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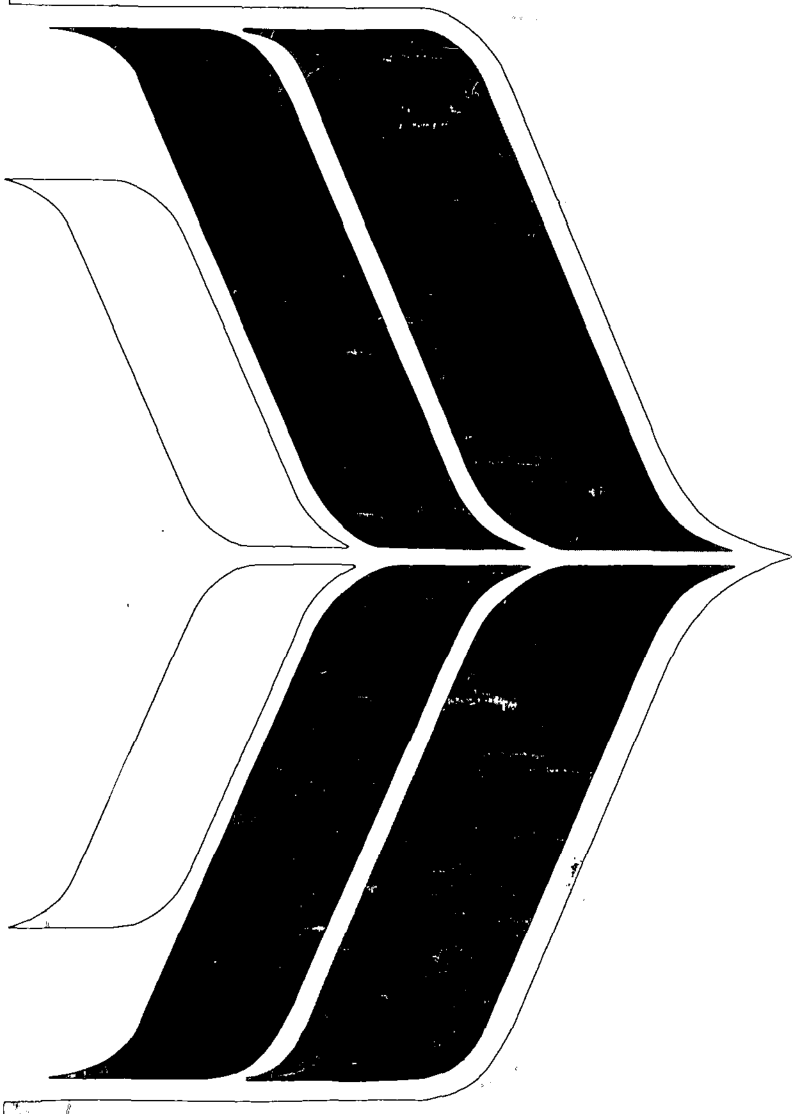

Water Treatment Plant Waste Management

HANDBOOK OF PRACTICE

WATER TREATMENT PLANT WASTE MANAGEMENT

by

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HANDBOOK OF PRACTICE
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TABLE OF CONTENTS

Page

List of Tables vii

List of Figures xi

Acknowledgements xix

Chapter

1. Introduction 1

1.1. Overview 1

1.2. Water Treatment 2

1.2.1. Coagulation Waste Streams 3
1.2.2. Softening Waste Streams 6
1.2.3. Wastes from Inorganic Removal Plants 9

1.3. Use of This Handbook 10

2. Waste Disposal 14

2.1. Regulations 14

2.1.1. Direct Discharge Regulations 17
2.1.2. Land Disposal Regulations 22

2.2. Direct Discharge to Receiving Streams ... 28

2.2.1. Determining Mass Changes 30
2.2.2. Benthic Impacts 34
2.2.3. Aluminum Toxicity 37
2.2.4. Toxicity of Non-Aluminum Compounds 47

2.3. Discharge to the Wastewater Plant 47

2.3.1. Equalization 48
2.3.2. Effects on Biolcgical Processes . 50
2.3.3. Sludge Handling Considerations .. 61

TABLE OF CONTENTS (con't)

Page

2.4. Landfill 63

2.5. Land Application 73

3. Characteristics of Water Treatment Plant Wastes 83

3.1. Types of Wastes Generated 84

3.2. Quantity of Wastes Generated 86

3.2.1. Solid/Liquid Wastes 86
3.2.2. Liquid Phase Wastes 102
3.2.3. Gas Phase Wastes 107

3.3. Physical Characteristics 107

3.3.1. Specific Resistance Test 108
3.3.2. Filter Leaf Test 115
3.3.3. Capillary Suction Time 120

3.4. Chemical Characteristics 124

3.4.1. Solid/Liquid Wastes 124
3.4.2. Liquid Phase Wastes 144

4. Solid/Liquid Waste Treatment 161

4.1. Overview of Processes and Applications .. 161

4.2. Overview of Pilot Studies 164

4.3. Sludge Thickening 170

4.3.1. Description 170
4.3.2. Design Considerations 174
4.3.3. Capital and Operating Costs 182

4.4. Sludge Conditioning 182

4.4.1. Description 182
4.4.2. Capital and Operating Costs 189

4.5. Sludge Pumping 194

4.5.1. Description 194
4.5.2. Capital and Operating Costs 194

TABLE OF CONTENTS (con't)

Page

4.6. Centrifuges 196

4.6.1. Description 196
4.6.2. Design Considerations 206
4.6.3. Capital and Operating Costs 208
4.6.4. Operating Considerations 213
4.6.5. Past Performance 214
4.6.6. Example Facility 219

4.7. Pressure Filter Press 229

4.7.1. Description 229
4.7.2. Design Considerations 231
4.7.3. Capital and Operating Costs 235
4.7.4. Operating Considerations 239
4.7.5. Past Performance 243
4.7.6. Example Facility 252

4.8. Vacuum Filter 261

4.8.1. Description 261
4.8.2. Design Considerations 266
4.8.3. Capital and Operating Costs 273
4.8.4. Operating Considerations 273
4.8.5. Past Performance 276

4.9. Belt Filter Press 281

4.9.1. Description 281
4.9.2. Design Considerations 283
4.9.3. Capital and Operating Costs 285
4.9.4. Operating Considerations 288
4.9.5. Past Performance 290
4.9.6. Example Facility 291

4.10. Sand Drying Beds 295

4.10.1. Description 295
4.10.2. Design Considerations 299
4.10.3. Capital and Operating Costs 312
4.10.4. Operating Considerations 314
4.10.5. Example Facility 317

4.11. Dewatering Lagoons 321

4.11.1. Description 321
4.11.2. Design Considerations 323
4.11.3. Capital and Operating Costs 324
4.11.4. Freeze-Thaw Considerations 324

TABLE OF CONTENTS (con't)

Page

4.12. Chemical Recovery 330

4.12.1. Lime Recovery 331
4.12.2. Coagulant Recovery 338

4.13. Design Examples 374

5. System Optimization For Solid/Liquid Waste Management 389

5.1. The Sludge Management Model 391

5.1.1. Start Up Program 391
5.1.2. Main Program 391
5.1.3. Output Program 393
5.1.4. Data Files 394
5.1.5. Cost Updating 395

5.2. Program Items 396

5.2.1. Water Treatment Process 396
5.2.2. Sludge Removal 399
5.2.3. Unthickened Sludge Pumping Station 400
5.2.4. Gravity Thickener 401
5.2.5. Thickened Sludge Pump Station ... 402
5.2.6. Sludge Dewatering 402
5.2.7. Sludge Conditioning 405
5.2.8. Final Disposal 406

5.3. Examples 407

References 424

Appendices

Development of Cost Curves 431

LIST OF TABLES

Number Page

1- 1 Guide to Locating Information on Wastes Within this Handbook 13

2- 1 Regulatory Acts Governing Water Plant Waste Disposal 15

2- 2 Example In-Stream Water Quality Guidelines and Standards 20

2- 3 Maximum Concentration of Contaminants For Characteristic of EP Toxicity 25

2- 4 Maximum Amount of Metal (lb/acre) Suggested For Agricultural Soils with Sewage Sludge 29

2- 5 Plankton Productivity, Chlorophyll Concentration and Assimilation Ratio in the Poquoson and Warwick Rivers During March and June, 1985 38

2- 6 Estimated 50% Survival Times of Brook Trout (Salvelinus fontinalis, Mitchill) in Synthetic Soft Water Solutions Containing Aluminum and Aluminum Complexing Ligands ... 42

2- 7 Chronic Aluminum Intoxication and Recovery For Trout 45

2- 8 Threshold Concentration For Various Metals in the Air Activated Sludge Process 52

2- 9 Stimulatory and Inhibitory Concentrations of Alkali and Alkaline - Earth Cations 55

2-10 Nutrient Content of Alum Sludge 75

2-11 Physical Properties and Plant Growth in Potting Media Amended with Alum Sludge 77

2-12 Analysis of Forrest Solids One Year After Application of Alum Sludge 80

3- 1 Major Water Treatment Plant Wastes 85

3- 2 Quantity of Backwash Water From Iron and Manganese Removal Plants 98

LIST OF TABLES (con't)

Number Page

3- 3 Regeneration of Cation Exchange Resins 106

3- 4 Specific Resistance For Various Chemical Sludges 114

3- 5 Reported Alum Sludge Characteristics 127

3- 6 Elemental Analysis of Alum Sludge From Oak Ridge, Tennessee 128

3- 7 Metal Concentrations in Durham, N.C. Alum Sludge and Their Source 130

3- 8 Metal Concentration in Tampa, Florida, Alum Sludge 131

3- 9 EP Toxicity Test Results For Alum Sludge 133

3-10 Partial Chemical Composition of Lime Sludges . 134

3-11 Summary of Radium Concentrations in Lime Softening Sludges and Backwash Waters 142

3-12 Theoretical Sludge Quantities and Theoretical and Measured Ra^{226} Concentrations 143

3-13 Example Ranges of Chemical Constituents of Ion Exchange Wastewater 145

3-14 Ra^{226} Concentrations in Ion Exchange Treatment Plant Wastewater 148

3-15 Crystal Lake's Water and Wastewater Barium Concentrations 152

3-16 Wastewater Characteristics at Palo Verde, AZ . 155

3-17 Wastewater Characteristics at Gila Bend, AZ .. 156

3-18 Fluoride Removal by Reverse Osmosis 158

3-19 Arsenic Removal by Reverse Osmosis 159

3-20 Nitrate Removal by Reverse Osmosis 160

4- 1 Range of Cake Solids Concentrations Obtainable 163

LIST OF TABLES (con't)

Number		Page
4- 2	Analysis of Centrifuge Operating Parameters ..	215
4- 3	Results of Centrifugation Evaluation	217
4- 4	Results of Basket Centrifuge Tests	218
4- 5	Raw and Finished Water Quality - Chesapeake, Virginia	220
4- 6	Monroe Plant - Typical Water Quality Grand Rapids, Michigan	226
4- 7	Pilot Plant Testing of Pressure Filter Systems on Alum Sludge	245
4- 8	Comparative Operational Data of Recessed Pressure Filters Treating Alum Sludge	247
4- 9	Pilot Scale Testing Results of Filter Presses on Alum Sludge	249
4-10	Typical Water Quality - Ann Arbor, Michigan ..	253
4-11	Summary of Water Quality - Fairfax County Water Authority	260
4-12	Results of Vacuum Filtration Evaluation	277
4-13	Average Performance and Operating Data by Vacuum Filter on Lime Softening Sludge	279
4-14	Leaf Test Results on Lime Sludge at Bismark, North Dakota	280
4-15	OWASA Raw Water Quality	292
4-16	Durham Water Quality	318
4-17	Water Plants Practicing Lime Recovery	333
4-18	Effectiveness of Centrifugation in Removing Impurities from $CaCO_3$ Sludge	336
4-19	Sludge Characteristics Applicable to Alum Recovery	348
4-20	Iron Coagulant Recovery Results - Athens, TN .	361

LIST OF TABLES (con't)

Number Page

4-21 Initial Conditions/Parameters of Raw Sludge City of Durham - Alum Recovery Tests 364

4-22 Recovered Alum and Residual Sludge Dewatering Parameters - City of Durham 366

4-23 Recovered Alum Quality 369

4-24 Water Treatment Plant Performance Comparison Using Recovered Alum - City of Durham 370

5- 1 Example 1 - Sludge Management Program 409

5- 2 Example 2 - Sludge Management Program 414

5- 3 Example 3 - Sludge Management Program 420

LIST OF FIGURES

Number		Page
1- 1	Waste Producing Processes In Coagulation Plants	4
1- 2	Waste Producing Processes In Softening Plants	7
2- 1	Mass Balance For Supply and Discharge	31
2- 2	Chronic Mortality by Aluminum Hydroxide to Benthic Invertebrates	36
2- 3	Solubility of Freshly Precipitated Aluminum Hydroxide	40
2- 4	50% Survival Time of White Sucker Fry (18.7 mm)	43
2- 5	Example Sludge Equalization for Sewer Discharge	49
2- 6	Response of Biomass to NaCl Shock Loading	53
2- 7	Effect of Water Treatment Plant Sludge (WTPS) on Settled Wastewater Suspended Solids	57
2- 8	Effect of Water Treatment Plant Sludge (WTPS) on Settled Wastewater BOD	58
2- 9	Phosphorus Removed Versus Water Treatment Sludge (WTPS) Added	59
2-10	Effect of Water Treatment Plant Sludge (WTPS) on the Combined Volume of Wastewater Sludge After 30 Minutes of Settling	62
2-11	Effect of Water Utility Sludge on Primary Sludge	64
2-12	Solids Concentration at Which a Sludge Becomes Handleable	66
2-13	Design Considerations in Constructing a Water Plant Waste Landfill	71
2-14	Typical Well Construction	72

LIST OF FIGURES (con't)

Number Page

2-15 Trench Method for Landfilling Water Plant Waste 74

2-16 Growth of Marigolds and Available Soil Phosphorus in Potting Media Amended with Alum Sludge 78

3- 1 Suspended Solids Versus Turbidity 89

3- 2 Backwash Basin Volume Required For 2 Filters and 10% Recycle 95

3- 3 Backwash Basin Volume Required For 10% Recycle at 24 MGD, 50 MGD 96

3- 4 Cycle Run Length as a Function of Body Feed For Diatomaceous Earth Plants 101

3- 5 Buchner Funnel Apparatus 109

3- 6 Time/Filtrate Volume Versus Filtrate Volume Plot Used in Specific Resistance Testing 111

3- 7 Use of Specific Resistance to Determine Optimum Chemical Dosage 113

3- 8 Specific Resistance Versus Polymer Dosage 116

3- 9 Time to Filter (TTF) Versus Polymer Dosage ... 117

3-10 Leaf Test Apparatus 118

3-11 The Optimization of Chemical Conditioning with the Filter Leaf Technique 121

3-12 Capillary Suction Time (CST) Apparatus 122

3-13 Capillary Suction Time (CST) Versus Polymer Dosage 123

3-14 Effect of Ca/Mg Ratio on Sludge Solids Concentration For Lime Sludges 135

3-15 Effect of Magnesium Concentration on the Specific Resistance of Softening Sludges ... 136

3-16 Removal of Inorganic Contaminants by Lime Softening 137

LIST OF FIGURES (con't)

Number		Page
3-17	Ra^{226} Decay Products	139
3-18	Proposed Correlation Between Ra^{226} and Total Hardness Removal Fractions For Lime Softening Plants	141
3-19	Variation of Ra^{226} Concentrations During Regeneration (10 - 15% NaCl Brine)	147
3-20	Barium and Hardness Concentrations in Spent Brine Well #8	149
3-21	Barium and Hardness Concentrations in Spent Brine Well #6	150
3-22	Sulfate and Nitrate in 6% Brine Discharge From System	153
4- 1	Sludge Handling Options	162
4- 2	Pilot Study Logic Diagram	165
4- 3	Continuous Flow Gravity Thickener	172
4- 4	Batch Thickening Tank Schematic	173
4- 5	Thickening Test in a Cylinder with Resulting Interface Height Versus Time Curve	175
4- 6	Effect of Cylinder Diameter on Settling Velocity	176
4- 7	Batch Thickening Test Plots	178
4- 8	Solids Flux Curve For Continuous Thickener	180
4- 9	Construction Cost For Gravity Sludge Thickeners	183
4-10	Operation and Maintenance Cost For Gravity Sludge Thickeners	184
4-11	Typical Sludge Conditioning Monomers and Polymers	186
4-12	CST Versus Polymer Dosage	188
4-13	Construction Cost For Polymer Feed Systems	190

LIST OF FIGURES (con't)

Number		Page
4-14	Operation and Maintenance Cost For Polymer Feed Systems	191
4-15	Construction Cost For Lime Feed Systems	192
4-16	Operation and Maintenance Cost For Lime Feed Systems	193
4-17	Construction Cost For Unthickened Sludge Pumping	195
4-18	Operation and Maintenance Cost For Unthickened Sludge Pumping	197
4-19	Construction Cost For Thickened Sludge Pumping	198
4-20	Operation and Maintenance Cost For Thickened Sludge Pumping	199
4-21	Solid Bowl Centrifuge	201
4-22	Basket Centrifuge Schematic	203
4-23	Centrifugation System	207
4-24	Typical Basket Centrifuge Installation	209
4-25	Typical Decanter Centrifuge Installation	210
4-26	Construction Cost For Decanter Centrifuges	211
4-27	Operation and Maintenance Cost For Decanter Centrifuges	212
4-28	Chesapeake Solids Handling Facility Process Schematic	221
4-29	Chesapeake Sludge - Feed and Sludge Cake Solids	223
4-30	Grand Rapids Solids Handling Facility - Process Schematic	227
4-31	Cutaway View of a Filter Press	236
4-32	Side View of a Filter Press	237
4-33	Typical Filter Press Installation	238

LIST OF FIGURES (con't)

Number Page

4-34 Construction Cost for Diaphragm Filter Press 240

4-35 Operation and Maintenance Cost For Diaphragm Filter Press 241

4-36 Sludge Plan Diagram - Erie County Water Authority, New York 246

4-37 Diaphragm Press Performance at Norfolk, Virginia 250

4-38 Ann Arbor Sludge - Feed and Cake Solids 255

4-39 Ann Arbor Solids Handling Facility - Process Schematic 257

4-40 Fairfax County Water Authority - James J. Corbalis Plant Schematic 259

4-41 Rotary Vacuum Filter 265

4-42 Rotary Vacuum Filter - Section 267

4-43 Horizontal Vacuum Filter Schematic 268

4-44 Vacuum Filtration System 269

4-45 Typical Vacuum Filter Installation 270

4-46 Rotary Vacuum Filter Cake Removal 272

4-47 Construction Cost For Vacuum Filters 274

4-48 Operation and Maintenance Cost for Vacuum Filters 275

4-49 The Three Basic Stages of a Belt Press 282

4-50 Typical Belt Filter Press Installation 286

4-51 Construction Cost For the Belt Filter Press .. 287

4-52 Operation and Maintenance Cost For Belt Filter Press 289

4-53 Orange Water and Sewer Authority - Sludge Facilities 294

LIST OF FIGURES (con't)

Number		Page
4-54	OWASA Belt Press - Solids Concentrations	296
4-55	Depth Required to Obtain Optimum Loading at Various Sludge Concentrations	303
4-56	Mean Annual Evaporation	309
4-57	Dry Cake Thickness Versus Depth Applied	310
4-58	Applications/Year Versus % Dry Solids	311
4-59	Construction Cost For Sand Drying Beds	313
4-60	Operation and Maintenance Cost For Sand Drying Beds	315
4-61	Durham Drying Bed Facilities - Process Schematic	320
4-62	Solids Production - Durham - Williams Water Plant	322
4-63	Construction Cost For Sludge Dewatering Lagoons	325
4-64	Mechanical Freeze-Thaw Process	327
4-65	Options of Lime Recovery	334
4-66	Percent Aluminum Recovered Versus Chemicals Added	343
4-67	Percent Aluminum Recovered Versus Sulfuric Acid Added	344
4-68	Initial Settling Velocity of Sludge Remaining After Acidification	345
4-69	Variation of Specific Resistance of Sludge Remaining Versus Percent Aluminum Recovered	347
4-70	Aluminum Dissolution as a Function of pH	349
4-71	Stoichiometric Acid Demand For Aluminum Dissolution	350
4-72	Rate of Aluminum Dissolution	351

LIST OF FIGURES (con't)

Number Page

4-73 Solids Reduction Versus Aluminum Dissolution . 353

4-74 Alum Recovery Flow Diagram - Quantification For Tampa, Florida 354

4-75 Flow Diagram of Liquid-Ion Exchange - Alum Recovery Process 355

4-76 Coagulant Recovery Solids Dewatering System - A.R. White and Co. 358

4-77 Sand Drying Bed Drainage For Acidified Sludge - Durham, N.C. 367

4-78 TOC Removal Using Recovered Alum - Durham, N.C. 371

4-79 TTHM Removal Using Recovered Alum - Durham, N.C. 372

4-80 Alum Recovery Process Flow Diagram - Durham, N.C. 373

FOREWORD

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The Research Foundation was created by the water supply industry as its center for cooperative research and development. The Foundation itself does not conduct research; it functions as a planning and management agency, awarding contracts to other institutions, such as water utilities, universities, engineering firms, and other organizations. The scientific and technical expertise of the staff is further enhanced by industry volunteers who serve on Project Advisory Committees and on other standing committees and councils. An extensive planning process involves many hundreds of water professionals in the important task of keeping the Foundation's program responsive to the practical, operational needs of local utilities and to the general research and development needs of a progressive industry.

All aspects of water supply are served by AWWARF's research agenda: resources, treatment and operations, distribution and storage, water quality and analysis, economics and management. The ultimate purpose of this effort is to assist local water suppliers to provide the highest possible quality of water, economically and reliably. The Foundation's Trustees are pleased to offer this publication as contribution toward that end.

The handling and disposal of water plant wastes continues to be a major problem for utilities of all sizes. This manual will be an invaluable resource for utility managers and design engineers by providing state-of-the-art information on all aspects of water plant waste management.

Jerome B. Gilbert
Chairman, Board of Trustees
AWWA Research Foundation

James F. Manwaring, P.E.
Executive Director
AWWA Research Foundation

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CHAPTER 1

INTRODUCTION

1.1. OVERVIEW

The purpose of this Handbook is to summarize procedures to characterize, design and utilize treatment/disposal systems for wastes produced at water treatment plants. The first Chapter presents the general categories of water plant wastes and the points in the treatment process at which they are typically generated.

Chapter 2 discusses the disposal of water treatment wastes including landfill, land application, sewer discharge and direct discharge to receiving streams. Included in this Chapter is a discussion of the regulations that are applicable to each of the ultimate disposal alternatives.

Chapter 3 presents the methodologies associated with characterizing the waste streams, including methods to determine waste quantities, test procedures for conditioning comparisons and chemical characteristics of the various types of wastes.

Chapter 4 presents the alternative processes involved in the handling and treatment of solid/liquid water plant wastes. Methods of treatment presented include thickening, pumping, conditioning, mechanical dewatering, non-mechanical dewatering and additional technologies. Included in this Chapter is a discussion of process theory, design considerations and examples, capital and operating costs, and a summary of past performance as reviewed from the literature or from direct utility contact.

The objective of Chapter 5 is to assist in the optimization of a sludge treatment system (considers solid/liquid wastes only). A methodology of selecting an efficient and cost effective system is presented. Due to the numerous alternatives available for sludge treatment and disposal systems, a computer program has been written including all the cost data presented in Chapter 4 to assist in the selection of the system. The computer program allows the user to "mix and match" treatment and disposal options to determine the relative cost of a particular application. Chapter 5 also discusses the intricacies of the computer program.

1.2. WATER TREATMENT

For the purposes of discussion of water plant wastes, treatment plants can be broadly divided into four general categories. First are those treatment plants that coagulate, filter and oxidize a surface water for removal of turbidity, color, bacteria, algae, some organic compounds and often iron and/or manganese. These plants generally use alum or iron salts for coagulation and produce two waste streams. The majority of the waste produced from these plants is sedimentation basin (or clarifier) sludge and filter backwash wastes. The second type of treatment plants are those that practice softening for the removal of calcium and magnesium by the addition of lime, sodium hydroxide and/or soda ash. These plants produce clarifier basin sludges and filter backwash wastes. On occasion, plants practice both of the above treatment technologies. Softening plant wastes can also contain trace inorganics such as radium that could affect their proper handling. The third type of plants are those that are designed to specifically remove trace inorganic substances, such as nitrate, fluoride, radium, arsenic, etc. These plants use processes such as ion exchange, reverse osmosis or adsorption. They produce liquid wastes or solid wastes, such as spent adsorption material. The fourth category of treatment plants are those that produce are air phase wastes which are produced during the stripping of volatile compounds.

1.2.1. Coagulation Waste Streams

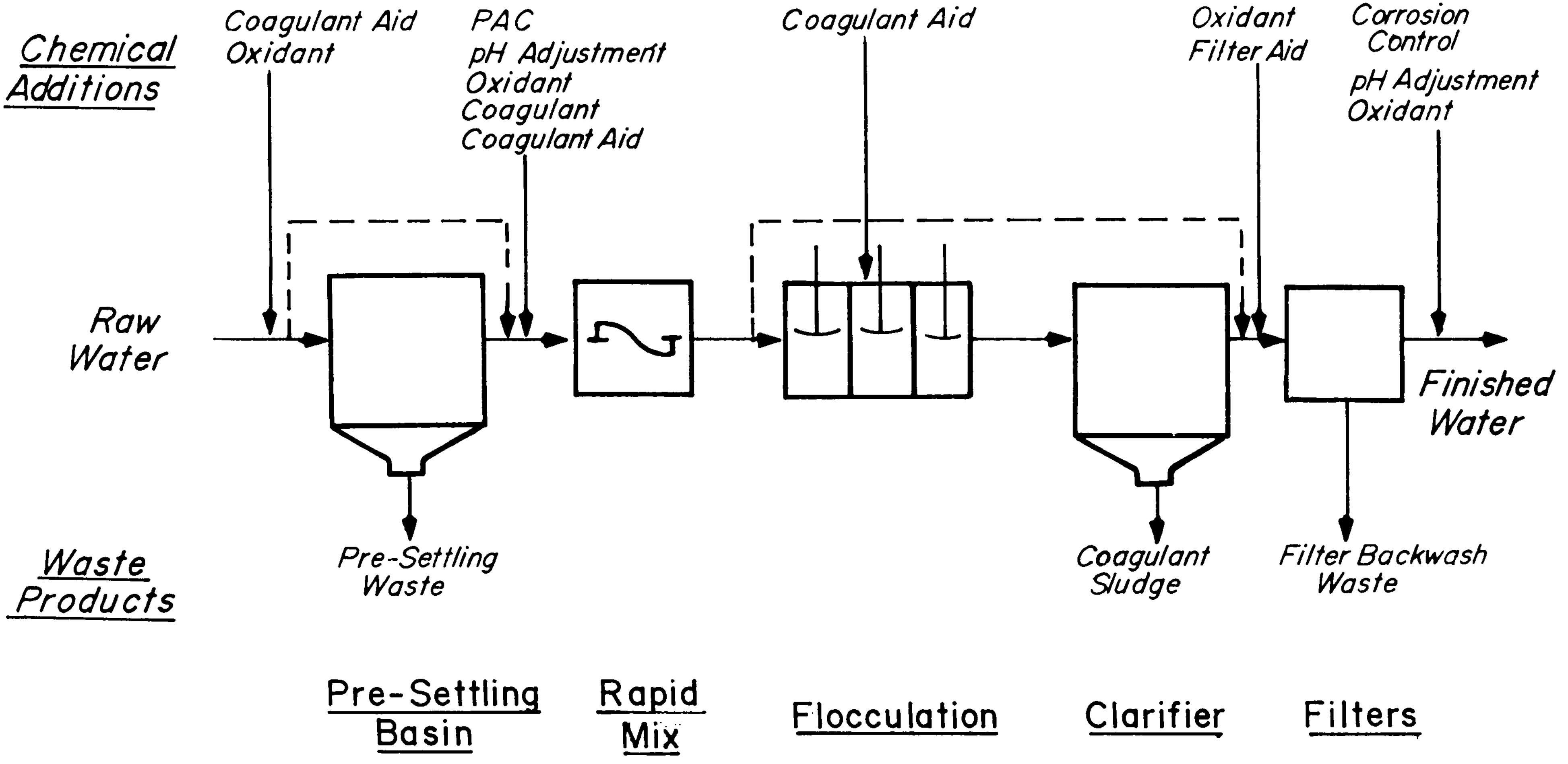
Coagulation of surface waters is by far the most commonly used water supply treatment technology. These waste streams make up the majority of the water plant wastes produced by the water industry. They are also some of the more difficult wastes to treat. Figure 1-1 shows a schematic of a conventional coagulation treatment process showing the typical waste products. Some water plants have a pre-sedimentation step. This is generally used only when the raw water source is high in settlable solids. Often no chemical is added prior to pre-sedimentation, although an oxidant or small amount of polymer may be added. It is generally accepted that as long as coagulant is not added, and therefore the solids are essentially only those settled from the raw water, then these solids can be discharged back to the watercourse on a controlled basis. Due to this handling, and the very site specific nature of this waste stream, pre-settling solids are not specifically addressed in this handbook.

The coagulation process itself generates most of the waste solids. Generally a metal salt (aluminum or iron) is added as the primary coagulant. In addition to the coagulant other solids producing chemicals such as powdered activated carbon, polymer, clay, lime, or activated silica may be used. These added chemicals will all produce waste solids. They are usually removed, along with the solids in the raw water, in a sedimentation tank or clarifier. In areas with very good raw water quality sedimentation basins are occasionally omitted and the solids removed by filtration only. This process, commonly known as direct filtration, is usually used for waters with low turbidity and requires low levels of coagulant. All solids removed in this process are collected with filter backwash water.

The quantity of solids produced depends on the raw water quality and chemical addition. Chapter 3 includes a discussion of how to determine the amount of solids produced. The

FIGURE 1-1

WASTE PRODUCING PROCESSES IN COAGULATION PLANTS



volume of sedimentation basin sludge produced depends on both the characteristics of the solids and the mechanism by which solids are removed from the basin. Many basins, particularly older ones, do not have mechanical means of removing the solids and must be manually cleaned. In these basins the solids are stored for extended periods of time and are allowed to accumulate to some predetermined level. Periodically the basin is drained and often washed out with a fire hose. Obviously for these basins, the cleaning frequency is a function of the volume of sludge produced and available storage volume in the basins. Manual cleaning results in batch production of sludge and makes subsequent sludge handling more difficult. In most situations it is desirable to retrofit the basins with continuous sludge removal equipment, which may be difficult to accomplish due to basin configurations. However, producing a fairly continuous and consistent flow of sludge to the sludge treatment process is often a critical factor in successful dewatering. Appropriate sludge removal in combination with flow equalization must be well planned.

The second major waste stream produced is from the batch process of backwashing the filters. The solids collected on the filters are those remaining after sedimentation or caused by the addition of a filter aid or formed by oxidation of perhaps iron or manganese. In a direct filtration process, these are the only solids produced. The volume is a function of the amount of water used for backwashing. This waste stream is produced at very high flow rates for short periods of time and again proper equalization is required.

Another waste product that is occasionally produced in a coagulation-based plant is spent granular activated carbon (GAC). GAC is sometimes used in the filters or post-filtration. When its use is for taste and odor removal, the carbon is disposed of after its capacity is exhausted. When its use is for continuous low-level organics removal, then the carbon is usually regenerated on-site, with essentially no waste stream.

1.2.2. Softening Waste Streams

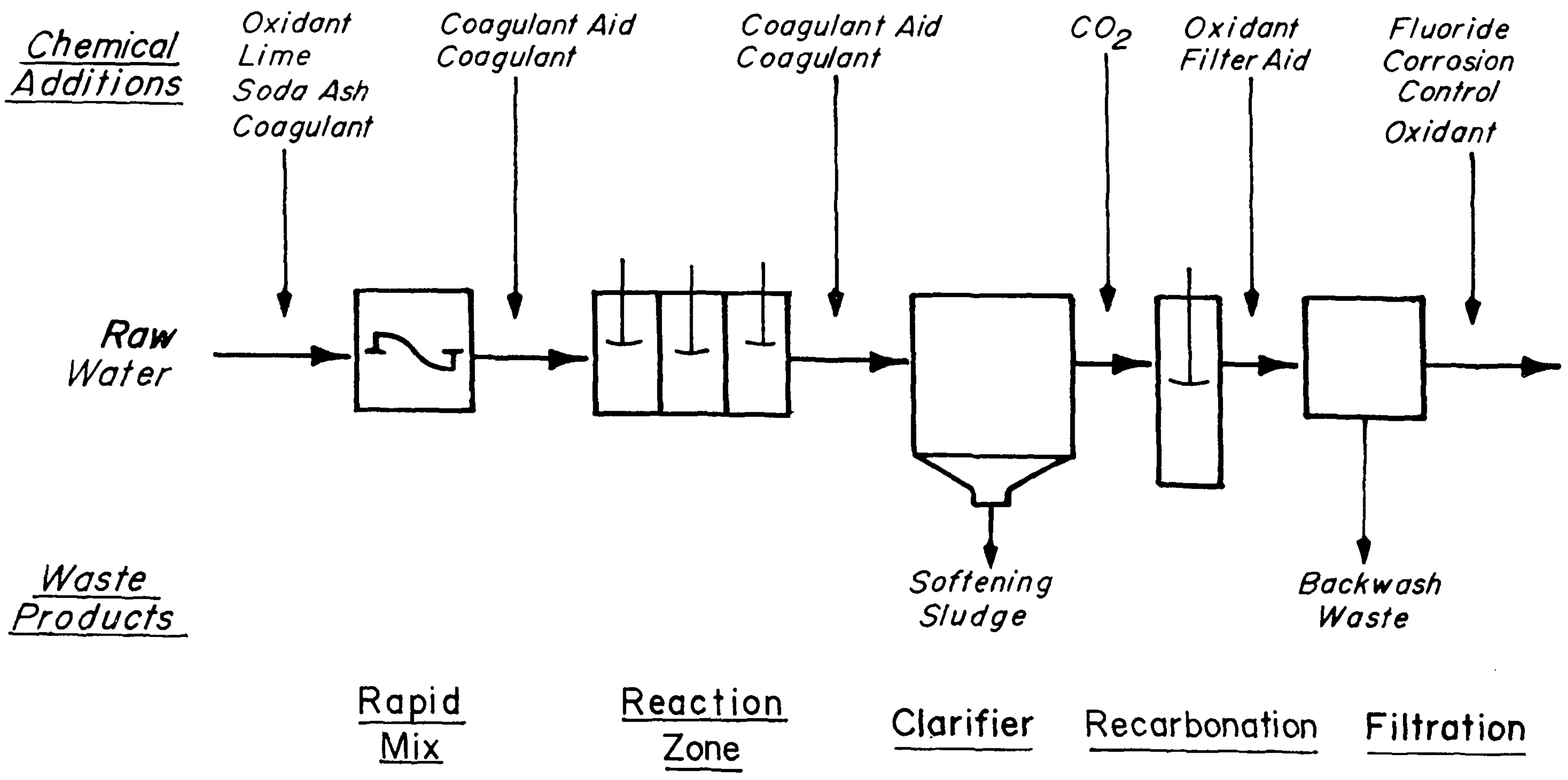
Wastes produced from softening plants represent the second major waste product produced by the water industry. Fortunately, they are generally more easily dewatered than are coagulant wastes, although the presence of some trace inorganics may make their proper disposal difficult. There are many variations of the softening process. Chemical addition, flow processes and the subsequent waste quantities and characteristics are all dependent on the raw water hardness and alkalinity constituents, and the desired finished water quality. Since softening is generally a process used to improve the chemical characteristics and aesthetics of the finished water rather than its potability, subjective decisions can be made as to the final desired quality. One of the factors that should enter into that decision process is the effects on sludge handling and costs.

Softening is accomplished either by chemical precipitation of the calcium and magnesium or by the use of ion exchange resins. The former, traditionally called lime/soda ash softening, is by far the most widely used softening process. In this method, lime is added for the removal of carbonate hardness supplemented with the use of soda ash for non-carbonate hardness removal if required. From the standpoint of sludge economics, it is desireable to leave as much magnesium hardness in the water as considered acceptable. Often the final magnesium hardness can be allowed to remain around 40 mg/l as $CaCO_3$, or slightly higher and not have an adverse effect on home water heaters. The less magnesium in the sludge, the easier it is to dewater.

Figure 1-2 is a rather simplified softening plant schematic. Several variations and complications of Figure 1-2 are used to obtain the desired water quality and minimize

FIGURE 1-2

WASTE PRODUCING PROCESSES IN SOFTENING PLANTS



costs. In softening plants there are usually two waste streams produced: the settled solids from the clarifier and the backwash wastes. Some plants will add a polymer or metal salt to aid in the removal of fine precipitates or color or turbidity present in the original water. Again, from a sludge viewpoint, the addition of metal salts should be held to a minimum as the presence of metal hydroxides could significantly increase sludge treatment costs. The use of polymers and slurry recirculation can help minimize the use of these coagulants.

In many plants the reaction zone and clarifier are combined into a single solids contact unit. In these plants sludge can be fairly uniformly withdrawn from the sludge blanket and a consistent suspended solids concentration and flow rate can be maintained. Plants that have separate clarifiers are often equipped with scrapers for sludge removal. Although not quite as easy to control as the sludge blanket units, the separate clarifiers can produce a fairly consistent sludge. As with coagulation plants, filter backwash water is produced at high flow rates for short periods of times. Filter backwash water may require equalization basins prior to treatment or discharge.

In water softening by ion exchange the water containing the hardness is passed through a column containing the ion exchange material. The hardness in the water exchanges with an ion from the ion exchange material. Generally, the ion exchanged with the hardness is sodium:

$$Ca(HCO_3)_2 + 2NaR = CaR_2 + 2NaHCO_3$$

where R represents the solid ion exchange material. By the above reaction, calcium (or magnesium) has been removed from the water and replaced by an equivalent amount of sodium; ie, two sodium ions for each divalent cation removed. The exchange results in essentially 100% removal of the hardness from the water until the exchange capacity of the ion exchange material is reached. When the ion exchange resin becomes saturated, "breakthrough" is said to have occurred because the hardness is no longer removed. At this point the ion exchange material is regenerated. During regeneration, the hardness is removed from the material by passing water containing a large amount of Na^+ through the column. The mass action of having so much Na^+ in the water will cause the hardness of the ion exchange material to enter the water and exchange with the sodium:

$$CaR_2 + 2NaCl = 2NaR + CaCl_2$$

The ion exchange material can now be used to remove more hardness.

This regenerant material is the wastewater stream which requires disposal. It contains the excess or left over NaCl, and the ions removed -- $CaCl_2$ and $MgCl_2$. Economics dictate that a readily available location be used for disposal of this brine. Therefore, most large plants that utilize ion exchange softening are located in coastal communities so that ocean brine disposal is practiced. Ion exchange has been used in small water supply systems in other parts of the country and wastes have most often been discharged to municipal wastewater systems, or to receiving streams. Two additional waste streams are also produced in conjunction with ion exchange. Prior to the use of the regenerant, the column is usually backwashed in an upflow mode to remove any suspended material. After regeneration the column is rinsed, which will produce a waste stream also high in dissolved solids.

1.2.3. Wastes from Inorganic Removal Plants

Removal of trace inorganic substances such as arsenic, barium, cadmium, chromium, fluoride, lead, nitrate, mercury, selenium, silver, radium and uranium is becoming more of a concern in the drinking water industry. Often hardest hit with removal of these constituents is the very small utility, with limited capital but with no other water source. The available processes for economically removing these substances have not been well defined, and the proper handling and disposal of resulting wastes can be critical to overall economic success. Unfortunately this area is a relatively new science for the water supply industry and little is known on actual applications.

Processes which may be applicable for removal of trace inorganics include lime softening, ion exchange, adsorption, reverse osmosis or electrodialysis. Depending upon concentrations and constituents in the waste streams, some of these residuals could be classified as hazardous wastes under RCRA regulations (Chapter 2 contains a discussion of RCRA regulations).

1.3. USE OF THIS HANDBOOK

The intent of this Handbook has been to bring together into one document that which is known regarding waste characteristics, handling and disposal within the water industry. The emphasis is on the applications of technology rather than the theory. The hope is that a user of the Handbook can save considerable time and effort in screening waste management alternatives by having readily available a compilation of current technology and practice. The Handbook includes information obtained from the published literature over the past 25 years. Another major source of information was direct utility contact. In many cases these contacts were site visits to obtain performance data, system design and operator experiences. Results of these contacts have been incorporated into the appropriate sections and reported as either "past performance" or developed in more detail as an "example facility."

The first technical chapter of this Handbook is on Disposal. Obviously, disposal is the last step in the waste management process. However, disposal is the driving force behind waste management practices. As such, disposal should be the first consideration in developing a treatment system. The available options should be fully explored and defined. If only one disposal option is available, it may define the whole treatment scheme. Multiple disposal options may help create treatment tradeoffs. Included within the disposal chapter is a discussion of appropriate regulations that can affect the water industry.

Following Disposal is the chapter on Waste Characterization. Methods and procedures are presented to help a utility gain insight into the properties of its wastes. Some of the procedures presented are also useful for routine process operation. Even for those utilities with no immediate plans for a waste treatment system, it is highly recommended that data collection become a routine part of the laboratory program.

Most of the Handbook deals with treatment technologies for solid/liquid wastes. The processes relative to sludge treatment have been addressed - pumping, thickening, conditioning, mechanical dewatering and non-mechanical dewatering. In each case is a discussion on how the technology can be evaluated for a specific waste. In addition, design and operational considerations that have been gained through experience at other utilities are presented including performance data available from pilot or operating installations. Finally, a case study of an operating facility is included as appropriate. Also included for each technology is a capital and operating cost curve. In the study and planning stages of alternative comparisons, preliminary cost evaluations can be extremely useful and help identify relative levels of capital and O & M expenditures for several alternatives.

The objective of the cost curves is to make the costs comparable between technologies. The curves are accurate for a study phase of paper screening of alternatives. As the evaluation moves to a pilot phase or preliminary design, site specific costing is needed. Often detailed cost estimates are higher than the sum of the appropriate cost curves here. That is often the case when costs for site constraints, earthwork, incorporating the sludge process into the existing water treatment system, yard piping, matching architecture, future expansion, operating flexibility, operator work shifts, etc. are all taken into account. Appendix A contains a description of how the cost curves were derived and an example comparison with an actual bid for construction by a contractor. Note that the costs are developed on a unit operations basis. For example, if a facility is using thickening, pumping, and filter press with lime conditioning then to obtain the total cost estimate, the appropriate curves must be added together. Appendix A will be helpful in using the cost curves. Section 4.13 illustrates the use of these cost curves.

The Handbook is organized from a unit operations approach and therefore data on the handling of a particular type of waste can be found throughout the Handbook. Table 1-1 is intended as a quick reference guide for obtaining information on a particular type of waste. This guideline table contains location information for Chapters 2 and 3. Chapters 4 and 5 pertain only to solid/liquid waste dewatering and are useful for evaluation of any of the solid/liquid wastes. Table 1-1 shows the location only where specific applicable information exists. For example, chemical characteristics of radium containing lime sludges is specifically discussed in Section 3.4 as shown in Table 1-1. Useful information for radium containing lime sludges could be found along with general lime sludge physical characteristics (Section 3.3) but this is not shown in Table 1-1 since radium wastes are not specifically addressed in that section.

TABLE 1-1

GUIDE TO LOCATING INFORMATION ON WASTES WITHIN THIS HANDBOOK

	SECTION							
Waste Producing Process	Regulations	Stream Discharge	Wastewater Discharge	Landfill	Land Application	Quantity Calcualtions or Estimates	Physical Characteristics & Conditioning Tests	Chemical Character-istics and Constituent Analysis
Coagulation	2.1, 2.1.1, 2.1.2., 3.4	2.2.1, 2.2.2, 2.2.3	2.3.1, 2.3.2, 2.3.3	2.4	2.5	3.2.1	3.3, 3.3.1, 3.3.2, 3.3.3	3.4
Softening - General	2.1, 2.1.1, 2.1.2	2.2.1	2.3.1, 2.3.2, 2.3.3	2.4	2.5	3.2.1	3.3, 3.3.1, 3.3.2, 3.3.3	3.4
Softening with Radium Removal	2.1, 2.1.1, 2.1.2	2.2.1	2.3.1	2.4	2.5			3.4
Filter Backwashing	2.1, 2.1.1	2.2.1	2.3.1, 2.3.2			3.2.1	3.3.1	
Ion Exchange	2.1, 2.1.1	2.2.1	2.3.1, 2.3.2			3.2.2		3.2, 3.4
Iron Removal	2.1, 2.1.1	2.2.1	2.3.1, 2.3.2			3.2.1		3.2.1
Slow Sand Filtration	2.1					3.2.1		3.2.1
Reverse Osmosis	2.1, 2.1.1	2.2.1	2.3.1, 2.3.2			3.2.2		3.4
GAC incl. GAC with Radon Removal	2.1, 2.1.2					3.2.1		
Pre-Coat (DE) Filt.	2.1					3.2.1		3.4
Activated Alumina	2.1					3.4		3.4
Air Stripping	2.1					3.2.3		

CHAPTER 2

WASTE DISPOSAL

2.1. REGULATIONS

The applicable regulations governing the disposal of water plant wastes can be broadly divided into two categories. First are those regulations associated with the Clean Water Act (CWA) which tend to limit the use of direct discharge of wastes into a watercourse as an acceptable technology. The second set of regulations are the Resource Conservation and Recovery Act (RCRA) and the Comprehensive Environmental Response, Compensation and Liability Act (CERCLA). These latter regulations primarily affect land disposal of water plant wastes. If the waste contains radioactivity, additional considerations apply. Since interpretation of the regulations as they affect water plant wastes is in a constant flux, and varies considerably from state to state, this discussion should be considered only as guidance that may be appropriate for safe and legal waste disposal. Local and state authorities should be contacted for specific regulations.

Within the provisions of the applicable regulatory statutes, several possible sections could apply to the disposal of water plant wastes. Table 2-1 summarizes the environmental statutes which affect water plant waste disposal. In discharging to a body of water, a permit must be obtained under the National Pollutant Discharge Elimination System (NPDES) as authorized under the Clean Water Act (CWA). The standards for the discharge permit are designed to protect aquatic and human life. This requirement to protect the environment led to the establishment of in-stream water quality criteria and standards. Criteria are defined as guidelines or goals established by EPA. Standards are the enforceable levels, generally established by the individual states. Allowable pollutant concentrations in a discharge can

TABLE 2-1

REGULATORY ACTS GOVERNING WATER PLANT WASTE DISPOSAL

Disposal Option	Applicable Regulations
Stream	NPDES (CWA) In-Stream Water Quality Criteria (CWA) Discharge Guidance Documents
Wastewater Plant	Pretreatment Standards (CWA)
Landfill	RCRA CERCLA State SW Requirements (RCRA) Low Level Radioactive Waste Requirements (State, NRC, DOT, EPA)
Land Application	Sludge Disposal Regulations (CWA) Low Level Radioactive Waste Requirements (State, NRC, DOT EPA)

Note:

CWA = Clean Water Act
RCRA = Resource Conservation and Recovery Act
CERCLA = Comprehensive Environmental Response, Compensation and Liability Act

be set to meet the in-stream water quality standards, the criteria levels or other levels as the individual states may deem appropriate for a specific water course. The act recognized that technology (coupled with economics) may not be sufficiently developed to allow all industrial dischargers to meet the desired in-stream levels. The act called for the development of guidance documents for industrial discharges, indicating the discharge levels which they should be able to meet. The CWA also required individual wastewater treatment plants to develop pretreatment standards to govern discharge of wastes into the sewer. These pretreatment regulations can effect the discharge of water plant wastes into the wastewater plant.

When considering landfilling of dewatered solids from water plants, RCRA requirements and the state's individual solid waste requirements govern. RCRA defines hazardous waste and establishes the guidelines for its safe treatment, storage and disposal. Even if a water plant waste is not considered hazardous, it will need to meet the individual state's requirements for solid waste disposal into an approved landfill. For some specialized water treatment plant wastes that contain low level radioactivity, regulations established by individual states may apply under guidance from the Nuclear Regulatory Commission (NRC).

The Clean Water Act authorized the establishment of criteria for land application of sludges. These criteria are designed for wastewater sludges and may only provide general guidelines for water plant wastes. Again, specialized wastes may be governed by requirements associated with low level radioactive waste disposal. In the next two sections are some considerations associated with the major regulations affecting water plant wastes.

2.1.1. Direct Discharge Regulations

The Water Pollution Control Act of 1965 (PL 84-660) required states to set standards for interstate waters and gave them authority to order treatment of wastes from water treatment plants. Some grant money was provided by this law for constructing water treatment projects, but most water treatment projects were assigned a low priority and little attention was given to the operating performance of such plants.

With the passage of the Clean Water Act (PL 92-500) and the establishment of EPA, a more formal procedure for controlling water treatment plant discharges was established. Water supply was formally declared as an industry. The implications of this decision were twofold:

- Construction grant monies offered public wastewater plants were not available to water treatment plants even if they were publicly held.
- A procedure for promulgating guidelines for discharges from water treatment plants was established.

The guidance document for the water supply industry divided water treatment plants into three categories:

1. Plants that use one of the following: coagulation, oxidation for iron and manganese removal or direct filtration.
2. Plants that use chemical softening procedures.
3. Plants that use a combination of the procedures in the above categories.

For each category, the best practical control technology was defined and allowable pH and total suspended solids limitations were established. The limits established for water plant discharges ranged from 5 to 10.8 lbs of solids per million gallons of water treated depending on plant capacity. Larger plants were held to lower solids discharge levels. This guidance document, however, did not progress beyond the draft guidance phase. It also did not address liquid phase waste discharges.

EPA has developed about 50 such guidance documents for various categories of industrial wastes. A 1985 presentation by EPA (2-1) confirms that there is no effluent guideline document for water treatment plants nor is there likely to be one. EPA has not established plans to publish a guidance document on water plant wastes. Therefore, discharge decisions are made either by the Regional EPA offices or by the individual states delegated to write their own permits. It is up to the permit writer to rule on the best available treatment technology for each plant on a case-by-case basis. According to EPA the primary criteria for allowance of direct discharge is to meet established in-stream water quality standards at the edge of the mixing zone:

> "In developing technology based limitations in permits, a controlled release of water clarifier sludge and filter backwash from water treatment plants in a manner which meets water quality standards may in appropriate circumstances be constituted to be technology based controls"(ref. 2-1)

Two key phrases are <u>controlled discharge</u> and <u>meets water quality standards</u>.

In-stream water quality criteria and standards are developed by individual states (with the use of some federal guidelines). Most states have classified each body of water for a designated use and set in-stream quality guidelines appropriately. Table 2-2 shows example in-stream water quality criteria and standards for several selected compounds (since standards vary from state to state only examples can be illustrated. The specific agency involved should be contacted). These quality criteria would apply to solid/liquid waste streams or liquid phase waste streams. In addition to meeting in-stream water quality standards, some states have established maximum allowable concentrations in the discharge. These limits generally apply if they are more stringent than the allowable discharge that will meet the in-stream water quality criteria. For example, Illinois does not allow a discharge of greater then 15 mg/l fluoride (F). Barium discharge is required to be less then 2 mg/l, even if the 1 mg/l in-stream standard could be met through dilution (see Section 2.2.1). Wisconsin has set maximum discharge levels of radium (soluble) in liquid wastes as follows:

$$\frac{Ra^{226}}{30} + \frac{Ra^{228}}{30} \leq 1 \text{ pCi/l}$$

These regulations apply to discharge from a water plant to a storm sewer or to a surface body of water. They also apply to an effluent from a wastewater treatment plant.

Discharge to wastewater treatment plants are generally governed by the individual plant's pretreatment regulations. There may also be some specific guidelines provided by state agencies. Wisconsin has limited the discharge of radium to a sewer as:

$$\frac{Ra^{226}}{400} + \frac{Ra^{228}}{800} \leq 1 \text{ pCi/l}$$

In addition, the total amount of radiation released to the sewer system cannot exceed 1.0 curie/year (see also the next Section for more detail on radioactive waste regulatory implications).

TABLE 2-2

EXAMPLE IN-STREAM WATER QUALITY GUIDELINES AND STANDARDS

	Guidelines			Example Standards
	Aquatic Life Chronic Criteria			Stream Used For Potable Water
	Fresh ug/l	Salt ug/l	Human Health* ug/l	mg/l
Arsenic (Dissolved)	72	63	2.2 ng/l	0.05
Barium				1.0
Beryllium	130		3.7 ng/l	
Cadmium	$e^{1.16}$ (ln(hardness))-3.841	12	10.0	0.01
Chloride				250
Chromium (hexavalent, dissolved)	7.2	54		
(trivalent, active)	$e^{0.819}$ (ln(hardness))+.537		170	
(TOTAL)				0.05
Copper	2.0	4(2y) 23(A)		1.0
Cyanide, free	4.2	0.57	20.0	
Fluoride				1.4
Hydrogen Sulfide	2.0	2.0		
Iron, total	1000			
soluble				0.3
Lead	$e^{1.34}$ (ln(hardness))-5.245	8.6	50	0.05
Manganese, total		100		
soluble				0.05
Mercury	0.00057	0.1	146 ng/l	0.002
Nickel (total)	$e^{0.76}$ (ln(hardness))+1.06	7.1	13.4	
Nitrate (as N)				10
Phenol	1.0	1.0	3,500	0.001
Selenium	35	54	10	0.01

TABLE 2-2 (con't)

EXAMPLE IN-STREAM WATER QUALITY GUIDELINES AND STANDARDS

	Guidelines			Example Standards
	Aquatic Life Chronic Criteria			
	Fresh ug/l	Salt ug/l	Human Health* ug/l	Stream Used For Potable Water mg/l
Silver	$.01e^{1.72\ (\ln(\text{hardness}))-6.52}$	0.023	50	0.05
Sulfate				250
TDS				500
Zinc	47	58	5,000	5.0
Aldrin	0.03	0.003	0.074 ng/l	
Chloride	0.0043	0.004	0.46 ng/l	
Endrin	0.0023	0.0023	1.0	0.0002
Heptachlor	0.0038	0.0036	0.28 ng/l	
Lindane	0.08	0.0016		0.004
Methoxychlor	0.03	0.03		0.10
Toxaphene	0.013	0.0007	0.71 ng/l	0.005
DDT	0.001	0.001	0.024 ng/l	0.1
Chloroform	1,240		0.19	
Radioactivity $Ra^{226+228}$				5 pCi/l
Gross Alpha Particle Activity (excl radon and uranium)				15 pCi/l

NOTE: Guideline values are from EPA, Water Quality Criteria Documents, Federal Register, Nov. 28, 1980 Vol. 45, No. 231, 79318. Standards are selected from various state regulations and do not reflect any one state's regulations.

*Values given are the ambient water quality criterion for protection of human health for non-carcinogens, and for carcinogens the value is the risk of one additional case of cancer in 1,000,000 persons.

2.1.2. Land Disposal Regulations

Land disposal regulations can apply to landfilling of solid wastes or land application of solid or liquid phase wastes. For landfilling of solid waste the waste needs to be classified in one of three categories:

1. Safe for normal landfilling as an industrial waste
2. Classifiable as a hazardous waste
3. Contains low level radioactivity

Obviously if a waste does not fit into category 2 or 3, it can be landfilled in an industrial waste landfill. Some states will allow the disposal of water plant wastes in a general sanitary landfill rather then an industrial waste landfill. However, very often in these states, requirements for the construction of a general sanitary landfill are as stringent as the requirements for the construction of an industrial waste landfill. These landfills are governed by the individual state requirements. As a minimum, these regulations will require that the waste cannot contain any free water (water that will drain by gravity). Other states will have specific regulations dealing with water plant wastes. For example, Virginia allows the disposal of water plant sludges in landfills if the solids concentration is over 20%. If the solids concentration is 20% to 35%, the sludge is to be mixed 6:1 (by volume) with solid waste; if the solids concentration is 35% to 60% they are to be mixed 4:1; and, if the solids concentration is over 60% the sludge can be disposed of without mixing in dry weather.

In order to classify a water plant waste as 'hazardous', hazardous must be defined. From an academic standpoint, Davis and Cornwell (2-2) have defined a hazardous waste as "any waste, or combination of wastes that poses a substantial danger, now or in the future, to human, plant, or animal life and which therefore cannot be handled or disposed of without special precautions." EPA has developed a more usable definition, by stating the ways in which a waste can be classified as hazardous: 1) by its presence on the EPA-developed lists, or 2) by evidence that the waste exhibits ignitable, corrosive, reactive or toxic characteristics. The regulations governing these definitions and the subsequent handling requirements are known as the Resource Conservation and Recovery Act of 1976, or RCRA. RCRA concerns the handling of wastes at currently operating facilities (such as water plants) and at facilities yet to be constructed. It was designed to meet disposal needs resulting from the Clean Water Act and the Clean Air Act. Those statutes require the removal of hazardous substances form air emissions and water discharges. Neither of these other statutes however assured that the disposal of the waste materials generated would be environmentally sound. RCRA was intended to provide that assurance. The five major elements of RCRA are:

1. Federal Classification of Hazardous Wastes
2. Cradle to Grave Manifest System
3. Federal Standards to be Followed by Generators, Treaters, Disposers, Storers of Hazardous Wastes
4. Enforcement of Federal Standards
5. Authorization of States to Obtain Primacy for Implementation of the Regulations.

So the major question is, are water plant wastes hazardous (as per current EPA definitions)? Water plant wastes are not on the developed list of specifically identified hazardous wastes, so that part of the definition does not apply. That leaves the properties of ignitability, reactivity, corrosivity or toxicity as a means of defining the waste material as hazardous. It is highly unlikely that water plant wastes will fail either of the first two. A waste is classified as corrosive if it has a pH less than or equal to 2 or greater than or equal to 12.5. It is possible that coagulant recovery side streams, perhaps filtrate from lime conditioning of sludge in a filter press, and brines from acid regeneration of ion exchange resins would fall outside these limits. The pH can be adjusted with appropriate neutralization.

Toxicity is evaluated by the EP toxicity test (40 CFR 261; 45 Federal Register Vol 45, page 33119, May 19, 1980). Basically the test is a measure of defined constituents that are present or will leach from the water plant waste. For a liquid waste the constituents are measured directly. For a solid waste the waste is held at pH 5.5 for several hours under defined procedures. If the liquid or extract from the waste contains concentrations greater then defined levels, then it is hazardous. Table 2-3 shows the currently defined contaminants for the EP toxicity test and their maximum allowed values. Those constituents and the levels are 100 times the drinking water MCL value. As this Handbook was being prepared, proposals were being considered to increase the number of constituents covered by the EP toxicity test, and to change the test method. It is likely that a continuing increase in the list will result.

Another set of regulations that could affect land disposal of water plant wastes is the Comprehensive Environmental Response, Compensation and Liability Act of 1980 (CERCLA). CERCLA provides authority for the removal of hazardous substances from improperly constructed or operated sites not in compliance with RCRA. The most noteworthy part of these regulations is that they allow clean up costs to be assessed against the user of the land disposal facility based on a volume use basis. The waste itself need not have directly caused the problem. For example, if a water utility disposed of its sludge at a private landfill that also accepted other industrial wastes which contaminated the

TABLE 2-3

MAXIMUM CONCENTRATION OF CONTAMINANTS FOR CHARACTERISTIC OF EP TOXICITY

Contaminant	Maximum Concen. (mg/l)
Arsenic	5.0
Barium	100.0
Cadmium	1.0
Chromium	5.0
Lead	5.0
Mercury	0.2
Selenium	1.0
Silver	5.0
Endrin (1,2,3,4,10,10-hexachloro-1, 7-epoxy-1,4,4a,5,6,7,8,8a-octahydro-1, 4-endo,endo-5,8-dimethano naphthalene)	0.02
Lindane (1,2,3,4,5,6-hexachlorocyclohexane, gamma isomer)	0.4
Methoxychlor (1,1,1-trichloro-2,2bis [p-methoxyphenyll] ethane)	10.0
Toxaphene ($C_{10}H_{1}0Cl_{8}$,technical chlorinated camphene, 67-69 percent chlorine)	0.5
2,4-D (2,4-dichlorophenoxyacetic acid)	10.0
2,4,5-TP Silvex (2,4,5-trichlorophenoxpropionic acid)	1.0

groundwater, the water utility could be liable for clean-up based on its volume use of the landfill, even if its sludge did not cause the problem. For this reason it is highly recommended that, if possible, the utilities use only landfills within their own governing jurisdiction.

Water plant wastes containing radium could come under the authority of three federal agencies: the Nuclear Regulatory Commission (NRC), the Environmental Protection Agency (EPA) and the Department of Transportation (DOT)[1]. As discussed below, however, none of the agencies directly regulates this type of waste. Currently, the ultimate authority for regulation of radium containing water plant wastes lies with the individual states.

The NRC is typically the initial contact with federal control of radioactive waste. This is the agency responsible for regulating the licensing, generation, containment and disposal of radioactive material. The NRC does not, however, regulate radium waste unless it is associated with nuclear fuel. The NRC has developed design criteria for the disposal of radium containing uranium mill tailings which can provide technical guidelines to utilities that must dispose of radium containing water plant wastes.

EPA is mandated to protect human health and the environment, therefore it has an interest in any waste potentially harmful to the environment. In this case EPA authority has been limited to avoid overlap with NRC authority. The extent of EPA involvement with radioactive wastes is shared responsibilities with NRC for uranium mill tailings. EPA was in the process of developing a guideline document for radioactive water plant waste disposal at the writing of this handbook.

The (DOT) regulates the shipment of any radioactive waste. DOT is a possible regulatory authority if the waste is

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[1] Discussion adapted from EPA document, "Disposal of Radium-Barium Sulfate Sludge," 908/6-82-009, Dec. 1982 (ref. 2-3).

shipped off site for disposal. The waste can be considered radioactive by DOT if 1) a state authority has designated the waste as radioactive, or 2) the radioactivity exceeds DOT established levels. DOT defines a radioactive waste as a material that has a specific activity of over 2,000 pCi/g. It is unlikely that any water plant sludge would exceed this level. However, if a state designates a waste as radioactive then DOT regulations apply. In such cases, shipment must be according to 49 CFR Part 172.392 which requires that the waste be packaged in leak-proof containers with acceptable levels of external radiation and transported in appropriately marked vehicles.

Specific regulation, therefore, other than transportation is left to the state agencies. South Dakota, for example, has regulations that require the water plant to be licensed as a generator of radioactive material.

Wisconsin has set the following criteria for landfilling of sludges containing radium:

- Solid waste containing 2 pCi/g (dry) or less of Ra^{226} can be landfilled in approved sanitary landfills.
- Solid waste containing greater than 2 pCi/g but less than or equal to 50 pCi/g of Ra^{226} can be disposed of in selectively approved sanitary landfills. The waste must be mixed with stabilizing solid waste so that the concentration of Ra^{226} averaged over any area of 100 m^2 will not exceed background levels by more then 5 pCi/g, averaged over any 15 cm thick soil below the surface.
- Solid waste containing over 50 pCi/g requires specific agency review.
- The radium containing waste should be disposed of in its own trench with separate liner and leachate collection/treatment system.

Illinois has similar regulations governing water plant wastes containing radium, with some additional restrictions to assure that the release of radon is less than 5 pCi/m^2·sec (see Section 3.4 for a discussion of radium/radon chemistry).

Land application of water plant wastes has generally not been governed by federal regulations. Guidelines for maximum metal applications for wastewater sludges are found in Table 2-4. These ranges may provide some insights for land application of water plant wastes (see Section 2.3 for use of these guidelines). Again, water plant wastes that contain radium may have special requirements. Wisconsin has temporarily halted land application of water plant wastes containing radium. Illinois allows the sludge to be used for soil conditioning on agricultural lands if the level of radium in the sludge is such that after the sludge is mixed with soil the incremental increase of the radium concentration in the soil does not exceed 0.1 pCi/g (see Section 2.3 for example calculation).

2.2. DIRECT DISCHARGE TO RECEIVING STREAMS

The direct discharge of water plant wastes into receiving bodies of water has traditionally been the most widespread disposal methodology. Over the past 15 years considerable debate has occurred on the continuance of this practice. While the debate has continued, very little direct research has been conducted on the environmental effects of water plant waste discharge.

This section focuses on situations where water plant waste discharges could have environmental effects and discusses the considerations that should be made in evaluating this disposal method. Many of the sections apply equally to solid/liquid wastes and to liquid phase wastes. Some additional emphasis has been placed on the discharge of alum sludges since that represents the greatest amount of water

TABLE 2-4

MAXIMUM AMOUNT OF METAL (LB/ACRE) SUGGESTED FOR AGRICULTURAL SOILS WITH SEWAGE SLUDGE

Metal	Soil Cation Exchange Capacity (meq/100 g)		
	<5	5 to 15	>15
	Maximum Amount of Metal (lb/acre)		
Pb	500	1,000	2,000
Zn	250	500	1,000
Cu	125	250	500
Ni	50	100	200
Cd	5	10	20

Source: EPA Sludge Treatment and Disposal (2-4) plant waste production and direct discharge of those wastes is still a common disposal method[2].

2.2.1. Determining Mass Changes

To assess the effects or determine the regulatory compliance of a discharge it is often necessary to calculate the in-stream concentration of the pollutant of interest. This is especially true for meeting in-stream water quality standards as presented in the previous section. The following equations can be used to estimate these in-stream concentrations or changes to in-stream concentration. They apply to solid/liquid wastes or liquid phase wastes. Caution should be exercised to distinguish between dissolved and non-dissolved pollutant concentrations.

If the water supply stream is the same as the water plant waste discharge stream as shown in Figure 2-1A then the following flux relationships exist for the river (F(1)) and the water treatment plant (F(2)):

$$F(1) = \frac{d(w_1)}{dt} = Q_1C_1 + Q_wC_w - QC_1 - Q_2C_2$$

$$F(2) = \frac{d(w_2)}{dt} = QC_1 + d(A) - Q_wC_w - (Q - Q_w)C_F$$

At steady state, and combining the two expressions results in:

$$Q_2C_2 = Q_1C_1 + d(A) - (Q - Q_w)C_F$$

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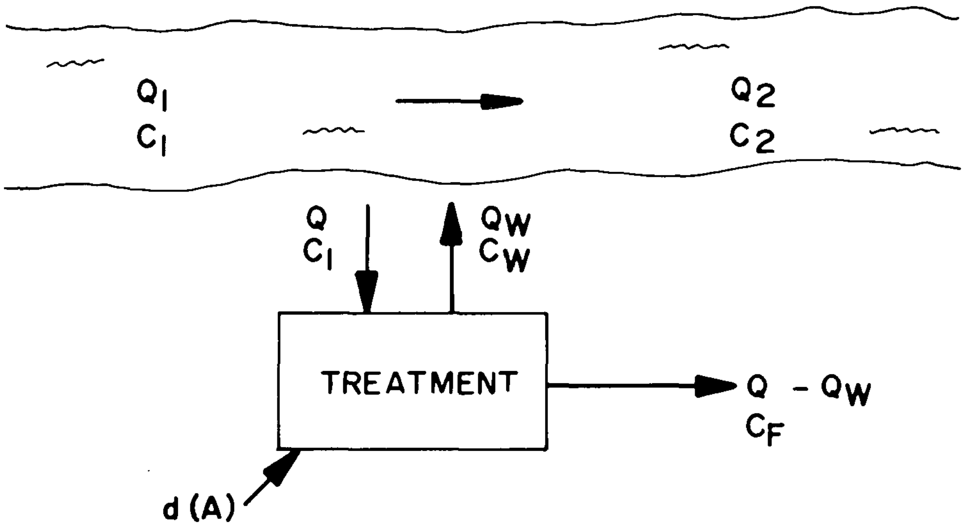
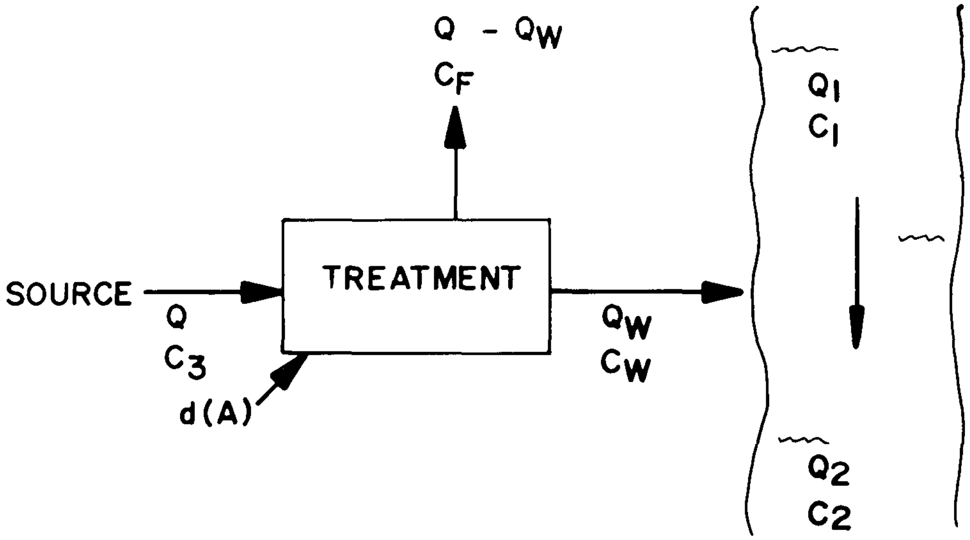
[2] Much of the information pertaining to direct discharge to receiving water courses is taken from the AWWA Research Division's Sludge Committee Report titled, "Research Needs for Alum Sludge Discharge" (ref. 2-5).



A. COMMON SOURCE



B. DIFFERENT SOURCES

FIGURE 2-1

MASS BALANCE FOR SUPPLY AND DISCHARGE

Reference: 2-5

where,

Q_1, Q_2 = Flow in river upstream (1) or downstream (2) of the water plant intake and the sludge discharge
Q = Flow into water treatment plant
Q_W = Flow of waste into the river
C_1, C_2 = Respective pollutant concentrations upstream and downstream of the plant
C_W = Concentration of pollutant in the waste stream
C_F = Finished water concentration of pollutant
d(A) = Change in pollutant due to chemical treatment, mass/time
w = Mass pollutant concentration
F = Flux change, mass concentration with time

or if the net change in mass is desired,

$$DW = Q_2C_2 - Q_1C_1 = d(A) - (Q - Q_W)C_F$$

If the discharge stream is not the same as the supply source as shown in Figure 2-1B, then the change in receiving stream concentration upstream and downstream of the discharge is:

$$Q_2C_2 - Q_1C_1 = QC_3 - (Q - Q_W)C_F + d(A)$$

where,

C_3 = Concentration of pollutant in the supply source

The change in pollutant due to treatment, d(A), is generally the chemical additions. For ion exchange systems the chemical additions would be the salt used for regeneration, which is known. For RO plants chemical addition is minimal. For coagulation and softening plants chemical additions are often reported as concentration with respect to the influent water plant flow,

$$d(A) = (dC_A)Q$$

d(A) would need to be found for the pollutant of interest. Some common examples are shown below:

Suspended solids : $dC_A = 0.44(\text{Alum}) + \text{PAC} + \text{Polymer}$

Total Aluminum (as AL) : $dC_A = 0.09(\text{Alum})$

Total Iron (as Fe) : $dC_A = 1.0\ (\text{Iron})$

Total Calcium (as $CaCO_3$) : $dCA = 2.0\ (\text{Ca Removed})$

The following example illustrates the calculation of a TDS change in the stream for the discharge of an ion exchange waste (see Section 3.2.2. for waste quantity calculations).

<u>Given</u>

Q = 1 mgd = 3.75×10^6 l/d
$d(A)$ = 2,100 lb NaCl/day = 9.53×10^8 mg/d
Q_W = 19,300 gpd = 7.24×10^4 l/day
$C_3 = C_F$ = 500 mg/l
C_1 = 100 mg/l
$Q_2 = Q_1$ = 10 mgd = 3.75×10^7 l/day

The above assumes $C_3 = C_F$, that is the weight of cations removed from the raw water equals the weight of sodium added to the finished water. It also assumes that the salt added is equal to the TDS in the discharge. Both assumptions are very close to correct (exact calculations are in Section 3.2.2.).

$$Q_2C_2 - Q_1C_1 = QC_3 - (Q - Q_w)C_F + d(A)$$

$$3.75 \times 10^7 C_2 - 3.75 \times 10^7(100) = 3.75 \times 10^6(500) - (3.75 \times 10^6 - 7.24 \times 10^4)500 + 9.53 \times 10^8$$

$$C_2 = 125 \text{ mg/l}$$

If the receiving stream had a 500 mg/l TDS limit for the in-stream water quality standard, the discharge would have no problem meeting this criteria. The procedure can be done for any of the in-stream water quality criteria.

The above relationships assume that homogeneity is reached instantaneously, which is not the case. The time, or distance downstream, for these average conditions to be reached is determined by the rates of diffusion and dispersion. Until the average conditions are achieved, the concentrations in certain portions of the water will be higher than the average while other portions will have lower concentrations.

2.2.2. Benthic Impacts

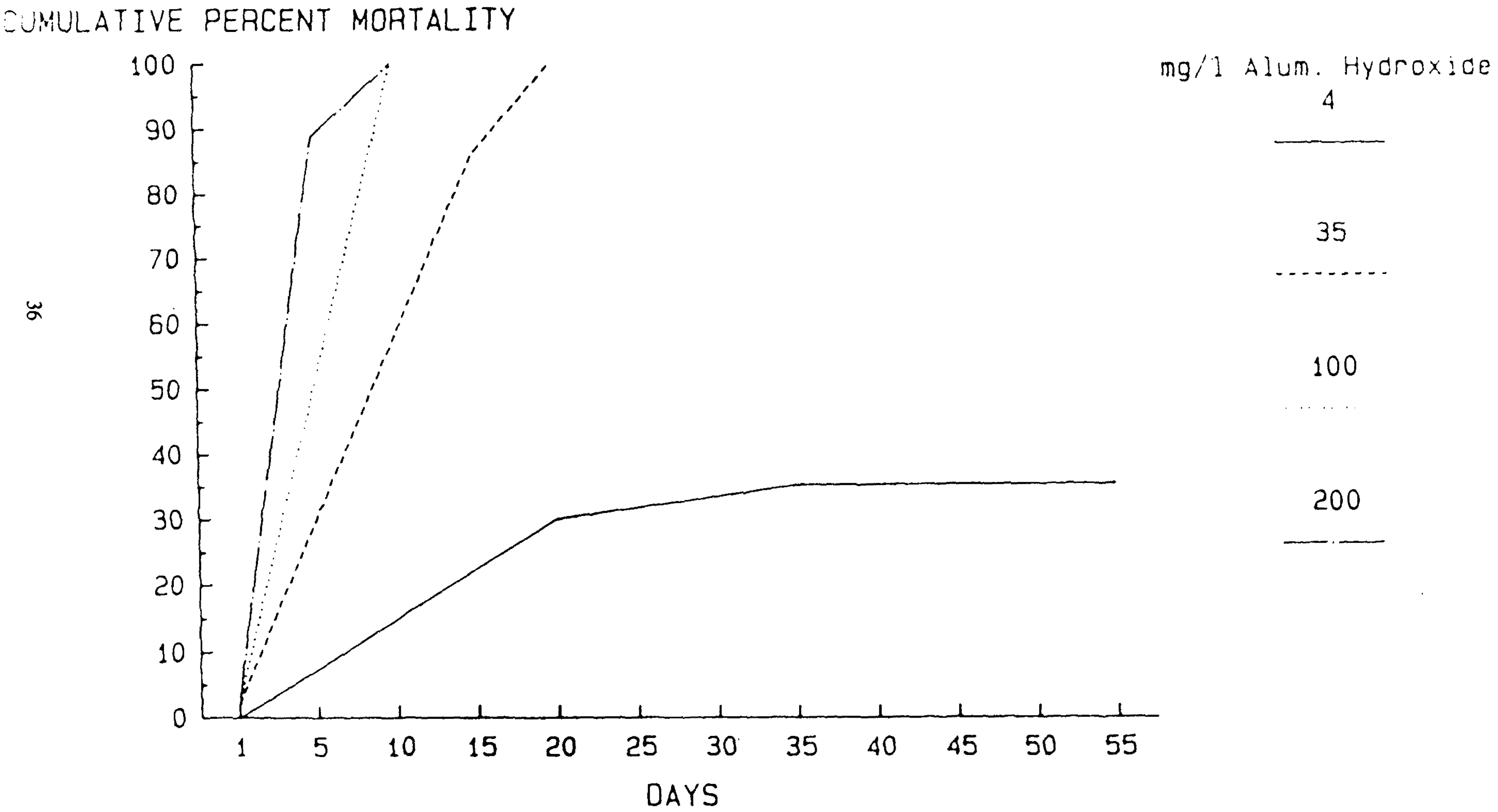
If the receiving stream or lake has a low velocity and the discharge is a solid/liquid waste then the above calculations and discussions do not apply. In this case the material contained in the residue discharge will tend to collect in a sludge deposit in the vicinity of the point of discharge. While very little work has been published on the benthic impacts of water plant waste deposits, it is likely that the ecosystem in the immediate vicinity of the sludge blanket will be impacted. In addition to the impact of the presence of the blanket itself, anaerobic conditions may develop, resulting in an increase in solubilization of metals, both from within the sludge and the bottom soil, a lowering of the pH, and a release of odors. One related research effort was conducted by Lamb and Bailey (2-6) for alum sludges. Their interest was on the effects of using alum to precipitate phosphorus in highly eutrophic lakes. The objective of their study was to determine the acute and chronic effects of aluminum hydroxide to Tanytarsus dissimilis. T. dissimilis is a representative of the chirocomidae, the family of organisms which occupies a significant portion of the benthic invertebrate community of lakes and important fish food organism. The acute tests were conducted by adding concentrations of alum to beakers at pH 7.8 of up to 960 mg/l (960 mg/l alum is equivalent to approximately 400 mg/l of suspended solids). The results of the acute bioassays indicated no apparent effects.

Mortalities in the chronic assays, however, were recorded at all alum doses. A simplified version of a graph of the results is shown in Figure 2-2. The results show a general increasing of mortality with an increase in alum sludge (aluminum hydroxide) although discrepancies within this conclusion existed. The time to reach 50% mortality was shortest at the 480 mg/l alum dose at about 4 days. The mortality time for 80 mg/l and 240 mg/l was not statistically different at about 9 days. The 10 mg/l dose showed 37% mortality at 55 days. The major discrepancy was the 960 mg/l dose which had a 50% mortality time of 23 days. The research concluded that there was some chemical toxicity at alum concentrations of 80, 240 and 480 mg/l and at least a stressed condition developed at 960 mg/l. They did not believe that the solid aluminum hydroxide itself was toxic, but rather some of the species, perhaps the negatively charged aluminates, tended to concentrate toxic materials. They also felt that a heavy floc layer on a lake bottom could inhibit pupae from reaching the surface and the deposited eggs from reaching the sediment.

Roberts and Diaz (2-7) conducted studies to quantify the effects of alum sludge discharge into a tidal stream at Newport News, Virginia. The streams that they studied were shallow with a low velocity. During their work phytoplankton

FIGURE 2-2

CHRONIC MORTALITY BY ALUMINUM HYDROXIDE TO BENTHIC INVERTEBRATES



Reference: 2-6

productivity was measured during an alum sludge discharge event. Table 2-5 shows clearly that production at the station nearest sludge discharge was suppressed. The authors concluded that a negative impact was associated with the alum sludge discharge. They explained the depression of production by the high suspended solids associated with the discharge. During sampling, light intensity was observed to decline virtually to zero at a depth of only a few millimeters. Thus they concluded, "this turbidity effect in itself would argue for the cessation of sludge discharge even in the absence of toxic effects". Net production of the benthic algae community was also negative during sampling periods, reflecting the low light intensity.

2.2.3. Aluminum Toxicity

Aluminum toxicity itself remains as one of the major concerns regarding the effects of alum sludge discharge. While research has not been conducted on aluminum toxicity related to alum sludge discharge, related research has been conducted in the acid rain field on the potential for aluminum toxicity. Because aluminum is such a ubiquitous element, the question of its toxicity is more difficult to resolve than that of the heavy metals (for which toxicity data were given in Table 2-2). Thus, although the toxicity of soluble aluminum has been demonstrated for a wide spectrum of plants and animals the overwhelming percentage of aluminum in the world is without measured toxic effect. Therefore, one must be able to distinguish among the forms of aluminum. The chemistry of aluminum in water is essentially that of aluminum hydroxide. It is characterized by being readily amphoteric and forms complex ions with other substances in water and polymerizes. When an aluminum salt of non-complexing characteristic is placed in water it undergoes the basic hydrolyses reactions:

TABLE 2-5

Plankton Productivity, Chlorophyll Concentration and Assimilation Ratio in the Poquoson and Warwick Rivers During March and June, 1985

Date	Carbon Uptake ($mg\ C/m^3/h$)		Chlorophyll a Concentration ($mg\ chlor\ a/m^3$)		Assimilation Ratio ($mg\ C/h/mg\ chlor\ a$)	
POQUOSON RIVER	Sta. 1	Sta. 6	Sta. 1	Sta. 6	Sta. 1	Sta. 6
03-05-84	1.4	18.9	0.85	8.7	1.6	2.4
03-06-84	10.4	30.5	3.7	8.5	2.8	3.6
06-18-84	<0.1	40.1	1.1	12.6	<0.1	3.2
06-19-84	11.7	88.7	62.9	15.8	0.2	5.6
WARWICK RIVER	Sta. 3	Sta. 6	Sta. 3	Sta. 6	Sta. 3	Sta. 6
03-22-84	6.1	53.6	3.7	12.4	1.6	4.3
06-21-84	3.1	85.3	25.5	16.9	0.1	5.0

Station 1 is near the point of discharge and Station 6 is downstream of the discharge.

Source: Roberts and Diaz (2-7)

$$Al(H_2O)_6^{3+} + H_2O = Al(H_2O)_5OH^{2+} + H_3O^+$$

$$Al(H_2O)_5OH^{2+} + H_2O = Al(H_2O)_4(OH)_2^+ + H_3O^+$$

$$Al(H_2O)_4(OH)_2^+ + H_2O = Al(H_2O)_3(OH)_3 + H_3O^+$$

and as the pH increases further the aluminate ion is formed

$$Al(H_2O)_3(OH)_3 + H_2O = Al(H_2O)_2(OH)_4^- + H_3O^+$$

which is the apparent limit of hydrolysis as the (-2) and (-3) species have not been reported. Given equilibrium constants for the above one can then plot the solubility of aluminum in the presence of solid aluminum hydroxide as a function of pH. Often plots of aluminum in equilibrium with crystalline aluminum hydroxide (Al_2O_3) are reported. However, the solubility of aluminum in the presence of colloidal aluminum hydroxide such as would exist in sludges is higher, and is shown in Figure 2-3. The diagram shows that at pH 5 the equilibrium concentration of soluble aluminum is about 0.135 mg/l. The diagram of Figure 2-3 is representative of kinetically reaching solids species of $Al(H_2O)_3(OH)_3$. As the solids crystallize further to Al_2O_3 (gibbsite) the solubility decreases. This crystallization to gibbsite may take a year or longer. During this time the pH will drop slightly as bound hydroxide becomes intimately incorporated with the crystal structure. Thus an organism in contact with freshly prepared sludge will be exposed to different soluble aluminum concentrations and different degrees of polymerized or crystallized solid species than an aged sludge. This, of course, further complicates the evaluation of alum sludge toxicity.

Aluminum is also capable of forming strong coordinate bonds with substances other than water. Complexes are formed with inorganic ligands such a fluoride and silicates. Insofar as complexing agents are present in water they will increase

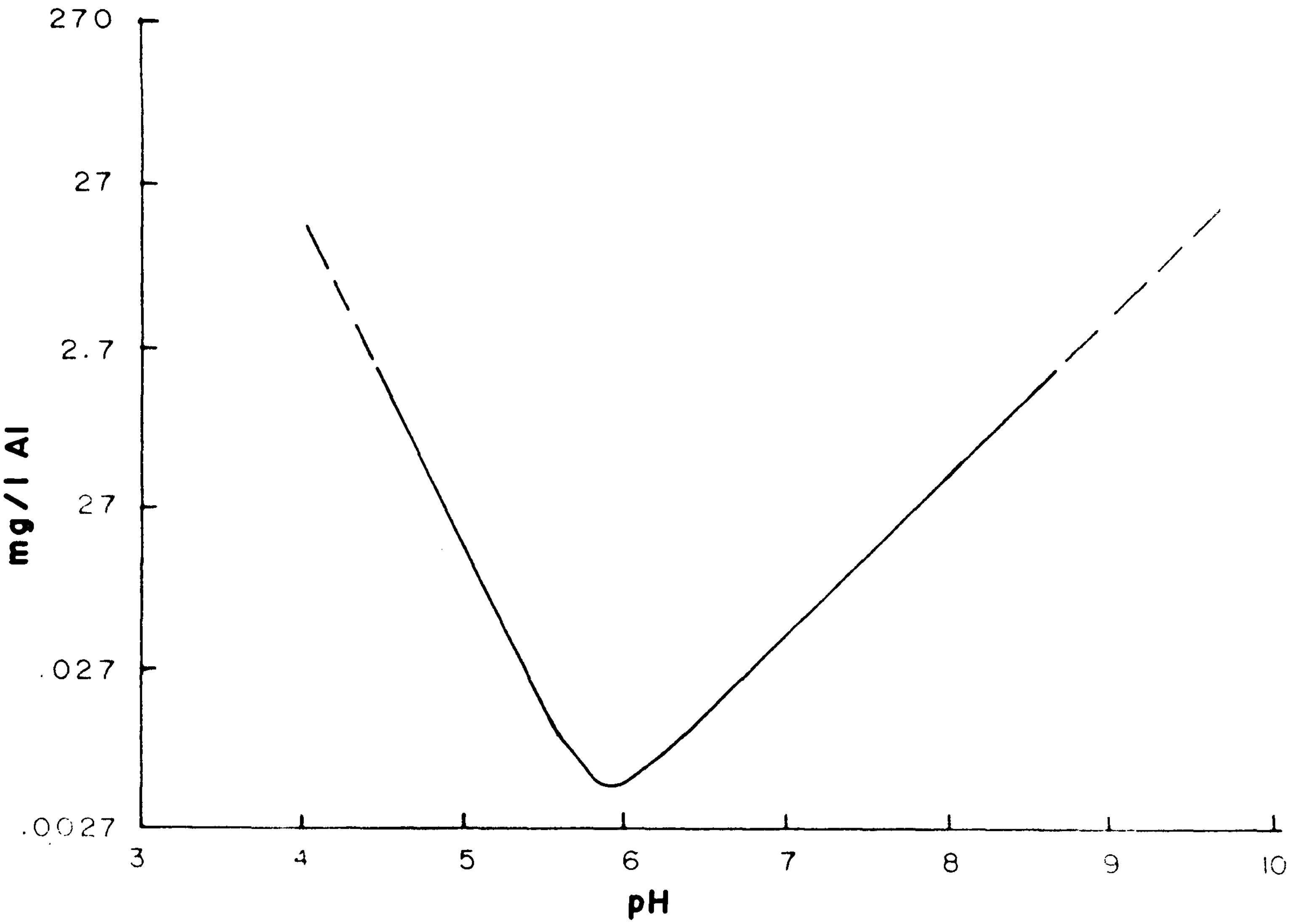


FIGURE 2-3

SOLUBILITY OF FRESHLY PRECIPITATED ALUMINUM HYDROXIDE the amount of dissolved aluminum in equilibrium with solid aluminum hydroxide. However, in natural waters the concentrations of these inorganic ligands is usually not high enough to appreciably effect aluminum solubility. Many organic materials of natural origin are capable of forming dissolved complexes with aluminum such as humic and fulvic acids, polyphenols, sugars and organic acids. All of these change the form of aluminum present and its solubility.

The following discussion presents some of the data on aluminum toxicity which may be pertinent to alum sludge discharge. Toxicity is discussed in terms of free aluminum ion, solid aluminum hydroxide (which has also been discussed with respect to benthic impacts) aluminate and soluble aluminum complexes.

Schofield and Trojnar (2-8) indicated that levels of aluminum may be the primary factor limiting survival of trout in the Adirondack lakes. They report brook trout mortality at aluminum levels of 0.2 to 0.5 mg/l. However, they did not distinguish between the type of aluminum species present. Driscoll et al. (2-9) conducted studies on the toxicity effects of different species of aluminum. They utilized synthetically prepared waters as well as waters obtained from the Adirondack lakes. Brook trout fry were exposed in soft, synthetic water to aluminum as the free ion and complexed with hydroxide, fluoride or citrate. The results are shown in Table 2-6. Survival of brook trout fry was significantly inhibited at total aluminum concentration of over 0.4 mg/l without the presence of complexing ligands. Fluoride somewhat increased survival and the addition of citrate significantly increased survival. They concluded that the inorganic aluminum forms seem to be the major species of concern in direct aluminum toxicity. They then conducted studies on white sucker fry (Figure 2-4), using natural waters and synthetic solutions. They estimated the 50% survival time as a function of aluminum concentration reported as free aluminum

TABLE 2-6

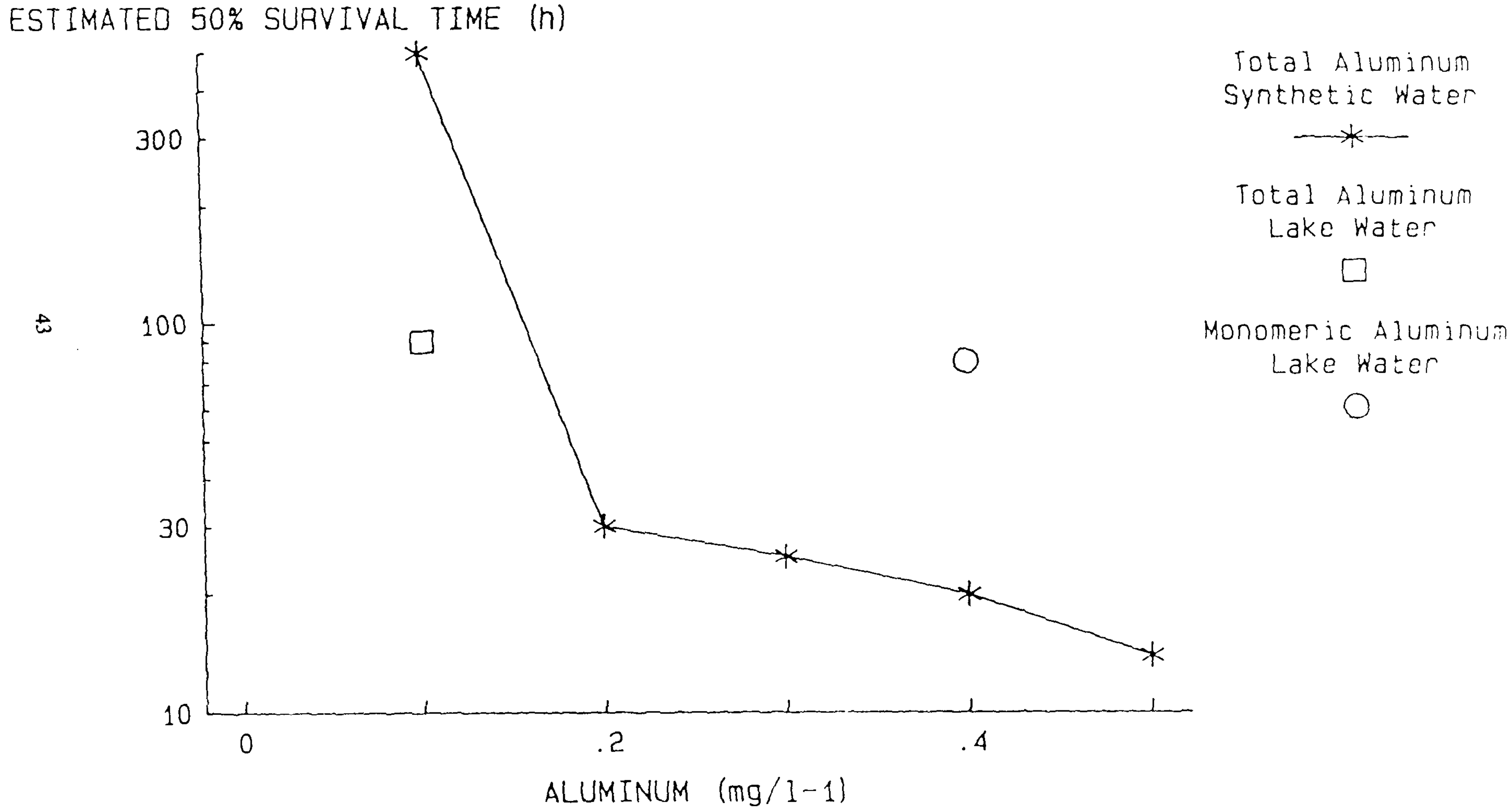
Estimated 50% Survival Times of Brook Trout (Salvelinus fontinalis, Mitchill) in Synthetic Soft Water Solutions Containing Aluminum and Aluminum Complexing Ligands

	Ligands	Total Aluminum Concentration	% Survival After 14 Days		Estimated 50% Survival Time (h)	
pH	Added	mg/l	Mean	Range	Mean	Confidence Limits
5.1	---	0.02	99	95-100	>500	---
	---	0.42	28	20- 30	115	88-150
	Fluoride	0.50	45	35- 45	202	136-302
	Citrate	~0.5	96	87-100	>500	---
4.4	---	0.01	95	80-100	>500	---
	---	0.48	42	27- 53	256	213-306
	Fluoride	0.47	71	67- 80	472	354-629
	Citrate	~0.5	82	40-100	>500	---

Source: Driscoll (2-9)

FIGURE 2-4

50% SURVIVAL TIME OF WHITE SUCKER FRY (18.7mm)



Reference: 2-9

ion and total aluminum. A significant decrease in the 50% survival time occurred near 0.2 mg/l aluminum ion concentration in both the synthetic and natural waters.

Freeman and Everhart (2-10, 2-11, 2-12) performed a partial chronic bioassay for aluminum using the growth rate of rainbow trout in continuous flow aquariums. They used waters containing 0.05, 0.52 and 5.2 mg/l total aluminum at various pH ranges. Due to the pH utilized, they primarily studied the effects of aluminum hydroxide and aluminate.

At pH 8, 90% of 5.2 mg Al per liter is suspended and 10% dissolved, as shown in Table 2-7. Feeding activity diminished within 24 hr, and gill hyperplasia (a swollen, congested condition) was evident within 5 days in many of the trout. The physical condition of all fish continued to deteriorate, with individuals suffering from inability to maintain equilibrium, general listlessness, loss of fright reaction, loss of negative phototaxis, darkening in coloration, and eventual death. Almost none of the trout exposed for the full 45 days survived, even after transfer to uncontaminated water. When exposed only to dissolved aluminum (0.52 mg/l) at the same pH, fish exhibited milder forms of the symptoms described above, with slower development of hyperplasia and loss of appetite. Mortality was greatly reduced, recovery was rapid and almost complete within 48 hr of transfer to uncontaminated water, and normal weight gain resumed. An aluminum level of 0.05 mg/l had no apparent effect. At pH 8.5, 32% of 5.2 mg/l aluminum is dissolved and at pH 9, 10% is dissolved. The above severe symptoms appeared in fish exposed to these conditions, but they appeared more rapidly. In addition fecal casts were evident. Exposure was terminated after 222 hr and 113 hr at the two pH conditions, respectively. On transfer of surviving trout to clean water, recovery was rapid in the first case and delayed, but apparently complete, in the second.

At pH 7, 99% of 5.2 mg/l aluminum was suspended. Symptoms were similar to those at pH 8, although they develop-

TABLE 2-7

CHRONIC ALUMINUM INTOXICATION AND RECOVERY FOR TROUT

	Al Concentration (Total Aluminum)									
	pH 7			pH 8			pH 8.5		pH 9	
	Control	0.52	5.2	Control	0.52	5.2	Control	5.2	Control	5.2
Exposure time (days)	45	45	45	45	45	45	9.25	9.25	4.7	4.7
Weight Change during bioassay (%)	+182	-3	-26	+192	+76	-18	+50	-32	+84	0
Mortality during bioassay (%)	2	44	58	0	8	77	0	51	0	67
Recovery time (days)	290	290	290	16	16	1	161	161	165	165
Weight gain during recovery relative to control weight gain (%)	100	164	195	100	101	--	100	159	100	141
Mortality during recovery (%)	3	5	16	--	0	83	5	3	2	2

Source: Freeman ((2-11) and Everhart and Freeman (2-10, 2-12) ed more slowly and mortality was slightly lower. Recovery in clean water was very slow. Fish exposed to 0.52 mg/l of aluminum at pH 7 (90% suspended) exhibited milder symptoms, but the mortality over 45 days was much higher than at pH 8 and recovery was very slow. The results of Freeman and Freeman and Everhart can be summarized as follows:

1. Aluminate ion is acutely toxic to trout at levels of 0.5 mg/l and greater. Recovery is rapid on transfer to uncontaminated water.
2. Aluminate ion also causes chronic injury, viz., gill hyperplasia. Recovery on transfer to clean water is rapid, perhaps because fish with extensive chronic damage will already have succumbed to acute effects.
3. Freshly precipitated aluminum hydroxide does not cause acute intoxication of fish in the usual sense, but it can cause chronic injury. The symptoms are similar to those induced by aluminates, but they are much slower in developing and recovery is very slow.

Hall et al. (2-13) conducted in-stream studies on the episodic effect of aluminum addition. They artificially added aluminum to streams and monitored biological changes. Some of their conclusions are summarized:

> These field and laboratory results suggest that episodic increases in Al can have significant biological and physical as well as chemical consequences in dilute, acidic surface waters. Macrobenthic community structure (e.g. distribution, abundance, and diversity) and function (e.g. trophic interactions and nutrient cycling) may be disrupted by episodic addition of soil-derived Al to streams

that become acidified during snowmelt. Moreover, the reduction of surface tension in streams or lakes receiving Al may represent a consequence of acidification that could affect aquatic biota and even some terrestrial forms by physically interfering with emergence, egg deposition, feeding, and mating behavior.

Clearly it is difficult to conclude which, if any of the above aluminum toxicity studies apply to the discharges of alum sludge. However, the evidence certainly suggests that water treatment plant managers should be aware of the potential problem and take precautions as appropriate.

2.2.4. Toxicity of Non-Aluminum Compounds

Most inorganic ions other than aluminum which could be discharged to a stream as part of a solid waste or a liquid phase waste have established chronic mortality levels and are regulated by state agencies. In Table 2-2 can be found mortality levels for such ions as iron, arsenic, barium, fluoride, nitrate, TDS and radioactivity. This Table, in conjunction with the mass balance calculations of Section 2.2.1, should be used to assess the toxicity of discharges of specific ions.

2.3. DISCHARGE TO THE WASTEWATER PLANT

Almost unanimously, water treatment plant operators would be happy to discharge their waste to the wastewater plant. In fact, disposal of water plant wastes to the wastewater plant is widely practiced and can be done very successfully. On the other hand, there have been failures with this method.

This section primarily discusses the effects of solid/liquid waste discharged to the wastewater plant. However, many liquid waste streams, including liquids resulting from the dewatering of solid/liquid wastes, are discharged to the sewer. The section on Effects on Biological Process considers the toxicity of components in liquid phase wastes and then discusses the effects of solid/liquid wastes on the biological process. Landfill and land application Sections deal only with solid wastes.

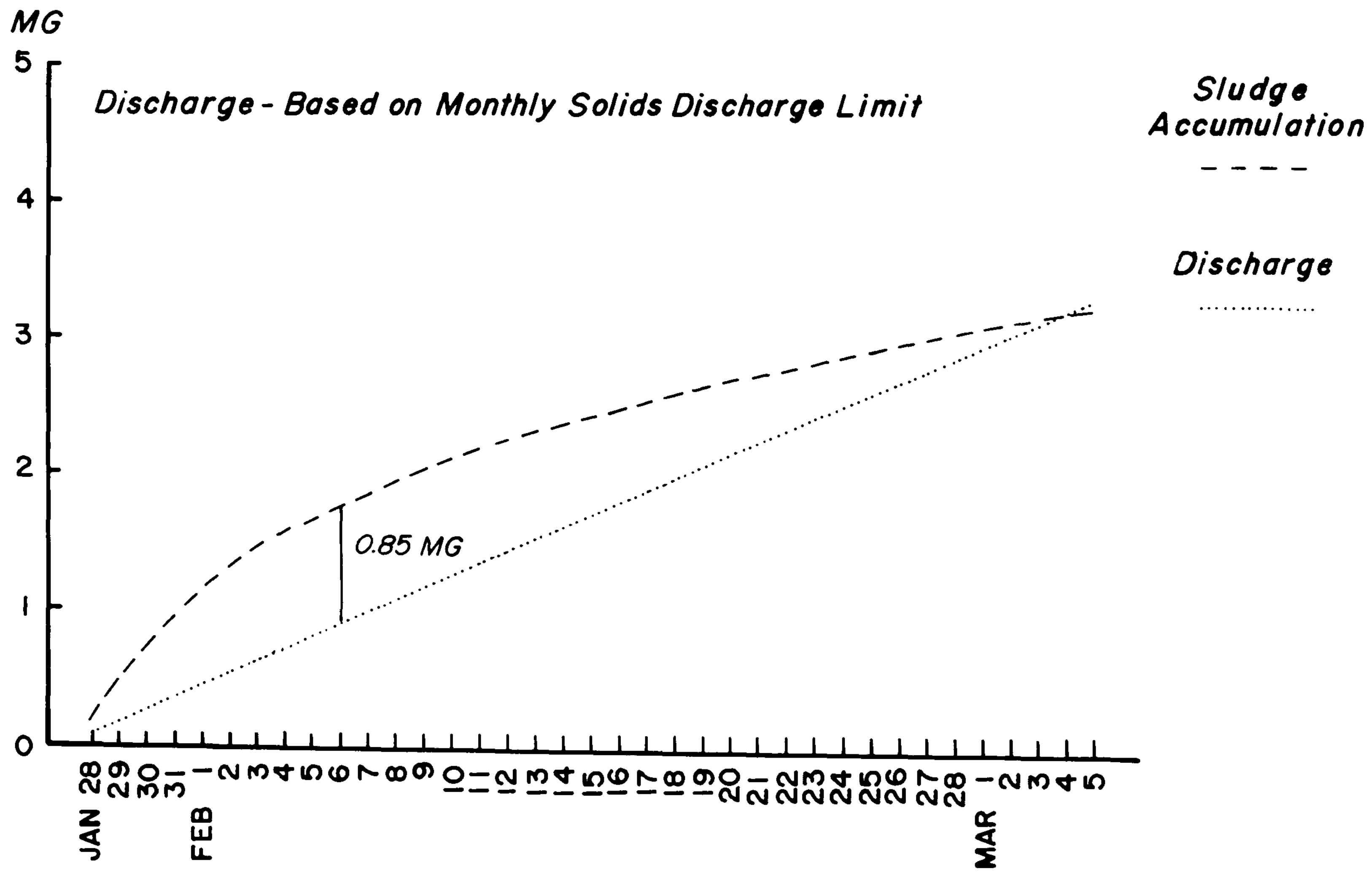
2.3.1. Equalization

Sludge from sedimentation basins can be withdrawn on a fairly continuous and uniform basis if the basins are equipped with sludge removal mechanisms. In this case it may be possible to directly discharge the sludge into the sewer system. Often, however, basins are cleaned on a discontinuous basis and equalization is required prior to discharge. Backwash water is produced at very high flow rates for short periods of time and equalization is nearly always required prior to sewer discharge. Regeneration wastes from ion exchange processes are produced only during the time of media regeneration and equalization may be needed.

It may also be that certain restrictions are placed on the discharge by the wastewater plant, such as time of day of discharge, instantaneous flow, maximum flow over a certain period of time or maximum solids discharge. Figure 2-5 shows an example mass storage diagram approach used for a water plant that had a maximum discharge limitation of solids. During peak solids production periods, storage was required. This particular storage volume was based on storing an average of 2% solids concentration. If a higher solids concentration is achieved then a smaller storage is required. This tradeoff usually results in the construction of a basin capable of decanting and thickening. Either a continuous flow or batch fill and draw thickener could be used, such as discussed in Section 4.3. It may be desirable to have the capability to mix the contents of the equalization/thickener tank and this should be considered in the design.

FIGURE 2-5

EXAMPLE SLUDGE EQUALIZATION FOR SEWER DISCHARGE



Equalization facilities for backwash water or similar high flow short duration productions prior to sewer discharge are similar to backwash equalization facilities discussed in Section 3.2.

Consideration should also be given to time of discharge. It may be desirable to discharge during periods of the day when sufficient flow is in the sewer to maintain desired velocities. On the other hand at certain times the sewer may flow full and a sludge discharge is undesirable. Generally, a velocity of about 2.5 fps should be maintained to prevent sedimentation of hydroxide sludge solids. Lime sludge may have settling velocities much higher than coagulant sludges, and it can be difficult to prevent its deposition in sewer lines. For the discharge of compounds toxic to the biological process, it may be necessary to equalize flows to allow for a continuous discharge or proper dilution.

2.3.2. Effects on Biological Processes

Possible effects, either beneficial or detrimental, of water plant wastes on the biological wastewater process are in the areas of toxicity to the biological processes, suspended solids removal or increases, BOD/COD removal or increases, and phosphorus removal.

Potential toxicity to the biological process itself primarily applies to the discharge of liquid phase wastes to the wastewater plant. Dissolved solids present in the liquid waste could be available in a form and present in a sufficient concentration to hinder the biological process. Defining the toxic effects of inorganic compounds on the biological wastewater treatment process is not a simple procedure. An initial shock load of a toxic compound can have an inhibitory effect on the biological process. However, with many compounds the microorganisms will adapt and adjust to the presence of the inorganic ion. Therefore, even if a pretreatment standard is being met, it is a good rule of thumb to equalize the discharge according to sewer flow patterns in order to provide a fairly uniform concentration of waste to the biological process.

Threshold concentrations of toxic ions in the biological process are the level above which a decrease in the COD removal efficiency of the biological process occurs. Table 2-8 shows data compiled by Srinath (in ref. 2-14) via a literature review on the threshold concentration for various metals in the activated sludge process. The metals listed in Table 2-8 can be present in liquid waste streams from water plants. However, several other compounds can also be present for which little data on the toxicity to the biological process are available. As discussed in Chapter 3, liquid wastes can contain fluoride, barium, nitrate, arsenic, radium and total dissolved solids. Nitrate loading rates could be a problem if the wastewater plant is required to denitrify.

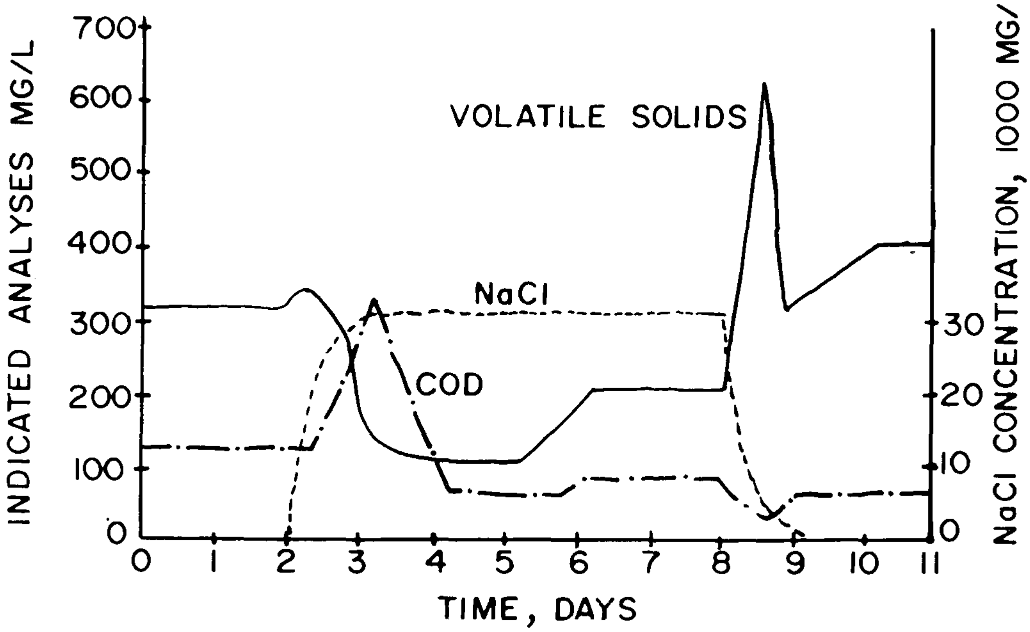
The effects of changing concentrations of salts in the wastewater due to slug discharge of high TDS water plant wastes is of considerable significance. A concentration change of 100 to 200 mg/l of NaCl is rather low and would have no impact on the biological process, while a concentration change of 35,000 mg/l NaCl (almost an undiluted discharge of ion-exchange waste) would cause considerable stress to the biological organisms. A mid-range change of 10,000 mg/l NaCl can actually stimulate growth. Figure 2-6 shows response curves of a completely mixed once through biological reactor to shock loadings of sodium chloride (2-15). Part A is a dose response to 30,000 mg/l NaCl. As the salt was added, the biomass concentration rose slightly and then decreased significantly. Correspondingly the effluent COD increased almost immediately after the salt addition. Within 2 days the substrate removal efficiency was regained. The volatile solids concentration was abnormally low but had partially recovered at day 6. Upon removal of the salt feed to the system the volatile solids immediately increased and leveled

TABLE 2-8

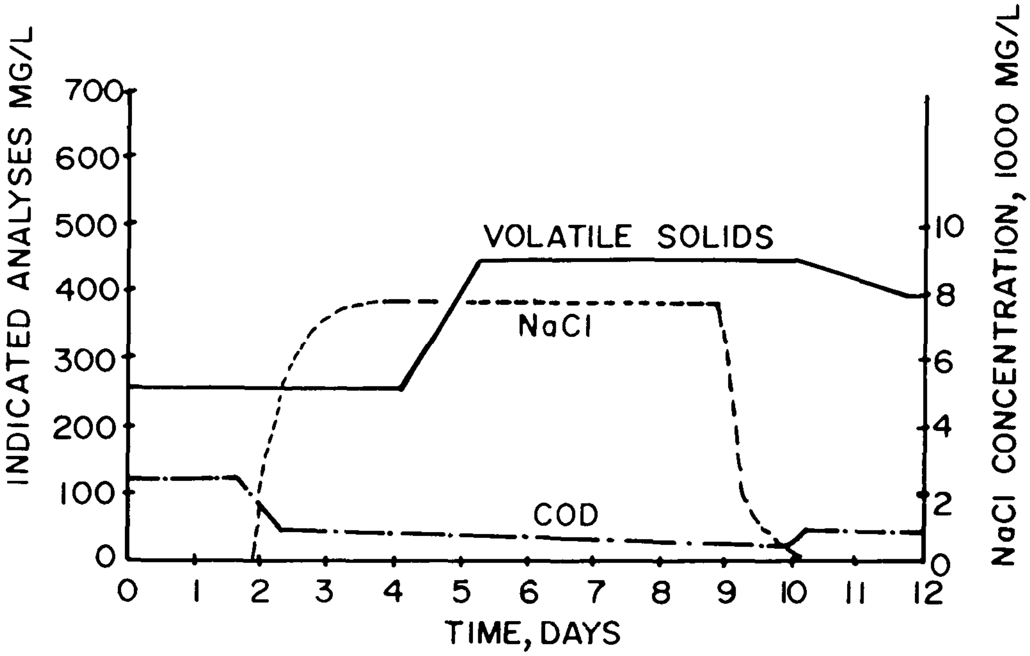
THRESHOLD CONCENTRATION FOR VARIOUS METALS IN THE AIR ACTIVATED SLUDGE PROCESS

Metal Ion	Concentration (mg/l)	Type A.S. Experiment
Silver	<.03	Carbonaceous
Vanadium	10.0	Carbonaceous
Zinc	2.0	Carbonaceous
	5.10	Carbonaceous
	1.0	Nitrification
Nickel	1.0 - 2.5	Carbonaceous
	1.0	Nitrification
	2.0	Nitrification
Chromium, $^{+6}$	10.0	Carbonaceous
	1.0	Nitrification
	10.0	Nitrification
	1.0	Nitrification
Chromium, $^{+3}$	10.0	Carbonaceous
Lead	10	Carbonaceous
Iron (Ferric)	15	Carbonaceous
Copper	1.0 - 10.0	Carbonaceous
	1.0	Nitrification
	2.0	Nitrification
Cadmium	1.0	Carbonaceous
	5.0	Nitrification

Source: E.G. Srinath (from reference 2-14)



NaCl DOSE=30,000 MG/L



NaCl DOSE= 10,000 MG/L

FIGURE 2-6

RESPONSE OF BIOMASS TO NaCl SHOCK LOADING

Reference: 2-15

off at a slightly higher concentration than prior to the shock loading. Part B of Figure 2-6 shows the effects of a shock loading of 10,000 mg/l NaCl. A significant increase in cell yield (about 75%) was sustained until the salt feed was stopped. The effluent COD also decreased during the period of 10,000 mg/l salt addition.

From this research it would appear that the effects of brine discharges from water treatment plants should have an insignificant impact on the wastewater treatment process as long as some precautions for equalization and dilution in the wastewater are followed. Toxicity of specific compounds within the waste streams should be carefully monitored and diluted sufficiently to eliminate toxic effects.

In addition to effects on the aerobic biological process, salts can affect anaerobic digestion. Toxicity is normally associated with the cation rather then the anion of the salt. Table 2-9 (2-16) shows the stimulatory and inhibitory concentrations for some cations of interest.

Little data are available on the effects of barium on the activated sludge process. Experimental data indicate that the soluble barium concentrations would have to exceed 50 mg/l (2-17) for toxicity to fresh water aquatic life to occur. In most wastewaters, sulfate or carbonate would be present to precipitate at least some of the barium as an insoluble, non-toxic compound. To precipitate 1 mg/l of barium requires about 0.7 mg/l of SO_4^{-2} or 0.23 mg/l of oxidized sulfide. Calculations should be conducted to assure the dilution or reduction of barium to an acceptable level.

Selenium can be found in small concentrations in ion exchange wastes from some groundwater treatment systems. Selenium is toxic to bacteria at 90 mg/l (2-17) and 180 mg/l for protozoan. It is unlikely these concentrations would be reached in a waste discharged to the sewer system.

Trivalent arsenic is highly toxic to invertebrates. Daphnia exhibit toxic symptoms at 4 mg/l arsenic (2-17). Some

TABLE 2-9

STIMULATORY AND INHIBITORY CONCENTRATIONS OF ALKALI AND ALKALINE - EARTH CATIONS

		mg/l	
Cation	Stimulatory	Moderately Inhibitory	Strongly Inhibitory
Sodium	100 - 200	3,500 - 5,500	8,000
Calcium	100 - 200	2,500 - 4,500	8,000
Magnesium	75 - 150	1,000 - 1,500	3,000

Source: McCarty (2-16)

of the lower order freshwater fish food organisms are affected at arsenic concentrations of 1.3 mg/l (2-17).

One of the more comprehensive studies on the effects to biological processes related to alum sludges was conducted by Rolan (2-18), in laboratory studies followed later by full scale work (2-19). Rolan added various concentrations of alum sludge to the raw wastewater and observed the impact on several performance criteria. An increase in alum sludge had a slight detrimental effect on the supernatant suspended solids quality resulting from primary sedimentation. Both turbidity and suspended solids increased by 25 to 40% as alum sludge was added at up to 200 mg/l of dry weight solids. Figure 2-7 shows the results for settled suspended solids concentrations from primary clarification. On the other hand, as Figure 2-8 shows, the settled BOD decreased slightly as the alum sludge dose was increased, possibly due to some adsorption of BOD onto the alum floc particles.

The work did find that the addition of alum sludge would lower the phosphorus content of the primary effluent. Figure 2-9 shows that about 1 mg/l of phosphorus was removed by 100 mg/l of alum sludge, probably by sorption onto the aluminum hydroxide matrix. Rolan conducted tests on the effects of sludge aging which showed a definite decrease in phosphorus adsorption capacity with an increase in sludge age from 6 hours to 2 months. It was concluded that while some phosphorus removal by the addition of alum sludge occurred, the removal itself was relatively insignificant as a treatment technology. Accordingly a series of tests was conducted to determine if water treatment plant sludge in conjunction with fresh alum would reduce the overall chemical costs for phosphorus removal. One series of samples was dosed with alum and another series was dosed with alum and water treatment plant sludge. The difference in final phosphorus concentration between the two was attributed to removal by the water treatment plant sludge. Contrary to the researcher's expecta-

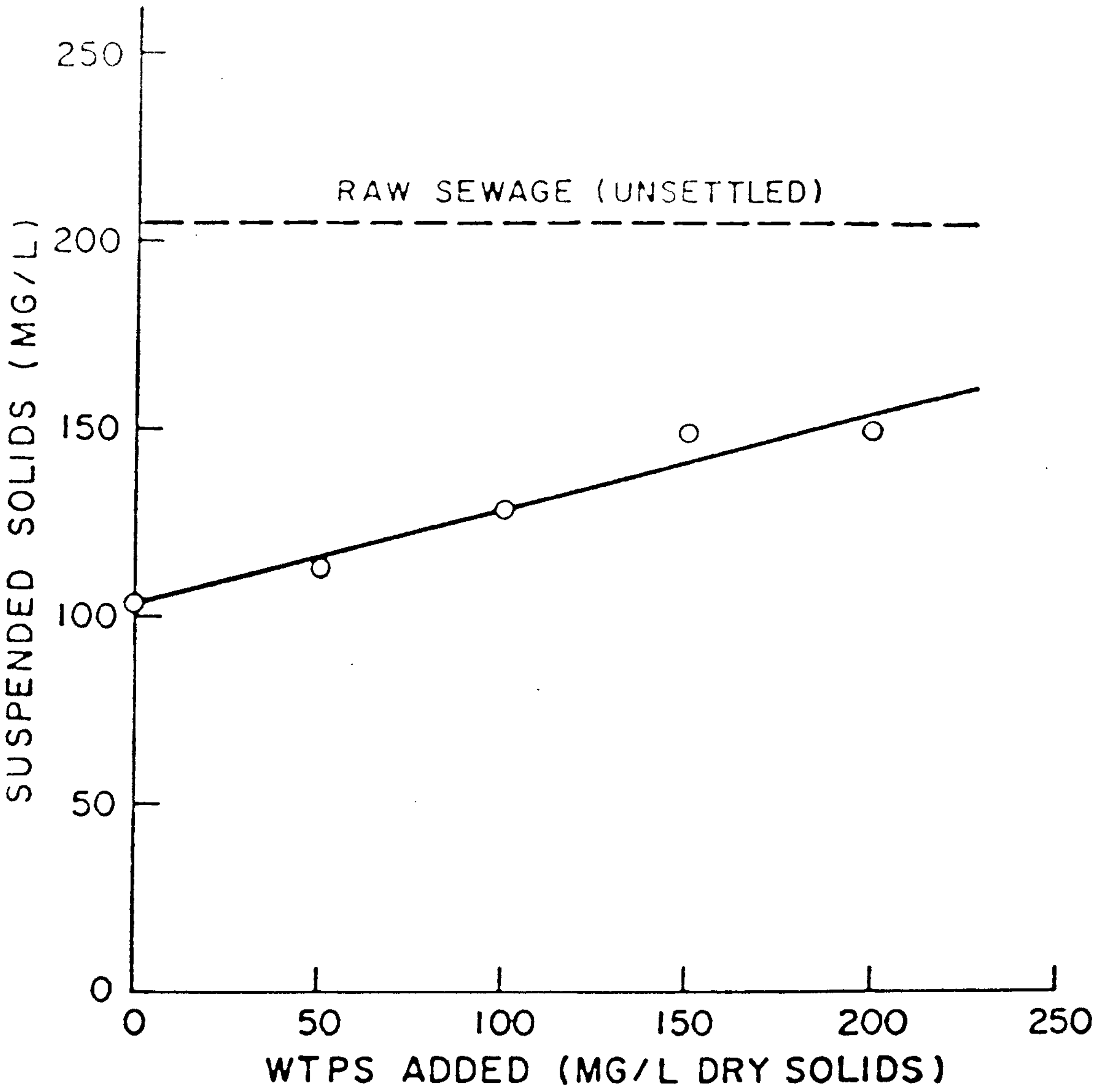


FIGURE 2-7

EFFECT OF
WATER TREATMENT PLANT SLUDGE (WTPS)
ON SETTLED WASTEWATER SUSPENDED SOLIDS

Reference: 2-18

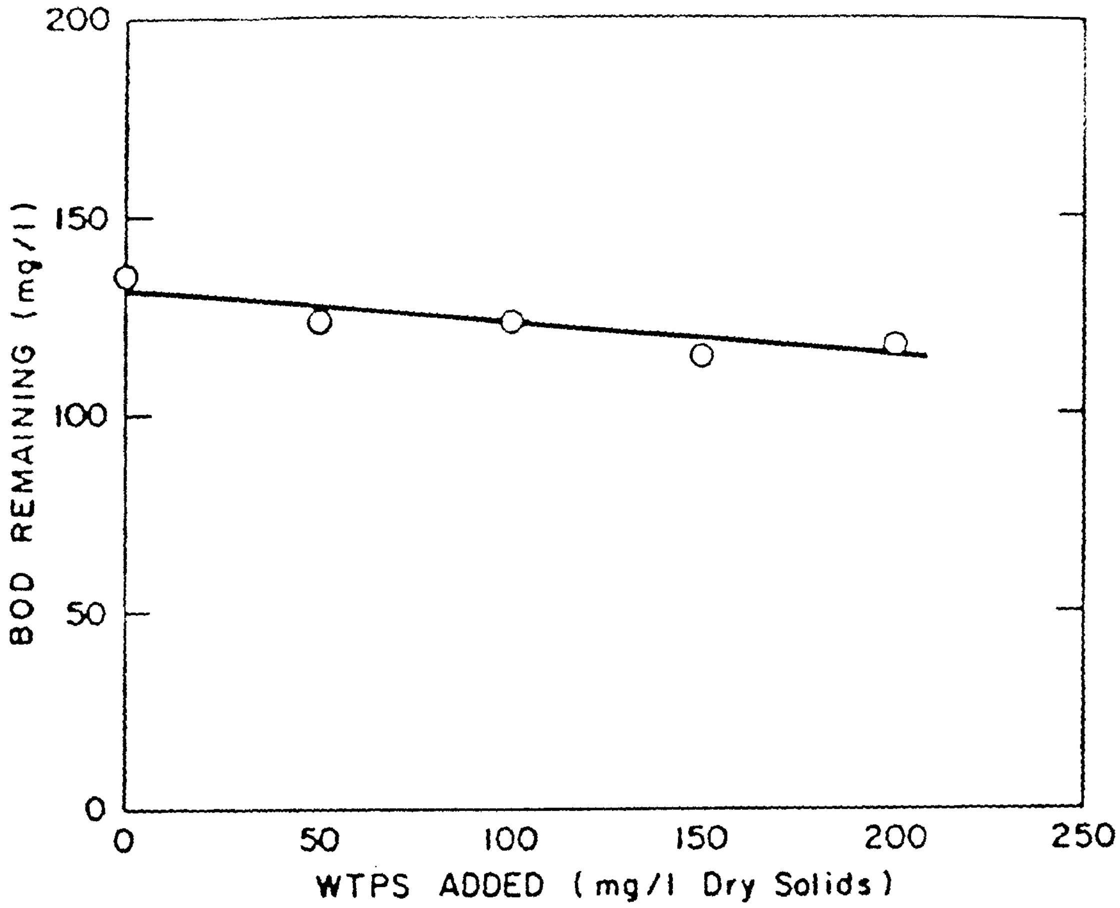


FIGURE 2-8

EFFECT OF
WATER TREATMENT PLANT SLUDGE (WTPS)
ON SETTLED WASTEWATER BOD

Reference: 2-18

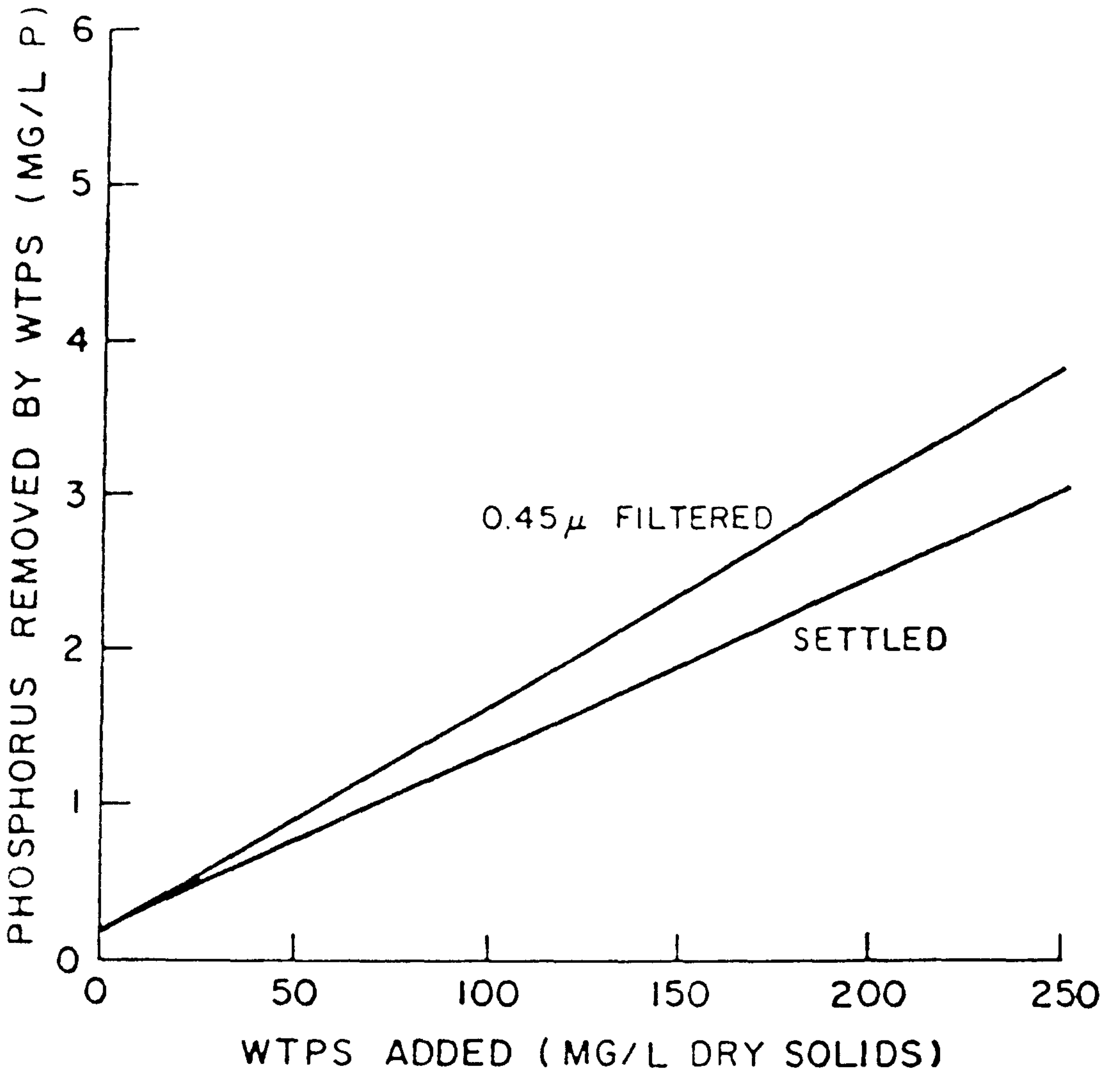


FIGURE 2-9

PHOSPHORUS REMOVED VERSUS
WATER TREATMENT PLANT SLUDGE (WTPS)
ADDED

Reference: 2-18

tions, these tests indicated that at alum doses required to achieve satisfactory phosphorus removal (i.e. Al:P molar ratio >1), very little benefit could be attributed to the water treatment plant sludge.

In later work by Rolan (2-19) full scale testing was conducted on the effects of alum sludge on the above process parameters. Alum sludge was added on a daily basis for about 60 days to a 7 mgd wastewater flow. The alum sludge was measured at a dose which increased the raw wastewater suspended solids concentration by 100 mg/l. No effects on plant finished BOD was found. However, final effluent suspended solids values increased from 20 mg/l to 40 mg/l. After stopping alum sludge discharge, suspended solids values returned immediately to the 20 mg/l range. Final effluent color also increased from about 45 to 50 Pt-Co units to 70-90 Pt-Co units. On the other hand, total phosphorus in the effluent decreased from 7 mg/l without alum sludge addition to 4 mg/l with the addition.

Salotto et al. (2-20) reported the results of a 200 gpd pilot plant investigation into the effects of water plant sludge on the activated sludge process. The water plant sludge was from a coagulation/softening process. The sludge contained 27% dry weight calcium, or about 67% $CaCO_3$ and only 0.75% Al, or less than 4% aluminum hydroxide species. Therefore, the sludge could be considered a lime softening sludge for comparison purposes with the results of Rolan. The activated sludge process was continually dosed with 200 mg/l dry weight solids of lime sludge. Although the pH of the sludge was 10, the raw wastewater pH was not significantly affected. The pH is one important parameter that would need to be closely monitored when discharging lime sludge to the sewer. No change in COD removal from influent to final effluent was found. Overall, the suspended solids removed was slightly better with the lime sludge addition, although the primary effluent suspended solids was slightly worse. The researchers explained this based on a carry-over of aluminum and magnesium hydroxide from the primary clarifier which eventually aided settling in the secondary clarifier. No change in phosphorus removal was found with the addition of 200 mg/l lime sludge.

Overall, it can be concluded that if the dosing of water utility sludge is equalized so that surges do not occur, and the dose is kept below 150 to 200 mg/l, no direct effect on the activated sludge process is likely to take place. If primary clarifiers are present, most of the water plant solids will be removed at that point. No change in overall BOD/COD or suspended solids removal would be expected, but should be monitored. If primary clarifiers are not present, then some adverse impacts may result: the activated sludge process will need to operate with a higher mixed liquor suspended solids concentration to maintain the desired mixed liquor volatile suspended solids concentration; overloading of the secondary clarifiers may occur, resulting in solids carry over.

With alum (and possibly iron) sludge some additional phosphorus removal may occur, depending upon the sludge dose. However, as a phosphorus removal process to achieve or help achieve normal phosphorus effluent goals, the sludge is of little or no benefit.

2.3.3. Sludge Handling Considerations

If a water utility waste that contains suspended solids is transferred to the wastewater plant, then additional sludge handling is necessary at the receiving facility. Considerations must be given to the unit processes of clarification, digestion, final dewatering and disposal.

Rolan conducted sludge volume tests to determine the additional primary clarifier sludge volume that would be expected after the addition of water plant alum sludge to raw wastewater. Figure 2-10 shows the results of the tests. The slope of the line is 1.2, indicating that the volume of the

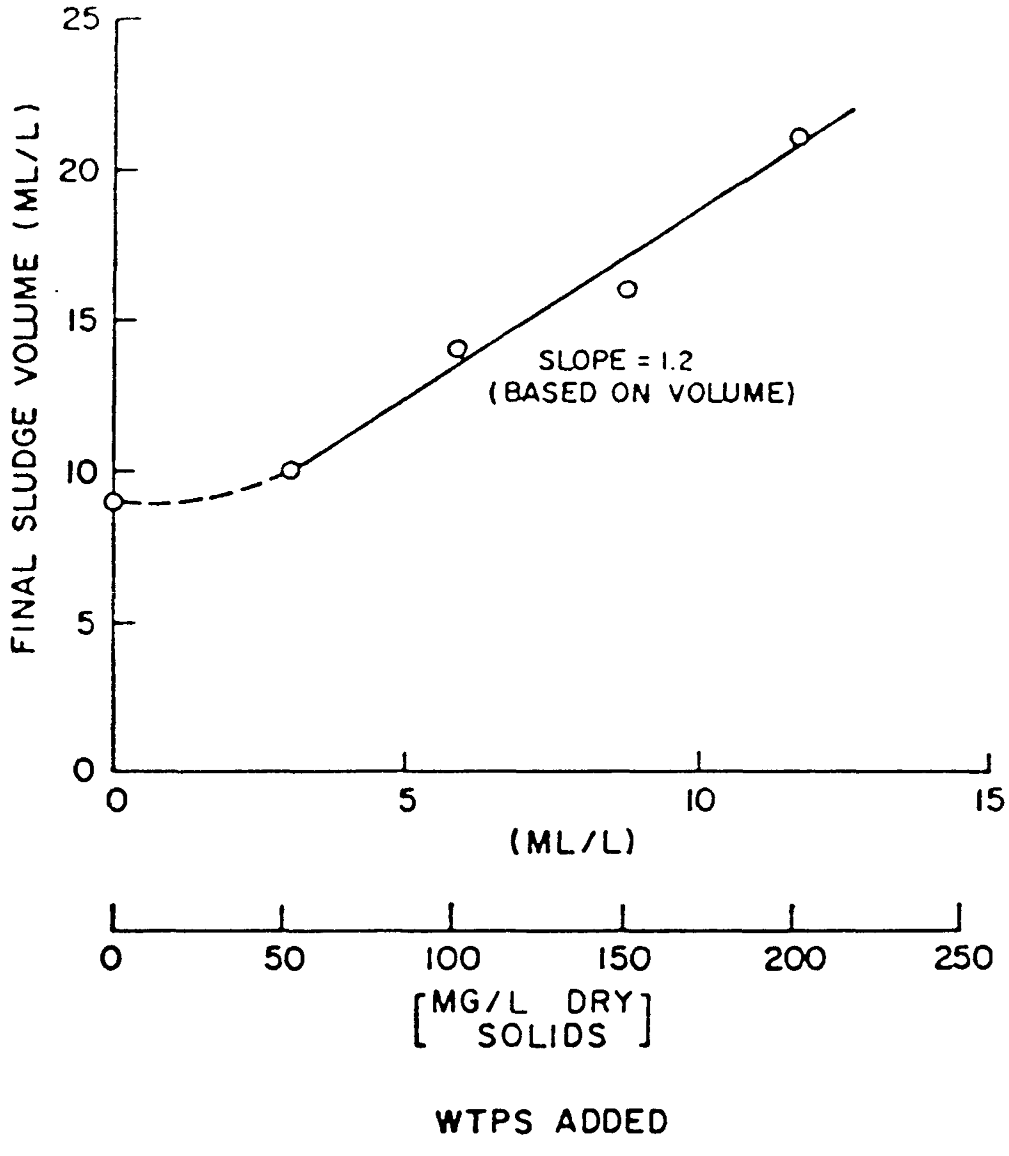


FIGURE 2-10

EFFECT OF
WATER TREATMENT PLANT SLUDGE (WTPS)
ON THE
COMBINED VOLUME OF WASTEWATER SLUDGE
AFTER 30 MINUTES OF SETTLING

Reference: 2-18

settled combined sludges is greater than if they had been settled separately. This increase in combined volume may be due to additional removal of fine solids caused by enmeshment within the water plant sludge. Overall, the addition of 200 mg/l of alum sludge solids approximately doubled the volume of primary clarifier sludge.

Digester sizing and performance must consider the additional volume and dry weight of sludge as well as the decrease in percent volatile matter. In the full scale tests by Rolan, raw sludge volume to the digesters increased by about 50% with the addition of 100 mg/l alum sludge (very close to that predicted in Figure 2-10). Percent solids to the digester increased from about 2.5% to 3.1%, while digested percent solids decreased from 4.5% to 4.2%. The percent volatile matter in the raw sludge decreased from 68% to 55%.

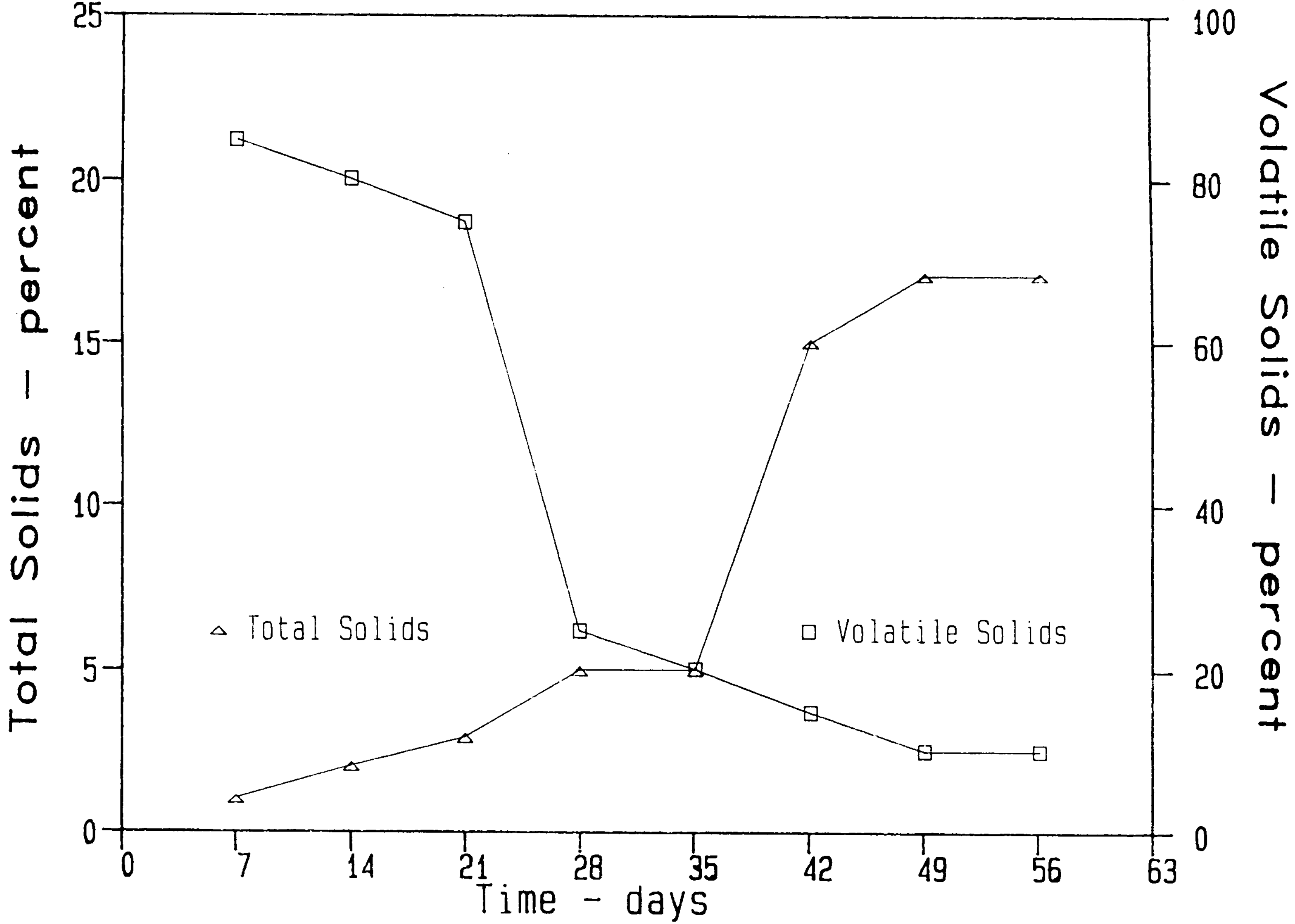
Similar results were obtained by Salotto (2-20) with the addition of lime sludges. While Salotto did not report primary sludge volume production, it may in fact be that lime sludges would decrease the primary sludge volume, since lime sludges settle much better than do alum sludges. However, as Figure 2-11 shows, the total solids concentration increased markedly with a corresponding sharp decrease in the percent volatile solids. The researchers felt that the results may be somewhat skewed because of the relatively high ratio of lime sludge solids to raw wastewater solids. Still they concluded that the digester capacity should be checked for the additional load caused by water plant sludge.

2.4. LANDFILL

For those utilities dewatering their solid/liquid waste streams, landfilling of the resultant solids is the most commonly used method for final disposal. The landfill itself can either be one that also (or primarily) accepts municipal refuse or one that is dedicated to the water plant waste. In the case where the sludge is disposed of in a municipal

FIGURE 2-11

EFFECT OF WATER-UTILITY SLUDGE ON PRIMARY SLUDGE



landfill, the utility often has little to do except determine the requirements for using the landfill. As discussed in Section 2.1.2, in some cases the landfill must be approved to accept industrial wastes in order to dispose of water plant sludge at the site. The principal constraint on using the landfill is usually the allowable solids concentration. While some landfill owners or state regulations will set a specific solids concentration, usually the requirements are more qualitative. The requirements may be stated as no free water, or must behave as a solid or must be handleable by earth moving equipment (also see Section 2.1.2.).

Some insights into the required solids concentration for a sludge to be handleable were provided in work by Calkins and Novak (2-21) as shown in Figure 2-12. They estimated a relationship between the solids concentration to which a sludge would gravity settle and the concentration at which the sludge becomes handleable. Coagulant sludges may only gravity thicken to a 3 to 4% solids concentration and therefore may be handleable at a 20% to 25% solids concentration. On the other hand, lime sludges may gravity thicken to a 40% solids concentration but not be handleable until a 60 to 70% solids concentration is achieved. Although some landfills have accepted a 12 to 15% alum sludge, often a 20% solids cake is a goal. Transportation and specific volume constraints, etc. may necessitate a higher concentration.

Caution should be taken in the landfilling of coagulant sludges because of the possible leaching of aluminum and other metals from the sludge. Municipal solid waste landfills are anaerobic, may produce volatile acids and hence have a pH in the vicinity of 5 to 5.5. This pH will allow for some dissolution of aluminum and other metals from the sludge. Landfills that are therefore equipped with liners and leachate collection systems are desireable. Also, some of the concerns associated with landfilling water plant wastes in a landfill owned by someone other than the governing jurisdiction of the

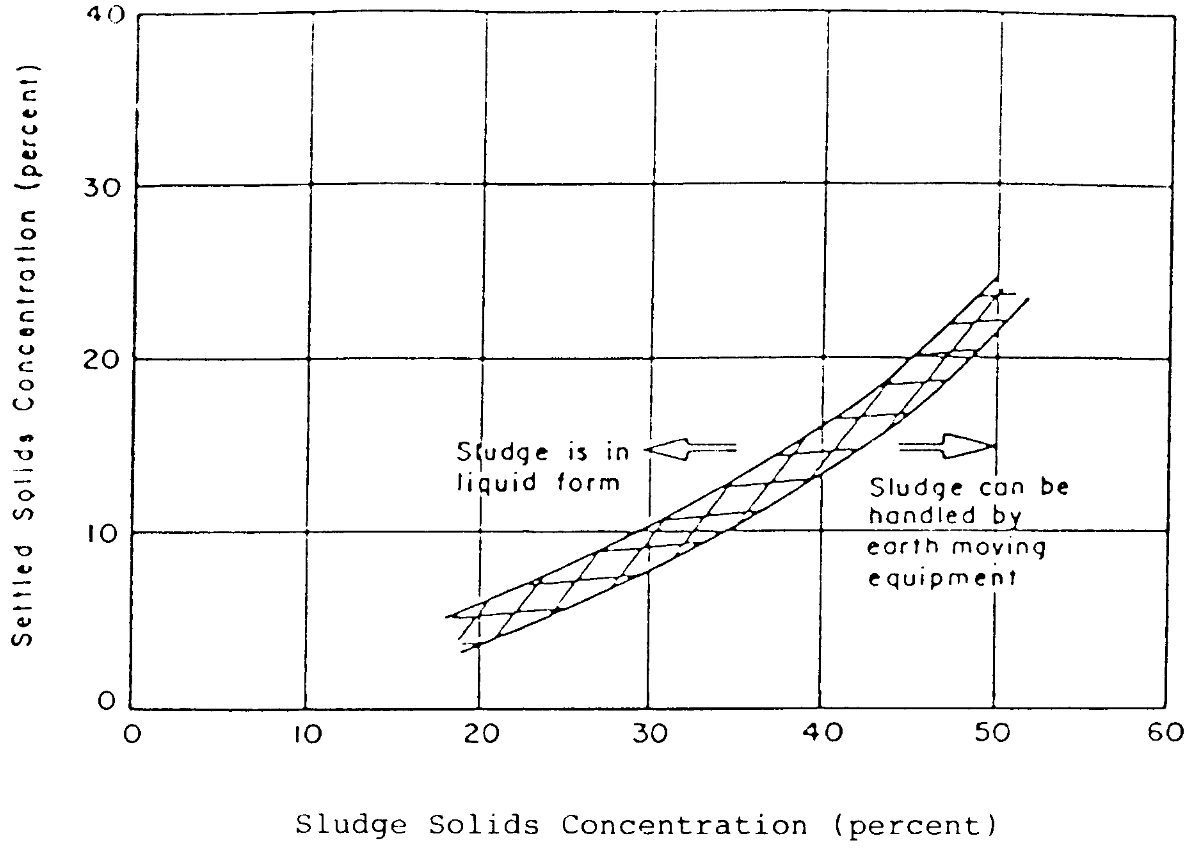


FIGURE 2-12

SOLIDS CONCENTRATION
AT WHICH A SLUDGE
BECOMES HANDLEABLE

Reference: 2-21

water plant have already been expressed in the section on regulations. Because of the possible legal implications of current regulations, it is not advisable to utilize one of these landfills.

Section 2.1.2. discussed example regulations regarding the safe disposal of sludges containing radium. As discussed in Section 3.4. on the chemical characteristics of radium, radium itself in the sludge can emit gamma rays when landfilled. However, the major concern is for the control of release of radon gas. Design of a landfill for radium sludges should at least keep radon emanation below 2 pCi/m^2-sec -- the standard set for uranium mill tailings. Radon release is a function of radium concentration, depth of sludge applied and depth of cover material. Another important factor is the moisture content of the sludge and cover material. Water has the effect of inhibiting radon flux. Detailed design considerations can be found in the Nuclear Regulatory Commission (NRC) report on uranium mill tailings (NUREG/CR-2340 and 1081). The following equation estimates the radon flux from a landfill:

$$F = RpE(LD)^{0.5}\tanh((L/D)^{0.5}d)10^4$$

where,

F = radon flux, pCi/m$^2\cdot$sec

R = concentration of Ra^{226} in the sludge, pCi/g dry weight

p = composite density of sludge, g/cm^3

E = emanation coefficient = 0.2

L = decay constant of radium = 2.1 x 10^{-6} sec^{-1}

D = bulk diffusion coefficient of radon, cm^2/sec

D = 0.106 exp(-.261m)

m = % moisture content

d = sludge depth, cm

$$\text{recall } \tanh x = \frac{e^{x} - e^{-x}}{e^{x} + e^{-x}}$$

Take for example a lime sludge at 40% moisture content containing 20 pCi/g dry weight. The density of lime sludge in this example is 1.3 g/cm^3. The radon flux can be found as the following for a 6 foot deep (183 cm) sludge depth:

$$D = 0.106 \exp(-.261(40))$$

$$D = 3.1 \times 10^{-6}$$

$$F = 20(1.3)(0.2)[(2.1 \times 10^{-6}(3.1 \times 10^{-6})]^{0.5} \tanh((2.1 \times 10^{-6}/3.1 \times 10^{-6})^{0.5}183)10^{4}$$

$$F = 5.2(2.55 \times 10^{-6})(0.59)(10^{4})$$

$$F = 0.08 \text{ pCi/m}^2\cdot\text{sec}$$

The radium concentration in most lime sludges as produced (see Section 3.4 for typical radium concentrations) would be low enough that the radon emanation would be below 2 pCi/m^2·sec. In cases where the radium is concentrated to higher levels, and as the sludge drys the radon release may exceed acceptable levels. In this case the waste may need to be disposed of in shallower depths or a sufficient cover material added to reduce the release of radon gas.

An extreme example for a radium containing lime sludge would be the disposal of a completely dried sludge (0% moisture) at an infinite depth. The following equation applies to an infinitely deep sludge. Considering a sludge containing 22 pCi/g Ra^{226}:

$$F = 10^{4}RpE(LD)^{0.5}$$

$$F = 10^{4}(22)(1.3)(0.2)(2.1 \times 10^{-6}(1.1 \times 10^{-1})^{0.5}$$

$$F = 27.5 \text{ pCi/m}^2\cdot\text{sec}$$

Therefore, as lime sludge dries more completely, the radon release could exceed a guideline of 2 $\mathrm{pCi/m^2 \cdot sec}$.

The amount of soil cover required to reduce the release to less than 2 $\mathrm{pCi/m^2 \cdot sec}$ can be calculated from the following equation (this equation has been simplified by assuming the porosity of the sludge and cover material are the same):

$$d_c = \left(\frac{D_c}{L}\right)^{.05} \left(\ln\left(\frac{2F}{F_d}\right) - \ln\left(\left(1+\left(\frac{D}{D_c}\right)^{.05}\right) + \left(1-\left(\frac{D}{D_c}\right)^{.05}\right)\left(\frac{F_d}{F}\right)^2\right)\right)$$

where,

D_c = bulk diffusion coefficient of radon release through cover material = 2.1×10^{-3} $\mathrm{cm^2/sec}$ for clay at 15% moisture based on the above equation for D.

F_d = desired radon release flux

F = radon release flux from sludge

d_c = depth of cover required, cm

therefore,

$$d_c = \left(\frac{2.1 \times 10^{-3}}{2.1 \times 10^{-6}}\right)^{0.5} \left(\ln\left(\frac{2(27.5)}{2}\right) - \ln\left(\left(1+\left(\frac{6.8 \times 10^{-2}}{2.1 \times 10^{-3}}\right)^{0.5}\right.\right.\right.$$

$$\left.\left.\left. + \left(1-\left(\frac{6.8 \times 10^{-2}}{2.1 \times 10^{-3}}\right)^{0.5}\right)\left(\frac{2}{27.5}\right)^2\right)\right)$$

$$\begin{aligned} d_c &= 31.6((3.31) - \ln(2.0 - 5.29 \times 10^{-3})) \\ &= 31.6(3.31 - 0.69) \\ d_c &= 82.8 \text{ cm} \end{aligned}$$

These equations can then be used for different radium concentrations, depths of application and type of cover material to determine the proper landfill methods.

Creating a dedicated landfill -- one that receives only the water plant sludge -- is also a widely practiced alternative. In this case, however, the utility must design and operate the landfill. Figure 2-13 shows some of the key features of a properly designed water plant waste landfill (2-2). Three downstream and one upstream monitoring wells should be installed. The wells should be screened to sample the perched groundwater table. Figure 2-14 shows construction of a typical monitoring well. PVC pipe is often used as it will perform as well as stainless steel in this situation, but is substantially cheaper. Some vinyl chloride and methylene chloride may leach from the PVC and show up in the water analysis in trace amounts, but as long as this is recognized appropriate data analysis can be made.

In order to establish a baseline of data, it is desirable to sample monthly for some period of time, such as 6 months prior to using the landfill and 3 months after its start up. Beyond this period twice per year or annual sampling is sufficient unless a problem is found. The constituents to be analyzed and frequency is also subject to good judgement. It may be reasonable to identify several species that can be routinely monitored (such as Aluminum, TDS, Sulfate, pH) and more detailed analysis on a less frequent basis. As an example of the parameters evaluated with sludge disposal, the following were required to be monitored in connection with alum sludge landfilling in Virginia: alkalinity, NH_3, As, Ba, Cd, Ca, Cl, Cr, Conductivity, Fecal Coliform, F, Hardness, Fe, Pb, Mg, Mn, Hg, NO_3, NO_2, O-P, TP, K, Se, Ag, Na, TKN, TOC, TSS, TS, TVSS, TVS.

If the groundwater table is likely to be above the bottom of the landfill, then the groundwater should be artificially lowered through the use of ditching and drains (with or

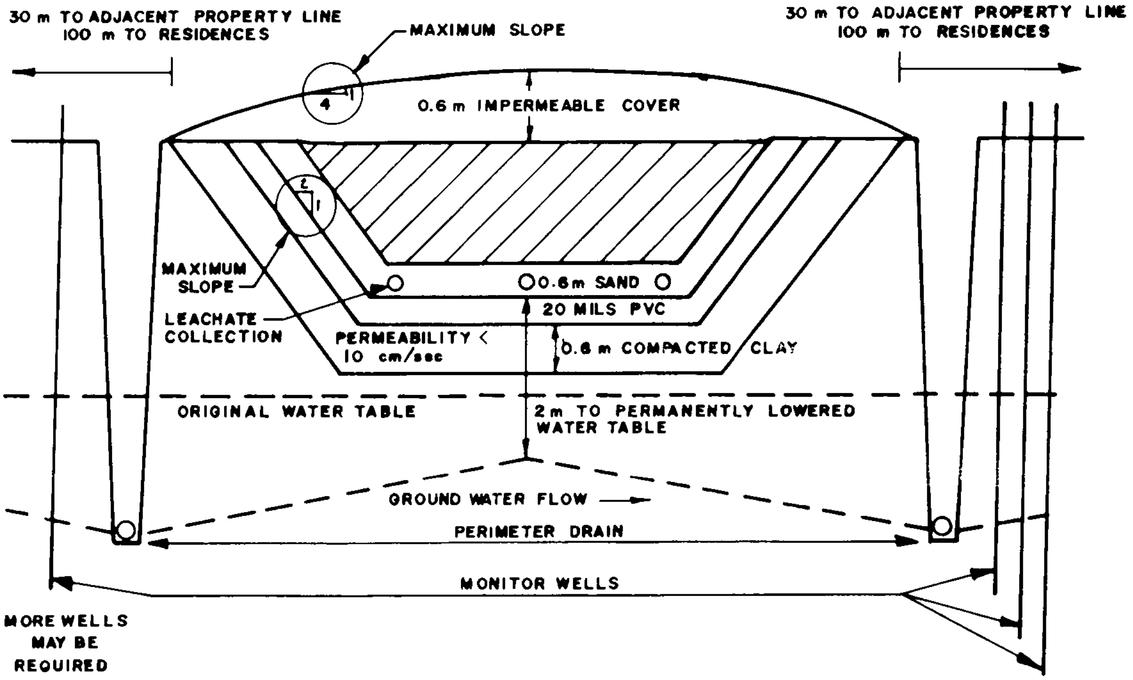


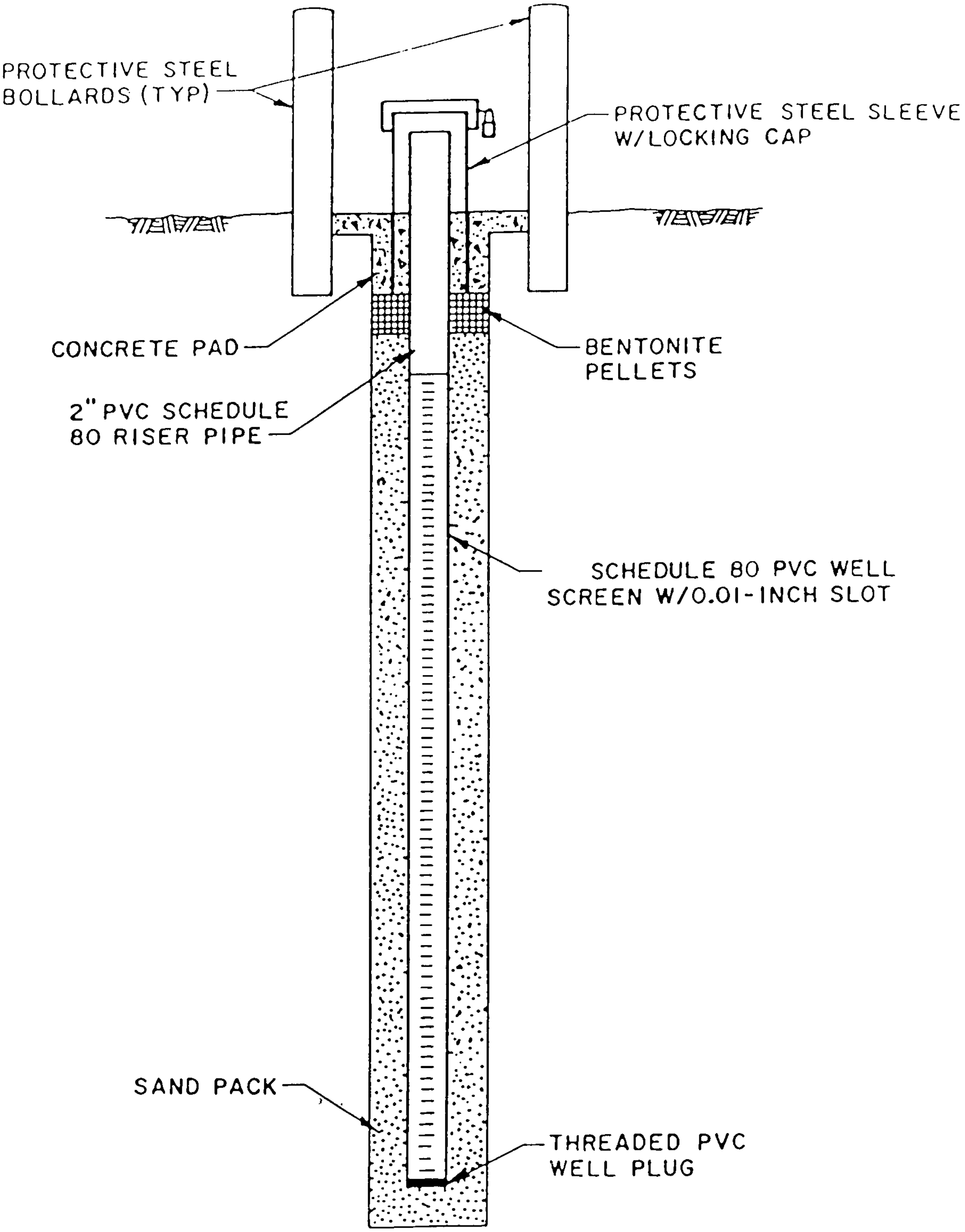
FIGURE 2-13

DESIGN CONSIDERATIONS IN CONSTRUCTING A WATER PLANT WASTE LANDFILL

Reference: 2-2

FIGURE 2-14

TYPICAL WELL CONSTRUCTION



NOT TO SCALE

without pumping) or well pointing. The bottom of the landfill should be lined and a leachate collection system utilized.

The trench method is the easy way to operate the landfill, such as is shown in Figure 2-15 (ref. 2-2). This method is convenient because the trucks can dump from the hard ground above the trench. Often the trench is only two to three truckloads wide (about 20 feet) so that dumping from the top without regular movement by bulldozing equipment is used. Unless the sludges exhibit high odors (usually not the case for dewatered water plant sludges) then daily covering is not required.

2.5. LAND APPLICATION

Land application is practiced to a limited extent for alum sludge disposal, more widely for lime sludge disposal and can be used for disposal by wastes from diatamaceous earth precoat plants. Alum sludges have little value as a soil conditioner, but there is some concern that is they will do some harm. The South Central Connecticut Regional Water Authority in conjunction with the Connecticut Agriculture Experiment Station (2-22) conducted studies on the effects of alum sludge applications. Table 2-10 shows the nutrient analysis conducted on the two alum sludges used in the study. As is typical, the alum sludges had essentially no nutrient value.

In preliminary greenhouse experiments, the researchers germinated ryegrass on mixtures of soil and alum sludge containing 0, 25, 50, 75 and 100% dried alum sludge. Although the sludge did not hinder seed germination, some inhibition of growth was noticed. The researchers then conducted several experiments with two objectives: 1) to substitute dried alum sludge for various constituents in potting soil mixtures and to measure their ability to support plant growth and, 2) to spray liquid alum sludge on forest plots and to measure the effects on the soil, litter decomposition and tree growth.

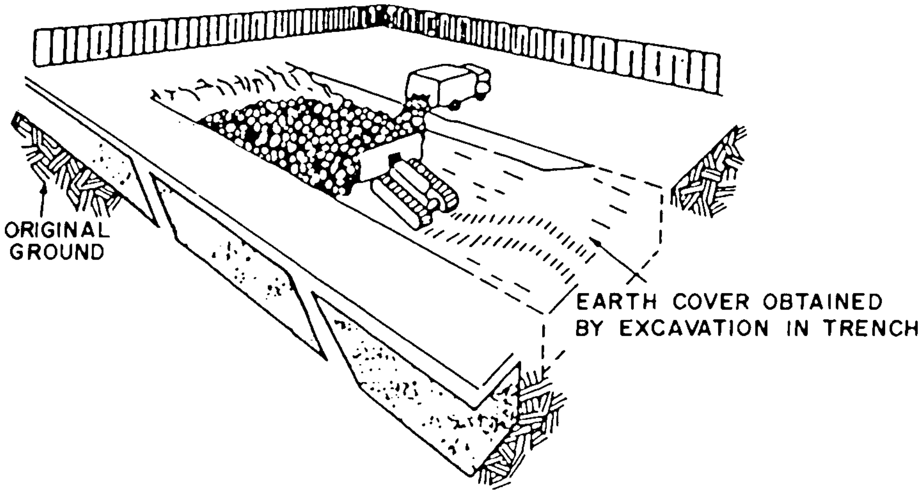


FIGURE 2-15

TRENCH METHOD FOR LANDFILLING
WATER PLANT WASTE

Reference: 2-2

TABLE 2-10

NUTRIENT CONTENT OF ALUM SLUDGE

Parameter	Saltonstall Conn mg/l	West River Conn mg/l
Copper	<0.02	0.17
Zinc	0.05	0.04
Iron	<0.02	0.34
Manganese	1.97	16.47
Nitrate N	0.08	0.05
Nitrite N	0.003	0.003
Total Phosphorus	0.13	0.13

Source: Bugbee and Frink (2-22)

To determine if dried alum sludge would be a suitable potting media, several combinations of alum sludge, soil, perlite and peat moss were prepared as shown in Table 2-11. Treatment 1 contained no alum. In treatments 2, 3 and 4 alum sludge replaced either soil, peat or perlite. In treatments 5, 6 and 7 alum sludge was added in different proportions to equal amounts of soil, perlite and peat. Lettuce seedlings were added to the soil mixtures with conventional amounts of fertilizers added. Pronounced differences in plant growth were noticed. Lettuce growing in media containing alum sludge took on a purple hue, usually associated with a phosphorus deficiency. The testing shown in Table 2-11 confirmed that phosphorus was less available to plants containing alum sludge. Deficiencies in plant-available phosphorus were probably due to phosphorus fixation by aluminum. Further experiments were conducted to determine the extent of the phosphorus deficiencies. In these tests marigolds were grown in soils containing increasing amounts of alum sludge. Conventional amounts of fertilizer were added, except that twice the normal amount of phosphorus was used. Figure 2-16 illustrates how plant growth and available phosphorus responded to various volume percentages of alum sludge. Significant declines in growth occurred at all levels of alum sludge addition. Available phosphorus declined correspondingly. The researchers concluded that:

1. Alum sludge can fix the available phosphorus in the soil thus causing a phosphorus deficiency in the plants.
2. Doubling the normal amount of phosphorus fertilizer addition did not overcome the phosphorus deficiency problem.

The second objective was to test the effects of liquid alum sludge applied to deciduous and coniferous forested

TABLE 2-11

PHYSICAL PROPERTIES AND PLANT GROWTH IN POTTING MEDIA AMENDED WITH ALUM SLUDGE

	Treatments %/V.				Physical Properties							
					%/V.				g/cc	g/plant	mg/l*	%
#	Soil	Perlite	Peat	Alum	Air at c.c.	Avail. H20	Unavail. H20	Total Pore S.	Bulk Density	Lettuce Growth	Soil P	Tissue P
1	33	33	33	0	7.7	45.6	16.9	70.2	0.52	5.9	95	.52
2	33	33	0	33	8.9	34.5	22.4	65.8	0.63	4.6	42	.14
3	33	0	33	33	6.6	38.7	24.1	69.4	0.60	4.5	39	.08
4	0	33	33	33	16.8	50.2	14.4	81.4	0.27	5.8	26	.09
5	25	25	25	25	6.8	45.3	20.0	72.1	0.54	3.5	34	.11
6	17	17	17	50	7.2	45.8	22.6	75.6	0.48	2.9	30	.08
7	11	11	11	67	13.9	40.3	23.7	77.9	0.43	2.0	33	.09

* Mean of six tests run throughout growth period

Source: Bugbee and Frink (2-22)

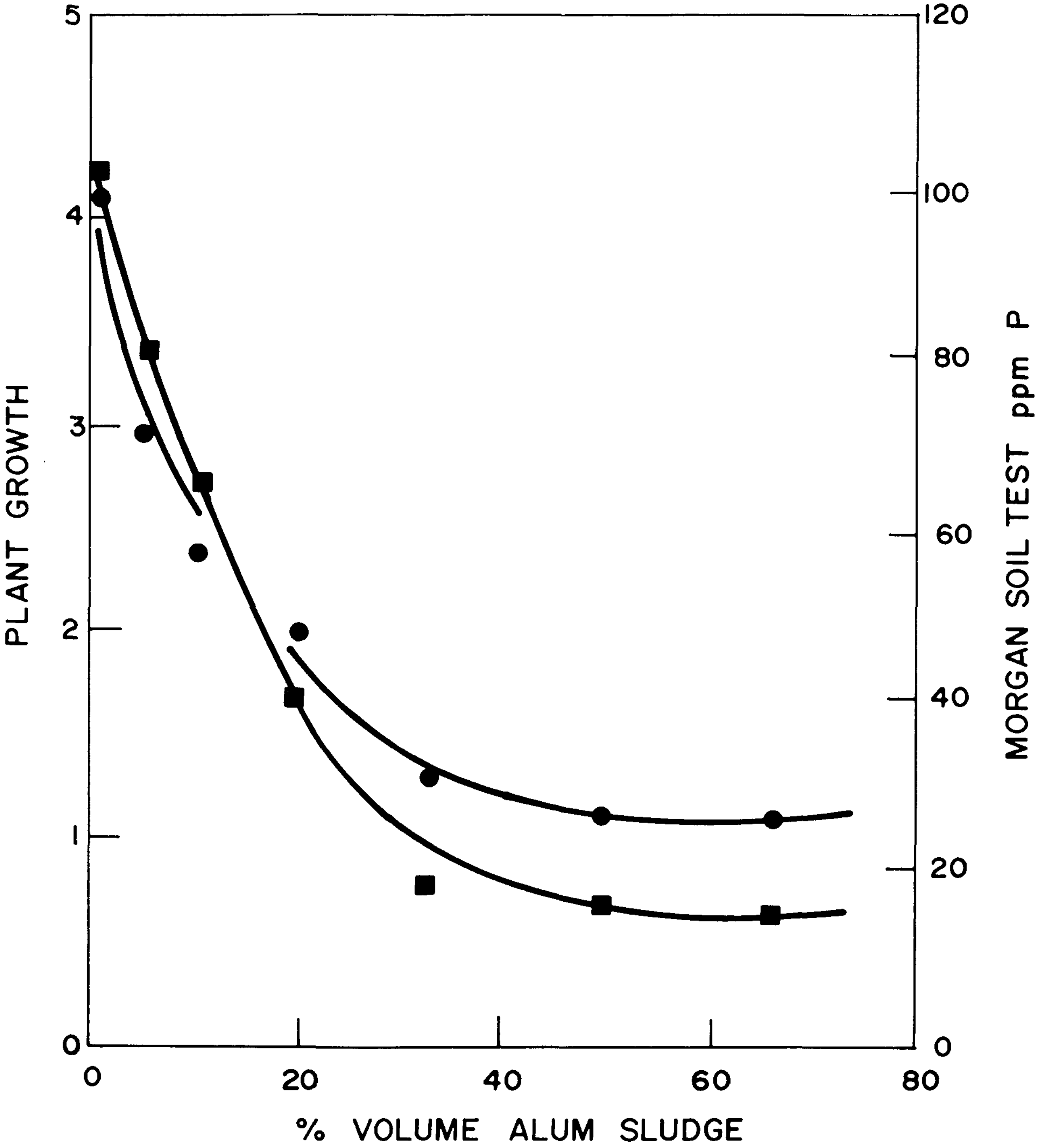


FIGURE 2-16

GROWTH OF MARIGOLDS (■) AND AVAILABLE SOIL PHOSPHOROUS (●) IN POTTING MEDIA AMENDED WITH ALUM SLUDGE

Reference: 2-22

lands. The deciduous forest used was predominantly sugar maple and the coniferous forest was Eastern hemlocks. Two 15 meter by 11 meter plots in forest were treated with 124,800 gal/acre of liquid alum sludge at a 1.5% suspended solids concentration (7.8 tons dry alum sludge/acre). Half the sludge was applied in the fall and half in the spring. About one year after the first application, the soil and trees were analyzed. Changes of availability of plant nutrients in the soil is shown in Table 2-12. There were no significant differences in nutrient availability or measured tree growth. Uptake of phosphorus appeared to be slightly less in the coniferous plots where alum was applied. Measurements would need to continue many years before conclusions regarding long term effects could be drawn. Note that the equivalent alum sludge loading to the forest lands was only about one-tenth that used in the previous potting experiments and fixation of phosphorus may not have occurred or may not have been measurable at the lower application.

Another consideration in the land application of water plant wastes could be the regulations regarding the limiting metal concentration. Although these regulations strictly apply to wastewater plant sludges, the limiting metal values deserve attention. Table 2-4 showed the guidelines for metal application rates to soils and Table 3-7 shows the metal content for one particular alum sludge. The allowable sludge loading to not exceed the metal loading can be found by

$$\text{Tons Sludge/Acre} = \frac{\text{lb/acre metal allowed}}{(\text{mg/kg in sludge})(0.002)}$$

Therefore, the following calculations would result for the particular sludge described in Table 3-7:

TABLE 2-12

ANALYSES OF FOREST SOLIDS ONE YEAR AFTER APPLICATION OF ALUM SLUDGE

		Concentration in Soil Extract (mg/kg)							
Plot	pH	NO_3^-	NH_4^+	P	K	Ca	Mg	Al	Mn
Deciduous Control									
Litter[a]	5.3	6 M	12 L[b]	12 L	60 L	500 L	75 H	10 L[b]	20 MH
0-5 cm	4.5	6 M	12 L	6 VL	30 VL	250 VL	15 L	125 H	5 L
5-10 cm	4.7	6 M	12 L	6 VL	30 VL	250 VL	12 L	125 H	5 L
Deciduous Alum									
Litter	6.2	5 M	12 L	12 L	60 L	500 L	50 MH	25 M	9 M
0-5 cm	4.8	6 M	12 L	6 VL	30 VL	250 VL	12 L	125 H	5 L
5-10 cm	5.0	6 M	12 L	6 VL	30 VL	250 VL	12 L	125 H	5 L
Coniferous Control									
Litter	4.6	3 L	12 L	9 L	45 L	425 L	35 M	20 M	10 M
0-5 cm	4.7	3 L	12 L	6 VL	30 VL	250 VL	20 ML	125 H	10 M
5-10 cm	4.8	3 L	12 L	6 VL	30 VL	250 VL	12 L	125 H	15 M
Coniferous Alum									
Litter	5.3	3 L	12 L	6 VL	30 VL	500 L	50 MH	50 MH	5 L
0-5 cm	5.1	3 L	12 L	9 L	45 L	425 L	25 M	125 H	5 L
5-10 cm	5.0	3 L	12 L	9 L	60 L	250 VL	20 M	125 H	5 L

VL = very low; L = low; ML = medium low; M = medium; MH = medium high; H = high; VH = very high

a - sample depth

b - low levels of these elements are optimum

Source: Bugbee and Frink (2-22)

Metal	Metal Limit lb/acre	Metal Content of sludge mg/kg	Total Sludge allowed tons/acre
Pb	500	1.2	200,000
Zn	250	68.	1,800
Cu	125	24.	2,600
Ni	50	2.4	10,000
Cd	5	0	-

It can be seen that for this particular alum sludge, zinc is the limiting metal, with an allowed sludge application of 1,800 tons/acre. Metal loadings would seldom be the limiting criteria for alum sludges at loading rates low enough to prevent phosphorus fixation.

In many farming regions, the application of nitrogen fertilizers causes a reduction in soil pH. Farmers normally apply sufficient quantities of lime in order to obtain the desired soil pH. In Ohio, for example, the lime requirement is about 3.1 tons/acre. Lime sludges from water plants are as effective as quarry limestone in neutralizing soils. The Ohio Department of Health (2-23) reported that liming materials typically available to farmers have a total neutralizing power (TNP) of 60 to 90. The department evaluated sludges from 7 utilities and found the TNP of lime sludges to range between 92 to 100, or better then that of commercially available materials. In Illinois, a calcium carbonate equivalent performed on several sludges indicated that the softening sludges were superior to agricultural limestones available locally. (2-24)

The sludge can be applied to farm lands by either spraying liquid sludge (<15% solids concentration) from a tank truck or by spreading and tilling dewatered lime sludge from a hopper-bed truck with a spinner device for spreading the sludge.

Transportation costs and farmer acceptance appear to be major drawbacks to more widespread use of land application of lime sludges. Consideration must also be given to the problem that lime sludge is produced continuously at the water plant, but is needed only seasonally by the farmer.

Special considerations apply to lime sludges containing radium. Illinois, for example, requires that the sludge be mixed with the soil so as to not increase the radium by more than 0.1 pCi/g. The maximum allowable application rate can be found by the following:

$$AR = 1{,}390\ Sd\ \frac{Ra}{R}$$

where,

AR = maximum allowed lime sludge application, tons/acre, dry weight

S = specific gravity of soil, for example = 2

d = depth of sludge/soil mixing, feet

Ra = allowed radium increases, pCi/g

R = radium in sludge, pCi/g dry weight

This is illustrated for the following example for a lime sludge containing 22 pCi/g radium, mixed with 6 inches of soil:

$$AR = 1{,}390(2)(0.5)\left(\frac{0.1}{22}\right)$$

$$AR = 6.3 \text{ tons/acre}$$

which exceeds the approximate desired loading for effective liming of 3 tons/acre. Therefore, this waste could be land applied without increasing the background radium by over 0.1 pCi/g.

CHAPTER 3

CHARACTERISTICS OF WATER TREATMENT PLANT WASTES

It has been said that there is no such thing as a typical water treatment plant. There is likewise no such thing as a typical water plant waste. A detailed analysis of waste characteristics should be conducted for each site. However, there are common characteristics among particular types of water plant wastes, and there are certainly common test methods that can be used to evaluate wastes.

In beginning to characterize water plant wastes there are four important areas that need to be addressed:

1. type of wastes generated
2. quantity of waste generated
3. classification by physical properties and dewatering characteristics
4. specific constituents in the waste streams, particularly as they may relate to proper disposal.

In this chapter sufficient guidance has been provided to enable the utility or design engineer to characterize their specific situation. Section 3.2 presents methods to estimate waste quantities (for solid/liquid wastes, liquid phase wastes and gas phase wastes) and suspended or dissolved components as may be appropriate. Typical values obtained from a cross-section of plants as were available in the literature or produced from direct contact have been included. The physical characteristics section is primarily directed at test methods to evaluate conditioning and dewaterability of solid/liquid phase waste streams. The chemical characteristics section presents example chemical analyses of solid/liquid wastes and liquid phase wastes. As appropriate, methods to estimate the chemical composition are presented.

3.1. TYPES OF WASTES GENERATED

Hydrolyzing metal salts or synthetic organic polymers are added in the water treatment process to coagulate suspended and dissolved contaminants and yield relatively clean water suitable for filtration. Most of these coagulants and the impurities they remove settle to the bottom of the settling basin where they become part of the sludge. These sludges are referred to as alum, iron or polymeric sludge according to which primary coagulant is used. These wastes account for approximately 70 percent of the water plant waste generated. The sludges produced in treatment plants where water softening is practiced using lime or lime and soda ash account for an additional 25 percent of the industry's waste production. It is therefore apparent that most of the waste generation involves water treatment plants using coagulation or softening processes. The above wastes are referred to as solid/liquid wastes in that the liquid waste (water) contains suspended solids. Other solid/liquid wastes produced in the water industry include all polymer coagulation wastes, wastes from iron or manganese removal plants, spent GAC, spent precoat filter media and wastes from slow sand filter plants.

The water industry also produces liquid phase wastes, referred to as such since the liquid phase (water) contains primarily dissolved solids which are within the liquid phase itself. These wastes include spent brine from ion exchange regeneration, reject water from reverse osmosis systems, reject water from electrodialysis plants and spent regenerant from specific adsorption media such as activated alumina. One type of gas phase waste is produced -- off-gasses from air stripping systems. The major types of treatment plant wastes are shown in Table 3-1 and will be discussed in subsequent sections as appropriate.

TABLE 3-1

MAJOR WATER TREATMENT PLANT WASTES

SOLID/LIQUID WASTES

1. Alum Sludges
2. Iron Sludges
3. Polymeric Sludges
4. Softening Sludges
5. Backwash Wastes
6. Spent GAC or Discharge from Carbon Systems
7. Slow Sand Filter Wastes
8. Wastes from Iron and Manganese Removal Plants
9. Spent Pre-Coat Filter Media

LIQUID PHASE WASTES

10. Ion-Exchange Regenerant Brine
11. Waste Regenerant from Activated Alumina
12. Reverse Osmosis Waste Streams

GAS PHASE WASTES

13. Air Stripping Off-Gasses

3.2. QUANTITY OF WASTES GENERATED

3.2.1. Solids/Liquid Wastes

The quantity of solid/liquid wastes (which are commonly referred to as 'sludge') generated from water treatment plants depends upon the raw water quality, dosage of chemicals used, the performance of the treatment process and the method of sludge removal.

One of the most difficult tasks facing the utility or engineer in planning and designing a sludge treatment process is determining the amount of waste to be handled. The waste quantity is usually determined as an annual average for a given design year and is a function of flow projections. As important as average production values is information on seasonal and monthly variations. It is not unusual for order of magnitude differences in sludge production to exist for different months of the year. Since determining quantities of waste produced requires a history of data compilation it is wise for a utility to begin collecting these data even if there are no immediate plans to implement a new waste management program. Such a data analysis process can even provide insights to options to ease burdens on existing systems and methods to increase efficiency of the water treatment process itself.

There are three methods to determine sludge waste quantities. None are completely accurate and therefore it is advisable to use and cross-check all three. Those methods are: calculations; coagulant mass balance analysis; and, field determination. Each is discussed below, beginning with the calculation procedure.

The amount of alum (or iron) sludge generated can be calculated fairly closely by considering the reactions of alum and/or iron in the coagulation process. Using an empirical relation to account for the sludge contribution from turbidity will improve the estimate and the contribution from other sources can be added as required.

When alum is added to water as aluminum sulfate, the reaction is typically represented by the simplified equation:

$$Al_2(SO_4)_3 \cdot 14H_2O + 6HCO_3^- = 2Al(OH)_3 + 6CO_2 + 14H_2O + 3SO_4^{-2}$$

If inadequate alkalinity is present, lime or sodium hydroxide is normally added to maintain the proper pH. If equilibrium were achieved the aluminum hydroxide, as shown, would be the predominate product. However, equilibrium is not normally obtained and a complex polymerized compound containing on the average 3 to 4 water molecules bound to the aluminum hydroxide is formed as the precipitate. This chemically bound water increases the sludge quantity, increases the sludge volume and also makes it more difficult to dewater since the chemically bound water can not be removed by normal mechanical methods (see Section 3.4.). The resulting aluminum hydroxide species has a molecular weight of 132 and 1 mg/l of alum added to water will produce approximately 0.44 mg/l of inorganic aluminum solids (3-1). Suspended solids present in the raw water produce an equivalent weight of sludge solids since they are non-reactive. It can be assumed that other additives such as polymer and powdered activated carbon produce sludge on a one to one basis. The amount of sludge produced in an alum coagulation plant for the removal of turbidity is then:

$$S = 8.34\ Q\ (0.44Al + SS + A)$$

where,

S = sludge produced (lb/day)
Q = plant flow (mgd)
Al = alum dose as 17.1% Al_2O_3 (mg/l)
SS = raw water suspended solids (mg/l)
A = additional chemicals added such as polymer, clay or activated carbon (mg/l)

If iron is used as the coagulant then the equation becomes

$$S = 8.34\ Q\ (2.9\ Fe + SS + A)$$

where the iron dose is expressed as mg/l of Fe^{3+} added or produced via Fe^{2+} oxidation (note that significant Fe^{2+} in the raw water will also produce sludge at a factor of 2.9 if it is oxidized). The above two equations should not be interpreted that iron produces several times the amount of sludge that alum produces. In the case of aluminum the dose is expressed as alum with a molecular weight of 594 and in the case of iron the dose is expressed as iron with a molecular weight of 55.8. In reality one mole of coagulating equivalent of iron produces about 20 to 25% more dry weight sludge than one mole of aluminum.

The above equations can then be used to track yearly or even daily variation changes in sludge weight produced. The difficulty in applying the relationships is that most plants do not routinely analyze raw water suspended solids concentrations. The logical correlation is to equate a turbidity unit to a suspended solids unit. Unfortunately the relationship is generally not necessarily 1 to 1:

$$SS,\ mg/l = b \cdot Tu$$

The value of b for low color, predominately turbidity removal plants can vary from 0.7 to 2.2. It may vary seasonally for the same raw water supply. A utility can therefore either continually measure suspended solids, or it may be possible to develop its particular correlation between turbidity and suspended solids. Figure 3-1 shows one such correlation, for a low color raw water source (3-1). Ideally, this correlation should be done weekly until information is learned as to seasonal variations in the suspended solids, turbidity

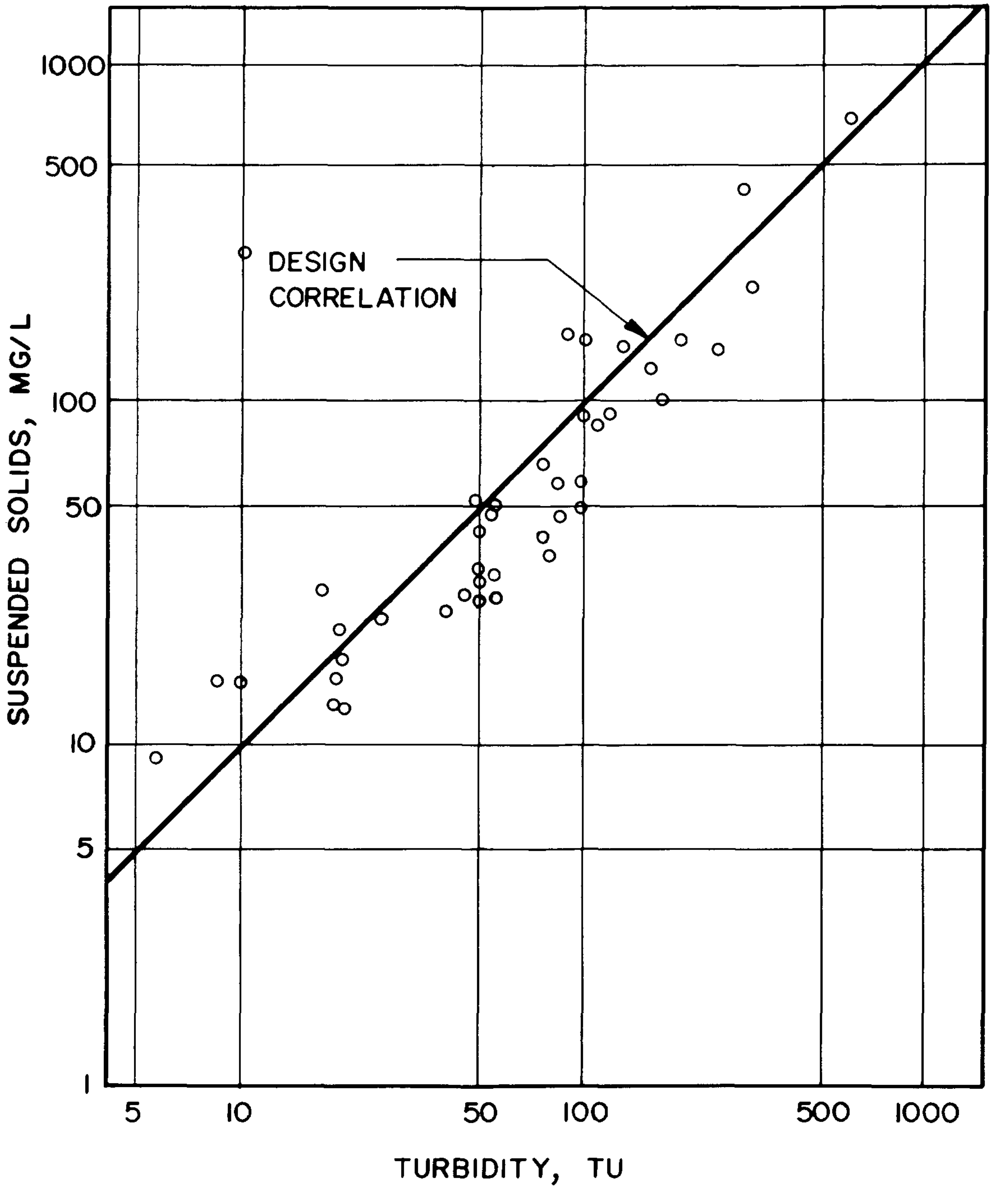


FIGURE 3-1

SUSPENDED SOLIDS VS TURBIDITY

relationship. After that a monthly correlation may be sufficient. Care needs to be used when conducting raw water suspended solids analysis. Since any filterable solids should be removed by coagulation, it is recommended that 0.45 um paper be used for the suspended solids analysis. For many raw waters less then 5 mg of weight may be in 1 liter of water and therefore, several liters may need to be filtered to obtain accurate results. Each individual utility will need to determine its exact procedure to assure themselves of accuracy. As will be discussed below, oven temperature is also important. For raw water suspended solids analysis, the standard temperature of 103°C is generally sufficient.

Another complication exits for raw water sources that contain a significant amount of color. Color can be a large contributor to the sludge production. For many waters, color is filterable on a 0.45 um filter and will therefore be measured in a suspended solids analysis. Values of b for colored raw waters can be as high as 20, but unless turbidity and color vary together, a correlation between SS and Tu will not exist.

The second method which can be used to estimate total sludge weight produced for coagulant sludges is to conduct a conservation of coagulant mass balance analysis. This method is based on conservation of the coagulant. That is, whatever is added in the coagulation process ends up in the sedimentation basin solids, backwash solids or finished water. The first step is to analyze the aluminum (or iron) content of the coagulant. As an approximation it can be assumed that dry weight alum is 9.1% aluminum. Since iron coagulants vary so much, the iron content must be determined or obtained from the manufacturer. A series of sludge, backwash solids and finished water samples are then collected and analyzed for aluminum (or iron). The pH of the solids is lowered to pH 1 and maintained for 10-15 minutes. The sample is then filtered and the filtrate analyzed for aluminum (or iron). This method will tend to solubilize the aluminum hydroxide but not solubilize the aluminum in the clays which may be present. A second aliquot of unacidified sludge is analyzed for suspended solids concentration. When conducting suspended solids tests on hydroxide sludges an oven temperature of 140°C should be used, rather than the normal 103°C. Care should be used to be certain the weight has stabilized. The total amount of sludge can then be computed.

For example, if a 10 mgd plant used 30 mg/l alum, the sludge was analyzed to contain 2,000 mg/l aluminum at 2.5% solids concentration, how much sludge is produced per day? Ignore backwash solids and finished water aluminum.

$$\begin{aligned} \text{Total Al added/day} &= 10 \text{ mgd} \times 8.34 \times 9.1\% \times 30 \text{ mg/l} \\ &= 228 \text{ pounds} \\ &= 1.03 \times 10^{8} \text{ mg} \end{aligned}$$

Since the sludge contains 2,000 mg/l Al there must be 51,756 l of sludge ($1.03 \times 10^{8}/2{,}000$) or 13,800 gpd. At 2.5% solids concentration there would be 2,877 pounds of dry weight sludge produced/day (13,800 x 8.34 x 2.5%). By collecting a series of sludge samples from various locations and depths from within a basin it is possible to formulate an accurate estimate of the amount of sludge produced. A complete mass balance should be done to include Al in the backwash solids and the finished water.

The third method for determining sludge quantities is through field determination. This is probably the most difficult method of all to obtain accurate results, unless the utility already has continuous sludge collection equipment with a monitoring system. If the utility already has a mechanical dewatering system then truck tare weights can be obtained and in conjunction with carefully conducted solids analyses, the amount of sludge produced can be determined. However, usually a utility needs to determine the amount of sludge produced before the installation of facilities and often the plant has manually cleaned sedimentation basins or does not have a way to continually measure sludge flow and dry weight.

In order to conduct an estimate by field determination, all basins should be cleaned. Then a specific period of time should be allowed for sludge to collect in the basins. By then using a "sludge judge" a cross-section of sludge depth can be made in each basin[3]. The sludge judge can also be used to collect a composite depth of sludge from several locations in the basins for suspended solids analyses. With the results of suspended solids analysis and the calculation of sludge volume in the basins a very rough projection of sludge production can be made. However, it is highly recommended that this procedure be supplemented by use of the other two procedures.

Through similar theoretical considerations a general equation has been developed (3-2) for plants that use a softening process with or without the use of alum, iron or polymer. The equation is

$$S = 8.34\, Q\, [2.0\, Ca + 2.6\, Mg + 0.44\, Al + 2.9\, Fe + SS + A]$$

where,

S = sludge production, lb/day
Ca = calcium hardness removed as $CaCO_3$ (mg/l)
Mg = magnesium hardness removed as $CaCO_3$ (mg/l)
Fe = iron dose as Fe (mg/l)
AL = alum dose as 17.1% Al_2O_3 (mg/l)
Q = plant flow, mgd
SS = raw water suspended solids (mg/l)
A = other additives (mg/l)

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[3] A sludge judge is a clear acrylic tube with a foot valve to allow collection of a cross section sample.

The above equations or prediction procedures will allow estimation of the dry weight of sludge produced. They will generally not predict the volume of sludge that will be produced.

Volumes and suspended solids concentrations of sludges leaving the sedimentation basins or clarifiers are a function of raw water quality, treatment and the sludge removal method. Sludges which are allowed to build up in basins, being cleaned only periodically by manual procedures, tend to compact and thicken in the bottom of the basins. There is often a stratification of solids with the heavier particles settling to the bottom and the hydroxide or lighter particles at the top. (This is one reason why it is very difficult to obtain representative sludge samples from within a basin. There will also be a distribution of particle types longitudinally). However, the actual concentration produced will depend upon the amount of water used to flush the solids out of the basin during cleaning. With increasing finished water quality standards there will be a trend to remove the solids as quickly as possible, generally with continuous collection equipment. In this case the solids concentrations will be lower since compaction height and time has been less. Solids concentrations for sludges produced with alum or iron coagulants and for low to moderate turbidity raw waters will be about 0.1 to 1.0%. The higher the coagulant to raw water solids ratio, the lower the solids concentration and the higher the sludge volume. Coagulant sludges from highly turbid raw waters may be in the 2 to 4% range and occasionally higher. Sludge volumes tend to be 0.1 to 3% of the raw water flow, with one survey (3-3) finding an average of 0.6%. Softening sludges will be more concentrated, usually as a function of the $CaCO_3$: $Mg(OH)_2$ ratio and the type of clarifier. Conventional sedimentation basins may only produce solids concentrations of 2 to 4%, whereas sludge blanket clarifiers can produce solids concentrations of up to 15%.

Sludge volumes will correspondingly vary considerably, from 0.5% to 5% of the water plant flow.

Records are usually available on backwash usage in order to estimate this volume of waste. Careful review and analysis is necessary to properly design a facility because of the high flow production over a short period of time. A significant seasonal and daily variation in backwash water usage will also often exist. The data on filter run length, number of backwashes per day, and sequencing of backwashes should be reviewed. Backwashing of two or three filters in a row as compared to backwashing throughout the day can have a major effect on design of equalization and thickening facilities. For example, Figures 3-2 and 3-3 show sizing considerations for a facility expanding from 24 mgd to 50 mgd and adding backwash recycle facilities. The backwash facilities are designed to allow 10% recycle (note that the analysis shown is only for maximum plant capacity and analysis should also be conducted for other flow conditions). Figure 3-2 shows the equalization facility volume requirement for 24 and 50 mgd plant flows and allowing all the filters to be backwashed sequentially. Ultimately, a 1,087,600 gallon tank for storage and equalization is required with this scenario. Alternatively, Figure 3-3 shows that only a 274,000 gallon tank is required if the backwashing is spread throughout the day with no more than two filters backwashed at a time. While this requirement reduces tankage, it may create an unacceptable constraint on the plant operations.

Wastes produced from plants designed solely for iron removal are similar in nature to iron coagulant wastes. Many of these plants consist of aeration for oxidation of the iron followed by a detention time for reaction and then filtration.

The dry weight of sludge produced is a direct function of the amount of iron removed:

$$2Fe^{2+} + 1/2\ O_2 + 11\ H_2O = 2Fe(OH)_3 \cdot 3H_2O + 4H^+$$

FIGURE 3-2

BACKWASH BASIN VOLUME REQUIRED

FOR 2 FILTERS & 10% RECYCLE

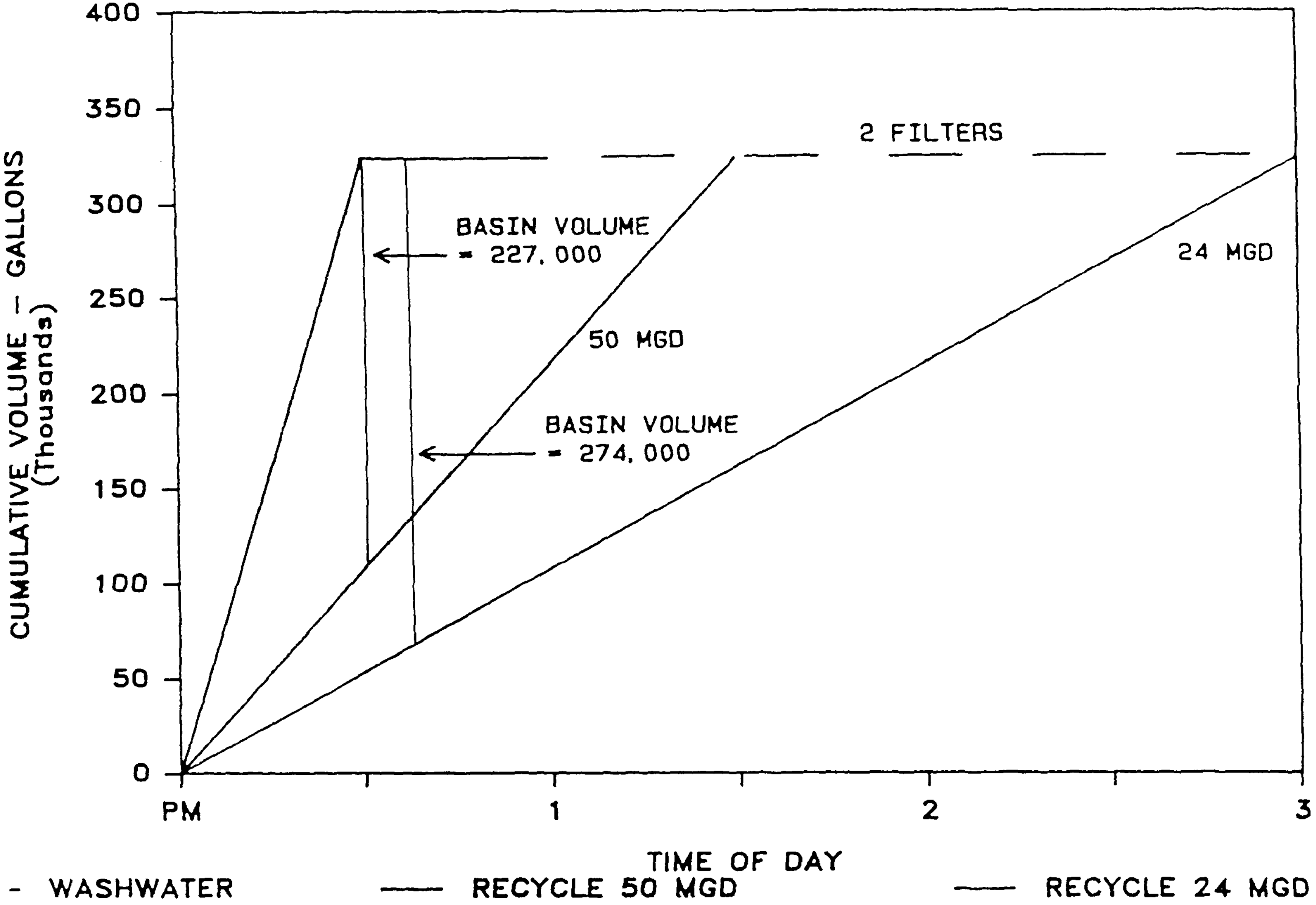
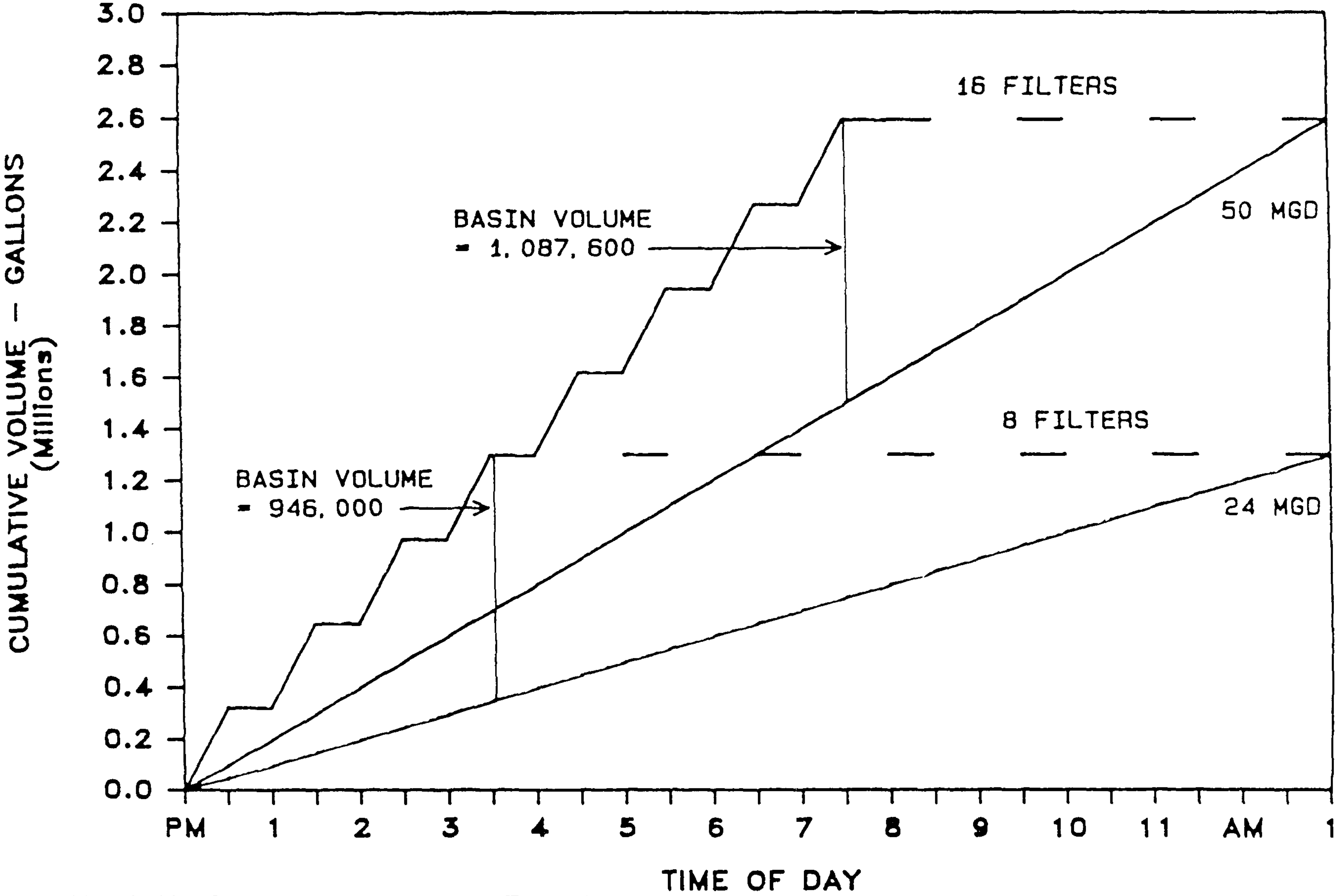


FIGURE 3-3

BACKWASH BASIN VOLUME REQUIRED

FOR 10% RECYCLE AT 24 MGD, 50 MGD



Every mg/l of Fe removed produces 2.9 mg of dry weight solids. The waste stream is produced during the backwashing of the filters. The flows and solids concentrations are very similar to filter backwash wastes discussed earlier.

Snoeyink (3-4) has reported volumes of backwash water produced from iron and manganese removal plants at six water plants in Iowa and Illinois as is shown in Table 3-2. From 3 to 38 gallons of backwash water are produced per 1,000 gallons of raw water filtered. Detention time after aeration does not seem to influence the amount of backwash water produced, nor does the raw water iron and manganese concentrations. Operating procedures seem to dictate the amount of wastewater produced. Low production of wastewater at Estherville is probably due to an extended filter run at the time of data collection. Adair's high wastewater production may be caused by a high solids loading and incomplete media cleaning. Therefore, frequent backwashing is required to maintain effluent quality.

Spent granular activated carbon (GAC) wastes are essentially equal in weight to the exhaustion of carbon, when regeneration is not practiced. This carbon may either be in the filters themselves or in a separate contact chamber following filtration. The water treatment plants individual wasting schedule or unexpected exhaustion of carbon dictates the amount and timing of GAC waste generation. Where carbon regeneration is practiced, the waste would normally occur as fines from backwash of the carbon columns, fines from the quench tank or debris from the scrubber.

In precoat filtration (such as diatamaceous earth filtration) the raw water containing the turbidity to be removed is passed through a uniform layer of filtering media that has been deposited (precoated) on a septum. As the water passes through the media, the turbidity is captured within the precoat media. As the filter run proceeds, additional filter media, called body feed, is continuously metered into the raw

TABLE 3-2

QUANTITY OF BACKWASH WATER FROM IRON AND MANGANESE REMOVAL PLANTS

PLANT	Raw Water Fe (mg/l)	Raw Water Mn (mg/l)	Detention Time After Aeration	Gal. Wastewater 1,000 gal Treated
Adair (9/74) Greensand (2 units)	0.5	0.01	140 min	15
Adair (5/75) (1 unit)	1.1	0.01	140 min	18
Adair (8/83) (1 unit)	--	--	140 min	38
Stuart (10/74) Anthracite (4 units)	0.94	0.01	27 hrs	8.5
Stuart (8/83)	--	--		7.6
Eldon (9/74) Anthracite (4 units)	2.0	0.01	7 min	6.7
*Estherville (10/74) Anthracite (2 units)	1.6	0.24	0 min	3.3
Holstein (10/74) Anthracite (4 units)	1.8	0.15	110 min	10.8
Herscher (4/75) Anthracite (2 units)	0.1 - 0.4	0.29 - 0.63	100 min	17

* Extended Filter Run

Source: Snoeyink (3-4)

water. When the pressure drop builds to the point that water passage is impractical, the filter media and the collected turbidity are cleaned off the system. Most plants using precoat filtration will have a raw water turbidity of under 10 TU. The amount of precoat material used will generally be in the range of 0.1 to 0.2 lb/sf of filter area. The body feed rate can vary between 1 to 10 mg/l of body feed per 1 mg/l raw water suspended solids (Recall that 1 TU does not equal 1 mg/l, and can be from about 0.7 to 2.1 mg/l.).

Techniques for cleaning the filter vary with the different kind of filter vessel and filter element. The most common methods involve (3-5)

- Sluicing the cake from the leaves with high-pressure external sprays directed on the exterior surface of the leaf,
- reversal of flow through tubular filter elements,
- draining the tank under differential air pressure, drying the cake, and then vibrating the leaves to dislodge the cake.

The third method is more commonly used in industries where the liquid being filtered is quite valuable and must be reclaimed, or where dry cake could be more economically handled than a slurry.

In sluice cleaning the entire septum area is covered by either stationary or oscillating fan-spray nozzles that deliver water under a minimum of 60 psig pressure, with sufficient flow to cut away and flush the spent cake off the septa.

When flow reversal is used to clean the filter elements, the velocity and volume of flow to the spent cake from the septa. In some cases, an air-bump technique is used, in which a volume of high-pressure air.

The waste stream from the precoat filtration process will contain the turbidity removed from the raw water, the precoat material and the body feed. The more body feed used, the longer the filter runs and hence the less precoat media used per gallon of water processed. Figure 3-4 shows an example plot between run length, pressure differential and body feed as a ratio to raw water suspended solids (3-5). A pressure differential of 30 psi is usually about the maximum economic level. For a raw water suspended solids of 10 mg/l, the following example could be generated for precoat media. It is assumed that the septum is coated at 0.1 - 0.2 lb/sf and the filtration rate is 2 gpm/sf.

body feed ratio	=	2:1	4:1	6:1
body feed, mg/l	=	20	40	60
filter run length, hrs	=	4	11	13
precoat media used, mg/l	=	94 - 188	34 - 68	29 - 58
		114 - 208	74 - 108	88 - 118

This particular plant would generate anywhere between 74 and 208 mg of waste per liter of water treated, or 617 to 1,734 lb/MG (plus the raw water suspended solids).

In general the precoat plant will produce waste according to the following

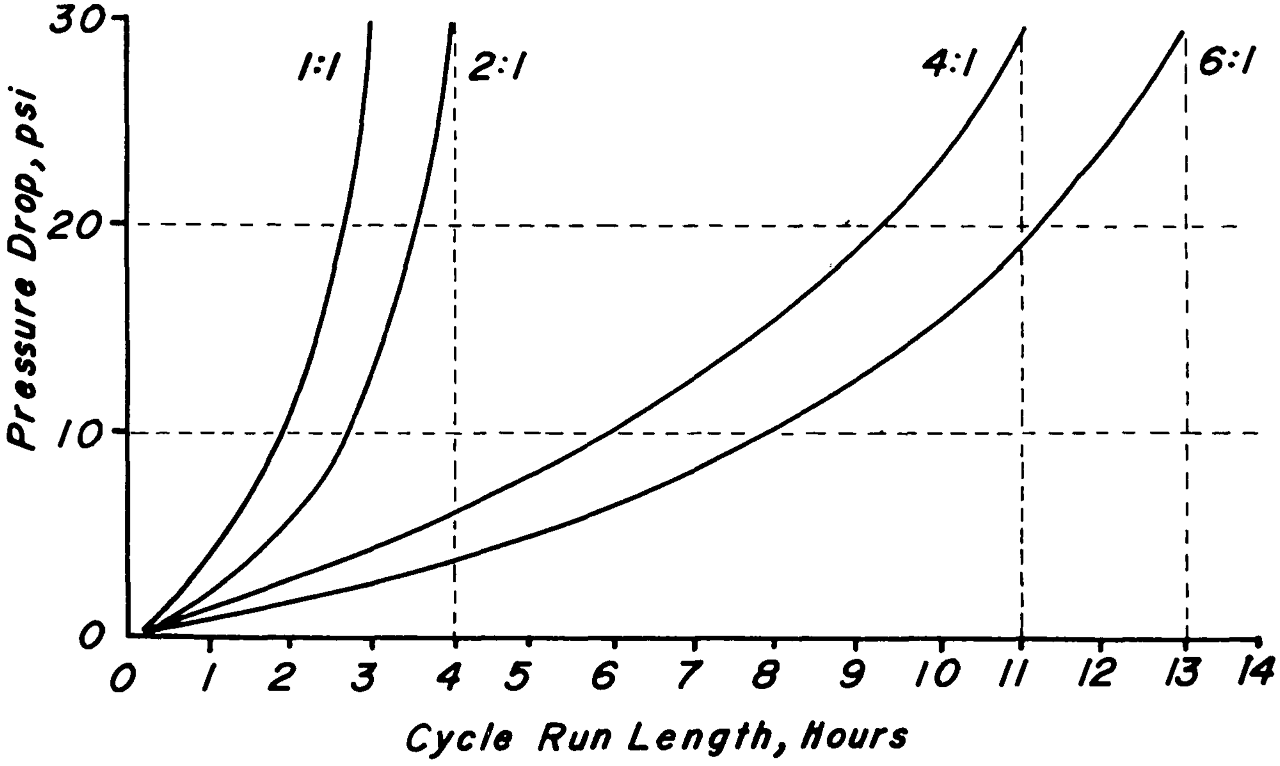
$$S = 8.34\ Q\left(BFR(SS) + \frac{PC(7566)}{FR(T)} + SS\right)$$

where,

S = lb/day of sludge production
Q = plant flow, mgd
BFR = body feed ratio
SS = raw water suspended solids, mg/l

FIGURE 3-4

CYCLE RUN LENGTH AS A FUNCTION OF BODY FEED FOR DIATOMACEOUS - EARTH PLANTS



PC = precoat application rate, lb/sf of filter
FR = filtration rate, gpm/sf
T = filter run length, hr

Slow sand filtration is again becoming a popular method of filtration. Generally it is used by smaller communities that treat a low turbidity, relatively clean raw water source. The slow sand filter is effective in removing Giardia cysts. During the filtration process a black organic detritus (commonly called a schmutzdecke) forms in the top layer of sand. This layer is generally thought to be a gelatinous biological layer which acts as a filtration media for fine particles. As the head loss through a slow sand filter exceeds the allowable level the sand bed must be scraped. The scraping operation generally involves lowering the water level in the filter sufficiently to allow removal of a layer of sand and the schmutzdecke. Empire, Colorado (3-6) reported that their bed needed to be cleaned once every 30 days when a 5 foot head loss is reached. During each cleaning about the top 0.2 in. of sand is removed, producing 14 f^3 of waste/per filter. During the 30 day period each filter treated a total of 3.5 MG, so that for this plant about 4 f^3 of waste are produced for each million gallons processed. The waste material is primarily sand and organic deposit. However, it also probably contains Giardia cysts and bacteria removed from the influent water and should be handled accordingly.

3.2.2. Liquid Phase Wastes

Ion exchange (IX) has been used for many years as a softening process. Most large plants in the past that utilized IX for softening were located near coastal areas so that brine wastes were discharged to the ocean. Many of these plants have been abandoned due to corrosion and high maintenance costs. However, IX is still being practiced by small treatment systems. Its chief advantage over lime softening is the ease of operation. In areas where disposal of lime sludges is a problem, IX may also be favorable if a suitable brine waste discharge point can be found. Wastes are produced from an ion-exchange process from the three phases of bed cleaning: backwash, regeneration and rinse. IX columns will normally be backwashed at a 5 to 6 gpm/sf rate for about 10 minutes, therefore generating 50 to 60 gallons of backwash waste for each square foot of contactor area. The regenerate, or brine waste volume depends upon the type of exchange media used, the amount of cation exchange capacity that is available and the efficiency of regeneration. Natural greensand filters have an exchange capacity of about 0.2 kg/f^3. (The units used for ion exchange are in terms of kg of hardness expressed as $CaCO_3$. Hardness refers to any di-valent or higher cation). The capacity of synthetic ion-exchange resins depends upon the regenerant driving force used, roughly as follows:

Salt Used lb/f^3	Exchange Capacity kg/f^3	Regenerant Use lb NaCl/kg removed
6	1.4 - 1.6	4.0
10	1.8 - 2.0	5.3
15	1.9 - 2.2	7.5

The theoretical salt (NaCl) demand for regeneration is 2.6 lb NaCl/kg hardness removed. Most plants will operate between 8 and 10 lb/f^3 regeneration, whereas small "homeowner" systems will operate at 6 lb/f^3. The strength of the initial regenerant solution will be normally between 8 to 18% NaCl. This results in the following waste production, considering only the regeneration step of bed cleaning:

Regenerant Use lb NaCl/kg Removed	Volume of Regenerant Used For Indicated Solution Strength, gal/kg Removed 6%	 10%	 12%	 16%
4.0	8.0	4.8	4.0	3.0
5.3	10.6	6.4	5.3	4.0
7.5	15.0	9.0	7.5	5.6

Therefore, for every kg of hardness removed approximately 3 to 15 gallons of brine waste are produced. This waste will contain the excess salt in the brine and the cations removed from the resin.

The rinse process is conducted in a downward flow mode for about 30 minutes and will use 20 to 40 gal/f^3 of resin volume. The following example will illustrate the calculation of IX waste production.

A 1-mgd plant with an initial hardness of 224 mg/l as $CaCO_3$ reduces the hardness to 85 mg/l using a synthetic IX resin. The regeneration rate is 6 lb salt/f^3 using a 10% salt solution.

Hardness Removed = 224 - 85 = 139 mg/l

Total Removed = 5.2×10^2 kg/day

The exchange capacity of the media is about 1.5 kg/f^3; therefore, every day 350 f^3 (520/1.5) of resin are exhausted. If the contactor operates at 6 gpm/sf then the surface area is about 115 f^2. The amount of backwash water produced at 50 gal/sf would be about 6,300 gallons. The concentrated brine waste produced would be 4.8 gal/kg or 2,500 gallons. The rinse water rate wastage would be about 10,500 gallons. Total waste flow therefore is 19,300 gallons for the 1 mgd plant example.

The TDS which would be in the waste can be estimated as the excess salt plus the cations removed. All of the chloride should be present and the excess sodium not used in exchange. The theoretical NaCl demand is 2.6 lb/kg, and NaCl is 40% sodium, therefore, the theoretical sodium demand is 1.04 lb/kg. In this example, 4 lb/kg of NaCl was supplied, or 1.6 lb/kg of sodium. The total computation is therefore:

Excess Sodium = (1.6 - 1.04) 520 = 291 lb
Chloride = .6(4) (520) = 1,248 lb
Hardness, as Ca[4] = .4(520) 1,000/454 = 458 lb
1,997 lb

The TDS of the regenerant waste solution itself would therefore be 96,700 mg/l. If blended with the rinse water the strength would be 18,600 mg/l.

Table 3-3 shows actual quantities of brine waste produced at 6 ion exchange plants (3-4). In the theoretical example 7.1 gal of regenerant waste was produced per f^3 of resin material, similar to that for the full scale plants shown in Table 3-3. Most of the full scale plants also used rinse water in the range of 20 to 40 gal/f^3 of resin. The total volume of wastewater produced for the two plants with similar hardness as the example is also very close to the calculated amount.

Water from reverse osmosis (RO) processes consists of the reject raw water containing the original constituents in that water that did not pass through the membrane. The volume of reject water is generally dependent on limiting the build up of a given ion to prevent membrane fouling. The reject water will often be in the range of 25 to 35% of the feed flow to the membrane process. The concentration of constituents in the reject will therefore be 3 to 4 times the concentration in the feed water.

---

[4] The conversion between Ca as $CaCO_3$ to as Ca is 0.4. 1,000/454 converts kg to lb.

TABLE 3-3

REGENERATION OF CATION EXCHANGE RESINS

PLANT	Gal. Wastewater/ 1,000 gal. Water Processed*	Raw Water Total Hardness (mg/l as $CaCo^3$)	Gal. Regenerant/ ft^3 resin	Gal. Rinse/ ft^3 resin	Concentration of Brine %	Gal. Processed/ ft^3 resin-cycle	Dosage (lb.salt/ft^3 resin)
Crystal Lake Plant #6	21.9	233	7.3	19.4	11	1,220	6.6
Crystal Lake Plant #8	17.2	244	5.1	19.0	15	1,400	6.5
Eldon	71.9	75	3.9	61.7	17	750	5.6
Grinnel	49.5	388	14.5	35.0	6	1,000	7.2
Holstein	53.5	885	5.7	19.7	14	475	6.6
Estherville	82.8	915	4.4	24.7	15	350	5.5

Note: The chemical characteristics of the backwash water did not show a large variation from the raw water, and therefore, they were not included in this analysis.

*Includes backwash water, regenerant waste and rinse water

Source: Snoeyink (3-4)

3.2.3. Gas Phase Wastes

Gas phase wastes are produced during the stripping of volatile gasses. Volatile gasses that may be removed during air/water contact include carbon dioxide, hydrogen sulfide, volatile organic compounds (including THMs) and radon. Gas removal from the water can take place through a natural water/air contact such as in a mix tank or sedimentation basin, through a natural draft aerator or through a forced air stripping system. For the first two systems it is nearly impossible to estimate an initial volume of air which contains the waste gas. For these systems only the quantity of gas contaminant itself can be estimated. For forced air systems, the gas volume as it leaves the stripping system is fairly well defined as a design parameter. Typical air:water ratios used may be 10:1 to 40:1, based on a volume to volume basis, or for every f^3 of water treated 10 to 40 f^3 of air are produced as a waste product. The volume of air can be estimated based on the water flow rate, if the air:water ratio (R) is known:

Air Waste, cfm = 92.8 RQ

Q = water flow, mgd

3.3. PHYSICAL CHARACTERISTICS

Physical characterization of water plant wastes is primarily directed at solid/liquid waste streams. Knocke (3-7) has divided physical properties of sludges into their 'macroproperties' and 'microproperties'. Macroproperties are such parameters as specific resistance, settling rates, and cake solids concentrations. Microproperties would include particle size distribution and density. Knocke has evaluated several of these microproperties and their effect on dewatering of alum and lime sludges. In general he found alum sludges that are predominantly aluminum hydroxide (low turbidity raw waters) had densities of 1.002 to 1.008 g/cm^3 and correspondingly dewatered to a 7 to 19% solids concentration by laboratory vacuum filtration. Alum sludges from higher turbidity raw waters had densities of 1.008 to 1.018 g/cm^3 and dewatered in the lab to a 20 to 34% solids concentration. The lime sludge evaluated had a density of 1.292 g/cm^3. An evaluation of floc shape for alum sludges showed the particles to be elliptical, with a ratio of minor to major axes between 0.5 to 0.75. The axes length for the major axes was in the range of 5 to 50 um for unconditioned alum sludges and 5 to 100 um for polymer conditioned sludges. Although a wide scatter in the data existed, he found that generally the larger the floc particle (longer axes length) the lower the specific resistance value. Polymer addition had the effect of increasing the particle size, resulting in a decrease in sludge specific resistance.

Tests which define macroproperties of sludges can be used to assist in the selection of dewatering aids and to determine relative ease of dewatering. They can be useful as an operating tool to determine conditioning doses on a routine basis. The four main tests are the specific resistance test, the time to filter test, the filter leaf test and the capillary suction time test. Each is described below.

3.3.1. Specific Resistance Test

The specific resistance test has been used to optimize sludge dewatering performance. The test is most valuable when evaluating chemical conditioning of sludge for full scale applications.

As shown in Figure 3-5, a simple Buchner Funnel apparatus can be used to determine the specific resistance of water plant sludges. Typically 100 ml portions of the sludges are added to the Buchner Funnel and the volume of filtrate generated at various times recorded. Based on the Carmen-Kozeny equation for flow through porous media an equation can

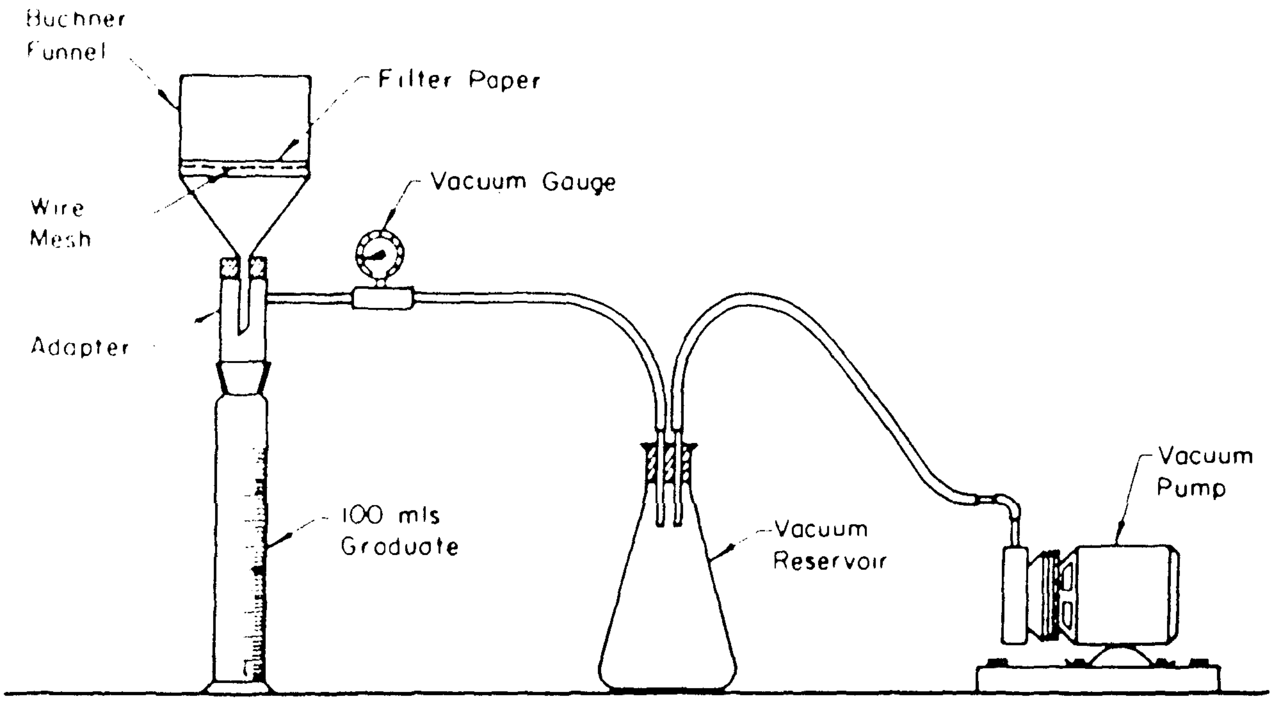


FIGURE 3-5

BUCHNER FUNNEL APPARATUS

be developed to describe the flow through a sludge cake and associated support media. The equation is written as:

$$\frac{t}{V} = \frac{uWR}{2PA^2} V + \frac{uR_r}{AP}$$

which is of the form:

$$\frac{t}{V} = bV + a$$

Therefore a plot of t/v vs V should give a straight line with a slope "b" and intercept "a" as shown in Figure 3-6. Since the slope of the line b equals:

$$b = \frac{uWR}{2PA^2}$$

then the specific resistance is

$$R = \frac{2bPA^2}{uW}$$

where,

R = specific resistance (sec^2/g) of sludge
b = slope of line (sec/cm^6)
P = vacuum applied (cm of water)
A = filter area (cm^2)
u = filtrate viscosity (poise)
w = dry weight of solids per volume of filtrate (g/cm^3)
R_r = specific resistance (sec^2/g) of filter media

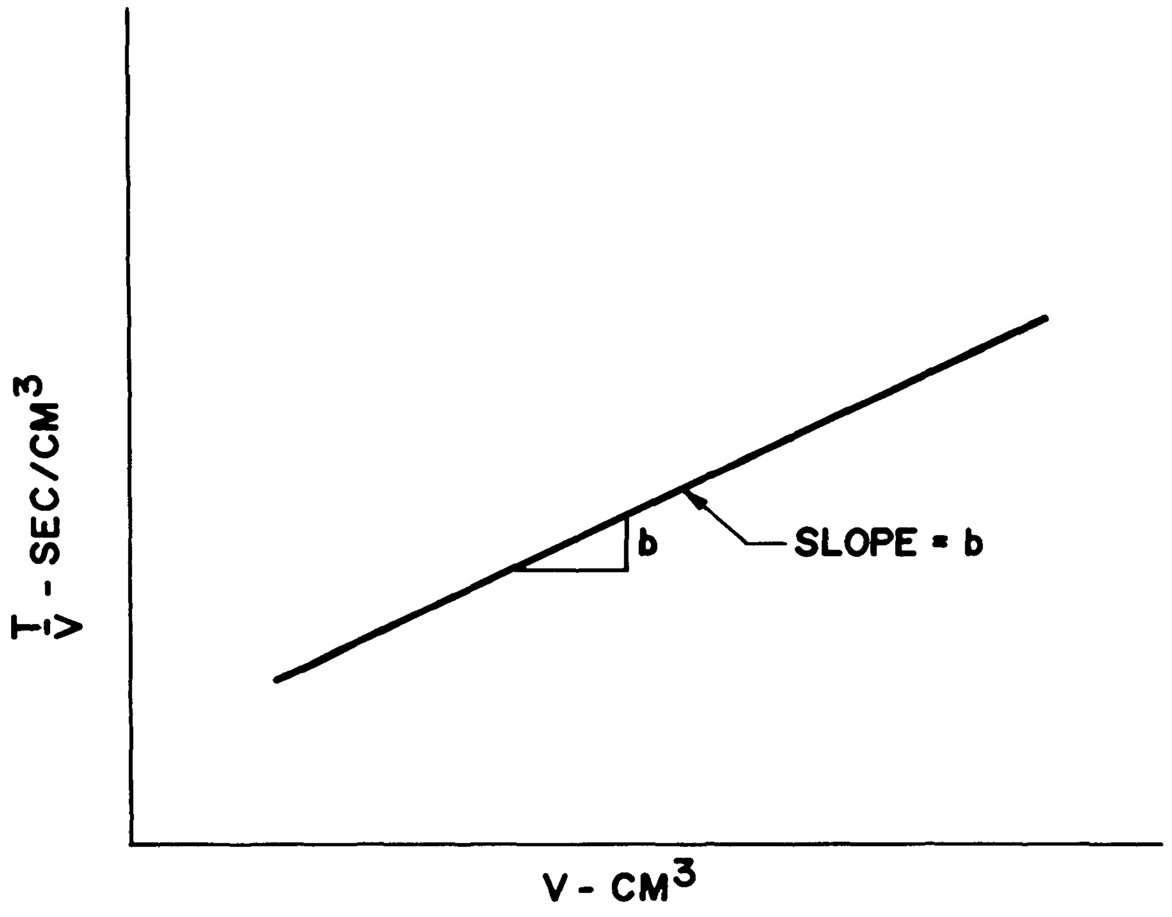


FIGURE 3-6

TIME/FILTRATE VOLUME VS. FILTRATE VOLUME PLOT USED IN SPECIFIC RESISTANCE TESTING

By repeating the specific resistance test for untreated sludge and sludges receiving various chemical conditioning it is possible to determine the optimum chemical treatment as shown in Figure 3-7. However, this is optimum only at the sludge solids concentration used in the experiment. Table 3-4 shows specific resistance values for several water plant sludges (3-8).

While specific resistance data could theoretically be used to size full scale equipment, this is not a recommended practice. The test can be useful for conditioning studies.

Most sludges are compressible and the degree of compression appears to depend on the vacuum applied and the geometry of the dewatering system. An empirical expression has been found to adequately relate specific resistance and the vacuum level. This expression states that

$$R = CP^{s}$$

where,

R = specific resistance
C = cake constant
s = coefficient of compressibility
P = vacuum applied

Both the cake constant, C, and the coefficient of compressibility, s, can be determined from a logarithmic plot of specific resistance vs the vacuum level. The coefficient of compressibility, s, is the slope of the straight line generated while the cake constant C is the intercept where P = 1. The coefficient of compressibility equals zero for an incompressible sludge. Hamon (3-9) found the coefficient of compressibility to vary from 0.6 to 0.8 for aluminum hydroxide sludges and 0.71 to 0.83 for ferric hydroxide sludges. Knocke (3-7) reported values of 0.97 for alum sludges and 0.8 for lime sludges.

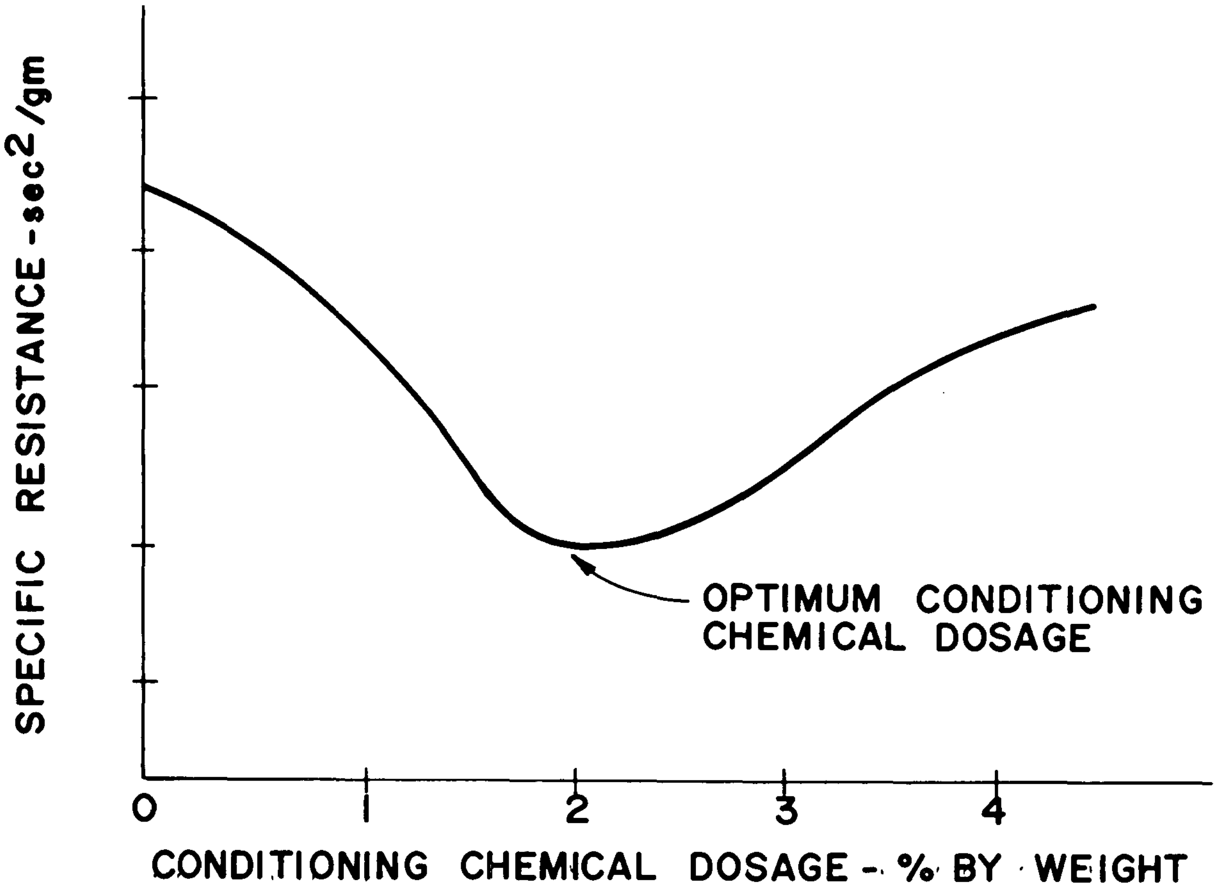


FIGURE 3-7

USE OF SPECIFIC RESISTANCE TO DETERMINE OPTIMUM CHEMICAL DOSAGE

TABLE 3-4

SPECIFIC RESISTANCE FOR VARIOUS CHEMICAL SLUDGES

Sludge	Location	Specific Resistance 10^6 (sec^2/g)
Lime and iron	Jefferson City	2.11
Lime and iron	Jefferson City	4.3
High magnesium softening sludge	Kansas City	5.49
Lime and alum	Boonville	5.83
Excess lime and alum backwash	Boonville	5.98
Lime and iron	Jefferson City	6.12
Lime and iron	Jefferson City	6.79
Lime and iron	Jefferson City	7.0
Softening	Kansas City	11.57
Excess lime and alum backwash	Boonville	13.2
Cationic flocculent	St. Joseph	14.1
Lime and iron	St. Louis	21.2
High magnesium softening sludge	Kansas City	25.1
Iron	St. Louis	40.8
Lime and alum	Boonville	53.4
Iron backwash	St. Louis Co.	76.8
Iron	St. Louis Co.	77.6
Cationic-flocculent backwash	St. Joseph	80.1
Iron backwash	St. Louis	121.8
Iron	St. Louis Co.	148.5
Alum	Moberly	164.3

Source: D.J. Calkins and J.T. Novak (3-8)

Figure 3-8 shows results of specific resistance as a function of polymer dose for alum sludge at Durham, N.C. The test was conducted using a 100 ml sample and using a 50 ml sample. The 100 ml sample gave better results, primarily since more accurate volume vs time data can be collected during the test procedure.

A simplification of the specific resistance test is the time to filter test (TTF). This test is set up with the same buchner funnel apparatus as the specific resistance test. The only data collected is the time for one-half of the volume to filter. This test is much faster to run and analyze the data than the specific resistance test and can provide useful information on the effects of conditioning procedures. Sample size can again be very important in obtaining good differentiation of results. Figure 3-9 shows test results on the TTF test for the same Durham sludge as the specific resistance test results above. It can be seen that the TTF test did predict the same optimal polymer dose as the specific resistance test.

3.3.2. Filter Leaf Test

The filter leaf test (Figure 3-10) duplicates on a laboratory scale vacuum filter operations as closely as possible. With the filter leaf test, the solids concentration of the sludge, vacuum level, filter media, cycle time, sludge conditioning, and submergence time (or percent filter submergence) may be varied. Sludge dewatering equipment scale up can be achieved by working with uniformly mixed representative sludges and duplicating the conditions to be employed in the full scale. The filter cloth of interest should be used and the cloth should be conditioned before collecting design or operating data.

Sludge samples can be prepared in a standard jar test apparatus in 2 liter batches, and gently transferred to a beaker for testing. The filter leaf containing the filter

FIGURE 3-8

SPECIFIC RESISTANCE VS POLYMER DOSAGE

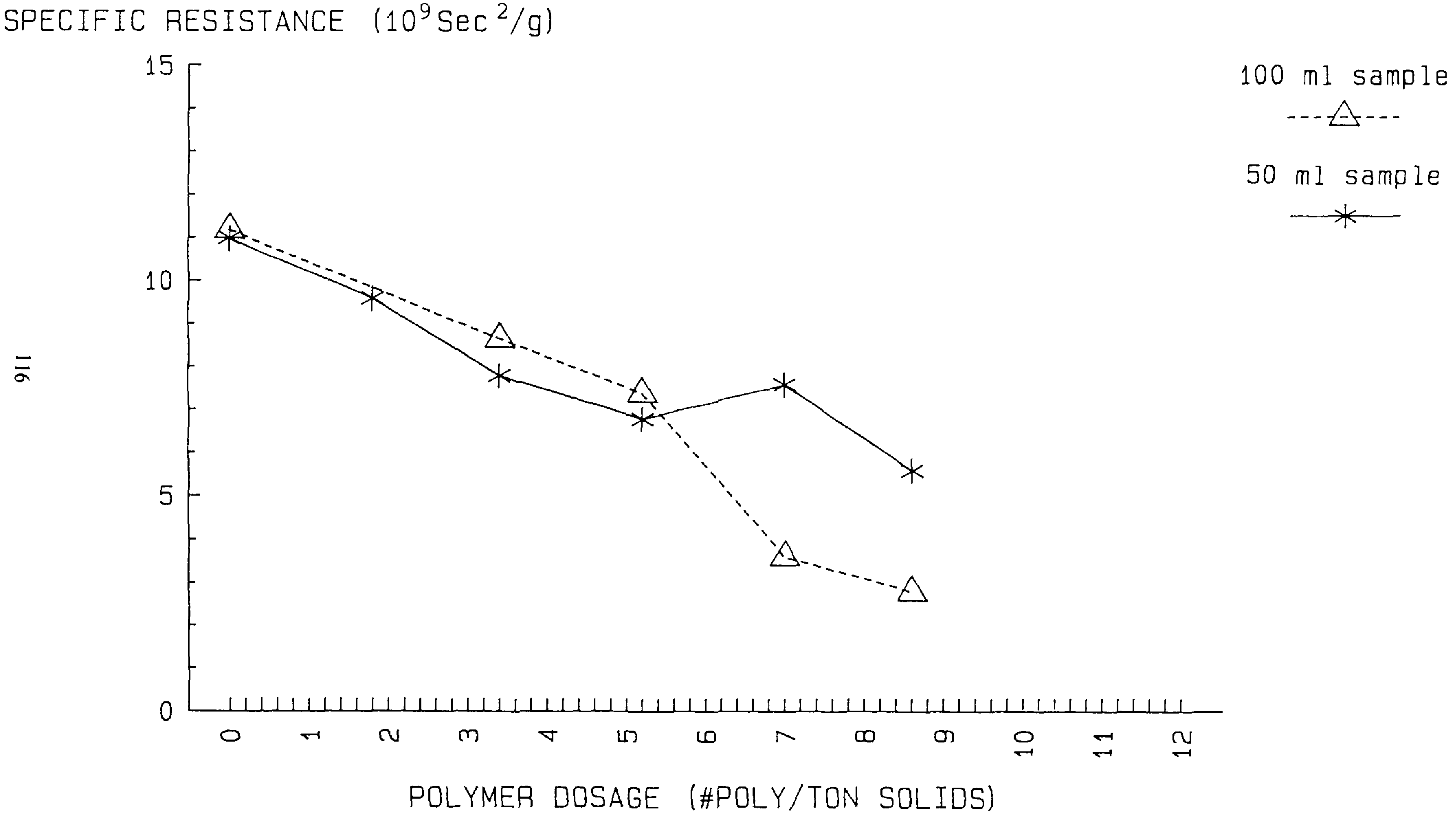
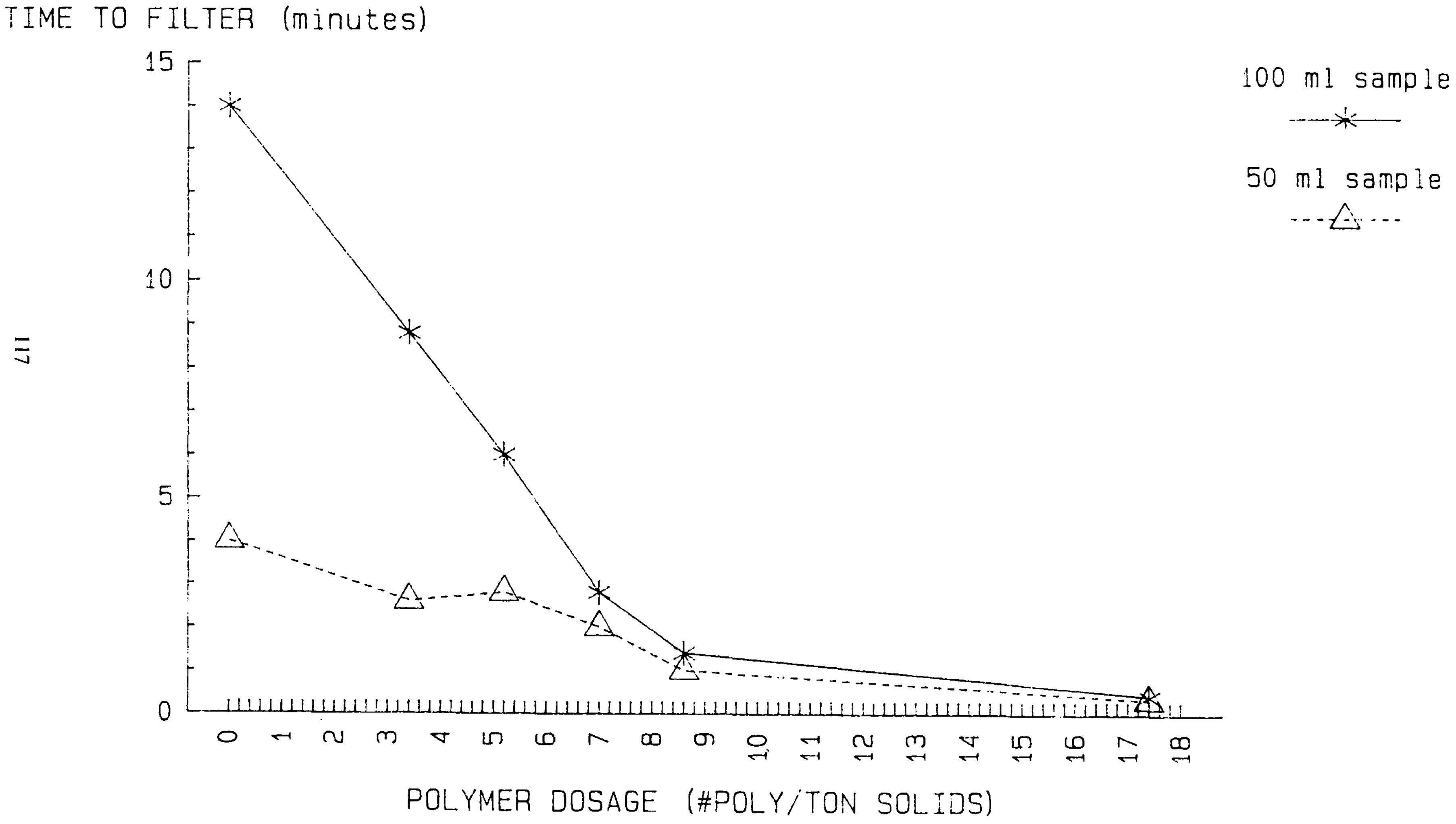


FIGURE 3-9

TIME TO FILTER (TTF) VS POLYMER DOSAGE



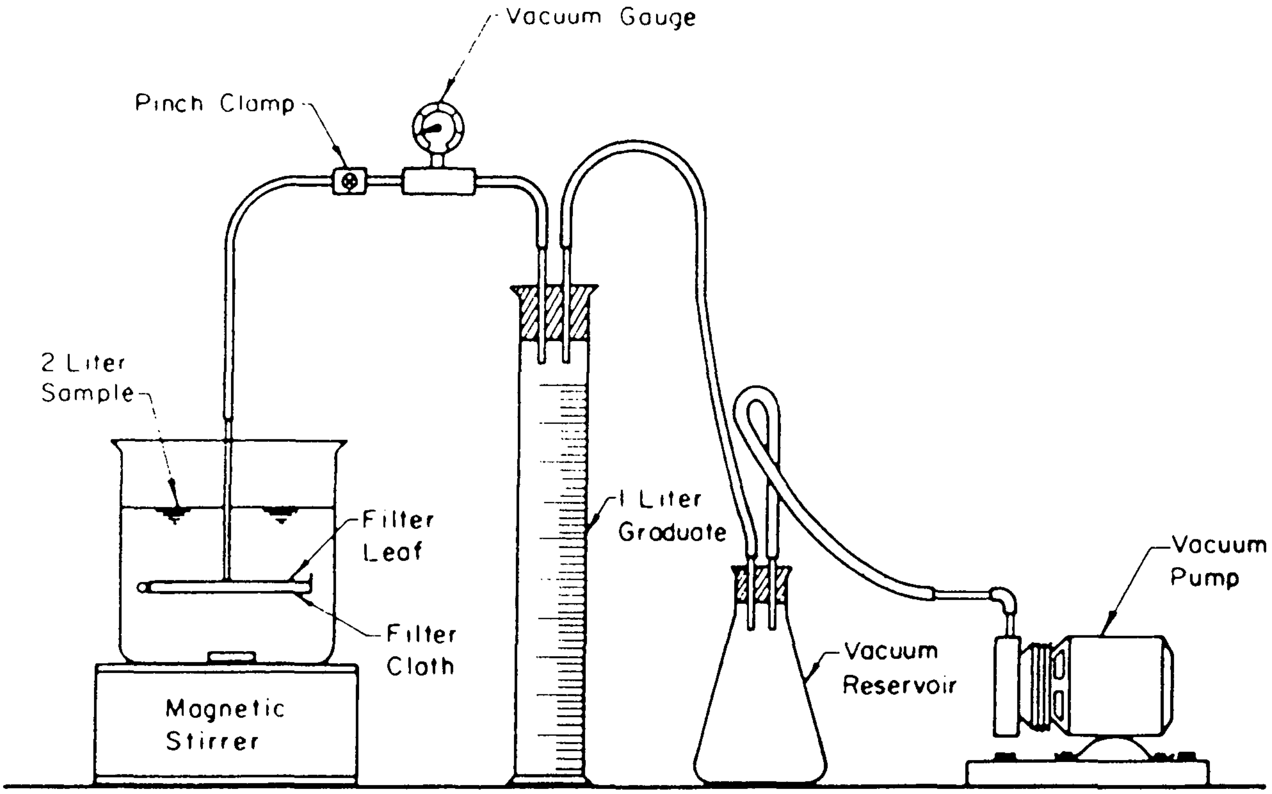


FIGURE 3-10

LEAF TEST APPARATUS

media to be used or evaluated is lowered into the well mixed sludge. The vacuum level and cake forming cycle should be the same as those used in the full scale equipment. At the end of the form time the filter leaf is gently removed from the sludge and allowed to dry in the atmosphere under the same vacuum level and drying time used in normal operations or design. At the end of the drying cycle, the filter cake thickness is measured and the solids removed from the filter media. The filtrate volume, wet and dry weight of solids recovered and the solids content of the cake are normally determined. The quality of the filtrate may also be of interest. From these test data the filter yield may be calculated as:

$$Y = \frac{W}{A\ T}$$

where,

Y = filter yield in dry solids produced per unit area per hour (lbs/sf/hr)
W = weight of dry cake formed during the test (lbs)
A = area of filter, sf
T = total cycle time, hr

The total cycle time includes the time the filter is submerged, the drying time and the cake removal time . The following example illustrates the calculation procedure:

Feed Solids Concentration = 4%
Cake Solids Concentration = 20%
Area of Filter Leaf = 0.1 sf
Total Cake Dry Weight = 0.028 lb

Cycle: 30 sec submerged
60 sec drying
30 sec off the filter
2-min cycle

Filtrate Solids Concentration = 500 mg/l

The filter yield can be calculated as follows:

$$\text{Filter Yield} = \frac{(0.028 \text{ lb}/0.1 \text{ ft })}{2 \text{ min} \times \frac{\text{hr}}{60 \text{ min}}} = 8.4 \text{ lb/sf/hr}$$

The experimental filter yields can be investigated for various sludge conditioning techniques (Figure 3-11) or vacuum filter operating modes to optimize design or operation.

3.3.3. Capillary Suction Time

The capillary suction time (CST) technique is one of the fastest and simplest tests to perform on the dewatering characteristics of sludges. The results are very useful for comparing conditioning methods or as an operator tool in determining polymer dose for full-scale dewatering devices.

The capillary suction time test is run on the apparatus shown in Figure 3-12. A representative sludge sample is placed into the cup in Figure 3-12. As the sludge is dewatered the liquid flows outward through a special blotter paper. As the liquid flows by the first probe it activates a timer and the timer stops when the flow reaches the second set of probes which are normally 1 cm away. Samples are conditioned by mixing a known concentration of polymer (or other conditioner) to the sludge. The sample is then poured into the tube of the CST apparatus. Figure 3-13 shows the CST results for the Durham sludge.

The two curves of Figure 3-13 also reflect the different tubes available from the manufacturer. It appeared for this sludge that the narrow tube diameter (taller tube) gave better

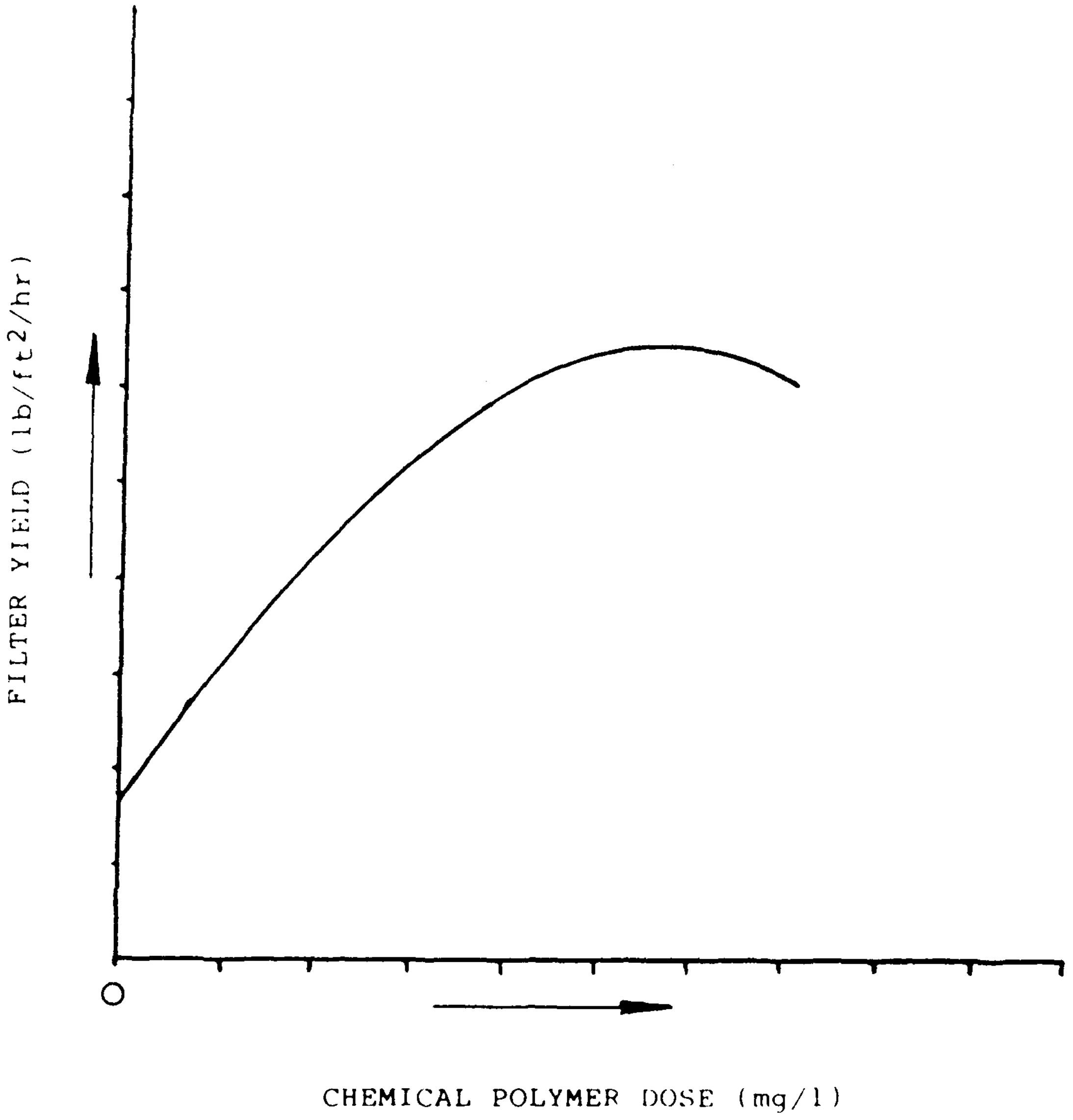


FIGURE 3-11

THE OPTIMIZATION OF CHEMICAL CONDITIONING WITH THE FILTER LEAF TECHNIQUE

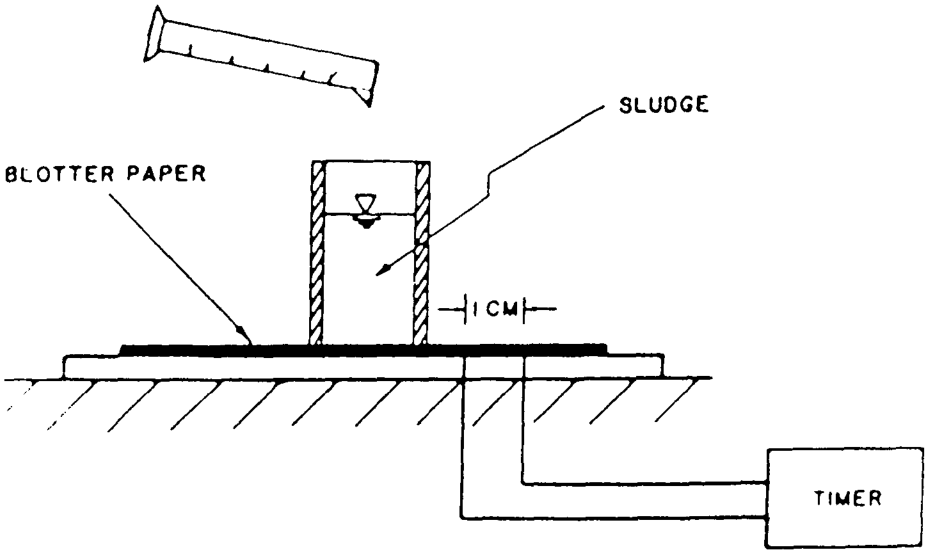
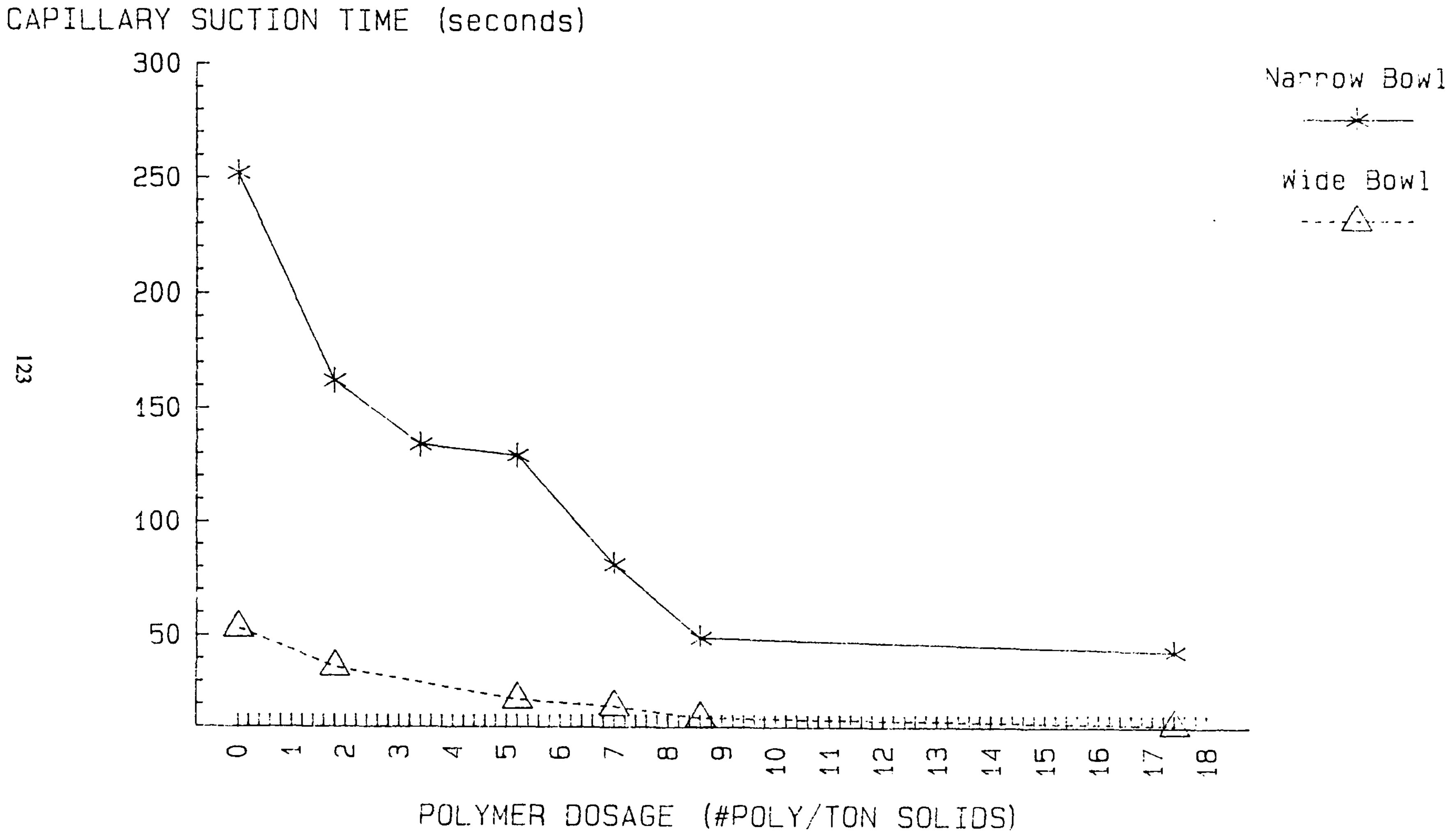


FIGURE 3-12

CAPILLARY SUCTION TIME (CST)
APPARATUS

FIGURE 3- 13

CAPILLARY SUCTION TIME (CST) VS POLYMER DOSAGE



defined results. Again the optimal polymer dose was about the same as predicted by the specific resistance and TTF test.

3.4. CHEMICAL CHARACTERISTICS

3.4.1. Solid/Liquid Wastes

The major constituent of any of the water plant wastes is water. In the case of suspended solids in the water the treatment objective is generally one of solid/water separation. When the waste contains predominantly dissolved solids in the water, either a phase change is required to remove the TDS, or the water and TDS are jointly disposed of in some acceptable manner.

Those wastes in which the water contains suspended solids are traditionally referred to as sludges. Vesilind (from ref. 3-7) classified the water content of wastewater sludges into four categories:

1. Free water is not held to sludge solids and can be removed by simple gravitational settling.
2. Floc water is trapped within sludge flocs and can be removed by mechanical dewatering.
3. Capillary water is held to sludge solids by surface tension and attractive forces and can be removed only by compaction and deformation of the sludge flocs.
4. Bound water is chemically bound to the individual floc particle and cannot be removed.

For chemical sludges Cornwell (3-10) has proposed three water classifications

1. Free water can be removed by drainage or low pressure mechanical methods.

2. <u>Hydrogen bound water</u> is attracted to the floc particle through hydrogen binding. The force of attraction of the water to the chemical floc is in the range of 0.13 kcal/mole.

3. <u>Chemically bound water</u> is bound through covalent bonds directly to the chemical floc.

Both aluminum and iron have a covalent number of 6, resulting in freshly precipitated flocs (after reaching equilibrium) of the form $Al(OH)_3 \cdot 3H_2O$ and $Fe(OH)_3 \cdot 3H_2O$. In the case of aluminum the chemically bound water is about 40%. Therefore, it is not likely that a mechanical device will dewater a sludge predominant in the chemical hydroxide floc to greater than a 60% solids concentration. In practice, dewatering is limited to achieving a 45 to 50% solids concentration. As the sludge ages the floc will slowly equilibrate to the oxide form (Al_2O_3 or Fe_2O_3), and solids concentrations in the 90% range can be achieved. Low pressure mechanical devices do not have sufficient energy to overcome hydrogen binding and hence vacuum filters, centrifuges and belt presses will generally only remove the free water and the water physically trapped within the floc particles.

Knocke and Wakeland (3-7) reported results where they used Vesilind's procedure to differentiate between the type of water. Using a centrifuge with increasing applied force, Knocke only found one type of disiluted water, with the solids concentration for alum sludge slowly approaching 15 to 17%. It is likely that the centrifuge only removed the free water and could not remove the hydrogen bound water. More energy would have been required, such as high-pressure equilibrium testing, to release the hydrogen bound water.

Chemical characterization of these sludges has previously only been of interest to researchers in the field or to those utilities considering chemical recovery. However, disposal concerns have increased the awareness of constituents of the chemical sludges. As the drinking water quality standards increase, so will the chemical levels in the waste streams. Analysis of wastes may need to be conducted based on the total concentration of chemical present or based on the characteristic EP Toxicity test. For liquid wastes, units of mg/l are appropriate and for solid wastes (dewatered sludges) units of mg/kg dry weight sludge would be appropriate. When trying to report results for comparative purposes it is useful to present all sludge data (dilute or concentrated) in terms of mg/kg dry weight.

Given and Spink (3-11) conducted a literature review and summarized their findings on alum sludge characteristics which is shown as Table 3-5. Schmitt and Hall (3-12) have reported an elemental analysis of alum sludge obtained from the Oak Ridge, Tennessee Water Treatment Plant as shown in Table 3-6. This sludge is characterized by a high percentage of non-hydrated metal species. Aluminum is 17,000 mg/kg or as the precipitated species about 8% of the solids are aluminum hydroxide. It is interesting to note that the rare earth elements were detected in the solids. In general, the rare earth elements of even atomic number were found in greater abundance than those of odd atomic number. This is in agreement with Haskins rule regarding the relative order of abundance of rare earth elements in the lithosphere. The analysis also shows as is typical of many east coast sludges, the presence of lead, zinc and manganese. This particular waste also showed 2.1 mg/kg of U^{238}. Cornwell (3-13) has reported metal analyses for sludges from Durham, North Carolina and Tampa, Florida (3-14) as shown in Tables 3-7 and 3-8. The Durham, North Carolina sludge is 38% aluminum hydroxide species, and in this case another 4% is iron hydroxide species. This waste shows some zinc and lead and a large amount of manganese. Most of this manganese is associated with colloidal manganese in the raw water since the plant does not oxidize prior to sedimentation. Also shown in Table

TABLE 3-5

REPORTED ALUM SLUDGE CHARACTERISTICS

Parameter	Concentration
Total Solids (TS)	0.1 to 27% by weight
Volatile Solids	10 to 35% of total solids
Suspended Solids	75 to 99% of total solids
pH	5.5 to 7.5
BOD	30 to 6,000 mg/l
COD	500 to 27,000 mg/l
Aluminum	4 to 11% of TS as Al (limited data)
Iron	6.5% of TS (one sample)
Manganese	<0.005 - 5% of TS
Arsenic	<0.04% of TS
Cadmium	<0.005% of TS
Individual Heavy Metals	<0.03% of TS
Total Kjeldahl Nitrogen	0.7 to 1,200 mg/l as N
Phosphate	0.3 to 300 mg/l as P
Total Plate Count	30 to >300,000 per ml

Source: Given and Spink (3-11)

TABLE 3-6

ELEMENTAL ANALYSIS OF ALUM SLUDGE FROM OAK RIDGE, TENNESSEE

Atomic No.	Element	Backwash Solids mg/kg	Sedimentation Basin Solids mg/kg	Blank
4	Be	0.29	<0.03	<0.02
5	B	1020	27	1
9	F	20	19	1
11	Na	39	10	4
12	Mg	833	1183	10
13	Al	Matrix	13500	6
14	Si	12500	17833	40
15	P	200	183	10
16	S	340	167	10
17	Cl	567	267	30
19	K	Matrix	1650	50
20	Ca	Matrix	3333	300
21	Sc	2	5	0.3
22	Ti	8500	6833	7
23	V	273	30	0.4
24	Cr	7260	200	4
25	Mn	1800	983	1
26	Fe	42733		40
27	Co	138	3	0.4
28	Ni	103	140	2
29	Cu	120	7	6
30	Zn	467	167	5
31	Ga	3	8	<0.3
32	Ge	0.9	2	<0.5
33	As	15	13	3
34	Se	1	1	<0.4
35	Br	6	13	<0.4
37	Rb	6	20	<0.3
38	Sr	200	93	0.4
39	Y	7	3	<0.2
40	Zr	80	40	<0.4
41	Nb	2	5	<0.2
42	Mo	4	<1	<0.9
44	Rw	<0.7	<1	<0.7
45	Rh	0.5	<3	<0.2
46	Pd	3	2	<0.9
47	Ag			
48	Cd	1	<1	<0.9
49	In	0.1	<0.4	<0.3

TABLE 3-6 (con't)

ELEMENTAL ANALYSIS OF ALUM SLUDGE FROM OAK RIDGE, TENNESSEE

Atomic No.	Element	Backwash Solids mg/kg	Sedimentation Basin Solids mg/kg	Blank
50	Sn	1	<1	<0.8
51	Sb	0.3	2	<0.5
52	Te	<0.9	<1	<0.9
53	I	1.9	6	<0.3
55	Cs	0.9	1	<0.3
56	Ba	367	333	<0.4
57	La	90	8	<0.3
58	Ce	97	28	<0.4
59	Pr	5	6	<0.3
60	Nd	113	18	<1
62	Sm	9	15	<1
63	Eu	4	2	<0.7
64	Gd	10	7	<2
65	Tb	0.8	1	<0.4
66	Dy	3	3	<1
67	Ho	0.6	1	<0.4
68	Er	1	<2	<1
69	Tm	0.6	<0.6	<0.4
70	Yb	2	<2	<1
71	Lu	0.6	<0.6	<0.4
72	Hf	3	<2	<2
73	Ta	9	17	20
74	W	5	<2	4
75	Re	<0.7	<0.9	<0.7
76	Os	<1	<1	<1
77	Ir	<0.7	<0.9	<0.7
78	Pt	2	<2	<1
79	Au	1	2	1
80	Hg	<2	<2	<2
81	Tl	2	<0.9	<0.7
82	Pb	50	47	1
83	Bi	0.6	<0.8	<0.5
90	Th	4	3	0.8
92	U	8	7	2

Source: Schmitt and Hall (3-12)

TABLE 3-7

METAL CONCENTRATIONS IN DURHAM, N.C. ALUM SLUDGE AND THEIR SOURCE

	Concentration of Metal in Commercial Alum		Concentration of Metal in 2.5% Sludge		Source of Metals
Metal	mg/l	ug metal/mg Al	mg/l - mg/kg	ug metal/mg Al	% from Alum
Cd	ND	0	ND - 0	0	---
Cr	9.5	0.2	0.6 - 24	0.3	66
Cu	0.1	0.002	0.6 - 24	0.3	0
Fe	1,160	18.4	292 - 11,680	146	12
Na	57	0.9	6.5 - 260	3.3	27
K	5.6	0.1	6.1 - 244	3.0	3
Mn	1.7	0.03	255 - 10,200	127	0
Ni	0.1	0.002	0.06 - 2.4	0.03	0
Pb	1.5	0.02	0.03 - 1.2	0.02	100
Zn	1.1	0.02	1.7 - 68	0.90	2
Ca	6.3	0.1	2.8 - 112	1.4	7
Mg	12.5	0.2	5.5 - 220	2.7	7
Al	63,000	---	1,970 - 78,800	---	---
Si	14.2	0.2	8.5 - 340	4.2	5
Ba	0.5	0.01	0.3 - 12	0.1	10
Ag	0.4	0.01	ND - 0	0	100
As	3.0	0.05	1.1 - 44	0.05	100
Se	ND	0	ND - 0	0	---
Hg	0.001	0	0.002 - 0.8	0	---

ND = Below detection limit

Source: Cornwell (3-13)

TABLE 3-8

METAL CONCENTRATION IN TAMPA, FLORIDA ALUM SLUDGE

	mg/l	mg/kg
Aluminum (Al)	850	170,000
Barium (Ba)	<1.0	---
Cadmium (Cd)	<0.01	---
Chromium (Cr)	0.35	70
Cobalt (Co)	0.08	16
Copper (Cu)	0.45	90
Iron (Fe)	33	62,400
Lead (Pb)	0.50	100
Magnesium (Mg)	12	2,360
Manganese (Mn)	0.34	68
Silver (Ag)	<0.01	---
Zinc (Zn)	0.11	22

Source: Cornwell (3-14)

3-7 is the estimated percent of the metal in the sludge which originates from the alum itself. The Tampa sludge analysis shows the characteristics of a highly colored, low turbidity raw water source. The solids are 80% aluminum hydroxide species and another 15% iron hydroxide. Nearly all of this sludge will dissolve at pH 1 to 2.

In Chapter 2, the importance of the EP Toxicity test in classifying a waste as hazardous according to RCRA requirements was discussed. Results for EP Toxicity tests for three different alum sludges is shown in Table 3-9. For these three sludges all concentrations are reported as essentially non-detectable and are well below the EP Toxicity criteria.

Total analyses for lime sludges are not reported in the literature. The wastes predominately consist of calcium carbonate and magnesium hydroxide. Table 3-10 shows partial analyses for 3 lime sludges.

Calkins and Novak (3-8) have investigated the importance of the calcium to magnesium ratio in the settling and dewatering characteristics of lime sludges and the AWWA sludge disposal committee has reported (3-2) that a sludge with a Ca:Mg ratio less than 2 will be difficult to dewater, whereas, a sludge with a Ca:Mg ratio greater than 5 will dewater relatively easily. A plot of Ca:Mg ratio versus the settled solids concentration and filter cake solids concentration is shown in Figure 3-14 (3-8). Similarly, high magnesium content of lime sludges adversely affects the specific resistance as shown in Figure 3-15.

Although specific inorganic concentrations are not reported in the literature for lime sludges, it is certain that they will be present to the extent that they are removed from the raw water. Figure 3-16 shows a generalized guideline of inorganic contaminants likely to be removed by lime softening for the indicated pH range (3-16). As can be seen, some removal of most of the inorganic contaminants will take place, with high removals for some of the compounds. It is

TABLE 3-9

EP TOXICITY TEST RESULTS FOR ALUM SLUDGE

Parameter, mg/l	Saltonstall Conn*	West River Conn*	Chesapeake Va**
Arsenic	<0.01	<0.01	<0.003
Barium	0.21	0.1	<0.1
Cadmium	<0.005	<0.005	0.005
Chromium	<0.01	<0.01	<0.05
Lead	<0.01	<0.01	<0.05
Mercury	<0.001	<0.001	<0.001
Selenium	0.09	<0.01	<0.007
Endrin	<0.01	0.01	<0.01
Lindane	<0.0002	<0.0002	<0.0002
Methoxychlor	<0.004	<0.004	<0.0001
Toxaphene	<0.1	<0.1	<0.002
2, 4-D	<0.005	<0.005	<0.001
2, 4, 5-T (Silvex)	<0.01	<0.01	<0.001

* Source: Bugbee and Frink (3-15)
** Source: City of Chesapeake, Va.

TABLE 3-10

PARTIAL CHEMICAL COMPOSITION OF LIME SLUDGES

	Boulder City Nevada	Miami Florida	Cincinnati Ohio
	(Percent By Weight)		
Silica, Iron Aluminum Oxide	2.6	1.5	4.4
Magnesium Oxide	87.2	93.0	88.1
Calcium Carbonate	7.0	2.8	2.3

Source: AWWA Sludge Committee Report (3-2)

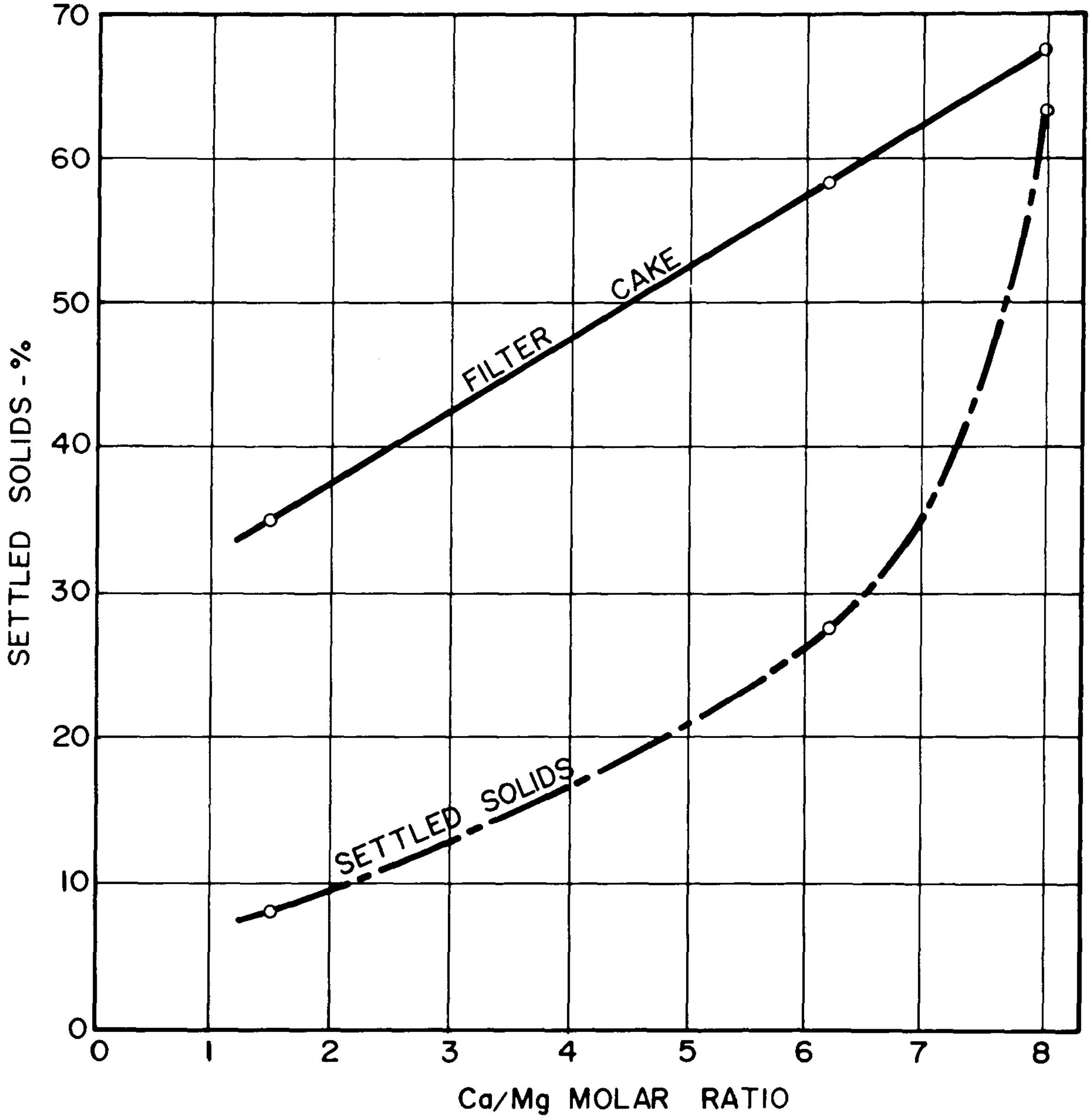


FIGURE 3-14

EFFECT OF Ca/Mg RATIO
ON SLUDGE SOLIDS CONCENTRATION
FOR LIME SLUDGES

Reference: 3-8

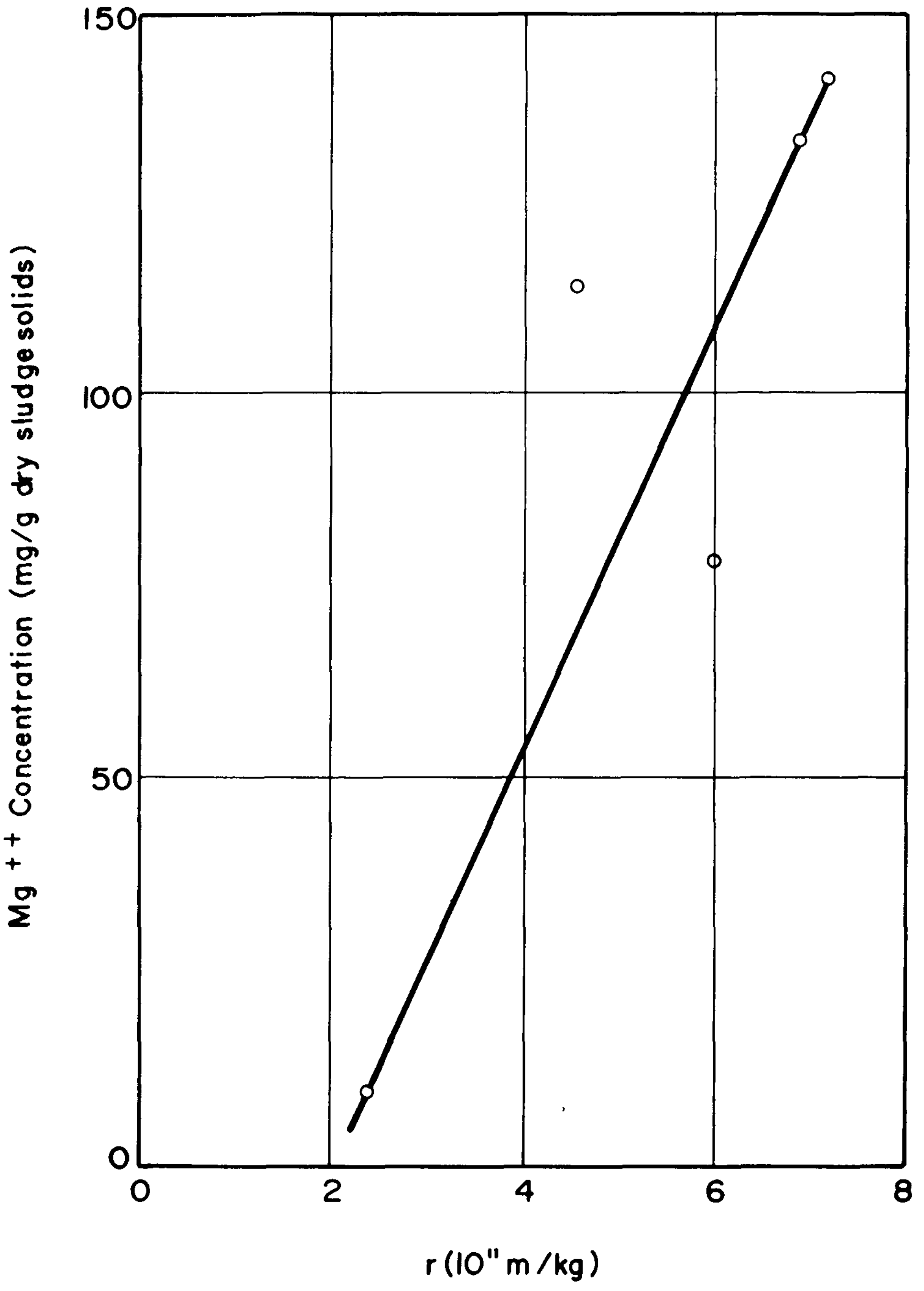


FIGURE 3-15

EFFECT OF MAGNESIUM CONCENTRATION ON THE SPECIFIC RESISTANCE OF SOFTENING SLUDGES

Reference: 3-8

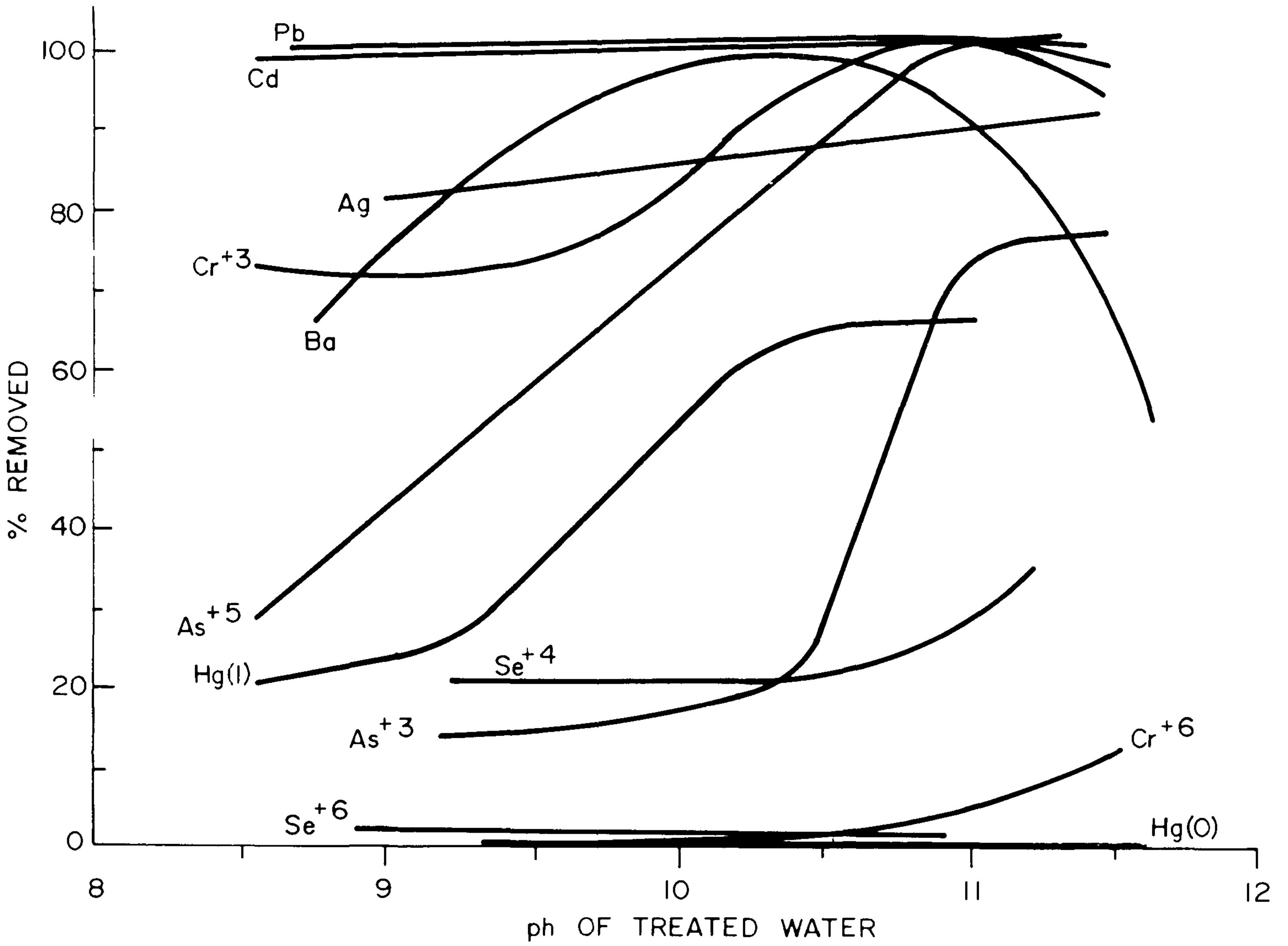


FIGURE 3-16

REMOVAL OF INORGANIC CONTAMINANTS BY LIME SOFTENING

Reference: 3-16

likely that most lime sludges will pass the EP Toxicity test procedure regardless of the concentration of constituent in the sludge. The EP Toxicity procedure is an extraction conducted at pH 5.5. However, the maximum amount of acetic acid to be used in the test procedure for pH reduction will generally not lower the pH of lime sludge due to its high buffer capacity and therefore metals will not leach.

Many groundwaters being treated for hardness removal also contain background concentrations of naturally occurring radium. Radium is a naturally occurring daughter product of U^{238}. Decay of U^{238} over millions of years passes through a series of elements eventually producing radium. The parent elements of radium are generally insoluble in water so that radium is often the first radioactive element which is found in drinking water supplies. Ra^{226} is the predominant species and has a half life of 1,500 years. Ra^{226} emits alpha and gamma rays, decaying to Rn^{222} (see Figure 3-17). Alpha, beta and gamma rays emitted from various intermediate products are all forms of ionizing radiation. In tissue, ionization can produce harmful cellular changes. Radon, with a half life of only 3.8 days, is a gas and thus will be released from water or sludge solids containing radium. The radon gas can then be inhaled by humans. The major threat to human health from radium daughters comes from breathing air containing radon and its very short lived daughters, which can accumulate as solids in the lungs. This exposes the lungs and other internal organs to continuous radiation. In addition the sludge can directly expose humans to gamma radiation from the decay of radium. However, for the most part the safe handling and disposal of radium containing wastes involves the prevention of radon exposure.

Measurement of radioactive components is expressed in curies or picocuries pCi (10^{-12} curies). A curie is the official unit of radioactivity, defined as exactly 3.70×10^{10} disintegrations per second. This decay rate is nearly

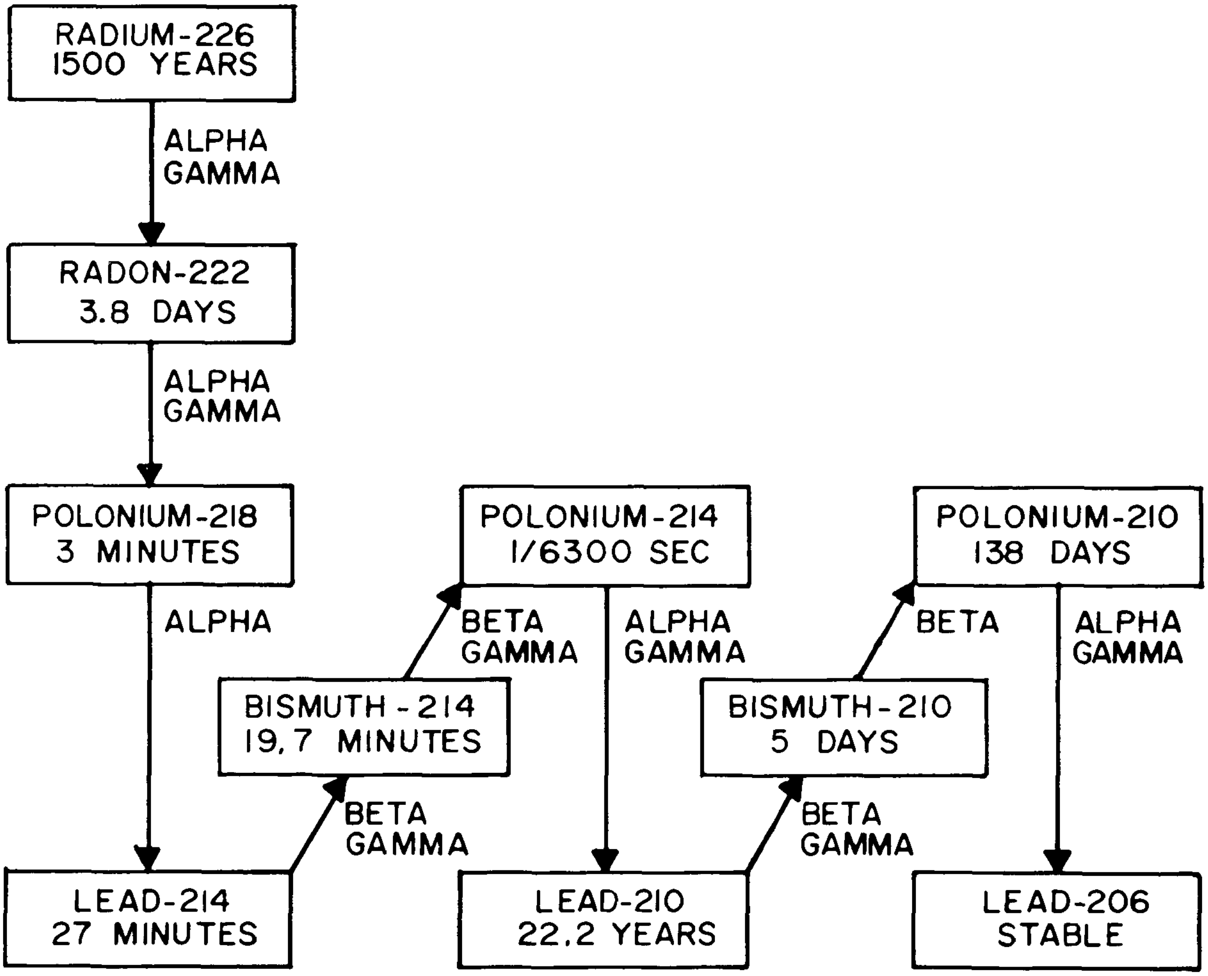


FIGURE 3-17

RADIUM-226 DECAY PRODUCTS

equivalent to that exhibited by one gram of radium in equilibrium with its disintegration products. About 0.0001 ml of radon per day at standard temperature and pressure is released from one gram of radium.

Radium is removed by lime softening increasingly as the percentage of hardness removed as shown in Figure 3-18[5]. Data reported by Snoeyink (3-4) for radium concentrations in various lime softening wastes are shown in Table 3-11. The sludge concentrations of Ra^{226} range from 1,000 to 11,000 pCi/l of sludge, Ra^{228} varies from 200 to 12,000 pCi/l. Since the radium is associated with the sludge solids, its concentration in the liquid stream is a function of the solids concentration. The concentration per gram of solids is 10 to 20 pCi/g for Ra^{226} and 1 to 11 PCi/g for Ra^{228}. Backwash water concentrations for Ra^{226} range from 6 to 50 pCi/l. Again, since this radium is associated with the solids it could be settled into a sludge waste. Snoeyink showed calculations to estimate the radium concentration in the sludge. The removal of radium would first be calculated by Figure 3-18 or determined by an actual mass balance within the water treatment plant. The equation presented in Section 3.2 for estimating quantities of lime sludge can then be used for total sludge production. The estimated pCi/g can then be determined. Table 3-12 shows estimates calculated by Snoeyink for four different wastes. The differences between measured and calculated are -37% to +37%. Considering the inaccuracy of obtaining a representative clarifier grab sample, and the assumption made for calculating theoretical sludge production, the differences are reasonable.

Granular activated carbon (GAC) can be used for removal of radon from waters. As shown in Figure 3-17 the radon collected on the GAC will quickly decay to the intermediate

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[5] Much of the discussion on inorganic wastes is taken from work by Snoeyink (3-4), whose assistance in providing data is greatly appreciated.

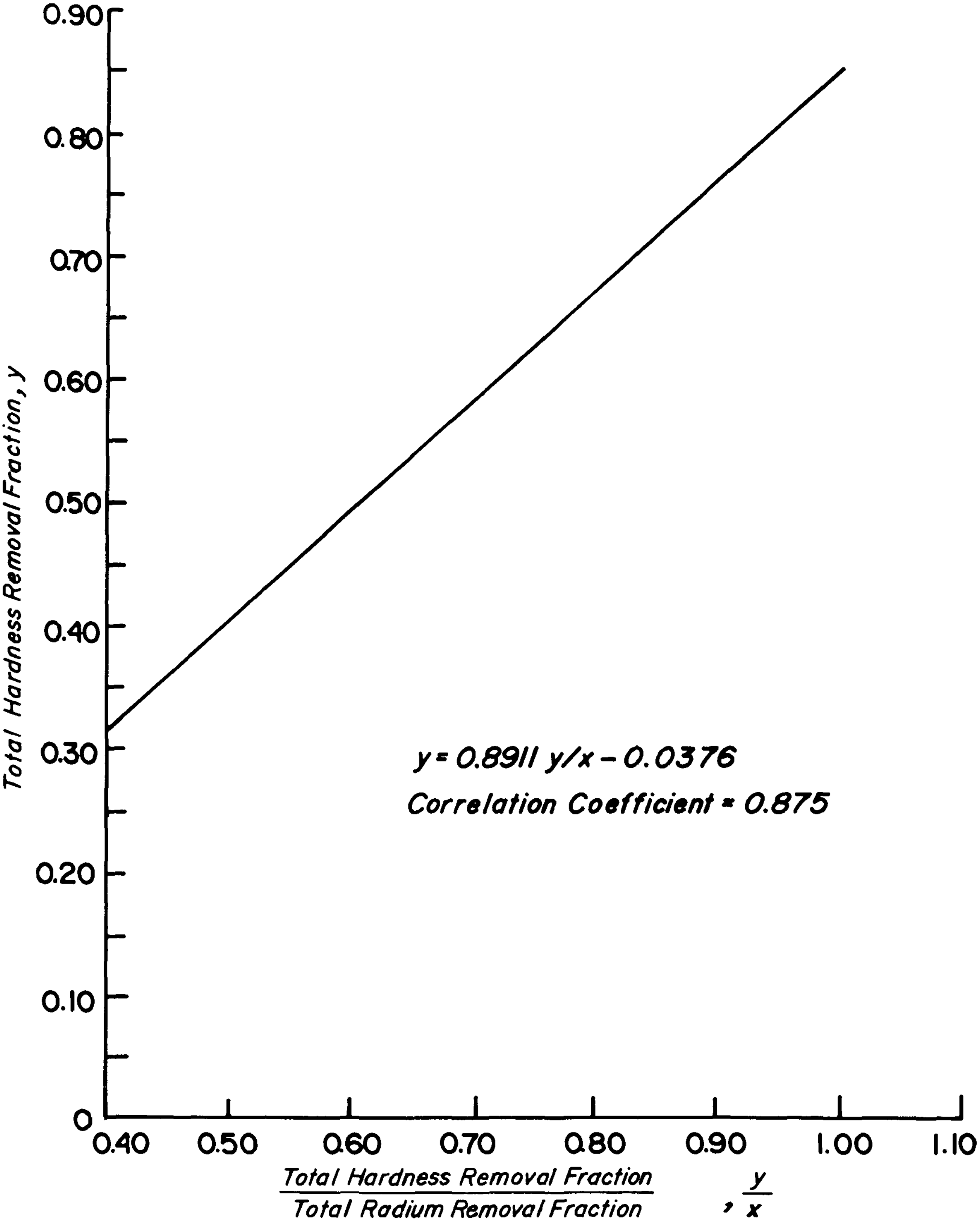


FIGURE 3-18

PROPOSED CORRELATION BETWEEN Ra 226 AND TOTAL HARDNESS REMOVAL FRACTIONS FOR LIME SOFTENING PLANTS

Reference: 3-4

TABLE 3-11

SUMMARY OF RADIUM CONCENTRATIONS IN LIME SOFTENING SLUDGES AND BACKWASH WATERS

LOCATION	% Solids	Ra^{226} (pCi/l)	Ra^{228} (pCi/l)	Ra^{226} pCi/g(dry)	Ra^{228} pCi/g(dry)
W. Des Moines, IA					
Lagoon Sludge	37.6	5,159	596	10.8	1.3
Clarifier Sludge*	1.6	<20	<40	<.02	<.04
Lagoon Sludge	NA	2,300	NA	NA	NA
Backwash Water	NA	6.3	NA	NA	NA
Bushnell, IL					
Clarifier Sludge	19	4,577	<45	21.6	<.21
Colchester, IL					
Clarifier Sludge	12.6	2,038	236	15.0	1.7
Backwash Water	0.23	<20	<39	NA	NA
Webster City, IA					
Sludge	NA	980	NA	NA	NA
Backwash Water	NA	50	NA	NA	NA
Peru, IL					
Backwash Water	NA	36.9	NA	NA	NA
Elgin, IL					
Active Lagoon Sludge	57.3	9,642	9,939	11.3	11.7
Inactive Lagoon Sludge	67.1	11,686	12,167	10.9	11.3
Clarifier Sludge	10.3	948	873	8.6	8.0
Backwash Water	0.051	<20	<40	<.02	<.04
Sludge*	NA	6,100	NA	NA	NA
Backwash Water	NA	18.3	NA	NA	NA

* Assume specific gravity = 1.0

Source: Snoeyink (3-4)

TABLE 3-12

THEORETICAL SLUDGE QUANTITIES AND THEORETICAL AND MEASURED Ra^{226} CONCENTRATIONS

Plant	Theoretical Sludge Production (Kg solids/day)	Theoretical Ra^{226} Concentration (pCi/g(dry))	Measured Ra^{226} Concentration* (pCi/g(dry))	% Diff.
W. Des Moines	3,955	10.0	10.8**	+8
Colchester	323	24.0	15*	-37.5
Bushness	712	15.8	21.6*	+36.7
Elgin	8,090	(-)	0.9*	NA

* Based on concentrations measured in the clarifier sludge

** Based on lagoon sludge

Source: Snoeyink (3-4)

product of lead-210, the first radon daughter with a significantly long half life. The lead-210 can collect to a concentration that the waste GAC must be handled as a low level radioactive waste. (See Section 2.1.2).

3.4.2. Liquid Phase Wastes

In Section 3.2 a method was presented for calculating the TDS of brine solutions resulting from ion exchange treatment. As was shown in those calculations the brine waste strength is very dependent upon the concentration of cations to be removed, the cation exchange capacity of the resin and the amount of regenerant used. Another important factor in comparing data is whether concentrations are reported as only in the regenerant waste itself or as diluted with rinse water and/or backwash water. Therefore, it is very difficult and probably misleading to discuss typical waste concentrations from IX plants. Each situation should be evaluated by the procedures discussed in the last section. Table 3-13 (3-4, 3-17, 3-18) shows a range of ion concentrations in liquid phase wastes resulting from ion exchange processes.

In addition to removal of hardness, ion exchange can be used for the removal of specific trace inorganic ions, such as radium and barium.

Singley (3-19) reported radium removal by ion exchange to range form 65% to 85% for plants with incomplete regeneration and 95% removal or better for plants practicing complete regeneration of their exchange resin. His survey of 8 ion exchange plants (4 in Iowa, 3 in Illinois and 1 in Florida) indicated radium removals of 84% to 97%, with raw water Ra^{226} concentrations ranging from 3.3 to 49 pCi/l. Brinck (3-20) found radium removals of 81% to 97% at seven ion exchange plants in Iowa and Illinois with raw water Ra^{226} concentrations ranging from 3.3 to 43 pCi/l. Schliekelman (3-18) performed an extensive study of Iowa water treatment plants and found ion exchange to remove 93% to 96% of Ra^{226} from raw

TABLE 3-13

EXAMPLE RANGES OF CHEMICAL CONSTITUENTS OF ION EXCHANGE WASTEWATER

Constituents (mg/l)	AWWA Report (3-17) Range of Averages	AWWA Report (3-17) One Plant's Data	Schiekelman (3-18) Range of Averages
TDS	15,000 - 35,000	15,656	54,000 - 130,000
Ca^{2+}	3,000 - 6,000	1,720	NA
Mg^{2+}	1,000 - 2,000	600	NA
Hardness (as $CaCO_3$)	NA	7,762	16,000 - 39,000
Na^+	2,000 - 5,000	3,325	NA
Cl^-	9,000 - 22,000	9,600	9,500 - 120,000

Source: Snoeyink (3-4)

waters containing 5.7 to 49 pCi/l. The exact concentration of Radium in the waste will vary significantly from plant to plant depending upon the regeneration practices as discussed previously. Generally, for a given radium concentration, as the hardness decreases the radium concentration in the waste increases. This is because less waste volume per kg of hardness removed is produced as the hardness decreases. Using the procedures of Section 3.2 it would be possible to estimate the average concentration of radium in the waste. One would have to compute this average based on the volume of brine plus rinse water since the waste radium would be distributed between these two volumes. Snoeyink reported data on the release of radium during the regeneration process from one plant as shown in Figure 3-19. Although during this regeneration the plant was not using sufficient salt for complete regeneration, it does show the profile and indicates that the majority of the radium was released during the rinse phase for this particular operation. Table 3-14 (3-18) shows data on radium average and peak concentrations in the ion exchange waste from four plants in Iowa.

The same factors that govern radium concentrations in the spent brine apply to barium in ion exchange wastes. Two ion exchange softening plants located in Crystal Lake, Illinois treat groundwater containing 9.5 mg/l and 16.1 mg/l Ba^{2+} at wells #6 and #8, respectively (Figures 3-20 and 3-21). At well #8, 153% of the barium removed by the ion exchange column during the service cycle was found in the regenerant wastewater (more than 100% barium removal from the column is possible if the previous regeneration was incomplete). The distribution of this barium was 1.8% in the backwash water, 3% in the brine, and 148% in the rinse water. The 10,000 gallons of wastewater produced from one regeneration cycle contained 30 kg of barium (800 mg/l).

At well #6, 84% of the barium placed on the column was removed during regeneration. Almost two percent was removed

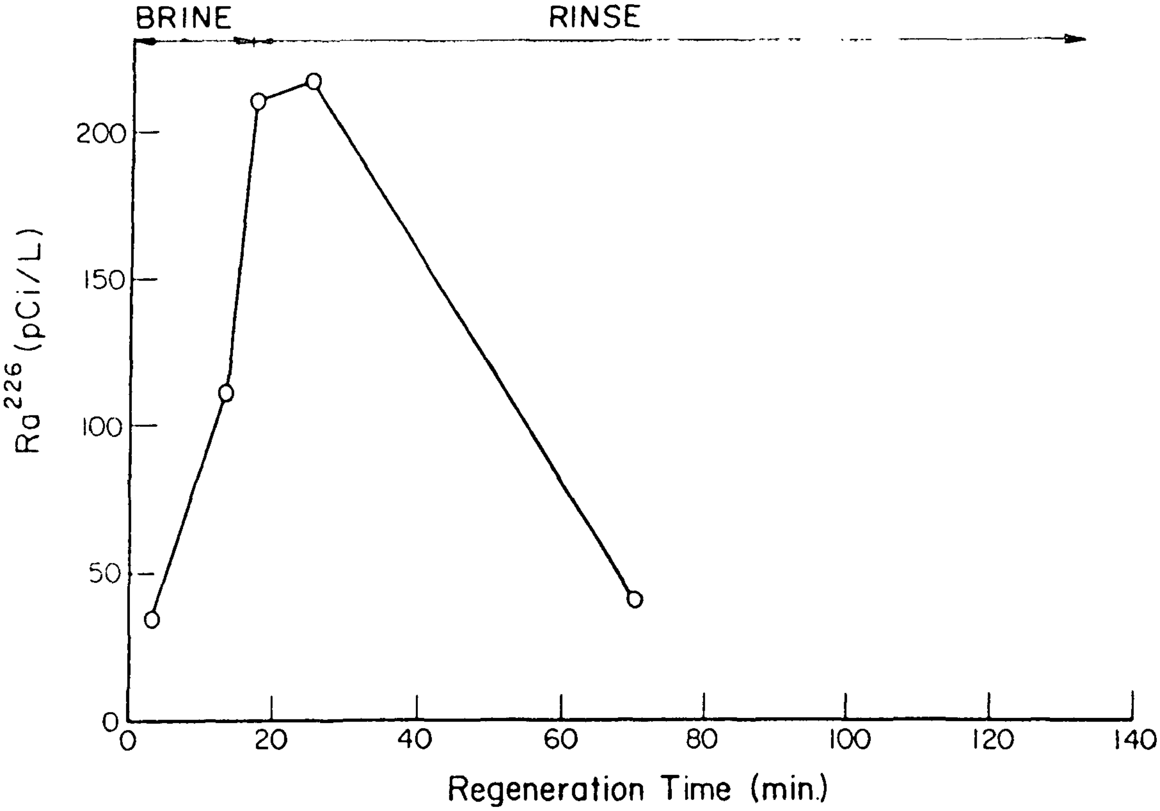


FIGURE 3-19

VARIATION OF Ra^{226} CONCENTRATIONS DURING REGENERATION (10-15% NaCl BRINE) ELDON, IA.

Reference: 3-4

TABLE 3-14

Ra^{226} CONCENTRATIONS IN ION EXCHANGE TREATMENT PLANT WASTEWATER

LOCATION	Average Brine + Rinse (pCi/l)	Average Brine + Rinse + Backwash (pCi/l)	Average for Peak 1/4 - 1/3 of Regeneration Cycle (pCi/l)	Peak Concentration in Wastewater (pCi/l)	Raw Water Concentration (pCi/l)
Eldon, IA	530	420	2,000	3,500	46
Estherville, IA	NA	52	114	320	5
Grinnell, IA	110	NA	260	320	6
Holstein, IA	175	NA	576	1,100	13

Source: Schliekelman (3-18)

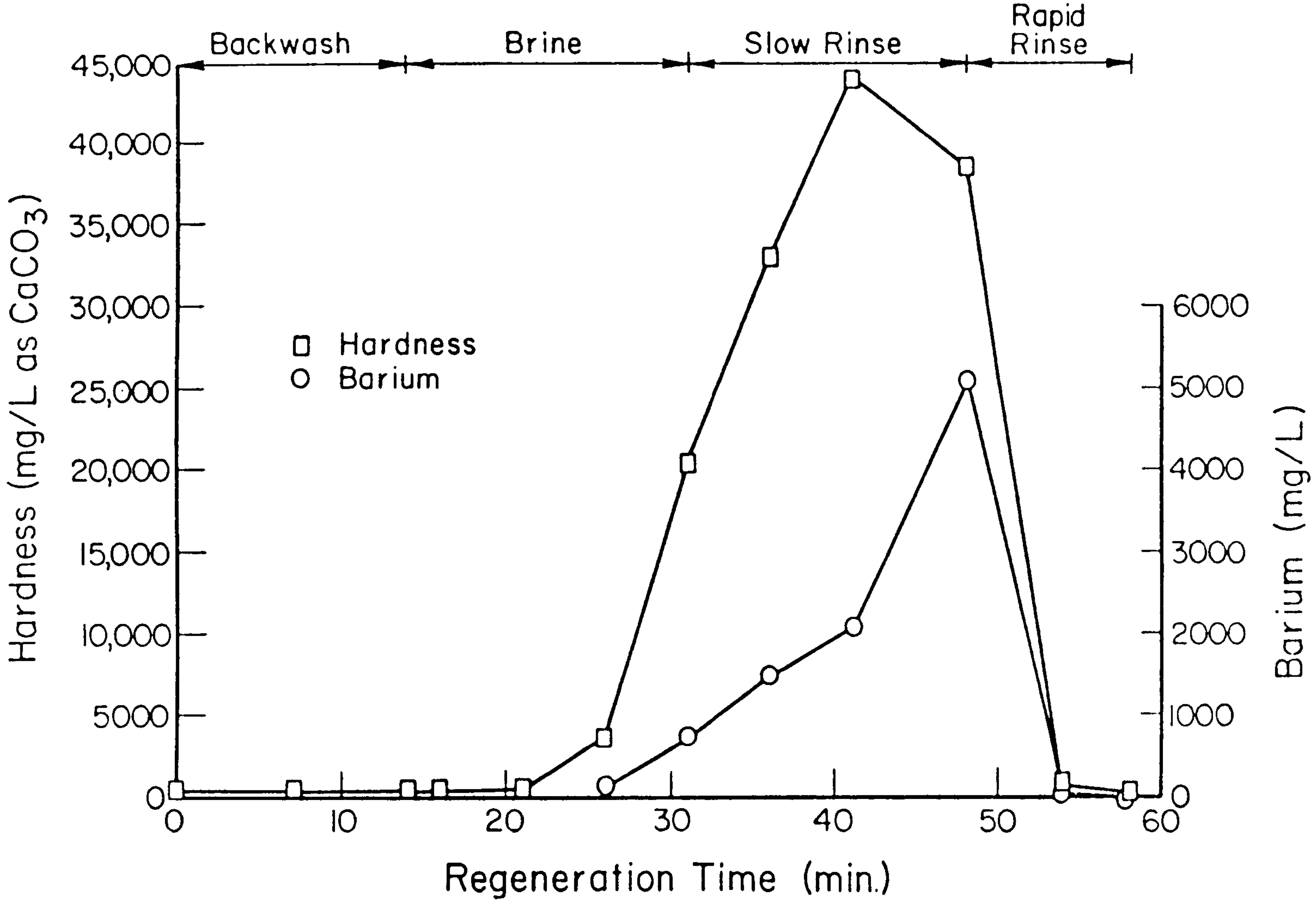


FIGURE 3-20

BARIUM AND HARDNESS CONCENTRATIONS IN SPENT BRINE
WELL 8, CRYSTAL LAKE, IL.

Reference: 3-4

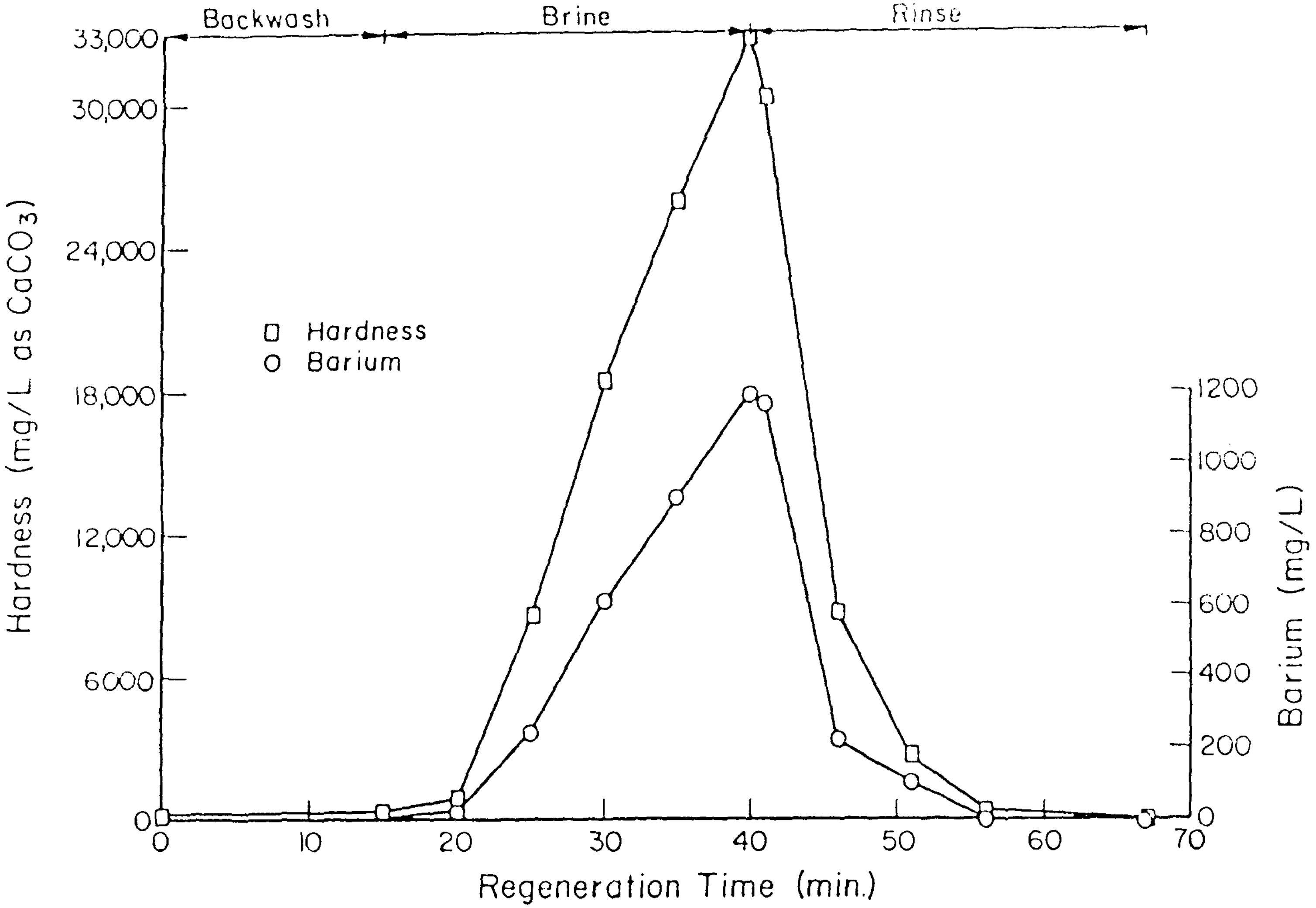


FIGURE 3-21

BARIUM AND HARDNESS CONCENTRATIONS IN SPENT BRINE WELL 6, CRYSTAL LAKE, IL.

Reference: 3-4

during backwash, 45% during the brine application and 37% during the rinse cycle. Six and one-half kg of barium were contained in the 9,500 gallons of wastewater generated from one regeneration cycle (180 mg/l) (3-4).

Sorg and Logsdon (3-21) found that the regeneration of the ion exchange resin in their study removed 85% of the barium placed on the column. Eighteen kg of barium was contained in the 9,250 gallons of wastewater generated. The average wastewater barium concentration in this plant was 534 mg/l, with a peak concentration of 6,000 mg/l Ba. The wastewater volume was 2.6% of the product water.

Barium concentrations in the water and wastewater at Crystal Lake are given in Table 3-15. Figures 3-20 and 3-21 show hardness and barium concentrations during regeneration for wells #6 and #8. By comparing these figures one can see that the peak hardness and barium concentrations coincide at well #6, but the peak hardness value occurs before the peak barium concentrations at well #8. This phenomenon may be caused by the greater amount of the barium on the column before regeneration at well #8 (78.4 g Ba/ft^3) compared to that at well #6 (40.4 g Ba/ft^3). More barium accumulates on the resin during the service cycle at well #8 because of a higher raw water barium concentration, and less regenerant is applied to the column per unit volume of water produced during the service cycle (3-4).

Ion exchange with strong base resins can be used to remove nitrate from water. Sodium chloride is used to regenerate the resin, so the spent brines are high in Na^+ and Cl^- as well as NO_3 and other anions that are removed from the water by the resin, such as sulfate. Some data from McFarland, California are shown in Figure 3-22 and indicate spent brine concentrations of SO_4^- as high as 30,000 mg/l and NO_3 as high as 6,000 mg/l (3-22 from ref. 3-4). Total waste volumes and average concentrations of the various ions in the total waste were not available.

TABLE 3-15

CRYSTAL LAKE'S WATER AND WASTEWATER BARIUM CONCENTRATIONS

	Ba in Raw Water (mg/l)	Peak Ba During Regeneration (mg/l)	Average Concentration in Brine and Rinse Water (mg/l)	Average Concentration in Backwash, Brine and Rinse Water (mg/l)
Well #6	9.5	1,197	328	153
Well #8	16.1	5,161	1,297	789

Source: Snoeyink (3-4)

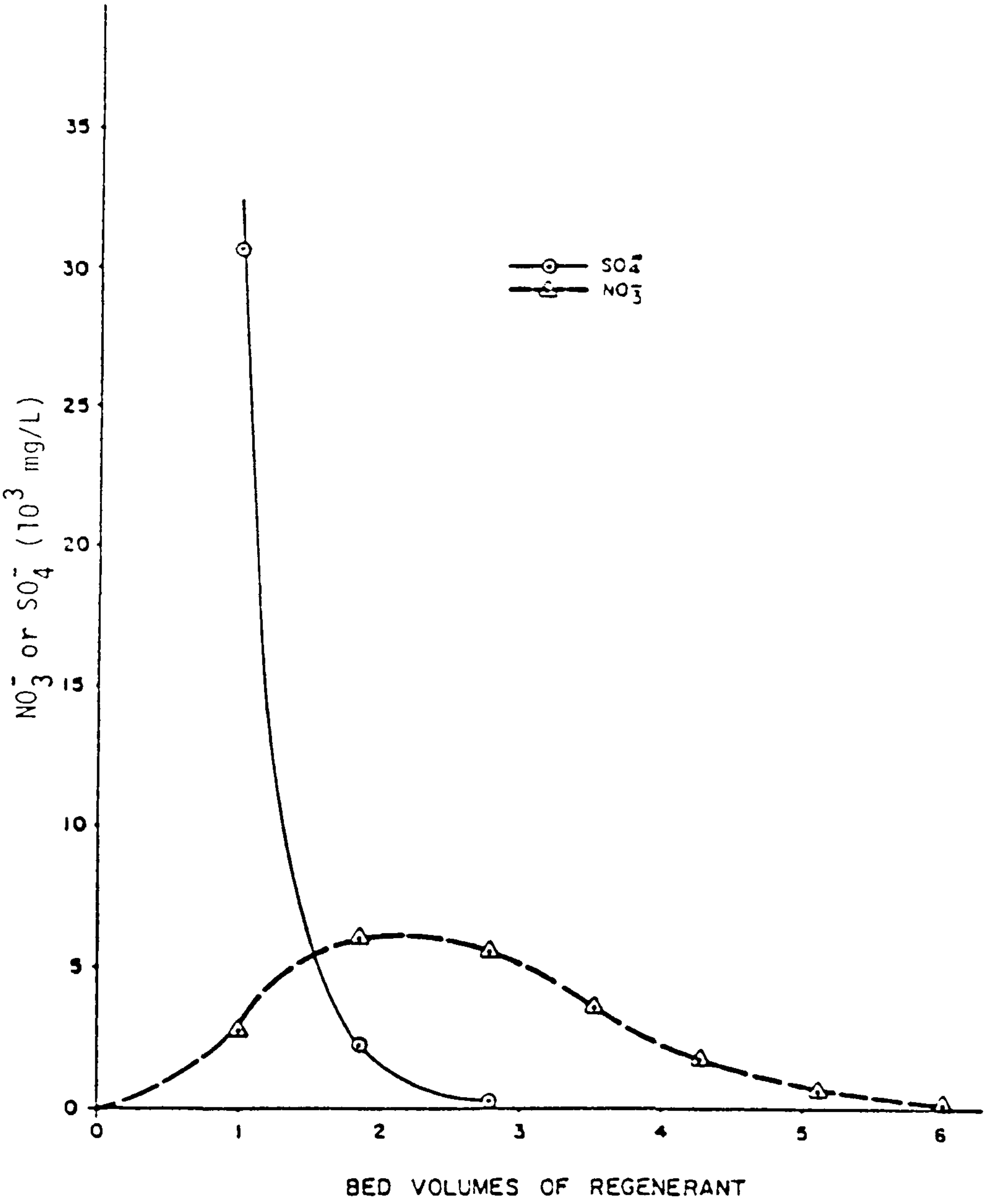


FIGURE 3-22

SULFATE AND NITRATE IN
6% BRINE DISCHARGE FROM SYSTEM

Reference: 3-4

Activated alumina is a solid absorbent which has good selectivity for fluoride, arsenic and selenium. The Palo Verde, AZ, defluoridation plant treats groundwater in three down-flow activated alumina pressure vessels. Treatment runs during the sampling period were terminated at an effluent concentration of 2 mg/l F^- (3-23 from ref. 3-4). Volumes and flow rates used in each of the regeneration steps, as well as contaminant concentrations, are detailed in Table 3-16. NaOH (2%) was used for regeneration. This plant produced 0.025 gallons of wastewater per gallon of water treated. The greatest concentration of F^- appeared in the upflow rinse.

A similar defluoridation treatment plant is operated at Gila Bend, AZ (3-23). Table 3-17 presents the volumes and flow rates used in each of the regeneration steps, along with the contaminant concentrations. The regenerations (both upflow and downflow) are done with 0.8 to 1% NaOH. The neutralization of the media is attained by using raw water adjusted to a pH of 3.2 with H_2SO_4. For one treatment run, 0.047 gallons of wastewater were produced per gallon of water treated.

In the last section it was pointed out that RO water streams are typically 25% to 35% of the feed water flow. Therefore, the concentration of TDS in the waste is less than for ion exchange plants.

Reverse osmosis (RO) systems are especially favorable for the treatment of water supplies with several contaminants that otherwise would require a combination of treatment methods. An RO system can be used to remove both anions and cations, as well as organic compounds, to produce an overall better quality of water. A good first approximation of the concentration of inorganic substances in RO reject water is:

$$\frac{\text{Concentration in raw water}}{\text{Fraction of water rejected}} = \text{Concentration in reject water}$$

TABLE 3-16

WASTEWATER CHARACTERISTICS AT PALO VERDE, AZ

	F^- mg/l	Al mg/l	As mg/l	$SO_4^{=}$ mg/l	Na mg/l	TDS mg/l	Mg mg/l	Turbidity NTU	Ca mg/l	pH
Backwash (760 gal; 10 gpm/ft^2)	1.8	230	0.210	220	225	880	2.8	17	53	5.8
Upflow Regeneration (150 gal; 2.4 gpm/ft^2)	3.2	13	0.017	184	750	1344	4.3	6.3	30	11.2
Upflow Rinse (1050 gal; 3.8 gpm/ft^2)	250	150	0.140	640	740	2629	1.8	20	6.9	10
Downflow Regeneration (150 gal; 2.4 gpm/ft^2)	23.5	11	0.030	156	410	1264	0.2	1.0	7.5	11.2
Neutralization (900 gal; 3.9 gpm/ft^2)	8.4	150	0.475	580	1240	2879	2.9	18	6.8	11.6
Wastewater Composite (mg/l)	91.5	156.4	0.246	469.2	743.5	2130	2.4	17	19.7	--

Source: Anderson (3-23) from (3-4)

TABLE 3-17

WASTEWATER CHARACTERISTICS AT GILA BEND, AZ

	F^- mg/l	Al mg/l	As mg/l	$SO_4^=$ mg/l	Na mg/l	TDS mg/l	Cl^- mg/l	Mg mg/l	Turbidity NTU	Silica mg/l	Ca mg/l	Cr mg/l	Fe mg/l
Backwash (10,000 gal; 400 gpm)	5.4	2.1	<.001	219	438	1278	590	0.4	300.0	9.5	33.0	0.013	14.5
Upflow Regeneration (4,800 gal; 200 gpm)	5.1	0.42	<.001	206	444	1310	590	0.4	140.0	7.7	32.0	0.016	NA
Rinse "A" (2,100 gal; 200 gpm)	6.1	0.22	0.076	739	741	2338	700	0.3	45.0	NA	30.5	0.026	NA
Rinse "B" (25,000 gal; 400 gpm)	192	193.3	0.49	721.5	1319	3704	605	<0.1	105.5	5.2	2.0	0.031	NA
Downflow Regeneration (2,700 gal; 200 gpm)	33.4	8.0	0.050	187	528	1453	580	NA	1.0	2.3	4.7	0.021	NA
Neutralization (85,700 gal; 400 gpm)	5.6	20.1	0.054	162.6	454	1380	570	<0.1	0.38	3.3	1.4	0.010	NA
Regenerant Composite	41.9	50.7	<0.13	285	624	1832	581	NA	49	NA	5.6	0.015	NA
Raw Water	5.0	0.08	0.014	174	432	1260	570	0.4	0.5	10.6	35.0	0.017	0.11

Source: Anderson (3-23) from (3-4)

For example, if there is 5 mg/l of F^- in the raw water and 25% reject water, the concentration in the waste should be:

$$\frac{5\ \text{mg/l}}{0.25} = 20\ \text{mg/l}$$

This calculation assumes 100% rejection of salts by the membrane; however, rejection is usually somewhat less than 100%, so the concentration in the reject water accordingly is somewhat less. A 400 psig RO system (high pressure) typically will reject from 90 to 95% of the TDS (3-4).

Data for fluoride removal are shown in Table 3-18. The concentration of F^- in the reject water is approximated fairly well by the above equation. Further, the process does an adequate job of reducing the F^- concentration in the product water to an acceptable level.

Similarly, data for arsenic removal by RO are given in Table 3-19, and nitrate removal in Table 3-20. The mass balance equation yields a reasonable approximation of arsenic concentrations in the reject water, but the NO_3 is poorly rejected by the cellulose acetate membrane and thus the approximation is not as good. The product water contains 31.9 mg/l NO_3 which is about one-third of the influent concentration.

TABLE 3-18

FLUORIDE REMOVAL BY REVERSE OSMOSIS

LOCATION	F^- Feed Concentration mg/l	F^- Product Concentration mg/l	F^- Reject Concentration mg/l	% Recovery	Type of System
Laboratory Study - Toray Membrane	6.6	0.62	24.54	75^1	small single element RO system
Laboratory Study - Filmtec Membrane	8.82	1.41	31.04	75^1	small single element RO system
Harbour Heights, FL					
Day: 1-5	4.72	0.31	14.08	74.4	two-stage system with spiral wound cellulose acetate membranes
6	7.06	0.60	21.85	74.4	
7-8	8.84	0.86	27.10	74.4	
9-10	12.56	0.94	37.36	74.4	
Sorrento Shores, FL^2	2.2	0.8	2.8	39	spiral wound, 425 psi, one-stage system
Venice, FL^2	2.2	0.8	4.0	54	hollow fiber, 400 psi, two-stage system
Bay Front TP, FL^2	0.4	0.1	0.5	28	hollow fiber, 200 psi, one-stage system
Spanish Lakes MHP, FL^2	0.4	0.2	0.7	31	spiral wound, 400 psi, two-stage system
Sarasota Bay MHP, FL^2	0.5	0.2	0.6	50	hollow fiber, 400 psi, one-stage system
Nokomis School, FL^2	0.65	<0.1	0.8	35 (design)	spiral wound, 200 psi, one-stage system

Source: Snoeyink (3-4)

1 - Values were adjusted to yield 75% recovery
2 - Full Scale operating facilities

TABLE 3-19

ARSENIC REMOVAL BY REVERSE OSMOSIS

LOCATION	Raw Water Concentration mg/l	Reject Water Concentration mg/l	Product Water Concentration mg/l	Reject Volume (gal/10,000) gal. treated
Harbour Heights, Fl				
Day: 26	4.3	13.6	1.55	2500
27	1.14	3.4	0.39	2500
29-30	2.12	5.5	0.64	2500
32-35	2.59	7.2	0.92	2500
36	2.80	7.0	0.76	2500
47-48	0.54	1.2	0.05	2500
50	1.00	3.4	0.05	2500
51	5.31	17.4	.08	2500
Laboratory Study - Toray Membrane	0.329	0.968	0.116	2500
Laboratory Study - Filmtec Membrane	0.057	0.175	0.018	2500
Sorrento Shores, FL	0.095	0.008	0.190	6100
Venice, FL	0.015	<0.005	0.025	4600
Bay Front TP, FL	0.008	<0.005	0.016	7200
Spanish Lakes MHP, FL	0.030	<0.005	0.075	6900
Sarasota Bay MHP, FL	0.020	<0.005	0.130	4200
Nokomis School, FL	0.015	<0.005	0.012	6900

Source: Snoeyink (3-4)

TABLE 3-20

NITRATE REMOVAL BY REVERSE OSMOSIS

Constituents (Well #3)	Pretreated Feed (mg/l)	Product (mg/l)	Brine (mg/l)
Ca	154	7.0	590
Mg	3.8	0.17	15
Na	92	11	345
K	3.6	0.5	12.8
CO_3^{2-}	0.0	0.0	0.0
HCO_3^-	7.80	5.2	45.9
Cl^-	92.8	6.0	346.9
SO_4^{2-}	380	5.0	1500
NO_3^-	93.0	31.9	270.2
F^-	0.06	0.03	0.13
Fe	<0.05	<0.05	0.08
Mn	<0.01	<0.01	0.01
As	<0.01	<0.01	<0.01
Cu	<0.01	<0.01	0.02
Zn	0.01	0.07	0.2
Hardness (as $CaCO_3$)	401.0	18.2	1538
Total Solids	823	64	3120
pH	5.2	5.6	5.9

Product Flow Rate = 13.0 gpm
Feed Pressure = 390 psig
Product Recovery = 74.3%
Spiral-Wound Cellulose Acetate Membrane
Wastewater = 38,700 gpd when operation is continuous.

Source: Guter (3-22) from ref. (3-4)

CHAPTER 4

SOLID/LIQUID WASTE TREATMENT

4.1 OVERVIEW OF PROCESSES AND APPLICATIONS

The treatment of solid/liquid wastes produced in water treatment processes involves the separation of the water from the solid constituents to the degree necessary for the selected disposal method. Therefore, the required degree of treatment is a direct function of the ultimate disposal method (see Chapter 2).

Water treatment sludges from a chemical coagulation process typically have a 0.5 to 2.0 percent solids concentration. These solids are difficult to gravity thicken to greater than a 3 to 4 percent solids concentration. Sludges resulting from lime softening can be removed from settling basins at solids concentrations as high as 10 percent and may gravity thicken to a 30 solids concentration.

There are several sludge treatment methodologies which have been practiced in the water industry. Figure 4-1 shows the most common sludge handling options available, listed by general categories of thickening, dewatering and disposal. In choosing a combination of possible treatment process trains, it is probably best to first identify the available disposal options and their requirements for a final cake solids concentrations. Most landfill applications will require a 'handleable' sludge and this may limit the type of dewatering devices which are acceptable. Methods and costs of transportation may affect the decision as to 'how dry is dry enough'. The criteria should not be to simply reach a given solids concentration but rather to reach a solids concentration of desired properties for the handling, transport and disposal options available.

Table 4-1 shows a generalized range of results which have been obtained for final solids concentrations from different dewatering devices for coagulant and lime sludges.

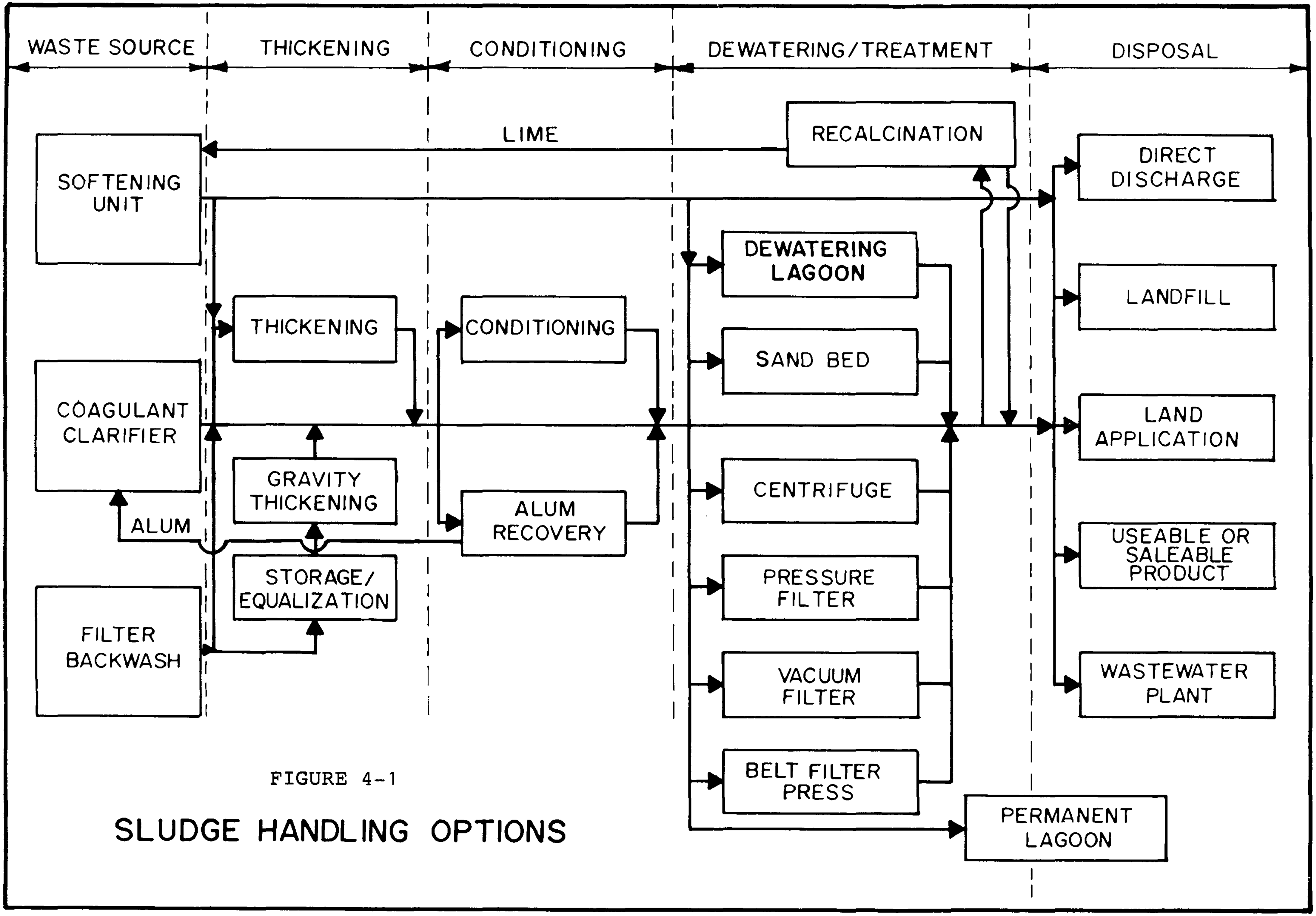


FIGURE 4-1

SLUDGE HANDLING OPTIONS

TABLE 4-1

RANGE OF CAKE SOLID CONCENTRATIONS OBTAINABLE

	% Solids Concentration	
	Lime Sludge	Coagulant Sludge
Gravity Thickening	15 - 30	3 - 4
Basket Centrifuge		10 - 15
Scroll Centrifuge	55 - 65	10 - 15
Belt Filter Press		10 - 15
Vacuum Filter	45 - 65	N/A
Pressure Filter	55 - 70	35 - 45
Sand Drying Beds	50	20 - 25
Storage Lagoons	50 - 60	7 - 15

4.2. OVERVIEW OF PILOT STUDIES

The best way to develop design data for sludge dewatering systems is to conduct on-site pilot tests. As applicable in each of the following treatment sections, specific pilot methods are discussed. In this section, an overview is presented of the considerations that must go into a pilot study for water plant waste treatment.

There are certain decisions that must be made prior to determining if a pilot study is needed (Figure 4-2). A first step is to ask what kind of sludge it is, how much there is and what are the available treatment options. Most of these steps could be pilot tested, although some are difficult to test on a small scale. One of the inherent difficulties in testing sludge options is that the pilot facility may require several other treatment steps in order to pilot one option. For example, a pressure filtration pilot plant may require that thickening and conditioning facilities be built on a pilot scale also. Land application of dried lime sludge cannot be tested without dewatering first, etc.

There are four reasons for conducting a pilot study. Studies should be considered if, 1) the technical feasibility is unknown; 2) the economic feasibility is unknown; 3) there may be a less expensive alternative available, or; 4) some process refinement or development of design data is needed. The technical feasibility of a process may be unknown if the method has not been utilized on large scale applications or not on wastes similar in nature to this particular type of water plant sludge. In this case, the pilot emphasis would be on several pertinent factors, such as: does it meet the treatment objectives, what variables affect performance, how much will it cost, what are the design and scale up factors. A technical feasibility pilot study can be quite involved and requires both research and engineering approaches.

The economic feasibility of a process may be unknown if it is technically proven elsewhere, but the factors affecting

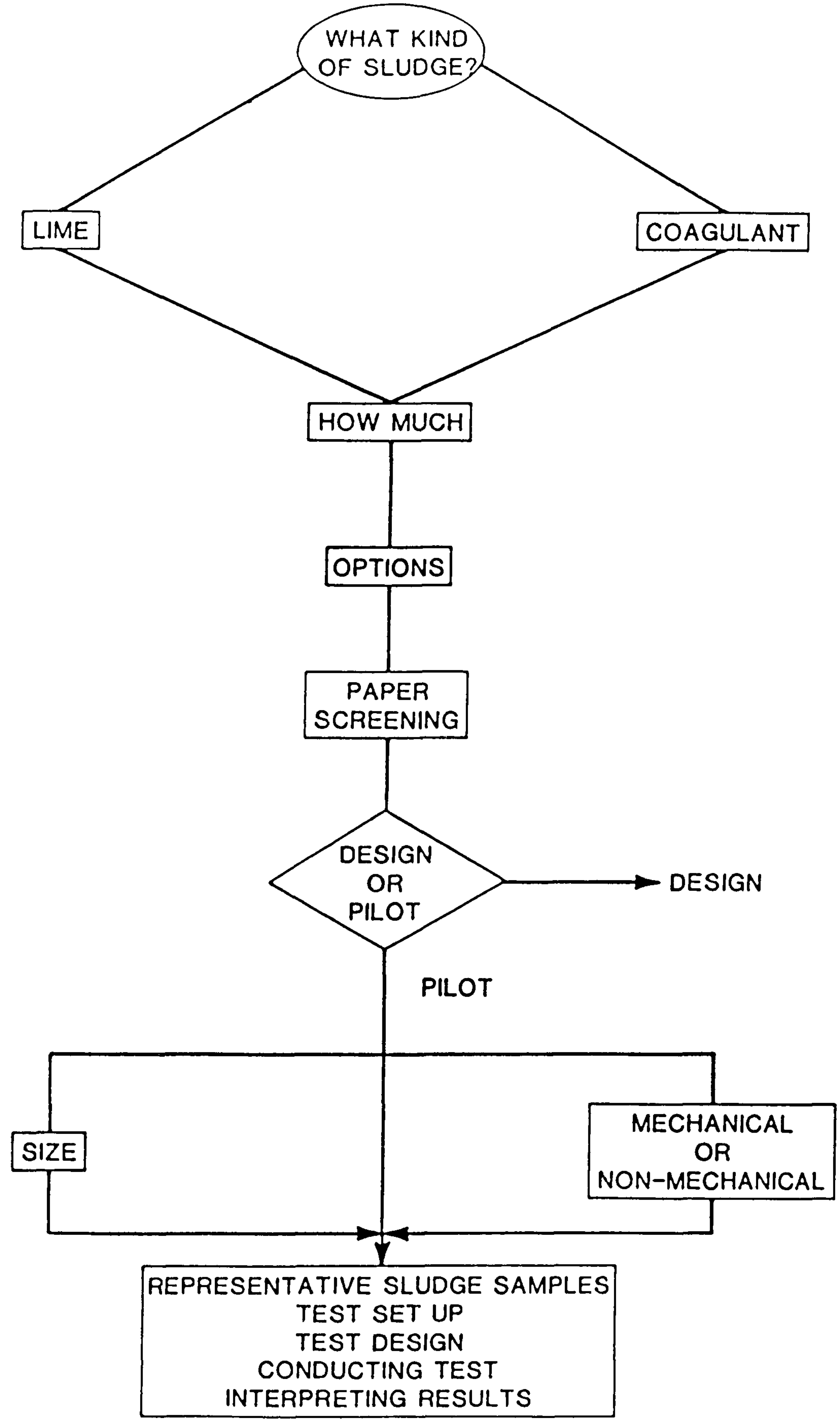


FIGURE 4-2

PILOT STUDY LOGIC DIAGRAM

cost for the particular application are not well defined. The emphasis of this type of study would be to vary as many factors as possible and to secure good operating cost data, such as operating time, chemicals and maintenance requirements. Variations in the sludge feed concentration or characteristics may also be tested to determine the effect on cost.

If it is known that a given alternative will work, but several options within the alternative are available (eg. different manufacturers) then a side-by-side study may be needed. Comparison studies require that as many variables as possible be held constant except for the process or equipment differences. The sludge feed should be the same to each and operating variables matched so that a true difference in performance can be evaluated.

Process refinement studies can be justified if either the design parameters need definition in order to ensure proper performance, or parameters affecting cost for final process evaluations are needed. These studies can also be very useful in optimization of an existing treatment process.

By assessing the alternatives in light of the above guidelines, a decision can be made as to whether a pilot study is warranted. Assuming the answer is affirmative a decision regarding the size of the study is needed. The choice is to conduct bench scale studies and/or on-site, larger scale pilot studies.

The terms bench and pilot studies have the implied meaning of size. While certainly size is generally associated with the difference between bench and pilot studies there really is a more important distinction. Pilot studies are of a size sufficient to produce reliable data for the design and costing of a full-scale installation, whereas bench studies do not provide that information. Bench studies have the advantages of allowing quick screening of various alternatives and testing of several variables on performance at relatively low costs. They do not provide reliable scale-up data and only limited design and cost information. Examples of very useful bench studies include Buchner funnel resistivity tests, filter leaf tests, capillary suction time tests, manufacturer dewatering equipment evaluations and conditioning studies (see Chapter 3).

In most cases the largest costs associated with pilot tests are labor and analytical testing and therefore sufficient funds should be appropriated for the labor and lab analysis necessary to gain the needed data. In addition to high costs, pilot studies often require large volumes of sludge feed material which may need special handling. Several weeks or even months may be needed to evaluate many variables. Examples of pilot studies for sludge treatment would include non-mechanical drying operations, such as a drying bed, manufacturer tests of mechanical drying equipment, and specialized studies such as coagulant recovery.

If the difference between a bench and pilot study is one of size required to provide design data, then the question of "how big is big enough" must be addressed. In some cases a small lab system may be large enough while in other cases even a major project would not answer all the design questions. Each type of process must be addressed on its own merit. As an example, in exploring tests of dewatering equipment each manufacturer should be consulted as to the type of equipment available for testing, the costs of rental for each size and their scale up factors. Some pilot dewatering facilities are production units and the manufacturer can use a 1:1 scale-up factor, whereas other units may need scale up factors of up to 2.5:1. Still other units may not have any scale-up factors since they are not prototype units. The utility should be aware of these differences and the potential impact on full scale costs when choosing pilot equipment.

A similar situation exists in non-mechanical treatment systems. A 10 ft x 10 ft sand drying bed study can give a great deal of information on loading rates and drying characteristics, but it is not large enough to evaluate potential operating problems such as non-uniform sludge application.

One of the difficulties in running bench or pilot studies on water plant sludges is obtaining representative sludge samples. Obviously, if the sample tested does not "represent" the sludge to ultimately be treated or does not reflect seasonal variations then the pilot study may not provide reliable results. The use of the methods presented in Chapter 3 can help assess if the sludge sample is representative. By knowing the alum dose, raw water suspended solids concentration and the solids concentration of the sludge sample, it is possible to estimate the aluminum content of the sludge. Measurement of the aluminum can then be used to check the sludge sample. The problem of obtaining a good sample is generally more acute with manually cleaned basins than with continuous withdrawal. With manually cleaned basins the sludge will often stratify with heavier solids on the bottom and lighter solids (with a higher aluminum hydroxide content) on the top. Similarly, for softening plants the calcium carbonate may be on the bottom and the magnesium hydroxide on the top. With a manual basin small samples can be collected by sampling at several depths and then blending the samples. For a pilot study it is probably best to collect a uniform sample in an external holding tank, such as a simple plastic swimming pool.

If the full scale installation is to include sludge thickening then provisions must be made to test various solids concentrations during the study. This can be accomplished by a batch fill and draw operation if other methods are not available. Of course, it would also be a good idea to collect information for the design of the thickener itself.

During the test it is easy to overlook important items. Pre-planning, creating good record logs and establishing a system of check points is paramount to successful studies.

The testing program should be divided into at least two phases. First, the complete program is established, starting with formulating the study objectives and desired results followed by identification of the key variables and tests required for data development. A fairly rigid schedule is set for the first half of the study where as many variables as possible are evaluated, one at a time. At the half-way point a complete data analysis should take place. The data are reduced to graphs and models as appropriate and attempts are made to draw the desired conclusions. The course of the testing program is often altered by this review process. In the situation that expensive equipment is being rented on a weekly basis, the review process may have to be scheduled for a weekend. This time to stop and assess the data is an integral part of any pilot program.

Also, during the testing program the engineering and water plant personnel should keep good records and observe what is going on in other parts of the water plant. It is very useful to keep a running log book in addition to any specific data sheets. In the log book is a sequential listing of operating conditions, unusual conditions, observations, starts and stops, etc. Many entries may seem unnecessary but no matter how detailed the information, there are always some gaps in the data or unexplained periods in operation.

The final aspect of any pilot study is interpretation of the results. A key aspect is to compare the results actually received to any projections that may be made as to what could have been accomplished. It is easy to say "we only got 15% solids <u>but</u> if we would have done something differently we could have gotten 25%." That may be true - but be careful and assign risks to any projections used in cost comparisons.

In summary, pilot studies of waste treatment can be valuable and are often required in the pre-design phase of the planning process. Their use should follow as much pre-screening as possible to eliminate non-viable alternatives. When pilot studies are used, proper definition of data needs is necessary, and if they are to provide design data then proper sizing is very important. Properly designed and operated pilot programs can result in significant cost and performance savings to the utility.

4.3. SLUDGE THICKENING

4.3.1. Description

After removal from a clarifier or sedimentation basin most water sludges can be further thickened in a gravity concentration tank. Thickening can be economically attractive in that it reduces the sludge volume and results in a more concentrated sludge for further treatment in the dewatering process. Some dewatering systems will perform more efficiently with higher solids concentrations. Thickening tanks can also serve as equalization facilities to provide a uniform feed to the dewatering step.

Although there are a few types of thickeners available on the market, the water industry almost exclusively uses gravitational thickening.

Sludge thickening is performed primarily for reduction in the volume of sludge which will require subsequent treatment and disposal. The relationship between the volume of sludge and the solids concentration is expressed as

$$V = \frac{M}{rsP}$$

where,

V = volume of sludge (m^3)

M = mass of dry solids (kg)

r = density of water, $10^3 kg/m^3$ (at 5^oC)

s = specific gravity of the sludge

P = percent solids expressed as a decimal (weight/weight)

An approximation for determining volume reduction based on percent solids is expressed as

$$\frac{V_2}{V_1} = \frac{P_1}{P_2} \quad \text{(approximate volume ratio)}$$

This is a quick and useful equation since the specific gravity of the sludge is not always known. Therefore, for thickening a 1 percent solids concentration sludge to 10 percent solids concentration a volume reduction of approximately 90 percent is achieved.

Gravity sludge thickeners are generally circular settling basins with either a scraper mechanism in the bottom (see Figure 4-3), or equipped with sludge hoppers (Figure 4-4). They may be operated as continuous flow or as batch 'fill and draw' thickeners. For continuous flow thickeners, the sludge normally enters the thickener near the center of the basin and is distributed radially. The settled water exits the thickener over a peripheral weir or trough and the thickened sludge is drawn off the basin. For tanks equipped with a scraper mechanism, the scrapper is located at the thickener bottom and rotates slowly. This movement directs the sludge to the draw-off pipe near the bottom, center of the basin. The slow rotation of the scraper mechanism also prevents bridging of the sludge solids. The basin's bottom is sloped to the center to facilitate collection of the thickened sludge.

Batch fill and draw thickening tanks are often equipped with bottom hoppers as was shown in Figure 4-4. In these tanks sludge flows into the tank, usually from a batch removal of sludge from the sedimentation basin, until the thickening tank is full. The sludge is allowed to quiescently settle and a telescoping decant pipe is used to remove supernatant. The

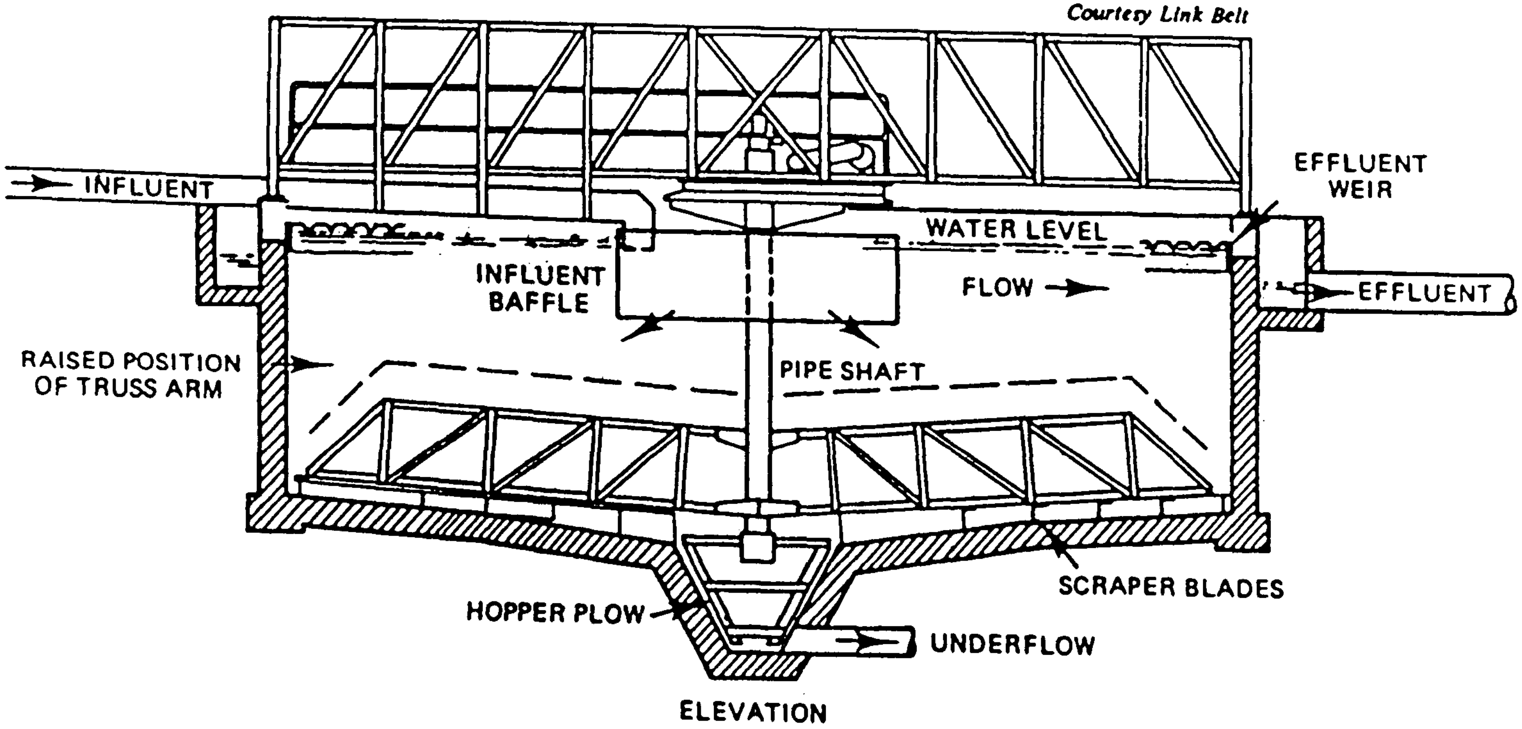


FIGURE 4-3

CONTINUOUS FLOW GRAVITY THICKENER

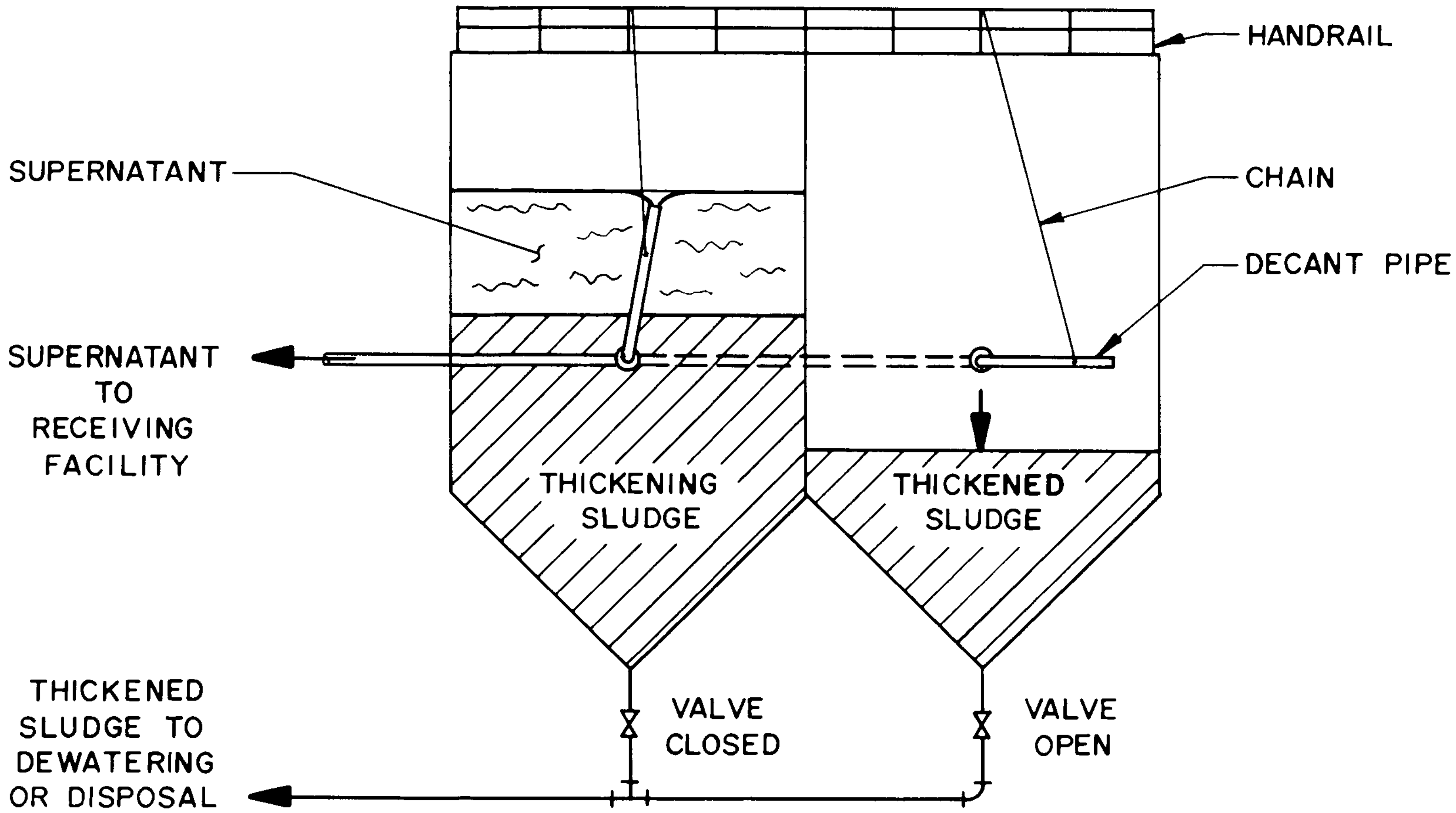


FIGURE 4-4

BATCH THICKENING TANK SCHEMATIC

decant pipe may be continually lowered as the solids settle until the desired solids concentration is reached or the sludge will not thicken further. The thickened sludge is then pumped out of the bottom hoppers to further treatment or disposal.

4.3.2. Design Considerations

Design of batch or continuous flow thickeners is usually accomplished based on previous experience of similar full scale installations or on laboratory settling tests. Small pilot scale thickener tests are very difficult to operate and the results obtained from them are not always reliable.

The common settling test used in the laboratory is conducted in a transparent cylinder filled with sludge and mixed to evenly distribute the solids. At time zero the mixing is stopped and the solids are allowed to settle. Water plant sludges from clarifiers and sedimentation basins will generally settle as a blanket with a well defined interface. By recording the height of the interface with time, a plot such as Figure 4-5 can be created. The free settling velocity is then determined as the slope of the straight line portion of the plot.

In considering the size of the test system, the cylinder diameter is probably the most critical factor. Vesilind (4-1), as shown in Figure 4-6, has evaluated the effects of various cylinder diameters. At low suspended solids concentrations (<0.4%) the smaller cylinders tended to underestimate the settling velocity which would result in a more conservative design. However, at suspended solids concentrations over 0.5% the smaller cylinders over-estimated the settling velocity. Vesilind went on to recommend four considerations in conducting lab thickening tests

1. The cylinder diameter should be large as possible; 8 inches is a practical compromise.

FIGURE 4-5

THICKENING TEST IN A CYLINDER WITH RESULTING INTERFACE HEIGHT VS TIME CURVE

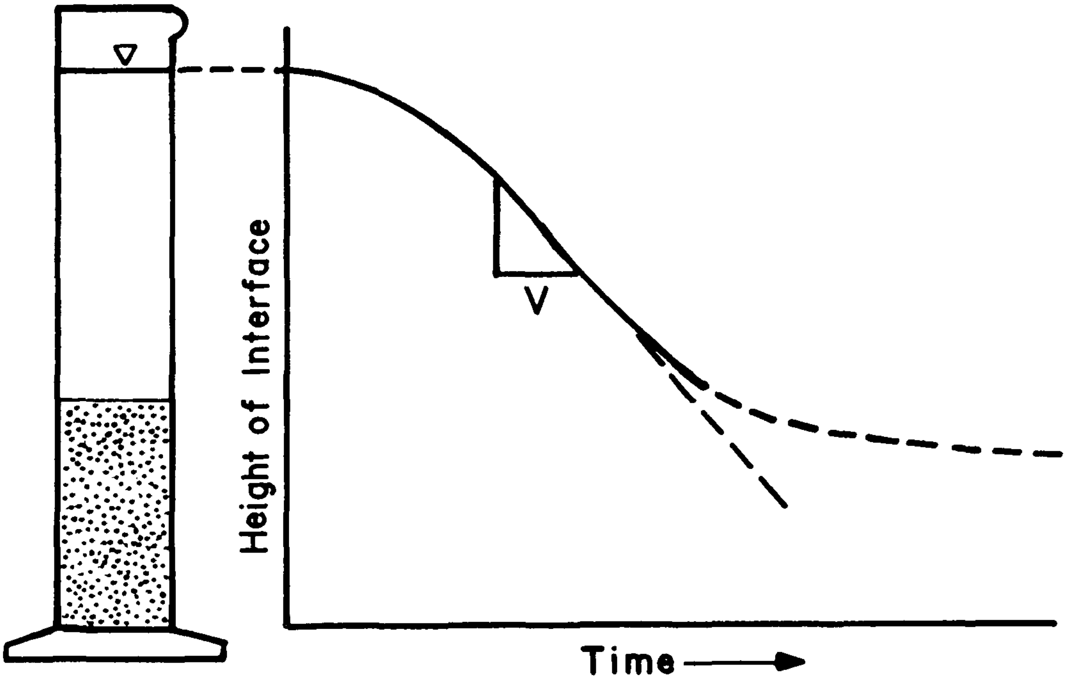
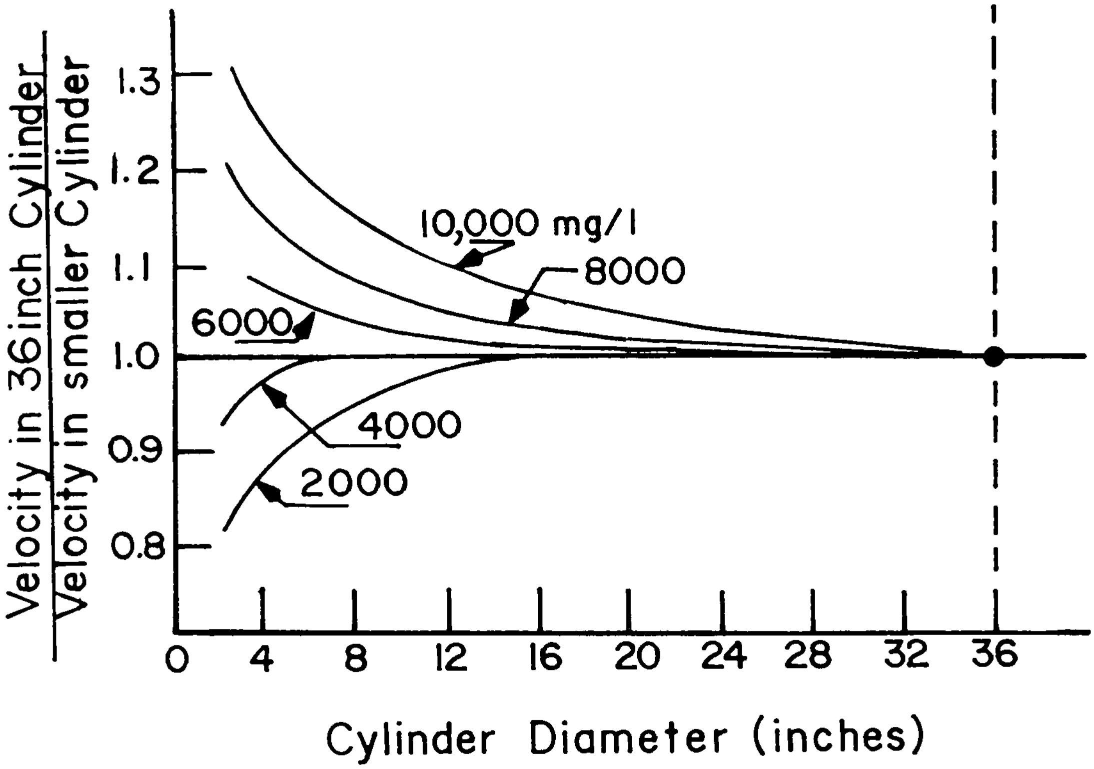


FIGURE 4-6

EFFECT OF CYLINDER DIAMETER ON SETTLING VELOCITY



2. The initial height should be the same as the prototype thickener depth. When this is not practical, 3 ft should be considered minimum.
3. The cylinder should be filled from the bottom.
4. The sample should be stirred throughout the test, but very slowly--0.5 rpm is a reasonable speed for an 8 inch cylinder. This slow stirring will help the test results of small cylinders better approach that of full-scale.

Having completed the settling test as described by Figure 4-5, the test is then repeated for several different initial suspended solids concentrations, resulting in Plot A of Figure 4-7. The solids flux, F, is then computed

$$F = vC_i$$

where,

F = solids flux, kg/m^2/hr
v = settling velocity, m/hr
C_i = initial suspended solids conc., kg/m^3

The flux curve will generally take the shape as shown in Figure 4-7B. The flux is zero at zero suspended solids concentration and the settling velocity approaches zero as the solids concentration increases, thereby driving the flux to zero and representing the maximum possible solids concentration to be achieved.

For batch fill and draw tanks the curves of Figure 4-7A can be directly used to estimate the settling time required and predict the thickened solids concentrations. Similarly, the curves of Figure 4-7B could be used to develop possible flux rates for the anticipated range of influent solids concentrations.

The results of several batch thickening tests plotted as A)interface velocity vs initial solids concentration, and B) solids flux vs initial solids concentration.

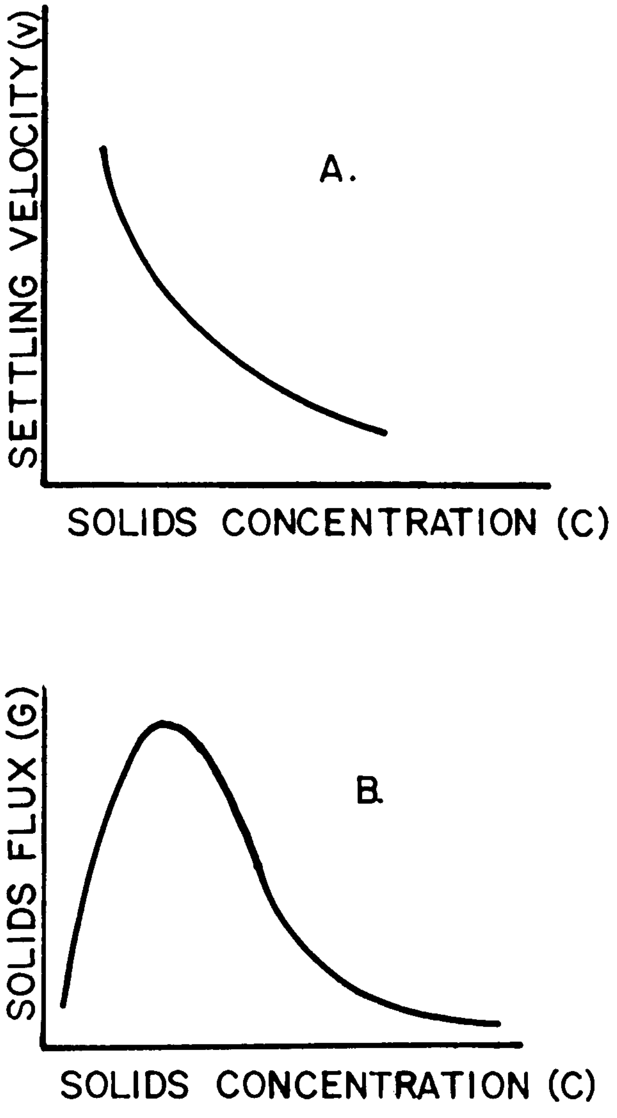


FIGURE 4-7

BATCH THICKENING TEST PLOTS

For continuous flow thickeners the solids move to the bottom of the tank not only due to the batch sedimentation discussed above, but also due to the velocity created by the underflow of sludge being removed from the thickener. The flux due to the sludge withdrawal is

$$F_u = v_u C_i$$

where,

F_u = flux due to underdrain withdrawal of sludge
v_u = downward velocity caused by sludge removal
C_i = solids concentration at a given layer in the thickener

The flux due to settling has already been defined and can be labeled for the continuous flow thickener as

$$F_B = v_i C_i$$

where,

F_B = flux due to solids settling
v_i = settling velocity of solids concentration C_i
C_i = solids concentration at a given layer in the thickener

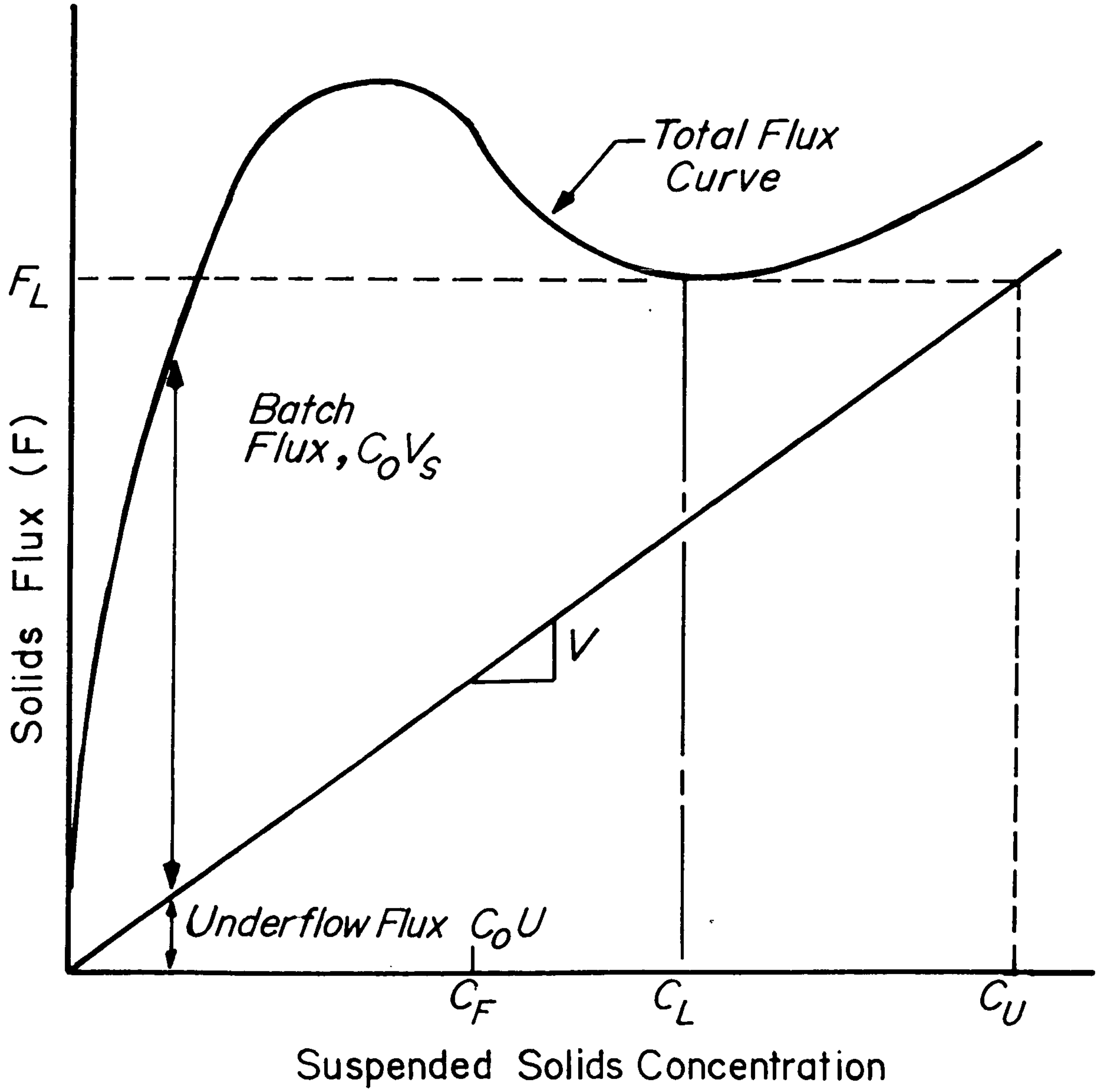
therefore, the total flux is

$$F = v_i C_i + v_u C_i$$

This is plotted as Figure 4-8 where the straight line is $v_u C_i$ and the batch flux is from Figure 4-7B. The minimum

FIGURE 4-8

SOLIDS FLUX CURVE FOR CONTINUOUS THICKENER



point of this flux curve occurs at the solids concentration layer which restricts performance, and thus the flux cannot be higher. Several methods are available to find this maximum flux, most of the methods having been described by Vesilind (4-1). One of the most used methods was developed by Coe and Clevenger (4-2).

Coe and Clevenger used mass and liquid balance equations to develop the following expression for thickener area:

$$A = \frac{Q_o C_o}{v_i} \left(\frac{1}{C_i} - \frac{1}{C_u}\right)$$

where,

A = thickener area (m^2)
Q_o = feed flow rate (m^3/d)
C_o = influent sludge solids conc. (kg/m^3)
v_i = settling velocity of solids C_i (m/sec)
C_i = design solids concentration of given layer within the thickener (kg/m^3)
C_u = desired underflow concentration (kg/m^3)

A number of trials is necessary using various values of v_i and C_i. The largest area calculated should be used as the minimum design thickener area.

Gravity thickeners can also be sized based on experience; i.e. the solids loading rates from similar plants. Lime sludges that are predominately composed of calcium carbonate can be expected to thicken to a 5 percent solids concentration in the sedimentation basin. These lime sludges can be expected to thicken in gravity thickeners to a 30 percent solids concentration at a loading rate of 20 to 40 lb/sf/day. Coagulant sludge loading rates would be in the range of 3 to 5 lb/sf/day. In the design of gravity thickeners the volume for storage of deposited sludge should also be provided.

4.3.3. Capital and Operating Costs

The cost curves for gravity thickeners are presented in this Section. The capital cost curve is shown in Figure 4-9. Thickener capital costs include the cost of the scraper mechanism and its installation and the cost of the circular reinforced concrete basin and appurtenances. The basin is assumed to have a 10 foot side wall depth. Effluent troughs and inbound weir baffles are included. Also, a center support column and steel half-span bridge are included under equipment costs. Typical excavation and site work for the basin and electrical and operation required for the operation of the equipment is also included.

Operating and maintenance costs are shown in Figure 4-10. The operating and maintenance costs include energy costs relative to the process scraper mechanism only and do not take into account the sludge pumping or chemical costs, as those costs are presented separately in other sections of this Chapter. The maintenance materials cost is for repair and replacement of the scraper mechanism and weir. The labor costs are for normal operation and maintenance of the process.

4.4. SLUDGE CONDITIONING

4.4.1. Description

Water sludge conditioning refers to the variety of chemical and physical techniques for altering sludge characteristics to make subsequent removal of water more efficient. There is no clear-cut, accepted conditioning method practiced for a given type of sludge. A conditioning agent that works well at one plant may not work at a similar plant. Sludge properties used for evaluating the effectiveness of conditioning agents include specific resistance, coefficient of compressibility, yield and capillary suction time (as discussed in Chapter 3).

FIGURE 4-9

CONSTRUCTION COST FOR GRAVITY SLUDGE THICKENERS

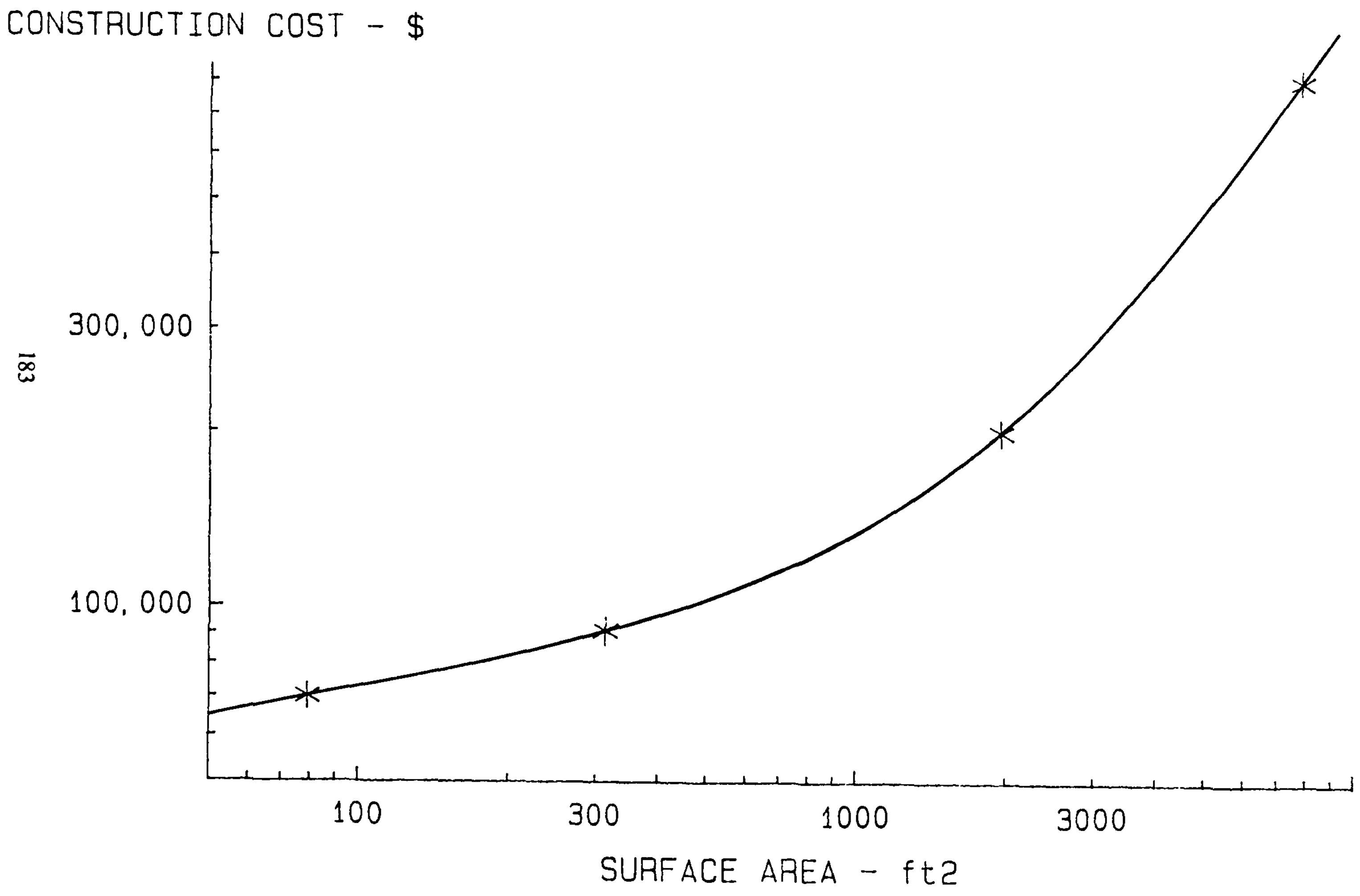
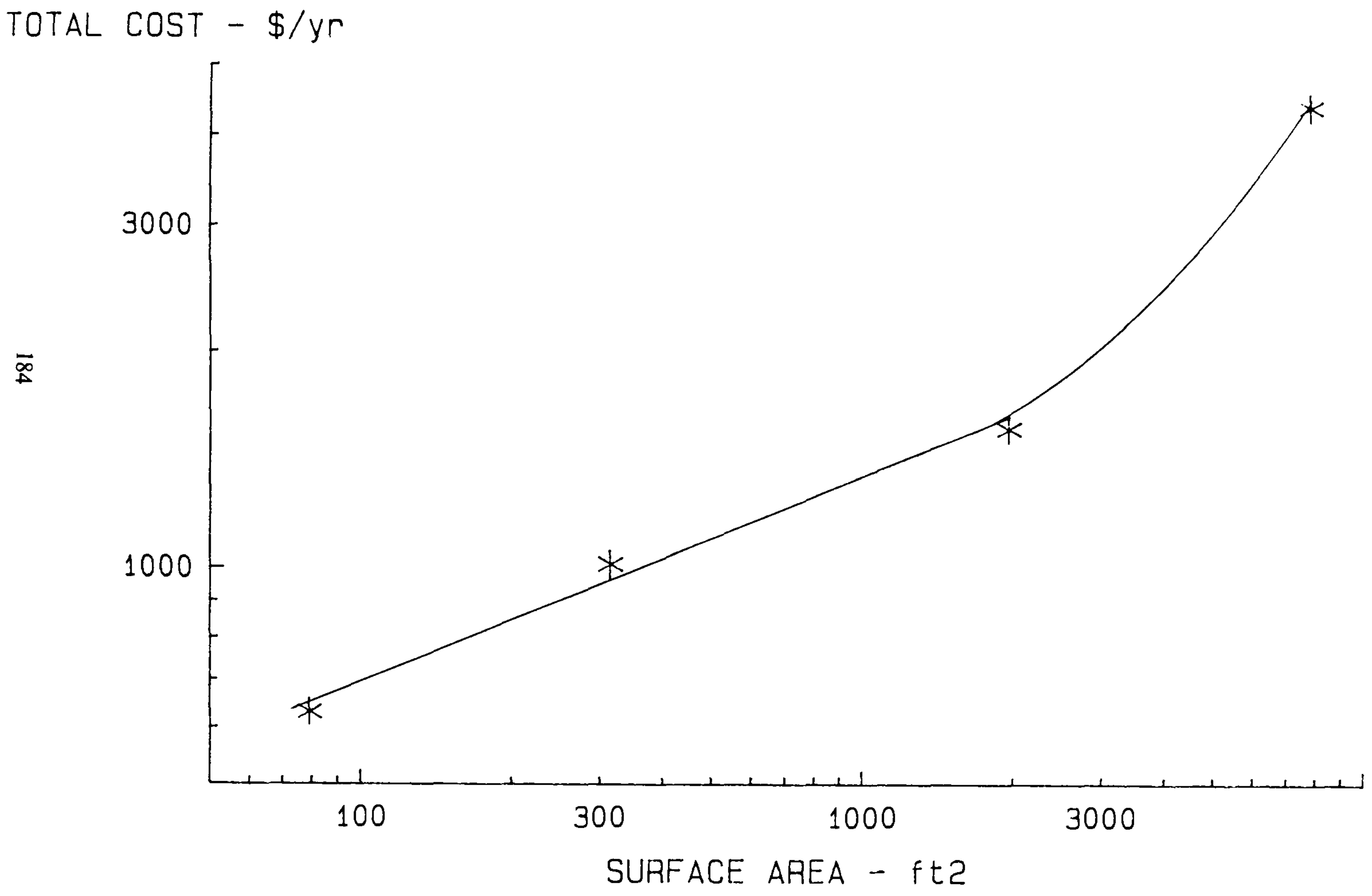


FIGURE 4-10

OPERATION AND MAINTENANCE COST FOR GRAVITY SLUDGE THICKENERS



Conditioning of water plant sludges is generally only applicable to hydroxide sludges and backwash wastes. Sludges from lime softening clarifiers are more easily dewatered and conditioning agents are seldom used. With hydroxide sludges conditioning agents are needed to either assist in the water removal processes or may be used to affect compressibility and minimize media clogging, such as in filter press operation. When conditioning is used for water/solids separation, polymers are generally the agent of choice. When the objective of conditioning is to prevent media clogging, lime has been traditionally utilized although recently polymers have successfully been used for this purpose.

Some of the common types of polymers that are used for sludge conditioning are shown in Figure 4-11 (4-3). Polymers vary in structural composition, molecular weight and charge density. For most cationic polymers the charge density is near 100% and the molecular weight of cationic polymers is generally less than anionic or nonionic polymers. Anionic polymers will vary in both charge density and molecular weight. Nonionic polymers have no charge density, but high molecular weights. As a broad generalization, for hydroxide sludge conditioning the higher the molecular weight of the polymer, coupled with a long carbon chain length the less dose is required for conditioning. Molecular weight may even be more important than the charge type or density (4-4).

Polymer addition has been useful, and in fact almost required for dewatering hydroxide sludges by either non-mechanical methods such as sand drying beds or mechanical methods such as centrifuges, belt filter presses and pressure filters. It appears that the primary mechanism is one of interparticle bridging such that the polymers form a porous matrix that permit water decant or drainage. It has been postulated (4-5) that the polymer does not alter the chemical structure of the hydroxide particles themselves.

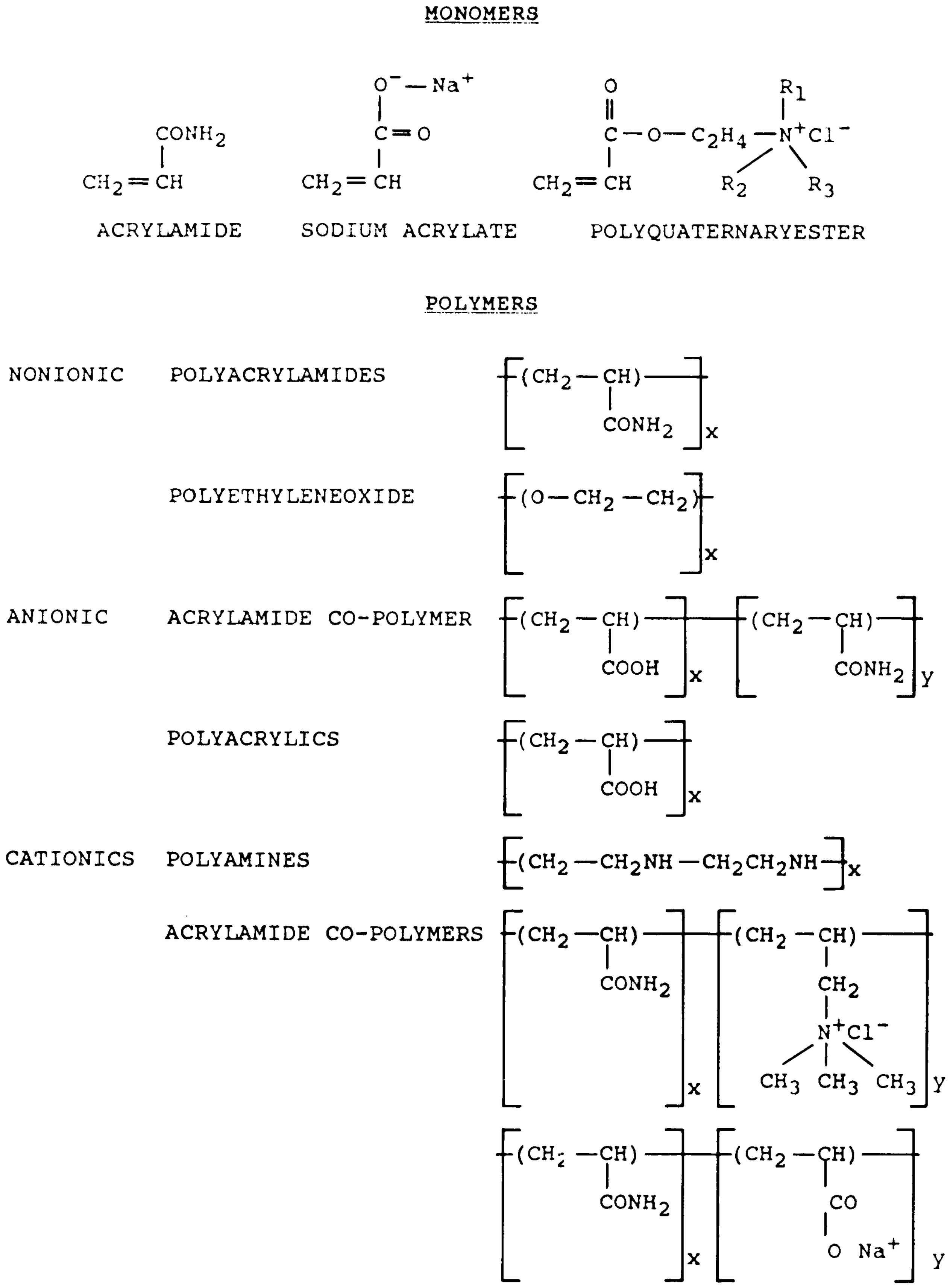


FIGURE 4-11

TYPICAL SLUDGE CONDITIONING MONOMERS AND POLYMERS

Reference: 4-3

Particular use of chemical agents is discussed within the specific dewatering application. However, some general comments are in order. When first selecting a polymer type, a series of screening tests is required. Generally, manufacturers of polymer will provide or inexpensively sell sample polymers which they believe will be suitable for the particular application. Many articles have appeared discussing quick visual methods of polymer screening as well as exhaustive test procedures. One of the best references is that prepared by the AWWA Research Foundation, "Procedures Manual For Selection of Coagulant, Filtration and Sludge Conditioning Aids in Water Treatment," and the techniques will not be repeated herein. With a little practice and experience it is possible to visually screen several polymers by simply adding increasing doses to small beakers of sludge and viewing the floc. While ideal doses do not look the same for different polymers and sludges, experience does allow rapid screening. To determine comparative doses and quantitive results, either the CST test or the specific resistance test is recommend. If CST equipment is available, the procedure is much faster. Figure 4-12 shows one such comparative plot for the effect of polymer type and dose on the conditioning of an aluminum hydroxide sludge.

Having selected a polymer and dose that is optimal, most utilities will want to competitively bid the purchase. Even if the polymer to be used is specified, several manufactures will want to bid what they consider an equal product. It is recommended that bid documents specify dollars versus performance. For example, a given polymer may cost twice per pound but reach the optimal conditioning at one-third the dose--not an uncommon situation. Bid prices should be based on the dollars to reach a specified test condition. In the Figure 4-12 it could be stated that purchase is being made for sufficient polymer to treat 650 tons/year of sludge and the polymer must attain a CST of 7. Bids will then be based on quantities and cost with appropriate guarantees. Of course, provisions

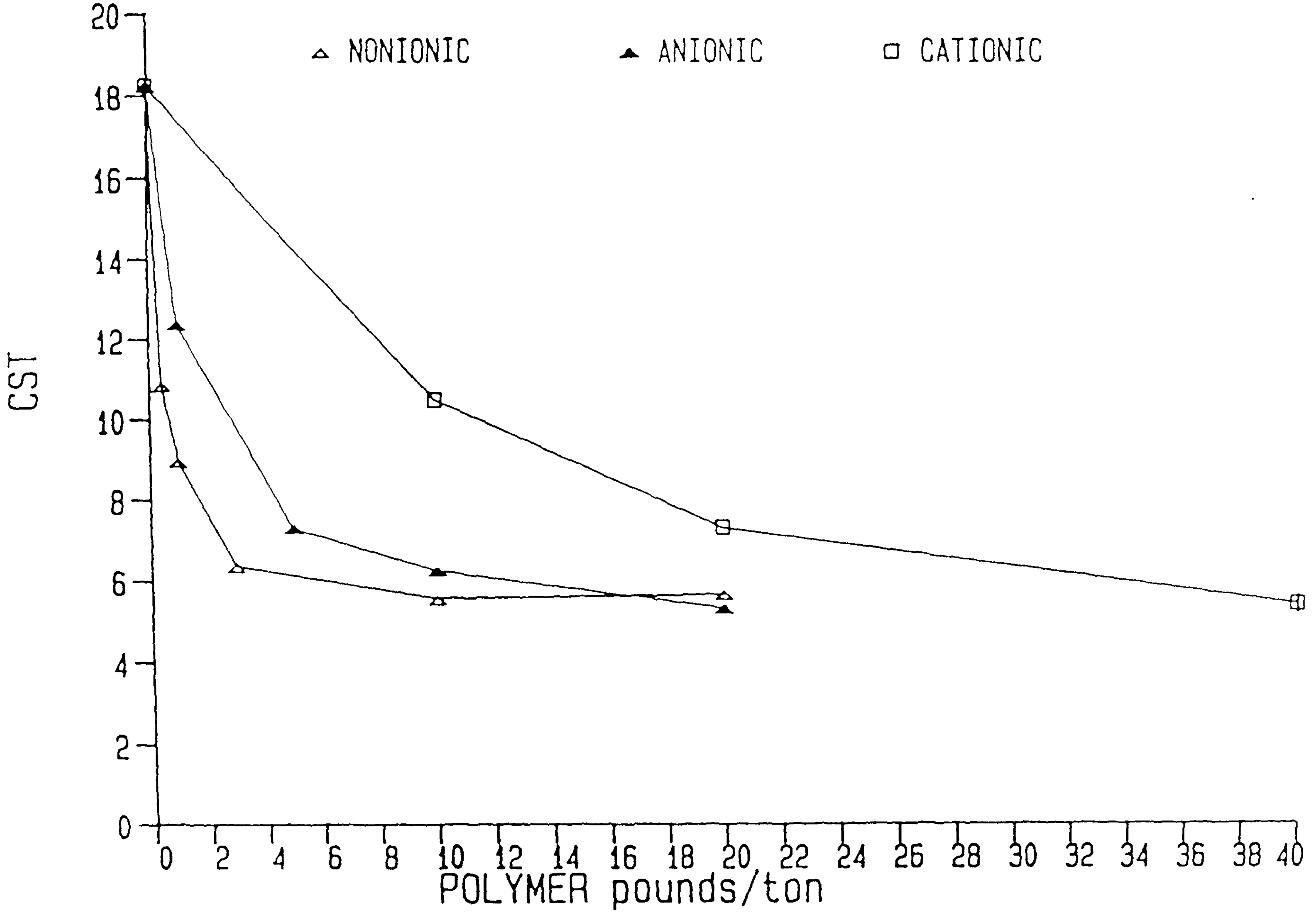


FIGURE 4-12

CST VERSUS POLYMER DOSAGE

must be allowed for manufacturers to demonstrate their product prior to bid.

4.4.2. Capital and Operating Costs

The cost curves for lime and polymer feed conditioning systems are presented in this Section.

The construction costs for the polymer feed system are shown in Figure 4-13.

The capital costs developed for polymer feed systems are based on feeding dry polymer directly to a storage hopper on a chemical feeder. The system is sized based on a 0.5 percent stock solution and 30 minutes of aging time. Piping, valves and instrumentation required for the proper operation of the feed system is included. A standby polymer feed pump is provided.

The operation and maintenance costs for the polymer feed system are shown in Figure 4-14. The O & M costs include the energy requirements for the feeder and metering pump, maintenance material costs and system labor requirements. Since the cost for the numerous polymers available are so variable, no polymer cost was included and must be added separately to the system cost.

A liquid polymer system is generally comparable in cost to a dry polymer system. The capital cost for the liquid system is usually lower but the operation and maintenance costs are usually higher, due to higher chemical purchase costs and increased horsepower.

The lime feed system was designed to feed hyrated lime. The system includes lime storage, dual feeder solution tanks with mixers, and dual metering pumps. The storage hopper was sized for 1 day of lime storage and was provided with dust collectors (see Figure 4-15).

The operation and maintenance costs for lime feed systems (Figure 4-16) included energy requirements for the mixers and metering pumps, maintenance materials costs and labor costs.

FIGURE 4-13

CONSTRUCTION COST FOR POLYMER FEED SYSTEMS

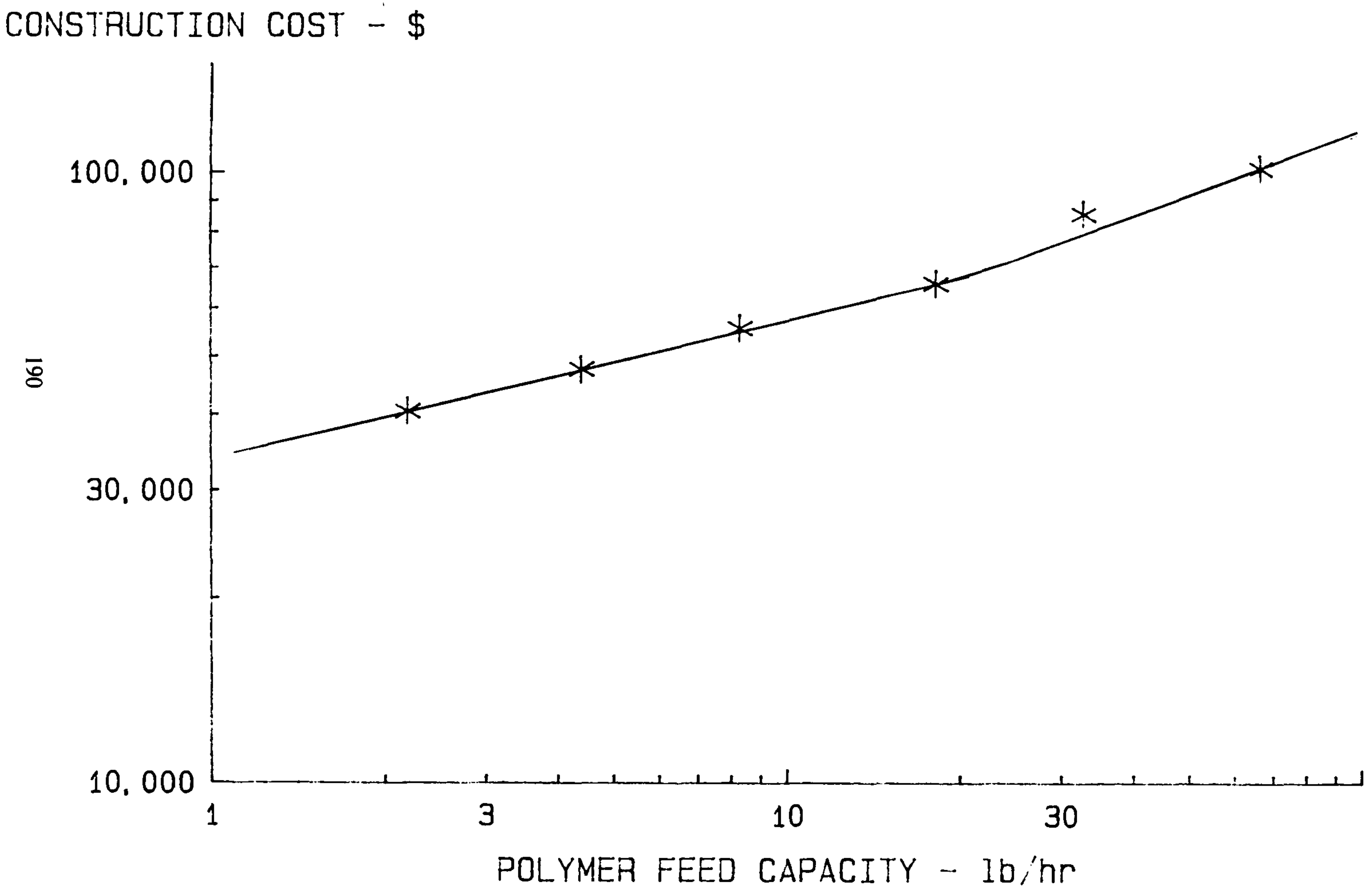


FIGURE 4-14

OPERATION AND MAINTENANCE COST FOR POLYMER FEED SYSTEMS

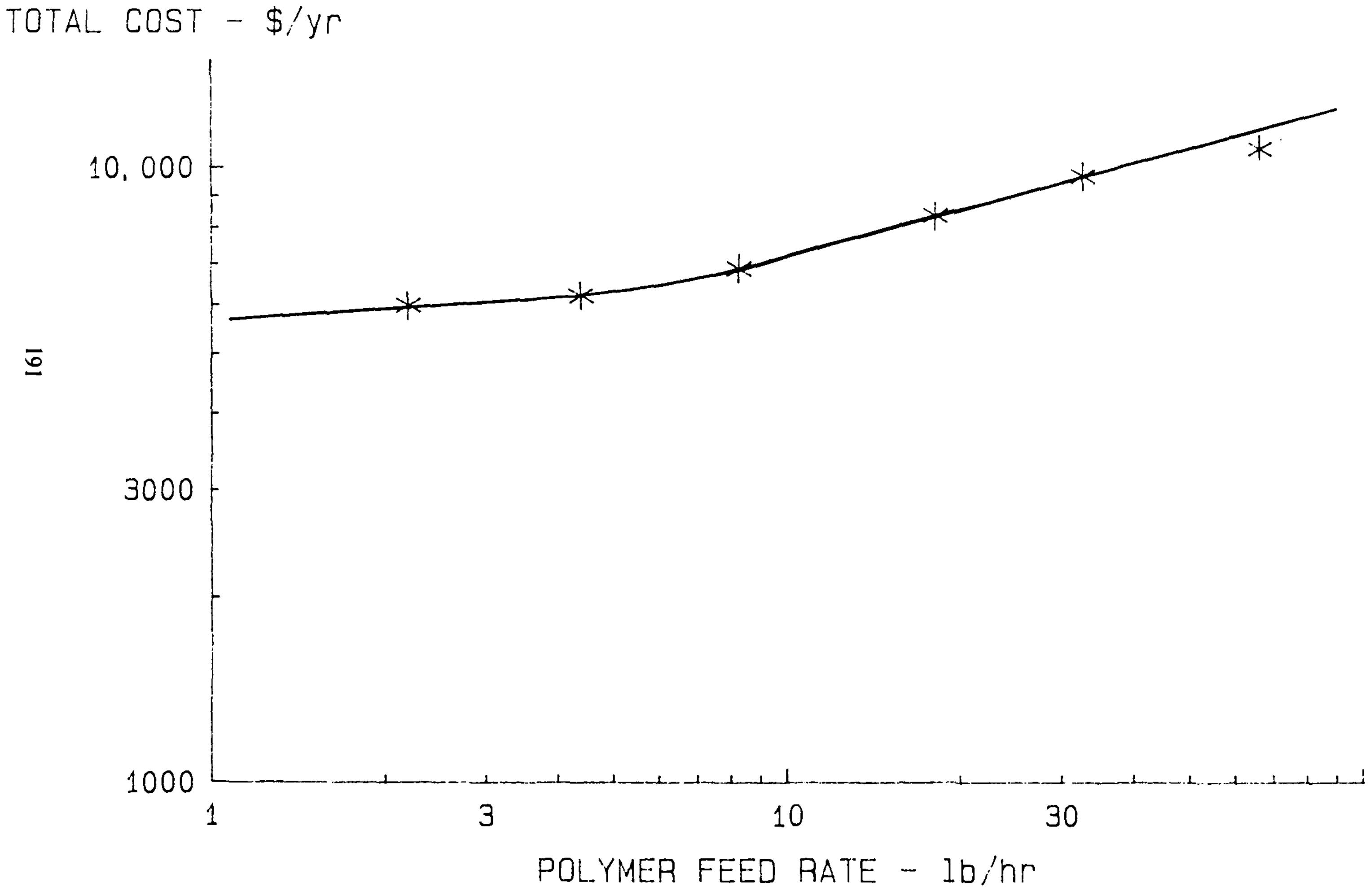


FIGURE 4-15

CONSTRUCTION COST FOR LIME FEED SYSTEMS

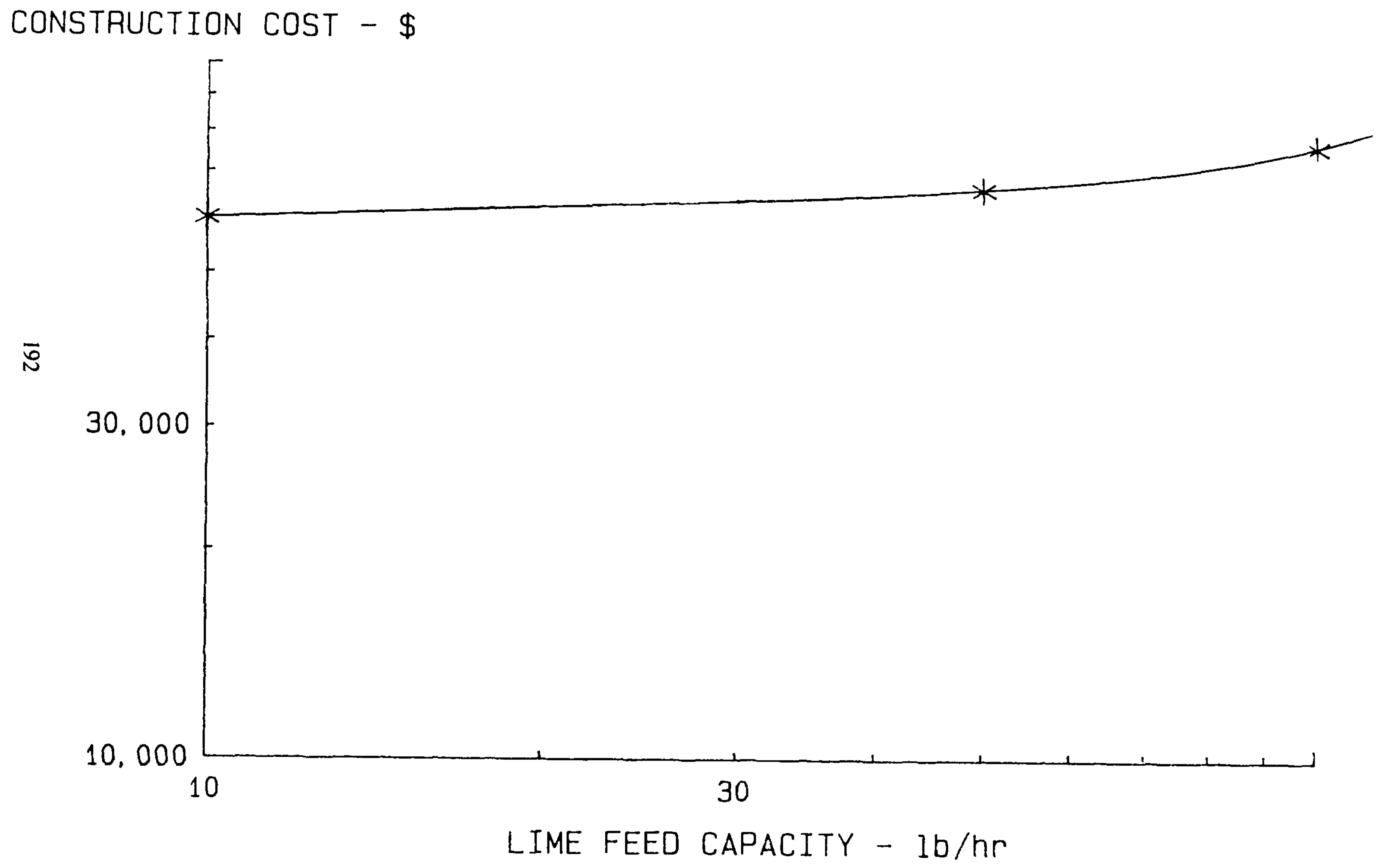
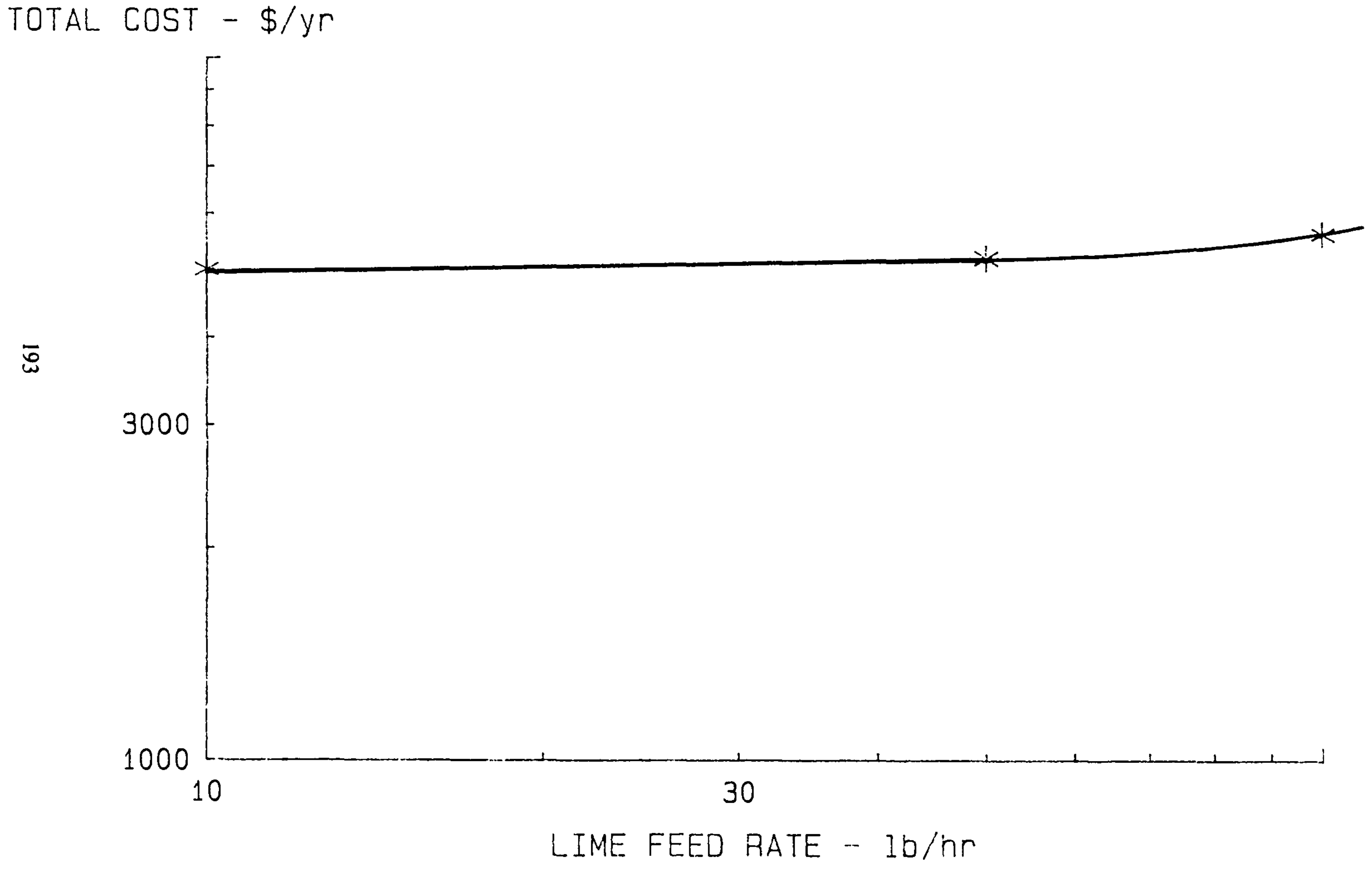


FIGURE 4-16

OPERATION AND MAINTENANCE COST FOR LIME FEED SYSTEMS



Labor and maintenance material costs were based on manufacturers recommendations and experience. The cost of lime itself is not included.

4.5. SLUDGE PUMPING

4.5.1. Description

The unthickened sludge coming from the sedimentation basin in a coagulation process is generally fairly dilute, ranging from 0.5 to 2 percent solids concentration for alum sludges. These sludges can be conveyed by gravity or siphoning from the sedimentation basin to a sludge pumping station. Pumping of unthickened sludge can usually be accomplished using centrifugal pumps. Sludges from lime clarifiers are much thicker and may need to be handled by 'thickened sludge pump stations', described below.

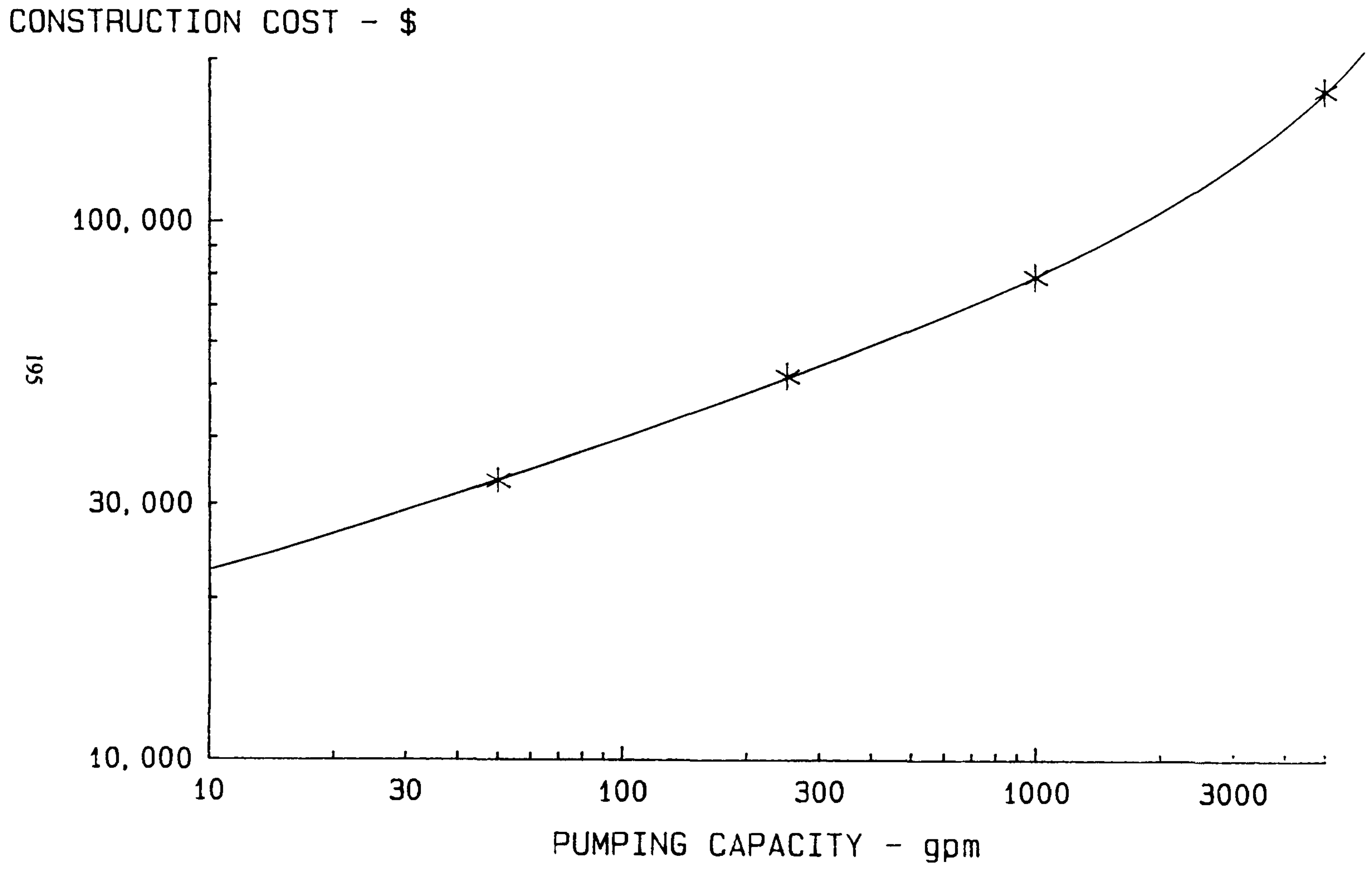
Sludge thickened in gravity thickeners can be expected to achieve a solids concentration of 3 to 4 percent for coagulant sludges and up to 30^+ percent for lime softening sludge. These thickened sludges generally can be pumped by progressive cavity pumps or other types of positive displacement pumps. Coagulant sludge in the 3 to 4 percent range can sometimes be pumped by centrifugal pumps and engineering judgement is needed as to which system is appropriate.

4.5.2. Capital and Operating Costs

Cost curves for the pumping of unthickened and thickened sludge are presented in this Section. Capital cost curves for pumping unthickened sludge are shown in Figure 4-17. Capital costs were developed for pumping stations of 12 feet depth. The pumps utilized were constant speed, submersible, centrifugal type with one standby pump. A precast concrete wet well was provided. Pipes and valves were sized for 5 feet per second velocity. Housing of the electrical equipment was provided over the wet well.

FIGURE 4-17

CONSTRUCTION COST FOR
UNTHICKENED SLUDGE PUMPING



Operating and maintenance cost curves for pumping unthickened sludge are shown in Figure 4-18. Operating and maintenance costs included process electrical energy, maintenance material and labor costs. Process electrical energy costs were based on pumping sludge against a total dynamic head of 30 feet. Maintenance materials and labor costs were developed for sustained system operation.

Capital cost curves for the pumping of thickened sludge are shown in Figure 4-19. Capital costs for pumping thickened sludge from gravity thickeners included progressive cavity pumps and motors, housing, piping, valving, electrical and instrumentation. Housing costs include a rectangular concrete structure below grade and above ground housing of the electrical and control equipment.

Operating and maintenance cost curves for pumping thickened sludge are shown in Figure 4-20. Operating and maintenance costs included process electrical energy, maintenance material and labor. Process electrical energy was derived for pumping a thickened sludge against a total dynamic head of 50 feet. Maintenance material and labor costs were for routine operation and maintenance.

4.6. CENTRIFUGES

4.6.1. Description

Centrifugation of sludge is basically a shallow depth settling process enhanced by applying centrifugal force. The basic physical principle of centrifugal force is that a moving body tends to continue in the same direction; if that body is forced to change directions, it resists the change and exerts a force against whatever is resisting it. In the case of centrifugal force, the force applied by the body is radially outward from the axis of rotation.

Centrifugation enhances settlement of the solids. In conventional settling tanks, the solids are acted on by the

FIGURE 4-18

OPERATION AND MAINTENANCE COST FOR UNTHICKENED SLUDGE PUMPING

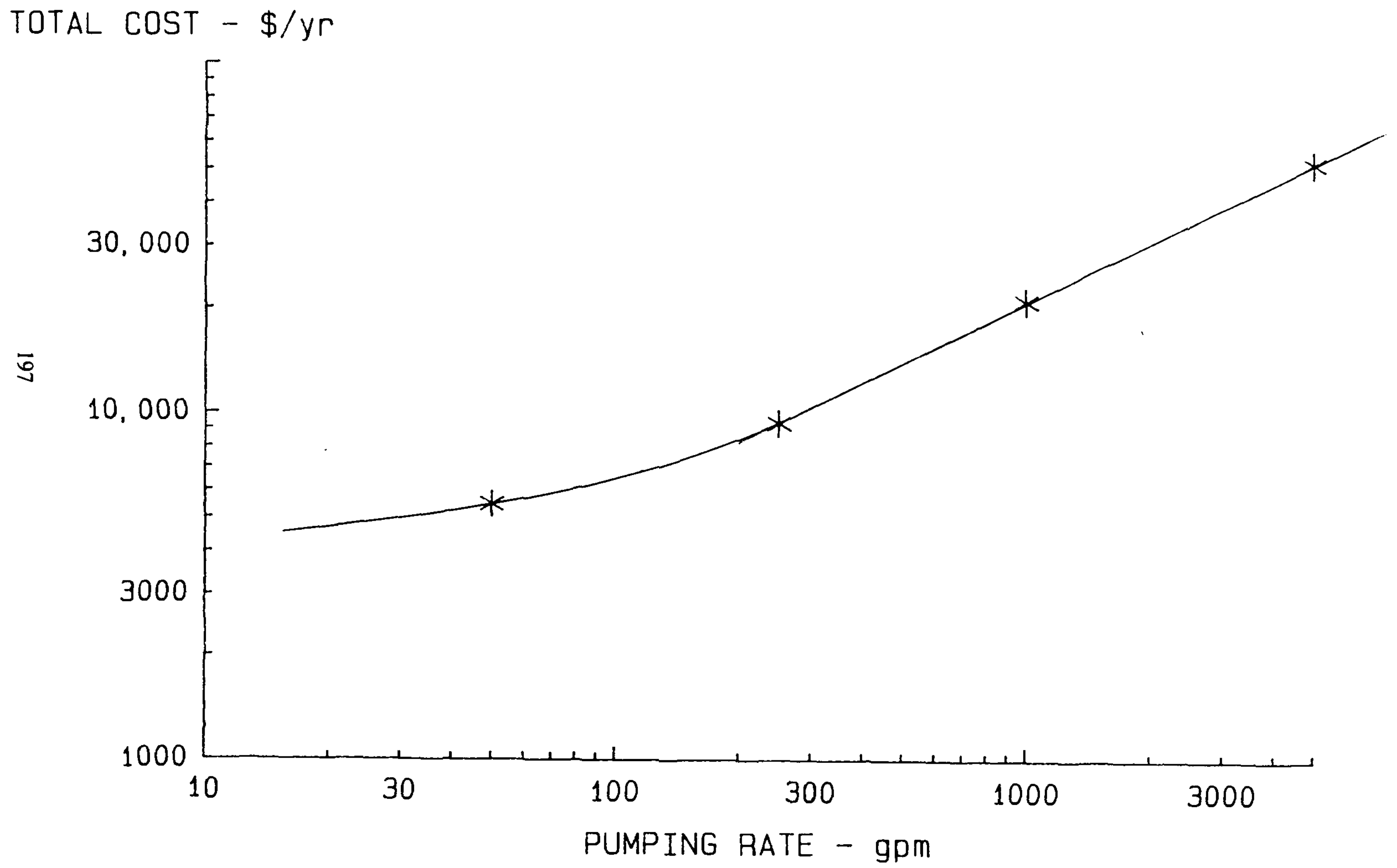


FIGURE 4-19

CONSTRUCTION COST FOR
THICKENED SLUDGE PUMPING

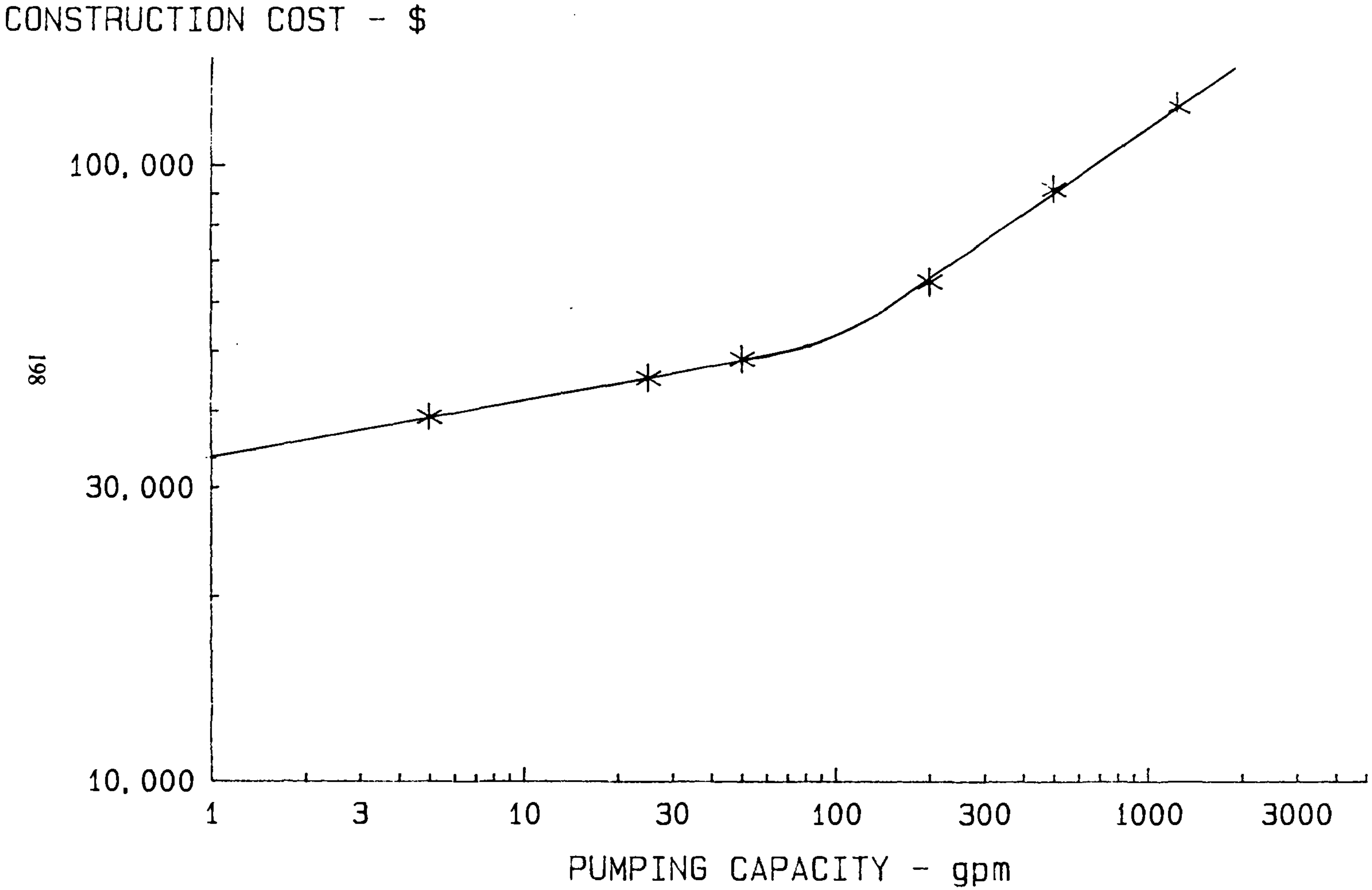
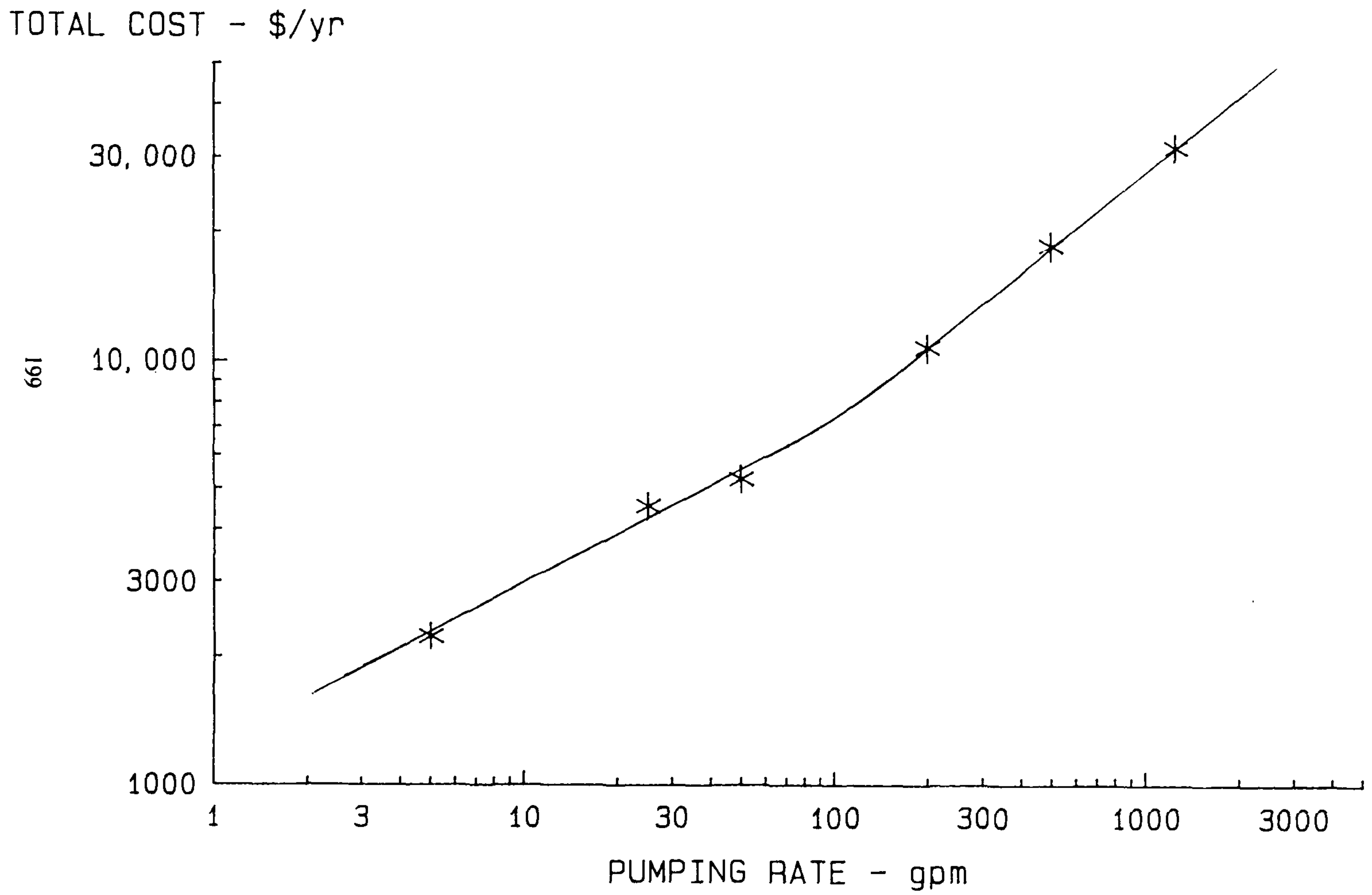


FIGURE 4-20

OPERATION AND MAINTENANCE COST FOR THICKENED SLUDGE PUMPING



force of acceleration due to gravity, g. In centrifugation, the applied force is rw^2, where r is the distance of the particle from the axis of rotation and w is the rotational speed. In modern centrifuges rw^2 may be 1500 to 4000 times the value of g.

The comparison of rw^2 to g has led to efforts by many to develop equations for centrifugation by substituting rw^2 for g. However, this substitution relates to discrete particles only and does not account for hindered settling and the effect of scrolling (moving the solids out of the bowl). These deficiencies in the theory of centrifugation limits the use of sedimentation theory as a basis for the design of centrifuges, and evaluation and design needs to be based on pilot studies.

Laboratory bottle centrifuge tests and small scale centrifuge tests have been used with some success. However, scale up from these small-scale tests have been best based on the development of empirical relationships and judgement. These procedures should only be used for initial process screening studies.

The two major types of centrifuges used for the dewatering of water plant sludge are the scroll-discharge, solid bowl decanter and the plow-discharge basket bowl centrifuge. The solid bowl centrifuge (also called scroll or decanter centrifuge) is a horizontal unit that utilizes a scroll conveyor inside the centrifuge bowl (see Figure 4-21). The unit is fed continuously with the solids settling against the bowl wall. The scroll rotates at a slightly different speed than the bowl and conveys the dewatered sludge to the small end of the centrifuge where it is discharged. The water is directed from the central axis of the centrifuge toward the centrifuge's large end where it is discharged. The water exits through adjustable weirs (level rings), which also control the pool depth.

Basket centrifuges are vertical units (orientation of the centerline) that operate as batch processes to thicken and/or

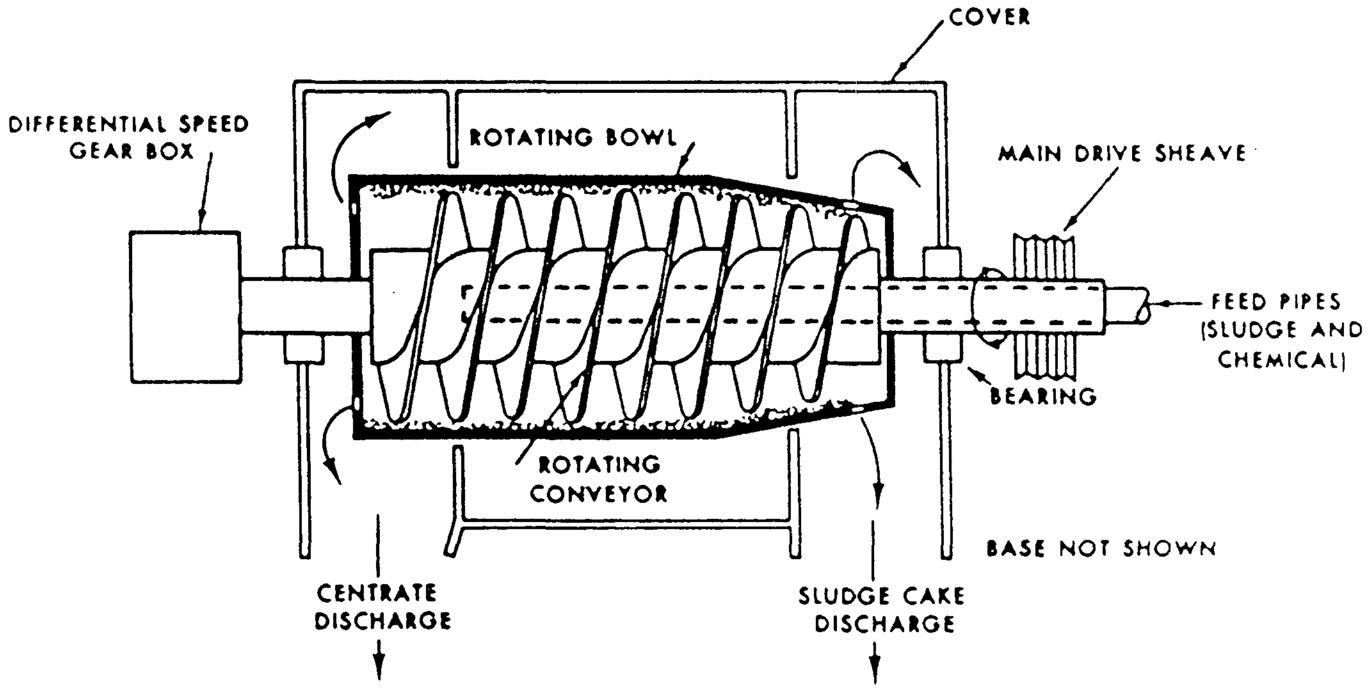


FIGURE 4-21

SOLID BOWL CENTRIFUGE

dewater sludges (see Figure 4-22). The batch cycle begins with the empty basket centrifuge being accelerated prior to feeding any sludge. A stationary inlet pipe extends through the centrifuge and feeds the sludge to the bottom of the centrifuge. Velocity dissipation of the sludge is accomplished by angling the feed pipe to the centrifuge floor. The centrate (the water from which the solids have been removed) is continuously withdrawn from the top of the centrifuge during the cycle. As the sludge feed flows upward through the centrifuge, the solids are continuously deposited along the centrifuge walls.

The feed flow is continued until the centrate loses its clarity or until a predetermined time has elapsed. Then the centrifuge will decelerate until reaching a preset rpm setting, at which time a plow will be activated. The plow (located near the center of the spindle shaft) travels to the outside wall of the centrifuge where it scrapes the sludge cake from the wall and releases it out the bottom of the centrifuge.

Some basket centrifuges contain skimmer mechanisms which precede the plow cycle. The skimmer moves into a preset position near the formed cake surface while the centrifuge is rotating at full speed. Its objective is to collect liquids remaining above the sludge cake surface. By directing the skimmer nozzle to the centrifuge wall, the skimmer can be used alone to discharge soft cakes in order to save time and power. Thus the skimmer adds operational versatility to the centrifuge dewatering process. Basket centrifuges are normally large units available in sizes up to 1200 mm (48 in.) diameter and 760 mm (30 in.) bowl depth. Although large in volume, the hydraulic pump rate to a single batch machine is only 1 to 3 L/sec (20-40 gpm).

The solid bowl decanter centrifuge is greatly favored over the basket centrifuge in the water industry. Since the basket centrifuge is a batch process, it is labor intensive

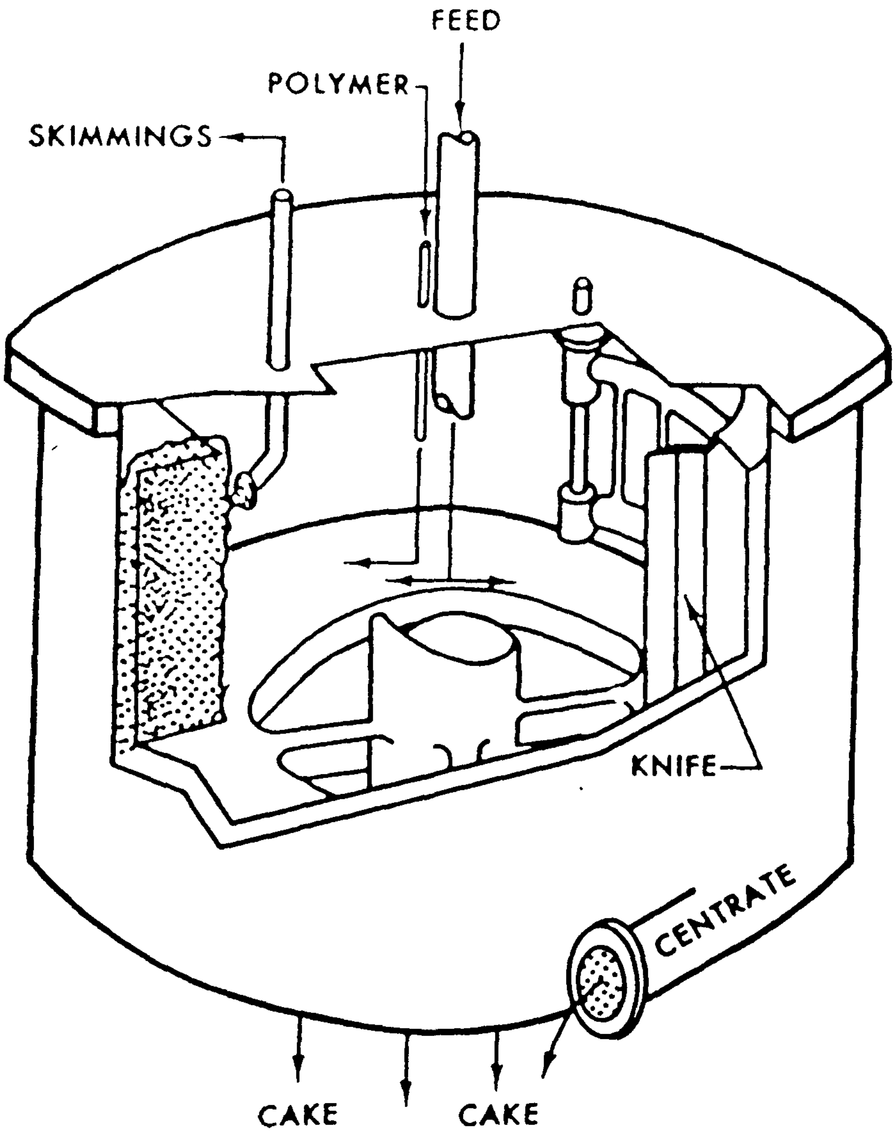


FIGURE 4-22

BASKET CENTRIFUGE SCHEMATIC

and requires continuous monitoring. Also, during the last 10 years the decanter centrifuges have been upgraded to apply centrifuge force up to 4,000 times g, of centrifugal force, which allows adequate dewatering in a continuous feed.

The best procedure for evaluation of centrifuges is pilot tests on prototype equipment. Tests should be conducted on a centrifuge exactly like that to be used in full-scale except smaller. Tests should be conducted for operational parameters of concern such as: feed flow rate, feed suspended solids concentration, and polymer conditioning, bowl speed, pool depth and scroll speed.

The best indicators of performance are the cake solids concentration and the centrate suspended solids concentration. It is preferred to obtain a pilot machine with a variable speed drive motor to alter the bowl speed and a machine where the pool depth can be easily changed. This will allow evaluation of machine variables as well as sludge characteristics.

Having found the optimal conditions for operation on a pilot machine, the problem of scale up to production units still remains. While this process is often left to the manufacturer, it is to the utilities and the engineers advantage to remain closely involved in the scale up considerations. Many times scale up will be done strictly assuming that liquid loading is the limiting parameter. However, with water plant sludge, solids loading can often be limiting. Therefore, full-scale units could be installed only to find their actual capacity is less than anticipated.

Scale up considering only liquid loading is often referred to as the 'Sigma concept.' The Sigma concept is based on Stokes Law describing the settling of discrete particles under the influence of gravity. Gravity is replaced by the centrifugal acceleration and the expression is integrated over the depth of the water pool. One then ends up with a term for the allowable flow through the centrifuge:

$$Q = \left(\frac{V\ w^2}{g \ln \frac{r_2}{r_1}}\right) \left(\frac{g(P_p - P)d^2}{18u}\right)$$

where,

v = volume of sludge/water in the pool
w = radial velocity of centrifuge, radians/sec
r_2 = radius from centerline of centrifuge to bowl
r_1 = radius from centerline of centrifuge to pool level
P_p = particle density
P = fluid density
d = particle diameter
u = viscosity

Note that the left hand term is made up of machine variables and the right hand term is sludge variables. Therefore, in scale up, if it is assumed that the sludge is the same for full scale as in the pilot studies

$$Q_2 = \frac{Q_1\ \Sigma_2}{\Sigma_1}$$

Where Q_2 is the allowable flow in the full scale centrifuge based on the optimal flow (Q_1) obtained in the pilot plant, and

$$\Sigma = \frac{V\ w^2}{g \ln \frac{r_2}{r_1}}$$

which are variables obtainable for the particular size pilot and full scale centrifuge.

An analysis of the solids loading limitation is known as the 'Beta concept', and is expressed as

$$Q_{s2} = \frac{Q_{s1} \beta_2}{\beta_1}$$

where Q_s is the solids throughput in units such as kg/hr and

$$\beta = \Delta W S N D z \Pi$$

where,

ΔW = bowl/conveyor differential speed

S = pitch of blades

N = number of leads

D = total bowl diameter

z = pool depth

again, all beta terms are made up of machine variables and set by the manufacturer for the unit of interest.

In scale-up the limiting conditioning should be calculated using both the sigma and beta concepts.

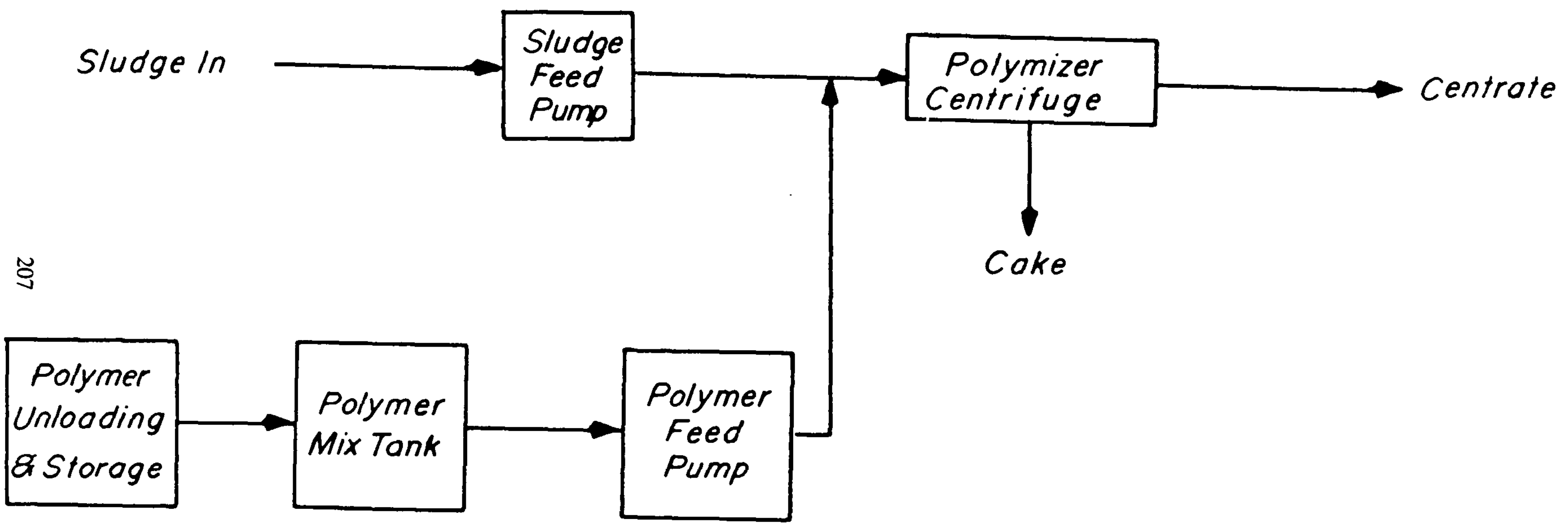
4.6.2. Design Considerations

The centrifugation system includes the centrifuge and for hydroxide sludges a sludge conditioning method. A schematic of the centrifugation system is shown in Figure 4-23.

An advantage to the centrifuge when compared to other dewatering methods is the small space requirement. A large centrifuge (200 to 700 gpm), may require an area of 400 sf. Additional space is required near the centrifuge for the following

FIGURE 4-23

CENTRIFUGATION SYSTEM



- Polymer storage, mixing tank and pumps
- Sludge feed pumps and piping
- Overhead hoist
- Proper operational and maintenance space

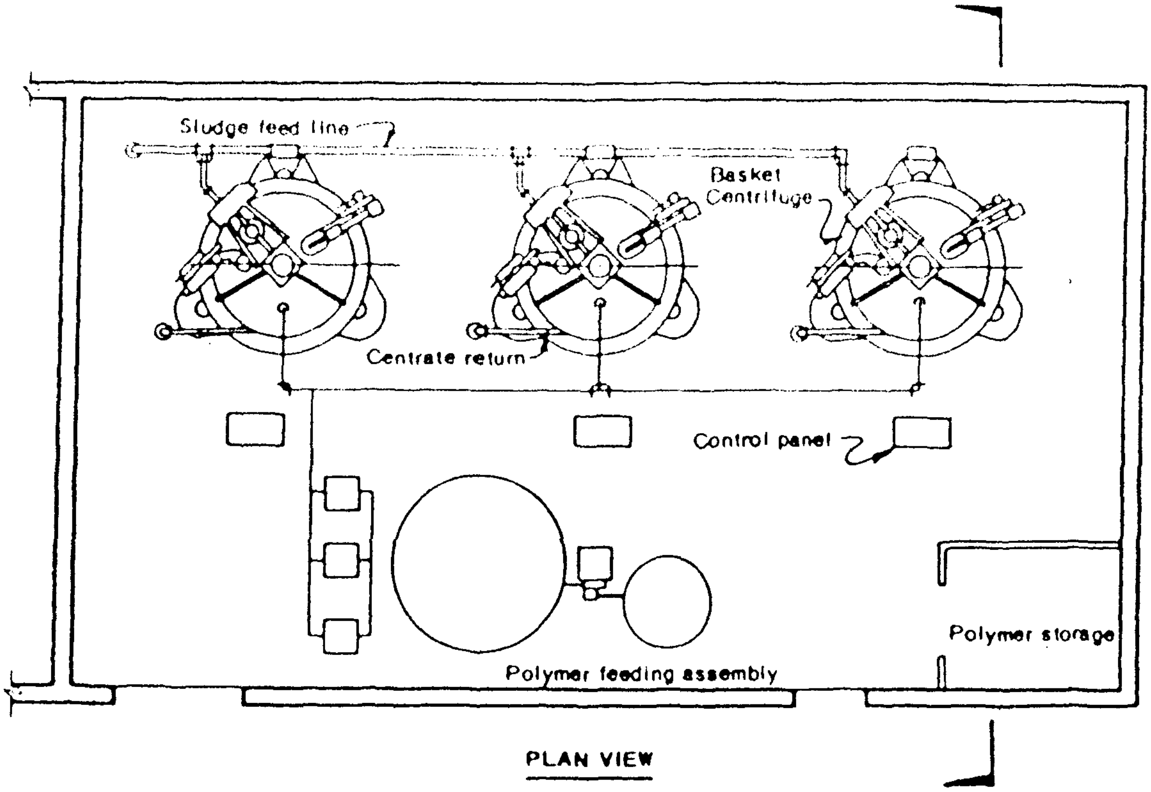
Centrifuges are often located on upper floors of the sludge building so that the cake may be discharged into trucks or hoppers below.

Foundations for centrifuges require special attention in order to support the heavy machine. The cost of the sludge building structure is usually increased when locating the centrifuge on the upper floors. Vibration from the machine may require special treatment. Typical installations for basket and decanter centrifuges are shown in Figures 4-24 and 4-25.

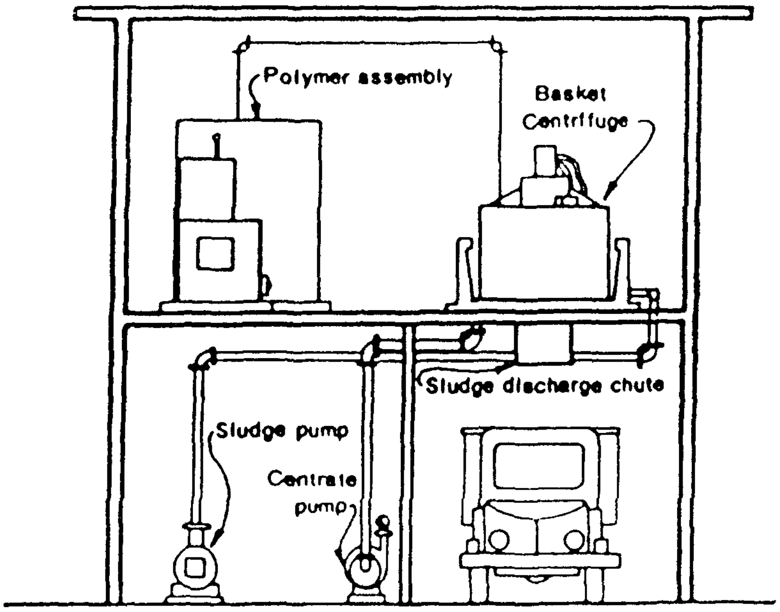
4.6.3. Capital and Operating Costs

Cost curves for decanter, solid bowl centrifuges are shown in Figures 4-26 and 4-27. The capital cost curve is shown in Figures 4-26. Capital costs include the cost for the manufactured equipment, the pipes and valves, electrical, instrumentation and housing. The centrifuge equipment cost includes the cost for the base centrifuge machine, the drive motor, the hydraulic backdrive, one centrate storage hopper, dual centrate pumps and flex connectors. The sludge feed pumps and filtrate pumps are not included in the system cost and can be obtained from the previous section on sludge pumping. Also, no sludge conditioning is included in the system costs; sludge conditioning costs can be added from the previous section in this Chapter. The housing size was derived from manufacturer's layout recommendations. Two story housing is provided.

The operating and maintenance cost curve is shown in Figure 4-27. Operating and maintenance costs for centrifuges include process energy, maintenance material and labor costs.



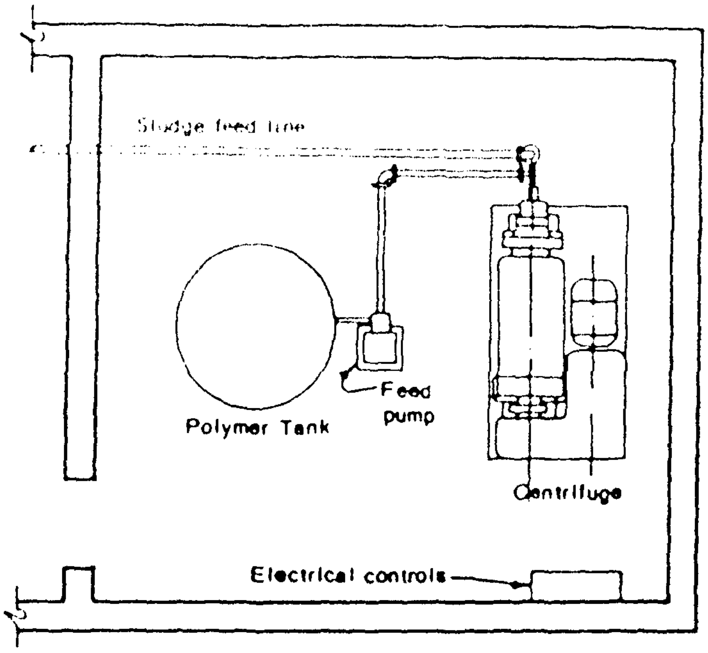
PLAN VIEW



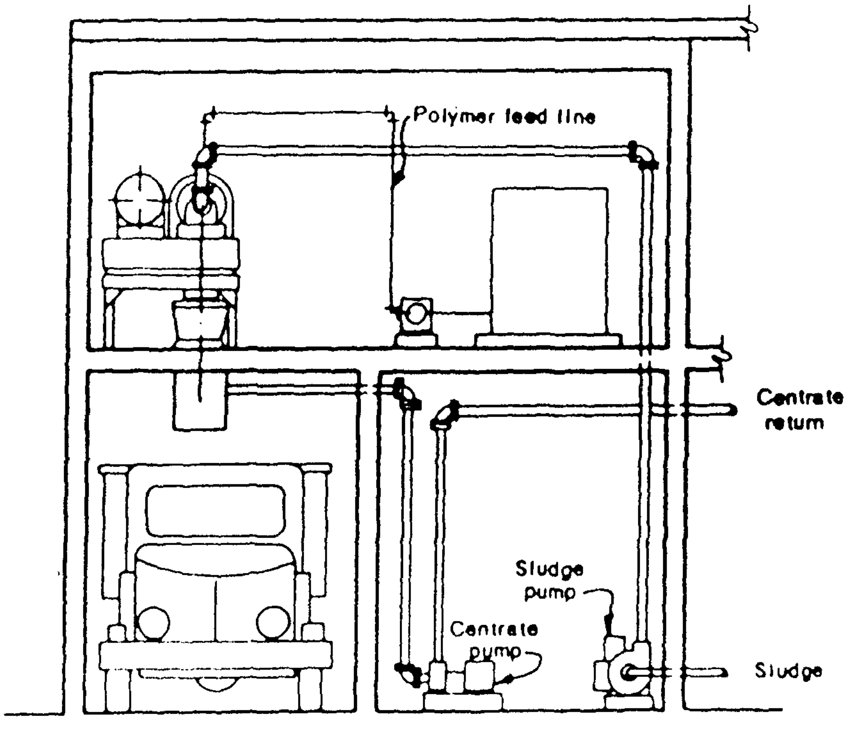
SECTION VIEW

FIGURE 4-24

TYPICAL BASKET CENTRIFUGE INSTALLATION



PLAN VIEW



SECTION VIEW

FIGURE 4-25

TYPICAL DECANTER CENTRIFUGE INSTALLATION

FIGURE 4-26

CONSTRUCTION COST FOR DECANTER CENTRIFUGES

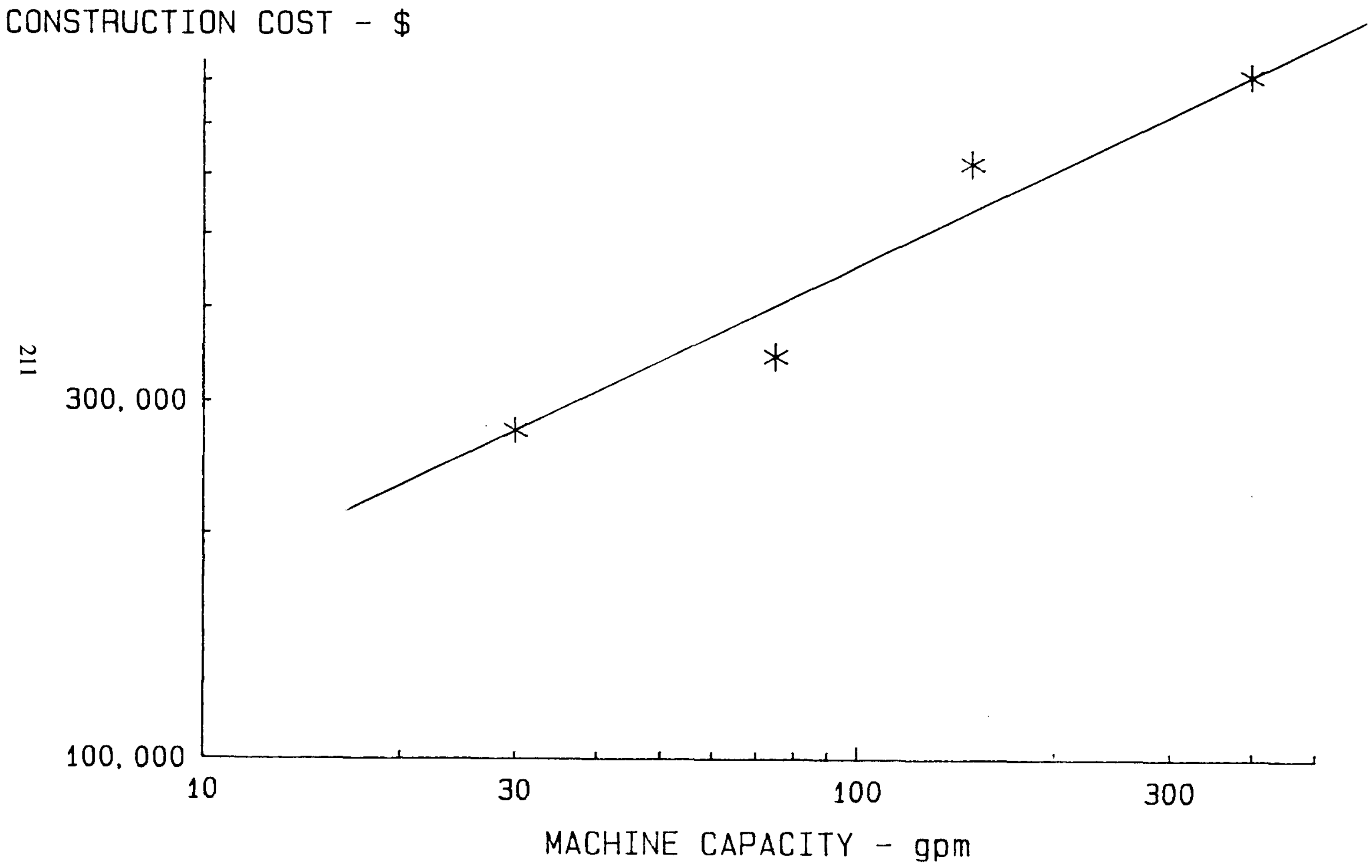
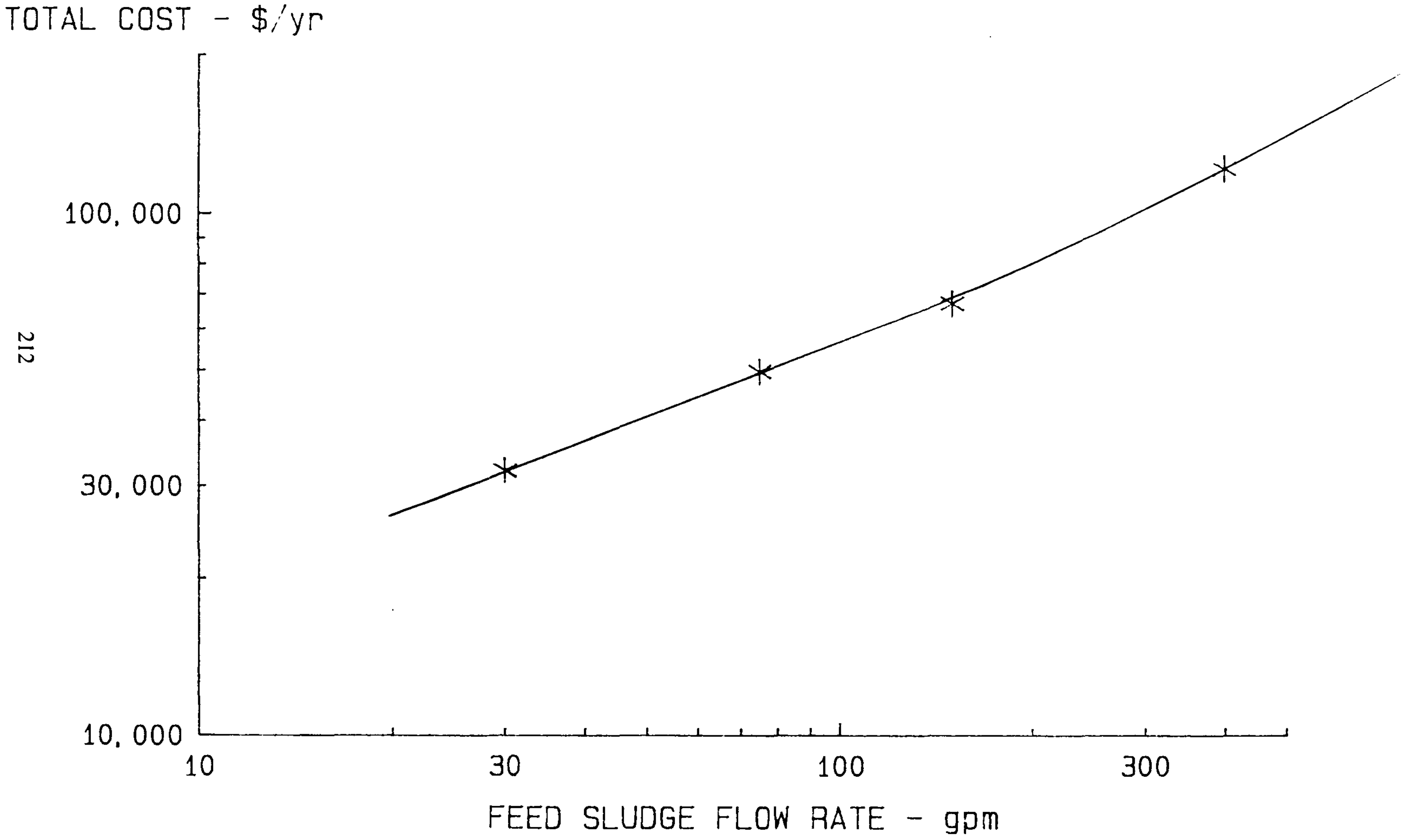


FIGURE 4-27

OPERATION AND MAINTENANCE COST FOR DECANTER CENTRIFUGES



Process energy costs include operating horsepower usage for main and back drive units. The process energy does not include energy required for feed sludge pumps or dewatered sludge handling. Maintenance material costs represent replacement parts, resurfacing of the scrolls and includes general maintenance.

Due to their scarcity in this field and overall unpopularity when compared to decanter centrifuges, basket centrifuge costs are not presented here. Generally, the basket centrifuge capital cost is essentially a straight line as a function of sludge flow rate. This is because basket centrifuges are available in limited sizes and as the flow increases the number of centrifuges and building size, etc. increases accordingly.

The difficulty in using the centrifuge cost curves given, is the curves were developed in terms of hydraulic (liquid) capacity. But as stated earlier, solids loading may well be the limiting criteria. For example, 50 gpm of sludge at 2% solids may be the same cost as 25 gpm at 4% although the curve would not reflect this. Therefore, clearly some knowledge of centrifuge performance is necessary from pilot plant tests. The cost curves have been developed for very weak feed sludges (1% or less), such that an assumption of hydraulic limitation was reasonable.

4.6.4. Operating Considerations

Operating considerations for centrifuges include feed flow rate, rotational speed, differential speed of the scroll, depth of settling zone and the controlled properties of the sludge (e.g., temperature, percent solids, etc.) To a point, the higher the rotational speed used and resulting "g" force applied, the dryer the cake that can be obtained from the centrifuge. However the wear and tear on the centrifuge is proportional to the cube of the bowl speed and thus precludes the use of excessively high "g" forces.

The reported cake dryness from solid bowl centrifuges varies depending on the many operating considerations. The feed rate has been seen to be a primary operational gauge of the ultimate performance. The best performance data has been obtained at 75 to 80 percent of the optimal hydraulic or solids capacity of the centrifuge. Lower polymer dosages are required and a dryer cake is achieved at this slightly lower than optimal capacity.

Another operational consideration is the solids concentration of the sludge feed. It has been demonstrated that for a particular set of centrifuge operational constraints, a well controlled feed concentration will produce consistently better results than a varying concentration. For centrifuges, it is usually advantageous to vary the hydraulic feed rate and to hold constant the solids loading rate for incoming sludges that have changing percent solids concentrations. This calls for an equalization/thickening facility prior to the centrifuge itself.

If the centrifuge is not to be used for any significant time (24 hours or more), the inside of the bowl needs to be washed down with significant quantities of water. If not washed, the solids remaining in the bowl will dry and possibly cause unbalanced operation.

4.6.5. Past Performance

Centrifuges have been utilized for treatment of alum sludges and lime softening sludges. Scroll centrifuges have generally been the chosen method when comparisons have been made to basket centrifuges. Nielsen et al. (4-6) conducted pilot studies comparing several dewatering alternatives including basket and scroll centrifuges. Conditions were evaluated under what was considered annual average (nominal) sludge production conditions and overload (high sludge production) conditions, as shown in Table 4-2. Under nominal loading conditions the scroll centrifuge produced a 15 percent

TABLE 4-2

ANALYSIS OF CENTRIFUGE OPERATING PARAMETERS

Parameters	Scroll Centrifuge	Basket Centrifuge
Solids loadings (dry Basis)		
Nominal		
Per unit	9,500	9,000
Total	19,000	19,000
Overload (overload conditions = 60,000 lb/day Per unit)	34,000	22,000
Total	68,000	66,000
Machine requirement		
Machine size	28 gpm	120-gal bowl
Assumed equipment life -- yr	20	20
Number of nominal load	2	2
Number of overload	2	3
Solids output		
Nominal		
Solids by weight -- percent	16	15
Gpd of wet cake	12,700	13,500
Overload		
Solids by weight -- percent	16	15
Gpd of wet cake	45,400	73,700
Polyelectrolyte dose -- lb/T		
Nominal load	3.0	4.5
Overload	8.0	5.0

Source: Nielson (4-6)

cake at a polymer dose of 3 lb/ton. The same result could be achieved with a basket centrifuge using a polymer dose of 4.5 lb/ton. Under high solids loading conditions the scroll centrifuge could still produce a 15% cake by increasing the polymer dose to 8 lb/ton, whereas, the basket centrifuge could only produce a 10% cake. Under high solids loading to the basket centrifuge a long cycle time was required to produce a 15% cake, hence making that alternative uneconomical.

Somewhat different results were obtained in work done by Westerhoff and Daly (4-7) treating an alum sludge. Using a scroll centrifuge as shown in Table 4-3 they obtained a 25% cake with a polymer dose of 3 lb/ton. With a basket centrifuge an 11% cake was achieved with no polymer. It is not clear from the reference whether the polymer was not tried with the basket centrifuge or if it was tried and did not improve the results.

Hagstrom and Mignone (4-8) reported results of testing 5 different alum sludges using a basket centrifuge. Their results are shown in Table 4-4. Polymer was used for conditioning but doses were not reported in comparable units. A general trend was evident of achieving a higher final cake solids concentration as the raw water turbidity increased.

Data reported for centrifugation of lime softening sludges appears to be limited to the use of scroll centrifuges. Albertson and Guidi (4-9) using a 10% feed concentration achieved a 57% final cake with 75% solids recovery (25% of the solids were lost in the centrate) reducing to a 52% cake with 90% recovery. The operational change made to achieve a higher capture was an approximate one-half decrease in the feed flow rate to the centrifuge (10 gpm to 5 gpm). Thus, a key variable in sizing the centrifuge is the desired solids capture percentage. At a 10% feed and even 90% capture, the centrate is still a 1% solids concentration--not acceptable for disposal but perhaps could be recycled to the treatment process. They also found that by using an

TABLE 4-3

RESULTS OF CENTRIFUGATION EVALUATION

Parameters	Scroll Centrifuge	Basket Centrifuge	
Coagulation Basin Sludge			
Feed concentration -- percent	3.0(2.7-3.1)	1.5	
Cycle time -- min	continuous	40	
Polyelectrolyte feed - lb/ton	3(2-4)	0	
Cake solids - percent	26(24-28)	11	
Filtrate suspended solids -- mg/l	540	150	
Filter Backwash Sludge			
Feed concentration -- percent	1.5(1.0-2.0)	1.2(1.0-1.5)	
Cycle time - min	continuous	20	80
Polyelectrolyte feed -- lb/ton	2.5(2.0-3.0)	0	1.0
Cake solids -- percent	18(18-22)	5(4-6)	10.0
Filtrate suspended solids -- mg/l	550	150	150

Source: Westerhoff and Daly (4-7)

TABLE 4-4

RESULTS OF BASKET CENTRIFUGE TESTS

	Raw Water Turbidity	Feed Solids %	Cake Solids %
Plant A	10	2 - 3.6	15 - 18
Plant B	22	1.7 - 2.8	18 - 22
Plant C	33	0.5 - 1.5	25 - 33
Plant D	30	1.1 - 4.4	30 - 35
Plant D	107	1.5 - 2.5	33 - 38

Source: Hogstrom and Mignone (4-8)

anionic polymer 90 to 100% solids capture could be achieved but the cake concentration was reduced to 45%. The AWWA Research Foundation report in 1969 (4-10) summarized operational data from four lime plants using centrifuges. It found final cake solids concentrations ranging from 55 to 65% with a centrate quality of 500 mg/l to 10,000 mg/l suspended solids concentration.

4.6.6. Example Facility

Chesapeake, Va. The City of Chesapeake, Virginia owns and operates a 10 mgd conventional treatment plant. The plant was constructed in 1978 and has produced 9.0 mgd of finished drinking water on average. Summer water production has been as high as 13 mgd.

The plant was originally constructed with rapid mixing, flocculation, settling and filtration processes. The source of water is the Northwest River which is typically very high in color and low in turbidity. The plant operates primarily as a color removal plant and has operated at a coagulation pH of 5.8 and an alum dose of 80 to 100 mg/l (as dry alum). Typical raw and finished water quality is shown in Table 4-5. An addition to the plant in 1984 included air stripping towers for control of trihalomethanes. The plant includes a solids handling facility consisting of a wetwell and sludge pumping facilities, screw pumps, backwash water clarifier, sludge thickener, and two centrifuges. A schematic of the sludge facility is shown in Figure 4-28.

Alum has been used as a coagulant in the plant and polymer as a coagulant aid. Alum sludge collected in the settling basins is removed continuously with travelling bridge sludge collectors. Sludge from the collectors is stored and thickened in a sludge thickener. Filter backwash is settled in a backwash clarifier designed to remove settleable solids from backwash water prior to recycle (or discharge to a lagoon). Sludge from the backwash clarifier is sent to the

TABLE 4-5

RAW AND FINISHED WATER QUALITY

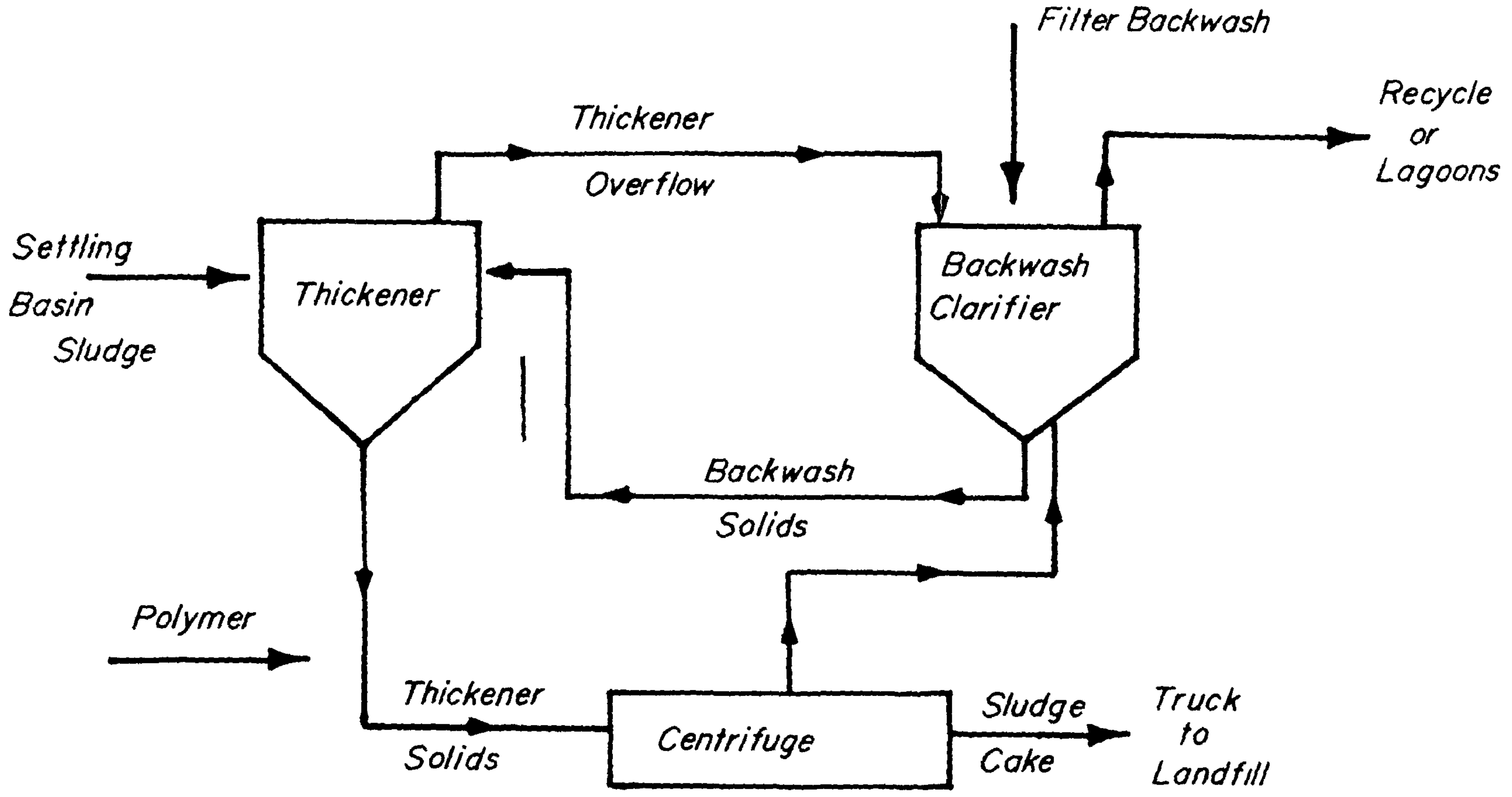
CHESAPEAKE, VIRGINIA

Parameter	Raw	Finished
Color	100 - 400 CU	0 - 10 CU
Total Organic Carbon (TOC)	20 - 40 mg/l	5 - 15 mg/l
TTHM 7-Day Formation Potential	2000 - 4000 ug/l	100 - 500 ug/l*
Chlorides	10 - 500 mg/l	10 - 500 mg/l
Turbidity	2 - 20 TU	.25 - .50 TU

* Without air stripping

FIGURE 4-28

CHESAPEAKE SOLIDS HANDLING FACILITY PROCESS SCHEMATIC



thickener and overflow from the thickener is sent to the centrifuges which have been utilized for sludge dewatering. Centrate from the centrifuge has been returned to the backwash clarifier.

The facility has produced dewatered sludge ranging from a 9% to 13% solids concentration. Average solids concentration has been 11%. Feed solids concentrations have ranged from 1% to 3% and have averaged 2%. Figure 4-29 shows actual sludge feed and cake solids concentration from the centrifuge operation. Solids production has ranged from 1,200 to 8,000 lbs per day.

Polymer has been added as a dewatering aid in the feed to the centrifuge. Polymer dose has varied from 10 to 40 lbs per ton of solids.

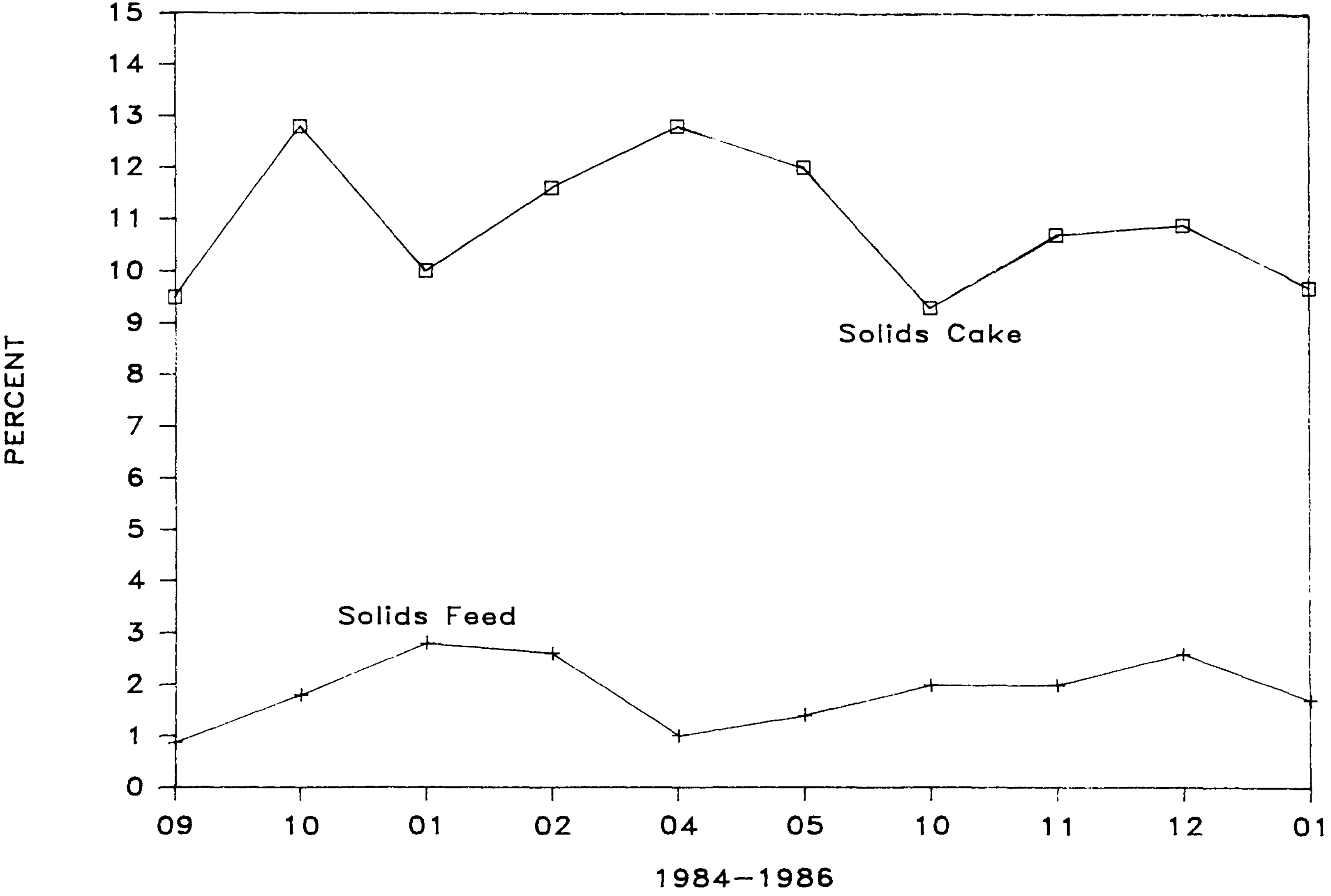
The time of operation of the solids handling facility and centrifuge has been primarily based on sludge production in the settling basins and backwash operation. Sludge has been continuously withdrawn from the settling basins and stored in the sludge thickener with the solids removed from the backwash clarifier. Operators periodically check sludge levels in the thickener and operate the centrifuges once a sludge depth of approximately 5 feet has accumulated in the thickener. Sludge is fed to the two centrifuges at a rate of approximately 20 to 30 gpm. The solids loading rate to the centrifuge has been approximately 120 pounds/hr per centrifuge.

Two centrifuges have been used to dewater the sludge. At start-up of the dewatering operation both centrifuges are used. When the sludge level begins to drop to about 3 feet in depth, one centrifuge is turned off and one centrifuge operated until most of the dewaterable solids are removed from the basin. Operationally, this point is seen by the "sloppy" cake from the centrifuge and checking of the sludge level in the thickener. During low water production periods (and low solids production) the facility has been operated approximately 2 to 3 days a week. Under maximum solids loading

FIGURE 4-29

CHEASAPEAKE SLUDGE

FEED AND SLUDGE CAKE SOLIDS



conditions the facility is operated almost continuously with the operation of one or two centrifuges as necessary.

The operators of the solids dewatering and centrifuge facility watch both the centrate color and turbidity, and cake solids concentration to make adjustments to the dewatering operation. Adjustments made by the operators include polymer dose, sludge pump feed rate and centrifuge rpm. The adjustments made to the dewatering facility has been primarily the responsibility of the operators and has been based on operational experience with the sludge over the years. Typically, adjustments in polymer feed and centrifuge speed have resolved any problems with sludge cake that is too wet to be hauled to the local landfill.

The disposal of the dewatered sludge has been to a municipal landfill owned and operated by the City. A separate portion of the landfill has been set aside for the alum sludge. The water utility owns and operates its own container trucks for hauling to the landfill.

Operational problems associated with the sludge facility have been minimal. The clogging of sludge feed lines or polymer feed lines with trash has been a problem periodically. The most significant problem the plant has experienced has been related to the water that is generated from the backwash clarifier and thickener and recycled to the plant rapid mix. During much of the year the raw water contains high levels of iron, manganese, and natural organic material. When sludge containing these materials is left for extended periods (2 to 3 days), the sludge becomes anaerobic and has resulted in very high levels of color, iron and manganese in the water being recycled. These high levels of color in the recycled water have been a continuous problem despite increasing oxidant dosage and feed points in the plant. The problem has been resolved by eliminating recycle into the plant and discharging the water after settling in lagoons located on the plant site (originally installed as emergency sludge storage lagoons.)

The lagoons discharge to the river upstream of the intake this but has not been a problem to date.

Grand Rapids, Mi. The City of Grand Rapids, Michigan owns and operates two water treatment plants, one of which (Monroe Filtration Plant) utilizes a centrifuge for sludge dewatering. The Monroe Plant operated as a softening and turbidity removal plant and treats surface water from the Grand River. On average, an alum dose of 15 mg/l and lime dose of 150 mg/l has been used. Typical raw and finished water quality is shown in Table 4-6.

The plant has operated primarily as a peaking plant during summer months to supplement a 60 mgd conventional coagulation plant located near Lake Michigan. The Monroe Plant has produced an average of 20 mgd of treated water and 40 mgd during peak periods.

A centrifuge was installed at the Monroe plant in the early 1960's after the direct discharge of sludge to surface water was no longer allowed and discharge to the wastewater plant proved unsuccessful. The centrifuge was selected by the City after testing conducted by manufacturers showed that the lime sludge could be successfully dewatered. The sludge facility has operated successfully for over 20 years and consists of a single centrifuge which has been fed sludge directly from the settling basins (See Figure 4-30). Solids have been removed from the basins by the use of a two stage mechanical cleaning mechanism and feed pumps which has allowed some flexibility in controlling the feed solids concentration. The feed solids concentration to the centrifuge have been maintained in the 8 to 10% range which has been optimal based on operational experience. Higher solids concentrations have resulted in pumping problems with the sludge feed pumps and 4-inch piping feeding the centrifuge. Sludge has been pumped directly into the centrifuge at rates of 30 to 100 gpm with a 10% solids concentration (about 1,600 lb/hr at 30 gpm). On average about 2,300 lb/MG of solids has been produced with the

TABLE 4-6

MONROE PLANT - TYPICAL WATER QUALITY

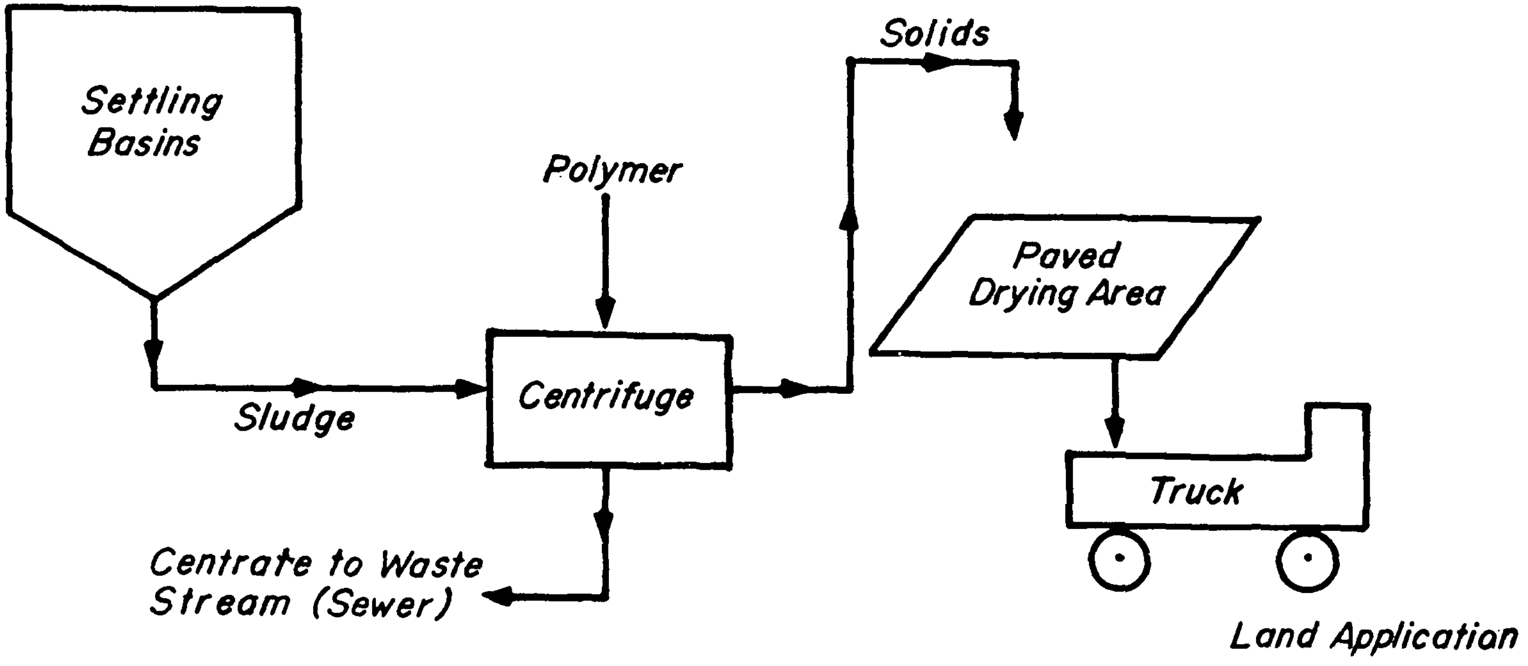
GRAND RAPIDS, MICHIGAN

1985 - 1986

	RAW	TAP
TOTAL HARDNESS (as $CaCO_3$)	239 mg/L	128 mg/L
TOTAL ALKALINITY (as $CaCO_3$)	186 mg/L	59 mg/L
PHENOLPHTHALEIN ALKALINITY (as $CaCO_3$)	1.6 mg/L	32 mg/L
NON-CARBONATE HARDNESS (as $CaCO_3$)	53 mg/L	71 mg/L
CHLORIDES	26 mg/L	32 mg/L
CALCIUM (as Ca)	61 mg/L	35 mg/L
MAGNESIUM (as Mg)	21.0 mg/L	9.9 mg/L
FLUORIDE (as F)	0.14 mg/L	0.21 mg/L
TOTAL IRON (as Fe)	0.49 mg/L	0.02 mg/L
COLOR	22 units	1 units
pH	8.2 units	10.1 units
TOTAL RESIDUE	394 mg/L	236 mg/L
SODIUM	12.1 mg/L	12.6 mg/L
BICARBONATE ALKALINITY (as HCO_3)	208 mg/L	0 mg/L
CONDUCTIVITY	557 umHos/cm	363 umHos/cm
TURBIDITY	14 NTU	0.20 NTU
CARBON DIOXIDE	3.0 mg/L	0 mg/L

FIGURE 4-30

GRAND RAPIDS SOLIDS HANDLING FACILITY PROCESS SCHEMATIC



plant operating for 8 to 12 hours at a 20 mgd rate during summer months.

Polymer has been added as a flocculant in the centrifuge at a dose of 2 to 4 lbs per ton of solids. Operational experience has shown that a sludge of 50 to 60% solids concentration could be produced without the use of a filter aid, however, the centrate contained from a 3 to 5% solids concentration. With the use of a polymer a 25% to 30% cake is typical with a 0.2 to 0.3% solids concentration in the centrate. The centrifuge has typically achieved over 95% solids capture with the use of polymer.

The solids produced from the centrifuge have been pumped to a drying area which consists of an asphalt slab located next to the plant. The sludge has been distributed to a depth of 18 to 24 inches over the area with a devise similar to a rotary spray irrigation system. This has allowed for spreading of the sludge over the drying area as it is pumped. The solids have been allowed to dry outside before being hauled by a local contractor. (The cost in 1986 for contractor hauling and disposal of the sludge has been $5.00 a ton). Since the operation takes place during summer months the sludge dries into "dry chunks" within a few weeks and can be easily loaded and hauled in a dump truck. The sludge has been used by local farmers as a liming material.

Operationally the centrifuge has been monitored by watching sludge solids in and out of the centrifuge and centrate water quality. Plant personnel have developed a laboratory vacuum filtration test using a buchner funnel to provide quick information on solids concentrations. A correlation of the "wet weight" of the solids to dry weight has shown a very good correlation and has provided operators with a quick method of checking process operations and effects of changing process conditions. Operators' primary responsibility during operation is to maintain feed solids concentrations and centrate quality. Changes in sludge basin scraper operation and feed pump rates have been used primarily for controlling the feed solids concentration. Machine parameters such as centrifuge speed are seldom changed. Polymer feed has been adjusted based on maintenance of a clear centrate from the centrifuge. Overall the system has performed well for the City and few problems have been encountered with the current dewatering and drying operation.

4.7. PRESSURE FILTER PRESS

4.7.1. Description

The basic concept of pressure filtration is the separation of water from a liquid sludge slurry using a positive pressure differential as the driving force. The pressure filter press is also called pressure filter, plate and frame press, recessed plate press, filter press and (with modification) diaphragm filter press. The pressure filter contains a series of filter plates supported by and contained in a structured frame. The plates are designed such that when two adjacent plates are brought together a compartment/chamber between the plates is formed to hold sludge. The plates are pushed tightly together, by hydraulic or electromechanical means, to make the compartment leakproof.

Lining the compartment is a cloth media which is porous enough to retain the sludge solids while releasing the water in the sludge. While the sludge is being pumped into the compartments, the solids are retained and the water released from the pressure filter press. The sludge pumping continues after the compartments are full (thus pressuring the compartment) until the solids concentration of the cake in the compartment is at an optimum value. Then the pumping is ceased, the plates separated and the retained sludge cake released by gravity for ultimate disposal.

The above describes one "cycle" of pressure filter press. At the end of one cycle the plates can be automatically realigned for loading the next batch of sludge.

During a filtration cycle, the pressure filter press will experience variations in sludge flow rate and pressure. At the beginning of the cycle the flow is at the maximum, and pressure at the minimum. At the end of the cycle the reverse is true and the cycle is terminated when the peak design pressure and the low feed flow rate is reached.

The original pressure filter press was the plate and frame type as described above. Improvements in the plate and frame press have included plate shifting mechanisms and better cycle controls.

A variation of the plate and frame filter press is the diaphragm filter press. The diaphragm filter press utilizes a flexible membrane/diaphragm across the face of each recessed plate to squeeze the forming cake. The filtration process proceeds the same as with the plate and frame press, only after the press is filled with sludge and the cake is forming, the diaphragm begins to squeeze. Diaphragm filter presses usually produce thinner cakes and consequently have shorter cycle times then plate and frame filter presses.

As with other mechanical dewatering devices it is very difficult to predict performance, design or cost from theoretical equations or laboratory analysis. State-of-the-art is such that pilot plant testing is necessary. Pilot filter presses are generally available in two types: those that simulate full scale performance and those that are in reality small scale production units. Simulation units have plates which are generally less than 1 sf of filter area and contain only 2 to 4 plates. They can provide information on conditioning requirements, final cake solids concentrations, filtrate quality and cycle time. They are obviously less expensive to use then prototype units due to lower sludge feed requirements, and lower shipping and rental fees. Manufacturers will often evaluate sludge shipped by the utility in their own lab using these units. Depending upon the total funds to be spent by the utility, results from these simulation units may be satisfactory. Generally, however, the sums of money to be spent for the full scale plant are sufficient to justify the cost of testing prototype units. In testing a prototype unit the utility should be sure that the equipment is essentially the same as the full scale unit. This includes such items as pressurization mechanisms (pump types, etc.), degree of automation, cleaning and wash methods and cake release methods. Beyond the mechanical and operational similarities, the general items to consider in pilot plant studies discussed earlier should also apply.

If a production unit has been pilot tested, then scale-up is fairly straightforward. The final results of pilot testing will be total solids to be treated (this is solids produced via water treatment plus solids added during conditioning, which can be as much as 30%), cake thickness (and therefore plate chamber volume required) and cycle time for one batch press operation. Scale up then is a matter of calculating the total plate volume requirements and picking the number and dimensions of appropriate units.

4.7.2. Design Considerations

As discussed above, the capacity of pressure filter presses is determined by the number of plates, the size of the plates and the thickness of the cake.

Plate and frame press manufacturers have three standardized chamber thicknesses: 25,30 and 40 mm (0.98, 1.18 and 1.57 in.). Filtration tests determine the most economical cake thickness.

The plates are provided in many designs, sizes, shapes and materials. Most plates are recessed and come in sizes from 0.5 to 2.0 m (2.0 to 7.0 ft). Plates can be circular, square, or rectangular. Plate materials include cast and ductile iron, coated steel and polypropylene. Materials for filter plates should be evaluated for strength, cost and mass affecting both handleability and structural support costs.

The number of plates in a filter press can affect performance. Poor distribution of sludge to remote chambers is possible when feeding well-conditioned sludge to a long press (greater than 80 plates).

Plate and frame filter presses are available operating at medium (700 k Pa or 100 psig) or high filtration pressures (1600 k Pa or 225 psig). Higher pressure usually results in a thicker cake with a higher solids concentration. However, research has determined that when higher pressures are applied to a compressible sludge, the solids become tightly packed which increases resistance and decreases filtration flow. Proper sludge conditioning is usually required to alter the compressible sludges characteristics.

The filter media lining each plate is almost exclusively of the monofilament type, ie., woven from a single filament rather than twisted together of many fibers, as with multifilament. The multifilament media has been found to blind in some applications due to swelling of the yarn. Filter media are rated by particle retention and permeability. Additional factors in the selection of filter cloth media are durability, cake release, blinding and chemical resistance. Air permeability is a measure of the openness of the filter cloth, determined by air flow through a unit area of media at a given pressure drop [for example, 800 $L/m^2/h$ at 0.1 kPa (100 cfm/sf at 0.5 in. H_2O)]. Air permeability is a measure of comparison between filter cloths. However, it doesn't represent the actual working permeability which is affected by strand swelling, solids impregnation and weave distortion. Also, it has been determined that early in the cake formation stage the cake itself acts as the filter and the filter media permeability is less important. Low permeability media (less than 250 $L/m^2/h$) are generally utilized on sludges which are relatively easy to dewater. While low permeability media yields high solids capture values, a greater tendency for media blinding, poor cake release and difficult cleaning would be expected. High permeability media (greater than 1500 L/m^2/h) are generally utilized on difficult to dewater sludges. High solids retention during early cake formation is forfeited for improved cake release and less media blinding.

The pumps feeding the sludge to the pressure filter press must be capable of operating under widely varying conditions. At the beginning of the filling cycle there is very little discharge pressure head, but substantial flow. Near the end of the filling cycle when the chamber is full and the sludge is being compressed, the pressure head is large and the flow small. It is also important that the flocs formed in conditioned sludges are not sheered while being pumped or pressurized. Generally positive displacement pumps (like plunger, piston or progressive cavity pumps) are better than centrifugal pumps in retaining the floc integrity. Some facilities have included two sets of pumps: one set of high capacity, low head pumps for filling the press and a second set of low capacity high head pumps for pressurization.

Pneumatic sludge feeding has been successfully demonstrated, but the initial capital cost is much greater than that for the pumped system. With the pneumatic system, the sludge is conditioned in a pneumatic tank and forced into the press with compressed air.

All filter media will eventually clog (blind) and washing is required. If the sludge conditions cause calcium carbonate to plate out on the media, an acid wash is required. The acid recirculatory system is by far the most common acid wash method. The acid dissolves the calcium carbonate and the recirculating motion dislodges loose objects from the media.

A second acid wash system which has been practiced is the acid soaking system. The press is simply filled with acid and allowed to stand, usually overnight. A disadvantage in the soaking system is that pockets of gas usually form at the top of the press and prevent the acid from coming in contact with all the media.

The washing process can be costly and time consuming. A precoat system can be added to a pressure filter press to help prevent premature blinding and prolong the period of time between washings. Precoat systems also assist in cake release by providing a film between the cake and the media.

In a precoat system, wash water is circulated through the press at high velocities at the beginning of the filtration cycle to dislodge and remove most particles on the media. The wash water is stored in a holding tank and can be used several times before needing to be discarded. After washing a mixture of water and a small amount of filter aid (usually either fly ash or diatomaceous earth) is circulated through the press to coat the media. The precoating step takes about 3 to 5 minutes.

The above described system is termed a dry material feed system, which is used primarily on the larger installations. The wet system is quite similar to the dry system except a precoat material preparation tank with mixer is utilized instead of a precoat tank.

A typical fly ash application rate is 4.8 kg/10 sq. meters (10 lbs/100 sf) compared to 2.9 kg/100 sq. meter (6 lbs/100 sf) for diatomaceous earth. Fly ash, however, is a more economical filter aid than diatomaceous earth, because the purchase price of diatomaceous earth is generally 10 to 20 times that of fly ash.

The precoat pumps are generally sized to pump at the rate of 0.2 to 0.3 $L/m^2/s$ (0.3 to 0.5 gpm/sf) of filter area. The filtrate storage tank is generally twice the volume of the press.

When dewatering hydroxide sludges with plate and frame filter presses the addition of a sludge conditioner has been the standard practice. The conditioner can either be a flocculating agent (like polymer), or a flocculating/bulking agent (like lime and fly ash). Lime has been used almost exclusively as the conditioning agent.

In conditioning of alum sludges, lime has been added in sufficient quantities to adjust the sludge pH to about 11. The amount of lime added is a function of the desired yield--the higher the lime dose the higher the yield. Lime doses can be as much as 20 to 30% by weight.

A decreased lime requirement has been demonstrated in the two stage conditioning system. A portion of the lime is initially added to the sludge, followed by polymer addition. After a short residence time, the supernatant is decanted and additional lime is added to raise the pH to about 11. Different conditioning systems and variations should be tested on the pilot scale before full scale implementation. (Additional discussion on sludge conditioning is included in the "Operating Considerations" Section of this discussion). Studies particulary with diaphragm presses have had success without any conditioner, provided a thin cake is produced.

The filter press system and appurtenant systems are shown as Figure 4-31, 4-32 and 4-33. Appurtenant systems include conditioning and precoat/filter aid systems.

If lime conditioning is utilized, the filtrate from the pressure filter press is operationally at a pH value of about 11^+, requiring special attention. This filtrate may not be of quality to be suitably returned to the head of the water treatment plant. These cases require the filtrate to be treated and/or disposed of separately. Another waste stream generated by a filter press operation is the acid wash waste. Generally, neutralization and discharge to a sewer system has been allowed.

4.7.3. Capital and Operating Costs

Cost curves for the diaphragm filter press are presented in this Section. While filter presses are still available without the diaphragm feature, diaphragms are becoming increasingly popular and only the costs for diaphragm presses are presented here.

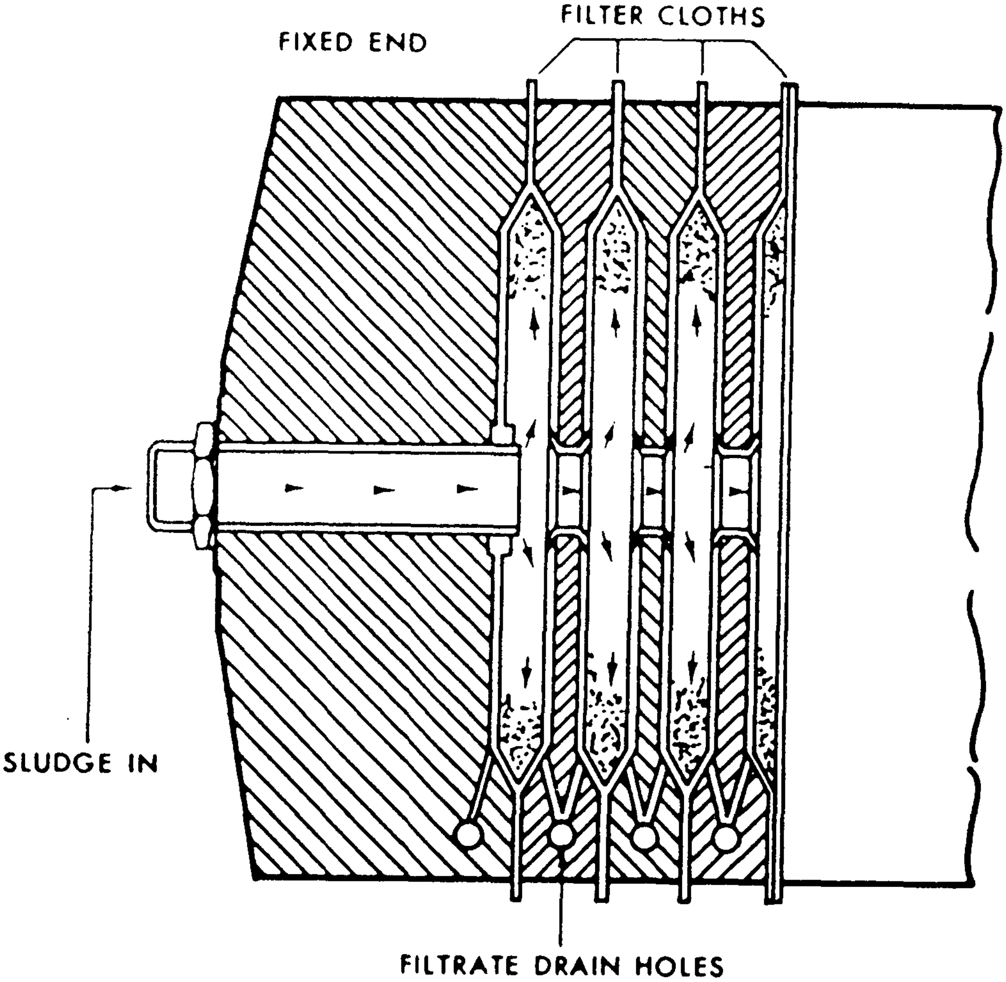


FIGURE 4-31

CUTAWAY VIEW OF A FILTER PRESS

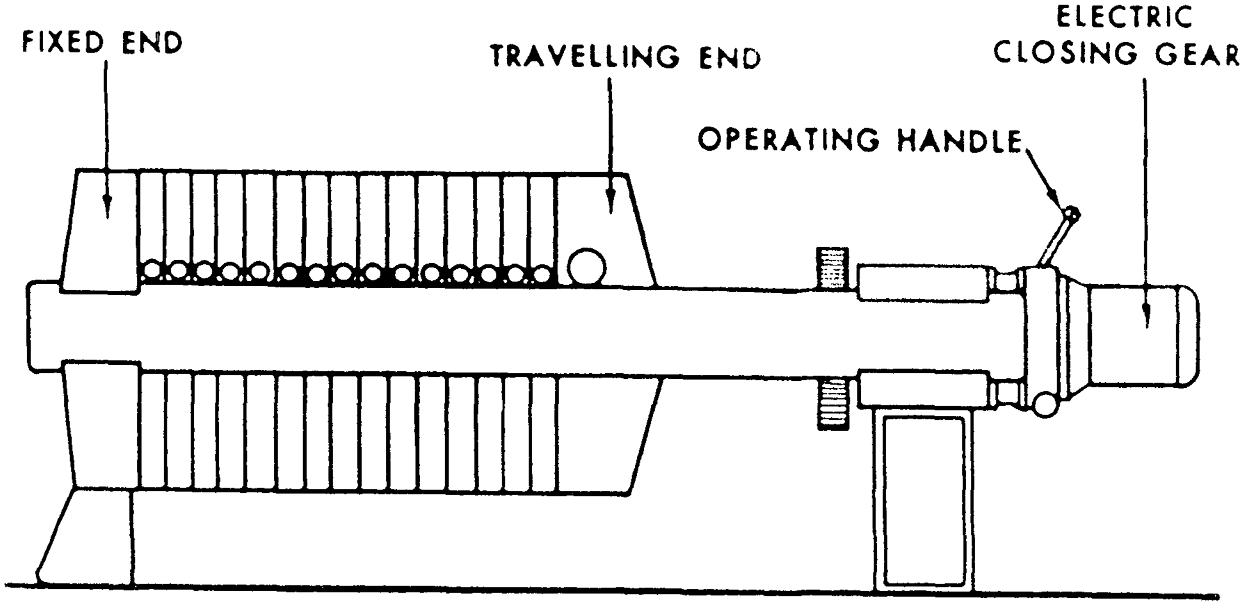
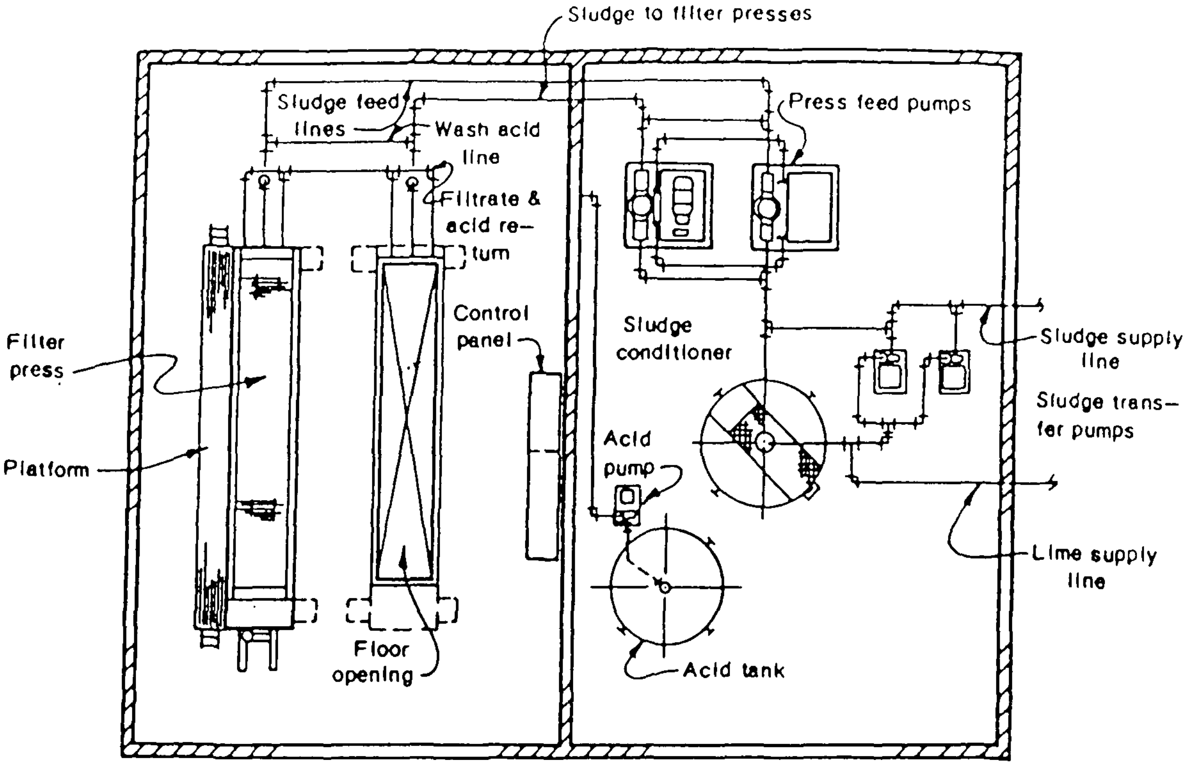
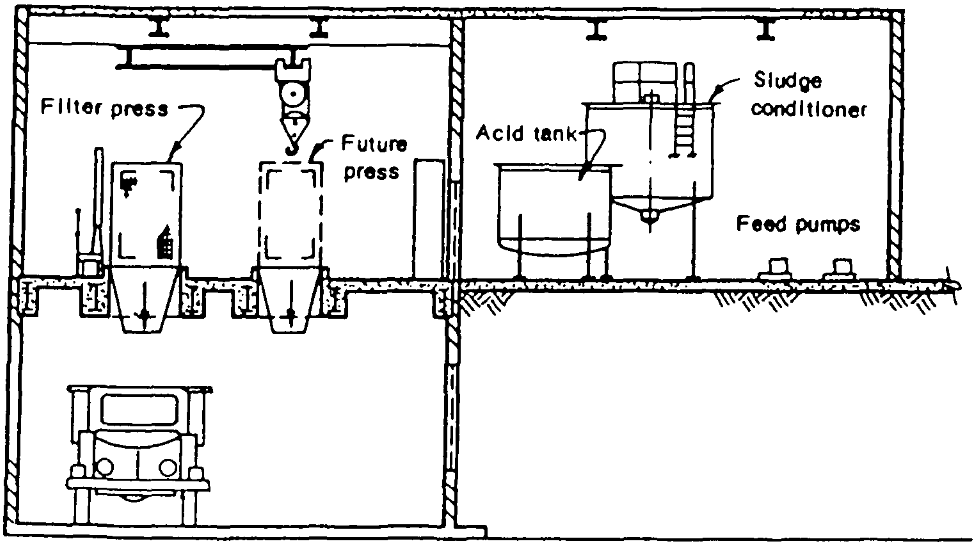


FIGURE 4-32

SIDE VIEW OF A FILTER PRESS



PLAN VIEW



ELEVATION VIEW

FIGURE 4-33

TYPICAL FILTER PRESS INSTALLATION

The construction cost curves are shown in Figure 4-34. The surface area is total area of press (both sides of plate). The construction costs include the costs associated with the equipment, labor, pipe and valves, electrical and instrumentation and two story housing. The equipment cost is broken down as filter press equipment, washer/shaker mechanism and ancillary equipment. The filter press equipment includes the cost for the diaphragm filter press and the hydraulic power unit. The washer/shaker mechanism is an added feature to aid in cake release. The washer mechanism includes wash water pump, tank and piping; the shaker is a plate shaker. The ancillary equipment includes feed pumps, sludge holding tank, a filtrate control valve and the air compressor system. Most modern filter press systems are recommended to use two types of feed pumps. Included in the ancillary equipment cost is the cost of one centrifugal pump for initial fill of the press and one progressive cavity pump for pressurized pumping. Standby pumping units can be added by including the cost of the appropriate sludge pumps per the previous section of this Chapter. Polymer and/or lime conditioning costs must be added if they are used.

The operation and maintenance cost curves are shown in Figure 4-35. The O & M costs include process energy, maintenance materials and labor costs. Process energy is predominately that consumed by the feed pumps, but also includes the plate shifting mechanism and ancillary equipment operation. Costs of conditioning chemicals and their systems are not included.

4.7.4. Operating Considerations

Probably the most important controllable factor that affects the rate of filtration after a particular pressure filter press is in operation is the conditioning of the sludge. The specific resistance test, the capillary suction time (CST) test and the high pressure filtration test can be

FIGURE 4-34

CONSTRUCTION COST FOR DIAPHRAGM FILTER PRESS

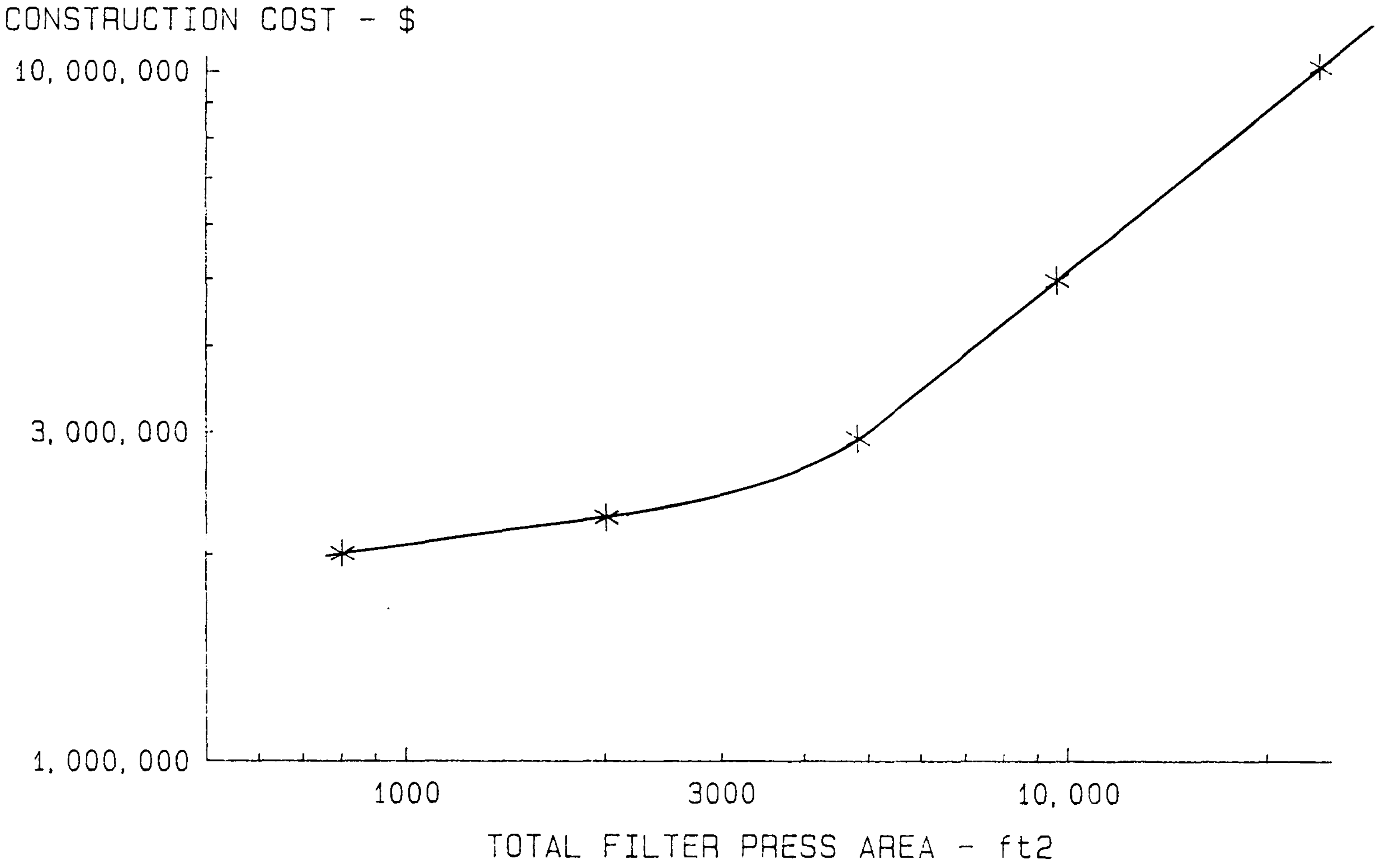
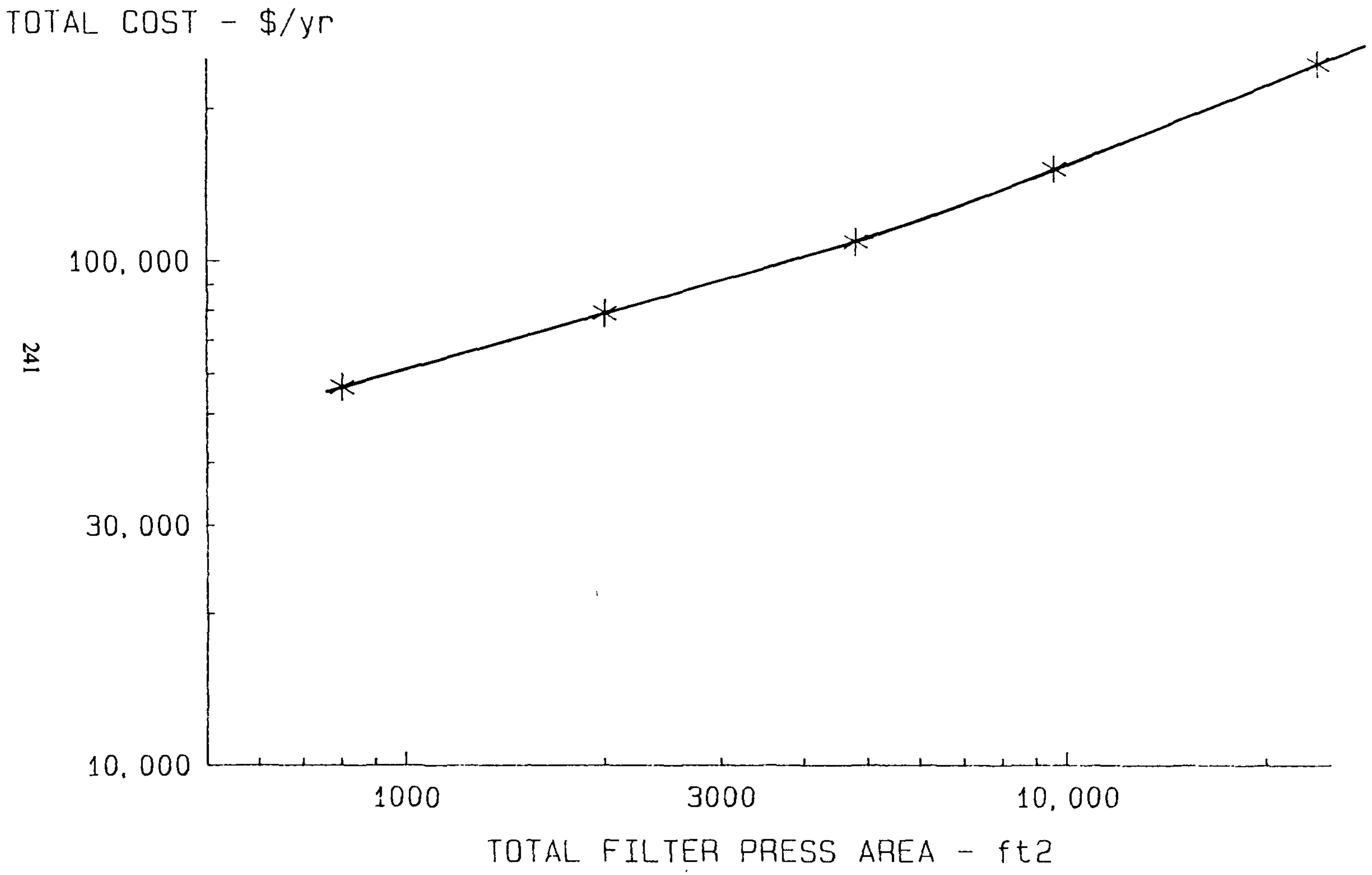


FIGURE 4-35

OPERATION AND MAINTENANCE COST FOR
DIAPHRAGM FILTER PRESS



used to measure the effectiveness of the conditioner used. Also it has been determined for pressure filtration, a compressibility coefficient of less than 1.0 is required. At values above 1.0 the specific resistance to filtration increases faster than the applied pressure differential, and a poor cake is formed. The addition of lime or other bulking agents can be added to reduce the sludge compressibility.

Proper maintenance of filter plates and media is critical in maintaining the design performance and preventing damage to the plates. Frequent cleaning of the media (perhaps as often as once every 8 to 10 cycles) will improve performance, reduce the time spent manually removing cake "stickers", and prolong the life of the media and possibly the plates themselves. Plates can be warped or broken if the media are nonuniformly blinded due to the pressure gradients created across the plate. Plate damage may be caused by uneven pressures in a chamber or between adjacent chambers. Media replacement may be required every 1000 to 4000 filter cycles.

Small amounts of leakage between plates is usually not cause for alarm. As the cake formation stage proceeds the leakage should reduce. Cleaning or removing of media wrinkles along the adjoining surfaces of the plates will generally end remaining leaks.

Filtrate cloudiness is a potential sign of poor press performance. At the beginning of the filter cycle some filtrate cloudiness is expected because the filtering cake has not yet formed. Continued cloudiness during the cycle may be a sign of a tear in the media or that the system pressure is too large or is pulsating.

Uneven filter cakes can be caused by a number of factors. Insufficient pressure or feed flow, or a clogged feed port can cause uneven filter cakes. Clogged feed ports can also create a differential pressure across a plate(one chamber full of sludge under pressure, the adjacent chamber empty) resulting in plate damage. A free filtering sludge in a bottom discharge press may experience rapid filtration at the chamber bottom resulting in an uneven buildup of cake along the full media surface. A top discharge press should be considered for this type of sludge.

Operational safety is generally provided by an electric light curtain which automatically shuts off the machine when an object crosses its path. This protects workers from injury when in the vicinity of the shifting plates. Photo-electric cells are stacked vertically in pairs across the front of the press. Each pair of cells emit a light beam which, when broken, activates the signal to stop the machine.

4.7.5. Past Performance

Filter presses have long been used to dewater wastewater sludges, but were first introduced to the water industry in the mid-1960's. As early as the mid-1800's filter presses were used successfully in England to dewater wastewater sludge without chemical pretreatment. In the U.S., the use of filter presses was first reported around the turn of the century in the northeast.

Although a mechanical process, the early filter presses were very labor intensive. In the early 1970's, filter press manufacturers began automating many of the manual tasks. Automatic plate shifting, cake discharge and washing features made the filter press more attractive. Also the machine capacity was greatly increased, making the presses more economical for the larger installations.

The major advantage of the pressure filter press when compared to the other mechanical dewatering equipment is the high solids concentration in the formed cake and the high clarity of the filtrate. Thus, filter presses have become increasingly attractive when cake disposal is a critical factor.

The first application of filter presses in water treatment in the United Sates was for the City of Atlanta in the

1960's. This was followed in the early 1970's by plants for the City of Houston, Erie County Water Authority, Jersey City Water Authority and Monroe County Water Authority. These units were generally recessed chamber filter presses operating at about 225 psi using lime as the conditioning agent. Precoating was often practiced using diatomaceous earth, fly ash or wood fiber.

Typical of results achieved for these units was pilot plant data reported in 1974 for Erie County Water Authority (4-7). In this study one unit was evaluated using a nylon filter cloth and requiring a precoat, while another unit used a polypropylene filter cloth without a precoat. Table 4-7 summarizes results of that testing. Cake concentrations are in the 35 to 40% solids concentration range depending on the feed solids concentrations and the conditioning method. The final flow schematic installed in Erie County, which included the precoat pressure filtration system, is shown in Figure 4-36. Comparative results of three full scale installations of the above type are shown in Table 4-8 (4-10).

Since the installation of the filter presses in the 1970's, comparisons are generally made between plate and frame presses, recessed chamber presses and diaphragm presses, or perhaps only the latter two. In 1985, Doe and Malmrose (4-11) reported on dewatering pilot plant studies conducted for Norfolk, Virginia. They compared results of two different diaphragm presses and one recessed chamber press. One of the diaphragm presses (Press A) had a single diaphragm while press B had two diaphragms. They converted results and reported them in terms of yield to produce a 30% solids concentration final cake. Yield is expressed as pounds of original sludge (without conditioning chemicals) which can be treated per hour per square foot of filter area (total filter area includes both sides of plates). Although this is not a usual method of reporting filter press performance, it does allow for a direct comparison between units and can be a useful method of data

TABLE 4-7

PILOT PLANT TESTING OF PRESSURE FILTER SYSTEMS ON ALUM SLUDGE

	PRECOAT UNIT		
	Coagulation Basin Sludge--Feed Concentration 4 per cent	Filter Backwash Sludge--Feed Concentration 2-3 per cent	Acidified Waste Residue--Feed Concentration 17 per cent
Cycle time - min	140	115	100
Lime dosage - per cent	16	36	18
Precoat - per cent	4	9	4
Cake solids - per cent	40	35	41
Filtrate suspended solids - mg/l	120	120	330
Filtrate pH	11 - 12	12	12

	NON-PRECOAT UNIT		
	Coagulation Basin Sludge Feed Concentration		Filter Backwash Sludge--Feed Concentration
	5.2 per cent	3.3 per cent	2.8 per cent
Cycle time - min	150	180	230
Lime dosage - per cent	14	12	18
Cake solids - per cent	37	35	31
Filtrate suspended solids - mg/l	30 - 60	30 - 60	30 - 60
Filtrate pH	11 - 12	11 - 12	11 - 12

Source: Westerhoff and Daly (4-7)

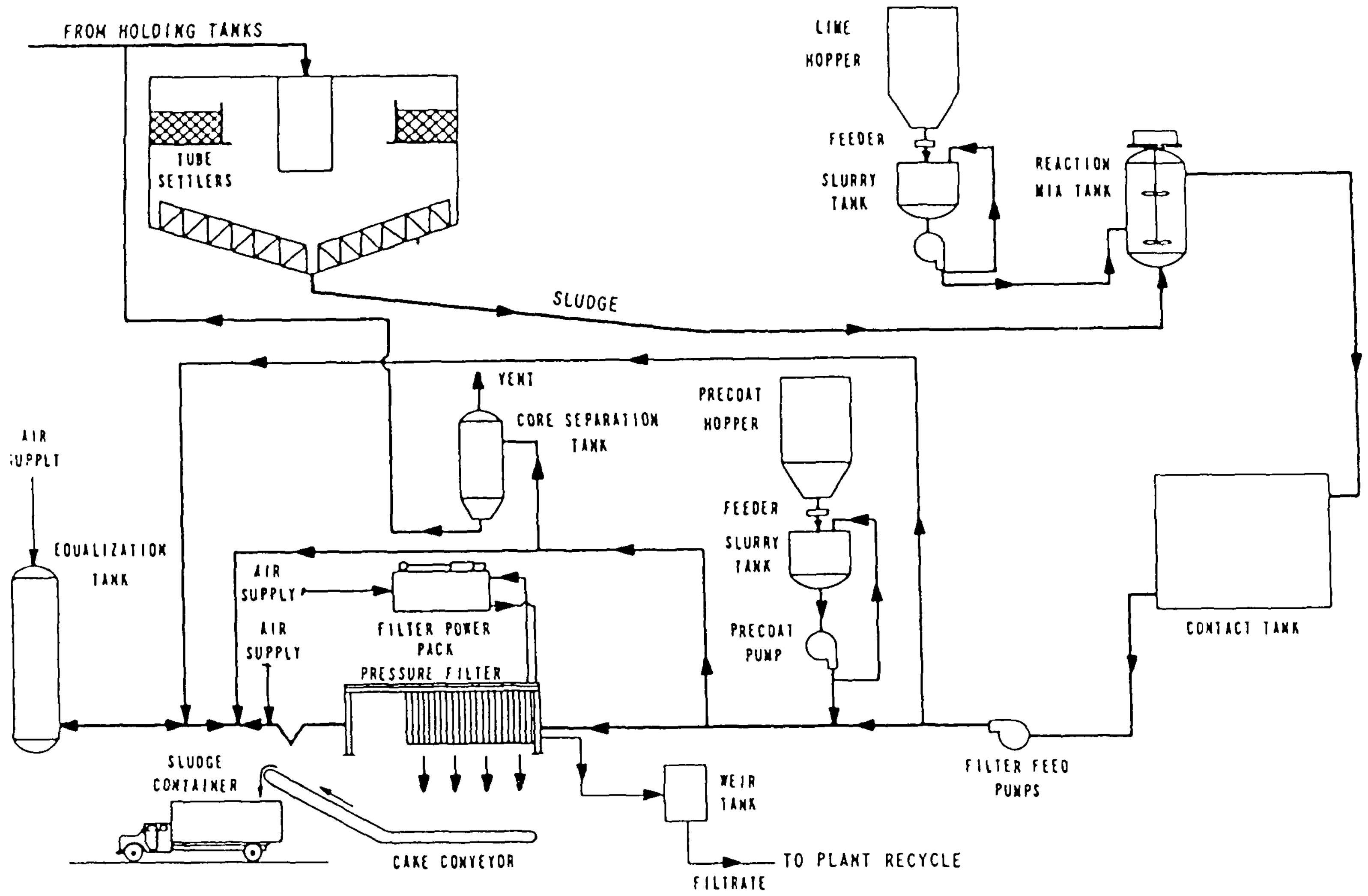


FIGURE 4-36

SLUDGE PLAN DIAGRAM
ERIE COUNTY WATER AUTHORITY, NY

TABLE 4-8

COMPARATIVE OPERATIONAL DATA OF RECESSED PRESSURE FILTERS TREATING ALUM SLUDGE

	Passaic Valley	Chattahoochee[1]	Hemphill[1]
Press Runs	40	49	76
Solids Processed (gal)	549,665	745,290	419,600
Avg. Gals Processed (each run)	13,741	15,394	5,521
Avg. Feed Solids %	3.8	5.0	1.5
Avg. Cake Dryness %	26	51.2	38.0
Total Solids Removed (tons)	87	233.1	383.2
Avg. Weight/Run (lbs)	4,360	9,514	10,084
Lime %	23	10	3
DE lbs/ton	50.57	13.56	18.1

[1] Atlanta, Georgia

Source: AWWARF report (4-10)

reduction. Table 4-9 shows the results of the testing. Only the diaphragm presses were able to achieve a 30% cake without any chemical conditioning. In general, the double diaphragm press required less conditioning agent than the single diaphragm press. Doe and Malmrose also reported cycle times for the various presses to reach the 30% cake. The recessed chamber press had an approximate 145 min cycle time compared to 55 min for press B. It should be noted that for many applications of filter presses, it is desired to achieve a higher percent solids concentration than 30%. Reaching a higher value would increase cycle times over that reported above and decrease yields. Additional conditioning agent may also be required.

Additional data on the performance of Press B at Norfolk show the relationships among some of the key operational variables. Figure 4-37 shows two example runs. Run 6 is for no conditioning at a 3% feed solids concentration, while run 8 is also approximately a 3% feed solids concentration but with 18% by weight lime conditioning. The definite relationship between squeeze time and final cake solids concentration is shown. In run 8 it appears that additional squeeze time would increase the cake solids concentration, whereas without conditioner the cake solids concentration seems to have maximized near 30%. Clearly as squeeze time increases, yield decreases. It should be noted that squeeze time is only one component of the cycle time. To it must be added the time to fill the chamber (called filtration time), the cake release time and cleaning time.

Data analysis by the manufacturer resulted in the development of empirical relationships for cake solids concentrations and yield based on the key operational parameters. The relationship was expressed as:

$$K_s = 21.1\ F_s^{0.079}\ F_t^{0.086}\ C_l^{0.003}\ S_t^{0.033}$$

TABLE 4-9

PILOT SCALE TESTING RESULTS OF FILTER PRESSES ON ALUM SLUDGE

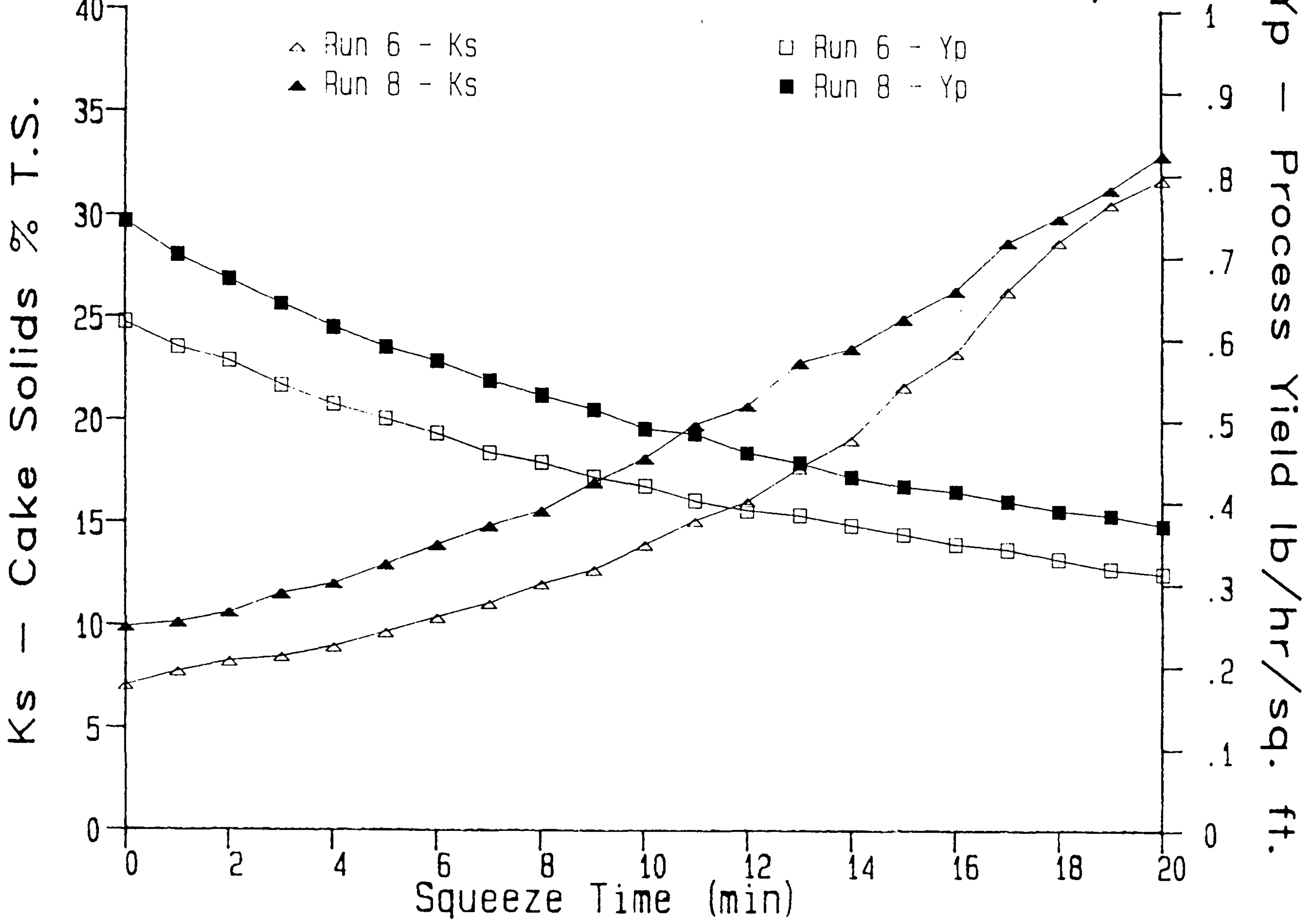
(Yields in lb/hr/sf)

Feed % Solids	Chemical Conditioning	Diaphragm Press A	Diaphragm Press B	Recessed Chamber Press
2	Lime	Yield - 0.25 40% Lime	Yield - 0.28 25% Lime	Yield - 0.21 30% Lime
2	None	Yield - 0.19	Yield - 0.14	---
2	Lime & Polymer	---	---	Yield - 0.31 30% Lime Polymer 6 lb/ton
2	Fly Ash & Polymer	Yield - 0.10 100% Fly Ash Polymer 2 lb/ton	Yield - 0.21 20% Fly Ash Polymer 1.6 lb/ton	Yield - 0.47 35% Fly Ash Polymer 5 lb/ton
4	Lime	Yield - 0.41 40% Lime	Yield - 0.43 20% Lime	Yield - 0.30 25% Lime
4	None	Yield - 0.31	Yield - 0.31	---

Source: Doe and Malmrose (4-11)

FIGURE 4-37

DIAPHRAGM PRESS PERFORMANCE AT NORFOLK, VA.



where,

K_s = cake solids concentration, %
F_s = feed solids concentration, %
F_t = filtration time, min
C_l = chemical lime dose, % by weight
S_t = squeeze time, min.

The correlation coefficient was a reasonably acceptable 0.82 and the sensitivity analysis clearly showed the relative importance of these parameters. Feed solids concentration was the most important variable, followed by filtration time (the time used to feed the chamber), chemical dose and squeeze time.

Similarly the process yield, Y_p was found with a 0.99 correlation coefficient as

$$Y_p = \frac{0.41 F_s^{1.024} C_l^{0.41}}{F_t^{0.37} S_t^{0.043}}$$

To determine the net yield the process yield must be adjusted to account for the cleaning cycle and turn around time simply to get the machine ready to be fed again (M_t):

$$Y = Y_p \left\{\frac{F_t + S_t}{F_t + S_t + M_t}\right\}$$

Filter presses have also been utilized in the dewatering of lime sludges although to a lesser extent than alum sludges. One of the first reported tests of lime sludges was reported by Burres et al. (4-12) on work done for Columbus, Ohio in the early 1970's. Pilot tests were conducted on several alternatives, which included filter pressing of lime sludge, containing about 10% by weight $Mg(OH)_2$. An initial 5% solids concentration sludge was thickened to 10% prior to feed to a recessed chamber filter press. Diatomateous earth was used as a precoat at a rate of about 3 lb/100 sf of media area. No conditioning was used, achieving a 50% final solids cake concentration.

In the early 1980's, Ann Arbor, Michigan installed a filter press to treat lime sludge. Based on extensive pilot studies the final press was designed to achieve a 50% solids concentration. After start-up (spring to summer months) performance was excellent, achieving a 65% final solids cake. However, as water temperatures fell, the achievable cake solids concentration dropped to 35% -- a very wet cake. This process of achieving a good solids cake in warm months and a poor cake in cold months has repeated itself, with the City investigating reasons for the poor winter performance and ways to improve it.

4.7.6. Example Facility

Ann Arbor, Mi. The City of Ann Arbor, Michigan owns and operates a split-treatment softening plant. The plant treats an average of 16 mgd and over 30 mgd during maximum day demand periods. The source of water for the plant has been a combination of surface water (Huron River) and groundwater at a ratio of approximately 85% surface water and 15% groundwater. The plant has operated primarily as a softening plant since both sources of water exhibit high hardness. Typical raw and finished water is shown in Table 4-10. The plant has used a filter press to dewater lime sludge since 1980. Prior to that time a calciner was used at the plant but was discontinued primarily due to cost considerations.

Filter backwash water in the plant has been equalized and recycled in the settling basins. Lime sludge produced in the plant has been continuously removed from settling basins and sent to sludge thickeners. A small amount of sludge has been continuously recycled which is typical of softening plants to

TABLE 4-10

TYPICAL WATER QUALITY

ANN ARBOR, MICHIGAN

	RAW WATER		FINISHED WATER
	River	Wells	
Total Hardness (as $CaCO_3$)	270 mg/l	500 mg/l	120 mg/l
Total Alkalinity (as $CaCO_3$)	200 mg/l	290 mg/l	64 mg/l
Turbidity	10 NTU	< 2 NTU	> 0.15
Color	40 units	0 units	5 units
pH	8.0	7.8	9.0
Total Organic Carbon (TOC)	10 mg/l	0 mg/l	2 mg/l
Total Dissolved Solids	400 mg/l	660 mg/l	250 mg/l

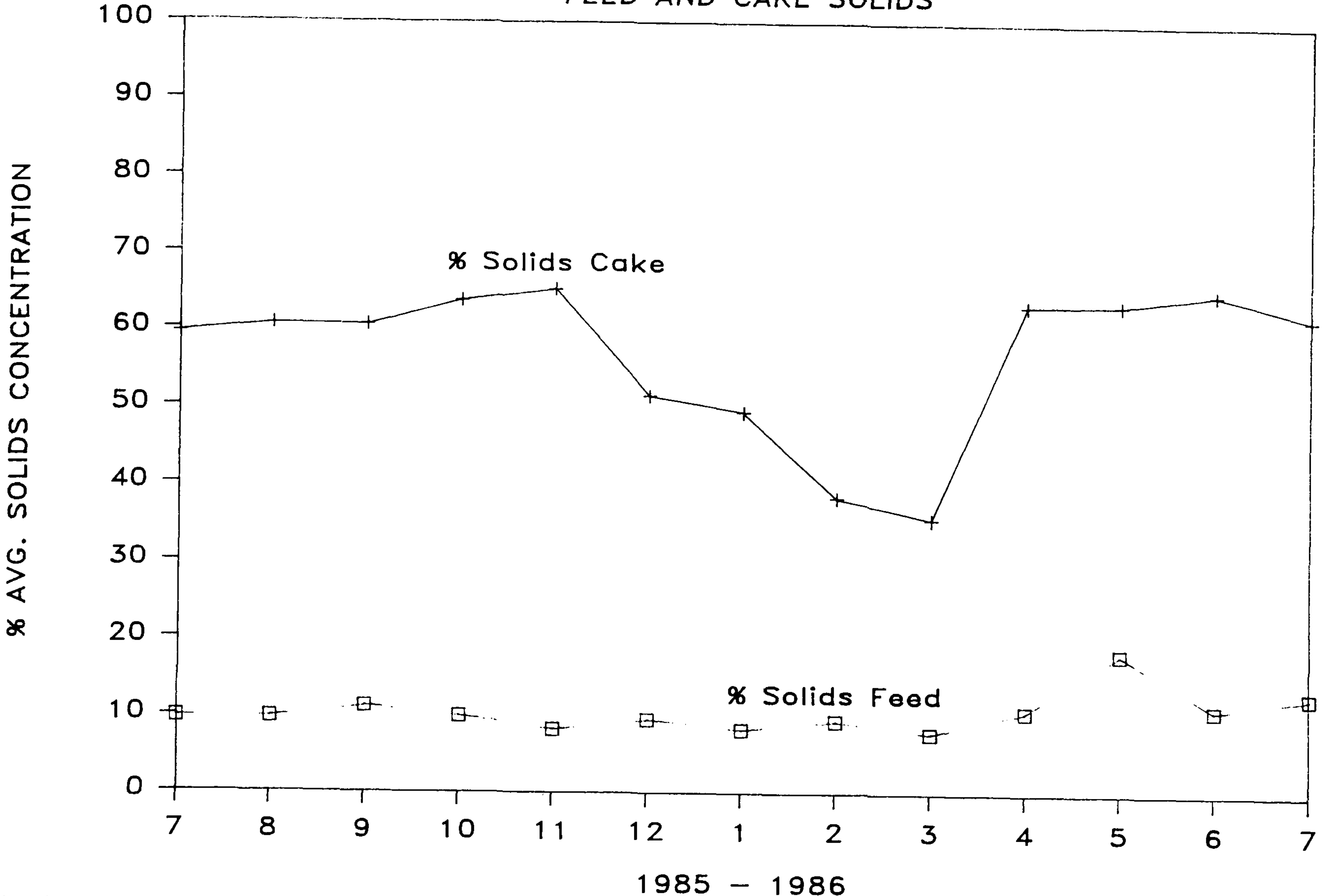
aid the softening precipitation process. Solids from the thickeners can be pumped to a sludge storage tank or fed directly to the filter press facility. The combined sludge storage and thickener capacity could provide up to 2 weeks of sludge storage if necessary. Normal operation has been to send sludge from the settling basins to the thickeners and then to a sludge storage tank prior to dewatering. Solids concentrations as high as 30% can be obtained from the thickeners; however, typical operation has been to keep the solids concentration in the 5 to 10% range for feed to the filter press operation. Solids production has been from 4,000 to 8,000 lbs/day of solids with peak production during the summer months.

The feed pumps to the filter presses include both low head, high capacity pumps for initial filling of the filter press and lower capacity, high pressure pumps for press operation. Sludge is fed to one of the two filter presses that are utilized during normal operation.

Each filter press has operated from 6 to 8 press cycles per day on average, equivalent to a cycle time of 120 to 160 minutes. The press has averaged a net yield of 1 lb/hr/sf and ranged from 0.5 to 1.35 lb/hr/sf. Ending a filter press cycle has been based on visual observation of filtrate flow and/or a pressure of 200 psi in the filter press. As mentioned earlier, the feed solids concentration has been in the 5 to 10% range into the press and ranged from 30% to 65% out of the press. Figure 4-38 shows actual solids feed concentrations and solids cake concentrations during a one year period. Experience has shown that much higher solids concentrations are obtainable with warm water temperatures than with cold water temperatures. Plant engineers and operators have evaluated press operating conditions, different filter cloths, and attempted to adjust cold winter water temperatures by blending warmer groundwater with cold river water in an attempt to achieve higher solids concentrations during winter

FIGURE 4-38

ANN ARBOR SLUDGE
FEED AND CAKE SOLIDS



months. The tests conducted to date have not shown any clear cut answers to achieve higher solids concentrations during winter months. Operationally the lower solids concentrations have not posed a significant problem because winter is a low water production period and the presses have been able to handle the sludge volumes generated.

The filtrate from the presses is returned to the raw water main to the plant. The sludge cake from the presses is dropped from the presses into trucks which are owned and operated by a sludge hauling contractor. The sludge is spread on nearby farms or is hauled to the local landfill for disposal. A schematic of the filter press facility in Ann Arbor is shown in Figure 4-39.

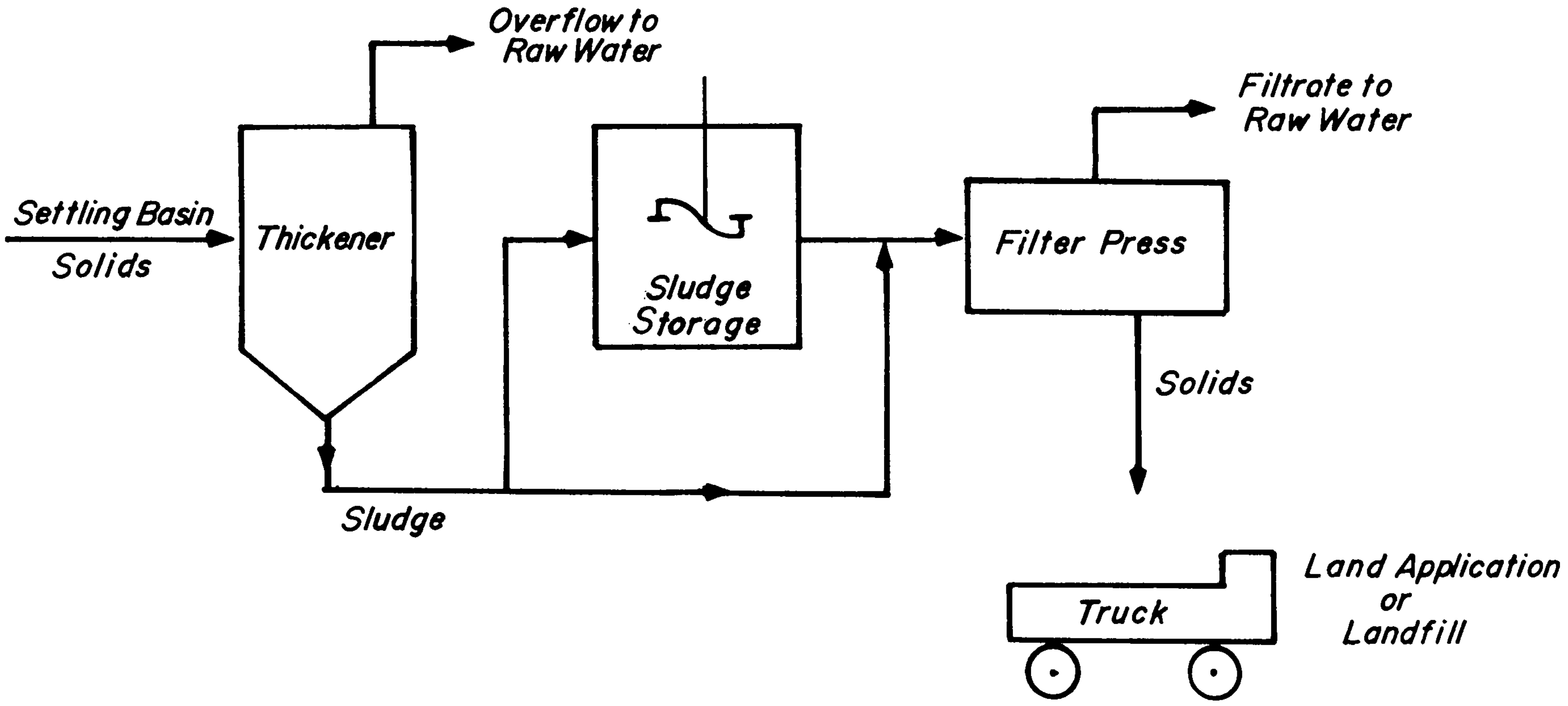
The time of operation of the filter press is generally 5 days a week and 16 hours per day. The operation has required the use of two full time operators for press operation. The operators are primarily responsible for the entire sludge facility operation.

An acid wash system which was included with the filter press allows for periodic cleaning of the press chambers and filter cloths. The wash process involves pumping of an acid solution through the press to dissolve solids buildup. An acid solution of 5% hydrochloric acid has been used for acid washing. The washing process takes approximately eight hours to complete and has been done 2 to 3 times each year. Disposal of the spent acid solution has been to the sanitary sewer system.

Operational problems have been minimal with the filter press. The lower solids concentrations during winter months has not posed a significant operations problem. Other communities that are contemplating the use of a filter press or other mechanical means of dewatering lime sludge may wish to pilot both winter and summer water conditions prior to final design based on the experience of the City of Ann Arbor.

FIGURE 4-39

ANN ARBOR SOLIDS HANDLING FACILITY
PROCESS SCHEMATIC



Fairfax County, Va. The Fairfax County Water Authority is a major purveyor of water in the rapidly growing region of Northern Virginia near Washington, D.C. The Authority provides water for approximately 550,000 people in Fairfax County, the City of Alexandria, Virginia and portions of Prince William County.

When the James J. Corbalis, Jr. Treatment Plant went on line in April 1982 the Fairfax County Water Authority also began operation of the alum sludge dewatering and disposal facility. As shown in the plant schematic of Figure 4-40 alum and fluoride are added to the raw water as it enters the plant through a control chamber and four rapid mix basins. Sodium hydroxide is fed for pH control and if required, powdered activated carbon can also be added. The average plant inflow is approximately 40 mgd and has the raw water quality characteristics shown in Table 4-11. The water typically receives an alum dosage of to mg/l, or 200 to 250 lbs/MG and the coagulated water flows into four tapered flocculation basins. The flocculated water flows through four sedimentation basins where the flocculated solids settle out and a scraper mechanism sweeps them to a discharge. The clarified water is then chlorinated and proceeds to eight high-rate filters.

The underflow from the sedimentation basins contains less than a 2% solids concentration and is pumped to a gravity thickener where the solids concentration is increased to an average of 8%. Thickened solids concentrations of 12 to 13% are occasionally achieved but not sustained. The thickened solids are then pumped into a reaction tank where approximately 4 to 10% by weight lime is added to raise the pH to 10.6 to 11.0 and the sludge then flows into conditioning tanks. Although polymer could be used, it rarely is because it reportedly makes the sludge 'spongey'. The conditioned sludge is then pumped through surge tanks and into one of two plate and frame filter presses that have been precoated with diatomaceous earth. Each press has 80 filters, 3,700 sf of

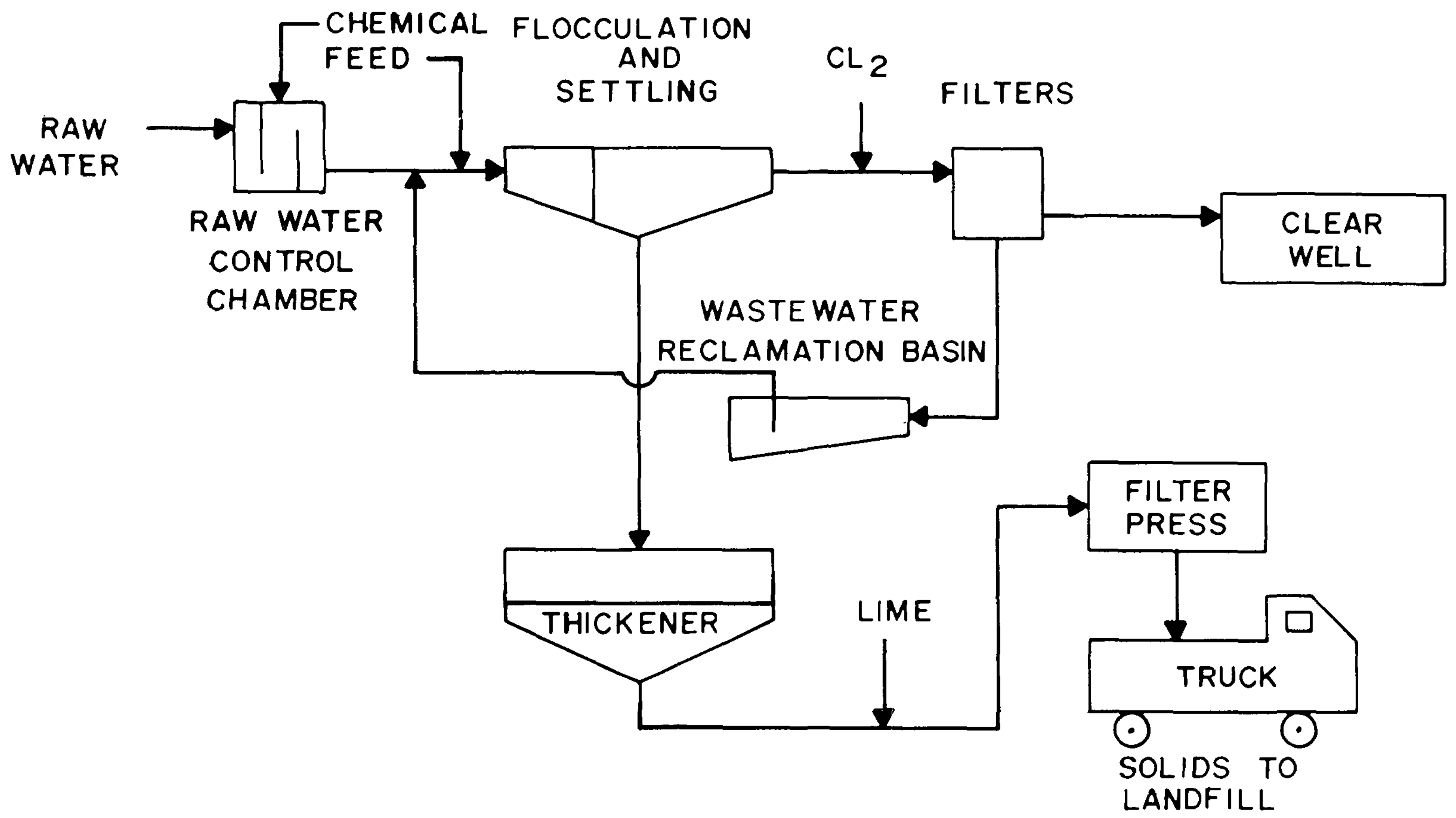


FIGURE 4-40

FAIRFAX COUNTY WATER AUTHORITY
JAMES J. CORBALIS PLANT SCHEMATIC

TABLE 4-11

SUMMARY OF RAW WATER CHEMICAL ANALYSIS

PERIOD OF 04/01/86 THROUGH 06/30/86
JAMES J. CORBALIS, JR. PLANT
FAIRFAX COUNTY WATER AUTHORITY

Parameter	Average
Alkalinity, Bicarbonate	67.2 mg/l
Alkalinity, Carbonate	12.8 mg/l
Alkalinity, Total	80.1 mg/l
Calcium	27.9 mg/l
Chemical Oxygen Demand	22.1 mg/l
Color	48.9 units
Dissolved Oxygen	7.8 mg/l
Magnesium	9.5 mg/l
pH	8.4 units
Solids, Fixed	140.0 mg/l
Solids, Volatile	47.9 mg/l
Solids, Total	191.7 mg/l
Solids, Total Dissolved	132.6 mg/l
Solids, Total Suspended	59.0 mg/l
Temperature	20.4 c
Threshold Odor	48.2 T.O.N.
Total Hardness	99.2 mg/l
Turbidity	36.23 T.U.

filter area and operates at 215 psi. The diatomaceous earth usage for precoat is 60 pounds per cycle or approximately 1.6 lbs per 100 sf. Each press cycle requires from 0.8 to 1.4 operational hours with values greater than 1.25 hours per cycle being typical. The solids handling facility is not run continuously since each filter could handle the entire sludge production. However, when the solids handling facility is in operation, a four man crew, including a supervisor, is required to handle all phases of operation and maintenance.

Each filter cake weighs approximately 150 lbs (wet weight) and each press contains 80 filter leaves. The production is therefore 12,000 lbs or 6 tons (wet weight) per cycle. In 1985 the facility averaged 2.87 cycles/day but did not operate every day. Typically 7-10 cycles are run per day during operation using both presses but as many as 18 cycles (including both presses) have been run during a 14 hour operation period. The filter cake typically has a 42 to 45% solids concentration although values as high as 48% have been achieved. The production rate of the press is about 0.65 lbs of original dry weight sludge per sf of filter area per cycle or a net yield of about 0.46 lb/hr/sf.

The pressed sludge cake is discharged to trucks and hauled to an on-site Fairfax County Water Authority Landfill. The press filtrate is discharged to the Fairfax County sanitary sewer after pH adjustment.

4.8. VACUUM FILTER

4.8.1. Description

In sludge filtration, pressure differential across a filter medium is required to force the water in the sludge through the medium while retaining the solids and ultimately forming a cake. The pressure differential in vacuum filtration is a vacuum applied to the downstream/receiving side of the medium.

The best way to express the conceptual theory of a vacuum filter press is the filter yield. The filter yield is defined as the mass of dry cake solids discharged from the filter media per hour per square foot of filter. Filter yield can be expressed as the product of filtrate production per unit area per unit time multiplied by the parameter w, mass of cake deposited per unit volume of filtrate. In the specific resistance equation presented in Chapter 3:

$$t = V\left(\frac{uR_r}{PA}\right) + \left(\frac{uR\ w}{2PA^2}\right)V^2$$

the term $V(uR_r/PA)$ relates to resistance caused by the filter medium. The value of this term is generally small when compared to the other term, $Y = \left(\frac{uR\ w}{2PA^2}\right)V^2$ and in most cases is considered to be negligible. The following expression for filter yield is developed:

$$Y = \left(\frac{2PwD}{uR\ t_c}\right)^{0.5}$$

where,

Y = filter cake yield, $kg/m^2 \cdot sec$

P = pressure (Ns/m^2)

w = feed solids concentration (kg/m^3)

D = drum submergence (fraction of the drum circumference below the sludge surface in the pan), dimensionless

u = viscosity (Ns/m^2)

R = specific resistance (m/kg)

t_c = cycle time (time for a complete revolution of the drum), sec

All variables in this expression are easily determined by the characteristics of the sludge and the equipment operational set-up, except the variable R. While this equation is theoretically correct, it has not proven accurate enough for design. One investigation (4-13) indicated the values of w and t_c had a more dramatic impact on the yield than the equation indicated. This same investigation generated the following expression for filter yield which was recommended for use in vacuum filter design:

$$Y = \left(\frac{2DP^{1-s}}{uC}\right)^{1/2} \frac{w^m}{t_c} n$$

where,

s = coefficient of compressibility (dimensionless)
m,n = dimensionless constants
C = specific resistance at unit pressure drop (also termed cake constant, see Section 3.3.1.).

other variables = same units as above

The constants m and n are determined by filter leaf testing. From this equation it is apparent that the filter yield is increased by increasing the drums submergence (D), vacuum pressure (P) (for normal pressure ranges) and feed solids concentration (w), and by decreasing the sludge's specific resistance (C) and cycle time (t_c). Many times efforts to increase filter yield will decrease the cake solids concentration.

Most vacuum filters employ a rotating drum with filter media on its surface. The drum is partially submerged (10 to 50 percent) in a vat of sludge. The sludge may be agitated to maintain the solids in suspension.

The drum revolves around a horizontal axis of rotation. A vacuum applied at the surface of the drum draws the filtrate through the media and cake to the collecting piping. The filtrate flow is controlled by a timing valve located at one end of the drum along the axis of rotation. A complete revolution of the drum is divided into three phases: cake pick-up or formation, cake drying and cake discharge. (See Figure 4-41A).

The cake formation stage takes place while the drum is submerged in the sludge vat. Wet sludge is collected on the filter media by the vacuum applied on the drum's surface.

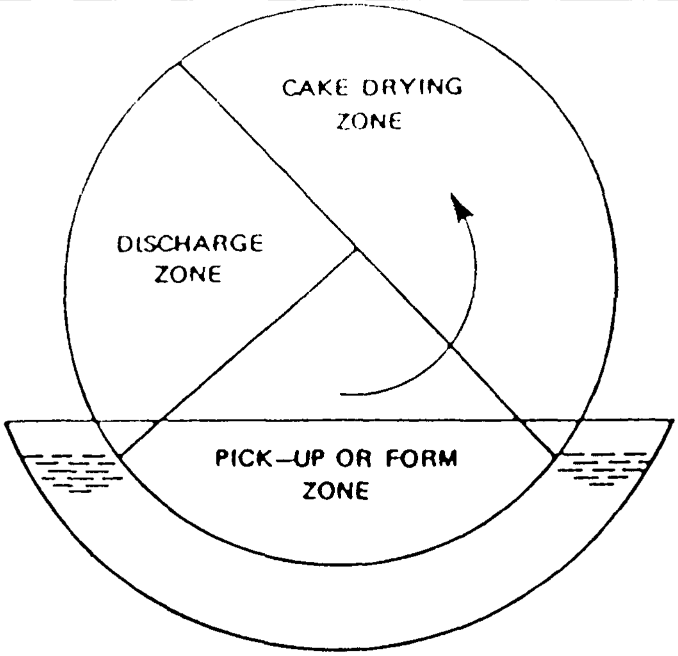
The cake drying stage begins when the sludge collected on the rotation drum surface leaves the vat and is exposed to air. The vacuum is continued and the air drawn through the sludge dewaters and assists in drying.

In the cake discharge phase, no vacuum is present and the cake is discharged from the press by various means depending on the type of vacuum press.

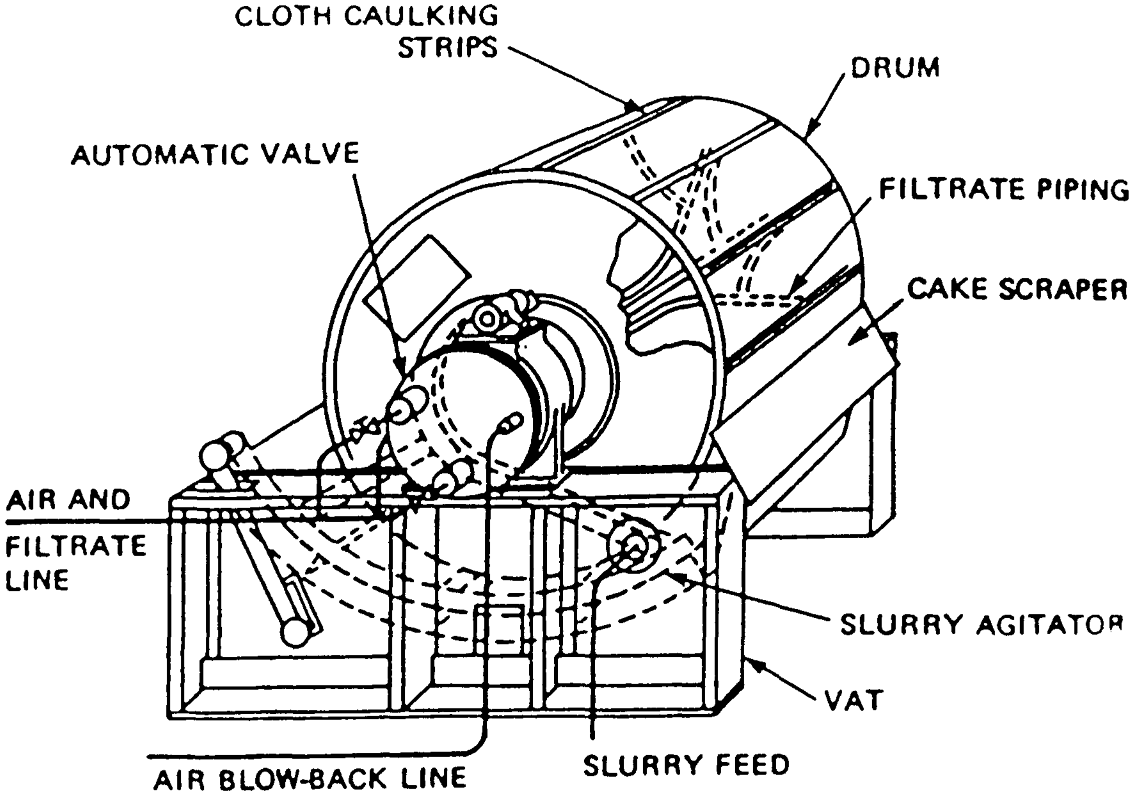
Washing of the filter media after cake discharge is performed on almost every vacuum filter. This washing removes the solid particles and conditioning agents which could clog the media openings and cause blinding. The washing is usually accomplished with a high pressure spray.

The original vacuum filter was the drum type (Figure 4-41B). In the drum vacuum filter the filter media is attached to and completely covers the drum surface. The formed cake is removed by a scraper plate angled very close to the rotating filter media and drum. The media is generally washed after cake discharge.

A relatively new variation of the vacuum filter is the top feed drum filter. In this type, the sludge is fed to the filter through a hopper located above the top of the drum. An advantage in this type over the conventional drum type is that sludge thickening is allowed in the hopper prior to application to the filter. Capital costs are also reduced since a



A. OPERATING ZONES



B. CUTAWAY VIEW

FIGURE 4-41

ROTARY VACUUM FILTER

smaller hopper is required and no sludge agitation equipment is needed.

Belt vacuum filters are quite similar to drum filters, but employ a nonattached cloth belt along the surface of the rotating drum (see Figure 4-42). During the discharge cycle, the belt is directed away from the drum by a series of rollers. As the belt turns at a sharp angle over the discharge roller, the cake breaks away. The belt can then be spray washed on both sides prior to its return to the drum face and the sludge slurry vat.

Horizontal vacuum filters (Figure 4-43) have also been utilized in the water treatment field. The filter media is fed from the top through a feed box designed to evenly distribute the sludge across the width of the filter. As the belt moves down the length of the unit a series of vacuum boxes draw the filtrate through the media into filtrate receivers. The cake is discharged at the end of the unit via the sharp turn of the filter media around the end roller. The media may be washed on the underside of the machine.

4.8.2. Design Considerations

A typical vacuum filtration system is shown in Figure 4-44. The housing facility and system installation are shown in Figure 4-45. Vacuum filtration, in general, is a continuous process; i.e. a continuous supply of sludge produces a continuous discharge of cake and filtrate.

Filtration aids are generally required to most economically dewater a coagulant sludge using a vacuum filter. Conventional filter aids include polymers and lime.

Precoating of the filter medium may be required to prevent blinding when dealing with a sludge which is difficult to dewater such as coagulant sludge. Precoating interrupts the continuous process because a preparation period is required to precoat the media prior to sludge filtration. In precoating, the drum rotates slowly (normally 5 to 12 revolu-

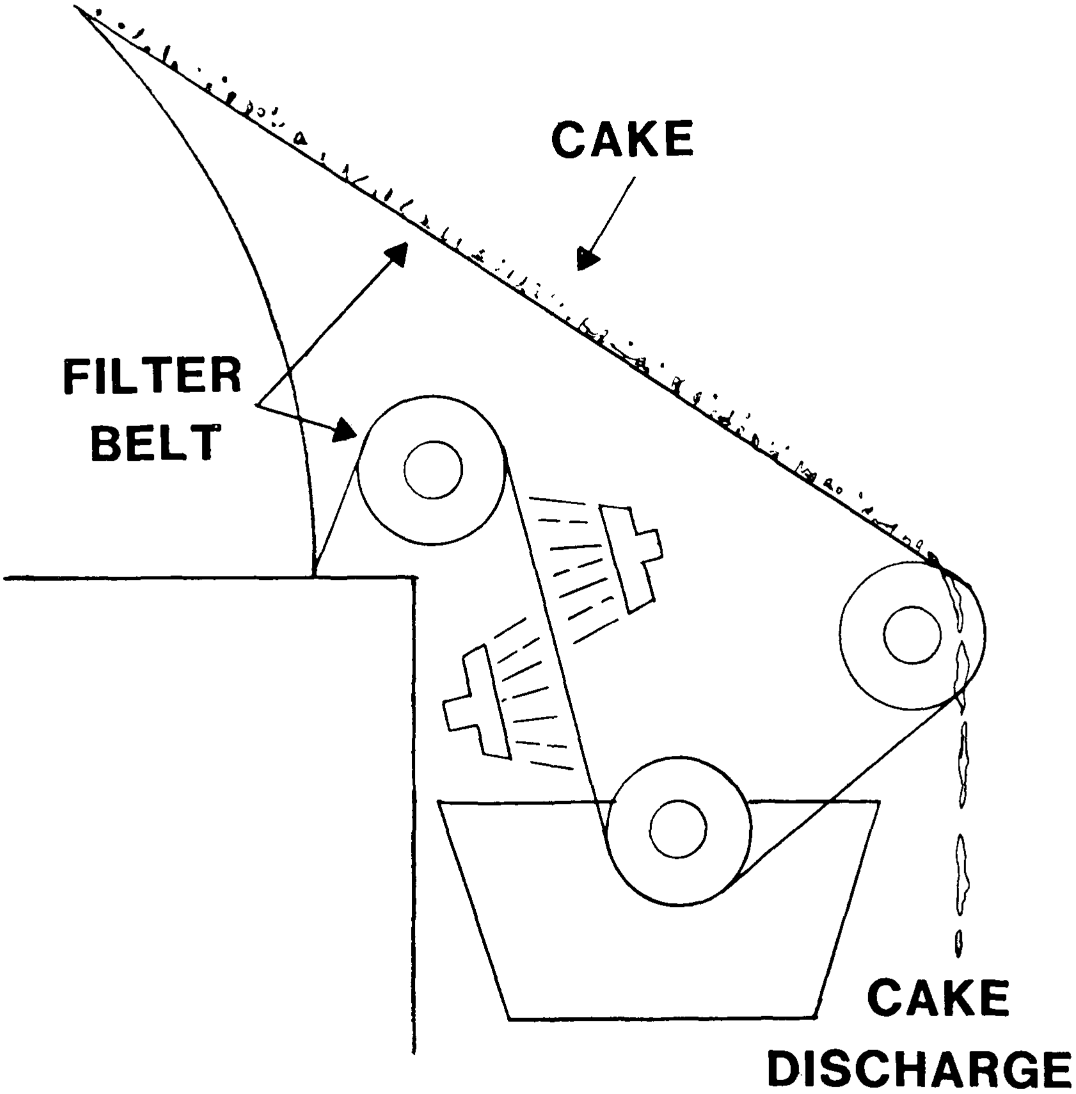


FIGURE 4-42

ROTARY VACUUM FILTER

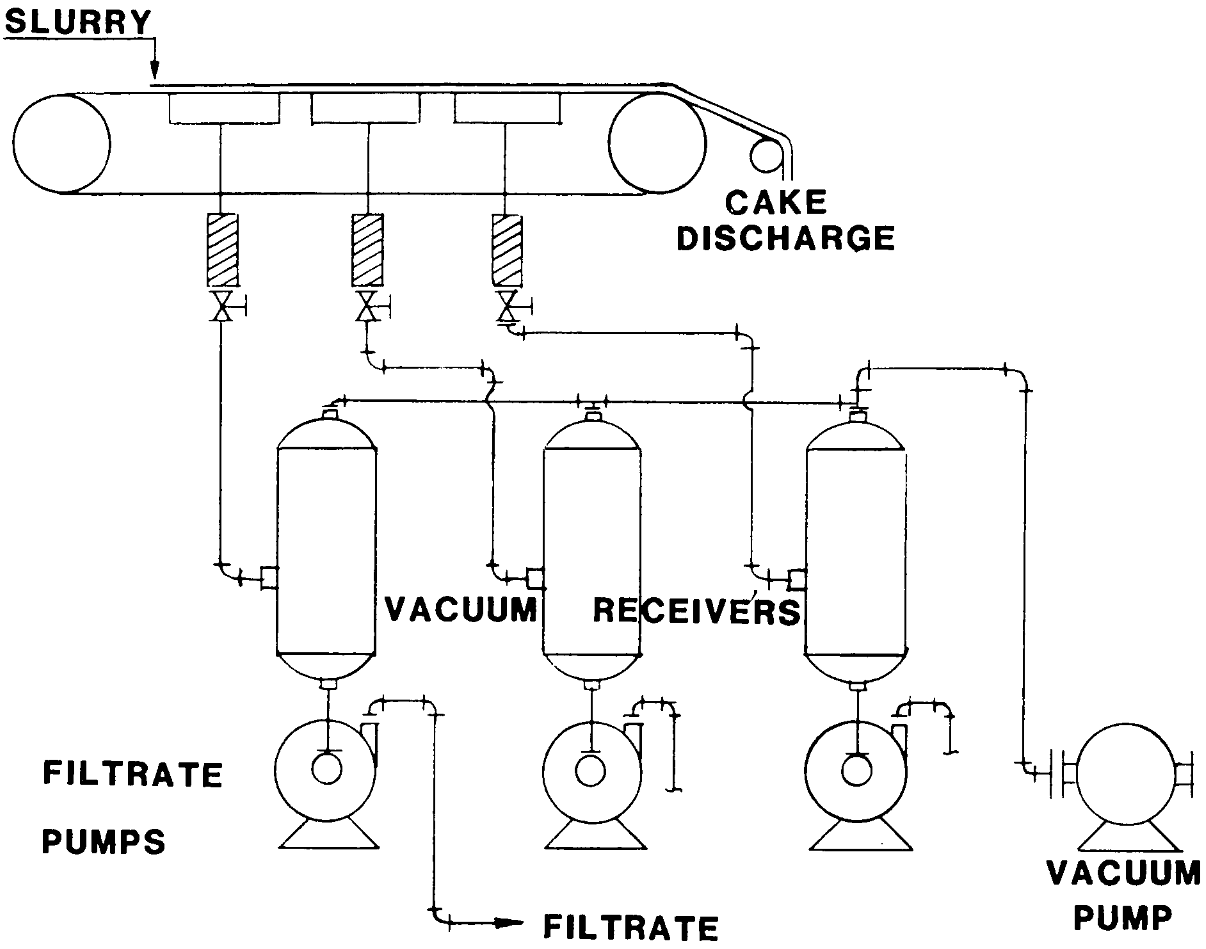
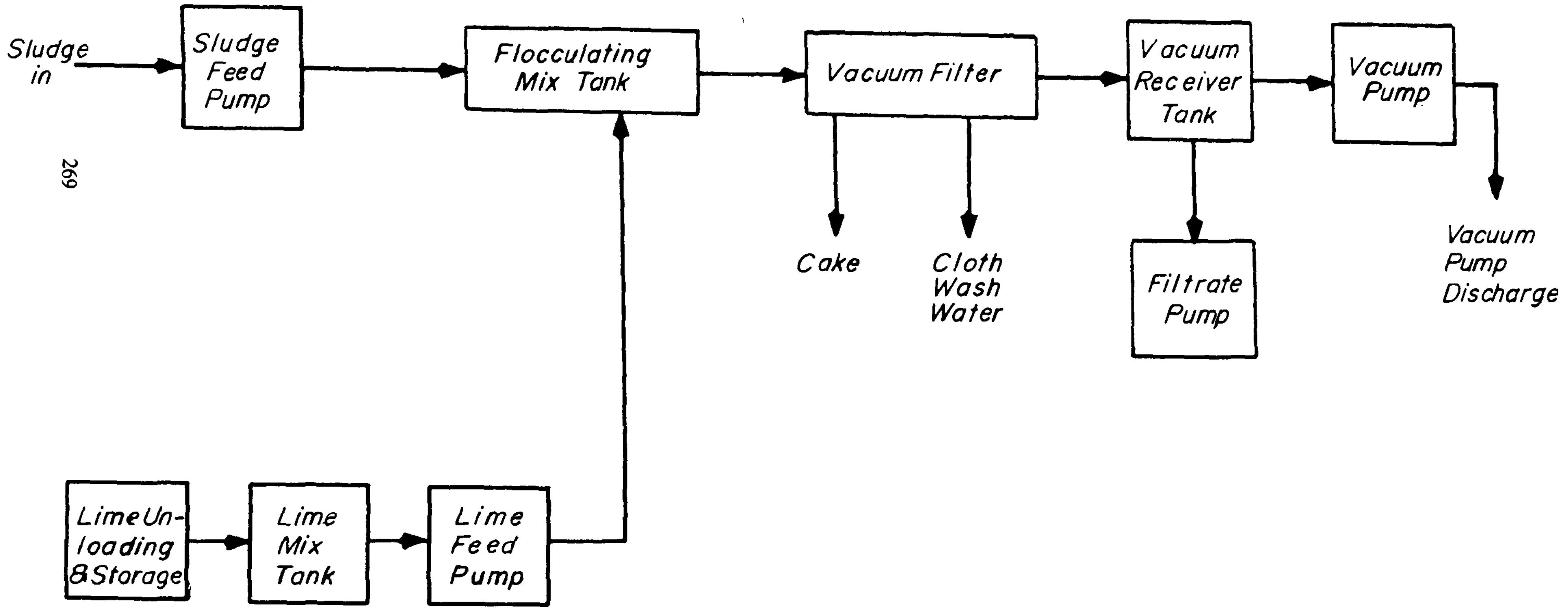


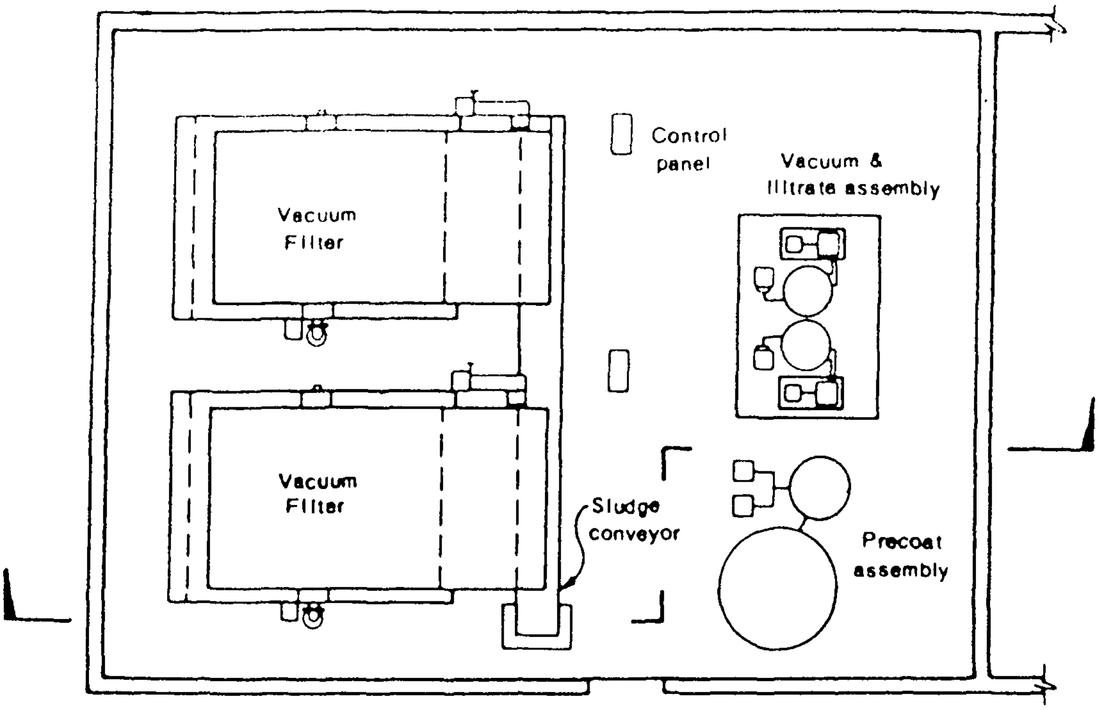
FIGURE 4-43

HORIZONTAL VACUUM FILTER SCHEMATIC

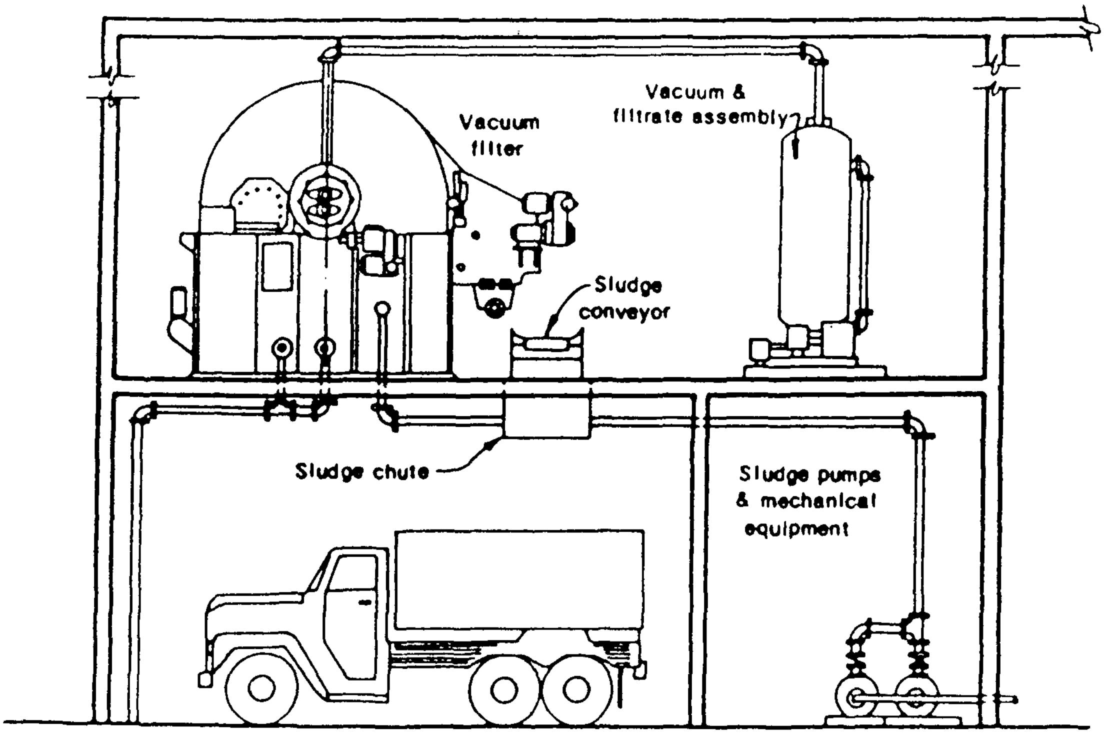
FIGURE 4-44

VACUUM FILTRATION SYSTEM





PLAN VIEW



ELEVATION VIEW

FIGURE 4-45

TYPICAL VACCUM FILTER INSTALLATION

tions per minute) in a vat of precoat material, (generally diatomaceous earth, fly ash, or lime) to obtain a precoat layer of 2 to 3 inches. Time to precoat is generally 50 to 60 minutes.

After forming the precoat cake, the filtering cycle then proceeds and the sludge cake accumulates against the outside of the surface of the precoat cake. An automatic knife blade continually advances at a preset rate of approximately 0.001 to 0.03 inches/minute. The knife removes the sludge cake and a small amount of the precoat material (Figure 4-46). When the precoat material is exhausted or reaches a predetermined minimum thickness, the filtration process is discontinued until new precoat is applied.

Cycle time is a critical design and operating consideration that can significantly affect the performance of the vacuum filter. The sludge cake will often begin cracking (therefore breaking the vacuum) at a drying time slightly less than the form time. Therefore, it is advantageous to establish the cycle time accordingly, which normally dictates a high submergence of the drum in the vat.

It has been shown that with longer form times, the cake thickness is increased while the yield is decreased. Optimizing both parameters, cake thickness and yield, has been accomplished using a cake thickness of 0.4 to 0.6 inches. Thinner cakes may crack early causing a vacuum break.

The vacuum system for a vacuum filter includes a vacuum pump and vacuum receiver. The vacuum pump supplies the necessary pressure differential across the filter media. The vacuum receiver is a tank which separates the filtrate from the air pulled by the vacuum pump during the cake drying stage. The air is allowed to continue through the receiver tank to the vacuum pump while the filtrate is stored for subsequent pumping by a filtrate pump.

Vacuum pumps are normally required to provide from 10 to 25 inches of Hg vacuum. Vacuum receivers are designed to

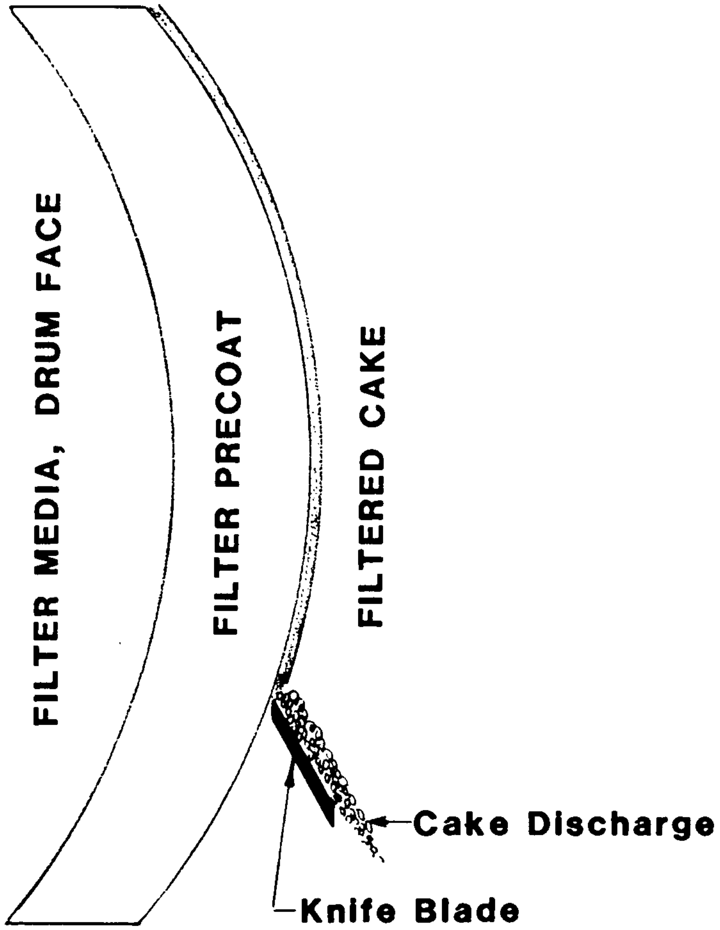


FIGURE 4-46

ROTARY VACUUM FILTER CAKE REMOVAL

provide an air velocity of up to 4 ft/sec and capacity enough for 2 to 3 minutes of air detention and 4 to 5 minutes of filtrate detention.

The sludge feed system includes the feed pump and the sludge vat. Feed pumps are required to feed sludge to the vacuum filter at the specified rate and are therefore generally of the positive displacement type. Feed pumps generally are controlled to maintain the liquid level in the sludge vat.

4.8.3. Capital and Operating Costs

The construction costs for vacuum filters are shown in Figure 4-47. The construction costs include the cost for the vacuum filter equipment, labor, pipes and valves, electrical and instrumentation and two story housing. The vacuum filter equipment includes the cost for the vacuum filter, the vacuum pump, vacuum receiver and a filtrate pump. Not included is the cost of the sludge feed pump which can be obtained from the sludge pumping cost curves presented earlier in this Chapter. Also, not included is sludge conditioning or additional sludge cake handling costs. For precoating, the lime conditioning system costs can be added to the base vacuum filter construction and O & M costs. Operating and maintenance costs are shown in Figure 4-48.

4.8.4. Operating Considerations

The operational variables for a vacuum filter include the cycle time, the sludge feed rate, the sludge level in the vat, the vacuum applied and choice of sludge conditioning and precoating agents. To some extent, the choice of filter media itself is an operating variable, but generally the selection is made during the design phase.

Cycle time can be varied by changing the rotational speed of the drum. Also, by changing the sludge level in the vat, the ratio of the formation time to total cycle time is varied.

FIGURE 4-47

CONSTRUCTION COST FOR VACUUM FILTERS

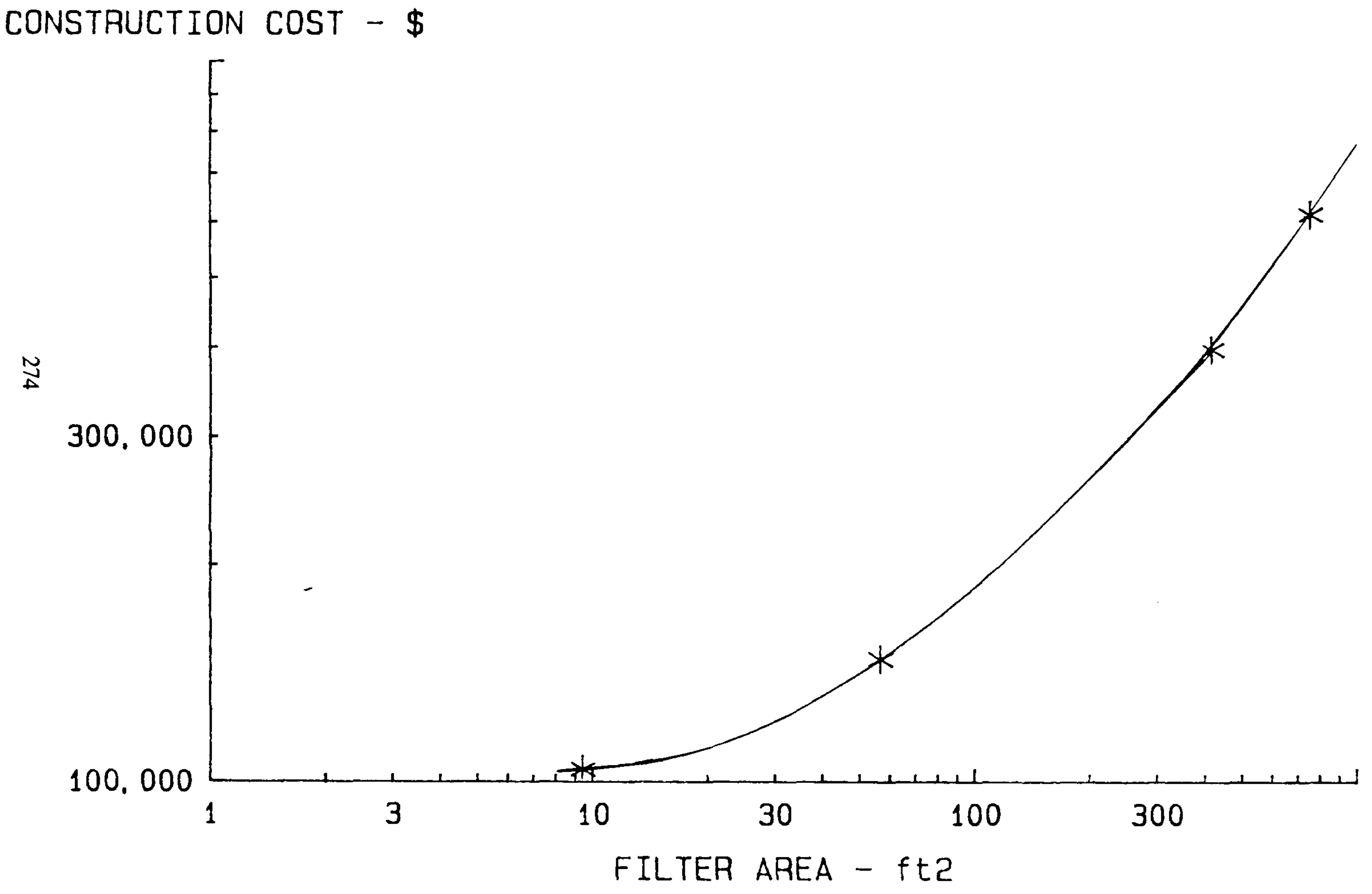
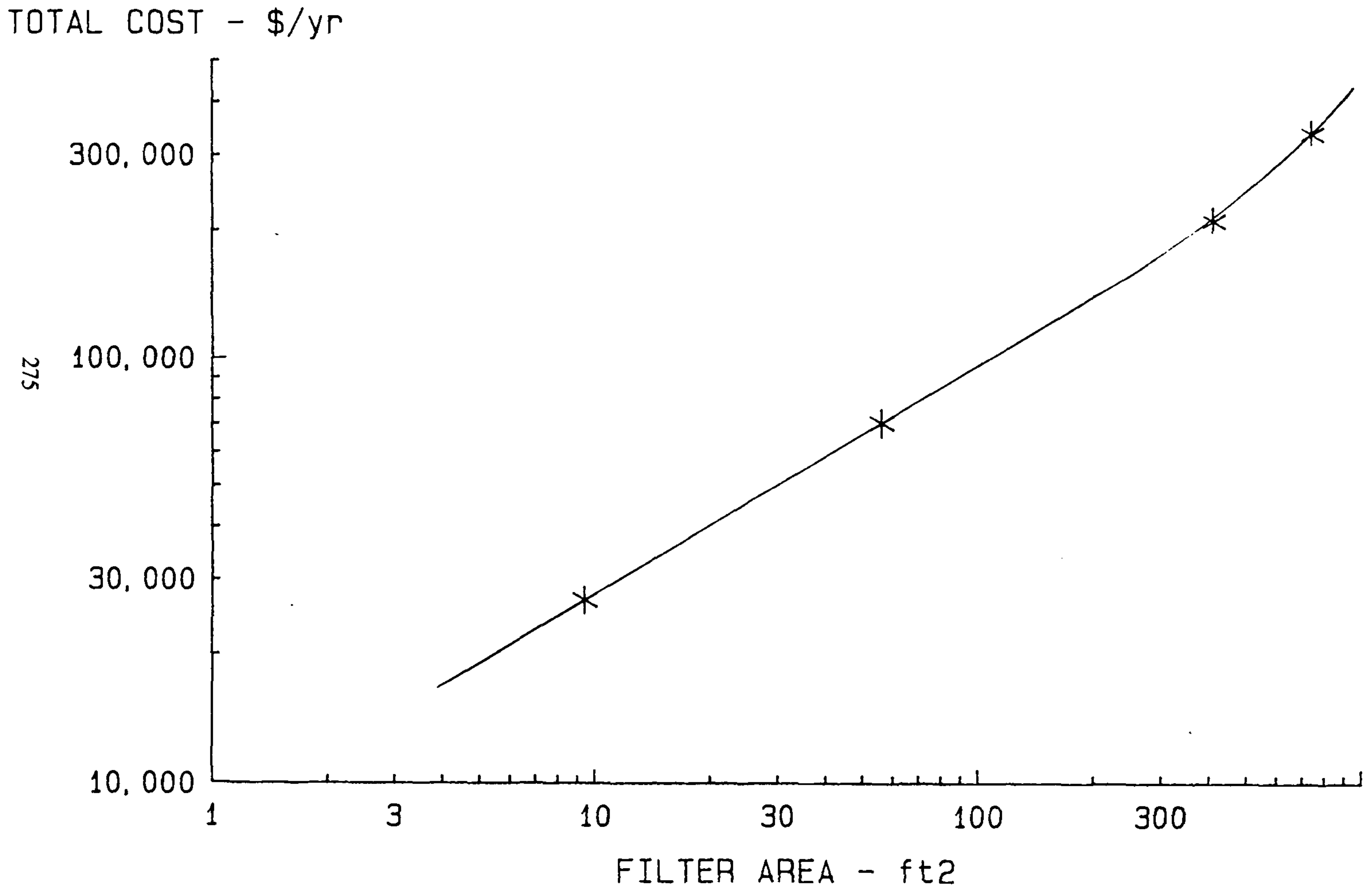


FIGURE 4-48

OPERATION AND MAINTENANCE COST FOR VACUUM FILTERS



The sludge vat is curved to match the drum's curvature and is designed to provide sludge to the vacuum filter. The sludge is slowly, agitated (11 to 15 strokes/minute) to maintain the solids in suspension and to ensure a homogenous mixture.

The best way to demonstrate the effects of varying the operational variables is to present and discuss data operated from tests performed for that purpose.

4.8.5. Past Performance

Vacuum filters have been evaluated on a pilot plant scale for use on alum sludges. Although some success has been achieved on the pilot scale no known full-scale application of vacuum filters for dewatering alum sludges exist. This has primarily been because of the high amounts of conditioning chemicals required and/or poor cake solids concentrations and poor yields.

Westerhoff and Daly (4-7) reported pilot plant work evaluating a precoat drum vacuum filter and a horizontal vacuum filter. Table 4-12 shows the results using the precoat vacuum filter treating coagulation basin sludge and filter backwash solids. The precoat material used was diatomaceous earth. The horizontal vacuum filter was only evaluated on acidified alum sludge and highly thickened filter backwash solids. Results of those tests are also shown in Table 4-12.

Nielsen et al. (4-6) evaluated a vacuum filter for their preliminary design for the East Bay Municipal Utility District. A rotary drum vacuum filter was used with and without a precoat. Based on laboratory filter leaf tests, they evaluated two types of belt material. The first belt was a monofilament polypropylene weave with an air permeability of 350 cfm/sf at 0.5-in. water pressure. This belt did not retain solids and was rejected after 3 trial runs. The second belt was a multifiliment polyester of 35 cfm/sf at 0.5-in. water pressure. A final cake solids concentration of 9% was

TABLE 4-12

RESULTS OF VACUUM FILTRATION EVALUATION

	Precoat Vacuum Filter 43 (tests)	Horizontal Vacuum Filter 37 (tests)
Coagulation Basin Sludge		
Feed Conc.[1] (%)	2 (2 - 5)	15 acidified
Flow rate (gal/hr/sq.ft.)	2 (2 - 4)	6
Precoat Dosage (%)	32 (30 - 35)	--
Lime Conditioning Agent (%)	--	60 (55 - 65)
Cake Solids (%)	34 (30 - 35)	37 (35 - 40)
Filt. Susp. Solids (mg/l)	10	140
Filtrate pH	7	12
Filter Backwash Sludge		
Feed Conc. (%)	1 - 2[2]	8
Flow rate (gal/hr/sq.ft.)	up to 5	6
Precoat Dosage (%)	45 (40 - 50)	--
Lime Conditioning Agent (%)	--	50
Cake Solids (%)	20	30
Filt. Susp. Solids (mg/l)	15	120
Filtrate pH	8	12
Filter Backwash Sludge[3]		
Feed Conc. (%)	--	13
Flow rate (gal/hr/sq.ft.)	--	6
Conditioning Agent	--	None
Cake Solids (%)	--	35
Filt. Susp. Solids (mg/l)	--	30

[1] All concentrations - % dry solids by weight.

[2] Less than 1% requires 200% precoat to obtain 10% solids.

[3] Sludge subjected to artificial freeze-thaw process prior to vacuum filtration.

Source: Westerhoff and Daly (4-7)

achieved with a polymer dose of 4 lb/ton at a yield of 5 lb/sf/hr. Cake solids concentration of 15% was achieved at a polymer dose of 2 lb/ton and a yield of 1 lb/sf/hr. Filter breakthroughs were reportedly frequent and constant operation attention was required. Precoat tests were conducted which increased the final cake concentration to 20%; however, the yield was only 0.1 to 0.2 lb/sf/hr.

Lime softening sludge dewatering by drum vacuum filters using multi-filament polypropylene medium is widely accepted. The most important factors affecting the dewatering of lime sludges are the feed solids concentration and the magnesium content.

A wide range of operating data on traveling belt medium is shown in Table 4-13. The feed concentration, the filter yield and operating vacuum shows close correlation. Increased vacuum and feed concentration improves filter yields while cake solids concentration shows only slight fluctuation.

Cycle time selection for vacuum filters on dewatering lime softening sludge will critically affect the filter's performance. Since both field and laboratory experience has shown that cake cracking occurs at a drying time just slightly less than the form time, a high submergence vat is necessary for obtaining maximum yields. Leaf test results on lime sludge at Bismarck, North Dakota are shown in Table 4-14. These data confirm that longer form time provides thicker cake, but decreases the yield. About 10.16 to 15.34 mm (0.4 to 0.6 inches) cake thickness is the best for yield and cake discharge. Thinner cake may crack early creating a loss of vacuum.

The first use of a vacuum filter in water treatment was the application of a belt vacuum filter to a lime sludge in Minot, N.D. A relatively high magnesium content sludge was successfully dewatered. In the late 1960's, Boca Raton, Florida tested and installed a belt vacuum filter on a highly calcium carbonate sludge. A 1 to 4% solids concentration

TABLE 4-13

AVERAGE PERFORMANCE AND OPERATING DATA BY VACUUM FILTER ON LIME SOFTENING SLUDGE

Feed Solids (%)	5 - 30
Cake Concentration (%)	40 - 70
Cake Yield ($kg/m^2/hr$)	0.8 - 4.0
(lb/sf/hr)	4 - 20
Filtrate Solids (mg/l)	950 - 1500
Solids Recovery (%)	95 - 99
Filter Speed (rpm)	0.2 - 0.5
Operating Vacuum (mm Hg)	381 - 635
(in. Hg)	15 - 25

Reference: "Operating and Pilot Data", Ingersoll Rand, Nashua, N.H. (1970 - 1977)

TABLE 4-14

LEAF TEST RESULTS ON LIME SLUDGE AT BISMARCK, N.D.*

Form Time (sec)	Cake Cracks (sec)	Filter Speed (rpm)	Cake Thickness (in)	Cake Thickness (mm)	Yield (lb/sf/hr)	Yield ($kg/m^2/hr$)	Cake Concentration (%)
30	20	1.10	0.40	10.16	48	9.6	35
60	45	0.55	0.50	12.70	35	7.0	36
90	68	0.37	0.70	17.78	28	5.6	35
120	97	0.27	0.90	22.86	23	4.6	34

* Feed Sludge - 14.8%

Reference: "Test Results on Lime Sludge at Bismarck, N.D.". Ingersoll Rand, Nashua, N.H. (1972)

sludge from the softening reactors was concentrated to 28 to 32% by a thickener. The vacuum filter could be loaded at rates of 60 lb/sf/hr resulting in a final cake concentration of 65%. The cake was further air dried prior to disposal.

Numerous applications have since been installed with similar results. The two primary factors affecting performance are the solids feed concentration and the magnesium hydroxide content.

4.9. BELT FILTER PRESS

4.9.1. Description

Belt filter presses use a combination of gravity draining and mechanical pressure to dewater sludges. A typical belt filter press consists of a chemical conditioning stage, a gravity drainage stage and a compression dewatering stage. (See Figure 4-49).

The dewatering process starts after the feed sludge has been properly conditioned, usually with polymer. The slurry enters the gravity drainage stage, where it is evenly distributed onto a moving porous belt. Readily drainable water passes through the belt as the slurry travels over the full length of the dewatering stage. Typically, 1 or 2 minutes are necessary to allow for the filtrate separation in the drainage stage.

Following gravity drainage, the partially dewatered sludge enters the compression dewatering stage. Here, the sludge is "sandwiched" between two porous cloth media belts which travel in a "S"-shape path over numerous rollers. Both belts operate under a specific tension which induces dewatering pressure onto the sludge. The "S"-shape path the sludge follows creates shear forces which assist in the dewatering process. The compressive and shear forces working on the sludge increase over the length of this dewatering stage. The final sludge cake is removed from the belts by blades.

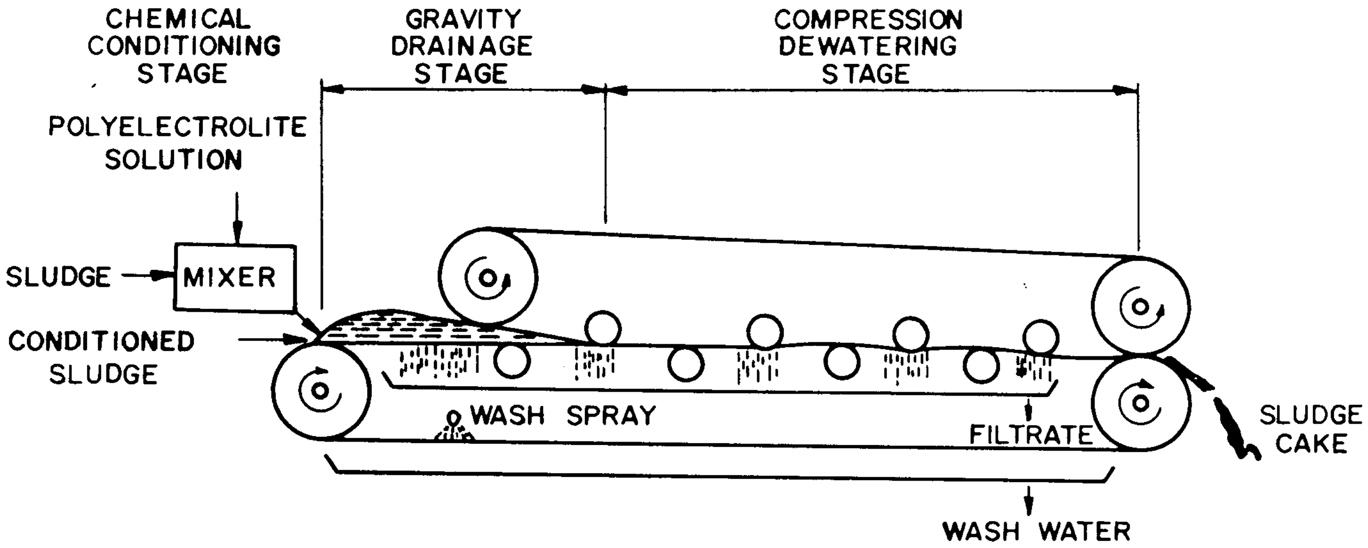


FIGURE 4-49

THE THREE BASIC STAGES OF A BELT PRESS

4.9.2. Design Considerations

Proper sludge conditioning is considered critical for obtaining acceptable dewatering results. A typical sludge conditioning unit consists of chemical conditioner storage, metering pumps, mixing equipment (chemical and chemical/sludge), controls and process piping.

In general, polymer is used for chemical conditioning. To achieve proper sludge conditioning, the polymer is first diluted to between 0.25 and 0.50% by weight before it is applied to the feed sludge. Next, the sludge and the polymer are thoroughly mixed. The required mixing time depends on sludge characteristics and type of polymer used.

Pumps associated with belt filter press operations include sludge feed pumps and chemical metering pumps. For sludge feed pumps, various types of equipment have been used such as piston pumps, progressive cavity pumps and rotary lobe pumps. However, constant rate pumps are preferred, because intermittent flow affects the belt filter press performance. Design considerations for sludge feed pumps should include hydraulic requirements, type of sludge to be dewatered and range of solids concentrations. Generally, each press has its own sludge feed pump to control the sludge loading. Standby pumping equipment is usually provided by either interconnecting piping between presses or by separate pumps.

Chemical metering pumps are associated with the sludge conditioning system to provide a proper polymer flow rate to the feed sludge. In general, positive displacement pumps (diaphragm or progressive cavity type) are used. Variable speed drives should be provided to control the output rate.

The design and selection of a belt filter press is often based on the throughput of the machine, i.e. the rate the sludge can be dewatered by the press. The throughput capacity can be limited either by the water in the sludge (hydraulically) or can be solids limited. A belt filter press having a particular type of belt at a particular width has a maximum loading capacity for a particular sludge. Generally the solids loading is considered the most critical factor and the throughput is expressed in terms of solids loading. The loading units are usually similar to a yield except expressed as belt width -- mass/width/time.

The structural main frame of belt filter presses consists of either steel channels, I-beams or tubing. Regardless of the type of frame used, a protective coating is generally applied. Corrosive resistant coatings include epoxy paints, hot-dipped galvanizing, and fiberglass envelopes.

Most belt filter presses are equipped with two porous cloth belts, namely the upper and lower belt. The sludge is supported and compressed between the two belts as it moves through the compression dewatering stage of the filter press. Belts are either of the seamed or endless type and should be able to withstand tensile strengths several times greater than the maximum operating tension. A wide variety of belt materials exist and selection should be based on sludge characteristics, solids capture and durability. In the compression stage, the belts are supported by rollers which are situated such that tension, shear and compressive forces are induced onto the sludge. Rollers are available in different types of materials including stainless steel. Protective coatings, such as rubber, should be considered dependent upon the corrosiveness of the environment. Maximum allowable roller deflection should be 1 mm (0.05 in.) at the roller mid-span.

Filtrate from gravity and compression dewatering processes, as well as wash water from the belt washing system, is collected under the belt filter press. From the collection housing units, drainage piping transports the water to a sump or floor drain. The floor area around the belt filter press slopes toward the drainage system to allow the unit to be washed down.

After the sludge has passed through the various dewatering stages of the press, the final sludge cake is removed from the belts by a "doctor blade". From this point, there are various methods available to remove the cake from the press location. One such option is direct discharge into a dump truck. However, press location and building layout are critical using this method. Other options include conveyors and hoppers, of which there are many different types available on the market. When selecting cake handling equipment the following criteria are to be considered: operating history, reliability, seasonal storage and disposal constraints (see Figure 4-50).

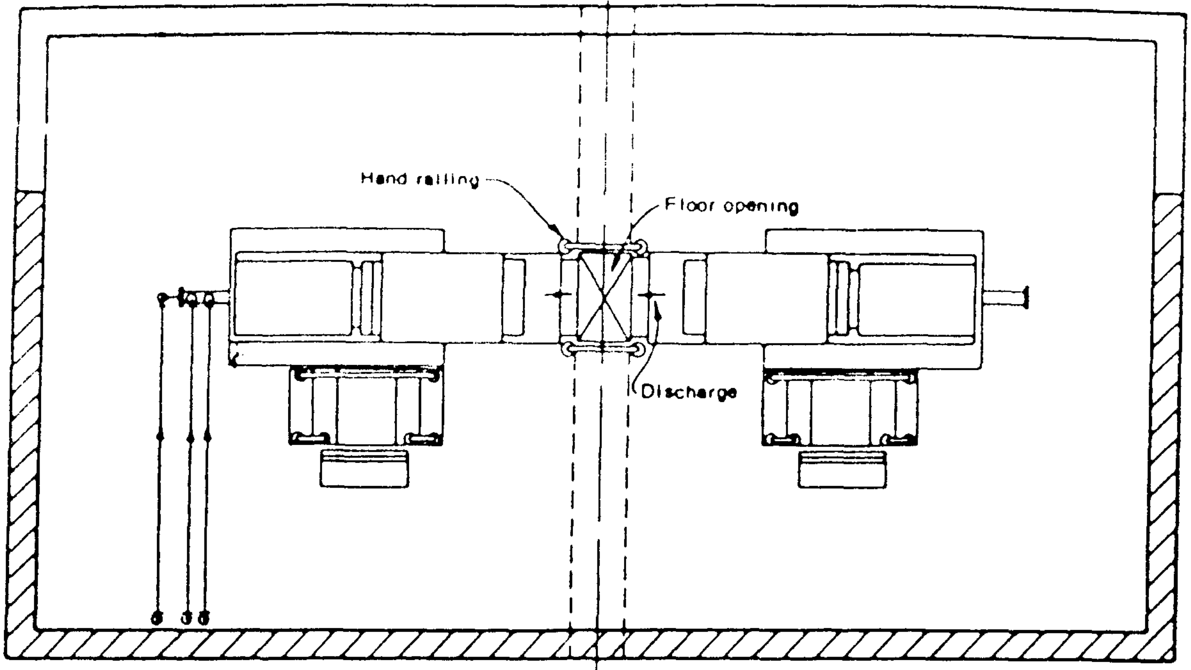
The travelling belt is generally cleaned by a belt washing system which sprays high pressure water on the belt. Since the belt is continuous, the belt washing system is located so as to wash the belt after cake discharge and prior to the next dewatering cycle. A spray nozzle is generally provided for each belt.

An adjustable belt tensioning system is provided for control of the pressure imposed by the belts on the sludge. The tensioning system can be either pneumatic, hydraulic ar mechanical. Belt life is a direct function of the belt operating tension. Automatic belt tension is sometimes provided.

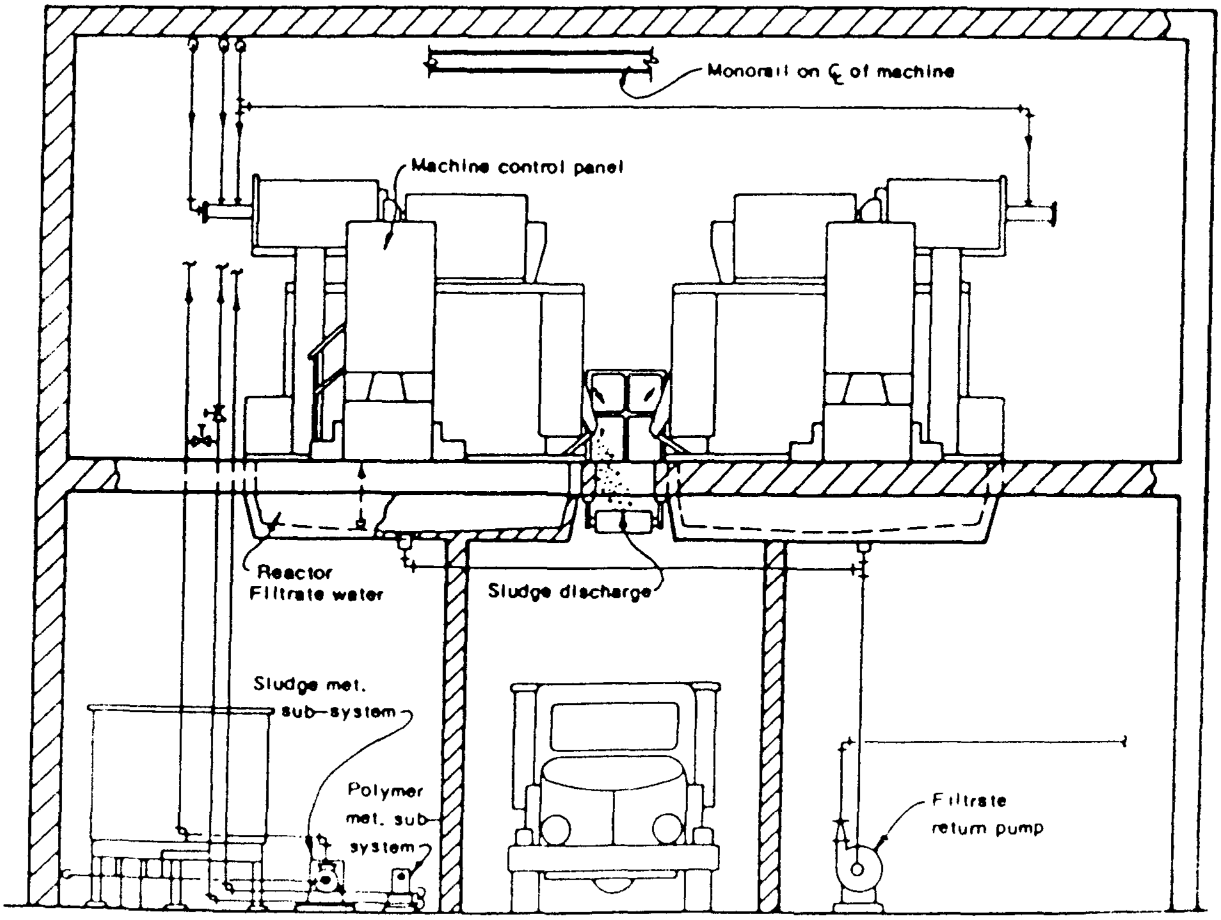
A belt tracking system is normally provided on a belt filter press to keep the belts tracking over the center of the rollers. Sensors and limit switches are employed to detect belt misalignment. A continuously adjustable roller realigns the belt by pneumatic, electric or hydraulic means.

4.9.3. Capital and Operating Costs

Capital cost curves for the belt filter press are shown in Figure 4-51. The construction costs include the belt filter press equipment, installation labor, pipe and valves, electrical and instrumentation and two story housing. The



UPPER LEVEL PLAN



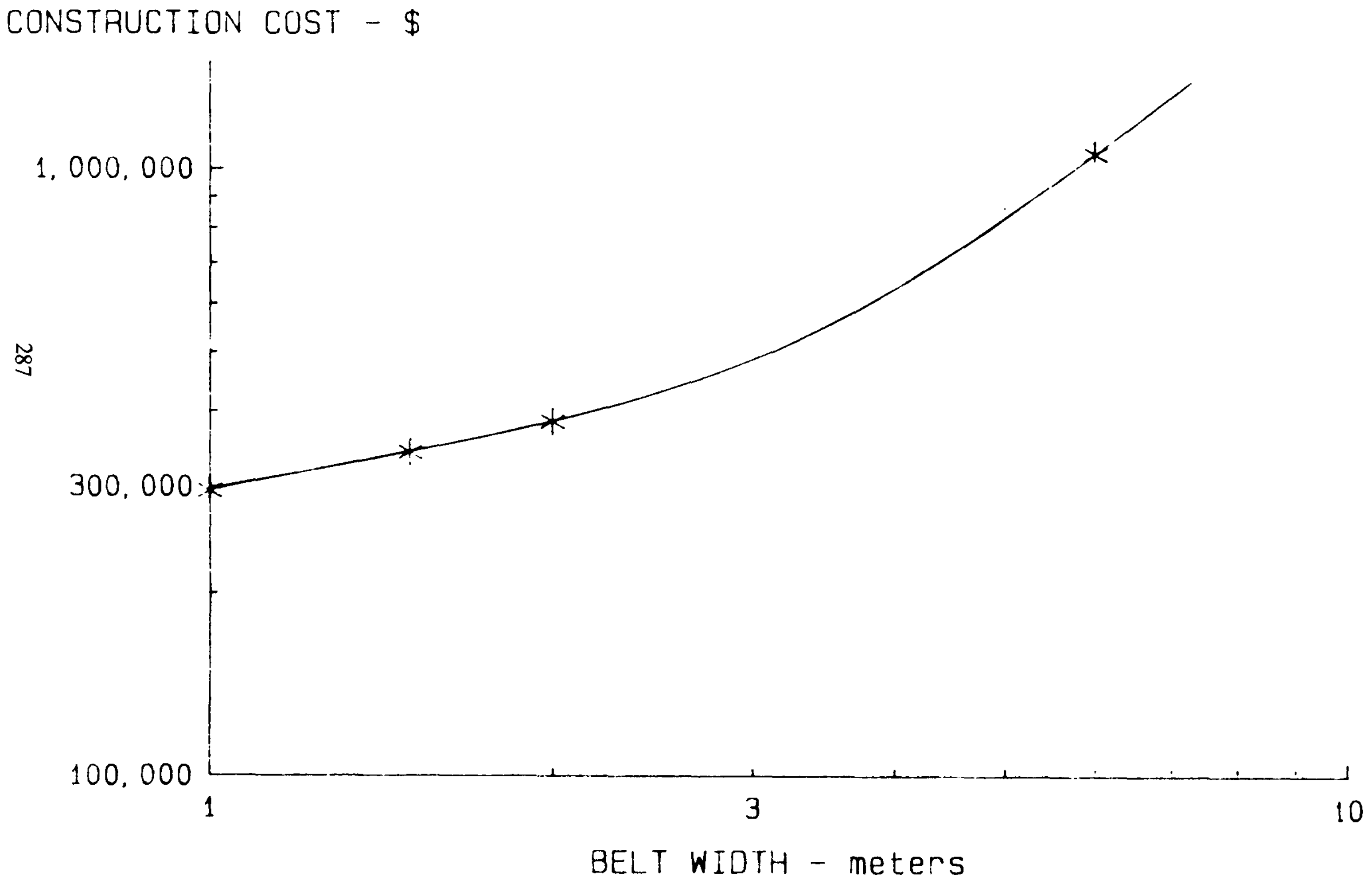
ELEVATION

FIGURE 4-50

TYPICAL BELT FILTER PRESS INSTALLATION

FIGURE 4-51

CONSTRUCTION COST FOR THE BELT FILTER PRESS



belt filter press equipment includes the cost for the belt filter press and the hydraulic power unit. Not included in the equipment cost is the cost of the sludge feed pump or the polymer conditioning system. The filtrate normally flows by gravity from the belt filter press, therefore a filtrate pump is not provided. The costs for the components not included can be added from the applicable preceding sections of this Chapter.

The operation and maintenance costs are shown in Figure 4-52. Operating and Maintenance costs include process energy, maintenance material and labor costs. Process energy costs were developed using total connected horsepower. Operation and maintenance costs for belt filter presses are very dependent on the sludge characteristics and may vary widely from sludge to sludge.

4.9.4. Operating Considerations

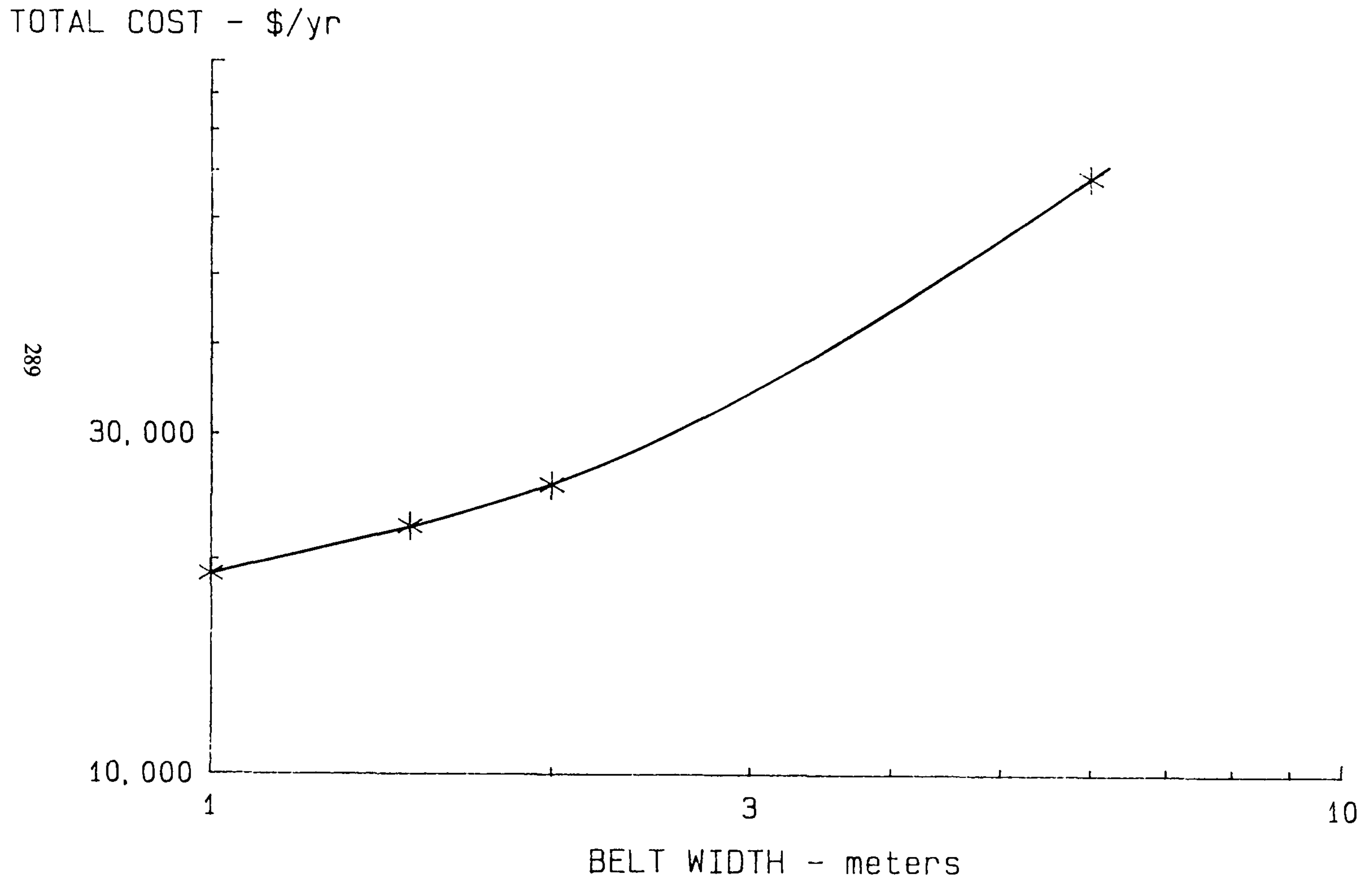
As with the other mechanical dewatering processes, the successful operation of the belt filter press should consider the solids concentration of the sludge cake produced, the quantity and quality of filtrate and wash water wastes, and the solids capture ratio. These parameters are dependent on the belt design and fabric, the type and conditioning of the sludge, and the pressure (both magnitude and time) applied to the sludge by the belts and rollers while in the press.

The belt tension is generally set by the manufacturer and while increasing the tension can slightly increase the cake solids concentration in some cases, the belt life can be significantly reduced. In most belt presses the pressure applied to the sludge by the rollers increases as the sludge passes through the cycle.

The belt speed determines the retention time of the sludge in the press and the time the sludge is subjected to pressure. The belt speed also dictates the throughput through the filter press. It is easily seen that while the belt speed

FIGURE 4-52

OPERATION AND MAINTENANCE COST FOR BELT FILTER PRESS



is the most controllable operational variable, it is also probably the most critical one.

While the majority of belt presses sold today have some automatic operating features, smaller installations may prefer manual presses. Although manual operation adds flexibility to the operation, protective system interlocks should be provided between each system to prevent mechanical damage.

An emergency "panic button" should be located near the primary operator station or control panel of the press to immediately shut down the press if activated. If the belt press is not to be used for a long period of time (a few days) the press should be properly prepared for the down time. The sludge should be purged from the pumps, pipes and press. A nonpotable or potable water supply (with backflow preventers) fed into the suction side of the sludge feed pump can purge the system. Finally, the belts can be sprayed clean with the high pressure wash sprayer.

4.9.5. Past Performance

The belt filter press was first introduced in the United States in the 1960's. The presses were quite similar in appearance to the Fourdrinier paper-making machine invented in 1799, which concentrates 0.5% paper slurry into a 20% solids cake.

In Europe, belt filter presses have experienced tremendous popularity in dewatering wastewater sludges and are the most common type of mechanical dewatering device in use. In the United States, more than 20 manufacturers sell belt filter presses. Their popularity primarily stems from their relatively low energy requirements when compared to the other mechanical dewatering equipment.

Belt filter presses have found application treating alum sludges when a high final cake concentration is not required. Westerhoff and Daly (4-7) reported on the performance of a pilot belt filter press treating alum sludge at the Erie

County Water Authority. A 2.5% feed solids concentration of sedimentation basin sludge was concentrated to 12% using 9 lb/ton polymer conditioning. Pre-thickened filter backwash sludge was concentrated from 1 to 2% to 10% using 12 lb/ton polymer.

The City of Raleigh, N.C. installed a belt filter press in the early 1980's to dewater its alum sludge. Sedimentation basin sludge and filter backwash water are gravity thickened to a 4.75% solids concentration. This is a relatively high concentration for gravity thickened alum sludges which helps the ultimate performance of the belt filter press. The City has two belt filter presses, each 2 m in belt width. The sludge is treated with about 5 to 6 lb/ton of polymer and then loaded to the presses at a rate of about 270 to 380 lb/ft/hr. Final cake solids concentration averages 22%. The filtrate solids concentration contains about a 1% solids concentration, equivalent to about 80% solids capture. The relatively low solids capture also helps contribute to the high cake solids concentration which is achieved. Some problems were encountered with short belt life, (only about 400 hours.) The belts were replaced with seamless fabric consisting of alternating left and right hand spirals joined by larger diameter cross-machined monofilament. This system makes the ends easy to join on the press and it becomes an endless fabric belt. Life of the belt has increased to about 2,000 hours.

4.9.6. Example Facility

The Orange Water and Sewer Authority (OWASA) located in Carborro, North Carolina owns and operates a 10 mgd conventional treatment plant. The source of water for the plant has been a nearby surface water reservoir. Typical raw water quality is shown in Table 4-15. The plant has used an alum dose of approximately 30 mg/l resulting in an average of over 2,000 lbs/day of alum sludge at a 10 mgd flow rate.

TABLE 4-15

OWASA RAW WATER QUALITY

Parameter	Typical Value	Range
Turbidity	30 ntu	4 - 150 ntu
Color	20 cu	10 - 25 cu
Alkalinity	30 mg/l	
Hardness (as $CaCO_3$)	25 mg/l	
Iron	0.5 mg/l	0.5 - 1.0
Manganese	0.15 mg/l	0.8 - 1.0 mg/l

In February of 1986, OWASA completed the construction of a solids handling facility for processing of both backwash water and settling basin solids. The facility was constructed to complement a land application co-disposal program in which alum sludge combined with sewage sludge has been applied to a 58 acre site. The belt filter press was constructed to provide the flexibility to dewater the sludge on-site and allow for disposal directly in a landfill during periods when land application was limited or not desirable due to weather conditions.

The sludge facility at OWASA included provisions for equalization and collection of backwash solids; storage and thickening of settling basin solids; and dewatering of solids with a belt filter press. A process schematic of the components of the sludge handling facility is shown in Figure 4-53. Backwash water from the plant has been stored in a 150,000 gallon storage tank for equalization of 2 or 3 filter backwashes. Water from the equalization basin has been pumped to a clarifier for removal of solids from the backwash water prior to recycle back to the water plant. The backwash water clarifier that was installed at the plant is a high rate, inclined plate type settler. The ability to add polymer as a flocculent aid was provided but good clarification performance has been achieved without it. The equalized flow through the clarifier has been 200 to 500 gpm. The supernatant has been returned to the treatment plant.

The solids from the backwash clarifier have been collected and thickened to a concentration of 2 to 4 percent. These solids have been sent to a solids storage basin providing up to 100,000 gallons of storage for both sedimentation basin sludge and clarifier solids. The solids storage basin has provided for both storage and additional thickening of the solids. Decant from the solids storage tank has been sent to the equalization basin.

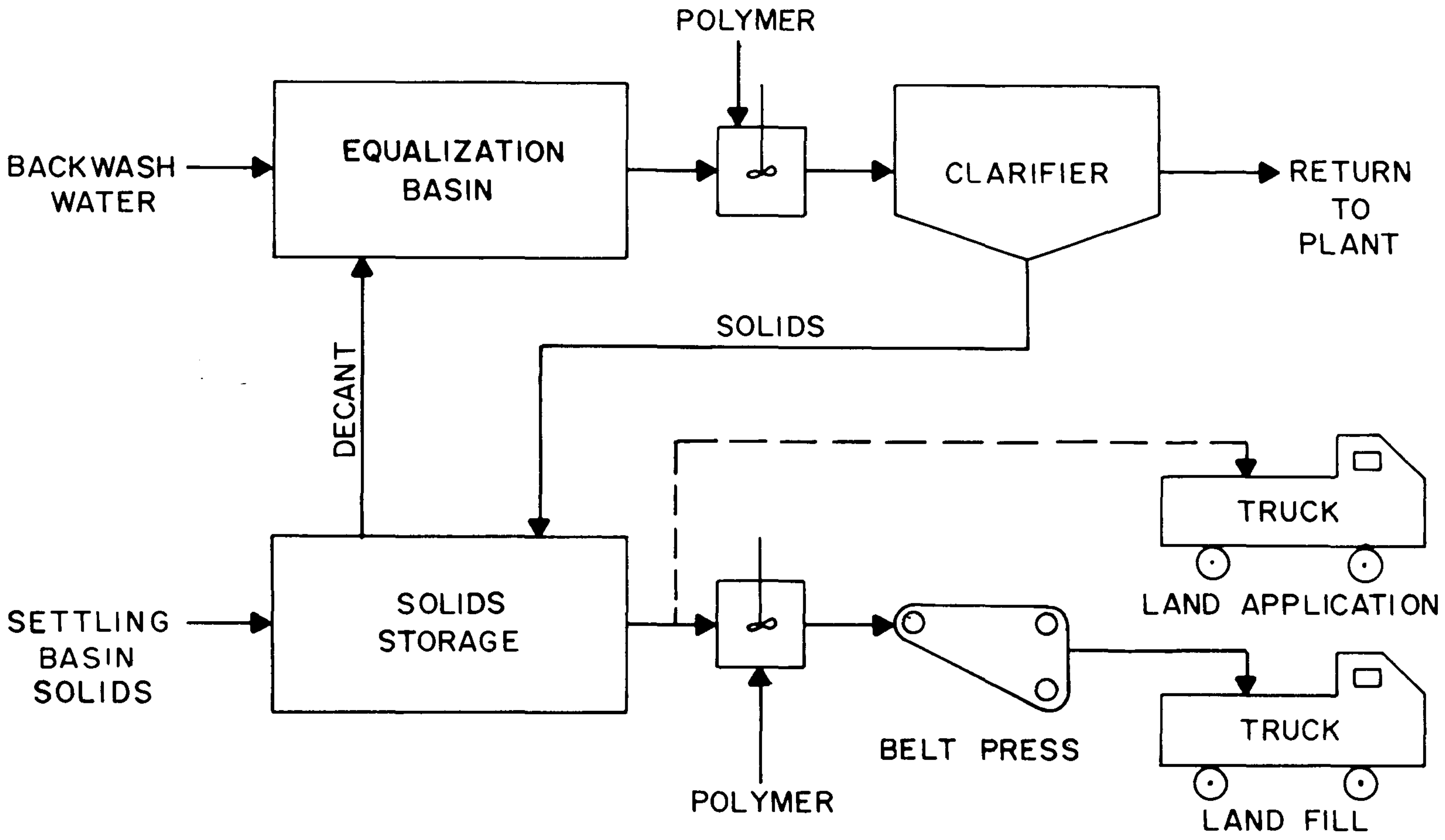


FIGURE 4-53

ORANGE WATER AND SEWER AUTHORITY
SLUDGE FACILITIES

The solids have been thickened prior to pumping from the solids storage tank to the belt press. Conditioning of the sludge with cationic polymer at a dose of about 2 lbs per ton of dry solids has provided for good drainage and dewatering of the sludge once applied to the belt press. Typical feed solids concentrations to the belt press have been from 2% to 4%. The resultant sludge cake solids concentration has generally been about 15%. Sludge feed rates to the belt press (designed to handle 500 to 800 lbs/hr of sludge on a dry weight basis) have been 50 to 80 gpm. Solids capture has ranged from 78 to 99%. A summary of typical operating data from March through August 1986 is shown in Figure 4-54. The equivalent solids loading rate is 80 to 130 lb/ft/hr.

The solids produced from the belt press have been trucked to a landfill for final disposal. During periods when the sludge has been land applied, all of the same equalization, clarification, and solids storage facilities have been utilized except for the belt press. When land application has been used, the thickened sludge has been pumped directly into a tanker truck with sewage sludge at a volume ratio of approximately 2:1 sewage sludge to alum sludge.

4.10. SAND DRYING BEDS

4.10.1. Description[6]

Developed initially for dewatering municipal wastewater sludges, sand drying beds have also been used for dewatering water treatment plant sludges. Sand drying beds operate on the simple principle of spreading the sludge out and letting it dry. As much water as possible is removed by drainage or decant and the rest of the water must evaporate until the desired final solids concentration is reached. Sand drying

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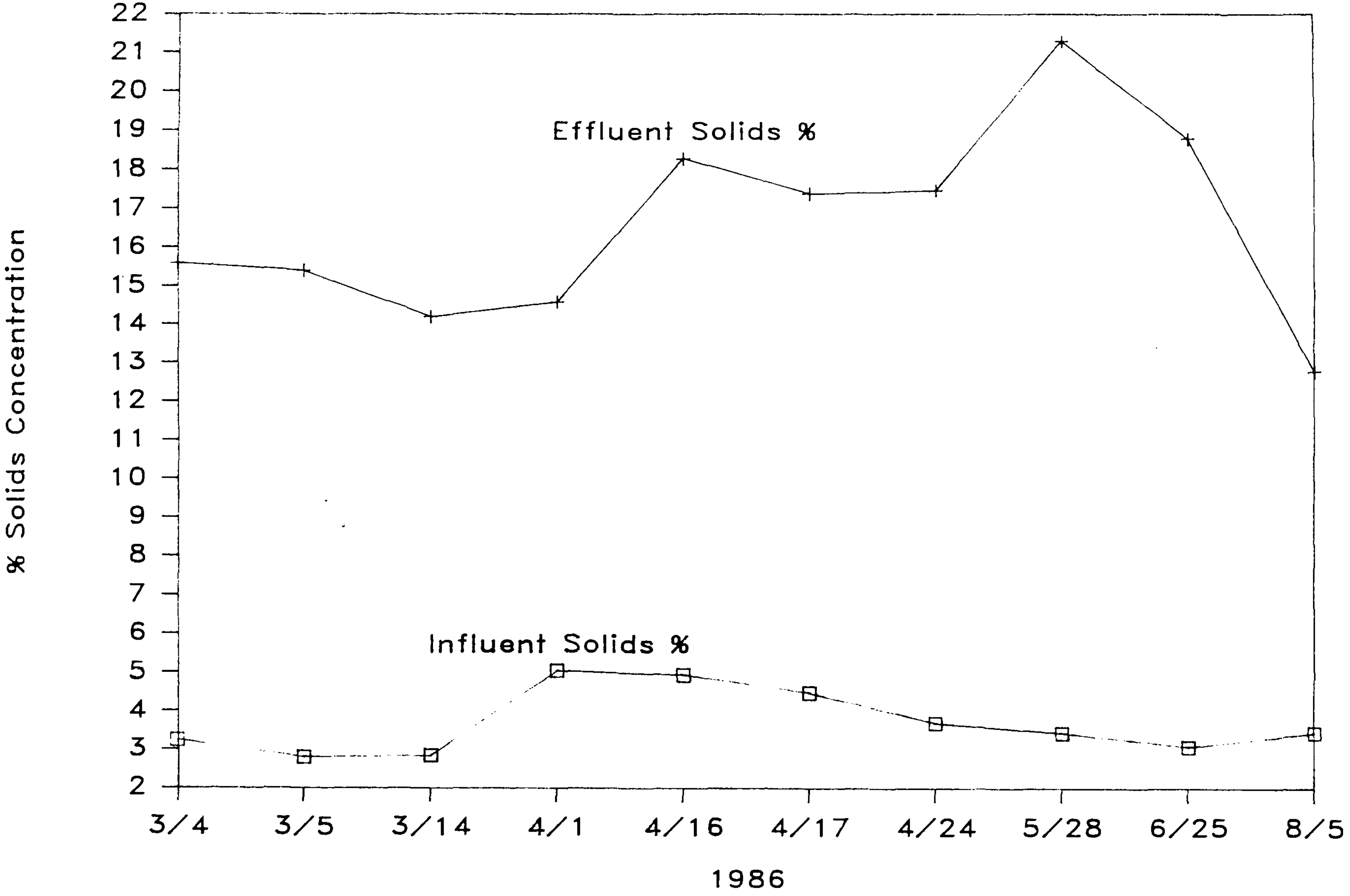
[6] Much of this section on sand drying beds was prepared by A.T. Rolan.

FIGURE 4-54

OWASA BELT PRESS

Solids Concentrations



beds have been built as easily as cleaning an area of land, dumping the sludge and hoping something happens to sophisticated automated drying systems.

Drying beds may be roughly categorized as follows:

1. Conventional rectangular beds with side walls, a layer of sand on gravel with underdrain piping to carry away the liquid. They are built either with or without provisions for mechanical removal of the dried sludge and with or without either a roof or a greenhouse-type covering.
2. Paved rectangular drying beds, with a center sand drainage strip with or without heating pipes buried in the paved section and with or without covering to prevent incursion of rain.
3. "Wedge-water" drying beds which include a wedge wire septum incorporating provision for an initial flood with a thin layer of water followed by introduction of liquid sludge on top of the water layer, controlled formation of cake and provision for mechanical cleaning.
4. Rectangular vacuum assisted drying beds with provision for application of vacuum to assist gravity drainage.

The dewatering of sludge on sand beds is accomplished by two major factors: drainage and evaporation. The removal of water from sludge by drainage is a two-step process. First, the water is drained from the sludge, into the sand and out the underdrains. This process may last a few days until the sand is clogged with fine particles or all the free water has drained away (4-1). Further drainage by decanting can occur once a supernatant layer has formed (if beds are provided with a means of removing surface water). Decanting can also be particularly important with sludges that do not crack for removal of rain. If the rain is not removed it can accumulate on the surface and slow the drying process. Providing for decanting is important for removal of free water released by chemical treatment of sludges applied to drying beds (4-14). The water remaining after initial drainage and decanting must be removed by evaporation.

During the first half of the twentieth century there was little progress in formulating drainage and drying relationships which took into account the parameters of sludge characteristics and quantity, and external factors such as evaporation potential of the air. Just as there are a number of dewatering methods, there are also a number of types of sludge. However, the differences are usually related to the treatment process. Since sludges with the same solids content but from different treatment plants will not dewater at the same rate, other drainage characteristics intrinsic to each sludge must be present (4-15).

The Water Pollution Control Federation Manual of Practice No. 20 (4-16) states that, "The prime objective in sand-bed sludge dewatering is to reduce the moisture content of the sludge cake to a level consistent with the means of sludge cake removal and ultimate disposal. Most well managed, existing plants, have standard procedures for the cycle of sludge dosing and removal of dried cake peculiar to each particular plant based on the type of sludge, the allowable moisture content of dried sludge cake, and the past experience with local climatological conditions. All of these factors combine to determine the drying time. The means of removing the sludge cake may be a controlling factor since the allowable moisture content is dependent on the type of removal equipment and the place or method of disposal."

The means of removing the sludge cake primarily controls the depth of sludge applied to the bed. The depth of sludge applied determines the dried cake thickness at the moisture content that permits the most economical sludge removal. The depth of sludge applied also affects the number of applications per year. The operating costs for sand drying beds are primarily related to the method for removal of sludge from the drying beds; i.e., labor, equipment, and sand replenishment. The most economical operation of a sand bed therefore is the method that minimizes the number of applications per year; i.e., the number of times a bed is cleaned, while obtaining the optimum thickness and moisture content of the dried sludge cake that is most economical to remove and provides for the minimum loss of sand. The mathematical models to be discussed later, can be used to develop data for optimizing the drying bed design and operation.

4.10.2. Design Considerations

The design of a sand drying bed is a function of: 1. the type of sludge to be dewatered, 2. the solids concentration of the applied sludge, 3. the depth of sludge applied, 4. the amount of water removed by decanting and drainage, 5. the evaporation rate -- which is affected by many environmental factors, 6. the type of sludge removal method used, and 7. the ultimate disposal method to be used. All of these factors need to be considered in order to determine the optimum design loading for a given location. Since most of these factors are very site specific, determination of the bed design loading must take into consideration localized differences.

Some of the interrelationships of the variables which should be considered are described below:

Sludge Characteristics - The type of sludge to be dewatered can significantly affect the area requirements for sand drying beds. Lime sludge dewaters much easier on drying beds than alum sludges; however, alum sludges also have significantly different dewatering characteristics from site to site. Chemical treatment with acids and polymers can significantly affect the dewatering characteristics of alum sludge and reduce the drying bed area requirements.

Solids Concentration - The initial dry solids concentration is one of the important factors in determining sizing of sand drying beds. Recent research has shown that for polymer and non-polymer treated alum sludges, bed sizing is very dependent upon the applied suspended solids concentration. For polymer treated sludges, a higher applied suspended solids concentration is generally associated with a lower bed area requirement.

Depth - There are many recommendations relative to the best depth to apply liquid sludge to sand drying beds. For polymer treated alum sludge sizing of beds is relatively independent of applied depth. The design consideration for applied depth would then be the depth of dried cake which is optimal for the removal method and the number of bed cleanings per year. For a given sludge concentration, if the application is comparatively shallow, the sludge dries more quickly but there might be such a small amount of dried cake that more labor would be required to remove a unit volume than if the application depth were thicker. More frequent applications may cause increased loss of sand as a result of the removal process. When the wet sludge is applied to the beds to a greater depth a longer time is required for drying but the thicker cake may be removed more economically.

Decanting and Drainage - A significant portion of the total water to be removed from sludge on sand drying beds is removed by decanting and drainage. The parameter of greatest importance with regard to drainage is not the drainage rate, but the percent of total water that is decanted and drained. Decanting can be very useful when polymers are used to improve sludge dewatering or to remove rain water during the drying cycle. This is particularly significant from a dewatering standpoint since the time required for evaporation is considerably longer than that required for decanting and drainage. Therefore, the total time the sludge must remain on the bed is controlled by the amount of water that must be removed by evaporation. As a result, the amount that can be removed by drainage and decanting must be maximized.

Inorganic and Organic Constituents - The quality of the raw water treated can significantly affect sludge characteristics. Inorganic constituents such as aluminum, iron and manganese can influence decisions about recycle of the decant and underdrainages from sand beds. Adequate consideration must be given to the operational impacts of recycling these and other constituents which may be released from the sludge. Heavy organic concentrations can cause taste and odor problems, if the filtrate is recycled. The recycle may cause increased production of chlorinated organics.

Climate - Regional climatic conditions greatly affect sludge dewatering on drying beds. The drying time is shorter in regions of greater sunshine, low rainfall, and low humidity. Southern localities, where the summers are longer, and arid regions where humidity is low, are more favorable for sludge drying bed use than northern localities. However, higher rainfall and higher humidity in many southern areas may have a significant detrimental effect on drying time. The prevalence and velocity of wind are also factors affecting evaporation rates from sludge beds. Some modification of design criteria may be desirable because of climatic conditions.

Alum sludge dewatering can be dramatically improved by the freeze-thaw cycle in colder climates which causes the rapid release of previously chemically bound water (see Section 4.11.4). Areas of heavy rainfall can create problems for sand drying beds, but with adequate provision for decanting rainwater and the use of polymers, drying beds can be effective when properly designed.

Loading Rate - Loading rate is a term commonly used in design of sand beds, usually expressed as pounds of dry sludge initially applied per square foot of sand filter area. It is determined as a function of suspended solids and applied depth:

$$IA(lb/sf) = \frac{(\%SS\ Applied)\ (62.4)\ (ft\ Applied\ Depth)}{100}$$

Figure 4-55 shows the simple relationship between suspended solids concentration and applied depth. For a given applied solids concentration the optimal loading is often in the 2 to 3 lb/sf range. However, the amount of sludge that can be processed per given bed area is dependent upon the applied solids concentration. Therefore, the loading rate by itself should not be used as a design parameter.

Design Basis - As with other methods of sludge handling, the design of sludge drying beds can be based on experience or by scale-up from laboratory tests.

Design manuals often include a table showing the required sludge bed areas in terms of population or quantity of water expected at the treatment plant. There has been very little data published on design standards for water plant sludges. Unfortunately, very little research has been done in the last 20 years on the design standards for sand drying beds for either wastewater or water treatment plant sludge.

Most data have previously been presented in terms of square feet of bed surface area required for dewatering on a per capita basis. This criterion is only valid for the characteristics of a particular waste and has no rational design basis if applied to water plant sludge. There is not an easily applied factor which can be used to design a sand drying bed. The best criteria would take into consideration climatic conditions (such as temperature, wind velocity, and precipitation), sludge characteristics, solids concentration and applied depth (4-17).

The currently accepted design criteria for sand drying beds are based on empirical data. There have been several attempts (4-18, 4-19, 4-20, 4-21, 4-22) at developing models to describe mathematically the complex relationships involved in the proper functioning of sand drying beds. Skinner (4-18)

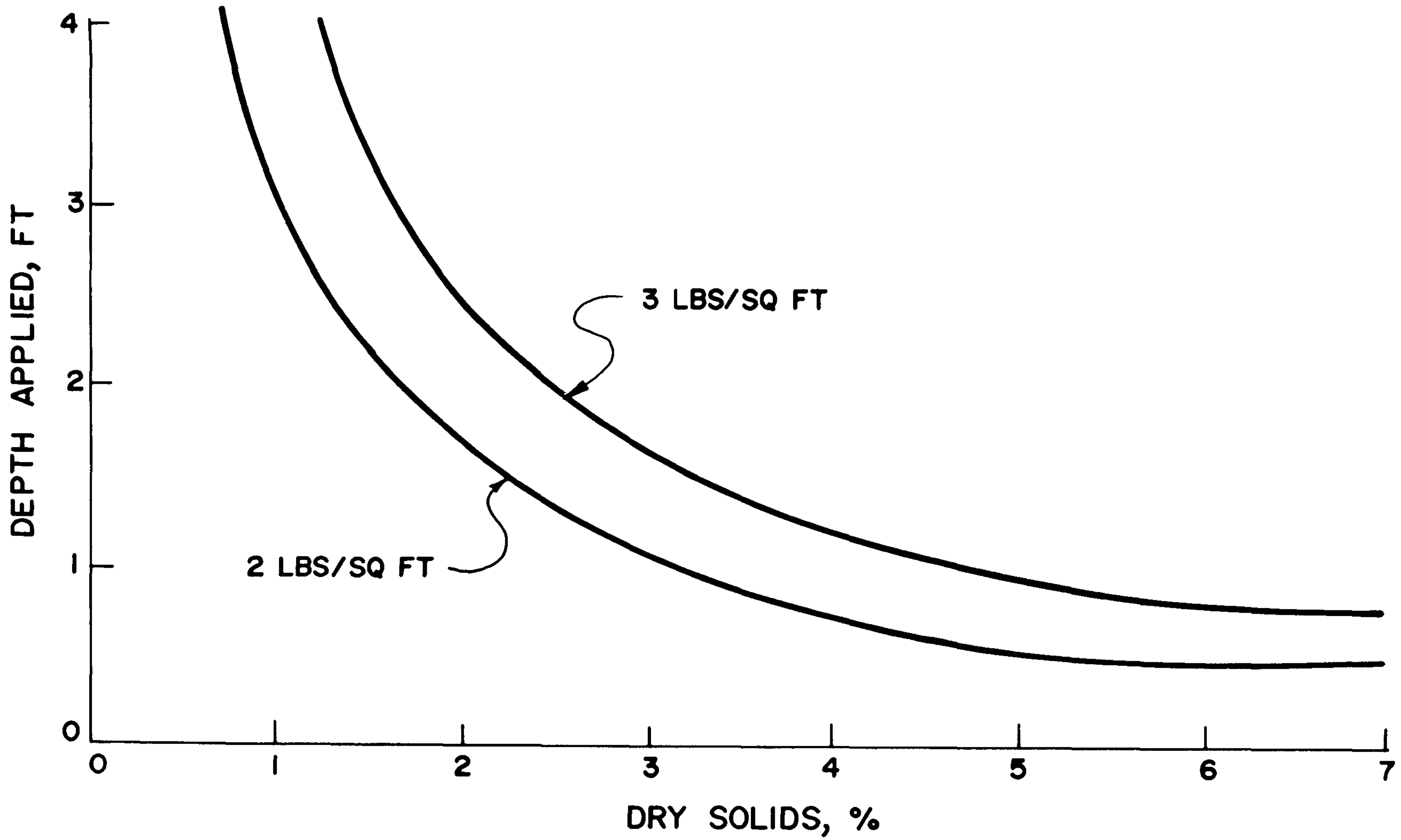


FIGURE 4-55

DEPTH REQUIRED TO OBTAIN OPTIMUM LOADING
AT VARIOUS SLUDGE CONCENTRATIONS

developed one of the first empirical equations for the computation of drying bed requirements for a wastewater treatment plant, as follows:

$$\text{Area/Capita} = \frac{(\text{Constant})(\text{Avg annual ppt - in})(\text{SSppm})}{(\text{No. of months})(\text{Mean Annual Temp})(\text{Mean wind vel mph})}$$

where the constant depends on the sewage treatment process. For covered beds the area calculated above is divided by 2. The formula points out the different factors that affect sludge drying.

Haseltine (4-23) developed one of the first empirical relationships describing the time required for sludge to dewater on sand drying beds. From field data collected at different treatment plants, he plotted the gross bed loading ($kg/m^2/day$) versus solids content of the sludge. The resulting equation was determined to be:

$$Y = 0.157\ S_0 - 0.286$$

where,

Y = gross bed loading of solids ($kg/m^2/day$)
S_0 = solids content (%) of applied sludge.

Since final moisture content of the sludge was also thought to be a determining factor of the sludge drainage and drying capacity of a sand bed, Haseltine defined the term "net bed loading" as the gross bed loading multiplied by the final water content of the dried sludge. The relationship between net bed loading and solids content was found to be

$$Z = 0.057\ S_0 - 0.082$$

where,

Z = net bed loading of solids (kg/m^2/day)

S_0 = solids content (%) of sludge charged to the bed

Although Haseltine's equations are strictly empirical, they have been used for the dimensioning of sludge drying beds.

A more recent attempt at developing a mathematical model for sand drying beds was developed by Rolan (4-20). A series of equations were developed which can be used to determine the design criteria for sand drying beds as well as to determine their optimum operation.

The relationships are based on being able to describe the two water loss mechanisms - drainage and evaporation - in terms useful for the design and operation of the drying beds. The drainage term also includes any water which is decanted from the beds. This water loss generally occurs over the first one to two days. The remainder of the water removed is due to evaporation.

Below are several definitions used by Rolan in development of the relationships:

IA = Initial application of sludge in pounds dry solids per square foot.

D(i) = Depth applied initially in inches.

DS(i) = Percent dry solids initially.

D(f) = Depth in inches at desired final dry solids concentration.

DS(f) = Percent dry solids concentration desired for final cake.

DD = Change in depth.

DD(u) = Change in depth due to loss of water to underdrain and decanting on bed.

DD(e) = Change in depth due to loss of water to evaporation.

P = Percent of volume applied to beds which passes through sand or is decanted, expressed as a decimal fraction.

T = Drying time in months.

E = Rate of evaporation in inches per month.

AA = Number of applications per year.

Y = Bed yield in pounds dry solids per square foot per year.

The initial loading of sludge (IA) in pounds per square foot for a given application can be calculated, based on the depth of sludge applied, D(i), and the dry solids content, DS(i), of the sludge applied.

$$IA = \frac{D(i)}{12 \text{ inches per foot}} \times \frac{62.4 \text{ pounds}}{\text{cubic feet}} \times \frac{DS(i)}{100}$$

The desired depth of the sludge at the time of removal is primarily dependent on: 1. the dry solids content desired for ultimate disposal, or 2. the depth or solids concentration required for efficient removal. The final depth, D(f), which is a function of the initial loading, IA, in pounds per square foot and the percent dry solids content desired, DS(f), for removal is found by:

$$D(f) = D(i) \times \frac{DS(i)}{DS(f)}$$

The change in depth, DD, is determined by subtracting the final depth, D(f), from the initial depth, D(i)

$$DD = D(i) - D(f)$$

The loss of moisture to the underdrain system and decanting is reflected in the rapid change in depth of sludge, DD(u), immediately following the application of sludge to the bed. The loss of moisture to the underdrain system is dependent on the type and depth of sludge applied as well as the solids content. Any water decanted from the bed would also be included in DD(u). In addition, polymer conditioning of sludges can have a significant impact on the percentage of moisture lost to the underdrains and decanting. The change in depth, DD(u), due to the loss of water to the underdrains can be calculated by:

$$DD(u) = D(i) \times P$$

if a value for the percentage loss (P) to the underdrains and decanting can be determined.

The change in depth due to evaporation accounts for the remaining loss in depth

$$DD(e) = DD - DD(u)$$

The time, T, required to accomplish the evaporation is dependent on the evaporation rate, E. Evaporation and the resultant changes in depth may not be linear because of changes in sludge characteristics prior to and following the formation of surface cracks. Also, bridging within the sludge cake may prevent compaction. However, for the purposes of these relationships, evaporation and depth change was assumed to be unaffected by these factors. Because evaporation rates exhibit seasonal variations, the annual average evaporation rate expressed in inches per month can be used in attempting to model the "average" performance of sand drying beds. Since winter evaporation rates can be significantly less than summer rates, analysis should be made for seasonal variations. The time for the sludge to dry, therefore, ignoring the one to two days of drainage time is given by:

$$T = DD(e)/E$$

The number of applications, AA, to each bed which can be accomplished in a year is therefore dependent on the evaporation time

$$AA = \frac{\text{12 months per year}}{T}$$

Finally, the bed yield, Y, in pounds per square foot per year is a function of the sludge applied and the number of applications per year.

$$Y = \frac{\text{cubic feet of sludge}}{\text{square foot to bed}} \times \frac{\text{pounds}}{\text{cu ft}} \times \frac{\text{percent solids}}{100} \times AA$$

$$= \frac{1' \times 1' \times D(i)/12}{\text{square foot}} \times \frac{\text{62.4 pounds}}{\text{cubic foot}} \times \frac{DS(i)}{100} \times AA$$

$$= IA \times AA$$

In order to determine the solids loading at a given percent dry solids and depth of applied sludge using these equations, an accurate estimate of the seasonal or average annual evaporation rate is needed. Figure 4-56 is a national evaporation map that can be used if local values are not available.

Figure 4-57 was obtained by using the model equations to calculate cake thickness at 25 percent dry solids and 65% for P for sludges for varying initial solids concentrations applied at 12, 18 and 24 inches in depth. Similarly, the model equations were also used to determine the maximum number of applications per year at various solids concentrations and depth of applied sludge as shown in Figure 4-58.

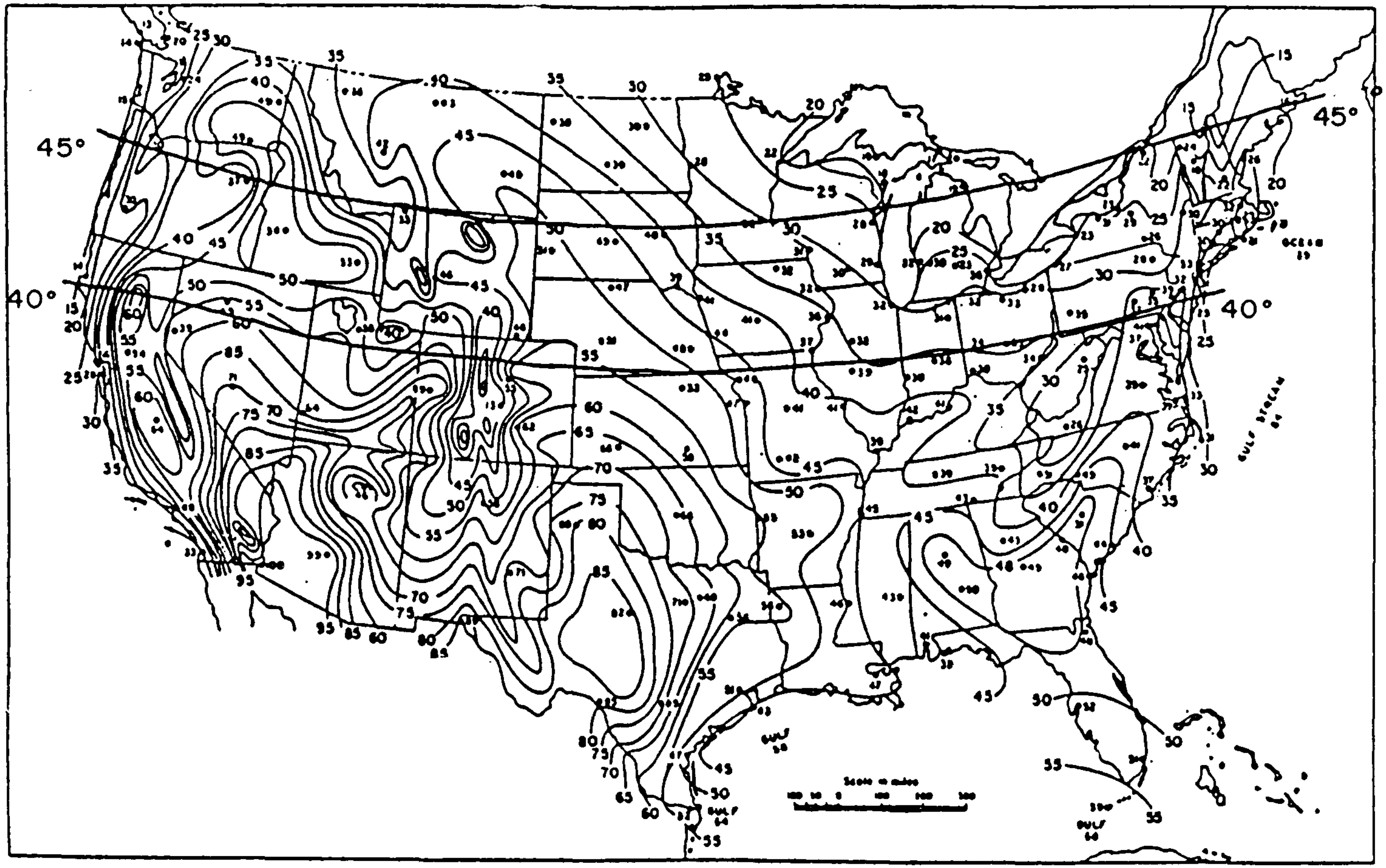


FIGURE 4-56

MEAN ANNUAL EVAPORATION

FIGURE 4-57

DRY CAKE THICKNESS VS DEPTH APPLIED

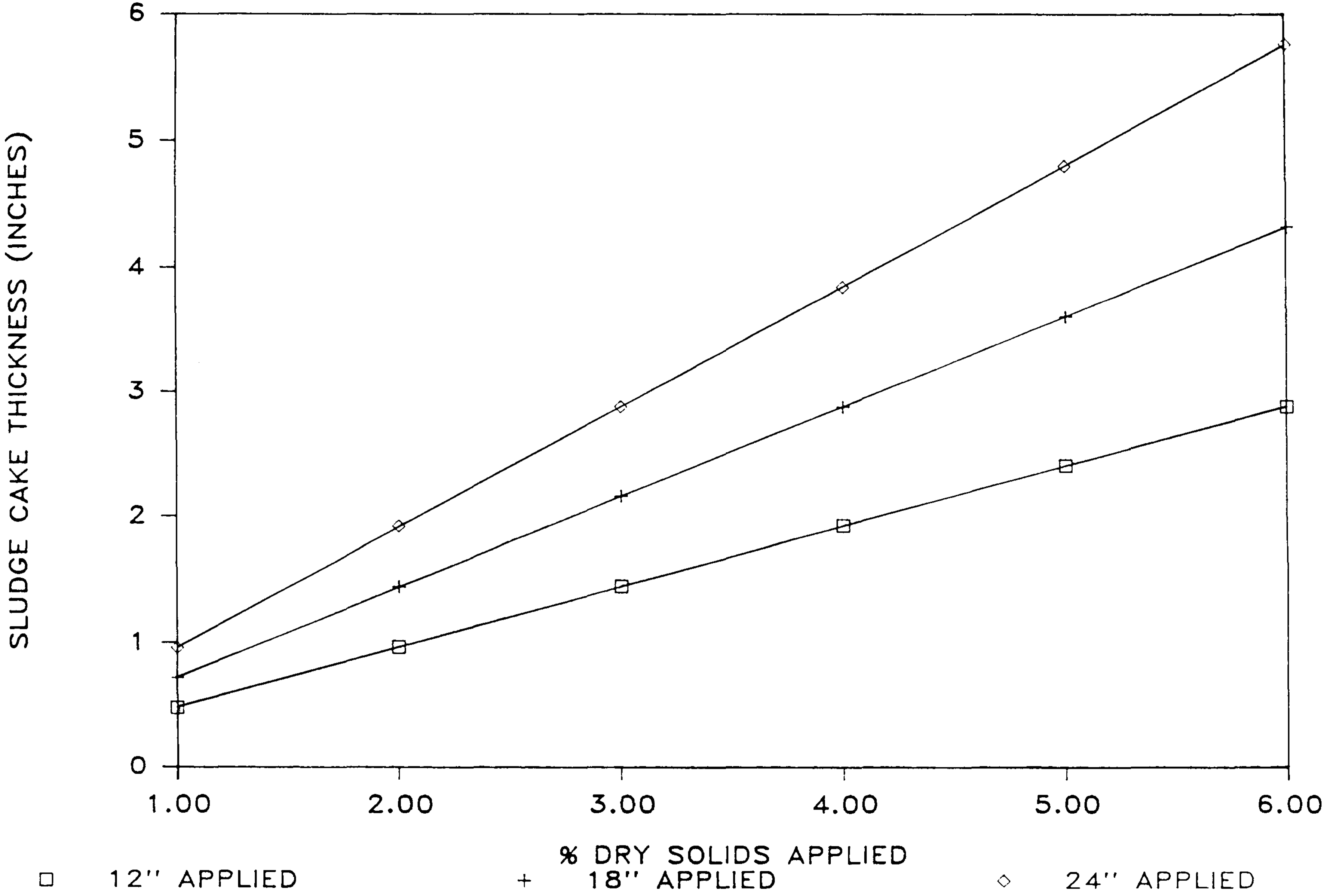
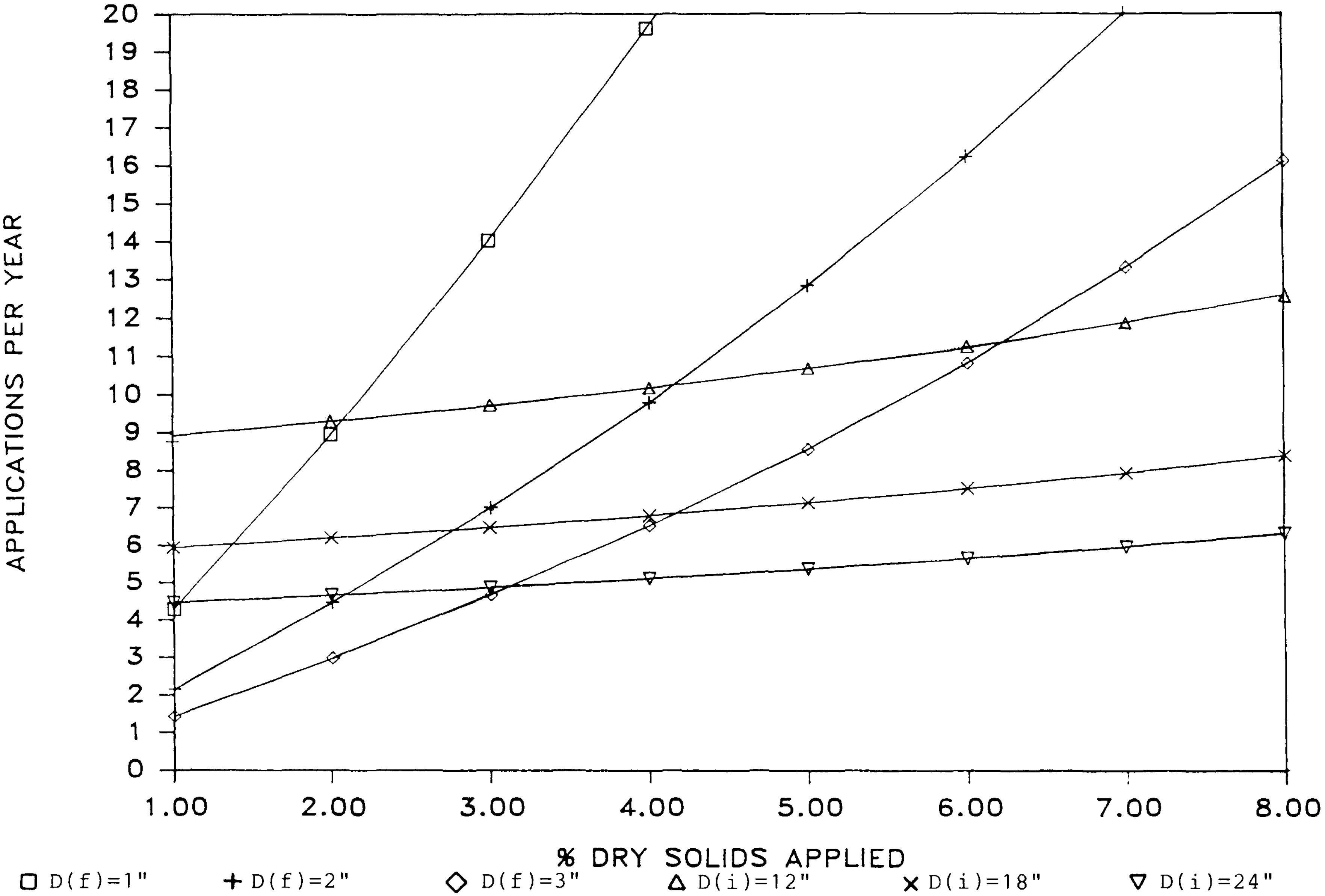


FIGURE 4-58

APPLICATIONS/YEAR VS % DRY SOLIDS



As an example of how these graphs could be used to optimize a drying bed operation, the following information was considered:

The City of Durham, North Carolina, uses a truck-mounted, vacuum-sludge removal system for cleaning sand drying beds. This unit has eliminated the labor-intensive problem commonly associated with cleaning sand drying beds. A properly designed sand drying bed can be cleaned by one man operating the vacuum unit from inside the driver's compartment using hydraulic controls. This unit works best with a dried sludge cake of 25 percent dry solids and a thickness of 3 to 4 inches. The expected solids content of the water plant sludge to be applied to the beds would be 2 to 3 percent dry solids.

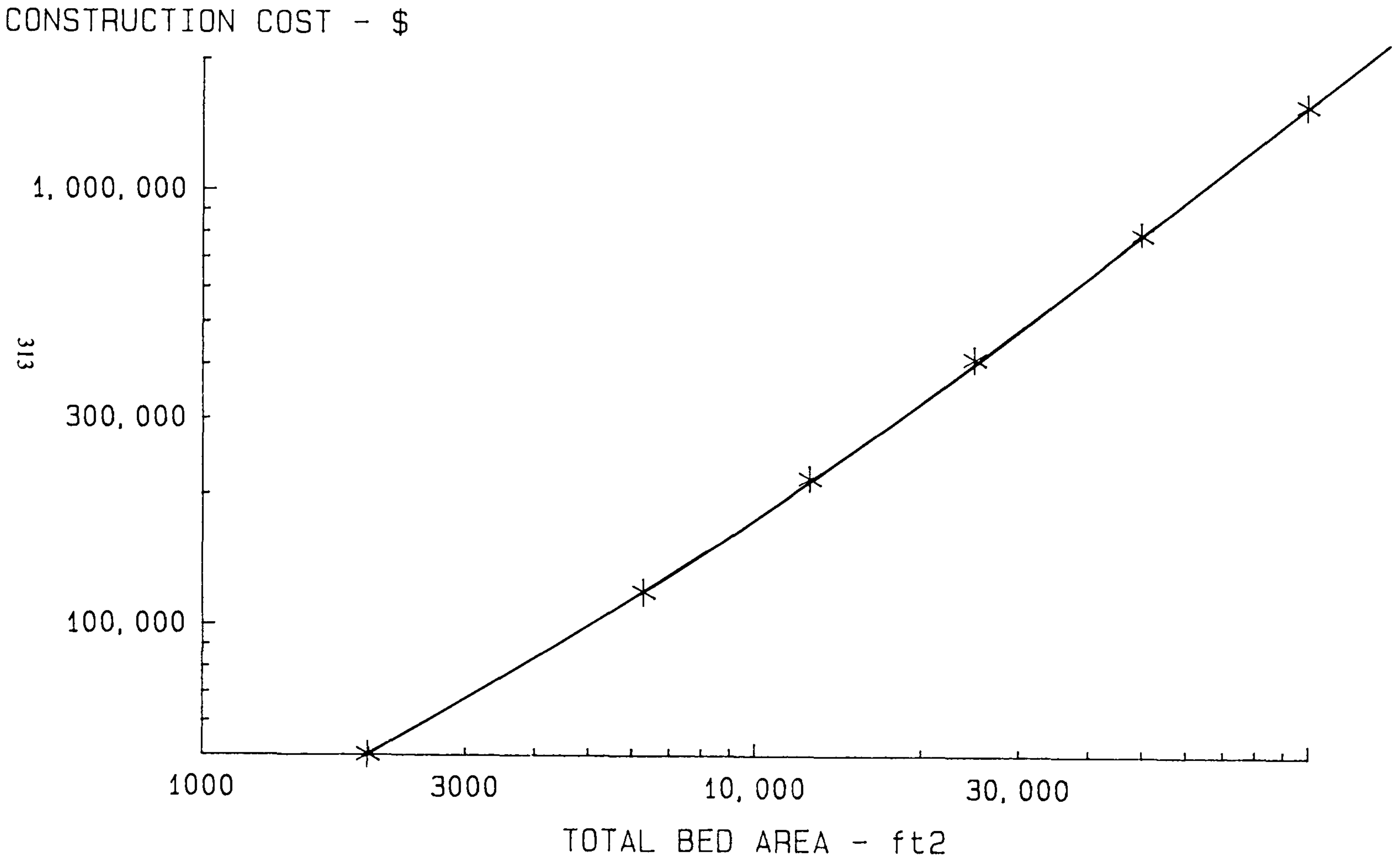
From Figure 4-57 it is apparent that for this example, sludge of 3 percent dry solids applied at 18 to 24 inches depth would produce the desired final sludge cake thickness. Figure 4-58 indicates that 18 to 24 inches of 2 to 3 percent dry solids sludge applied would mean 4 to 6 applications (and cleanings) per year. It should also be noted however, that the commonly suggested sludge application depth of 8 to 12 inches would not be practical for this example because of the thin cake produced and the greater number of applications (and cleanings) per year required in order to obtain the needed bed yield in pounds per square foot per year.

4.10.3. Capital and Operating Costs

Cost curves for sand drying beds are included in this section. Construction costs are shown in Figure 4-59. Construction costs included excavation and backfill, concrete walls and floor, granular media and pipes and valves. Installation labor is included under each item of work. Excavation work was performed such that finished grade on top of sand matched initial ground elevation. Concrete walls were assumed at 8 inches thickness; concrete floor, 6 inches thickness. The sand layer was calculated as 18 inches thick

FIGURE 4-59

CONSTRUCTION COST FOR SAND DRYING BEDS



with gravel supporting layer and underdrain media. The feed pipe was sized as 6 inches ductile iron pipe. The underdrains were 6 inch perforated PVC pipe. The collection piping was 6 inch PVC for the 2,000 sf bed and 12 inch PVC for the 6,300 sf bed and larger beds.

The operation and maintenance costs are shown in Figure 4-60.

The operation and maintenance costs include fuel, maintenance material and labor. All O & M costs are relative to the removal of dried cake from the beds and bed preparation for the next application of sludge. The fuel costs are for a front end loader. Maintenance material costs were calculated assuming one-quarter inch of sand to be place 20 times per year. Labor costs for sand drying bed operation were based on staffing, requirements at several known installations of comparable size.

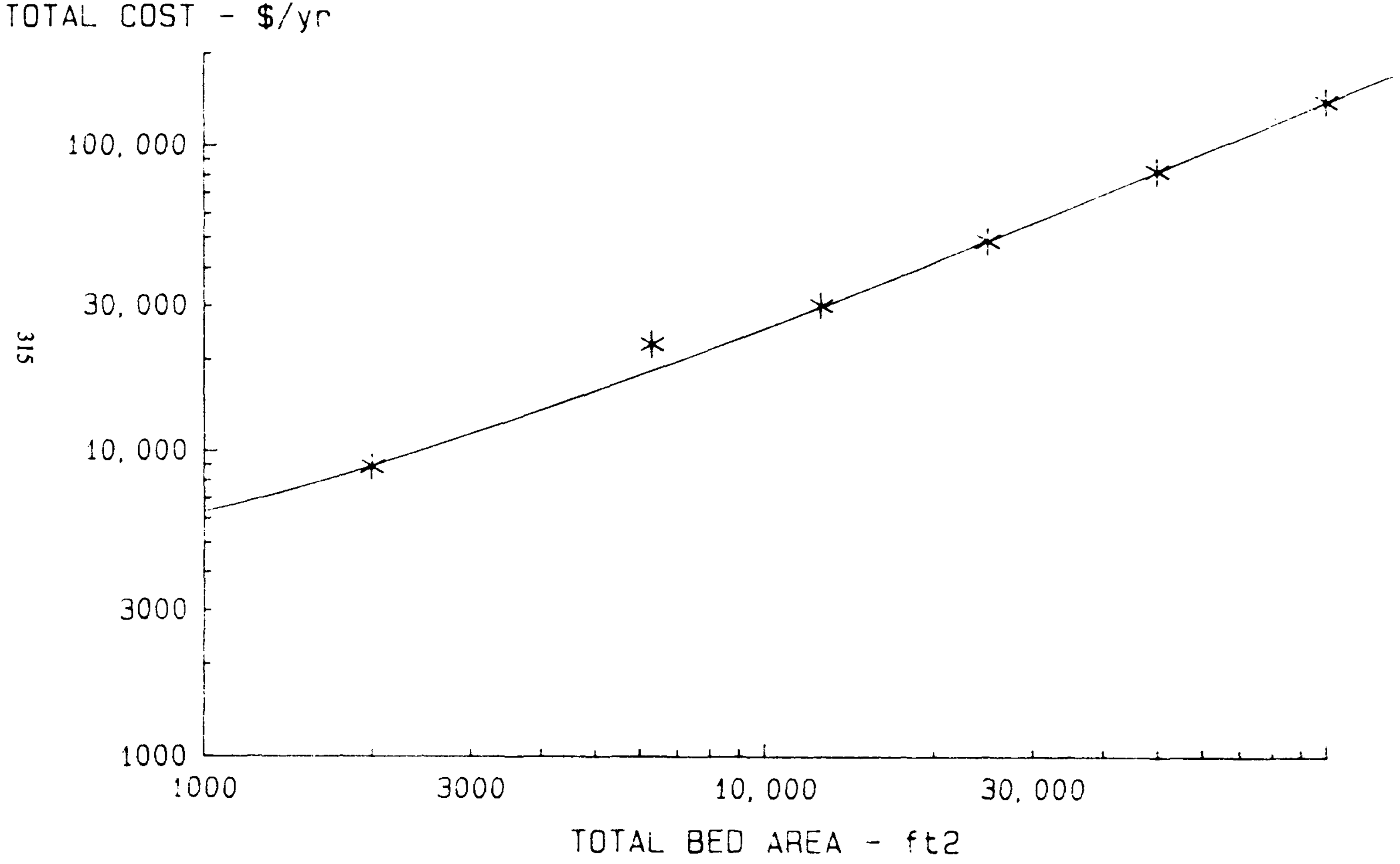
4.10.4. Operating Considerations

The prime objective in sand bed sludge dewatering is to reduce the moisture content of the sludge cake to a level consistent with the means of sludge cake removal and ultimate disposal. Most well-managed existing plants have standard procedures for the cycle of sludge dosing and removal of dried cake based on the type of sludge, the allowable moisture content of the dried sludge cake, and local climatological conditions. The allowable moisture content, which is dependent on the place or method of disposal and the type of removal equipment may be the controlling factor for most drying bed operations.

Rarely are mechanical dewatering devices designed without provision for chemical feed. It is only recently, however, that designers have begun to make the same provision for sludge drying beds. New designs should include chemical conditioning to offset unpredictable weather conditions and variable sludge characteristic (low percent solids, poor

FIGURE 4-60

OPERATION AND MAINTENANCE COST FOR SAND DRYING BEDS



drainability, or changing chemical characteristics). Conditioning also may be useful in improving the sludge drainage rate and thereby increase the capacity of drying beds.

The capillary suction time (CST) meter and the time to filter test (TTF) may be used for comparative evaluation of both polymer type and dosage. The best economic evaluation makes comparisons based on the grams of chemical added per kilogram of dry sludge solids (lb/ton of dry solids), rather than on a parts per million dosage.

In addition, optimum dosages should be determined with care, because their effectiveness can be hampered by both under and overdosing. The net and gross sludge bed loadings for chemically treated and untreated beds should be compared in laboratory tests and under actual field conditions. Blinding of the sand may result if excessive amounts of chemicals are used.

When the use of polymers is planned, the bed operation should take into account the angle of repose of applied sludge. With polymer treatment and the rapid release of free water, the uniform distribution of sludge on the sand surface can be affected by the angle of repose of the sludge--often as much as 1 inch vertical for each 10 feet horizontal. This problem can be resolved, however, by providing multiple points of addition, providing for partitioning of beds, or limiting the size of each bed.

Equipment limitations (i.e. type of pumps and dimensions of beds) may require the operator to apply a more dilute sludge to minimize problems with pumping and distribution of sludge on the bed. By applying a thinner sludge at a high rate, the angle of repose effect can be reduced. The free water which could have otherwise been removed by decanting prior to applying sludge to the bed can be easily removed through filtration and decanting on the bed. A uniform layer of sludge is essential to uniform drying and efficient utilization of drying bed area. A trade-off is that thinner sludge may blind the sand at a lower net drainage, thereby increasing the bed area requirements.

Increased labor costs have made manual sludge removal economically feasible only in the smallest plants. During manual removal, hand tools are used to lift the dried cake from the sand. One of the best tools is a shovel-like fork with several tines, approximately 25 mm (1 in.) apart. For best results, the cake must be dried sufficiently (generally 25 to 30% dry solids) so that it cracks and begins to peel away from the sand. The cake then can be lifted from the sand with a relatively small loss of sand. The removed cake may be forked directly into wheelbarrows, small trucks or small rubber-tired wagons. The sand surface should not be required to take wheel loads; concrete treadways or appropriate temporary planking can be used instead.

Many plants now use mechanical removal equipment consisting of either front-end loaders or truck-mounted vacuum removal systems, thereby reducing the labor requirements for sludge removal to a minimum. Because of the cost of operating mechanical removal equipment, the dry cake thickness and moisture content must be optimized. Generally, a dry solids content of 15 to 25% is sufficient for mechanical removal.

4.10.5. Example Facility

The City of Durham, North Carolina owns and operates two water treatment plants one of which (Williams Water Treatment Plant) utilizes sand drying beds for alum sludge dewatering. The source of water for the plant is a nearby surface water reservoir. The plant has utilized alum as a coagulant at an annual average dose of 30 mg/l alum. The monthly average dose is 15 to 45 mg/l. Typical raw and finished water quality is shown in Table 4-16.

Sand drying beds have been used for sludge dewatering at the plant since 1978. The sludge handling facilities associated with the beds include decant tanks, sludge pumping and

TABLE 4-16

DURHAM WATER QUALITY

Turbidity (NTU)	Average	Typical Range
Raw	43.0	8.0 - 140
Settled	5.0	1.0 - 15.0
Finished	0.3	0.1 - 1.4
Color (cu)		
Raw	41.0	20.0 - 60.0
Finished	<5.0	<5.0 - 5.0
pH		
Raw	6.7	6.3 - 6.8
Settled	5.5	5.0 - 6.0
Finished	7.0	6.7 - 7.7
Total Alkalinity (mg/l)		
Raw	15.2	6.0 - 25.0
Finished	17.6	9.0 - 24.0
Total Hardness (mg/l $CaCO_3$)		
Raw	20.0	10.0 - 26.0
Finished	22.0	12.0 - 26.0

polymer feed equipment. A schematic of the sludge handling facilities and drying beds is shown in Figure 4-61. The facility has both a batch solids thickening (decant tanks) and dewatering process. The settling basins were constructed with hopper bottoms equipped with manually operated plug valves. Sludge from the settling basin having a solids concentration from 0.5 to 1.0% is sent to one of two 77,000 gallon decant tanks. Thickening of the sludge in the decant tanks for several days to a week has resulted in a settled solids concentration of 3% to 5%. Decant water from the tanks is returned to the raw water terminal reservoir for recycle to the water treatment plant. The solids from the decant tanks are pumped to the drying beds. A nonionic polymer at a dose of 12 to 30 lbs per ton has been used to aid in dewatering. The polymer is fed on the suction side of the sludge pumps to provide mixing prior to discharge onto the drying beds.

The sludge is applied to four drying beds each 5,100 square feet in surface area at a loading of approximately 2 to 3 lbs of solids per square foot. The sludge depth generally varies from one to two feet. The solids have been allowed to dry to a concentration of 15 to 20% before being removed from the beds. Sludge is removed from the beds by the use of vacuum trucks which also haul the sludge to the local City owned landfill.

One full time operator has been assigned to the facility and has been responsible for solids transfer, thickening, decant of supernatant water, polymer feed operations and general facility maintenance. Plant laboratory personnel have been responsible for the selection of polymer dosage based on the performance of capillary suction time (CST) testing. Based on operational experience a polymer dose which would provide a CST of 15 seconds or lower has been desired to provide good dewatering characteristics. Generally, the thickened sludge without polymer addition has a CST of over 150 sec. Polymer conditioned sludge has generally had a CST in the 6 to 15 sec range.

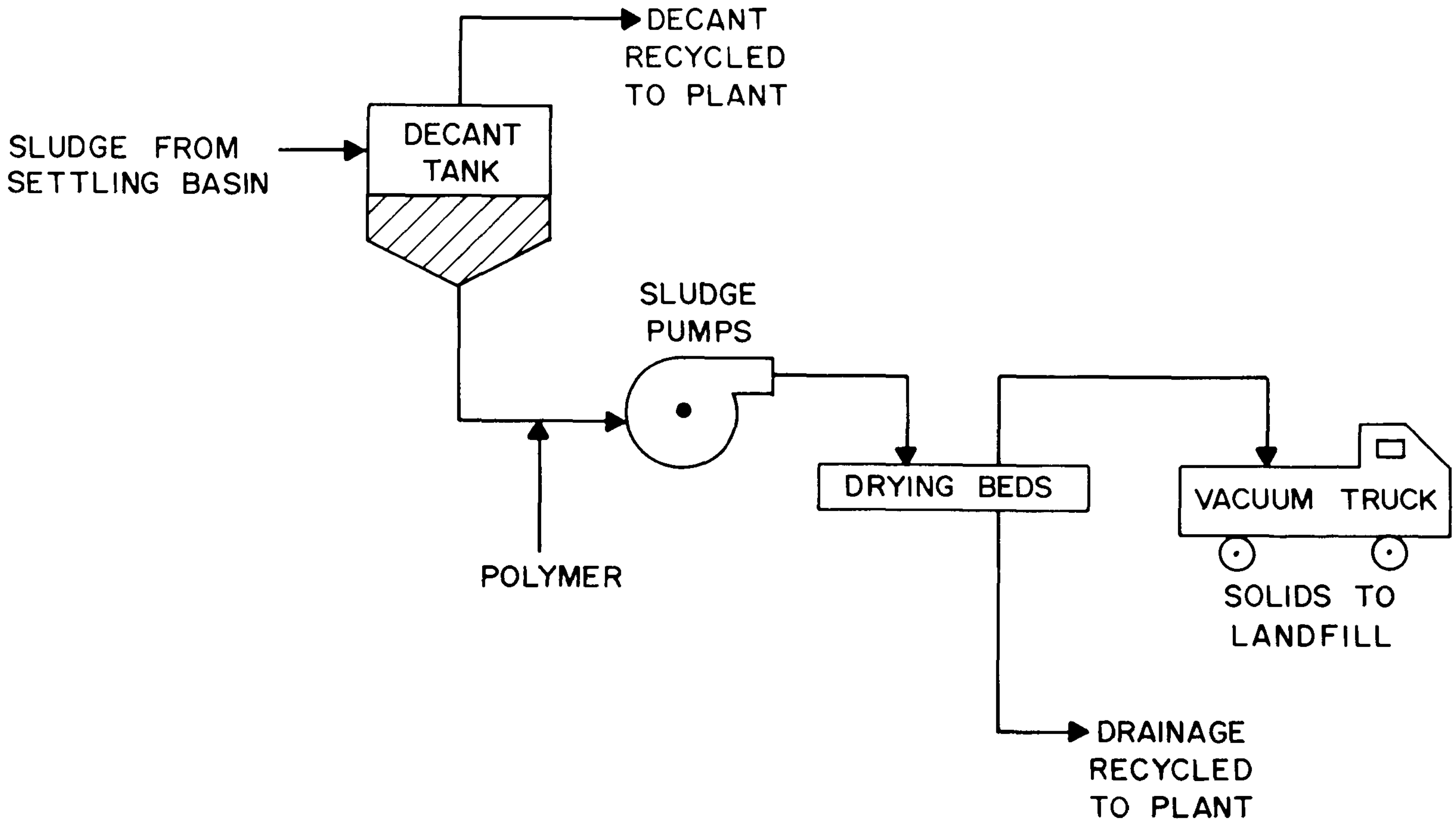


FIGURE 4-61

**DURHAM DRYING BED FACILITIES
PROCESS SCHEMATIC**

The amount of sludge produced at the plant has been estimated based on plant tests and correlation of alum dose and turbidity. Approximately 650 tons per year (1.78 TPD) of solids on average has been produced. The monthly average of solids production has ranged from 0.5 to 3.9 tons per day based on historical records (see Figure 4-62).

The drying time on the beds has been a function of climate. During warm weather the beds have been unloaded after one to two weeks of drying time. During winter months the drying time has been one to two months. One of the most frequent operational problems with the beds has been the need to dry for long periods during winter months. Typically, sludge has been hauled by tanker trucks to nearby lagoons during high solids loading periods or during winter months when the drying beds can not keep up. The City has planned construction of more drying beds to handle all sludge production peaks.

4.11. DEWATERING LAGOONS

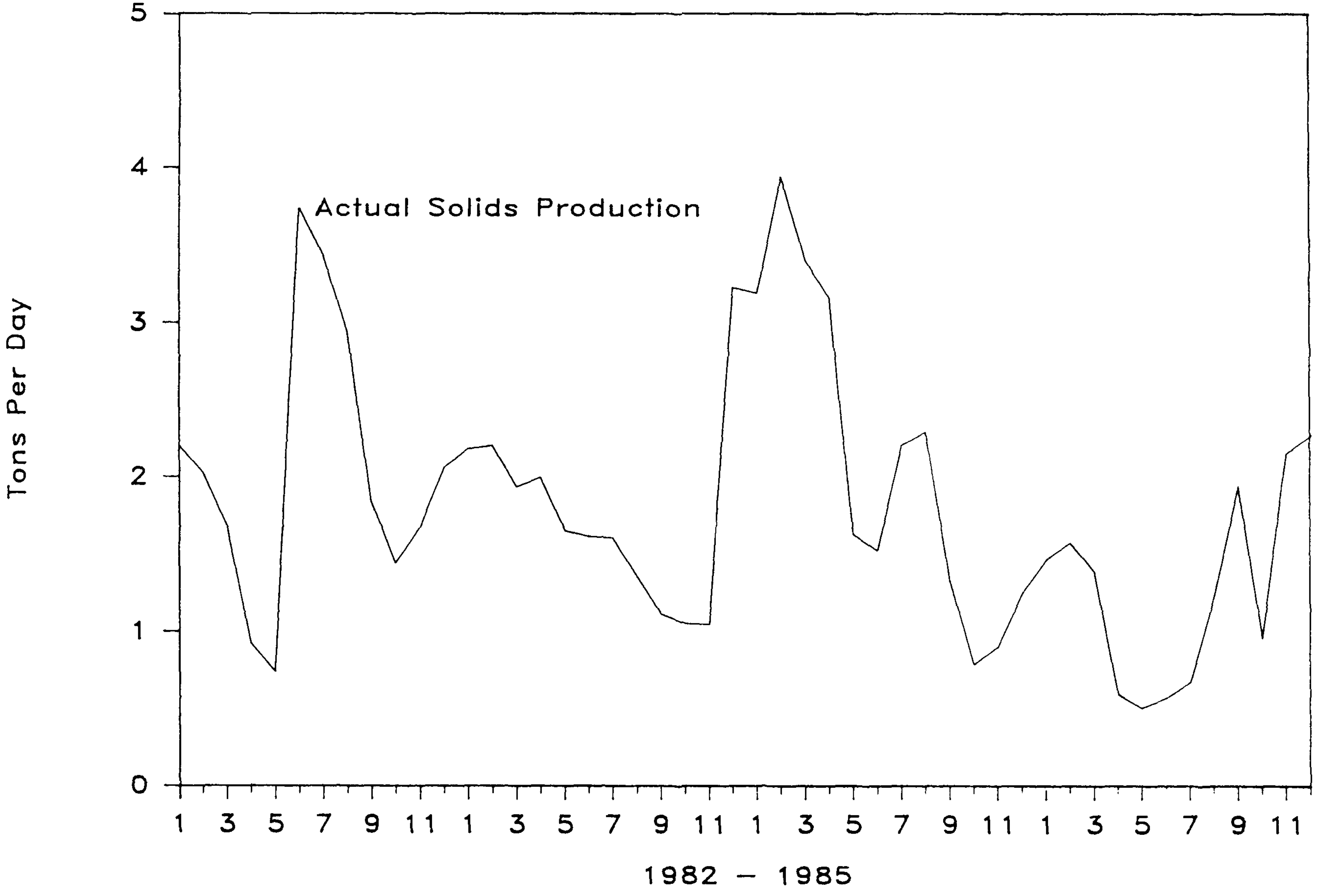
4.11.1. Description

Lagoons can either be constructed as storage lagoons or dewatering lagoons. Storage lagoons are designed to store and collect the solids for some predetermined amount of time. They will generally have decant capabilities but no underdrain system. Storage lagoons should be equipped with sealed bottoms to protect the groundwater. Once the storage lagoon is full or decant can no longer meet discharge limitations it must be abandoned or cleaned. To facilitate drying, the standing water may be removed by pumping, leaving a wet sludge. Coagulant sludges can only be expected to reach a 7 to 10% solids concentration in storage lagoons. The remaining solids must be either cleaned out wet or allowed to evaporate. Depending upon the depth of the wet solids, evaporation can take years. The top layers will often form a crust preventing evaporation of the bottom layers of sludge.

FIGURE 4-62

SOLIDS PRODUCTION

DURHAM – WILLIAMS WATER PLANT



The primary difference between a dewatering lagoon and a storage lagoon is that a dewatering lagoon has a sand and underdrain bottom, similar to a drying bed. Dewatering lagoons can be designed to achieve a dewatered sludge cake. The advantage of a dewatering lagoon over a drying bed is that storage is built into the system to assist in meeting peak solids production or to assist in handling sludge during wet weather. The disadvantage is that the bottom sand layers can blind with multiple loadings, thereby increasing the required surface area as compared to conventional drying beds. Polymer treatment can be useful in preventing this sand blinding.

4.11.2. Design Considerations

Storage lagoons, which are generally earthen basins, have no size limitations but have been designed in areas from 0.5 to 15 acres, ranging in depth from 4 to 20 or more feet. Storage and dewatering lagoons may be equipped with inlet structures designed to dissipate the velocity of the incoming sludge. This minimizes turbulence in the lagoons and help prevent carryover of solids in the decant. The lagoon outlet structure is designed to skim the settled supernatant and is sometimes provided with flash boards to vary the draw-off depths. Any design of a storage lagoon must consider how the sludge will be ultimately removed unless the site is to be abandoned.

The basis for design of dewatering lagoons is essentially the same as that for sand drying beds. The difference is that the applied depth is higher and the number of applications per year is greatly reduced. During the pilot study phase, careful consideration should be given to the effect that continual or multiple loading has on the volume of water removed by decanting and drainage. The surface area required for a dewatering lagoon will be equal to or greater than that required for a sand drying bed.

4.11.3. Capital and Operating Costs

Cost curves for sludge storage lagoons only are included in this Section. Construction costs are shown in Figure 4-63. Construction costs include excavation and sitework, concrete inlet and outlet structures and pipes and valves. All costs are installed costs. The excavation cost was derived assuming the excavated material could be used for construction of a peripheral berm and that the cuts balanced the fills. The depth of the lagoon was assumed to be 10 feet. The inlet structure included slope protection but no flow distribution. The outlet structure included decant drawoff capability and decant outlet piping and valving. No select material was assumed for the lagoon liner. If a natural clay layer is not present costs for a bottom liner should be added. No underdrain system was included in the costs. If the lagoon is to be used as a continuous dewatering lagoon with decant and underdrains, the sand dry bed cost curves are more appropriate to use.

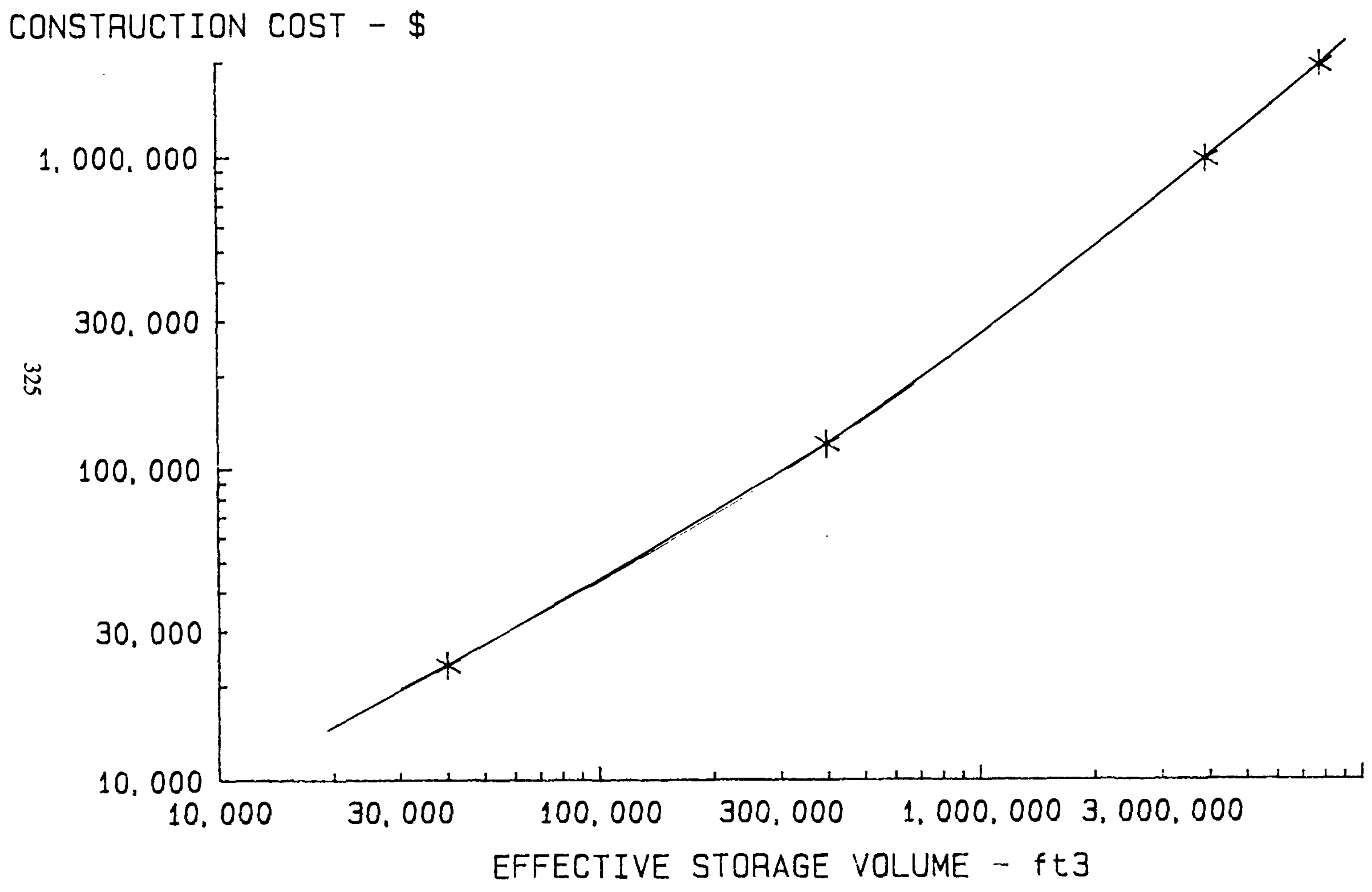
Operating and maintenance costs for the storage lagoon have not been presented. This is because sludge removal is dependent upon the individual design and cake dryness. However, these costs can be very significant and any cost analysis should consider how the lagoons will eventually be cleaned, perhaps subsequently dewatered, and the solids disposed of.

4.11.4. Freeze-Thaw Considerations

Freeze-thaw dewatering of alum sludges is generally a modification of sand bed or dewatering lagoon drying, although freezing can be accomplished by mechanical refrigeration. The freezing process dehydrates the sludge particles by freezing the water that is closely associated with them. As the sludge is chilled, the particles are first concentrated by selective freezing of the water. Next the solids tend to separate from

FIGURE 4-63

CONSTRUCTION COST FOR SLUDGE DEWATERING LAGOONS

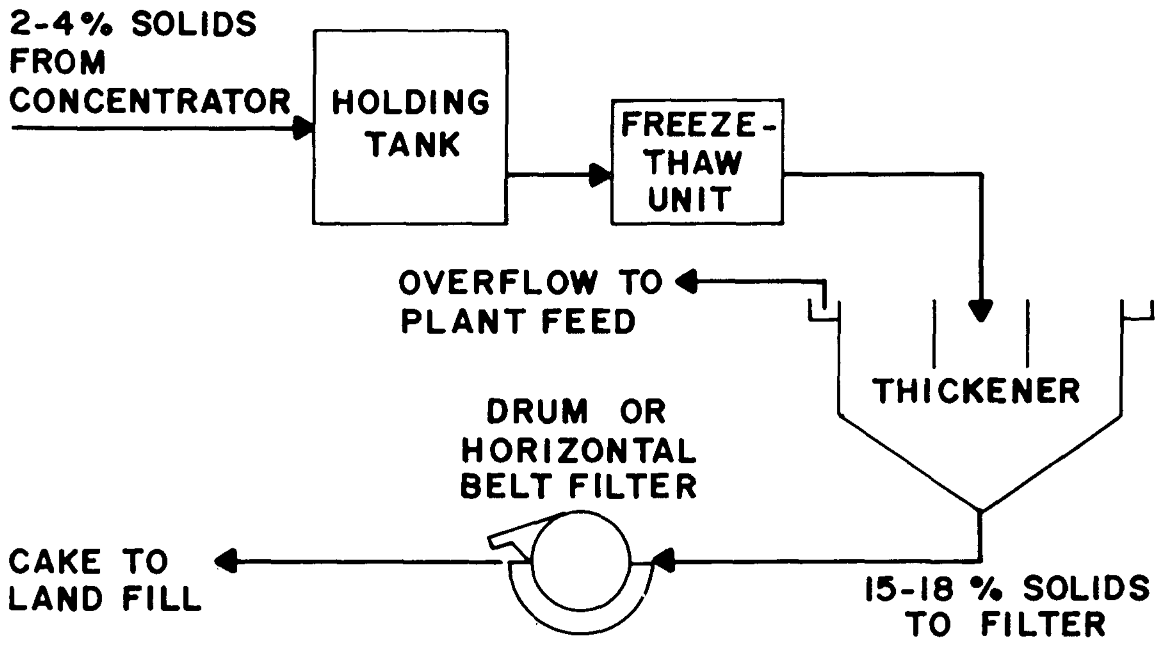


their internal water by the freezing of the particles themselves. The solid mass, when thawed, forms granual-shaped particles. This coarse material readily settles and retains its shape and size. The residue dewaters rapidly and makes suitable landfill material.

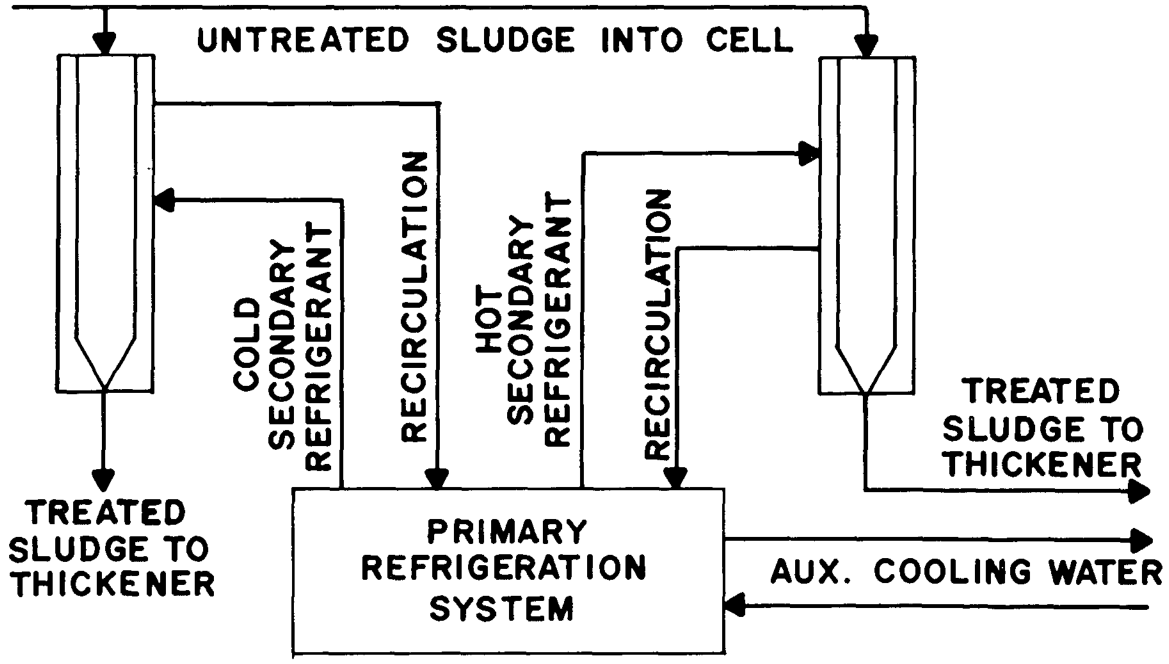
Wilhelm and Silverblatt (4-24) reported comparative information on the dewatering of three coagulant sludges prior to and after freeze treatment. Without freeze treatment the alum sludge was dewatered with a precoat vacuum filter. The freeze-treated particles readily settled to a 17 to 22% solids concentration and were further dewatered by vacuum filtration without a precoat. Some of the results are summarized below:

	VF Before Freeze Treatment		VF After Freeze Treatment		Max. Settled Conc. Per Cent Solids by Wt.	
Sludge	Loading Rate lb/h/sf	% Cake Solids	Loading Rate lb/h/sf	% Cake Solids	Before Treatment 1-4 days Settling	After Treatment 1-5 days Settling
1	0.5-0.7	20	30-60	34	2-4	20
2	0.2	18	50-120	25	1.5-3	19
3	0.5-0.8	21	50-150	33	2.5-5	22

Figure 4-64 is a schematic of a mechanical freeze-thaw system. The process must be designed to freeze the sludge completely and allow sufficient time to dehydrate the particles. These requirements naturally lend themselves to batch systems where large volumes of sludges are alternately frozen and thawed. It appears that quick freezes on the order of a few minuets or less of freezing time do not allow the solids to concentrate sufficiently between the large, relatively pure ice crystals. However, the main shortcoming of fast-freeze-type processes is that they tend toward incomplete freezes. Because the last water to freeze is the water that is closely associated with



A. FREEZE-TREATMENT PROCESS FLOW DIAGRAM



B. SCHEMATIC OF FREEZE-THAW UNIT

FIGURE 4-64

MECHANICAL FREEZE-THAW PROCESS

Reference: 4-24

the particles, incomplete freezing does not dehydrate the majority of the particles. It has been found that even when only 10 to 20 percent of the particles are untreated, the effects of freeze treatment are almost eliminated.

After freeze treatment, the solids may be concentrated in a relatively small thickener and then further dewatered by vacuum filtration or by natural drainage and evaporation. The slurry after treatment can be sent directly to a lagoon or drying bed. In the refrigeration unit, the sludge flows into a bank of deep rectangular pans and is frozen by a cold secondary refrigerant, such as glycol or a brine solution, which is recirculated around the freezing pans. When the sludge is completely frozen, the cold refrigerant is drained and pumped to a second bank of freezing pans to continue the freezing process. The sludge is thawed by recirculating a hot solution of glycol or brine around the frozen pans. Auxiliary cooling water may also be used in direct contact with the ice to speed the thawing process.

Freezing times are dependent on the temperature of the secondary refrigerant and the thickness of the freezing pans. Because refrigeration systems become much more expensive at lower temperatures, the practical operating temperature range for the secondary refrigerant is from 5F to 25F (-15C to-3.89C). Adequate treatment of the sludge is obtained with 20 to 60 min freezing times. The design range of pan thicknesses is from 1/2 to 2 inc. (1.3 to 5 cm) when the pans are frozen from both sides.

In a natural freeze-thaw system the sludge is collected in a lagoon or on a drying bed. Ideally, the lagoon should be equipped with underdrains. As much water is removed as possible. The sludge is then allowed to freeze in the winter and thaw in the spring. The water released by the freeze-thaw cycle is removed through the underdrains. The addition of an underdrain will increase the dewatering characteristics of the lagoon prior to, and after, freeze-thawing. Pilot scale lagoon systems can be utilized to evaluate this method's effectiveness and establish design parameters. If required, elimination of rain and snow from the lagoon can be accomplished by constructing a roof cover. Freezing must take place prior to a snow cover.

The potential advantages of a freeze-thaw lagoon system are as follows:

1. It is insensitive to variations in sludge quality.
2. No conditioning is required.
3. Minimum operator attention is needed.
4. It is a natural process in cold climates.
5. A solids cake is more acceptable at landfill.
6. Sludge is easily worked with conventional equipment.

Farrell et al. (4-25) conducted studies in an attempt to develop design guidelines for a natural freeze-thaw lagoon system. Small scale experiments showed that the sludge freezes at the same rate as water, that is no unusual inhibition of freezing was experienced. Therefore, freezing rate information collected for water should be directly applicable to sludge freezing. They conducted experiments on the degree of freezing in a Minnesota winter, with and without snow cover and evaluated the change in specific resistance. The results are summarized below:

Sample	Degree Frozen %	Specific Resistance $Sec^2/g \times 10^{-8}$	Suspended Solids %
Control	0	10.6	2.1
No Snow Cover	100	0.3	9.5
Snow Cover (12-21 in)	40-70	5.0	5.3

The specific resistance of the sample without snow cover was reduced by a factor of about 30. However, the samples with a snow cover were only partially frozen and the specific resistance was reduced by only a factor of 2. The depth to which the sludge can be applied depends upon the number of freeze days. A climate such as Chicago may freeze a depth of 45-in. while a Cincinnati climate may only freeze a 1-in. depth of sludge.

Several natural freeze-thaw installations are located in New York State (4-26). At the 36-mgd alum coagulation plant of the Metropolitan Water Board of Oswego County, filter backwash is discharged to lagoons that act as decant basins. Thickened sludge is pumped from the lagoons to special freeze-thaw basins in layers about 18 inches thick. The sludge has never been deeper than 1 foot during freezing because of additional water losses. The 1-foot sludge layer becomes about 3 inches of dried material after freeze-thaw. The treated sludge has been allowed to accumulate in the basins so that ultimate disposal has not been a problem.

At the Akron (New York) Water Treatment Plant (1.5-mgd capacity), the sedimentation basins are cleaned in the spring and fall and the sludge pumped to the thickener where it is removed every three or four weeks to three drying beds. The overall dimensions for the combined beds are approximately 50 ft x 30 ft. The sludge has never been applied more than 1 foot thick, which dries to about 4 inches of solids. Sludge is removed from the drying beds during the summer and fall as it becomes dry. Some sludge that is discharged in the fall is frozen and exhibits very good dewatering and handling characteristics, like a fine sand.

4.12. CHEMICAL RECOVERY

Actual practice of the recovery of chemicals from water plant sludges has centered around the production of lime from lime softening sludges and the recovery of alum or iron from coagulant sludges. The objective of chemical recovery is generally a combination of producing the recovered chemical at a price less than the commercial price (thus representing a chemical cost savings to the plant) and at reducing the quantity of waste product requiring treatment or disposal (thus saving on sludge handling costs). Processes for recovering chemicals from both types of sludge have and are being utilized, however each has found only limited application. With the current available technologies it is probable that less use of lime recovery will be made in the future and more use of coagulant recovery. However, more stringent disposal regulations may significantly increase the use of both lime and coagulant recovery. The application of each is discussed below.

4.12.1. Lime Recovery

Lime recovery is accomplished by the process of recalcination. In the basic process of recalcination, the lime sludge, consisting of primarily $CaCO_3$, is dewatered and burned producing calcium oxide (quicklime).

In 1938, in a discussion of a paper by W.W. Aultman (as reported in ref. 4-27) on reuse of lime in softening, Charles P. Hoover of Columbus, Ohio called attention to problems associated with the likely success of recalcination for many water suppliers. He pointed out the problems for producing an acceptable product for plants with high magnesium or suspended particles. He added that Miami, with the largest softening plant in the Country using well water, had the best chance to accomplish recalcination of lime at a reasonable cost. Early in 1938, Aultman conducted classical studies on the recalcining of lime for the Metropolitan Water District of Southern California.

In 1941, A.P. Black who was employed as a consultant for the Miami Water Plant first suggested to the Miami Water and

Sewer Board that they consider recalcination. A plant was eventually completed in December 1948, which was the first full scale recalcination plant. In this process, the sludge was first thickened in circular gravity thickener with sludge scrapers to about a 20% solids concentration. This slurry was dewatered by centrifugation to a 66% solids concentration. The dewatered cake was fed to a rotary kiln where the cake was heated to 2100 ^{O}F. The stack gasses contained about 25% carbon dioxide which was sent to the recarbonation basin of the softening process for pH reduction. Since that time several plants have employed recalcination, as shown in Table 4-17 (4-28).

As discussed in Chapter 3, every mole of lime added to remove a mole of calcium carbonate hardness, produces 2 moles of $CaCO_3$ sludge:

$$CaO + Ca(HCO_3)_2 = 2CaCO_3 + H_2O$$

In recalcination, this calcium carbonate is then dewatered and heated to produce lime

$$CaCO_3 + heat = CaO + CO_2$$

Therefore, for every original mole of CaO added to the softening process, theoretically 2 moles of CaO are produced in recalcination. In actual practice only about 20% excess lime is produced.

As recalcination has been applied, various alternatives have been utilized as shown in Figure 4-65. One of the problems which had inhibited more wide-spread use of recalcination was that impurities in the sludge either made the recovery of lime inefficient or the resulting product was not of high quality. These contaminants which are not volatilized during calcination will increase with recycle and reuse, causing problems both in the slaking process and in efficient

TABLE 4-17

WATER PLANTS PRACTICING LIME RECOVERY

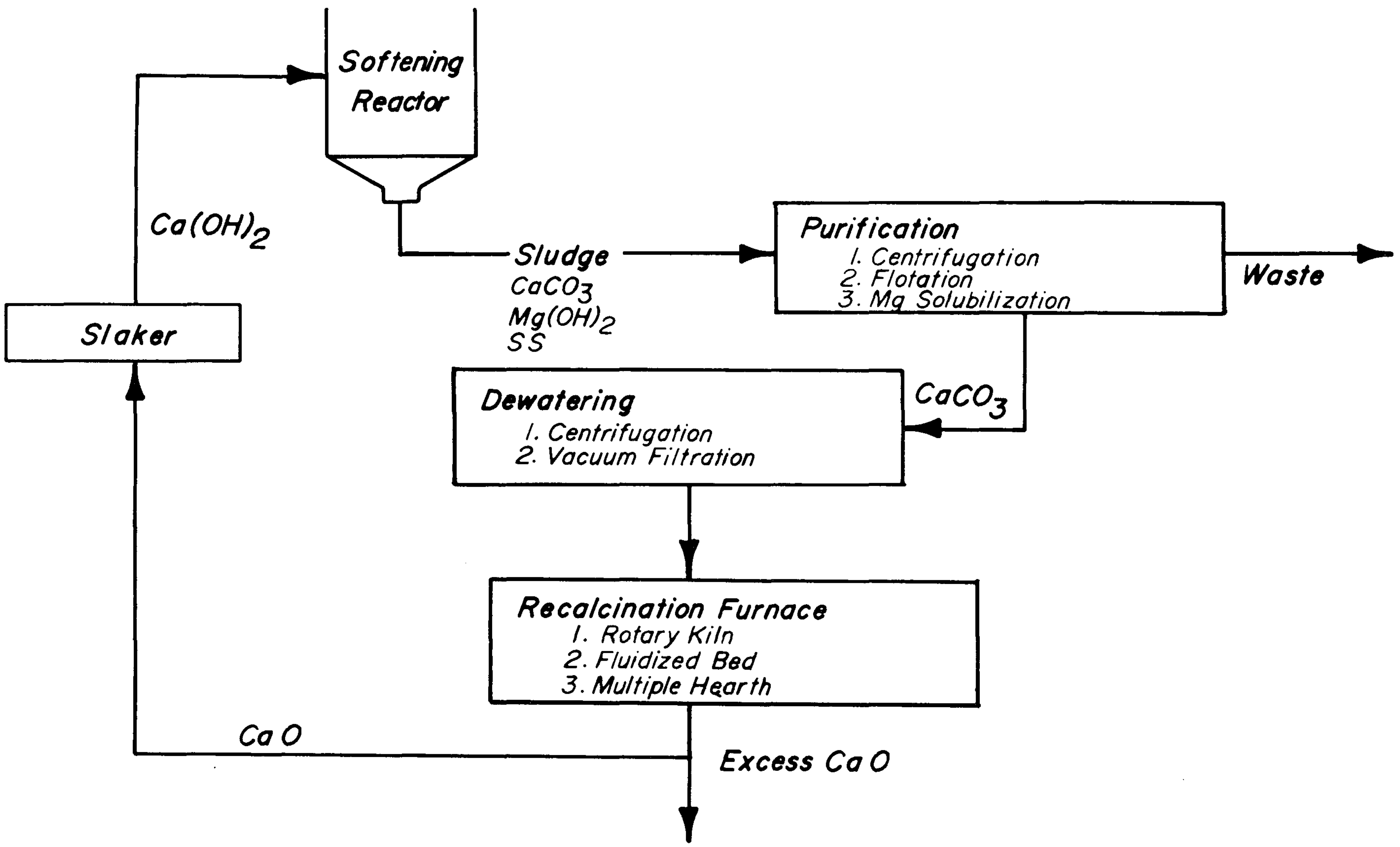
City	Date Installed	Production[a] of CaO (ton/day)	Type of Calciner
Miami, FL	1948	90	Rotary Kiln
Salina, KS	1954	--	Flash Calcination
Lansing, MI	1954	16	Fluidized Reactor
Dayton, OH	1960	92	Rotary Kiln
Ann Arbor, MI	1968	8	Fluidized Reactor
St. Paul, MN	1968	25	Fluidized Reactor
Melbourne, FL	1978	16	Multiple Hearth Furnace

[a] Average actual daily production.

Source: Thompson (4-28)

FIGURE 4-65

OPTIONS OF LIME RECOVERY



calcination. The primary impurity present in groundwaters affecting calcination is magnesium and sometimes silica. Surface waters will also have suspended solids and coagulant hydroxides if the latter are used in the treatment process.

Figure 4-65 shows that the first step for many plants practicing recalcination is a purification process. The most common method of eliminating impurities from the calcium carbonate sludges is one or two stage centrifugation. Centrifugation uses the specific gravity difference between the calcium carbonate and the impurity to make the separation. Primarily this procedure has been used to separate magnesium hydroxide from the calcium carbonate, although at least one study separated silt by this procedure. In both cases the calcium carbonate is heavier and moves to the wall of the centrifuge while the magnesium hydroxide or silt has a lower specific gravity and is lost in the centrate. The primary disadvantage of this method is that some calcium carbonate is also lost in the centrate, depending upon the amount of impurity present and the required degree of classification. Table 4-18 (4-29) shows data for the separation achieved for one such sludge. This particular sludge was fairly low in impurities with an initial $CaCO_3$ purity of about 85%. A 30% reduction in magnesium was achieved, 40% reduction in iron and an 80% reduction in aluminum. The loss of $CaCO_3$ to the centrate was only about 10%, again reflective of the relatively pure original sludge. The purified cake for the example shown is equivalent to about 96% $CaCO_3$. It has been estimated that at least a 91% grade of $CaCO_3$ is needed to be suitable for feed to the recalcination step.

When the magnesium content is higher, a higher degree of separation is needed and more of the $CaCO_3$ is lost. In these cases it may be appropriate to remove the magnesium by selective dissolution. This is accomplished by mixing the CO_2 from the recalcination stack gasses with the sludge. For example, Lansing, Michigan lowers the pH of the sludge to 9.0

TABLE 4-18

EFFECTIVENESS OF CENTRIFUGATION IN REMOVING IMPURITIES FROM $CaCO_3$ SLUDGE

	Feed to Centrifuge	Cake from Centrifuge	% Reduction in Cake
Moisture, %	85.4	33	
Solids Concentration, %	14.6	67	
Magnesium, %	1.9	1.3	30%
Iron as Fe_2O_3, %	0.85	0.49	42%
Aluminum as Al_2O_3, %	8.0	1.8	78%

Source: Sheen and Lammers (4-29)

using the CO_2 and thereby reduces the magnesium oxide content of the cake from 3.5% to 1.8%. Solubilization takes place via the reaction

$$Mg(OH)_2 + 2CO_2 = Mg(HCO_3)_3$$

The separated magnesium carbonate can then be precipitated with lime, dewatered and disposed of. Alternately the liquid could be heated to 35 to 45°C using heat recovery from the recalcination process to produce magnesium trihydrate which is usable as a coagulant:

$$Mg(HCO_3)_2 + 2H_2O + \text{heat} = MgCO_3 \cdot 3H_2O + 2CO_2.$$

Thompson (4-28) has further proposed that the magnesium trihydrate could be burned at 550°C to produce magnesium oxide which may be a saleable product:

$$MgCO_3 \cdot 3H_2O + \text{heat} = MgO + CO_2 + H_2O.$$

Following purification, if needed, and dewatering of the calcium carbonate, the cake is flash dried and burned. Available furnace types include a rotary kiln, flash calciner, fluidized bed and multiple hearth.

The economics of applying the process primarily depend upon the cost of fuel necessary to calcinate the sludge. The fuel consumption is in the range of 8,500,000 to 12,000,000 BTU per ton of CaO produced. No. 2 fuel oil has a heat value of 141,000 BTU/gal, so that 60 to 90 gallons of No. 2 fuel oil are required per ton of CaO produced. As fuel oil prices rise the cost of producing the lime can quickly exceed the cost of purchase.

An additional factor which should be considered in the economics of lime recalcination are disposal costs. Particularly those plants that have radium in the lime sludge may find disposal very difficult and recalcination a viable alternative. However, caution should be exercised as data are not available on the fate of radium in the calcination process.

The removal mechanism for radium in lime softening is unknown. If the radium is in some way associated with the magnesium hydroxide then it may be removed from the calcium carbonate during the purification step. However, if it exists as a precipitate such as $RaCO_3$ (in which case it has a high specific gravity) or is sorbed with the $CaCO_3$ itself then it will remain in the cake during centrifugation and hence be calcined. Based on physical data, the radium could vaporize during calcination. Radium has a boiling point of 2060°F, which is very close to the 2100°F used for calcination. A plant considering recalcination with radium in the sludge, should consider the fate of radium, both to eliminate a build-up and to assure proper disposal of the waste streams.

4.12.2. Coagulant Recovery

Methods for recovery of coagulants from water treatment plants sludges have been investigated since the turn of the century. Most of the studies have focused on alum recovery, although methods have also been investigated to recover iron and to produce a magnesium coagulant (the latter was discussed in the last section).

Past Performance. As outlined by Roberts and Roddy (4-30) the earliest attempt to reclaim alum sludge was made by Jewel, who in 1903 patented a process for water treatment and for reclaiming the coagulant by reacting the aluminum hydroxide with sulfuric acid. Mathis (4-31), in 1923, was issued a patent for basically the same process as developed by Jewel. "Black Laboratories," of Orlando, Florida, in 1951 suggested the use of an alum sludge recovery process utilizing the sulfur dioxide gas from boiler stacks as a source of sulfuric acid. Some of the first reported alum recovery research in the water treatment field was by Palin (4-32). Palin's work was conducted at the Whittle Dene Waterworks of Newcastle, England. In his first set of experiments filter washwater was treated with 0.05% and 0.1% (by volume) sulfuric acid. Chloride was added to oxidize the color present in the dilute recovered alum solution. This recovered alum was then used in conjunction with commercial grade alum in order to determine the amount of commercial grade alum needed to coagulate the raw water and lower the color of the finished water to 10 Hazen. Palin found that in treating raw water 28 ppm commercial alum was needed, while treating raw water plus 3% by volume recovered alum only 11 ppm commercial alum was needed. Despite this large reduction in alum dosage the cost of acid used was higher than the cost of alum saved. Palin reported superior results when the sludge was charred at 400°C before acid treatment. It was found that 1 ton of oven dried sludge would yield 2 tons of aluminum sulfate cake (14% Al_2O_3) upon addition of about 0.9 tons of 98% H_2SO_4.

In Tampa, Roberts and Roddy (4-30) studied the recovery of alum in both pilot and full scale processes. The alum sludge was thickened by settling for 3 hours. The solids content reached 1% in pilot plant scale and as high as 2% on full scale operation. The sludge samples were reacted with enough sulfuric acid to convert the aluminum hydroxide to aluminum sulfate. The amount of acid used varied depending upon the alkalinity of the raw water. The pH range for complete aluminum dissolution was between 1.5 and 2.5 for highly alkaline and less alkaline waters, respectively. After the reclaimed alum was recycled ten times, there was no reported reduction in finished water quality. It was estimated that chemical costs could be reduced by 70% using the acid recovery method. However, the process was never utilized due to subsequent concerns for handling the large volume of dilute alum and the recycle of organics which originate in the highly colored raw water.

Isaac and Vahidi (4-33) in 1961 studied alum recovery for a method of sludge disposal. Isaac tested both alkaline and acid methods of aluminum recovery. He found that aluminum recovery with caustic soda was not very satisfactory. It was also observed that organic matter bound with the aluminum hydroxide, especially organic color, was much more soluble in alkali than in acid. It was therefore decided to use the acidic method for aluminum recovery. Aluminum was recovered from fresh sludge and from anaerobically digested sludge. Tests were then conducted to determine the volume occupied by the sludge after acid treatment. At a pH of 2.5, corresponding to 79% aluminum recovery, a 74% volume reduction of sludge was obtained. Recovered alum was usually about 75% as efficient as fresh alum in reducing color, although one test resulted in an efficiency of 89%. The researchers concluded that the pH should be lowered to about 3.0 for a recovery of about 60% to 65% of the aluminum since the organic color was not dissolved to an excessive extent at this pH.

In laboratory experiments Webster (4-34) found that if sulfuric acid were added to alum sludge to depress the pH value to about 2.4, a clustering effect of the floc particles took place with extremely rapid settling of the insoluble matter. The supernatant liquor contained the alum, representing about 80% recovery. A pilot plant for alum recovery was then constructed. Good coagulation was not obtained with recycled alum that had been recovered at a pH of below 3.0. Webster concluded that the alum reduced the pH of the raw water below the range for acceptable color removal. Therefore, the pH of the sludge was reduced to 3.5 for alum recovery and reuse. No detrimental effects resulted from continued recycling of the alum recovered at a pH of 3.5.

Streicher (in ref. 4-10) conducted pilot tests to determine the usefulness of acid recovery of aluminum followed by filter pressing the remaining sludge. The pH was reduced to 1.5 to 2.5 by sulfuric acid. He found that when the ratio of $Al(OH)_3(s)$ to other suspended matter in the sludge was high, considerably less than stoichiometric amounts of sulfuric acid were required. If the ratio were low, more than stoichiometric amounts of acid were needed. Acid treatment resulted in reduction of the sludge volume to less than 10% of the original volume, and the solids concentration of the settled sludge reached 20%. The alum recovery was 80% to 93%. With the use of a filter press the remaining sludge was concentrated to 40% to 50% solids concentration.

Fulton (4-35, 4-36) described an alum recovery system scheduled to be put into operation in 1974 at Jersey City, New Jersey. The process consisted of thickening, acid addition and filter pressing of the resulting sludge. The acid recovery could be bypassed and only the filter press used if necessary. For a 100 mgd plant, the savings were estimated at $4.60 per million gallons when alum recovery was used. An alum recovery of 90% was estimated. The process has not been utilized.

Westerhoff (4-37) in 1973 conducted a 15-week pilot plant study to determine the effect of recycling alum recovered from waste alum sludge by an acidic process. The pH of the sludge was reduced to 2.0 for conversion of aluminum hydroxide to aluminum sulfate. The main purpose of the study was to evaluate potential contaminant build-up in the recycled alum. Measurements were made on total microscopic count, coliform, hardness, alkalinity, cyanide, fluoride, phenol, dissolved solids, nitrates, sulfates, chlorides and several metals such as copper, lead and zinc. Throughout the study final water analysis for the pilot plant using recycled alum and for the full scale plant using fresh alum were essentially the same, indicating that impurities were not built-up by recycling alum.

Westerhoff and Daly (4-7) conducted a complete study of various alum sludge dewatering facilities. They tested pressure filtration with and without alum recovery, centrifugation, rotary vacuum filtration, horizontal vacuum filtration with and without alum recovery, coagulation, filter press and freezethaw. The studies showed alum recovery followed by horizontal vacuum filtration to be workable process warranting economic evaluation. The recovery of alum varied from 50% to 90%. Coagulation basin sludge was thickened from an initial 4% to 6% solids to a final 21% solids by acid treatment. After filtration the solids content was 37%. However, because of the low alum dosage used for raw water turbidity removal, the most economical method of alum sludge treatment was determined to be pressure filtration without alum recovery.

Chen (4-38) conducted a series of laboratory studies to characterize water plant sludges after acid addition for alum recovery and to evaluate the effectiveness of the recovered alum. Figure 4-66 are the results from evaluating several acids and bases on the chemical demand to achieve a given aluminum dissolution and Figure 4-67 shows the use of acid to dissolve the aluminum from several different sludges. As shown, acids dissolve the aluminum in an amount very close to that predicted by stoichiometry while bases show relatively poor dissolution of the aluminum. Since sulfuric acid and hydrochloric acid both act stoichiometrically and hydrochloric acid is more expensive, sulfuric acid has been the preferred chemical to accomplish dissolution. In Chen's work the acid dose was based on stoichiometry rather than reaching a particular pH. Chen conducted studies on the settleability of the residual sludge which remains after the acid addition for aluminum dissolution. He conducted standard 1-liter graduated cylinder tests as discussed in Chapter 3 and determined the unhindered settling velocity of the solids. Figure 4-68 shows the settling velocities obtained for several different sludges as a function of the amount of aluminum dissolution that occurred. Although the data are insufficient for design of a continuous flow thickener, they do show the very rapid separation that takes place between the solids and the dissolved aluminum solution.

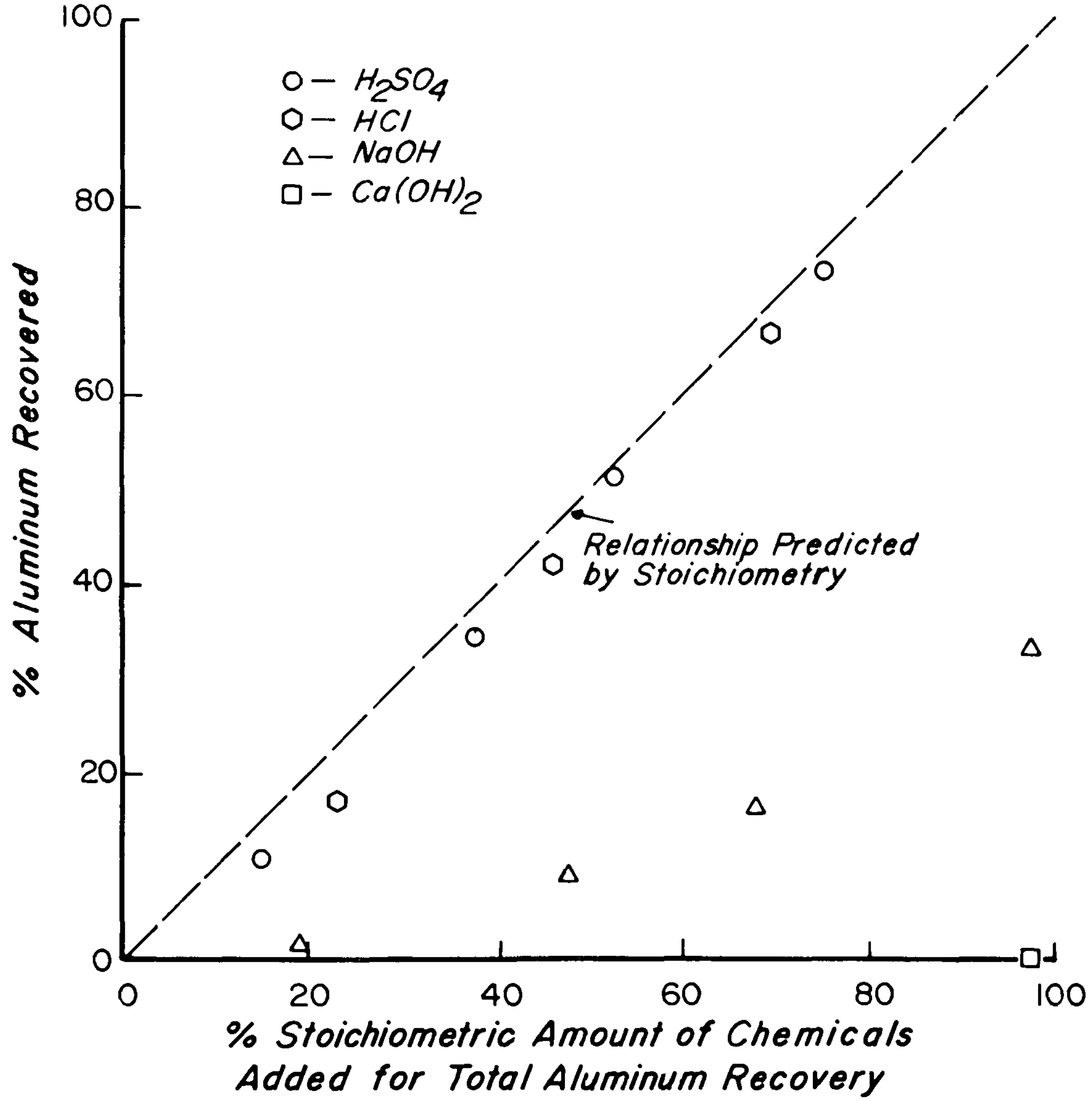


FIGURE 4-66

PERCENT ALUMINUM RECOVERED VS CHEMICALS ADDED

Reference: 4-38

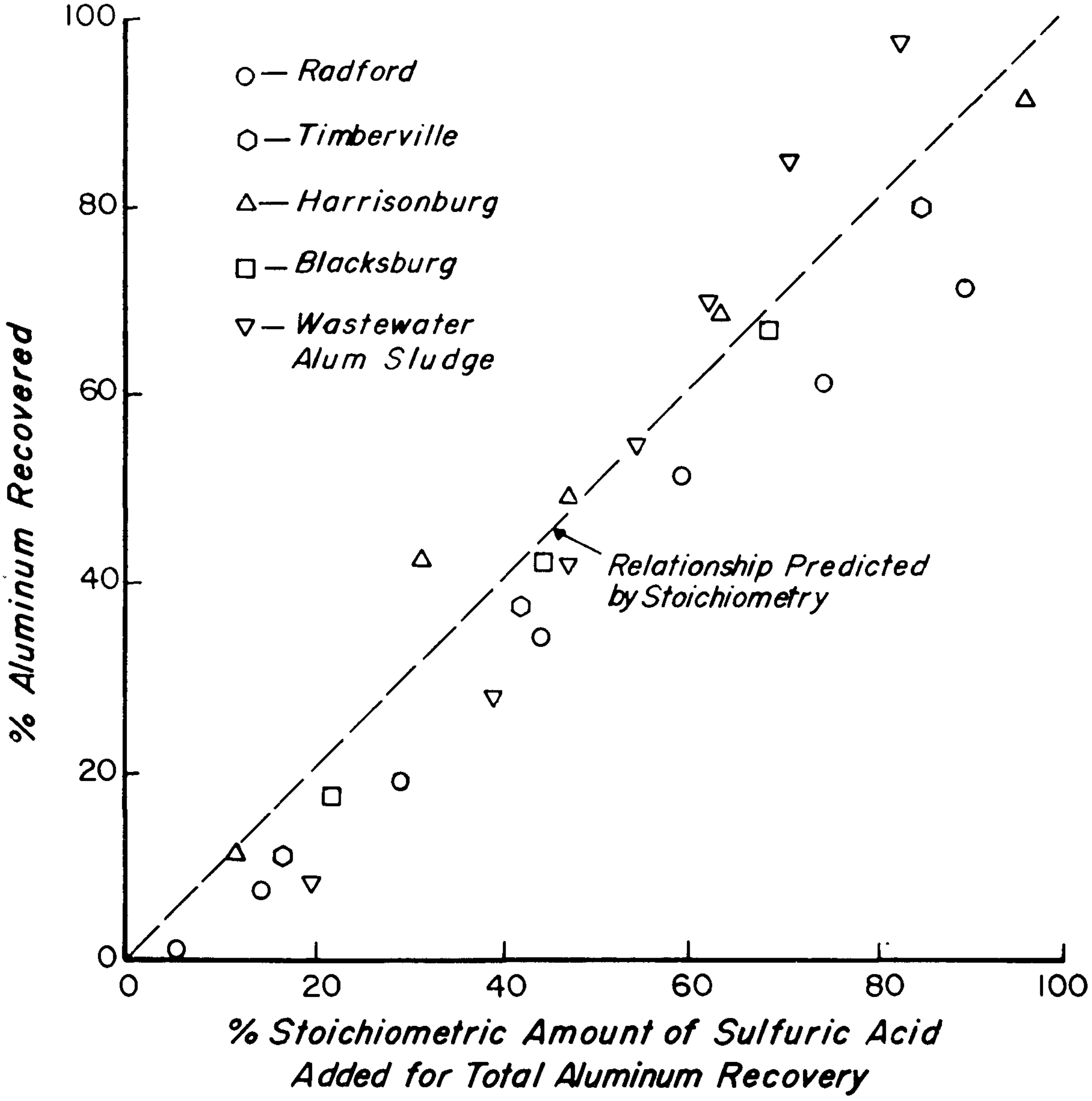


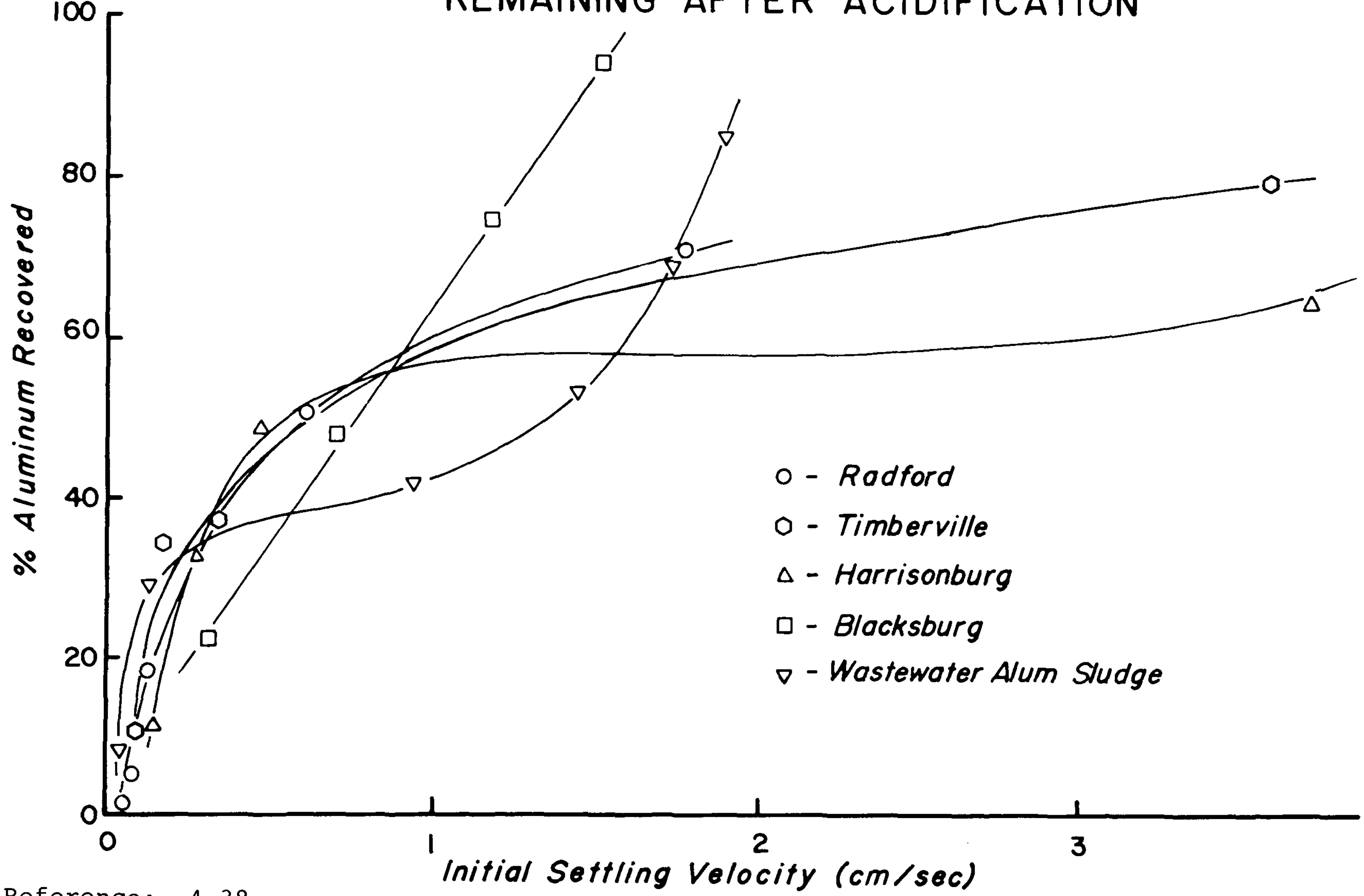
FIGURE 4-67

PERCENT ALUMINUM RECOVERED VS. SULFURIC ACID ADDED

Reference: 4-38

FIGURE 4-68

INITIAL SETTLING VELOCITY OF SLUDGE REMAINING AFTER ACIDIFICATION



Reference: 4-38

Figure 4-69 shows the resulting specific resistance (test described in Chapter 3) for these same sludges again as a function of aluminum dissolution. For 3 of the sludges a significant reduction in specific resistance was obtained, with the minimum values occurring near 60 to 80% aluminum dissolution.

Cornwell and Susan (4-39) reported similar laboratory experiments on sulfuric acid dissolution of aluminum and the characteristics of the remaining solids. They conducted studies on five different sludges; characteristics of the sludges shown in Table 4-19. The sludge obtained from Washington was high in non-dissolvable solids associated with the raw water suspended solids content, with a relatively low aluminum hydroxide content as shown by the dissolvable inorganic solids concentration. The sludges from Indianapolis and Concord are both from raw waters with low color and turbidities in the 40 TU range. Concord uses a higher ratio of alum to turbidity as reflected by the percentages shown. Tampa ia a highly colored low turbidity raw water and Moline uses alum and a high dose of lime in its treatment process. Figure 4-70 shows aluminum dissolution as a function of pH. Except for two of the sludges maximum dissolution was obtained at about pH 2. Figure 4-71 shows the acid demands to achieve the aluminum dissolution. As with the work by Chen, 1.5 moles of sulfuric acid are required per mole of aluminum dissolved. The percentages of aluminum content of the sludges was determined by a total digestion of the sludge and therefore reflects aluminum associated with the precipitated aluminum hydroxide as well as aluminum complexed with the naturally occurring clay particles. The aluminum associated with the clay is generally not dissolved in the pH range used for alum recovery.

Tests were also conducted on the kinetics of aluminum dissolution for two of the sludges as shown in Figure 4-72. Equilibrium was reached after about 15 minutes of mixing for the sludges shown.

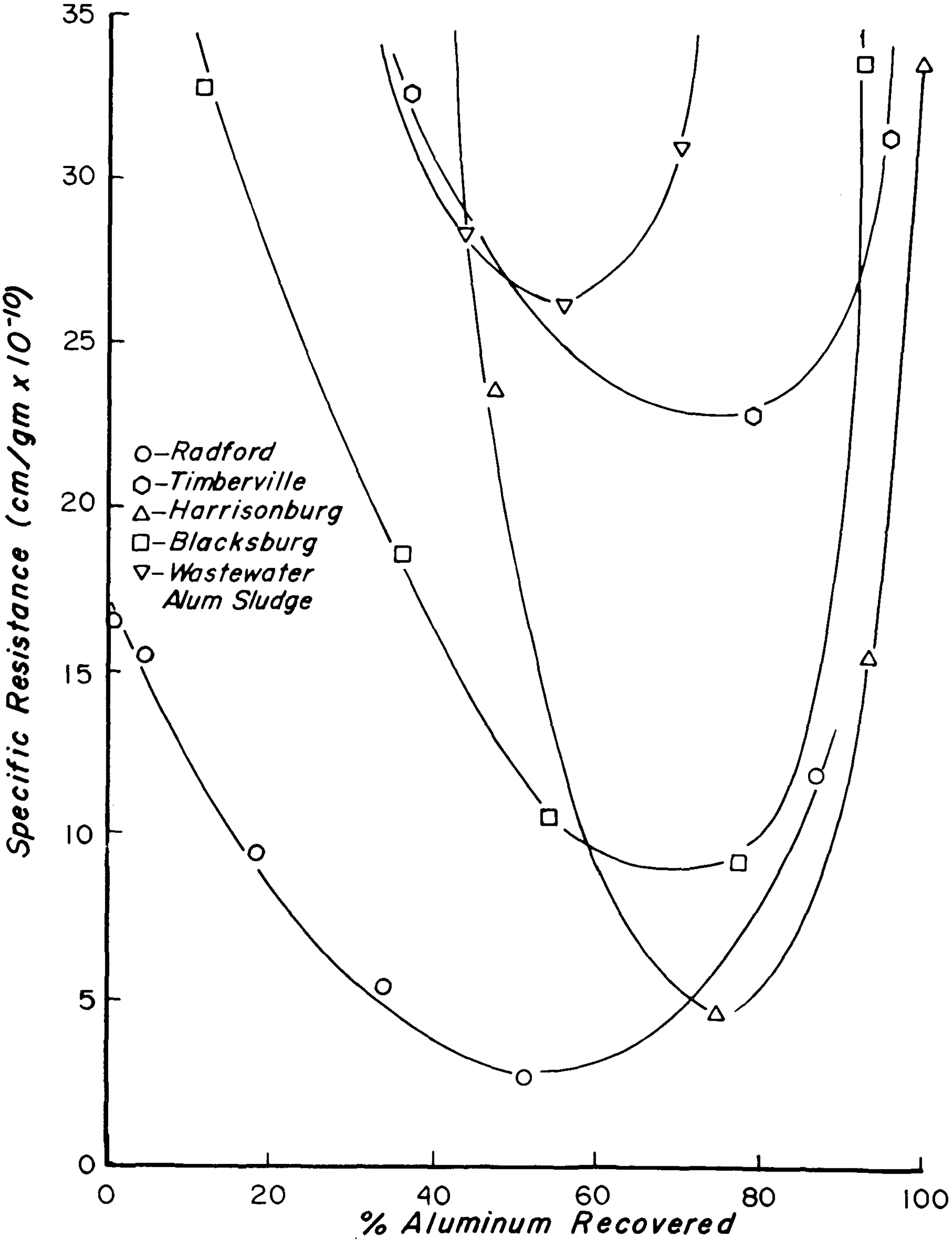


FIGURE 4-69

VARIATION OF SPECIFIC RESISTANCE OF SLUDGE REMAINING VS. PERCENT ALUMINUM RECOVERED

Reference: 4-38

TABLE 4-19

SLUDGE CHARACTERISTICS APPLICABLE TO ALUM RECOVERY

Parameter	Washington	Indianapolis	Concord	Tampa	Moline
Dissolvable Inorganic Solids, %	9	26	36	61	7
Non-Dissolvable Inorganic Solids, %	49	52	18	6	79
Dissolvable Organic Solids, %	26	12	18	25	2
Non-Dissolvable Organic Solids, %	16	9	28	8	13
Solids Concentration, %	12.1	1.7	1.7	1.6	1.7

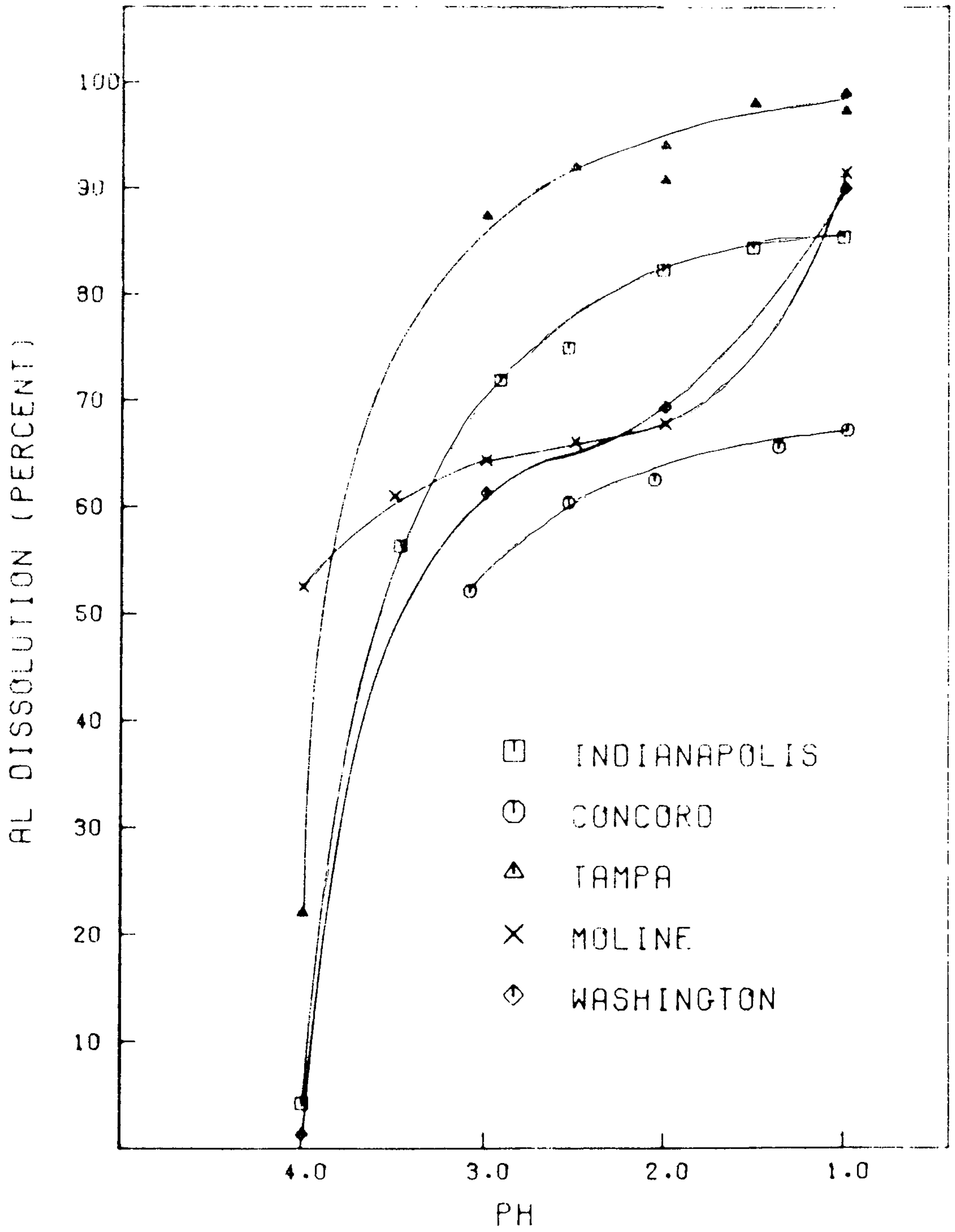


FIGURE 4-70

ALUMINUM DISSOLUTION
AS A FUNCTION OF pH

Reference: 4-39

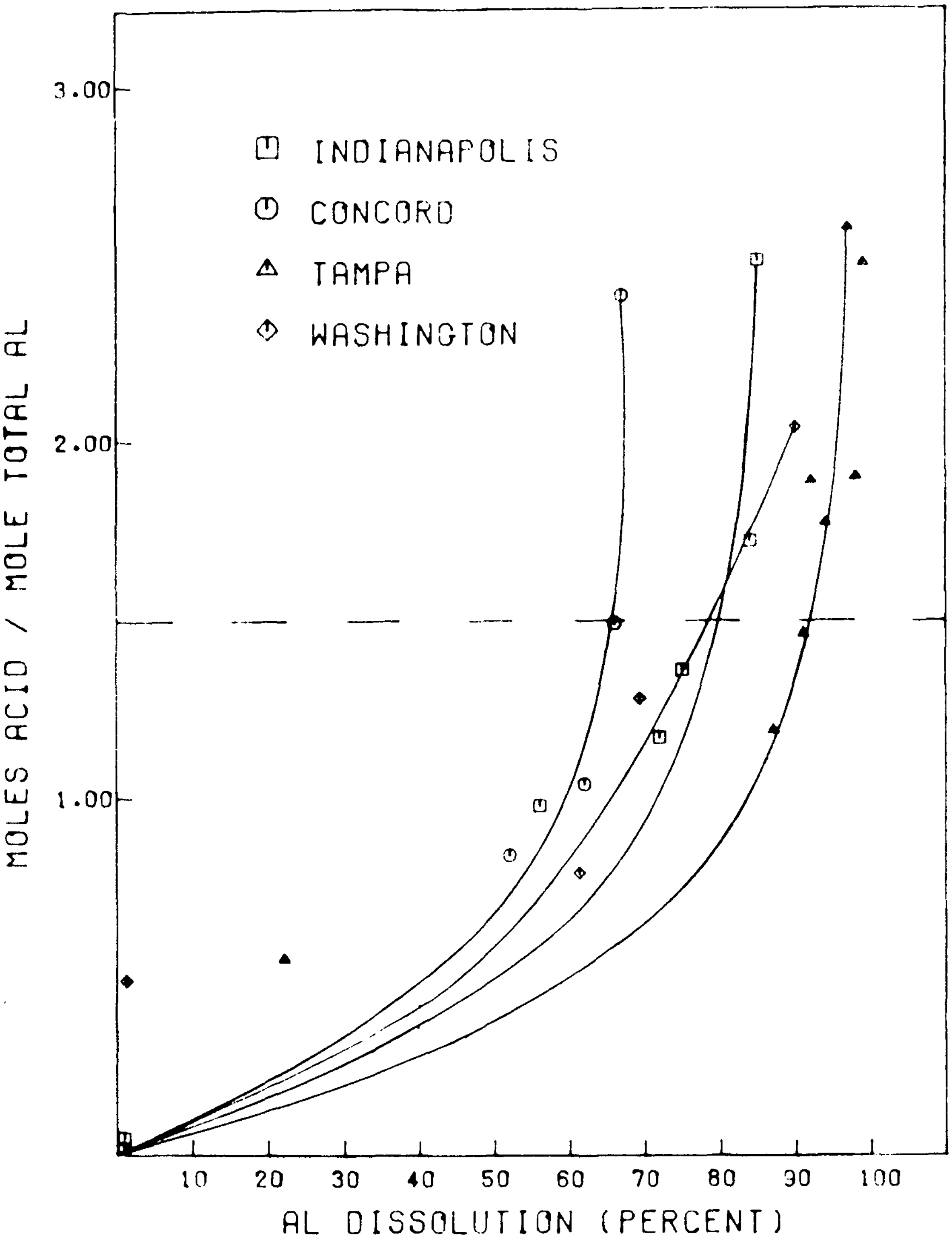


FIGURE 4-71

STOICHIOMETRIC ACID DEMAND
FOR ALUMINUM DISSOLUTION

Reference: 4-39

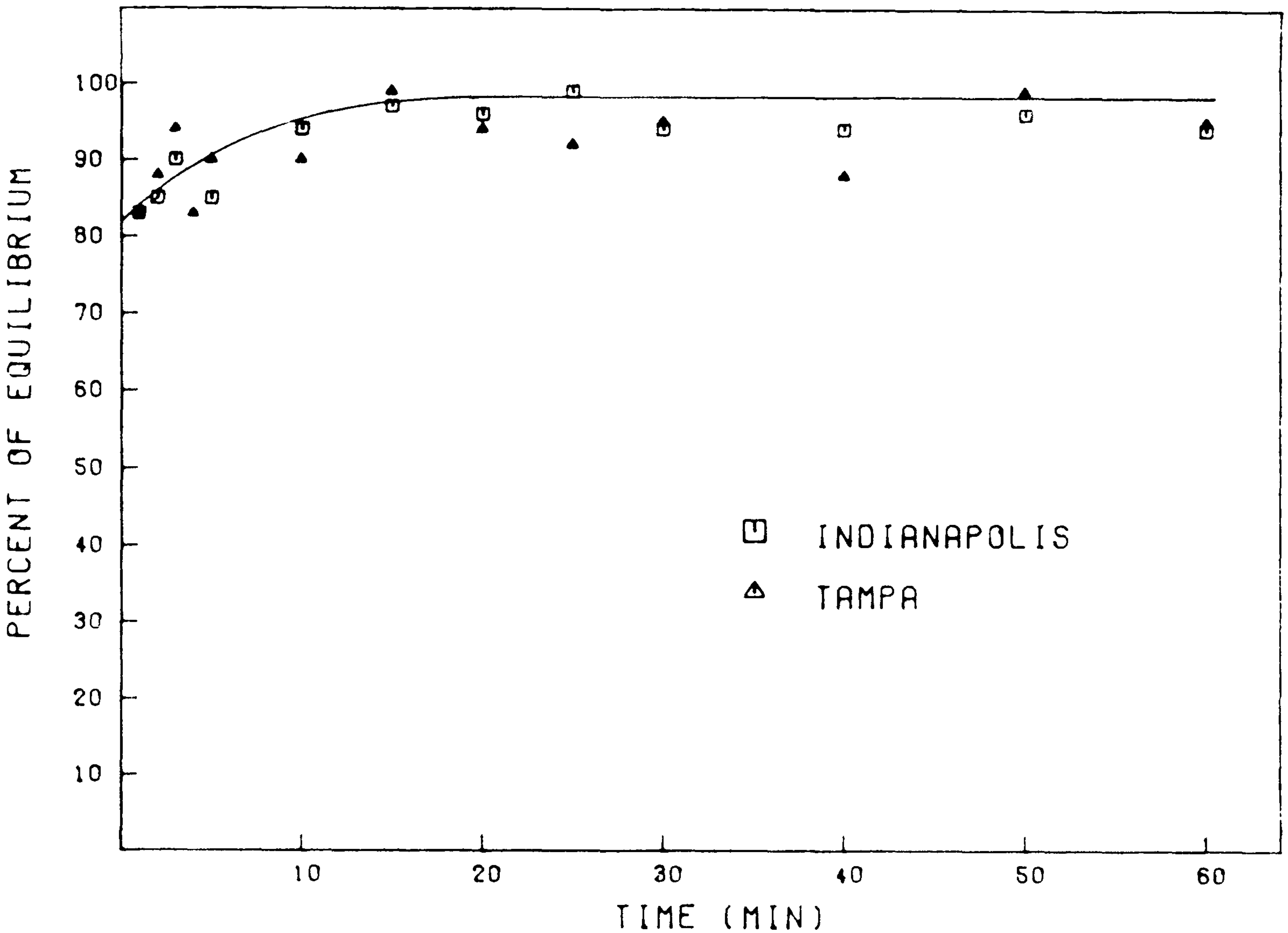


FIGURE 4-72

RATE OF ALUMINUM DISSOLUTION

Reference: 4-39

Figure 4-73 shows the dry weight sludge reduction obtained following aluminum dissolution. This of course is reflective of the reduction in weight of sludge which would require treatment and disposal after recovery.

Following the above testing, Susan (4-40) proposed a full scale flow diagram as shown in Figure 4-74, with a mass balance shown for Tampa. The raw sludge would be thickened and mixed with sulfuric acid. A thickener/sedimentation tank would be used to separate the remaining solids from the recovered alum. The residual solids could be dewatered by several options, with sand drying beds the option proposed by Susan.

Lindsey and Tongkasame (4-41) developed a method to purify the reclaimed alum which results from acidifying alum sludge by using ultrafiltration (UF). UF is a pressure driven membrane process. The theory of operation is that by applying 20 to 60 psi to the reclaimed alum that the water molecules and small aluminum molecules would pass through the membrane while larger color molecules would not pass through. In this way color would be removed from the recovered alum solution. Little results were presented on the exact aluminum passage and color (TOC) rejection, however the implication was that the results were favorable. Concern did exist, however, as to whether the flux rate could be maintained at an acceptable level to be economical.

Cornwell (4-42) conducted a series of bench and pilot scale tests on utilizing liquid-ion exchange to recover, purify and concentrate the alum. The theory was to preferentially remove the aluminum from the sludge, thereby leaving any impurities with the solids. The objective was to also concentrate the recovered alum to a concentration near that of commercial liquid alum. Two types of equipment and flow diagrams were evaluated. The basic process as pilot tested at Tampa, Florida is shown in Figure 4-75. The first step in the recovery process is "extraction". Extraction is the operation

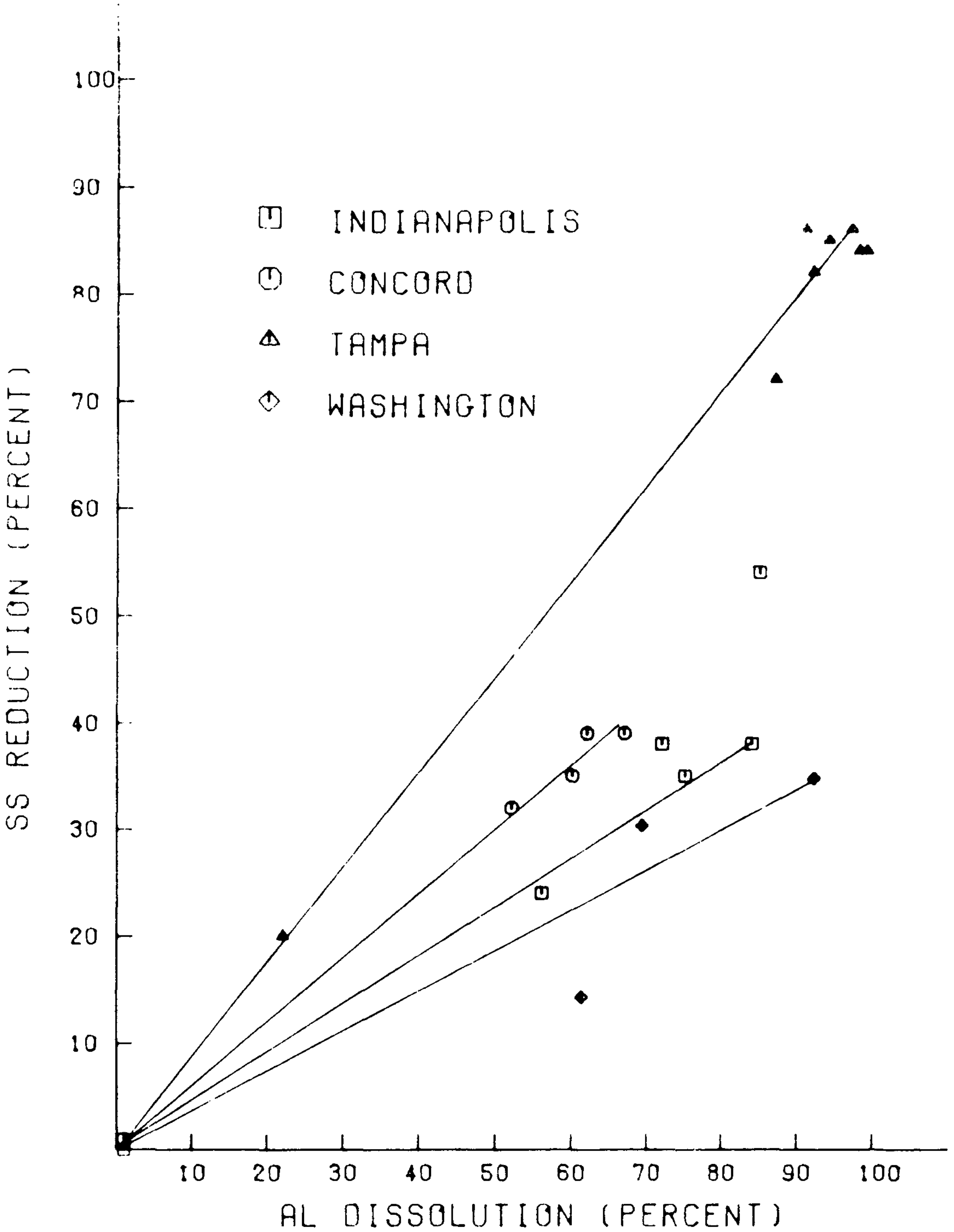


FIGURE 4-73

SOLIDS REDUCTION VERSUS ALUMINUM DISSOLUTION

Reference: 4-39

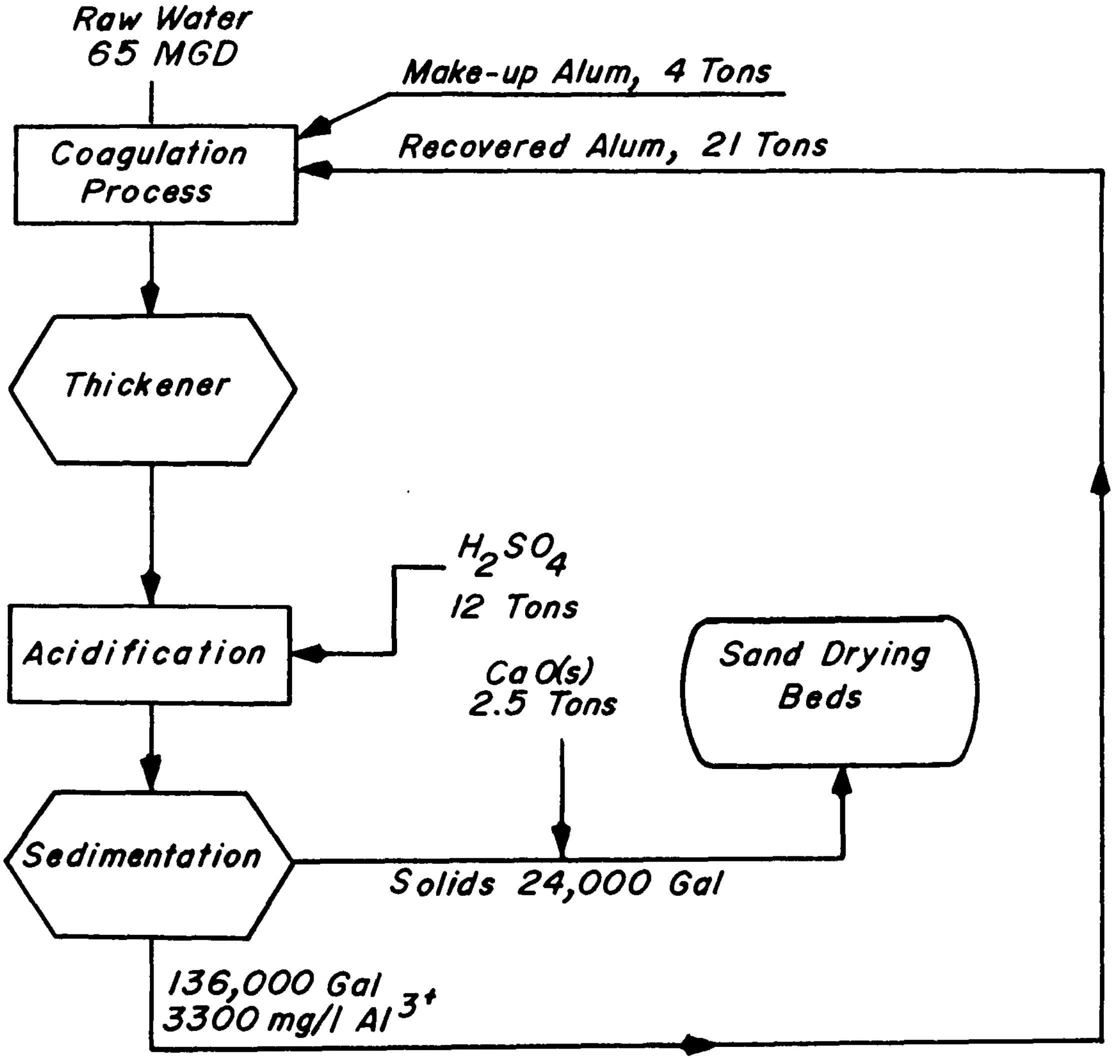


FIGURE 4-74

ALUM RECOVERY FLOW DIAGRAM
QUANTIFICATION FOR TAMPA, FLORIDA

Reference: 4-40

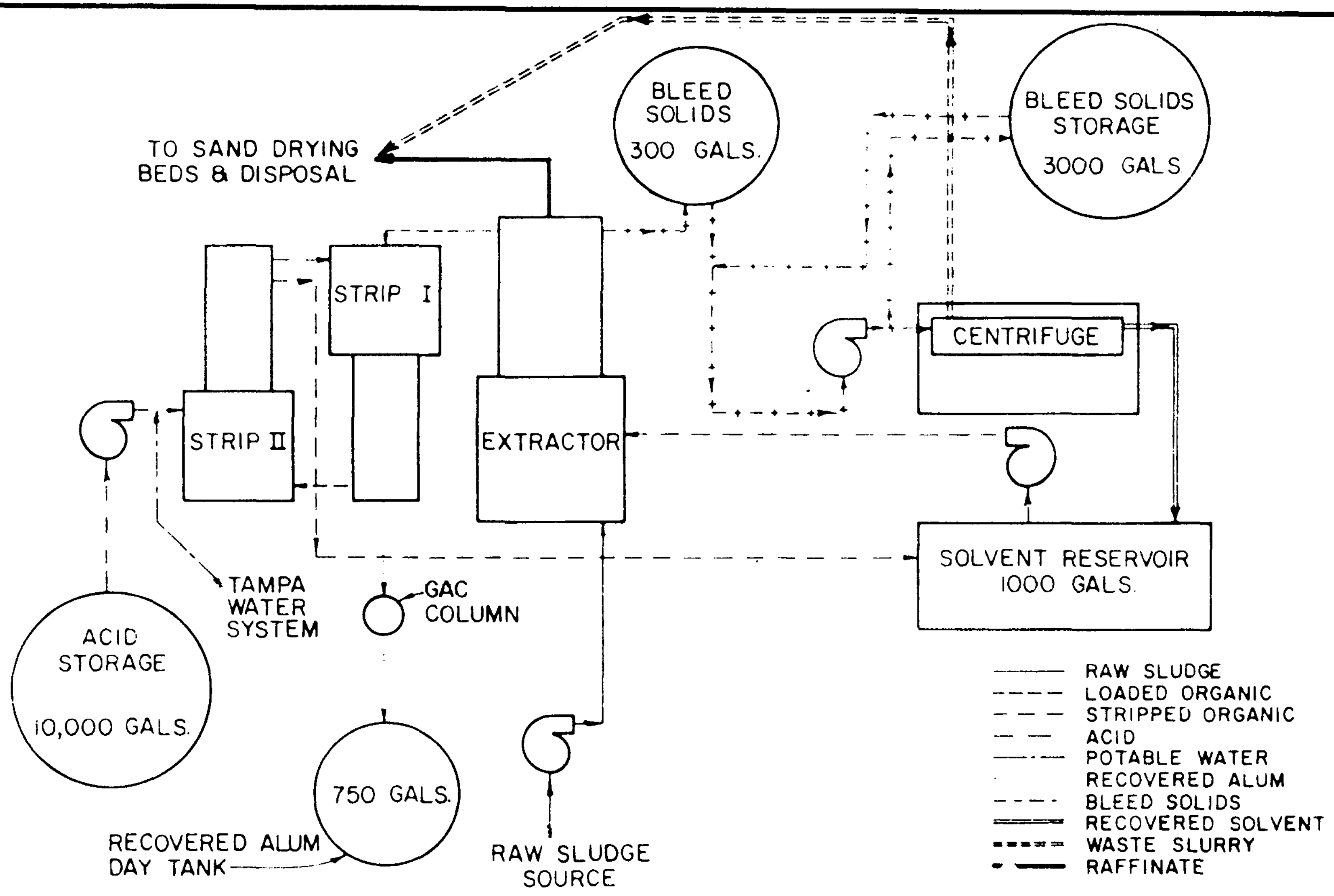


FIGURE 4-75

FLOW DIAGRAM OF LIQUID-ION EXCHANGE ALUM RECOVERY PROCESS

Reference: 4-42

in which the sludge is contacted with a solvent to selectively extract the aluminum from the sludge. In this equipment setup a rapid mixer was used to contact the sludge and solvent, followed by a shallow settling tank for separation of the aluminum rich solvent from the remaining solids and water. The solvent consists of approximately 15% v/v (volume to volume) octyl acid phosphate, 2% v/v tributyl phosphate and 83% v/v inert solvent. During phase separation in the settler 3 phases form. The top layer is the aluminum rich solvent called extract, the middle layer is thickened residual solids called bleed solids and the bottom layer is water, called raffinate. The extract and raffinate were removed from the settler via a dual weir system. The bleed solids were siphoned from the interface and stored. These solids consists of solvent, water and the residual solids remaining after aluminum extraction. The bleed solids were treated by a centrifuge to remove and recover the solvent (which is aluminum rich extract). The water and residual solids flow from the centrifuge to sand drying beds. The extract was stored in the solvent reservoir. During stripping sulfuric acid is added to remove the aluminum from the solvent, thereby producing recovered aluminum sulfate (alum) and regenerating the solvent for use in the extraction stage. The recovered alum received a final polishing step by passing through a GAC column. Results of approximately 500 hours of pilot plant testing (at an average sludge feed rate of 4 gpm) showed an aluminum recovery of 91% with a standard deviation of 5%. The recovered alum was essentially of equal or better quality then commercial liquid alum. Difficulties associated with operation of the process primarily centered on the centrifugation step for recovery of the solvent from the bleed solids. Modifications to the centrifuge at the pilot facility appeared to alleviate these problems, but testing was too short for definitive conclusions to be drawn. The process would primarily be applicable only if the acidification process did not produce an acceptable quality of recovered alum.

Pigeion, et al. (4-43) evaluated iron recovery from sludges produced when iron was the coagulant. They investigated a combination of acid and sulfide to solubilize the iron. The purpose of the sulfide was to reduce ferric iron to the more soluble ferrous form:

$$2Fe(OH)_3 + 3Na_2S = 2FeS + S + 6NaOH.$$

The results did show that sodium bisulfate addition would enhance the recovery of iron and that iron recovery could be achieved at a higher pH than using acid alone. However caution would need to be utilized for the control of H_2S and the effectiveness of the recovered ferrous iron as a coagulant.

The first reported full scale installation of alum recovery in the water treatment field was at the Higashimurayama plant in Tokyo, with start-up in 1964. Their objective was to minimize the amount of final sludge cake requiring disposal. In their case a 30 to 40% reduction in sludge dry weight was achieved using alum recovery via sulfuric acid dissolution. The process consisted of thickening, acid addition in a two-stage rapid mix tank, separation of the recovered alum from the residual solids in another thickener tank, lime addition to the residual solids followed by vacuum filters and in subsequent plants filter pressing.

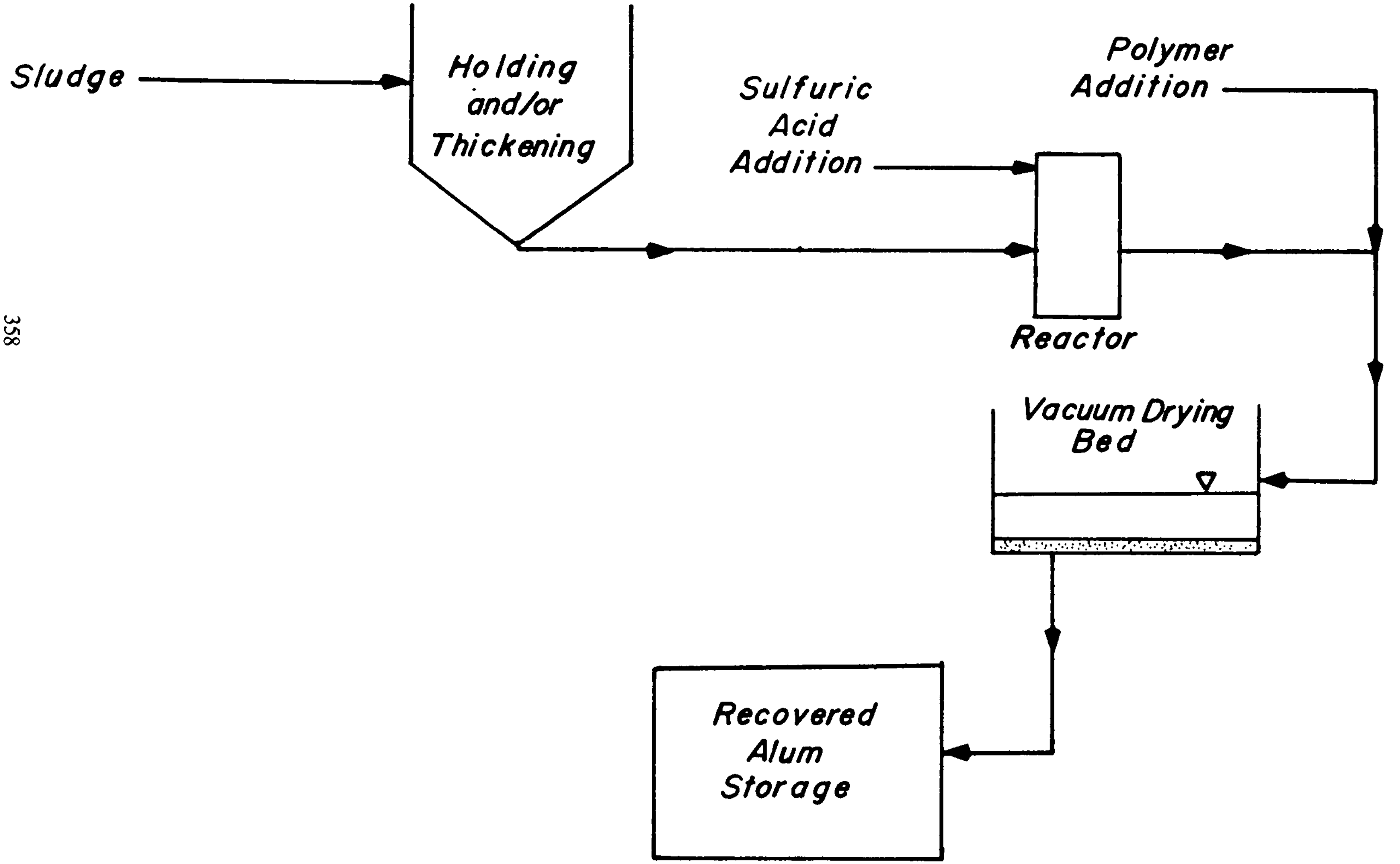
Full-Scale U.S. Installations. There are three operating plants utilizing coagulant recovery. Two recover alum and one recovers iron. All of these plants use the process developed and patented by A.R. White of A.R. White and Co. and follow a flow diagram similar to that shown in Figure 4-76.

The Richmond County, North Carolina Water Treatment Plant began practicing alum recovery in the fall of 1985. This is a 2 mgd plant with a raw water turbidity of 15 and a true color of 40. The alum dose using a commercial alum is 16 mg/l. Backwash water and sludge from the sedimentation basin are

FIGURE 4-76

COAGULANT RECOVERY SOLIDS DEWATERING SYSTEM

A.R. WHITE AND CO.



collected in a 40-ft diameter thickener. Solids are collected in the thickener in order to produce enough solids for the recovery process. This storage volume also helps meet peak sludge production and allows additional operator flexibility.

The thickened solids concentration is 3 to 4%. These solids are pumped to a mix tank where sulfuric acid is added to reduce the pH to about 2.5. Polymer is added to the reduced pH sludge at a dose of about 3 lb/ton. Vacuum drying beds are used to dewater the solids and to separate the reclaimed alum. The recovered alum is stored in a 8,000 gallon storage tank where it is pumped to the rapid mix tank for use as a coagulant. The percentage of alum recovery achieved for one detailed five days test was 94%. This percentage recovery includes not only the actual amount of aluminum recovered, but also considers that the recovered alum solution is more effective a coagulant then the commercial alum. Therefore the 94% represents the amount of commercial alum that can be replaced with the amount of alum produced. Significant testing at this site has shown the recovered alum to be about 20% more effective as a coagulant then commercial alum.

The Montgomery County Water Authority in North Carolina has a similarly operating alum recovery process. This is a 5.5 mgd plant using approximately 15 mg/l of commercial alum. This plant has found operating results similar to that described above for the Richmond County plant.

The Athens Utilities Board in Athens, Tennessee operates the only iron recovery plant. This is approximately a 6 mgd plant with a raw water turbidity of 18 TU and using a ferri-floc dose of 30 mg/l. Backwash water is recycled to the head of the plant so that all the solids are collected in the sedimentation basins. Sedimentation basin sludge at a solids concentration of 2 to 7% is pumped to a mixed reaction tank where sulfuric acid is added to reduce the pH to about 1.6. Polymer is added at a dose of 8 pounds per ton of dry solids prior to dewatering on a vacuum drying bed. The acidification process results in a 50 to 60% reduction in dry weight solids requiring dewatering and handling for disposal. Approximately a 20% make-up volume of commercial ferrifloc is needed when the process is at steady state, for a net effective recovery of 80%. Table 4-20 shows some of the key annual average data. Also shown in Table 4-20 are the operating cost for chemicals, dewatering and sludge haul for the iron recovery process as compared to data prior to use of the process. A comparison of these key cost items shows about a 50% reduction in annual cost. Approximately one-third of the savings is attributable to recovery of the iron itself. The remaining cost savings is due to the 50 to 60% reduction in solids that require dewatering and haul and the improved dewaterability of the acidified solids.

Implementation Considerations. The largest full scale alum recovery process applied to date is currently under design in Durham, North Carolina. The process will treat a plant flow of 22 mgd and a solids production of 800 tons per year. This recovery facility is scheduled for operation by the summer of 1988. The design was based on extensive pilot and full scale testing. Results of two full scale tests are briefly discussed below, both for the purpose of presenting the results as well as to describe the type of data that a utility should collect when considering implementation of an alum recovery system (4-44).

Three technical or performance factors are important in the implementation of an alum recovery and reuse system:

1. Sludge Characteristics and Reduction
2. Dewaterability of the Remaining Sludge
3. Coagulant Recovery and Reuse

1. Sludge Characteristics and Reduction. Acid demand should be determined as required to achieve a given degree of

TABLE 4-20

IRON COAGULANT RECOVERY RESULTS

ATHENS, TENNESSEE

ANNUAL AVERAGE

	Using Iron Recovery	Without Iron Recovery
A. OPERATING RESULTS		
Raw Water Turbidity	18 TU	18 TU
Average Flow, Q	1.8 mgd	1.8 mgd
Commercial Ferrifloc Required	7 mg/l	29 mg/l
$Ca(OH)_2$ Dose	23 mg/l	28 mg/l
Cl_2 Dose	2.4 mg/l	1.2 mg/l
H_2SO_4 Dose (mg/l of Q)	12 mg/l	0 mg/l
Solids Requiring Treatment/Haul	31 Tpy	85 Tpy
Dewatering Polymer Dose (equivalent to produced solids)	8 lb/ton	14 lb/ton
% Dry Solids From Vacuum Bed	25	17
Volume of Solids for Haul	121 cy/yr	496 cy/yr
B. ANNUALIZED COSTS		
Ferrifloc	$ 2,166/yr	$ 8,977/yr
Lime	4,488	5,350
Chlorine	1,780	890
H_2SO_4	1,405	0
Polymer	546	1,190
Labor to Fill/Clean Sludge Bed	8,840	18,720
Haul Truck	485	1,984
Power (vacuum bed)	1,000	1,500
TOTAL ANNUAL OPERATING COST	$20,710	$38,611

coagulant recovery, a given degree of solids reduction and the effects of acid dose on dewaterability of the remaining solids. Generally, coagulant recovery and solids reduction correspond directly to each other and as one is maximized the other is also. The exception is for hard waters where a lower pH may begin to precipitate $CaSO_4$, thereby increasing the solids production. Dewaterability as shown earlier in the work by Chen may not be minimal at the maximum point of alum recovery, and this should be evaluated and tradeoffs considered. Acid demands can be determined in the lab by titrating the sample with 1+1 H_2SO_4. However, it should be noted that the aluminum dissolution is time dependent, and therefore about 15 minutes of mixing should be allowed between each acid reading. Aliquotes would then be taken after the 15 minute time period and analyzed for suspended solids, dissolved aluminum and one or more of the dewaterability parameters (Chapter 3) if desired.

Utilizing the equation of Chapter 3 for the sludge production it is possible to estimate the solids reduction to be achieved following alum recovery. Recall that equation as

$$S = 8.34Q\ (0.44AL + bTU + A)$$

or percent reduction is

$$\text{Percent Reduction} = \frac{100(0.44)AL}{0.44\ AL + bTU + A}$$

For the City of Durham, the following annual averages apply:

$AL = 25$ mg/l
$TU = 25$ mg/l
$b = 0.74$
$A = 0$

and therefore,

$$\text{Percent Reduction} = \frac{100(0.44)25}{0.44(25) + .74(25)}$$

$$= 37\%$$

or a 37% reduction in dry weight solids would be expected. In practice this reduction may be slightly higher due to dissolution of organic compounds that precipitated or sorbed onto the alum floc, or may be lower if incomplete dissolution of aluminum hydroxide occurs. Table 4-21 shows the results from two full scale tests at Durham. The first sample was lowered to pH 2.1 with 550 gallon of 93% sulfuric acid and the second sample to pH 2.0 with 825 gallons of 93% sulfuric acid. The acid demand in each case was 0.67 tons of acid per ton of alum dissolved. This corresponds to 2.0 moles of H_2SO_4 per mole of aluminum dissolved and compares to the stoichiometric amount of 1.5 to 1. About one-half of the excess demand was accounted for by the dissolution of iron and the resulting production of ferric sulfate. Ferric sulfate in itself is a good coagulant which at least partially accounts for why some recovered alum coagulants perform better than commercial alum. Also shown in Table 4-21 are the solids characteristics. The August test reduced the solids from 14,000 pounds to 6,600 pounds, a 53% reduction. The September test reduced the solids from 21,500 pounds to 15,600 pounds, a 27% reduction. A theoretical calculation showed that the first test should have had a 42% solids reduction and the second test a 39% solids reduction. Based on the difficulty of determining solids production on a full scale, the calculations may well be more accurate then the test results. In any event the average reduction would still be expected to be in the 35 to 40% range.

TABLE 4-21

INITIAL CONDITIONS/PARAMETERS OF RAW SLUDGE

CITY OF DURHAM

ALUM RECOVERY TESTS

	Test 1 (August)	Test 2 (September)
pH	7.13	6.64
Volume (gal.)	70,000	77,000
Suspended Solids Concentration (%)	2.4	3.35
Dry Weight of Solids (#)	14,000	21,510

ACID TREATED SLUDGE

	Test 1 (August)	Test 2 (September)
pH	2.1	2.0
H_2SO_4 (93% conc.) Added (gal.)	550	825
Ton Acid/Ton Alum Dissolved	0.67	0.68
Dry Weight of Solids Remaining (#)	6,600	15,610
% Solids Reduction	53	27

Source: Bishop (4-44)

2. Dewaterability. For the City of Durham, dewatering was accomplished on sand drying beds. Following the acid reaction, the solids were allowed to settle overnight. After removal of the supernatant alum, the remaining solids were raised to pH 3.5 with sodium hydroxide. A pH of 3.5 was selected for two reasons. First raising the pH of the solids from 2.0 to 3.5 would allow landfilling of the solids without concern for the solids being classified as hazardous by the corrosivity test. Secondly, pH 3.5 is below the point that significant precipitation of the remaining dissolved aluminum would occur. This means aluminum hydroxide would not form to interfere with dewatering and the dissolved aluminum could be recovered from the underdrain and decant of the sand beds. Table 4-22 shows the parameters as applied to the sand bed. In the first test a polymer was not utilized that successfully reduced the CST of the sludge, where in the second test a polymer was used to reduce the CST. Figure 4-77 shows the drainage results for the two tests, with the second test clearly showing better performance. The results indicated that only about one-third of the bed area would be required with alum recovery. Based on annual average this would save the City the construction of about 40,000 to 60,000 sf of bed area.

3. Alum Recovery and Reuse. Detailed lab study at Durham showed that essentially 100% aluminum dissolution is achieved at pH 2. Therefore, the amount of alum that can be reclaimed is primarily dependent upon the efficiency of separation between the solids and liquid. With the system used at Durham the sludge is acidified in a large tank. Following acidification the solids settle and the decant is removed as alum. Alum can also be collected as decant and underdrain from the drying beds. Table 4-22 shows the volumetric recovery from each of the tests. About 75% volumetric recovery was achieved. In the first test a higher volume would have been recovered from the beds if a successful

TABLE 4-22

RECOVERED ALUM AND RESIDUAL SLUDGE DEWATERING PARAMETERS

CITY OF DURHAM

ALUM RECOVERY TESTS

		Test 1 (August)	Test 2 (September)
1.	DEWATERING PARAMETERS		
	Volume Applied to Bed (gal.)	22,000	29,000
	pH after Neutralization	3.6	3.5
	Solids Concentration (%)	3.6	5.2
	Dry Weight of Solids Applied to Bed (#)	6,600	12,580
	# NaOH/Ton Solids Neutralized	83	103
	Polymer		Cationic
	Polymer Dosage (# polymer/ton solids)	38	
	Drying Bed Area Required (sf)	2,575	5,150
	Bed Loading Rate (#/sf)	2.6	2.4
2.	RECOVERED ALUM PARAMETERS		
	Total Sludge Volume (gal.)	70,000	77,000
	Volume Alum Supernatant (gal.)	48,000	41,000
	Volume Alum From Beds (gal.)	7,000	16,200
	Total Alum Volume (gal.)	55,000	57,200
	% Alum Recovery	79	74
	Aluminum Concentration (mg/l)	2,000	2,700
	# Alum Recovered	8,860	13,850

Source: Bishop (4-44)

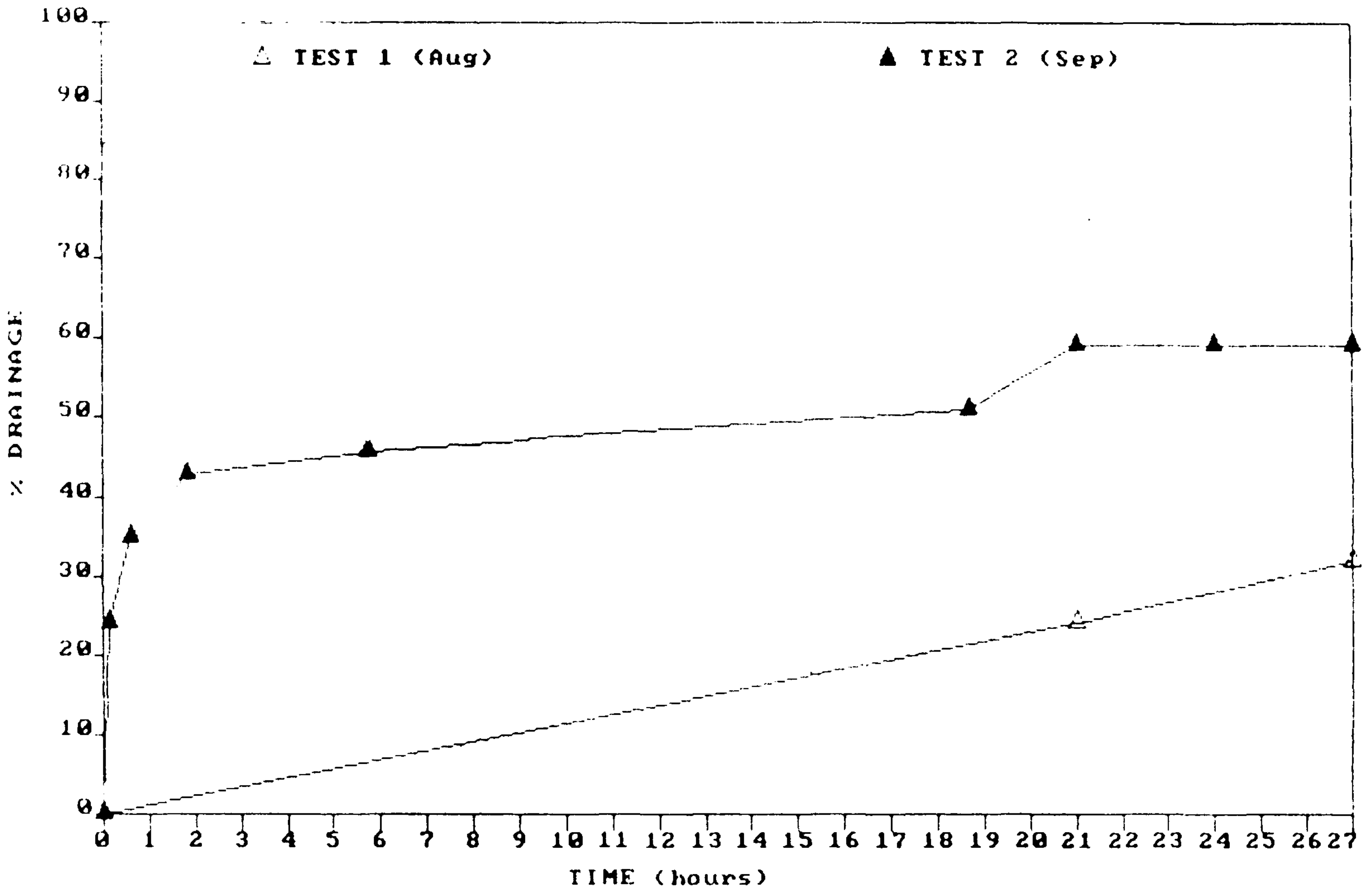


FIGURE 4-77

SAND DRYING BED DRAINAGE FOR ACIDIFIED SLUDGE
DURHAM, NC

polymer was used. Table 4-23 shows the metal concentrations of the recovered alum compared to the commercial alum used at Durham. Also shown is the concentration of metal in the alum divided by the aluminum concentration, expressed as ug metal per mg aluminum. It can be seen that fairly consistently the metals fed to the raw water would be higher for a given aluminum dose using recovered alum. However, except for iron and manganese, the dilution factor reduces the metals to below the MCL or goal values even if no removal occurred during treatment. Iron will act as a coagulant and is not a problem. Manganese should be closely monitored.

During the full scale tests the plant was split into two 11 mgd trains. One train used commercial alum and the second train used recovered alum. Essentially all normal operating parameters were identical. Of particular interest was the finished water metal concentrations, TOC and TTHMFP. Table 4-24 compares the finished water metal concentrations for the recovered alum and commercial alum sides of the plant. No differences were detected. As shown in Figures 4-78 and 4-79 the TOC and TTHMFP concentrations were generally higher on the recovered alum side. It is not known if this was due to a carry over of organics with the recovered alum or reflects a less then ideal alum dose on the recovered alum side as measured by organic removal. These are parameters that should be closely monitored.

Figure 4-80 shows the final process schematic that is being designed at Durham. Solids will be collected from the settling basins in batch thickeners. There the sludge will be thickened to 2 to 4% solids concentration. Acid will then be added to the thickener and mixed to effect aluminum dissolution. The remaining solids will settle overnight and the liquid alum will be decanted by gravity to an alum storage tank. The pH of the remaining solids will be raised to pH 3.5 with sodium hydroxide. The solids will then be pumped to the drying beds with polymer added in-line during the transfer.

TABLE 4-23

RECOVERED ALUM QUALITY

CITY OF DURHAM

	Commercial Alum		Recovered Alum	
Metal	mg/l	ug metal/mg Al	mg/l	ug metal/mg Al
Cd	ND	0	ND	0
Cr	9.5	0.2	0.6	0.3
Cu	0.1	0.002	0.6	0.3
Fe	1,160	18.4	292	146
Na	57	0.9	6.5	3.3
K	5.6	0.1	6.1	3.0
Mn	1.7	0.03	255	127
Ni	0.1	0.002	0.06	0.03
Pb	1.5	0.02	0.03	0.02
Zn	1.1	0.02	1.7	0.90
Ca	6.3	0.1	2.8	1.4
Mg	12.5	0.2	5.5	2.7
Al	63,000	---	1,970	---
Si	14.2	0.2	8.5	4.2
Ba	0.5	0.01	0.3	0.1
Ag	0.4	0.01	ND	0
As	3.0	0.05	1.1	0.05
Se	ND	0	ND	0
Hg	0.001	0	0.002	0

ND = below detection limit

TABLE 4-24

WATER TREATMENT PLANT PERFORMANCE COMPARISON USING RECOVERED ALUM

CITY OF DURHAM

TEST 1, AUGUST, 1985

all units ug/l

Rec = Recovered Alum
Comm = Commercial Alum

FINISHED WATER

		8/16		8/17		8/18	
Metal	Raw Water	Rec	Comm	Rec	Comm	Rec	Comm
Cd	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Cr	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Cu	<5.0	<5.0	<5.0	<5.0	5.8	<5.0	5.7
Fe	310	<50	<50	<50	<50	<50	<50
Na	5,450	5,570	5,800	5,780	5,800	5,760	5,800
K	1,830	1,850	1,840	1,980	1,880	1,970	1,910
Mn	840	10	40	40	40	40	<10
Ni	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Pb	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Zn	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Ca	5,550	5,600	5,400	5,600	5,500	5,550	5,400
Mg	2,650	2,650	2,700	2,700	2,600	2,600	2,650
Al	1,400	400	400	400	400	400	400
Si	5,000	4,600	4,800	4,950	4,800	4,500	4,200
Ba	85	30	35	30	30	35	45
Ag	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
As	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Se	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0	<5.0
Hg	<0.5	<0.5	<0.5	<0.5	<0.5	<0.5	<0.5

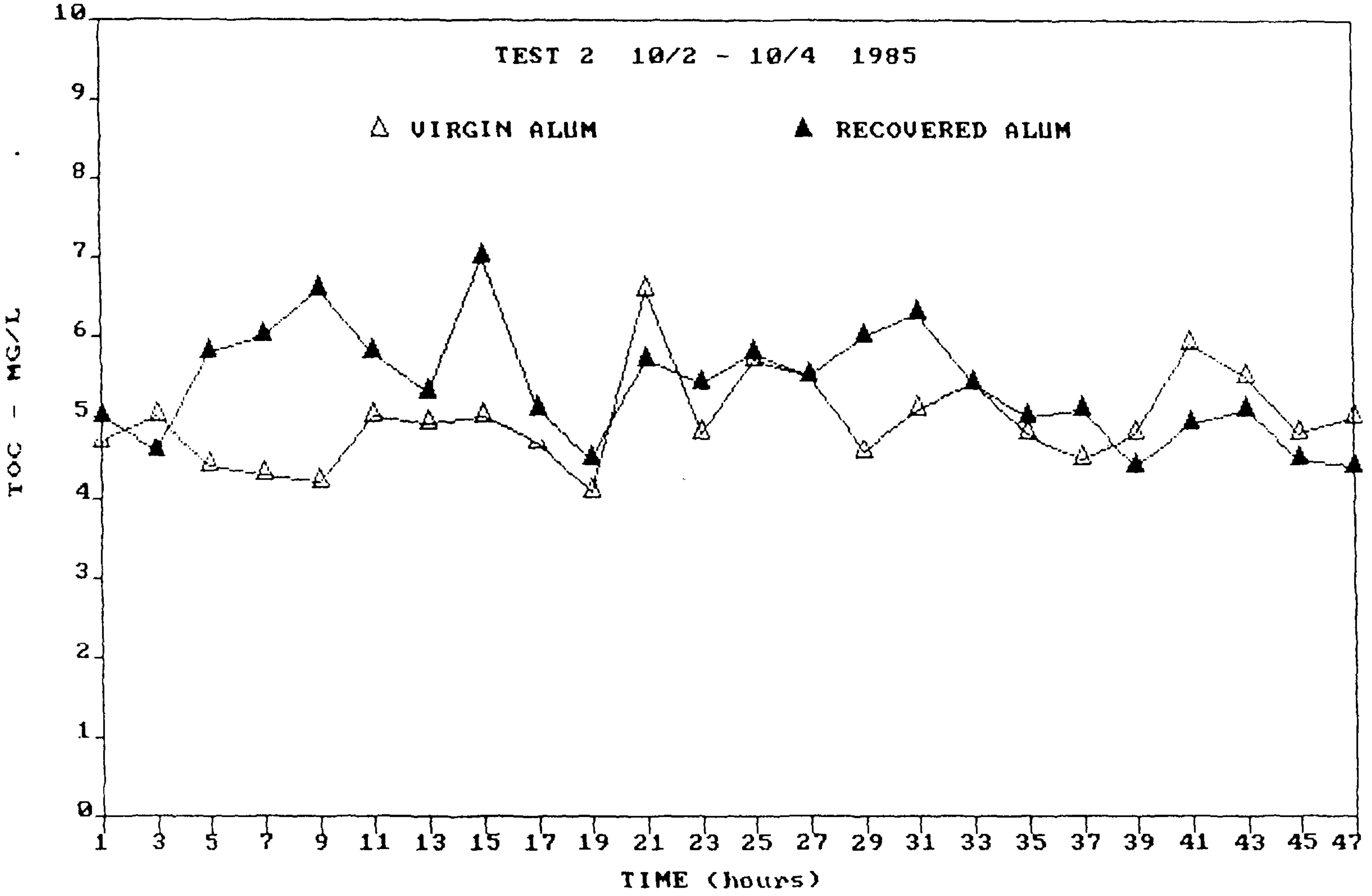


FIGURE 4-78

TOC REMOVAL USING RECOVERED ALUM
DURHAM, NC

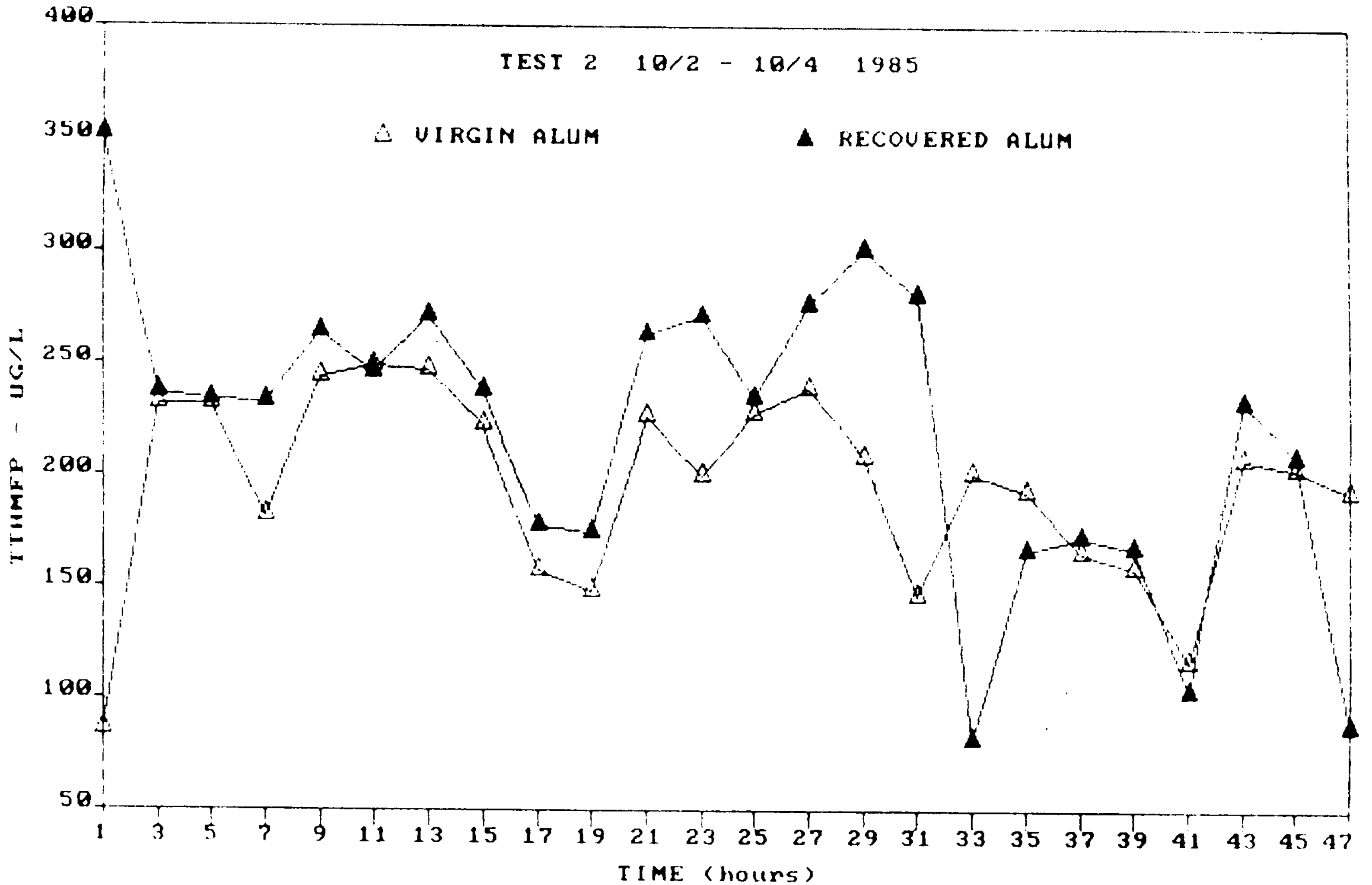


FIGURE 4-79

TTHM REMOVAL USING RECOVERED ALUM
DURHAM, NC

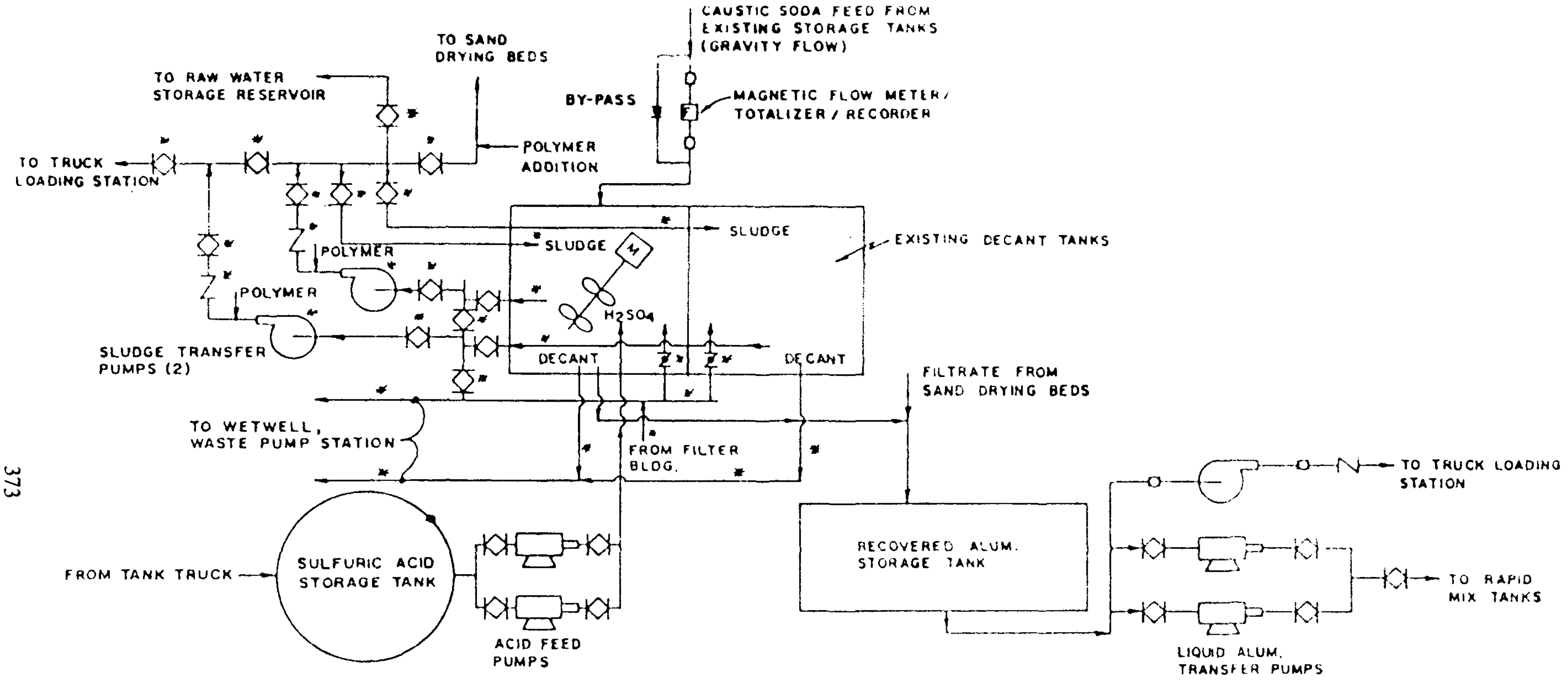


FIGURE 4-80

ALUM RECOVERY PROCESS FLOW DIAGRAM
DURHAM, NC

Decant and underdrain from the beds for the first 48 hours will also flow by gravity to the alum storage tank. Flexibility also exists to use the recovered alum at the wastewater plant to meet required phosphorus limitations. The recovered alum can be trucked to the wastewater plant where storage tanks are being located, or fed directly to the sewer.

4.13. DESIGN EXAMPLES

This section of the handbook provides design calculation examples and costs for the preliminary evaluation of sludge handling and dewatering alternatives. The following examples have been included to show some of the considerations which should be included with the evaluation of costs for sludge handling facility alternatives.

Alum Sludge

A water treatment plant using alum as a coagulant needs to dispose of basin sludge in a local landfill. The plant has operated for 10 years with direct discharge of its alum sludge but now must provide dewatering to meet local landfill requirements of 30% total solids concentration. The settling basins have continuous sludge removal equipment and produce a sludge solids concentration of approximately 1% throughout the year. A summary of typical monthly alum dose and raw turbidity is shown below.

	Turbidity (NTU)	Alum Dose (mg/l) as Alum	Daily Flow (mgd)
January	3	70	6 mgd
February	3	70	6 mgd
March	6	90	6 mgd
April	10	90	7 mgd
May	15	90	8 mgd
June	12	120	10 mgd
July	10	120	10 mgd
August	7	100	10 mgd
September	7	90	9 mgd
October	6	90	7 mgd

November	5	80	7 mgd
December	4	70	6 mgd

It is desired to evaluate the preliminary capital and O & M cost of dewatering sludge with drying beds versus a mechanical filter press operation. Regardless of which dewatering method is employed, sludge thickening and pumping facilities will be needed.

The following is an example of the preliminary sizing and costing of the sludge facilities.

Sludge Production

Sludge production at the plant has not been monitored however raw water alum dose and turbidity data are available.

Based on the equation for estimation of sludge production from Chapter 3:

$$S = 8.34\ Q\ (0.44\ Al + SS + A)$$

where,

S = Sludge in lb/day
Q = Plant Flow (mgd)
Al = Alum Dose as 17.1% Al_2O_3 (mg/l)
SS = Raw Water Suspended Solids (mg/l)
A = Additional Chemicals Added (mg/l)

the amount of alum sludge produced (lb/day) can be calculated. It is assumed that the concentration of suspended solids (SS) in the above equation is equal to 0.75 raw turbidity where:

$$SS\ mg/l = 0.75 \times NTU$$

Since no other chemicals are added the above equation becomes:

$$S = 8.34\ Q\ (0.44\ Al + 0.75\ NTU)$$

This equation results in the following estimates of monthly sludge production at the plant.

Monthly	Sludge Produced (lb/day)
January	1600
February	1600
March	2100
April	2600
May	3200
June	5100
July	4900
August	4000
September	3300
October	2500
November	2200
December	1600
Average	2900 lb/day

An estimate of the sludge flow assuming it is 1% solids would be:

$$Q\ (\text{mgd}) = \frac{\text{lb/day}}{8.34 \times 10{,}000}$$

The maximum sludge flow based on the sludge production information above shows that the month of June (5100 lb/day) would average 0.061 mgd (42.5 gpm). Average sludge flow would be 0.035 mgd (24.0 gpm).

Thickener Design

It is desired to provide a thickener which will thicken sludge from 1% to 4% during the maximum flow month. Based on batch thickening tests conducted on sludge from the plant, a mass loading of 0.12 lb/sf/hr would provide for a 4% solids concentration.

The required size of the thickener is then:

$$\text{Area (sf)} = \frac{5100 \text{ lb/day } (1 \text{ day}/24 \text{ hr.})}{0.12 \text{ lb/sf/hr}}$$

$$= 1770 \text{ sf}$$

In order to determine the cost of the thickener, Figure 4-9 reveals a cost of $180,000 for a circular concrete basin with 12 foot sidewater depth. This cost includes sludge removal equipment and other appurtenances such as influent baffles and effluent weirs.

The O & M costs for the thickener are estimated to be $1,600/year based on costs presented in Figure 4-10.

Thickened Sludge Pumping

The pumping of sludge from the basins to the drying beds or filter press will require the construction of sludge pumping facilities to transfer sludge from the thickener to the dewatering facility. The size of the pumping equipment is based on the expected maximum thickened sludge pump rate. Based on 1% influent and 4% underflow solids concentration the thickened sludge pump capacity can be calculated based on the approximate volume ratio:

$$\frac{V_2}{V_1} = \frac{P_1}{P_2}$$

where,

V_1 = 42.5 gpm
P_1 = 1%
P_2 = 4%

The expected sludge flow from the thickener would be:

$$V_2 \text{ (gpm)} = \frac{42.5\ (1\%)}{4\%}$$

$$= 10.6 \text{ gpm (at 4 percent solids)}$$

The design sludge pumping rate should be higher than this rate in that the facility will only operate for two shifts per day. Under this operating scenario the sludge flow becomes:

$$\text{Design Flow} = 10.6 \text{ gpm} \times \frac{24 \text{ hrs}}{16 \text{ hrs}}$$

$$= 16 \text{ gpm of } 4\% \text{ solids}$$

The cost of a pumping facility for sludge is shown in Figure 4-19. The estimated cost for the sludge pumps is $43,000 including connecting piping, valving electrical equipment and instrumentation. (This construction cost curve was used because it includes the cost of separate dry well, wet well construction). The O & M cost for the pumps is shown in Figure 4-20 and has been estimated to be $4,200/year.

Filter Press

The size of a filter press to dewater the sludge from the thickener is based on obtaining a desired sludge cake solids concentration of 30%. The overall process yield for a filter press can be estimated from the equation (from Section 4.7):

$$Y = Y_p \left(\frac{F_t + S_t}{F_t + S_t + M_t}\right)$$

where,

Y = Overall Yield for Press (lb/hr/sf)

Y_p = Process Yield (lb/hr/sf)
F_t = Filtration Time (min)
S_t = Squeeze Time (min)
M_t = Machine Time (min)

Assume based on pilot testing that a 30% solids cake can be produced with a process yield of 0.4 lb/hr/sf and that the following typical values from pilot tests apply for F_t, S_t, and M_t.

$$F_t = 120 \text{ minutes}$$

$$S_t = 18 \text{ minutes}$$

$$M_t = 15 \text{ minutes}$$

The overall process yield is then:

$$Y = 0.40 \text{ lb/hr/sf } \left(\frac{120 \text{ min} + 18 \text{ min}}{120 \text{ min} + 18 \text{ min} + 15 \text{ min}}\right)$$

$$= 0.36 \text{ lb/hr/sf}$$

In order to determine the size of the press needed for solids dewatering assume that the press would handle maximum month solids production during 7 days a week and two 8 hour shifts each day.

As presented earlier, the solids to the facility from the thickening process during June is 5100 lb/day on a monthly average. Converting this to two shifts (or 16 hours a day operation):

Design Solids Loading Rate to Process =

$$16 \text{ gpm} \times 60 \frac{\text{min}}{\text{hr}} \times 8.34 \frac{\text{lb}}{\text{gal}} \times 4\% = 320 \text{ lb/hr}$$

The filter press size is based on the total filter press area required to process 320 lb/hr of solids. The resulting filter press size can then be calculated as follows:

$$\text{Press Size (sf)} = \frac{\text{Sludge Processed}}{\text{Y}}$$

$$= \frac{320 \text{ lb/hr}}{0.36 \text{ lb/hr/sf}} = 889 \text{ sf}$$

The cost estimate for a filter press facility including the press, building, ancillary equipment and housing is shown in Figure 4-34. The estimated 1986 construction cost is $2,050,000. The annual O & M cost for this facility is shown in Figure 4-35 and is estimated at $58,000 per year.

Sand Drying Beds

The use of sand drying beds requires knowledge of local evaporation rates. Beds are usually sized to handle the worst case drying conditions (usually winter and spring months) when evaporation rates are much lower than during summer months. Assume for this example that the local evaporation rates based on historical records are as follows:

Winter/Spring Evaporation = 3.0 inches/month

Summer Evaporation = 7.0 inches/month

Annual Average Evaporation = 4.5 inches/month

Also, assume that (based on pilot tests) that the initial loading depth to the beds will be 12 inches for good drainage performance and the solids concentration from the thickeners will be 4% solids. Typically a polymer is added to sludge prior to application to the beds and for this example assume that a dose of 10 lb per ton of polymer is required to provide good drainage on the beds. In order to compare the beds to the filter press option, assume that the final desired cake solids concentration is 30%.

In the calculation of the required surface area for drying bed sludge dewatering, the following definitions apply:

IA = Initial application of sludge in pounds dry solids per square foot.
D(i) = Depth applied initially in inches.
DS(i) = Percent dry solids initially.
D(f) = Percent dry solids concentration desired for final cake.
DD = Change in depth.
DD(u) = Change in depth due to loss of water to underdrain and decanting on bed.
DD(e) = Change in depth due to loss of water to evaporation.
P = Percent of volume applied to beds which passes through sand or is decanted, expressed as a decimal fraction.
T = Drying time in months.
E = Rate of evaporation in inches per month.
AA = Number of applications per year.
Y = Bed yield in pounds dry solids per square foot per year.

The initial loading of sludge (IA) in pounds per square foot for a given application can be calculated, based on the depth of sludge applied, D(i), and the dry solids content, DS(i), of the sludge applied.

$$IA = \frac{D(i)}{12 \text{ inches per foot}} \times \frac{62.4 \text{ pounds}}{\text{cubic feet}} \times \frac{DS(i)}{100}$$

$$= \frac{12 \text{ inches}}{12} \times 62.4 \text{ lb/cf} \times \frac{4}{100}$$

IA = 2.5 lb/sf loading onto the beds.

The final depth can be determined from:

$$D(f) = D(i) \times \frac{DS(i)}{DS(f)}$$

$$= 12 \text{ inches} \times \frac{4\%}{30\%}$$

$$D(f) = 1.6 \text{ inches}$$

The resulting total change in sludge depth is then:

$$DD = D(i) - D(f)$$

$$= 12 \text{ inches} - 1.6 \text{ inches}$$

$$DD = 10.4 \text{ inches}$$

The amount of depth lost to the underdrain based on P = 50% (determined from pilot tests) can be calculated by:

$$DD(u) = D(i) \times P$$

$$= 12 \times 0.50$$

$$DD(u) = 6 \text{ inches}$$

The time, T, to achieve a 30% solids concentration is dependent on the evaporation rate. Assume that the winter evaporation rate of 3.0 inches per month occurs from January through early May. A review of the sludge production calculated earlier shows that May is the highest sludge production month (3,200 lb/day) during this low evaporation period.

The change in depth required due to evaporation is then:

$$DD(e) = DD - D(u)$$

$$= 10.4 \text{ inches} - 6 \text{ inches}$$

$$DD(e) = 4.4 \text{ inches}$$

and the time required is then

$$\begin{aligned} T &= \frac{DD(e)}{E} \\ &= \frac{4.4 \text{ inches}}{3.0 \text{ inch/month}} \\ T &= 1.5 \text{ months} \end{aligned}$$

The number of applications, AA, to the beds which can be accomplished under these solids loading and evaporation conditions is then:

$$\begin{aligned} AA &= \frac{12 \text{ month per year}}{T} \\ &= \frac{12}{1.5} \\ AA &= 8 \text{ applications per year} \end{aligned}$$

The equivalent yield, Y, in pounds per square foot per year is then:

$$\begin{aligned} Y &= IA \times AA \\ &= 2.5 \text{ lb/sf} \times 8\text{/year} \\ &= 20 \text{ lb/sf/year} \end{aligned}$$

In order to determine the bed area required, the annual equivalent solids production based on May solids production is:

Annual Solids Equivalent

$$\begin{aligned} &= 3{,}200 \text{ lb/day} \times 30 \text{ day/month} \times 12 \text{ month/year} \\ &= 1{,}152{,}000 \text{ lb/year} \end{aligned}$$

and,

$$\text{Bed Area} = \frac{1{,}152{,}000 \text{ lb/year}}{20 \text{ lb/sf/year}} = 57{,}600 \text{ sf}$$

The estimated 1986 construction cost for 57,600 sf of sand drying beds can be estimated from Figure 4-59 and is $811,000. The annual O&M cost for the beds can be estimated from Figure 4-60 and is $90,000 per year.

It is typical for most alum sand drying bed operations to add a polymer to the sludge prior to addition to the beds. Assuming a dose of 10 lb of polymer per ton of solids, as mentioned earlier, the polymer feed system size can be estimated as follows.

The dose required is 10 lb per ton and the maximum month solids generation was 5,100 lb per day in June.

The resulting maximum required polymer feed rate can then be calculated as follows:

Polymer Dose = 10 lb/ton of solids applied

Solids production during a 16 hour, 7 day a week operation is:

$$\text{Solids Produced} = 5{,}100 \text{ lb/day} \times \frac{24 \text{ hr}}{16 \text{ hr}} \times \frac{1 \text{ day}}{24 \text{ hr}}$$

$$= 319 \text{ lb/hr equivalent production}$$

The amount of polymer feed rate can then be calculated as:

Polymer feed rate (lb/hr)

$$= 319 \frac{\text{lb solids}}{\text{hr}} \times 10 \frac{\text{lb poly}}{\text{ton solids}} \times \frac{1 \text{ ton}}{2{,}000 \text{ lb}}$$

$$= 1.6 \text{ lb/hr}$$

The 1986 construction cost for the polymer feed system is shown in Figure 4-13 and is estimated at $38,000. The annual O&M cost for the polymer feed system from Figure 4-14 is estimated to be $5,800 per year. These costs do not include the cost of the polymer itself. The polymer cost can be estimated assuming a delivered price of $1.30/lb (which may vary significantly depending on type and manufacturer) and an average annual use based on average annual sludge production. The estimated annual polymer cost is as follows:

Average Annual Sludge Production = 2,900 lb/day

$$\text{Solids Ton/day} = \frac{2{,}900 \text{ lb}}{2{,}000 \text{ lb/ton}} = 1.45 \text{ ton/day}$$

Daily Polymer Use = 1.45 ton/day x 10 lb poly/ton solids

= 14.5 lb poly/day

Annual Cost = 14.5 lb poly/day x $1.30/lb x 365 days/yr

= $6,880/yr

Cost Summary

A summary of the costs for the construction and O&M of a filter press facility and a sand drying bed facility for the alum sludge design examples is as follows. Additional costs for hauling and disposal of the dried sludge are not included.

Dewatering Method	1986 Construction Cost	Annual O&M Cost	Annual Chemical Cost
FILTER PRESS			
a. Thickener	$ 180,000	$ 1,600/yr	--
b. Sludge Pumping	$ 43,000	$ 4,200/yr	--
c. Filter Press	$2,050,000	$ 58,000/yr	--
	$2,273,000	$ 63,800/yr	

SAND DRYING BEDS

a.	Thickener	\$ 180,000	\$ 1,600/yr	--
b.	Sludge Pumping	\$ 43,000	\$ 4,200/yr	--
c.	Drying Beds	\$ 811,000	\$ 90,000/yr	--
d.	Polymer Feed	\$ 38,000	\$ 5,800/yr	\$6,880/yr
		\$1,072,000	\$102,600/yr	\$6,880/yr

These cost estimates can then be used in a Present Worth (PW) analysis or Annual Equivalent Cost (AEC) analysis to evaluate the most cost effective solution to sludge handling. For example, the present worth of the filter press facility based on a 20 year facility life and 8% interest:

$$\begin{aligned} PW &= \$2{,}723{,}000 + 63{,}800(P/A)^{8\%}_{20} \\ &= \$2{,}273{,}000 + 63{,}800(9.818) \\ &= \$2{,}899{,}000 \end{aligned}$$

For the Sand Drying Beds:

$$\begin{aligned} PW &= \$1{,}072{,}000 + 102{,}600(P/A)^{8\%}_{20} \\ &= \$1{,}072{,}000 + 102{,}600(9.818) + 6{,}880(9.818) \\ &= \$2{,}147{,}000 \end{aligned}$$

The results of this analysis is that the sand drying bed operation would be the least cost alternative based on present worth analysis.

Lime Sludge

A softening plant using groundwater as a source of supply removed 150 mg/l (as $CaCO_3$) of calcium hardness and 35 mg/l (as $CaCO_3$) of magnesium hardness with lime.

Water production is fairly constant throughout the year at a rate of 25 mgd. It is desired to evaluate the cost for dewatering of the lime sludge with a vacuum filter. Sludge will be drawn directly from the settling basins at a 10% concentration.

The quantity at sludge produced (from Chapter 3) can be estimated from

$$S = 8.34\ Q[2.0\ Ca + 2.6\ Mg]$$

where,

Q = Plant Flow, mgd
Ca = Calcium Hardness Removed as $CaCO_3$ (mg/l)
Mg = Magnesium Hardness Removed as $CaCO_3$ (mg/l)
S = lb per day of Solids Produced.

The daily sludge production is then

$$S = 8.34(25\text{ mgd})(2.0 \times 150\text{ mg/l} + 2.6 \times 35\text{ mg/l})$$

$$= 81{,}523\text{ lb per day}$$

It is desired to process this sludge over 7 days a week and two 8 hour shifts. The equivalent design loading rate is then:

$$\text{Design Solids Loading} = 81{,}523\text{ lb/day}\left(\frac{1\text{ day}}{16\text{ hr}}\right)$$

$$= 5{,}095\text{ lb/hr}$$

Based on pilot tests conducted to evaluate design parameters it was determined that a 40 lb/sf/nr loading rate would provide the required solids concentration of 50% minimum.

The required vacuum filter area is then calculated as:

$$\text{Filter Area} = \frac{5095 \text{ lb/hr}}{40 \text{ lb/hr/sf}} = 127 \text{ sf}$$

The 1986 construction cost from Figure 4-47 is estimated to be $210,000. The estimated annual O&M cost for this facility is $110,000/year.

CHAPTER 5

SYSTEM OPTIMIZATION FOR SOLID/LIQUID WASTE MANAGEMENT

One of the most difficult tasks for a utility or consultant in determining the most efficient waste treatment and disposal plan is to "put it all together." The many options available, all the way from changing coagulants to reduce the quantity of sludge, to interrelated recovery systems, requires extensive evaluations to determine the best system.

Chapter 4 presented cost curves on the various treatment and disposal options to be considered by a utility or consultant. Chapter 3 presented methods to estimate sludge production quantities. With this information the critical scenarios can be developed -- how much sedimentation basin sludge and volume is there, how should it be removed from the sedimentation tanks (is it better to have continuous removal equipment followed by a small thickener, or is batch, manual cleaning acceptable with a larger holding tank?); how much backwash water is produced (which all interrelates to filter run lengths and filter wash sequencing); should we change coagulants which may cost more for chemicals but reduces the sludge volume; are holding tanks necessary, and if so, what size; what dewatered solids concentration is needed for a given disposal method and how does that relate to the dewatering method needed?

Obviously a whole list of such questions can be created and should be answered for proper system optimization. Some answers are easy and some combinations ridiculous. However, most are very legitimate combinations which are difficult to quickly evaluate. The approach presented here allows the sludge management process to be integrated into the complete water treatment process to help determine the overall solution.

A computer program compatible with an IBM/PC has been written to allow evaluations of optimum sludge management systems[7]. The program was developed to allow the user to create the sludge treatment and disposal system as desired, with the program then showing the anticipated results and cost of implementing such a system. It becomes an interactive program whereby the user can use the results and make changes in the system to develop an improved system in subsequent runs. The program is a tool which allows several combinations to be evaluated quickly, so that the user can determine which systems deserve further attention.

In order to be a total system optimization, the sludge management program has been combined with a water treatment process program. This process program provides output on sludge and backwash quantities for different plant operating conditions. In this manner the user can also evaluate the effects of different coagulants, different lime softening treatment considerations and backwashing operations on sludge management decisions.

An example of an input sequence to the sludge management program would be to create the system desired: manual cleaning of basins once per 3 months, holding basin to equalize flow, filter press, landfill at a site 5 miles away. Output from the program would include sludge characteristics at all stages in the system, number of units required, chemical demands and capital and operational costs. The user would then create multiple situations, comparing the end results. The objective is not that the user can now design a system, but rather he has better insight as to which combinations warrant testing or detailed evaluation.

The following sections of this chapter give a description of the program, limitations and usefulness, and assumptions made in the program development.

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[7] Program is available by contacting the authors.

5.1. THE SLUDGE MANAGEMENT MODEL

The sludge management model consists of three main programs with the file names, SLUDGE, WATER, and RESULT. All three programs work interactively to evaluate a complete sludge handling/treatment process. The user should be aware that all data entries need to be made with lower case letters, otherwise error messages may appear.

5.1.1. Start Up Program

To run the model, the user first needs to load BASIC from the DOS diskette, because all three programs are written in BASIC language. Next, SLUDGE is to be loaded which in turn executes the complete model. All subsequent programs will be executed automatically during the course of the program. Of course, the model can be made self loading with the DOS diskette. For this, the user should refer to the IBM Disk Operation System manual. SLUDGE acts as an introduction to the user and it briefly outlines the main program, WATER. Also, SLUDGE allows the user to execute a subroutine to update the cost equations for the various types of equipment. A detailed description of cost updating is presented in Section 5.1.5. Finally, SLUDGE also asks whether data files are to be used in the main program. The first time the model is run, no data files have yet been created, and thus the answer is no ("n"). Data files are more thoroughly discussed in Section 5.1.4.

5.1.2. Main Program

Immediately following SLUDGE, the main program, WATER, is automatically executed. WATER performs all the calculations required for the specific sludge handling process the user wants to "design". Each item requires a certain amount of input data which are outlined in Section 5.2. To aid a user who does not have the availability of specific data, say an alum sludge loading rate for a gravity thickener, the computer displays common ranges of design values following many input items. Before attempting to "design" a system, Section 5.2 of this chapter needs to be reviewed such that all the required input data are readily available.

As one is providing input data for WATER, occasional mistakes may occur. The program contains two features that allow a user to make corrections without starting all over. First of all, those items which require a fairly large amount of data ask the user whether changes are needed at the end of the input sequence. If corrections are necessary, the program returns to the first input item of the particular system one is working on and displays the current input value followed by asking what the new value should be. If no correction is required for a particular item, the user enters "s". For example:

	Current Value	New Value
·raw turbidity	10	s
·turbidity/solids conversion factor	1	2

Here, the raw turbidity remains the same, but the conversion factor is changed from 1 to 2.

The second method for making changes is by simply running the particular item again and inputting the correct values. This option is available for those items which require very little input data. After data has been entered, the bottom of the screen contains a message like:

```
a-continue   b-return to sludge pumping   c-return to treatment
```

If the user, for example, just completed the sludge pumping sequence, but an error was made, simply enter "b" which returns the program to the beginning of sludge pumping.

5.1.3. Output Program

Following all the calculations in the WATER program, the computer loads the third program, RESULT, from the diskette which deals primarily with the output data.

First a summary table of the sludge characteristics is displayed on the screen. This table lists all the sludge treatment and handling items which were selected with their corresponding sludge flow rates and percent solids. An example may look like this:

```
                    SUMMARY OF SLUDGE HANDLING
                    --------------------------

                  Inflow Rates               Outflow Rates
            sludge     sludge  solids  sludge     sludge  solids
item      (lbs/day) (gal/day) (%)    (lbs/day) (gal/day)  (%)
----------------------------------------------------------------
Sed. Basins                             7672.8     92000  1.000
Unthick.P.S. 7672.8     92000    1.000  7672.8     92000  1.000
Thickener    7672.8     92000    1.000  7672.8     23000  4.000
Thick. P.S.  7672.8     23000    4.000  7672.8     23000  4.000
Belt Press   7672.8     23000    4.000  7672.8      6133 15.000

* NOTES:  1..  Sedimentation basins are cleaned mechanically
          2..  Sludge conditioning is used prior to mechanical
               dewatering
```

Next, a summary table of all the construction and annual operation and maintenance costs which are associated with each sludge handling item is printed on the screen. Also, a total construction and O & M cost is shown as well as how many units are required of the various types of equipment.

```
               SUMMARY CONSTRUCTION AND O & M COST
               -----------------------------------

                       construction    annual o & m     number of
item                       cost            cost           units
----------------------------------------------------------------
UNTHICK.SLUDGE P.S.       81278           12378             1
GRAVITY THICKENER        130478           34198             1
THICK. SLUDGE P.S.        24532            4773             1
BELT FILTER PRESS        312847           46041             1
```

```
CONDITIONING              44408           8532            1
LANDFILL                                 15030
                         ------          ------
                         ------          ------
     TOTAL               593543          120952
```

After these tables have been reviewed by the user, the computer asks whether a complete hard copy output of all the input and output data is desired. The printed output contains the two previously discussed summary tables and complete input and resulting output data for each sludge handling item. Examples of this type of output can be found in Section 5.3 where some example problems are presented.

5.1.4. Data Files

After a complete sludge handling process has been put together, the user has the opportunity to save the input data and results on a diskette. It is recommended that a separate disk is used to store these data files. At the end of the output program, the computer asks if the user wants to save the particular system that was just "designed". This is very useful if, for example, total cost comparisons are to be made between systems utilizing different types of dewatering equipment. After a simulation is performed, the user can save it under a given file name prior to executing the next simulation. Once all simulations have been processed, print outs can be produced of each simulation for a detailed comparison.

> NOTE: After a simulation has been saved, the user can directly return to the main program using the previously entered data up to the point where changes need to be made. This can save a considerable amount of time when performing simulations for comparison.

The set-up of the data file sequence is fairly straight forward. At the end of the output program, the results of each selected sludge handling process are saved. File names, such as MODEL1.DAT or RUN1.DAT, need to be specified for each simulation. The same file name must be used when loading the data file in the SLUDGE program. When using data files, the main program simply prints the summary table for each sludge handling item again. If no changes have to be made to a particular item, press "a" (continue) and the results of the next item are printed. If, however, changes have to be made, the same routine as described is Section 5.1.2 for correcting input errors is used.

Unfortunately, the use of data files has some limitations. Since all sludge treatment and handling items are based on the sludge quantities and percent solids determined in the previous item, any change to a particular process effects all the processes that follow. For example, if the underflow percent solids from a gravity thickener is changed from 4% to 3%, this in turn results in a larger sludge flow rate (in gal/day) which affects the thickened sludge pumping station, sludge dewatering and final disposal. Thus, in other words, data files are only good up to the point where changes need to be made. From there on, all data needs to be entered as usual.

5.1.5. Cost Updating

One of the main objectives of the computer model is to determine total capital and operating and maintenance costs for the various sludge handling systems. For each particular item the model calculates these costs utilizing equations derived from the cost curves shown in Chapter 4 of the handbook. These equations need to be updated over a period of time to adjust for increases in cost for building materials, labor, power, etc.

As mentioned in Section 5.1.1, the cost updating routine can be initiated in the SLUDGE program. Before using the cost updating routine, the user should refer to Appendix A of the handbook which discusses the various methods for adjusting the cost curves. The same methods are used in the computer model for updating the cost equations.

5.2. PROGRAM ITEMS

The following sections are organized in the same manner in which the program is set up, and they discuss the various aspects of each sludge handling/treatment item. In the program, the user has the opportunity to select any combination of items in order to "design" a custom made sludge management process.

5.2.1. Water Treatment Process

The first item the program addresses is the water treatment process. This is where the initial sludge volume and characteristics are established based on raw water conditions and chemical additives.

The model can simulate processes which apply to coagulation/filtration plants or to water softening plants. A third option under the water treatment process that can be utilized is sludge quantity input. Here, the user can bypass the water treatment process and enter actual sludge quantities directly into the program. After a specific water treatment process is simulated with the computer model, a summary table is printed on the screen. This table includes all the input data and resulting sludge quantities.

The following is a more detailed description of each option under the water treatment process. It discusses each item's purpose and the required input data.

Option 1: Coagulation

To determine the daily quantity of sludge produced at a filtration plant, the following data needs to be entered:

- Raw Water Turbidity
- Turbidity Solids Conversion Factor (to convert raw turbidity (ntu) to suspended solids (mg/l) - see Chapter 3 of the handbook.)
- Sludge Percent Solids in the Sedimentation Basins
- Chemical Additives (alum or iron, polymer, carbon), (mg/l)
- Plant Flow (mgd)
- Sedimentation Basin Surface Area (sf)
- Precipitate Settling Velocities (cm/min) (obtained from jar tests @ 80% and 20% remaining) or Applied Turbidity to the Filters
- Number of Filters
- Filter Loading Rate (gpm/sf)
- Filter Run Length (hours)
- Whether Filter Backwash is Recycled or Not

With these data, the computer calculates the sludge production in lbs/day, the sludge flow rate in gal/day, and the backwash solids and flow rate in lbs/day and gal/day respectively. Also, if filter backwash is recycled back to the head of the plant, the recycle flow rate is expressed in gal/min. For filter backwash, the program assumes that 20 gpm/sf over a duration of 15 minutes is required.

Option 2: Lime Softening

To simulate a water softening plant, the following data needs to be entered:

- Calcium (mg/l as ion)
- Magnesium (mg/l as ion)
- HCO_3 (mg/l as ion)
- CO_2 (mg/l as ion)
- Final Hardness (mg/l as $CaCO_3$)
- Plus all Required Data Points Listed for Water Filtration Plants (Option 1)

In addition to calculating sludge and backwash items discussed in Option 1, the program also determines the required lime and soda ash dosages and expresses them in mg/l as $CaCO_3$. In some cases, no soda ash may be required to obtain the desired final hardness resulting in zero being printed for the soda ash dosage.

Option 3: Sludge Quantity Input

This option provides the user with a method to enter actual sludge quantities into the computer model and requires the following input data:

- Plant Capacity (mgd)
- Sedimentation Basin Surface Area (sf)
- Sludge Production (lbs/day)
- Sludge Solids Concentration (%)

With the above data, the program determines the sludge flow rate in gal/day and then continues directly to sedimentation basin cleaning procedure (see Section 5.2.2). This method can be useful if, for example, a utility has a good feel for their daily sludge production and, therefore, does not need to simulate the water treatment process.

5.2.2. Sludge Removal

The model offers the user two methods for sludge removal from the sedimentation basins. One method is mechanical cleaning in which either scrapers, vacuum system or similar, continuously removes sludge from the bottom of the basins. With this method, the user has the option to control the rate of sludge removal by specifying the number of hours per day and days per week the sludge removal equipment operates. If the user declines to control the removal rate, the program assumes that the removal equipment operates 24 hours per day, 7 days a week.

The second method is a batch removal procedure in which a sedimentation basin is shut down periodically (say every 30 or 60 days) and completely drained. The remaining sludge on the bottom is then forced out through drains using fire hoses.

The batch removal sequence in the program is set up with an equalization basin such that a continuous sludge flow occurs to the dewatering facilities. The user has the option to include a sludge transfer pump station between the sedimentation basins and the equalization basin. Sludge is removed from the equalization basin with an unthickened sludge pump station as described in Section 5.2.3.

To determine the sizes of the above described units for the manual removal system, the user first of all needs to establish the quantity of sludge that will be removed from the sedimentation basins, by entering the following data:

- Depth of Sedimentation Basins (feet)
- Number of Sedimentation Basins
- Number of Operating Days before Cleaning Basins

Based on the previously determined sludge production rate, the volume of sludge in each basin is calculated by the program.

Next, to find the sludge flow rate from the sedimentation basins, the user needs to enter how many days and hours per day are required to clean a single basin. With this information, the size and associated construction and O & M costs for the transfer pump station (if selected) are determined.

Finally, to calculate the size of the equalization basin it is assumed that all sludge is removed over a period equal to the number of operating days of the sedimentation basins. Also, the equalization basins are assumed to be 25 feet deep. The user has to input the sludge percent solids leaving the equalization basin as well as whether the sludge outflow will be continuous or controlled over a certain number of days per week. Construction and operation and maintenance costs for the equalization basin are based on the cost curves developed for a gravity thickener.

At the end of the batch removal routine, a summary table appears on the screen, indicating the required sizes of the units and their associated costs.

For both methods of sludge removal from the sedimentation basins, the user should note that if the sludge flow rate is not continuous, annual operation and maintenance costs are reduced since the facilities do not operate 365 days a year.

5.2.3. Unthickened Sludge Pumping

After the sludge has been removed from the sedimentation basins, it requires dewatering prior to final disposal. In some instances, the sludge can flow by gravity to specific dewatering facilities, but in most cases a pump station is required.

If a pump station is selected to be part of the sludge handling process, the computer calculates the pump station's required size and the construction and annual operation and maintenance costs based on the previously determined sludge flow rate.

5.2.4. Gravity Thickener

The first type of dewatering process that can be chosen is a gravity thickener. If a gravity thickener is included in the sludge treatment process, the program first checks whether an unthickened sludge pump station precedes the thickener. If a pump station was not selected, the program asks whether it should return to the unthickened sludge pumping sequence or continue with the gravity thickening sequence. A user may, for example, omit cost calculations for a pump station because a method already exists to get the sludge to the thickener.

Next, the current sludge flow rate and percent solids are displayed on the screen, along with two methods for determining thickener size and outflow conditions. The first method deals with the relationship between solids flux and suspended solids concentration (see Section 4.3 of the handbook). The input requirements for this method are:

1. Sludge Underflow Percent Solids (%)
2. Sludge Solids Flux (lb/sf/day)

The second method is based on familiarity with similar treatment plants where gravity thickeners are in use. The program asks the user to input the sludge loading rate (lb/sf/hr) and the underflow percent solids concentration.

After the required input data have been provided, the computer calculates the resulting outflow sludge quantities and characteristics as well as the required surface area of the thickener. Based on the surface area, the computer determines the number of thickeners required, total construction cost, and annual operation and maintenance cost. At the end of the gravity thickening sequence, a summary table appears on the screen which includes inflow and outflow characteristics, thickener size, and associated costs.

Also, during the sludge thickening routine, the user has the chance to adjust the sludge flow rate from the thickener to subsequent dewatering facilities in order to meet specific operating schedules. If the user selects to adjust the current sludge flow rate, the hours per day and days per week during which sludge is removed from the thickener need to be entered.

5.2.5. Thickened Sludge Pump Station

Following the gravity thickener, the sludge volume can be further reduced by using mechanical and/or non-mechanical dewatering processes. In some instances, the thickened sludge can flow by gravity to the next dewatering process, but again a pump station is usually required. The program utilizes similar calculations as for an unthickened sludge pump station to determine construction and operation and maintenance costs.

5.2.6. Sludge Dewatering

The final dewatering option consists of three methods to choose from:

1. Mechanical Dewatering
2. Non-Mechanical Dewatering
3. Combination of Both

If method 1 or 2 is selected, the computer immediately proceeds to the specific part of the model assigned to mechanical or non-mechanical dewatering. A description of each part follows shortly. If the user desires to employ both mechanical and non-mechanical dewatering (select method 3), the computer returns by asking what percentage of the total sludge flow is assigned to each dewatering process. This method is provided if a utility desires to use say belt filter presses for 40% of the total sludge flow and sand drying beds for the remaining 60% of the sludge. If method 3 is selected, the program first addresses mechanical dewatering then non-mechanical.

Option 1: Mechanical Dewatering

If mechanical dewatering is selected as part of the sludge dewatering process, the screen clears and the following Table appears:

EQUIPMENT	RESULTING PERCENT SOLIDS ALUM SLUDGE	 LIME SLUDGE
1...Decanter Centrifuge	10 - 15	55 - 56
2...Filter Press	35 - 45	55 - 70
3...Vacuum Filter	---	45 - 65
4...Belt Press	10 - 15	---

The above Table indicates the four dewatering options that are available to the user. Also, a typical range of the final percent solids that can be obtained with each option are indicated based on the type of sludge that is treated. The input data requirements depend on which type of equipment is simulated and is summarized below:

1. Decanter Centrifuge
   - Final Solids Concentration
2. Filter Press
   - Final Solids Concentration
   - Squeeze Time (min)
   - Process Yield (lb/sf/hr)
   - Filtration Time (min)
   - Cleaning Time (min)
3. Vacuum Filter
   - Final Solids Concentration
   - Sludge Loading Rate (lb/sf/hr)
4. Belt Filter Press
   - Final Solids Concentration
   - Sludge Loading Rate (lb/ft/hr)

The resulting percent solids can be selected from the above Table or the user can input a solids concentration based on experience and familiarity with the sludge and the type of equipment. For the other input data, Chapter 4 of the handbook provides useful information. Sludge conditioning usually precedes all of the available mechanical dewatering options. A discussion of the conditioning sequence is addressed in Section 5.2.7.

After the required input data have been entered, the screen clears and a summary table of mechanical dewatering results appears. The table includes inflow and outflow sludge quantities and characteristics, number of dewatering units that are required, annual sludge production and the construction and annual operation and maintenance costs. Also, preconditioning equipment and O & M costs are included (if selected).

Option 2: Non-Mechanical Dewatering

Non-mechanical dewatering options consist of storage lagoons and sand drying beds. After the non-mechanical dewatering sequence has been executed, the screen clears and the following table appears:

	RESULTING SLUDGE % SOLIDS	
	ALUM SLUDGE	LIME SLUDGE
1...Storage Lagoons	7 - 15	50 - 60
2...Sand Drying Beds	20 - 25	50

The resulting sludge percent solids listed in the above Table are commonly obtainable concentrations for alum and lime sludges.

The first set of input data requested by the computer are the type of dewatering method that is used, the resulting sludge percent solids, and whether pre-conditioning of the sludge is desired.

If storage lagoons are selected as a method of dewatering, the next input item is how many months the lagoons operate before being shut down for settled sludge removal. The program determines the total required lagoon volume by assuming that 60 percent of the lagoon is filled with sludge at the time it is shut down. Construction and annual operation and maintenance costs are determined based on the lagoon volume and the number of lagoons required.

If sand drying beds are used as a dewatering method, the user has to enter the following input data:

1. The percent volume loss of the sludge to the underdrains and decant system.
2. The initial sludge depth applied to the drying beds.
3. The rate of evaporation expressed in inches per month.

With these data the computer calculates the final sludge depth, the time the sludge needs to remain on the beds and the total required surface area. Construction and annual operation and maintenance costs are then calculated based on the required drying bed surface area.

Finally, for both lagoons and sand drying beds, a summary table is printed on the screen which displays inflow sludge characteristics, sizes of lagoons or sand drying beds, annual sludge production and associated costs.

Also included in this summary are the costs for the sludge preconditioning equipment (if selected). These costs and type of conditioning chemical are determined in a separate subroutine as described in the next section.

5.2.7. Sludge Conditioning

The sludge conditioning sequence is a subroutine to the sludge dewatering process and is commonly used to increase final solids concentrations. It is noted that conditioning chemicals are generally only added to alum sludges and not to lime sludges.

The computer allows the user to select either polymer or lime as a conditioning chemical. Next, the required chemical feed rate needs to be entered in lbs per ton of sludge as well as the chemical cost by weight. Based on the chemical feed rate, the construction and annual operation and maintenance costs are calculated. The chemical cost is evaluated on an annual basis and added to the annual feed equipment O & M cost. As mentioned earlier, the associated costs for the conditioning equipment is included in the summary table of the selected sludge dewatering equipment.

5.2.8. Final Disposal

Even though there are several methods to permanently dispose of water treatment sludge, the most commonly acceptable method is disposal in a sanitary landfill. Prior to calculating annual operational cost for hauling dewatered sludge to a landfill, the final sludge characteristics and the annual tonnage (wet weight) is displayed on the screen. If for sludge dewatering method 3 was selected (using both mechanical and non-mechanical dewatering), the sludge characteristics resulting from both methods is displayed.

If a user wants to determine the cost for hauling and disposing sludge in a landfill, the round trip mileage to the landfill and the tipping fee per ton (delivered wet weight) need to be entered as data. In some cases a utility is not charged for a tipping fee because their own municipal landfill is used. In this case, zero could be entered for a tipping fee. Disposal costs are determined by assuming $0.18 per mile per wet ton of sludge for hauling, plus $2.40 per ton of sludge for loading and unloading, plus the tipping per ton at the landfill.

5.3. EXAMPLES

The following are several example simulations performed by the model. No attempt has been made to provide an example of every possible combination, since the list would be endless. Instead, these examples allow a first time user to gain familiarity with the model and eventually become comfortable enough to perform some of their own, more unique, simulations.

Example 1:

It is desired to simulate a sludge handling system consisting of the following:

- Coagulation Treatment Process
- Mechanically Cleaned Sedimentation Basins
- Unthickened and Thickened Sludge Pump Stations
- Gravity Thickener
- Mechanical Dewatering with Filter Press
- Sludge Disposal in Sanitary Landfill

The required input data are as listed below, followed by the results.

- Treatment Process (Coagulation) : 1
- Raw Turbidity : 10 ntu
- Turbidity Conversion Factor : 1.25
- Sludge Percent Solids : 0.85
- Chemicals - Alum : 12 mg/l
  - Polymer : 1.0
- Plant Flow : 15 mgd
- Sed. Basin Surface Area :35,000 sf
- Applied Turbidity : 2
- Number of Filter : 4
- Filter Loading Rate : 4 gpm/sf

- Filter Run Length : 24 hours
- Backwash Recycle : "y"
- Basin Cleaning Procedure : "b"
- Unthickened Sludge Pump Station : "y"
- Gravity Thickener : "y"
- Adjust Sludge Flow Rate : "y"
- Days Per Week Pumping : 5
- Hours Per Day Pumping : 8
- Method to Determine Thickener Size: 2
- Sludge Loading Rate : 0.12 lbs/sf/hr
- Underflow Solids Concentration : 4%
- Thickened Sludge Pump Station : "y"
- Sludge Dewater Method : 1
- Mechanical Dewatering Method : 2
- Resulting % Solids : 40%
- Sludge Pre-Conditioning : "y"
- Squeeze Time : 18 minutes
- Process Yield : 0.40 lb/hr/sf
- Filtration Time : 100 minutes
- Cleaning Time : 15 minutes
- Pre-Conditioning Chemical : lime (2)
- Chemical Feed Rate : 150 lbs/ton
- Chemical Cost Per Ton of Chemical : $75
- Landfill disposal : "y"
- Round Trip Miles to Landfill : 50
- Tipping Fee Per Ton : 10

A hard copy of all the results are presented in Table 5-1.

Example 2:

In this example, it is desired to analyze whether drying beds are more economical to use than the filter press in example 1. To simulate this, we use the data we entered for example 1 is used.

TABLE 5-1

```
                    ******************************
                    *                            *
                    *  SLUDGE MANAGEMENT RESULTS *
                    *                            *
                    ******************************



                         SUMMARY OF SLUDGE HANDLING
                         --------------------------

                  inflow conditions                    outflow conditions
               sludge     sludge     solids           sludge     sludge     solids
   item      (lbs/day)  (gal/day)    ( % )          (lbs/day)  (gal/day)    ( % )
-------------------------------------------------------------------------------------
SED. BASINS                                           2474.5      34906     0.850
UNTHICK. P.S.   2474.5      34906    0.850            2474.5      34906     0.850
THICKENER       2474.5      34906    0.850           10392.8      31153     4.000
THICKENED P.S. 10392.8      31153    4.000           10392.8      31153     4.000
FILTER PRESS   10392.8      31154    4.000           10392.8       3115    40.000

* NOTES : 1.. Sedimentation basins are cleaned mechanically
          2.. Sludge conditioning is used prior to mechanical dewatering



                   SUMMARY CONSTRUCTION AND O & M COST
                   -----------------------------------

                         construction     annual o & m     number of
         item               cost              cost            units
        -------------------------------------------------------------------

        UNTHICK.SLUDGE P.S.    26351             4742           1
        GRAVITY THICKENER     132862              965           1
        THICK. SLUDGE P.S.     50469             4162           1
        FILTER PRESS         2214866            64822           1
        CONDITIONING           63635             5471.          1
        LANDFILL DISPOSAL                       24187
                           =========          =======
             TOTAL           2488183           104348



================================================================================
```

TABLE 5-1 (con't)

```
                         WATER TREATMENT PROCESS


Plant data :
     Plant flow                              =        15   mgd
     Sed. basin surface area                 =     35000   sq.ft
     Number of filters                       =         4
     Filter loading rate                     =       4.0   gpm/sf
     Filter run time                         =        24   hours
     Backwash recycle                        =     542.5   gal/min

Raw water data :
     Raw turbidity                           =      10.0   ntu
     Conversion factor                       =         1
     Raw suspended solids                    =      12.5   mg/l

Treatment :
     Alum dosage                             =      12.0   mg/l
     Polymer dosage                          =       1.0   mg/l
     Carbon dosage                           =       0.0   mg/l
     Applied turbidity                       =      2.00   ntu

Sludge production :
     Sludge production                       =    2474.5   lbs/day
     Sludge production                       =     34906   gal/day
     Sludge percent solids                   =     0.850   %

     Backwash solids                         =    250.20   lbs/day
     Backwash flow                           =    781245   gal/day


===========================================================================


                      SED. BASIN CLEANING PROCEDURE


Basins are cleaned mechanically

     Sludge dry solids rate                  =    2474.5   lbs/day
     Sludge flow rate                        =     34906   gal/day
     Sludge percent solids                   =     0.850   %

      Note : Sludge removal occurs  24  hours per day,  7  days per week


===========================================================================
```

TABLE 5-1 (con't)

```
                         UNTHICKENED SLUDGE PUMP STATION


      Pump station capacity                 =      24.2   gal/min
      Total construction cost               = $  26351
      Total annual oper. & maint. cost      = $   4742


=============================================================================


                              GRAVITY THICKENER


    Inflow conditions :
         Sludge dry solids rate             =   2474.5  lbs/day
         Sludge flow rate                   =    34906  gal/day
         Sludge percent solids              =    0.850  %

    Outflow conditions :
         Sludge dry solids rate             =  10392.8  lbs/day
         Sludge flow rate                   =    31153  gal/day
         Sludge percent solids              =    4.000  %

    Thickener design :
         Diameter                           =     34      ft
         Number of thickeners               =     1
         Sludge loading rate                =    .12       lbs/sf-hr

    Economic analysis :
         Total construction cost            = $ 132862
         Total annual oper. & maint. cost   = $    965

         Note :  Thickener outflow is adjusted to  8  hours per day  5
                  days per week.


=============================================================================


                          THICKENED SLUDGE PUMP STATION


         Pump station capacity              =  64.9  gal/min
         Number of pump stations            =     1

         Total construction cost            = $  50469
         Total annual oper. & maint. cost   = $   4162
```

TABLE 5-1 (con't)

```
                          MECHANICAL DEWATERING


  Sludge feed conditions :
       Sludge dry solids rate                  =  10392.8 lbs/day
       Sludge flow rate                        =    31154 gal/day
       Sludge percent solids                   =    4.000 %

  Final sludge conditions :
       Sludge volume per operating day         =       99 cu.ft
       Sludge percent solids                   =   40.000 %

  Sludge dewatering equipment :
       Type of equipment                       =   filter press
       Sqeeze time                             =     18.0  minutes
       Process yield                           =     0.40  lb/hr/sq.ft.
       Actual press yield                      =     0.35  lb/hr/sq.ft.
       Filter press size                       =   3660.6  sq.ft.
       Number of units                         =   1
       Annual sludge production (dry wt.)      =   452.09  tons
       Total construction cost                 = $2214866
       Total annual oper. &  maint.  cost      = $  64822

  Sludge conditioning equipment:
       Conditioning chemical                   =   lime
       Chemical feed rate                      =   150        lbs/ton
       Chemical cost per ton                   = $ 75
       Total construction cost                 = $    63635
       Total annual oper. & maint. cost        = $     5471


========================================================================


                             FINAL SLUDGE DISPOSAL


  Sanitary landfill disposal :
       Annual sludge production ( wet weight )    =     1130.22  ton
       Sludge percent solids                      =      40.000  %
       Equivalent annual sludge volume            =       36235  cu.ft

       Round trip miles to landfill               =        50
       Tipping fee at landfill per ton            = $      10
       Total annual disposal cost                 = $ 24187


========================================================================
```

After the results of example 1 are printed out, we return to the main program and proceed until the dewatering results appear on the screen. Now enter "b" (which places us at the beginning of the dewatering sequence).

Next select "2" (non-mechanical dewatering) followed by "2" (sand drying beds). Now the following data are requested:

- Resulting Sludge % Solids : 25%
- Sludge Conditioning : "y"
- Percent Volume Loss to Underdrains : 50%
- Depth of Sludge Applied to Beds : 18 inches
- Rate of Evaporation : 3 inches/month
- Conditioning Chemical (Polymer) : 1
- Chemical Feed Rate : 4 lbs/ton
- Chemical Cost per Ton of Chemical : $2,000
- Sludge Landfill Disposal : "y"
- Round Trip Miles : 50 miles
- Tipping Fee :$10/ton

The results of example 2 are presented in Table 5-2.

<u>Example 3:</u>

A small utility wants to upgrade their 12 mgd water treatment plant by adding an equalization basin, a gravity thickener with associated pump stations and a dewatering facility with belt filter presses.

Previously, the three sedimentation basins were manually cleaned after 30 days. Sludge production rates are estimated to be 1,900 lbs/day at a 0.8% solids concentration.

To determine the anticipated capital and O & M costs, the user should start off with selecting Option 3 of the water treatment process (sludge quantity input) and enter the following data:

TABLE 5-2

```
                    ******************************
                    *                            *
                    *  SLUDGE MANAGEMENT RESULTS *
                    *                            *
                    ******************************



                      SUMMARY OF SLUDGE HANDLING
                      --------------------------

                 inflow conditions                  outflow conditions
               sludge     sludge     solids       sludge     sludge     solids
    item     (lbs/day)  (gal/day)    ( % )      (lbs/day)  (gal/day)    ( % )
------------------------------------------------------------------------------
SED. BASINS                                       2474.5      34906     0.850
UNTHICK. P.S.    2474.5      34906    0.850       2474.5      34906     0.850
THICKENER        2474.5      34906    0.850      10392.8      31153     4.000
THICKENED P.S.  10392.8      31153    4.000      10392.8      31153     4.000
DRYING BEDS     10392.8      31154    4.000      10392.8       4985    25.000

* NOTES : 1.. Sedimentation basins are cleaned mechanically
          2.. Sludge conditioning is used prior to non-mechanical dewatering



                 SUMMARY CONSTRUCTION AND O & M COST
                 -----------------------------------


                         construction     annual o & m     number of
         item                cost             cost            units
        --------------------------------------------------------------

        UNTHICK.SLUDGE P.S.    26351            4742            1
        GRAVITY THICKENER     132862             965            1
        THICK. SLUDGE P.S.     50469            4162            1
        SAND DRYING BEDS      650591           70095            1
        CONDITIONING           36197            7660            1
        LANDFILL DISPOSAL                      38699
                             =========        =======
             TOTAL            896470          126323



===============================================================================
```

TABLE 5-2 (con't)

```
                         WATER TREATMENT PROCESS


 Plant data :
      Plant flow                              =        15  mgd
      Sed. basin surface area                 =     35000  sq.ft
      Number of filters                       =         4
      Filter loading rate                     =       4.0  gpm/sf
      Filter run time                         =        24  hours
      Backwash recycle                        =     542.5  gal/min

 Raw water data :
      Raw turbidity                           =      10.0  ntu
      Conversion factor                       =         1
      Raw suspended solids                    =      12.5  mg/l

 Treatment :
      Alum dosage                             =      12.0  mg/l
      Polymer dosage                          =       1.0  mg/l
      Carbon dosage                           =       0.0  mg/l
      Applied turbidity                       =      2.00  ntu

 Sludge production :
      Sludge production                       =    2474.5  lbs/day
      Sludge production                       =     34906  gal/day
      Sludge percent solids                   =     0.850  %

      Backwash solids                         =    250.20  lbs/day
      Backwash flow                           =    781245  gal/day


=============================================================================


                         SED. BASIN CLEANING PROCEDURE


 Basins are cleaned mechanically

      Sludge dry solids rate                  =    2474.5  lbs/day
      Sludge flow rate                        =     34906  gal/day
      Sludge percent solids                   =     0.850  %

       Note : Sludge removal occurs  24  hours per day,  7  days per week


=============================================================================
```

TABLE 5-2 (con't)

```
                    UNTHICKENED SLUDGE PUMP STATION


   Pump station capacity                  =      24.2   gal/min
   Total construction cost                = $  26351
   Total annual oper. & maint. cost       = $   4742


==========================================================================


                         GRAVITY THICKENER


 Inflow conditions :
      Sludge dry solids rate             =   2474.5  lbs/day
      Sludge flow rate                   =    34906  gal/day
      Sludge percent solids              =    0.850  %

 Outflow conditions :
      Sludge dry solids rate             =  10392.8  lbs/day
      Sludge flow rate                   =    31153  gal/day
      Sludge percent solids              =    4.000  %

 Thickener design :
      Diameter                           =    34     ft
      Number of thickeners               =    1
      Sludge loading rate                =   .12      lbs/sf-hr

 Economic analysis :
      Total construction cost            = $ 132862
      Total annual oper. & maint. cost   = $    965

      Note :  Thickener outflow is adjusted to  8  hours per day  5
               days per week.


==========================================================================


                    THICKENED SLUDGE PUMP STATION


   Pump station capacity                 =  64.9  gal/min
   Number of pump stations               =     1

   Total construction cost               = $  50469
   Total annual oper. & maint. cost      = $   4162

==================================        ================================
```

TABLE 5-2 (con't)

```
                        NON-MECHANICAL DEWATERING



Sludge feed conditions :
     Sludge dry solids rate                 =  10392.8  lbs/day
     Sludge flow rate                       =    31154  gal/day
     Sludge percent solids                  =    4.000  %

Resulting sludge characteristics :
     Annual sludge production (dry wt.)     =   452.09  tons
     Sludge percent solids                  =   25.000  %

Sludge dewatering :
     Dewatering method                      =   sand drying beds
     Required sludge drying time            =      61   days
     Total surface area required            =   40493   sq.ft
     Total construction cost                = $ 650591
     Total annual oper. & maint. cost       = $ 70095

Sludge conditioning equipment :
     Conditioning chemical                  =   polymer
     Chemical feed rate                     =   4       lbs/ton
     Chemical cost per ton                  = $ 2000
     Total construction cost                = $  36197
     Total annual oper. & maint. cost       = $   7660


==========================================================================


                            FINAL SLUDGE DISPOSAL



Sanitary landfill disposal :
     Annual sludge production ( wet weight )    =    1808.35  ton
     Sludge percent solids                      =     25.000  %
     Equivalent annual sludge volume            =      57976  cu.ft

     Round trip miles to landfill               =        50
     Tipping fee at landfill per ton            = $      10
     Total annual disposal cost                 = $   38699


==========================================================================
```

- Plant Capacity : 12 mgd
- Sedimentation Basin Surface Area : 20,000 sf
- Sludge Production : 1,900 lbs/day
- Sludge Percent Solids : 0.8%

Next, the user proceeds to batch removal sequence and enters data as follows:

- Depth of Sedimentation Basins : 15
- Number of Sedimentation Basins : 3
- Number of Operating Days before Cleaning Basins : 30
- Number of Days Required to Clean a Basin : 1
- Number of Hours Required Per Day Per Basin : 12
- Do you want a Sludge Transfer Pump Station : "y"
- Sludge Percent Solids Leaving Equalization Basin : 1.25%
- Do you want to Control Equalization Basin Outflow : "y"
- Days Per Week Pumping Sludge From Basin : 5
- Hours Per Day Pumping Sludge From Basin : 12

Next, the unthickened sludge pump station, the thickener (assume sludge loading of 0.3 lb/sf/hr and 2% underflow) and the thickened pump station are entered.

Final sludge dewatering is accomplished in a mechanical process using a belt filter press with an assumed loading rate of 100 lb/hr/f producing a sludge with a 15% solids concentration. The sludge is preconditioned with polymer at a rate of 5 lbs/ton.

The final sludge cake is disposed of in the nearby landfill. A round trip is 50 miles and the tipping fee is estimated at $30/ton.

Using the sludge management model, the results are obtained as shown in Table 5-3.

TABLE 5-3

```
                         *******************************
                         *                             *
                         *  SLUDGE MANAGEMENT RESULTS  *
                         *                             *
                         *******************************



                              SUMMARY OF SLUDGE HANDLING
                              --------------------------

                     inflow conditions                    outflow conditions
                 sludge     sludge     solids          sludge     sludge     solids
    item       (lbs/day)  (gal/day)    ( % )         (lbs/day)  (gal/day)    ( % )
-----------------------------------------------------------------------------------
SED. BASINS                                           38000.0     569544    0.800
TRANSFER STA.    38000.0     569544    0.800          38000.0     569544    0.800
EQUAL. BASIN     38000.0     569544    0.800           5320.0      51031    1.250
UNTHICK. P.S.     5320.0      51031    1.250           5320.0      51031    1.250
THICKENER         5320.0      51031    1.250           5320.0      31894    2.000
THICKENED P.S.    5320.0      31894    2.000           5320.0      31894    2.000
BELT PRESS        5320.0      31894    2.000           5320.0       4253   15.000

* NOTES : 1.. Sedimentation basins are cleaned manually
          2.. Sludge conditioning is used prior to mechanical dewatering

                        SUMMARY CONSTRUCTION AND O & M COST
                        -----------------------------------


                           construction    annual o & m    number of
            item               cost            cost          units
      --------------------------------------------------------------------

      TRANSFER PUMP STA.       71664           1779            1
      EQUALIZATION BASIN      436630           2690            1
      UNTHICK.SLUDGE P.S.      35981           4302            1
      GRAVITY THICKENER       171705           1561            1
      THICK. SLUDGE P.S.       47774           5011            1
      BELT FILTER PRESS       327740          21354            1
      CONDITIONING             35193           7556            1
      LANDFILL DISPOSAL                       95808
                             =========       =======
           TOTAL              1126687         140061



==========================================================================
```

TABLE 5-3 (con't)

```
                        INITIAL SLUDGE CHARACTERISTICS


      Plant flow                              =          12  mgd
      Sed. basin surface area                 =       20000  sf
      Sludge production                       =        1900  lbs/day
      Sludge production                       =       28477  gal/day
      Sludge percent solids                   =       0.800  %


=============================================================================


                        SED. BASIN CLEANING PROCEDURE


  Basins are cleaned manually through batch process :

  Sedimentation basins :
       Depth of sed. basins                      =   15          ft.
       Number of sed. basins                     =   3
       Number of operating days of basins        =   30
       Sludge dry solids flow rate out           =   38000.0   lbs/day
       Sludge flow rate out                      =    569544   gal/day

  Sludge transfer station   :
       Number of pump stations                   =   1
       Pump station capacity                     =      791.0  gal/min
       Construction cost                         = $    71664
       Operation and maintenance cost            = $     1779

  Equalization basin :
       Number of basins                          =   1
       Equalization basin diameter               =   74          ft.
       Construction cost                         = $   436630
       Operation and maintenance cost            = $     2690

  Equalization basin outflow :
       Sludge dry solids flow rate               =       5320  lbs/day
       Sludge flow rate                          =      51031  gal/day
       Sludge percent solids                     =      1.250  %

        Note : Sludge removal from equalization basin occurs  12
               hours per day,  5  days per week


=============================================================================
```

TABLE 5-3 (con't)

```
                          UNTHICKENED SLUDGE PUMP STATION


      Pump station capacity                  =      70.9   gal/min
      Total construction cost                = $   35981
      Total annual oper. & maint. cost       = $    4302


==========================================================================


                             GRAVITY THICKENER


  Inflow conditions :
       Sludge dry solids rate                =   5320.0  lbs/day
       Sludge flow rate                      =    51031  gal/day
       Sludge percent solids                 =    1.250  %

  Outflow conditions :
       Sludge dry solids rate                =   5320.0  lbs/day
       Sludge flow rate                      =    31894  gal/day
       Sludge percent solids                 =    2.000  %

  Thickener design :
       Diameter                              =    44      ft
       Number of thickeners                  =    1
       Sludge loading rate                   =   .3       lbs/sf-hr

  Economic analysis :
       Total construction cost               = $ 171705
       Total annual oper. & maint. cost      = $    1561


==========================================================================


                          THICKENED SLUDGE PUMP STATION


      Pump station capacity                  =  44.3  gal/min
      Number of pump stations                =     1

      Total construction cost                = $  47774
      Total annual oper. & maint. cost       = $   5011


==========================================================================
```

TABLE 5-3 (con't)

```
                              MECHANICAL DEWATERING


    Sludge feed conditions :
         Sludge dry solids rate                    =   5320.0 lbs/day
         Sludge flow rate                          =    31894 gal/day
         Sludge percent solids                     =    2.000 %

    Final sludge conditions :
         Sludge volume per operating day           =      203 cu.ft
         Sludge percent solids                     =   15.000 %

    Sludge dewatering equipment :
         Type of equipment                         =   belt filter press
         Sludge loading rate                       =    100.0  lb/ft/hr
         Belt filter press size                    =     1.00  meter
         Number of units                           =    1
         Annual sludge production (dry wt.)        =   347.13  tons
         Total construction cost                   = $ 327740
         Total annual oper. &  maint.  cost        = $  21354

    Sludge conditioning equipment:
         Conditioning chemical                     =   polymer
         Chemical feed rate                        =   5       lbs/ton
         Chemical cost per ton                     = $ 2000
         Total construction cost                   = $    35193
         Total annual oper. & maint. cost          = $     7556


=============================================================================


                                   FINAL SLUDGE DISPOSAL


    Sanitary landfill disposal :
         Annual sludge production ( wet weight )       =     2314.20  ton
         Sludge percent solids                         =      15.000  %
         Equivalent annual sludge volume               =       74193  cu.ft

         Round trip miles to landfill                  =          50
         Tipping fee at landfill per ton               = $        30
         Total annual disposal cost                    = $ 95808


=============================================================================
```

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APPENDIX A

DEVELOPMENT OF COST CURVES

COST CURVES

Cost curves for the primary processes associated with a water sludge management system are presented in Chapter 4 of the Handbook. The data points for the cost curves were generated from Tables included at the end of this Appendix. The format used in developing the cost curves is similar to that used in "Estimating Water Treatment Costs - Volume 2 Cost Curves Applicable to 1 to 20 mgd Treatment Plants" by the consulting firm of Culp/Wesner/Culp for and published by the U.S. Environmental Protection Agency (EPA-600/2-79-1626). The costs generated herein were generated independent of the costs in the EPA manual. No comparison of costs between this Manual or the EPA manual has been attempted or was intended.

Each unit operation has been costed independently with the intent of being able to "mix and match" various components to generate the most cost effective/workable sludge management system. In some cases, the process stands alone and its costs are not influenced by other accompanying processes, such as gravity thickeners. However in other cases, the process performance can be improved if a process is enhanced with the addition of another component: for example, vacuum filters enhanced by a lime (precoat) system.

Sludge pumping costs are generated for use in transporting the sludge from one process to the other when needed. For instance, unthickened sludge pumping represents the cost in pumping the sludge to the thickener. Thickened sludge pumping costs are for transporting the post-thickener sludges. Some process pumping requirements are included in the individual process costs: such as filter press pumping of feed sludge and filtrate.

Hauling costs are not included in the operation and maintenance cost for the removal of dried sludge cake from the mechanical dewatering process. Mechanical dewatering methods assume cake discharge from the second story of the sludge building directly to a receiving truck below. Separate hauling costs may be added to the selected sludge systems if desired.

The total cost for each sludge management process is presented in two cost curves: a construction curve and an operation and maintenance curve. The construction cost curve represents the initial capital expense which would be incurred for the sludge management process. The operation and maintenance curve represents the annual expense which would be incurred for the operation and maintenance of the sludge management process.

CONSTRUCTION COST CURVES

The purpose of the construction cost curves is to allow for comparison of the initial capital costs for the alternative sludge management systems. Construction cost curves are given in Chapter 4 for each sludge management process discussed. Included in the curves are the costs for the equipment, interconnecting piping and valves, the installation labor and housing costs.

Equipment costs were derived from manufacturers' cost estimates based on actual and conceptual designs, and published data. A 10 percent contingency was applied to manufacturers' furnished equipment costs to cover specialties and extras normally encountered.

Piping and valving costs were generated to represent the expense for furnishing and installing the piping and valving required to connect the process equipment. The piping and valving was laid out to extend outside the housing limits of the process equipment. The piping and valving costs were derived from manufacturer's recommendations, as a percent of equipment costs and from experience of the authors. The piping and valving costs include installation.

The installation labor costs include the total cost for installation of the process equipment. Installation of the other items listed, such as building costs and piping and valving are included under that item, unless stated otherwise. The installation labor costs were based on manufacturers' recommendations, actual installations, hourly estimates for conceptual construction crews and/or as a percent of the equipment cost.

The building costs represent the construction cost for housing the process equipment. Building costs were assumed as $75/square foot of 1st floor building and $50/square foot for a 2nd floor. The square footage requirements were based on manufacturers' recommendations. No special foundation systems or pilings were included.

The subtotals of the construction costs were increased by a factor of 20 percent to account for Contractor's overhead and profit.

OPERATION AND MAINTENANCE COST CURVES

The purpose of the operation and maintenance cost curves is to provide comparisons on the annual operating and maintenance costs of the alternative sludge management systems. An equitable comparison of alternative systems can be made by summing the present worth values of the operating and maintenance costs with the construction costs. Included in the operating and maintenance costs are the annual costs for the process power, the maintenance materials and labor.

The process power costs represent the annual electrical power cost for the process equipment. The assumption for run time per day for the process equipment is stated for each process (usually assumed as 16 hours per day). The unit cost for energy was calculated as 7 cents per kilowatt-hour. Electric motor energy usage was assumed to be the same as the motor rating. No building energy costs were included.

The maintenance material costs represent the annual costs for the material required to maintain the process equipment. The maintenance materials costs were based on manufacturers' estimates and percent of equipment costs.

The operation and maintenance labor costs represent the total annual labor cost to operate and maintain the process equipment. The labor rate was assumed at $16.50 per hour, which includes the employee salary and benefits. The labor costs were derived from hourly estimates made for daily operation and maintenance of the equipment, experience and manufacturers' recommendations.

Operating and maintenance costs do not include the energy costs associated with the building itself. That is, the costs for heating, cooling, ventilating and lighting the building are not included. These costs were not included because they are so variable depending on the part of the country, climate, hours of operation and the owner's preference. No chemical costs are included in any of the O & M curves.

UPDATING THE COST CURVES

The equipment costs used to generate the cost curves in this handbook are for June, 1986. There are numerous methods available to assist in updating costs from one time period to another.

A large majority of cost updating in the past has been achieved using cost indices. The Engineering News Record (ENR) Construction Cost Index (CCI) and the Building Cost Index have been extensively used to update construction and building costs. To use either of these, the index value is obtained for the period in question (generally today) and divided by the index for the period for which the prices are known. This factor is then applied to the building or construction cost to be updated and the new cost is established.

Another approach to cost updating is to break down the costs into components for updating by individual category. In addition to the ENR indices, the Bureau of Labor Statistics (BLS) publishes a wide range of indices relative to many of the cost categories. The BLS have maintained the Consumer Indices (BLS designation below) and more recently the Standard Industrial Classification (SIC) Index. The construction costs presented in this Handbook were developed in such a way that a categorical updating procedure can be applied. Some of the indices which may be of interest along with the category which they can be applied follow:

Cost Component		June 1986 Index Value
Manufactured Equipment	BLS General Purpose Machinery and Equipment - Code 114	325.6
Labor	ENR Wage Index (Skilled Labor)	3863.06
Excavation & Sitework	ENR Wage Index (Skilled Labor)	3863.06
Concrete	BLS Concrete Ingredients - Code 132	339.3
Electrical & Instrumentation	BLS Electrical Machinery and Equipment - Code 117	256.9
Pipe & Valves	SIC 34943 - Metal Valves for Piping Systems and Equipment, Except Plumbing and Heating Valves	104.2
Housing	ENR Building Cost Index	2492.67

The indices given above have been selected for general presentation purposes. Other indices may be chosen at the discretion of the user. It is recognized that the SIC Indices are sometimes more specific and are becoming more popular. Substitution of certain SIC indices or a different categorical breakdown may be beneficial for certain applications.

Updating of the Operation and Maintenance Cost Curves can be done by updating the unit price items used for their development. By factoring the actual rate per kilowatt-hours of power to the \$.07/kwh assumed, the annual energy cost can be updated. Maintenance material costs can be updated using the Producer Price Index for Finished Goods over the current index of 289.3. Labor can be updated based on wage rates factored over the \$16.50/hour used.

MODIFYING THE COST CURVES FOR A SPECIFIC INSTALLATION

There are many scenarios which can be developed for use of and modification to the given cost curves. The cost curves, as prepared, can provide a means of comparing alternative sludge management systems on a comparable basis. That is, similar assumptions went into each process for system layout, operation and maintenance. The generation of a total cost to represent a particular system may require modification to the cost curves based on a different set of assumptions.

Modifications which could be required include the number of shifts of operation (16 hours, 2 shifts assumed). Also the chemical cost of polymer or lime conditioning usage needs to be added to the cost. The cost of heating, ventilation, cooling and lighting the sludge building in a particular system may be added. Also some cost additions associated with the handling and storage of dewatered sludge may be needed to represent the cost at a particular installation.

It is obvious that not all assumptions made in this Handbook will meet the needs of each system. Some modification to the curves is expected. However, it is anticipated the cost curves as presented will amply serve the task of comparison of the sludge management processes presented.

COST CURVE VERIFICATION

Example facilities were used to verify many of the cost curves presented in this Handbook. One such example facility is the City of Norfolk, Virginia Solids Collecting and Handling Facility at the Moores Bridges Water Purification Plant. The major capital items associated with the solids collecting and handling facilities included the following:

a. 2 separate solid transfer pump stations to transport unthickened sludge from the sedimentation basins to the new thickeners (200 gpm each)
b. two new Thickener Tanks (3,318 ft^2 each - 14' S.W. depth)
c. thickened sludge pump station (200 gpm each)
d. polymer feed system (128 gph)
e. the Dewatering Facility including the building, the diaphragm filter press (150 chambers, 1,560 mm^2 each plate, 6,016 sf total), feed pumps, wash system and appurtenances

The project construction began in October, 1985 and was completed in January, 1987. The awarded contract amount was 5.6 million dollars. Not all of the items of work however, could be applied to the cost curves presented in this Handbook. The buildings and tanks were placed on piles due to unsuitable subsurface conditions, which is an additional cost to these curves presented herein. Also, the work in the sedimentation basin is special to this project. The dewatering building contained an elevator, bathrooms, a lunch room, control room and a large storage room which are extra costs. The comparison between Norfolk's facility costs and this Handbook's cost curves was made possible when the City furnished a copy of the schedule of values on the Contractors Pay Requisition printout. The summary of that comparison is presented below:

Item	Norfolk Bid	Handbook Values	Figure within Handbook
Sludge Transfer Pumps	$ 46,600	$ 49,000	4-17
Thickeners	$ 600,200	$ 653,700	4-9
Thickened Sludge Pumps	$ 43,000	$ 65,000	4-19
Polymer Feed System	$ 112,500	$ 120,000	4-13
Filter Press Facility	$3,335,900	$3,447,500	4-34
TOTAL	$4,168,200	$4,335,200	

TABLE A-1

CONSTRUCTION COST FOR GRAVITY THICKENERS

Diameter (ft)	10	20	50	4-50's
Surface Area (ft^2)	79	314	1963	7852
Eqpt., Elec. & Instr.	33,000	38,300	81,400	325,600
Concrete	6,380	13,940	46,380	185,500
Excavation	1,240	2,480	6,190	24,950
Labor	6,600	7,660	16,280	85,120
Piping/Valves	11,000	13,200	16,500	49,500
Subtotal	$58,220	$75,580	$166,750	$670,670
(+20% Contr. OH&P)	11,640	15,120	33,350	134,130
Total	$69,860	$90,700	$200,100	$804,800

TABLE A-2

OPERATION AND MAINTENANCE COST FOR GRAVITY THICKENERS

Diameter (ft) Surface Area (ft^2)	10 79	20 314	50 1963	4.50's 7852
Power	340	460	690	2,740
Maint. Materials	140	260	460	990
Labor	150	290	410	640
Total - Annual O&M Cost	$630	$1,010	$1,560	$4,370

TABLE A-3

CONSTRUCTION COST FOR DRY POLYMER SYSTEM

Polymer Feed Rate (#/Hr)

	2.2	4.4	8.3	18.2	32.9	66.7
Equipment	19,400	22,550	26,700	32,100	39,600	45,230
Labor	4,270	4,850	5,600	8,580	6,120	9,050
Pipe & Valves	1,900	2,140	2,400	2,730	2,950	3,390
Elec. & Instr.	1,970	1,970	1,970	2,170	2,540	2,900
Housing	6,200	8,130	9,720	11,250	18,200	24,000
Subtotal	$33,740	$39,640	$46,390	$54,830	$71,410	$ 84,570
(+20% Contr. OH&P)	6,750	7,930	9,280	10,970	14,280	16,910
Total	$40,490	$47,570	$55,670	$65,800	$85,690	$101,480

TABLE A-4

OPERATION AND MAINTENANCE COST FOR DRY POLYMER SYSTEM

	Polymer Feed Rate (#/Hr)					
	2.2	4.4	8.3	18.2	32.9	66.7
Power	$ 805	$ 805	$ 805	$1,380	$1,380	$ 1,610
Maint. Materials	420	450	690	890	1,190	1,450
Labor	4,760	4,950	5,382	6,100	7,130	7,690
Total O & M Cost for Dry Polymer System	$5,985	$6,205	$6,877	$8,370	$9,700	$10,750

TABLE A-5

CONSTRUCTION COST FOR LIME FEED SYSTEMS

	System Capacity (#/hr.)		
	10	50	100
Equipment	$26,400	$30,800	$36,850
Labor	5,300	5,850	6,820
Pipe & Valves	2,400	2,540	2,680
Elec. & Instr.	2,200	2,350	2,490
Housing	13,520	13,000	14,800
Subtotal	$49,200	$55,540	$63,640
(+20% contr. OH&P)	9,960	11,100	12,730
Total	$59,780	$66,640	$76,370

TABLE A-6

OPERATION AND MAINTENANCE COST FOR LIME FEED SYSTEMS

	System Capacity (#/hr.)		
	10	50	100
Power	$1,035	$1,035	$1,150
Maintenance Materials	160	240	310
Labor	3,800	3,920	4,165
Total Annual O&M Cost	$4,995	$5,195	$5,625

TABLE A-7

CONSTRUCTION COST FOR UNTHICKENED SLUDGE PUMPS

Pumping Rate (gpm)	50	250	1000	5000
Equipment*	$ 6,120	$ 9,640	$13,050	$ 33,070
Elec. & Instrumentation	3,060	4,430	5,480	12,570
Excavation $ Sitework	920	1,090	2,460	4,960
Concrete Wetwell & Slab**	1,970	4,810	9,490	23,680
Labor***	3,140	4,880	6,100	7,480
Pipe & Valves	4,120	10,200	17,600	28,130
Housing	8,250	8,250	11,880	36,390
Subtotal	$27,580	$43,300	$66,000	$146,280
+20% Contr. OH & P	5,520	8,660	13,210	29,260
Total	$33,100	$51,960	$79,270	$175,540

* Includes pump manufacturers standard controls.

** Installed Costs.

*** Equipment installation only, other items of installation included in cost of that item.

TABLE A-8

OPERATION AND MAINTENANCE COST FOR UNTHICKENED SLUDGE PUMPS

Pumping Rate (gpm)	50	250	1000	5000
Power*	$ 920	$1,530	$ 6,130	$18,400
Maintenance & Material	2,700	4,510	8,660	22,470
Labor	1,815	3,300	6,270	11,720
Total Annual O & M Cost	$5,435	$9,340	$21,060	$52,590

* Assume 1 pump operating 16 hours/day, 7 days/week.

TABLE A-9

CONSTRUCTION COST FOR THICKENED SLUDGE PUMPS

Pumping Rate (gpm)	5	25	50	200	500	1250
Pump/Motor Assembly	$ 4,540	$ 8,380	$ 9,990	$16,800	$31,790	$ 49,390
Elec. & Instrumentation	1,440	1,560	1,810	2,710	3,210	3,630
Excavation & Sitework	4,320	4,320	4,320	4,320	4,420	4,530
Concrete Dry Well & Slab	19,820	19,820	19,820	19,820	21,660	23,470
Labor	1,860	2,930	3,550	8,820	13,170	20,540
Pipe & Valves	580	700	940	1,370	1,470	2,270
Subtotal	$32,560	$37,710	$40,430	$53,840	$75,720	$103,830
+20% Contr. OH & P	6,510	7,540	8,090	10,770	15,140	20,770
	$39,070	$45,250	$48,520	$64,610	$91,460	$124,600

TABLE A-10

OPERATION AND MAINTENANCE COST FOR THICKENED SLUDGE PUMPS

Pumping Rate (gpm)	5	25	50	200	500	1250
Power*	$310	$1,530	$1,530	$3,070	$7,670	$15,330
Maintenance Materials	1,090	1,760	1,990	4,550	6,290	9,700
Labor	830	1,240	1,730	3,140	4,620	6,600
Total Annual O & M Cost	$2,230	$4,530	$5,250	$10,760	$18,580	$31,630

* Assume 1 pump operating 16 hours/day, 7 days/week.

TABLE A-11

CONSTRUCTION COST FOR DECANTER CENTRIFUGE

	Feed Sludge Flow Rate (gpm)			
	30	75	150	400
Equipment	143,000	176,000	330,000	440,000
Installation Labor	28,600	35,200	66,000	88,000
Piping/Valves	15,730	20,240	39,600	57,200
Elec./Instr.	11,400	12,300	14,850	17,600
Housing	30,000	43,500	67,500	74,300
Subtotal	$228,730	$287,240	$517,950	$677,100
(+20% Contr. OH&P)	45,750	57,450	103,590	135,420
Total	$274,480	$344,690	$621,540	$812,520

TABLE A-12

OPERATION AND MAINTENANCE COST FOR DECANTER CENTRIFUGE

	Feed Sludge Flow Rate (gpm)			
	30	75	150	400
Power*	$ 9,600	$12,580	$25,160	$ 51,460
Maintenance Materials	4,400	6,180	8,760	17,520
Labor	18,070	30,940	33,300	52,800
Total Annual O&M Cost	$32,070	$49,700	$67,220	$121,780

* Based on operating 16 hours/day, 7 days/week.

TABLE A-13

CONSTRUCTION COST FOR DIAPHRAGM FILTER PRESS

Area, Ft^2	802	2006	4814	9629	24,072
*Volume, Ft^3	83.6	209	502	1003	2,508
Filter Press	$ 358,000	$ 495,000	$ 891,000	$1,760,000	$ 4,235,000
Washer/Shaker Mechanism	93,000	99,000	104,500	209,000	418,000
Ancillary Equip.	220,000	247,500	258,500	330,000	495,000
Labor	225,000	230,000	320,000	580,000	1,250,000
Pipe & Valves	140,000	150,000	160,000	225,000	325,000
Elec. & Instr.	160,000	160,000	160,000	225,000	350,000
Housing (2 story)	475,000	500,000	550,000	825,000	1,400,000
Subtotal	$1,671,000	$1,881,500	$2,444,000	$4,154,000	$ 8,473,000
(+20% Contr. OH&P)	334,200	376,300	488,800	830,800	1,694,600
Total	$2,005,200	$2,257,800	$2,932,800	$4,984,800	$10,167,600

* Pre-diaphragm volume

TABLE A-14

OPERATION AND MAINTENANCE COST FOR DIAPHRAGM FILTER PRESS

Area, Ft^2	802	2,006	4,814	9,629	24,072
Volume Ft^3	83.6	209	502	1,003	2,508
Power	$ 3,100	$ 7,400	$ 15,100	$ 29,800	$ 57,900
Maintenance Materials	3,890	5,800	11,600	23,100	55,100
Labor	$49,600	$66,000	$ 82,500	$ 99,000	$132,000
Total Annual O&M	$56,590	$79,200	$109,200	$151,900	$245,000

TABLE A-15

CONSTRUCTION COST FOR VACUUM FILTERS

Filter Area (Ft.^{2})	9.4	56.5	415	754
Vacuum Filter	$ 33,080	$ 44,000	$110,000	$132,000
Labor	16,600	22,500	56,100	61,400
Pipe & Valves	5,200	5,700	7,900	12,700
Elec. & Instr.	3,800	4,300	11,500	35,500
Housing	28,130	46,880	145,500	270,000
Subtotal	$ 86,810	$123,380	$331,000	$511,600
(+20% Contr. & OH&P)	17,360	24,680	66,200	102,320
Total	$104,170	$148,060	$397,200	$613,920

TABLE A-16

OPERATION AND MAINTENANCE COST FOR VACUUM FILTERS

Filter Area (Ft.^{2})	9.4	56.5	415	754
Power*	$ 9,200	$23,020	$ 30,660	$ 38,430
Maintenance Materials	4,400	16,100	73,000	119,700
Labor	13,200	31,350	107,250	181,500
Total	$26,800	$70,470	$210,910	$339,630

* Based on operating 16 hours/day, 7 days/week.

TABLE A-17

CONSTRUCTION COST FOR BELT FILTER PRESS

	Belt Width (meters)			
	1.0	1.5	2.0	6.0 (3 @ 2.0)
Equipment	104,500	126,500	143,000	429,000
Installation Labor	20,900	25,300	28,600	85,800
Pipe & Valves	16,500	20,600	27,500	66,000
Elec. & Instr.	20,000	22,000	25,000	60,000
Housing	84,100	90,100	96,100	240,300
Subtotal	$246,000	$284,500	$320,200	$ 881,100
+ 20% Contr. OH&P)	49.200	56,900	64,000	176,220
Total	$295,200	$341,400	$384,200	$1,057,300

TABLE A-18

OPERATION AND MAINTENANCE COST FOR BELT FILTER PRESS

	Belt Width (meters)			
	1.0	1.5	2.0	6.0 (3 @ 2.0)
Power	2,400	3,600	4,800	14,400
Maintenance Material	4,000	4,500	5,000	15,000
Labor	12,700	14,200	15,700	39,300
Total-Annual O $ M Cost	$19,100	$22,300	$25,500	$68,700

TABLE A-19

CONSTRUCTION COST FOR SAND DRYING BEDS

Area, Ft^2	2,000	6,300	12,600	25,000	50,000	100,000
Excavation	$ 4,840	$ 11,240	$ 22,490	$ 49,470	$ 89,950	$ 189,890
Concrete	20,270	51,750	97,100	187,760	369,050	731,830
Media	4,030	11,484	22,970	45,940	91,880	183,760
Pipe & Valves	12,710	24,470	38,820	61,400	112,580	214,390
Subtotal	41,850	98,880	181,380	344,570	663,460	1,319,870
+20% Contr. OH&P	$ 8,370	$ 19,780	$ 36,280	$ 68,910	$ 132,690	$ 263,970
Total	$50,220	$118,660	$217,660	$413,480	$ 796,150	$1,583,840

TABLE A-20

OPERATION AND MAINTENANCE COST FOR SAND DRYING BEDS

Area, Ft^2	2,000	6,300	12,600	25,000	50,000	100,000
Fuel	$ 120	$ 380	$ 760	$ 1,500	$ 3,000	$ 6,000
Maintenance Materials	490	1,550	3,100	6,160	12,320	24,640
Labor	$8,250	$20,600	$26,400	$41,250	$67,800	$110,060
Total O $ M Cost	$8,860	$22,530	$30,260	$48,910	$83,120	$140,700

TABLE A-21

CONSTRUCTION COST FOR SLUDGE DEWATERING LAGOONS

Storage Volume (ft.^{3})	40,000	400,000	4,000,000	8,000,000
Excavation & Sitework	$ 7,400	$ 74,050	$666,670	$1,333,330
Concrete Inlet & Outlet Structures	8,420	16,840	84,200	168,400
Pipe & Valves	3,580	8,940	59,620	111,780
Sub Total	$19,400	$ 99,830	$810,490	$1,613,510
+20% Contr. OH & P	3,880	19,970	162,100	322,700
Total	$23,280	$119,800	$972,590	$1,936,210

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