calculations represent kinetic limit performance for the process and this may be lowered in a real system due to one or more of the engineering design or unit specific parameters including mixing, carrier rate, flue gas temperature and velocity gradients, and staged injection. The influence of each of these parameters has been the subject of considerable research and development and some work in this area is being continued. Mixing effectiveness of grid injectors has been increased such that it is possible to achieve 95 to 100 percent of the kinetic limit performance. Wall injector mixing can now be optimized for a specific geometry by utilizing a field-validated three-dimensional flow modeling technique (Hurst, 1985).

The field testing of wall injectors was performed on a 100 x 10^6 kcal/hr gas and oil fired boiler in Kawasaki, Japan. In this boiler, it was desirable to optimize performance in the range of 30 to 50 percent $DeNO_x$ since it is unnecessary to achieve high NO_x reduction except during pollution emergencies. The modeling results were verified by field testing with a maximum variation between calculation and data of ±7 percent. Thus, in the 80 to 100 percent load range, this boiler may operate with a single zone of injectors making adjustments to carrier and NH_3 rate as required to meet emission targets (Hurst, 1985).

The subject of ammonia breakthrough with the Thermal $DeNO_x$ process is extremely complicated because it depends upon the interaction of numerous factors such as flue gas concentrations of NO_x, H_2O, O_2, SO_3, HCl, particulates, the time-temperature relationship of the flue gas, the effectiveness of mixing NH_3 with flue gas, high temperature materials which the flue gas contacts within and downstream of the injection zone, the type of flue gas heat recovery equipment, and the type of flue gas scrubbing equipment. Because of its complexity, ammonia breakthrough must be evaluated for a particular firing system, and there are very few generalizations which can be made. Since breakthrough is linked to a certain degree to $DeNO_x$ performance, fired equipment in which the time-temperature relationship is favorable to achieving high $DeNO_x$ performance will also typically exhibit low NH_3 breakthrough. In any case, the placement of the injectors for a given time-temperature relationship and their mixing effectiveness are of primary importance in

minimizing NH_3 breakthrough, and this is within the control of the $DeNO_x$ system design. In some cases, factors not under the control of the $DeNO_x$ system design combine to make high removal efficiency with low NH_3 breakthrough incompatible.

Laboratory experiments have shown that in the lower temperature flue gas, SO_3 combines with H_2O to produce sulfuric acid starting at a flue gas temperature as high as 430 C. Above 430 C the sulfur is present as SO_2 only. When the flue gas temperature drops to the region of approximately 320 C, ammonia combines with sulfuric acid to form ammonium bisulfate in the reaction:

$$NH_3 + H_2SO_4 \rightleftarrows NH_4\ HSO_4$$

As flue gas temperature is further reduced, the ammonia bisulfate reacts with additional ammonia to form ammonium sulfate in the reaction:

$$NH_3 + NH_4\ HSO_4 \rightleftarrows (NH_4)_2\ SO_4$$

Equlibrium data for each of these reactions are plotted in Figure 3-31. The following conclusions can be drawn from these curves (Hurst, 1985).

- First, the curves intersect at high NH_3 concentrations, one would conclude that at high NH_3 concentrations, in the flue gas relative to SO_3, the problem of bisulfate formation can be significantly curtailed.

- Second, as the NH_3 concentration is reduced, the curves diverge and thus show that the bisulfate compound exists over a wider temperature difference before moving to the sulfate. Therefore, lower breakthrough in the case of NH_3 injection processes is actually more detrimental to fouling/corrosion from bisulfate except in the limit when very little or no NH_3 breakthrough occurs. In such case, the amount of deposit is curtailed simply by the unavailability of NH_3.

- Third, the initial formation temperature of bisulfate increases with increases of NH_3, SO_3 and H_2O concentrations. This is consistent with previous data. However, the equilibrium temperature between the sulfate and bisulfate also increases with increases in these constituent concentrations, but at a faster rate.

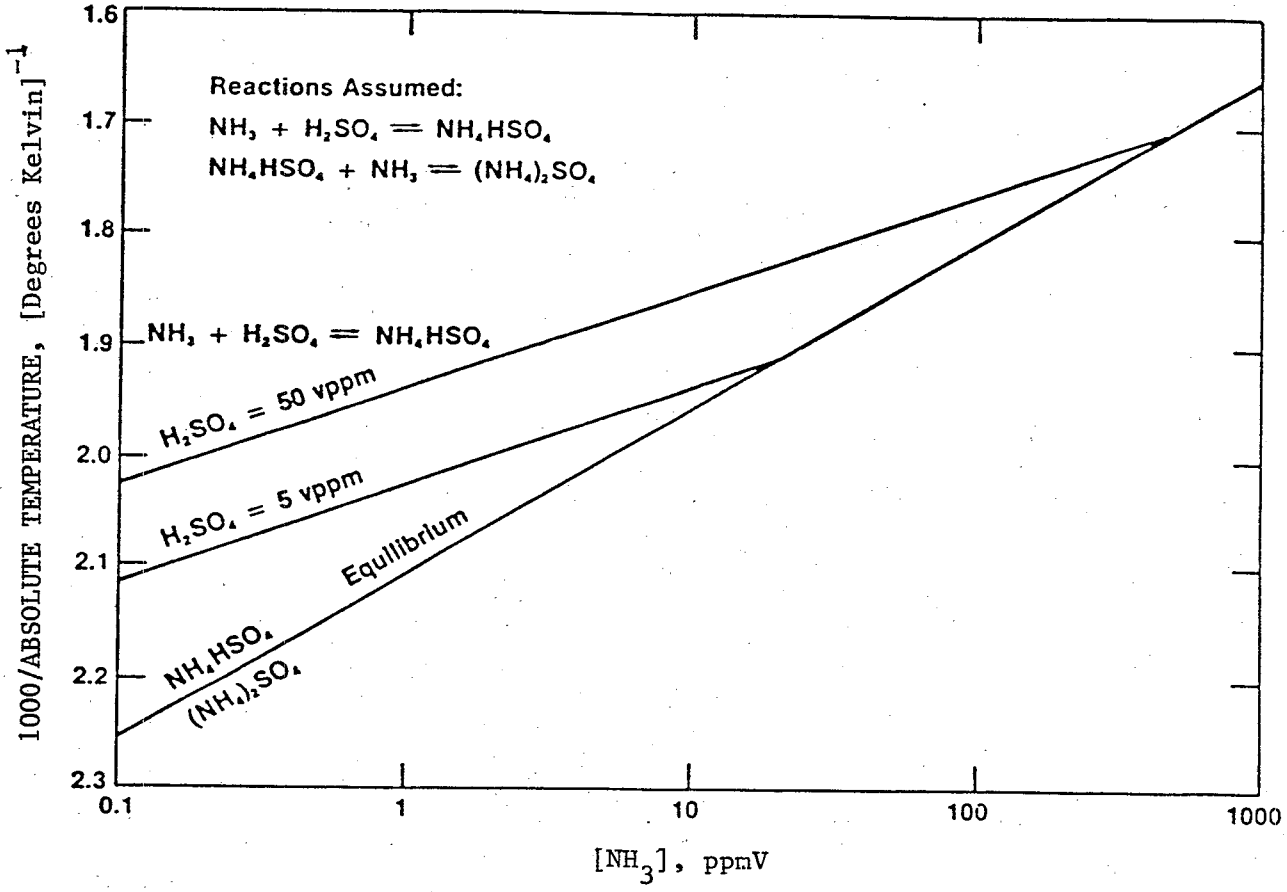


Figure 3-31. Ammonium bisulfate/sulfate equilibrium with field operating conditions overplotted (Hurst, 1985).

In most cases, severe fouling clearly occurs in a region where significant bisulfate formation would be expected as a result of the concentrations of NH_3 and SO_3 and the flue gas temperature. Low fouling cases generally satisfy one or more of the following four criteria:

1. Excess NH_3 so that NH_4HSO_4 formation is curtailed at any temperature, i.e., operating condition falls below the equilibrium curve between bisulfate and sulfate.

2. Temperature at the outlet of the air heater is at or above temperature where significant NH_4HSO_4 is formed.

3. Temperature at the point of NH_3 injection is lower than formation temperature of $(NH_4)_2SO_4$.

4. Excess NH_3 is very low so that deposit amount is insufficient to cause a fouling problem.

Use of the data from Figure 3-31 and the low fouling assessment criteria permits the accurate design of features in the boiler and NH_3 injection system to eliminate or minimize the problem of sulfate fouling air emissions.

3. DeNOx Costs - Based on statements by Exxon (Hurst, 1985), the latest version of thermal $DeNO_x$ technology requires much lower investment than the older technology and is substantially less expensive than catalyzed ammonia injection (SCR) processes. At the same plant in Kawasaki, Japan; a boiler was equipped with Thermal $DeNO_x$, the performance of which was discussed above, and a fired heater of similar heat input was equipped with SCR. The cost of the SCR facility was approximately $2M and the cost of the Thermal $DeNO_x$ facility was approximately $0.4M or 20 percent of the SCR cost.

Cost information for the improved Thermal $DeNO_x$ process applied to utility boilers is summarized in Tables 3-21 and 3-22 and compared to previous estimates using the older technology. In Table 3-21 note that investment costs for the improved technology are only 45 to 55 percent of the older technology. The annual revenue requirements have been calculated in accordance with the standard U.S. utility procedures. The lower costs here reflect lower NH_3 consumption and a much lower carrier air requirement. In addition, lower capital investment results in lower indirect costs.

TABLE 3-21. TOTAL CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS FOR "IMPROVED" THERMAL $DeNO_x$ APPLIED TO COAL-FIRED BOILERS (Hurst, 1985)

	Investment		Annual Requirements	
	k$	$/kW	k$	Mills/kWh
Older Technology				
500 MW NOi = 300 ppmv	11792	23.58	4577	1.31
500 MW NOi = 600 ppmv	17157	34.31	8125	2.32
Improved Technology				
500 MW NOi = 300 ppmv	6526	13.05	3159	0.90
500 MW NOi = 600 ppmv	7755	15.51	4712	1.35

NOi = Initial NO_x

TABLE 3-22. COST EFFECTIVENESS FOR "IMPROVED" THERMAL $DeNO_x$ APPLIED TO COAL-FIRED BOILERS (Hurst, 1985)

	Older Technology	
	Mills/kWh	$/lb NO_x Removed
500 MW NOi = 300 ppmv	3.07	0.97
500 MW NOi = 600 ppmv	5.26	0.83
	Improved Technology	
500 MW NOi = 300 ppmv	2.03	0.57
500 MW NOi = 600 ppmv	2.94	0.41

NOi = Initial NO_x

Cost effectiveness results shown in Table 3-22 have been provided on two bases, mills/kWh and \$/lb NO_x removed. The removal efficiency is 70 percent for the "older" technology and 80 percent for the "improved" technology.

The NO_x emission levels of 300 and 600 ppmv (0.4 and 0.8 lb/MBtu) used in Tables 3-21 and 3-22 span the range of most coal-fired boilers using other than the recently developed low-NO_x burners. Thermal $DeNO_x$ is more cost effective (\$/lb NO_x removed) on boilers which have higher initial NO_x levels.

The investment costs developed for the "older" technology compare favorably with 1982 cost estimates prepared by the Electric Power Research Institute which are also based on "older" technology. These estimates are presented in Table 3-23 along with estimates from the same reference for two SCR processes. The two SCR processes include the Kawasaki Heavy Industries (KHI) process and the Hitachi-Zosen process. The KHI process is the least costly of the two, but is still approximately three times the cost of Thermal $DeNO_x$. If the published annual operating cost for the KHI process are escalated to a 500 MW plant, this amounts to \$15M for NO_x = 300 ppmv and \$19.3M for NO_x = 600 ppmv at 90 percent NO_x reduction. Using the investment and operating cost data for KHI, it is possible to calculate the cost effectiveness. The results of such calculations are presented in Table 3-24. A comparison between these cost effectiveness data and those for the "improved" Thermal $DeNO_x$ process reveals that Thermal $DeNO_x$ is approximately four to five times more cost effective than SCR.

B. Selective Catalytic Reduction (SCR)--

1. Process Description -- The SCR concept is similar to Thermal $DeNO_x$ but the critical temperature window is eliminated, the process can take place at a lower temperature and therefore is less affected by load changes. Offsetting these advantages is the cost of a catalyst system which is subject to trace metals contamination which can shorten its lifetime and add significantly to system cost. This can pose a serious limitation in shale oil combustion systems where trace metal content is high.

In this process, gaseous ammonia diluted through either air (i.e., below the flammability limit) or steam is injected, through a grid system, into the flue gas stream at the exit of the boiler or other combustor. The

TABLE 3-23. TOTAL CAPITAL INVESTMENT FOR THERMAL $DeNO_x$ AND SELECTIVE CATALYTIC REDUCTION FOR A COAL-FIRED 500 MW BOILER (Hurst, 1985)

	Investment $/kW	
	NOi = 300 ppmv	NOi = 600 ppmv
Thermal $DeNO_x$	19	25
Kawasaki, Heavy Ind. (SCR)	69	83
Hitachi-Zosen (SCR)	101	109

NOi = Initial NO_x

TABLE 3-24. COST EFFECTIVENESS COMPARISON OF IMPROVED THERMAL $DeNO_x$ VS KAWASAKI INDUSTRIES SCR FOR A COAL-FIRED 500 MW BOILER (Hurst, 1985)

	Cost Effectiveness			
	NOi = 300 ppmv		NOi = 600 ppmv	
	Mills/kWh	$/lb NO_x Removed	Mills/kWh	$/lb NO_x Removed
Thermal $DeNO_x$	2.03	0.57	2.94	0.41
KHI (SCR)	9.86	2.43	12.53	1.54

NOi = Initial NO_x

ammonia/flue gas mixture then travels to catalytic reactors where the ammonia and NO_x interact resulting in a substantial NO_x reduction at the exit of the reactor. The primary reactions are:

$$NO + NH_3 + 1/4\ O_2 \rightarrow N_2 + 1\ 1/2\ H_2O$$

$$NO_2 + 2NH_3 + 1/2\ O_2 \rightarrow 1\ 1/2\ N_2 + 3H_2O$$

As can be seen in the above equations, the stoichiometric NH_3:NO_x mole ratio is approximately 1:1 considering that NO_x at the boiler outlet is generally assumed to be 95 percent NO and 5 percent NO_2. In actual practice this ratio may be 0.9 to 1.0 for a 90 percent removal efficiency based on the inlet NO_x concentration, assuming that a suitable catalyst volume has been provided to achieve this efficiency. If adequate catalyst volumes are not provided to achieve high efficiencies, the result of increasing the NH_3:NO_x mole ratio will be ammonia slip or breakthrough which is simply the flow of excess ammonia past the reactor(s) and on to the downstream equipment.

The ideal process temperature range is generally limited to 310 C to 400 C. Below the lower temperature there is danger of the formation of ammonium sulfates and bisulfates which can attack the catalyst surface. If the low temperature operation is continuous (i.e., 12 hours or more), this attack can permanently damage the catalyst. An economizer flue gas bypass is typically provided to raise the economizer exit flue gas temperate at low loads to an acceptable level and thereby overcome this problem in boiler applications. The economizer outlet temperature rarely exceeds 400 C at full load conditions. However, if exceeded, the NH_3 may disassociate and NO_x reduction process will be impaired; if temperatures exceed 500 C to 550 C the catalyst may sinter and be permanently damaged.

The NO_x reduction efficiency of this process is dependent upon the space velocity selected for the catalyst. The space velocity is basically the total flue gas volume per hour divided by the bulk volume of the catalyst, noting that the flue gas and catalyst volumes are generally stated in standard cubic feet or normal cubic meters for this purpose. The units of space velocity are reciprocal hours. In general, reasonably low space velocities

and/or area velocities are required for high process efficiencies. This means that increased catalyst quantities are required for high NO_x reduction efficiencies (e.g., 90 percent). It is important to note that increased quantities of catalyst are also required to control ammonia breakthrough. That is, not only must a sufficient quantity of catalyst be provided to achieve the required NO_x reduction efficiency, but additional catalyst is generally required to insure the oxidation of excess ammonia used in the process to prevent excessive ammonia slip (i.e., typically less than 10 ppm). The specific relationship of these indices and NO_x reduction efficiency is related to the proprietary catalyst used by the various SCR manufacturers.

The catalyst used in these processes is either solid or surface coated catalyst on a substrate. Both are subject to erosion damage from fly ash in coal fired power plant flue gases, and the efficiency of the catalyst can be impacted over the long term (1 to 3 years). The loss of catalyst effectiveness will probably require the use of an excess of catalyst to maintain the required process efficiency over an extended period of time. If this is the case, the capital and operating costs for the SCR process will be increased accordingly.

These catalysts also have the tendency to oxidize SO_2 to SO_3 which tends to react with the NH_3 carryover and the form ammonium bisulfate at the exit of the reactor in the reaction:

$$NH_3 + SO_3 + H_2O \rightarrow NH_4HSO_4$$

As mentioned in the Thermal $DeNO_x$ discussions, ammonium bisulfate has the tendency to condense in the intermediate baskets of a standard regenerative air preheater. This makes their removal by conventional soot blowing difficult. The result, particularly on high sulfur fuels, can be significant fouling of these preheaters which will cause high draft losses and possibly unit load limitations. Frequent air heater water washings may be required. Japanese experience with SCR systems has led to the development of combined intermediate/cold end air heater baskets which have a heat transfer surface configuration which is conducive to cleaning by soot blowing. This combined

section is longer than its conventional counterparts (the separate intermediate and cold end baskets in a standard regenerative air heater). The result is more expensive air heaters which may reduce the ammonium bisulfate fouling. However, on high sulfur fuels, the effectiveness of this accommodation is still questionable.

2. System Experience -- A number of gas and oil fired boilers in Japan have been equipped with SCR. In the U.S., by 1983, only the Southern California Edison Plant at Huntington Beach, CA had been equipped with a system. The full scale Kawasaki SCR system was attached to one of two exhaust ducts from the 215 MW Bacock & Wilcox unit. It was designed for a removal efficiency of 90 percent and a maximum ammonia slip of 10 ppmv and a pressure drop of 30 cm water column. After more than two years of testing, operating results have shown that the desired NO_x removal and pressure drop performances were obtained. The ammonia breakthrough constraint has not been satisfied but there were uncertainties in the measurements (Kerry and Weir, 1985). Operation of the system showed that the catalytic reaction of ammonia plus NO_x behaved as a first order reaction with removal dependent on ammonia addition. Thus, NO_x removal was controlled by ammonia addition. As ammonia addition was increased to increase removal, ammonia slip increased. Operation also showed that ammonia was adsorbed by the catalyst. The amount appeared to be a function of the ammonia flow rate. The unit has been run 70 percent of the time on gas and 30 percent on 0.25 percent sulfur oil. After 11,000 hours of activity, only a 5 percent loss of catalyst activity was reported.

The overall SCR operational experience on coal fired utility boilers is minimal. In Japan, Takahara Unit 1 (250 MW) is now testing the SCR process on coal. In the U.S., three coal fired pilot plants are in operation, these are:

- The EPRI Arapahoe Test Facility at the Arapahoe Power Plant of the Public Service Company of Colorado.
  Size - 5000 scfm (2.5 MW)
  Supplier - Kawasaki Heavy Industries

- The U.S. EPA Pilot Plant at the Mitchell Station, Georgia Power Company
  Size - 0.5 MW
  Supplier - Hitachi Ship Building and Engineering Company

- The U.S. EPA Pilot Plant at the Big Bend Station of the Tampa Electric Company
  Size - 0.6 MW (Combined SO_x/NO_x Removal)
  Supplier - UOP Inc.

3. SCR Costs -- A comprehensive study of SCR economics for coal fired utility boilers was made by Damon et al. (1985). The results of three other studies conducted for the EPA, EPRI and TVA were analyzed and placed on an identical basis for a 500 MW net capacity. The results are shown in Table 3-25 (Damon et al., 1985).

TABLE 3-25. SCR COSTS (Damon, et al., 1985)†

Report	$/Net/kW	$/ACFM	$/Nm^3/min	First Year Mills/kWhr*	Levelized Mills/kWhr*
EPRI/Stearns-Rogers	94	15.93	560	6.87	12.59
EPRI/U.S. EPA/TVA	79	15.97	560	6.14	11.44
U.S. EPA/TVA	65	13.21	410	5.11	9.53

* Based on a one year catalyst guarantee life
† Costs are in 1983 dollars.

As can be seen, the capital costs developed for the EPRI reports on a $/kW (net) basis are somewhat different. The reason for this variation is the difference (approximately 19 percent) in the adjusted flue gas flows used for these reports. However, on $/Nm^3/min basis, the two EPRI reports show identical costs. The two TVA reports use nearly identical flue gas flows but have significantly different capital costs. This results in the significantly reduced catalyst requirements assumed in the U.S. EPA/TVA report. All three of these studies used Hitachi-Zosen's catalyst NOXNON 500 Type 3 in their cost analyses. The lower costs of this latter report could only be accepted when this reduced catalyst loading is demonstrated to be viable in achieving a 90 percent NO_x reduction efficiency while at the same time controlling NH_3 slip to 10 ppm or less.

The Southern California Edison plant described above shows a similar result of 11.3 mils/kWhr (Kerry and Wier, 1985) at 100 percent capacity. In

terms of cost effectiveness, \$/ton NO_x (as NO_2) removed, the cost ranges from \$16,700/ton at a 30 percent capacity factor to \$6,400/ton at 100 percent capacity factor.

3.3 SULFUR COMPOUND CONTROL

Sulfur from shale oil processing operations is emitted in three basic forms:

- Inorganic reduced - H_2S
- Inorganic oxidized - SO_2/SO_3 (SO_x)
- Organic reduced - CS_2, COS, MeSH, thiophenes, etc.

The principal source of reduced sulfur emissions is the retort off-gas in which the organic fraction of the total sulfur content ranges from one to fifteen percent as discussed in Section 2.0. Most sulfur removal systems are unable to adequately scrub the organic sulfur compounds which are insoluble in water. This is a serious concern which has not as yet been adequately addressed by project proponents. The activated carbon-hypochlorite process, as presented in Section 3.3.5, appears to provide for the removal of organic sulfur. Claims of organic sulfur removal have also been made for the Lo-Cat process, but there is minimal experience in long term application of either of these technologies on removal of organic sulfur species.

The oxidized sulfur is emitted by primarily combustion processes including incinerators of the retort tailgas. Using conventional flue gas desulfurization (FGD) on the incinerator off-gas is another way to ensure the removal of the organic sulfur.

The EPA has detailed many sulfur removal processes in their Control Technology Appendices for Pollution Control Technical Manuals (Ondich, 1983). In this section, those processes that are relevant to shale oil processes are presented. Since those technologies are included in the EPA Appendices, they will be summarized with any pertinent comments regarding applicability. Processes not appearing in the EPA Appendices will be presented in greater detail, where information is available.

3.3.1 Stretford Process (H_2S)

A. Process Description--

The process chemistry of the Stretford technology is based on the absorption of H_2S in an alkaline scrubbing solution and subsequent liquid-phase oxidation of the captured H_2S to elemental sulfur. The Stretford liquor is a dilute solution of sodium carbonate (Na_2CO_3), sodium metavanadate ($NaVO_3$), and sodium salts of the 2:6 and 2:7 isomers of anthraquinone disulfonic acid (ADA). The solution is maintained at a pH range of 8.5 to 9.5 and at a temperature of approximately 43 C. Citric acid is used as a complexing agent to keep quadravalent vanadium dissolved in the alkaline solution. Anti-foaming agents are used when required usually as the result of condensed hydrocarbon in the solution.

The process chemistry is complex and poorly understood. However, in a simplified manner, the process of removing H_2S and converting it to elemental sulfur is a six step process as follows:

1. The H_2S is absorbed in the alkaline Stretford solution in a suitable gas/liquid contactor.

2. The H_2S reacts with the sodium carbonate to form sodium hydrosulfide and sodium bicarbonate:

$$H_2S + Na_2CO_3 + NaHS + NaHCO_3 \qquad \text{(R-11)}$$

3. The hydrosulfide reacts with sodium metavanadate to form elemental sulfur, a quadravalent vanadium salt, and sodium hydroxide:

$$2NaHS + 4NaVO_3 + H_2O \longrightarrow Na_2V_4O_9 + 2S + 4NaOH \qquad \text{(R-12)}$$

4. The quadravalent vanadium salt reacts with ADA to regenerate the sodium metavanadate:

$$Na_2V_4O_9 + 2NaOH + H_2O + 2ADA \longrightarrow 4NaVO_3 + 2ADA \cdot 2H \qquad \text{(R-13)}$$

5. The sodium hydroxide and sodium bicarbonate reaction products further react to form sodium carbonate:

$$NaOH + NaHCO_3 \longrightarrow Na_2CO_3 + H_2O \qquad \text{(R-14)}$$

6. The reduced ADA reacts with oxygen to regenerate the ADA:

$$2ADA \cdot 2H + O_2 \longrightarrow 2ADA + 2H_2O \qquad \text{(R-15)}$$

The overall process reaction can be written as the oxidation of H_2S to elemental sulfur:

$$2H_2S + O_2 \rightarrow 2S + 2H_2O \qquad \text{(R-16)}$$

Several side reactions that form nonregenerable compounds are possible in the Stretford process. These consist primarily of oxidized sulfur compounds such as sodium sulfate and sodium thiosulfate. The nonregenerable compounds can build up in the system and eventually impede the performance of the process by interferring with the principal chemical reactions. These compounds must be removed from the process either by purging them from the system or by recovering them in a regeneration system.

The oxidized sulfur compounds form when the dissolved oxygen in the process liquor is too high, which occurs when the pH is too low. H_2S absorption also is reduced at low pH. The high CO_2 found in retort offgas can reduce the pH unless the process pH level is maintained by sodium bicarbonate addition. A simplified process flow diagram of an EPA pilot plant is presented in Figure 3-32. Reactions (13) through (15) take place in the reaction tank. The air which provides the oxygen for Reaction (15) also causes the elemental sulfur to float as a heavy froth on the liquid surface.

The sulfur froth overflows from the oxider to a froth tank while the sulfur-free reactivated Stretford solution flows through a cooling tower before falling into a surge tank and being recirculated to the absorber. The cooling tower is designed to evaporate sufficient water to maintain the system in water balance and to control the solution temperature which is normally kept at 305 to 314 K. A solution heater (not shown in Figure 3-32) usually is provided in case additional heat is required to promote evaporation of water to maintain the water balance.

The sulfur can be recovered from the froth in several ways, depending on the desired end product, total sulfur production, and utilities cost. For small sulfur capacities, simple filtration of the sulfur froth may be economic. For larger sulfur production rates, one or more stages of centrifuging (or filtering) followed by heating may be used. In this mode of

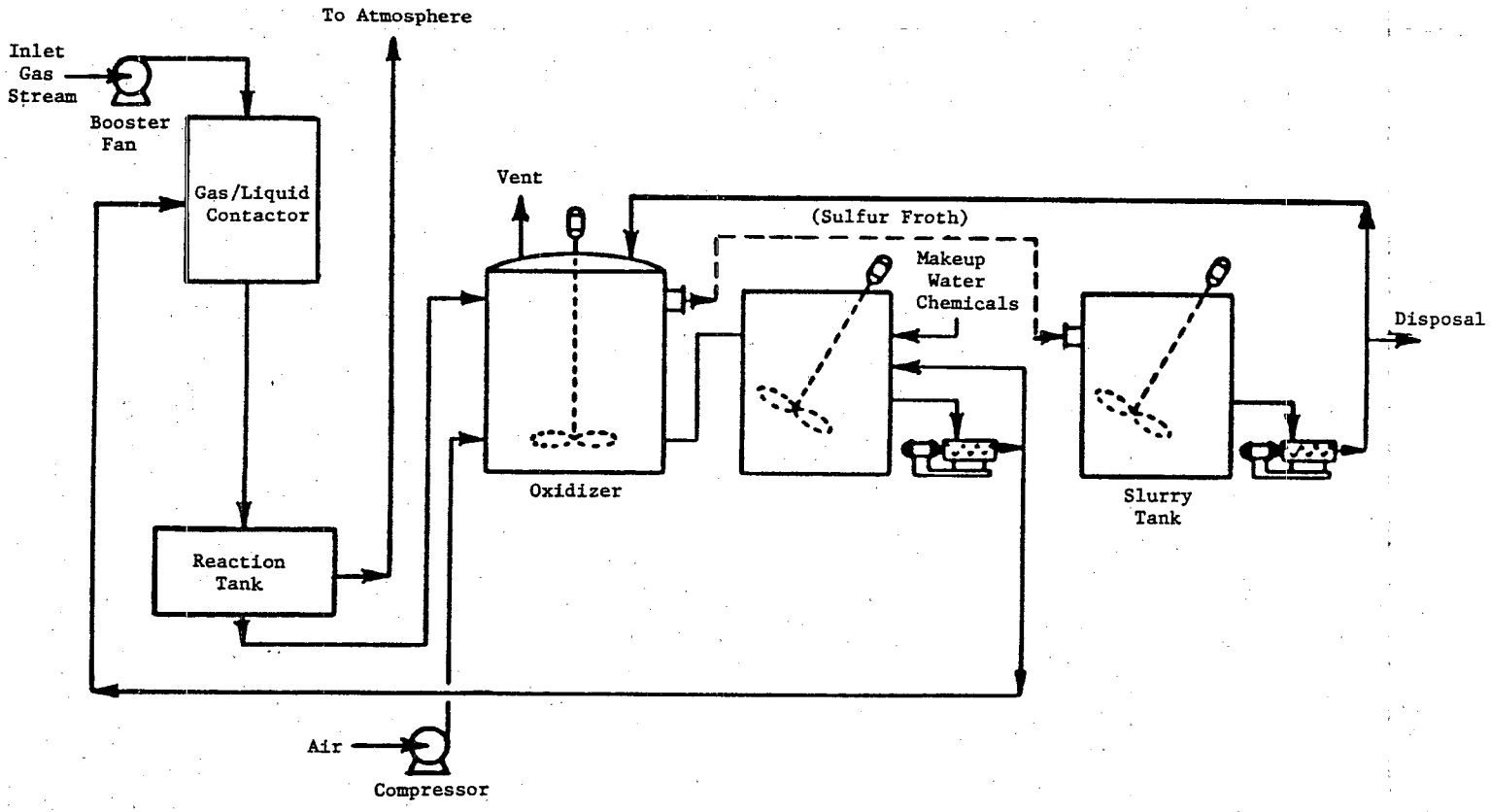


Figure 3-32. Simplified flow diagram of the Stretford pilot plant. (Taback, H.J., et al., 1985a)

operation, the sulfur cake is reslurred with steam condensate to reduce the Stretford reagent loss. The slurry is pumped through a sulfur melter and then into a steam-jacketed separator, where molten sulfur is recovered and the remaining dilute Stretford solution is returned to the system. Any wash water and reslurry water provide additional water load to the cooling tower duty. In an alternative mode of operation (i.e., direct autoclaving), the sulfur froth is pumped through a sulfur melter into an autoclave, where the phases are separated and the hot Stretford solution is returned to the surge tank. This mode of operation has the advantages of simplicity of operation, lower capital investment, and reduction of water evaporation load on the cooling tower.

B. Process Applicability--

Over 100 Stretford units are currently in operation worldwide, with capacities ranging from 2.7×10^9 to 5.4×10^{12} Nm^3/D and sulfur removal rates ranging from 0.45 to 82 Mg/D (Ondich, 1983). Applications include purification of Claus unit tail gases, amine regenerator off-gases, coke oven gases, coal gasification product streams, low Btu gas streams, oil gasification SNG streams, reformed petroleum products, and vent gases from geothermal steam turbines. Stretford sulfur recovery units also are included in the designs of at least four Lurgi coal gasification facilities in the U.S. (i.e., projects sponsored by Tenneco Coal Gasification, Nokota Company, WyCoal Gas, Inc., and Great Plains Gasification Associates) (Beychok & Rhodes, 1981).

Inlet concentrations of H_2S as low as 300 ppmv can be processed in a Stretford unit. The practical upper limit for H_2S inlet concentration is 15 percent, although in theory a unit could be developed for higher concentrations (Ondich, 1983).

The serious limitation of the Stretford process is the removal of organic sulfur. Tests performed with the EPA mobile Stretford plant have shown no organic sulfur removal (Taback, et al., 1985a). Ondich (1983) reports a private communication indicating 90 percent removal of methyl mercaptan but then states that carbonyl sulfide and carbon disulfide are not removed by the Stretford process. It is the concensus among the shale oil process designers that the Stretford process cannot be counted on for organic sulfur removal.

C. Process Performance--

When properly operated H_2S removal efficiency can be very high. Emissions of less than 1 ppmv H_2S can be obtained (Ondich, 1983). Any hydrogen cyanide in the feed stream is converted to sodium thiocyanate which is a contaminant to the Stretford solution.

Although there are many installations where the Stretford process is employed, actual operating data are limited. Stream characterization of most of the effluent streams, including trace and minor constituents, are lacking. Data from a design for a Stretford unit that would treat lean acid gas from the Rectisol unit at an El Paso Natural Gas Burnham coal gasification facility are shown in Table 3-26. Typical compositions of the Stretford absorber offgas from several Beavon sulfur removal plants which utilize the Stretford process are presented in Table 3-27. The recovered sulfur normally consists of 99.5 percent sulfur with small amounts of contaminants such as vanadium salts and sodium thiocyanate. Organic sulfur compounds and the lower hydrocarbons (e.g., containing five carbon atoms or less) essentially appear in the purified gas stream. Higher hydrocarbons as well as ammonia will be present in the oxidizer vent stream, if present in the absorber feed gas.

D. Process Reliability--

With so many constituents in the solution, it is not surprising that many Stretford users have reported startup problems such as low removal efficiency, foaming, clogging, etc. Because these problems are treated on a confidential basis, there are little published data on their cause.

The EPA reported foaming problems attributed to organic vapor in the inlet gas. Low sulfur removal efficiencies also were experienced. These were attributed to poor gas to liquid contact. Both problems appeared to be solvable (Taback, et al., 1985a). Preliminary information from the SASOL Lurgi coal gasification facility in South Africa indicated that biological growth due to high carbon dioxide levels in the feed was primarily responsible for plugging problems in the Stretford absorption towers. Conventional absorbers consist of packed towers. Plugging normally occurs in the bottom 2 to 3 meters of packing after 6 to 12 months operation [2]. The use of a

TABLE 3-26. DESIGN DATA FOR A STRETFORD UNIT TREATING RECTISOL OFFGAS AT EL PASO NATURAL GAS BURNHAM COAL GASIFICATION FACILITY (Ondich, 1983)

Component	Raw Gas	Absorber Offgas	Solvent Purge
CO, %	96.0	99.0	
H_2S, ppmv	7400	8	
COS, ppmv	77	75	
CS, ppmv	2	2	
HCN, ppmv	30	0	
CO, %	0.17	0.16	
CH_4, %	0.53	0.52	
C_2H_4, %	0.22	0.22	
C_2H_6, %	0.30	0.29	
H_2, %	0.43	0.42	
H_2O, %	1.6	4.32	80.0
$Na_2 S_2O_3$, %			10.8
NaSCN, %			4.4
$NaVO_3$, %			0.7
ADA, %			1.1
$NaHCO_3 + Na_2CO_3$, %			3.0

TABLE 3-27. TYPICAL COMPOSITIONS OF STRETFORD ABSORBER OFFGAS IN BEAVON SULFUR REMOVAL PLANTS (Ondich, 1983)

Component	Facility A	Facility B	Facility C	Facility D	Facility E
H_2S, ppmv	5	<1	<0.1-7	<1	<1
SO_2, ppmv	<20	<1	<5	<0.4	<1
COS, ppmv	50	30	7-23	5	15
CS_2, ppmv	20	9	1	0.5	0.1
CH_3SH, ppmv	<5	NR	NR	NR	NR
CO, ppmv	NR	500	565	250	670
CH_4, ppmv	NR	200	221	NR	NR
N_2 + AR, %	NR	89	NR	NR	NR
CO_2, %	NR	7	5.7	NR	NR
H_2, %	NR	3	5.6	NR	NR

NR - Not reported

venturi scrubber followed by a short conventional absorber will result in appreciable reductions in investment and maintenance costs since plugging is limited to the venturi scrubber which can be more easily cleaned than a conventional absorber (Ondich, 1983). Stretford designs with two venturi scrubbers in parallel (one spare) followed by a short absorber can conceivably eliminate plant shutdown for absorption equipment cleanout. Plugging of sulfur froth lines is greatly reduced by using long-radius elbows and flushing the lines before shutdown. Some provision must be made for sour gas disposal during absorber cleanouts and any mechanical failures.

E. Process Economics--

Investment and operating costs are affected by operating pressure, hydrogen sulfide content of feed gas, and disposition of sulfur products. The capital investment costs for a Stretford unit are presented as a function of sulfur removal capacity in Figure 3-33. The capital investment costs were obtained by EPA from studies recently performed by Catalytic, Inc., Oak Ridge National Laboratory, and Black & Veatch [Ondich, 1983]. From data provided by J. F. Pritchard Co. to Black & Veatch, an equation was developed relating capital cost to sulfur recovered and flow rate as follows:

$$\left(\frac{A}{36.1}\right)^{0.6} \left(\frac{B}{37.5}\right)^{0.18} (7.8) = C$$

where: A = sulfur removed, kmol/hr

B = feed gas flow rate, m^3/min per kmol sulfur removed per hour

C = total capital cost less interest during construction, $\$10^6$ (1980)

Operating requirements for the Stretford process (presented in Table 3-28) were extracted from information provided by the Ralph M. Parsons Co. to Black & Veatch with respect to the Beavon sulfur removal process. Other cost data also were reviewed (Ondich, 1983).

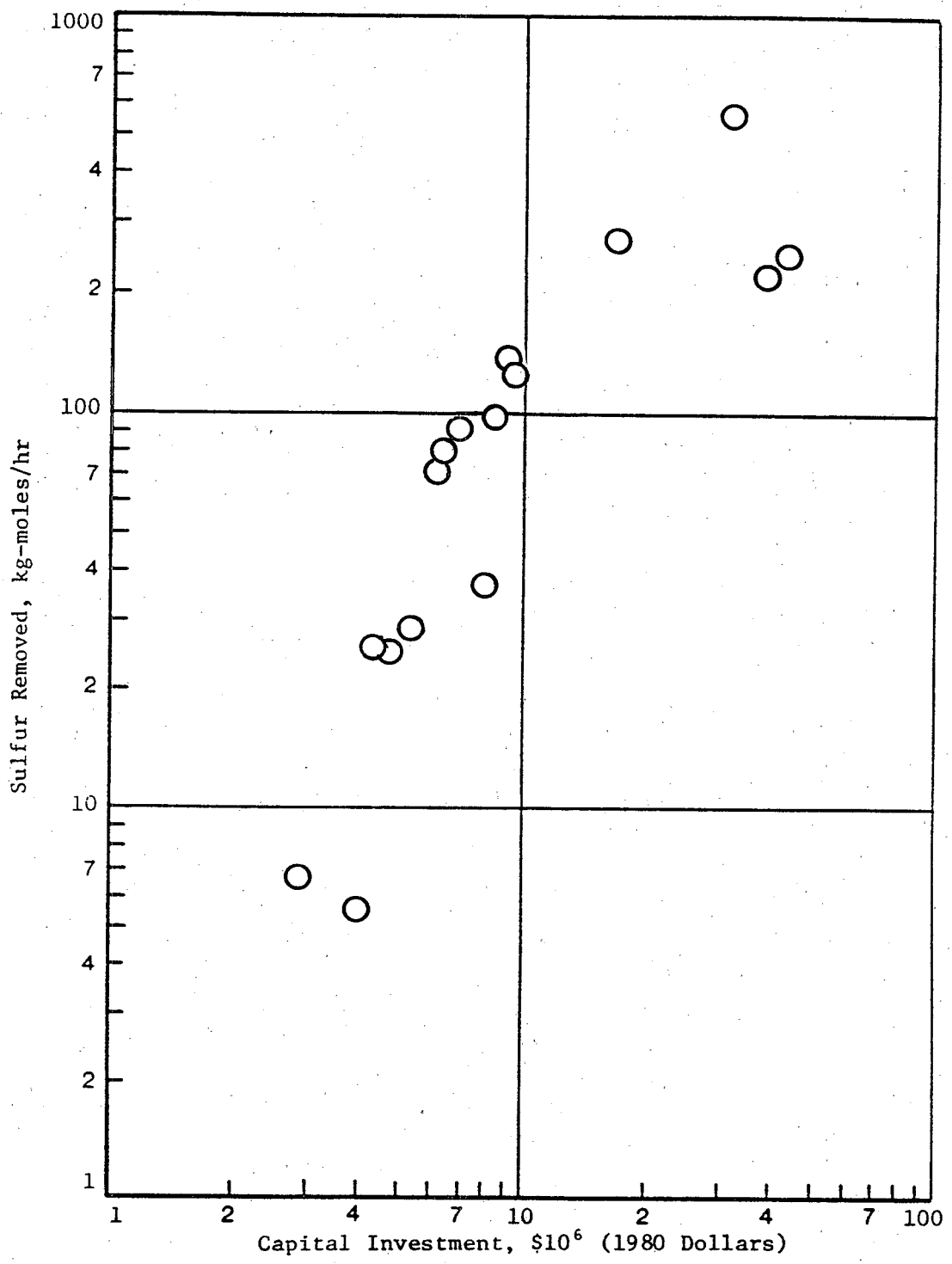


Figure 3-33. Stretford capital investment cost (less interest during construction.) (Adapted from Ondich, 1983)

TABLE 3-28. OPERATING REQUIREMENTS FOR THE STRETFORD PROCESS[a]
(Ondich, 1983)

For 36.1 kmol/hr (27 Mg/day) unit		
cooling water	=	5.70 m^3/min
electricity	=	610 kW
For 41 kmol/hr (31 Mg/day) unit		
chemicals	=	$350,000/year
steam	=	7.30 Mg/hr (16,000 lb/hr)

[a] Data extracted from information provided to Black & Veatch by the Ralph M. Parsons Company for the Beavon Sulfur Removal Process, Nov. 1980.

3.3.2 Lo-Cat Process (Reduced Sulfur Compounds)

A. Process Description--

The Lo-Cat process is similar to the Stretford process except that chelated iron is used in place of sodium metavanadate as the principal agent for producing elemental sulfur. The details of the chemical scrubbing solution are proprietary with ARI Technologies, Inc. Palatine, Illinois.

The process using ARI-310 catalytic reagent is the third generation of a process which was originated by Humphreys and Glasgow in London. Their Chelated Iron Process used ethylene diamine tetra acetic acid (EDTA) to hold iron in solution. The process was incorporated in an oil refinery in Landarcy, Wales, in 1964, but was not a success because of the instability of the chelated iron compound.

ARI developed the second generation process which overcame the problem of iron precipitation at high pH, by the addition of a polyhydroxylated sugar or "Type B" chelating agent to the EDTA or "Type A" chelate. An installation was made at the Plateau Refining Company in Bloomfield, New Mexico in 1977. This was followed by the installation of six small units, and two refinery

process units which produced 7 to 15 long tons of sulfur per day. These units performed well with respect to H_2S removal, but higher than expected degradation of the organic chelating agents was experienced. ARI-310, developed in 1982, improved stability.

In the Lo-Cat process, the chelated iron acts as a catalyst to the overall reaction:

$$2H_2S + O_2 \rightarrow 2H_2O + 2S \downarrow$$

A schematic of the process is shown in Figure 3-34. In the absorber, the iron reacts with sulfide ions, accomplishing the oxidation in the electro-chemical sense by accepting electrons according to the half reaction:

$$2Fe^{+++} + S^{=} \rightarrow 2Fe^{++} + S \downarrow$$

No oxygen is involved in the production of the sulfur.

In the oxidizer, the iron reacts with dissolved oxygen and is regenerated for reuse according to the equation:

$$2Fe^{++} + O_2 + 2H_2O \rightarrow 2Fe^{+++} + 4(OH)^{-}$$

These reactions are specific for the production of sulfur. However, a significant byproduct is formed whenever sulfide ions and oxygen are brought together. Thiosulfate is produced when sulfide ions "leak" out of the absorber and get into the oxidizer. This happens in the Stretford process because these catalysts have low activity and finish the reaction slowly. The Lo-Cat reaction is faster which eliminates this problem. Thiosulfate is also produced when oxygen dissolved in the oxidizer is not consumed before the solution is carried into the absorber. This occurs in the Lo-Cat process because there is an equilibrium between the oxidation state of the iron and the oxygen content of the solution. (Hardison, 1984a). Lo-Cat plants operate with approximately 1 ppm of oxygen in the reoxidized solution, so there is little byproduct thiosulfate formation.

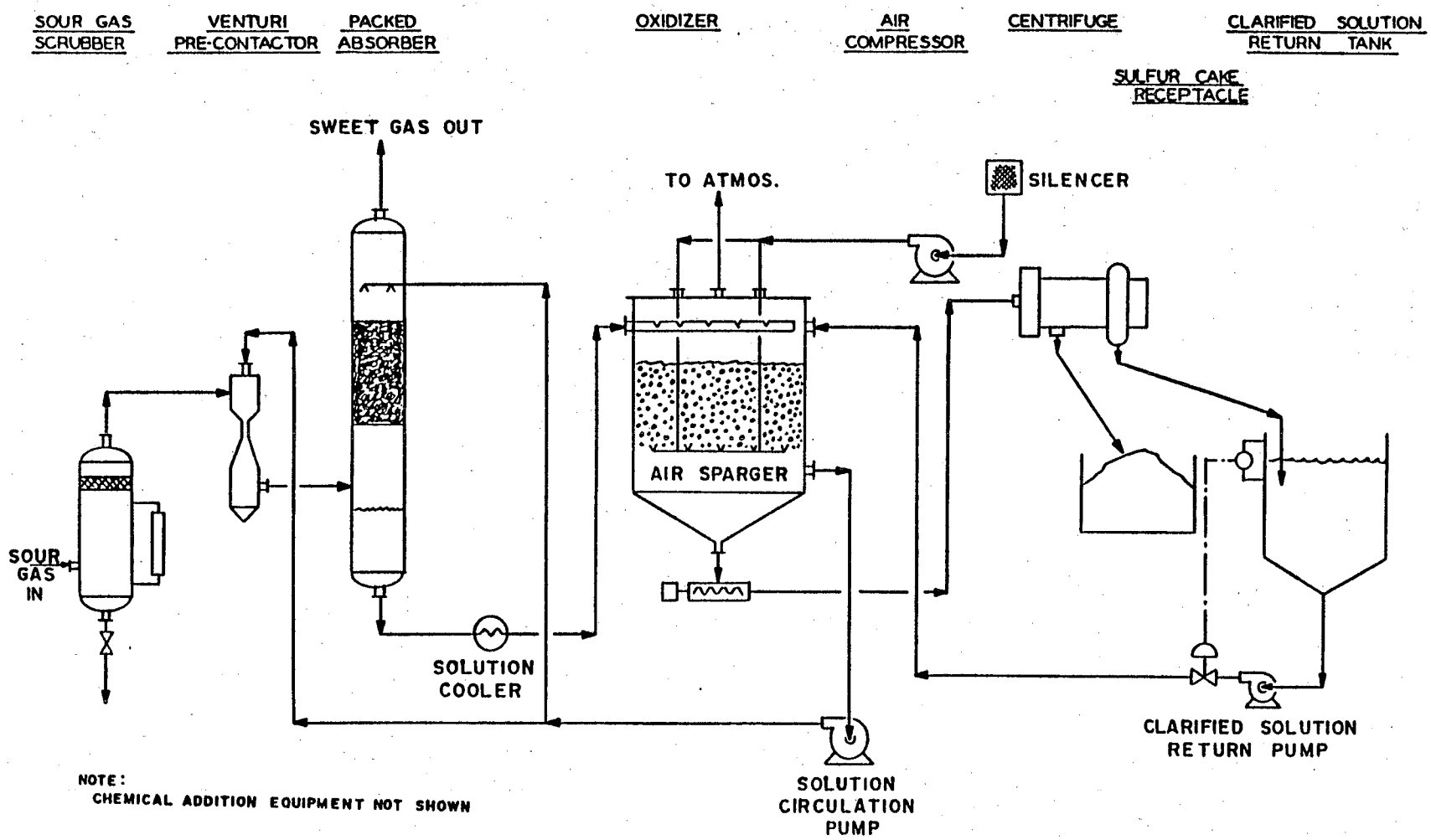


Figure 3-34. Flow scheme for conventional Lo-Cat® unit (Hardison, 1984a). Reprinted by permission of ARi Technologies, Inc., Palatine, IL.

The rapid reaction of the Lo-Cat solution with H_2S producing elemental sulfur requires careful consideration in designing the gas/liquid contactor. A packed tower such as used for the Stretford process can become clogged with sulfur in a Lo-Cat process. A venturi contactor alone, or preceding the packed tower (as shown in Figure 3-34) should be considered for this process.

B. Process Applicability--

The Lo-Cat process is intended for the same applications as the Stretford. Its only use on a shale oil retort was at Geokinetics (Hardison and Lekas, 1984). The design conditions at this plant were as follows:

Flow	600 Nm^3/min
Pressure	1×10^{-4} Pa
Moisture	10 to 12 vol%
Temperature	40 C (maximum)
Gas Composition	(Dry Basis)
H_2S	0.220 (2000 ppmV maximum)
CO_2	22.17
O_2	1.97
CO	5.31
N_2	58.76
H_2	9.25
COS	0.008
CS_2	--
C_1	1.44
C_2	0.41
C_3	0.29
C_4	0.10
C_5	0.04
C_6+	0.05
	100.00

The Lo-Cat system at Geokinetics was assembled from used hardware available from an abandoned Stretford plant. The major items of refurbished equipment used included:

Absorber Vessel. A 23 cm x 18 m column packed with 9 m 5 cm diameter stainless steel pall rings, estimated to give an efficiency of 99.3 percent minimum.

Two Oxidizer Vessels. Each 4 m x 4 m vessel with an air sparger and 55° cone bottom which were coated internally with epoxy.

Two Air Blowers. Each 40 Nm^3/min unit.

Two Solution Circulation Pumps. Each 750 gpm centrifugal pump.

The potential suitability of the Lo-Cat process for cleaning the gas and for production of elemental sulfur was demonstrated. During the first two months of operation, (February 5 through March, 1984), H_2S loading at the inlet averaged 1600 ppm and was decreased to about 2 ppm by the Lo-Cat unit with gas flows at approximately the 400 Nm^3/min "normal" rate. Catalyst solution circulation rate was maintained at approximately 6000 L/min and catalyst solution strength was 1000 to 2000 ppm iron content (Hardison and Lekas, 1984).

Consumption of catalyst concentrate (ARI-310 is furnished in liquid form with approximately 18,000 ppm of active Fe) has been minimal on units with operating melters, but is discarded with the sulfur when a melter is not used. In this case, the discard rate constituted the greatest single operating expnese, and reached approximately 6.5 gallons per hour of catalyst solution. ARI-310M solution is a chelate-rich concentrate which is added to replace chelate compounds which degrade slowly by oxidation. ARI-310SA is a startup additive used to bring the solution to equilibrium conditions quickly. The cost of chemicals for operation of the Lo-Cat unit from January through September 1984 are shown in Table 3-29 (the sulfur production was not measured but has been estimated at 240 days x 0.85 onstream factor x 1.25 metric ton/day = 256 metric tons for the project).

TABLE 3-29. CHEMICAL COSTS FOR LO-CAT
(Hardison, 1984b)

Chemical	Total Purchased	Original Fill	Operating Supply	Operating Cost $/metric ton
ARI-310 Catalyst Concentrate	$44,400	$23,100	$21,300	$ 83.20
ARI-310SA Startup Additive	6,300	6,300	--	--
ARI-310M Makeup Solution	8,500	--	8,450	33.10
ARI-400 Biochem	1,450	--	1,450	5.70
Antifoam	290	--	290	1.10
	$61,000	$29,364	$31,500	$123,00

The initial operation of the plant was carried out with the ammonia absorption and stripping towers in operation. During this period, considerable difficulty was experienced with fouling of the ammonia wash tower by tar carried over from the retorts through a marginal entrainment separator vessel. The Lo-Cat absorber operated at high efficiency (>99 percent) without plugging during this period.

Because the Lo-Cat unit is tolerant of high ammonia concentrations, the ammonia wash system was taken out of service. Subsequently, the packing in the Lo-Cat absorber was fouled repeatedly with oily sulfur and several shutdown periods were required to dump and clean the packing. The combination of tar and condensed sulfur could not be flushed successfully through the two-inch pall rings.

The scrubber was modified to include a venturi prescrubber to reduce the tar and sulfur load on the packed column, but the venturi did not correct the plugging problem and it was frequently necessary to bypass some gas around the absorber to keep the pressure drop down to the 0.5 psi available.

Several conclusions can be drawn with regard to packed columns in this service:

- In general, packed columns in slurry service should have provision for cleaning without interfering with normal operation. A spare column is recommended where 100 percent onstream factor is essential.

- Oil and tars must be separated from the gas stream before it reaches the sulfur-removing process.

- Low density packings, which have given satisfactory service in several other Lo-Cat units, are recommended over the conventional pall rings used in the Geokinetics unit.

The plugging problem was mitigated, but not eliminated by replacing the lower packed bed with a low pressure drop venturi.

The Lo-Cat solution functioned well and remain active and resisting contamination. When the plant was operated for a period of several hours without reoxidation, the catalyst solution turned black (normal color dark brown or coffee color) and showed redox potentials below -300 mv, representing totally reduced iron. On aeration, the catalyst solution regained full activity.

C. Process Reliability--

The reliability of an ARI Lo-Cat system in a retort gas application has not be adequately stressed. The process chemistry appears to be reasonably stable and resistant to the normal contaminants. Undoubtedly, it is desirable to rid the process gas of oil, tar, moisture and particulate matter before it is delivered to the unit. This is equally true for the Stretford process.

The chelated iron solution is corrosive to many varieties of steel. Manufacturer's material recommendations and warranties are important in selecting materials which contact the solution.

The selection of a contactor may involve a tradeoff between removal efficiency and maintenance. Dense packed towers that achieve high removal efficiency may clog. Solutions to this problem include providing a second packed tower to permit continuous operation while one tower is being cleaned

or to add one or more venturi contactors to remove most of the sulfur before the packed tower does the final gas polishing.

D. Process Economics--

The cost of a basic Lo-Cat plant similar to the one shown in Figure 3-34 which ARI Technologies refer to as an autocirculation unit (Hardison, 1984a) is presented in Table 3-30:

TABLE 3-30. CAPITAL COSTS FOR LO-CAT PLANT

Capacity	Cost $			
Metric Ton/day Sulfur	Design	Equipment	Centrifuge	Total
0.1	10,000	110,000	N/A	120,000
0.5	10,000	160,000	N/A	170,000
1.0	10,000	230,000	120,000	360,000
3.0	10,000	350,000	120,000	480,000

Another cost basis for Lo-Cat (Table 3-31) was provided by Hydrocarbon Processing (1984) as follows:

TABLE 3-31. COST BASIS FOR LO-CAT
(Hydrocarbon Processing, 1984)

Gas Flow, MMscfd	2	5	100
(MNM^3/day)	(57)	(142)	(2,800)
H_2S Conc., ppmV	6,000	500	1,000
Sulfur Prod. Metric Ton/day	0.5	0.1	3.8
Capital Cost, $1,000	600	350	3,000
Net Op. Cost, $/MCF	0.06	0.008	0.0002
($/$Mm^3$)	(2)	(0.3)	(0.007)

A final data point is the operating cost for a Lo-Cat plant in a Texaco gas field processing 10 MMscfd (280 MNm^3/day) (the sulfur concentration was unspecified) of gas containing 99.6 percent CO_2 and water (Hardison, Oil and Gas Journal, 1984). The annual operating costs are:

Chemical	
Catalyst Concentrate	$10,000
Catalyst Makeup	7,700
Buffer (KOH)	9,100
Electrical	13,700
TOTAL	$40,500

3.3.3 Unisulf Process (H_2S)

The Unisulf process is a homogeneous aqueous catalytic process which converts absorbed H_2S to elemental sulfur using sparged air. In operation, it resembles the Stretford process giving the same high sulfur yields ($\geq$ 99.9%);

however, Union Oil, its developer and only user in the U.S., claims that the Unisulf process has a much lower chemical consumption. Union Oil's shale retort off-gas located at their Parachute Shale Oil Project in Colorado, now undergoing startup, utilizes the Unisulf process on the retort offgas.

A. Process Description--

The Unisulf process is a homogeneous catalytic process for oxidizing H_2S to sulfur which is similar to the Stretford process in concept and application but it operates without a bleed stream. Although it can be used to recover sulfur from gas streams rich in H_2S, it is expected to find broadest application in recovering over 99.9% of the sulfur from gas streams containing less than about 10 mole percent H_2S. The Unisulf process employs an aqueous solution composition which absorbs H_2S and converts it to elemental sulfur by oxidation with air. The absorbing solution is also regenerated and recycled back to the absorber. Typical components of the Unisulf solution are as follows:

- Sodium Carbonate and Sodium Bicarbonate
- Vanadium Complex
- Thiocyanate Ions
- Carboxylate Complexing Aent
- A Water Soluble Aromatic Compound
- Water

A flow sheet of the Unisulf process is shown in Figure 3-35. Acid gas is fed to a venturi scrubber and absorber which operates in series to remove over 99.9 mole percent of the H_2S. The treated acid gas leaves the absorber odor free, usually containing less than 1.0 ppmv H_2S. The absorber also serves as a reaction vessel to oxidize the absorbed H_2S to particles of elemental sulfur in the reaction:

$$2V^{+5} + H_2S + 2\ OH^- \rightarrow 2V^{+4} + 2H_2O + S$$

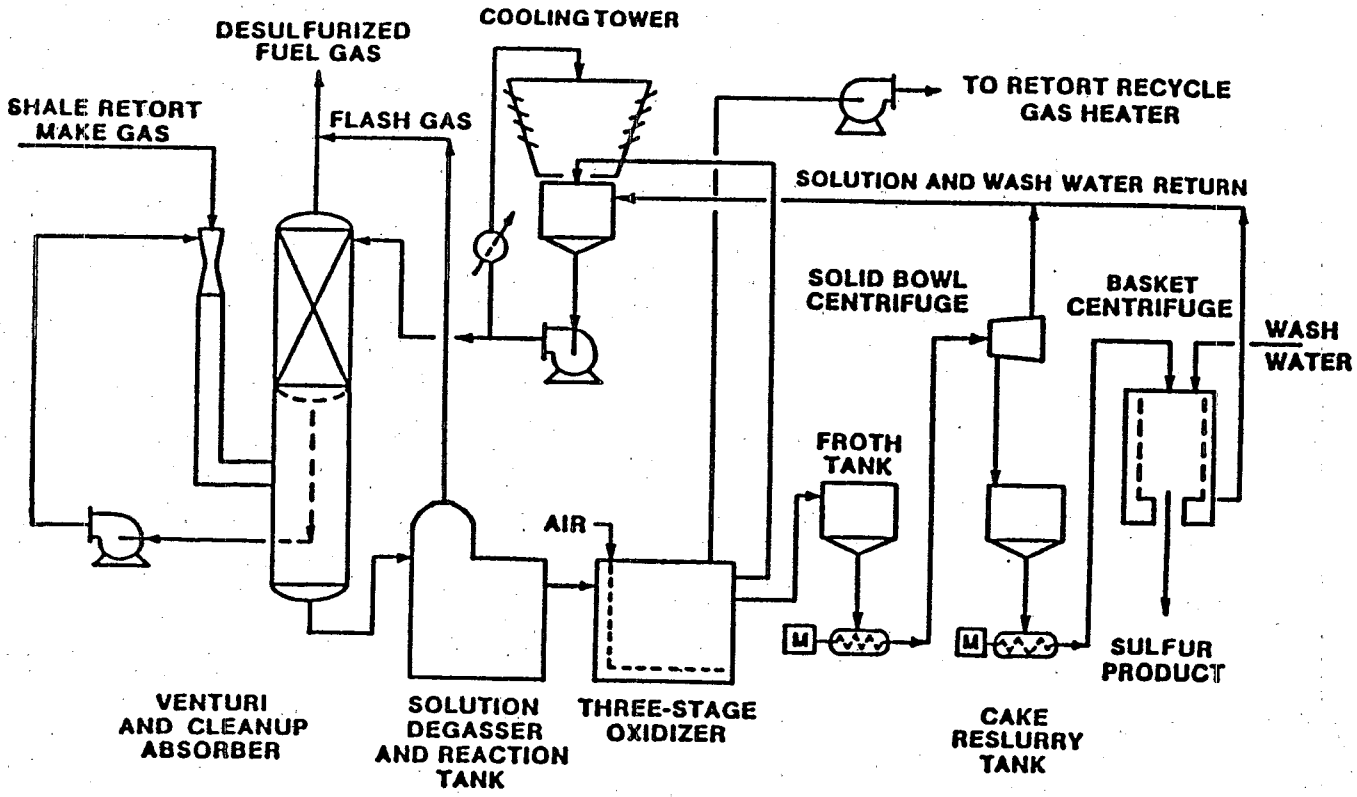


Figure 3-35. Unisulf Process (after Hass, 1984).

After a short residence time in the absorber, the slurry is transferred to an oxidizing vessel which uses sparged air to oxidize the reduced vanadium ions from V^{+4} to V^{+5} in the reaction:

$$2V^{+4} + H_2O + 1/2\ O_2 \rightarrow 2V^{+5} + 2OH^-$$

Simultaneously, the sulfur particles adhere to air bubbles and float to the surface of the aqueous solution by a froth flotation technique. The sulfur froth overflows through a weir into a receiving tank forming a slurry. The solids are later mechanically separated from the mother liquid by either filtering or centrifuging. The filtrate is pumped to a balance tank which also receives the regenerated solution leaving the oxidizer. The balance tank is used mainly as a inventory vessel for the regenerated solution which is recycled back to the absorber and venturi scrubber. The properties of the unisulf absorber solution and the sulfur which is formed are different than those of the Stretford process and therefore cause an improved absorber operation and a more efficient separation of sulfur from the regenerated solution than for the Stretford process.

The effects of design and operating parameters on the recovery of sulfur provided by the Unisulf plant are summarized in Figure 3-36. It shows that an efficient vertical basket centrifuge (121.9 cm diameter x 76.2 cm depth) can process 6500 kg/day of sulfur from a feed slurry containing five percent (wt) sulfur received directly from the oxidizer. The centrifuge cycle includes alternating washing and dewatering steps. To process greater quantities of sulfur, the slurry would have to be concentrated above five percent sulfur. The entire process operates at ambient temperature, 20 to 50 C (70 to 120°F).

B. Process Applicability--

The Unisulf process has been adapted to treat shale oil retort gas, Rectisol expansion gas, Claus plant tail gas, and other nonconventional gases. The first commercial application of the Unisulf process was made at the SASOL I facility located at Sasolburg, Republic of South Africa, beginning in December 1981. In these facilities, the sulfur recovery plants treat a Rectisol plant tail gas containing about 1.3 mole percent H_2S in CO_2. A typical off-gas composition includes the following percentages:

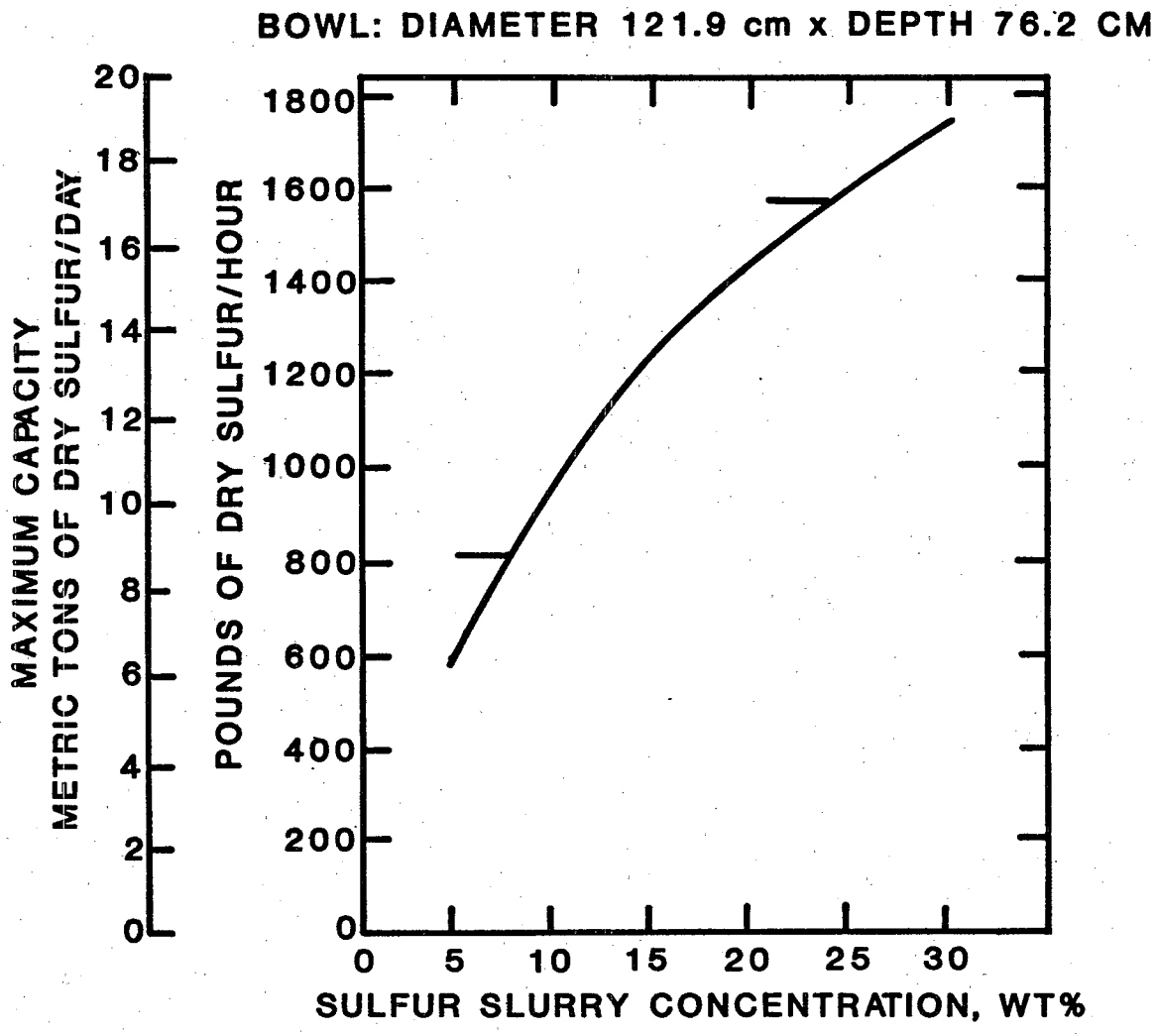


Figure 3-36. Centrifuge capacity for separating sulfur from aqueous feed slurry (after Hass, 1984).

CO_2	97
H_2S	1.3
Saturated $C_1 - C_5$	1.0
Unsaturated $C_1 - C_5$	0.2
CO	0.5
	100

plus traces (up to 50 ppmv) of HCN. In this application, the Unisulf process recovered about 15 t/d of sulfur. During a 160-day test, over 2200 metric tons of sulfur were recovered. The only major problem encountered was the presence of contaminant oxygen in the feed gas. In the Rectisol plant, rich methanol solvent is expanded in several stages to release acid gases. The last two stages operate at subatmospheric pressure, thus requiring vacuum blowers to pump the acid gas to the Unisulf process. Unless the seals on these blowers are properly maintained (purged with an inert gas), air can leak into the acid gas. Oxygen and H_2S can then react under basic conditions to form thiosulphate.

A processing scheme was developed by Union which minimizes thiosulphate buildup if the contaminant oxygen concentration in the expansion gas can be controlled below a specified low level. Subsequent operation has proved that with these modifications, the Unisulf solution can be used without thiosulphate production even though the feed gas does contain a low level of contaminant oxygen. This apparently eliminates the need for purge streams and disposal of spent solution.

The retort offgas composition at the Unisulf plant in Parachute Creek includes the following percentages (Hass, et al., 1984).

Saturated $C_1 - C_5$	37
H_2	24
CO_2	19
Unsaturated $C_1 - C_5$	8
CO	6
H_2S	5
C_6+	0.6
N_2	0.4
	100

plus approximately 150 ppmV of HCN.

The Unisulf process is being used along with the Beavon Sulfur Removal Process to treat Claus plant offgas at Union's shale oil upgrading plant in Colorado. The Beavon process converts all sulfur compounds in the Calus tailgas (SO_x, S_x, COS, CS_2, etc.) to H_2S. The inlet gas composition to the Unisulf plant includes the following percentages:

N_2	90
CO_2	6
H_2	3
H_2S	1
	100

plus CO and saturated hydrocarbons of 500 ppmv each.

Further applications of the Unisulf process are expected including recovering sulfur from sour gas streams leaving tar sands operations, sewage disposal plants, and geothermal operations (Hass, et. al., 1984).

C. Process Performance and Economics--

At this time no commercial data are available for the Unisulf process performance or costs. The manufacturer expects that the Unisulf process should have the same capital costs and high sulfur recovery efficiency (99 plus percent) as reported earlier for the Stretford process, however, the Unisulf plant has a lower operating cost.

3.3.4 Alkaline Scrubbing

Based on some pilot scale tests on offgas from an in-situ shale oil retort the concept of alkaline scrubbing of H_2S appears to be viable for this application (Taback et al., 1985a). Much more development is required to prove the concept and produce reliable hardware. However, in view of the potential economic advantages, this process is presented here as an option to be considered as a part of a future oil shale process plant development.

The process schematic for removing H_2S from a gas stream that is rich in CO_2 and that may also contain NH_3 is shown in Figure 3-37. The process gas enters the alkaline scrubber where the H_2S any NH_3 present and some CO_2 are absorbed by the scrubber liquid. This process is selective so that more H_2S is absorbed compared to its concentration than CO_2. The measure of the H_2S

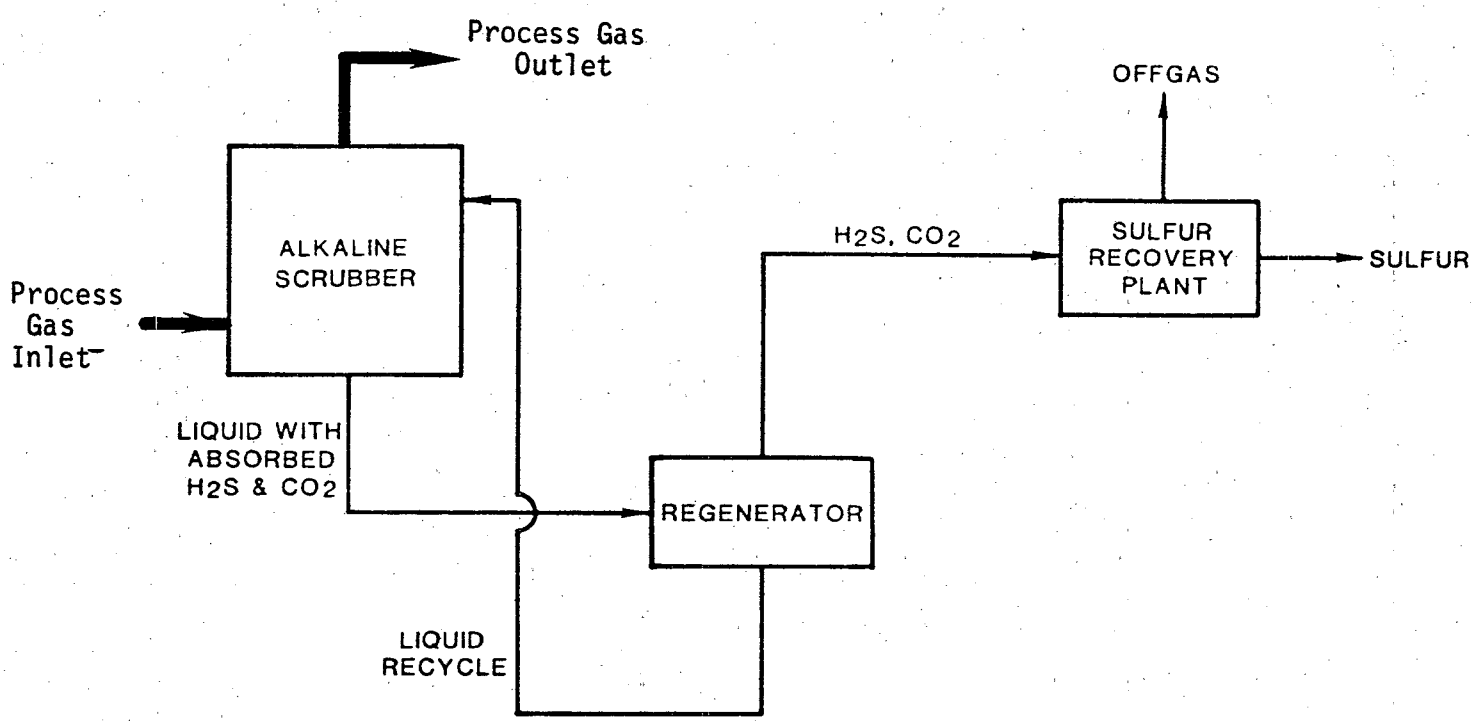


Figure 3-37. Alkaline scrubbing process schematic for H_2S removal. (Taback, et al., 1985a).

selectivity is the percent reduction of H_2S divided by the percent reduction of CO_2. Depending on the initial concentrations of H_2S and CO_2 in the process gas, selectivities of over 40 are desired.

The next step in the process is to distill off the absorbed H_2S and CO_2 from the scrubber liquid in the regenerator. The gaseous H_2S and CO_2 (with an enriched amount of H_2S) are sent to a sulfur recovery process of which the Claus is probably the most cost effective. The Claus process prefers the highest possible H_2S concentration. The process might work with 8 percent H_2S but 15 percent is needed for confidence and 25 percent or higher is desired (Lovell et al., 1982).

For a typical in-situ retort the H_2S concentration might be 1500 ppmv and the CO_2 concentration 22 percent. The CO_2 concentration is 150 times greater than the for H_2S in the process gas. To enrich the H_2S concentration the following H_2S selectivities* are required (Taback et al., 1985a).

Sulfur Plant (Claus) Inlet gas H_2S Percentage	Scrubber Selectivity* Required
8	12
15	23
25	38

Thus a selectivity of over 40 is desired. This is undoubtedly the worst case. If the H_2S concentration is higher, such as in an indirectly heated retort gas where it might reach 4 percent, then a much lower selectivity is acceptable.

The alkaline scrubber removes the NH_3 as well and it may be recovered an appropriate point in the regeneration process. The recovered NH_3 may be used elsewhere in the plant for NO_x reduction or incinerated with proper staging to prevent NO_x formation.

*Selectivity is defined rigorously by Taback, et al. (1985a). Qualitatively, selectivity is a measure of the preference of the process to scrub H_2S over CO_2.

While it is important that the scrubber achieve the required selectivity to accommodate the claus plant, the importance of overall sulfur removal must be maintained. Most process changes which improve selectivity worsen sulfur recovery. Venturi and tray tower scrubbers were investigated in the EPA's pilot program. A plot of selectivity versus removal efficiency for both types of scrubbers is presented in Figure 3-38. The tower favors removal while the venturi favors selectivity. Taback et al. suggests a multi-stage scrubber. One example presented combines a venturi and tower as shown in Figure 3-39. The result is 95 percent H_2S removal efficiency and 37 selectivity. Multiple tower stages where the gas passes through fresh solution at each stage are shown in Figure 3-40. The removal efficiency is estimated at 99 percent, but the selectivity is only 22.

3.3.5 Activated Carbon and Hypochlorite Process

A variation of the alkaline scrubbing process for H_2S removal has been developed by the Pulp & Paper Research Institute of Canada (PPRIC) and Teller Environmental Systems, Inc. (TESI) (Prakacs, 1983, and Teller and Amberg, 1985). The process consists of alkaline absorption of the acid gases in the presence of activated carbon and with sodium hypochlorite added to the solution to improve the solubility of the organic sulfur gases. The particular process commercialized by TESI includes provision for particulate nucleation which results in particulate removal as well as absorption of the acid gases. Unlike other H_2S removal systems, this one has the capability of removing organic sulfur species (mercaptans, thiophenes, etc.) which is important in achieving the low total sulfur emissions needed to comply with PSD regulations. This process has the capability of reducing the total reduced sulfur (TRS) emissions to the level of 2 - 5 ppm and particulate emissions to the level of 0.06 to 0.18 g/Nm^3, dry.

A. Process Design--

A schematic of the activated-carbon hypochlorite process is shown in Figure 3-41. Retort is first treated in a venturi scrubber which serves to wet the particulate for collection within the packed bed section. The gas then flows to a packed bed section where it is contacted with a caustic slurry containing oxygenated activated carbon and sodium hypochlorite plus the reduced sulfur species removed from the gas phase.

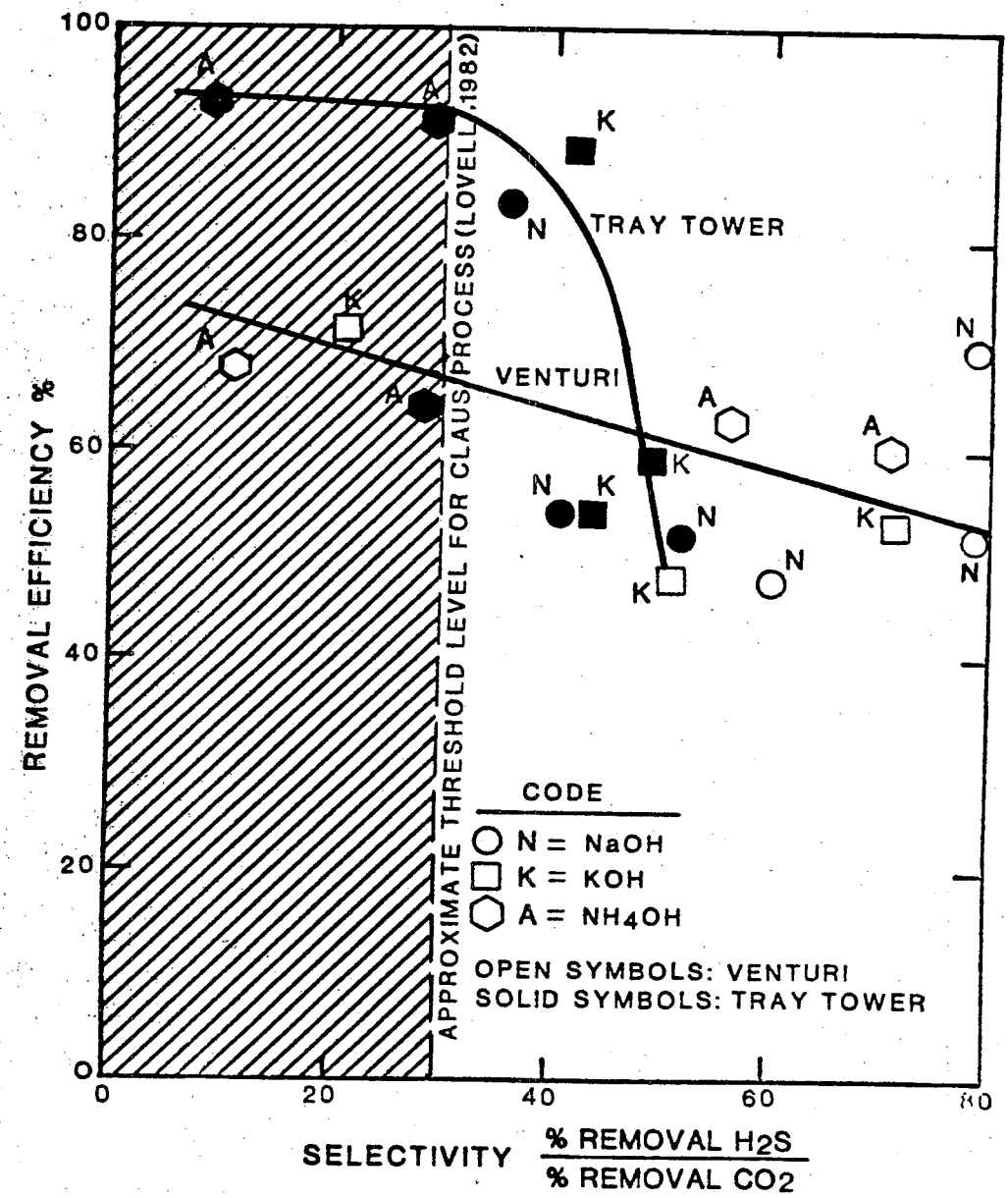
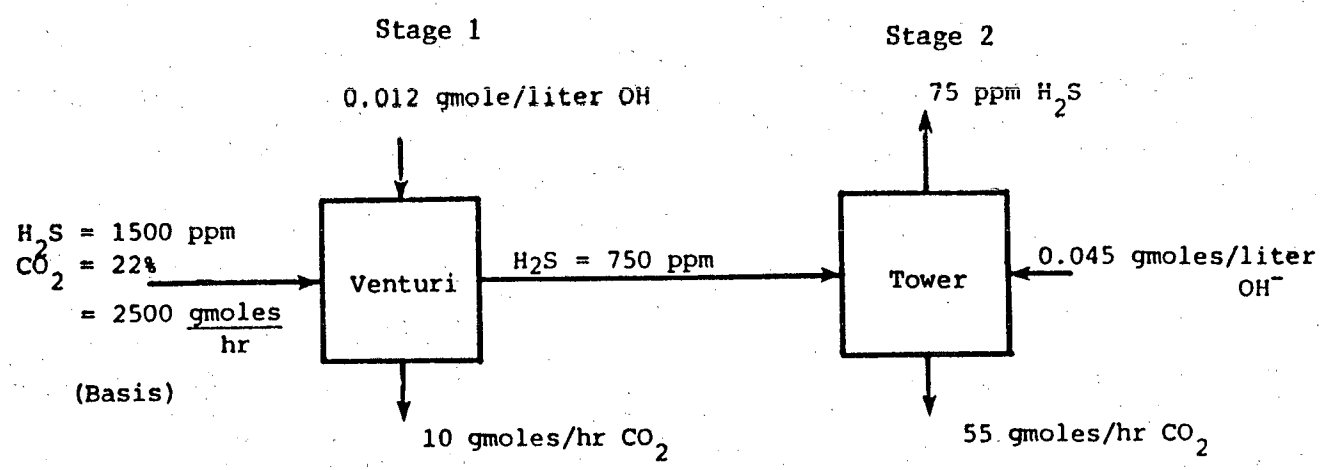


Figure 3-38. Removal efficiency vs. selectivity for alkaline scrubber (Taback, et al., 1985a).



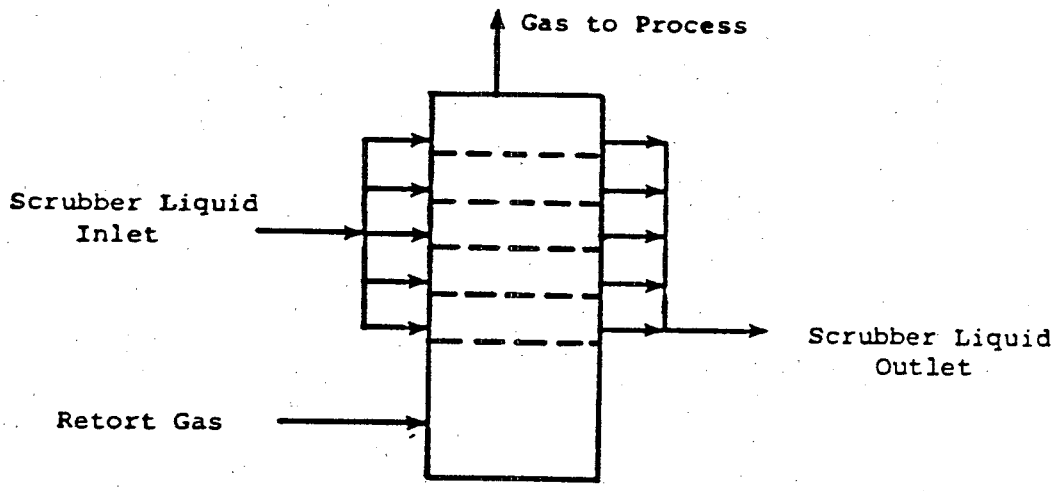
	Position	H_2S, ppm	CO_2, %	CO_2 gmoles		OH^-, gmoles/liter
			Gas	Liquid		
	Inlet	1500	22	2500		
Outlet	Stage 1-Venturi	750	21.9	2490		0.012
Outlet	Stage2-Tower	75	21.4	2435		0.045
				65		

$$H_2S \text{ removal Efficiency} = \frac{1500 - 75}{1500} = 95\%$$

$$CO_2 \text{ Removal Efficiency} = \frac{2500 - 2435}{2500} = 2.6\%$$

Selectivity = 37

Figure 3-39. Two-stage alkaline scrubber; venturi and tray tower. (Taback, H.J., et al., 1985a).



	Inlet (Assumed)	Stage 1 Exit	Scrubber Exit
H_2S Conc., ppm	1500	150	15
CO_2 Conc., %	22	21.5	21.0

H_2 removal eff. = $\frac{1500-15}{1500} = 99\%$

CO_2 removal eff. = $\frac{22-21}{22} = 4.5\%$

Selectivity = $\frac{99}{4.5} = 22$

Figure 3-40. Tray tower scrubber with isolated liquid inlets. (Taback et al, 1985a)

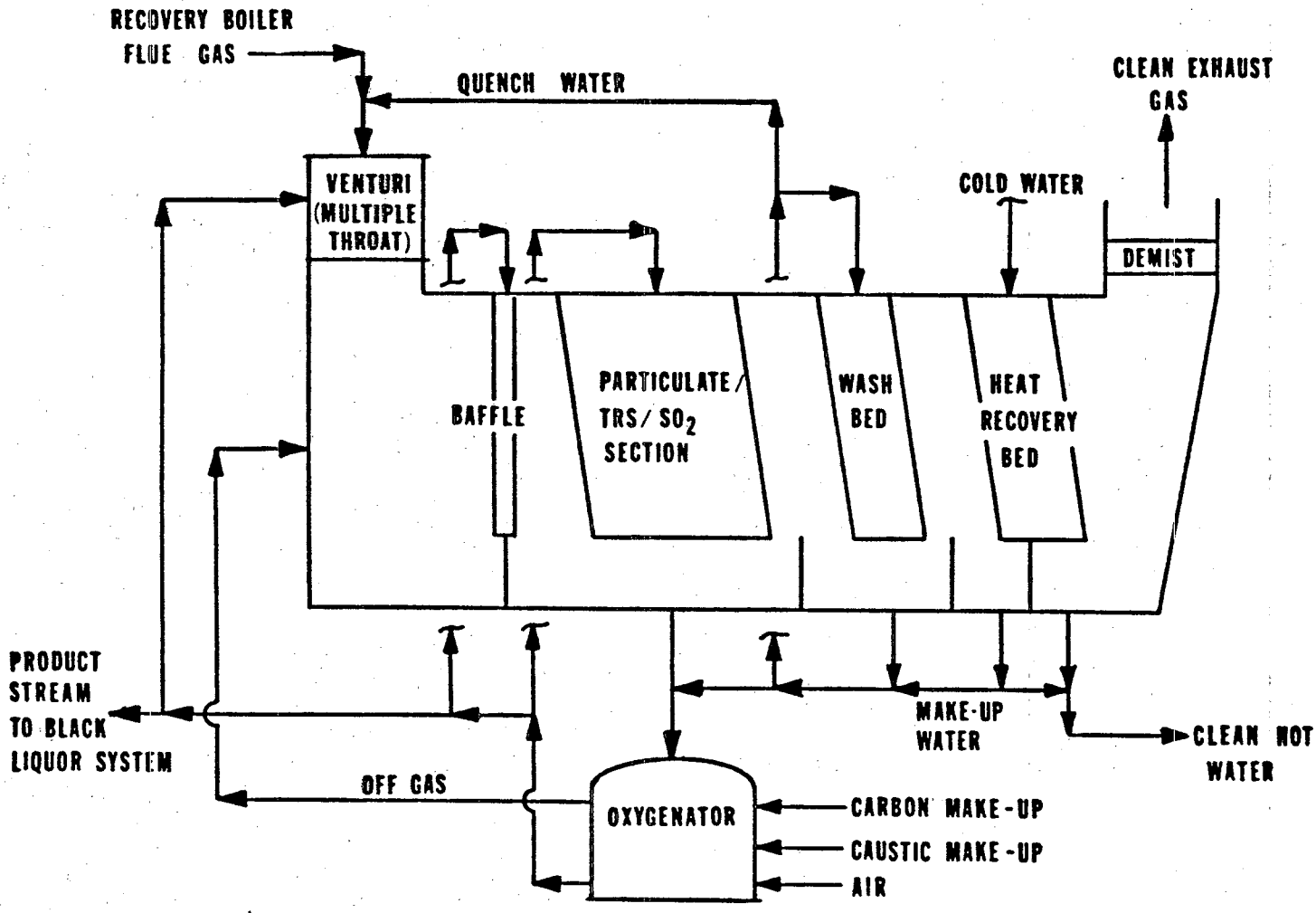


Figure 3-41. Teller System Flow Sheet. 20-30% Total Dissolved Solids. (Teller, 1985b). Reprinted by permission of Research-Cottrell, Somerville, NJ.

The spent slurry is regenerated in a separate vessel by providing oxygen which is adsorbed on the catalyst surface for subsequent reaction in the packed bed. A bleed stream containing approximately 25 wt percent solid sodium thiosulfate ($Na_2S_2O_2$) is removed for disposal or additional processing and sale.

B. H_2S Removal--

The principal steps involved in the removal of H_2S are (a) absorption in the liquid phase, (b) partial adsorption on the surface of the activated carbon, and (c) oxidation.

The most important chemical reactions of the process can be summarized in the following simplified fashion when NaOH is used as the make-up alkali.

$$2NaOH + CO_2 \rightleftarrows Na_2CO_3 + H_2O \qquad (1)$$

$$Na_2CO_3 + CO_2 + H_2O \rightleftarrows 2NaHCO_3 \qquad (2)$$

$$Na_2CO_3 + H_2S \rightleftarrows NaHCO_3 + NaHS \qquad (3)$$

$$NaHCO_3 + H_2O \rightleftarrows NaHS + CO_2 + H_2O \qquad (4)$$

Reactions (1) and (2) govern the (initial) chemical composition and, more importantly, the pH of the scrubing medium. At 12-16 percent by volume of CO_2 in the flue gas, and the typical 70-75 C steady-state temperatures in the scubber, the equilibrium molar proportion of $NaHCO_3$ to Na_2CO_3 is approximately 70:30, resulting in a pH of 9.0 ±0.2.

Reactions (3) and (4) have unfavorable equilibria for efficient absorption of H_2S. The degree of removal of H_2S by the process as a whole depends primarily on an oxidation reaction, which can shift the equilibria of Reactions (3) and (4) towards essentially complete absorption of the H_2S:

$$2NaHS + 2O_2 \xrightarrow{\text{act. carbon}} Na_2S_2O_3 + H_2O \qquad (5)$$

Reaction (5) takes place partly in the absorption stage of the process, in the presence of 2-3 percent oxygen. Pilot studies showed that when activated carbon in concentrations of 0.1-0.4 percent by weight is used as the adsorbent/catalyst, 100-120 ppm of H_2S can be removed with 99 percent

efficiency in the scrubber itself without a separate aeration step. At higher H_2S concentrations, aeration of the scrubbing medium in a separate oxidizer is required before recycle to the scrubber.

The residence (aeration) time and the aeration rate requirements are a function of the inlet gas H_2S concentration as well as on the required H_2S concentration in the exit gas. The latter can be estimated using the following theoretical correlations:

$$p_eH_2S = K[(Na_2S)H^+] \qquad (6)$$

where:

p_eH_2S = equil. partial pressure of H_2S, atm.

$K = 77 \times 10^6$ at 70 C

$[Na_2S]$ = Sodium sulfide concentration, g mol/L of Na_2S equiv.

$[H^+]$ = hydrogen ion con., g mol/L

Equation 6 can be used to develop plots which provide useful information for the design. For a given design, the equation can be used for optimizing the scrubber and the oxidizer.

As indicated by the family of curves in Figure 3-42, a desired H_2S concentration in a scrubber outlet gas can be achieved by increasing the pH (more alkali added) or by reducing the residual sulfide concentration (more intensive aeration). Data from both pilot- and commercial-scale installations indicate that the theoretical correlations represented in Figure 3-42 can be approached very closely in properly designed and operated systems.

As should be apparent from Reactions (3), (4) and (5), the overall efficiency of the process of removing the H_2S from the flue gas depends on providing:

1. Sufficient absorption capacity in the scrubber.
2. Adequate aeration capacity for oxidizing the primary absorption product of H_2S, via., the HS^- ion, into $Na_2S_2O_3$.

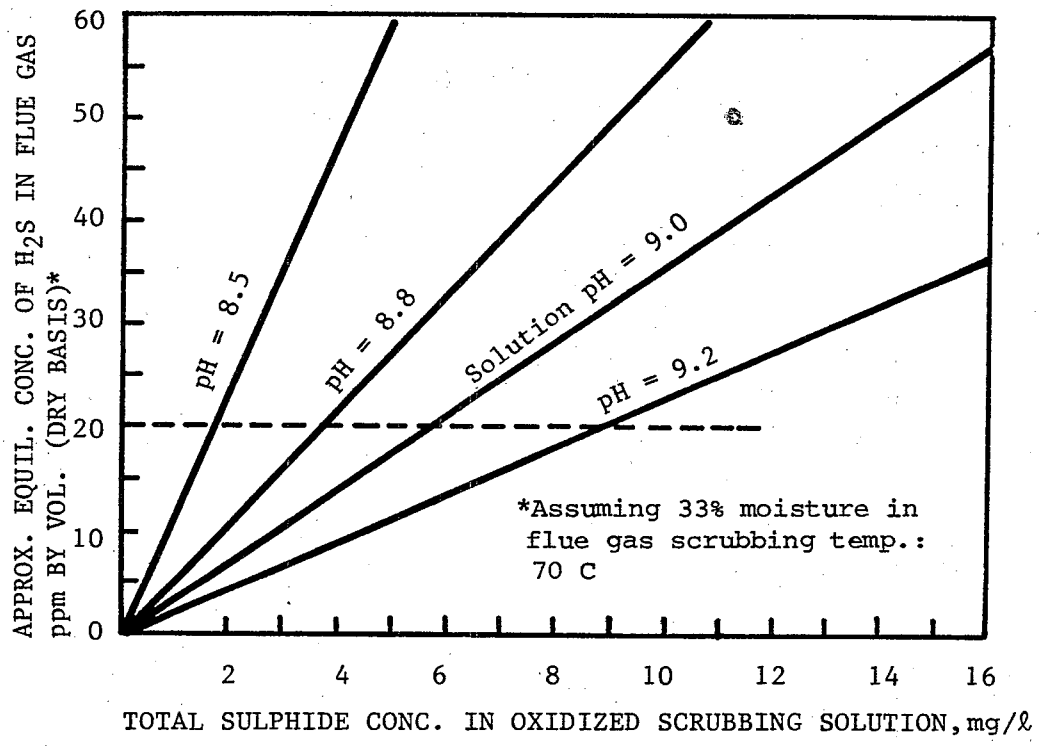


Figure 3-42. Equilibrium H_2S concentration in flue gas, as a function of $S^{=}$ concentration in oxidized scrubbing solution (Prahacs, 1983). Reprinted by permission of the Pulp and Paper Research Inst. of Canada.

Because of the extremely complicated nature of the absorption and subsequent simultaneous liquid phase and catalytic solid-surface-oxidation reactions involved, a universally applicable predictive theoretical model of the overall process has not been developed as yet. The problem of providing reliable predictive models for the three principal physical types and configurations of absorption-oxidation systems tested in a pilot-scale and/or commercial scale is made more difficult by the fact that there is a variable rate and degree of catalyst contamination that occurs when scrubbing real retort gases.

Not withstanding the above problems in arriving at a universally applicable predictive model of the process, sufficient empirical data and correlations were obtained for design purposes.

The following is an illustration of how some of the empirical design parameters were developed. A parameter called the Number of Transfer Units (NTU) for the removal of H_2S in the pilot-scale studies was used as measure of performance and is defined by the following formula:

$$NTU = \ln \frac{pH_2S_i}{pH_2S_o}$$

where:

pH_2S_i = partial pressure of H_2S in the inlet gas

pH_2S_o = partial pressure of H_2S in the outlet gas.

Pilot-scale studies (Prahacs, 1983) were carried out in a turbulent contact absorber with a cross-section of 0.3 x 0.3 m and three to five absorption stages with a height of 1.5 m each. When calculating the apparent NTU/stage for various loads (at fixed gas flows but varying inlet conentrations of H_2S) the values varied between 0.1 to 1.6/stage. However, when correlations between the intensity and efficiency of oxidation vs. the NTU/stage were reexamined, it was found that with adequate oxidation, NTU values of 0.95 to 1.6 could always be achieved. This indication of the oxidation step being the performance-limiting operation has been confirmed in commercial installations.

Within the observed range of NTU's under conditions of adequate oxidation and other variables kept essentially constant, the liquid-to-gas mass flow ratios (L/G) proved to be governing the attainable absorption efficiencies. This can be particularly critical in the case of high but variable inlet concentrations of H_2S to the absorber. Increasing the L/G from 10 to 16 under conditions of 400-1000 ppm of H_2S inlet concentrations can bring about approximately proportional increase in NTU's. In actual practice, packing heights (or travel path through the packing, in the case of the cross-flow scrubbers) in the range of 10-15 ft (providing for gas-liquid contact times of 1-2 seconds), were shown to be capable of absorbing H_2S in concentrations of 600 to 1200 ppm by vol. (dry basis) with $98\text{-}99^+$ percent efficiency. This performance could be achieved only if the other operating parameters, both in terms of the chemical composition of the scrubbing medium and the air supply to the oxidizer, were maintained at appropriate levels.

C. Organic Sulfur Gas Removal--

Laboratory studies and limited pilot-scale tests have been conducted in which a dilute sodium hypochlorite solution was used as a scrubbing agent for efficient "polishing" removal of dimethyl sulfide and various mixtures of organic sulfur compounds (Teller, 1985a). These tests were aimed at having an alternative process configuration (with a last hypochlorite stage) for achieving ≤ 1 ppm total odoriferous sulfur emission levels, regardless of the original inlet conentrations of the organic sulfur compounds. Removal efficiencies of 97-98 percent were obtained for methyl and dimethyl mercaptans at inlet concentrations of 10 to 20 ppm with hypochlorite dosages of 13-18 grams/gram of organic sulfur removal. This approach was found to be technically feasible, although considerably more expensive in terms of chemical cost than the use of alkaline suspensions of activated carbon alone. Also, the possible emission of traces of chlorinated organic compounds might need to be checked before implementing this suplementary processing step.

The removal of CH_3SH is achieved in a fashion similar to H_2S, but the overall reaction mechanism is less well understood. The following represents an illustration of the likely principal reactions involved:

$$CH_3SH + NaHCO_3 \rightleftarrows CH_3\ SNa + SNa + CO_2 + H_2O \qquad (7)$$

$$2CH_3SNa + 1/2O_2 + H_2O \rightarrow (CH_3)_2S_2 + 2NaOH \qquad (8)$$

With inlet concentrations of 2-10 ppm of CH_3SH and 0-5 ppm of $(CH_3)_2S_2$, the general experience in the commercial installations is that both of these organic compounds can be removed nearly quantitatively. This, presumably is due primarily to physical absorption and adsorption of these sulfur compounds in the scrubbing medium, possibly followed by other oxidation reactions, e.g., to sulfonic acid. The removal of $(CH_3)_2S_2$, if present in significant concentrations, is only partial. Again, it is probably by physical absorption in the solution and adsorption on the carbon, with the possibility of oxidation to sulfoxides. The removal efficiency of $(CH_3)_2S_2$ or any of the organic sulfur compounds can be improved, if necessary, simply by the introduction of additional activated carbon.

D. Process Performance--

1. Contactor -- Two types of packed bed scrubbers have been used for his process: a countercurrent absorber and a cross flow absorber (Prahacs, 1983). These contactors are shown in Figure 3-43. The principal difference between the two with regard to removal efficiency is the concentration profile of the caustic-activated carbon slurry through the packed bed.

To remove the H_2S from the retort gas it is first necessary to absorb the H_2S into the liquid phase and then react the H_2S with $CO_3^{=}$ to form HS^-. The physical absorption and reaction of H_2S can be described using the absorption - chemical reaction model described by the equation:

$$K_1a = K_ia\ [1 + C_1\ (q/Cai)]^{0.5}$$

where:

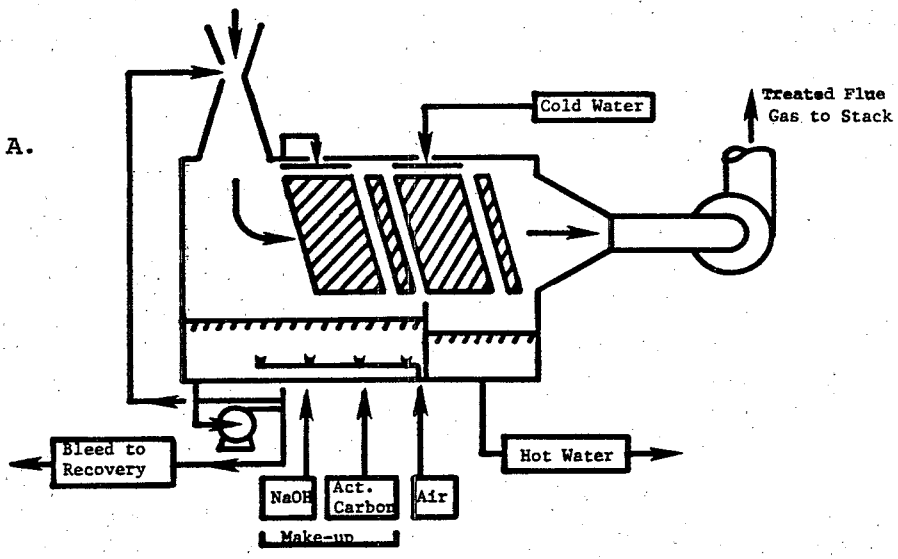
q = concentration of free caustic

C_ai = partial pressure of H_2S in the gas phase converted to concentration in the liquid

K_ia = transfer coefficient for physical absorption

K_1a = transfer coefficient for chemically enhanced absorption

A.



Schematic diagram of one version of the PPRIC/BCRC scrubbing process using a TESI cross-flow scrubber (Prahacs, 1983). Reprinted by permission of the Pulp and Paper Research Inst. of Canada.

B.

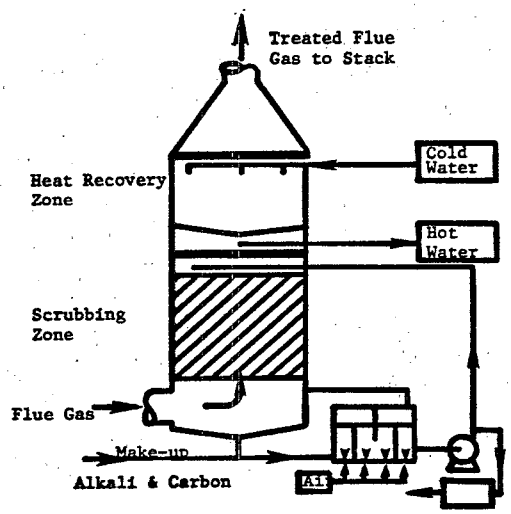


Figure 3-43. Schematic diagram of the PPRIC/BCRC scrubbing process in a packed counter-current scrubber designed by Flakt Canada (Prahacs, 1983). Reprinted by permission of the Pulp and Paper Research Inst. of Canada.

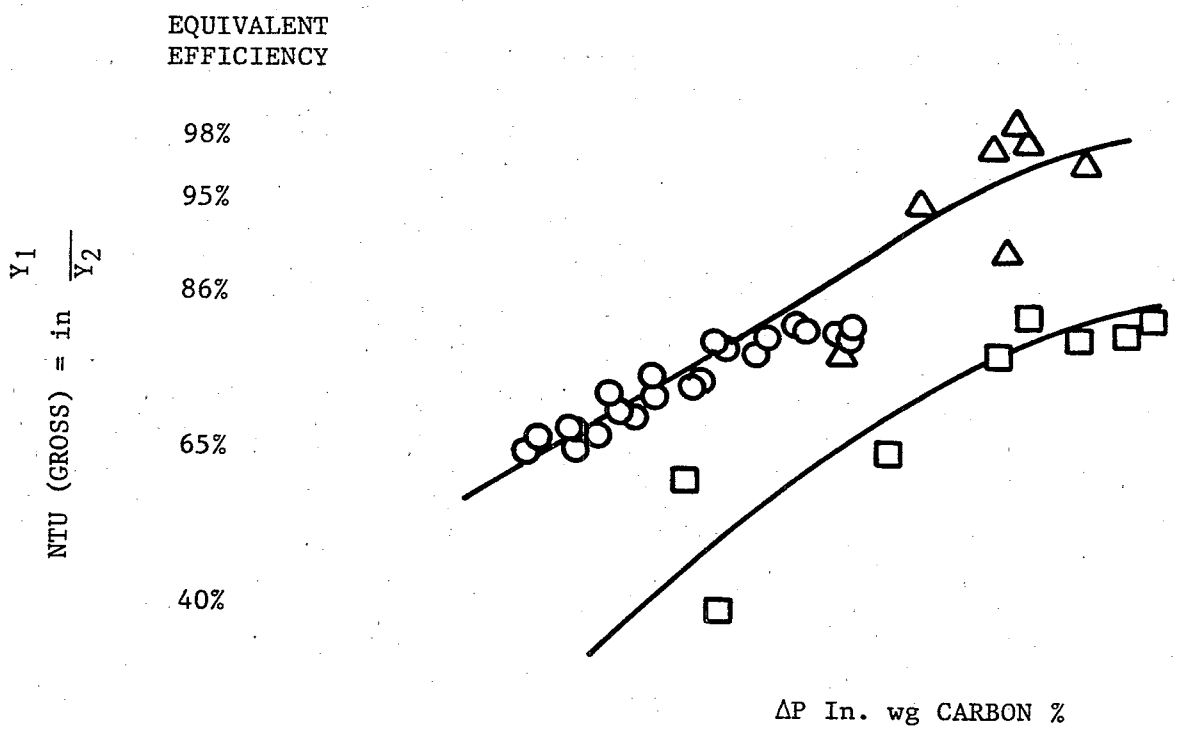
The crossflow contactor has significant advantages over a countercurrent contactor for maximizing the absorption rate of H_2S by controlling the caustic concentration - H_2S gas concentration profile. The countercurrent contactor has a high caustic concentration at the top of the tower where the H_2S concentration is low and the concentration term ($q/C_a i$) is very high. However, at the bottom of the tower, the caustic has been consumed by both the H_2S and CO_2 and the concentration term ($q/C_a i$) approaches zero. The net effect is that the effective K_1a for the tower is the average of a high (top) and low (bottom) value. With the cross flow contactor the caustic concentration can be maintained at a high value for all concentrations of H_2S and the overall K_1a is consistently high for the entire contactor.

In addition to controlling the caustic concentration, the cross flow contactor allows variation of the liquid/gas ratio to provide adequate caustic to regions of the contactor with high H_2S concentrations. Other advantages of the cross flow contactor claimed by the manufacturer are lower pressure drop (less than one-third of countercurrent contactors), stable operation (no flooding), no build-up of solids at gas/liquid support plates, and 30 - 60 percent lower liquid flow rates.

Extensive pilot-plant studies have been conducted by both PPRIC and Teller Environmental Systems (Prahacs, 1983 and Teller and Amberg, 1985). Test results from two of these pilot studies for the crossflow contactor and a turbulent contact absorber (TCA) are shown in Figure 3-44. At 100 ppm inlet H_2S concentration, the crossflow contactor resulted in 98+ percent total sulfur removal efficiency while the TCA had 86 percent removal efficiency (Teller and Amberg, 1985).

The performance of the TESI system at inlet H_2S concentrations ranging from 100 ppmV to 1000 ppmV is presented in Figure 3-44. Additional test results are shown in Figures 3-45 and 3-46 for high (300 - 500 ppmV H_2S) and low (10 - 20 ppmV H_2S) inlet gases. The outlet H_2S concentration is stable at less than 4 ppmV regardless of the initial gas concentration.

2. Activated Carbon Concentration -- In the contactor, the path length for diffusion, and hence the diffusion rate, is determined by the bulk concentration of the activated carbon and the mode of suface renewal which determines the carbon concentration at the boundary layer. The PPRIC patents



○	Boiler #1 - 7.5 ft. Bed Nucleator	~10.	0.03-0.06
□	Bhatia - TCA 5 stages	~18	0.5
△	Boiler #2 -~7.5 ft. Bed Nucleator	~11	0.03 - 0.06

TRS INLET ppm, Y_1

Figure 3-44. Kraft recovery boilers TRS removal data (Teller, 1985a). (Reprinted by permission of Research-Cottrell, Somerville, NJ)

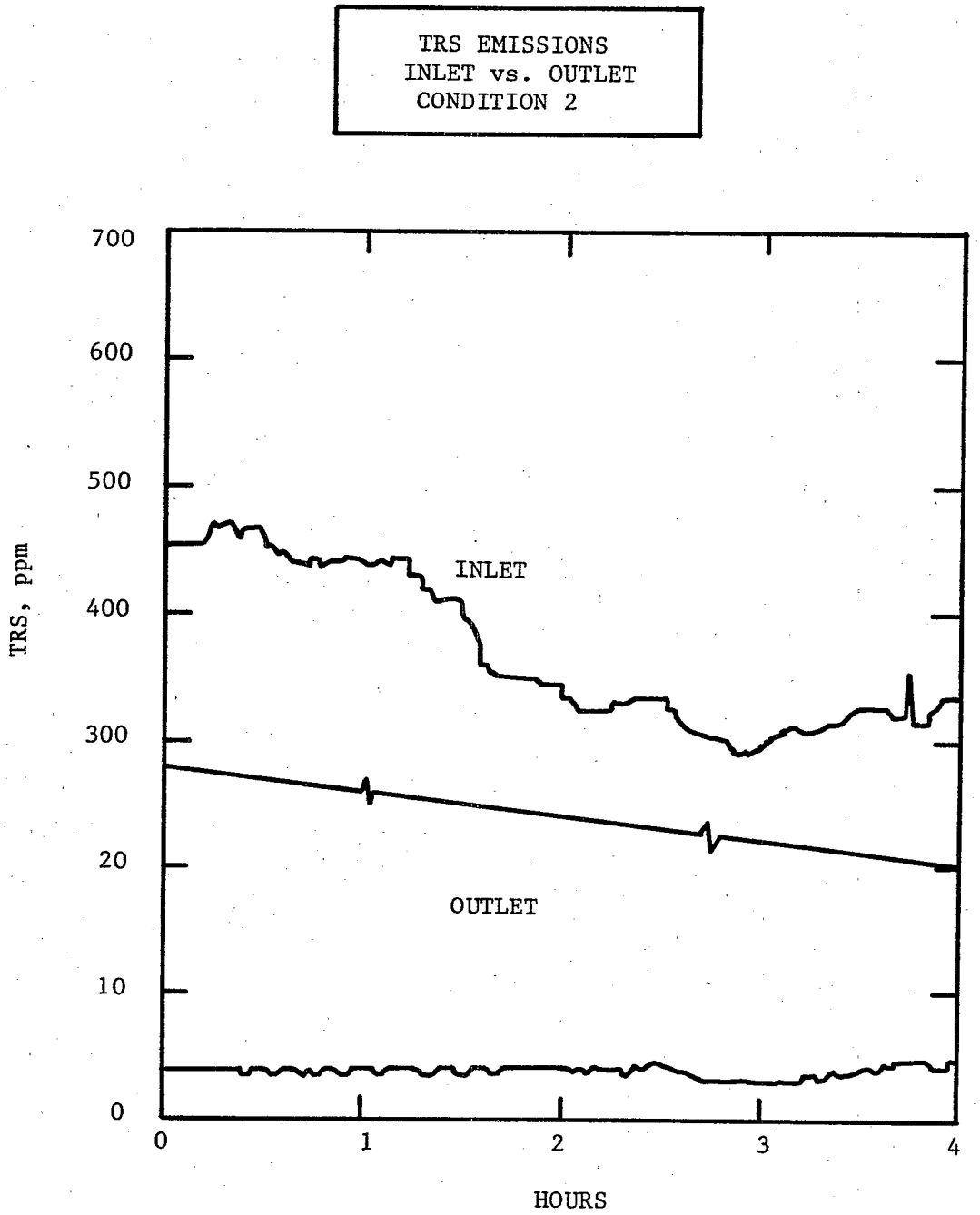


Figure 3-45. TRS emissions inlet vs. outlet condition 2. (Teller, 1985a). (Reprinted by permission of Research-Cottrell, Somerville, NJ)

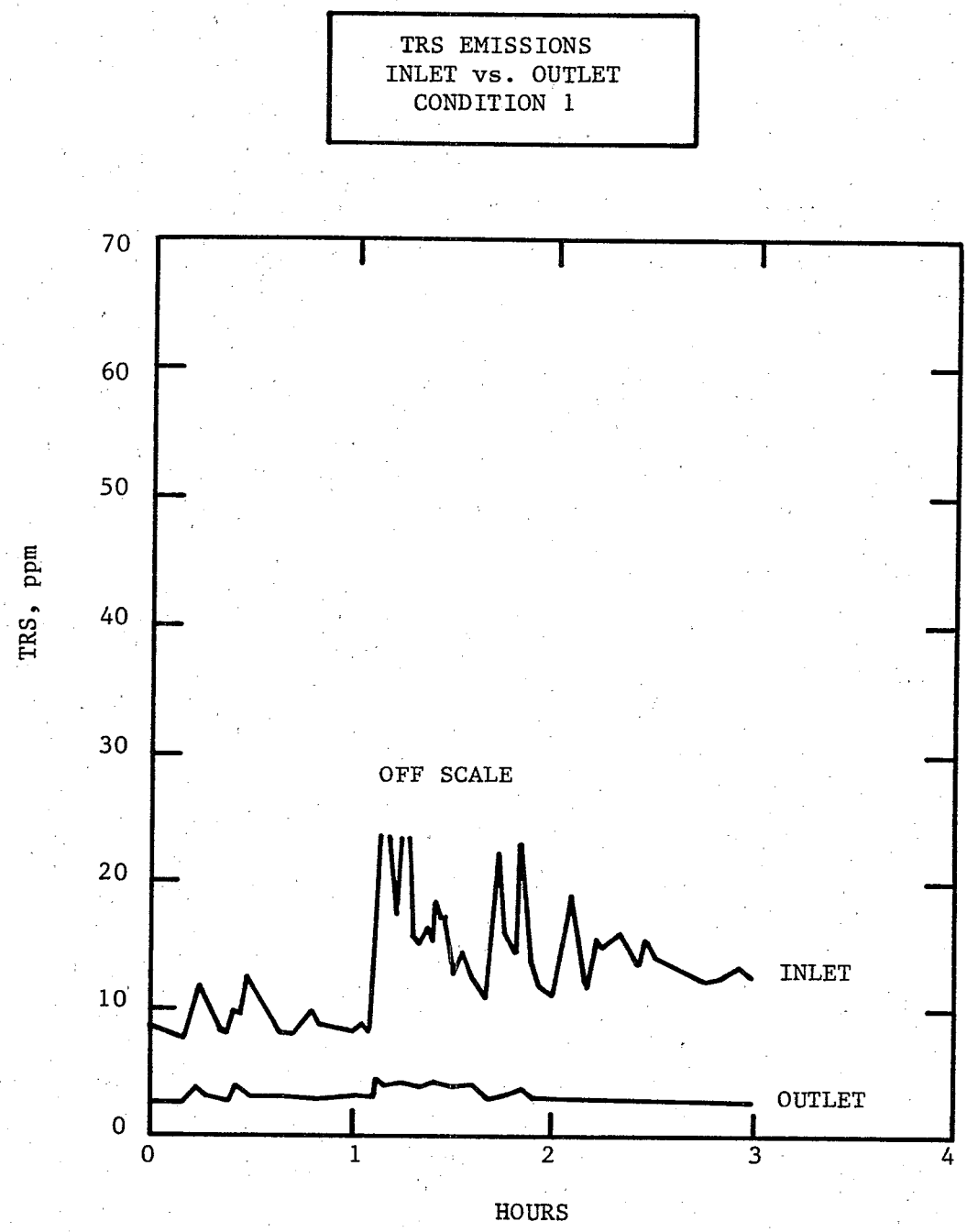


Figure 3-46. TRS emissions inlet vs. outlet condition 1 (Teller, 1985a). (Reprinted by permission of Research-Cottrell, Somerville, NJ)

using a countercurrent contactor call for 0 - 10 percent carbon and laboratory data indicate 0.5 percent is sufficient. Tests conducted with the TESI cross flow scrubber indicated at 0.05 to 0.10 percent is sufficent due to the effective laminar surface renewal that is inherent in the crossflow scrubber.

The bulk concentration of carbon and surface renewal characteristics also determine the kinetics of adsorption of H_2S or the sodium salt (NaHS) on the carbon surface.

The effect of the carbon concentration on removal efficiency is shown in Figure 3-47. There is an increase in removal efficiency from 97.2 percent to 99.1 percent by increasing the carbon concentration from the range of 0.03 - 0.07 percent to 0.08 - 0.13 percent.

3. Particulates -- The TESI system can also include two means for particulate control. The first is a dry venturi scrubber and the second is a wet nucleation scrubber. The reader is referred to Section 3.1.1D for a description of the dry venturi system.

The particulate collection of a crossflow scrubber is due to two factors. First, the dry venturi increases the size of the particulates by using solid particles (i.e. 20 μm) as the targets for the small particles (i.e. 0.5 μm) resulting in large particles in the gas stream to the scrubber rather than the initial small ones. Consequently, an expected result would be to increase the particulate removal of any scrubber since removal efficiency is directly related to particle size.

In addition to the scrubbing action of the gas-liquid contact, when the gas stream contacts the liquid, it is cooled to the adiabatic saturation temperature and then proceeds to condense water. The condensation process is generally aided by the presence of condensation sites and the large particles from the dry venturi provide these condensation (or nucleation) sites. Table 3-32 shows performance data for commercial systems utilizing the nucleation process for particulate control. These systems do not have a dry venturi upstream and are processing the small diameter particles indicated (from 0.04 to 10 μm).

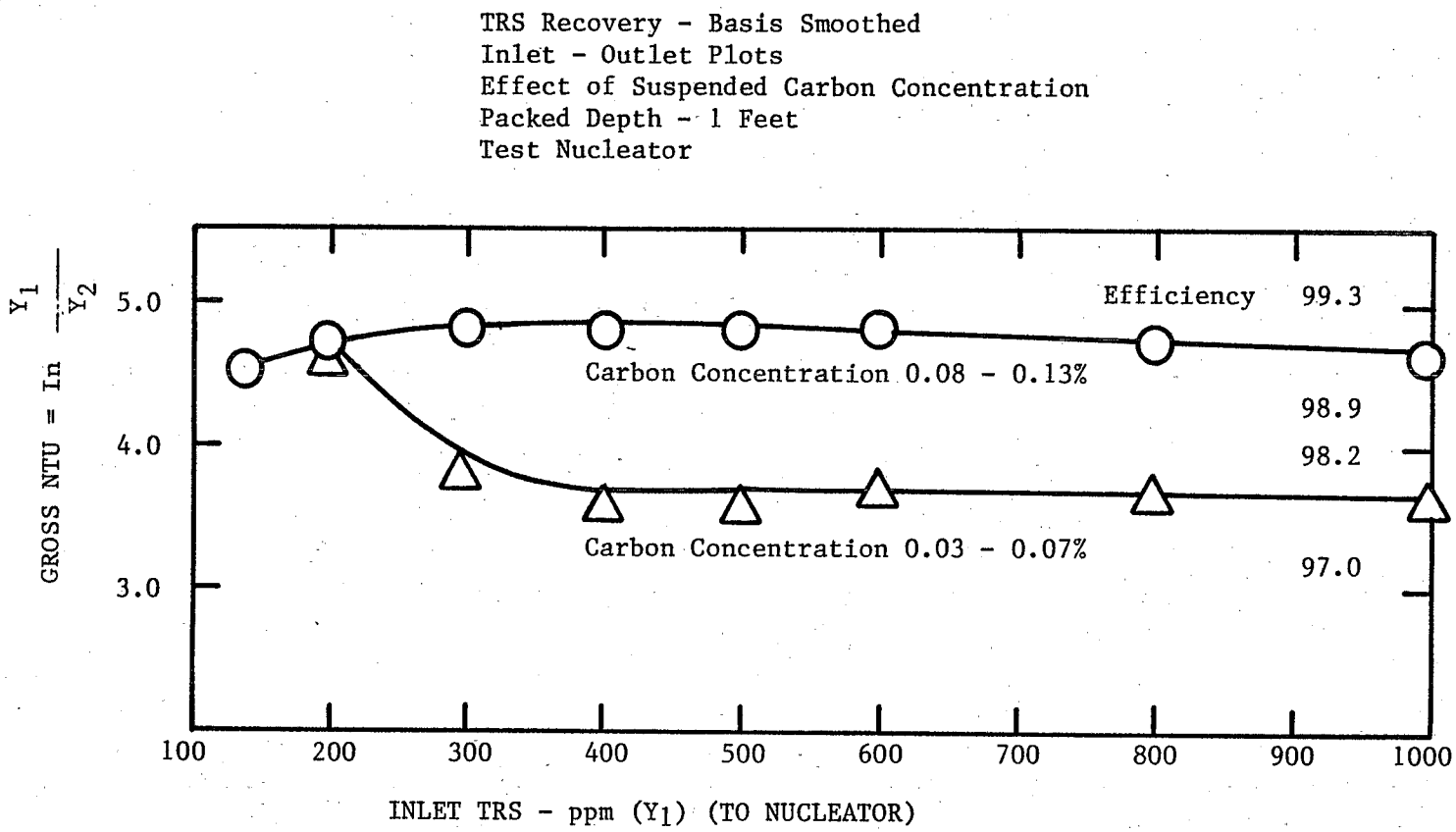


Figure 3-47. Kraft boiler with direct contact evaporation (Teller, 1985a). (Reprinted by permission of Research-Cottrell, Somerville, NJ)

TABLE 3-32. COMMERCIAL NUCLEATION PROCESS APPLICATIONS

(Teller Environmental Systems, Inc.)

(Teller, 1979)

	Gas Inlet Conditions							Gas Outlet		
			Particulate			Gases				
Application	Flow ACFM	Temp. °F DB/DP	Type	gm/ncm	Size Microns	Type	Conc. ppm	Part. gm/ncm	Gas ppm	System ΔP in w.g.
Recovery Boiler SO_2 - Partic.	200,000	300/160	Hydrophilic	0.8-2	0.2-3	SO_2	1500	0.06-0.13	<1	11
Recovery Boiler TRS - Partic.	150,000 -300,000	300/160	Hydrophilic	0.7-3.5	0.2-3	TRS SO_2	<1500 <300	0.03 -0.07	<10 <3	13
Fertilizer HF,SiF - Partic.	20,000 -100,000	170/150	Hydrophilic	0.05-0.3	0.05-3	HF SiF	<1500 <1500	0.01	<3 <1	6
Jet Engines SO_2 - Partic.	500,000- 3,000,000	1200	Hydrophobic	0.02 -0.3	0.04- 0.2	SO_2	<50	0.004 -0.009	<3	2
Org. Chem. Inc. HC_x - Partic.	150,000	1500	Hydrophilic Hydrophobic	0.1- 0.6	0.02 -10	HC_x	2000	<0.02	<10	6

The kinetics of isothermal nucleation is highly dependent on the population density of the water molecules with a desirable level above 25 percent by volume.

4. Operating Experience -- Teller Environmental Systems, Inc. (TESI) has installed eleven full scale systems for control of H_2S from black liquor recovery boilers (Teller, 1985b). The operating performance for these eleven systems is presented in Table 3-33.

The TRS outlet conentrations range from 3 to 43 ppm with the majority of installation having exit concentrations at 10 ppm or less. The exit particulate concentrations range from 7 to 730 mg/Nm^3 with most of the facilities having exit concentrations less than 130 mg/Nm^3.

5. Costs -- The cost for a full scale TESI system has been estimated at \$8.40/kg-hr of gas (1984 dollars) based on a gas flow rate of 5×10^6 kg/hr (Teller, 1985b).

3.3.6 Claus Process (H_2S)

The Claus process has been used extensively for four decades in refineries, gas plants and coking operations where large quantities of H_2S are produced. The feed stream to the Claus plant should be at least 25 percent H_2S although reliable operation is possible at as low as 15 percent concentration and with great effort even lower. The plants typically are capable of 95 percent elemental sulfur recovery.

The EPA Control Technology Appendices for Pollution Control Technical Manuals [Ondich, 1983] provides a comprehensive discussion of this process including many important design and application factors. A summary of that discussion is presented here.

A. Process Description--

The Claus sulfur removal process is a vapor-phase, dry , high temperature process in which hydrogen sulfide-rich gases from acid gas removal processes are catalytically reacted with sulfur dioxide (produced by air oxidation of hydrogen sulfide) to form elemental sulfur. Modern Claus units produce a high-quality, salable sulfur, as well as steam. The Claus unit offgas generally contains several thousand ppmv sulfur and may require further processing to reduce emissions of sulfur species (e.g., SCOT, Beavon, or Wellman-Lord processes).

TABLE 3-33. TELLER RECOVERY BOILER SYSTEM PERFORMANCE
(Teller, 1985b)

Installation	ACFM	m^3/s	DBT/DO	Inlet PPMDV			Outlet PPMDV		
				TRS	SO_2	Part. mg/Nm^3	TRS	SO_2	Part. mg/Nm^3
1	125,000	60	161	30-80	0.4	690-1150	3-10	---	46
2	186,940(1)	90	270/150	17	1500	1830-2520	<5	<1	68-137
3	205,000(6)	100	162	105-250	---	---	2-10(3)	---	---
	210,240(7)	100	---	255-970	---	---	7-43	---	---
4	319,000	150	300/160	240-450	100-250	1600-10300	3-20	<1	30
5	169,400	80	165	250	50	1830-9200	5-10	1	114
6	275,800(2)	130	225/147.5	8-20	---	1370	4-11	---	34
7	154,100	70	300/164	N/A	N/A	5700	N/A	N/A	450-730
8	325,000	150	410/154	200-1200	100	N/A	5	0	7
9	244,200	120	296/160	N/A	N/A	2380-3440	N/A	1	40
10	387,000(5)	180	---	---	---	---	---	---	46
11	350,000(5)	170	---	---	---	---	---	---	---

(1) No oxygenation
(2) No carbon
(3) At design liquid rates
(4) Test conducted during precipitator malfunction inlet 2-4 GR/DSCF
(5) Startup - 1983
(6) Softwood
(7) Hardwood
(8) No Venturi
N/A - Not available

A typical Claus unit (Figure 3-48) consists of a thermal reaction furnace and one or more catalytic stages, depending on the desired level of sulfur in the treated tail gas. The overall Claus reaction is given by:

$$3H_2S + 1.5O_2 = 3H_2O + 3S$$

The overall Claus reaction actually takes place in two steps. Sulfur dioxide is formed by oxidation of a portion of the hydrogen sulfide in the thermal stage of a Claus unit:

$$H_2S + 1.5O_2 = SO_2 + H_2O$$

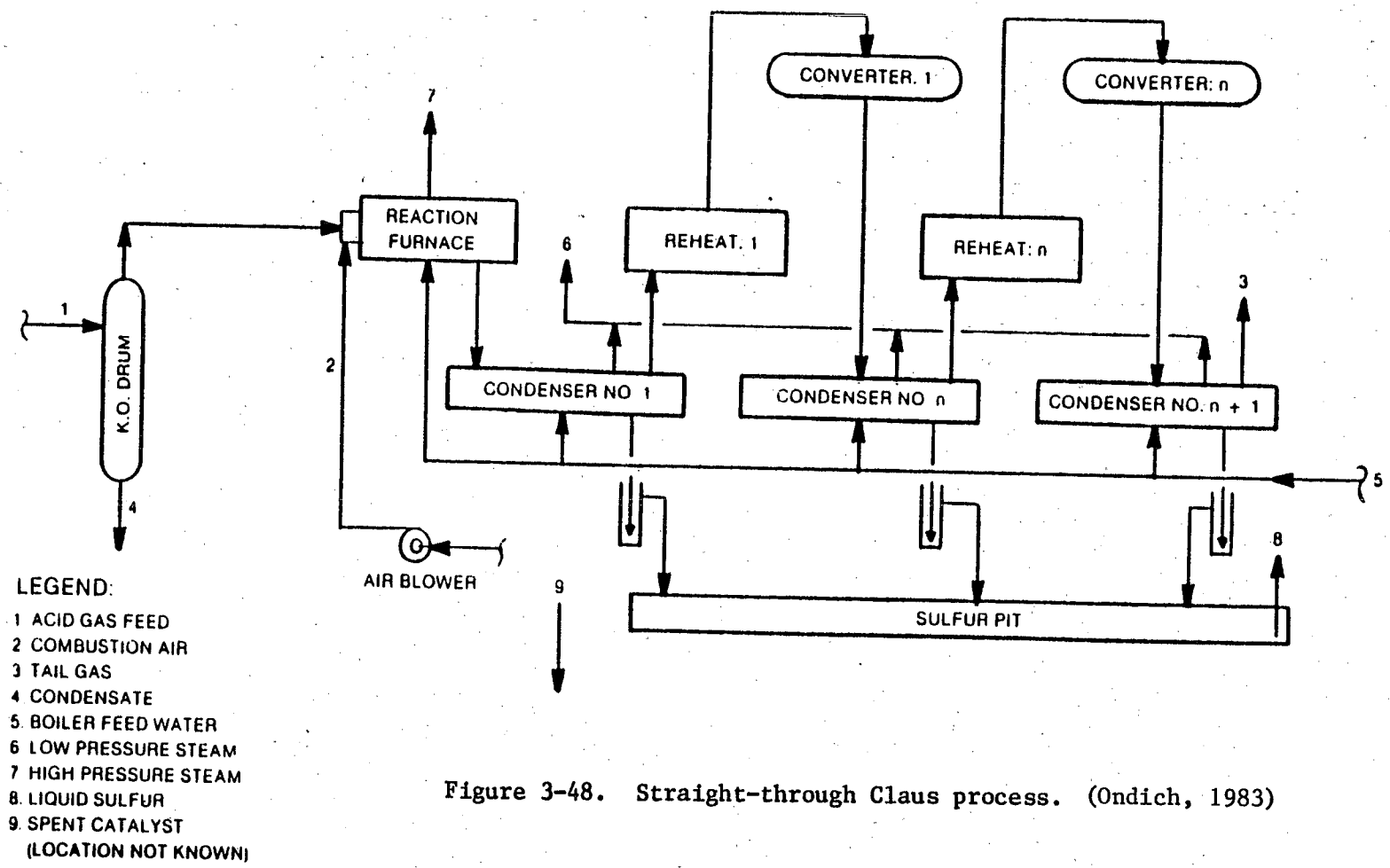
The prevailing combustion conditions (1370 to 1800 K and pressure slightly above atmospheric) also result in the formation of sulfur vapor from hydrogen sulfide and sulfur dioxide with a 50 to 70 percent yield.

$$2H_2S + SO_2 = 3\ S\downarrow + 2H_2O$$

The process gas is cooled in a steam-producing waste heat boiler and the sulfur is condensed and removed. Additional formation of sulfur from the remaining hydrogen sulfide and sulfur dioxide is promoted by either bauxite or alumina catalyst in one or more catalytic stages. Each catalytic reactor is preceded by a process gas reheat unit and followed by a sulfur condenser.

The Claus process is controlled by regulating the flow of combustion air to the reaction furnace so that only about one-third of the hydrogen sulfide is oxidized to sulfur dioxide. This provides the optimum quantity to react with the hydrogen sulfide to produce elemental sulfur. However, high carbon dioxide contents (such as those encountered in shale oil retort offgas often dilute the acid gas to the point where stable combustion of one-third of the hydrogen sulfide is very difficult. Therefore, Claus plant designs are generally based on one of three combustion configurations, depending on the carbon dioxide concentration in the feed gas: "straight-through," "split-stream," and the "sulfur burning" mode.

For carbon dioxide levels below about 30 percent by volume (corresponding to hydrogen sulfide levels above 40 percent by volume), the "straight-



LEGEND:

1 ACID GAS FEED
2 COMBUSTION AIR
3 TAIL GAS
4 CONDENSATE
5 BOILER FEED WATER
6 LOW PRESSURE STEAM
7 HIGH PRESSURE STEAM
8 LIQUID SULFUR
9 SPENT CATALYST (LOCATION NOT KNOWN)

Figure 3-48. Straight-through Claus process. (Ondich, 1983)

through" process is generally used (i.e., the entire feed gas stream is sent to the Claus furnace, Figure 3-48). At carbon dioxide concentrations above about 30 percent by volume (corresponding to hydrogen sulfide concentrations below about 40 percent by volume), free flame combustion with stoichiometric air becomes unstable and the "split-stream" process must be used. In this design approximately one-third of the acid gas is completely oxidized to form sulfur dioxide and the remaining two-thirds is bypassed around the combustion chamber. When the acid gas contains less than 5 to 10 percent by volume hydrogen sulfide (corresponding to very high carbon dioxide levels) combustion temperatures are too low to sustain the thermal sulfur reactions and side reactions become troublesome, especially those involving hydrocarbons.

At very low hydrogen sulfide levels, the "sulfur burning" mode may be used, in which liquid sulfur is recycled to the combustion chamber to supply heat and sulfur dioxide for reaction with hydrogen sulfide downstream in the process.

The presence of hydrocarbons, carbon dioxide, ammonia, and other impurities in the feed complicate the reduction process with the formation of carbonyl sulfide and carbon disulfide in the combustion zone and the partial hydrolysis of thes compounds to hydrogen sulfide in the catalytic reactors. Carbonyl sulfide and carbon disulfide are slow to react over the Claus catalyst and, therefore, represent sulfur that is not totally recovered as elemental sulfur. The presence of ammonia in the feed gas gives rise to secondary reactions which may disturb the operation of the units. There are limited data available on the maximum allowable concentration of these various contaminants. However, decreased sulfur removal efficiencies appear to be due to decreased hydrogen sulfide concentrations rather than the presence of impurities in the acid gas feed. The presence of hydrogen cyanide in the acid gas can lead to excessive equipment corrosion and catalyst deactivation via formation of thiocyanates.

B. Process Applicability--

More than 15 million tons of sulfur are recovered annually from treatment of acid gases by the Claus process. There are at least 170 Claus plants in the U.S. which are used in a wide variety of industries including petroleum, natural gas, and coke production. Applicability of the Claus

process to coal conversion process gas purification systems has not been entirely established. A Claus plant is featured in the design of the Hygas pilot plant at Chicago, Illinois for processing the acid gas stream from a diglycol amine unit. A number of firms are experienced designers of Claus plants in the U.S. including: Black, Sivals & Bryson, Inc.; Ford, Bacon & Davis, Inc.; and the Ralph M. Parsons Company.

C. Process Performance--

Modern Claus units are capable of greater than 95 percent sulfur recovery. However, the sulfur removal efficiency is dependent on certain factors such as: the number of catalytic conversion stages, the inlet feed stream composition, the operating temperatures, catalyst maintenance procedures, maintenance of the proper stoichiometric ratio of hydrogen sulfide to sulfur dioxide, and operating capacity factor.

The Claus reaction is reversible and is limited by chemical equilibrium. Sulfur recovery is enhanced by removing heat since the desired reactions in the Claus process are exothermic. Operation at the lowest possible temperature in the catalytic reactors without condensing sulfur vapor on the catalyst is desired, but adequate reaction rates must be maintained. The percentage of sulfur recovery increases with the number of catalytic stages and the acid gas feed hydrogen sulfide concentration as indicated in Table 3-34. The addition of a fourth catalytic stage can give recoveries above 99 percent. However, the Claus reaction rate is significantly reduced at high sulfur conversion because of the lower temperatures imposed by thermodynamics and, especially, the lower hydrogen sulfide and sulfur dioxide concentrations. Claus plants in the U.S. generally have two or three catalytic stages.

TABLE 3-34. SULFUR RECOVERY VARIATION RELATIVE TO ACID GAS FEED COMPOSITION AND NUMBER OF CATALYTIC STAGES (Ondich, 1983)

Mole Percent H_2S In Acid Gas Feed	Number of Catalytic Stages	Percent Sulfur Recovery
15	2	90
50	2	93
90	1	85
90	2	94
90	3	97

Sulfur recovery is also dependent upon catalyst performance. Very active catalysts which resist aging are required to maintain fast reaction rates, especially at lower temperatures. The formation of sulfate on the surface of this catalyst is the most significant cause of activity loss, but inorganic or organic deposits also lead to catalyst deactivation. Because sulfate formation (called sulfation) is limited by equilibrium conversion, it increases with low hydrogen sulfide concentrations, high sulfur trioxide, high oxygen concentrations, and low temperatures. Thus, a catalyst in the second and third catalytic conversion stages is more susceptible to deactivation by sulfation, even though these later stages require a more active catalyst to maintain reasonable reaction rates at the low temperatures. Catalyst life generally varies from 2 to 5 years depending on plant operation and contaminants in the feedstock.

Deviation above or below the 2:1 stoichiometric ratio of hydrogen sulfide to sulfur dioxide results in a loss of Claus plant efficiency. At less than stoichiometric amounts of air, the sulfur yield falls due to reduced flame temperatures. At greater than stoichiometric amounts of air, it declines due to oxidation of elemental sulfur. Operation of a Claus plant below capacity may also impair sulfur recovery efficiency. Although the impact of these process variables represents, in most cases, a loss of only 1 to 3 percent efficiency, at the relatively high efficiencies typical of Claus

operations (95 percent), a 1 to 3 percent efficiency loss represents a 20 to 60 percent increase in uncontrolled sulfur emissions.

Actual operating data are shown in Table 3-35. The treated gas from the Claus process generally contains several thousand ppmv (dry basis) of sulfur which is present as hydrogen sulfide, sulfur dioxide, carbonyl sulfide, carbon disulfide, and sulfur vapor. Depending on the exact nature of the acid gas feed and the operating conditions, combined carbonyl sulfide and carbon disulfide levels as high as 5000 ppmv may exist in the tail gas.

The "split-stream" process efficiently suppresses the formation of carbon sulfides. However the primary reaction products in the split-stream process are sulfur dioxide and water with virtually no elemental sulfur being formed (Ondich, 1983). This can be a serious limitation since an FGD system would be required to capture the sulfur and prevent air pollution.

D. Process Reliability--

Available information indicates no special maintenance problems or unusually hazardous conditions created by the Claus process. Principal problems result from lack of feed or from upsets in feed rate and composition caused by operating problems in upstream units. Plant shutdowns generally are due to plugging problems. Carryover of hydrocarbons present in the acid gas feed may lead to coke formation in the catalyst beds. High concentrations of carbon dioxide and ammonia in the acid feed gas (e.g., greater than 30 percent carbon dioxide and 500 ppmv ammonia) can result in precipitation of ammonium salts through the Claus unit. Hydrogen cyanide in the acid gas can produce high molecular weight compounds that plug the Claus unit even more severely than the ammonium salts. Hydrogen cyanide also can lead to excessive equipment corrosion.

E. Process Economics--

The cost of a Claus plant varies as a function of the acid gas hydrogen sulfide content and the daily sulfur production capacity. The presence of inert diluents (e.g., carbon dioxide, nitrogen, and air) and combustible contaminants (e.g., ammonia, hydrocarbons, carbon monoxide, hydrogen cyanide, carbonyl sulfide, and carbon disulfide) increase the initial investment cost of the plant. The combustible impurities also affect the operating costs.

TABLE 3-35. REPORTED CLAUS PLANT OPERATING DATA
(Ondich, 1983)

	Plant A[a]	Plant B[a]	Plant C[b]	Plant D[a]	Plant E[c]	Plant F[b]	Plant G[a]	Plant H[d]
Daily Sulfur Production 10^3 kg)	360	230	120	410 (4 units)	N.R.*	185	90	N.R.
Number of Catalytic Stages	2	2	3	2	3	3	2	2
Process Variation	Straight-Through	Straight-Through	Straight-Through	Straight-Through	Straight-Through	Straight-Through	Straight-Through	Split-Stream
Acid Gas Feed Source	CS_4 Plant	CS_4 Plant	Refinery	Refinery	Coke Oven	Sulfinol Process	Refinery	Gas Plant
Acid Gas Feed Quantity (Nm^3/hr)	12,350	6,640	2,570	7,200	N.R.	8,490	4,500	N.R.
Acid Gas Feed Composition (mol %)								
B_3S	96.3	95.5	90	83.5	75.11	68	50.3	8.79
N_2	0.2	0.4		0.4	3.48		1.0	0.22
CO_2	0.2	0.6	8	11.5	18.20	30	46.1	90.67
CH_4	3.2	3.4	2	0.5	2.07	2	2.5	0.32
H_2O				4.0				
CH_2	0.2	0.5			0.29			
NH_3				0.1				
HCN					0.45			
SO_2					0.05			
H_2					0.33			
Tail Gas Quantity (Nm^3/hr)	33,380	21,220	7,260	40,710	N.R.	15,630	10,050	N.R.
Tail Gas Composition (mol %)								
H_2S	0.9	1.0	1.0	0.8	0.22	0.40	0.7	0.20
SO_3	0.4	0.5	0.5	0.4	0.13	0.24	0.3	0.12
S	0.3	0.3		0.3			0.2	
CO_2	1.1	1.3	3.5	4.4	11.57	17.5	21.8	79.18
N_2	64.2	64.3	65.0	60.2	87.40	79.0	51.5	19.67
H_2O	33.1	32.6		33.9			25.5	
CS_2					0.11			Trace
COS					0.15			0.03
H_2					0.42			0.41
CO						2.5		0.10
CH_4								0.28
Sulfur Conversion Efficiency, %	95.5	94.0	95.3	90.0	98.5	98.3	94.7	95.6

[a] Stauffer Chemical Co. (4)
[b] Shell Oil Co. (4)
[c] Bethlehem Steel Corp. (3), data are given after incineration
[d] Home Oil Co. (Personal communication with C.E. Loiselle, December 13, 1979)
[e] Not reported

The Claus capital investment costs, as a function of both hydrogen sulfide concentration in the Claus feed gas and the kg moles of sulfur removed in the Claus process, are presented in Figures 3-49 and 3-50. Figure 3-49 is based on data for Claus feed streams containing 15 and 50 volume percent hydrogen sulfide and includes installed costs only. Figure 3-50 is an extension of Figure 3-49 and presents total capital costs, which include installed costs as well as contingency costs, engineering design costs, and contractor's fees.

The utility and labor requirements estimated for the Claus process are presented in Table 3-36.

TABLE 3-36. OPERATING REQUIREMENTS FOR THE CLAUS PROCESS (Ondich, 1983)

Utilities	
Steam (generated), 10^3 kg/kg mole sulfur removed	0.10 at 0.7 MPa 0.01 at 4 MPa
Boiler feedwater, m^3/kg mole sulfur removed	0.125
Electric power, kW/kg mole sulfur removed	3.0
Process Materials	
Chemicals, $ (first quarter 1980)/kg mole sulfur	0.012
Manpower	
Labor, manhours/yr	1577

3.3.7 SCOT (Shell Claus Offgas Treating) Process (H_2S)

The SCOT process is designed to reduce the emissions of sulfur species from Claus plant tail gas. The system process is discussed in Section 3.3.6. The EPA Control Technology Appendices for Pollution Control Manuals

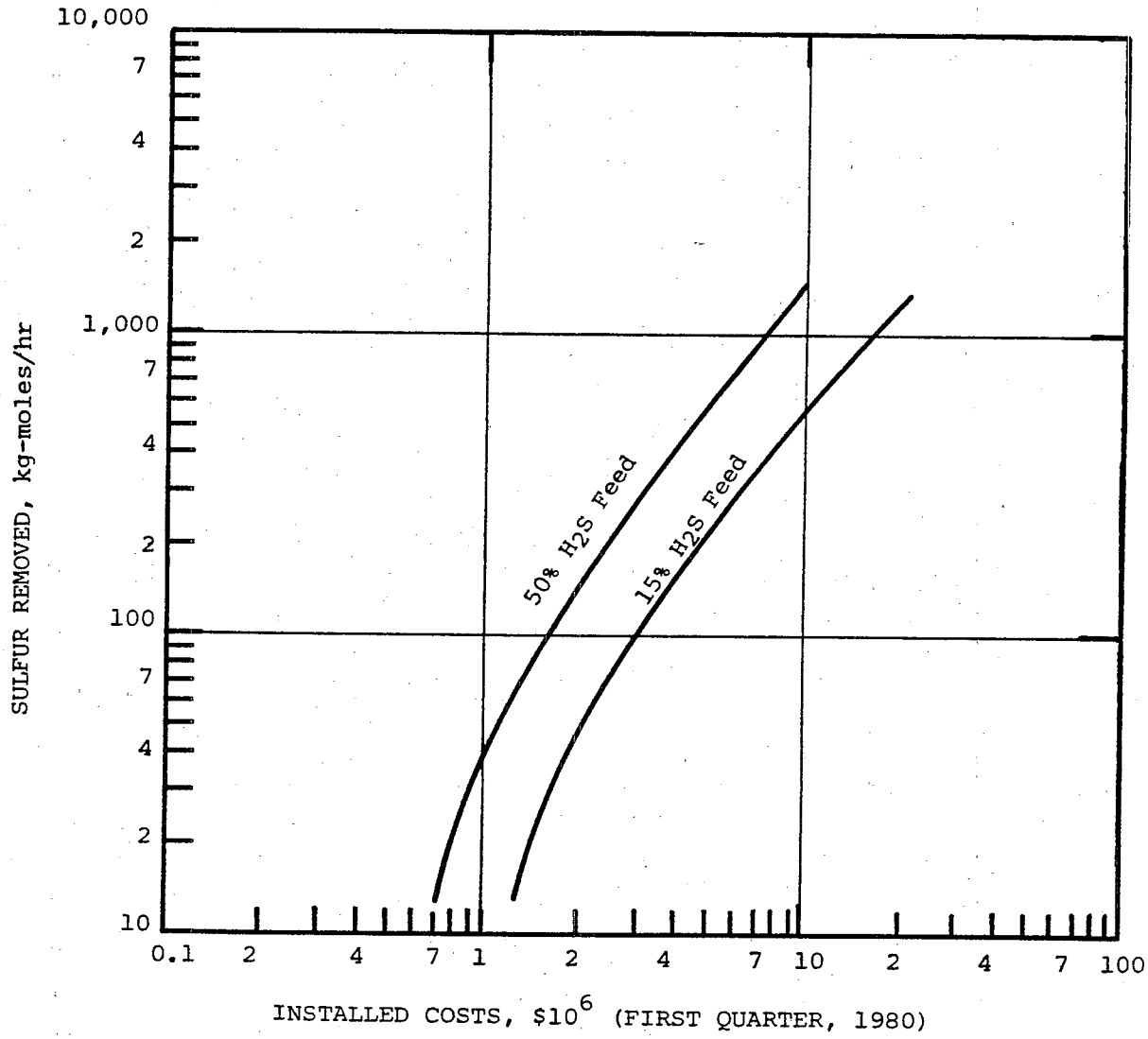


Figure 3-49. Claus plant installed costs (Ondich, 1983).

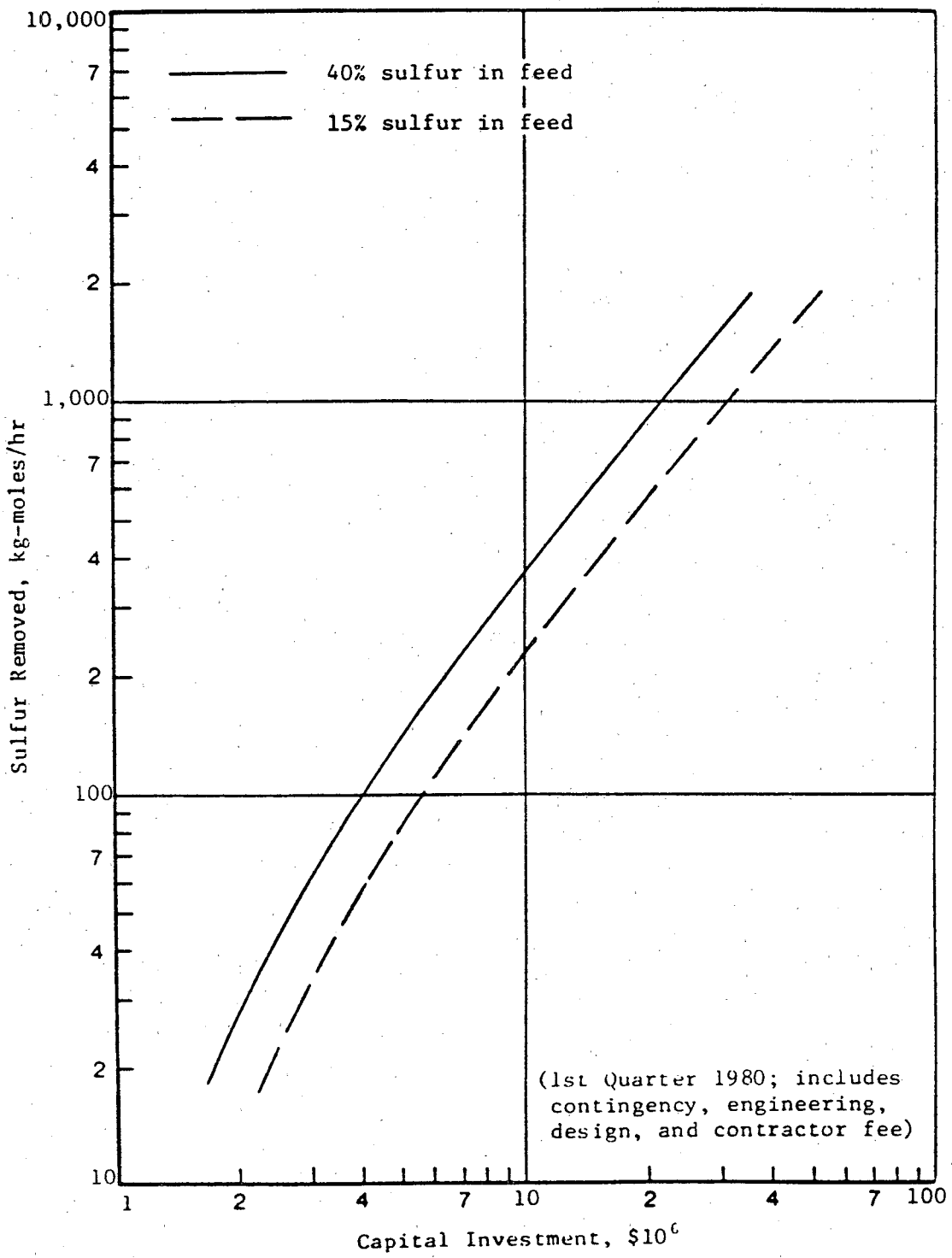


Figure 3-50. Estimated Claus plant capital investment cost (Ondich, 1983)

[Ondich, 1983] covers this process detail. That information is summarized in the following subsections.

A. Process Description--

A typical flow diagram for the SCOT process is presented in Figure 3-51. The process essentially consists of two sections: a reduction section and an alkanolamine absorption section. In the reduction section, all sulfur compounds and any free sulfur in the Claus offgas are converted into hydrogen sulfide over a cobalt-molybdate catalyst at a temperature of about 570 K in the presence of a reducing gas. The reducing gas can be hydrogen, a mixture of hydrogen/carbon monoxide mixtures of hydrogen/hydrocarbon mixtures supplied from an outside source, or can be generated by substoichiometric combustion of a fuel gas in the direct heater preceding the reduction reactor.

In the presence of hydrogen and steam, the following reactions take place in the reduction reactor:

$$SO_2 + 3H_2 = H_2S + 2H_2O$$

$$S_3 + 8H_2 = 8H_2S$$

$$COS + H_2O = CO_2 + H_2S$$

$$CS_2 + 2H_2O = CO_2 + 2H_2S$$

When carbon monoxide also is present as a reducing agent, the following additional reactions may occur:

$$SO_2 + 3CO = COS + 2CO_2$$

$$8S + 8CO = 8COS$$

$$CO + H_2O = CO_2 + H_2$$

$$CO + H_2S = COS + H_2$$

The reactor effluent is subsequently cooled in a heat exchanger and a quench tower. The excess condensate from the quench, which contains a small

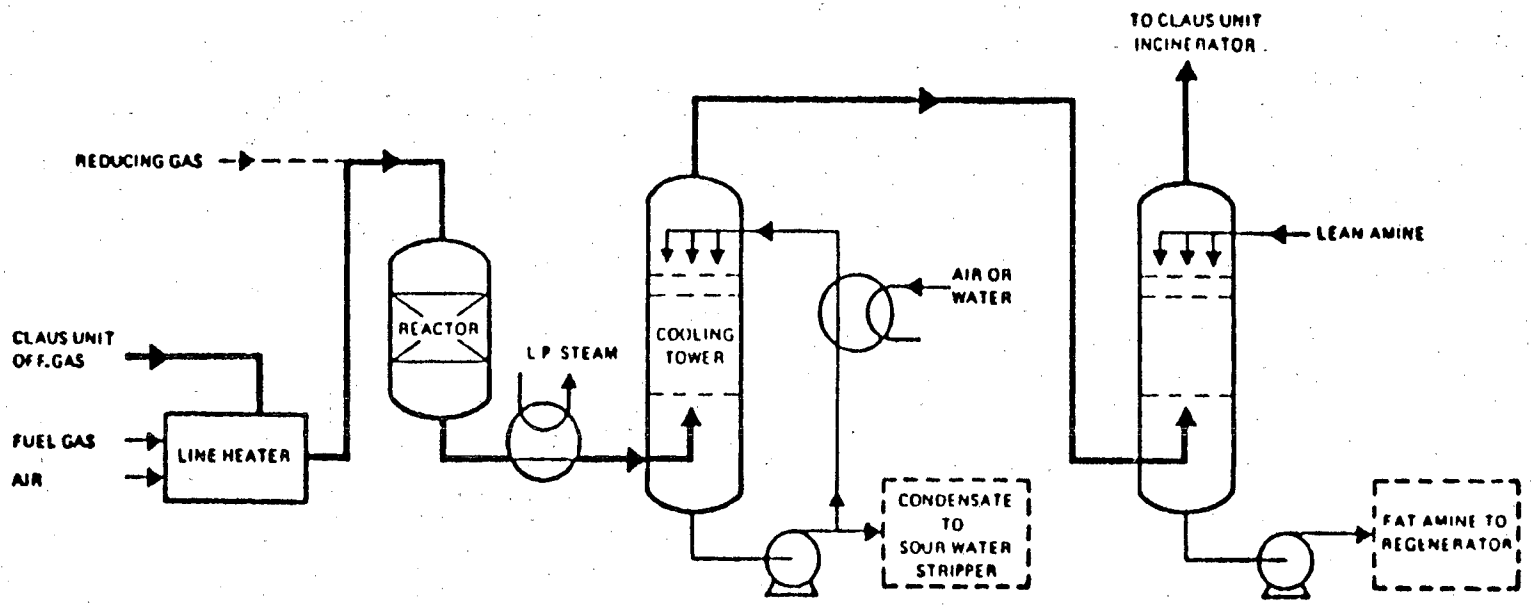


Figure 3-51. Shell Claus offgas treating (SCOT) process schematic flow diagram. (Ondich, 1983)

amount of hydrogen sulfide and possibly NH_3, is sent to a sour water stripper. Any H_2S and NH_3 present is released and recycled to the Claus feed.

The cooled gas, which contains up to 3 mole percent of hydrogen sulfide and up to 20 mole percent of carbon dioxide, is countercurrently scrubbed with an alkanolamine solution in an absorption column. The treated gas, containing traces of hydrogen sulfide and carbonyl sulfide, is burned. The rich amine solution is regenerated in a conventional stripper, and the desorbed hydrogen sulfide is recycled to the front of the Claus plant for recovery of additional sulfur.

The selection of the type of solvent used in SCOT absorbers is largely determined by the composition of the Claus tail gas, more specifically its CO_2 content. For amine treating processes using MEA (monoethanolamine) or DEA (diethanolamine) or the ADIP solvents based on DIPA (diisopropanolamine) and MDEA (methyldiethanolamine), the rate of H_2S absorption is high and practically the same, since the absorption is gas-film-controlled. Therefore, to achieve a given purity of the treated gas the same number of absorption trays is required for these four solvents.

Three different levels of integration of the solvent systems have been applied, viz. add-on, common regeneration, and cascaded lineup. The term add-on refers to a self-contained SCOT unit with a fully independent solvent system, including solvent regeneration which as such can be "added-on" to one or more Claus units. Most SCOT units built so far are of the add-on type. However, in recent SCOT designs there is a distinct trend towards the application of integration, in particular the cascaded line-up.

In the common generation and cascaded line-up systems the SCOT solvent section is integrated with the sovent section of the primary gas treating unit. An integrated system all treating units must use the same type of solvent.

In the common regeneration system the loaded SCOT solvent is regenerated together with the fat solvent stream(s) from the primary gas treaters in one common regenerator - usually considered part of the primary gas treaters.

Owing to the low H_2S partial pressure in a SCOT absorber, the solvent leaving this column is only partly loaded with H_2S. The loading capacity left can be utilized for the removal of H_2S from a gas with a high H_2S partial pressure. In the cascaded line-up this further loading takes place by introducting the SCOT "fat" solvent into a primary gas absorber. From this column the solvent is routed to the (again common) solvent regenerator.

B. Process Applicability--

The SCOT process is applicable for treatment of tail gases from Claus plants operating under a variety of conditions. As discussed previously, factors such as the Claus tail gas pressure and composition, available reducing gas, and desired level of integration with upstream processes influence the overall SCOT design.

The first two commercial SCOT units began operation in the fall of 1973. By late 1980, 32 plants in the U.S. and 25 plants outside the U.S. were in operation. Approximately 8 plants in the U.S. and 7 plants outside the U.S. are in various stages of planning, design, and construction. These plants are located mostly in petroleum refineries and range in size from 3 to 2100 Mg/day of sulfur recovered in the Claus unit. The largest SCOT unit in the U.S., at 850 Mg/day of sulfur, is located in the Shell Oil Plant at Eustace, Texas.

C. Process Performance--

The vent gas from the SCOT absorber typically contains 200 to 500 ppmv hydrogen sulfide. In Table 3-37, typical compositions of Claus and SCOT unit gas streams are presented. For this example case, the SCOT offgas contains 300 ppmv H_2S, 10 ppmv COS, and 1 ppmv CS_2. The low concentration of COS could be the result of the low CO_2/H_2O ratio (less than 0.1) in the Claus offgas, leading to favorable conditions for the hydrolysis of COS in the SCOT reactor. The calculated equilibium COS concentration for this case, however, is even lower at 1 ppmv. This is in conflict with the observation that the concentration of COS approaches thermodynamic equilibrium.

Two other sets of sulfur emission data are presented in Table 3-38. However, none of the data presented provides information on the concentrations of COS in the SCOT offgas. Also, all three plants started up prior to 1978

TABLE 3-37. TYPICAL COMPOSITIONS OF GAS STREAMS IN CLAUS AND SCOT UNITS (Ondich, 1983)

(Basis: 94% sulfur recovery in Claus unit)

Composition, % Volume	Claus Intake	Claus Offgas	SCOT Offgas	Incinerated SCOT Offgas
H_2S	89.9	0.85	0.03	<10 ppm vol
SO_2	--	0.42	--	0.02
S_3 vapor & mist	--	0.05	--	--
COS	--	0.05	10 ppm vol	--
CS_2	--	0.04	1 ppm vol	--
CO^2	--	0.22	--	--
CO_2	4.6	2.37	3.05	4.42
HC^2(MW: 30)	0.5	--	--	--
H_2	--	1.60	0.96	--
H_2O	5.0	33.10	7.00	9.84
N_2	--	61.30	88.96	83.94
O_2	--	--	--	1.78
Total	100.0	100.00	100.00	100.00
Temperature, K	310	410	310	920
Pressure, kPa	147	127	98	98
Gas quantity, mole relative	1	3.0	2.2	3.5

TABLE 3-38. SULFUR CONCENTRATION IN SCOT OFFGAS (Ondich, 1983)

Plat Location	Sulfur Concentration in SCOT Offgas	
	Before Incineration	After Incineration
Shell/Deer Park, Texas		160 - 350 ppmv, average 200 ppmv on the basis of 50% excess air
Gulf/Port Arthur, Texas	517 ppmv	267 ppmv
Texaco/Port Arthur, Texas	200 to 500 ppmv H_2S average 300 ppmv H_2S	

and were probably not designed to meet the current New Source Perforamnce Standards (NSPS) for petroleum refinery Claus sulfur recovery plants.

D. Process Reliability--

The SCOT process is generally considered to be highly reliable and easy to operate. Three sets of process reliability data have been collected (Ondich, 1983). These data indicate that:

- At the Shell/Deer Park, Texas facility, the SCOT unit has 99.5 percent on-stream time (excluding three weeks scheduled maintenance shutdown every three or four years).

- At the Gulf/Port Arthur, Texas facility, typical on-stream time for the SCOT unit ranges from 97.9 percent in 1977 and 1978 to 99.9 percent in 1980. Due to a heater tube failure in 1979 and a process blower failure in early 1981, the on-stream time during these two periods is considerably lower.

- At the Texaco/Port Arthur, Texas facility, on-stream time for the SCOT unit is 99.16 percent. Operational problems causing down-time include: (1) leaking amine exchanger, (2) high incinerator temperature, (3) blower failure, and (4) malfunction and repair of regulators, control valves, and emergency shut-down instruments.

E. Process Economics--

The capital investment cost for a SCOT unit designed to treat the tail gas from a 150 Mg/D Claus unit operating at 94 percent sulfur recovery was quoted by Shell Development Company to be $4 million based on mid-1979 dollars. In 1973 Shell provided capital investment costs for SCOT units; $1.8 million for a 100 Mg/D unit, $3.2 million for a 200 Mg/D unit, and $7.2 million for a 1000 Mg/D unit. Generally the capital investment for the add-on SCOT unit is approximately equal to the capital investment required for the preceding Claus unit. For the integrated SCOT unit (sharing amine regeneration and sour water stripping facilities), the capital investment for the SCOT unit equals about 75 percent of that for the Claus unit (Ondich, 1983).

In Figure 3-52, the capital investment costs for the SCOT unit are presented as a function of the capacity of the preceding Claus unit.

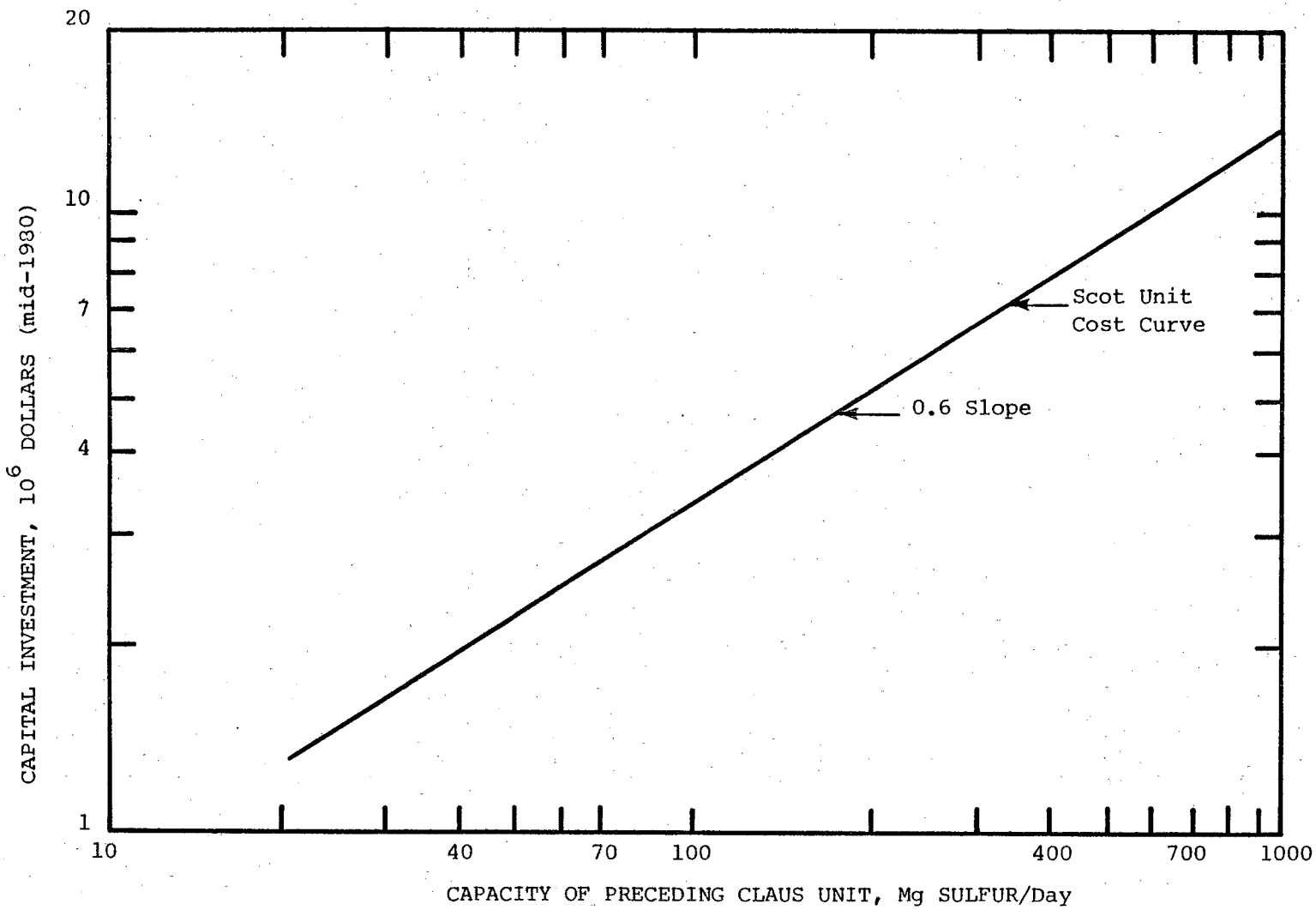


Figure 3-52. Capital investment cost for Scot units. (Ondich, 1983).

3.3.8 Flue Gas Desulfurization (Wet and Dry)

Flue gas desulfurization (FGD) techniques are used widely for removing sulfur oxides (SO_x) and other acid gases from the exhaust gases of combustion processes. In any shale oil recovery operation where conventional boilers, process heaters, etc. are used with fuel containing sulfur, FGD, in one form or another, can be readily employed to limit SO_x emissions. Where retorted shale is used as fuel in fluidized-bed combustors (FBC), the SO_x emission should be low. The use of FGD on these FBC exhaust gases may not be cost effective but could be used if the SO_x levels are excessive.

Retort offgas with high organic sulfur content also may be treated by incinerating the organics in the offgas and then scrubbing SO_x created with an FGD system. Any H_2S in the offgas also will be converted to SO_x by the incineration process.

Incineration followed by FGD is not usually the most cost effective technique for removing reduced sulfur compound (i.e., H_2S, COS, CS, MeSH, etc.). The gas volume increases by 1.5 to 3 times in the incineration which increases the scrubber size proportionally. Therefore, the reduced sulfur removal processes, discussed in the preceding section should be employed if possible.

However, to date, scrubbing processes which will remove over 90 percent of the reduced organic sulfuric compounds (COS, CS_2, MeSH, etc.) have not been used in full-scale operations with retort gas. Therefore, some shale oil processing plants with a high organic sulfur content in their retort gas are being designed to use incineration and FGD.

The alkaline/carbon/hypochlorite scrubber, discussed in Section 3.3.5, could be the alternative to incineration/FGD if it proves to have the high reduced organic sulfur removal that is expected.

FGD is widely discussed in the literature. The following is a brief recap of the technology as extracted from the Pollution Control Technical Manual Appendices (Ondich, 1983).

A. Calcium-Based, Wet Systems--

Calcium based wet FGD systems are the most prevalent type of SO_2 control processes used for coal-fired utility generating stations. Lime,

limestone, and alkaline ash systems represent approximately 80 percent of the FGD capacity currently operating, under construction, or being planned. Almost all of these processes are nonregenerable or throwaway processes. The major impetus for the use of nonregenerable systems is economic. The cost of limestone is such that discarding the resulting sludge is the cheapest SO_2 control mechanism available.

The history of lime and limestone FGD systems since their initial application in the early 1970s has not always been favorable. Numerous problems have plagued commercial systems throughout the development of this technology. One of the major causes of problems is the relatively low solubility of most calcium compounds present in FGD systems. Because of the low solubility of calcium carbonate and calcium sulfite, high liquid to gas (L/G) ratios are needed for adequate SO_2 removal. The low solubility of calcium sulfate has often caused scale formation in a system either due to undersized reaction tanks or chemical upsets. Another problem is the high solubility of calcium chloride in this chemical system which results in high liquid phase chloride concentrations that can cause corrosion.

The FGD system with a lime/limestone type of scrubbing solution is the generic process offered by many different vendors. Each vendor may have unique features incorporated in their design; however, the overall process chemistry and equipment are generally very similar. Both lime and limestone produce the same reaction products, although differences exist in the operating parameters. A general flow diagram is shown in Figure 3-53. Hydrated lime (CaO) is slaked onsite to form calcium hydroxide slurry. This slurry reacts with SO_2 to form calcium sulfite and calcium sulfate. The limestone chemistry is similar; however, carbon dioxide is also generated. The limestone, often gravel size or larger, is crushed in ball mills to produce a reagent slurry.

In the typical system, the flue gas is contacted with a slurry of calcium sulfur salts and the reagent. The gas may or may not contain fly ash, depending on the absorber design. Venturi and turbulent contact absorbers (TCA) have been used for combined SO_2 and particulate control while spray towers, perforated trays, and other low pressure drop contactors are often used for just SO_2 removal. A large volume of slurry (relative to the gas

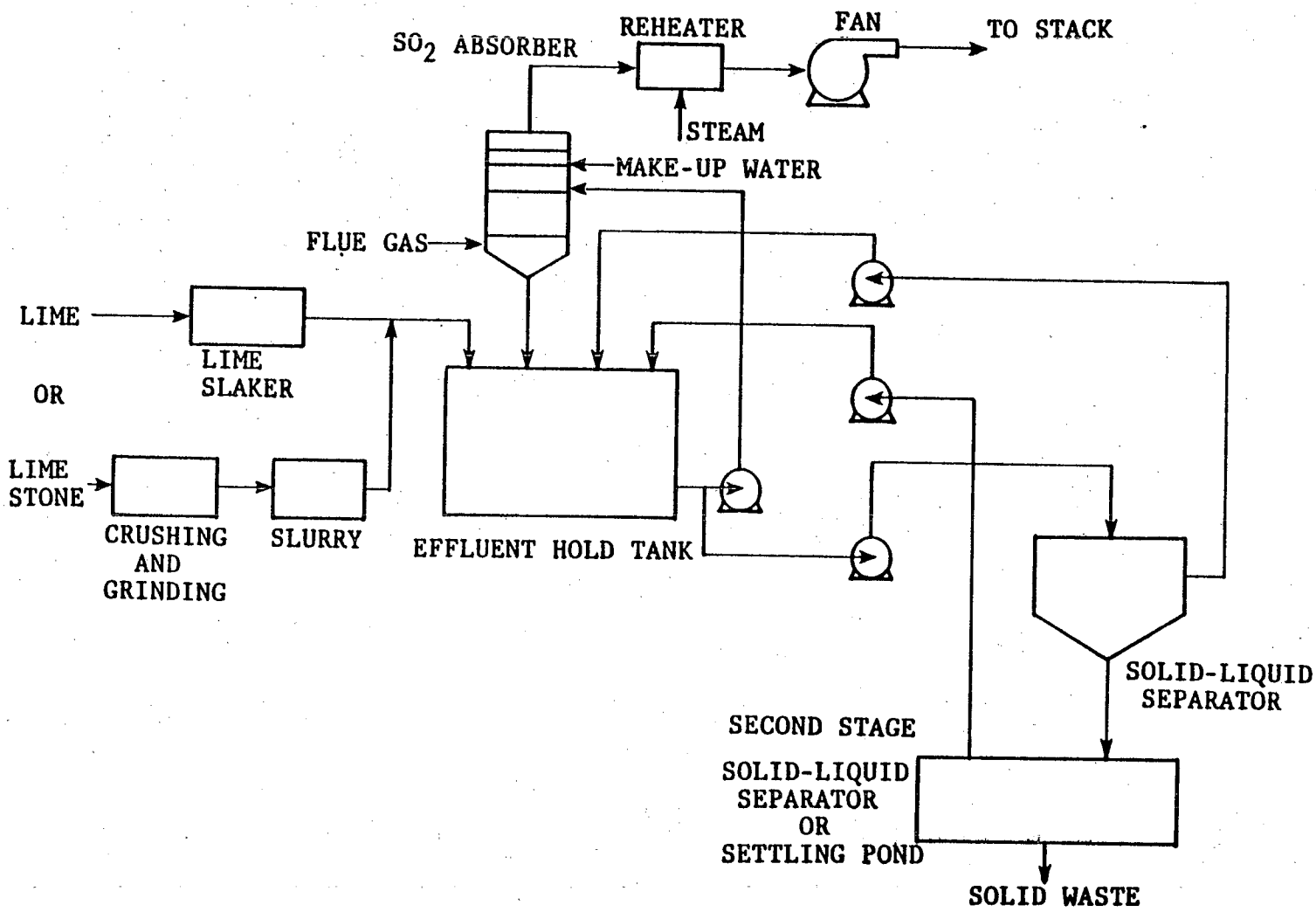


Figure 3-53. Typical process flow diagram for lime/limestone scrubbing. (Ondich, 1983).

KVB72 807430-2031

volume) is sprayed or dispersed in some manner in the contactor, saturating the flue gas and removing the SO_2. Typical L/G ratios for 5.3-16 m^3 liquid per thousand actual m^3 of gas. The scrubbed gas is then passed through mist eliminators and is often reheated to restore buoyancy before being discharged.

The SO_2-rich liquor typically drains into a large tank where neutralization and precipitation reactions occur. The alkaline reagent is added to this tank to maintain the desired system pH. When lime is used as the reagent, the feed liquor pH is typically 8-9 and the absorber effluent is 5-7. In a limestone system, because it is buffered by the carbonate species, the pH usually ranges between 5 and 6. Operating at this lower pH tends to increase the sulfite oxidation rate. If a super-saturated condition is reached without adequate gypsum seed crystals present in the slurry, hard scale can form on the absorber internals, disrupting operation. In order to obtain scale-free operation, the reaction tank is either sized to maintain the gypsum concentration low enough to prevent scale formation, or the oxidation is forced to near completion, which results in adequate crystal precipitation sites to prevent scaling.

A small portion of the recirculating slurry is removed to control the suspended solids concentration. This stream can be thicknened, filtered, or centrifuged before it is ultimately disposed.

B. Sodium-Based, Wet Systems--

Approximately ten percent of commercial FGD systems are based on sodium processes, of which there are three types, dual alkali, sodium carbonate, and Wellman-Lord. All of these processes absorb SO_2 in the same manner. The makeup sodium reagent typically is either caustic (sodium hydroxide) or soda ash (sodium carbonate). Trona, a mineral form of sodium carbonate, also is used.

The sodium compounds used in these processes are relatively expensive compared to reagents such as lime or limestone. However, sodium compounds have certain advantages. The primary advantage is that they are much more soluble than their respective calcium salts. As a result, the liquid phase alkalinity of the absorbing liquid permits very efficient SO_2 removal at low L/G ratios. The predominant reactive species is the sulfite ion which neutralizes the sulfurous acid formed. Absorber effluent is treated with makeup reagent to convert the bisulfites back to sulfites.

The high solubility of sodium salts, although good for SO_2 removal, creates a disposal problem. Since the salts are soluble, direct disposal can only be accomplished economically in arid areas where evaporation ponds can be employed. The waste, if it is not oxidized to sulfate, also could place a large chemical oxygen demand on any receiving waters.

The dual alkali and Wellman-Lord systems solve the soluble effluent disposal problem in two different manners. In the dual alkali system, lime is reacted with the absorber effluent to produce a sludge which is discarded. This reaction also converts the sodium sulfite back into caustic, which is returned to the system. In the Wellman-Lord system,the bisulfite containing effluent is steam stripped to form sodium sulfite and gaseous SO_2. The SO_2 is then processed into sulfur or sulfuric acid. The Wellman-Lord selection is dependent on the available market for either of its byproducts.

The dual alkali process uses alkaline sodium compounds to achieve efficient SO_2 removal, and lime to regenerate the sodium-based scrubbing liquor. The flue gas normally is treated for particulate removal before it reaches the system. A presaturator can be included to prevent chloride buildup in the recirculating liquor when treating high chloride coals. The dual alkali process has two major chemical steps, absorption and regeneration. Sodium sulfite is the major alkaline species present in the liquid phase.

In the absorption step, caustic and soda ash react with SO_2 to form sodium sulfite, which, upon additional absorption of SO_2, forms sodium bisulfite. Oxidation of the sulfite ion to sulfate also occurs.

A portion of this recirculating clear liquor stream is sent to a reaction tank where lime is added to precipitate the sulfur species. The solids are separated in a thickener and concentrated in a vacuum filter before disposal The regenerated sodium compounds are then recycled to the absorber. Alkali losses in the filter cake moisture are replaced by the makeup sodium reagent. Most systems include some type of filter cake washing system to minimize the sodium loss.

The oxidation reaction of sulfite to sulfate requires a treatment scheme to remove the inactive sulfate from the clear liquor. Two mechanisms exist, the selection of which depends on the relative sulfate to sulfite

concentrations. Both calcium sulfate and calcium sulfite are relatively insoluble compounds, although gypsum (a form of calcium sulfate) is about an order of magnitude more soluble than calcium sulfite. Consequently, unless the sulfate to sulfite ion ratio in the solution is the same as the ratio of solubility products of the two compounds, only one species should precipitate. When this ratio is high, gypsum will precipitate as lime is added. This is called a "dilute" system because the concentration of the active sodium sulfite is fairly low. In a "concentrated" system, both calcium sulfite and sulfate precipitate and the liquor is subsaturated with respect to gypsum by a coprecipitation mechanism. The sulfite and sulfate form a solid solution in the crystal lattice structure. This mechanism maintains the subsaturated gypsum condition.

The Wellman-Lord process uses an aqueous solution of Na_2SO_3 to remove SO_2. The sulfite ion is oxidized to form sulfate. The $NaHSO_3$ solution from the absorber is thermally decomposed in a steam-heated evaporator to regenerate sodium sulfite.

Most of the water vapor from the overhead SO_2-H_2O mixture is condensed, and the SO_2-rich gas is further concentrated to about 85 percent SO_2 by steam stripping of the condensate. The SO_2-rich stream is suitable for processing into elemental sulfur or sulfuric acid. A slurry of Na_2SO_3 crystals is formed in the evaporator, redissolved with the condensate from the evaporator overhead, and recycled to the absorber as regenerated scrubbing solution.

Accumulation of sulfate in the Wellman-Lord system is prevented by a continuous purge treatment of the $NaHSO_3$ solution from the absorber. Earlier installations used a refrigerated purge treatment section to crystallize and removal $NaSO_4$. More recent developments indicate that a high temperature purge treatment requires less energy and capital. Sodium losses from the purge treatment are made up by addition of NaOH or Na_2CO_3 to the recirculating scrubbing solution.

C. Spray Dryer Systems--

Wet-reagent, dry-scrubbing systems, known as spray dryers, use a solution or slurry, which is evaporated to dryness by the flue gas, to remove SO_2.

Spray drying is a relatively new FGD process rapidly gaining utility acceptance. In spray dryer systems, hot flue gas from the boiler or air preheater passes into a reaction vessel with a residence time of 5 to 10 seconds. Here it is contacted with a slurry, paste, or solution of alkaline material. The flue gas is humidified adiabatically to within 30°K of its saturation temperature by evaporation of the slurry solution. The slurry generally is dried to less than one percent free moisture. These salts, along with fly ash, are entrained in the flue gas and carried to the particulate control section. This is typically a baghouse, although ESPs can also be used. Some of the solids drop to the floor of the dryer and are combined with the baghouse catch. Collected solids may be recycled through the reagent preparation system along with solids from the particulate collector. The SO_2 capture reactions take place both during and following the drying process, continuing even on the particulate collector surfaces.

There are three advantages of wet-reagent, dry-scrubbing processes over wet processes. First, the flue gas leaving the SO_2 absorber is not saturated with water vapor. This usually prevents moisture condensation and subsequent corrosion in downstream equipment. Second, waste material is handled dry, rather than in a slurry or a liquor. Third, particulate control is achieved at no additional pressure drop. For throwaway applications, dry solids disposal can be easier. One disadvantage of these systems is the somewhat lower SO_2 removal capabilities. Commercial operating experience also is less with the dry systems than with the wet systems.

There are several types of spray dryer system designs being offered. All involve the following four steps:

- reagent preparation
- atomization
- SO_2 absorption-water evaporation, and
- particulate removal

In addition, solids recycle is integrated into most of the current system designs. During reagent preparation, the alkaline material is dissolved or slurried in water. Limestone has not been shown to be sufficiently reactive for this application. Both lime and soda ash are

reactive and cost effective for spray dryer applications. At this time, however, most spray drying systems are lime based.

Atomization is accomplished in either a rotary atomizer or with nozzles, In a rotary atomizer, liquid is fed into a rotating wheel. The liquid is accelerated, and is atomized at the wheel's edge forming a spray of droplets. The spray leaves the wheel horizontally at an angle of about 180°. Droplet size is dependent on the wheel speed, fluid viscosity, and feed rate, with wheel speed being the most important variable.

In a nozzle atomizer, either the slurry or solution is fed under high pressure (hydraulic) or a separate high pressure (two-fluid) medium is supplied. In a nozzle, the feed is ejected from the orifice as a high speed film which disintegrates into droplets. In a large module, multiple nozzles can be installed.

SO_2 absorption and water evaporation occurs simultaneously, generally within a second or two after the droplets and gas enter the dryer. In a properly designed system, the dried product must be free flowing while the chamber itself remains dry and free of deposits. This is achieved by controlling: (1) the particle size of the atomized feed, and (2) the dryer outlet temperature. The dry salt mixture produced is usually about 70 percent anhydrous sulfite and 30 percent anhydrous sulfate.

Particulate removal is accomplished either in a baghouse or precipitator. Bag collectors offer sites for additional SO_2 removal with unreacted alkalinity. Up to 10 percent of the total SO_2 removal occurs in the baghouse. Bag fabrics can be blinded by wetting, therefore operation must remain well above the saturation temperature. On the other hand, ESP collectors do not assist with the SO_2 absorption; but, they are less senstive to condensation. Therefore, the spray dryer can be operated as close as 11 C above the saturation temperatures which causes higher SO_2 removal in the reactor.

Some spray drying systems also incorporate solids recycle to increase reactant utilization and in some cases also take advantage of fly ash alkalinity. Spent solids removed from the bottom of the spray dryer are sent to the sorbent preparation area (e.g., slurry tank). Most system designs reslurry the product solids in a loop separate from the fresh sorbent

preparation loop, and then combine the two slurries just upstream of the atomizer. This allows closer control of the recycle slurry pH, which has been found to impact the overall system performance.

D. Dry Sorbent Injection System--

The newest FGD concept is to inject a dry sodium sorbent powder ahead of the particulate collection filter (i.e., baghouse). This system offers the industry a potential for substantial cost savings as well as design and operational simplicity compared to wet scrubbing systems or even compared to spray dryers. It also requires less water than wet or spray dry type systems, an important consideration in regions where water is in short supply. This system is especially attractive for boilers firing western U.S. coals subject to 70 percent SO_x removal New Source Performance Standards.

The system involves the injection of Trona, Nahcolite or any other naturally-occuring forms of sodium bicarbonate and sodium carbonate. This dry powder is introduced into the gas stream at a point between one and two seconds of flow time before the gas enters the baghouse. The gas temperature at the point of injection and in the baghouse is important to the removal efficiency achieved. The nominal temperature is 150 C (300°F) and the optimum must be established for the specific sorbent selected.

The sorbent selection is the most important aspect of the system design. Nahcolite is most effective. It can achieve a 70 percent removal efficiency with a normalized stoichiometric ratio (NSR) of approximately 0.75. Trona can achieve 70 percent removal efficiency with an NSR of 1.3 and soda ash has proven to be esssentially ineffective in tests to date (Muzio and Sonnichsen, 1984).

This process is still in development under sponsorship of the Electric Power Research Institute (EPRI). Demonstration tests on a 22 MW coal fired boiler were conducted at the Public Service of Colorado's Cameo Unit 1 (Muzio and Sonnichsen, 1984).

In late 1985, EPRI was planning a project to demonstrate the process at full scale for SO_2 control on coal-fired utility boilers. The project was to be co-sponsored by City Utilities of Colorado Springs and the FMC Corporation. The demonstration would be conducted on Unit 1 of the R.D. Nixon plant of City Utilities.

The primary disadvantage of this process is the disposal of the water-soluble byproducts, sodium sulfite and sulfate. Because certain areas in the Colorado and Utah areas may have suitable disposal sites for these byproducts, this process may have an application for shale oil processing in those states.

Figure 3-54 is a simplified flow schematic of a coal-fired power plant equipped with dry sodium injection for SO_2 control. With the dry sodium injection processes tested thus far, the reagent material is fed into the flue gas stream ahead of a baghouse and downstream of the system air heater. In the ductwork, the reagent particles decompose to sodium carbonate (Na_2CO_3) forming an open, porous microstructure, which then reacts with SO_2 to form either Na_2SO_3 or Na_2SO_4. Unreacted sodium carbonate deposited on the fabric in the baghouse can also react with SO_2 to provide further SO_2 removal. The decomposition process is apparently important to the efficiency of the process, particularly at temperatures between 120 and 200 C. Typically 70 to 80 percent of the total SO_2 removal occurs in the baghouse and 20 to 30 percent in the upstream ductwork, although these ratios can vary considerably with injection temperature and, perhaps, particle size.

From the work completed and that currently is in progress, the following parameters have been identified as being important:

- Sodium sorbent type
- Sodium/sulfur injection ratio
- Particle size
- Injection method (sorbent dispersion)

A list of sodium based reagents which have been tested, thus far, in dry SO_2 removal systems is shown in Table 3-39. Sodium sesquicarbonate is a synthetic material normally obtained by refining Trona ore. It results as an intermediate step in the manufacture of soda ash. Trona ore is dissolved in a liquor and impurities are removed. Crystalization then results in a pure sodium sesquicarbonate product. Trona is a naturally occurring sodium sesquicarbonate. It is water soluble and decomposes with heat to yield sodium carbonate, carbon dioxide, and water. Tremendous deposits of Trona, located in the State of Wyoming are currently being mined and refined into soda ash for use in a great number of commercial products.

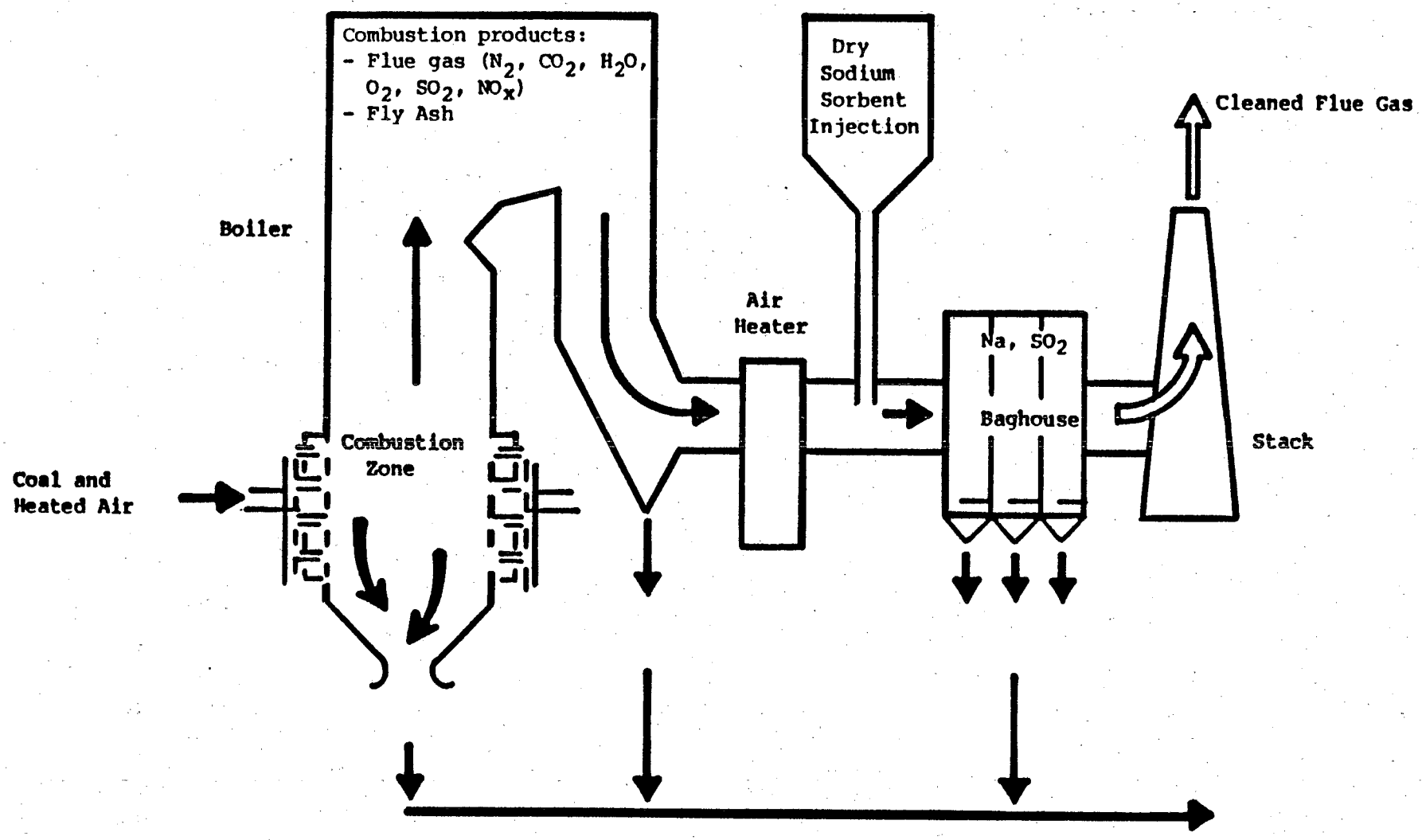


Figure 3-54. Schematic of a coal-fired power plant with dry-sodium injection for SO_2 control. (KVB, Inc., 1985).

TABLE 3-39. SODIUM REAGENT MATERIALS
(KVB, Inc., 1985)

Sorbent	Formula	Approx. Purity	Comments
Sodium Bicarbonate	$NaHCO_3$	98%	Synthetic Material
Nahcolite	$NaHCO_3$	76%	Natural Material Contains high level of insolubles
Sodium Sesquicarbonate	$NaCO_3 \cdot NaHCO_3 \cdot 2H_2O$	98%	Synthetic Material
Trona	$Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$	88%	Natural Material Contains high levels of insolubles
Sodium Carbonate	$NaHCO_3$	98%	Synthetic Material
Sodium Tribicarbonate	$NaHCO_3 \cdot 3NaHCO_3$	93%	Synthetic Material

In the dry sodium scrubbing process, the following series of reactions are postulated to take place to produce sodium carbonate for the removal of SO_2:

Bicarbonate (Nahcolite)

$$2NaHCO_3 \rightarrow Na_2CO_3 + CO_2 + H_2O$$

Sesquicarbonate (Trona)

$$2[Na_2CO_3 \quad NaHCO_3 \quad 2H_2O] \rightarrow 3Na_2CO_3 + 5H_2O + CO_2$$

Tribicarbonate

$$2[Na_2CO_3 \quad 3NaHCO_3] \rightarrow 5Na_2CO_3 + 3CO_2 + 3H_2O$$

Then for sulfation:

$$2Na_2CO_3 + 2SO_2 + O_2 \rightarrow 2Na_2SO_4 + 2CO_2$$

or

$$2Na_2CO_3 + 2SO_2 \rightarrow 2NaSO_3 + 2CO_2$$

The decomposition step, which for Nahcolite at a temperature of 300°C occurs in approximately one half second, generally leads to an increase in the specific surface area of the sorbent. The surface area of Nahcolite can increase from about 0.5 to 5 M^2/g, of Trona from 1.5 to 3 M^2/g, and of sesquicarbonate from 1.0 to 2.5 M^2/g during decomposition.

In Figure 3-55, typical results comparing the SO_2 removal achieved with various sodium compounds as a function of the NSR are shown. At an NSR

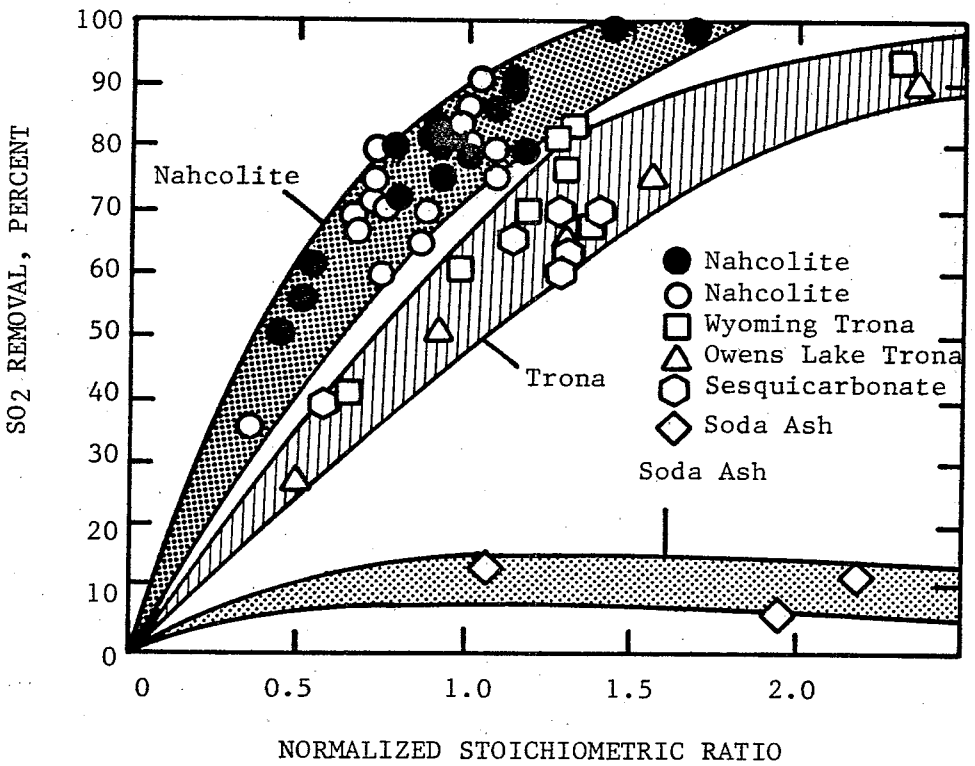


Figure 3-55. SO_2 removal as a function of normalized stoichiometric ratio. (KVB, Inc., 1985).

of 1, Nahcolite typically produces a 70 to 85 percent removal and Trona or sodium sesquicarbonate 40 to 60 percent. Soda ash (sodium carbonate) also has been tried but discontinued when it produced an unacceptable removal efficiency of ten percent.

In the temperature range normally encountered in baghouse systems used with boilers (125 C to 170 C), the flue gas temperature can affect sorbent usage although the effect can vary with reagent type. For sodium bicarbonate, there is a general increase in sorbent utilization with increasing temperature at the point of injection.

This improvement in utilization appears to be related to the rate of thermal decomposition of the sodium bicarbonate and the availability of SO_2 in the vicinity of the freshly decomposed sorbent surface. For Nahcolite the concern is at low temperatures. At a temperature of 130 C, the SO_2 removal efficiency using Nahcolite can be 50 percent lower than it is at 150 C.

With a Trona reagent, the baghouse temperature can be as low as 120 C without a detrimental effect on SO_2 removal. The capability of Trona to remain reactive at low temperatures can probably be attributed to its faster reaction rate relative to Nahcolite. In general, with injection of sodium sesquicarbonate or Trona, the effect of increasing the temperature within the 125 C to 170 C window has been found to have little or no effect (Muzio and Sonnichsen, 1984).

The testing performed thus far has indicated that the sorbent particle size is important. The influences of particle size is on the distribution and dropout of the particles in the baghouse. The measured effect of particle size on the SO_2 removal efficiency is shown in Figure 3-56. A significant influence of particle size was seen, although from this figure it is not clear if the influence or particle size is due directly to a change in reaction rate or is an indirect effect due to increase particle dropout in the ductwork and baghouse. Current indications are that the decrease in sodium utilization for larger paticles is due primarily to the dropout of large particles directly into the baghouse hopper without depositing on the filter bags. This conclusion is important when considering injection design criteria, reagent pulverization requirements and possibly baghouse inlet transition and hopper designs. The carrying velocity into a baghouse compartment hopper needs to be

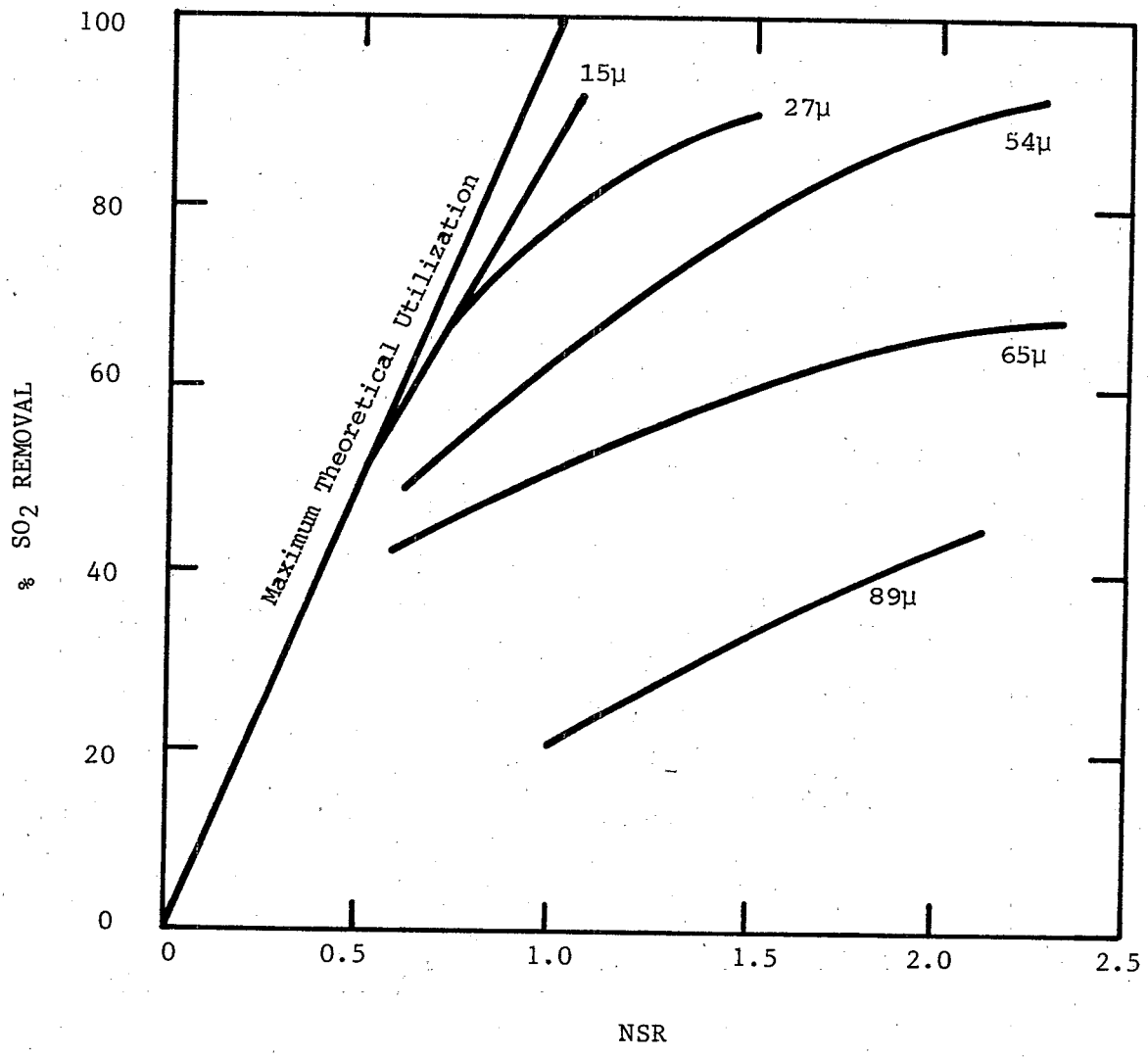


Figure 3-56. Sodium utilization as a function of NSR and particle size. All bicarb materials supplied by one chemical company. (KVB, Inc., 1985).

sufficient to keep a reagent particle from falling out; and the reagent distribution needs to remain somewhat uniform across the tubesheet. This underlines the importance of a good aerodynamic modeling effort relating to the injection system and baghouse designs, especially at lower boiler steam loads where flue gas flow rates are substantially lower than design specifications.

On tests conducted to date, the addition of sodium sorbents have had no adverse effects on baghouse performance. During the EPRI tests at the Public Service of Colorado, Cameo Station, the baghouse collection efficiency remained at 99.96 percent when Nahcolite and Trona were injected at quantities required for 70 percent SO_2 removal. No appreciable increase in baghouse pressure drop characteristics or bag cleanability where observed during sorbent injection while operating in the pressure- or time-initiated cleaning modes.

During earlier pilot tests, the only problem observed with increased baghouse pressure drop when injecting very fine sorbents (MMD = 8 μm). When using this material, the cleaning frequency had to be increased, after which satisfactory operation could be maintained (Bland, 1985).

E. Process Performance and Economics--

Two commercially available systems, limestone and Wellman-Lord, were selected as being representative of FGD process by Ondich (1983). Other FGD processes may offer certain technical or cost advantages, but the two selected processes are believed to be "state-of-the-art" as far as achievable levels of SO_2 control are concerned.

FGD costs for boilers in shale oil plants will depend upon the amount of sulfur emissions control required to meet NSPS standards. This may vary depending upon the amount of sulfur in the fuel. For an Illinois No. 6 type bituminous coal having a sulfur content of 3 to 3.5 percent approximately 90 percent removal is generally required. However, for a lower sulfur fuel, a smaller amount of removal may be necessary.

FGD cost data have been developed by the EPA for electric utility steam generating units ranging in size from 25 MW to 1000 MW (Ondich, 1983). The cost variations are principally governed by: 1) size of the boiler, 2)

fuel used, 3) averaging time over which the plant must meet the SO_2 limitation, and 4) level of control maintained.

Table 3-40 presents estimated capital investments and annual operating costs for the limestone and Wellman-Lord FGD processes applied to a 500 MW boiler burning a 3.5 percent sulfur coal and achieving 90 percent SO_2 control. These estimates are in 1983 dollars. Capital investment estimates include direct and indirect costs, such as purchase of equipment, installation labor and material, engineering costs, land required for sludge, etc. Annual operating costs include fuel costs, operating labor, and overhead expenses for safety.

TABLE 3-40. COSTS OF SO_2 CONTROL FOR 90 PERCENT REMOVAL
Basis: 500 MW unit burning a 3.5 percent sulfur coal
(Adapted from Ondich, 1983)

	Capital Investment, $/kW	Annual Operating Costs, $/kWh
Limestone Scrubber	170	0.005
Wellman-Lord Scrubber	160	0.005

In order to estimate costs for shale oil steam and power generating units, it can be useful to evaluate costs in terms of dollars per flue gas flow rate. For the limestone and Wellman-Lord FGD process, the capital costs for a 500 MW eletric unit were both estimated to be $1,100 per kmol/hr. Similarly, annual operating costs for both were estimated to be approximately $250 per kmol/hr. When applying these costs to shale oil plants, adjustments for size can be made based upon the six-tenths rule. Costs for FGD units requiring less than 90 percent control can be estimated by bypassing part of the flue gas and treating the remainder for 90 percent removal so as to achieve the required SO_2 removal.

3.3.9 SOx Absorption in Spent Shale (ASSP)

The ability of combusted carbonate-containing spent shale to absorb SO_2 gives rise to a novel concept for controlling sulfur emissions in oil shale plants. This concept is referred to as ASSP which stands for Absorption on Spent Shale Process by Van Zanten and Haas (1986).

The ASSP concept has several potential advantages over conventional sulfur removal technologies:

- The sorbent is cheap and inherently abundant in oil shale plants.
- The process requires combustion of the spent shale which is already incorporated into several of the retorting technologies or which would be a useful add-on to recover residual carbon values.
- Since non-H_2S compounds are converted to SO_2 by combustion, ASSP could represent a more efficient removal relative to gas sweetening processes which only remove H_2S.

The ASSP concept uses a fluidized transport system to combust either raw or retorted shale, thereby providing the vehicle for converting sulfur compounds to SO_2 and absorbing the SO_2 in the shale matrix. The concept envisions either a conventional dense-phase fluidized bed or a dilute-phase fluidized bed or a dilute-phase contactor (lift pipe). Key elements of the process are shown in Figure 3-57.

In an engineering assessment of the ASSP concept, J&A Associates (Van Zanten and Haas, 1986) evaluated three types of retorting processes:

- Direct heated
- Indirect heated
- Indirect heated with combustion integrated into the process

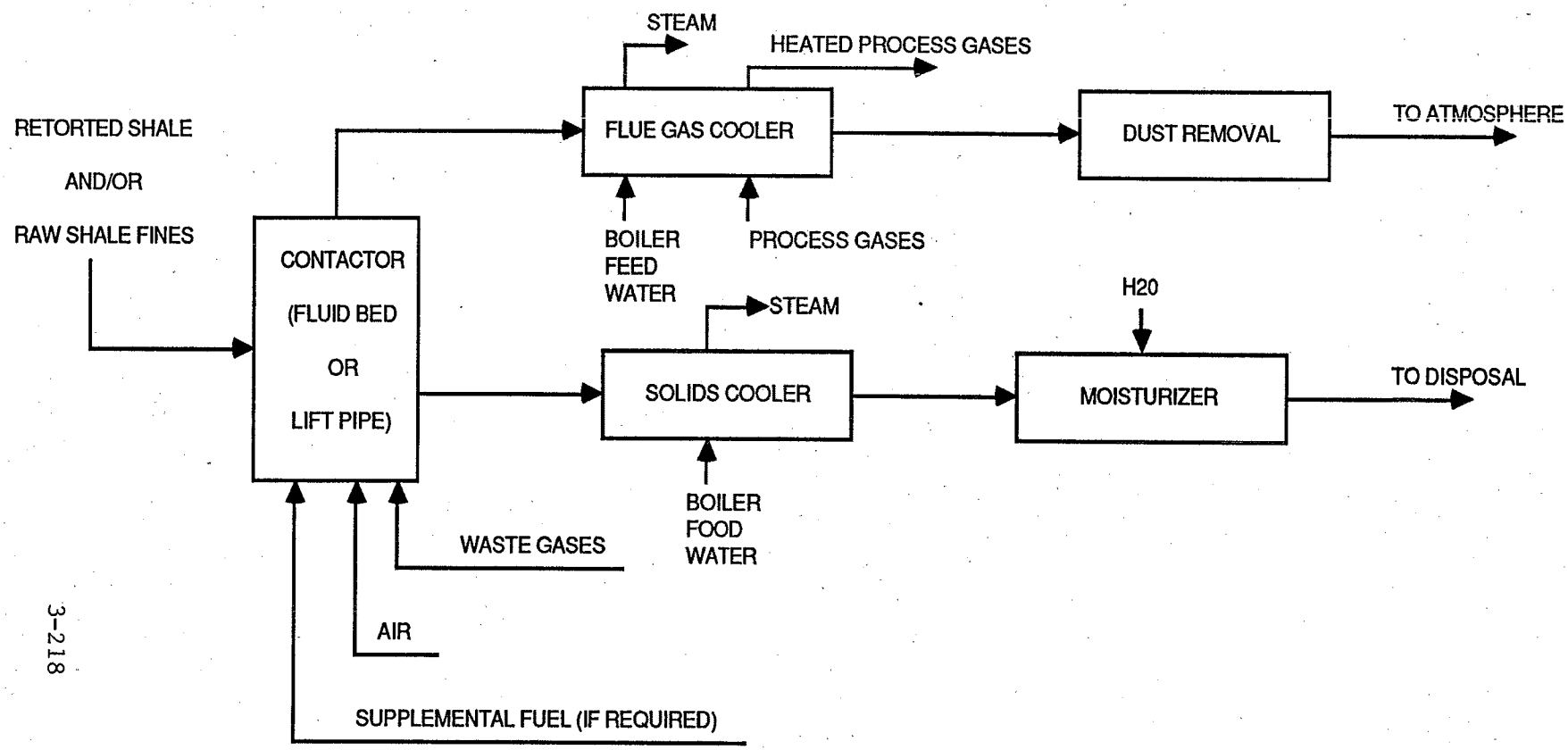


Figure 3-57. ASSP process flow diagram (Van Zanten, et al;, 1986).

Specific retorting technologies and sites were selected as representative of these three retort types as follows:

Retort Type	Process	Site
Direct heated	MIS with Unishale C	Cathedral Bluffs (Tract C-b)
Indirect heated	Unishale B	Union Oil (Parachute Creek)
Integral combustor	Lurgi	Rio Blanco (Tract C-a)
Integral combustor	Unishale C	Union Oil (Parachute Creek)

The study assumed that amine absorption is used to remove acid gases from indirect heated retort gases and that regenerated acid gases are burned in the ASSP combustor. MIS gases were assumed to be processed in the ASSP combustor without pretreatment.

For comparison purposes, conventional sulfur removal processes were evaluated:

- Direct heated - Case A; Unisulf + Flue gas desulfurization on combusted modified in-situ gases
  Case B; Unisulf + Stretford on modified in-situ gases
- Indirect heated - Unisulf on Unishale B gases
- Integral combustor - DEA + Stretford on Lurgi gases
  - Unisulf on Unishale C gases

Major equipment costs were taken from EPA Pollution Control Technical Manuals (PCTM). ASSP equipment was sized and costs factored from in-house data and PCTM's. Costs were factored to the first quarter of 1985.

The results of the cost study showed changes in incremental capital and operating costs for ASSP relative to conventional processing in Table 3-41 below.

TABLE 3-41. COST COMPARISON FOR ASSP
(VanZanten, et al., 1985)

Retort Type	Direct Heated Case A, Case B		Indirect Heated	Integral Combustor	
Retorting Process	MIS/Unishale C		Unishale B	Lurgi	Unishale C
ASSP Incremental Cap. Cost, $\$10^6$	-71.2	-63.2	+90.2	-13.0	-32.1
ASSP Incremental Ann. Oper. Cost, $\$10^6$/yr	+10.83	+12.07	-19.21	- 2.29	-1.56

These cost comparisons show that the best potential for application of ASSP are those processes which already have a spent shale combustor integrated into the retorting process (e.g., Lurgi, Unishale C, Chevron STB, and Tosco HSP). Capital and operating cost savings for Unishale C and Lurgi cases are primarily a result of deleting the Unisulf and Stretford plants.

Economics for the indirect and direct heated retorts are good to marginal. Factors which will affect the economics are:

- How effectively combustor heat can be utilized (simple steam raising is the best desirable).
- The value of steam.
- The use of fast or circulating fluid beds to reduce investment in combustor equipment.

Pilot plant tests were performed in an 46 cm (18-in) diameter, bubbling fluidized-bed combustor of the type which is integrated into the retort process. A total of 44 individual tests were performed. Variables evaluated were combustor temperature, solids residence time, gas residence time, oxygen concentration, inlet gas sulfur concentration, staged combustion, and raw shale injection. Over the entire range of conditions tested, emissions of primary pollutants were:

Component	Range
SO_2	1-38 ppmV
NO_x	80-670 ppmV
CO	0.05 - 1.80 vol%
Trace hydrocarbon	51-8465 ppmV

Key findings of the tests were:

- SO_2 emissions were controlled easily to low levels at virtually all conditions tested, probably as a result of the high Ca/S ratios used.

- NO_x emissions were primarily sensitive to oxygen concentration (Figure 3-58). Reasonably good NO_x control could be obtained with flue gas oxygen concentrations below about 3 volume percent. The lowest NO_x concentrations were seen at O_2 levels approaching zero but at the expense of higher CO and trace hydrocarbon emissions.

- CO and trace hydrocarbon emissions were primarily sensitive to flue gas oxygen concentration (Figure 3-59). Good control of both could be obtained at O_2 levels about 2 volume percent.

Emissions of NO_x move in a direction opposite to SO_2, CO and trace hydrocarbon emissions. Thus, finding a set of operating conditions which minimize all four represents a compromise. One test was run which produced nearly optimum results.

Conditions for this test were:

Bed temperature	937°K
Solids residence time	9.4 min
Gas residence time	0.9 sec
Gas superficial velocity	1.3 m/sec
Flue gas O_2	2.6 vol%
Ca/S mole ratio	10.3
Raw shale/spent shale ratio	1:36

At these conditions, the following result were obtained.

SO_2	11 ppmV
NO_x	160 ppmV
CO	0.27 vol%
Trace hydrocarbon	388 ppmV
Combustion efficiency	89%

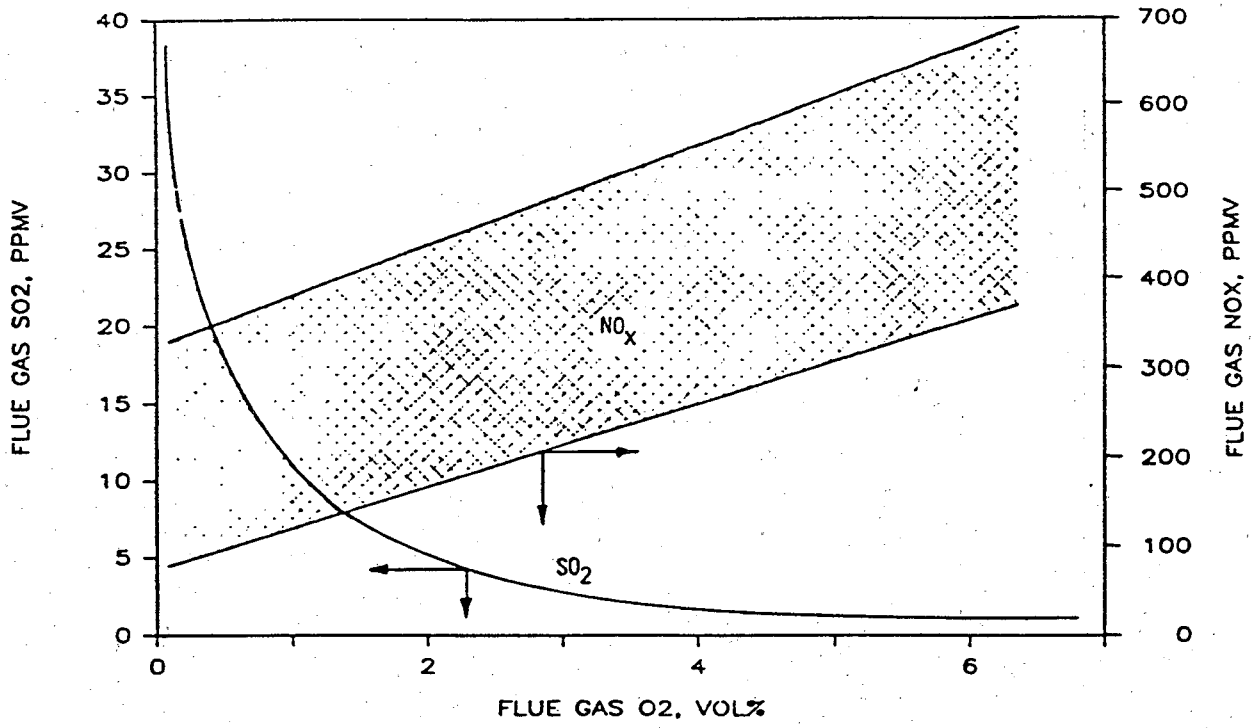


Figure 3-58. Effect of flue gas oxygen on SO_2 and NO_x emissions. (VanZanten, et al., 1986).

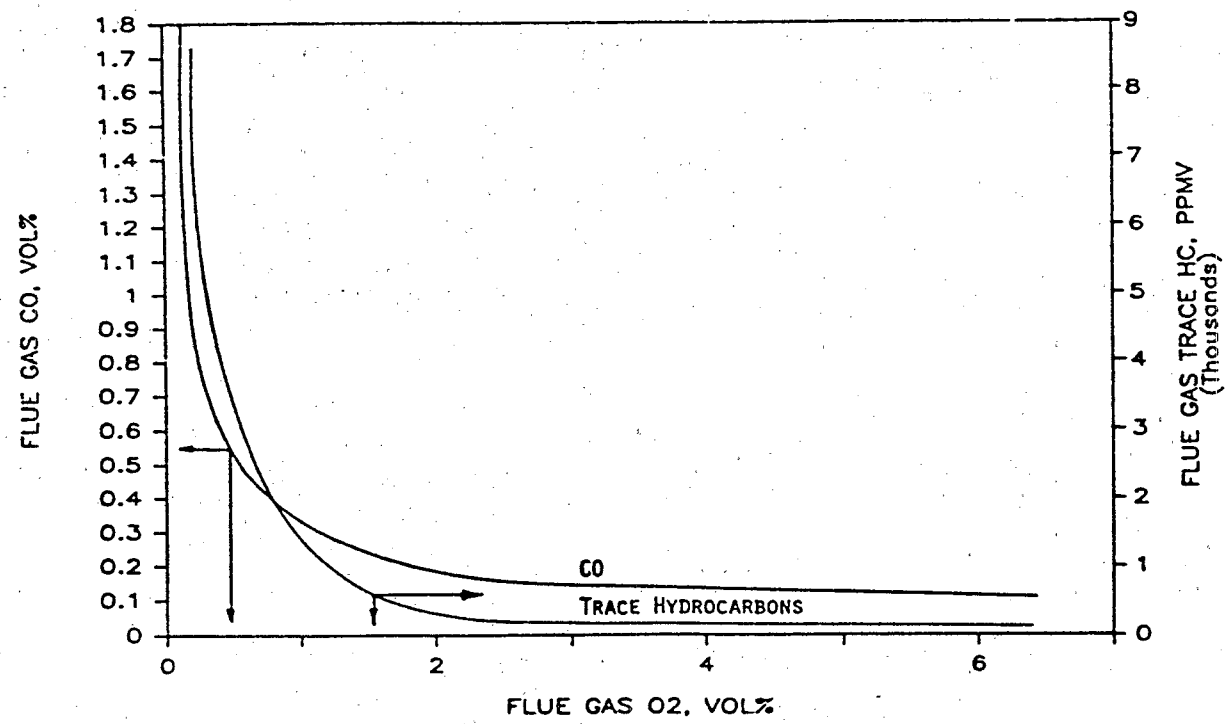


Figure 3-59. Effect of flue gas oxygen on CO and trace hydrocarbon emissions. (VanZanten, et al., 1986).

During selected tests, both combustor flue gas and retort gas were sampled and analyzed for selected trace elements -- mercury, cadmium, arsenic, lead, beryllium and fluorine. Solids streams also were analyzed for trace elements in an attempt to determine where trace elements go. One run was performed where a spike solution of mercury and cadmium was added to the combustor.

Results of the trace element tests indicated some relative trends with regard to emissions but because of the short duration of the sampling, no hard conclusions can be reached which would allow extrapolation of result to long term steady-state operations. Some of the key observations were:

- Lead, beryllium and fluorine were found to have low volatility. That is, of the amounts present in raw shale, only very small percentages were volatilized to the gas streams.

- Arsenic was found in significant concentrations in the retort gas (100-400 ppmV) although the amount of arsenic found represented less than 15 percent of that in the raw shale.

- So little mercury was present in the raw shale that mercury emissions could not be characterized with high accuracy. Mercury emissions were very low except during the spike indicating that mercury, if present in higher concentrations in the raw shale, could possibly pose emissions problems.

- Although significant amounts of cadmium was found in the gases at higher retort and combustor temperatures, emissions represented less than 10 percent of cadmium present in raw shale.

There is some evidence that mercury and cadmium introduced to the combustor during the spike test condensed within the retort equipment and revolatilized over time. However, because of the limited number of samples taken, the validity of any conclusions drawn would be questionable. Longer term steady-state operations would have to be studied to determine the fate of mercury and cadmium with more certainty.

3.4 ENGINE EMISSIONS OF CO AND HC

An oil shale recovery plant using an above-ground or modified in-situ (MIS) retort requires extensive mining. The primary gaseous emissions from the mining operation are carbon monoxide, hydrocarbons and nitrogen oxides from diesel engines.

One method for control of diesel emissions in underground mines that has gained fairly wide acceptance is the use of catalytic mufflers. These systems have been found effective in reducing emission levels of carbon monoxide (CO), unburned hydrocarbons (HC), and odor.

The results of an experimental program carried out by the Bartlesville Energy Research Center (BERC) (Marshall, et al;, 1978) on the effectiveness of various catalysts on reducing CO and HC emissions are presented in this section.

Four noble metal catalytic converters (two units each) and two engines were used in these tests. Three of the converters are standard production models and are representative of a significant fraction of those used in underground mines; the fourth converter is a prototype. The engines were mounted on test stands and coupled to eddy-current dynamometers. The engines were operated over a wide range of speed and load conditions. Converter and engine designations are given in Tables 3-42 and 3-43, respectively. All fuel used in these tests was obtained from the same supplier. Fuel property data are given in Table 3-44.

TABLE 3-42. CATALYTIC CONVERTERS (Marshall, et al. 1978)

Designation	Catalyst Bed Construction	Manufacturer	Models
A	Pelletized	Universal Oil Products	Prototypes
B	Pelletized	Oxy Catalyst	DV101, 149D
C	Monolithic	Oxy Catalyst	OC-4, OC-5
D	Monolithic	Engelhard	PTX-5DF, PTX-6DF

TABLE 3-43. ENGINES TESTED (Marshall, et al., 1978)

Code	Engine Type	Nominal Power bhp
1	Direct-injection, 4-stroke cycle, air-cooled	50
2	Indirect-injection, 4-stroke cycle, water-cooled	100

TABLE 3-44. FUEL PROPERTIES (Marshall, et al., 1978)

Specific gravity	0.845
Gravity (°API)	36.0
Sulfur (pct)	0.24
FIA analysis, pct:	
Aromatics	34
Olefins	2
Paraffins	64
Cetane index	49.7
Distillation, °F:	
10 pct evaporated	420
50 pct evaporated	490
90 pct evaporated	580

The exhaust gas was continuously monitored before and after each catalyst system for determination of CO, carbon dioxide (CO_2), sulfur dioxide (SO_2), nitric oxide (NO), nitrogen dioxide (NO_2), HC (total), and smoke (opacity). Batch samples were collected for subsequent determination of sulfur trioxide (SO_3), aldehydes, HC character, particulate mass loading, and particulate size.

The results of tests with the four catalysts showed significant differences in the oxidation efficency of various catalytic reactors. For a given catalyst design, the reduction in CO levels was fairly consistent for the two engines (Figures 3-60, 61). At low power modes (low exhaust temperature) the catalysts had minimal effect on CO emissions. At high power modes (high exhaust temperature) the catalysts exhibited a major influence on CO levels -- from 60 to 90 percent reduction at full load. The exhaust temperature at which 50 percent of the CO was oxidized ranged from about 350°F for catalysts A to greater than 700°F for catalyst C. Although there were large differences in oxidation efficiency, all of the catalysts effected major reductions in CO levels for those engine operating modes where the highest levels of CO were produced. Peak and average emission rates of CO and HC are given in Table 3-45. Note that the peak CO emission rate exceeded 5 g/hr/rated hp (the "acceptable" maximum emission rate of CO) only in those cases without any exhaust treatment. The value of 5 g CO/hr/rated hp is significant because at this level the quantity of ventilation air necessary to dilute the exhaust CO_2 to its threshold limit value (TLV) (0.5 percent) is sufficient for dilution of CO to its TLV (50 ppm).

The acceptable maximum emission rate can be computed for any toxicant for which an acceptable TLV level is specified. Values of acceptable maximum emission rate and TLV for various toxicants are given in Table 3-46. The emission rate of CO_2 was selected as the basis upon which the acceptable maximum emission rates are calculated because (1) CO_2 emission rate is relatable to fuel rate or power output, (2) CO_2 is the only toxicant necessarily produced from the complete combustion of a hydrocarbon fuel, and (3) the quantity of CO_2 is the toxicant that defines the minimum ventilation air rate required to dilute the exhaust to acceptable levels.

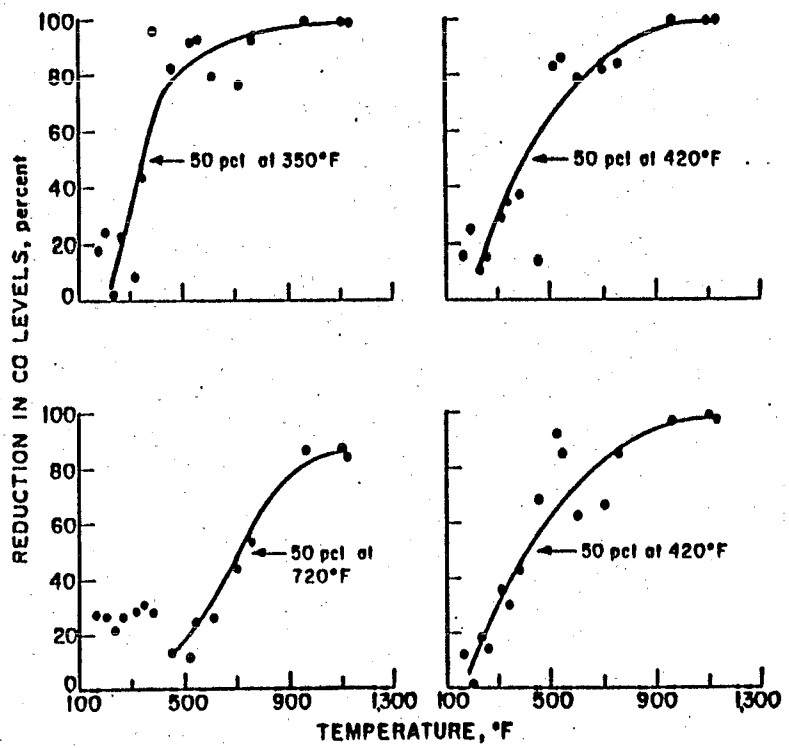


Figure 3-60. Catalyst efficiency for oxidation of CO vs exhaust temperature, engine 1-various speeds and loads. (Marshall, et al., 1978).

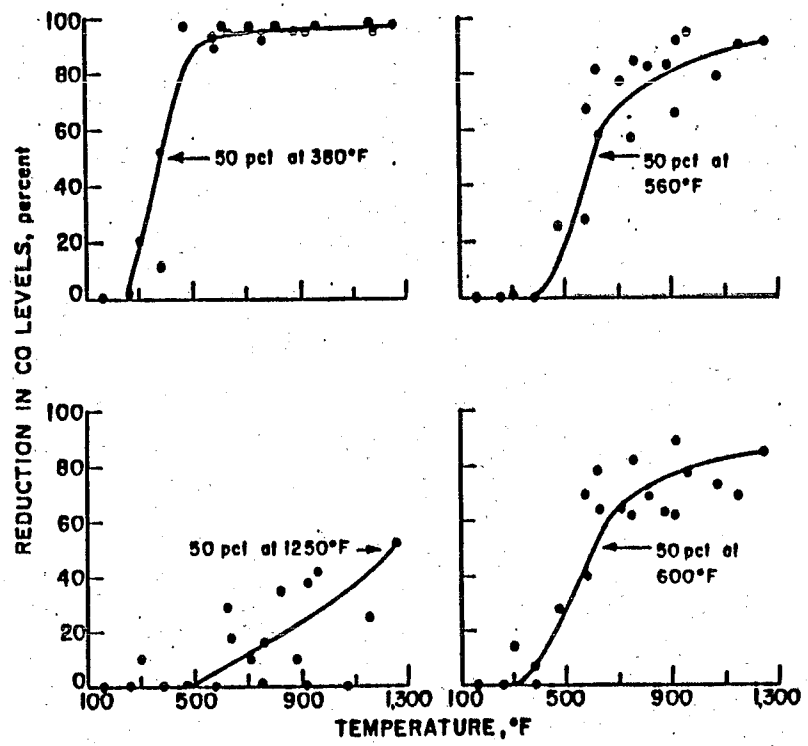


Figure 3-61. Catalyst efficiency for oxidation of CO vs exhaust temperature, engine-2 various speeds and loads. (Marshall, et al., 1978).

TABLE 3-45. EMISSION RATES OF CO AND HC, g/hr/rated hp
(Marshall, et al., 1978)

Exhaust Treatment System	Peak Rates CO	Peak Rates HC	Average Rates CO	Average Rates HC
ENGINE 1				
No catalyst	18.7	1.5	3.4	1.0
Catalyst A	1.5	0.8	0.5	0.3
Catalyst B	1.5	0.8	0.6	0.4
Catalyst C	2.6	0.9	1.0	0.5
Catalyst D	2.2	0.9	0.7	0.5
ENGINE 2				
No catalyst	5.7	4.1	2.6	1.5
Catalyst A	2.6	2.9	0.3	0.3
Catalyst B	3.5	2.9	1.1	0.8
Catalyst C	4.4	3.8	2.4	1.3
Catalyst D	2.8	3.4	1.1	0.9

TABLE 3-46. ACCEPTABLE MAXIMUM EMISSION RATES
(Marshall, et al., 1978)

Constituent	TLV, ppm	Acceptable Maximum Emission Rate[1] mg/hr/Rated hp
CO	50	5,000
NO	25	2,500
NO_2	5	800
HCHO	2	200
Acrolein	0.1	20
SO_2	5	1,100
SO_3 (as H_2SO_4)	0.25	80

[1] Acceptable maximum emission rate, mg/hp/hr = 3.5 x TLV x molecular weight.

Based on: fuel = CH_2
brake specific fuel consumption = 0.5 lb/hp/hr

The catalysts effected a 10- to 50-pecent reduction in peak emission rates of HC. The reduced efficiency (in comparison to oxidation of CO) is a consequence of the rather high emissions of HC at light load conditions (and correspondingly low exhaust temperatures). The results of a detailed examination of the hydrocarbon by gas liquid chromatography indicate that the catalysts tend to preferentially oxidize the unsaturated hydrocarbons. The remaining high-molecular-weight hydrocarbons are, as a result, predominately saturated.

The effect of the catalysts on exhaust odor intensity as determined by a panel of odor judges was similar to the effects on CO and HC; that is, at light load (low exhaust temperature) the reduction in odor intensity was significantly less than at heavy load conditions. The effects of load and catalyst on odor intensity are shown in Figure 3-62. Peak and average odor intensity data are given in Table 3-47. The effect on peak odor intensity was much less pronounced for Engine 2, primarily owing to the rather high odor intensity at light loads where the catalyst had little effect.

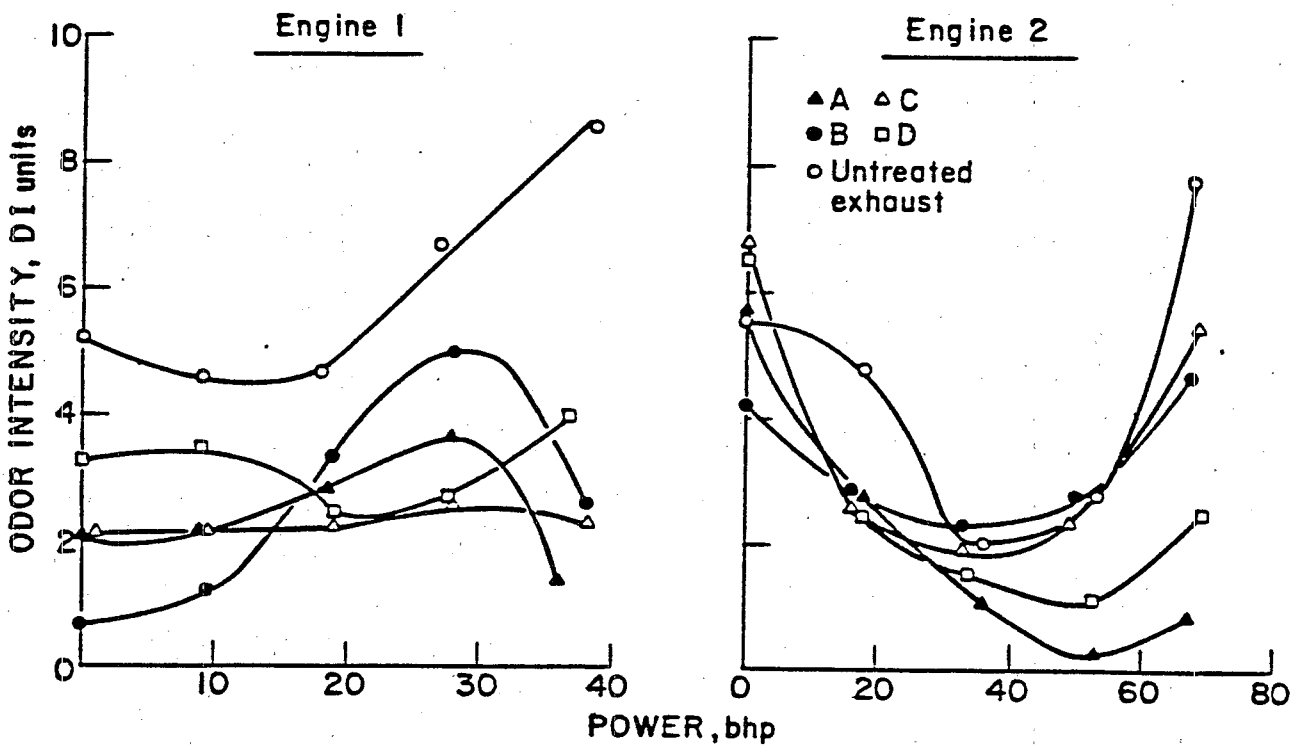


Figure 3-62. Effects of load and catalyst on odor intensity. (Marshall, et al., 1978).

TABLE 3-47. EXHAUST ODOR INTENSITY (Marshall, et al., 1978)

Exhaust Treatment System	Odor Intensity[1] Peak	Average[2]
ENGINE 1		
No catalyst	9.8	7.0
Catalyst A	3.9	2.2
Catalyst B	5.0	2.9
Catalyst C	4.7	3.1
Catalyst D	4.9	3.5
ENGINE 2		
No catalyst	8.9	6.1
Catalyst A	5.7	2.1
Catalyst B	8.9	4.1
Catalyst C	8.4	4.8
Catalyst D	7.5	3.9

[1] Dilution rate = 400/1.

[2] Diesel intensity (DI) units.

None of the catalytic reactors had any significant effect on emissions of NO, NO_2, or particulates. These findings with respect to NO_2 should not be taken as definitive, however, because there is some question as to the validity of the results obtained by the sampling and analytical technique employed in this investigation. An ambient temperature chemiluminescent analyzer was used to determine exhaust levels of NO and NO_2. To prevent condensation of water in the detector, the exhaust sample was directed through a water knock-out trap. Results of other work indicate that this procedure might yield erroneous information as to absolute levels of NO_2. This analytical technique does, however, appear to yield valid information concerning relative levels of NO_2; that is, NO_2 levels as determined in a system with minimum perturbation of the sample -- no knockout trap -- are roughly linear with the levels determined with the system used in these experiments. Therefore, the conclusion that the catalysts had no major effect on NO_2 levels appears to be reasonable

The particulate emission rate was measured directly (by a filtration-gravimetric technique) and indirectly (plume opacity). The results obtained by both techniques indicated that the catalysts generally had little or no influence on particulate emission rate. The physical characteristics of the particulates (size, size range, and distribution) were also unaffected by the various catalytic reactors. The particulate matter, which had the appearance of carbon black, had a size median diameter (number basis) of about 0.04 to 0.8 μm and a mass median diameter of 0.128 μm. The particulate size was independent of engine, fuel, speed, and load.

The aldehyde emission levels were determined by three analytical methods: total aldehydes by dinitrophenyl hydrazone, formaldehyde by chromotropic acid, and acrolein by hexyl resorcinol. The results showed that total aldehyde (RCHO) levels were typically highest at light load conditions and decreased with increasing load. The formaldehyde component accounted for 20 to nearly 100 percent of the total. Engine operating conditions influenced acrolein level in the same manner as total aldehydes -- that is, levels of acrolein generally decreased with increasing load.

The effects of the various catalysts on emissions of aldehydes were similar to effects on CO and HC -- catalyst A showed highest overall efficiency, and catalyst C showed the lowest. The results are summarized in Table 3-48. Because peak emission rates of aldehydes occurred at light load (low exhaust temperature), the catalysts had little influence, except for one case. The peak aldehyde emission rate was reduced to below the acceptable maximum (200 mg/hr/rated hp); however, in only one case (catalyst A, engine 1) was the peak acrolein emission rate below 20 mg/hr/rated hp. Average emission rates of acroelin were reduced substantially. The average ambient level of acrolein would be well below its TLV with a ventilation rate of 50 cfm/rated hp. This ventilation rate is approximately equal to that necessary to limit ambient CO_2 level to its TLV.

TABLE 3-48. ALDEHYDE EMISSIONS, mg/hr/rated hp
(Marshall, 1978)

Exhaust Treatment System	Peak Rates RCHO*	Peak Rates HCHO†	Peak Rates Acrolein	Average Rates RCHO*	Average Rates HCHO†	Average Rates Acrolein
ENGINE 1						
No catalyst	106	35	41	59	21	15
Catalyst A	86	36	16	39	7	6
Catalyst B	93	38	21	40	15	8
Catalyst C	--	32	24	--	13	10
Catalyst D	99	52	44	39	14	7
ENGINE 2						
No catalyst	243	175	124	111	78	32
Catalyst A	134	51	34	27	9	3
Catalyst B	87	38	42	47	14	10
Catalyst C	225	171	56	103	57	10
Catalyst D	103	49	45	40	28	8

* RCHO = Total Aldehydes
† HCHO = Formaldehyde

For all toxicants discussed thus far, the influence of the catalytic reactors was either beneficial or not significant. Emissions of one toxic material, however, are adversely influenced by catalysts -- SO_3 or sulfric acid (H_2SO_4). Diesel fuel contains significant levels of sulfur, which is oxidized to SO_2 in the combustion process. Although SO_2 is a toxicant, emissions can be easily controlled to tolerable levels by limiting the level of sulfur in the fuel. Even with a moderately low sulfur fuel (about 0.2 percent), however, the catalyst tend to cause oxidation of the SO_2 to SO_3. Unfortunately, the catalysts that display the greatest efficiency for oxidation of CO, aldehydes, and HC also tend to produce the greatest SO_2 conversion to SO_3. Moreover, the catalysts tend to "store" the sulfur during light-load (low temperature) operating conditions and then release much as SO_3 at heavy loads. This is evidenced by data indicating that greater quantities of sulfur are emitted as SO_2 and SO_3 than are present in the fuel that goes

into the converter at a specific time. The results of the SO_3 determinations are given in Table 3-49. The acceptable maximum value for the specific emission rate of H_2SO_4 is 80 mg/hr/rated hp. Note that for catalyst A, ventilation air flow rate of 100 cfm/rated hp would not be sufficient to dilute the H_2SO_4 to less than 0.25 ppm -- even for the "average" emission rates. Thus, it would appear that the use of catalytic reactors for emissions reduction is not toally satisfactory when using high-sulfur fuel. Additional exhaust treatment would be necessary to remove sulfates produced by the catalysts. Although this probably could be achieved with a water scrubber system installed downstream of the catalyst, this aspect of emissions control was not examined in this study.

TABLE 3-49. SULFATE[a] EMISSIONS, mg/hr/rated hp
(Marshall, et al., 1978)

Exhaust Treatment System	Peak Rates	Average Rates
ENGINE 1		
No catalyst	102	22
Catalyst A	1,044	182
Catalyst B	473	41
Catalyst C	220	53
Catalyst D	283	63
ENGINE 2		
No catalyst	47	14
Catalyst A	1,011	315
Catalyst B	702	139
Catalyst C	114	30
Catalyst D	354	96

(1) Almost entirely H_2SO_4

SECTION 4.0

PSD PERMIT APPLICATION ANALYSIS

A shale oil recovery plant encompasses a large number of widely varying types of unit operations ranging from solids crushing and handling associated with typical mining operations to gas and liquid processing and clean-up systems associated with refineries. A significant source of data for determining the environmental impact of each type of process are the applications for Prevention of Significant Deterioration (PSD) permits filed by the various developers. During the investigation into the various emission control processes for a number of the proposed facilities in the West, the data submitted in these PSD applications were entered into a computerized data base.

The purpose of this task was to determine and evaluate average values and trends for the emission rates of the criteria pollutants for each type of process. This task was complicated considerably by the fact that in each permit application a different format for presenting the essential emissions data was used. The data from each permit application were entered on a standard format so that sort routines on the various performance parameters and unit processes could be made. These parameters were evaluated for consistency and trends. For example, the NO_x emissions from vehicles for below ground mining should be similar for each of the facilities using similar mining techniques and this was found to be true. However, in other cases, such as carbon monoxide emissions from combustion, some inconsistencies in the reported data were obvious when the values were compared.

The results of these analyses are shown in Tables C-1 through C-4 in Appendix C - Summary of PSD Permit Application and Unit Process Analysis. A summary of the specific control techniques presented in the PSD Applications used for each unit process is listed in Table 4-1. To permit proper sorting for evaluation, each process was categorized by two levels -- primary and secondary. The primary categories are defined as:

1. below ground mining

TABLE 4-1. SOURCES AND CONTROLS FOR PSD APPLICATIONS

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
BELOW GROUND ACTIVITIES-MINING				
(OPERATIONS:POLLUTANTS)				
DRILLING: PARTICULATES				
UEF (UNITS)	-	0.68 (KG/HOLE)	0.009 (KG/T)	0.0045 (KG/T)
CONTROL & EFF. (%)	-	WATER INJECTION (75)	WATER SPRAY (80)	WET SUPP. & BAF. SET (87.5)
BLASTING: PARTICULATES				
UEF (UNITS)	0.107 (T/DAY)	23 (KG/BLAST)	-	0.005 (KG/T)
CONTROL & EFF. (%)	-	-	-	BAF. SET. (50)
: CO				
UEF (UNITS)	0.87 (T/DAY	0.096 (KG/KG)	0.01 (KG/KG)	0.01 (KG/KG)
CONTROL & EFF. (%)	-	NONE	NONE	NONE
: NOx				
UEF (UNITS)	0.066 (T/DAY)	0.0007 (KG/KG)	0.0054 (KG/KG)	0.0007 (KG/KG)
CONTROL & EFF. (%)	-	-	NONE	NONE
REMOVAL: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
CONVEYING: PARTICULATES				
UEF (UNITS)	0.025 (KG/T)	0.1 (KG/T)	-	0.1 (KG/T)
CONTROL & EFF. (%)	BAGHOUSE (99.5) CHEM. SPRAY (85)	ENC. CYCLONE (93) BAGHOUSE (99.9)	-	BAGHOUSE & BAF. SET (99.5)

Table 4-1. Continued

PROJECT:		CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:		RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):		1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:		CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
TRANSFERING:	PARTICULATES				
	UEF (UNITS)	0.075 (KG/T)	-	9.2 (G/NM3)	-
	CONTROL & EFF. (%)	BAGHOUSE (99.5) CHEM. SPRAY (85)	-	BAGHOUSE (99.7)	-
PRIMARY CRUSH & SCREEN:	PARTICULATES				
	UEF (UNITS)	-	0.02 (KG/T)	13.7 (G/NM3)	0.25 (KG/T)
	CONTROL & EFF. (%)	-	BAGHOUSE (99.9)	WETTED FAN SCRUBBER (99.9)	BAGHOUSE & BAF. SET (99.5)
SECONDARY CRUSH & SCREEN:	PARTICULATES				
	UEF (UNITS)	-	-	-	0.75 (KG/T)
	CONTROL & EFF. (%)	-	-	-	BAGHOUSE & BAF. SET (99.6)
IN-MINE VEHICLES:	PARTICULATES				
	UEF (UNITS)	-	-	26 E-04 (KG/L)	26 E-04 (KG/L)
	CONTROL & EFF. (%)	-	-	-	WET SUPP. & BAF. SET (87.5)
:	CO				
	UEF (UNITS)	-	0.197 (KG/HR)	11 E-03 (KG/L)	11 E-03 (KG/L)
	CONTROL & EFF. (%)	-	NONE	NONE	NONE
:	HC				
	UEF (UNITS)	-	0.073 (KG/HR)	34 E-04 (KG/L)	34 E-04 (KG/L)
	CONTROL & EFF. (%)	-	NONE	NONE	NONE
:	NOx				
	UEF (UNITS)	-	0.91 (KG/HR)	51 E-03 (KG/L)	51 E-03 (KG/L)
	CONTROL & EFF. (%)	-	NONE	NONE	NONE
:	SO_2				
	UEF (UNITS)	-	0.06 (KG/HR)	37 E-04 (KG/L)	37 E-04 (KG/L)
	CONTROL & EFF. (%)	-	NONE	NONE	NONE

UEF = UNCONTROLLED EMISSION FACTORS
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO. 1982	BECHTEL PETROLEUM INC. 1981
BELOW GROUND ACTIVITIES-MINING				
(OPERATIONS:POLLUTANTS)				
DRILLING: PARTICULATES				
UEF (UNITS)	0.009 (KG/T)	0.009 (KG/T)	0.68 (KG/HOLE)	0.009 (KG/T)
CONTROL & EFF. (%)	WATER SPRAY (88)	WET SUPP. & BAF SET (80)	DUST COLL. & WET SUPP. (90)	WET SUPP. & BAF. SET (80)
BLASTING: PARTICULATES				
UEF (UNITS)	2.3 (KG/BLAST)	0.005 (KG/T)	0.0013 (KG/T)	-
CONTROL & EFF. (%)	PREVENT OVERSHOOTING (N/A)	BAF. SET (50)	(SETTLING 94.3)	-
: CO				
UEF (UNITS)	-	0.01 (KG/KG)	-	0.01 (KG/KG)
CONTROL & EFF. (%)	-	MINIMIZE FREQUENCY (N/A)	-	NONE
: NOx				
UEF (UNITS)	-	0.054 (KG/KG)	-	0.0054 (KG/KG)
CONTROL & EFF. (%)	-	MINIMIZE FREQUENCY (N/A)	-	NONE
REMOVAL: PARTICULATES				
UEF (UNITS)	0.0185 (KG/T)	-	0.0185 (KG/T)	-
CONTROL & EFF. (%)	MINIMIZE FALL DISTANCE (N/A)	-	(SETTLING 94.3)	-
CONVEYING: PARTICULATES				
UEF (UNITS)	-	-	0.1 (KG/T)	9 (G/NM3)
CONTROL & EFF. (%)	-	-	ENCLOSED (70)	BAGHOUSE (99.7)
TRANSFERING: PARTICULATES				
UEF (UNITS)	0.0001 (KG/T)	-	-	0.00045 (KG/T)
CONTROL & EFF. (%)	BAGHOUSE (99.7)	-	-	WET SUPP. (90)
PRIMARY CRUSH & SCREEN: PARTICULATES				
UEF (UNITS)	0.01 (KG/T)	7 (G/NM3)	0.04 (KG/T)	0.01 (KG/T)
CONTROL & EFF. (%)	VENT DROPOUT & BAGHOUSE (99.7)	BAGHOUSE (99.5)	ENCLOSED & WATER (70)	WET SUPP. & BAF. SET (80)
SECONDARY CRUSH & SCREEN: PARTICULATES				
UEF (UNITS)	-	-	-	14 (G/NM3)
CONTROL & EFF. (%)	-	-	-	BAGHOUSE (99.7)
IN-MINE VEHICLES: PARTICULATES				
UEF (UNITS)	-	26 E-04 (KG/L)	-	26 E-04 (KG/L)
CONTROL & EFF. (%)	-	TUNE ENGINES (N/A)	-	NONE

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO. 1982	BECHTEL PETROLEUM INC. 1981
: CO				
UEF (UNITS)	-	11 E-03 (KG/L)	-	11 E-03 (KG/L)
CONTROL & EFF. (%)	-	TUNE ENGINES (N/A)	-	NONE
: HC				
UEF (UNITS)	-	34 E-04 (KG/L)	-	34 E-04 (KG/L)
CONTROL & EFF. (%)	-	TUNE ENGINES (N/A)	-	NONE
: NOx				
UEF (UNITS)	-	51 E-03 (KG/L)	-	51 E-03 (KG/L)
CONTROL & EFF. (%)	-	TUNE ENGINES (N/A)	-	NONE
: SO_2				
UEF (UNITS)	-	37 E-04 (KG/L)	-	37 E-04 (KG/L)
CONTROL & EFF. (%)	-	TUNE ENGINES (N/A)	-	NONE

UEF = UNCONTROLLED EMISSION FACTORS
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
MATERIAL HANDLING & PREPARATION-SURFACE				
SURFACE SOILS-REMOVAL: PARTICULATES				
UEF (UNITS)	226 (G/M2)	189 (G/M2)	0.27 (KG/M2-MONTH)	-
CONTROL & EFF. (%)	NONE	-	WATER SUPP. (N/A)	-
-LOAD: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
-HAUL: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
-WIND: PARTICULATES				
UEF (UNITS)	0.076 (KG/T)	0.027 (KG/M2-YR)	-	0.00096 (KG/M2-DAY)
CONTROL & EFF. (%)	CHEM. SPRAY (50)	WATER (50) MULCH (85)	-	VEG. COVER (75)
RAW SHALE-CRUSH & SCREEEN: PARTICULATES				
UEF (UNITS)	97 (G/DAY)	0.06 (KG/T)	16 (G/NM3)	0.005 (KG/T)
CONTROL & EFF. (%)	BAGHOUSE (99.5)	BAGHOUSE (99.9)	BAGHOUSE (99.7)	BAGHOUSE (99.9)
SECONDARY & TERTIARY C & S: PARTICULATES				
UEF (UNITS)	97 (G/DAY)	0.08 (KG/T)	-	0.75 (KG/T)
CONTROL & EFF. (%)	BAGHOUSE (99.5)	BAGHOUSE (99.9)	-	BAGHOUSE (99.9)
-CONVEY & TRANSF: PARTICULATES				
UEF (UNITS)	2 (T/DAY)	0.1 (KG/T)	0.0001 (KG/T)	0.1 (KG/T)
CONTROL & EFF. (%)	BAGHOUSE (99.5)	ENC. CYC. (93) BAGHOUSE (99.9)	WET SUPP. (90)	BAGHOUSE (99.8)
-HAUL: PARTICULATES				
UEF (UNITS)	-	0.72 (KG/VMT)	-	-
CONTROL & EFF. (%)	-	CHEM. SUPP. (85)	-	-
-DUMP: PARTICULATES				
UEF (UNITS)	0.0063 (T/DAY)	-	-	-
CONTROL & EFF. (%)	CHEM. SPRAY (85)	-	-	-
-STORAGE (LIVE): PARTICULATES				
UEF (UNITS)	-	-	0.0002 (KG/T)	0.0013 (KG/T)
CONTROL & EFF. (%)	-	-	WET SUPP. (50) DROP CHUTE	COMPACTION & ENC. (99.5)

Table 4-1. Continued

PROJECT:		CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:		RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):		1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:		CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
(WIND):	PARTICULATES			(75)	
	UEF (UNITS)	0.0072 (T/DAY)	0.19 (T/YR)	0.2 (KG/T)	-
	CONTROL & EFF. (%)	NONE	SHELTER (50)	WATER (50)	-
(MAINTENANCE):	PARTICULATES				
	UEF (UNITS)	-	-	0.0002 (KG/T)	-
	CONTROL & EFF. (%)	-	-	WET SUPP. (50) DROP CHUTE	-
-STORAGE (DEAD):	PARTICULATES			(75)	
	UEF (UNITS)	-	-	-	0.0013 (KG/T)
	CONTROL & EFF. (%)	-	-	-	NONE
(WIND):	PARTICULATES				
	UEF (UNITS)	0.0078 (G/SEC)	0.026 (T/YR)	0.182 (KG/YR)	0.00035 (KG/T)
	CONTROL & EFF. (%)	NONE	NONE	CHEMICAL (95)	NONE
RETORT FEED-SILOS:	PARTICULATES				
	UEF (UNITS)	-	-	7 (G/NM3)	0.1 (KG/T)
	CONTROL & EFF. (%)	-	-	BAGHOUSE (99.7)	BAGHOUSE (99.1)
-CONVEYOR:	PARTICULATES				
	UEF (UNITS)	0.3 (G/SEC)	-	-	-
	CONTROL & EFF. (%)	N/A	-	-	-
-DISCHARGE:	PARTICULATES				
	UEF (UNITS)	-	-	-	0.1 (KG/T)
	CONTROL & EFF. (%)	-	-	-	WET SCRUBBER (95)

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
\- = NOT INDICATED

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
OETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
MATERIAL HANDLING & PREPARATION-SURFACE				
SURFACE SOILS-REMOVAL: PARTICULATES				
UEF (UNITS)	7 (KG/SCRAPER-HR)	-	-	-
CONTROL & EFF. (%)	-	-	-	-
-LOAD: PARTICULATES				
UEF (UNITS)	-	-	0.06 (KG/T)	-
CONTROL & EFF. (%)	-	-	NONE	-
-HAUL: PARTICULATES				
UEF (UNITS)	-	-	0.23 (KG/VMT)	-
CONTROL & EFF. (%)	-	-	N/A (85)	-
-WIND: PARTICULATES				
UEF (UNITS)	N/A	-	0.015 (KG/M2-YR)	-
CONTROL & EFF. (%)	VEG. COVER & WIND BREAK (50)	-	N/A (50)	-
RAW SHALE-CRUSH & SCREEEN: PARTICULATES				
UEF (UNITS)	-	7 (G/NM3)	-	16 (G/NM3)
CONTROL & EFF. (%)	-	BAGHOUSE (99.5)	-	BAGHOUSE (99.7)
SECONDARY & TERTIARY C & S: PARTICULATES				
UEF (UNITS)	-	7 (G/NM3)	-	-
CONTROL & EFF. (%)	-	BAGHOUSE (99.5)	-	-
-CONVEY & TRANSF: PARTICULATES				
UEF (UNITS)	0.1 (KG/T)	0.00045 (KG/T)	-	-
CONTROL & EFF. (%)	COVERED (99)	WET SUPP. (90)	-	-
-HAUL: PARTICULATES				
UEF (UNITS)	-	1.2 (KG/VMT)	0.05 (KG/VMT)	-
CONTROL & EFF. (%)	-	CHEM. SUPP. (85)	WATER & WET CHEM. (85)	-
-DUMP: PARTICULATES				
UEF (UNITS)	-	-	0.02 (KG/T)	-
CONTROL & EFF. (%)	-	-	WET SUPP. (90)	-
-STORAGE (LIVE): PARTICULATES				
UEF (UNITS)	0.0065 (KG/T)	0.0045 (KG/T)	LOAD IN 0.0115, LOAD-OUT	0.0045 (KG/T)
CONTROL & EFF. (%)	WATER & DROP CHUTE (97.4)	WET SUPP. (90)	0.00025 (KG/T) N/A (75)	LOWERING WELL (75)
(WIND): PARTICULATES				
UEF (UNITS)	0.0008 (KG/T)	0.16 (KG/T)	0.015 (KG/M2-YR)	0.16 (KG/T)
CONTROL & EFF. (%)	WATER (50)	CHEM. SUPP. (95)	N/A	WATERING (50)
(MAINTENANCE): PARTICULATES				

Table 4-1. Continued

PROJECT:		SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER
LOCATION:		CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
OETORTED SHALE OIL (CUBIC METERS PER DAY):		7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:		HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
	UEF (UNITS)	-	-	-	0.01 (KG/T)
	CONTROL & EFF. (%)	-	-	-	WATERING (50)
-STORAGE (DEAD):	PARTICULATES				
	UEF (UNITS)	-	-	-	-
	CONTROL & EFF. (%)	-	-	-	-
(WIND):	PARTICULATES				
	UEF (UNITS)	0.0138 (KG/T)	-	-	0.16 (KG/T)
	CONTROL & EFF. (%)	POLYMER (90)	-	-	WATERING (50)
RETORT FEED-SILOS:	PARTICULATES				
	UEF (UNITS)	-	7 (G/NM3)	-	7 (G/NM3)
	CONTROL & EFF. (%)	-	BAGHOUSE (99.5)	-	BAGHOUSE (99.7)
-CONVEYOR:	PARTICULATES				
	UEF (UNITS)	-	-	-	0.003 (KG/T)
	CONTROL & EFF. (%)	-	-	-	WET SUPP. (90)
-DISCHARGE:	PARTICULATES				
	UEF (UNITS)	-	-	-	-
	CONTROL & EFF. (%)	-	-	-	-

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued.

PROJECT:		CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:		RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):		1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:		CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
MATERIAL HANDLING & PREPARATION -SURFACE (CONT)					
FINES-CONVEYOR & TRANSFER:	PARTICULATES				
	UEF (UNITS)	0.0035 (KG/T)	-	9 (G/NM3)	0.1 (KG/T)
	CONTROL & EFF. (%)	BAGHOUSE (99.5)	-	BAGHOUSE (99.7)	BAGHOUSE (99.1) SCRUBBER 99.9)
-STORAGE:	PARTICULATES				
	UEF (UNITS)	0.82 (T/DAY)	-	-	0.004 (KG/T)
	CONTROL & EFF. (%)	BIN FILTER (99)	-	-	WET SUPP. (75)
(WIND):	PARTICULATES				
	UEF (UNITS)	-	-	-	0.00096 (KG/M2-DAY)
	CONTROL & EFF. (%)	-	-	-	WET SUPP. (75)
(MAINTENANCE):	PARTICULATES				
	UEF (UNITS)	-	-	-	7 (KG/SCRAPER-HR)
	CONTROL & EFF. (%)	-	-	-	WET SUPP. (75)
RETORTED SHALE-CONVEYER:	PARTICULATES				
	UEF (UNITS)	0.82 (G/SEC)	0.1 (KG/T)	-	0.1 (KG/T)
	CONTROL & EFF. (%)	-	12% WATER (99)	-	BAGHOUSE (99.9)
-TRANSFER:	PARTICULATES				
	UEF (UNITS)	13 (T/DAY)	-	0.00001 (KG/T)	0.1 (KG/T)
	CONTROL & EFF. (%)	SCRUBBER (98)	-	WATER SPRAY (90)	WET SCRUBBER (99.7)
-HAUL:	PARTICULATES				
	UEF (UNITS)	0.0035 (KG/T)	0.72 (KG/VMT)	-	-
	CONTROL & EFF. (%)	NONE	CHEM. SUPP. (85)	-	-
-DUMP:	PARTICULATES				
	UEF (UNITS)	-	0.00004 (KG/T)	-	-
	CONTROL & EFF. (%)	-	-	-	-
-STORAGE:	PARTICULATES				
	UEF (UNITS)	1 (G/SEC)	-	0.00001 (KG/T)	0.00075 (KG/T)
	CONTROL & EFF. (%)	NONE	-	LOWER BOOM (75)	WET SUPP. (90)
(WIND):	PARTICULATES				
	UEF (UNITS	0.71 (KG/M2-YR)	0.0143 (KG/M2-YR)	0.126 (KG/M2-YR)	0.00096 (KG/M2-DAY)
	CONTROL & EFF. (%)	WATER SPRAY (50)	WATERING (50)	COMP. & CHEM. (85)	WET SUPP. (90)
(MAINTENANCE):	PARTICULATES				
	UEF (UNITS)	-	-	7 (KG/SCRAPER-HR)	7 (KG/SCRAPER-HR)
	CONTROL & EFF. (%)	-	-	WATER (50)	WET SUPP. (90)

Table 4-1. Continued.

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
COMBUSTION SOURCES				
POWER GENERATION: PARTICULATES				
UEF (UNITS	5 (G/SEC)	13 (NG/J)	-	6 (NG/J)
CONTROL & EFF. (%)	NONE	BAGHOUSE (N/A)	-	NONE
: CO				
UEF (UNITS)	8 (G/SEC)	N/A	-	7 (NG/J)
CONTROL & EFF. (%)	NONE	OXYGEN CONTROL (N/A)	-	NONE
: HC				
UEF (UNITS)	15 (G/SEC)	-	-	1 (NG/J)
CONTROL & EFF. (%)	NONE	-	-	NONE
: NOx				
UEF (UNITS	338 (G/SEC)	300 (NG/J)	-	2454 (NG/J)
CONTROL & EFF. (%)	NONE	LOW NOX BURNER (N/A)	-	H2O WASH (97)
: SO_2				
UEF (UNITS)	50 (G/SEC)	82 (NG/J)	-	2531 (NG/J)
CONTROL & EFF. (%)	NONE	LIMESTONE SCRUBBER (N/A)	-	STRETFORD (97.1)

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UNITA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
MATERIAL HANDLING & PREPARATION -SURFACE (CONT)				
FINES-CONVEYOR & TRANSFER:	PARTICULATES			
	UEF (UNITS)	-	0.02 (KG/T)	-	0:.022 (KG/T)
	CONTROL & EFF. (%)	-	WET SUPP. (90)	-	WET SUPP. (90)
-STORAGE:	PARTICULATES				
	UEF (UNITS)	-	0.02 (KG/T)	-	0.013 (KG/T)
	CONTROL & EFF. (%)	-	WET SUPP. (90)	-	WATERING (50)
(WIND):	PARTICULATES				
	UEF (UNITS)	-	1.4 (KG/M2-YR)	-	1.4 (KG/M2-YR)
	CONTROL & EFF. (%)	-	COMP. & CHEM. SUPP. (85)	-	WATERING (50)
(MAINTENANCE):	PARTICULATES				
	UEF (UNITS)	-	7 (KG/SCRAPER-YR)	-	-
	CONTROL & EFF. (%)	-	WATER (50)	-	-
RETORTED SHALE-CONVEYER:	PARTICULATES				
	UEF (UNITS)	-	0.0001 (KG/T)	0.0115 (KG/T)	-
	CONTROL & EFF. (%)	-	INC. CONTROL-MOIST 15%	BAGHOUSE (97.5)	-
-TRANSFER:	PARTICULATES				
	UEF (UNITS)	N/A	0.0001 (KG/T)	-	0.00015 (KG/T)
	CONTROL & EFF. (%)	VENTURI SCRUBBER (98.4)	INC. CONTROL-MOIST 15%	-	MOIST. 20% WATER
-HAUL:	PARTICULATES				
	UEF (UNITS)	-	-	0.89 (KG/VMT)	1 (KG/VMT)
	CONTROL & EFF. (%)	-	-	WATER & CHEM. (85)	NONE
-DUMP:	PARTICULATES				
	UEF (UNITS)	-	-	0.02 (KG/T)	0.00015 (KG/T)
	CONTROL & EFF. (%)	-	-	WET SUPP. (90)	-
-STORAGE:	PARTICULATES				
	UEF (UNITS)	-	0.0001 (KG/T)	-	0.00015 (KG/T)
	CONTROL & EFF. (%)	-	INC. CONTROL-MOIST 15%	-	15% WATER
(WIND):	PARTICULATES				
	UEF (UNITS	0.108 (KG/M2-YR)	0.00096 (KG/M2-DAY)	0.0029 (KG/M2-YR)	1.9 (KG/M2-YR)
	CONTROL & EFF. (%)	CHEM. SUPP. (83)	CHEM. SUPP. (85)	N/A (50)	COMP. & CHEM. SUPP. (85)
(MAINTENANCE):	PARTICULATES				
	UEF (UNITS)	-	7 (KG/SCRAPER-YR)	7 (KG/HR)	7 (KG/SCRAPER-HR)
	CONTROL & EFF. (%)	-	WATER SPRAY (50)	N/A (90)	WATERING (50)

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UNITA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
COMBUSTION SOURCES				
POWER GENERATION: PARTICULATES				
UEF (UNITS	-	2.2 (KG/HR)	7 (NG/J)	13 (NG/J)
CONTROL & EFF. (%)	-	NONE	GAS VS LIQUID FUEL (70)	-
: CO				
UEF (UNITS)	-	2.5 (KG/HR)	64 (NG/J)	27 E-05 (KG/NM3) & 60 E-(KG/L)
CONTROL & EFF. (%)	-	NONE	- -	
: HC				
UEF (UNITS)	-	0.44 (KG/HR)	4 (NG/J)	2 E-06 (KG/NM3) & 12 E-05 (KG/L)
CONTROL & EFF. (%)	-	NONE	GAS VS LIQUID FUEL (70)	-
: NOx				
UEF (UNITS	-	29.9 (KG/HR)	86 (NG/J)	86 (NG/J)
CONTROL & EFF. (%)	-	NONE	LOW NOX BURNER (17)	-
: SO_2				
UEF (UNITS)	-	-	-	43 E-05 (KG/NM3) & 34 E-06 (KG/L)
CONTROL & EFF. (%)	-	-	-	-

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
COMBUSTION SOURCES (CONT)				
FLARE: PARTICULATES				
UEF (UNITS)	0.0154 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: CO				
UEF (UNITS)	0.031 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: HC				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
: NOx				
UEF (UNITS)	0.45 (G.SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: SO_2				
UEF (UNITS)	2.5 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
RETORTING-GAS TREATMENT PLANT: PARTICULATES				
UEF (UNITS)	5.1 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: CO				
UEF (UNITS)	8.6 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: HC				
UEF (UNITS)	1.5 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: NOX				
UEF (UNITS)	125.8 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
: SO2				
UEF (UNITS)	53.5 (G/SEC)	-	-	-
CONTROL & EFF. (%)	NONE	-	-	-
HEATERS & LIFT PIPES: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
: CO				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
: HC				
UEF (UNITS)	-	-	-	-

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
CONTROL & EFF. (%)	-	-	-	-
: NOx				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
: SO_2				
UEF (UNITS)	-		-	-
CONTROL & EFF. (%)	-	-	-	-
-TOSCO ELUTRIATORS: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
-TOSCO MOISTURERS: PARTICULATES				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
\- = NOT INDICATED

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
COMBUSTION SOURCES (CONT)				
FLARE: PARTICULATES				
UEF (UNITS)	-	24 E-05 (KG/NM3)	22 NG/J)	24 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
: CO				
UEF (UNITS)	-	27 E-05 (KG/NM3)	7 (NG/J)	27 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
: HC				
UEF (UNITS)	-	48 E-06 (KG/NM3)	1 (NG/J)	48 E-06 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
: NOx				
UEF (UNITS)	-	37 E-04 (KG/NM3)	58 (NG/J)	37 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	-	-
: SO_2				
UEF (UNITS)	-	22 E-04 (KG/NM3)	-	22 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
RETORTING-GAS TREATMENT PLANT: PARTICULATES				
UEF (UNITS)	-	N/A	-	24 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	BAGHOUSE (90)	-	-
: CO				
UEF (UNITS)	-	-	-	27 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
: HC				
UEF (UNITS)	-	-	-	48 E-06 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-
: NOx				
UEF (UNITS)	-	-		22 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	-		-
: SO_2				
UEF (UNITS)	-	N/A	-	10-20 PPMV
CONTROL & EFF. (%)	-	LIME SLURRY (90)	-	TAIL GAS UNIT (<250 PPM)
HEATERS & LIFT PIPES: PARTICULATES				
UEF (UNITS)	-	N/A	-	148 (G/NM3)
CONTROL & EFF. (%)	-	CYC. (93.8) SCRUBBER (99.5.)	-	CYC. (93.8) SCRUBBER (99.5.)
: CO				
UEF (UNITS)	-	-	-	99 E-05 (KG/L) & 27 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	-

Table 4-1. Continued

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
: HC				
UEF (UNITS)	-	N/A	-	0.37 (KG/T)
CONTROL & EFF. (%)	-	TREATED GAS & INCINERATION (85)	-	INCINERATION (85)
: NOx				
UEF (UNITS)	-	-	-	61 E-04 (KG/L) & 34 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	-	-	LOW NOX BURNERS (N/A)
: SO_2				
UEF (UNITS)	-	N/A	-	55 E-06 (KG/L) & 22 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	TREATED GAS & INCINERATION (85)	-	-
-TOSCO ELUTRIATORS: PARTICULATES				
UEF (UNITS)	-	-	-	1 % FEED
CONTROL & EFF. (%)	-	-	-	CYC. (93.8) SCRUBBER (99.5)
-TOSCO MOISTURERS: PARTICULATES				
UEF (UNITS)	-	9 (G/NM3)	-	9 G/NM3)
CONTROL & EFF. (%)	-	VENTURI SCRUBBER (95.5)	-	VENTURI SCRUBBER (95.5)

UEF = UNCONTROLLED EMISSION FACTOR
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
COMBUSTION SOURCES (CONT)				
UPGRADING-HEATERS: PARTICULATES				
UEF (UNITS)	-	-	0.0028 (KG/T)	6 (NG/J)
CONTROL & EFF. (%)	-	-	BAGHOUSE (99.7)	NONE
: CO				
UEF (UNITS)	-	N/A	85 E-05 (KG/NM3)	7 (NG/J)
CONTROL & EFF. (%)	-	OXYGEN CONTROL (N/A)	NONE	NONE
: HC				
UEF (UNITS)	-	-	34 E-05 (KG/NM3)	1 (NG/J)
CONTROL & EFF. (%)	-	-	NONE	NONE
: NOx				
UEF (UNITS)	-	9 (NG/J)	34 E-04 (KG/NM3)	99 (NG/J)
CONTROL & EFF. (%)	-	LOW NOX BURNERS (N/A)	NONE	NONE
: SO_2				
UEF (UNITS)	-	17 (NG/J)	34 E-03 (KG/NM3)	0.26 (NG/J)
CONTROL & EFF. (%)	-	SCRUBBERS, HYDROTREATING (N/A)	STRETFORD (99.7)	STRETFORD (N/A)
MISCELLANOUS				
VALVES, FLANGES, PUMPS, ETC.: HC				
UEF (UNITS)	-	0.0114 (KG/HR)	0.143 (KG/NM3)	N/A
CONTROL & EFF. (%)	-	MAINTENANCE (70)	MAINTENANCE (50)	DOUBLE SEALS (N/A)
API SEPARATORS: HC				
UEF (UNITS)	-	24 E-06 (KG/L)	-	-
CONTROL & EFF. (%)	-	COVERED (95)	-	-
STORAGE-CRUDE SHALE OIL: HC				
UEF (UNITS)	-	N/A	-	2 (T/YR)
CONTROL & EFF. (%)	-	FLOATING ROOF (N/A)	-	FLOATING ROOF. SEALS (N/A)
-HYDROTREATED SHALE OIL: HC				
UEF (UNITS)	N/A	N/A	N/A	3 (T/YR)
CONTROL & EFF. (%)	CONE ROOF (N/A)	FLOATING ROOF (N/A)	FLOATING ROOF (N/A)	FLOATING ROOF. SEALS (N/A)
-FUEL OIL: HC				
UEF (UNITS)	N/A	N/A	N/A	0.30 (KG/DAY) FILL LOSS ONLY
CONTROL & EFF. (%)	CONE ROOF (N/A)	FIXED ROOF (N/A)	FLOATING ROOF (N/A)	FIXED ROOF (N/A)
-DIESEL FUEL: HC				
UEF (UNITS)	N/A	N/A	N/A	11 E-07 (KG/L) FILL LOSS ONLY
CONTROL & EFF. (%)	CONE ROOF (N/A)	FIXED ROOF (N/A)	FIXED ROOF (N/A)	BURIED (N/A)

Table 4-1. Continued

PROJECT:	CATHEDRAL BLUFFS	CLEAR CREEK	COTTONWOOD	PARAHO UTE FACILITY
LOCATION:	RIO BLANCO COUNTY, COLORADO	GRAND VALLEY, COLORADO	GREEN RIVER BASIN, UTAH	UINTA BASIN, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	1908	15,900	5004	6678
BIBLIOGRAPHY REFERENCE:	CATH. BLUFFS CO., 1982	ENV. & RESEARCH TECH., 1982	MAGIC CIRCLE ENERGY, 1982	PARAHO DEVELOPMENT, 1981
-GASOLINE: HC				
UEF (UNITS)	N/A	-	N/A	60 E-05 (KG/L) FILL LOSS ONLY
CONTROL & EFF. (%)	CONE ROOF (N/A)	-	FIXED ROOF (N/A)	BURIED (N/A)
-NAPHTHA: HC				
UEF (UNITS)	-	-	-	-
CONTROL & EFF. (%)	-	-	-	-
COOLING TOWERS: HC				
UEF (UNITS)	-	72 E-06 (KG/L)	-	-
CONTROL & EFF. (%)	-	-	-	-
ON-SITE VEHICLES-HDT: PARTICULATES				
UEF (UNITS)	0.084 (KG/DAY)	2.1 (KG/VMT)	16 E-04 (KG/L)	0.54 (KG/VMT)
CONTROL & EFF. (%)	NONE	CHEM. SUPP. (85)	MAINTENANCE (N/A)	WATER (50)
-LDT: PARTICULATES				
UEF (UNITS)	-	0.86 (KG/VMT)	0.77 (G/VMT)	0.027 (KG/VMT)
CONTROL & EFF. (%)	-	CHEM. SUPP. (85)	MAINTENANCE (N/A)	WATER (50)
-BUS: PARTICULATES				
UEF (UNITS)	-	2.1 (KG/VMT)	-	0.54 (KG/VMT)
CONTROL & EFF. (%)	-	CHEM. SUPP. (85)	-	WATER (50)

UEF = UNCONTROLLED EMISSION FACTORS
N/A = NOT AVAILABLE
- = NOT INDICATED

Table 4-1. Continued.

PROJECT:	SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:	CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):	7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:	HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
COMBUSTION SOURCES (CONT)				
UPGRADING-HEATERS: PARTICULATES				
UEF (UNITS)	132 (KG/DAY)	0.0136 (KG/HR)	7 (NG/J)	24 E-03 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	GAS VS LIQUID FUEL (70)	-
: CO				
UEF (UNITS)	118 (KG/DAY)	0.15 (KG/HR)	18 (NG/J)	27 E-05 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	GAS VS LIQUID FUEL (60)	-
: HC				
UEF (UNITS)	19 (KG/DAY)	0.027 (KG/HR)	4 (NG/J)	48 E-06 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	GAS VS LIQUID FUEL (70)	-
: NOx				
UEF (UNITS)	999 (KG/DAY)	1.8 (KG/HR)	86 (NG/J)	37 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	LOW NOX BURNER (35)	-
: SO_2				
UEF (UNITS)	327 (KG/DAY)	4.2 (KG/HR)	343 (NG/J)	22 E-04 (KG/NM3)
CONTROL & EFF. (%)	-	NONE	PSA GAS/FUEL GAS MIX (99)	-
MISCELLANOUS				
VALVES, FLANGES, PUMPS, ETC.: HC				
UEF (UNITS)	-	11 E-02 (KG/NM3)	80 E-03 (KG/NM3)	14 E-02 (KG/NM3)
CONTROL & EFF. (%)	-	MAINTENANCE (50)	INSPECT. & MAINT. (50)	MAINTENANCE (50)
API SEPARATORS: HC				
UEF (UNITS)	-	-	60 E-05 (KG/L)	-
CONTROL & EFF. (%)	-	-	COVERED (96)	-
STORAGE-CRUDE SHALE OIL: HC				
UEF (UNITS)	-	92 E-05 (KG/NM3-YR)	N/A	N/A
CONTROL & EFF. (%)	-	FLOATING ROOF (N/A)	FLOATING ROOF. SEALS (N/A)	FLOATING ROOF (N/A)
-HYDROTREATED SHALE OIL: HC				
UEF (UNITS)	-	92 E-05 (KG/NM3-YR)	N/A	N/A
CONTROL & EFF. (%)	-	FLOATING ROOF (N/A)	FLOATING ROOF. SEALS (N/A)	FLOATING ROOF (N/A)
-FUEL OIL: HC				
UEF (UNITS)	-	-	N/A	N/A
CONTROL & EFF. (%)	-	-	VAPOR COLL. & COMB. (100)	FIXED ROOF (N/A)
-DIESEL FUEL: HC				
UEF (UNITS)	-	-	-	N/A
CONTROL & EFF. (%)	-	-	-	FIXED ROOF (N/A)

Table 4-1. Continued.

PROJECT:		SAND WASH	SYNTANA	UNION FACILITY	WHITE RIVER PROJECT
LOCATION:		CENTRAL UINTA BASIN, UTAH	UINTA COUNTY, UTAH	PARACHUTE CREEK, COLORADO	VERNAL, UTAH
RETORTED SHALE OIL (CUBIC METERS PER DAY):		7155	9063	14,310	16,890
BIBLIOGRAPHY REFERENCE:		HOLMES, J.G., 1981	QUINTANA MINERALS, 1984	UNION OIL CO., 1982	BECHTEL PETROLEUM, 1981
-GASOLINE:	HC				
	UEF (UNITS)	-	-	-	-
	CONTROL & EFF. (%)	-	-	-	-
-NAPHTHA:	HC				
	UEF (UNITS)		-	N/A	N/A
	CONTROL & EFF. (%)	-	-	FLOATING ROOF. SEALS (N/A)	FLOATING ROOF (N/A)
COOLING TOWERS:	HC				
	UEF (UNITS)	-	-	72 E-05 (KG/L)	-
	CONTROL & EFF. (%)	-	-	MONITOR (90)	-
ON-SITE VEHICLES-HDT:	PARTICULATES				
	UEF (UNITS)	-	-	0.54 (KG/VMT)	-
	CONTROL & EFF. (%)	-	-	-	-
-LDT:	PARTICULATES				
	UEF (UNITS)	-	-	0.15 (KG/VMT)	-
	CONTROL & EFF. (%)	-	-	WATER (50)	-
-BUS:	PARTICULATES				
	UEF (UNITS)	-	-	0.54 (KG/VMT)	-
	CONTROL & EFF. (%)	-	-	-	-

UEF = UNCONTROLLED EMISSION FACTORS
N/A = NOT AVAILABLE
- = NOT INDICATED

2. above ground mining
3. retorting and gas utilization upgrading, and
4. miscellaneous

There are 54 secondary categories and these are noted for each process listed in Table 4-1. These categories are described in Appendix C.

The following discussion evaluates the emission rates for the criteria pollutants for each unit process reported by the seven projects included in this analysis. Details of the individual processes are presented in Section 5.0, Process Analysis.

It is important for the reader to note that this analysis is based on data presented in the PSD permit applications and is not necessarily representative of either the final permitted emission rates or the actual operating emission rates. In some cases, the final permitted emission rates were obtained and a comparison made between the permit application and the final permit. This analysis is presented at the end of Section 4.0.

4.1 PARTICULATE EMISSIONS

The PSD application data for particulate emissions from each of the seven developers are presented in Table 4-2. The control of particulate emissions from oil shale mining operations is similar to other mining processes. The reader is referred to Section 3 for a detailed analysis of these emissions sources, emission factors and control techniques. The following discussion will present data analysis based on the PSD Applications. This information is presented to show the degree of variability and estimated emission ranges for particulates and to emphasize particular processes that can have a significant affect on the overall emission rate.

TABLE 4-2. CONTROLLED PARTICULATE EMISSIONS*
kg/1000 m^3 of Oil

	Mining & Material Handling			Process			
Project	Below	Above	Total	Retort & Upgrade	Misc.	Total	Facility
Cathedral Bluffs	89	190	280	53	1	54	330
Clear Creek	49	135	184	660	11	671	855
Cottonwood	48	104	152	8	13	21	173
Paraho	64	217	281	72		72	353
Syntana	49	98	147	60	30	90	237
Union B	70	55	125	14	42	56	181
White River	42	56	98	115		115	213
Average	59	122	181	140	20	160	341

*Based on PSD Permit Applications

4.1.1 Mining - Below Ground (Primary Category 1)

Below ground mining emissions varied from 42 kg/1000 m^3 of oil to 89 kg/1000 m^3 with an average value of 59 kg/1000 m^3 of oil. These emissions are the result of the unit operations listed in Table 4-3.

TABLE 4-3. CONTROLLED BELOW GROUND PARTICULATE EMISSIONS††

	Secondary Category*	Average kg/1000 m^3 Of Oil	Avg. % of Total Facility PM Emissions
Drilling	1	8	5
Blasting	2	14	7
Conveying and transfer	3,4,5	6	5
Primary & secondary crushing	6,7	14	8
Vehicle & engine emissions	8	18	10
Total		60	35

††Based on PSD Permit Applications
* See Appendix C for category definition

A. Sources--

The major sources for below ground mining emissions are blasting, crushing and vehicles/engines which account for 14 - 18 kg/1000 m^3 of oil each or 7 - 10 % of the total facility particulate emissions. Less significant sources are drilling and transfer operations which account for 6-8 kg/1000 m^3 of oil and approximately 5 percent each of the total facility particulate emissions.

B. Controls--

Emission factors, control techniques and control percentages are shown in Table 4-1. Typically, water sprays, wet suppression and baffled settling are used to control the below ground emissions from blasting and drilling. There are considerable differences in the assumed control percentages ranging from 50 to 94 percent for the baffled settling. The primary crushing and transfer operations mostly used baghouses for control with 99+ percent control efficiency. Other options used are a wetted fan scrubber (99 percent) and wet suppression (80 percent). The emission factors vary from 0.02 to 1.5 lb/ton. The particulate emission from below-ground vehicles and engines are based on the emission factor of 22 lb/1000 gal of diesel fuel.

4.1.2 Above-Ground Mining Emissions (Primary Category 2)

Above-ground particulate mining emissions varied from 55 kg/1000 m^3 of oil to 217 kg/1000 m^3 of oil with an average value of 122 kg/1000 m^3 of oil. The above-ground mining emission rates and the unit processes are shown in Table 4-4.

A. Sources--

The major source of above-ground particulate emissions is the handling and disposal of spent shale. These unit processes (categories 30 - 36) account for 22 percent of the total facility particulate emissions and amount to 54 kg/1000 m^3 of oil. The handling of the raw shale (categories 13 - 25) account for 18 percent of the total facility particulate emissions and amount to 50 kg/m^3 of oil. The removal and hauling of surface soils account for 7 percent of the toal facility particulate emissions and amount to 14 kg/1000 m^3 of oil.

TABLE 4-4. CONTROLLED ABOVE-GROUND MINING AND MATERIAL HANDLING PARTICULATE EMISSIONS††

Unit Process	Secondary Category*	Average g/1000 m^3 of Oil	Percent of Total Facility PM Emissions
Surface Soils Removal, Hauling & wind	9,10,11,12	14	7
Secondary & tertiary crushing	13, 14	20	9
Conveying & screening	15, 16	20	9
Storage	17 - 25	10	4
Fines storage	26,27,28,29	7	3
Spent shale matls. handling Conveying, transfer, stacking, haul, storage,	30 - 36	54	22
Total		125	44

*See Appendix C for category definition
††Based on PSD Permit Applications

4.1.3 Retort Operations and Retort Gas Combustion

Particulate emissions from the retort operations and gas combustion varied from 8 kg/1000 m^3 of oil to 660 kg/1000 m^3 of oil with an average value of 140 kg/1000 m^3 of oil. The major sources and emission rates are presented in Table 4-5.

TABLE 4-5. CONTROLLED PARTICULATE EMISSIONS FROM RETORT OPERATION††

Unit Process	Secondary Category*	Average kg/1000 m^3 of Oil	Percent of Total Facility PM Emissions
Steam generation	37	20	7
Retort heating	40	25 - 120	10 - 25
Total		45 - 140	17 - 32

*See Appendix C for category definitions
††Based on PSD Permit Applications

A. Sources—

The major sources for particulate emissions from the retort are from the steam generation (Category 37) from burning the retort off-gas and fuel combustion to provide heat for the retort (Category 40). The steam generation from burning the off gas produces 20 kg/1000 m^3 of oil of particulates and averages eight percent of the total facility particulate emissions.

The retort heating produces 20 - 120 kg/1000 m^3 of oil and averages 25 percent of the total facility particulate emissions. The retort heating values are high due to the inclusion of the Clear Creek Fluidized bed spent shale combustor which produces an extremely high particulate emission of 640 kg/1000 m^3 of oil. The average particulate emission for the facilities not using the fluidized bed spent shale combustor is 25 kg/1000 m^3 oil, or about ten percent of the total facility particulate emissions.

B. Controls—

The controls used depended on the fuel used. When burning gaseous fuel an emission factor of 0.015 lb/MMBtu and no controls were used. When the combustion processes results in contact between the flue gas and a solid (i.e. burning spent shale and/or fines or a Tosco retort requiring a ball heater) a baghouse was used with an emission factor of 0.03 lb/MMBtu.

4.1.4 Upgrading and Miscellaneous

Particulate emissions from upgrading and miscellaneous above-ground sources varied from 20 kg/1000 m^3 of oil to 62 kg/1000 m^3 of oil with an average value of 40 kg/1000 m^3 of oil. The major sources are shown in Table 4-6.

A. Sources—

The upgrading process produces 20 kg/1000 m^3 of particulate and averages seven percent of the total facility particulate emissions. The major sources for above ground miscellaneous emissions are vehicles/engines and fugitive dust which both produce 10 kg/1000 m^3 of oil and an average of five percent of the total facility emissions.

TABLE 4-6. CONTROLLED PARTICULATE EMISSIONS FROM UPGRADE AND MISCELLANEOUS ABOVE-GROUND OPERATIONS††

Unit Process	Secondary Category*	Average kg/1000 m^3 of Oil	Percent of Total Facility PM Emissions
Upgrading heaters	43	20	7
Vehicles/engines	53	10	5
Fugitive dust	54	10	5
Total		40	17

††Based on PSD Permit Applications
* See Appendix A for category definitions

B. Controls--

Controls used are primarily wet suppression to control fugitive dust.

4.2 TOTAL FACILITY GASEOUS EMISSIONS

Total gaseous emission summaries for the criteria pollutants carbon monoxide, hydrocarbons, nitrogen oxides and sulfur oxides are presented in Table 4-7. The following section discusses the variations for the facility totals. A discussion on the individual process emission rates follows in Section 4.3.

TABLE 4-7. TOTAL FACILITY CONTROLLED GASEOUS EMISSIONS SUMMARY††
kg/1000 m^3 of Oil

	CO	HC	NO_x	SO_x
Cathedral Bluffs	250	90	3200	710
Clear Creek (fluid bed)	11300	230	5800	550
Cottonwood (fluid bed)	250	110	4300	800
Paraho	260	50	1700	650
Syntana	230	150	1200	290*
Union	500	250	1200	450
White River	210	210	1100	160†
Avg.	290	160	1700	510
Avg./flb	5800		5000	

††Based on PSD Permit Applications
* and † Corrected values - refer to Section 4.4

A. Carbon Monoxide--

The carbon monoxide emissions for facilities not using a fluidized bed combustor are fairly consistent ranging from 200 to 500 kg/1000 m^3 of oil with an average value of 290 kg/1000 m^3 of oil. The two facilities using the fluidized bed differ widely on their estimates for carbon monoxide emissions. This variation is due to the different applications of the fluidized bed. The Clear Creek facility uses the fluidized bed to burn spent shale. It is critical to control closely the temperature in the combustor to limit the reactions of dolomite and calcite and the formation of both thermal and fuel NO_x emissions. The low combustion temperature and the low carbon concentration in the char result in the high CO emissions.

The Cottonwood facility uses the fluidized bed to burn fines, spent char and retort gas. The presence of the retort gas and the raw shale with its relatively high carbon content result in improved control of the combustion conditions resulting in lower CO emissions.

B. Hydrocarbons--

The hydrocarbon emissions are also fairly consistent ranging from 50 to 250 kg/1000 m^3 of oil with an average value of 160 kg/1000 m^3 of oil. The two low values, 50 and 90 kg/1000 m^3 of oil for Cathedral Bluffs and Paraho-Ute are due to their low estimates for storage and above ground vehicles and engines.

C. Nitrogen Oxides (NO_x)--

Nitrogen oxide emissions estimates for the five facilities without a fluidized bed shale combustor range from 1100 to 3200 kg/1000 m^3 of oil with an average value of 1700 kg/1000 m^3 of oil. The highest value, 3200 kg/1000 m^3 of oil for Cathedral Bluffs, is due to the higher temperatures and higher gas flow rate typical of an in-situ retort. Nitrogen oxide emissions for the two facilities using a fluidized bed shale combustor were estimated in the range from 4300 to 5800 kg/1000 m^3 of oil with an average value of 5000 kg/1000 m^3 of oil. These high emissions are due to the combustion of the nitrogen in the shale.

It should be noted that these estimates from the PSD permit application do not include provision for fuel nitrogen present after a water wash system has been used to remove ammonia. The fuel NO_x emission rate is determined by the efficiency of ammonia removal, presence of other nitrogen compounds or (if fluidized bed) nitrogen content of shale fines or carbonaceous spent shale. The effect of residual ammonia and organic nitrogen on the NO_x emission rate was estimated based on the analysis described in Section 5.2.2-C and the increased NO_x emissions are shown in Table 4-8.

TABLE 4-8. INCREASE IN NO_x EMISSIONS FROM FUEL NITROGEN kg/1000 m^3 of Oil

Project	Type of Retort Heating	Thermal NO_x	Fuel NO_x	Percent Increase
Clear Creek	Solids recycle	5800	500	9
Cathedral Bluff -	In-situ w/union above ground retort	3200	4500	140
Parahute	Direct combustion	1700	2700	160

The effect of the fuel NO_x varies with the type of process and the retort gas production. In the case of the solids recycle process with low retort gas production rate of high Btu gas, the fuel related NO_x from incomplete ammonia removal and organic nitrogen compounds only increase the NO_x emissions by nine percent. The direct combustion process with higher gas production of low Btu gas results in an increase of 2700 kg/1000 m^3 of oil in NO_x emissions for a 160 percent increase over the expected value. In the case of the in-situ process, the higher retort temperatures and gas production result in an increase of 4500 kg/1000 m^3 of oil. This value is based on the relative oil production rates for the in-situ and above ground (Union) process.

The reader also should note that these estimates are based on a number of assumptions including retort gas flow rate, nitrogen partitioning to the retort gas, organic nitrogen content of the retort gas and expected ammonia removal rates. Consequently, these values are rough estimates and are presented primarily to illustrate the potential impact of the fuel related nitrogen content of the retort gas on the facility nitrogen oxide emissions.

D. Sulfur Oxides (SO_x)--

Sulfur oxide emissions range from 160 kg/1000 m^3 of oil to 800 kg/1000 m^3 of oil with an average value of 510 kg/1000 m^3 of oil. The two facilities with the lowest SO_x emissions are White River (160 kg/1000 m^3 of oil) and Syntana (290 kg/1000 m^3 of oil), both of which are using the circular grate Superior above ground retort.

1. Organic sulfur -- In Section 2.2.1.A, the concentration of organic sulfur species in the retort gas was described. The presence of these organic sulfur species at higher concentrations than assumed results in a considerable increase in sulfur oxide emissions. In all cases, the PSD applications assume no significant organic sulfur content in the retort gas. The effect of various organic sulfur levels on the facility SO_x emissions are shown in Table 4-9.

TABLE 4-9. EFFECT OF ORGANIC SULFUR IN RETORT GAS ON SULFUR OXIDE EMISSIONS

Organic Sulfur Percent	Potential Increase in SO_x Emissions From Organic Sulfur, kg/1000 m^3 of Oil
15	7600
10	5200
5	2800
1	940

These levels of increased SO_x emissions are substantial and greater than the total estimated facility emissions even at the one percent level for organic sulfur. Processes that utilize a flue gas desulfurization after combustion of the retort gas avoid this problem by burning all of the sulfur gases to SO_x prior to clean-up.

As discussed above for the organic nitrogen effect on NO_x emissions, this analysis for the effect of organic sulfur levels on SO_x emissions is based on a number of assumptions including sulfur partitioning to the gas, organic sulfur content and the limited effectiveness of an H_2S cleanup system

on organic sulfur compounds. The estimates are made to indicate the nature of the problem rather than to indicate that these would be expected values.

4.3 EMISSION SOURCE BREAKDOWN

4.3.1 Carbon Monoxide

The carbon monoxide emissions for the four major process categories are shown in Table 4-10.

TABLE 4-10. CONTROLLED CARBON MONOXIDE EMISSIONS[††]
kg/1000 m^3 of Oil

	Mining & Material Handling			Processing		
	Below	Above	Total	Retort	Upgrade	Total
Primary Category No.	1	2		3	4	
Project						
Cathedral Bluffs	150			60	40	250
Clear Creek (fluid bed)	150	8		11100	0	11300
Cottonwood (fluid bed)	130			100	30	250
Paraho	180			60	20	260
Syntana	155			30	50	230
Union B	120	8		210	160	500
White River	120			60	40	220
Average	140	8		90	60	300

[††]Based on PSD Permit Applications

The below-ground mining emissions are in relatively close agreement ranging from 120 to 180 kg/1000 m^3 of oil with an average value of 140 kg/1000 m^3 of oil. The small above-ground mining emissions are due to vehicles involved in soil and spent shale removal and hauling.

The carbon monoxide emissions from the retorting processes not using the fluidized bed spent shale combustor vary from a low of 30 kg/1000 m^3 of oil to a maximum of 210 kg/1000 m^3 of oil with an average value of 90 kg/1000 m^3 of oil. As discussed previously, the wide variation in the two

facilities using fluidized bed combustion is due to the different applications.

The upgrading process carbon monoxide emissions vary from 20 to 160 kg/1000 m^3 of oil. The high value for the Union facility, 160 kg/1000 m^3 of oil, is due to a higher value for above-ground vehicles (65 kg/1000 m^3 of oil) than the other projects. The following discussion presents data on the specific sources and controls for carbon monoxide emissions.

A. Below-Ground Mining (Category 1)--

Below-ground emissions of carbon monoxide result from blasting and below ground vehicles. The data presented in the Specific Process Sort, Table C-4, Appendix C were used to determine the average emission rates for each specific process. The specific process emission rates of carbon monoxide are shown in Table 4-11.

TABLE 4-11. CONTROLLED CARBON MONOXIDE EMISSIONS FOR BELOW GROUND MINING††

Process	Secondary Category No.	Average kg/1000 m^3 of oil	% of Total Carbon Monoxide Facility Emissions
Blasting	1	70	25
Vehicles	8	80	30
Total		150	55

††Based on PSD permit applications

1. Sources -- The primary sources are blasting (70 kg/1000 m^3 of oil) and below ground vehicles (80 kg/1000 m^3 of oil). These two sources account for 55 percent of the total facility carbon monoxide emissions (excluding the fluidized bed facility with its large CO emission rate).

2. Controls -- Vehicles account for 25 percent of the total reported facility emissions for carbon monoxide. Consequently, a substantial improvement in vehicle controls (e.g. catalytic mufflers) could reduce these

estimated emission rates. Blasting emissions have all been estimated at the minimum number of blasts; consequently, no additional controls are available.

B. Above Ground Mining (Category 2)--

Carbon monoxide emissions from above ground mining operations are less than 10 kg/1000 m^3 of oil and are not significant, amounting to less than 3 percent.

C. Retort & Gas Utilization (Category 3)--

Carbon monoxide emissions from retort and gas utilization are shown in Table 4-12.

TABLE 4-12. CONTROLLED CARBON MONOXIDE EMISSIONS FROM RETORT AND GAS COMBUSTION††

	Average kg/1000 m^3 of oil	Percent of Total
Standard combustion or fluidized bed with retort gas	90	35
Fluidized bed for spent shale combustion-recycle solids	11000	98

††Based on PSD Permit Applications

The carbon monoxide emissions from the retort and gas combustion are due to incomplete oxidation and are directly dependent on the type of combustion and amount of retort gas produced. For processes that do not use a fluidized-bed combustor, the CO emissions can be expected to be approximately 90 kg/1000 m^3 of oil which represent 35 percent of the total facility carbon monoxide emissions.

The projects using fluidized-bed combustion vary considerably in their estimated carbon monoxide emissions. The Cottonwood facility uses a fluidized-bed to burn fines and spent shale along with the retort off-gas. The estimated CO emissions for the fluidized-bed combustor are 96 kg/1000 m^3 of oil. The Clear Creek facility uses a fluidized-bed to burn spent shale and provide heat for the retort as recycled solids. The estimated CO emission rate is 11000 kg/1000 m^3 of oil. The additional high value fuel (retort gas

and fines) burned in the Cottonwood fluidized-bed results in more complete combustion and hence, the lower CO emission rates.

D. Upgrade (Category 4)--

The carbon monoxide emissions from upgrading operations are shown in Table 4-13.

TABLE 4-13. CARBON MONOXIDE EMISSIONS FROM UPGRADING OPERATIONS††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total CO Emissions
Heaters, boilers, etc.	43	40	14
Vehicles	53	25	7
Total		65	21

††Based on PSD Permit Applications

1. Sources -- The major sources for carbon monoxide emissions from upgrading processes are the various combustors for heaters, boilers, hydrogen reformers, etc. (40 kg/1000 m^3 of oil) and in plant vehicles (25 kg/1000 m^3 of oil).

2. Controls -- Promotion of complete combustion in either the furnace heaters or vehicles will reduce carbon monoxide emissions.

4.3.2 Hydrocarbons

Hydrocarbon emission rates are shown in Table 4-14.

TABLE 4-14. CONTROLLED HYDROCARBON EMISSIONS[††]
kg/1000 m^3 of Oil

	Mining & Material Handling		Process		
	Below	Above	Retort	Upgrade	Total
Primary Category No.	1	2	3	4	
Project					
Cathedral Bluffs	40		50	8	100
Clear Creek	20		0	210	230
Cottonwood	20		3	90	110
Paraho	30		10	10	50
Syntana	20		60	70	150
Union B	20		10	220	250
White River	30		70	120	220
Avg.	25		35	100	160

[††]Based on PSD Permit Applications

The below-ground mining hydrocarbon emissions vary from 20 to 40 kg/1000 m^3 of oil with an average value of 25 kg/1000 m^3 of oil. The above-ground mining hydrocarbon emissions were not presented. The above-ground vehicles do emit some hydrocarbons, however.

Hydrocarbon emissions from retort and gas utilization varied from 0 to 70 kg/1000 m^3 of oil with an average value of 35 kg/1000 m^3 of oil. The facilities using a Tosco II retort, Syntana and White River, both have relative high hydrocarbon emission rates, 60 and 70 kg/1000 m^3 of oil, due to the direct contact preheating of the raw shale. The Cathedral Bluffs facility has a high value due to the combustion of large quantities of low heating value retort gas. The facilities using retorts other than the Tosco II or in-situ have hydrocarbon emissions from retort and gas utilization of less than 10 kg/1000 m^3 of oil.

Hydrocarbon emissions from the upgrading process varied from 8 to 220 kg/1000 m^3 with an average value of 160 kg/1000 m^3.

A. Below-Ground Mining- (Category #1)--

Below ground mining hydrocarbon emissions sources are shown in Table 4-15.

TABLE 4-15. CONTROLLED HYDROCARBON EMISSIONS FROM BELOW GROUND MINING††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of oil)	Percent of Total HC Emissions
Vehicles	8	20	21

††Based on PSD Permit Applications

1. Sources -- The major source for hydrocarbon emissions from below-ground mining are vehicles and engines. The emission rate is 20 kg/1000 m^3 of oil and amounts to approximately 21 percent of the total facility hydrocarbon emissions.

2. Controls -- Additional controls on the vehicles (e.g. catalytic mufflers) could reduce these hydrocarbon emision rates.

B. Above Ground Mining -(Category #2)--

No significant hydrocarbon emissions result from the above-ground mining operations. The EPA's AP-42 emission factor for hydrocarbons from IC engines is approximately one hundredth of the CO emissions from that source type.

C. Retort and Gas Combustion (Category #3)--

The hydrocarbon emission rates for retort and gas combustion are shown in Table 4-16.

TABLE 4-16. CONTROLLED HYDROCARBON EMISSION FROM RETORT AND GAS COMBUSTION††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of oil)	Percent of Total HC Emissions
Gas combustion	37	16	10
Retort heating			
Not Tosco II	40	4	3
Tosco II	40	50	30

††Based on PSD Permit Applications

1. Sources -- The major sources for hydrocarbon emission from retort and gas combustion are the retort gas used for power generation and to provide

heat to the retort. The gas combustion for power generation produces 16 kg/1000 m^3 of oil amounting to 10 percent of the total facility hydrocarbon emissions. The hydrocarbon emissions from combustion to provide heat to the retort is dependent on the type of retort. Retort processes other than the Tosco II process produce relatively low amounts (4 kg/1000 m^3 of oil) of hydrocarbon emissions. The Tosco II retort has two direct contact heating processes — an "elutriator" to heat the ceramic balls and raw shale preheater. Fuel gas is burned and the hot combustion gases are mixed directly with the recycled ceramic balls to provide the heat for retorting and raw shale for preheat. An incinerator is provided for control of the hydrocarbon emissions; however, the HC emissions for this process are still quite high at 60 - 70 kg/1000 m^3 of oil.

2. Controls -- Controls for the combustion processes have been included in the plant design, primarily by promoting complete combustion. The controls for the Tosco II process include an incinerator on the raw shale preheat system. Additional controls would be ineffective.

D. Upgrading (Category #4)--

The hydrocarbon emission rates for upgrading operations are shown in Table 4-17.

TABLE 4-17. CONTROLLED HYDROCARBON EMISSIONS FROM UPGRADING OPERATIONS††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total HC emissions
Heaters	43	6	3
Fugitive	44	100	50
Storage	46	40	20
Vehicles	53	4	2
Total		150	75

††Based on PSD Permit Applications

1. Sources -- The major sources for hydrocarbon emission from upgrading are heaters, fugitive emissions, storage and vehicles. These emissions total 75 percent of the total facility hydrocarbon emission with the major

contributors being the fugitive emissions (50 percent) and storage (20 percent).

2. Controls -- No additional controls are available to reduce these emissions.

4.3.3 Nitrogen Oxides

The nitrogen oxide emission rates are shown in Table 4-18.

TABLE 4-18. CONTROLLED NITROGEN OXIDE EMISSIONS[††] (kg/1000 m^3 of Oil)

	Mining & Material Handling		Process		
	Below	Above	Retort	Upgrade	Total
Primary Category #	1	2	3	4	
Cathedral Bluffs	500		1400	1300	3200
Clear Creek (fluid bed)	140	30	5600	10	5800
Cottonwood (fluid bed)	270		4000	10	4300
Paraho	470		930	300	1700
Syntana	300		400	500	1200
Union B	230	20	570	350	1200
White River	340		460	260	1100
Average (no fuid bed)	320	25	750	390	1700
(with fluid bed)			4800		5500

[††]Based on PSD Permit Applications

Below ground mining emissions vary from 140 to 500 kg/1000 m^3 of oil with an average value of 320 kg/1000 m^3 of oil. Above-ground NO_x emissions (primarily from vehicles) are minimal and were reported by only two facilities.

Retort and gas combustion NO_x emissions vary from 400 to 5600 kg/1000 m^3 of oil. The large variation is due to the difference in the processes. The facilities that use the fluidized bed combustion of the spent shale (Clear Creek and Cottonwood) have high NO_x emission rates (4000 - 5600 kg/1000 m^3 of oil) due to the combustion of the fuel related nitrogen in the shale. The

other facilities have NO_x emission rates from <400 to 1400 kg/1000 m^3 of oil which are determined by the combustion of the retort off gas and retort heaters. The high value for the Cathedral Bluffs facility is due to the higher gas production rate of the in-situ process.

The upgrading NO_x emissions are fairly consistent with the exception of the Clear Creek and Cottonwood facilities. NO_x emission rates from these latter two facilities (10 kg/1000 m^3 of oil) are unexplainedly one to two orders of magnitude lower than the other five facilities.

A. Below-Ground Mining (Category #1)--

The nitrogen oxide emission rates for below-ground mining operations are shown in Table 4-19.

TABLE 4-19. CONTROLLED NITROGEN OXIDE EMISSIONS FROM BELOW-GROUND MINING AND MATERIAL HANDLING††

Process	Secondary Category No.*	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total NO_x Emissions
Blasting	2	20	1
Vehicles	8	300	20
Total		320	21

††Based on PSD Permit Applications
* See Appendix C for category definition

1. Sources -- The major source for below ground mining emissions are vehicle/engines producing 300 kg/1000 m^3 of oil amounting to 20 percent of the total facility NO_x emissions. Blasting is a relatively minor source for NO_x emissions.

2. Controls -- Additional controls on the vehicles could reduce the overall plant NO_x emissions.

B. Above-Ground Mining (Category #2)--

Only two facilities reported NO_x emissions from above-ground mining operations amounting to 20 - 30 kg/1000 m^3 of oil or less than one percent of the total facility NO_x emissions.

C. Retort and Gas Combustion (Category #3)--

The NO_x emissions from oil shale retorting and other combustion devices in an oil shale plant are presented in Table 4-20.

TABLE 4-20. CONTROLLED NITROGEN OXIDE EMISSIONS FROM RETORT AND OTHER GAS COMBUSTION††
(Not Including Organic Nitrogen & Residual NH_3)

Process	Secondary Category No.*	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total NO_x emissions
Gas combustion	37	550	30
Retort heaters			
(no fluid bed)	40	220	14
(with fluid bed)	40	4800	95
Total			
(no fluid bed)		770	44
(with fluid bed)		5400	95+

††Based on PSD Permit Applications
* See Appendix C for category definitions

1. Sources -- The major sources for NO_x emissions from retort and other gas combustion are due to the need to provide heat for the retort or power generation from the retort off gas. The NO_x emission rate is determined by the type of combustion (thermal NO_x) and the nitrogen content of the fuel (fuel NO_x). The PSD permit applications consistently assumed that fuel NO_x emissions are negligible based on effective fuel gas cleanup. However, the presence of organic nitrogen compounds reduce the removal efficiency of the nitrogen clean-up process (water wash for ammonia removal). Therefore, the actual expected NO_x emissions will be greater than calculated in the PSD permits (See Section 4.2).

Combustion of the retort gas produces NO_x emissions of 550 kg/1000 m^3 of oil or 30 percent of the total facility NO_x emissions. The NO_x emission rate for the Cathedral Bluffs Plant is high at approximately 1000 kg/1000 m^3 of oil due to the higher temperatures in an in-situ retort producing higher quantities of retort gas.

The NO_x emission from providing heat for the retort depends on the type of combustion. The facilities not using a fluidized bed spent shale combustor produce NO_x emissions for retort heat of approximately 220 kg/1000 m^3 of oil and account for 14 percent of the total facility NO_x emissions. The facilities using the fluidized bed spent shale combustor produce considerably more NO_x (4800 kg/1000 m^3 of oil) due to the nitrogen in the spent shale.

2. Controls -- Additional controls to reduce NO_x emissions from these combustion processes are staged combustion or selective catalytic reduction (SCR). The SCR can only be applied on relatively clean gases and would not be suitable for the fluidized bed combustor flue gas.

Staged combustion could be applied on either the clean (gas and oil) fuels or the fluidized bed combustors. The reader is referred to Section 3.2.1 for a description of the application of staged combustion on the fluidized bed (or cascading bed) combustor.

D. Upgrading (Category #4)--

The nitrogen oxide emissions from upgrading operations are presented in Table 4-21.

TABLE 4-21. CONTROLLED NITROGEN OXIDE EMISSIONS FROM UPGRADING OPERATIONS††
(Not Including Organic Nitrogen and Residual NH_3)

Process	Secondary Category No.*	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total NO_x Emissions
Heaters	43	450	25
Vehicles	53	10	>1
Total		460	26

††Based on PSD permit applications
* See Appendix C for category definitions

1. Sources -- The major source for NO_x emissions from upgrading operations are the heaters and furnaces. These units produce 450 kg/1000 m^3 of oil and account for 25 percent of the total facility NO_x emissions. Above-ground vehicles add a minimal amount of NO_x emissions amounting to less than one percent of the total.

2. Controls -- Staged combustion and ammonia injection are two controls that could reduce these NO_x emissions.

4.3.4 Sulfur Oxides

The sulfur oxide emission rates are shown in Table 4-22.

TABLE 4-22. CONTROLLED SULFUR OXIDE EMISSIONS[††] (kg/1000 m^3 of Oil)

Primary Category No.	Mining & Material Handling		Process		
	Below	Above	Retort	Upgrade	Total
	1	2	3	4	
Cathedral Bluffs	30		330	350	710
Clear Creek	10	10	540		550
Cottonwood	20		780		800
Paraho	30		600	20	650
Syntana	20		270	10	300*
Union B	20	10	390	30	450
White River	20		130	10	160†
Average	20	10	430	15§	475§

††Based on PSD permit applications

* - Corrected value - reported values for SO_x emissions for the retort and other gas combustion were low by a factor of 5 (see section 4.4 below).

† - Corrected value - reported values for SO_x emissions for the retort and gas combustion were low by a factor of 3 (see section 4.4 below).

§ - Value for Cathedral Bluffs upgrading emissions not included.

Below ground mining emissions of sulfur oxides are relatively small, ranging from 10 to 30 kg/1000 m^3 of oil and averaging 15 kg/1000 m^3 of oil. Above-ground mining emissions of sulfur oxides are minimal.

The sulfur oxide emissions from the retort and other gas combustion has wide variation ranging from 130 to 780 kg/1000 m^3 with an average value of 430 kg/1000 m^3. The variation is due to the differences in the retort

processes and the clean-up methods. These will be discussed further in Section 4.3.4.C, below.

The sulfur oxide emissions from the upgrading operations are relatively small (10 - 30 kg/1000 m^3 of oil) for all facilities other than the Cathedral Bluffs facility. The value of 350 kg/1000 m^3 is much higher than the other facilities due to the fact that the facility is relying on the flue gas desulfurization (FGD) unit to reduce the SO_x emissions and the flue gases from the upgrading heaters are not included in that FGD system. Consequently, the sulfur oxide emissions are considerably higher reflecting the high sulfur level in the retort off gas after the H_2S removal.

A. Below Ground Mining (Category #1)--

The emission rates for sulfur oxides from below ground mining are shown in Table 4-23.

TABLE 4-23. CONTROLLED SULFUR OXIDE EMISSIONS FROM BELOW-GROUND MINING AND MATERIAL HANDLING††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total SO_x Emissions
Vehicles/engines	8	20	5

††Based on PSD permit applications

1. Sources -- The only source for below-ground mining emissions of sulfur oxides is vehicles producing 20 kg/1000 m^3 of oil of emissions which amount to 5 percent of the total facility sulfur oxide emissions.

2. Controls -- The only control technique would be to reduce the sulfur content of the fuel used in the engines.

B. Above Ground Mining--

Sulfur oxide emissions from above ground vehicles is minimal.

C. Retort and Gas Utilization--

Emissions of sulfur oxides from the retort and gas utilization are a function of:

- type of retort
- gas production rate
- efficiency of the fuel gas cleanup
- post-combustion flue gas desulfurization

The effect of the various process designs and clean-up systems is discussed extensively in Section 5, Process Analysis.

The emissions of sulfur oxides for the individual operations are presented in Table 4-24.

TABLE 4-24. CONTROLLED SULFUR OXIDE EMISSIONS FROM RETORT AND OTHER GAS COMBUSTION††

Process	Secondary Category No.	Emission Rate (kg/1000 m^3 of Oil)	Percent of Total SO_x Emissions
Retort gas combustion	37	230	42
Retort heater	40	200	35
Total		430	77

††Based on PSD permit applications

1. Sources -- The primary sources of SO_x emissions from retort and gas utilization are the combustion of the retort gas (230 kg/1000 m^3 of oil; 42 percent of the total facility SO_x emissions) and the retort heater (200 kg/1000 m^3 of oil; 35 percent of the total facility emissions.)

2. Controls -- Additional controls can be used to reduce these sulfur oxide emissions. The reader is referred to Section 3 for a complete discussion on the various sulfur oxide control techniques that can reduce these emissions.

3. Organic sulfur -- As discussed above, the PSD Applications do not consider the presence of organic sulfur gases. The effect of including the expected emissions from combustion of these organic sulfur compounds varies with the type of retort. For example, sulfur oxide emissions from the solids recycle retort used by Clear Creek would be expected to increase by nine percent; the direct combustion retort (Paraho) emissions would increase by as much as 150 percent. The Cathedral Bluffs facility has included a post

combustion FGD and therefore, their estimated SO_x emissions would not increase.

4.4 PSD PERMIT APPLICATION DATA INCONSISTENCIES

During this evaluation of the PSD permit application data, a number of inconsistencies were found in the reported facility emissions. In some cases reasonable explanations were found; in other cases no explanations were found. The more significant of these items are discussed below for the purpose of informing future permit applicants in estimating their own emissions.

4.4.1 Carbon Monoxide

The combustion related carbon monoxide emissions for the White River and Syntana facility as reported in their PSD permit applications are shown in Table 4-25.

TABLE 4-25. CONTROLLED CO EMISSIONS FROM WHITE RIVER AND SYNTANA[††]

	Energy Consump. MW	Reported Emissions kg/1000 m^3 Oil	Corrected Emissions kg/1000 m^3 Oil
White River			
Retort Tosco ball heater	140	8.8	8.8
Retort recycle heater	280	18.1	18.1
Upgrade reformer	630	40.5	40.0
Utility steam	710	314.0	22.0
Total	1760	382.0	89.0
Syntana			
Upgrade FGD	20	0.7	1.8
Retort Tosco ball heater	83	5.1	11.0
Retort Superior	166	4.0	8.6
Retort FGD Steam	285	6.8	14.5
Upgrade Hydro reformer	478	7.4	15.9
Total (as reported)	1030	24.0	51.8

[††]Based on PSD permit application

Since these two facilities are similar in concept, the wide variation in carbon monoxide emissions (382 kg/1000 m^3 of oil for White River compared to 24 kg/1000 m^3 of oil for Syntana) for the steam/generation process appeared suspect. The steam boiler emissions for White River seemed to be too high. The high value was found to be due to a different definition of the emission factor. Generally, the project applicants used natural gas emission factors when estimating emissions from retort gas combustion. In all cases except for White River, an assumption was made that the emission factor was actually based on the weight of pollutant per unit of energy consumed rather than weight of pollutant per unit of volume. The definition is important when computing emissions for burning low Btu gas (with 100 Btu/cubic ft) rather than emissions from burning high Btu gas (with 1000 Btu/cubic ft). In this instance it leads to a 10 fold difference in the emissions.

In this case of carbon monoxide emissions from the steam boiler, the White River applicant used an emission factor of 17 lb/mscf; which is the same as 0.017 lb/MMBtu for natural gas with a heating value of 1000 Btu/scf. However, when burning low Btu gas the other facilities assumed the emission factor would be the same 0.017 lb/MMBtu which is 1.7 lb/mscf for 100 Btu/scf gas. The corrected values for the White River facility are shown in Table 4-25.

The problem faced by all applicants is that there is little emission data on the combustion of low Btu gas. The CO emissions depend a great deal on the specific burner design. The low energy content of the gas will usually result in a lower flame temperature and probably a higher CO emission per unit of energy throughput. Using the CO emission factors for 1000 Btu/mscf gas may result in predicted emissions lower than actual but it is probably a closer prediction than using the emission factor based on gas volume throughput.

In checking the difference between the White River and Syntana facility, it also was discovered that the emissions in the Syntana facility permit application are all 1/3 of the value calculated from the emission factor and the heat input rate. This was true for <u>all</u> emissions from the Syntana facility. The (reported and corrected) combustion related emissions of CO from the Syntana facility are also shown in Table 4-25.

Since the CO emissions from all of the gases burned except those in the Tosco elutriator and lift pipe were based on the same 0.026 kg/MW (0.017 lb/MMBtu) emission factor, the total emissions based on energy consumption should balance. However this was not the case.

The total CO emission rated (excluding the Tosco ball heater) is:

(1030 MW total - 83 MW Tosco) x 0.026 kg/MW = 21 kg/hr = .57 ton/day

The CO emission rate reported, as totals, were (.239 - .052) or .187 tons/day.

Further investigation of the PSD application table of emissions showed that, although the values given are indicated as being totals, they are actually values per operating unit. Where there are two units, the "totals" must be doubled to arrive at the actual emission rate. This type of presentation is extremely misleading.

The corrected carbon monoxide emissions are 52 kg/1000 m^3 of oil rather than 24 kg/1000 m^3 of oil as indicated in the PSD application. With these corrections, both facilities have CO emission rates 0.05 kg/1000 m^3 of oil/MW based on the total heat rate.

4.4.2 Sulfur Oxides (SOx)

The SO_x emissions also were inconsistent. The Syntana emission data appeared to be 1/3 of the proper value and the analysis was similar to that discussed above for carbon monoxide. The White River project SO_x value reported in their summary emissions table was 1/5 of the actual value. This discrepancy was determined as follows:

The White River PSD permit states that the retort gas contains 175 ppmw sulfur in low Btu gas. The SO_x emission rate is:

$$\frac{[\text{Heat rate, MW}]}{\text{Heating Value MW/m}^3} \times [\text{Density (kg/m}^3)] \times [\text{ppm}] = \text{kg sulfur/hr}$$

$$\frac{1760 \text{ MW}}{0.002140 \text{ MW-hr/m}^3} \quad (1.2 \text{ kg/m}^3) \quad (175 \text{ kg sulfur}/10^6 \text{ kg}) = 172 \text{ kg sulfur/hr}$$

172 kg sulfur/hr = 5.4 kg moles sulfur/hr x 64 kg SO_2/kg mole sulfur = 344 kg SO_x/hr = 8260 kg SO_x/day

Assuming that all of the fuel sulfur is converted to SO_x, this would be the proper estimate of SO_x emissions. The reported value is 1630 kg/day. Using the emission factor reported in the White River PSD Permit Application of 0.0037 kg/m^3 gives the following:

0.0037 kg/m^3 x 820,000 m^3/hr = 360 kg SOx/hr = 8600 kg SO_x/day

which agrees with the calculation based on their reported concentration of 175 ppm.

These discrepancies seem to reveal a need to have a standardized approach for the PSD permit application to allow comparative evaluation of the various processes.

4.5 ACTUAL PERMITTED EMISSION RATES

The previous discussion has focused on the emission rates estimated by the developers in their PSD permit applications. The actual permits issued indicated some significant variations from the applications.

Table 4-27 shows the application and actual permitted emission rates. Notable differences result from:

1. Requirement for selective catalytic reduction for NO_x control on the fluidized-bed combustor at the Clear Creek plant.

2. Use of catalytic converters for mine vehicles to decrease CO and HC emissions at Paraho.

3. Revision of emission estimates at Cathedral Bluffs and Cottonwood facilities.

In most cases the permit application data was consistent with the actual permitted values.

TABLE 4-27. COMPARISON OF PERMIT APPLICATION AND ACTUAL PERMITTED EMISSION RATES ($kg/1000\ m^3$ of Oil

	Particulate		SOx		NOx		CO		HC	
	Application	Permitted	A	P	A	P	A	P	A	P
Cathedral Bluffs	330	235 (1)	710	627 (2)	3200	1300 (3)	250	250	90	90
Clear Creek	855	N/A	550	550	5800	1500 (4)	11300	11300	230	230
Cottonwood	173	730 (5)	800	800	4300	3700	250	230	110	140
Paraho	350	350	650	620	1700	1400 (6)	260	150 (7)	50	20 (7)
Syntana	237	N/A	290	N/A	1200	N/A	230	N/A	150	N/A
Union B	181	N/A	450	N/A	1200	N/A	500	N/A	250	N/A
White River	213	213	160	160	1100	1100	210	210	210	210

N/A - Not available.

(1) - Particulate emissions for spent shale distribution, raw shale storage, spent shale conveyor were not included.

(2) - SO_x emissions from incinerator reduced from 127 to 83 $kg/1000\ m^3$ of oil.

(3) - Major reduction due to decrease by 73.5 percent in estimated NO_x emissions from retort gas combustion in steam boilers reformer, recycle gas heater and oil charge heat. No explanation available.

(4) - NO_x emissions from retort combustor decreased by application SCR to exhaust gas.

(5) - Large increase due to increases in particulate emission from blasting, conveying, transfer and fluidized-bed combustor.

(6) - NO_x reduced by 30% for reformer furnace.

(7) - Addition of catalytic converter on mine vehicles providing 85% control CO and 90% control of HC accounts for major reduction.

SECTION 5.0

PROCESS ANALYSIS

A wide variety of processing schemes can be used to recover oil from shale. The purpose of this section is to investigate some of the many possible variations to determine the net effect on criteria pollutant emissions.

The retort processes (Section 2), air pollution control equipment (Section 3) and the data provided by the Prevention of Significant Deterioration (PSD) permit applications (Section 4) have been described previously. In this section, the information presented in the previous sections will be used to evaluate the overall facility emissions for various retort processes and pollution control equipment.

BACKGROUND

A shale oil recovery plant is quite complex involving many varied operations. The possible processing combinations are presented in Figure 5-1.

Our primary concern in this analysis is emissions of nitrogen and sulfur oxides and particulate. Emissions of carbon monoxide and hydrocarbons are generally consistent for all processes with the following exceptions. The combustion of the spent shale can produce very high carbon monoxide emissions and this will be discussed below for that particular process. The only process that results in high hydrocarbon emissions is the Tosco II retort with its direct heating of raw shale and ceramic balls with flue gas.

In this section the methodology used to evaluate the various processes is presented, followed by an evaluation of the sulfur and nitrogen gases produced by the retort for the various processes. An analysis of the emissions for five typical processes being considered for full scale development is also included.

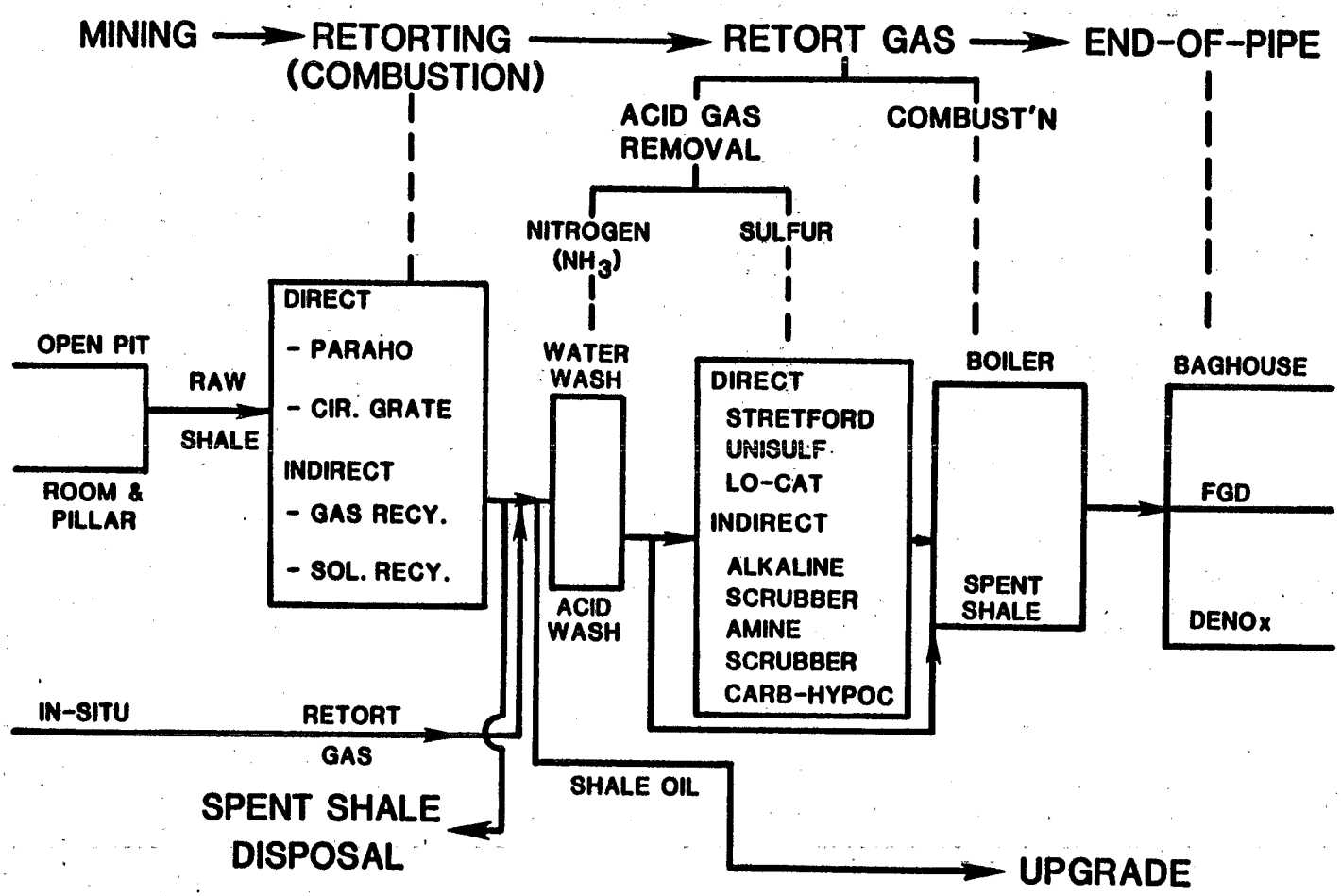


Figure 5-1. Possible processing combinations.

For each analysis, a base case scenario is presented to determine the criteria pollutant emissions. Two alternative processing schemes are then considered to reduce these emissions to their lowest levels.

5.1 METHODOLOGY

To evaluate the many process variations it was first necessary to determine what process combinations are feasible. Using the information presented in the literature and PSD applications, the potential unit operations for each process category was prepared. Figure 5-1 presents these alternatives. The basic design parameters that affect the pollutant emissions for each process are shown in Table 5-1.

For each of the unit operations, design parameters were applied as indicated by either the retort process conditions described in Section 2, the pollution control performance as described in Section 3 or the reported emissions from the PSD applications as described in Section 4. The following discussion presents the design parameters used in the analysis.

TABLE 5-1. DESIGN PARAMETERS
(Taback, H.J., et al., 1986)

Operation	Design Parameter
Mining	Type of mining
	open pit
	room and pillar
	in situ
Retort	Retort Gas Produced, m^3/m^3 of oil
	Heating Value of Retort Gas, kJ/m^3
	Partitioning of sulfur and nitrogen
Product Recovery	None- The product recovery process has no significant effect on emissions.
NH_3 removal	NH_3 exit concentration, ppm
	organic nitrogen content of retort gas
H_2S removal	H_2S exit concentration, ppm
	organic sulfur content of retort gas
	organic sulfur gas removal efficiency, %
Gas Utilization	Boiler - dilution ratio (dry gas/fuel)
	Spent Shale Combustion
	exit concentrations for NO_x, SO_x & CO
End of Pipe Controls	Particulate-Baghouse-exit loading, g/m^3 of oil
	Sulfur - FGD - exit SO_x, ppm
	Nitrogen
	Ammonia injection - exit NO_x
	Staged Combustion - exit NO_x & CO

5.1.1 Mining

The choice of mining technique determines the emission rates from the mining operation. The values used were developed from the PSD permit applications analysis, Section 4, for room and pillar mining and from the literature for open pit mining (PCTM-Lurgi-Open Pit Mining). The emission rates used are shown in Table 5-2.

TABLE 5-2. CONTROLLED EMISSION RATES FOR MINING††

Type of Mining	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Open pit	370	50	470	40	410
Room & pillar	150	20	350	20	180
Room & pillar with Catalytic converters on Engines	15	2	350	20	180

††Based on PSD permit applications

5.1.2 Retort

A. Gas Volume & Heating Value--

The retort gas produced for the various types of retorts was determined from the information presented in the PSD applications. The gas rates are shown in Table 5-3.

TABLE 5-3. RETORT GAS PRODUCTION RATES††

Type of Retort	Gas Produced m^3 of gas/m^3 of Oil	Heating Value Gram-Calories/Liter
In-situ	7000	9000
Direct heating	1800	9000
Indirect heating	180	90000

††Based on PSD permit applications

The in-situ process produces the highest retort gas flow rate due to the combined effect of the higher retort temperatures converting more of the kerogen to gas and the higher sweep gas flow required to provide adequate oxygen to burn the shale. The direct heating retort has similar conditions (i.e. high temperatures and requirement for adequate oxygen for combustion) but to a lesser degree than the in-situ retort and, consequently, has lower retort gas flow rates. The indirect heating process has the lowest retort gas flow rate due to the lower retort temperatures and low gas flow rate with no dilution required to provide oxygen for combustion within the retort.

The heating value of the retort gas is determined by the amount of dilution gas; in-situ and direct heated retorts produce low heating value gas at 9000 gram-calories/liter (100 Btu/scf) and the indirect heated retort produces high heating value gas at 90000 gram-calories/liter (1000 Btu/scf).

B. Sulfur Gases--

1. Species and Distribution -- The sulfur in the raw shale is partitioned to the spent shale, oil and retort gas. The significant variations in raw shale sulfur content, percentage of sulfur partitioned to the gas phase and the chemical structure of these sulfur gases result in the gas clean-up strategy being quite complex.

The sulfur emission problem can be solved either by removing the sulfur prior to combustion or by addition of a flue gas desulfurization process after combustion. As the combustion process dilutes the pollutant concentrations and increases the gas flow rate, the economically preferred technique is sulfur removal prior to combustion.

The form of the sulfur in the gas is extremely important when considering the sulfur removal processes. The sulfur recovery processes that have been considered for cleaning the retort gas prior to combustion are not effective in removing the organic sulfur compounds which can amount to as much as 10 - 16 percent of the total sulfur in the retort gas. Consequently, the effectiveness of these clean-up processes is dependent on the relative amounts of organic sulfur to H_2S. Even high efficiencies of H_2S removal (99%) are not sufficient to reduce the sulfur emissions below the 850 kg/1000 m^3 of oil (0.3 lb/bbl) regulatory level for Colorado.

The results of the calculation to determine the net sulfur emissions from burning the retort gas after the sulfur (H_2S) removal are shown in Table 5-4. For raw shale with 1.0 wt % sulfur where 30 percent of the sulfur in the raw shale is evolved with the retort gas, the net sulfur emissions range from 1000 kg/1000 m^3 oil (0.33 lb/bbl) for retort gas with 99 percent of the total sulfur as H_2S to 7600 kg/1000 m^3 oil (2.66 lb/bbl) for retort gas with 85 percent of the total sulfur as H_2S. Even with 0.5 wt% sulfur in the

raw shale the retort gas must contain greater than 95 percent of the sulfur as H_2S to approach the regulatory limit of 850 kg/1000 m^3 oil (0.3 lb/bbl).

TABLE 5-4. EFFECT OF ORGANIC SULFUR ON SULFUR EMISSIONS

Basis: sulfur wt % = 1; raw shale at 30 gal/ton; 28 % of sulfur in retort gas

H_2S, %	85	90	95	99
Organic sulfur gas of total sulfur %	15	10	5	1
Sulfur as H_2S, kg/1000 m^3 oil	200	220	230	230
Organic S, kg/1000 m_3 oil	3600	2400	1200	240
Total S, kg/1000 m^3 oil	3800	2600	1400	500
Total S (as SO_x), kg/1000 m^3 oil	7600	5200	2800	1000

Essentially, the above analysis indicates that the H_2S removal process must be capable of removing a significant amount of the organic sulfur gases to meet the regulatory limit of 850 kg/1000 m^3 of oil 0.3 lb sulfur (as SO_x) /bbl). To avoid the costly alternative of adding an "end of pipe" flue gas desulfurization, two alternatives can be considered. The first, activated carbon-hypochlorite H_2S removal process, is an improvement on the H_2S scrubbing process to remove organic sulfur species. The second is the solids recycle process which limits sulfur gas emissions by the chemistry of the retort and combustion process.

The activated carbon-hypochlorite H_2S removal process described in Section 3 can be designed to remove the organic sulfur compounds as well as the H_2S and, therefore, is suitable for removal of the sulfur gases prior to combustion eliminating the need for more expensive post combustion control. This process can remove 99+ percent of the H_2S and 90-98+ percent of the organic sulfur gases (Teller 1985b). This results in a net sulfur removal efficiency of 99 percent and sulfur emissions (SO_x) of 500 kg/1000 m^3 oil (0.17 lb/bbl) even when the organic sulfur gases are 15 percent of the total sulfur.

The recycle solids process controls the emissions of sulfur gases by the chemistry of the process. The sulfur emissions from the recycle solids

process represent only one percent of the total sulfur inlet primarily as H_2S in the retort gas. Therefore, the amount of organic sulfur is minimal and the H_2S removal processes alone are sufficient to reduce the sulfur emissions below the regulatory limit. The above discussion is summarized in Table 5-5.

TABLE 5-5. SUMMARY OF ESTIMATED SO_x EMISSION RATES

Basis: 1 wt% sulfur oil shale;
30% partitioned to gas phase
15% organic sulfur

	Control Technique	SO_x, kg/1000 m^3 of oil
First Generation—no solids recycle	H_2S removal	7700
First Generation—no solids recycle	H_2S removal and flue gas desulfurization	850
First generation - no solids recycle	Enhanced H_2S and organic sulfur gas removal	570
Second Generation - with solids recycle	Standard H_2S removal	100

2. Design Conditions — The H_2S removal process determines the residual H_2S and organic sulfur in the retort gas. The processes considered are:

1 - Direct or indirect conversion - considered the same for performance of H_2S removal:

H_2S exit concentration = 50 ppm
organic sulfur - assumed at 5 % of total sulfur in retort gas - no removal

2 - Activated carbon-hypochlorite process

H_2S exit concentration = 10 ppm
organic sulfur assumed at 5% of total sulfur in retort gas - 90% removal

3 - Spent shale combustion

SO_x exit concentration = 20 ppm

C. Nitrogen Gases--

1. Species and Distribution -- The removal of nitrogen gases is also difficult to predict due to the degree of variability in nitrogen content in shale, partitioning between gas, oil and spent shale and chemical form of the gaseous nitrogen species.

The presence of organic nitrogen species presents the same problem for limiting fuel related NO_x emissions as that described above for the SO_x emissions; namely that the removal processes generally considered are not effective in reducing the organic nitrogen content of the retort gas.

The primary nitrogen removal technique considered is removal of ammonia from the retort gas by a water wash absorption tower followed by an ammonia recovery stripper. The outlet ammonia concentration is determined by the effectiveness of the ammonia absorber. At atmospheric pressure the equilibrium exit partial pressure for ammonia at 50 C is 0.5 mm Hg (Perry and Green, 1984).

The nitrogen content of the retort gas is determined by:

The exit gas ammonia concentration (0.5 mm Hg)
The amount of retort gas produced by the retort (Table 5-3)
The amount of nitrogen partitioned to the retort gas (Table 2-8)
The percentage of nitrogen present as organic nitrogen compounds (Table 2-9)

Based on this information, an analysis of the effect of organic nitrogen on the overall NO_x emissions was made for three types of retorts. These results are shown in Table 5-6.

TABLE 5-6. ESTIMATED EMISSIONS OF NITROGEN OXIDE FROM COMBUSTION OF RETORT GAS

Type of Retort	Direct Combustion	Solids Recycle	In-situ
Retort gas, m^3/m^3 oil	1800	180	7000
Nitrogen in gas, %	30	10	50
N from NH_3, kg/1000 m^3 oil	890	90	3500
Organic N, kg/1000 m^3 oil	380	13	640
Total N, kg/1000 m^3 oil	1300	100	4100
Total NO_x, kg/1000 m^3 oil	2800	214	8800

(a) Solids recycle retort — has the lowest fuel-NO_x emissions (from fuel nitrogen) due to the combined effect of the low retort gas production (180 m^3/m^3 oil) and the low partitioning of nitrogen to the gas phase (10 percent). Consequently, since the exit ammonia concentration is based on an equilibrium value partial pressure, the lower retort gas flow rates results in less NH_3 in the exhaust. The resulting fuel-NO_x emissions from burning this retort gas are 214 kg/1000 m^3 oil.

(b) Direct heated retort -- This type of retort (e.g. Paraho) produces 1800 m^3 gas/m^3 of oil and partitions 30 percent of the nitrogen to the gas phase. Therefore, the NH_3 removal process has 10 times as much gas to process (compared to the solids heated retort) and a higher concentration of NH_3 (and associated organic nitrogen compounds). Consequently, using the same exit partial pressure of 0.5 mmHg NH_3 the resulting fuel-NO_x emissions from NH_3 and organic nitrogen combustion are 2700 kg/1000 m^3 oil.

(c) In-situ -- The in-situ retort produces the highest retort gas flow, 7000 m^3/m^3 oil with the highest partitioning of nitrogen to the gas phase (50 percent). Therefore, the ammonia removal is least efficient and the resulting fuel-NO_x emissions from NH_3 and organic nitrogen are 8800 kg/1000 m^3 oil.

2. Combustion of Spent Shale -- As mentioned previously, the combustion of the carbonaceous shale is desirable for recovery of the energy content.

The nitrogen reactions resulting from spent shale combustion were described in Section 2.1.1.

The combustion of the spent shale to obtain the extremely low sulfur emissions can result in large nitrogen emissions due to the fuel nitrogen. While the inorganic nitrogen burns slowly, it does burn to some degree and any residual organic nitrogen is also burned to NO_x. In the PSD permit applications NO_x emissions were based on the conversion of approximately 15 percent of the nitrogen in the spent shale to NO_x in the combustor. The principal of NO reduction with carbon in a staged combustor can be applied to reduce NO_x emissions.

3. Design Conditions -- The method for ammonia removal determines the amount of fuel nitrogen in the cleaned retort gas. Since the gas is eventually burned, this fuel-nitrogen is emitted as NO_x unless a staged combustion burner is used. The design parameters for considering the various ammonia clean-up schemes are:

1 - Water wash - NH_3 in exit gas based on pp = 0.5 mm Hg
- organic nitrogen based on 2 percent of nitrogen in retort gas and no removal with water wash
- thermal NO_x from retort gas rate, heating value and 0.2 lb/MMBtu

2 - Acid wash - same as water wash except NH_3 exit conc = 10 ppm

5.1.3 Gas Utilization and End-Of-Pipe

The End-Of-Pipe controls are those either added after combustion of the retort gas to remove particulate, NO_x and SO_x or incorporated as part of the combustion process as in staged combustion for NO_x control.

A. Particulate--

For particulate control, two alternatives were considered. The first is the base case using a standard baghouse. The second control technique is the combined dry venturi-baghouse. The dry venturi-baghouse combination provides for particulate control that is somewhat independent of type of particulate. The applicants for PSD permits all considered a minimum

particulate exit loading of 0.07 g/m^3 (0.03 gr/scf) which was based on standard technology within the limits of the unknowns associated with oil shale particulate. By capturing the small particles on larger target particles of specified physical properties, the dry venturi eliminates the major uncertainties in designing baghouses with respect to particulate type and size.

Two alternative design conditions were considered, based on the face velocity in the baghouse of 0.5 and 1.5 ft/sec (see Figure 3-3).

particulate loading g/m 3	face velocity m/sec
0.02	0.5
0.001	1.5

B. Sulfur Oxides--

If the H_2S (and organic sulfur) removal is not sufficient to reduce the sulfur emissions to an acceptable level, a post combustion flue gas desulfurization system must be added. This could be a wet or dry scrubber.

The second sulfur control technique is the use of a spent shale combustor. The combustion of the spent shale has two important advantages; 1) recovery of the energy value of the char and 2) reduction of the sulfur oxide emissions due to the scrubbing nature of the spent shale. However, the spent shale combustion also has two distinct disadvatages; 1) high emissions of NO_x from the nitrogen in the spent shale and 2) high emissions of CO due to incomplete combustion. These emissions (NO_x and CO) will be discussed in the following section.

1 - Flue Gas Desulfurization -- 50 ppm SO_x exit concentration

2 - Combustion of spent shale with retort gas

- 10 ppm SO_x
- 300 ppm NO_x
- 1000 ppm CO

C. Nitrogen Oxides--

Two controls were considered for reducing NO_x emissions. The first is ammonia injection which has been applied successfully in utility boilers and could be used when the retort gas is burned in a conventional boiler for steam and electrical generation.

The second NO_x control considered is staged combustion which has particular advantages due to the ability to adequately control the fuel related NO_x. The staged combustion could be applied to either the conventional boiler or the spent shale combustor.

There is a tradeoff between the NO_x and CO emissions in the spent shale combustor. As indicated above, the NO_x and CO emissions are quite high at 300 and 1000 ppm, respectively. Higher combustor temperatures increase the NO_x emissions but decrease the CO emissions. In the range of 600 - 800 C the NO_x emissions can range from 250 ppm to over 600 ppm while the CO emissions can vary from 200 ppm to over 1000 ppm at the lower temperatures with low excess oxygen.

In addition, the control of the fuel related NO_x depends on low excess oxygen (actually sub-stoichiometric combustion) which increases the CO emissions even higher.

The fluidized-bed combustor has limitations in process control which result in the inability to provide adequate staging for NO_x control. However, the cascading bed combustor is designed as a staged device. Consequently, combustion conditions can be controlled at each stage of the process, alternating between fuel rich and fuel lean zones to reduce the NO formed to N_2 and complete the combustion of the CO formed to CO_2.

There are no test results to support this hypothesis; however, the chemistry and engineering of the cascading bed combustor indicate that proper design could result in minimal NO_x and CO emissions.

1 - NH_3 injection - 20 ppm NOx exit concentration

2 - cascading bed combustor - 50 ppm NOx exit concentration
- 50 ppm CO

5.2 RESULTS

The results from the analysis for all five cases are shown for particulates, nitrogen oxides and sulfur oxides in Tables 5-7, 5-8 and 5-9, respectively. The following discussion presents the individual analysis for each case.

5.2.1 Direct Combustion - Case #1

A. Base Case--

The base case for this process consists of open pit mining with a Paraho type retort followed by water wash for ammonia removal and a Stretford direct sulfur conversion plant. The retort gas is then burned in a boiler to produce steam and electricity followed by a baghouse for particulate control.

The schematic for this case is shown in Figure 5-2. The expected emissions for this process for the base case and two alternatives are shown in Table 5-10.

B. Alternate #1--

Alternate 1 scheme uses room and pillar mining in place of an open pit mine, the activated carbon-hypochlorite process for H_2S and organic sulfur removal prior to combustion, an acid wash in addition to the water wash for ammonia removal and a dry venturi/baghouse combination with a face velocity of 0.5 m/sec (1.5 ft/sec).

The use of a room and pillar mine in place of an open pit mine reduces CO and HC emissions by 42 and 13 percent, respectively. NO_x emissions from mining vehicles are also reduced from 471 kg/1000 m^3 of oil to 345 kg/1000 m^3 of oil.

The activated carbon-hypochlorite process for H_2S and organic sulfur removal reduces SO_x emissions from combustion of the retort gas from 2600 kg/1000 m^3 of oil to 160 kg/1000 m^3 of oil and results in an overall facility reduction of SO_x emissions of 93 percent.

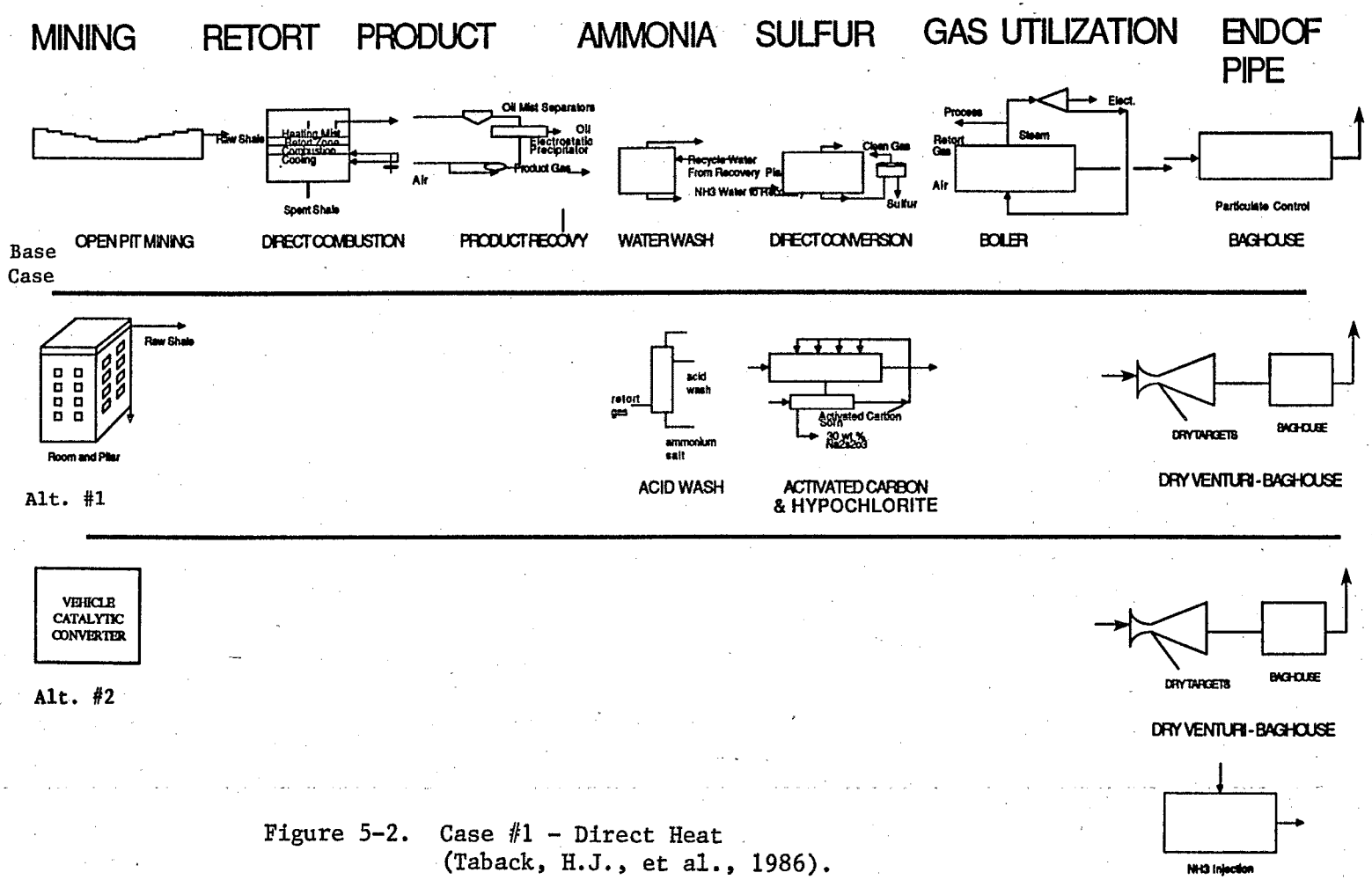


Figure 5-2. Case #1 - Direct Heat
(Taback, H.J., et al., 1986).

TABLE 5-7. CONTROLLED PARTICULATE EMISSIONS
(Taback, H.J., et al., 1986)

Case Process	HHV cal/ℓ	Retort Gas ft^3/bbl	Retort Gas m^3/m^3 Oil	Combustion Dilution Ratio	Gas Flow m^3/m^3 Oil	Base Part. Loading g/m^3	Base kg/1000 m^3 of Oil	Alt. #1 Face Velocity = 1.5 ft/sec Part. Loading g/m^3	Alt. #1 kg/1000 m^3 of Oil	Alt. #2 Face Velocity = 3.5 ft/sec Part. Loading g/m^3	Alt. #2 kg/1000 m^3 of Oil
1. Direct combustion	9000	10000	1761	2	3521	0.07	246	0.02	70	0.001	4
2. Direct combustion circular	9000	10000	1761	2	3521	0.07	246	0.02	70	0.001	4
3. Indirect combust-gas	90000	1000	176	15	2641	0.07	185	0.02	53	0.001	3
4. Indirect combust-solids-recy.	9000	1000	176	15	2641	0.07	185	0.02	53	0.001	3
fluidized bed comb.		53000	9331	1	9331	0.07	653	0.02	187	0.001	9
TOTAL		54000	9507		11972	0.07	838	0.02	24	0.001	12
5. Modified in-situ[a] above	9000	30000	5282	2	10563	0.07	739	0.02	211	0.001	11
ground-ind. gas recy.	90000	400	70	15	1056	0.07	74	0.02	21	0.001	1
TOTAL		30400	5352		11619		813		232		12

(a) Based on 60% of oil produced

TABLE 5-8. CONTROLLED SULFUR OXIDE EMISSIONS FROM RETORT GAS COMPOSITION
(Taback, H.J., et al., 1986)

Case Process	Stretford, Unisulf, LoCat, etc.					Activated Carbon and Hypochlorite Process				
	H_2S In Offgas (Base Case) ppm	Organic S In Offgas (5% of H_2S) kg/1000 m^3 Oil	SO_x Emissions From H_2S, kg/1000 m^3 Oil H_2S	SO_x(a) Emissions Total, kg/1000 m^3 Oil	SO_x(a) Emissions Total ppm	H_2S In Offgas ppm	Removal Effic. for Organic Sulfur, %	SO_x Emissions From H_2S, kg/1000 m^3 Oil	SO_x(a) Emissions Total kg/1000 m^3 Oil	SO_x Emissions Total ppm
1. Direct combustion	50	1200	199	2600	736	10	90	40	160	45
2. Direct combustion-circular grate	---(a)	0		370	175	---(b)			106	50
3. Indirect combustion-gas recycle	50	1200	20	2400	914	10	90	4	124	47
4. Indirect combustion-solid recycle	50	43(c)	20	106	40	10	90	4	8	3
+ fluidized	20			224	24				224	24
5. Modified in-situ (and) above ground-direct	50	720	600	2030	192	10	90	119	191	18
comb.	50	480	8	970	916	10	90	2	50	47
combined process	50	1200	600	3000		10	90	121	240	

(a) Total SO_2 emission from combustion of H_2S and organic sulfur in the offgas

(b) No H_2S produced in this process

(c) Low organic sulfur due to solids capture of H_2S in retort

TABLE 5-9. CONTROLLED NITROGEN OXIDE EMISSIONS
(Taback, H.J., et al., 1986)

Case Process	NH_3 Part. Press. = 0.5mmHg Exit kg/1000 m^3 Oil	Organic Nitrogen 2% of N	Fuel NO_x kg/1000 m^3 of Oil	Thermal NO_x kg/1000 m^3 of Oil	Total NO_x kg/1000 m^3 of Oil	Exit Conc NO_x ppm	Alt. #1 Acid Wash NH_3, ppm = 10 Organic N = 2% Fuel NO_x kg/1000 m^3 of Oil	Total NO_x kg/1000 m^3 of Oil	Exit Conc NO_x ppm		Alt. #2 NH_3 Injection Exit Conc NO_x ppm	Total NO_x kg/1000 m^3 Oil
1. Direct combustion	790	256	2242	570	2812	799	586	1156	328	NH_3-inj	20	85
2. Direct combustion-circular	790	256	2242	570	2812	799	586	1156	328	NH_3-inj	20	85
3. Indirect combust-gas recycle	79	85	351	570	921	349	186	756	286	NH_3-inj	20	63
4. Indirect combust-solid recycle	79	85	351	570	921	349	186	756	286	NH_3-inj	20	63
fluidized	300[a]		3359	0	3359	360		3359	360	Stgd/cbn	50	560
			0	0	4281			4115				623
lift gas					3457			3291				486
5. Modified in-situ-	2370	258	5632	1710	7342	695	665	2375	225	NH_3-inj	20	254
above ground-in	32	34	141	228	369	349	74	302	286	NH_3-inj	20	25
												279

(a) Based on 15% conversion of nitrogen in spent shale to NO_x

TABLE 5-10. TOTAL FACILITY CONTROLLED EMISSIONS
Case 1 - Direct Combustion
(Paraho type retort)
(Taback, H.J., et al., 1986)

Base Conditions: H_2S-Stretford (50 ppm), NH_3-water wash
Particulate loading = 0.07 g/m^3

	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Open pit	370	50	471	38	410
Retort gas	125	35	2812	2593	246
Upgrade	25	150			
Total	520	235	3283	2631	656

Alternate #1 SO_x-H_2S removal - activated carbon and hypochlorite
NO_x-NH_3 removal - water and acid wash
PM - dry venturi and baghouse
Room and pillar mining

	CO	HC	NO_x	SO_x	PM
Room and pillar	150	20	345	20	181
Retort gas	125	35	1156	159	70
Upgrade	25	150			
Total	300	205	1501	179	251
Reduction (from Base Case)	220	30	1782	2451	405
% Reduction	42	13	54	93	62

Alternate #2 NO_x - ammonia injection
PM - dry venturi/baghouse
HC and CO - vehicles/catalytic converter

	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	85	159	4
Upgrade	2	60			
Total	142	97	430	179	185
Reduction (from Alt. #1)	158	108	1071	0	67
% Reduction	52	53	71	0	27

The use of an acid wash in addition to the water wash for ammonia removal reduces NO_x emissions from combustion of the retort gas from 2800

kg/1000 m^3 of oil to 1200 kg/1000 m^3 of oil and results in an overall facility reduction (combined effect of room & pillar mining and the acid wash for ammonia removal) of 54 percent.

The dry venturi/baghouse for an end-of-pipe particulate control reduces particulate emissions from combustion of the retort gas from 246 kg/1000 m^3 of oil to 70 kg/1000 m^3 of oil and the combined effect of room and pillar mining and the dry venturi/baghouse reduces the overall facility particulate emissions by 62 percent.

C. Alternate #2--

The Alternate 2 scheme uses catalytic converters on the mining vehicles and engines to reduce CO and HC emissions, ammonia injection to reduce NO_x from combustion of the retort gas and the dry venturi/baghouse with a space velocity of 1.0 m/sec (3.5 ft/sec) for additional particulate control.

The catalytic converters reduce the CO and HC emissions from the mining operations by 90 percent and result in an overall facility reduction of 52 percent for CO and 53 percent for HC.

The use of ammonia injection reduces the NO_x emissions from 1200 kg/1000 m^3 of oil to 85 kg/1000 m^3 of oil for retort gas and results in an overall facility reduction of 71 percent.

The dry venturi/baghouse with the higher space velocity reduces particulate emissions from combustion of the retort gas to 4 kg/1000 m^3 of oil and results in an overall facility reduction of 27 percent.

5.2.2 Direct Combustion - Circular Grate - Case # 2

A. Base Case--

The base case for this process consists of room and pillar mining with a circular grate retort (e.g., Superior, Allis Chalmers, Dravo) followed by a boiler and a baghouse. As this type of retort provides combustion of a portion of the retort gas, the sulfur is converted to sulfur oxides prior to the boiler and no H_2S removal is proposed. The retort contains a water wash for ammonia and sulfur oxide removal as an integral part of the process.

The schematic for this case are shown in Figure 5-3. The emissions expected for this schematic and the two alternatives are shown in Table 5-11.

TABLE 5-11. TOTAL FACILITY CONTROLLED EMISSIONS
Case 2 - Direct Combustion
(Circular grate retort)
(Taback, H.J., et al., 1986)

Base Case	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Room and pillar	150	20	345	20	181
Retort gas	125	35	2812	370	246
Upgrade	25	150			
Total	300	205	3157	390	427
Alternate #1 SO_x - Flue Gas Desulfurization					
NO_x-NH_3 removal - water and acid wash					
PM - dry venturi/baghouse					
HC and CO - vehicles/catalytic converter					
	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	1156	106	70
Upgrade	2	60			
Total	142	97	1501	126	251
Reduction (from Base Case)	158	108	1656	264	176
% Reduction	52	53	52	68	41
Alternate #2 NO_x - ammonia injection					
PM - dry venturi/baghouse					
	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	85	106	4
Upgrade	2	60			
Total	142	97	430	126	185
Reduction (from Alt. #1)	0	0	1226	0	66
% Reduction	0	0	74	0	26

B. Alternate #1--

The first alternate scheme uses catalytic converters for the mining engines and vehicles to reduce CO and HC, flue gas desulfurization after

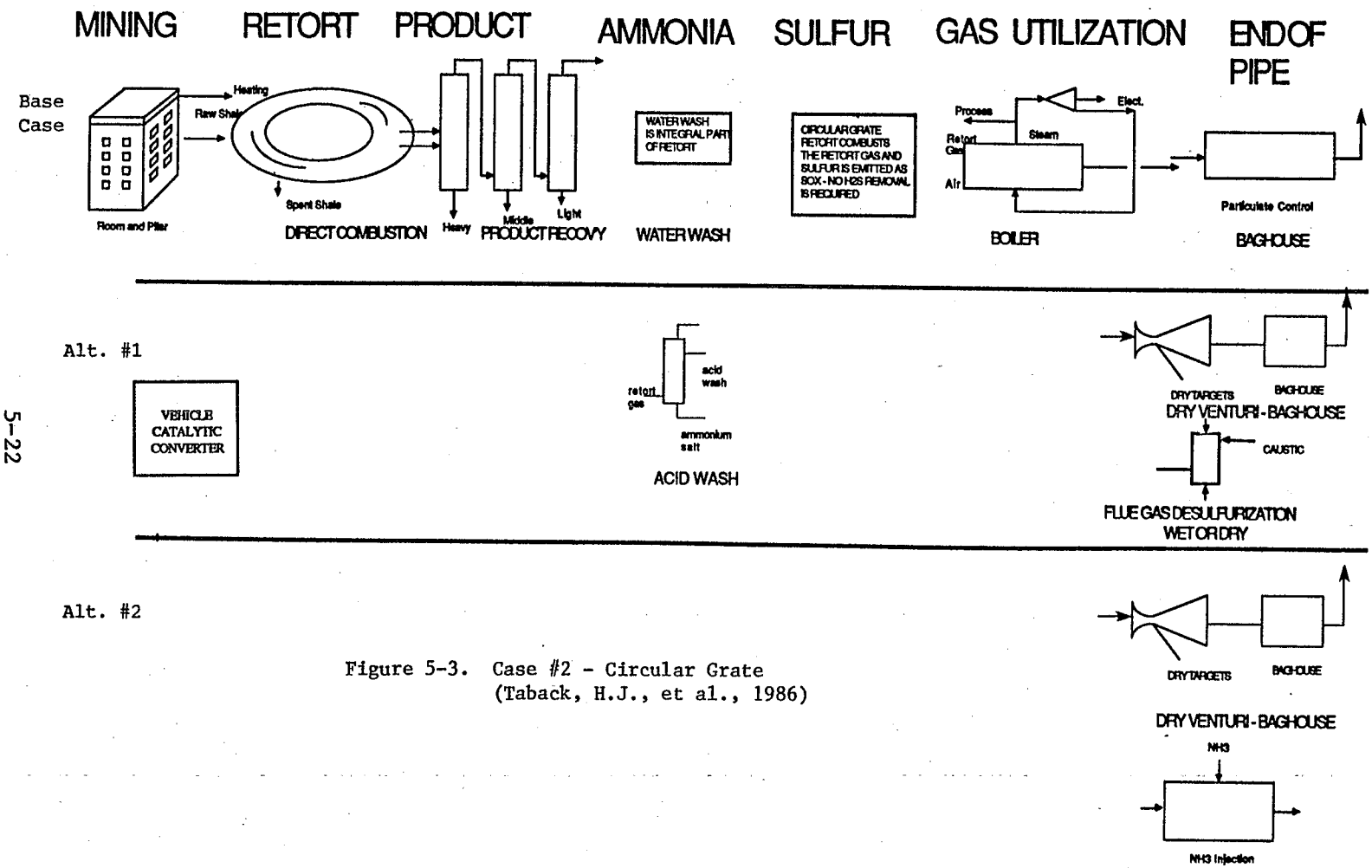


Figure 5-3. Case #2 - Circular Grate
(Taback, H.J., et al., 1986)

combustion to reduce the SO_x emissions, an acid wash for additional ammonia removal and a dry venturi/baghouse for particulate control.

The CO and HC emissions from mining vehicles and engines are reduced by 90 percent and the overall facility emission is reduced by 52 and 53 percent, respectively.

The acid wash reduces the NO_x emissions from retort gas combustion from 2800 kg/1000 m^3 of oil to 1200 kg/1000 m^3 of oil and the overall facility reduction is 52 percent.

The flue gas desulfurization reduces the SO_x emissions from retort gas combustion from 370 kg/1000 m^3 to 106 kg/1000 m^3 and reduces the overall facility SO_x emission rate by 68 percent.

The dry venturi/baghouse with a space velocity of 0.5 m/sec (1.5 ft/sec) reduces particulate from retort gas combustion from 246 kg/1000 m^3 of oil to 70 kg/1000 m^3 of oil and results in an overall facility reduction in particulate emissions of 41 percent.

C. Alternate #2--

The second alternative scheme uses ammonia injection for post combustion NO_x control and the dry venturi/baghouse with a space velocity of 1 m/sec (3.5 ft/sec).

The NO_x emissions from combustion of the retort gas are reduced from 1156 kg/1000 m^3 to 85 kg/1000 m^3 of oil and the overall facility NO_x emission is reduced by 74 percent.

The particulate emissions from combustion of the retort gas are reduced from 70 kg/1000 m^3 to 4 kg/1000 m^3 and the overall facility particulate emission is reduced by 26 percent.

5.2.3 Indirect Combustion - Gas Recycle- Case # 3

A. Base Case--

The base case for this process consists of open pit mining with an indirect combustion - gas recycle retort followed by a water wash for

ammonia removal and an direct conversion process for H_2S removal. The retort gas is burned in a boiler followed by a baghouse for particulate control.

The schematic for this process is shown in Figure 5-4. The emission rates for the three alternatives are presented in Table 5-12.

TABLE 5-12. TOTAL FACILITY CONTROLLED EMISSIONS
Case 3 - Indirect Combustion
Gas Recycle Retort (Union B)
(Taback, et al., 1986)

Base Case	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Open pit	370	50	471	38	410
Retort gas	125	35	2812	2593	246
Upgrade	25	150	------ No Data		----------
Total	520	235	3283	2631	656
Alternate #1	SO_x-H_2S removal - activated carbon and hypochlorite				
	NO_x-NH_3 removal - water and acid wash				
	PM - dry venturi and baghouse				
	Room and pillar mining				
	CO	HC	NO_x	SO_x	PM
Room and pillar	150	20	345	20	181
Retort gas	125	35	1156	159	70
Upgrade	25	150	-------No Data		----------
Total	300	205	1501	179	251
Reduction (from Base Case)	220	30	1782	2451	405
% Reduction	42	13	54	93	62
Alternate #2	NO_x - ammonia injection				
	PM - dry venturi/baghouse				
	HC and CO - vehicles/catalytic converter				
	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	85	159	4
Upgrade	2	60	------- No Data		--------
Total	142	97	430	179	185
Reduction (from Alt. #1)	158	108	1071	0	67
% Reduction	52	53	71	0	27

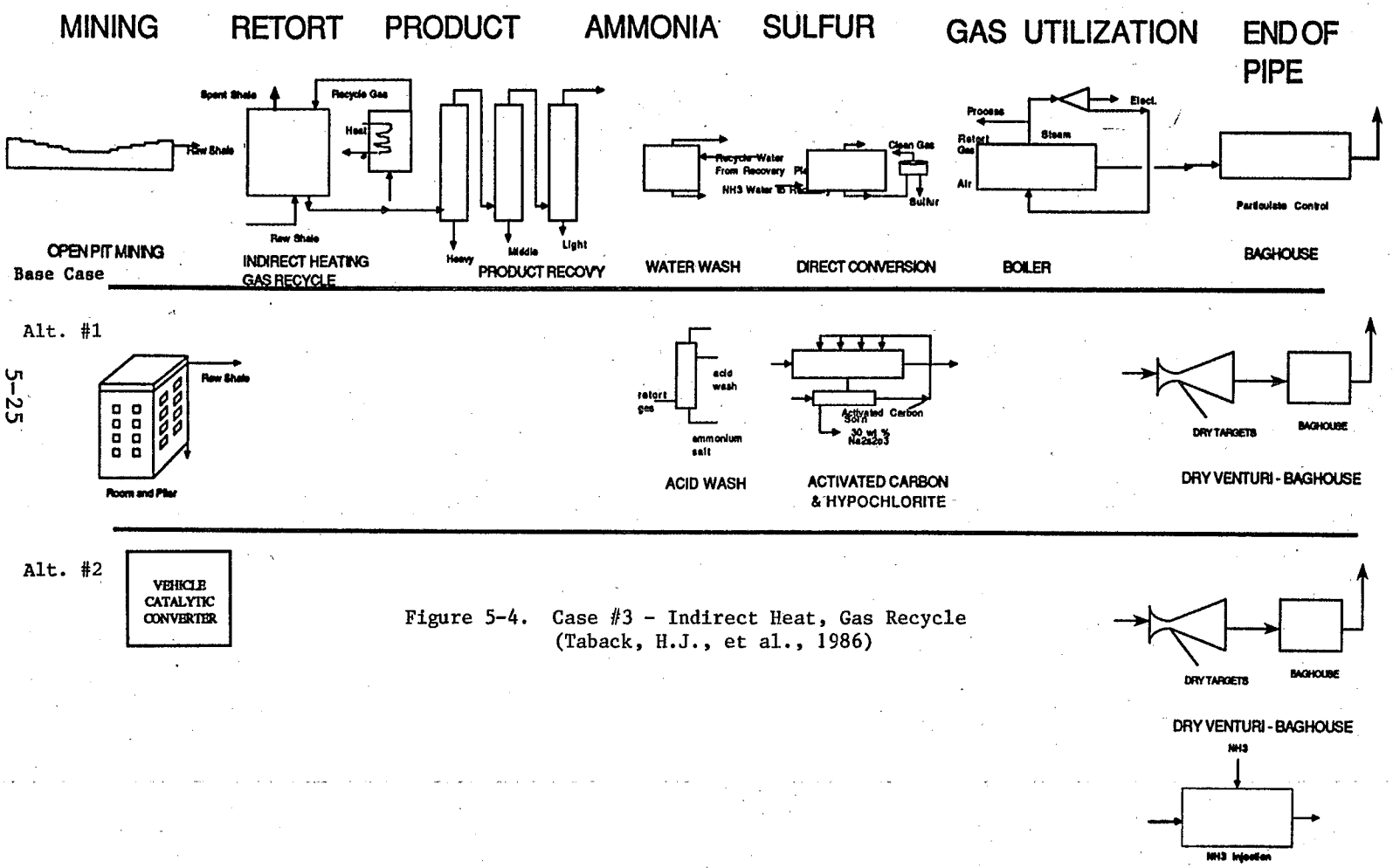


Figure 5-4. Case #3 - Indirect Heat, Gas Recycle
(Taback, H.J., et al., 1986)

B. Alternate #1--

The first alternative consists of using an acid wash for ammonia removal, activated carbon and hypochlorite process for H_2S and organic sulfur removal, the dry venturi/baghouse for particulate control and a room and pillar mine.

The NO_x emissions are reduced from 2812 to 1156 kg/1000 m^3 of oil from combustion of the retort gas and the overall facility NO_x emissions are reduced by 54 percent.

The SO_x emissions from combustion of the retort gas are reduced from 2593 to 159 kg/1000 m^3 of oil and the overall facility SO_x emissions are reduced by 93 percent.

The particulate emissions are reduced from 246 to 70 kg/1000 m^3 of oil and the overall facility particulate emissions are reduced by 62 percent.

C. Alternate #2--

The second alternate process scheme uses ammonia injection for post combustion NO_x control and the dry venturi/baghouse with the high space velocity, and catalytic converters on vehicles.

The NO_x emissions from combustion of the retort gas are reduced from 1156 to 85 kg/1000 m^3 of oil and the overall facility NO_x emissions are reduced by 71 percent.

The particulate emissions from combustion of the retort gas are reduced from 70 to 4 kg/1000 m^3 of oil for an overall facility reduction of 27 percent.

5.2.4 Indirect Combustion - Solids Recycle - Fluidized Bed Spent Shale Combustor - Case # 4

A. Base Case--

The base case is room and pillar mining and an indirect combustion solids recycle retort with a fluidized bed spent shale combustor. The retort gas is burned and the flue gas is treated by a baghouse.

The schematic for this process is shown in Figure 5-5. The emission rates for the three alternatives are presented in Table 5-13.

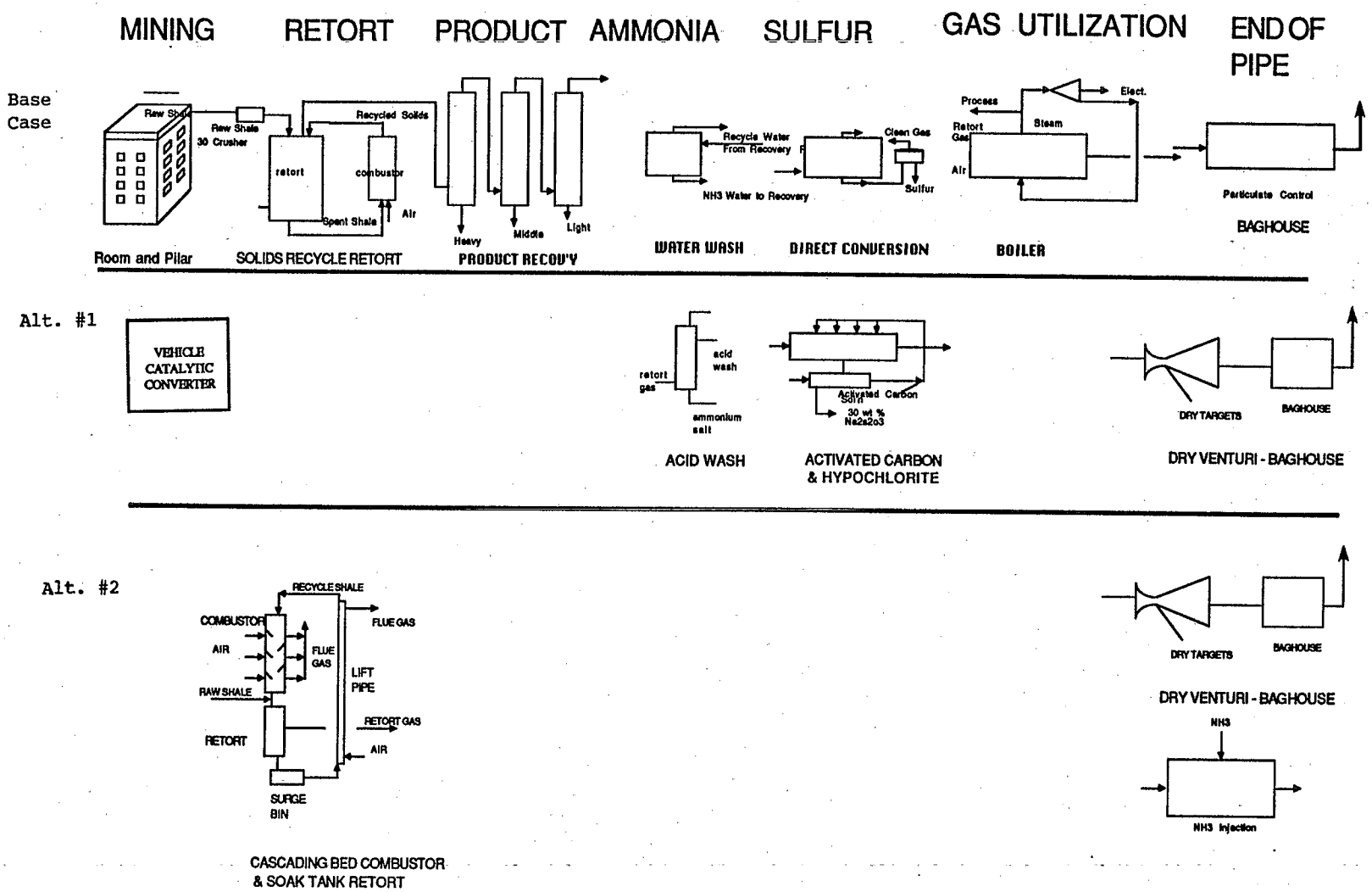


Figure 5-5. Case #4 Solids Recycle.
(Taback, H.J., et al., 1986)

TABLE 5-13. TOTAL FACILITY CONTROLLED EMISSIONS
Case 4 - Indirect Combustion, Solids Recycle
(Fluidized bed combustor)
(Taback, et al., 1986)

Base Case	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Room and pillar	150	20	345	20	181
Retort gas	125	35	921	106	185
Combustion	11197		3359	224	653
Upgrade	25	150	---------No Data	----------	
Total	11497	205	4626	350	1019

Alternate #1 SO_x-H_2S removal - activated carbon and hypochlorite

NO_x-NH_3 removal - water and acid wash

PM - dry venturi/baghouse

HC and CO - vehicles/catalytic converter

	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	756	8	53
Combustion	11197		3359	224	187
Upgrade	2	60	-----No Data	----------	
Total	11340	97	4460	252	420
Reduction (from Base Case)	158	108	166	98	599
% Reduction	1	53	4	28	59

Alternate #2 NO_x - ammonia injection for retort staged combustion for combustor gas cascading bed spent shale combustor

PM - dry venturi

	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas	125	35	63	8	3
Combustion	423		423	169	9
Upgrade	2	60	------ No Data	----------	
Total	565	97	831	197	193
Reduction (from Alt. #1)	10775	0	3629	55	227
% Reduction	95	0	81	22	54

B. Alternate #1--

The first alternate uses an acid wash to remove ammonia from the retort gas, activated carbon and hypochlorite process to remove H_2S and organic sulfur from the retort gas and the dry venturi/baghouse for particulate control on the burned retort gas and the exhaust gas from the spent shale combustion unit, and catalytic converters on mine vehicles to reduce CO and HC levels by 90 percent.

The acid wash reduces the NO_x emissions associated with the retort gas from 921 kg/1000 m^3 of oil to 756 kg/1000 m^3 of oil. However, because the NO_x emissions from the spent shale combustor are so high (3400 kg/1000 m^3 of oil) the reduction in facility NO_x emissions is only 4 percent.

The activated carbon H_2S removal process reduces the SO_x emissions from combusting the retort gas from 106 kg/1000 m^3 of oil to 8 kg/1000 m^3 of oil. The overall facility SO_x emission is reduced by 28 percent.

The dry venturi/baghouse with a space velocity of 0.5 m/sec (1.5 ft/sec) reduces particulate from the retort gas and combustor flue gas from 838 kg/1000 m^3 to 240 kg/1000 m^3. The overall facility particulate emission is reduced by 59 percent.

C. Alternate #2--

The second alternative uses ammonia injection for NO_x reduction from burning the retort gas and staged combustion for NO_x reduction from the spent shale combustor.

A fluidized bed is essentially a single stage device where high turbulence provides equal concentrations of reactants throughout the bed. This fact makes it difficult to design a single fluidized bed unit to provide the temperature and excess oxygen control necessary to reduce the NO formed in the combustion process. The staging effect could be achieved by using a series of fluidized beds followed by combustion chambers; however, this concept would result in multiple units requiring solids transfer and would be quite complicated. Therefore, to utilize this NO_x control, a cascading bed combustor as described in Section 2.4.9 is used in place of the fluidized bed. As the cascading bed is essentially a staged device, it provides easy access and means for control of individual stage temperatures and

concentrations. Therefore, the NO reduction can be achieved without the excessive CO emissions that are typical of the fluidized bed combustor.

The dry venturi/baghouse with the higher space velocity is also applied.

Ammonia injection for the combusted retort gas reduces the NO_x emissions from 756 to 63 kg/1000 m^3 of oil. The staged combustion on the cascading bed spent shale combustor reduces the NO_x emissions from 3359 to 423 kg/1000 m^3 (based on an exit concentration of 50 ppm NO_x). The overall facility NO_x emission is reduced by 81 percent.

5.2.5 Modified In-situ With Indirect Combustion, Gas Recycle Above Ground Process - Case # 5

A. Base Case--

The base case consists of a modified in-situ retort supported by an indirect combustion - gas recycle above ground retort. The in-situ retort produces 60 percent of the oil.

The gas from the in-situ retort is cleaned with the standard water wash for ammonia removal and either direct or indirect conversion of H_2S for sulfur gas control. The combusted gas is passed through a baghouse for particulate control. The schematic for this process is shown in Figure 5-6. The expected emissions for three alternatives are presented in Table 5-14.

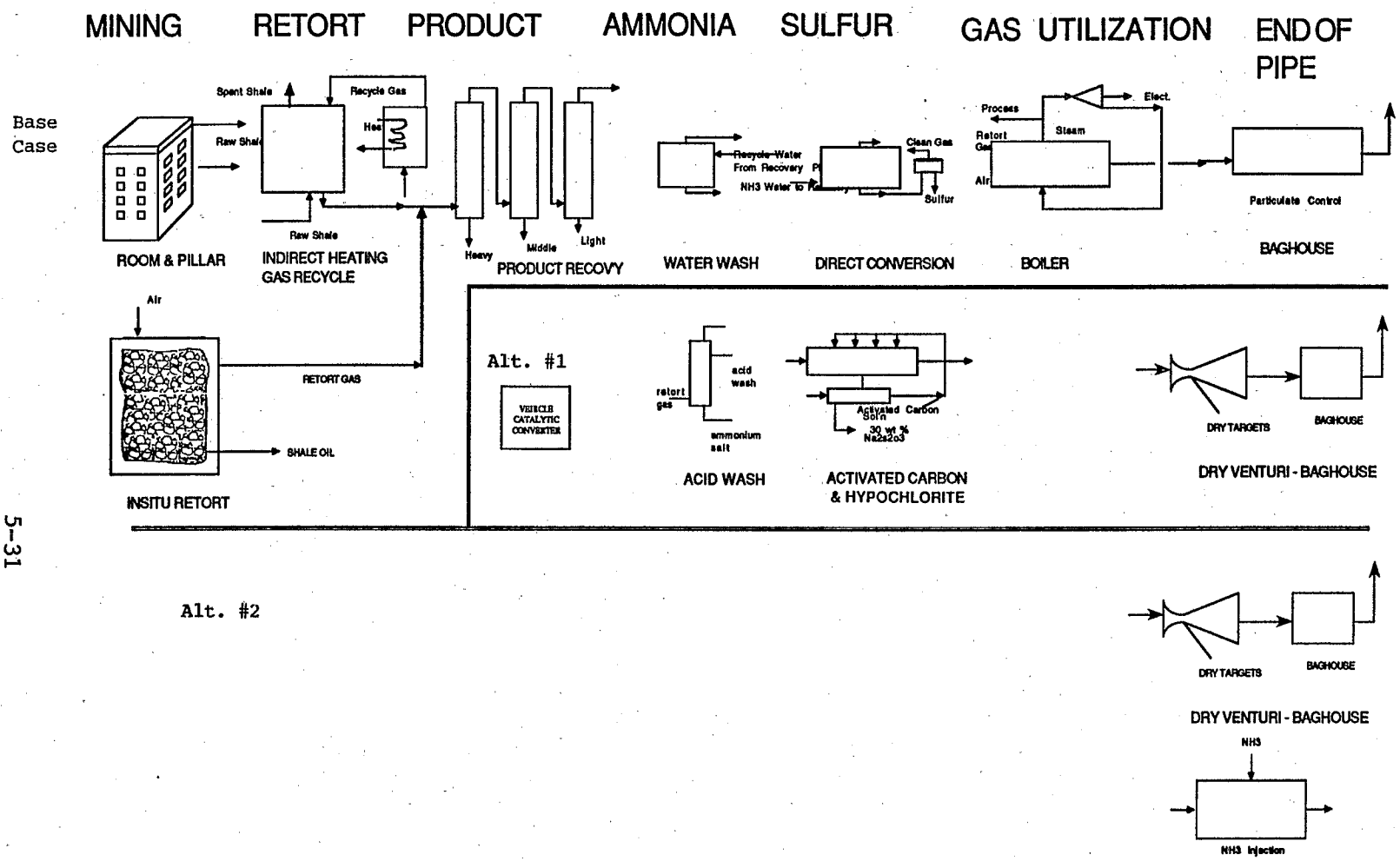


Figure 5-6. Case #5 Modified In-Situ (Indirect Heat Above-Ground). (Taback, H.J., et al., 1986)

TABLE 5-14. TOTAL FACILITY CONTROLLED EMISSIONS
Case 5 - Modified In-situ
Indirect Combustion Above-Ground

Base Case	Emissions, kg/1000 m^3 of Oil				
	CO	HC	NO_x	SO_x	PM
Room and pillar	150	20	345	20	181
Retort gas in-situ	72	35	7342	2033	739
Above-ground	53		369	968	74
Upgrade	25	150			
Total	300	205	8056	3020	994

Alternate #1 SO_x-H_2S removal - activated carbon

NO_x-NH_3 removal - water and acid wash

PM - dry venturi/baghouse

HC and CO - vehicles/catalytic converter

	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas in-situ	72	35	2375	191	211
Above-ground	53	0	302	50	21
Upgrade	2	60			
Total	142	97	3022	261	413
Reduction (from Base Case)	158	108	5034	2760	581
% Reduction	52	53	62	91	58

Alternate #2 NO_x - ammonia injection

PM - dry venturi/baghouse

HC and CO - vehicles/catalytic converter

	CO	HC	NO_x	SO_x	PM
Room and pillar	15	2	345	20	181
Retort gas in-situ	72	35	254	191	11
Above-ground	53	0	25	50	1
Upgrade	2	60			
Total	142	97	624	261	193
Reduction (from Alt. #1)	0	0	2398	0	221
% Reduction	0	0	79	0	53

B. Alternate 1--

The first alternate process scheme uses catalytic converters for mine vehicles, an acid and water wash for ammonia removal, the activated carbon-hypochlorite process of H_2S and organic sulfur removal and the dry venturi/baghouse for particulate control.

The acid wash reduces the NO_x emissions from burning the in-situ retort gas from 7342 to 2375 kg/1000 m^3 and reduces the NO_x emissions from burning the above ground retort gas from 369 to 302 kg/1000 m^3. The higher effect on the in-situ retort gas is due to relatively high gas volume produced by the in-situ retort which results in high ammonia emission rates from either the water wash or acid wash towers. The overall facility NO_x emission is reduced by 62 percent.

The activated carbon-hypochlorite process reduces the SO_x emissions from 3020 to 261 kg/1000 m^3 of oil for a 91 percent reduction.

The dry venturi/baghouse reduces the particulate emissions associated with burning the retort gases (in-situ & above ground) from 813 to 230 kg/1000 m^3 of oil. The overall facility reduction in particulate emissions is 58 percent. The use of catalytic converters on mine vehicles reduces the CO and HC levels from mining by 90 percent, and the total CO and HC levels by 52 and 53 percent respectively.

C. Alternate #2--

The second alternate control scheme uses ammonia injection for post combustion NO_x control and the dry venturi/baghouse with the higher space velocity for particulate control.

NO_x emissions from combustion of the retort gas are reduced from 2677 to 279 kg/1000 m^3 of oil for an overall facility reduction of 79 percent. Particulate emissions from combustion of the retort gas are reduced from 232 to 12 kg/1000 m^3 of oil for an overall facility reduction of 53 percent.

The above information for Case #5 is summarized in Table 5-14.

5.2.6 Summary

The following discussion evaluates the differences between the five cases and the effect on the overall facility emissions.

A. Carbon Monoxide and Hydrocarbon Emissions--

Carbon monoxide emissions for the base case conditions are reasonably consistent with the exception of case #4, the indirect combustion, solids recycle with a fluidized bed spent shale combustor which has extremely high CO emissions. The alternate 1 conditions reduce the CO emissions by approximately 90 percent due to the addition of catalytic converters to the mine vehicles and engines. However, this has little effect on Case #4 CO emissions. The alternate 2 conditions, using a cascading bed spent shale combustor with its superior means for controling combustion conditions at each stage of the combustion process reduces the CO emissions to 570 kg/1000 m^3 of oil; however, the facility CO emissions for Case #4 are still considerably higher than the other four cases. No additional means of control seems feasible.

The hydrocarbon emissions are consistent for all of the cases considered. The addition of the catalytic converters reduce these hydrocarbon emissions from the mining operations by 90 percent and no further reduction seems feasible.

B. Particulates, Nitrogen and Sulfur Oxides (PM, NO_x, SO_x)--

1. Retort Gas Combustion -- The emissions from combustion of the retort gas for particulates, nitrogen and sulfur oxides are shown in Figures 5-7, 5-8 and 5-9.

These figures indicate that there is wide variation in emission levels for the five processes based on the PSD permit application proposed technology (Base Case conditions). For particulates, the emission levels vary from 200 to 800 kg/1000 m^3 of oil; for nitrogen oxides the emission levels vary from 1000 to 8000 kg/1000 m^3 of oil; for sulfur oxides the emission levels vary from 350 to 3000 kg/1000 m^3 of oil.

The first alternative considered was the use of the activated carbon enhanced H_2S removal process, an acid wash for improved ammonia removal and the addition of a dry venturi - baghouse for post combustion particulate control. Referring to Figures 5-7, 5-8 and 5-9, the emission levels for alternate #1 show considerably less variation, particularly for sulfur oxides (ranging from 100 to 250 kg/1000 m^3 of oil) and particulates (range from 50 to

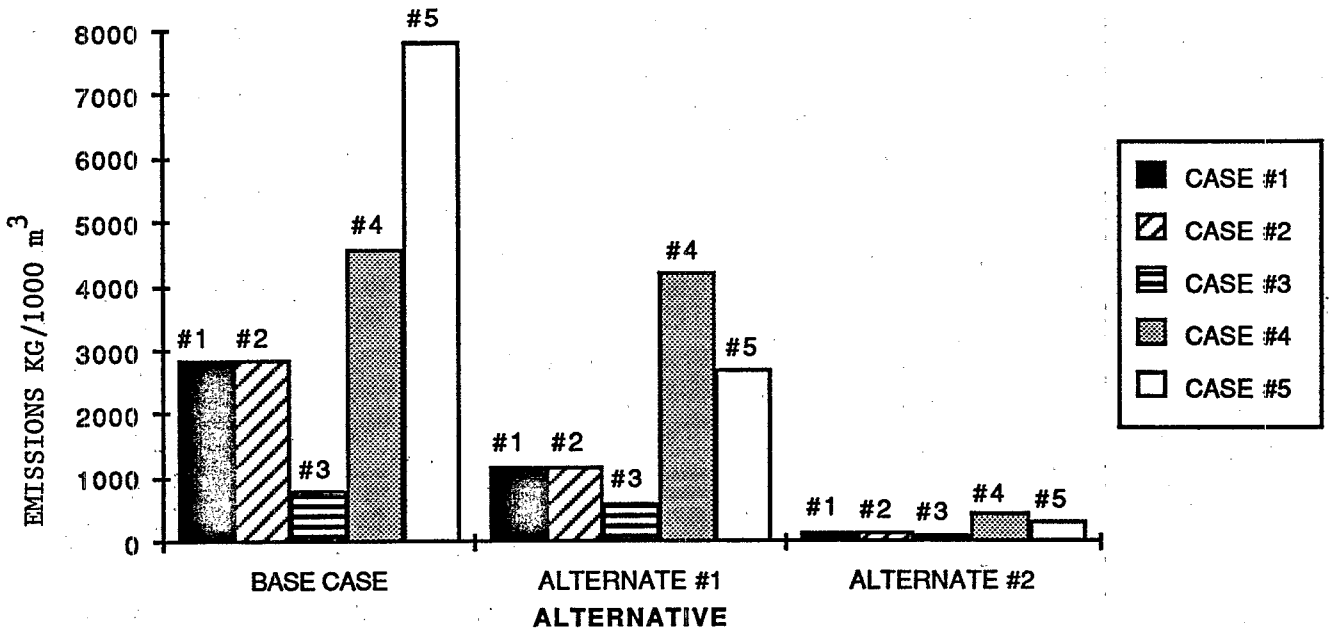


Figure 5-7. Nitrogen oxide emissions from retort gas combustion. Summary for five cases. (Taback, H.J., et al., 1986)

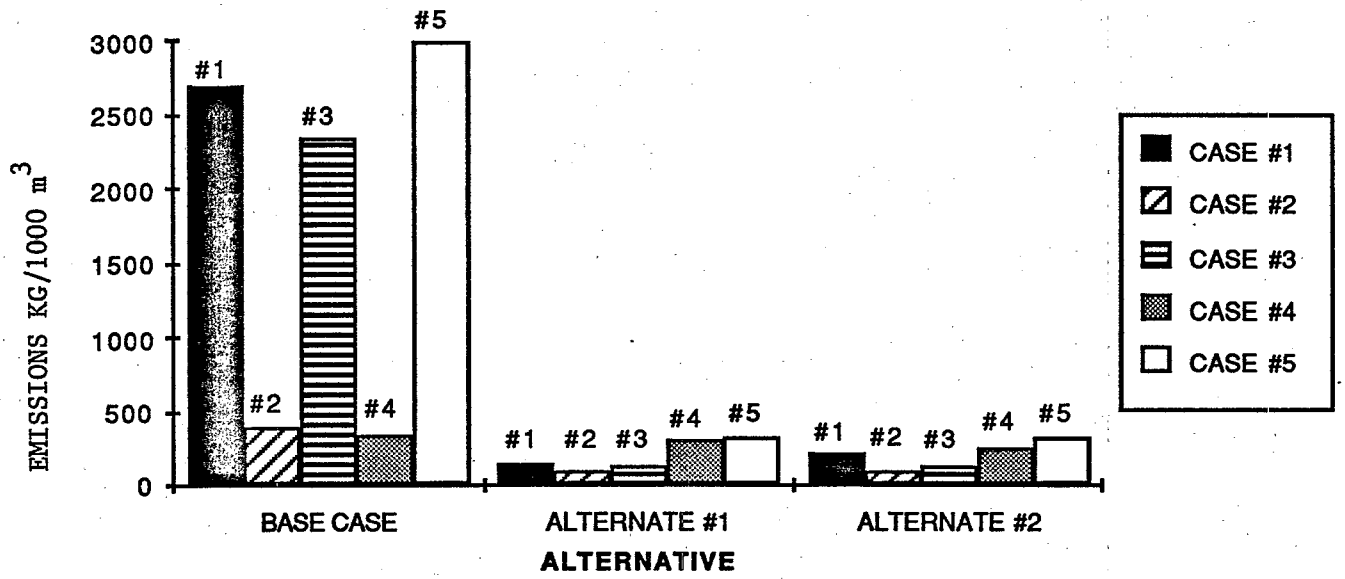


Figure 5-8. Sulfur oxide emissions from retort gas combustion. Summary for five cases. (Taback, H.J., et al., 1986)

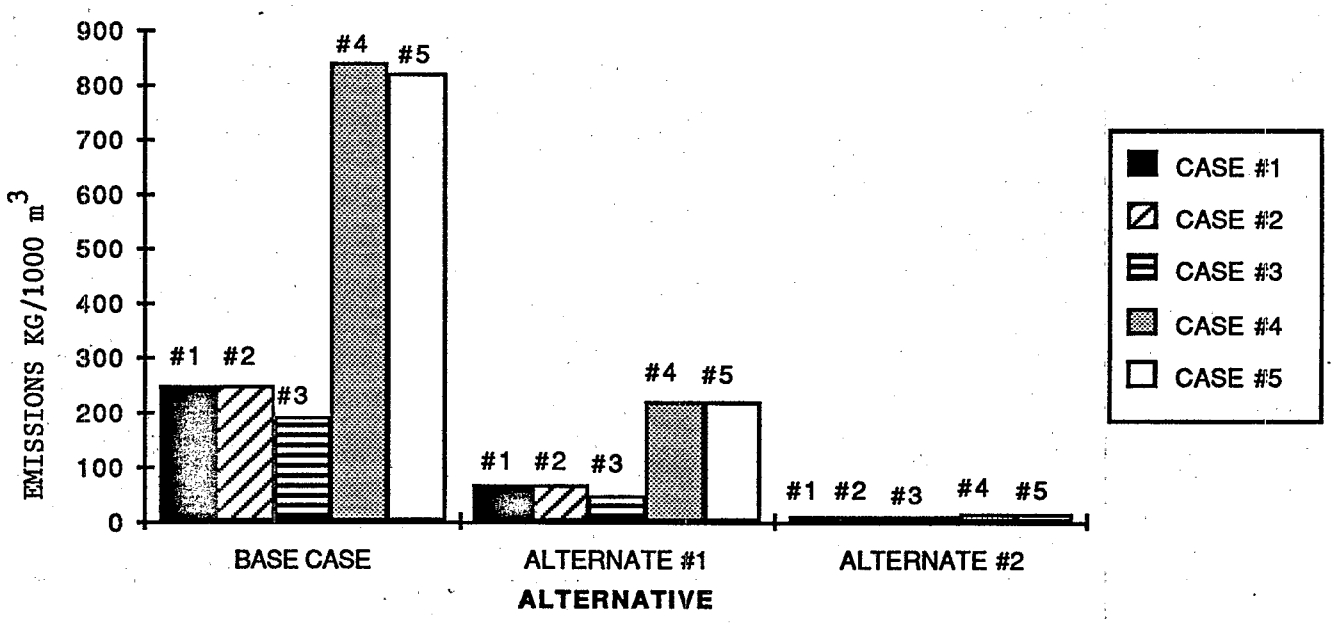


Figure 5-9. Particulate emissions from retort gas combustion. Summary for five cases. (Taback, H.J., et al., 1986)

200 kg/1000 m^3 of oil). The variation of nitrogen oxide emissions is still considerable, ranging from 1000 to 4000 kg/1000 m^3 of oil. Essentially, the acid wash only removes the residual ammonia without affecting the organic nitrogen content and has no effect on the thermal NO_x; therefore, there is relatively little improvement in the NO_x emission rate.

The second alternative considered was ammonia injection for NO_x control from boiler and/or furnace combustion, staged combustion for control of NO_x emission from the spent shale combustor and the dry venturi - baghouse with an increased space velocity which improves collection performance at the expense of increased pressure drop. Again, referring to Figures 5-7, 5-8 and 5-9, it is apparent that the addition of these controls essentially levels the performance of all five processes.

2. Total Facility Emissions -- The particulate, nitrogen oxide and sulfur oxide emission levels for alternate #2 conditions along with the total facility emissions are shown in Figures 5-10, 5-11 and 5-12. The particulate emissions (Figure 5-9) still show variation from 4 to 12 kg/1000 m^3 of oil. However, the absolute value is considerably less than the particulate emissions from the mining and solids handling operations and the total facility is essentially equivalent for all five cases ranging from 180 to 200 kg/1000 m^3 of oil.

The nitrogen oxide emissions, Figure 5-11, range from 75 to 500 kg/1000 m^3 of oil. While this is still a significant variation, again the absolute magnitude of the values is such that the net variation in the total NO_x emission for the five facilities is less than 2 to 1 ranging from 400 to 800 kg/1000 m^3 of oil.

The sulfur oxide emissions, Figure 5-12, range from 100 to 250 kg/1000 m^3 of oil and are essentially the same for the total facility as there are no other significant sources of sulfur emissions.

The basic conclusion derived from the above analysis is that, although the air emission levels for the different retort processes with controls proposed in the PSD permit applications can vary considerably, sometimes by as much as two orders of magnitude, the application of control techniques that

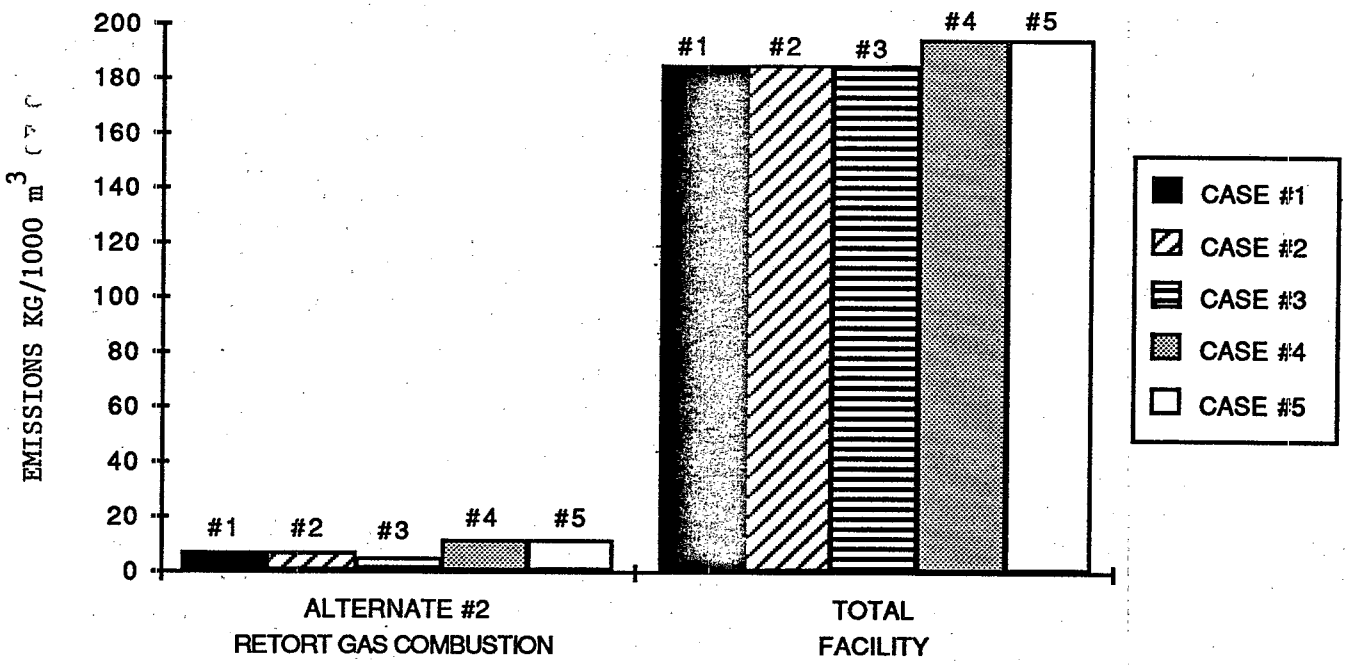


Figure 5-10. Particulate emissions for total facility. Summary for five cases. (Taback, H.J., et al., 1986).

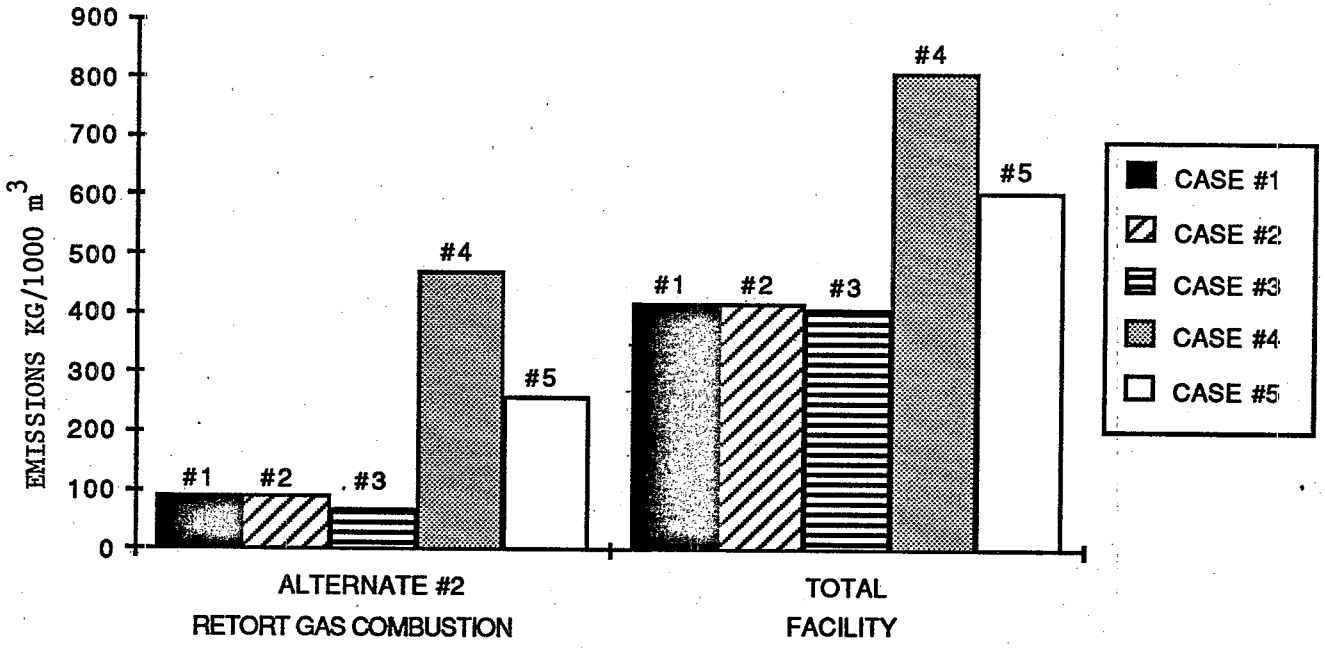


Figure 11. Nitrogen oxide emissions for total facility. Summary for five cases (Taback, H.J., 1986).

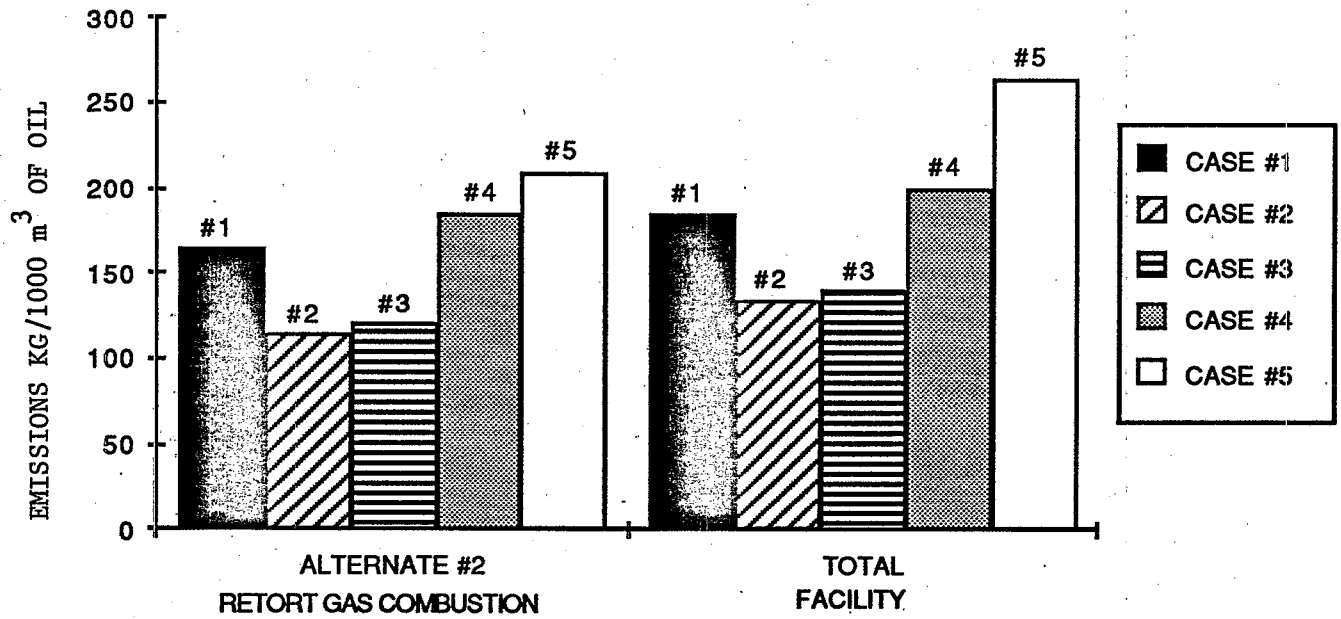


Figure 12. Sulfur oxide emissions for total facility. Summary for five cases (Taback, H.J., et al., 1986).

are either improvements over proposed technology or more suitable for a specific application, result in similar emission levels for all five processes considered. This statement does need to be qualified by the fact that some of the control techniques considered have not been applied specifically to the oil shale recovery process. However, these techniques have been proven at the full scale level in various other difficult control applications.

are either improvements over proposed technology or more suitable for a specific application, result in similar emission levels for all five processes considered. This statement does need to be qualified by the fact that some of the control techniques considered have not been applied specifically to the oil shale recovery process. However, these techniques have been proven at the full scale level in various other difficult control applications.

SECTION 6.0

REFERENCES

Agarwal, A. K. Assessment of Solid Waste Characteristics and Control Technology for Oil Shale Retorting. EPA-600/7-86-019, NTIS PB86 198371, 1986.

Anonymous. Parachute Creek Shale Oil Program. Union Oil Company Brochure, 1983, 12 pp.

Anonymous. Clear Creek Shale Oil Project - Joint Venture. PSD Permit Application for Upgrading in Grand Valley. Environmental Research & Tech., Fort Collins, CO, 1982.

Anonymous. Permit Application. Paraho-UTE Facility. Prevention of Significant Deterioration. Utah Air Conservation Program. Paraho Development Corp., Grand Junction, CO, November 1981.

Anonymous. Perspective on the Emerging Oil Shale Industry. EPA Environmental Assessment, EPA-625/9-81-005, NTIS PB83-171769, 1983.

ARCO Coal Company. Black Thunder Haul Road Supply, 1980.

Ball, B. C. "An investigation into the Potential Economics of Large-Scale Shale Oil Production." American Chemical Society Series. Vol. 1981, No. 163, pp. 195-221, 1981.

Banks, C. E., et al. "Simulated Dewatering Requirements at an Oil Shale Surface Mine. Piceance Creek Basin, Colorado." Mineral Industries Bulletin, Vol. 21, No. 2, 1978.

Barke, W. L., et al. Evaluation of Processes for Liquefaction and Gasification of Solid Fossil Fuels. Volume 2 - Oil Shale and Tar Sands: Mining and Liquid Recovery. Stanford Research Inst. Menlo Park, CA, Project No. ECC-2896, 1975, 267 pp.

Bartok, W., et al., Basic Kinetic Studies and Modeling of NO Formation in Combustion Processes, AIChE Symposium Series 68 (126): 30, 1972.

Bates, E. R., and K. Jakobson. "Status of EPA's Pollution Control Guidance Document For Oil Shale." Fourteenth Oil Shale Symposium, Golden, CO, 1981.

Bates, E. R., and T. L. Thoem. Environmental Perspective on the Emerging Oil Shale Industry. EPA Report No. EPA-600/2-80-205b, NTIS PB 81 186942, 1981.

Bates, E. R., W. W. Liberick, and J. Burckle. Oil Shale: Potential Environmental Impacts and Control Technology. Environmental Research Brief, U.S. EPA Industrial Environmental Research Laboratory, Cincinnati, OH, Report No. EPA-600/D-84-036, NTIS No. PB-84-190743, 1985, 14 pp.

Baughman, G. L., and T. A. Sladek. "Shale Oil Recovery Methods." Mining Engineering, Vol. 33, No. 1,1981, pp. 43-47.

Beychok, M. R. and W. J. Rhodes. Comparison of Environmental Design Aspects of Some Lurgi-Based Synfuels Plants, 6th Symposium on Environmental Aspects of Fuel Conversion Technology, Denver, CO, October 1981.

Bland, V. "Evaluation of Sodium Sorbent Utilization in Flue Gas Desulfurization." KVB Report 72-806930-2020, Interim Report Research Project 1682-2 for Electric Power Research Institute, Palo Alto, CA, 1985. Unpublished draft report.

Booker, J. D. "Oil Shale Retorting in the First Commercial Plants." Proceedings of the 16th Intersociety Energy Conversion Engineering Conference. Technologies for the Transition. ASME Publ., Vol. 1, 1981, pp. 47-51.

Bourcier, D. R., et al. "Preliminary Evaluation of Heavy Metals in Dustfall in the Intermountain West Region." 14th Annual Conference Trace Substances in Environmental Health, Columbia, MO, 1980.

Braun, R. L., et al. Mathematical Modeling of the Cascading-Bed Retorting System and Comparisons of Model Calculations with LLNL Pilot Retort Data, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Bureau of Reclamation, U.S. Department of Interior. Laboratory and Field Studies of Soil Stabilizers, Engineering Research Center, Denver, CO, 1982.

Burnham, A. K. Review and Comparison of Chemistry of In-Situ and Aboveground Processing, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Calvert, S. "How to Choose a Particulate Scrubber. Chemical Engineering. 84(18): pp. 54-68, 1977.

Calvert, S., J. Goldschmid, D. Leith, and D. Metha. Wet Scrubber System Study, Volume I - Scrubber Handbook. EPA-R2-72-118a, NTIS No. PB213016, 1976.

Capital and Operating Costs of Selected Air Pollution Control Systems. Prepared by Gard, Inc. EPA-450/5-80-002, 1978.

Cathedral Bluffs Shale Oil Co. PSD Permit Application. Prevention of Significant Deterioration. (Amendment), 1982.

Cavanaugh, E. C., et al. Atmospheric Pollution Potential from Fossil Fuel Resource Extraction, On-Site Processing, and Transportation. Radian Corporation, Austin, TX, Report No. EPA-600/2-76-064, NTIS No. PB-252649, 1976, 292 pp.

Cena, R. J., General Description for a One Ton/Day Rapid Pyrolysis, Solid Recycle Retort, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Cha, C. Y., and D. Chazin. "A Survey of Current Technologies for Production of Oil from Oil Shale by In-Situ Retorting Processes: Their Technical and Economic Readiness and Requirements for Further Developments." AIChE Symposium Series. Vol. 78, No. 216, 1982, pp. 1-17.

Chappell, W. R., and D. D. Runnells. "Toxic Trace Elements and Oil Shale Production." 11th Annual Conference on Trace Substances in Environmental Health: Abstract, 1977.

Chevron Shale Oil Company. Clear Creek Shale Oil Project. Joint Venture Mine and Retort Facilities. PSD Permit Application. Vol. 1, Environmental Research & Technology, Fort Collins, CO, 1982.

Cieslewicz, W. J. "Selected Topics of Recent Estonian-Russian Oil Shale Research and Development." Colorado School of Mines Quarterly. Vol. 66, No. 1, 1971.

Colorado School of Mines: Proceedings of the First Five Oil Shale Symposia, 1964-68. Vol. 59, No. 3, 1964, 911 pp.

Committee on Synthetic Fuel Safety. Safety Issues Related to Synthetic Fuels Facilities, NTIS PB82-258682, 1982.

Cooperative Inst. for Research in Environmental Science, Boulder, CO. Environmental Chemistry of Oil Shale Development. Interim Technical Progress Report, January 1 - August 1, 1983. Report No. DOE/ER/60121-1, 1983. 11 pp.

Cotter, J. E., et al. Fugitive Dust at the Paraho Oil Shale Demonstration Retort and Mine. EPA-600/7-79-208, NTIS No. PB80-122591,1979.

Cox, C. H., and G. L. Baughman. "Oil Sands: Resource, Recovery, and Industry." Mineral & Energy Resources, Vol. 23, No. 4, 1980, Publ., Colorado School of Mines.

Damon, J. E., et al. "Economics of SCR Post Combustion NO_x Control Processes," In Proceedings of the 1982 Joint Symposium on Stationary Combustion NO_x Control,Vol. II, EPA-600/9-85-022b, NTIS No PB85-235612, 1985.

Daum, K. A., et al. "Analysis of Sulfur Control Strategies for the Oil Shale Industry." J. Air Pollution Control Assoc. Vol. 32, No. 4, 1982. pp. 391-392.

Dennis, R., ed., Handbook on Aerosols. TID-26608. Technical Information Center, Energy Research and Development Administration, Tennessee, 1976.

Denver Research Institute, Pollution Control Technical Mannual: Lurgi Oil Shale Retorting with Open Pit Mining, EPA-600/8-83-005, NTIS PB83-200204, 1983.

Denver Research Institute. Pollution Control Technical Manual: Modified "In-Situ" Oil Shale Retortig Combined with Lurgi Surface Retorting, EPA 600/8-83-004, NTIS PB83-200121, 1983.

Denver Research Institute. Pollution Control Technical Manual: TOSCO 2 Oil Shale Retorting with Underground Mining, EPA-600/8-83-003, NTIS PB83-200212, 1983.

Department of Energy, Office of Technology Impacts, Environmental Control Costs for Oil Shale Processes, Washington, DC, October 1979.

Detailed Emissions Calculations for Oil Shale Mining and Ore Handling and Construction - Related Activities Associated with the Clear Creek Shale Oil Project. Appendix 2, 1982.

Diaz, J. C. and R. L. Braun. Mathematical Modeling of Oil Shale Retorting in a Fluidized-Bed Pyrolyzer and Lift-Pipe Combustor, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Domahidy, G., et al. "Air Pollution control for Oil Shale Applications." Proceedings from the 1983 Eastern Oil Shale Symposium, pp. 271-280.

Duir, J. H., B. A. Christolini, and C. F. Griswold. "Oil Shale Retorting Technology." Chemical Engineering Progress. Vol. 79, No. 2, 1983. pp. 45-50.

Duir, J. H., et al. "Union Oil Shale Retorting Technology." Chemical Engineering Progress. Vol. 6, No. 3, 1982, p. 264.

Dunn, D. W., T. A. Bonner, and S. C. Cheng. Oxides of Nitrogen/Ammonia Control Technology for Oil Shale Retort Emissions (Final Report). Monsanto Research Corp., Dayton, OH, Report No. EPA-600/2-84-078, NTIS No. PB84-171453, 1984, 89 pp.

Dyni, J. R. "Lacustrine Oil Shales and Stratigraphy of Part of the Kishenehn Basin, Northwestern Montana." Mineral & Energy Resources. Vol. 26, No. 4, 1983, Publ., Colorado School of Mines, Golden, CO.

Dziegiel, H. T., et al. The Thermal DeNOx Demonstration Project. In: Proceedings of the 1982 Joint Symposium on Stationary Combustion NO_x Control, Volume II, EPA-600/9-85-022b, NTIS No. PB85-235612, 1985.

Edgar, T. F., et al. Environmental Effects of In-Situ Gasification of Texas Lignite. EPA Report EPA-600/7-81-035, NTIS PB81-171654, 1981.

Edward, M. S. H_2S Removal Process for Low-Btu Coal Gasification. Oak Ridge National Laboratory, ORNL/TM 6077, January 1979.

Environmental Protection Agency. Manual for Methods of Quickly Vegetating Soils of Low Productivity, Construction Activities. Office of Water Programs, Applied Technology Division, Washington, DC., 1975.

Environmental Protection Agency. Iron and Steel Plant Open Dust Source Fugitive Emission Evaluation, EPA-600/2-79-103, NTIS No. PB299385, 1979.

Environmental Protection Agency. Technology Assessment Report for Industrial Boiler Applications: Particulate Collection. EPA-600/7-79-178h, NTIS No. PB80-176365, December 1979.

Environmental Protection Agency. Air Pollution Engineering Manual, Second Edition. AP-40, May 1973.

Environmental Protection Agency. Capital and Operating Costs of Selected Air Pollution Control Systems. EPA-450/5-80-002, December 1978.

Environmental Protection Agency. Development of Emission Factors for Fugitive Dust Sources, EPA-450/3-74-037, NTIS No. PB238262, 1974.

Environmental Protection Agency. Compilation of Air Pollutant Emission Factors, AP-42, 1985.

Environmental Protection Agency. Extended Evaluation of Unpaved Road Dust Suppressants in the Iron and Steel Industry. Prepared by Midwest Research Institute, EPA-600/2-84-027, NTIS No PB84-154350, 1984.

Environmental Protection Agency. Control of Emissions from Lurgi Coal Gasification Plants. EPA-450/2-78-012, NTIS No. PB279012, March 1978.

Evans, R. J., et al. "Development of a Sampling Train for Environmental Assessment of Coal Gasification and Oil Shale Hydro Retorting Processes." 182nd American Chemical Society National Meeting, New York, NY, August 23-28, 1981, Paper No. 182.

Fisher, S. T. "Electrical Induction Heating for the In-Situ Processing of Coal, Oil Shale and Oil Sand." International Electrical, Electronics Conference and Exposition, 1977.

Folsom, B. A., et al., "The Effects of LBG Composition and Combustor Characteristics on Fuel NO_x Formation, "Journal of Engineering for Power, 1980, Vol. 102, p. 459.

Fosdick, G. E., and M. W. Legatski. "Prediction of Air Quality Impacts for a Commercial Shale Oil Complex." AIChE Symposium Series. Vol 73, No. 165, 1975, pp. 281-290.

Fox, J. P., et al. "Mercury Emissions from a Simulated In-Situ Oil Shale Retort." Eleventh Oil Shale Symposium Proceedings. Publ., Colorado School of Mines, 1978, pgs 55-75.

Fruchter, J. S., et al. "Elemental Partitioning in Above Ground Oil Shale Retort Pilot Plant." J. Environmental Science Technology. Vol. 14, No. 11, 1980, pp. 1374-1381.

Frye, K. N. "Status of the Synfuel Program." Trans. American Nucl. Society. Vol. 38, 1981, pp. 33-34.

Fuel and Mineral Resources, Inc., Reston, VA. Critical Review, Comparative Evaluation, Cost Updates, and Baseline Data Development Services in Oil Shale Mining, In-Situ Liquefaction, and Above Ground Retorting Processes from the Environmental, Permitting, and Licensing Viewpoints. Vol. III Emission Source Identification and Source-Specific Pollution Control Applications. Part 1 Report No. DOE/ET/12072-TIS, NTIS No. DE83003775, 1981, 252 pp.

Garcia, L. L., H. F. Schulte, and H. J. Ettinger. "Industrial Hygiene Study at the Anvil Points Oil Shale Facility." American Ind. Hyg. Assoc. J. No. 42, 1981.

Gary, J. H. (Ed). Seventeenth Oil Shale Symposium Proceedings. Publ., Colorado School of Mines, 1984, 440 pp.

Glass, J. W. and J.O.L. Wendt. Mechanisms Governing the Destruction of Nitrogeneous Species During the Fuel-Rich Combustion of Pulverized Coal, 19th Symposium on Combustion, The Combustion Institute, 1982, pp. 1243-1251.

Green, L. "Hydroretorting of Oil Shale with Nuclear Process Heat." ASME Energy Meeting, 1976.

Habenicht, C. H., et al. Steam Stripping and Reverse Osmosis for the Treatment of Oil Shale Wastewaters: Report of a Field Test. DOE Laramie Energy Technology Center, Project No. 5-31343, 1980, 88 pp.

Hang, H. S. Control of NO_x from Coal-Fired Boilers: Combustion Modification Techniques, DOE Report ANL/ECT-3, Appendix D. Argonne National Laboratory, Argonne, IL, 1981.

Hardison, L. C. Minimizing Gas Treating Costs with LO-CAT© for H_2S Removal, Gas Processors Assn. Regional Meeting, Amarillo, TX, October 1984a.

Hardison, L. C. and J. Lekas. Application of the LO-CAT© Hydrogen Sulfide Oxidation Process to Geokinetics In-situ Oil Shale Retorting, Eastern Oil Shale Symposium, Lexington, KY, 1984b.

Hardison, L. C. Catalytic Gas-Sweetening Process Selectivity Converts H_2S to Sulfur, Treats Acid Gas, Oil & Gas Journal, June 1984c.

Harstein, A. M. and B. M. Harneg. "Oil Shale RD and D Program Overview." American Chemical Society, Division Fuel Chem., Prepr. Vol. 25, No. 3, 1980, pp. 1-11.

Hass, R. H., et al. Recent Developments in Selectox and Unisulf Sulfur Recovery Technologies, Sulfur-84 International Conference, Alberta, Canada, June 1984 (Union Oil Co., Brea, CA).

Haug, R. L. "Fluidized Bed Combustion of Sludge Derived Fuel," City of Los Angeles, Bureau of Engineering, 1981.

Hayhurst, A. N. and I. M. Vince. Nitric Oxide Formation from N_2 in Flames: the Importance of "Prompt" NO, Progress in Energy Combustion Science 6, pp. 35-51, 1980.

Henderson, K. B. Investigation of The Geokinetics Horizontal In-Situ Oil Shale Retorting Process. Quarterly Report, January - March 1984. Geokinetics, Inc., Salt Lake City, UT, Report No. DOE/LC/10787-132, NTIS No. DE84016422, 1984, 53 pp.

Henschel, D. B. Environmental Monitoring Reference Manual for Synthetic Fuels Facilities. EPA-600/8-83-027, NTIS No. PB83-251850, 1983.

Hesketh, H.E. Fine Particle Collection Efficiency Related to Pressure Drop, Scrubbant and Particle Properties, and Contact Mechanism. Journ. Air Poll. Control Assoc. 24(10): 939-942, 1974.

Hodgson, A. T., M. J. Pollard, and N. J. Brown. "Mercury Emissions from a Modified In-Situ Shale Retort." J. Atmosphere Environment, Vol. 18, No. 2, 1984, pp. 247-254.

Holmes, J. G. Application for Prevention of Significant Deterioration Permit. Sand Wash Shale Oil Facility. TOSCO Development Corp., Denver, CO, 1981.

Hughes, E. E., et al. Oil Shale Air Pollution Control (Final Report). Stanford Research Inst., Menlo Park, CA, Report No. EPA-600/2-75-009, NTIS No. PB-242858, 1975, 114 pp.

Hunter, S. C. and S. S. Cherry. "NOx Emissions from Petroleum Industry Operations," API Publication 4311, American Petroleum Inst., Washington, D.C., The Combustion Institute, 1979.

Hurst, B. E. "Thermal $DeNO_x$ Technology Update," In: Proceedings: 1985 Symposium on Stationary Combustion NO_x Control, Volume 1, EPA-600/9-86-021a, NTIS No. PB86-225042, 1986.

Hutchins, J. S., W. W. Krech, and M. W. Legatski. "Environmental Aspects of a Commercial Oil Shale Operation." AIME Environmental Qualification Conference, Washington, D. C., 1971, Paper No. EQC 25.

Hydrocarbon Processing, International Edition, Vol. 63, No. 4, Gulf Publishing Co., Houston, TX, April 1984.

International Science & Technology Inst., Washington, D.C. Assessment of Oil Shale Technology in Brazil. Final Technical Report, October 27, 1980 - July 27, 1981. Report No. DOE/ER/30010-01, NTIS No. DE81027574, 1981, 143 pp.

Jackson, R. M. "The Union Oil Company of California. Parachute Creek Oil Shale Program." 30th Annual Petroleum and Chemical Industry Conference, Denver, CO, 1983.

Jones, J. B., et al. "Paraho Oil Shale Retorting Processes." Handbook of Synfuels Technology. Publ., McGraw-Hill, New York, NY, 1984, pp. 4.63-4.93.

Kane, G. C., and R. J. Long. "Western Oil Shale Development." Trans. American Nucl. Society. Vol. 38, No. 37, 1981.

Kerry, H. A. and A. Weir, Jr. "Operating Experience on So. California Edison's 197.5 MW Selective Catalytic Reduction $DeNO_x$ System." In: Proceedings: 1985 Symposium on Stationary Combustion NO_x Control, EPA-600/9-86-021a, NTIS PB86-225042, 1986.

Kirkpatrick, L. W. "Air Pollution Aspects of Proposed Oil Shale Development in Northwestern Colorado." Quarterly of the Colorado School of Mines. Vol. 69, No. 2, 1974, pp. 103-108.

Kleive, J. R., G. D. Rawlings, and J. R. Hoeflein. Assessment of Oil Shale Retort Wastewater Treatment and Control Technology: Phases I and II. EPA 600/7-81-081, NTIS PB 81-187288, 1981.

Kronenburger, L. "Control of Air Pollutant Emissions from Oil Shale Facilities." Oil Shale: The Environmental Challenges, III Colorado School of Mines Press, Golden, CO, 1983.

Knudsen, C. W., and J. B. Dicks. "Synthetic Fuels: Coal and Oil Shale Readiness." Proceedings ASME Winter Annual Meet. Technical Economics, Synfuels and Coal Energy Symposium, 1983, 1984, pp. 139-145.

KVB, Inc., "Environmental Control Implications of Generating Electric Power from Coal, 1977, Technology Status Report, Appendix D, Assessment of NO_x Control Technology for Coal-Fired Utility Boilers," DOE Report ANL/ECT-13, Argonne National Laboratory, Argonne, IL.

KVB, Inc., "Baghouse Injection of Sodium Compounds for SO_2 Control in a Full-Scale Demonstration," Vol. I, Technical Proposal, KVB8672-02594, 1985.

Lappi, R. L., et al. Overview of Environmental Impacts of Large-Scale Surface Mining of Oil Shale: Piceance Basin, Colorado. EPA Environmental Research Brief, No. EPA-600/D-82-283, 1982.

Lawrence Livermore National Laboratory, 5th Briefing on Oil Shale Research, Livermore, CA, October 1985.

Link, J. M., et al. Hydraulic Transportation of Oil Shale (Final Report). Colorado School of Mines Research Institute, 1977.

Loper, D. R. "Chevron Program in Shale Oil Development." 1981 Winter Meeting of the American Nuclear Society, 1981.

Lovell, R. J., S. W. Dylewski, and C. A. Peterson. Control of Sulfur Emissions from Oil-Shale Retorts. IT-Enviroscience, Knoxville, TN, EPA 600/7-82-016, NTIS No. PB82-231945, 1982, 192 pp.

Magee, E. M. Evaluation of Pollution Control in Fossil Fuel Conversion Processes. Exxon Research & Engineering Company, Linden, N. J. EPA-600/2-76/101, NTIS No. PB255842, 1976, 306 pp.

Magee Circle Energy Corp., Utah Cottonwood Washington PSD (Permit Application), Oklahoma City, OK, April 1982.

Mallon, R. G., The Design, Construction and Operation of a Fluidized Bed Retort Using Solid Recirculation, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Marshall, W. F., et al., Effects of Catalytic Reactors on Diesel Exhaust Composition, Bartlesville Research Center, Dept. of Energy, Bartlesville, OK, 1978.

McLendon, T. R., and R. A. Martel. "Combustion Retorting of Oil Shale at Low Void Volumes." Proceedings of the 18th Intersociety Energy Conversion Engineering Conference. Vol. 2, 1983, pp. 502-507.

Merrow, E. W. Review of Cost Estimation in New Technologies: Implications for Energy Process Plants. Rand Corp. Report, NTIS No. RAND/R/2481/DOE, 1979, 128 pp.

Meyer, E. G. Oil Shale Energy Source Development. Final Report. Report No. DOE/LC/10758-T1, NTIS No. DE83006602, 1983, 21 pp.

Midwest Research Institute. Iron and Steel Plant Open Source Fugitive Emission Control Evaluation. Prepared for U.S. Environmental Protection Agency, EPA-600/2-83-110, NTIS No. PB84-110568, 1983.

Mobile-Pacific Oil Shale Environmental Impact Statement. Vol. 1-Text. (Final Report), Bureau of Land Management, U.S. Department of Interior, 1984.

Muzio, L. J. and T. W. Sonnichsen. Dry SO_2 Particulate Removal for Coal-Fired Boilers, Vol. 2:22-MW Demonstration Using Nahcolite, Trona and Soda Ash, CS-2894, Vol. 2, Electric Power Research Inst., Research Project 1682-2, Final Report, June 1984.

National Research Council, Washington, D. C. Review of the 1982 Department of Energy Health and Environmental Effects Assessments on Coal Liquefaction and Oil Shale Technologies. DOE Contract No. DE-AC01-82ER60066, NTIS No. PB84-194422, 1983, 72 pp.

Office of Air Quality Planning and Standards. EPA, Research Triangle Park. Source Category Survey: Oil Shale Industry Phase 1. Report No. EPA-450/3-81-010, NTIS No. PB82-120114, 1981, 105 pp.

Office of Technology Impacts. Department of Energy, Washington, D. C. Environmental Control Costs for Oil Shale Processes. Report No. DOE/EV-0055, 1979, 601 pp.

Oh, M.S., F. D. Fujimoto, and R. W. Taylor. Ammonia Release from Shale Combustion and Fischer Assay Retorting, 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, October 1985.

Oil Shale 1982: A Technology and Policy Primer. Colorado Energy Research Institute, Colorado School of Mines, 1981.

Ondich, G.G. "Control Technology Appendices for Pollution Control Technical Manuals," EPA-600/8-83-009, NTIS PB83-214734, 1983.

Ondov, J. M. Physical and Chemical Characteristics of Fine Particles in Off-Gases from Coal Combustion and Oil Shale Retorting. Lawrence Livermore National Laboratory, Livermore, CA, NTIS PB-DE81026540, 1981.

Paraho Development Corp., PSD Permit Application - Paraho UTE Facility, Grand Junction, CO, November 1981.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module. Critical Items Identification, Task 11, Report No. DOE/ET/14103-T2, NTIS No. PBDE82003727, 1981, 39 pp.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module: Environmental Compliance Plant, Topical Report, Task 6. Report No. DOE/ET/14108/T10, NTIS No. PBDE84016078, 1981, 81 pp.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module: Environmental, Safety and Health Plan. Topical Report, Task 8, Park 1, Design and Safety Program. Report No. DOE/ET/14103-T19, NTIS No. PBDE84015998, 1981, 279 pp.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module, Environmental, Safety and Health Requirements. Task 5. Report No. DOE/ET/14103-T6, NTIS No. PBDE84016090, 1981, 107 pp.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module: NEPA Compliance. Topical Report, Task 7, Part 1. Report No. DOE/ET/14103-T14, NTIS No. PBDE84016088, 1981, 314 pp.

Paraho Development Corp., Grand Junction, CO. Paraho Oil Shale Module: Trade Off Studies, Topical Report, Task 12. Report No. DOE/ET/14103-T9, NTIS No. PBDE84016076, 1981, 108 pp.

PEDCo Environmental, Inc. Improved Emission Factors for Fugitive Dust from Western Surface Coal Mining Sources. EPA-600/7-84-048, NTIS No. PB84-170802, 1984.

PEDCo Environmental, Inc. Cost Effectiveness of Dust Controls Used on Unpaved Mine Roads. Prepared for Bureau of Mines, U.S. Department of the Interior.

PEDCo Environmental, Inc. Fugitive Dust Control Techniques at Hazardous Waste Site. Field Sampling Report No. 1. Sampling Results for Exposed Area Testing. Prepared for U.S. Environmental Protection Agency, Solid and Hazardous Waste Research Division, Cincinnati, OH, 1984a.

PEDCo Environmental, Inc. Fugitive Dust Control Techniques at Hazardous Waste Site. Field Sampling Report No. 2. Sampling Results for Exposed Area Testing. Prepared for U.S. Environmental Protection Agency, Solid and Hazardous Waste Research Division, Cincinnati, OH, 1984b.

PEDCo Environmental, Inc. Fugitive Dust Control Techniques at Hazardous Waste Site. Field Sampling Report No. 3. Sampling Results for Exposed Area Testing. Prepared for U.S. Environmental Protection Agency, Solid and Hazardous Waste Research Division, Cincinnati, OH, 1984c.

PEI Associates, Inc. Handbook - Dust Control at Hazardous Wates Sites, EPA-540/2-85-003, 1985.

Perry, R. H. and D. W. Green. Perry's Chemical Engineers' Handbook, 6th ed., McGaw-Hill Book Co., San Francisco, 1984.

Pershing, D. W. and J.O.L. Wendt. "Pollutant Control Through Staged Combustion of Pulverized Coal: Phase I," CONPENNSV Report FE-1817-2, 1976(a).

Pershing, D. W. and J.O.L. Wendt. "Pulverized Coal Combustion: The Influence of Flame Temperature and Coal Combustion on Thermal and Fuel NO_x," 16th Symposium (International) on Combustion, 1976(b). Pub. Combustion Institute, Pittsburgh, PA.

Pershing, D. W. and J.O.L. Wendt. "Relative Contributions of Volatile Nitrogen and Char Nitrogen to NO_x from Pulverized Coal Flames," 83rd National AIChE Meeting, 1977.

Peters, J.M. Predicting Efficiency of Fine-Particle Collectors. Chemical Engineering, Volume 80, No. 9, 99-102, 1973.

Peterson, E. J., et al. Approach to Oil Shale Health and Environmental Research: A Concept Paper. Los Alamos National Lab., N.M., Report No. DOE/LA-9805, NTIS No. DE8400640p, 1983, 27 pp.

Peterson, K. K. (Ed) Oil Shale--The Environmental Challenges. Proceedings of Int'l Symposium, August 11-14, 1980, Vail, CO, Publ., Colorado School of Mines, Golden, CO, 1980, 380 pp.

Peterson, K. K. (Ed) Oil Shale--The Environmental Challenges II. Proceedings of Int'l Symposium, August 10-13, 1981, Vail, CO, Publ., Colorado School of Mines, 1981, 392 pp.

Peterson, K. K. (Ed) Oil Shale--The Environmental Challenges III. Proceedings of Int'l Symposium, August 9-12, 1982, Vail, CO, Publ., Colorado School of Mines, Golden, CO, 1982, 261 pp.

Pforzheimer, H. "Oil Shale Up-Date." American Chemical Society, Division Pet. Chem., Prepr, Vol. 23, No. 4, 1978, 1360 pp.

Phillips, T. T. Aspen Simulation of Oil Shale Retort Off-Gas Cleanup with Venturi Scrubbers. Los Alamos National Laboratory, N.M., Report No. DOE/LA-10044-MS, NTIS No. PBDE84011018, 1984, 15 pp.

Pohl, J. H. and A. F. Sarofim. Devolatilization and Oxidation of Coal Nitrogen, Proc. of the 16th Symp. on Combustion pp. 475-489, 1977.

Prahacs, S., Control of Kraft Recovery Furnace Emissions with Simultaneous Heat Recovery Using the PPRRIC/BERC Scrubbing Process, Pulp & Paper Research Institute of Canada, Report MR37, Pointe Claire, P.Q., July 1983.

Pritchard, J. F. Co. Data obtained through J. Geick, Black & Veatch, Consulting Engieers, Kansas City, MO, 1980.

Private communication from K.E. Buttke, Southport Equipment Corporation to Bob Bakshi of TRW. January 26, 1982.

Quintana Minerals Corp., PSD Permit Application for a Commercial Shale Oil Facility, Syntana, UT, undated (approx. 1982).

Rammler, R. W., et al. "Lurgi-Ruhrgas Process." Handbook of Synfuels Technology. Publ., McGraw-Hill, New York, NY, 1984, pgs 4.5-4.44.

"Research on Petroleum, Natural Gas, and Oil Shale: The University's Role" Proceedings of the NSF-RANN/ERDA Workshop at Colorado School of Mines, 1975.

Rex, R. C., and P. A. Lynch. Environmental Analysis of the Eastern Shale Hydroretorting Data Base. Hycrude Corp., Chicago, IL, Report No. DOE/LC/11039-1617, NTIS No. PBDE84009255, 1984, 274 pp.

Roeck, D. R. Combustion of Oil Shale in Fluidized-Bed Combustors - An Overview. (Final Report). EPA-600/7-82-045, NTIS No. PB82-249889, 1982.

Rosbury, K. D., and R. A. Zimmer. Cost-Effectiveness of Dust Controls Used on Unpaved Haul Roads. Vol. I - Results, Analysis, and Conclusions. U.S. Bureau of Mines, Minneapolis, MN, 1983.

Rosbury, K. D., and R. A. Zimmer. Cost-Effectiveness of Dust Controls Used on Unpaved Haul Roads. Vol. II - Data. U.S. Bureau of Mines, Minneapolis, MN, 1983.

Schirmer, R. M., and D. A. Nickerson. "Low-NO_x Burner for High-Nitrogen Fuels." Colorado School of Mines, Oil Shale 14th Symposium, Golden, CO, 1981.

Schmidt, R. A. Oil Shale and Tar Sands Processing into Fuels for Power Generation: A Critical Assessment. Electric Power Research Inst., Palo Alto, CA, 1976, 45 pp.

Severson, S.D., F.A. Horney, D.S. Ensor, and G.R. Markowski. Economic Evaluation of Fabric Filtration Versus Electrostatic Precipitation for Ultrahigh Particulate Collection Efficiency. FP-775, 1978.

Shih, C. C., and J. E. Cotter. Technological Overview Reports for Eight Oil Shale Recovery Processes. EPA-600/7-79-075, NTIS PB295665, 1979.

Sickles, J. C., et al. "Investigation of the Atmospheric Chemistry of Emissions from Fuel Conversion Facilities." 71st APCA Annual Meeting and Exhibition, Houston, TX, 1978.

Sievers, R. E. Environmental Chemistry of Oil Shale Development. Technical Progress Report, January 1 - November 30, 1983. Cooperative Inst. for Research in Environmental Science, Bolder, CO, Report No. DOE/ER/60121-T1, NTIS No. PBDE84004081, 1983, 75 pp.

Sitnai, O. "Process Concept of Retorting of Julia Creek Oil Shale." Energy Progress. Vol. 4, No. 2, 1984, pp. 100-105.

Sklarew, D. S., et al. "Trace Nitrogen-Containing Species in the Offgas from 2 Oil Shale Retorting Pocesses." Environ. Science Technology. Vol. 18, No. 8, 1984, pp. 592-600.

Sklarew, D. S., and D. J. Hayes. "Trace Sulfur-Containing Species in the Offgas from 2 Oil Shale Retorting Processes." Environ. Science Technology. Vol. 18, No. 8, 1984, pp. 600-603.

Sladek, T. A. "Recent Trends in Oil Shale, Part 1: History, Nature, and Reserves." Mineral Industries Bulletin. Vol. Q7, No. 6, 1974, Publ., Colorado School of Mines.

Sladek, T. A. "Recent Trends in Oil Shale, Part 3: Shale Oil Refining and Some Oil Shale Problems." Mineral Industries Bulletin. Vol. 18, No. 2, 1975, Publ., Colorado School of Mines.

Smith, J. W. "Oil Shale Resources of the United States." Mineral & Energy Resources, Vol. 23, No. 6, 1980, Publ., Colorado School of Mines.

Smith, J. W. "We Need Oil From Oil Shale." Mineral & Energy Resources. Vol. 27, No. 5, 1984, Publ., Colorado School of Mines.

Sohn, H. Y. "Oil Shale Retorting Processes." Health Implications of New Energy. Ann Arbor Science Publ., Ann Arbor, MI, 1980, pp. 415-426.

Stevens, A. L., et al. "Occidental Modified In-Situ (MIS) Process." Handbook of Synfuels Technology. Publ., McGraw-Hill, New York, NY, 1984, pp. 4.95-4.119.

Sullivan, R. F. "Refining Shale Oil." American Petroleum 1978 Proceedings. Vol. 57, 1978, pgs 199-210.

Sung, R. D., and C. H. Prien. "Evaluation of the Pollution Abatement Technologies Available for Treatment of Wastewater from Oil Shale Processing." AIChE Symposium Series. Vol. 73, No. 166, 1977, pp. 323-332.

"Synthetic Liquid Fuels from Oil Shale, Tar Sands, and Coal: Proceedings of the Hydrocarbons Symposium." Colorado School of Mines Quarterly. Vol. 65, No. 4, 1970.

Taback, H. J., et al., "Alkaline and Stretford Scrubbing Tests for H_2S Removal from In-situ Oil Shale Retort Offgas," EPA-600/7-85-029, NTIS PB 85 246965, 1985a.

Taback, H. J., et al., "Alkaline Scrubbing of In-situ Oil Shale Retort Offgas at Geokinetics," 18th Oil Shale Symposium Proceedings, Colorado School of Mines Press, Golden, CO, 1985b.

Taback, H. J., et al., "Processing In-situ Oil Shale Retort Offgas with a Stretford Plant at Geokinetics," 18th Oil Shale Symposium Proceedings, Colorado School of Mines Press, Golden, CO, 1985c.

Taback, H. J., et al., "The Effect of Oil Shale Recovery Processes on Air Emissions," 19th Oil Shale Symposium Proceedings, Colorado School of Mines, Golden, CO, 1986.

Taylor, R. W., et al. "Nitric Oxide (NO) Emissions from Combustion of Retorted Oil Shale," 18th Oil Shale Symposium Proceedings, Colorado School of Mines, 1985a.

Taylor, R. W., et al. "Nitric Oxide (NO) from Combustion of Retorted Shale," Shale Oil, 5th Briefing on Oil Shale Research, Lawrence Livermore Laboratory, CA, 1985b.

Teller, A. J. and H. R. Arnberg. "Considerations in the Design for TRS & Particulate Recovery from Effluents of Kraft Recovery Furnaces," a paper received from A. J. Teller, Teller Environmental Systems, Inc., Shrewsbury, MA, 1985a.

Teller, A. J. "Hot Water Production from Recovery Boiler Gases," a paper received from Teller Environmental Systems, Shrewsbury, MA, May 1985b.

Teller, A. J., "Simultaneous Acid Gas and Particulate Recovery," In: Symposium on the Transfer and Utilization of Particulate Control Technology, Volume 2. EPA-600/7-79-044b, NTIS PB 295227, 1979.

Teller Environmental Systems, Inc. Dry Venturi (a schematic concept drawing), Shrewsbury, MA, 1986.

Teller, A. J., Teller Environmental Systems, Inc. Incineration - Resource Recovery Flue Gas Emission Control, presented at the Acid Gas and Dioxin Control for Waste to Energy Facility Symposium, Washington, DC, November 1985c.

Teller, A. J., Teller Environmental Systems, Private Communication, 1985d.

Thompson, R. E., et al. "Retrofit NO_x Control Guidelines for Coal-Fired Utility Boilers," Research Project 2154-1, Final Report for the Electric Power Research Institute, Palo Alto, CA, 1985.

Tomany, J.P. Air Pollution: the Emissions, the Regulations, and the Controls. American Elsevier Publishing Co., Inc., 1975.

Tosco Development Corp., PSD Permit Application, Sand Wash Shale Oil Facility, June 1981.

Travis, C. C., et al. Exposure Assessment Methodology and Reference Environments for Synfuel Risk Analysis. Oak Ridge National Laboratory, TN, Report No. DOE ORNL/TM-8672, NTIS No. PBDE 84004128, 1983, 89 pp.

TRC Environmental Consultants. Coal Mining Emission Factor Development and Modeling Study, 1981a.

TRW Energy Systems Group, Oil Shale Data Book, prepared for U.S. Department of Energy, NTIS PB 80-125636, June 1979.

Union Oil Co. of California, Union Energy Mining Division: PSD Permit Application for Union's Phase II Oil Shale Mining and Retort Facility, 1982.

Union Oil Co. of California, Union Energy Mining Division: PSD Permit Application for Union's Phase II Shale Oil Upgrading Plant Facility, 1982.

University of Wyoming Research Corp., Laramie, WY. Oil Shale Tar Sands, and Underground Coal Gasification. Quarterly Progress Report, July - September 1983. Report No. DOE/FE/60177-1534, NTIS No. DE84008069, 1983, 92 pp.

U.S. Army Engineer Waterways Experiment Station. Materials Evaluated as Potential Soil Stabilizers. Vicksburg, MS, 1977.

U.S. Army Corps of Engineer, Sacramento District. Getty and Cities Service Shale Oil Projects. Environmental Impact Statement (Final Report), 1984.

Van Zanten, K. D., et al. "Control of Sulfur Emissions from Oil Shale Retorting Using Spent Shale Absorption," 19th Oil Shale Symposium Proceedings, Colorado School of Mines Press, Golden, CO, 1986.

Vatavik, W.M., and R. B. Neveril. Part X: Estimating Size and Cost of Venturi Scrubbers. Chemical Engineering, 1981, pp 93-96.

Waples, D. W. "C/N Ratios in Source Rock Studies." Mineral Industries Bulletin. Vol. 20, No. 5, 1977, Colorado School of Mines, Golden, CO.

Weber, J. H. "Oil Shale Processing Technology Above Ground and Below Ground, Sampling Strategy and Characterization of Potential Emissions from Synfuel Production." Symposium on Sampling Strategy and Characterization of Potential Emissions from Synfuel Production, Austin, TX, June 6, 1974.

Weiss, H. "Retorting of Oil-Shale - Background, Status and Potential of the Lurgi Ruhrgas (LR) Process." American Chemical Society Abstracts. Vol. 183, March 1982, p. 26.

Wendt, J.O.L. "Fundamental Coal Combustion Mechanisms and Pollutant Formation in Furnaces", Prog. Energy Combust. Sci., Vol. 6, pp. 201-222, 1980.

White, J.J. Role of Electrostatic Precipitators in Particle Control: A Retrospective and Prospective View. Jour. Air Poll. Control Assoc. Vol. 25, No. 2., 1975.

White River Shale Project. Prevention of Significant Deterioration of Air Quality - Oil Shale Tracts Ua and Ub. Bechtel Petroleum, Inc., San Francisco, CA, Job #14188, August 1981.

Wong, C. M., R. W. Crawford, A. K. Burnham, and P.E. Miller. From Bench-Top to Pilot Scale: Determination of Sulfur Gases by Triple-Quadripole Mass Spectrometry (TQMS), 5th Briefing on Oil Shale Research, Lawrence Livermore National Laboratory, 1985.

APPENDIX A

TRACE ELEMENTS

Western oil shale from the Green River Formation is a fine-grained, sedimentary rock that contains about 20 wt% organic material in a mineral matrix composed primarily of dolomite and calcite. Also present in trace quantities are compounds of certain metals considered to be hazardous to human health. These may be in the form of sulfides, as substitution products in major mineral phases, or as organometallic compounds (Desborough et al. 1976; Fox et al. 1982; Fish 1983). These trace metallic compounds are mobilized and released to the environment when oil shale is processed.

These hazardous compounds are released as dust, in gaseous form or as contaminants in product or waste streams. Fugitive dust and particulates, largely raw and retorted shale fines and suspended soil material, are produced by mining, crushing, screening, haulage, retorted shale disposal and other material handling operations.

Trace elements present in the raw shale are redistributed during retorting among the oil, retort gases, process water and retorted shale. The low to medium Btu gas produced during retorting may be subsequently used for power generation or as a heat source. Depending upon an element's mineral residence, the retort operating conditions and subsequent offgas handling, some of the trace metals may be emitted from the retort plant as gases and particulates. Trace metals may also be released during the refining of the crude shale oil. However, such emissions are expected to be similar to those from conventional refineries. This source has not been specifically studied for shale oil refineries and is thus not discussed here.

Table A-1 summarizes and compares emission factors and rates for those operations that are unique to the oil shale industry, namely, retorting and material handling. The database and assumptions explicit in these estimates are discussed at greater length in subsequent sections. These same trace elements may also be emitted from other conventional emission sources at an oil shale facility, including industrial oil- and coal-fired boilers and diesel engines; however, emissions from those sources are expected to be low

TABLE A-1. SUMMARY OF TRACE ELEMENT EMISSION FACTORS AND EMISSION RATES FOR RETORTING AND SOLIDS HANDLING

	Emission Factors		Emission Rate Per 10,000 BPD Capacity		
	Retorting(1) (mg/bbl)	Materials Handling(2) (mg/bbl)	Retorting(1) (ton/yr)	Materials Handling(2) (ton/yr)	Total (ton/yr)
As	>13-770	1.8	>0.05-3.1	0.01	>0.06-3.1
B	--	2.6	--	0.01	0.01
Ba	<24-<292	18	<0.1-<1.2	0.1	0.1
Be	--	0.090	--	0.004	>0.004
Br	>33	0.16	>0.1	0.001	>0.1
Cd	0-1070	0.029	0-4.3	0.0001	0-4.3
Cl	1300	6.9	5.2	0.03	5.2
Co	<0.3-<3	0.33	<0.001-<0.04	0.001	0.001
Cr	<6-151	1.2	<0.02-0.6	0.005	0.005-0.6
Cu	<8	1.2	<0.03	0.005	0.005
F	--	49	--	>0.2	>0.2
Fe	<1040-<4000	720	<4.2-<16	2.9	2.9
Hg	39-208	0.0032	0.2-0.8	0.0001	0.2-0.8
Mo	<1-6	0.90	<0.004-0.02	0.004	0.004-0.02
Mn	<6-<60	11	<0.02-<0.2	0.04	0.04
Ni	<5	0.79	<0.02	0.0032	0.0032
Pb	<31	0.79	<0.1	0.0032	0.0032
Sb	<1	0.071	<0.004	0.0003	0.0003
Se	>15-940	0.082	>0.06-1.8	0.0003	0.06-1.8
Th	<1	0.21	<0.004	0.0008	0.0008
U	<0.3-<2	0.15	<0.001-<0.008	0.0006	0.0006
V	<18	4.0	<0.07	0.016	0.016
Zn	<12-<17	2.7	<0.05-<0.07	0.01	0.01

(1) Assumes no trace elements are removed in criteria-pollutant control technology and that no special technology is installed to reduce trace elements.

(2) Assumes particulate and fugitive emissions are controlled as specified in company permit applications.

compared to retorting and material handling operations. They would be similar to emissions reported for these elements in other industries. Therefore, these conventional sources are not covered here, and the reader is referred to the following references: for boilers (Krishnan and Hellwig 1982 and Littlejohn 1984); for diesel emissions (NRC 1981).

Table A-1 shows that retorting is the principal source of trace element emissions, particularly of the volatile elements As, Br, Cd, Cl, Hg and Se. Although no data have been reported for F, it is anticipated that it would be emitted during retorting because it is chemically similar to the other halides that have been reported in retort gases (Br, Cl and I). Material handling, which releases about 30 g/bbl of particles, contributes greater than 1 mg/bbl of As, Cl and F, as well as the nonvolatile elements B, Ba, Cr, Cu, Fe, Mn, V and Zn.

Of those elements that are emitted in significant quantities, only Pb, Be, F and Hg are covered in the PSD permit applications for various projects. Table A-2 compares PSD emissions with emissions calculated in this study.

The most striking discrepancies between the PSD permit applications and this study occur for Hg. The PSD permit applications assumed that little or no Hg would be emitted from the retort, typically arguing that Hg is released in the elemental form and therefore would condense out or be otherwise removed in the retort system. However, some studies reviewed here indicate that the majority of the retort Hg is present as organomercurials that are not expected to condense or be otherwise lost. Therefore, it is possible that PSD applications, as presently filed, greatly underestimate Hg emissions.

This section reviews available literature on hazardous, noncriteria trace elements released to the atmosphere during the production of crude oil from Green River oil shale. Emission factors are estimated and compared with those used by industry in PSD applications.

TABLE A-2. COMPARISON OF PERMIT APPLICATIONS ESTIMATES OF Be, F, Hg AND Pb EMISSIONS WITH THOSE DETERMINED IN THIS STUDY (tons/yr)

	Be		F		Hg		Pb	
	PSD (tons/yr)	This Study (tons/yr)	PSD (tons/yr)	This Study (tons/yr)	PSD (tons/yr)	This Study (tons/yr)	PSD (tons/yr)	This Study (tons/yr)
Cathedral Bluffs	0	>0.005	7.8	>0.2	0.003	0.2-1	0.15	0.004
Clear Creek	(2)	>0.04	(2)	>2	(2)	2-8	(2)	0.03
Cottonwood	0.001	>0.02	0.93	>1	0.05	1-5	(2)	0.02
Paraho	(1)	>0.02	2.9	>1	(1)	1-3	(2)	0.01
Syntan	(1)	>0.02	1.45	>1	(1)	1-4	0.03	0.01
Union B	0.0024	>0.04	1.45	>2	(1)	2-7	(2)	0.03
White River	(1)	>0.04	1.8	>2	(1)	2-8	0.46	0.03

(1) Negligible

(2) Not reported

RETORTING

The principal source of trace element emissions at an oil shale facility is the retort itself. During retorting, raw shale is pyrolyzed to release organic vapors, and retorted shale may be combusted to recover energy. The high processing temperatures, from 500 to 700 C in surface retorts and up to 1200 C in in-situ retorts, can mobilize some of the trace elements in the raw shale and redistribute them among gases, oil, process water and retorted shale.

The majority of the trace elements are present in the gaseous state in unprocessed retort gases, which contain very low particulate concentrations. These gases would only reach the environment as a result of leaks or during upset conditions. However, most developers will generate power from the large quantities of low Btu gas produced during retorting. Combustion of these gases to produce power or in a flare to control emergency gas purges can generate significant quantities of respirable particulates that are highly enriched in volatile trace elements, including As, Br, Hg and Se.

The quantity of trace elements volatilized during retorting has been studied by making direct, gas phase measurements and by mass balances in which gas phase emissions are calculated by difference. These studies have revealed that trace element emissions during retorting are primarily controlled by retorting temperature, retort configuration, and operation of the raw shale bed. Raw shale mineralogy, heating rate and/or retorting atmosphere may also affect trace element partitioning.

This section summarizes what is presently known (August 1985) about trace element emissions from oil shale retorts. Emphasis is placed upon Hg, Cd, As and Se because they are volatile in oil shale retorts and may be present in the untreated gases at relatively high concentrations. Available data for other less volatile elements are also summarized.

The vast majority of the trace element data reviewed here was developed using small laboratory retorts or pilot scale demonstration facilities in which gases were not processed and treated as presently planned

for commercial facilities. Therefore, the applicability of this information to commercial oil shale plants is highly uncertain.

Retort gases at a commercial plant will be processed and treated to remove particulates, SO_x, NO_x and hydrocarbons. In most proposed commercial processes, the gases are compressed, heavier hydrocarbons are removed (the shale oil) and treated gases are mixed with natural gas and burned. Therefore, the information presented here is typically for uncontrolled emissions unless otherwise noted.

Ranges are reported for emission factors and are presumed to apply equally to all types of processes, unless specifically excepted.

Mercury

Because Hg and its compounds are the most volatile among the trace elements that occur in oil shale, many studies have been conducted on its distribution during retorting. These studies have demonstrated that Hg is volatilized from the shale between 160 and 380 C, well in advance of shale pyrolysis and oil evolution, and is redistributed among the retorted shale, shale oil, water and gases. From 23 percent up to 100 percent of the Hg originally present in the raw shale is partitioned into the gases, depending upon raw shale mineralogy and retort operation.

Mercury concentrations of from less than 0.2 up to 8200 $\mu g/m^3$ have been reported in retort gases, and emission rates of up to 60 $\mu g/min$ have been measured at field in-situ retorts. Emission factors of from 39 to 208 mg/bbl have been determined for five separate retorts based on continuous offgas measurements.

In uncombusted gases, the majority of this Hg is present as organomercurials, principally dimethylmercury, while in combusted gases, elemental Hg predominates. Particulate Hg concentrations in uncombusted gases are low and typically are 1-2 percent of the total Hg, while in combusted gases, they account for over 30 percent of the total Hg. Reported concentrations of Hg and its compounds in shale gases typically exceed OSHA

and American Congress of Governmental and Industrial Hygienists (ACGIH) standards for workroom air, and thus may be environmentally significant.

Mass Balance Studies. Most investigations of the distribution of Hg in oil shale retorts have used mass balances in which the amount of Hg present in retort inputs (raw shale) are compared with those in retort outputs (retorted shale, water, oil, gas). In fact, most early evidence for the emission of Hg from retorts came indirectly from such studies by measuring all outputs except the gas stream and then calculating the gas phase by difference. Subsequent studies in which Hg in the gas stream was directly measured substantiated these earlier estimates.

Average Hg mass balances for four types of oil shale retorts are summarized in Table A-3 as a percentage of raw shale Hg. The imbalance is the percentage of raw shale Hg that is not recovered in the measured products and thus includes experimental error. The gas phase is rarely measured in such studies, and it is typically assumed to be equal to the imbalance. This approach has been validated for Hg in several studies (Fox 1985a; Hodgson et al. 1982) by making continuous gas phase measurements.

These studies (Table A-3) demonstrate that the majority of the Hg is mobilized from the raw shale and redistributed among the products. A small amount, from 1 percent to about 10 percent, is nonvolatile and is retained in the retorted shale at temperatures up to 1000 C. Although higher percentages, from 18 to 33 percent, have been reported in the retorted shale, these are due to incomplete retorting and/or deposition of volatilized Hg on cool raw or retorted shale (Fox et al. 1977; Hodgson et al. 1982; Fruchter et al. 1980), as discussed later. The majority of the Hg, from 70 percent up to 100 percent, is redistributed to the gases, although some may be subsequently retained on cool retorted shale (e.g., Paraho). Lesser quantities are partitioned to the oil and water.

Hodson et al. (1982) and Olsen et al. (1985) used a 5.5 kg, externally heated retort to study the effect of shale temperature, heating rate and shale type on Hg partitioning. A typical Hg emission profile from these studies is shown in Figure A-1. They found that Hg evolution in western shale starts at 160 C, peaks at 190 to 240 C and terminates at 250 to 320 C, well in advance

TABLE A-3. MERCURY MASS DISTRIBUTION PATTERNS FOR SOME OIL SHALE RETORTS(1)
(Percent of Total Hg in Raw Shale)

	LETC 20-kg Simulated In-Situ Retort (CSR)			LLL Simulated In-Situ Retort		Paraho Direct	Small Lab. Retorts		
							LBL	BPNL	Fisher Assay
	4 Runs T=540 C (2)	7 Runs T=540 C (3)	Run CS-69 T=760 C (4)	6 Runs T=868-1025 (3)	Runs 5-11 T=1003 C (5)	1 Run T=500 C (6)	2 Runs T=500 C (7)	1 Run T=500 C (8)	1 Run T=500 C (9)
Retorted shale	29.4	18.0	12.5	1.0	7.0	32.8	1.8	5.0	1.9
Shale oil	<7.0	<7.1	8.3	9.2	6.0	28.2	5.8	0.4	24.9
Water	<0.7	1.6	2.5	1.0	0.9	0.01	0.1	3.0	15.5
Gas	--	--	72.2	--	--	23.0	92.2	108.0	--
Total	29.4	26.7	95.5	11.2	13.9	84.0	99.9	116.4	42.3
Imbalance(10)	-70.4	-73.3	-4.5	-88.8	-86.1	-16.0	-0.1	+16.4	-57.7

(1) Abbreviations used in this table are:

LETC = Laramie Energy Technology Center; T = temperature; LLL = Lawrence Livermore Laboratory; LBL = Lawrence Berkeley Laboratory; BPNL = Battelle Pacific Northwest Laboratory.

(2) Fox et. al. (1977).

(3) Fox (1980).

(4) Fox (1985).

(5) Fruchter et. al. (1978).

(6) Fruchter et. al. (1980).

(7) Hodgson et. al. (1982)

(8) Olsen et. al. (1985)

(9) Donnell and Shaw (1980)

(10) This is the % of raw shale Hg that was not recovered in the products and represents a system loss when it is negative. A positive value signifies that more Hg was recovered than was charged to the system.

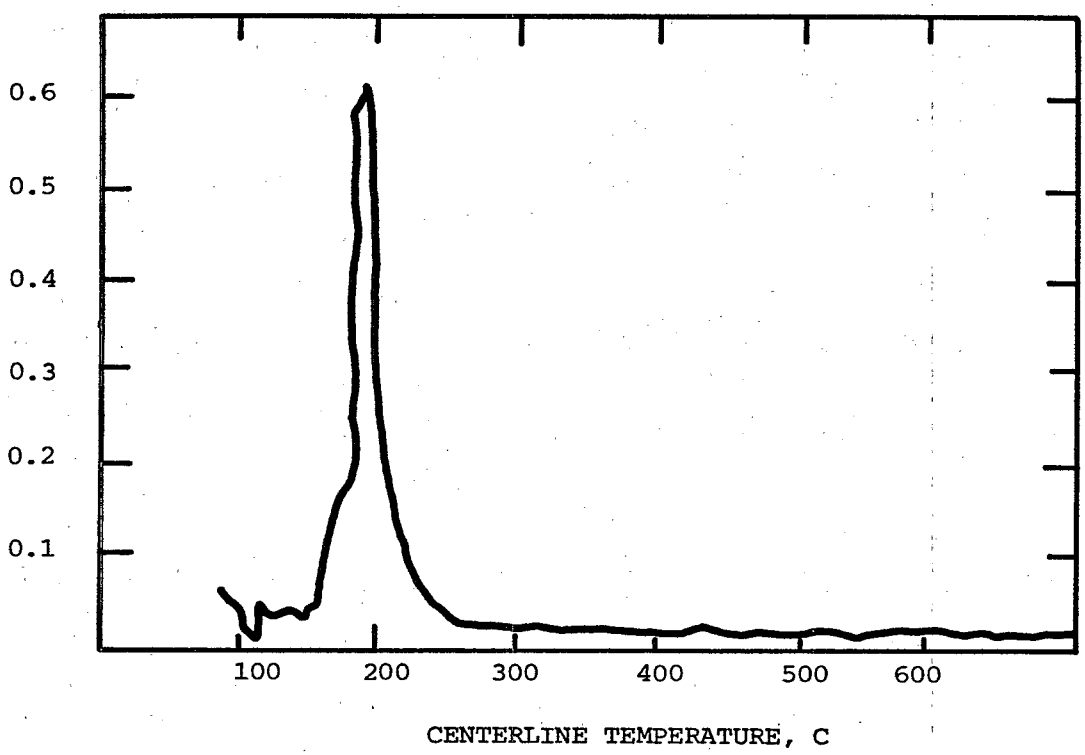


Figure A-1. Evolution of Hg into the offgas of the Battelle Pacific Northwest (BPNL) 5.5-Kg laboratory retort for a heating rate of 2 C/min to 750 C for a Green River oil shale (Olsen et al. 1985).

of shale pyrolysis and oil production. This is similar to the H_2S evolution profile from organic sulfur (Wong 1983).

They also found that from 2 percent up to 14 percent of the Hg originally present in the raw shale was not volatilized at 500 to 750 C and remained in the retorted shale. This is consistent with other studies in which maximum retorting temperatures reached 760 to 1025 C (Fox 1985a; Fox 1980, Fruchter et al. 1978). Therefore, a small fraction of the Hg in raw shale is nonvolatile at retorting temperatures and is not redistributed during retorting or emitted to the atmosphere.

Thus, the retorting temperature of commercial processes, which range from 500 C to 1200 C, should have no significant effect on the quantity of Hg that is mobilized from the raw shale, since volatilization is essentially complete by 320 C. Differences in the amount of nonvolatile Hg are probably due to differences in raw shale mineralogy, while the amount actually present in the gases depends on other aspects of retort operation, as discussed below.

Concentrations. Direct measurements of total Hg in the offgases from several types of retorts are summarized in Table A-4. In the majority of these studies, total Hg was continuously measured throughout the runs using Zeeman atomic absorption spectroscopy (Girvin and Fox 1981). All of the measurements were made in uncontrolled offgas following the condensation of heavy hydrocarbons (i.e., the shale oil).

Mercury concentrations in uncombusted retort shale gases range from less than 0.2 up to 8200 $\mu g/m^3$. Generally, much higher concentrations have been measured in gases from small laboratory retorts than from the much larger pilot and field retorts (Table A-4). These differences are partially due to Hg losses by condensation and adsorption in the extensive offgas plumbing systems of these larger field retorts (Hodgson et al. 1982, 1984) and differences in gas flow rates.

Laboratory retorts are carefully constructed and operated to minimize losses by condensation and adsorption. However, it is possible that when a commercial retort reaches steady state with respect to Hg, it may well behave

TABLE A-4. TOTAL MERCURY CONCENTRATIONS AND EMISSION FACTORS FOR SOME OIL SHALE RETORTS[1]

	Retorting Temp. (C)	Hg Offgas Concen. ($\mu g/m^3$)	Hg Emission Factor (mg/bbl oil)	Number of Measurements	Reference
LABORATORY RETORTS					
LETC 20-Kg retort[3]	760	<0.2-8200	81	Contiuously throughout 72-hr run by ZAA	Fox (1985a)
LBL 5.5-Kg retort[2]	500	<0.2-4000	208	Continuously throughout 5-hr run by ZAA	Hodgson et. al. (1982)
BPNL 5.5-Kg retort[2]	500	<0.2-4000	68	Continuously throughout run by ZAA	Olsen et. al. (1985)
PILOT RETORTS					
LLL 6000-Kg retort[3]	945	<42-540	--	25 hrs of continuous data by ZAA	Girvin et. al. (1980)
LETC 9000-Kg retort[3]					
Before offgas burner		5.0	--	1 sample using selective absorption traps	Fruchter, et. al. (1978)
After offgas burner		2.0	--		
FIELD EXPERIMENTS					
Rio Blanco Retort 1 (18x18x122m)	>1000	59-182	102	42 hrs continuous data over 35 days by ZAA	Hodgsen et. al. (1984)
Occidental Retort 6	>1000	10-26	--	13 discrete samples on separate days using Au-bead absorption tubes	U.S. DOE (1980)
Paraho direct (10.5 ton/hr)	750				
Thermal oxidizer		4.5-12	39	5 samples using Au-coated bead traps	Fruchter et al. (1979)
Recycle gas		<20-75		3 samples using Au-coated bead traps	"
Geokinetics					
Demister	>1000	1.4-2.7	--	2 samples	Rinaldi et al. (1981)
Incinerator		0.2	--	1 sample	

(1) The following abbreviations are used in this table:
LETC = Laramie Energy Technology Center; LLL = Lawrence Livermore Laboratory; LBL = Lawrence Berkeley Laboratory; BPNL = Battelle Pacific Northwest Laboratory; ZAA = Zeeman Atomic Absorption Spectroscopy.
(2) Fischer-Assay type retort in which the shale bed is uniformly heated with an external Lindberg heater.
(3) Laboratory simulations of in-situ processes.

much as a laboratory retort, since Hg losses by condensation and adsorption would no longer be significant.

Differences between laboratory and field Hg concentrations are also due to short duration sampling of highly variable offgas emissions, which results in low concentrations for pilot and field in-situ retorts. Laboratory studies are also typically conducted in an inert atmosphere with sweep gas volumes limited to rates required to remove the oil product from the reaction zone.

Very little information is available on Hg concentrations in combusted retort gases. The three studies reported in Table A-4 indicate that concentrations range from 0.2 to 75 $\mu g/m^3$, typically at least a factor of three lower than in corresponding uncombusted gases. Natural gas combustion air and combustion by-products reduce stack gas concentrations, compared with uncombusted retort gas. Particulate Hg is formed during combustion, as discussed below, and some probably settles out.

Emission Factors. Mercury emission factors were estimated from the raw data for the Hg mass balances summarized in Table A-3 and the concentrations presented in Table A-4. The resulting factors (Table A-4) range from 39 to 208 mg of Hg per barrel of oil produced, which scales to 14.2 to 75.9 metric tons per year for a 1,000,000 bbl/day industry. This wide range in emission factors is primarily due to variations in raw shale Hg content and system losses.

Emission Profiles. The temporal distribution of Hg in offgases varies for surface and in-situ processes. Mercury is uniformly emitted from surface processes, while for vertical modified in-situ (VMIS) retorts (i.e., Occidental, Rio Blanco), it is nonuniformly emitted in a pulse during the final one-third of each retort burn (Figure A-2).

The nonuniform distribution typical of VMIS retorts is attributed to their unique operation (Fox 1985a; Fox et al. 1978). A VMIS retort consists of a stationary bed of rubblized shale through which a reaction zone is vertically propagated by sweep gases that are continuously introduced at the top of the retort. Mercury present in the raw shale is released at 160-320 C,

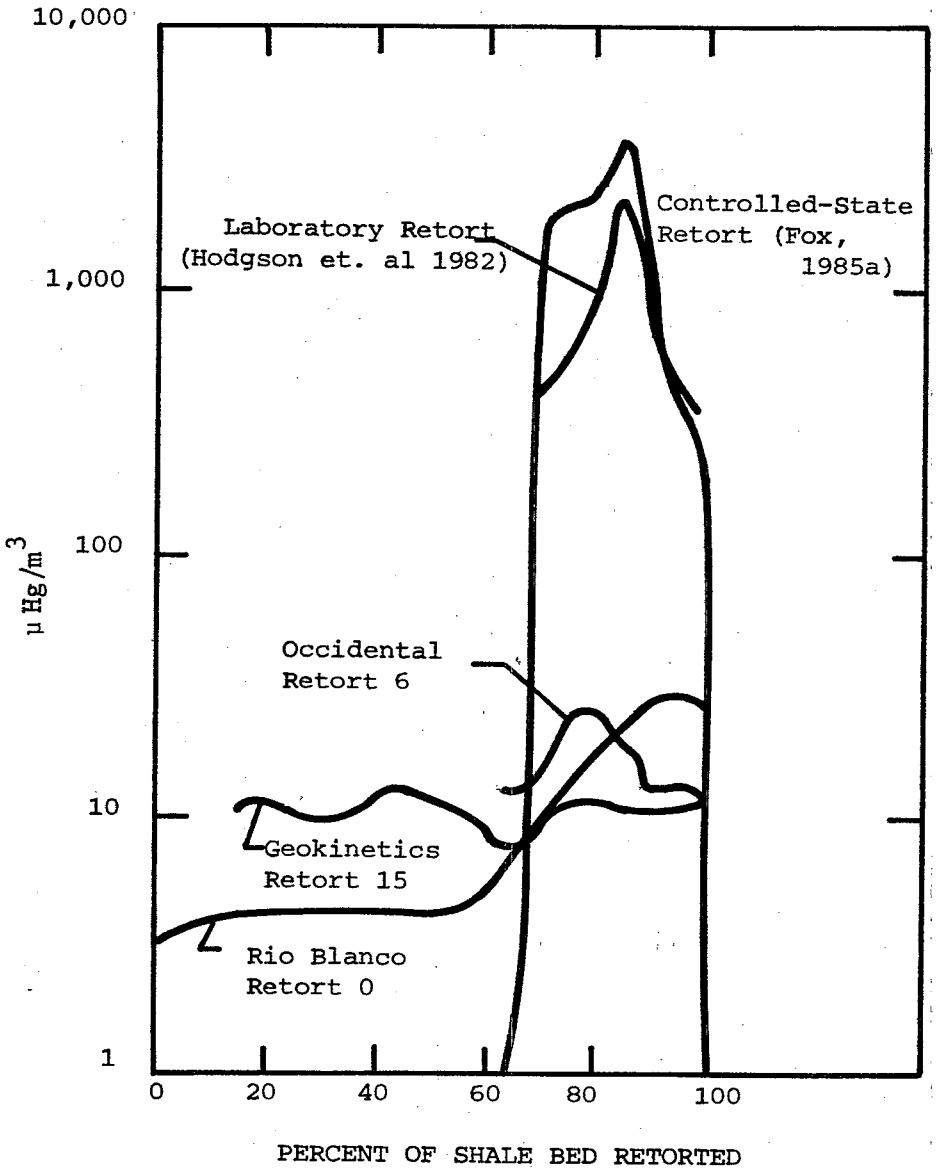


Figure A-2. Mercury emissions from in-situ retorts as a function of percent burn complete (Frucheter et. al. 1983, except as noted).

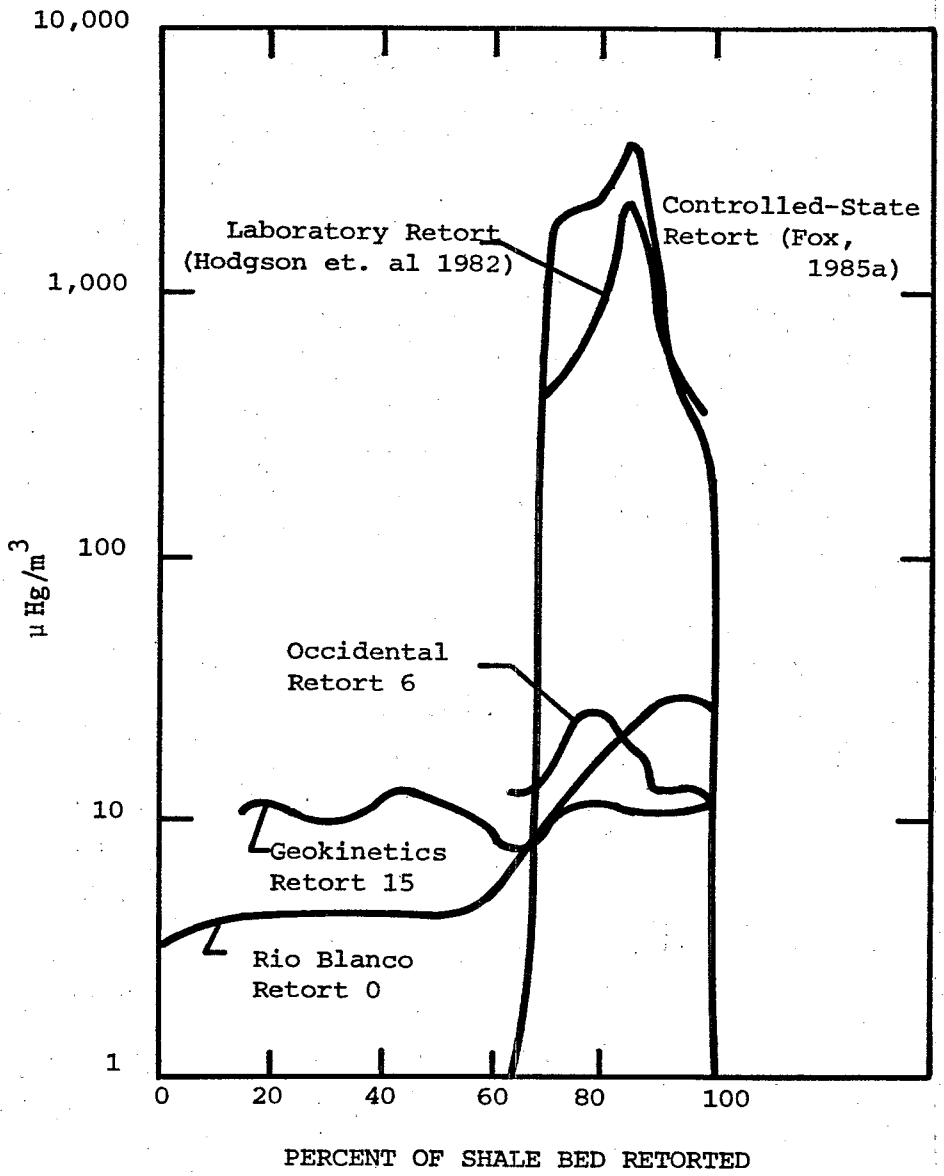


Figure A-2. Mercury emissions from in-situ retorts as a function of percent burncomplete (Frucheter et. al. 1983, except as noted).

well in advance of pyrolysis. This Hg is carried down the bed by the sweep gases and is eventually deposited on the cool shale ahead of the reaction zone. This deposited Hg is subsequently revolatilized when the bed heats up.

This process of successive volatilization and deposition continues during retorting, retarding the release of Hg from the retort. When the shale bed is saturated with Hg and/or the retort gets sufficiently hot, Hg is volatilized and swept out of the retort. If the shale bed is saturated with Hg before the reaction zone reaches the bottom of the retort, Hg will be uniformly released at a low level prior to the appearance of the pulse. This behavior has been observed during two field experiments (Occidental, Rio Blanco) and in laboratory simulations of the in-situ process (Figure A-2).

Additional substantiating evidence (Fox 1985a) was also provided by studying the Hg profile along a partially retorted shale bed (Figure A-3). Individual 0.5 ft segments of the bed were separately recovered and analyzed for Hg, Cd and other elements. Both Hg and Cd accumulated in cool zones below the point at which retorting had been terminated. Mercury accumulation is evident in the peak at zone 17.

The above-discussed emission profile was not observed at the Geokinetics retort (Figure A-2), a horizontal true in-situ retort, nor at Rio Blanco Retort 1 (Hodgson et al. 1984). Hodgson's measurements at Rio Blanco Retort 1 were made during the latter half of the run and did not span a long enough time period to observe a Hg peak. Hodgson observed large temporal variations in the Hg emission rate, both within and between days, presumably due to large variations in retorting conditions used in an attempt to control retorting. Additional variability is expected from natural variations in raw shale mineralogy (Hodgson et al. 1984).

The cause of the relatively uniform Hg emissions from the Geokinetics retort is uncertain. Since it is a distinct technology and differs substantially from VMIS, it is not unreasonable to anticipate a different emission profile. Mercury may have been deposited downstream of the exit gas piping, in unretorted regions of the retort, or in the overburden above the retort.

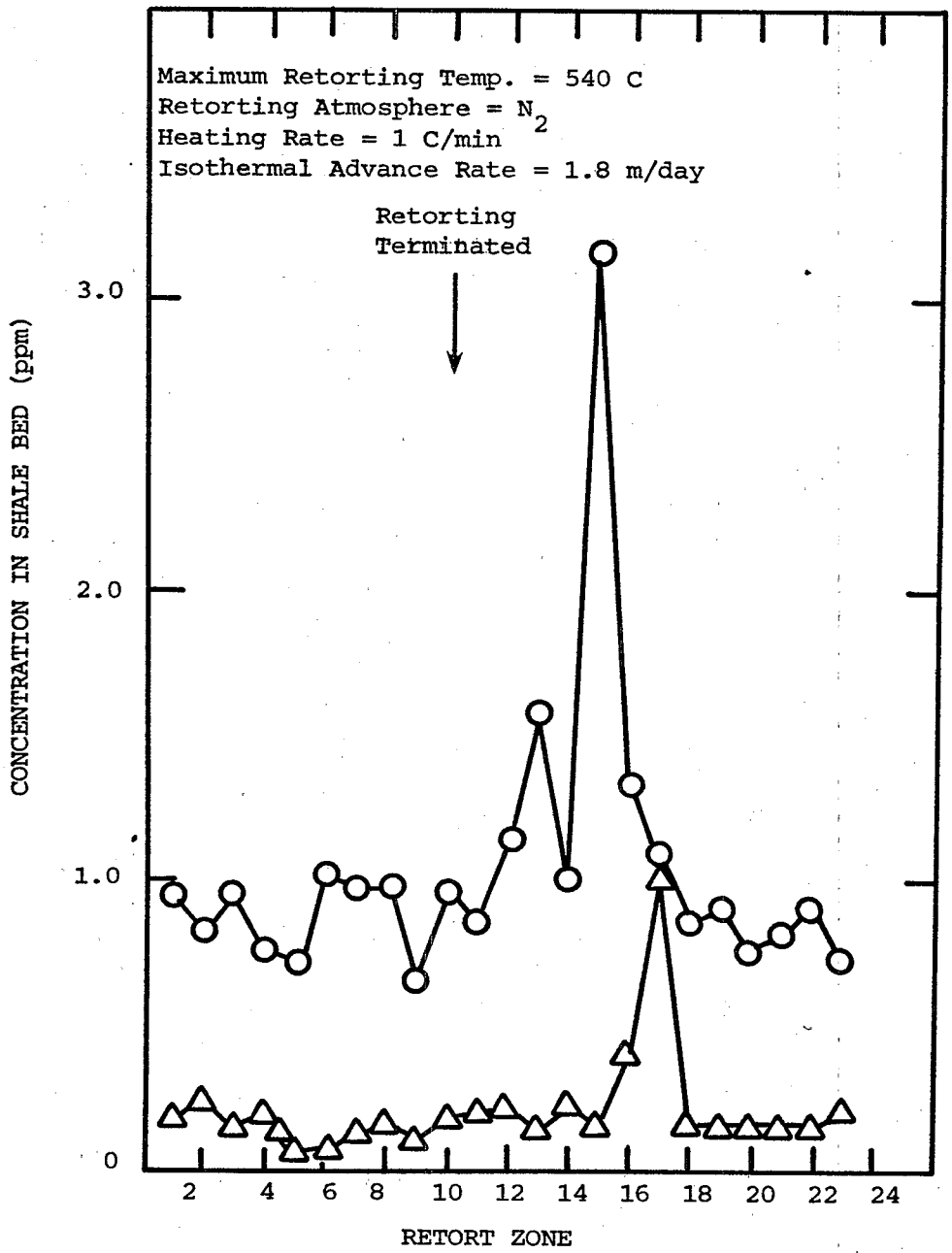


Figure A-3. Variation of Hg and Cd along the shale bed in an interrupted run of the 20-Kg controlled state retort. Zones are numbered consecutively from the top to the bottom of the retort. Each zone is 15 cm (0.5 ft) in length (Fox 1980).

The nonuniform emission profile typical of most in-situ retorts (Figure A-2) is not expected for suface retorts, which are operated as plug flow reactors with co- or countercurrent gas flow (i.e., Union B, Paraho). The shale bed continuously moves while the reaction zone is stationary. Mercury emission profiles for these retorts are expected to be relatively uniform, with minor fluctuations in offgas Hg concentrations due to differences in the Hg content of the raw shale feed. Although no long-term monitoring has been conducted at a surface retort, several samples collected over a four-month period by Fruchter et al. (1979) definitely displayed this type of behavior.

However, Hg emissions from these types of processes are reduced by a process similar to that occurring in in-situ retorts. If retort gases contact retorted shale that is not subsequently reheated, Hg will be deposited on the shale. For example, in the Paraho process, the shale moves downward, countercurrent to recycle gases, which move upward. Retorted shale and recycle gases meet near the bottom of the retort, providing opportunity for adsorption of gas-phase constituents onto the solids.

The Hg mass balance for the Paraho direct-mode process (Table A-3) provides clear evidence for the occurrence of this mechanism. In that study, only 23 percent of the Hg present in the raw shale was found in the retort gases, an unprecedentedly low amount, while 72 to 100 percent has been found in the gases from other retorts (Table A-3). In contrast, 33 percent of the Hg was found in the retorted shale, while 2 to 12 percent has been found in most other retorted shales. (The high values of 18 and 29 percent in Table A-3 are due to incomplete retorting of the shale bed and Hg accumulation in partially retorted shale in the bottom of the retort.) Apparently, Hg in the recycle gas in the Paraho retort is deposited on cool retorted shale in the bottom of the retort. A similar mechanism has been postulated to control offgas Hg concentrations for the Union B process (UOC 1985).

Mercury Speciation. The majority of the Hg measured in uncombusted shale gases exists in the gaseous state. Hodgson et al. (1982) reported that the particulate Hg concentrations in two offgas samples from the Lawrence Berkeley Laboratory 5.5 kg retort were 5 and 14 $\mu g/m^3$, contributing 1.1 and

1.7 percent of the total Hg. Similarly, Fruchter et al. (1978) reported that the particulate Hg concentration in a single offgas sample from the LETC 10-ton retort was 0.15 $\mu g/m^3$, or 3.0 percent of the total Hg. Work reported by Ondov et al. (1982) suggests that the particulate fraction is mostly oil mist.

Some of the gaseous Hg is converted into particulate Hg when the gases are combusted. Mercury volatilized during retorting is converted into elemental and oxide forms and condenses, either homogenously or on the surfaces of even smaller ash particles. Similar behavior has been widely reported for Hg from many other combustion sources (e.g., Gladney et al. 1976).

Fruchter et al. (1978) found that 32 percent of the Hg in the stack gases from an offgas burner on the LETC 10-ton retort was in the particulate form (0.6 $\mu g/m^3$), while only 3 percent of the Hg was in the particulate form before the burner (0.15 $\mu g/m^3$). He also later reported (Fruchter et al. 1979) that Hg in dust from the Paraho thermal oxidizer was enriched by a factor of 140 times, relative to retorted shale, indicating significant condensation of Hg in the thermal oxidizer.

The specific Hg compounds present in the gases have been observed to vary in both time and by source (Table A-5). Organomercurials dominate in untreated gases (Hodgson et al. 1982, 1984; Olsen et al. 1985), while elemental Hg and particulate Hg dominate in combusted stack gases (Fruchter et al. 1978, 1979). However, early in retorting, before hydrocarbons are released into the gas stream (i.e., during startup of a commercial plant), elemental Hg dominates. Later, when hydrocarbon vapors are present, organomercurials, primarily dimethylmercury and diethylmercury, predominate (Olsen et al. 1985); traces of elemental Hg, mercuric chloride, methylmercury chloride, di-n-propylmercury, and methylethylmercury may also be present.

These alkyl Hg compounds are widely acknowledged to be the most toxic Hg species (WHO 1976; ACGIH 1983), and even small releases of untreated retort gases, such as from process leaks or during upset conditions, could result in health effects. The concentrations of alkyl Hg compounds reported by Olsen

TABLE A-5. GASEOUS MERCURY SPECIES IN SOME RETORT GASES
($\mu g/m^3$)

		In-Situ Retorts			Surface Retort	Laboratory Retorts				
						5.5 Kg LBL(6)	5.5 Kg BPNL(7)			
						Untreated Gas				
	Boiling Point(2) (°C)	Rio Blanco Retort 1(3) Untreated Gas	LETC 10-ton(4) Before Offgas Burner	LETC 10-ton(4) After Offgas Burner	Paraho Direct(5) Thermal Oxidizer	Downside of Peak	Onset of Hg Evolution(8)	Hg Peak	Downside of Hg Peak	End of Hg Evolution
Hg°	357	30-94	(4)	(4)	4.5- 9.4	0.0	183	0.68	0.62	N.D.
$HgCl_2$	304		1.0	0.1	<0.1-<0.3		N.D.(1)	N.D.	N.D.	N.D.
CH_3HgCl	Vol. 100		0.8	0.3	<0.2-<0.4		N.D.	N.D.	N.D.	N.D.
$(CH_3)_2Hg$	95		3.0(4)	0.9(4)	<0.1-<0.4		23	N.D.	3350	24.5
$(C_2H_5)_2Hg$	159	66-138	(1)	--	--	1400	N.D.	N.D.	45.2	12.7
$(n\text{-}C_3H_7)_2Hg$	190		--	--	--		N.D.	N.D.	11.4	N.D.
$(CH_3)(C_2H_4)Hg$			--	--	--		N.D.	N.D.	11.4	N.D.

1) A dash (--) signifies the species was not analyzed and N.D. (not detected) signifies that it was looked for, but below the detection limit.

2) Dean (1979) and CRC (1964).

3) Hodgson et. al. (1984).

4) Fruchter et. al. (1978). The $(CH_3)Hg$ includes HgO.

5) Fruchter et. al. (1979).

6) Hodgson et. al. (1982).

7) Olsen et. al. (1985).

8) Typical Hg evolution peak is shown in Figure A-1.

et al. (1985) greatly exceed the threshold limit value of 10 $\mu g/m^3$ established for alkyl Hg compounds in workroom air (ACGIH 1983).

The organomercurials in the untreated gas would be combusted in most commercial processes and converted into elemental Hg and HgO, which are considerably less toxic. Only traces of methylmercury chloride and dimethylmercury have been observed in combusted retort gases (Fruchter et al. 1978).

Pollution Control. The foregoing sections have demonstrated that from 39 to 208 mg of Hg may be emitted for every barrel of oil produced (Table A-3). For each 10,000 BPD of production capacity, this corresponds to an emission rate of 0.2 to 0.8 ton/yr of Hg.

However, shale gases will be variously processed and treated to remove the criteria pollutants prior to discharge to the atmosphere. Commercial oil shale plants will include technology to reduce particulates, SO_x, NO_x and nonmethane hydrocarbons, as discussed in other chapters of this report. The effectiveness of this technology in reducing Hg emissions from oil shale retorts has not been studied experimentally.

Some studies have used thermodynamic calculations to predict the effect of various pollution control devices on offgas Hg concentrations (U.S. DOE 1979; Cathedral Bluffs 1981). These analyses, which characteristically conclude that essentially all of the offgas Hg is removed in the pollution control devices, assume that the Hg is present in its elemental form. In fact, the majority is present as highly volatile organomercurials (Table A-5) that would not condense or precipitate out in the pollution control devices.

Very few studies have experimentally determined the effect of pollution control devices on trace element emissions. The authors are not aware of any studies that demonstrate that gaseous Hg or very small particulates that are highly enriched in trace elements are removed in any of the pollution control technologies proposed for oil shale plants.

It is widely acknowledged that particulate reduction technology, such as electrostatic precipitators, do not remove very fine combustion

particulates that may be highly enriched in trace elements (e.g., Kaakinen 1974; Gladney et al. 1976; Klein et al. 1975; Krishnan and Hellwig 1982). Scrubbers are likewise ineffective (Mansour and Jones 1978; Kaakinen 1974). A sizeable (and presently unquantified) fraction of trace element emissions from oil shale retorts may be present in this form.

It is significant to note that all of the six proposed commercial oil shale projects that have submitted PSD applications (Table A-1) have estimated that Hg emission rates are considerably less than discussed herein and that no gaseous Hg is emitted; the major source of Hg emissions in these PSD applications is from fugitive dusts, which are primarily raw and retorted shale fines.

Company estimates of Hg emission rates are based on Hg measurements in untreated gases from pilot plants and/or on the assumption that the majority of the Hg occurs as elemental Hg and thus would condense out in the pollution control devices and offgas plumbing. However, in all cases where the authors were able to examine the supporting data, the claim of no gaseous Hg emissions is unsubstantiated. Technology conventionally used to control Hg emissions in other industries have been summarized and reviewed by Sittig (1976).

Cadmium

Very few studies have been conducted on the distribution of Cd during oil shale retorting, primarily because Cd is difficult to accurately measure in oil shale materials. Cadmium concentrations of from 1 $\mu g/m^3$ to over 1000 $\mu g/m^3$ have been directly measured in retort gases. Since these concentrations greatly exceed the ACGIH recommended limit of 50 $\mu g/m^3$ on Cd and its compounds in work-room air, these concentrations may be environmentally significant.

Mass balance studies reveal that for most commonly encountered retorting conditions, an average of from 12 percent to 29 percent of the Cd originally present in the raw shale is lost from the retort, presumably in the retort gases. If losses to the offgas plumbing system are not considered, this corresponds to the emission of from 0 to 175 mg of Cd per barrel of oil produced. Generally, Cd losses increase with increasing retorting temperature, and thus are expected to be higher for in-situ retorts.

Mass Balance Studies. Cadmium mass balances for three types of oil shale retorts are summarized in Table A-6. None of these studies directly measured Cd in the gases. However, the imbalances or system losses, which are the percent of the raw shale Cd that was not recovered in the retorted shale, oil and water, provide an upper-limit estimate of the amount of Cd that may be distributed to the gases.

These studies demonstrate that the majority of the Cd is either nonvolatile and retained in the retorted shale, or it is lost from the retort, presumably to the gases. An average of 12 to 33 percent of the Cd was not recovered in these studies and was presumed to be volatilized from the raw shale. Therefore, the potential exists for Cd emission from virtually all commercial processes.

However, some of the volatilized Cd may have a short residence time in the gas. Hodgson (1985) found that 14 percent of the Cd originally present in the raw shale had condensed in the gas sampling lines, while in another study, Fox (1980) found that Cd was removed from the gases at <390 C and deposited on the partially retorted shale (Figure A-3).

In contrast, in the modified Fischer Assay retort (Goodfellow and Atwood 1974), all of the Cd was recovered in the retorted shale in two separate studies using different analytical methods (Fox 1985b; Shrendrikar and Faudel 1978). This is believed to be due to the rapid heating rates, from 2 to 14 C/min, and the relatively short time (20 mins) that the retort is held at its maximum temperature of 500 C. If this is the correct interpretation, then little or no Cd may be volatilized from fluidized bed combustors, such as those proposed for the Union C, Lurgi and Chevron processes.

These studies (Table A-6), which reported Cd losses for a range of retorting conditions, clearly demonstrate that the fraction of the Cd lost from the retort increases with increasing retorting temperature. The data in Table A-6, except that for the Fischer Assay, is plotted in Figure A-4 as a function of maximum retorting temperature. This not only demonstrates that the percentage of the Cd lost and presumably volatilized increases with temperature, but also that the magnitude of the loss is inversely related to the capacity of the retort.

TABLE A-6. CADMIUM MASS DISTRIBUTION PATTERNS FOR SOME OIL SHALE RETORTS(1)
(Percent of Total Cd in Raw Shale)

	LETC 20-kg Simulated In-Situ Retort		LLL Simulated In-Situ Retort	Small Lab. Retorts Fischer Assay		LBL 5.5 kg
	4 Runs T=540 C (2)	7 Runs T=540 C (3)	4 Runs(a) T=494-1003 C (2) (3)	1 Run T=500 C (4)	3 Runs T=500 C (5)	1 Run T=500 C (6)
Retorted shale	80.4	70.6	87.8	103	100	65.4
Shale oil	<0.3	<0.3	0.3	<0.3	0.0	1.3
Water	<0.01	0.003	<0.01	<0.2	0.0	<0.03
Gas	--	--	--	--	--	--
Total	80.4	70.6	88.1	103	100	66.7
Imbalance(7)	-19.6	-29.4	-11.9	+3.0	0.0	-33.3

(1) Abbreviations used in this table are:

LETC = Laramie Energy Technology Center; T = temperature; LLL = Lawrence Livermore Laboratory; LBL = Lawrence Berkeley Laboratory; BPNL = Battelle Pacific Northwest Laboratory.

(2) Fox et. al. (1977).

(3) Fox (1980).

(4) Fox (1985).

(5) Shendrikar and Faudel (1978).

(6) Hodgson (1985).

(7) When this value is positive, it indicates that more Cd was recovered in the products (i.e., shale oil, retort gas, water, retorted shale, than was charged to the system in the raw shale. The loss includes Cd deposited in the offgas plumbing and volatilized to the gases, a well as experimental error.

This inverse relationship is analagous to the one previously noted for Hg (gas concentration decreases as capacity of retort increased, Table A-4) and likely has a common source, namely system losses. Cadmium may deposit within the retort, either in the shale bed or offgas system, due to nonuniform heating and/or sweep gas flow. In small laboratory retorts, such as the 5.5-kg retort used by Hodgson, retorting can be precisely controlled, and relatively uniform temperatures and sweep gas flow rates can be maintained throughout the retort cross-section and length. Any Cd volatilized within such a retort would probably be released from the retort into the offgases, and the only losses would occur in the offgas and sampling system. However, as the retort is scaled up, bed heterogenities and/or poor process control can cause nonuniform bed temperatures and sweep gas flow rates, which allow the Cd to deposit in cool regions along the bed or on the retort walls.

Convincing evidence for the occurrence of such a trapping mechanism in larger retorts has been presented by Fox (1980). Cadmium profiles through a partially retorted shale bed (Figure A-3) indicate that Cd is readily removed from the gases at temperatures of less than 390 C and is deposited on the shale bed ahead of the reaction zone. Some Cd also may be deposited behind the reaction zone (see peak at zone 13 in Figure A-3).

Concentrations. Very few direct measurements of Cd in shale gases have been attempted, and available measurements are believed to greatly underestimate actual offgas Cd concentrations. However, the available measurements do substantiate that Cd is volatilized from in-situ retorts and emitted in the untreated gases. Some Cd may also be emitted from surface retorts, but the evidence is less conclusive than for in-situ retorts.

Hodgson (1985) used an atomic absorption spectrometer equipped with a single-slot burner to continuously measure Cd in gases from the LBL 5.5-kg retort charged with retorted shale previously heated to 500 C. Cadmium concentrations of up to 1000 $\mu g/m^3$ were observed. However, these concentrations may be low. The Cd mass balance demonstrated a low recovery and simultaneous measurements using an impinger technique were higher. The Cd was evolved from the shale over a temperature range of 775 to 950 C and peaked

at 890 C. However, 33 percent of the Cd had been previously volatilized in an earlier run at 500 C in which the gases were not monitored.

Cadmium concentrations in discrete samples of offgases from two separate in-situ retorts have also been reported (Rinaldi et al. 1981; Ondov et al. 1982). Rinaldi et al. (1981) reported gaseous Cd concentrations of about 1 $\mu g/m^3$ in each of three samples taken at the demister inlet and outlet and the incinerator outlet of a Geokinetics field burn. The concentration of Cd in offgas particulates from the LLL 6000-kg simulated in-situ retort was 2.7 $\mu g/g$ (Ondov et al. 1982). However, neither investigator demonstrated that his sampling and analytical methods were suitable for collecting and accurately quantifying Cd in shale gases. Furthermore, a few grab samples cannot be presumed to represent a lengthy retorting experiment, particularly in light of the nonuniform emission profiles documented for these retorts (Fox 1985a).

Emission Factors. Cadmium emission factors were estimated from raw data for the Cd mass balances summarized in Table A-6, adjusted for a 14 percent loss to system plumbing (Hodgson 1985). The resulting factors, shown in Figure A-4, range from 0 to 1070 mg/bbl and average 285 mg/bbl, which scales to 104 metric tons/year for a 1,000,000 bbl/day industry.

There is presently inadequate information to develop factors for specific retorting processes. However, based on the information reviewed here, the lower end of this range (0 to 260 mg/bbl) might be used for surface retorts and the upper end (175 to 1070 mg/bbl) might be used with the higher temperature in-situ processes. It is presently uncertain whether a fluidized bed combustor would contribute additional Cd emissions.

Emission Profiles. The temporal distribution of Cd in offgases has not been measured but is expected to be similar to the Hg profiles discussed previously and summarized in Figure A-2. The nonuniform emission of Hg for in-situ retorts is due to its accumulation along the shale bed, ahead of the reaction zone, as shown in Figure A-3. This retards the release of Hg, causing a nonuniform emission profile (Figure A-2), as discussed previously. Since Cd also accumulates along the shale bed (Figure A-3), it is also expected to be nonuniformly emitted.

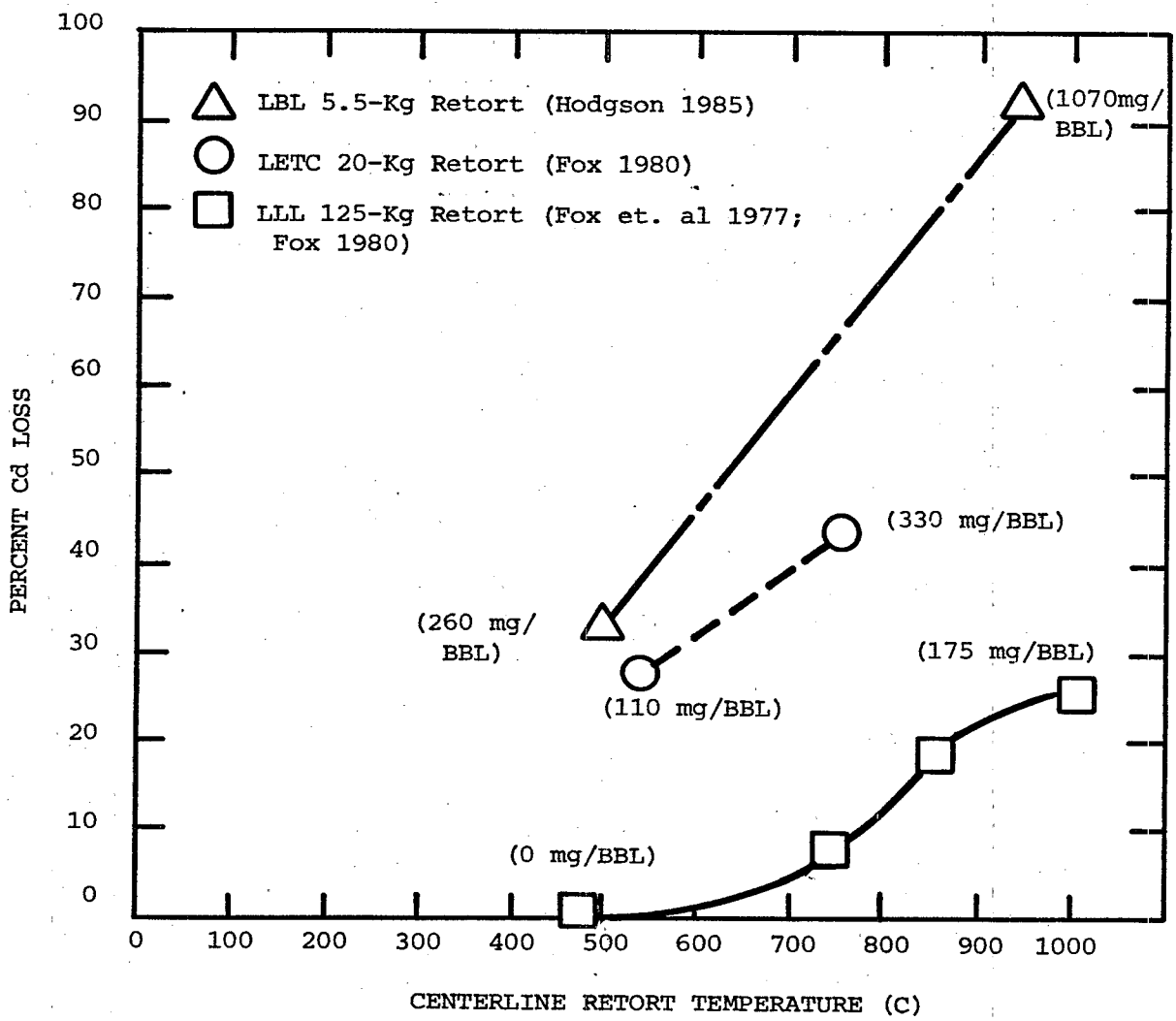


Figure A-4. Effects of maximum retorting temperature on Cd loss from oil shale retorts. Values in parentheses are Cd emission factors in mg Cd volatilized per barrel of oil produced.

Cadmium Speciation. The Cd species present in shale gases have not been determined. However, they are expected to include some of the same species reported for Hg since both Cd and Hg are Group IIb metals and have similar chemistries. In particular, both metals are known to form organometallics of the type R_2M and RR'M, where R and R' are alkyl groups and M is the metal (Hagihara et al. 1968). However, the thermal stability of Cd compounds is also the lowest among the organometallics of the Group IIb metals, and the only organocadmium compound that is stable under ambient conditions is dimethylcadmium.

Pollution Control. The foregoing sections indicate that from 0 up to 1070 mg Cd may be released to the gases for every barrel of shale oil produced (Figure A-4). For each 10,000 BPD of production capacity, this corresponds to the emission of up to 4 ton/yr of Cd. The effectiveness of criteria-pollutant control technology in reducing these emissions is unknown and cannot even be speculated upon due to limited information.

No emission or ambient air quality standards exist for Cd, and it is not addressed in any existing oil shale PSD permit. Therefore, no control technology has been required to remove Cd. However, the potential magnitude of Cd emissions and the well-established toxicity of Cd and its compounds indicate that further study of this element is warranted.

Selenium

Selenium concentrations of 4.2 $\mu g/m^3$ to over 18 $\mu g/m^3$ have been reported in the gases from in-situ retorts. Since these are well below the OSHA standard of 200 $\mu g/m^3$ on Se and its compounds in workroom air, these concentrations probably are not environmentally significant. Selenium has not been detected in gases from surface processes. Emission factors of >15 to 940 mg/bbl have been estimated for in-situ retorts.

However, some mass balance studies suggest that much higher Se concentrations could occur in gases from in-situ retorts. For retorting temperatures of 900 C to over 1000 C, from 8 to 27 percent of the Se is unaccounted for in the products and presumed to be lost to the gases.

Mass Balance Studies. Selenium mass balances for four types of oil shale retorts are summarized in Table A-7. These studies demonstrate that the majority of the Se, from 69 percent up to 100 percent, is nonvolatile and is retained in the retorted shale. Little or no Se is apparently volatilized at 500 C since near-zero mass balances were obtained for both the Paraho direct and Fischer Assay retorts. At temperatures greater than about 890 C in in-situ retorts, from 8 percent up to 27 percent of the Se originally present in the raw shale (Fox 1985b) is not accounted for in the products, suggesting that it is volatilized and emitted from the retort.

Selenium behavior at temperatures from >500 C up to about 870 C is uncertain because mass balances within this range have positive imbalances due to a sampling and/or analytical problem. This obscures any Se losses since the positive imbalance is greater than the loss.

However, Ondov et al. (1982) measured a Se concentration of 9 $\mu g/m^3$ in the gases from one of the same retorts studied by Fox (1980) operated at 840 C. This corresponds to the volatilization of about 0.1 percent of the raw shale Se. This low distribution is much too small to detect in mass balance studies.

Concentrations. Although Se has been directly measured in retort gases and particulates, the available measurements are believed to underestimate actual offgas Se concentrations. However, the available measurements do substantiate that Se is volatilized from in-situ retorts and is present in the untreated gases. There is currently no evidence that Se is emitted from surface retorts.

Ondov et al. (1982) used coconut charcoal absorption tubes to trap gaseous Se emitted during a run of LLL's 6000-kg simulated in-situ retort. Adsorbed vapors were analyzed by neutron activation analysis. Three discrete samples were taken during this run at 59 hr, 101 hr and 121 hr following retort ignition. The first sample was collected when the average centerline temperature was 842 C and lean shale was being retorted; it contained 8.9 $\mu g/m^3$ of Se. The next two samples were taken when the average centerline temperature was 945 C and rich shale was being retorted. The capacity of the charcoal tube for Se was exceeded for both of these samples, and only minimum

TABLE A-7. SELENIUM MASS DISTRIBUTION PATTERNS FOR SOME OIL SHALE RETORTS[1]
(Percent of Total Se in Raw Shale)

	LETC 20-kg Simulated In-Situ Retort	LLL Simulated In-Situ Retort		Paraho Direct	Fischer Assay
	11 Runs T = 540-760 C (2)	3 Runs T = 494-868 C (2)	6 Runs T = 887-1025 C (2)	1 Run T = 750 C (3)	3 Runs T = 500 C (4)
Retorted shale	121	123	69.2	95.8	99.9
Shale oil	5.9	3.3	4.4	3.8	0.0
Water	0.7	0.6	0.6	1.9	0.0
Gas	--	--	--	<0.8	--
Total	127.6	126.9	74.2	101.5	99.9
Imbalance[5]	+27.6	+26.9	-25.8	+1.5	-0.1

(1) The following abbreviations are used in this section:

LETC = Laramie Energy Technology Center; T = temperature; LLL = Lawrence Livermore Laboratory; LBL = Lawrence Berkeley Laboratory; BPNL = Battelle Pacific Northwest Laboratory

(2) Fox (1980).

(3) Fruchter et.al. (1980).

(4) Shrendikar and Faudel (1978).

(5) When this value is positive, it indicates that more Se was recovered in the products than was charged to the system. This generally indicates a sampling or analytical problem. When the value is negative, less Se was recovered in the products than was charged in the raw shale. The loss includes Se deposited in the offgas plumbing and volatilized to the gases as well as all experimental error.

values of 18 $\mu g/m^3$ and 14 $\mu g/m^3$ were reported. The actual concentrations could be higher. During this run, Se concentrations in the particulates were 0.1 to 2.5 $\mu g/g$.

In another study at a Geokinetics in-situ retort, Se concentrations of 0.5 and 4.2 $\mu g/m^3$ were measured at the demister outlet and the incinerator outlet, respectively. Corresponding particulate loadings at the demister inlet and incinerator outlet were 5 and 10 $\mu g/g$, respectively. The accuracy or representativeness of these samples is unknown. However, they are of the same order of magnitude as those reported by Ondov et al. (1982) at the LLL simulated in-situ retort.

Emission Factors. Selenium emission factors were estimated from raw data for the Se mass balances reported by Fox (1980) (Table A-7) and from the gas measurements reported by Ondov et al. (1982).

Ondov's measurements indicate that for a retorting temperature of 842 C in lean shale, 12 mg of Se was emitted per barrel of oil produced. For a retorting temperature of 945 C in rich shale, over 9 mg of Se was emitted per barrel of oil. The combined emission for the entire run is >15 mg/bbl. This is estimated to be about 0.3 percent of the Se originally present in the raw shale.

The six Se mass balances reported by Fox (1980) for the LLL simulated in-situ retorts indicate that from 420 to 940 mg of Se was lost from the retort for each barrel of oil produced. These values are over an order of magnitude higher than values calculated from Ondov's direct gas phase measurements. It is highly probable that some of the Se lost in Fox's studies was deposited in cool regions within the retort and offgas plumbing system, as previously reported for both Cd and Hg, and was not present in the gas stream. It is also possible that Ondov's values are very low, due both to nonuniform emissions from in-situ retorts and the reported saturation of the collection device.

Thus, for in-situ retorts, the Se emission factor ranges from greater than 15 to 940 mg per barrel of oil, based on available data. The emission

factor for surface retorts is believed to be near zero. Additional experimental studies are required to improve these estimates.

Arsenic

Mass Balance Studies. The volatility of As and its compounds is similar to that of Se and much less than Hg and Cd. Mass balance studies are not useful for estimating the amount of As volatilized and potentially emitted because less than 1 percent of the raw shale As is lost, which is considerably less than the experimental error in such studies. Additionally, the As mass balances reported in most studies (Fox 1980, Wildeman and Meglen 1978; Fox et al. 1977) are flawed by high positive imbalances resulting from an analytical problem.

Concentrations and Emission Factors. Only two studies have reported enough information to reliably estimate the amount of As volatilized and emitted from oil shale retorts. In the first such study, Fruchter et al. (1980) completed an As mass balance for the Paraho semiworks retort. Total As ranged from 13 to 55 $\mu g/m^3$ in the thermal oxidizer stack gases and from 120 to 155 $\mu g/m^3$ in the recycle gases. Since these, as well as other reported As concentrations, are well below established OSHA and ACGIH standards for workroom air, they are not believed to be environmentally significant. This is equivalent to about 0.07 percent of the As present in the raw shale and corresponds to an emission rate of 330 mg/hr and an emission factor of 65 mg/bbl.

In another study, Ondov et al. (1982) measured the total As concentration in three samples of gas from a run of the LLL 6000-kg simulated in-situ retort, as previously described above for Se. Arsenic concentrations ranged from 5.5 to greater than 17 $\mu g/m^3$; corresponding As concentrations in particulates were 0.3 to 2.5 $\mu g/g$. This is equivalent to an emission factor for the entire run of greater than 13 mg of As per barrel of oil and corresponds to the emission of about 0.01 percent of the As originally present in the raw shale. This value, which is an order of magnitude or more lower than emission factors for the Paraho and Union B retorts, is believed to be

low, as previously discussed above for Se. The higher temperatures of the LLL retort should result in higher emissions.

Higher As concentrations, from 11 to 380 $\mu g/m^3$, have been reported by UOC (1985) for the Unisulf offgases from their Brea pilot plant (simulates the Union B retort). Concentrations in Unisulf gases are reduced when natural gas and air are added prior to combustion, resulting in stack concentrations of 0.1 to 28 $\mu g/m^3$. This corresponds to an emission rate of 12 g/hr and an emission factor of 740 mg/bbl. This is estimated to be about 0.01 percent of the As originally present in the raw shale.

Significantly higher concentrations of gaseous As were reported by Rinaldi et al. (1980) for a Geokinetics in-situ retort. The As concentration at the demister inlet was 130 $\mu g/m^3$ while at the demister outlet, it was 38 $\mu g/m^3$. The concentration dropped to 0.4 $\mu g/m^3$ at the incinerator stack. These data are consistent with those reported by Fruchter et al. (1979) for the Paraho retort. Corresponding As concentrations in particulates were 40 $\mu g/g$ at the demister inlet and 140 $\mu g/g$ at the demister outlet. Apparently, gaseous As is converted into particulate As, as reported by Fruchter et al. (1978) for another in-situ retort. This is common to most combustion sources.

Arsenic Speciation. Fruchter has reported As speciation data for gases from several oil shale retorts (U.S. DOE 1980; Fruchter et al. 1983; Fruchter et al. 1979; Fruchter et al. 1978). These measurements (Table A-8) indicate that gaseous arsenic trioxide and arsine occur in all samples. Traces of arsine have also been variously reported by the U.S. DOE (1980) for the Paraho direct mode retort (0.17-0.46 $\mu g/m^3$) and by Rinaldi et al. (1981) for a Geokinetics in-situ retort (<5 to 5 $\mu g/m^3$).

Traces of organoarsenic compounds, including methylarsine and dimethylarsine, were reported in uncombusted stack and recycle gases from both in-situ and surface processes. These compounds were not found in the thermal oxidizer gases, presumably because they are combusted and converted into arsenic trioxide.

Arsenic particulate concentrations (Table A-8) are very low (0.4-1.2 $\mu g/m^3$) in uncombusted retort gas, constituting from 1 to 3 percent of the

TABLE A-8. ARSENIC SPECIES IN SOME RETORT GASES[1]

	Field In-situ Retorts		Simulated In-situ		Paraho Direct Mode[5]	
	Occidental Retort 6 Stack Gas[2]	Rio Blanco Retort 1 Untrd. Gas[3]	Before Off-gas Burner[4]	After Offgas Burner[4]	Thermal Oxidizer	Recycle
Arsenic						
As_2O_3	<0.8 - 1.8	2-40			12-36	112-149
AsH_3	0.85- 3.5	<0.03-0.5			<0.2-0.3	1.8-2
CH_3AsH_2			15.0	1.4	<0.2	2.8-3
$(CH_3)_2AsH$	<0.4 -<0.9	N.D.			<0.2	1.0-1.7
Other Organic As					<0.2	0.5
Particulate As	--	--	0.4	2.5	19	1.2
Total As	--	--			12.3-55	120-155

(1) The following abbreviations are used in this table:

LETC = Laramie Energy Technology Center; N.D. = Not Detected.

(2) U.S. DOE (1980); Stack gas compressed and demisted prior to venting. Temperature in stack was 78± 3C. AsH_3 values reported as near blanks and thus not significant.

(3) Fruchter et. al. (1983).

(4) Fruchter et. al. (1978) (LETC 10-ton retort)

(5) Fruchter et. al. (1979).

total As, and are somewhat higher (2.5-19 $\mu g/m^3$) in flared gases, contributing from 35 to 64 percent of the total As. Arsenic volatilized during retorting is converted into elemental and oxide forms which condense, either homogenously or on the surface of even smaller ash particles. This is substantiated by the enrichment of As in thermal oxidizer ash (Table A-9) and is consistent with numerous studies of other combustion sources (i.e., Gladney et al. 1976).

Other Trace Elements

Concentrations of other trace elements measured in gases from several retorts are summarized in Table A-9, together with the data previously discussed for Hg, Cd, Se and As. This summary indicates that the only other elements that have been detected in shale gases are the halogens, Br (>25 $\mu g/m^3$) and Cl (>980 $\mu g/m^3$), and Cr, Fe, Mo, Pb and Sb. Neither of the halogens exceed OSHA or ACGIH limits for workroom air. Flourine may also be emitted from in-situ shale retorts and should be measured.

Of the remaining elements which have been detected (Cr, Fe, Mo, Pb, Sb), only Sb compounds are expected to be volatile at retorting temperatures (CRC 1968; Dean 1979). The presence of the other nonvolatile elements is most likely due to sample or system contamination. Nevertheless, additional study of these elements is probably warranted.

Radon, a radioactive gas, has also been reported in gases from two separate burns of Rio Blanco retorts (Fruchter et al. 1983). Average Rn concentrations during these two runs were 6.7 pCi/L and 28.3 pCi/L. This corresponds to an emission factor of 6 μCi of Rn per barrel and an emission rate of 3.3 $\mu Ci/s$.

When shale gases are combusted, many of the above-discussed gaseous trace elements are converted into condensed phases, in processes similar to the production of ash in power plant boilers. The resulting fine particulates are enriched in the volatile trace elements, notably As, Cl, Br, Cd, Hg and Se. Some of these fine particulates would be removed in electrostatic precipitators or scrubbers, but the majority of them will probably be emitted.

TABLE A-9. CONCENTRATION OF GASEOUS AND PARTICULATE TRACE ELEMENTS IN RETORT OFFGASES[1]

	Paraho Direct Mode[2]				In-Situ Retorts			
					Geokinetics[3]		LLL 6-ton[4]	
	Thermal Oxidizer Dust ($\mu g/m^3$)	Enrichment Ratio (Dust ÷ Retorted Shale)	Recycle Gas ($\mu g/m^3$)	Thermal Oxidizer ($\mu g/m^3$)	Gas ($\mu g/m^3$)	Particulate ($\mu g/g$)	Gas ($\mu g/m^3$)	Particulate ($\mu g/g$)
As	520	8.8	120-155	13-55	0.4, 130	140, 40	5.5->17	0.3-2.4
B	--	--	<30	--	--	<1280	--	--
Ba	430	0.7	<900-<4000	<800-<200	--	--	<44-<80	12-78
Br	2.8	3.5	--	--	--	--	>11->25	0.12-5.1
Cd	--	--	--	--	0.9, 1	--	--	2.7
Cl	--	--	--	--	--	--	>450->980	40-3000
Co	--	--	<2-<300	<3-<7	--	10, 2	<0.6-<3.6	0.046-0.17
Cr	--	--	<7-<50	<3-<100	--	--	<70-114	1.5-13.6
Cu	--	--	<5-<100	<20	--	--	--	--
Fe	1.91%	0.8	<20-<500	<20-<30	--	<2, 50	<1900-<5100	21-280
Hg	4.2	140	<20-75	4.5-12	0.2, 1.4	--	<42-540[4]	--
Mn	380	1.0	<20-<100	<20-<100	--	--	<12-<52	1.3-2.5
Mo	--	--	<40	--	--	<11, 2	<1.5-4.2	1.2-22
Ni	--	--	<1-<200	<20-<70	--	--	--	2.1-3.7
Pb	--	--	<70-<300	<30-<80	20, 18	<23, 50	--	--
Sb	5.0	1.9	<3-<20	<6-<10	4, 2.1	--	<1.5-<4	0.1-4.0
Se	41	18	<70	--	4.2, N.D.	10, 5	8.9->18	0.1-2.5
Th	--	--	<2-<200	<2-<100	--	--	--	0.31-0.98
U	7.0	1.4	<2-<50	<2-<40	--	--	<0.6-<1.6	0.10
V	140	1.1	<40-<400	<70-<100	--	--	--	--
Zn	120	1.5	<200	<10-<100	--	--	<33-<44	2.4

(1) The following abbreviations are used in this table: LLL = Lawrence Livermore Laboratory.

(2) Fruchter et. al. (1979).

(3) Rinaldi et. al. (1981).

(4) Ondov et. al. (1982).

Fruchter analyzed a single sample of ash from the Paraho thermal oxidizer and computed enrichment ratios relative to retorted shale. The dust composition and enrichment ratios are reported in the first two columns of Table A-9. These data rather dramatically demonstrate this enrichment phenomenon. Note that As, Br, Hg and Se, all of which were found to be volatilized in this review, are significantly enriched in the oxidizer dust. The enrichment factor for Sb also suggests that it may be volatilized.

Retort emission factors are summarized and ranked in Table A-10. Interestingly, when emissions are normalized to oil yields (the emission factor), Cl emerges as the element emitted in the greatest quantities, followed by Cd, Se, As, Hg, Cr, Br and Mo.

MATERIAL HANDLING

Because oil shale is a low grade fuel source that contains about a half a barrel of oil per ton of rock, large amount of rock material must be handled at an oil shale plant. For each barrel of oil that is produced in an above-ground retort, about 2 tons of rock must be mined, crushed, sieved, hauled and loaded into the retort. About 80 percent of these 2 tons remains after the oil has been removed. This retorted shale must be cooled, unloaded from the retort, hauled, placed and compacted in a disposal pile. Additional solid material, primarily mine overburden, soil used to reclaim the disposal pile, and superficial material disturbed during construction must also be handled. At in-situ retorts, for each barrel of in-situ oil that is produced, about 2 tons of raw shale must be mined and either disposed of on the surface or processed in an aboveground retort.

The handling of these enormous quantities of material generates large amounts of fugitive dust that must be controlled. The majority of the fugitive dust at an aboveground retort is raw and spent shale fines and suspended soil material. At an in-situ retort, the majority is raw shale fines and suspended soil material.

Oil shales are enriched in some trace elements, notably As, B, Cd, Mo and Se, compared to average crustal material. They also contain rather high concentrations of silica. The next two sections summarize what is presently

TABLE A-10. SUMMARY OF UNCONTROLLED TOTAL (GASEOUS AND PARTICULATE) TRACE ELEMENT EMISSION FACTORS FOR OIL SHALE RETORTS (mg/bbl)

Element	Retort Emission Factor[1] (mg/bbl oil)
Cl	1300
Cd	0-1070
Se	>15-940
As	>13-770
Hg	39-208
Cr	<6-151
Br	>33
Mo	<1-6
Fe	<1040-<4000
Ba	<24-<292
Pb	<31
Zn	<12-<17
V	<18
Cu	<8
Mn	<6-<60
Co	<0.3-<3
U	<0.3-<2
Ni	<5
Sb	<1
Th	<1
Rn	6 uCi/bbl

(1) Based on data reported by Fruchter et. al. (1979, 1980) and Ondov, et. al. (1982), except as otherwise presented in the text (Hg, Cd, As, Se).

known about trace elements and silica in fugitive dusts produced by solid material handling at oil shale plants.

Trace Elements

Several investigations have reported the composition of fugitive dust at the Anvil Points site as operated by Paraho (Cotter et al. 1979; Cotter et al. 1978; Fruchter et al. 1979; Garcia et al. 1981; Hargis et al. 1983). Most of these studies and other company data are summarized in the Paraho data compendium prepared by the U.S. DOE (1980). All of these studies are generally consistent. The following discussion is largely based on the studies reported by Cotter et al. (1979), Fruchter et al. (1979) and U.S. DOE (1980b) because they are the most comprehensive.

At Anvil Points, retorting, crushing and retorted shale disposal generated the largest quantity of fugitive dust. The dusts had a bimodal distribution, and the majority of the particles were either >7 μm or <1.2 μm. The respirable fraction, defined as particles less than 3.3 μm in diameter, generally constituted greater than 50 percent of the fugitive dust.

The chemical characteristics of fugitive dust from several different locations at the Anvil Points site are very similar to those of raw and retorted shales. However, As, Cr, V, Sb and Co were enriched in the respirable fraction of most samples (Fruchter et al. 1979). This is probably due to the presence of thermal oxidizer dust (Table A-9). Trace elements associated with thermal oxidizer dust are accounted for in the retort emission factors.

Because shale-plant fugitive dusts from material handling are primarily raw and retorted shale fines, emission factors for this source can be reliably calculated from raw and/or retorted shale composition data. Trace element emission factors for fugitive dusts produced by material handling are summarized in Table A-11. These were estimated by multiplying the average particulate emission rate from all material handling operations (63.5 lb/1000 bbl) by estimated average dust concentrations. We estimated dust composition by multiplying the average raw shale composition in column 1 of Table A-11 by 1.3 to convert it to a retorted shale basis. This provides

TABLE A-11. SUMMARY OF CONTROLLED TRACE ELEMENT EMISSION FACTORS FOR FUGITIVE DUSTS GENERATED BY MATERIAL HANDLING OPERATIONS AT OIL SHALE FACILITES

	Mahogany Zone Raw Shale(1) (ppm or $g/10^6 g$)	Retorted Oil Shale(3) (ppm or $g/10^6 g$)	Controlled Emission Factor(4) (mg/bbl)
As	48	53	1.8
B	70	--	2.6
Ba	478	609	18
Be	2.4(2)	--	0.090
Br	4.2(2)	--	0.16
Cd	0.77	0.81	0.029
Cl	184(2)	--	6.9
Co	8.8	12.3	0.33
Cr	33	46	1.2
Cu	33	51	1.2
F	1,300	--	49
Fe	19,200	26,900	720
Hg	0.086	0.03	0.0032
Mo	24	30	0.90
Mn	289	396	11
Ni	21	31	0.79
Pb	21	30	0.79
Sb	1.9	28	0.071
Se	2.2	2.8	0.082
Th	5.7	8.4	0.21
U	4.1	6.2	0.15
V	106	136	4.0
Zn	72	87	2.7

(1) Average for 125 samples from two cores from the Naval Oil Shale Reserve, as reported by Giauque, et. al. (1980).

(2) Average of 12 Mahogany-zone composite samples as reported by Poulson et. al. (1977).

(3) Average for 25 samples of retorted shale, as reported by Fox (1980).

(4) Computed from average particulate emissions for permitted mining and retorted shale disposal emissions for six plants (63.5 lb/1000 bbl) and Mahogany Zone raw shale composition (Column 1) converted to a retorted shale basis or factor = (raw shale) (1.3) (63.5) (37.48×10^{-2} g bbl/lb 10^3 bbl).

an upper limit for dust composition and also tends to significantly overestimate volatile trace element emissions associated with particulates.

A process- and site-specific dust concentration could also be estimated from a knowledge of the fraction of the dust that is raw and retorted shale and soil. Extensive and reliable chemical composition data exist for raw shales (Giauque et al. 1980; Tuttle et al. 1983; Dean et al. 1981; Desborough et al. 1976), retorted shales (Fruchter et al. 1979; Fox 1980; U.S. DOE 1980) and soils (Dean et al. 1979) from oil shale regions and can be used for such estimates. Typical retorted shale composition data are provided in Table 11 for convenience.

Silica

Silica (quartz) in airborne dust is a respiratory health concern. Some evidence exists that prolonged exposure to silica-bearing shale oil dust can cause pneumoconiosis.

In an industrial hygiene study at the Anvil Points Mine (Hargis et al. 1983), the free silica content in 10 dust samples from the crusher and retort areas ranged from 7 to 9 percent. A single sample taken in the mine area had a silica content of 4 percent. Assuming a maximum quartz content of 10 percent, the threshold limit values for total and respirable dust are 2.5 and 0.8 $\mu g/m^3$. The majority of the dust samples in most studies have exceeded these values. However, it is important to realize that this was an experimental facility with no dust control, and workers were not exposed for long periods.

These silica concentrations correspond to an emission rate of 25 to 57 lbs/day of SiO_2 and an emission factor of 1.2 to 2.6 g/bbl of SiO_2.

APPENDIX A

REFERENCES

ACGIH, "Threshold Limit Values for Chemical Substances and Physical Agents in the Work Environment with Intended Changes for 1983-84," American Congress of Governmental Industrial Hygienists, Cincinnati, OH, 2nd Printing, 93 pp., 1983.

CRC, Handbook of Chemistry and Physics, R. C. Weast (Ed.), The Chemical Rubber Co., Cleveland, OH, 49th Ed., 1968.

Cathedral Bluffs Shale Oil Co., Prevention of Significant Deterioration, 1981.

Cotter, J. E., et al. "Sampling and Analysis Research Program at the Paraho Shale Oil Demonstration Plant," U.S. EPA Report EPA-600/7-78-065, NTIS No. PB 284027, 71 pp., 1978.

Cotter, J. E., et al. "Fugitive Dust at the Paraho Oil Shale Demonstration Retort and Mine," U.S. EPA Report EPA-600/7-79-208, NTIS No. PB80122591, 78 pp., 1979.

Dean, J. A., (Ed.) Lange's Handbook of Chemistry, McGraw-Hill Book Co., New York, 12th Ed., 1979.

Dean, W. E., et al. "Geochemical Variation in Soils in the Piceance Creek Basin," Western Colorado, Geological Survey Bulletin 1479, 47 pp., 1979.

Dean, W. E., et al. "Geochemical and Mineralogical Analysis of U.S. Geologic Survey Oil-Shale Core CR-2, Piceance Creek Basin," Western Colorado, U.S. Geological Survey Open File Report 81-596, 25 pp., 1981.

Desborough, G. A., et al. "Concentration and Mineralogical Residence of Elements in Rich Oil Shales of the Green River Formation, Piceance Creek Basin, Colorado, and the Uinta Basin, Utah -- A Preliminary Report," Chemical Geology, Vol. 17, pp. 13-26, 1976.

Donnell, J. R. and V. E. Shaw. "Mercury in Oil Shale from the Mahogany Zone of the Green River Formation, Eastern Utah and Western Colorado," Journal of Research of the U.S. Geological Survey, Vol. 5, No. 2, pp. 221-226, 1977.

Fish, R. H. "Organometallic Geochemistry, Isolation and Identification of Organoarsenic and Inorganic Arsenic Compounds from Green River Formation Oil Shale," in Geochemistry and Chemistry of Oil Shales, F. P. Miknis and J. F. McKay (Eds.), ACS Symposium Series 230, American Chemical Society, Washington D.C., pp. 423-432, 1983.

Fox, J. P. The Partitioning of Major, Minor and Trace Elements During Simulated In-situ Oil Shale Retorting. Lawrence Berkeley Laboratory Report LBL-9062, Berkeley, CA, 1980.

Fox, J. P., A. T. Hodgson, and D. C. Girvin. Trace Elements in Oil Shale Materials in Energy and Environmental Chemistry, Fossil Fuels, Vol. 1, Keith, L. H. (Ed.), Ann Arbor Science Publishers, Ann Arbor, MI, p. 69-102, 1982.

Fox, J. P. "Distribution of Mercury During Simulated In-situ Oil Shale Retorting," Environmental Science and Technology, Vol. 19, No. 4, pp. 316-322, 1985a.

Fox, J. P. Unpublished mass balance data for the Fischer assay retort, Fox Consulting, Berkeley, CA, 1985b.

Fox, J. P, et. al. "The Partitioning of As, Cd, Cu, Hg, Pb and Zn During Simulated In-situ Oil Shale Retorting," 10th Oil Shale Symposium Proceedings, Colorado School of Mines Press, Golden, CO, pp. 223-237, 1977.

Fox, J. P., et. al. "Mercury Emissions from a Simulated In-situ Oil Shale Retort," 11th Oil Shale Symposium Proceedings, Colorado School of Mines Press, Golden, CO, pp. 55-75, 1978.

Fruchter, J. S., et al. "High-Precision Trace Element and Organic Constituent Analysis of Oil Shale and Solvent-Refined Coal Materials," in Analytical Chemistry of Liquid Fuel Sources, P. C. Uden, S. Siggia, and H. B. Jensen (Eds.), ACS Advances in Chemistry Series 170, American Chemical Society, Washington, D.C., pp. 255-281, 1978.

Fruchter, J. S. et. al. Source Characterization Studies at the Paraho Semiworks Oil Shale Retort, Battelle Pacific Northwest Laboratory Report PNL-2945, Richland, WA, 70 pp., 1979.

Fruchter, J. S., et al. "Elemental Partitioning in an Above-ground Oil Shale Retort Pilot Plant," Environmental Science and Technology, Vol. 14, No. 11, pp. 1374-1381, 1980.

Fruchter, J. S. et al. "Potential Air Emissions from Oil Shale Retorting," in Oil Shale, the Environmental Challenges III, K. K. Peterson (Ed.), Colorado School of Mines Press, Golden, CO, pp. 139-164, 1983.

Garcia, L. L., H. F. Schulte, and J. J. Ettinger, J. J. "Industrial Hygiene Study at the Anvil Points Oil Shale Facility," American Industrial Hygiene Association Journal, Vol. 42, pp. 796-804, 1981.

Giauque, R. D., et al. Characterization of Two Core Holes from the Naval Oil Shale Reserve Number 1, U.S. EPA Report EPA-600/7-81-024, NTIS PB81 167736, 176 pp., 1981.

Girvin, D. C. and J. P. Fox. On-line Zeeman Atomic Absorption Spectroscopy for Mercury Analysis in Oil Shale Gases, U.S. EPA Report EPA-600/7-80-130, 95 pp., 1980.

Girvin, D. C., et al. "On-Line Measurement of Trace Elements in Oil Shale Offgases by Zeeman Atomic Absorption Spectroscopy," in Energy and Environment Division Annual Report 1979, Lawrence Berkeley Laboratory Report LBL-10486, pp. 5-29 -- 5-33, 1980.

Gladney, E. S., et al. "Composition and Size Distribution of In-stack Particulate Material at Coal-Fired Power Plant," Atmospheric Environment, Vol. 10, pp. 1071-1077, 1976.

Goodfellow, L. and M. T. Atwood. "Fischer Assay Oil Shale Procedures of Oil Shale Corporation," Quarterly of the Colorado School of Mines, Vol. 69, No. 2, pp. 205-219, 1974.

Hagihara, N. et al. Handook of Organometallic Compounds, W. A. Benjamin, Inc., New York, 1044 pp., 1968.

Hargis, K. M., et al. "Aerosol Sampling and Characterization in the Developing U.S. Oil Shale Industry, in Aerosols in the Mining and Industrial Work Environments, Vol. 2, Characterization, V. A. Marple and B. Y. H. Liu (Eds.), Ann Arbor Science Publishers, Ann Arbor, MI, pp. 481-499, 1983.

Hodgson, A. T. Unpublished data from a laboratory study of Cd partitioning in a 5.5-kg Fischer Assay-type retort, Lawrence Berkeley Laboratory, Berkeley, CA, 1985.

Hodgson, A. T., et al. Mercury Mass Distribution During Laboratory and Simulated In-situ Oil Shale Retorting, Lawrence Berkeley Laboratory Report LBL-12908, 39 pp., 1982.

Hodgson, A. T., et al. "Mercury Emissions from a Modified In-situ Oil Shale Retort", Atmospheric Environment, Vol. 18, No. 2, pp. 247-253, 1984.

Kaakinen, J., Trace Element Study in a Pulverized-Coal-Fired Power Plant, Ph.D. Dissertation, University of Colorado, Boulder, CO, 186 pp., 1974.

Klein, D. H., et al. "Pathways of Thirty-Seven Trace Elements Through Coal-Fired Power Plants," Environmental Science and Technology, Vol. 9, pp. 973-979, 1975.

Krishnan, E. R. and G. V. Hellwig. "Trace Emission from Coal and Oil Combustion," Environmental Progress, Vol. 1, No. 4, pp. 290-295, 1982.

Littlejohn, R. F. "Emission of Trace Elements from Coal-Fired Industrial Boilers. A Survey of Relevant Literature," Energy Research, Vol. 8, pp. 375-386, 1984.

Mansour, M. N. and D. G. Jones. Emission Characteristics of Paraho Shale Oil as Tested in a Utility Boiler, Electric Power Research Institute Report EPRI AF-709, 1978.

National Research Council (NRC), Health Effects of Exposure to Diesel Exhaust, National Academy Press, Washington, D.C., 169 pp., 1981.

Olsen, K. B., et al. "Partitioning and Chemical Speciation of Volatile Trace Elements During Inert Gas Oil Shale Retorting," Proceedings of First Annual Oil Shale/Tar Sands Contractors Meeting, U.S. Dept. of Energy, Morgantown, WV, 1985.

Ondov, J. M., et al. Measurements of Potential Atmospheric Pollutants in Offgases from the Lawrence Livermore National Laboratory's 6-Tonne Retort, Experiment L-3, Lawrence Livermore Laboratory Report UCRL-53265, 44 pp., 1982.

Poulson, R. E., et al. Minor Elements in Oil Shale and Oil-Shale Products, Laramie Energy Research Center Report LERC/RI-77/1, 1977.

Rinaldi, G., et al. Environmental Characterization of Geokinetics In-situ Oil Shale Retorting Technology, U.S. EPA Report EPA-600/7-81-021 (NTIS PB 81-163-727), 1981.

Shendrikar, A. D. and G. B. Faudel. "Distribution of Trace Metals During Oil Shale Retorting," Environmental Science and Technology, Vol. 12, No. 3, pp. 332-334, 1978.

Sittig, M. Toxic Metals. Pollution Control and Worker Protection, Noyes Data Corp., Park Ridge, NJ, pp. 349, 1976.

Tuttle, M. L., et al. "Inorganic Geochemistry of Mahogany Zone Oil Shale in Two Cores from the Green River Formation," in Geochemistry and Chemistry of Oil Shales, F. P. Miknis and J. F. McKay (Eds.), ACS Symposium Series 230, American Chemical Society, Washington, D.C., pp. 249-267, 1983.

Union Oil Company, Air Quality Technical Report, Parachute Creek Shale Oil Program, Phase II, Woodward-Clyde Consultants, Walnut Creek, CA, 1985.

Union Oil Company, PSD Permit Application for Union's Phase II Oil Shale Mining and Retort Facility, 1982.

U.S. DOE, Environmental Control Costs for Oil Shale Processes, U.S. Department of Energy Report DOE/EV-0055, Washington, D.C., 450 pp., 1979.

U.S. DOE, Environmental Research on a Modified In-situ Oil Shale Process," U.S. Department of Energy Report DOE/EV-0078, Washington, D.C., 94 pp., 1980a.

U.S. DOE, Paraho Environmental Data, U.S. Department of Energy Report DOE/EV-0086, Washington, D.C., 1980b.

Wildeman, T. R. and R.R. Meglen, "Analysis of Oil Shale Materials for Element Balance Studies," in Analytical Chemistry of Liquid Fuel Sources, P. C. Uden, S. Siggia, and H. B. Jensen (Eds.), ACS Advances in Chemistry Series 170, American Chemical Society, Washington, D.C., pp. 195-212, 1978.

Wong, C. M. Quantitative Analysis and Kinetics of Trace Sulfur Gas Species from Oil Shale Pyrolysis by Triple Quadrupole Mass Spectrometry, Lawrence Livermore Laboratory Report UCRL-89361, 1983.

World Health Organization (WHO), Environmental Health Criteria 1. Mercury, Geneva, 131 pp., 1976.

APPENDIX B

CODISPOSAL EMISSIONS

Codisposal is the combining of two or more waste streams for disposal. For oil shale development, it refers to the simultaneous disposal of wastewaters and low volume solid streams with the retorted shale. These materials would be blended with the retorted shale prior to disposal and/or added at the pile. The types of materials that have been considered for codisposal and their approximate quantities are summarized in Table B-1. Wastewaters and raw shale rejects together can comprise over 30 percent by weight of the total disposed material. The balance of the solids would comprise up to 1 percent by weight of the disposed material.

Among those materials proposed for codisposal (Table B-1), the only ones that presently appear to be environmentally important with respect to air emissions are the wastewaters. Much of the present controversy surrounding the issue of codisposal involves these waters and, in recent years, the term "codisposal" has been used exclusively to designate the disposal of wastewaters with the retorted shale (e.g., Persoff et al 1984; Hawthorne 1984). These waters include the process streams which may comprise over half of the water codisposed with the retorted shale.

The principal environmental concerns attributed to codisposal are its effect upon leachate composition and its contribution to air emissions. The wastewaters proposed for codisposal include process waters that contain high concentrations of organic carbon, dissolved solids and several trace elements. Many of these compounds are individually unregulated, and they include a number of substances that are malodorous and toxic. Those compounds that are volatile could be emitted directly to the atmosphere during cooling and wetting or from the pile surface, while those that are soluble may be leached by precipitation percolating through the pile.

These potential emissions and leachates may be controlled by pretreating one or more of the wastewaters to remove volatile and/or soluble materials. The design options include steam stripping and biological treatment. Various end-of-stream control technologies may also be used.

TABLE B-1. SOLID WASTE GENERATION AT A TYPICAL COMMERCIAL OIL SHALE FACILITY (Heistand 1985)

Material	Weight Percent
Retorted Shale	85
Various Wastewaters	<5 → 20
Raw Shale Rejects (1)	<2 → 10
Water Treatment Sludges (2)	0.8
Flue Gas Desulfurization Chemicals (2)	0.3
Off-Specification Byproducts (3)	0.1
Biological Sludges (2)	0.05
Oily Solids	0.04
Scrap and Garbage	0.004
Oil Upgrading Catalysts	0.003
Fuel Gas Cleanup Chemicals	<0.001

(1) Includes dusts from air pollution control devices, raw shale feed preparation fine rejects and low-grade ores.

(2) This material is primarily (about 90 wt. %) water.

(3) Primarily sulfur and coke (for projects using the TOSCO II retort).

Those that have been proposed for codisposal air emissions include Venturi scrubbers and flaring of flue gases.

B.1 CODISPOSAL OF WASTEWATERS

The balance of this section discusses the codisposal of wastewaters and retorted shale. Relatively large volumes of water are required for retorted shale disposal ranging from 5 to 25 weight percent. This water is used to cool the hot retorted shale, moisturize the retorted shale to facilitate compaction and control dust at the pile surface. Various wastewaters are employed for this purpose to minimize the use of fresh water.

The codisposal of wastewaters with retorted shale is divided into three phases, as follows:

- Phase I: Cooling of retorted shale from 500-750 C to 300 C;
- Phase II: Wetting retorted shale to optimum moisture content during which the temperature is reduced from 300 C to about 80 C; and,
- Phase III: Disposal in a landfill.

These three phases correspond to discrete periods when environmental impacts may occur. The following subsections define and describe Phase I through Phase III codisposal processes.

B.1.1 Phase I Cooling

In most projects that have been proposed, cooling and moisturization are carried out in separate process units. Following the extraction of oil, the retorted shale is at pyrolysis (500 C) or combustion (>500-750 C) temperatures, depending upon the type of process. This hot material must be cooled prior to disposal to permit safe handling and to prevent fires in the pile. This is typically done with water in indirect processes (e.g., TOSCO II, Union B) and with flowing gases (i.e., air and recycle gas) in direct processes (e.g., Paraho, Superior). Those processes that use gas

cooling are not of interest here because water is not used for cooling and codisposal is not involved until the moisturization phase.

Two types of processes for cooling hot retorted shale have been proposed in the literature. The first, the rotating drum steam generator, is used with the TOSCO II process (e.g., BLM 1975; C-b 1976; TOSCO 1982). Hot spent shale is introduced into conventional rotating drum coolers where its temperature is reduced by tumbling and by water sprays. The use and design of such coolers has been described in the literature (Perry and Chilton 1973).

The second, water immersion quenching, has been proposed for the Superior and Union B retorts, which are designed with the cooler attached to the retort to achieve a positive seal. In the Union B retort, hot retorted shale falls by gravity down chutes through a shaft cooler where it is cooled by water sprays. Steam generated during quenching and cooling is condensed, scrubbed and returned to the shaft cooler, while noncondensable vapors are recycled to the retort or flared (Duir et al 1983; UNOCAL 1986; Deering et al 1984, 1985).

Phase I cooling should not have a significant impact on the environment so long as it occurs in a closed system. Most of the water used in cooling is flashed to steam, condensed and recycled. Dust is removed by high efficiency baghouses or scrubbers. Condensible vapors stripped from the pores of the retorted shale or released when the wastewaters contact the hot retorted shale are scrubbed from the gas stream and become part of the recycled condensate; noncondensable vapors may be recycled to the retort or flared. Contaminants in the cooling waters that are not volatile under cooling conditions, primarily inorganic substances and gases trapped in retorted shale pores, may be subsequently volatilized and/or leached.

B.1.2 Phase II Moisturization

Additional water may be added to retorted shale prior to disposal in the pile. This water serves to further cool the retorted shale, typically from about 300 C to 80-90 C and to bring the codisposed solids (Table B-1) to their "optimum moisture content". This quantity of water, which may range from 7 to 25 weight percent, represents the amount of water required to

compact the material to a specified dry density with the least compactive effort.

Two processes have been proposed for moisturizing retorted shale prior to disposal. The first, rotating drum moisturization, has been proposed for the TOSCO II and Lurgi processes (RBOSC 1981; TOSCO 1982). This process operates similar to the rotating drum steam generator described above, except the steam produced is not used to generate electricity. The second, water immersion, has been proposed for use with the Union B process. In this process, the retorted shale and other solids are blended with wastewater in a pug mill (UNOCAL 1986). The gases and solids resulting from moisturization are typically routed through a Venturi scrubber in which wastewaters are used to scrub vapors and entrained solids from the flue gases.

Phase II moisturization is not utilized in all proposed commercial projects. In the Paraho-Ute Project (Paraho 1982), the majority of the retorted shale would be disposed in a relatively loose fill encased in a blanket of highly compacted, impervious retorted shale. The only water used in retorted shale disposal is added at the pile and therefore is part of the Phase III disposal process, as discussed below.

Phase II moisturization may lead to significant environmental consequences if appropriate control technology is not installed. Organic compounds associated with the wastewaters may be volatilized and emitted at the scrubber. The temperatures during Phase II moisturization (<300 C) are too low to volatilize any trace elements, with the possible exception of arsenic which has been found in volatile organic forms in process waters.

B.1.3 Phase III Disposal

The cooled retorted shale is transferred by conveyor belt or by truck to the disposal site where the solids are placed by traveling stacker (movable conveyor) or truck. Rubber-tired compaction equipment compacts the retorted shale to the specified density, typically over 85 lb/ft^3. The pile itself will be constructed in lifts or windrows of 6 to 24 inches. Each individual lift will be compacted to a specified field density, and when a portion of the pile reaches its final configuration, the surface will be reclaimed.

Additional water may be sprayed on the retorted shale during transit and as needed at the pile to replace water lost by evaporation and for dust control. Since retorted shale surface temperatures are high (65 to 90 C) on placement and evaporation rates in the region are high (>50 in/yr), up to 30 weight percent of the moisturization water may be evaporated during placement and compaction and must be replaced. Higher quality surface or ground waters, rather than wastewaters, have been proposed for this use by some developers due to potential worker exposure to wastewaters. Those projects that do not use wastewaters for retorted shale cooling and moisturization prior to disposal (e.g., Paraho 1982) may add 5 to 10 weight percent wastewater to control dust at the pile.

The retorted shale will cool during transport, placement and compaction during which time organic compounds may be volatilized. As each successive lift is added, additional material may volatilize. This volatilization process is expected to continue until the pile surface is reclaimed. The soil layer proposed by most developers is expected to sorb and trap any volatile emissions from underlying retorted shale.

B.2 CODISPOSAL AIR EMISSIONS

When wastewaters contact hot retorted shale, volatile compounds present in the waters can be emitted as gases (Phases I and II). Nonvolatile compounds are transferred to the solids and emitted with particulates. The moisturized solids continue to release vapors when disposed in the pile (Phase III) until the pile surface is reclaimed. Such emissions are not unique to oil shale operations, and the emission of organic vapors from disposal sites and soil surfaces has been widely reported (e.g., from pesticide applications, solvent disposal). The phases referred to here (I, II and III) are defined in Section B.1.

From an air emission standpoint, the compounds of greatest concern are the volatile organic ones, particularly those containing nitrogen. These compounds comprise over 95 weight percent of the emissions from codisposal, the balance (5 weight percent) arising from codisposed organics associated with particulates. Most of the trace elements and inorganic ions present in wastewaters are not volatile at the temperatures involved in codisposal. The

organic compounds, however, include many with high vapor pressures at the temperatures prevalent in codisposal (80 C-300 C), which are malodorous (e.g., alkylpyridines) and toxic (e.g., nitriles). These substances may be emitted with codisposal flue gases and are regulated as nonmethane hydrocarbons under the Clean Air Act. However, many of the individual compounds of concern are not regulated.

The nonvolatile organic compounds, which may be present on particulates from codisposal operations, include a number of mutagenic and carcinogenic compounds, including the polynuclear aromatic hydrocarbons (PNAs) and possibly nitrated and aminated PNAs. These substances could pose an inhalation hazard for workers in the vicinity of the wetting and disposal operations. Organics associated with particulates may be emitted during all three of the codisposal phases.

B.2.1 Characterization of Codisposal Air Emissions

Phase I, II and III gaseous emissions have been simulated and characterized in laboratory studies (Hawthorne 1984; Hawthorne et al 1985) and modelled (Persoff et al 1984).

B.2.1.1 Laboratory Studies. Hawthorne (1984) simulated Phase I and II emissions by heating 2 grams of retorted shale to four temperatures (80 to 450 C) and contacting it with 0.2 mL of oil shale process water. The vapors were collected in a Tenax trap and analyzed and quantitated by gas chromatography/mass spectrometry. All of the samples used in these experiments were from a run of Laramie Energy Technology Center's 150-ton simulated in situ retort.

The results of Hawthorne's Phase II simulation are summarized in Figure B-1, which plots the mass concentration of total organic carbon in micrograms of TOC per milliliter of process water (μg/mL) as a function of retorted shale temperature. Considerably more organic carbon was volatilized from the gas condensate than from the retort water. Gas condensates have a higher fraction of volatile compounds because they are condensed from the gas stream, while retort waters are aqueous extracts of the oil. Figure B-1 can be used to estimate Phase I and II codisposal emissions.

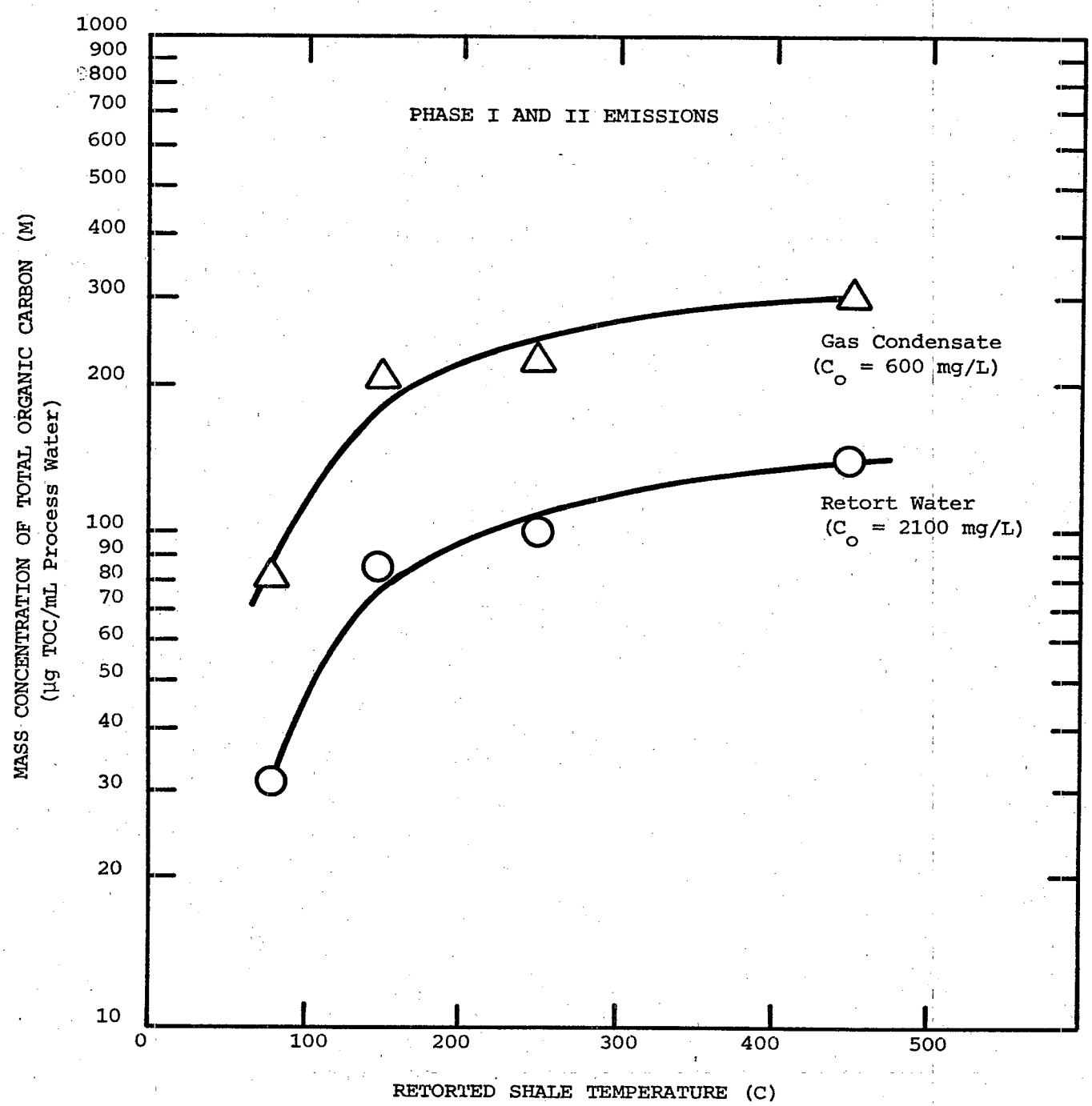


Figure B-1. Mass concentration of total organic carbon as a result of cooling and moisturizing retorted shale with process waters. (Adapted by J. Fox from Hawthorne, 1984).

The classes of compounds found by Hawthorne in these experiments are summarized in Table B-2. The most abundant classes in descending order of concentration are: pyridines, phenols, ketones, anilines, pyrroles, nitriles and quinolines. These classes contain a number of malodorous (e.g., pyridines) and toxic (e.g., nitriles, phenols) compounds. Since many of them are irritants to the eyes and upper respiratory tract (e.g., ketones, pyridines), worker exposure should be considered. Nonmethane hydrocarbons are very minor components of the emissions, contributing less than 2 percent of the emitted organic carbon. The specific compounds volatilized during Phase II wetting were essentially the same as those found in the air space above oil shale process water samples (Hawthorne and Sievers 1984; Hunter et al 1985).

Hawthorne also quantitated Phase III emissions that would occur from an 80 C pile surface during lift construction. This was simulated by measuring organic carbon emissions over time from a 14 weight percent codisposed mixture. An 80 C, 2g sample of retorted shale was contacted with 0.30 mL of process water and mixed, and emissions were quantitated as the sample cooled over a 49 hr period. The results of this simulation are plotted in Figure B-2 (solid lines), together with an extrapolation to 100 C (dashed line), based on temperature relationships from Figure B-1. Figure B-2 can be used to estimate the quantity of organic carbon emitted from the pile surface during construction.

Hawthorne found that the organic carbon emission rate following moisturization decreased logarithmically with time. At the end of 49 hours, a period equivalent to that required to construct a lift in many commercial processes, the total organic carbon emission rate was negligible, amounting to 0.014 to 0.11 μg C/mL/hr.

Thus, it may be anticipated that the emission rate from a given area of the pile will be cyclical. The emission rate will increase when each new lift is added and thereafter decrease logarithmically over time until the next lift is added, and so on. When the final pile depth is reached, the surface will be covered with top soil and reclaimed. Emissions following reclamation have not been studied but will probably be negligible due to: (1) logarithmic decay over time noted by Hawthorne (1984); and (2) sorption of vapors by

TABLE B-2. CLASSES OF ORGANIC COMPOUNDS EMITTED AS A RESULT OF COOLING AND MOISTURIZING RETORTED SHALE WITH PROCESS WATERS (μg EMITTED/mL OF PROCESS WATER) (Hawthorne et al, 1985)

	MASS CONCENTRATION (μg/mL)							
	Gas Condensate				Retort Water			
	Retorted Shale Temperature (°C)				Retorted Shale Temperature (°C)			
COMPOUND CLASS	450	250	150	80	450	250	150	80
Total alkylpyridines	242	184	157	62	88	66	70	23
Total alkylanilines	20	15	16	4.4	10	7.6	5.6	0.8
Total alkylquinolines	8.2	5.8	4.2	0.4	6.5	4.2	3.0	0.8
Total alkylpyrroles	20	16	17	8.2	7.3	5.1	3.5	1.3
Total alkylnitriles	13	11	10	4.1	14	9.6	5.3	1.4
Total alkylphenols	73	58	50	12	33	24	16	4.3
Total ketones	45	39	41	23	21	19	15	10
Total alkylthiophenes	3.7	2.3	2.0	1.3	5.5	2.2	1.0	0.6
Total alkylbenzenes	0.6	0.5	0.4	0.3	3.4	0.5	0.2	0.1
Total identified emitted compounds (M_{tot})	427	332	299	116	191	139	119	42
% of emitted compounds identified	84	87	88	85	78	79	85	82
Total emitted organic carbon	295	222	201	80	140	100	84	31

(1) Percent of emitted compounds identified was determined by dividing the total integrated area of the chromatographic peaks of the identified species by the total integrated area of all peaks in the flame ionization detector chromatogram.

(2) Total emitted organic carbon was estimated by calculating the amount of total emitted species as hydrocarbons and multiplying this result by the quantity 0.84μg of C/1.00μg of hydrocarbon.

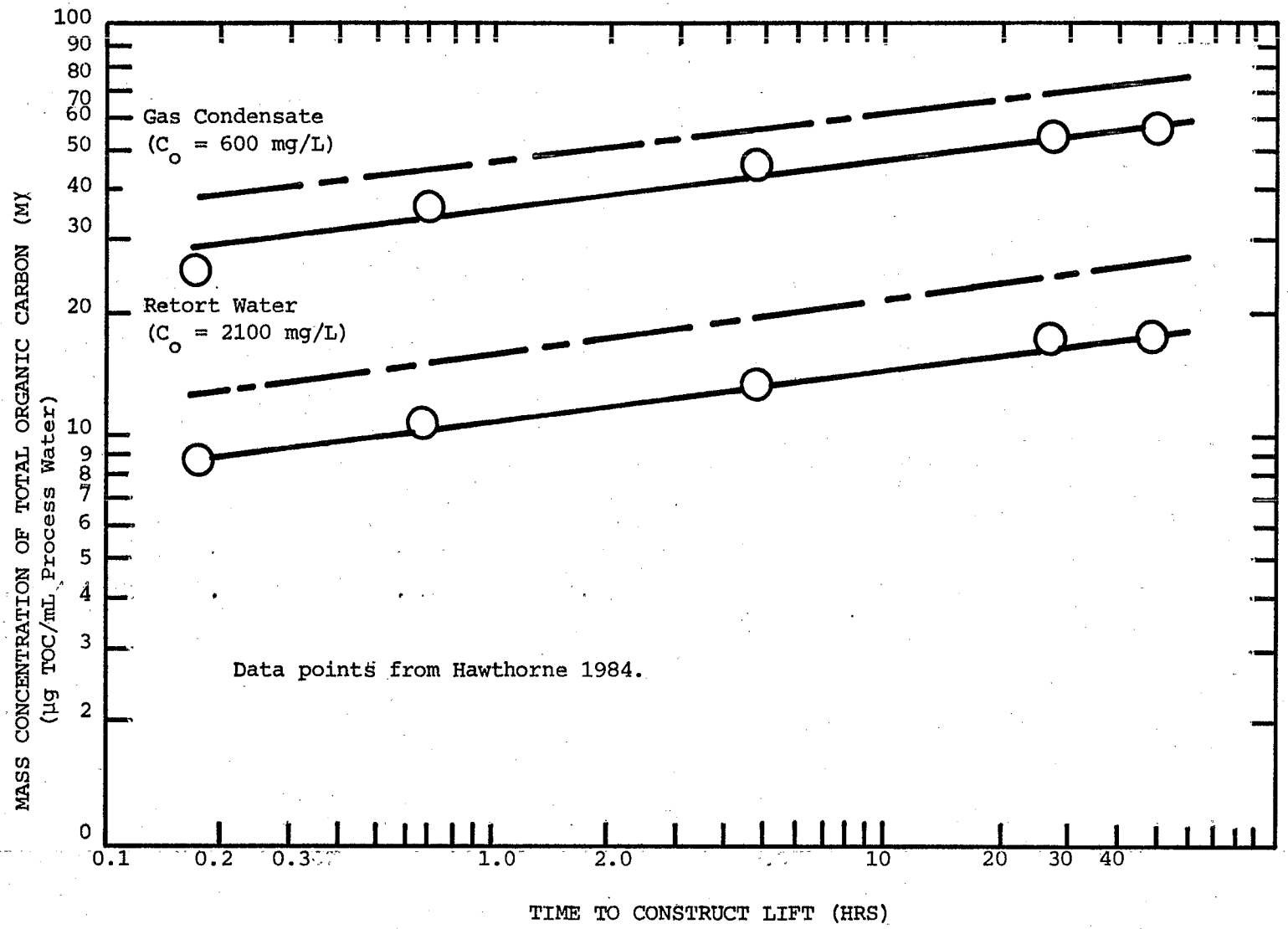


Figure B-2. Mass concentration of total organic carbon at the pile surface, subsequent to cooling and moisturizing of retorted shale with process water. (Adapted by J. Fox from Hawthorne, 1984).

organic matter in soils (e.g., Spencer et al 1982; Lokke 1984). However, microbially mediated reactions may occur within the pile, which would release byproduct gases (e.g., Wildung and Garland 1985; Hassler et al 1984).

B.2.1.2 Mathematical Model. Emissions from Phase III codisposal operations have also been estimated by mathematically modelling the major processes that occur in the near-surface pile environment (Persoff et al 1984). They divided Phase III emissions into three processes which were separately modelled, as follows: (1) expulsion of gases from internal pore space during compaction; (2) diffusion of vapors to the lift surface following compaction; and (3) emission of particulates. They used their model to estimate the emission rates for six nitrogen heterocycles from a 50,000 BPD facility in which the retorted shale was moisturized to 15 weight percent with untreated Occidental gas condensate and compacted to 85 lb/ft^3 at a temperature of 85 C (a worst-case).

The factors that determine the actual emission rate of a particular compound depend on the chemical content of the waste water, the spent shale, interaction between the water and spent shale and the partitioning of the compounds between the solid, liquid and gaseous phase.

Persoff et al (1984) used a mathematical model to calculate the partition coefficients and interstitial void volumes to estimate daily emission rates of pyridine, methyl dimethly and trimethly pyridines, quinolines, and pyrroles.

Many simplifying assumptions were made and research should be undertaken to develop actual data. This analysis was not intended to determine all singularly accurate values but rather to illustrate the framework of the interpretation of the laboratory and field data as they become available. The assumptions used are summarized in Table B-3.

A. Removal of Volatile Solutes by Absorption on Spent Shale

Nitrogen compounds can be removed from the waste water by absorption onto the spent shale which would decrease its concentration and eventual emission rate. An evaluation was made of the absorption potential of collidine (2,4,6-trimethylpyridine). A simulated in-situ spent shale from the

TABLE B-3. BASIC ASSUMPTIONS USED FOR PHASE II EMISSION CALCULATIONS (Persoff et al, 1984)

Production rates:	oil: 50,000 bbl/day spent shale: 1.4 ton/bbl oil
Moisturization:	15% moisture (i.e., 15g water/100g spent shale); moisturizing water is Oxy-6 condensate
Compacted density:	85 lb/ft^3 (1.36 metric $tons/m^3$)
Height of compacted lift:	0.5m (1 lift/day)
Active disposal area:	93,000 m^2
No sorption of solutes to solid phase.	
Disposal shale temperature:	85 C

Laramie Research Center (LRC) 10-ton retort was used. The primary data can be fit by a Langmuir isotherm:

$$C/Q = (1/A\ B) + (C/A)$$

where:

C = equilibrium phase concentration of collidine (μg/ml)

Q = equilibrum solid phase concentration of collidine (μg/g)

A = sorption capacity (μg/g)

B = constant (ml/μg)

Using values for A and B of 0.274 μ/g and 0.029 ml/g, respectively, for concentrations of collidine less than 100 μg/ml, the degree of absorption that would occur would be less than 1.4 μg/ml. Therefore, removal of collidine by absorption would be minimal. Similar data analyses are needed for the other compounds of interest at the relevant temperatures. However, for this analysis, the removal of volatile solutes by absorption on spent shale is considered negligible.

B. Emission of Compounds During Compaction: Expulsion of Gas From Pore Space

Estimation of Gas Composition in Pore Space--

To estimate emission by expulsion of inertial air, both the composition of the expelled air and its volume must be determined. The expelled air will have equilibrated with defined masses of solids spent shale and moisturized water. Because solute removal by absorption would apparently be minimal, as discussed above, only equilibration of air and moisturizing water need be considered. Partial pressure that would be exerted by each constituent in the gas phase would be equal to the product of its mole fraction in the liquid phase, its liquid phase activity coefficient, and the vapor pressure of the pure compound. These values for the heterocyclic nitrogen compounds from an Oxy-6 gas condensate and the resulting head space composition are shown in Table B-4.

TABLE B-4. NITROGENOUS HETEROCYCLES IN OXY-6 GAS CONDENSATE AND ESTIMATED HEADSPACE COMPOSITION (Persoff et al, 1984)

Compound	Mole Fraction x 10^6 [a]	Activity Coefficient [b]	Vapor Pressure of Pure Compound (85 C)	Mass Concentration in Headspace [c] mg/m^3
Pyridine	3.35	12.29	294	56
Methylpyridines	4.25	64.99	183	276
Dimethylpyridines	7.91	358.79	114	2036
2-, 4-, 6-Trimethyl-pyridines	10.41	2054.33	71	10806
Quinolines	1.64	1918.22	<4	83
Pyrroles	1.32	12.29 [d]	183	12

[a] Based on pooling all isomers.

[b] Activity coefficient calculated by UNIFAC.

[c] Mass concentration at standard temperature and pressure (i.e., 0 C and 1 atm).

[d] Activity coefficient assumed to be same as pyridine.

Volume of Gas Expelled During Compaction--

Once the compositon of the gas in the pore spaces has been calculated (for this example, head space over pure waste water was assumed) the calculation of emissions by expulsion of inertial air requires a value for the volume expelled. The volume of gas expelled can be estimated directly from the bulk density of spent shale during emplacement and after compaction (0.865 and 1.36 metric tons/m^3, respectively) and from the solid density of the spent shale (2.5 metric tons/m^3). Using these values, it can be shown that 0.356 m^3 of inertial gas is expelled during compaction of the spent shale generated from production of one barrel of oil.

C. Emission of Compound After Compaction: Diffusion to Surface of Lift

After compaction it can be assumed that the pile will remain hot for 24 hours until the next lift is in place. Although each lift will cool by radiation and evaporation of water, such cooling is limited because the shale is placed on a previously compacted lift which is still above ambient temperature. Spent shale that is black (i.e., contains residual char) will gain heat from solar radiation. Surface temperatures of 65 C have been observed on sunny summer days for black spent shale that has been previously been cooled to ambient temperature. Thus, 65 C can be taken as the lower temperature limit for cooling a spent shale on hot days. It is approriate to consider hot weather conditions for codisposal scenarios studies since air pollution management is generally directed at limiting severe episodes rather than annual average concentrations.

For the diffusion of compounds (e.g., nitrogenous heterocycles), the following assumptions were made: (i) the surface of the disposal pile is a zero-concentration boundary; (ii) the concentration of nitrogenous heterocycles is uniform throughout the pile immediately after compaction; (iii) and the bottom of the lift is a zero-flux boundary.

An analytical solution based on on a three-phase matrix to determine an "effective diffusivity" was used to estimate the fraction of each solute that would diffuse to the surface of the pile during 24 hours. The results of these calculations are summarized in Table B-5.

TABLE B-5. ESTIMATED LOSS OF NITROGENOUS HETEROCYCLES VIA DIFFUSION (Persoff et al, 1984)

Compound	Diffusivity $m^2/sec \times 10^{-5}$			Henry's Constant[b]	Fraction Lost From Disposal Pile in 1 Day[a]
	Gas Phase	Liquid Phase	Effective[a]		
Pyridine	1.36	0.000321	0.00461	4.75	0.148
Methylpyridines	1.21	0.000283	0.0121	14.54	0.231
Dimethylpyridines	1.10	0.000255	0.0325	44.85	0.377
2-, 4-, 6-Trimethyl-pyridines	1.01	0.000233	0.0813	135.15	0.590
Quinolines	1.02	0.000247	0.00734	10.35	0.181
Pyrroles	1.47	0.000355	0.00281	2.59	0.115

[a] Method of calculation shown in Persoff et al (1984)

[b] Henry's constant = (vapor pressure of pure compound)(activity coefficient)/(total pressure)

Two important effects are not considered in this analysis. The first is evaporation of water from the pile. Considerable water loss may be expected at 85 C. Evaporation of water will cool a pile while also making the remaining liquid phase more concentrated. The second effect that also contributes to considerable volatilization is "wicking". As the moisture content in the upper region of the pile decreases relative to he lower region of the pile, the resulting moisture potential gradient causes moisture to move upward toward the surface. This results from capillarity. This flow of water within the pile causes an advective transport to the surface in addition to the diffusive transport modeled.

The calculated emission rates for nitrogeneous heterocycles that would result if spent shale were moisturized with Oxy-6 is condensate (undiluted and untreated except for ammonia removal) are summarized in Table B-6. Emissions from particulates were calculated directly using the value of 2 grams of fugitive dust per barrel of oil (Crawford et al 1977) and assuming that the dust contained 15 percent moisture which in turn had the concentrations of nitrogeneous heterocycles are shown in Table B-6. The emissions from the particulates are minimal when compared with those from diffusion and expulsion.

B.2.1.3 Comparison of Facility Emissions. Persoff's estimates of Phase III gaseous emissions are compared in Table B-6 with those calculated from Hawthorne's experimental data using the procedures detailed below. Overall, Persoff's model predicts TOC emission rates that are about a factor of two higher than those measured by Hawthorne (1984). Individual compound and class emission rates estimated from these two methods deviate by as much as a factor of seven. When one realizes that Persoff's model was not calibrated or validated with authentic data and that all variables were estimated from procedures such as those presented in Lyman et al (1982), it is truly remarkable that the two sets of estimates are as close as they are.

The cause for the discrepancy between Hawthorne's laboratory simulation and Persoff's model predictions is not known precisely, but inaccurate input data is certainly an important factor. In general, it is believed that both methods tend to overestimate Phase III emissions. Hawthorne's estimates are probably high because he did not compact his

TABLE B-6. COMPARISON OF PHASE III CODISPOSAL GASEOUS EMISSION RATES CALCULATED BY TWO INDEPENDENT METHODS

COMPOUND OR CLASS	MASS EMISSION RATE (lb/hr) hg/1000 m^3 of oil		
	Hawthorne (1984)	Persoff et al (1984)	Ratio (P/H)
Pyridine	7.0	2.2	0.3
Methylpyridines	10.3	5.2	0.5
Dimethylpyridines	12.5	20.2	1.6
2-, 4-, 6-Trimethylpyridines	11.0	51.0	4.6
Quinolines	.3	1.9	7.0
Pyrroles	2.9	.6	0.2
TOTAL	43.9	81.0	1.8

(1) Both sets of estimates were calculated for a 50,000 BPD facility in which the retorted shale is moisturized to 15 weight percent with a gas condensate and compacted to 85 lb/ft^3 at a temperature of 85 C.

samples, and he continuously flushed them with air during the experiment, both of which would tend to increase emissions compared to the field case.

Persoff's estimates are also probably high because his model did not consider evaporation, which occurred in Hawthorne's simulation and which is significant in the arid high mountain desert terrain of oil shale regions. Free-water surface evaporation in many areas exceed 50 in/yr. This evaporation will cool the pile surface, reducing emissions that strongly depend on temperature. Gas expulsion may also not be as significant as suggested by Persoff's model, which assumed that all of the interstitial gases was expelled. In fact, gases may be trapped within internal pores, and only those pores near the surface of a lift may contribute. Although Persoff's model did not consider adsorption, there is presently no evidence that it is important for nitrogen heterocycles at the pH and temperatures common in codisposal (e.g., Zachara et al 1986; McGowan and Sorini 1985; Routson and Li 1980).

B.2.2 Summary

Codisposal of process waters is estimated to release about 188 kg/1000 m^3 of oil of organic vapors to the atmosphere. This is greater than the average facility (all sources) hydrocarbon emission rate of 160 kg/1000 m^3 of oil calculated from seven oil shale PSD applications. Codisposal of the waters would contribute an additional 5.7 kg/1000 m^3 of oil of organic material to normal particulate emissions, which average 340 kg/1000 m^3 of oil for the same seven projects (Taback and Goldstick 1986).

APPENDIX B

REFERENCES

Anonymous. Shale Oil Project (C-b), Detailed Development Plan and Related Materials, Vol. 1, Submitted to Area Oil Shale Supervisor, Grand Junction, CO (1976).

Bureau of Land Management (BLM), Final Environmental Statement, Proposed Development of Oil Shale Resources by the Colony Development Operation in Colorado, U.S. Department of Interior (1975).

Crawford, K. W., C. H. Prien, L. B. Baboolal, C. C., Shih, A. A., Lee. "A Preliminary Assessment of the Environmental Impacts from Oil Shale Developments." EPA/600/7-77-069, NTIS PB-272283, 1977.

Deering, R. F., R. O. Dhondt, and J. E. Hines. "Process for Cooling, Depressurizing and Moisturizing Retorted Oil Shale," U.S. Patent 4,461,673 (1984).

Duir, J. H., C. G. Griswold, and B. A. Christolini. "Oil Shale Retorting Technology," CEP, February 1983, p. 45-50

Hassler, R. A., D. A. Klein, and R. R. Meglen, R. R. "Microbial Contributions to Soluble and Volatile Arsenic Dynamics in Retorted Oil Shale," J. Environ. Qual., Vol. 13, No. 3, 1984. p. 466.

Hawthorne, S. B. "The Emission of Organic Compounds from Shale Oil Wastewaters," Ph.D. Thesis, Department of Chemistry, University of Colorado, Boulder, CO, 1984.

Hawthorne, S. B. and R. E. Sievers. "Emission of Organic Air Pollutants from Shale Oil Wastewaters," Environ. Sci. Technol., Vol. 18, No. 6, 1984. p. 992.

Hawthorne, S. B., R. E. Sievers, and R. M. Barkley. "Organic Emissions from Shale Oil Wastewaters and their Implications for Air Quality," Environ. Sci. Technol., Vol. 19, No. 10, 1985. p. 992.

Heistand, R. N. "Estimating Solid Wastes from Oil Shale Facilities," Proceedings of the Eighteenth Oil Shale Symposium, Colorado School of Mines, Golden, Colorado, 1985. p. 291.

Hunter, L., P. Persoff, P. and C. G. Daughton. "Identification and Correlation of Volatile Components in Oil Shale Retort Wastewaters," in Chemistry for Protection of the Environment, Leuven, Belgium, September 10-13, 1985.

Lokke, H. "Sorption of Selected Organic Pollutants in Danish Soils," Ecotox. Environ. Safety, Vol. 8, 1984. p. 395.

Lyman, W. J., W. F. Reehl, and D. H. Rosenblatt, D. H. (Eds.) "Handbook of Chemical Property Estimation Methods," McGraw-Hill Book Company, New York, 1982).

McGowan, L. J. and S. S. Sorini. "The Effect of Residual Carbon on Adsorption of Organic Compounds by Retorted Oil Shale," in Proceedings of the Eighteenth Oil Shale Symposium, Colorado School of Mines Press, Golden, CO, 1985. p. 317.

Paraho Development Corp., Paraho-Ute Project Technical Report, prepared by Paraho Development Corp. for Bureau of Land Management in conjunction with the Uintah Basin Synfuels Environmental Impact Statement, 1982.

Perry, R. H. and C. H. Chilton. "Chemical Engineers' Handbook, 5th Edition," McGraw-Hill Book Company, New York, 1973.

Persoff, P., L. Hunter, L. and C. G. Daughton. "Atmospheric Emissions from Codisposed Oil Shale Wastes, A Preliminary Assessment," Lawrence Berkeley Laboratory Report, LBID-890, 1984.

Rio Blanco Oil Shale Company, Modification to the Detailed Development Plant, Lurgi Demonstration Project, Tract C-a, submitted to the USGS Deputy Conservation Manager-Oil Shale, 1981.

Routson, R. C. and S. W. Li. "Collidine Sorption on a Silt Loam Soil and a Spent Shale," Soil Sci., Vol. 130, No. 5, 1980. p. 233.

Spencer, W. F., W. J. Farmer, and W. A. Jury. "Review: Behavior of Organic Chemicals at Soil, Air, Water Interfaces as Related to Predicting the Transport and Volatilization of Organic Pollutants," Environ. Toxic. Chem., Vol. 1, 1982. p. 17.

Tosco Development Corporation (TOSCO), Project Description, Technical Report, San Wash Shale Oil Project, Uintah County, UT, Report prepared for Utah Bureau of Land Management, 1982.

UNOCAL, Draft Environmental Monitoring Plan, Phase I Project (Unishale B), Vols. I and II, prepared by UNOCAL Energy Mining Division for Department of Treasury, May 1986.

Wildung, R. E. and T. R. Garland. "Microbial Development on Oil Shale Wastes: Influence on Geochemistry in Soil Reclamation Processes, Microbiological Analyses and Applications," R. L. Tate III and D. A. Klein (Eds.), Marcel Dekker, Inc., New York, 1985. p. 107.

Zachara, et al. "Quinoline Sorption to Subsurface Materials: Role of pH and Retention of the Organic Cation," Environ. Sci. Technol., Vol. 20, No. 6, 1986. pp. 620-627.

APPENDIX C

SUMMARY OF PSD PERMIT APPLICATION EMISSIONS

These are the data reported in the applications for Prevention of Significant Deterioration (PSD) Permits. The data are presented in six tables as follows:

Table C.1 - Gaseous Emissions

The data are sorted by (1) Project, (2) Pollutant and (3) General Process. This sort provides the total emissions for each project; i.e., total emissions of CO from Cathedral Bluffs. The information is used to determine and compare total facility emissions.

Table C.2 - Particulate Emissions

The data are sorted by (1) Project and (2) General Process. This sort provides the total particulate emissions for each project. The information is used to determine and compare total facility emissions.

Table C.3 - Gaseous Emissions

The data are sorted by (1) Pollutant, (2) Process and (3) Project. This sort provides emissions for each process within the project. The information is used to determine and compare emission rates for each general process category (mining, retort, etc.) within each facility; i.e., NO_x emissions from mining process for each project.

Table C.4 - Particulate Emissions

The data are sorted by (1) General Process and (2) Project. This sort provides particulate emissions for each process within the project. The information is used to determine and compare emission rate for each general process category (mining, retort, etc.) within each facility.

Table C.5 - Gaseous Emissions

The data are sorted by (1) Pollutant, (2) Specific Process and (3) Project. This sort provides emissions for each process. The information is used to determine and compare average value for specific processes; i.e., CO emissions from blasting and particulate emissions from underground vehicles.

Table C.6 - Particulate Emissions

This data is sorted by (1) Specific Process and (2) Project. This sort provides emissions for each specific process. The information is used to determine and compare average values for specific processes.

Each category is used for the data base sort routines and is described below.

C.1 PROJECT

The projects included in the data base are:

Project Name	Process Description
Cathedral Bluffs	Oxydental MIS/Lurgi modified in-situ with above ground Lurgi solids recycle retort
Clearcreek Oil Shale	Above ground Chevron - Fluidized Bed Retort and spent shale combustion with solids recycle
Cottonwood	Above ground circular grate retort with fluidized bed combustion of spent shale and fines
Paraho	Above ground Direct Heat Retort
Union B	Above ground Indirect Heat - Gas Recycle
Syntana	Above ground T 3 Process - direct heat - semi-batch
White River	Above ground circular grate (Superior) with Union B and Tosco II

C.2 POLLUTANT

The pollutants included are:

Carbon Monoxide (CO)
Hydrocarbons (C)
Nitrogen Oxides (NO_x)
Sulfur Oxides (SO_x)
Particulates (PM)

C.3 GENERAL PROCESS

The General Process categories:

Category Description

1. mining
2. retort
3. gas utilization
4. upgrading, storage

C.4 SPECIFIC CATEGORIES

The Specific Categories are described in Table 4-1 in the text and indicated in the various tables.

TABLE C-1. GASEOUS EMISSIONS

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
Catherdral Bluffs	mining-below	mine shaft vents		CO	.30	8	1	158.37	63.86
Catherdral Bluffs	utility	steam boilers		CO	.05	37	3	23.78	9.59
Catherdral Bluffs	retort	flares		CO	.00	38	3	.18	.07
Catherdral Bluffs	retort	sponge oil reboiler		CO	.01	40	3	3.33	1.34
Catherdral Bluffs	retort	recycle gas heater		CO	.05	40	3	25.21	10.16
Catherdral Bluffs	upgrade	incinerator		CO	.00	43	4	.67	.27
Catherdral Bluffs	upgrade	H2 Recycle Heater		CO	.00	43	4	.90	.36
Catherdral Bluffs	upgrade	H2 Charge Heater		CO	.01	43	4	2.85	1.15
Catherdral Bluffs	upgrade	oil charge heater		CO	.01	43	4	4.66	1.88
Catherdral Bluffs	upgrade	reformer		CO	.05	43	4	28.06	11.31
Catherdral Bluffs	mining-below	mine shaft vents		HC	.08	8	1	39.47	42.09
Catherdral Bluffs	utility	steam boilers		HC	.08	37	3	42.80	45.64
Catherdral Bluffs	retort	flares		HC		38	3		
Catherdral Bluffs	retort	sponge oil reboiler		HC	.00	40	3	.57	.61
Catherdral Bluffs	retort	recycle gas heater		HC	.01	40	3	4.52	4.82
Catherdral Bluffs	upgrade	H2 Recycle Heater		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	incinerator		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	H2 Charge Heater		HC	.00	43	4	.57	.61
Catherdral Bluffs	upgrade	oil charge heater		HC	.00	43	4	.81	.86
Catherdral Bluffs	upgrade	reformer		HC	.01	43	4	4.76	5.07
Catherdral Bluffs	upgrade	storage		HC		43	4		
Catherdral Bluffs	mining-below	mine shaft vents		NOx	.91	8	1	475.12	15.00
Catherdral Bluffs	utility	steam boilers		NOx	1.89	37	3	989.24	31.24
Catherdral Bluffs	retort	flares		NOx	.00	38	3	2.62	.08
Catherdral Bluffs	retort	sponge oil reboiler		NOx	.09	40	3	48.04	1.52
Catherdral Bluffs	retort	recycle gas heater		NOx	.70	40	3	369.06	11.65
Catherdral Bluffs	upgrade	incinerator		NOx	.02	43	4	9.04	.29
Catherdral Bluffs	upgrade	H2 Recycle Heater		NOx	.02	43	4	12.84	.41
Catherdral Bluffs	upgrade	H2 Charge Heater		NOx	.07	43	4	39.00	1.23
Catherdral Bluffs	upgrade	oil charge heater		NOx	.12	43	4	62.78	1.98
Catherdral Bluffs	upgrade	reformer		NOx	2.21	43	4	1159.03	36.60
Catherdral Bluffs	mining-below	mine shaft vents		SOx	.06	8	1	29.96	4.19
Catherdral Bluffs	utility	steam boilers		SOx	.28	37	3	147.44	20.64
Catherdral Bluffs	retort	flares		SOx	.02	38	3	8.09	1.13
Catherdral Bluffs	retort	sponge oil reboiler		SOx	.04	40	3	20.45	2.86
Catherdral Bluffs	retort	recycle gas heater		SOx	.30	40	3	156.95	21.97
Catherdral Bluffs	upgrade	H2 Recycle Heater		SOx	.01	43	4	5.23	.73
Catherdral Bluffs	upgrade	H2 Charge Heater		SOx	.03	43	4	17.60	2.46
Catherdral Bluffs	upgrade	oil charge heater		SOx	.05	43	4	28.54	3.99
Catherdral Bluffs	upgrade	incinerator		SOx	.24	43	4	126.98	17.78
Catherdral Bluffs	upgrade	reformer		SOx	.33	43	4	173.12	24.23

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-above	blasting		CO	.05	2	1	2.97	.03
Clear Creek Shale Oil	mining-below	blasting		CO	.36	2	1	22.66	.20
Clear Creek Shale Oil	mining-below	crushing/screening	primary	CO	.72	6	1	45.03	.40
Clear Creek Shale Oil	mining-below	vent		CO	1.23	8	1	77.62	.69
Clear Creek Shale Oil	mining-above	top soil	removal	CO	.03	9	2	1.83	.02
Clear Creek Shale Oil	mining-above	top soil	load	CO	.00	10	2	.17	.00
Clear Creek Shale Oil	mining-above	raw shale haul		CO	.03	16	2	2.11	.02
Clear Creek Shale Oil	mining-above	spent shale haul		CO	.06	32	2	3.94	.03
Clear Creek Shale Oil	utility	steam superheat		CO	.58	37	3	36.41	.32
Clear Creek Shale Oil	retort	char combustion		CO	176.04	40	3	11071.92	98.26
Clear Creek Shale Oil	retort	char combustion	coal gri	CO	.05	40	3	2.97	.03
Clear Creek Shale Oil	retort	TEG Concentrator		CO	.00	43	4	.11	.00
Clear Creek Shale Oil	mining-above	vehicles	road mai	CO	.01	53	4	.70	.01
Clear Creek Shale Oil	mining-below	crushing/screening	primary	HC	.11	6	1	7.19	3.08
Clear Creek Shale Oil	mining-below	vent		HC	.17	8	1	10.96	4.70
Clear Creek Shale Oil	mining-above	top soil	removal	HC	.01	9	2	.57	.24
Clear Creek Shale Oil	mining-above	raw shale haul		HC	.01	16	2	.68	.29
Clear Creek Shale Oil	mining-above	spent shale haul		HC	.01	32	2	.51	.22
Clear Creek Shale Oil	upgrade	fugative		HC	3.26	44	4	204.89	87.80
Clear Creek Shale Oil	upgrade	storage		HC	.13	46	4	7.99	3.42
Clear Creek Shale Oil	mining-above	vehicles	road mai	HC	.01	53	4	.57	.24
Clear Creek Shale Oil	mining-above	blasting		NOx	.02	2	1	1.43	.02
Clear Creek Shale Oil	mining-below	blasting		NOx	.03	2	1	1.71	.03
Clear Creek Shale Oil	mining-below	crushing/screening	primary	NOx	.75	6	1	47.03	.82
Clear Creek Shale Oil	mining-below	vent		NOx	1.30	8	1	81.61	1.42
Clear Creek Shale Oil	mining-above	top soil	removal	NOx	.20	9	2	12.44	.22
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.00	10	2	.06	.00
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.01	10	2	.74	.01
Clear Creek Shale Oil	mining-above	raw shale haul		NOx	.19	16	2	11.99	.21
Clear Creek Shale Oil	mining-above	spent shale haul		NOx	.14	32	2	8.73	.15
Clear Creek Shale Oil	utility	steam superheat		NOx	1.87	37	3	117.57	2.04
Clear Creek Shale Oil	retort	char combustion		NOx	86.90	40	3	5465.19	94.85
Clear Creek Shale Oil	retort	char combustion	coal gri	NOx	.15	40	3	9.53	.17
Clear Creek Shale Oil	retort	TEG Concentrator		NOx	.00	43	4	.29	.00
Clear Creek Shale Oil	mining-above	vehicles	road mai	NOx	.05	53	4	3.42	.06
Clear Creek Shale Oil	mining-below	crushing/screening	primary	SOx	.07	6	1	4.34	.78
Clear Creek Shale Oil	mining-below	vent		SOx	.12	8	1	7.48	1.35
Clear Creek Shale Oil	mining-above	top soil	removal	SOx	.01	9	2	.86	.15

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-above	top soil	load	SOx	.00	10	2	.06	.01
Clear Creek Shale Oil	mining-above	raw shale haul		SOx	.01	16	2	.68	.12
Clear Creek Shale Oil	mining-above	spent shale haul		SOx	.01	32	2	.51	.09
Clear Creek Shale Oil	utility	steam superheater-sou		SOx	.49	37	3	30.82	5.55
Clear Creek Shale Oil	retort	char combustion	coal gri	SOx	.05	40	3	2.97	.53
Clear Creek Shale Oil	retort	char combustion		SOx	8.06	40	3	506.80	91.33
Clear Creek Shale Oil	retort	TEG Concentrator		SOx	.00	43	4	.11	.02
Clear Creek Shale Oil	mining-above	vehicles	road mai	SOx	.00	53	4	.29	.05
Paraho-Ute	mining-below ground	blasting		CO	.52	2	1	77.39	29.81
Paraho-Ute	mining-below ground	mobile equipment		CO	.67	8	1	99.67	38.39
Paraho-Ute	upgrade	package boiler		CO	.01	37	3	.83	.32
Paraho-Ute	retort	power generation		CO	.39	37	3	58.57	22.56
Paraho-Ute	upgrading	hydrotreater feed fur		CO	.01	43	4	.98	.38
Paraho-Ute	upgrading	reformer furnace		CO	.15	43	4	22.18	8.54
Paraho-Ute	mining-below ground	mobile equipment		HC	.21	8	1	30.98	63.26
Paraho-Ute	upgrade	package boiler		HC	.00	37	3	.16	.33
Paraho-Ute	retort	power generation		HC	.07	37	3	10.26	20.95
Paraho-Ute	upgrading	hydrotreater feed fur		HC	.00	43	4	.16	.33
Paraho-Ute	upgrading	reformer furnace		HC	.03	43	4	3.91	7.99
Paraho-Ute	upgrading	storage	crude sh	HC	.01	46	4	.82	1.66
Paraho-Ute	upgrading	storage (day)	crude sh	HC	.01	46	4	1.22	2.48
Paraho-Ute	upgrading	storage	hydrotre	HC	.01	47	4	1.05	2.15
Paraho-Ute	upgrading	storage	fuel oil	HC	.00	48	4	.04	.08
Paraho-Ute	upgrading	storage	diesel &	HC	.00	49	4	.37	.75
Paraho-Ute	mining-below ground	blasting		NOx		2	1		
Paraho-Ute	mining-below ground	mobile equipment		NOx	3.17	8	1	473.97	27.44
Paraho-Ute	upgrade	package boiler		NOx	.02	37	3	3.74	.22
Paraho-Ute	retort	power generation		NOx	6.25	37	3	936.25	54.21
Paraho-Ute	upgrading	hydrotreater feed fur		NOX	.09	43	4	13.18	.76
Paraho-Ute	upgrading	reformer furnace		NOX	2.00	43	4	300.03	17.37
Paraho-Ute	mining-below ground	mobile equipment		SOx	.23	8	1	34.31	5.25
Paraho-Ute	upgrade	package boiler		SOx	.05	37	3	7.07	1.08
Paraho-Ute	retort	power generation		SOx	3.97	37	3	595.18	91.11
Paraho-Ute	upgrading	hydrotreater feed fur		SOx	.00	43	4	.03	.00
Paraho-Ute	upgrading	reformer furnace		SOx	.11	43	4	16.65	2.55
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
syntana	mining-below	vehicles		CO	.57	8	1	62.58	34.03
syntana	retort	FGD	steam bo	CO	.15	37	3	16.14	8.78
syntana	retort	FGD	superior	CO	.09	40	3	9.51	5.17
syntana	retort	Tosco ball heater & 1		CO	.05	40	3	5.17	2.81
syntana	upgrade	FGD	hydrotre	CO	.01	43	4	1.53	.83

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
syntana	upgrade	FGD	hydrogen	CO	.16	43	4	17.87	9.72
syntana	upgrade	hydrogen	furnace	CO	.23	43	4	25.33	13.78
syntana	mining-below	vehicles		HC	.18	8	1	19.42	13.10
syntana	retort	FGD	steam bo	HC	.03	37	3	2.83	1.91
syntana	retort	FGD	superior	HC	.02	40	3	1.79	1.21
syntana	retort	Tosco ball heater & 1		HC	.51	40	3	55.87	37.67
syntana	upgrade	FGD	hydrotre	HC	.00	43	4	.29	.20
syntana	upgrade	FGD	hydrogen	HC	.03	43	4	3.20	2.16
syntana	upgrade	hydrogen	furnace	HC	.04	43	4	4.71	3.17
syntana	upgrade	oil storage tanks		HC	.03	46	4	3.12	2.11
syntana	upgrade	fugitive emissions		HC	.52	54	4	57.07	38.48
syntana	mining-below	blasting		NOx	.23	2	1	25.03	2.17
syntana	mining-below	vehicles		NOx	2.70	8	1	297.37	25.79
syntana	retort	FGD	steam bo	NOx					
syntana	utility	steam		NOx		37	3		
syntana	retort	retort indirect heate		NOx		39	3		
syntana	retort	FGD	superior	NOx	.99	40	3	108.80	9.44
syntana	retort	Tosco ball heater & 1		NOx	.57	40	3	62.58	5.43
syntana	upgrade	FGD	hydrotre	NOx	.16	43	4	17.77	1.54
syntana	upgrade	FGD	hydrogen	NOx	1.90	43	4	210.16	18.23
syntana	upgrade	hydrogen	furnace	NOx	2.20	43	4	242.30	21.02
syntana	mining-below	vehicles		SOx	.20	8	1	21.53	3.74
syntana	retort	FGD	steam bo	SOx	.91	37	3	100.13	17.38
syntana	utility	steam		SOx		37	3		
syntana	retort	claus plant		SOx		39	3		
syntana	retort	FGD	claus pl	SOx	.51	39	3	56.27	9.77
syntana	retort	retort indirect heate		SOx		39	3		
syntana	retort	FGD	superior	SOx	.53	40	3	57.97	10.06
syntana	retort	Tosco ball heater & 1		SOx	.40	40	3	44.46	7.72
syntana	upgrade	FGD	hydrotre	SOx	.09	43	4	9.51	1.65
Union	mining-below	drilling		CO	.00	1	1	.09	.02
Union	mining-below	blasting		CO	1.14	2	1	79.71	16.09
Union	mining-below	raw shale removal/sca	engines/	CO	.55	8	1	38.75	7.82
Union	mining-above	raw shale	topsoil	CO	.01	11	2	.43	.09
Union	mining-above	spent shale	hauling	CO	.05	32	2	3.77	.76
Union	mining-above	spent shale	grooming	CO	.01	36	2	.50	.10
Union	utility	steam		CO	2.66	37	3	185.67	37.49
Union	comb.-retort	sponge oil stripper		CO	.03	40	3	2.44	.49
Union	retort	gas recycle heater		CO	.33	40	3	23.13	4.67
Union	upgrade	fractionater	reboiler	CO	.01	43	4	.38	.08
Union	upgrade	dearseniter	purge he	CO	.00	43	4	.32	.06

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m 3 oil	% total
Union	upgrade	dearseniter	charge h	CO	.02	43	4	1.65	.33
Union	upgrade	unicracker	charge h	CO	.03	43	4	2.09	.42
Union	upgrade	steam boiler	no air p	CO	.51	43	4	35.38	7.14
Union	upgrade	steam boiler	with air	CO	.51	43	4	35.38	7.14
Union	upgrade	reformer furnace		CO	.26	43	4	18.01	3.64
Union	mining-above	vehicles/engines		CO	.97	53	4	67.57	13.64
Union	mining-below	drilling		HC	.00	1	1	.03	.01
Union	mining-below	blasting		HC		2	1		
Union	mining-below	raw shale removal/sca	engines/	HC	.18	8	1	12.87	5.14
Union	mining-above	raw shale	topsoil	HC	.00	11	2	.14	.06
Union	mining-above	spent shale	hauling	HC	.02	32	2	1.24	.49
Union	mining-above	spent shale	grooming	HC	.00	36	2	.16	.06
Union	utility	steam		HC	.05	37	3	3.80	1.52
Union	comb.-retort	sponge oil stripper		HC	.00	40	3	.30	.12
Union	retort	gas recycle heater		HC	.06	40	3	4.26	1.70
Union	upgrade	fractionater	reboiler	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	purge he	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	charge h	HC	.00	43	4	.32	.13
Union	upgrade	unicracker	charge h	HC	.01	43	4	.38	.15
Union	upgrade	steam boiler	no air p	HC	.01	43	4	.70	.28
Union	upgrade	steam boiler	with air	HC	.01	43	4	.70	.28
Union	upgrade	reformer furnace		HC	.05	43	4	3.36	1.34
Union	upgrade	storage tanks		HC	1.10	46	4	77.08	30.81
Union	upgrade	oil storage		HC	.02	46	4	1.46	.58
Union	upgrade	miscellaneous	table 4-	HC	.84	46	4	58.85	23.52
Union	mining-above	vehicles/engines		HC	.11	53	4	7.36	2.94
Union	upgrade	fugative		HC	1.10	54	4	77.08	30.81
Union	mining-below	drilling		NOx	.01	1	1	.42	.04
Union	mining-below	blasting		NOx	.29	2	1	20.23	1.70
Union	mining-below	raw shale removal/sca	engines/	NOx	2.97	8	1	207.74	17.48
Union	mining-above	raw shale	topsoil	NOx	.03	11	2	2.31	.19
Union	mining-above	spent shale	hauling	NOx	.31	32	2	21.46	1.81
Union	mining-above	spent shale	grooming	NOx	.05	36	2	3.42	.29
Union	utility	steam		NOx	3.54	37	3	247.31	20.81
Union	comb.-retort	sponge oil stripper		NOx	.42	40	3	29.22	2.46
Union	retort	gas recycle heater		NOx	4.14	40	3	289.16	24.33
Union	upgrade	fractionater	reboiler	NOx	.05	43	4	3.30	.28
Union	upgrade	dearseniter	purge he	NOx	.04	43	4	2.79	.23
Union	upgrade	dearseniter	charge h	NOx	.20	43	4	14.14	1.19
Union	upgrade	unicracker	charge h	NOx	.26	43	4	18.07	1.52
Union	upgrade	steam boiler	no air p	NOx	.56	43	4	39.25	3.30
Union	upgrade	steam boiler	with air	NOx	.67	43	4	47.05	3.96
Union	upgrade	reformer furnace		NOx	3.22	43	4	225.12	18.94
Union	mining-above	vehicles/engines		NOx	.25	53	4	17.69	1.49

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
Union	mining-below	drilling		SOx	.00	1	1	.02	.63
Union	mining-below	blasting		SOx	.03	2	1	2.38	1.27
Union	mining-below	raw shale removal/sca	engines/	SOx	.19	8	1	13.00	5.07
Union	mining-above	raw shale	topsoil	SOx	.00	11	2	.16	6.97
Union	mining-above	spent shale	hauling	SOx	.02	32	2	1.31	20.26
Union	mining	dumping shale		SOx	.04	33	2	2.85	20.90
Union	mining-above	spent shale	grooming	SOx	.00	36	2	.24	22.80
Union	utility	steam		SOx	2.02	37	3	141.16	23.43
Union	comb.-retort	sponge oil stripper		SOx	.29	40	3	20.39	25.33
Union	retort	gas recycle heater		SOx	3.34	40	3	233.74	25.33
Union	upgrade	fractionater	reboiler	SOx	.01	43	4	.76	27.23
Union	upgrade	dearseniter	purge he	SOx	.01	43	4	.63	27.23
Union	upgrade	dearseniter	charge h	SOx	.05	43	4	3.36	27.23
Union	upgrade	unicracker	charge h	SOx	.06	43	4	4.25	27.23
Union	upgrade	steam boiler	no air p	SOx	.11	43	4	7.55	27.23
Union	upgrade	steam boiler	with air	SOx	.11	43	4	7.55	27.23
Union	upgrade	reformer furnace		SOx	.15	43	4	10.15	27.23
Union	mining-above	vehicles/engines		SOx	.02	53	4	1.17	33.56
Utah-cottonwood	mining-below ground	blasting		CO	.40	2	1	80.33	31.51
Utah-cottonwood	mining-below ground	vehicles-combustion e		CO	.24	8	1	46.97	18.42
Utah-cottonwood	retort	fluid bed combustor		CO	.48	37	3	95.57	37.49
Utah-cottonwood	upgrade	fugitive	on site	CO	.03	53	4	6.85	2.69
Utah-cottonwood	upgrade	fugitive	on site	CO	.13	53	4	25.21	9.89
Utah-cottonwood	mining-below ground	vehicles-combustion e		HC	.07	8	1	14.51	13.36
Utah-cottonwood	retort	fluid bed combustor		HC	.02	37	3	3.08	2.84
Utah-cottonwood	upgrade	product storage		HC	.11	46	4	22.49	20.70
Utah-cottonwood	upgrade	fuel oil storage tank		HC		48	4		
Utah-cottonwood	upgrade	fugitive		HC	.34	54	4	68.55	63.11
Utah-cottonwood	mining-below ground	blasting		NOx	.22	2	1	43.88	1.02
Utah-cottonwood	mining-below ground	vehicles-combustion e		NOx	1.12	8	1	223.05	5.19
Utah-cottonwood	retort	fluid bed combustor		NOx	20.12	37	3	4020.34	93.53
Utah-cottonwood	upgrade	fugitive	on site	NOx	.06	53	4	11.24	.26
Utah-cottonwood	mining-below ground	vehicles-combustion e		SOx	.08	8	1	16.14	2.02
Utah-cottonwood	retort	fluid bed combustor		SOx	3.91	37	3	781.58	97.98
White River Shale Proj	mining-below	blasting		CO	.87	2	1	51.58	24.37
White River Shale Proj	mining-below	vehicles		CO	1.10	8	1	65.17	30.89
White River Shale Proj	utility	steam		CO	.44	37	3	26.33	12.50
White River Shale Proj	retort	Tosco ball heater & 1		CO	.15	40	3	8.97	4.25
White River Shale Proj	retort	recycle gas heater	union b	CO	.32	40	3	18.80	8.90

TABLE C-1. Continued

Project	general process	specific process	addition	pollu	m-tons/d	cat1	cat2	kg/1000m^3 oil	% total
White River Shale Proj	upgrade	reformer furnace		CO	.69	43	4	40.62	19.23
White River Shale Proj	mining-below	vehicles		HC	.34	8	1	20.31	9.38
White River Shale Proj	utility	steam		HC	.32	37	3	19.13	8.83
White River Shale Proj	retort	Tosco ball heater & 1		HC	.83	40	3	49.00	22.62
White River Shale Proj	retort	recycle gas heater	union	HC	.06	40	3	3.81	1.76
White River Shale Proj	upgrade	reformer furnace		HC	.14	43	4	8.22	3.79
White River Shale Proj	upgrade	valves,flanges,pumps,		HC	1.02	44	4	60.17	27.78
White River Shale Proj	upgrade	crude storage		HC	.46	46	4	27.29	12.60
White River Shale Proj	upgrade	crude shale oil stora		HC	.48	46	4	28.69	13.24
White River Shale Proj	mining-below	blasting		NOx	.47	2	1	27.94	2.63
White River Shale Proj	mining-below	vehicles		NOx	5.24	8	1	310.42	29.28
White River Shale Proj	utility	steam		NOx	4.77	37	3	282.59	26.65
White River Shale Proj	retort	gas treatment plant		NOx	.07	39	3	4.14	.39
White River Shale Proj	retort	Tosco ball heater & 1		NOx	1.02	40	3	60.17	5.68
White River Shale Proj	retort	recycle gas heater	union	NOx	2.01	40	3	119.27	11.25
White River Shale Proj	upgrade	reformer furnace		NOx	4.32	43	4	255.73	24.12
White River Shale Proj	mining-below	vehicles		SOx	.38	8	1	22.51	14.18
White River Shale Proj	utility	steam		SOx	1.63	37	3	96.70	60.93
White River Shale Proj	retort	gas treatment-claus t		SOx	.36	39	3	21.06	13.27
White River Shale Proj	retort	retort 1		SOx		40	3		
White River Shale Proj	retort	Tosco ball heater & 1		SOx	.10	40	3	6.18	3.89
White River Shale Proj	retort	recycle gas heater	union	SOx	.07	40	3	3.92	2.47
White River Shale Proj	upgrade	reformer furnace		SOx	.14	43	4	8.33	5.25

TABLE C-2. PARTICULATE EMISSIONS

project	general process	specific process	additional desc	cati	cate	pol	m-ton/d	kg/1000m³ oil	% total
Cathedral Bluffs	mining-below	mine shaft vents	0	8	1	pm	.17	88.46	26.54
Cathedral Bluffs	mining-above	reclaim drawhole	0	9	2	pm	.01	2.71	.81
Cathedral Bluffs	mining-above	raw shale	crushing buildi	13	2	pm	.00	1.62	.49
Cathedral Bluffs	mining-above	raw shale	screening plant	13	2	pm	.00	1.43	.43
Cathedral Bluffs	mining-above	raw shale	transfer house	15	2	pm	.02	9.04	2.71
Cathedral Bluffs	mining-above	raw shale	conveyor-stacke	15	2	pm	.05	25.21	7.56
Cathedral Bluffs	mining-abo	raw shale storage	5 day	18	2	pm	.01	3.80	1.14
Cathedral Bluffs	mining-abo	raw shale storage	5 year	21	2	pm	.01	5.42	1.63
Cathedral Bluffs	mining-above	raw shale	conveyor-retort	24	2	pm	.00	2.38	.71
Cathedral Bluffs	mining-abo	spent shale	stacker conveyo	30	2	pm	.05	25.21	7.56
Cathedral Bluffs	mining-above	spent shale	conveyor	30	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-above	spent shale	transfer house	31	2	pm	.02	9.04	2.71
Cathedral Bluffs	mining-abo	spent shale	conveyer discha	31	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-above	spent shale	transfer	31	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-abo	spent shale	distribution	33	2	pm	.07	38.52	11.56
Cathedral Bluffs	mining-abo	spent shale storag	0	34	2	pm	.01	2.85	.86
Cathedral Bluffs	comb-utility	steam boilers	0	37	3	pm	.03	14.27	4.28
Cathedral Bluffs	retort	incinerator	0	40	3	pm	.00	.38	.11
Cathedral Bluffs	comb.-retort	sponge oil reboile	0	40	3	pm	.00	1.90	.57
Cathedral Bluffs	comb.-retort	recycle gas heater	0	40	3	pm	.03	14.74	4.42
Cathedral Bluffs	comb-upgrade	reformer furnace	0	43	4	pm	.03	16.98	5.09
Cathedral Bluffs	upgrade	H2 Recycle heater	0	43	4	pm	.00	.57	.17
Cathedral Bluffs	upgrade	H2 charge heater	0	43	4	pm	.00	1.71	.51
Cathedral Bluffs	upgrade	Oil charge heater	0	43	4	pm	.01	2.85	.86
Cathedral Bluffs	mining-abo	fugative dust	haul roads	54	4	pm	.00	1.43	.43
Clear Creek Shale Oil	mining-below	drilling	inter. waste	1	1	PM	.00	.26	.13
Clear Creek Shale Oil	mining-below	drilling	0	1	1	PM	.01	.74	.38
Clear Creek Shale Oil	mining-below	blasting	inter. waste	2	1	PM	.10	6.22	3.19
Clear Creek Shale Oil	mining-below	blasting	0	2	1	PM	.26	16.27	8.35
Clear Creek Shale Oil	mining-below	primary crushing	0	6	1	PM	.00	.21	.11
Clear Creek Shale Oil	mining-below	vehicles-vent	0	8	1	PM	.40	25.34	13.01
Clear Creek Shale Oil	mining-above	surface soils	removal-drill	9	2	PM	.01	.74	.38
Clear Creek Shale Oil	mining-above	surface soils	removal	9	2	PM	.13	8.28	4.25
Clear Creek Shale Oil	mining-above	surface soils	removal-blast	9	2	PM	.25	15.58	8.00
Clear Creek Shale Oil	mining-above	surface soils	haul	11	2	PM	.36	22.83	11.72
Clear Creek Shale Oil	mining-above	surface soils	wind	12	2	PM	.00	.21	.11
Clear Creek Shale Oil	mining-above	raw shale	crushing	13	2	PM	.00	.29	.15
Clear Creek Shale Oil	mining-above	raw shale	crushing-2nd	13	2	PM	.01	.68	.35
Clear Creek Shale Oil	mining-above	raw shale	crushing-3rd	14	2	PM	.01	.91	.47
Clear Creek Shale Oil	mining-above	raw shale	conveying	15	2	PM	.02	1.13	.58
Clear Creek Shale Oil	mining-above	raw shale	conveying	15	2	PM	.81	51.08	26.23
Clear Creek Shale Oil	mining-above	int waste	dump	16	2	PM	.01	.34	.18
Clear Creek Shale Oil	mining-above	int waste	haul	16	2	PM	.15	9.36	4.81
Clear Creek Shale Oil	mining-above	raw shale	dump	17	2	PM	.12	7.53	3.87
Clear Creek Shale Oil	mining-above	raw shale	wind errosion	19	2	PM	.00	.02	.01
Clear Creek Shale Oil	mining-above	raw shale	wind	22	2	PM	.04	2.57	1.32
Clear Creek Shale Oil	mining-above	spent shale	conveying	30	2	PM	.18	11.59	5.95

TABLE C-2. Continued

project	general process	specific process	additional desc	cat1	cate	pol	m-ton/d	kg/1000m 3 oil	% total
Clear Creek Shale Oil	mining-above	spent shale	haul	32	2	PM	.02	1.37	.70
Clear Creek Shale Oil	comb-retort	retort gas	0	37	3	pm			
Clear Creek Shale Oil	comb-retort	TEG Concentrator	0	37	3	PM	.00	.23	.03
Clear Creek Shale Oil	comb-utility	steam superheat	0	37	3	PM	.14	8.96	1.36
Clear Creek Shale Oil	comb-retort	char combustion	feed bins(sourc	40	3	PM			
Clear Creek Shale Oil	comb-retort	char combustion	coal grinding	40	3	PM	.09	5.71	.87
Clear Creek Shale Oil	comb-retort	char combustion	0	40	3	PM	10.24	643.77	97.74
Clear Creek Shale Oil	mining-above	vehicles	0	53	4	PM	.17	10.50	5.39
Clear Creek Shale Oil	mining-above	coal	0	54	4	PM	.01	.68	.35
cottonwood	mining-below ground	mining	0	1	1	pm	.11	21.76	12.50
cottonwood	mining-below ground	transfer	transfer points	5	1	pm	.05	9.61	5.52
cottonwood	mining-below ground	raw shale	primary crushin	6	1	pm	.03	5.26	3.02
cottonwood	mining-below ground	vehicles-combustio	0	8	1	pm	.06	11.42	6.56
cottonwood	mining-above ground	raw shale	crusher-lump br	13	2	pm	.02	4.81	2.76
cottonwood	mining-above ground	raw shale	screening	13	2	pm	.03	5.60	3.22
cottonwood	mining-above ground	raw shale	surge bin	15	2	pm	.02	3.23	1.85
cottonwood	mining-above ground	raw shale	storage-load in	18	2	pm	.00	.30	.17
cottonwood	mining-above ground	raw shale	storage-reclaim	18	2	pm	.00	.60	.34
cottonwood	mining-above ground	raw shale	storage(live)wi	18	2	pm	.03	6.89	3.96
cottonwood	mining-above ground	raw shale	storage(dead)wi	21	2	pm	.01	2.77	1.59
cottonwood	mining-above ground	raw shale	retort feed bin	23	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort feed sil	23	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort feed con	24	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort discharg	25	2	pm	.00	.79	.46
cottonwood	mining-above ground	fines	FBC discharge b	26	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	conveyor-transf	26	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	FBC feed bin	27	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	storage silo	27	2	pm	.01	1.09	.63
cottonwood	mining-above	spent shale	conveyor	30	2	pm			
cottonwood	mining-above	spent shale	disposal(load i	33	2	pm			
cottonwood	mining-above	spent shale-fines	loading/dumping	33	2	pm			
cottonwood	mining-above	spent shale-fines	storage bin	34	2	pm	.01	1.12	.64
cottonwood	mining-above	spent shale	disposal-wind	35	2	pm	.03	5.01	2.87
cottonwood	mining-above	spent shale	disposal-groomi	36	2	pm	.33	66.55	38.23
cottonwood	com-retort	shale fluid bed co	0	40	3	pm	.04	8.20	4.71
cottonwood	comb-upgrade	fugitive	paved roads	53	4	pm	.01	2.18	1.25
cottonwood	comb-upgrade	fugitive	unpaved roads	53	4	pm	.06	11.24	6.46
Paraho-Ute	mining-below ground	mining	0	1	1	pm	.04	5.71	1.70
Paraho-Ute	mining-below ground	blasting	0	2	1	pm	.17	25.48	7.30
Paraho-Ute	mining-below ground	conveying	0	4	1	pm	.03	5.10	1.50
Paraho-Ute	mining-below ground	crushing/screening	primary	6	1	pm	.08	12.16	3.60
Paraho-Ute	mining-below ground	crushing/screening	secondary	7	1	pm	.08	12.16	3.60
Paraho-Ute	mining-below ground	mobile equipment	0	8	1	pm	.02	3.06	1.00
Paraho-Ute	mining-above ground	surface soils	wind	12	2	pm	.07	10.19	2.40
Paraho-Ute	mining-above ground	raw shale	tertiary crush/	14	2	pm	.24	36.35	10.50

TABLE C-2. Continued

project	general process	specific process	additional desc	cati	cate	pol	m-ton/d	kg/1000m^3 oil	% total
Paraho-Ute	mining-above ground	raw shale	trans. fr. live	15	2	pm	.01	1.77	.50
Paraho-Ute	mining-above ground	raw shale	convey/transfer	15	2	pm	.05	7.34	2.20
Paraho-Ute	mining-above ground	raw shale	sample & weigh	15	2	pm	.10	14.27	4.20
Paraho-Ute	mining-above ground	raw shale	screening/trans	15	2	pm	.16	24.60	7.10
Paraho-Ute	mining-above ground	raw shale	live storage	18	2	pm	.00	.14	.04
Paraho-Ute	mining-above ground	raw shale	emergency stora	21	2	pm	.11	16.10	4.70
Paraho-Ute	mining-above ground	raw shale	retort feed	24	2	pm	.04	5.38	1.60
Paraho-Ute	mining-above ground	fines	conveyor-transf	26	2	pm	.01	.82	.40
Paraho-Ute	mining-above ground	fines	to bin	26	2	pm	.01	1.14	.40
Paraho-Ute	mining-above ground	raw shale	fines transfer/	26	2	pm	.01	1.79	.50
Paraho-Ute	mining-above ground	fines	storage	26	2	pm	.11	16.50	3.80
Paraho-Ute	mining-above ground	spent shale	conveying	30	2	pm	.01	1.30	.37
Paraho-Ute	mining-above ground	spent shale	to bin	30	2	pm	.02	2.28	.70
Paraho-Ute	mining-above ground	spent shale	conveying-A	30	2	pm	.03	4.24	1.30
Paraho-Ute	mining-above ground	spent shale	retort overflow	31	2	pm	.04	5.38	1.60
Paraho-Ute	mining-above ground	spent shale	retort discharg	31	2	pm	.23	34.24	10.00
Paraho-Ute	mining-above ground	spent shale	storage	34	2	pm	.22	33.50	7.90
Paraho-Ute	upgrade	package boiler	0	37	3	PM	.00	.33	.09
Paraho-Ute	comb-retort	power generation	0	37	3	PM	.35	51.70	15.00
Paraho-Ute	upgrading	hydrotreater feed	0	43	4	pm	.01	.82	.30
Paraho-Ute	upgrading	reformer furnace	0	43	4	PM	.13	19.57	5.70
syntana	mining-below	mining	0	1	1	pm	.13	13.91	5.87
syntana	mining-below	blasting	0	2	1	pm	.18	19.32	8.16
syntana	mining-below	vehicles	0	8	1	pm	.14	15.31	6.46
syntana	mining-above	raw shale	primary crushin	13	2	pm	.01	1.42	.60
syntana	mining-above	raw shale	secondary crush	13	2	pm	.29	32.14	13.57
syntana	mining-above	raw shale	storage	18	2	pm	.03	3.47	1.47
syntana	mining-above	raw shale	storage-wind	22	2	pm	.04	4.77	2.01
syntana	mining-above	raw shale	retort feed-tos	24	2	pm	.00	.18	.07
syntana	mining-above	raw shale	conveyor feed	24	2	pm	.00	.35	.15
syntana	mining-above	fines transfer	0	26	2	pm	.01	.85	.36
syntana	mining-above	fines	storage-wind	27	2	pm	.02	2.64	1.12
syntana	mining-above	fines	maintenance	29	2	pm	.05	5.87	2.48
syntana	mining-above	spent shale	storage-load-tr	34	2	pm	.02	1.89	.80
syntana	mining-above	spent shale	storage-wind	35	2	pm	.16	17.62	7.44
syntana	mining-above	spent shale	storage-mainten	36	2	pm	.24	26.33	11.12
syntana	utility	steam	0	37	3	pm			
syntana	retort	FGD	steam boiler	37	3	pm	.05	5.89	2.49
syntana	retort	retort indirect he	0	39	3	pm			
syntana	retort	FGD	superior heater	40	3	pm	.03	3.48	1.47
syntana	retort	Tosco ball heater	0	40	3	pm	.34	37.73	15.93
syntana	retort	Tosco moisturizer	0	42	3	pm	.06	6.13	2.59
syntana	upgrade	FGD	hydrotreater	43	4	pm	.00	.53	.22
syntana	upgrade	FGD	hydrogen reform	43	4	pm	.06	6.61	2.79
syntana	mining-above	fugative	truck traffic	53	4	pm	.28	30.44	12.85

TABLE C-2. Continued

project	general process	specific process	additional desc	cati	cate	pol	m-ton/d	kg/1000m 3 oil	% total
Union	mining-below	drilling	0	1	1	pm	.00	.34	.19
Union	mining-below	blasting	0	2	1	PM	.01	.50	.28
Union	mining-below	removal	0	3	1	pm	.03	2.31	1.28
Union	mining-below	conveying	0	4	1	PM	.17	11.59	6.40
Union	mining-below	raw shale crushing	primary	6	1	PM	.07	4.63	2.56
Union	mining-below	crushing	primary	6	1	PM	.63	43.78	24.18
Union	mining-below	raw shale removal/	engines/vehicle	8	1	pm	.10	7.15	3.95
Union	mining-above	raw shale	topsoil hauling	11	2	pm	.01	.49	.27
Union	mining-above	raw shale	crushing/second	13	2	PM	.12	8.61	4.76
Union	mining-above	raw shale	crushing/tertia	14	2	PM	.41	28.71	15.85
Union	mining-above	raw shale	conveying/trans	15	2	PM	.01	.59	.32
Union	mining-above	raw shale storage	load out	18	2	PM	.00	.06	.04
Union	mining-above	storage silos	0	18	2	PM	.01	.50	.27
Union	mining-above	raw shale storage	load in	18	2	PM	.01	.77	.43
Union	mining-above	raw shale storage	wind erosion	19	2	PM	.00	.19	.10
Union	mining-above	retort pad	wind errosion	19	2	pm	.01	.36	.20
Union	mining-above	spent shale convey	0	30	2	pm	.08	5.55	3.07
Union	mining-above	spent shale	stacking	31	2	pm	.02	1.54	.85
Union	mining-above	spent shale	hauling	32	2	pm	.05	3.16	1.75
Union	mining-above	spent shale	truck dumping	33	2	pm	.03	1.98	1.09
Union	mining-above	spent shale dispos	0	34	2	PM			
Union	mining-above	spent shale-storag	wind errosion	35	2	pm	.01	.57	.32
Union	mining-above	spent shale	grooming/compac	36	2	pm	.03	1.93	1.07
Union	utility	steam	0	37	3	pm	.09	6.09	3.36
Union	retort	sponge oil strippe	0	39	3	pm	.01	.61	.34
Union	retort	gas recycle heater	0	40	3	PM	.10	7.31	4.03
Union	mining-above	vehicles/engines	light duty	53	4	pm	.04	2.64	1.46
Union	mining-above	vehicles/engines	road maint/acce	53	4	pm	.05	3.54	1.96
Union	mining-above	vehicles/engines	diesel	53	4	pm	.15	10.62	5.86
Union	mining-above	fugative dust	0	54	4	PM	.36	24.95	13.78
White River Shale Project	mining-below	mining	0	1	1	pm	.26	15.31	7.17
White River Shale Project	mining-below	crushing	primary	6	1	pm	.14	8.49	3.98
White River Shale Project	mining-below	crushing	secondary	7	1	pm	.04	2.63	1.23
White River Shale Project	mining-below	vehicles	0	8	1	pm	.27	15.85	7.42
White River Shale Project	mining-above	conveyor surge bin	0	15	2	pm	.01	.86	.40
White River Shale Project	mining-above	retort feed silos	three	18	2	pm	.05	2.69	1.26
White River Shale Project	mining-above	raw shale stockpil	load/groom/wind	18	2	pm	.12	6.93	3.25
White River Shale Project	mining-above	conveyor feed-unio	0	24	2	pm	.01	.51	.24
White River Shale Project	mining-above	conveyor feed-supe	transfer	24	2	pm	.05	2.69	1.26
White River Shale Project	mining-above	fines reclaim conv	0	26	2	pm	.00	.17	.08
White River Shale Project	mining-above	fines-storage	reclaim/wind/tr	27	2	pm	.07	4.41	2.06
White River Shale Project	mining-above	fines	storage/conveyo	27	2	pm	.14	8.54	4.00
White River Shale Project	mining-above	spent shale discha	un/tos/sup	30	2	pm	.01	.37	.17
White River Shale Project	mining-above	spent shale convey	0	30	2	pm	.05	3.22	1.51
White River Shale Project	mining-above	spent shale dispos	load in	34	2	pm	.02	1.07	.50
White River Shale Project	mining-above	spent shale dispos	wind erosion	35	2	pm	.05	2.79	1.31
White River Shale Project	mining-above	spent shale dispos	grooming	36	2	pm	.36	21.60	10.12

ABLE C-2. Continued

project	general process	specific process	additional desc	cat1	cate	pol	m-ton/d	kg/1000m^3 oil	% total
White River Shale Project	utility	steam	0	37	3	pm	.72	42.55	19.93
White River Shale Project	retort	Tosco ball heater	0	40	3	pm	.44	26.06	12.21
White River Shale Project	retort	recycle gas heater	union	40	3	pm	.19	11.28	5.29
White River Shale Project	retort	elutriator	0	41	3	PM	.05	3.09	1.45
White River Shale Project	retort	Tosco elutriators	and moisturizer	41	3	pm	.13	7.79	3.65
White River Shale Project	retort	processed shale mo	0	42	3	pm			
White River Shale Project	upgrade	reformer furnace	0	43	4	pm	.41	24.55	11.50

TABLE C-3. GASEOUS EMISSIONS

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
Catherdral Bluffs	mining-below	mine shaft vents		CO	.30	8	1	158.37	63.86
					.30			158.37	63.86
Clear Creek Shale Oil	mining-above	blasting		CO	.05	2	1	2.97	.03
Clear Creek Shale Oil	mining-below	blasting		CO	.36	2	1	22.66	.20
Clear Creek Shale Oil	mining-below	crushing/screening	primary	CO	.72	6	1	45.03	.40
Clear Creek Shale Oil	mining-below	vent		CO	1.23	8	1	77.62	.69
Paraho-Ute	mining-below ground	blasting		CO	.52	2	1	77.39	29.81
Paraho-Ute	mining-below ground	mobile equipment		CO	.67	8	1	99.67	38.39
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
syntana	mining-below	vehicles		CO	.57	8	1	62.58	34.03
Union	mining-below	drilling		CO	.00	1	1	.09	.02
Union	mining-below	blasting		CO	1.14	2	1	79.71	16.09
Union	mining-below	raw shale removal/scal	engines/ve	CO	.55	8	1	38.75	7.82
Utah-cottonwood	mining-below ground	blasting		CO	.40	2	1	80.33	31.51
Utah-cottonwood	mining-below ground	vehicles-combustion eq		CO	.24	8	1	46.97	18.42
White River Shale Projec	mining-below	blasting		CO	.87	2	1	51.58	24.37
White River Shale Projec	mining-below	vehicles		CO	1.10	8	1	65.17	30.89
Clear Creek Shale Oil	mining-above	top soil	removal	CO	.03	9	2	1.83	.02
Clear Creek Shale Oil	mining-above	top soil	load	CO	.00	10	2	.17	.00
Clear Creek Shale Oil	mining-above	raw shale haul		CO	.03	16	2	2.11	.02
Clear Creek Shale Oil	mining-above	spent shale haul		CO	.06	32	2	3.94	.03
Union	mining-above	raw shale	topsoil ha	CO	.01	11	2	.43	.09
Union	mining-above	spent shale	hauling	CO	.05	32	2	3.77	.76
Union	mining-above	spent shale	grooming/c	CO	.01	36	2	.50	.10
Catherdral Bluffs	utility	steam boilers		CO	.05	37	3	23.78	9.59
Catherdral Bluffs	retort	flares		CO	.00	38	3	.18	.07
Catherdral Bluffs	retort	sponge oil reboiler		CO	.01	40	3	3.33	1.34
Catherdral Bluffs	retort	recycle gas heater		CO	.05	40	3	25.21	10.16
					.10			52.50	21.17
Clear Creek Shale Oil	utility	steam superheat		CO					

TABLE C-3. Continued

TABLE C-3. Continued

Project	general process	specific process	additional pollu		m-ton/d	cat	cat	kg/1000m^3 oil	% total
Clear Creek Shale Oil	retort	char combustion		CO	176.04	40	3	11071.92	98.26
Clear Creek Shale Oil	retort	char combustion	coal grind	CO	.05	40	3	2.97	.03
Paraho-Ute	upgrade	package boiler		CO	.01	37	3	.83	.32
Paraho-Ute	retort	power generation		CO	.39	37	3	58.57	22.56
syntana	retort	FGD	steam boil	CO	.15	37	3	16.14	8.78
syntana	retort	FGD	superior h	CO	.09	40	3	9.51	5.17
syntana	retort	Tosco ball heater & li		CO	.05	40	3	5.17	2.81
Union	utility	steam		CO	2.66	37	3	185.67	37.49
Union	comb.-retort	sponge oil stripper		CO	.03	40	3	2.44	.49
Union	retort	gas recycle heater		CO	.33	40	3	23.13	4.67
Utah-cottonwood	retort	fluid bed combustor		CO	.48	37	3	95.57	37.49
White River Shale Projec	utility	steam		CO	.44	37	3	26.33	12.50
White River Shale Projec	retort	Tosco ball heater & li		CO	.15	40	3	8.97	4.25
White River Shale Projec	retort	recycle gas heater	union b	CO	.32	40	3	18.80	8.90
Catherdral Bluffs	upgrade	incinerator		CO	.00	43	4	.67	.27
Catherdral Bluffs	upgrade	H2 Recycle Heater		CO	.00	43	4	.90	.36
Catherdral Bluffs	upgrade	H2 Charge Heater		CO	.01	43	4	2.85	1.15
Catherdral Bluffs	upgrade	oil charge heater		CO	.01	43	4	4.66	1.88
Catherdral Bluffs	upgrade	reformer		CO	.05	43	4	28.06	11.31
Clear Creek Shale Oil	retort	TEG Concentrator		CO	.00	43	4	.11	.00
Clear Creek Shale Oil	mining-above	vehicles	road maint	CO	.01	53	4	.70	.01
Paraho-Ute	upgrading	hydrotreater feed furn		CO	.01	43	4	.98	.38
Paraho-Ute	upgrading	reformer furnace		CO	.15	43	4	22.18	8.54
syntana	upgrade	FGD	hydrotreat	CO	.01	43	4	1.53	.83
syntana	upgrade	FGD	hydrogen r	CO	.16	43	4	17.87	9.72
syntana	upgrade	hydrogen	furnace	CO	.23	43	4	25.33	13.78
Union	upgrade	fractionater	reboiler	CO	.01	43	4	.38	.08
Union	upgrade	dearseniter	purge heat	CO	.00	43	4	.32	.06
Union	upgrade	dearseniter	charge hea	CO	.02	43	4	1.65	.33

TABLE C-3. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
Union	upgrade	unicracker	charge hea	CO	.03	43	4	2.09	.42
Union	upgrade	steam boiler	no air pre	CO	.51	43	4	35.38	7.14
Union	upgrade	steam boiler	with air p	CO	.51	43	4	35.38	7.14
Union	upgrade	reformer furnace		CO	.26	43	4	18.01	3.64
Union	mining-above	vehicles/engines		CO	.97	53	4	67.57	13.64
Utah-cottonwood	upgrade	fugitive	on site ve	CO	.03	53	4	6.85	2.69
Utah-cottonwood	upgrade	fugitive	on site ve	CO	.13	53	4	25.21	9.89
White River Shale Projec	upgrade	reformer furnace		CO	.69	43	4	40.62	19.23
Catherdral Bluffs	mining-below	mine shaft vents		HC	.08	8	1	39.47	42.09
Clear Creek Shale Oil	mining-below	crushing/screening	primary	HC	.11	6	1	7.19	3.08
Clear Creek Shale Oil	mining-below	vent		HC	.17	8	1	10.96	4.70
Paraho-Ute	mining-below ground	mobile equipment		HC	.21	8	1	30.98	63.26
syntana	mining-below	vehicles		HC	.18	8	1	19.42	13.10
Union	mining-below	drilling		HC	.00	1	1	.03	.01
Union	mining-below	blasting		HC		2	1		
Union	mining-below	raw shale removal/scal	engines/ve	HC	.18	8	1	12.87	5.14
Utah-cottonwood	mining-below ground	vehicles-combustion eq		HC	.07	8	1	14.51	13.36
White River Shale Projec	mining-below	vehicles		HC	.34	8	1	20.31	9.38
Clear Creek Shale Oil	mining-above	top soil	removal	HC	.01	9	2	.57	.24
Clear Creek Shale Oil	mining-above	raw shale haul		HC	.01	16	2	.68	.29
Clear Creek Shale Oil	mining-above	spent shale haul		HC	.01	32	2	.51	.22
Union	mining-above	raw shale	topsoil ha	HC	.00	11	2	.14	.06
Union	mining-above	spent shale	hauling	HC	.02	32	2	1.24	.49
Union	mining-above	spent shale	grooming/c	HC	.00	36	2	.16	.06
Catherdral Bluffs	utility	steam boilers		HC	.08	37	3	42.80	45.64

TABLE C-3. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
Catherdral Bluffs	retort	flares		HC		38	3		
Catherdral Bluffs	retort	sponge oil reboiler		HC	.00	40	3	.57	.61
Catherdral Bluffs	retort	recycle gas heater		HC	.01	40	3	4.52	4.82
Paraho-Ute	upgrade	package boiler		HC	.00	37	3	.16	.33
Paraho-Ute	retort	power generation		HC	.07	37	3	10.26	20.95
syntana	retort	FGD	steam boil	HC	.03	37	3	2.83	1.91
syntana	retort	FGD	superior h	HC	.02	40	3	1.79	1.21
syntana	retort	Tosco ball heater & li		HC	.51	40	3	55.87	37.67
Union	utility	steam		HC	.05	37	3	3.80	1.52
Union	comb.-retort	sponge oil stripper		HC	.00	40	3	.30	.12
Union	retort	gas recycle heater		HC	.06	40	3	4.26	1.70
Utah-cottonwood	retort	fluid bed combustor		HC	.02	37	3	3.08	2.84
White River Shale Projec	utility	steam		HC	.32	37	3	19.13	8.83
White River Shale Projec	retort	Tosco ball heater & li		HC	.83	40	3	49.00	22.62
White River Shale Projec	retort	recycle gas heater	union	HC	.06	40	3	3.81	1.76
Catherdral Bluffs	upgrade	H2 Recycle Heater		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	incinerator		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	H2 Charge Heater		HC	.00	43	4	.57	.61
Catherdral Bluffs	upgrade	oil charge heater		HC	.00	43	4	.81	.86
Catherdral Bluffs	upgrade	reformer		HC	.01	43	4	4.76	5.07
Catherdral Bluffs	upgrade	storage		HC		43	4		
Clear Creek Shale Oil	upgrade	fugative		HC	3.26	44	4	204.89	87.80
Clear Creek Shale Oil	upgrade	storage		HC	.13	46	4	7.99	3.42
Clear Creek Shale Oil	mining-above	vehicles	road maint	HC	.01	53	4	.57	.24
Paraho-Ute	upgrading	hydrotreater feed furn		HC	.00	43	4	.16	.33
Paraho-Ute	upgrading	reformer furnace		HC	.03	43	4	3.91	7.99
Paraho-Ute	upgrading	storage	crude shal	HC	.01	46	4	.82	1.66
Paraho-Ute	upgrading	storage (day)	crude shal	HC	.01	46	4	1.22	2.48
Paraho-Ute	upgrading	storage	hydrotreat	HC	.01	47	4	1.05	2.15
Paraho-Ute	upgrading	storage	fuel oil	HC	.00	48	4	.04	.08
Paraho-Ute	upgrading	storage	diesel & g	HC	.00	49	4	.37	.75

TABLE C-3. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
syntana	upgrade	FGD	hydrotreat	HC	.00	43	4	.29	.20
syntana	upgrade	FGD	hydrogen r	HC	.03	43	4	3.20	2.16
syntana	upgrade	hydrogen	furnace	HC	.04	43	4	4.71	3.17
syntana	upgrade	oil storage tanks		HC	.03	46	4	3.12	2.11
syntana	upgrade	fugitive emissions		HC	.52	54	4	57.07	38.48
Union	upgrade	fractionater	reboiler	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	purge heat	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	charge hea	HC	.00	43	4	.32	.13
Union	upgrade	unicracker	charge hea	HC	.01	43	4	.38	.15
Union	upgrade	steam boiler	no air pre	HC	.01	43	4	.70	.28
Union	upgrade	steam boiler	with air p	HC	.01	43	4	.70	.28
Union	upgrade	reformer furnace		HC	.05	43	4	3.36	1.34
Union	upgrade	storage tanks		HC	1.10	46	4	77.08	30.81
Union	upgrade	oil storage		HC	.02	46	4	1.46	.58
Union	upgrade	miscellaneous	table 4-7?	HC	.84	46	4	58.85	23.52
Union	mining-above	vehicles/engines		HC	.11	53	4	7.36	2.94
Union	upgrade	fugative		HC	1.10	54	4	77.08	30.81
Utah-cottonwood	upgrade	product storage		HC	.11	46	4	22.49	20.70
Utah-cottonwood	upgrade	fuel oil storage tanks		HC		48	4		
Utah-cottonwood	upgrade	fugitive		HC	.34	54	4	68.55	63.11
White River Shale Projec	upgrade	reformer furnace		HC	.14	43	4	8.22	3.79
White River Shale Projec	upgrade	valves,flanges,pumps,s		HC	1.02	44	4	60.17	27.78
White River Shale Projec	upgrade	crude storage		HC	.46	46	4	27.29	12.60
White River Shale Projec	upgrade	crude shale oil storag		HC	.48	46	4	28.69	13.24
Catherdral Bluffs	mining-below	mine shaft vents		NOx	.91	8	1	475.12	15.00
Clear Creek Shale Oil	mining-above	blasting		NOx	.02	2	1	1.43	.02
Clear Creek Shale Oil	mining-below	blasting		NOx	.03	2	1	1.71	.03
Clear Creek Shale Oil	mining-below	crushing/screening	primary	NOx	.75	6	1	47.03	.82
Clear Creek Shale Oil	mining-below	vent		NOx	1.30	8	1	81.61	1.42
Paraho-Ute	mining-below ground	blasting		NOx		2	1		
Paraho-Ute	mining-below ground	mobile equipment		NOx	3.17	8	1	473.97	27.44
syntana	mining-below	blasting		NOx	.23	2	1	25.03	2.17
syntana	mining-below	vehicles		NOx	2.70	8	1	297.37	25.79

TABLE C-3. Continued

Project	general process	specific process	additional pollu		m-ton/d	cat	cat	kg/1000m^3 oil	% total
Union	mining-below	drilling		NOx	.01	1	1	.42	.04
Union	mining-below	blasting		NOx	.29	2	1	20.23	1.70
Union	mining-below	raw shale removal/scal	engines/ve	NOx	2.97	8	1	207.74	17.48
Utah-cottonwood	mining-below ground	blasting		NOx	.22	2	1	43.88	1.02
Utah-cottonwood	mining-below ground	vehicles-combustion eq		NOx	1.12	8	1	223.05	5.19
White River Shale Projec	mining-below	blasting		NOx	.47	2	1	27.94	2.63
White River Shale Projec	mining-below	vehicles		NOx	5.24	8	1	310.42	29.28
Clear Creek Shale Oil	mining-above	top soil	removal	NOx	.20	9	2	12.44	.22
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.00	10	2	.06	.00
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.01	10	2	.74	.01
Clear Creek Shale Oil	mining-above	raw shale haul		NOx	.19	16	2	11.99	.21
Clear Creek Shale Oil	mining-above	spent shale haul		NOx	.14	32	2	8.73	.15
Union	mining-above	raw shale	topsoil ha	NOx	.03	11	2	2.31	.19
Union	mining-above	spent shale	hauling	NOx	.31	32	2	21.46	1.81
Union	mining-above	spent shale	grooming/c	NOx	.05	36	2	3.42	.29
Catherdral Bluffs	utility	steam boilers		NOx	1.89	37	3	989.24	31.24
Catherdral Bluffs	retort	flares		NOx	.00	38	3	2.62	.08
Catherdral Bluffs	retort	sponge oil reboiler		NOx	.09	40	3	48.04	1.52
Catherdral Bluffs	retort	recycle gas heater		NOx	.70	40	3	369.06	11.65
Clear Creek Shale Oil	utility	steam superheat		NOx	1.87	37	3	117.57	2.04
Clear Creek Shale Oil	retort	char combustion		NOx	86.90	40	3	5465.19	94.85
Clear Creek Shale Oil	retort	char combustion	coal grind	NOx	.15	40	3	9.53	.17
Paraho-Ute	upgrade	package boiler		NOx	.02	37	3	3.74	.22
Paraho-Ute	retort	power generation		NOx	6.25	37	3	936.25	54.21
syntana	retort	FGD	steam boil	NOx	1.71	37	3	188.84	16.38
syntana	utility	steam		NOx		37	3		
syntana	retort	retort indirect heater		NOx		39	3		
syntana	retort	FGD	superior h	NOx	.99	40	3	108.80	9.44
syntana	retort	Tosco ball heater & li		NOx	.57	40	3	62.58	5.43
Union	utility	steam		NOx	3.54	37	3	247.31	20.81
Union	comb.-retort	sponge oil stripper		NOx	.42	40	3	29.22	2.46

TABLE C-3. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
Union	retort	gas recycle heater		NOx	4.14	40	3	289.16	24.33
Utah-cottonwood	retort	fluid bed combustor		NOx	20.12	37	3	4020.34	93.53
White River Shale Projec	utility	steam		NOx	4.77	37	3	282.59	26.65
White River Shale Projec	retort	gas treatment plant		NOx	.07	39	3	4.14	.39
White River Shale Projec	retort	Tosco ball heater & li		NOx	1.02	40	3	60.17	5.68
White River Shale Projec	retort	recycle gas heater	union	NOx	2.01	40	3	119.27	11.25
Catherdral Bluffs	upgrade	incinerator		NOx	.02	43	4	9.04	.29
Catherdral Bluffs	upgrade	H2 Recycle Heater		NOx	.02	43	4	12.84	.41
Catherdral Bluffs	upgrade	H2 Charge Heater		NOx	.07	43	4	39.00	1.23
Catherdral Bluffs	upgrade	oil charge heater		NOx	.12	43	4	62.78	1.98
Catherdral Bluffs	upgrade	reformer		NOx	2.21	43	4	1159.03	36.60
Clear Creek Shale Oil	retort	TEG Concentrator		NOx	.00	43	4	.29	.00
Clear Creek Shale Oil	mining-above	vehicles	road maint	NOx	.05	53	4	3.42	.06
Paraho-Ute	upgrading	hydrotreater feed furn		NOX	.09	43	4	13.18	.76
Paraho-Ute	upgrading	reformer furnace		NOX	2.00	43	4	300.03	17.37
syntana	upgrade	FGD	hydrotreat	NOx	.16	43	4	17.77	1.54
syntana	upgrade	FGD	hydrogen r	NOx	1.90	43	4	210.16	18.23
syntana	upgrade	hydrogen	furnace	NOx	2.20	43	4	242.30	21.02
Union	upgrade	fractionater	reboiler	NOx	.05	43	4	3.30	.28
Union	upgrade	dearseniter	purge heat	NOx	.04	43	4	2.79	.23
Union	upgrade	dearseniter	charge hea	NOx	.20	43	4	14.14	1.19
Union	upgrade	unicracker	charge hea	NOx	.26	43	4	18.07	1.52
Union	upgrade	steam boiler	no air pre	NOx	.56	43	4	39.25	3.30
Union	upgrade	steam boiler	with air p	NOx	.67	43	4	47.05	3.96
Union	upgrade	reformer furnace		NOx	3.22	43	4	225.12	18.94
Union	mining-above	vehicles/engines		NOx	.25	53	4	17.69	1.49
Utah-cottonwood	upgrade	fugitive	on site ve	NOx	.06	53	4	11.24	.26
White River Shale Projec	upgrade	reformer furnace		NOx	4.32	43	4	255.73	24.12
Catherdral Bluffs	mining-below	mine shaft vents		SOx	.06	8	1	29.96	4.19

TABLE C-3. Continued

Project	general process	specific process	additional pollu		m-ton/d	cat	cat	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-below	crushing/screening	primary	SOx	.07	6	1	4.34	.78
Clear Creek Shale Oil	mining-below	vent		SOx	.12	8	1	7.48	1.35
Paraho-Ute	mining-below ground	mobile equipment		SOx	.23	8	1	34.31	5.25
syntana	mining-below	vehicles		SOx	.20	8	1	21.53	3.74
Union	mining-below	drilling		SOx	.00	1	1	.02	.63
Union	mining-below	blasting		SOx	.03	2	1	2.38	1.27
Union	mining-below	raw shale removal/scal	engines/ve	SOx	.19	8	1	13.00	5.07
Utah-cottonwood	mining-below ground	vehicles-combustion eq		SOx	.08	8	1	16.14	2.02
White River Shale Projec	mining-below	vehicles		SOx	.38	8	1	22.51	14.18
Clear Creek Shale Oil	mining-above	top soil	removal	SOx	.01	9	2	.86	.15
Clear Creek Shale Oil	mining-above	top soil	load	SOx	.00	10	2	.06	.01
Clear Creek Shale Oil	mining-above	raw shale haul		SOx	.01	16	2	.68	.12
Clear Creek Shale Oil	mining-above	spent shale haul		SOx	.01	32	2	.51	.09
Union	mining-above	raw shale	topsoil ha	SOx	.00	11	2	.16	6.97
Union	mining-above	spent shale	hauling	SOx	.02	32	2	1.31	20.26
Union	mining	dumping shale		SOx	.04	33	2	2.85	20.90
Union	mining-above	spent shale	grooming/c	SOx	.00	36	2	.24	22.80
Catherdral Bluffs	utility	steam boilers		SOx	.28	37	3	147.44	20.64
Catherdral Bluffs	retort	flares		SOx	.02	38	3	8.09	1.13
Catherdral Bluffs	retort	sponge oil reboiler		SOx	.04	40	3	20.45	2.86
Catherdral Bluffs	retort	recycle gas heater		SOx	.30	40	3	156.95	21.97
Clear Creek Shale Oil	utility	steam superheater-sour		SOx	.49	37	3	30.82	5.55
Clear Creek Shale Oil	retort	char combustion	coal grind	SOx	.05	40	3	2.97	.53
Clear Creek Shale Oil	retort	char combustion		SOx	8.06	40	3	506.80	91.33
Paraho-Ute	upgrade	package boiler		SOx	.05	37	3	7.07	1.08
Paraho-Ute	retort	power generation		SOx	3.97	37	3	595.18	91.11
syntana	retort	FGD	steam boil	SOx	.91	37	3	100.13	17.38

TABLE C-3. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
syntana	utility	steam		SOx		37	3		
syntana	retort	claus plant		SOx		39	3		
syntana	retort	FGD	claus plan	SOx	.51	39	3	56.27	9.77
syntana	retort	retort indirect heater		SOx		39	3		
syntana	retort	FGD	superior h	SOx	.53	40	3	57.97	10.06
syntana	retort	Tosco ball heater & li		SOx	.40	40	3	44.46	7.72
Union	utility	steam		SOx	2.02	37	3	141.16	23.43
Union	comb.-retort	sponge oil stripper		SOx	.29	40	3	20.39	25.33
Union	retort	gas recycle heater		SOx	3.34	40	3	233.74	25.33
Utah-cottonwood	retort	fluid bed combustor		SOx	3.91	37	3	781.58	97.98
White River Shale Projec	utility	steam		SOx	1.63	37	3	96.70	60.93
White River Shale Projec	retort	gas treatment-claus ta		SOx	.36	39	3	21.06	13.27
White River Shale Projec	retort	retort li		SOx		40	3		
White River Shale Projec	retort	Tosco ball heater & li		SOx	.10	40	3	6.18	3.89
White River Shale Projec	retort	recycle gas heater	union	SOx	.07	40	3	3.92	2.47
					2.16			127.86	80.57
Catherdral Bluffs	upgrade	H2 Recycle Heater		SOx	.01	43	4	5.23	.73
Catherdral Bluffs	upgrade	H2 Charge Heater		SOx	.03	43	4	17.60	2.46
Catherdral Bluffs	upgrade	oil charge heater		SOx	.05	43	4	28.54	3.99
Catherdral Bluffs	upgrade	incinerator		SOx	.24	43	4	126.98	17.78
Catherdral Bluffs	upgrade	reformer		SOx	.33	43	4	173.12	24.23
					.67			351.47	49.20
Clear Creek Shale Oil	retort	TEG Concentrator		SOx	.00	43	4	.11	.02
Clear Creek Shale Oil	mining-above	vehicles	road maint	SOx	.00	53	4	.29	.05
					.01			.40	.07
Paraho-Ute	upgrading	hydrotreater feed furn		SOx	.00	43	4	.03	.00
Paraho-Ute	upgrading	reformer furnace		SOx	.11	43	4	16.65	2.55
syntana	upgrade	FGD	hydrotreat	SOx	.09	43	4	9.51	1.65
Union	upgrade	fractionater	reboiler	SOx	.01	43	4	.76	27.23
Union	upgrade	dearseniter	purge heat	SOx	.01	43	4	.63	27.23
Union	upgrade	dearseniter	charge hea	SOx	.05	43	4	3.36	27.23
Union	upgrade	unicracker	charge hea	SOx	.06	43	4	4.25	27.23
Union	upgrade	steam boiler	no air pre	SOx	.11	43	4	7.55	27.23
Union	upgrade	steam boiler	with air p	SOx	.11	43	4	7.55	27.23
Union	upgrade	reformer furnace		SOx	.15	43	4	10.15	27.23
Union	mining-above	vehicles/engines		SOx	.02	53	4	1.17	33.56

TABLE C-3. Continued

Project	general process	specific process	additional pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
White River Shale Projec	upgrade	reformer furnace	SOx	.14	43	4	8.33	5.25
				.14			8.33	5.25

TABLE C-4. PARTICULATE EMISSIONS

project	general process	specific process	additional desc	cati	cate	pol	m-ton/d	kg/1000m 3 oil	% total
Cathedral Bluffs	mining-below	mine shaft vents	0	8	1	pm	.17	88.46	26.54
							.17	88.46	26.54
Clear Creek Shale Oil	mining-below	drilling	inter. waste	1	1	PM	.00	.26	.13
Clear Creek Shale Oil	mining-below	drilling	0	1	1	PM	.01	.74	.38
Clear Creek Shale Oil	mining-below	blasting	inter. waste	2	1	PM	.10	6.22	3.19
Clear Creek Shale Oil	mining-below	blasting	0	2	1	PM	.26	16.27	8.35
Clear Creek Shale Oil	mining-below	primary crushing	0	6	1	PM	.00	.21	.11
Clear Creek Shale Oil	mining-below	vehicles-vent	0	8	1	PM	.40	25.34	13.01
cottonwood	mining-below ground	mining	0	1	1	pm	.11	21.76	12.50
cottonwood	mining-below ground	transfer	transfer points	5	1	pm	.05	9.61	5.52
cottonwood	mining-below ground	raw shale	primary crushin	6	1	pm	.03	5.26	3.02
cottonwood	mining-below ground	vehicles-combustio	0	8	1	pm	.06	11.42	6.56
Paraho-Ute	mining-below ground	mining	0	1	1	pm	.04	5.71	1.70
Paraho-Ute	mining-below ground	blasting	0	2	1	pm	.17	25.48	7.30
Paraho-Ute	mining-below ground	conveying	0	4	1	pm	.03	5.10	1.50
Paraho-Ute	mining-below ground	crushing/screening	primary	6	1	pm	.08	12.16	3.60
Paraho-Ute	mining-below ground	crushing/screening	secondary	7	1	pm	.08	12.16	3.60
Paraho-Ute	mining-below ground	mobile equipment	0	8	1	pm	.02	3.06	1.00
syntana	mining-below	mining	0	1	1	pm	.13	13.91	5.87
syntana	mining-below	blasting	0	2	1	pm	.18	19.32	8.16
syntana	mining-below	vehicles	0	8	1	pm	.14	15.31	6.46
Union	mining-below	drilling	0	1	1	pm	.00	.34	.19
Union	mining-below	blasting	0	2	1	PM	.01	.50	.28
Union	mining-below	removal	0	3	1	pm	.03	2.31	1.28
Union	mining-below	conveying	0	4	1	PM	.17	11.59	6.40
Union	mining-below	raw shale crushing	primary	6	1	PM	.07	4.63	2.56
Union	mining-below	crushing	primary	6	1	PM	.63	43.78	24.18
Union	mining-below	raw shale removal/	engines/vehicle	8	1	pm	.10	7.15	3.95
White River Shale Project	mining-below	mining	0	1	1	pm	.26	15.31	7.17
White River Shale Project	mining-below	crushing	primary	6	1	pm	.14	8.49	3.98
White River Shale Project	mining-below	crushing	secondary	7	1	pm	.04	2.63	1.23
White River Shale Project	mining-below	vehicles	0	8	1	pm	.27	15.85	7.42
Cathedral Bluffs	mining-above	reclaim drawhole	0	9	2	pm	.01	2.71	.81
Cathedral Bluffs	mining-above	raw shale	crushing buildi	13	2	pm	.00	1.62	.49
Cathedral Bluffs	mining-above	raw shale	screening plant	13	2	pm	.00	1.43	.43
Cathedral Bluffs	mining-above	raw shale	transfer house	15	2	pm	.02	9.04	2.71

TABLE C-4. Continued

project	general process	specific process	additional desc	cati	cate	pol	m-ton/d	kg/1000m 3 oil	% total
Cathedral Bluffs	mining-above	raw shale	conveyor-stacke	15	2	pm	.05	25.21	7.56
Cathedral Bluffs	mining-abo	raw shale storage	5 day	18	2	pm	.01	3.80	1.14
Cathedral Bluffs	mining-abo	raw shale storage	5 year	21	2	pm	.01	5.42	1.63
Cathedral Bluffs	mining-above	raw shale	conveyor-retort	24	2	pm	.00	2.38	.71
Cathedral Bluffs	mining-abo	spent shale	stacker conveyo	30	2	pm	.05	25.21	7.56
Cathedral Bluffs	mining-above	spent shale	conveyor	30	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-above	spent shale	transfer house	31	2	pm	.02	9.04	2.71
Cathedral Bluffs	mining-abo	spent shale	conveyer discha	31	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-above	spent shale	transfer	31	2	pm	.04	20.93	6.28
Cathedral Bluffs	mining-abo	spent shale	distribution	33	2	pm	.07	38.52	11.56
Cathedral Bluffs	mining-abo	spent shale storag	0	34	2	pm	.01	2.85	.86
Clear Creek Shale Oil	mining-above	surface soils	removal-drill	9	2	PM	.01	.74	.38
Clear Creek Shale Oil	mining-above	surface soils	removal	9	2	PM	.13	8.28	4.25
Clear Creek Shale Oil	mining-above	surface soils	removal-blast	9	2	PM	.25	15.58	8.00
Clear Creek Shale Oil	mining-above	surface soils	haul	11	2	PM	.36	22.83	11.72
Clear Creek Shale Oil	mining-above	surface soils	wind	12	2	PM	.00	.21	.11
Clear Creek Shale Oil	mining-above	raw shale	crushing	13	2	PM	.00	.29	.15
Clear Creek Shale Oil	mining-above	raw shale	crushing-2nd	13	2	PM	.01	.68	.35
Clear Creek Shale Oil	mining-above	raw shale	crushing-3rd	14	2	PM	.01	.91	.47
Clear Creek Shale Oil	mining-above	raw shale	conveying	15	2	PM	.02	1.13	.58
Clear Creek Shale Oil	mining-above	raw shale	conveying	15	2	PM	.81	51.08	26.23
Clear Creek Shale Oil	mining-above	int waste	dump	16	2	PM	.01	.34	.18
Clear Creek Shale Oil	mining-above	int waste	haul	16	2	PM	.15	9.36	4.81
Clear Creek Shale Oil	mining-above	raw shale	dump	17	2	PM	.12	7.53	3.87
Clear Creek Shale Oil	mining-above	raw shale	wind errosion	19	2	PM	.00	.02	.01
Clear Creek Shale Oil	mining-above	raw shale	wind	22	2	PM	.04	2.57	1.32
Clear Creek Shale Oil	mining-above	spent shale	conveying	30	2	PM	.18	11.59	5.95
Clear Creek Shale Oil	mining-above	spent shale	haul	32	2	PM	.02	1.37	.70
cottonwood	mining-above ground	raw shale	crusher-lump br	13	2	pm	.02	4.81	2.76
cottonwood	mining-above ground	raw shale	screening	13	2	pm	.03	5.60	3.22
cottonwood	mining-above ground	raw shale	surge bin	15	2	pm	.02	3.23	1.85
cottonwood	mining-above ground	raw shale	storage-load in	18	2	pm	.00	.30	.17
cottonwood	mining-above ground	raw shale	storage-reclaim	18	2	pm	.00	.60	.34
cottonwood	mining-above ground	raw shale	storage(live)wi	18	2	pm	.03	6.89	3.96
cottonwood	mining-above ground	raw shale	storage(dead)wi	21	2	pm	.01	2.77	1.59
cottonwood	mining-above ground	raw shale	retort feed bin	23	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort feed sil	23	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort feed con	24	2	pm	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort discharg	25	2	pm	.00	.79	.46
cottonwood	mining-above ground	fines	FBC discharge b	26	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	conveyor-transf	26	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	FBC feed bin	27	2	pm	.01	1.09	.63
cottonwood	mining-above ground	fines	storage silo	27	2	pm	.01	1.09	.63
cottonwood	mining-above	spent shale	conveyor	30	2	pm			
cottonwood	mining-above	spent shale	disposal(load i	33	2	pm			

TABLE C-4. Continued

project	general process	specific process	additional desc	cat1	cate	pol	m-ton/d	kg/1000m^3 oil	% total
cottonwood	mining-above	spent shale-fines	loading/dumping	33	2	pm			
cottonwood	mining-above	spent shale-fines	storage bin	34	2	pm	.01	1.12	.64
cottonwood	mining-above	spent shale	disposal-wind	35	2	pm	.03	5.01	2.87
cottonwood	mining-above	spent shale	disposal-groomi	36	2	pm	.33	66.55	38.23
Paraho-Ute	mining-above ground	surface soils	wind	12	2	pm	.07	10.19	2.40
Paraho-Ute	mining-above ground	raw shale	tertiary crush/	14	2	pm	.24	36.35	10.50
Paraho-Ute	mining-above ground	raw shale	trans. fr. live	15	2	pm	.01	1.77	.50
Paraho-Ute	mining-above ground	raw shale	convey/transfer	15	2	pm	.05	7.34	2.20
Paraho-Ute	mining-above ground	raw shale	sample & weigh	15	2	pm	.10	14.27	4.20
Paraho-Ute	mining-above ground	raw shale	screening/trans	15	2	pm	.16	24.60	7.10
Paraho-Ute	mining-above ground	raw shale	live storage	18	2	pm	.00	.14	.04
Paraho-Ute	mining-above ground	raw shale	emergency stora	21	2	pm	.11	16.10	4.70
Paraho-Ute	mining-above ground	raw shale	retort feed	24	2	pm	.04	5.38	1.60
Paraho-Ute	mining-above ground	fines	conveyor-transf	26	2	pm	.01	.82	.40
Paraho-Ute	mining-above ground	fines	to bin	26	2	pm	.01	1.14	.40
Paraho-Ute	mining-above ground	raw shale	fines transfer/	26	2	pm	.01	1.79	.50
Paraho-Ute	mining-above ground	fines	storage	26	2	pm	.11	16.50	3.80
Paraho-Ute	mining-above ground	spent shale	conveying	30	2	pm	.01	1.30	.37
Paraho-Ute	mining-above ground	spent shale	to bin	30	2	pm	.02	2.28	.70
Paraho-Ute	mining-above ground	spent shale	conveying-A	30	2	pm	.03	4.24	1.30
Paraho-Ute	mining-above ground	spent shale	retort overflow	31	2	pm	.04	5.38	1.60
Paraho-Ute	mining-above ground	spent shale	retort discharg	31	2	pm	.23	34.24	10.00
Paraho-Ute	mining-above ground	spent shale	storage	34	2	pm	.22	33.50	7.90
syntana	mining-above	raw shale	primary crushin	13	2	pm	.01	1.42	.60
syntana	mining-above	raw shale	secondary crush	13	2	pm	.29	32.14	13.57
syntana	mining-above	raw shale	storage	18	2	pm	.03	3.47	1.47
syntana	mining-above	raw shale	storage-wind	22	2	pm	.04	4.77	2.01
syntana	mining-above	raw shale	retort feed-tos	24	2	pm	.00	.18	.07
syntana	mining-above	raw shale	conveyor feed	24	2	pm	.00	.35	.15
syntana	mining-above	fines transfer	0	26	2	pm	.01	.85	.36
syntana	mining-above	fines	storage-wind	27	2	pm	.02	2.64	1.12
syntana	mining-above	fines	maintenance	29	2	pm	.05	5.87	2.48
syntana	mining-above	spent shale	storage-load-tr	34	2	pm	.02	1.89	.80
syntana	mining-above	spent shale	storage-wind	35	2	pm	.16	17.62	7.44
syntana	mining-above	spent shale	storage-mainten	36	2	pm	.24	26.33	11.12
Union	mining-above	raw shale	topsoil hauling	11	2	pm	.01	.49	.27
Union	mining-above	raw shale	crushing/second	13	2	PM	.12	8.61	4.76
Union	mining-above	raw shale	crushing/tertia	14	2	PM	.41	28.71	15.85
Union	mining-above	raw shale	conveying/trans	15	2	PM	.01	.59	.32
Union	mining-above	raw shale storage	load out	18	2	PM	.00	.06	.04
Union	mining-above	storage silos	0	18	2	PM	.01	.50	.27
Union	mining-above	raw shale storage	load in	18	2	PM	.01	.77	.43
Union	mining-above	raw shale storage	wind erosion	19	2	PM	.00	.19	.10

TABLE C-4. Continued

project	general process	specific process	additional desc	cat1	cate	pol	m-ton/d	kg/1000m 3 oil	% total
Union	mining-above	retort pad	wind errosion	19	2	pm	.01	.36	.20
Union	mining-above	spent shale convey	0	30	2	pm	.08	5.55	3.07
Union	mining-above	spent shale	stacking	31	2	pm	.02	1.54	.85
Union	mining-above	spent shale	hauling	32	2	pm	.05	3.16	1.75
Union	mining-above	spent shale	truck dumping	33	2	pm	.03	1.98	1.09
Union	mining-above	spent shale dispos	0	34	2	PM			
Union	mining-above	spent shale-storag	wind errosion	35	2	pm	.01	.57	.32
Union	mining-above	spent shale	grooming/compac	36	2	pm	.03	1.93	1.07
							.79	55.02	30.38
White River Shale Project	mining-above	conveyor surge bin	0	15	2	pm	.01	.86	.40
White River Shale Project	mining-above	retort feed silos	three	18	2	pm	.05	2.69	1.26
White River Shale Project	mining-above	raw shale stockpil	load/groom/wind	18	2	pm	.12	6.93	3.25
White River Shale Project	mining-above	conveyor feed-unio	0	24	2	pm	.01	.51	.24
White River Shale Project	mining-above	conveyor feed-supe	transfer	24	2	pm	.05	2.69	1.26
White River Shale Project	mining-above	fines reclaim conv	0	26	2	pm	.00	.17	.08
White River Shale Project	mining-above	fines-storage	reclaim/wind/tr	27	2	pm	.07	4.41	2.06
White River Shale Project	mining-above	fines	storage/conveyo	27	2	pm	.14	8.54	4.00
White River Shale Project	mining-above	spent shale discha	un/tos/sup	30	2	pm	.01	.37	.17
White River Shale Project	mining-above	spent shale convey	0	30	2	pm	.05	3.22	1.51
White River Shale Project	mining-above	spent shale dispos	load in	34	2	pm	.02	1.07	.50
White River Shale Project	mining-above	spent shale dispos	wind erosion	35	2	pm	.05	2.79	1.31
White River Shale Project	mining-above	spent shale dispos	grooming	36	2	pm	.36	21.60	10.12
							.94	55.85	26.16
Cathedral Bluffs	comb-utility	steam boilers	0	37	3	pm	.03	14.27	4.28
Cathedral Bluffs	retort	incinerator	0	40	3	pm	.00	.38	.11
Cathedral Bluffs	comb.-retort	sponge oil reboile	0	40	3	pm	.00	1.90	.57
Cathedral Bluffs	comb.-retort	recycle gas heater	0	40	3	pm	.03	14.74	4.42
							.06	31.29	9.39
Clear Creek Shale Oil	comb-retort	retort gas	0	37	3	pm			
Clear Creek Shale Oil	comb-retort	TEG Concentrator	0	37	3	PM	.00	.23	.03
Clear Creek Shale Oil	comb-utility	steam superheat	0	37	3	PM	.14	8.96	1.36
Clear Creek Shale Oil	comb-retort	char combustion	feed bins(sourc	40	3	PM			
Clear Creek Shale Oil	comb-retort	char combustion	coal grinding	40	3	PM	.09	5.71	.87
Clear Creek Shale Oil	comb-retort	char combustion	0	40	3	PM	10.24	643.77	97.74
							10.47	658.67	100.00
cottonwood	com-retort	shale fluid bed co	0	40	3	pm	.04	8.20	4.71
							.04	8.20	4.71
Paraho-Ute	upgrade	package boiler	0	37	3	PM	.00	.33	.09
Paraho-Ute	comb-retort	power generation	0	37	3	PM	.35	51.70	15.00
							.35	52.03	15.09
syntana	utility	steam	0	37	3	pm			
syntana	retort	FGD	steam boiler	37	3	pm	.05	5.89	2.49
syntana	retort	retort indirect he	0	39	3	pm			

TABLE C-4. Continued

project	general process	specific process	additional desc	catl	cate	pol	m-ton/d	kg/1000m^3 oil	% total
syntana	retort	FGD	superior heater	40	3	pm	.03	3.48	1.47
syntana	retort	Tosco ball heater	0	40	3	pm	.34	37.73	15.93
syntana	retort	Tosco moisturizer	0	42	3	pm	.06	6.13	2.59
							.48	53.23	22.47
Union	utility	steam	0	37	3	pm	.09	6.09	3.36
Union	retort	sponge oil strippe	0	39	3	pm	.01	.61	.34
Union	retort	gas recycle heater	0	40	3	PM	.10	7.31	4.03
							.20	14.00	7.73
White River Shale Project	utility	steam	0	37	3	pm	.72	42.55	19.93
White River Shale Project	retort	Tosco ball heater	0	40	3	pm	.44	26.06	12.21
White River Shale Project	retort	recycle gas heater	union	40	3	pm	.19	11.28	5.29
White River Shale Project	retort	elutriator	0	41	3	PM	.05	3.09	1.45
White River Shale Project	retort	Tosco elutriators	and moisturizer	41	3	pm	.13	7.79	3.65
White River Shale Project	retort	processed shale mo	0	42	3	pm			
							1.53	90.77	42.53
Cathedral Bluffs	comb-upgrade	reformer furnace	0	43	4	pm	.03	16.98	5.09
Cathedral Bluffs	upgrade	H2 Recycle heater	0	43	4	pm	.00	.57	.17
Cathedral Bluffs	upgrade	H2 charge heater	0	43	4	pm	.00	1.71	.51
Cathedral Bluffs	upgrade	Oil charge heater	0	43	4	pm	.01	2.85	.86
Cathedral Bluffs	mining-abo	fugative dust	haul roads	54	4	pm	.00	1.43	.43
							.04	23.54	7.06
Clear Creek Shale Oil	mining-above	vehicles	0	53	4	PM	.17	10.50	5.39
Clear Creek Shale Oil	mining-above	coal	0	54	4	PM	.01	.68	.35
							.18	11.19	5.74
cottonwood	comb-upgrade	fugitive	paved roads	53	4	pm	.01	2.18	1.25
cottonwood	comb-upgrade	fugitive	unpaved roads	53	4	pm	.06	11.24	6.46
							.07	13.42	7.71
Paraho-Ute	upgrading	hydrotreater feed	0	43	4	pm	.01	.82	.30
Paraho-Ute	upgrading	reformer furnace	0	43	4	PM	.13	19.57	5.70
							.14	20.38	6.00
syntana	upgrade	FGD	hydrotreater	43	4	pm	.00	.53	.22
syntana	upgrade	FGD	hydrogen reform	43	4	pm	.06	6.61	2.79
syntana	mining-above	fugative	truck traffic	53	4	pm	.28	30.44	12.85
							.34	37.58	15.86
Union	mining-above	vehicles/engines	light duty	53	4	pm	.04	2.64	1.46
Union	mining-above	vehicles/engines	road maint/acce	53	4	pm	.05	3.54	1.96
Union	mining-above	vehicles/engines	diesel	53	4	pm	.15	10.62	5.86
Union	mining-above	fugative dust	0	54	4	PM	.36	24.95	13.78
							.60	41.74	23.05
White River Shale Project	upgrade	reformer furnace	0	43	4	pm	.41	24.55	11.50
							.41	24.55	11.50
							25.78*	2345.58*	800.00*

TABLE C-4. Continued

Project	general process	specific process	additional	pollu	m-ton/d	cat	cat	kg/1000m^3 oil	% total
Union	mining-below	drilling		CO	.00	1	1	.09	.02
Clear Creek Shale Oil	mining-above	blasting		CO	.05	2	1	2.97	.03
Clear Creek Shale Oil	mining-below	blasting		CO	.36	2	1	22.66	.20
Paraho-Ute	mining-below ground	blasting		CO	.52	2	1	77.39	29.81
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
Union	mining-below	blasting		CO	1.14	2	1	79.71	16.09
Utah-cottonwood	mining-below ground	blasting		CO	.40	2	1	80.33	31.51
White River Shale Projec	mining-below	blasting		CO	.87	2	1	51.58	24.37
Clear Creek Shale Oil	mining-below	crushing/screening	primary	CO	.72	6	1	45.03	.40
Catherdral Bluffs	mining-below	mine shaft vents		CO	.30	8	1	158.37	63.86
Clear Creek Shale Oil	mining-below	vent		CO	1.23	8	1	77.62	.69
Paraho-Ute	mining-below ground	mobile equipment		CO	.67	8	1	99.67	38.39
syntana	mining-below	vehicles		CO	.57	8	1	62.58	34.03

TABLE C-5. GASEOUS EMISSIONS

Project	general process	specific process	additional	pollu	m-tons/d	cat	cat	kg/1000m^3 oil	% total
Union	mining-below	drilling		CO	.00	1	1	.09	.02
					.00			.09	.02
Clear Creek Shale Oil	mining-above	blasting		CO	.05	2	1	2.97	.03
Clear Creek Shale Oil	mining-below	blasting		CO	.36	2	1	22.66	.20
Paraho-Ute	mining-below ground	blasting		CO	.52	2	1	77.39	29.81
syntana	mining-below	blasting		CO	.41	2	1	45.76	24.88
Union	mining-below	blasting		CO	1.14	2	1	79.71	16.09
Utah-cottonwood	mining-below ground	blasting		CO	.40	2	1	80.33	31.51
White River Shale Projec	mining-below	blasting		CO	.87	2	1	51.58	24.37
Clear Creek Shale Oil	mining-below	crushing/screening	primary	CO	.72	6	1	45.03	.40
Catherdral Bluffs	mining-below	mine shaft vents		CO	.30	8	1	158.37	63.86
Clear Creek Shale Oil	mining-below	vent		CO	1.23	8	1	77.62	.69
Paraho-Ute	mining-below ground	mobile equipment		CO	.67	8	1	99.67	38.39
syntana	mining-below	vehicles		CO	.57	8	1	62.58	34.03
Union	mining-below	raw shale removal/sc	engines/ve	CO	.55	8	1	38.75	7.82
Utah-cottonwood	mining-below ground	vehicles-combustion		CO	.24	8	1	46.97	18.42
White River Shale Projec	mining-below	vehicles		CO	1.10	8	1	65.17	30.89
Clear Creek Shale Oil	mining-above	top soil	removal	CO	.03	9	2	1.83	.02
Clear Creek Shale Oil	mining-above	top soil	load	CO	.00	10	2	.17	.00
Union	mining-above	raw shale	topsoil ha	CO	.01	11	2	.43	.09
Clear Creek Shale Oil	mining-above	raw shale haul		CO	.03	16	2	2.11	.02
Clear Creek Shale Oil	mining-above	spent shale haul		CO	.06	32	2	3.94	.03
Union	mining-above	spent shale	hauling	CO	.05	32	2	3.77	.76
Union	mining-above	spent shale	grooming/c	CO	.01	36	2	.50	.10
Catherdral Bluffs	utility	steam boilers		CO	.05	37	3	23.78	9.59
Clear Creek Shale Oil	utility	steam superheat		CO	.58	37	3	36.41	.32
Paraho-Ute	upgrade	package boiler		CO	.01	37	3	.83	.32
Paraho-Ute	retort	power generation		CO	.39	37	3	58.57	22.56
syntana	retort	FGD	steam boil	CO	.15	37	3	16.14	8.78
Union	utility	steam		CO	2.66	37	3	185.67	37.49

TABLE C-5. Continued

Project	general process	specific process	additional	pollu	m-tons/d	cat	cat	kg/1000m^3 oil	% total
Utah-cottonwood	retort	fluid bed combustor		CO	.48	37	3	95.57	37.49
White River Shale Projec	utility	steam		CO	.44	37	3	26.33	12.50
					4.75			443.29	129.04
Catherdral Bluffs	retort	flares		CO	.00	38	3	.18	.07
Catherdral Bluffs	retort	sponge oil reboiler		CO	.01	40	3	3.33	1.34
Catherdral Bluffs	retort	recycle gas heater		CO	.05	40	3	25.21	10.16
Clear Creek Shale Oil	retort	char combustion		CO	176.04	40	3	11071.92	98.26
Clear Creek Shale Oil	retort	char combustion	coal grind	CO	.05	40	3	2.97	.03
syntana	retort	FGD	superior h	CO	.09	40	3	9.51	5.17
syntana	retort	Tosco ball heater &		CO	.05	40	3	5.17	2.81
Union	comb.-retort	sponge oil stripper		CO	.03	40	3	2.44	.49
Union	retort	gas recycle heater		CO	.33	40	3	23.13	4.67
White River Shale Projec	retort	Tosco ball heater &		CO	.15	40	3	8.97	4.25
White River Shale Projec	retort	recycle gas heater	union b	CO	.32	40	3	18.80	8.90
Catherdral Bluffs	upgrade	incinerator		CO	.00	43	4	.67	.27
Catherdral Bluffs	upgrade	H2 Recycle Heater		CO	.00	43	4	.90	.36
Catherdral Bluffs	upgrade	H2 Charge Heater		CO	.01	43	4	2.85	1.15
Catherdral Bluffs	upgrade	oil charge heater		CO	.01	43	4	4.66	1.88
Catherdral Bluffs	upgrade	reformer		CO	.05	43	4	28.06	11.31
Clear Creek Shale Oil	retort	TEG Concentrator		CO	.00	43	4	.11	.00
Paraho-Ute	upgrading	hydrotreater feed fu		CO	.01	43	4	.98	.38
Paraho-Ute	upgrading	reformer furnace		CO	.15	43	4	22.18	8.54
syntana	upgrade	FGD	hydrotreat	CO	.01	43	4	1.53	.83
syntana	upgrade	FGD	hydrogen r	CO	.16	43	4	17.87	9.72
syntana	upgrade	hydrogen	furnace	CO	.23	43	4	25.33	13.78
Union	upgrade	fractionater	reboiler	CO	.01	43	4	.38	.08
Union	upgrade	dearseniter	purge heat	CO	.00	43	4	.32	.06
Union	upgrade	dearseniter	charge hea	CO	.02	43	4	1.65	.33
Union	upgrade	unicracker	charge hea	CO	.03	43	4	2.09	.42
Union	upgrade	steam boiler	no air pre	CO	.51	43	4	35.38	7.14
Union	upgrade	steam boiler	with air p	CO	.51	43	4	35.38	7.14
Union	upgrade	reformer furnace		CO	.26	43	4	18.01	3.64
White River Shale Projec	upgrade	reformer furnace		CO	.69	43	4	40.62	19.23
Clear Creek Shale Oil	mining-above	vehicles	road maint	CO	.01	53	4	.70	.01
Union	mining-above	vehicles/engines		CO	.97	53	4	67.57	13.64
Utah-cottonwood	upgrade	fugitive	on site ve	CO	.03	53	4	6.85	2.69
Utah-cottonwood	upgrade	fugitive	on site ve	CO	.13	53	4	25.21	9.89
					1.14			100.32	26.22
Union	mining-below	drilling		HC	.00	1	1	.03	.01
Union	mining-below	blasting		HC		2	1		

TABLE C-5. Continued

Project	general process	specific process	additional pollu		m-tons/d	cat	cat	kg/1000m 3 oil	% total
Clear Creek Shale Oil	mining-below	crushing/screening	primary	HC	.11	6	1	7.19	3.08
Catherdral Bluffs	mining-below	mine shaft vents		HC	.08	8	1	39.47	42.09
Clear Creek Shale Oil	mining-below	vent		HC	.17	8	1	10.96	4.70
Paraho-Ute	mining-below ground	mobile equipment		HC	.21	8	1	30.98	63.26
syntana	mining-below	vehicles		HC	.18	8	1	19.42	13.10
Union	mining-below	raw shale removal/sc	engines/ve	HC	.18	8	1	12.87	5.14
Utah-cottonwood	mining-below ground	vehicles-combustion		HC	.07	8	1	14.51	13.36
White River Shale Projec	mining-below	vehicles		HC	.34	8	1	20.31	9.38
Clear Creek Shale Oil	mining-above	top soil	removal	HC	.01	9	2	.57	.24
Union	mining-above	raw shale	topsoil ha	HC	.00	11	2	.14	.06
Clear Creek Shale Oil	mining-above	raw shale haul		HC	.01	16	2	.68	.29
Clear Creek Shale Oil	mining-above	spent shale haul		HC	.01	32	2	.51	.22
Union	mining-above	spent shale	hauling	HC	.02	32	2	1.24	.49
Union	mining-above	spent shale	grooming/c	HC	.00	36	2	.16	.06
Catherdral Bluffs	utility	steam boilers		HC	.08	37	3	42.80	45.64
Paraho-Ute	upgrade	package boiler		HC	.00	37	3	.16	.33
Paraho-Ute	retort	power generation		HC	.07	37	3	10.26	20.95
syntana	retort	FGD	steam boil	HC	.03	37	3	2.83	1.91
Union	utility	steam		HC	.05	37	3	3.80	1.52
Utah-cottonwood	retort	fluid bed combustor		HC	.02	37	3	3.08	2.84
White River Shale Projec	utility	steam		HC	.32	37	3	19.13	8.83
Catherdral Bluffs	retort	flares		HC		38	3		
Catherdral Bluffs	retort	sponge oil reboiler		HC	.00	40	3	.57	.61
Catherdral Bluffs	retort	recycle gas heater		HC	.01	40	3	4.52	4.82
syntana	retort	FGD	superior h	HC	.02	40	3	1.79	1.21
syntana	retort	Tosco ball heater &		HC	.51	40	3	55.87	37.67
Union	comb.-retort	sponge oil stripper		HC	.00	40	3	.30	.12
Union	retort	gas recycle heater		HC	.06	40	3	4.26	1.70
White River Shale Projec	retort	Tosco ball heater &		HC	.83	40	3	49.00	22.62
White River Shale Projec	retort	recycle gas heater	union	HC	.06	40	3	3.81	1.76

TABLE C-5. Continued

Project	general process	specific process	additional	pollu	m-tons/d	cat	cat	kg/1000m^3 oil	% total
Catherdral Bluffs	upgrade	H2 Recycle Heater		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	incinerator		HC	.00	43	4	.14	.15
Catherdral Bluffs	upgrade	H2 Charge Heater		HC	.00	43	4	.57	.61
Catherdral Bluffs	upgrade	oil charge heater		HC	.00	43	4	.81	.86
Catherdral Bluffs	upgrade	reformer		HC	.01	43	4	4.76	5.07
Catherdral Bluffs	upgrade	storage		HC		43	4		
Paraho-Ute	upgrading	hydrotreater feed fu		HC	.00	43	4	.16	.33
Paraho-Ute	upgrading	reformer furnace		HC	.03	43	4	3.91	7.99
syntana	upgrade	FGD	hydrotreat	HC	.00	43	4	.29	.20
syntana	upgrade	FGD	hydrogen r	HC	.03	43	4	3.20	2.16
syntana	upgrade	hydrogen	furnace	HC	.04	43	4	4.71	3.17
Union	upgrade	fractionater	reboiler	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	purge heat	HC	.00	43	4	.06	.03
Union	upgrade	dearseniter	charge hea	HC	.00	43	4	.32	.13
Union	upgrade	unicracker	charge hea	HC	.01	43	4	.38	.15
Union	upgrade	steam boiler	no air pre	HC	.01	43	4	.70	.28
Union	upgrade	steam boiler	with air p	HC	.01	43	4	.70	.28
Union	upgrade	reformer furnace		HC	.05	43	4	3.36	1.34
White River Shale Projec	upgrade	reformer furnace		HC	.14	43	4	8.22	3.79
Clear Creek Shale Oil	upgrade	fugative		HC	3.26	44	4	204.89	87.80
White River Shale Projec	upgrade	valves,flanges,pumps		HC	1.02	44	4	60.17	27.78
Clear Creek Shale Oil	upgrade	storage		HC	.13	46	4	7.99	3.42
Paraho-Ute	upgrading	storage	crude shal	HC	.01	46	4	.82	1.66
Paraho-Ute	upgrading	storage (day)	crude shal	HC	.01	46	4	1.22	2.48
syntana	upgrade	oil storage tanks		HC	.03	46	4	3.12	2.11
Union	upgrade	storage tanks		HC	1.10	46	4	77.08	30.81
Union	upgrade	oil storage		HC	.02	46	4	1.46	.58
Union	upgrade	miscellaneous	table 4-7?	HC	.84	46	4	58.85	23.52
Utah-cottonwood	upgrade	product storage		HC	.11	46	4	22.49	20.70
White River Shale Projec	upgrade	crude storage		HC	.46	46	4	27.29	12.60
White River Shale Projec	upgrade	crude shale oil stor		HC	.48	46	4	28.69	13.24
Paraho-Ute	upgrading	storage	hydrotreat	HC	.01	47	4	1.05	2.15
Paraho-Ute	upgrading	storage	fuel oil	HC	.00	48	4	.04	.08
Utah-cottonwood	upgrade	fuel oil storage tan		HC		48	4		
Paraho-Ute	upgrading	storage	diesel & g	HC	.00	49	4	.37	.75
Clear Creek Shale Oil	mining-above	vehicles	road maint	HC	.01	53	4	.57	.24
Union	mining-above	vehicles/engines		HC	.11	53	4	7.36	2.94

TABLE C-5. Continued

Project	general process	specific process	additional pollu		m-tons/d	cat	cat	kg/1000m^3 oil	% total
syntana	upgrade	fugitive emissions		HC	.52	54	4	57.07	38.48
Union	upgrade	fugative		HC	1.10	54	4	77.08	30.81
Utah-cottonwood	upgrade	fugitive		HC	.34	54	4	68.55	63.11
Union	mining-below	drilling		NOx	.01	1	1	.42	.04
Clear Creek Shale Oil	mining-above	blasting		NOx	.02	2	1	1.43	.02
Clear Creek Shale Oil	mining-below	blasting		NOx	.03	2	1	1.71	.03
Paraho-Ute	mining-below ground	blasting		NOx		2	1		
syntana	mining-below	blasting		NOx	.23	2	1	25.03	2.17
Union	mining-below	blasting		NOx	.29	2	1	20.23	1.70
Utah-cottonwood	mining-below ground	blasting		NOx	.22	2	1	43.88	1.02
White River Shale Projec	mining-below	blasting		NOx	.47	2	1	27.94	2.63
Clear Creek Shale Oil	mining-below	crushing/screening	primary	NOx	.75	6	1	47.03	.82
					.75			47.03	.82
Catherdral Bluffs	mining-below	mine shaft vents		NOx	.91	8	1	475.12	15.00
Clear Creek Shale Oil	mining-below	vent		NOx	1.30	8	1	81.61	1.42
Paraho-Ute	mining-below ground	mobile equipment		NOx	3.17	8	1	473.97	27.44
syntana	mining-below	vehicles		NOx	2.70	8	1	297.37	25.79
Union	mining-below	raw shale removal/sc	engines/ve	NOx	2.97	8	1	207.74	17.48
Utah-cottonwood	mining-below ground	vehicles-combustion		NOx	1.12	8	1	223.05	5.19
White River Shale Projec	mining-below	vehicles		NOx	5.24	8	1	310.42	29.28
Clear Creek Shale Oil	mining-above	top soil	removal	NOx	.20	9	2	12.44	.22
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.00	10	2	.06	.00
Clear Creek Shale Oil	mining-above	top soil	load	NOx	.01	10	2	.74	.01
Union	mining-above	raw shale	topsoil ha	NOx	.03	11	2	2.31	.19
Clear Creek Shale Oil	mining-above	raw shale haul		NOx	.19	16	2	11.99	.21
Clear Creek Shale Oil	mining-above	spent shale haul		NOx	.14	32	2	8.73	.15
Union	mining-above	spent shale	hauling	NOx	.31	32	2	21.46	1.81
Union	mining-above	spent shale	grooming/c	NOx	.05	36	2	3.42	.29
Catherdral Bluffs	utility	steam boilers		NOx	1.89	37	3	989.24	31.24

TABLE C-5. Continued

Project	general process	specific process	additional	pollu	m-tons/d	cat	cat	kg/1000m 3 oil	% total
Clear Creek Shale Oil	utility	steam superheat		NOx	1.87	37	3	117.57	2.04
Paraho-Ute	upgrade	package boiler		NOx	.02	37	3	3.74	.22
Paraho-Ute	retort	power generation		NOx	6.25	37	3	936.25	54.21
syntana	retort	FGD	steam boil	NOx	1.71	37	3	188.84	16.38
syntana	utility	steam		NOx		37	3		
Union	utility	steam		NOx	3.54	37	3	247.31	20.81
Utah-cottonwood	retort	fluid bed combustor		NOx	20.12	37	3	4020.34	93.53
White River Shale Projec	utility	steam		NOx	4.77	37	3	282.59	26.65
Catherdral Bluffs	retort	flares		NOx	.00	38	3	2.62	.08
syntana	retort	retort indirect heat		NOx		39	3		
White River Shale Projec	retort	gas treatment plant		NOx	.07	39	3	4.14	.39
Catherdral Bluffs	retort	sponge oil reboiler		NOx	.09	40	3	48.04	1.52
Catherdral Bluffs	retort	recycle gas heater		NOx	.70	40	3	369.06	11.65
Clear Creek Shale Oil	retort	char combustion		NOx	86.90	40	3	5465.19	94.85
Clear Creek Shale Oil	retort	char combustion	coal grind	NOx	.15	40	3	9.53	.17
syntana	retort	FGD	superior h	NOx	.99	40	3	108.80	9.44
syntana	retort	Tosco ball heater &		NOx	.57	40	3	62.58	5.43
Union	comb.-retort	sponge oil stripper		NOx	.42	40	3	29.22	2.46
Union	retort	gas recycle heater		NOx	4.14	40	3	289.16	24.33
White River Shale Projec	retort	Tosco ball heater &		NOx	1.02	40	3	60.17	5.68
White River Shale Projec	retort	recycle gas heater	union	NOx	2.01	40	3	119.27	11.25
Catherdral Bluffs	upgrade	incinerator		NOx	.02	43	4	9.04	.29
Catherdral Bluffs	upgrade	H2 Recycle Heater		NOx	.02	43	4	12.84	.41
Catherdral Bluffs	upgrade	H2 Charge Heater		NOx	.07	43	4	39.00	1.23
Catherdral Bluffs	upgrade	oil charge heater		NOx	.12	43	4	62.78	1.98
Catherdral Bluffs	upgrade	reformer		NOx	2.21	43	4	1159.03	36.60
Clear Creek Shale Oil	retort	TEG Concentrator		NOx	.00	43	4	.29	.00
Paraho-Ute	upgrading	hydrotreater feed fu		NOX	.09	43	4	13.18	.76
Paraho-Ute	upgrading	reformer furnace		NOX	2.00	43	4	300.03	17.37
syntana	upgrade	FGD	hydrotreat	NOx	.16	43	4	17.77	1.54
syntana	upgrade	FGD	hydrogen r	NOx	1.90	43	4	210.16	18.23
syntana	upgrade	hydrogen	furnace	NOx	2.20	43	4	242.30	21.02
Union	upgrade	fractionater	reboiler	NOx	.05	43	4	3.30	.28
Union	upgrade	dearseniter	purge heat	NOx	.04	43	4	2.79	.23
Union	upgrade	dearseniter	charge hea	NOx	.20	43	4	14.14	1.19
Union	upgrade	unicracker	charge hea	NOx	.26	43	4	18.07	1.52
Union	upgrade	steam boiler	no air pre	NOx	.56	43	4	39.25	3.30
Union	upgrade	steam boiler	with air p	NOx	.67	43	4	47.05	3.96
Union	upgrade	reformer furnace		NOx	3.22	43	4	225.12	18.94
White River Shale Projec	upgrade	reformer furnace		NOx	4.32	43	4	255.73	24.12

TABLE C-5. Continued

Project	general process	specific process	additional pollu		m-tons/d	cat	cat	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-above	vehicles	road maint	NOx	.05	53	4	3.42	.06
Union	mining-above	vehicles/engines		NOx	.25	53	4	17.69	1.49
Utah-cottonwood	upgrade	fugitive	on site ve	NOx	.06	53	4	11.24	.26
Union	mining-below	drilling		SOx	.00	1	1	.02	.63
Union	mining-below	blasting		SOx	.03	2	1	2.38	1.27
Clear Creek Shale Oil	mining-below	crushing/screening	primary	SOx	.07	6	1	4.34	.78
Catherdral Bluffs	mining-below	mine shaft vents		SOx	.06	8	1	29.96	4.19
Clear Creek Shale Oil	mining-below	vent		SOx	.12	8	1	7.48	1.35
Paraho-Ute	mining-below ground	mobile equipment		SOx	.23	8	1	34.31	5.25
syntana	mining-below	vehicles		SOx	.20	8	1	21.53	3.74
Union	mining-below	raw shale removal/sc	engines/ve	SOx	.19	8	1	13.00	5.07
Utah-cottonwood	mining-below ground	vehicles-combustion		SOx	.08	8	1	16.14	2.02
White River Shale Projec	mining-below	vehicles		SOx	.38	8	1	22.51	14.18
Clear Creek Shale Oil	mining-above	top soil	removal	SOx	.01	9	2	.86	.15
Clear Creek Shale Oil	mining-above	top soil	load	SOx	.00	10	2	.06	.01
Union	mining-above	raw shale	topsoil ha	SOx	.00	11	2	.16	6.97
Clear Creek Shale Oil	mining-above	raw shale haul		SOx	.01	16	2	.68	.12
Clear Creek Shale Oil	mining-above	spent shale haul		SOx	.01	32	2	.51	.09
Union	mining-above	spent shale	hauling	SOx	.02	32	2	1.31	20.26
Union	mining	dumping shale		SOx	.04	33	2	2.85	20.90
Union	mining-above	spent shale	grooming/c	SOx	.00	36	2	.24	22.80
Catherdral Bluffs	utility	steam boilers		SOx	.28	37	3	147.44	20.64
Clear Creek Shale Oil	utility	steam superheater-so		SOx	.49	37	3	30.82	5.55
Paraho-Ute	upgrade	package boiler		SOx	.05	37	3	7.07	1.08
Paraho-Ute	retort	power generation		SOx	3.97	37	3	595.18	91.11
syntana	retort	FGD	steam boil	SOx	.91	37	3	100.13	17.38

TABLE C-5. Continued

Project	general process	specific process	additional pollu		m-tons/d	cat	cat	kg/1000m 3 oil	% total
syntana	utility	steam		SOx		37	3		
Union	utility	steam		SOx	2.02	37	3	141.16	23.43
Utah-cottonwood	retort	fluid bed combustor		SOx	3.91	37	3	781.58	97.98
White River Shale Projec	utility	steam		SOx	1.63	37	3	96.70	60.93
Catherdral Bluffs	retort	flares		SOx	.02	38	3	8.09	1.13
syntana	retort	claus plant		SOx		39	3		
syntana	retort	FGD	claus plan	SOx	.51	39	3	56.27	9.77
syntana	retort	retort indirect heat		SOx		39	3		
White River Shale Projec	retort	gas treatment-claus		SOx	.36	39	3	21.06	13.27
Catherdral Bluffs	retort	sponge oil reboiler		SOx	.04	40	3	20.45	2.86
Catherdral Bluffs	retort	recycle gas heater		SOx	.30	40	3	156.95	21.97
Clear Creek Shale Oil	retort	char combustion	coal grind	SOx	.05	40	3	2.97	.53
Clear Creek Shale Oil	retort	char combustion		SOx	8.06	40	3	506.80	91.33
syntana	retort	FGD	superior h	SOx	.53	40	3	57.97	10.06
syntana	retort	Tosco ball heater &		SOx	.40	40	3	44.46	7.72
Union	comb.-retort	sponge oil stripper		SOx	.29	40	3	20.39	25.33
Union	retort	gas recycle heater		SOx	3.34	40	3	233.74	25.33
White River Shale Projec	retort	retort		SOx		40	3		
White River Shale Projec	retort	Tosco ball heater &		SOx	.10	40	3	6.18	3.89
White River Shale Projec	retort	recycle gas heater	union	SOx	.07	40	3	3.92	2.47
Catherdral Bluffs	upgrade	H2 Recycle Heater		SOx	.01	43	4	5.23	.73
Catherdral Bluffs	upgrade	H2 Charge Heater		SOx	.03	43	4	17.60	2.46
Catherdral Bluffs	upgrade	oil charge heater		SOx	.05	43	4	28.54	3.99
Catherdral Bluffs	upgrade	incinerator		SOx	.24	43	4	126.98	17.78
Catherdral Bluffs	upgrade	reformer		SOx	.33	43	4	173.12	24.23
Clear Creek Shale Oil	retort	TEG Concentrator		SOx	.00	43	4	.11	.02
Paraho-Ute	upgrading	hydrotreater feed fu		SOx	.00	43	4	.03	.00
Paraho-Ute	upgrading	reformer furnace		SOx	.11	43	4	16.65	2.55
syntana	upgrade	FGD	hydrotreat	SOx	.09	43	4	9.51	1.65
Union	upgrade	fractionater	reboiler	SOx	.01	43	4	.76	27.23
Union	upgrade	dearseniter	purge heat	SOx	.01	43	4	.63	27.23
Union	upgrade	dearseniter	charge hea	SOx	.05	43	4	3.36	27.23
Union	upgrade	unicracker	charge hea	SOx	.06	43	4	4.25	27.23
Union	upgrade	steam boiler	no air pre	SOx	.11	43	4	7.55	27.23
Union	upgrade	steam boiler	with air p	SOx	.11	43	4	7.55	27.23
Union	upgrade	reformer furnace		SOx	.15	43	4	10.15	27.23
White River Shale Projec	upgrade	reformer furnace		SOx	.14	43	4	8.33	5.25
Clear Creek Shale Oil	mining-above	vehicles	road maint	SOx	.00	53	4	.29	.05
Union	mining-above	vehicles/engines		SOx	.02	53	4	1.17	33.56
					415.10*			36042.71*	3051.45*

TABLE C-6. PARTICULATE EMISSIONS

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-below	drilling	inter. waste	1	1	.00	.26	.13
Clear Creek Shale Oil	mining-below	drilling	0	1	1	.01	.74	.38
cottonwood	mining-below ground	mining	0	1	1	.11	21.76	12.50
Paraho-Ute	mining-below ground	mining	0	1	1	.04	5.71	1.70
syntana	mining-below	mining	0	1	1	.13	13.91	5.87
Union	mining-below	drilling	0	1	1	.00	.34	.19
White River Shale Project	mining-below	mining	0	1	1	.26	15.31	7.17
Clear Creek Shale Oil	mining-below	blasting	inter. waste	1	2	.10	6.22	3.19
Clear Creek Shale Oil	mining-below	blasting	0	1	2	.26	16.27	8.35
Paraho-Ute	mining-below ground	blasting	0	1	2	.17	25.48	7.30
syntana	mining-below	blasting	0	1	2	.18	19.32	8.16
Union	mining-below	blasting	0	1	2	.01	.50	.28
Union	mining-below	removal	0	1	3	.03	2.31	1.28
Paraho-Ute	mining-below ground	conveying	0	1	4	.03	5.10	1.50
Union	mining-below	conveying	0	1	4	.17	11.59	6.40
cottonwood	mining-below ground	transfer	transfer poi	1	5	.05	9.61	5.52
Clear Creek Shale Oil	mining-below	primary crush	0	1	6	.00	.21	.11
cottonwood	mining-below ground	raw shale	primary crus	1	6	.03	5.26	3.02

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
Paraho-Ute	mining-below ground	crushing/scre	primary	1	6	.08	12.16	3.60
Union	mining-below	raw shale cru	primary	1	6	.07	4.63	2.56
Union	mining-below	crushing	primary	1	6	.63	43.78	24.18
White River Shale Project	mining-below	crushing	primary	1	6	.14	8.49	3.98
Paraho-Ute	mining-below ground	crushing/scre	secondary	1	7	.08	12.16	3.60
White River Shale Project	mining-below	crushing	secondary	1	7	.04	2.63	1.23
CB	mining-below	mine shaft ve	0	1	8	.17	88.46	26.54
Clear Creek Shale Oil	mining-below	vehicles-vent	0	1	8	.40	25.34	13.01
cottonwood	mining-below ground	vehicles-comb	0	1	8	.06	11.42	6.56
Paraho-Ute	mining-below ground	mobile equipm	0	1	8	.02	3.06	1.00
syntana	mining-below	vehicles	0	1	8	.14	15.31	6.46
Union	mining-below	raw shale rem	engines/vehi	1	8	.10	7.15	3.95
White River Shale Project	mining-below	vehicles	0	1	8	.27	15.85	7.42
CB	mining-above	reclaim drawh	0	2	9	.01	2.71	.81
Clear Creek Shale Oil	mining-above	surface soils	removal-dril	2	9	.01	.74	.38
Clear Creek Shale Oil	mining-above	surface soils	removal	2	9	.13	8.28	4.25
Clear Creek Shale Oil	mining-above	surface soils	removal-blas	2	9	.25	15.58	8.00
Clear Creek Shale Oil	mining-above	surface soils	haul	2	11	.36	22.83	11.72
Union	mining-above	raw shale	topsoil haul	2	11	.01	.49	.27

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
Clear Creek Shale Oil	mining-above	surface soils	wind	2	12	.00	.21	.11
Paraho-Ute	mining-above ground	surface soils	wind	2	12	.07	10.19	2.40
CB	mining-above	raw shale	crushing bui	2	13	.00	1.62	.49
CB	mining-above	raw shale	screening pl	2	13	.00	1.43	.43
Clear Creek Shale Oil	mining-above	raw shale	crushing	2	13	.00	.29	.15
Clear Creek Shale Oil	mining-above	raw shale	crushing-2nd	2	13	.01	.68	.35
cottonwood	mining-above ground	raw shale	crusher-lump	2	13	.02	4.81	2.76
cottonwood	mining-above ground	raw shale	screening	2	13	.03	5.60	3.22
syntana	mining-above	raw shale	primary crus	2	13	.01	1.42	.60
syntana	mining-above	raw shale	secondary cr	2	13	.29	32.14	13.57
Union	mining-above	raw shale	crushing/sec	2	13	.12	8.61	4.76
Clear Creek Shale Oil	mining-above	raw shale	crushing-3rd	2	14	.01	.91	.47
Paraho-Ute	mining-above ground	raw shale	tertiary cru	2	14	.24	36.35	10.50
Union	mining-above	raw shale	crushing/ter	2	14	.41	28.71	15.85
CB	mining-above	raw shale	transfer hou	2	15	.02	9.04	2.71
CB	mining-above	raw shale	conveyor-sta	2	15	.05	25.21	7.56
Clear Creek Shale Oil	mining-above	raw shale	conveying	2	15	.02	1.13	.58
Clear Creek Shale Oil	mining-above	raw shale	conveying	2	15	.81	51.08	26.23
cottonwood	mining-above ground	raw shale	surge bin	2	15	.02	3.23	1.85
Paraho-Ute	mining-above ground	raw shale	trans. fr. 1	2	15	.01	1.77	.50
Paraho-Ute	mining-above ground	raw shale	convey/trans	2	15	.05	7.34	2.20

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m 3 oil	% total
Paraho-Ute	mining-above ground	raw shale	sample & wei	2	15	.10	14.27	4.20
Paraho-Ute	mining-above ground	raw shale	screening/tr	2	15	.16	24.60	7.10
Union	mining-above	raw shale	conveying/tr	2	15	.01	.59	.32
White River Shale Project	mining-above	conveyor surg	0	2	15	.01	.86	.40
Clear Creek Shale Oil	mining-above	int waste	dump	2	16	.01	.34	.18
Clear Creek Shale Oil	mining-above	int waste	haul	2	16	.15	9.36	4.81
Clear Creek Shale Oil	mining-above	raw shale	dump	2	17	.12	7.53	3.87
CB	mining-abo	raw shale sto	5 day	2	18	.01	3.80	1.14
cottonwood	mining-above ground	raw shale	storage-load	2	18	.00	.30	.17
cottonwood	mining-above ground	raw shale	storage-recl	2	18	.00	.60	.34
cottonwood	mining-above ground	raw shale	storage(live	2	18	.03	6.89	3.96
Paraho-Ute	mining-above ground	raw shale	live storage	2	18	.00	.14	.04
syntana	mining-above	raw shale	storage	2	18	.03	3.47	1.47
Union	mining-above	raw shale sto	load out	2	18	.00	.06	.04
Union	mining-above	storage silos	0	2	18	.01	.50	.27
Union	mining-above	raw shale sto	load in	2	18	.01	.77	.43
White River Shale Project	mining-above	retort feed s	three	2	18	.05	2.69	1.26
White River Shale Project	mining-above	raw shale sto	load/groom/w	2	18	.12	6.93	3.25
Clear Creek Shale Oil	mining-above	raw shale	wind errosio	2	19	.00	.02	.01
Union	mining-above	raw shale sto	wind erosion	2	19	.00	.19	.10
Union	mining-above	retort pad	wind errosio	2	19	.01	.36	.20
CB	mining-abo	raw shale sto	5 year	2	21	.01	5.42	1.63
						.01	5.42	1.63

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
cottonwood	mining-above ground	raw shale	storage(dead	2	21	.01	2.77	1.59
Paraho-Ute	mining-above ground	raw shale	emergency st	2	21	.11	16.10	4.70
Clear Creek Shale Oil	mining-above	raw shale	wind	2	22	.04	2.57	1.32
syntana	mining-above	raw shale	storage-wind	2	22	.04	4.77	2.01
cottonwood	mining-above ground	raw shale	retort feed	2	23	.00	.79	.46
cottonwood	mining-above ground	raw shale	retort feed	2	23	.00	.79	.46
CB	mining-above	raw shale	conveyor-ret	2	24	.00	2.38	.71
cottonwood	mining-above ground	raw shale	retort feed	2	24	.00	.79	.46
Paraho-Ute	mining-above ground	raw shale	retort feed	2	24	.04	5.38	1.60
syntana	mining-above	raw shale	retort feed-	2	24	.00	.18	.07
syntana	mining-above	raw shale	conveyor fee	2	24	.00	.35	.15
White River Shale Project	mining-above	conveyor feed	0	2	24	.01	.51	.24
White River Shale Project	mining-above	conveyor feed	transfer	2	24	.05	2.69	1.26
cottonwood	mining-above ground	raw shale	retort disch	2	25	.00	.79	.46
cottonwood	mining-above ground	fines	FBC discharg	2	26	.01	1.09	.63
cottonwood	mining-above ground	fines	conveyor-tra	2	26	.01	1.09	.63
Paraho-Ute	mining-above ground	fines	conveyor-tra	2	26	.01	.82	.40
Paraho-Ute	mining-above ground	fines	to bin	2	26	.01	1.14	.40
Paraho-Ute	mining-above ground	raw shale	fines transf	2	26	.01	1.79	.50
Paraho-Ute	mining-above ground	fines	storage	2	26	.11	16.50	3.80
syntana	mining-above	fines transfe	0	2	26	.01	.85	.36

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
White River Shale Project	mining-above	fines reclaim	0	2	26	.00	.17	.08
cottonwood	mining-above ground	fines	FBC feed bin	2	27	.01	1.09	.63
cottonwood	mining-above ground	fines	storage silo	2	27	.01	1.09	.63
syntana	mining-above	fines	storage-wind	2	27	.02	2.64	1.12
White River Shale Project	mining-above	fines-storage	reclaim/wind	2	27	.07	4.41	2.06
White River Shale Project	mining-above	fines	storage/conv	2	27	.14	8.54	4.00
syntana	mining-above	fines	maintenance	2	29	.05	5.87	2.48
CB	mining-abo	spent shale	stacker conv	2	30	.05	25.21	7.56
CB	mining-above	spent shale	conveyor	2	30	.04	20.93	6.28
Clear Creek Shale Oil	mining-above	spent shale	conveying	2	30	.18	11.59	5.95
cottonwood	mining-above	spent shale	conveyor	2	30			
Paraho-Ute	mining-above ground	spent shale	conveying	2	30	.01	1.30	.37
Paraho-Ute	mining-above ground	spent shale	to bin	2	30	.02	2.28	.70
Paraho-Ute	mining-above ground	spent shale	conveying-A	2	30	.03	4.24	1.30
Union	mining-above	spent shale c	0	2	30	.08	5.55	3.07
White River Shale Project	mining-above	spent shale d	un/tos/sup	2	30	.01	.37	.17
White River Shale Project	mining-above	spent shale c	0	2	30	.05	3.22	1.51
CB	mining-above	spent shale	transfer hou	2	31	.02	9.04	2.71
CB	mining-abo	spent shale	conveyer dis	2	31	.04	20.93	6.28
CB	mining-above	spent shale	transfer	2	31	.04	20.93	6.28
Paraho-Ute	mining-above ground	spent shale	retort overf	2	31	.04	5.38	1.60
Paraho-Ute	mining-above ground	spent shale	retort disch	2	31	.23	34.24	10.00

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
Union	mining-above	spent shale	stacking	2	31	.02	1.54	.85
Clear Creek Shale Oil	mining-above	spent shale	haul	2	32	.02	1.37	.70
Union	mining-above	spent shale	hauling	2	32	.05	3.16	1.75
CB	mining-abo	spent shale	distribution	2	33	.07	38.52	11.56
cottonwood	mining-above	spent shale	disposal(loa	2	33			
cottonwood	mining-above	spent shale-f	loading/dump	2	33			
Union	mining-above	spent shale	truck dumpin	2	33	.03	1.98	1.09
CB	mining-abo	spent shale s	0	2	34	.01	2.85	.86
cottonwood	mining-above	spent shale-f	storage bin	2	34	.01	1.12	.64
Paraho-Ute	mining-above ground	spent shale	storage	2	34	.22	33.50	7.90
syntana	mining-above	spent shale	storage-load	2	34	.02	1.89	.80
Union	mining-above	spent shale d	0	2	34			
White River Shale Project	mining-above	spent shale d	load in	2	34	.02	1.07	.50
cottonwood	mining-above	spent shale	disposal-win	2	35	.03	5.01	2.87
syntana	mining-above	spent shale	storage-wind	2	35	.16	17.62	7.44
Union	mining-above	spent shale-s	wind errosio	2	35	.01	.57	.32
White River Shale Project	mining-above	spent shale d	wind erosion	2	35	.05	2.79	1.31

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
cottonwood	mining-above	spent shale	disposal-gro	2	36	.33	66.55	38.23
syntana	mining-above	spent shale	storage-main	2	36	.24	26.33	11.12
Union	mining-above	spent shale	grooming/com	2	36	.03	1.93	1.07
White River Shale Project	mining-above	spent shale d	grooming	2	36	.36	21.60	10.12
CB	comb-utility	steam boilers	0	3	37	.03	14.27	4.28
Clear Creek Shale Oil	comb-retort	retort gas	0	3	37			
Clear Creek Shale Oil	comb-retort	TEG Concentra	0	3	37	.00	.23	.03
Clear Creek Shale Oil	comb-utility	steam superhe	0	3	37	.14	8.96	1.36
Paraho-Ute	upgrade	package boile	0	3	37	.00	.33	.09
Paraho-Ute	comb-retort	power generat	0	3	37	.35	51.70	15.00
syntana	utility	steam	0	3	37			
syntana	retort	FGD	steam boiler	3	37	.05	5.89	2.49
Union	utility	steam	0	3	37	.09	6.09	3.36
White River Shale Project	utility	steam	0	3	37	.72	42.55	19.93
syntana	retort	retort indire	0	3	39			
Union	retort	sponge oil st	0	3	39	.01	.61	.34
CB	retort	incinerator	0	3	40	.00	.38	.11
CB	comb.-retort	sponge oil re	0	3	40	.00	1.90	.57
CB	comb.-retort	recycle gas h	0	3	40	.03	14.74	4.42
Clear Creek Shale Oil	comb-retort	char combusti	feed bins(so	3	40			
Clear Creek Shale Oil	comb-retort	char combusti	coal grindin	3	40	.09	5.71	.87

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
syntana	mining-above	fugative	truck traffi	4	53	.28	30.44	12.85
Union	mining-above	vehicles/engi	light duty	4	53	.04	2.64	1.46
Union	mining-above	vehicles/engi	road maint/a	4	53	.05	3.54	1.96
Union	mining-above	vehicles/engi	diesel	4	53	.15	10.62	5.86
CB	mining-abo	fugative dust	haul roads	4	54	.00	1.43	.43
Clear Creek Shale Oil	mining-above	coal	0	4	54	.01	.68	.35
Union	mining-above	fugative dust	0	4	54	.36	24.95	13.78

TABLE C-6. Continued

project	general process	specific proc	additional d	cat	cat	m-ton/d	kg/1000m^3 oil	% total
Clear Creek Shale Oil	comb-retort	char combusti	0	3	40	10.24	643.77	97.74
cottonwood	com-retort	shale fluid b	0	3	40	.04	8.20	4.71
syntana	retort	FGD	superior hea	3	40	.03	3.48	1.47
syntana	retort	Tosco ball he	0	3	40	.34	37.73	15.93
Union	retort	gas recycle h	0	3	40	.10	7.31	4.03
White River Shale Project	retort	Tosco ball he	0	3	40	.44	26.06	12.21
White River Shale Project	retort	recycle gas h	union	3	40	.19	11.28	5.29
White River Shale Project	retort	elutriator	0	3	41	.05	3.09	1.45
White River Shale Project	retort	Tosco elutria	and moisturi	3	41	.13	7.79	3.65
syntana	retort	Tosco moistur	0	3	42	.06	6.13	2.59
White River Shale Project	retort	processed sha	0	3	42			
CB	comb-upgrade	reformer furn	0	4	43	.03	16.98	5.09
CB	upgrade	H2 Recycle he	0	4	43	.00	.57	.17
CB	upgrade	H2 charge hea	0	4	43	.00	1.71	.51
CB	upgrade	Oil charge he	0	4	43	.01	2.85	.86
Paraho-Ute	upgrading	hydrotreater	0	4	43	.01	.82	.30
Paraho-Ute	upgrading	reformer furn	0	4	43	.13	19.57	5.70
syntana	upgrade	FGD	hydrotreater	4	43	.00	.53	.22
syntana	upgrade	FGD	hydrogen ref	4	43	.06	6.61	2.79
White River Shale Project	upgrade	reformer furn	0	4	43	.41	24.55	11.50
Clear Creek Shale Oil	mining-above	vehicles	0	4	53	.17	10.50	5.39
cottonwood	comb-upgrade	fugitive	paved roads	4	53	.01	2.18	1.25
cottonwood	comb-upgrade	fugitive	unpaved road	4	53	.06	11.24	6.46

TECHNICAL REPORT DATA
(Please read Instructions on the reverse before completing)

1. REPORT NO.	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Air Pollution Control Alternatives For Shale Oil Production Operations		5. REPORT DATE December, 1987 6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S) H. J. Taback, R. J. Goldstick		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS KVB,* Inc., 18006 Skypark Boulevard, Irvine, CA 92714		10. PROGRAM ELEMENT NO. 11. CONTRACT/GRANT NO. 68-03-3166
12. SPONSORING AGENCY NAME AND ADDRESS EPA, Office of Research and Development Air and Energy Engineering Research Laboratory Research Triangle Park, NC 27711		13. TYPE OF REPORT AND PERIOD COVERED Final Dec. 85-Dec.87 14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES AEERL Project Officer is Edward R. Bates (EPA, HWERL, Cincinnati,OH 45268, 513-569-7774). (*) Under Subcontract to Metcalf and Eddy, Inc., Wakefield, MA 01880.

16. ABSTRACT The available air emission data and air pollution control technology data for the production of shale oil are consolidated, evaluated and presented in a manner to be useful to project developers in preparing environmental impact statements and permit applications as well as to their respective regulatory approval agencies. The processes covered include subsurface and surface mining; raw shale sizing and handling; all of the various retorting schemes, both in situ and above-ground; spent shale combustion; spent shale disposal and product upgrading. Air pollution control technologies covered include most of the traditional processes for nitrogen oxide (NO_x), sulfur compounds, particulate, volatile organic compounds (VOC) and carbon monoxide (CO) control along with recently-developed processes such as: Catalytic mufflers for vehicles, and staged combustion for NO_x; caustic-charcoal-sodium hypochlorite scrubbing, spent shale absorption and dry sorbent injection for organic sulfur and SO_x; and improved filterbag materials, moving bed granular filters and dry venturis for fine particulates.

A conclusion is reached that if state-of-the-art control technology is applied, the overall emission levels per unit of oil produced will be essentially the same for in situ or above-ground retorts and for retorts that are either directly or indirectly heated.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Oil Shale Retorting Desulfurization NO_x Control Scrubbers	Pollution Control Stationary Sources Air Emissions	13B 08G 07A 07D 07B 13I

18. DISTRIBUTION STATEMENT Release to Public	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 573
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (Rev. 4-77) PREVIOUS EDITION IS OBSOLETE