EPA 815-P-01-001
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TECHNOLOGIES AND COSTS FOR REMOVAL OF ARSENIC FROM DRINKING WATER

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STANDARDS AND RISK MANAGEMENT DIVISION
OFFICE OF GROUND WATER AND DRINKING WATER
UNITED STATES ENVIRONMENTAL PROTECTION AGENCY
WASHINGTON, D.C.

APRIL 1999

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Under Contract with the USEPA No. 68-C6-0039
Delivery Order 13

ACKNOWLEDGEMENTS

This document was prepared by the United States Environmental Protection Agency, Office of Ground Water and Drinking Water under the guidance of the Standards and Risk Management Division, Targeting and Analysis Branch. The Task Order Project Officer was Mr. Amit Kapadia.

Technical consultants played a significant role in the preparation of this document. The primary technical consultant was International Consultants, Inc. with significant subcontract assistance provided by Malcolm Pirnie, Inc. The Project Manager was Ron Braun of International Consultants. The International Consultants technical support team was lead by Chris Hill, Technical Project Leader, and included Helen Owens, Dr. Dave Jorgenson, Roger Azar and Paul Harvey. The Malcolm Pirnie technical support team was lead by Zaid Chowdhury with project support from Anne Jack, Amparo Flores and Spyros Papadimas. Additional technical assistance and consultation were provided by: Gary Amy, University of Colorado - Boulder; Joe Chwirka, CH2MHill - Albuquerque; Dennis Clifford, University of Houston; Marc Edwards, Virginia Polytechnic Institute; Janet Hering, California Institute of Technology; Jeff Robinson, Indiana-American Water Company; and Mark Weisner, Rice University.

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LIST OF ACRONYMS

AA	activated alumina
AWWA	American Water Works Association
AWWARF	American Water Works Association Research Foundation
BLS	Bureau of Labor Statistics
BV	bed volume
C/F	coagulation/filtration
CFR	Code of Federal Regulations
D/DBP	Disinfectant/Disinfection By-Product
DBP	disinfection by-product
DD	direct discharge
DMAA	dimethyl arsenic acid
DOC	dissolved organic carbon
DWRD	Drinking Water Research Division
EBCT	empty bed contact time
ED	electrodialysis
EDR	electrodialysis reversal
ENR	Engineering News Record
EP	evaporation ponds and drying beds
EPA	United States Environmental Protection Agency
Fe/Mn	iron/manganese
ft	feet
GAC	granular activated carbon
GFH	granular ferric hydroxide
gpd	gallons per day
gpm	gallons per minute

GT	gravity thickening
HD	hazardous waste landfill disposal
HDPE	high-density polyethylene
ID	indirect discharge
IOCS	iron oxide coated sand
IX	ion exchange
kgal	thousand gallons
kgpd	thousand gallons per day
kWh	kilowatt hour
LA	land application
lb/acre	pounds per acre
LS	lime softening
MCL	Maximum Contaminant Level
MCLG	Maximum Contaminant Level Goal
MD	mechanical dewatering
MDL	minimum detection limit
MF	microfiltration
mg/kg	milligram per kilogram
mg/L	milligrams per liter
MGD or mgd	million gallons per day
MMAA	monomethyl arsenic acid
MWCO	molecular weight cut-off
MWDSC	Metropolitan Water District of Southern California
NF	nanofiltration
NOM	natural organic matter
NIPDWR	National Interim Primary Drinking Water Regulation
NPDES	National Pollutant Discharge Elimination System

NPDWR	National Primary Drinking Water Regulation
O&M	operations and maintenance
OGWDW	Office of Ground Water and Drinking Water
POE	point-of-entry
POTW	public-owned treatment works
POU	point-of-use
ppb	parts per billion
ppm	parts per million
PPI	Producer Price Index (for Finished Goods)
psi	pounds per square inch
psig	pounds per square inch gauge
PVC	polyvinyl chloride
RCRA	Resource Conservation and Recovery Act
RO	reverse osmosis
scf	standard cubic feet
SD	sanitary landfill disposal
SDWA	Safe Drinking Water Act
sf	square feet
SL	storage lagoons
SMI	Sulfur-Modified Iron
SOC	synthetic organic compound
sq ft	square feet
TBLL	Technically Based Local Limits
TCLP	Toxicity Characteristic Leaching Procedure
TDS	total dissolved solids
TOC	total organic carbon
TSS	total suspended solids

TWG	Technologies Working Group
UF	ultrafiltration
UV_{254}	ultraviolet 254
WET	Whole Effluent Toxicity
wk	week
yr	year
μg/L	micrograms per liter

EXECUTIVE SUMMARY

BACKGROUND

In 1976 EPA issued a National Interim Primary Drinking Water Regulation (NIPDWR) for arsenic at 50 parts per billion (ppb or μg/L). Under the 1986 amendments to the Safe Drinking Water Act (SDWA), Congress directed EPA to publish Maximum Contaminant Level Goals (MCLGs) and promulgate National Primary Drinking Water Regulations (NPDWRs) for 83 contaminants, including arsenic. As EPA missed the statutory deadline for promulgating an arsenic regulation, a citizens' group filed suit to compel EPA to do so; EPA entered into a consent decree to issue the regulation. The EPA Office of Ground Water and Drinking Water (OGWDW) held internal workgroup meetings throughout 1994, addressing risk assessment, treatment, analytical methods, arsenic occurrence, exposure, costs, implementation issues, and regulatory options before deciding in early 1995 to defer the regulation to better characterize health effects and treatment technology. When Congress reauthorized the SDWA on the August 6, 1996, section 1412(b)(12)(A) was added. This addition specifies in part, that EPA propose a NPDWR for arsenic by January 1, 2000 and issue a final regulation by January 1, 2001.

The purpose of this document is to characterize the ability of arsenic removal technologies and to estimate costs for treatment technologies that can be used by utilities to meet regulatory standards. This document was originally published in 1993 as *Treatment and Occurrence of Arsenic in Potable Water Supplies* (Malcolm Pirnie, 1993a). Design criteria from the 1993 document have been re-evaluated and modified in accordance with the most recent research and input from a panel of experts. The design criteria established were used to develop treatment costs for arsenic removal. These costs will be used by EPA to determine national costs for various arsenic regulatory scenarios.

Costs were developed using the W/W Cost Model (Culp/Wesner/Culp, 1994), the Water Model (Culp/Wesner/Culp, 1984), and the *Very Small Systems Best Available Technology Cost Document* (Malcolm Pirnie, 1993b). For some technologies (e.g., membranes), published data of operating plants were used to estimate costs, as the models were judged inadequate or out of date. Where appropriate, vendors and equipment manufacturers were contacted to assess the accuracy of the cost models, and, when necessary, costs were modified to reflect the input from these sources.

ARSENIC PROPERTIES AND REMOVAL TECHNOLOGIES

Arsenic occurs in two primary forms; organic and inorganic. Organic species of arsenic are predominantly found in foodstuffs, such as shellfish, and include such forms as monomethyl arsenic acid (MMAA), dimethyl arsenic acid (DMAA), and arseno-sugars. Inorganic arsenic occurs in two valence states, arsenite (As III) and arsenate (As V). As(III) species consist primarily of arsenious acid (H_3AsO_3) in natural waters. As(V) species consist primarily of $H_2AsO_4^-$ and $HAsO_4^{2-}$ in natural waters (Clifford and Lin, 1995). Most natural waters contain the inorganic forms of arsenic. Moreover, natural groundwaters contain the more toxic (among the inorganic species) form As(III) as reducing conditions prevail. In natural surface waters, however, As(V) is the dominant species. Arsenite is removed less efficiently because it predominantly occurs in the uncharged (H_3AsO_3) state in source waters with a pH of less than 9.0. The dominant arsenate forms are anionic species, $H_2AsO_4^{2-}$ and $HAsO_4^-$.

Arsenic removal is dependent upon the ionic form present and water chemistry. As a result, identification of the ionic form is necessary for selection and design of an arsenic removal process. All technologies discussed in this document remove arsenate more effectively than arsenite. Therefore, if arsenite is the predominant species present, oxidation to arsenate may be required to achieve the desired removal.

Source water pH plays a significant role in determining the removal efficiency of a particular technology. Most processes are relatively unaffected by pH in the range of 6.5 to 9.0. However, activated alumina studies have shown the optimum pH for arsenic removal to be between 5.5 and 6.0. Reverse osmosis processes may require pH adjustment to prevent precipitation of salts on the membrane surface.

Co-occurrence of inorganic contaminants, such as sulfate and silica, as well as suspended solids, can cause interference with arsenic removal. Sulfate is preferentially adsorbed over arsenic by ion exchange processes. This preference can result in another phenomenon known as peaking, which occurs when arsenic is displaced on the resins by the sulfate causing effluent concentrations in excess of the influent levels.

This document evaluates arsenic removal technologies for drinking water and the costs associated with those technologies. Specifically, the following treatment processes are discussed:

Precipitative processes, including coagulation/filtration (C/F), direct filtration, coagulation assisted microfiltration, enhanced coagulation, lime softening (LS), and enhanced lime softening;

Adsorption processes, including activated alumina (AA), and iron oxide coated sand (IOCS);

Ion exchange (IX) processes;

Membrane filtration, including microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), reverse osmosis (RO), and electrodialysis reversal (EDR);

Alternative treatment processes, including biological processes, granular ferric hydroxide, sulfur-modified iron and iron filings, and greensand filtration;

Point-of-entry (POE) and point-of-use (POU) devices.

Many of these processes were evaluated to develop cost curves for the technologies. The above list also includes some experimental technologies which could not be costed out at this time. Discussions of each of these technologies are included in this document for future considerations of these processes for arsenic removal:

Iron oxide coated sand;
Greensand filtration;
Iron filings and sulfur-modified iron; and
Granular ferric hydroxide.

DEVELOPMENT OF DESIGN CRITERIA AND TREATMENT COSTS

Three cost models were used in cost development: the *Very Small Systems Best Available Technology Cost Document* (Malcolm Pirnie, 1993), hereafter referred to as the VSS model; the Water Model (Culp/Wesner/Culp, 1984); and the W/W Cost Model (Culp/Wesner/Culp, 1994). Curve fitting analysis was conducted on the modeled cost estimates including the utilization of transition flow regions to provide better estimates within the breakpoints between models. The following flow ranges have been established for each model and transition flow region:

VSS - 0.015 to 0.100 mgd
Transition 1 - 0.100 to 0.270 mgd
Water Model - 0.27 to 1.00 mgd
Transition 2 - 1 to 10 mgd
W/W Cost Model - 10 to 200 mgd

Flow categories were developed to provide adequate characterization of costs across each of the flow regions presented above. A minimum of four data points were generated for each of the flow regions, with the exception of the transition regions, where cost estimates are based upon a linear regressions between the last data point of the previous region and the first data point of the following region. Table ES-1 presents the design and average flows, and cost models used in this process.

The arsenic species present can greatly affect the removal efficiency of the selected treatment process. Pre-oxidation may be necessary to convert arsenite to arsenate. This document presents two pre-oxidation alternatives; chlorination and potassium permanganate feed. Costs are presented for dosages of 1.5, 3.0 and 5.0 mg/L for both pre-oxidation technologies. Other oxidation technologies, such as hydrogen peroxide feed and ozonation, may also be effective, but are not considered as typical oxidation processes for this contaminant.

Design criteria for the technologies selected for the development of cost curves are shown in Table ES-2. These design criteria were developed by consulting various engineers and experts convened by the American Water Works Association (AWWA) in 1994 (Frey, et al., 1997).

Table ES-1

Flows Used in the Cost Estimation Process

Design Flow (mgd)	Average Flow (mgd)	Cost Model
0.010	0.0031	VSS
0.024	0.0056	VSS
0.087	0.024	VSS
0.10	0.031	VSS
0.27	0.086	Water
0.45	0.14	Water
0.65	0.23	Water
0.83	0.30	Water
1.0	0.36	Water
1.8	0.7	W/W Cost
4.8	2.1	W/W Cost
10	4.5	W/W Cost
11	5	W/W Cost
18	8.8	W/W Cost
26	13	W/W Cost
51	27	W/W Cost
210	120	W/W Cost
430	270	W/W Cost

Shaded rows represent data used in the estimation of costs with the transition regions.

TABLE ES-2

Design Criteria and Key Assumptions

Process Name	Process Design Criteria
Coagulation/Filtration	Ferric chloride dose = 25 mg/L Polymer dose = 1 mg/L Lime dose = 25 mg/L (as CaO) for pH adjustment Waste flow = 1% of treated flow at 1% solids Package plants for small systems. On-site construction for large systems. Rapid mix = 1 minute, Flocculation = 20 minutes, and Sedimentation = 2.5 hours using rectangular tanks Dual media gravity filters at 5 gpm/sf
Enhanced Coagulation	Additional ferric chloride dose = 10 mg/L Additional lime dose = 10 mg/L All other specifications same as conventional coagulation/filtration.
Direct Filtration	Ferric chloride dose = 10 mg/L Polymer dose = 1 mg/L Lime dose = 10 mg/L (as CaO) for pH adjustment All other criteria same as above except sedimentation. No sedimentation basins are provided.
Coagulation Assisted Microfiltration	Ferric chloride dose = 10 mg/L All other specifications are standard microfilter specifications.
Lime Softening	Lime dose = 250 mg/L. Hydrated lime & soda ash for small systems. Quick lime with on-site slaking for large systems. CO_2 = 35 mg/L (liquid carbon dioxide) for recarbonation. Waste flow = 2% of total flow at 1% solids Package plant for small systems. On-site construction for large systems. Rapid mix = 1 minute, Flocculation = 20 minutes, and Sedimentation = 1000 gpd/sf using circular tanks Dual media gravity filters at 5 gpm/sf
Enhanced Lime Softening	Additional lime dose = 50 mg/L Additional CO_2 dose = 35 mg/L All other specifications same as above.

TABLE ES-2 (cont.)

Design Criteria and Key Assumptions

Process Name	Process Design Criteria
Ion Exchange	Regeneration frequency = 300, 500, 700, 1500 or 2500 bed volumes Resin replacement of 25% per year Regenerant = 15 pounds NaCl/cf of resin Expansion during regeneration = 50% Regeneration cycle = 10 minutes backwashing at 2.5 gpm/sf, 10 minutes regeneration at 0.5 gpm/sf
Activated Alumina	Regeneration frequency = 500, 2000, 5000, 10000, 25000, or 50000 bed volumes NaOH = 50 mg/L H2SO4 = 70 mg/L Replacement activated alumina = 1.5% per regeneration Regenerant = 1 pound of 50% NaOH and 0.2 pound of H2SO4 per 1,000 gallons of water Regeneration cycle = 35 minute regeneration at 2.5 gpm/sf, 30 minute rinse at 5gpm/sf Backwashing = 8 to 9 gpm/sf for 10 minutes
Microfiltration	none: based on plant survey information
Ultrafiltration	Hollow fiber membrane modules Skid-mounted ultrafiltration racks Automated system Continuous operation Membrane life expectancy of 4 years Membrane cleaning costs not included
Nanofiltration	none: based on plant survey information
Reverse Osmosis	Adequate pretreatment performed prior to the membrane process Spiral-wound cellulose acetate membrane elements Influent TDS of 10,000 mg/L Influent temperature between 65 and 95 degrees F Acid addition to prevent membrane fouling Single-pass system Operating Pressure of 400 to 450 psi Feed water recovery of 75% Brine disposal costs not included Membrane life expectancy of 3 years Membrane cleaning performed once per month

Capital costs are presented in April 1998 dollars. Appropriate *Engineering News Record* (ENR) and Bureau of Labor Statistics (BLS) cost indices were used for cost computation. The Producer's Price Index for Finished Goods were used adjusting operations and maintenance (O&M) cost estimates.

Capital and O&M cost curves and equations are presented for each technology discussed in Chapter 3. Capital costs are expressed as total cost (M-$); O&M estimates are expressed in dollars per year. For the ion exchange and activated alumina processes, a range of O&M costs have been cited since costs will vary with the number of bed volumes treated between regeneration, which is a function of several water quality parameters, such as ambient sulfate level, and initial and target arsenic concentrations.

Capital and O&M cost estimates were compared with actual data presented in *Evaluation of Full-Scale Treatment Technologies at Small Drinking Water Systems* (ICF and ISSI, 1998). It was found that the estimates presented in this document are reasonable. Capital cost estimates were routinely conservative, but followed the general trends seen in actual data. O&M estimates typically represented an approximate average of the real world costs. Actual data was not available for all technologies, and comparisons are not presented for some of the technologies discussed.

RESIDUALS HANDLING AND DISPOSAL ALTERNATIVES

Each of the treatment technologies presented in this document will produce residuals, either solid or liquid streams, containing elevated levels of arsenic. It is important to address residuals characteristics when selecting an arsenic removal technology. Handling and disposal costs can be significant, and if a waste stream happens to be hazardous the implications are even greater. This document evaluates typical characteristics of residuals produced by each of the treatment technologies presented, and discusses appropriate handling and disposal methods. Specifically, the following handling and disposal methods are discussed:

- Residuals handling
  - Gravity thickening;
  - Mechanical dewatering, including centrifuges and filter presses;
  - Non-mechanical dewatering, such as evaporation ponds and storage lagoons;

- Disposal alternatives
  - Direct discharge to receiving water;
  - Discharge to sanitary sewer for treatment at a wastewater treatment plant;
  - Land application;
  - Sanitary landfill disposal; and
  - Hazardous landfill disposal.

There are a number of factors which can influence residuals handling and disposal costs. The primary factor affecting capital cost is the size of the water system, i.e., population and water needs. All other costs are directly proportional to these factors. The amount of waste generated plays a significant role in determining the handling and disposal method to be utilized. Many handling methods are impractical for large water systems because of land requirements. However, some handling methods require expensive process equipment which may make them more suitable to large water systems. Similarly, waste disposal methods requiring large capital investments may make them impractical for small water systems.

Many handling and disposal methods require extensive oversight which can be a burden on small water systems. Generally, labor intensive technologies are more suitable to large water systems. Transportation can also play a significant role in determining appropriate handling and disposal options. If off-site disposal requires extensive transportation, alternative disposal methods should be evaluated. Complex handling and disposal methods usually require more maintenance. When evaluating handling and disposal methods, it is generally best to select that option which will require the least amount of oversight and maintenance.

Residuals handling and disposal costs can be difficult to estimate. There are a number of factors which affect capital and O&M costs, and disposal costs can be largely regional. EPA has published two manuals for estimating residuals handling and disposal costs; *Small Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993a), and *Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993b). Both present a variety of handling and disposal options, applications and limitations of those technologies, and capital and O&M cost equations. Residuals handling and disposal costs are not included in this document. The references listed above can be used to generate such costs.

POINT-OF-ENTRY AND POINT-OF-USE TREATMENT OPTIONS

Centralized treatment is not always a feasible treatment option, for example in areas where each home has a private well or centralized treatment is cost prohibitive. In these instances, point-of-entry (POE) and point-of-use (POU) treatment options may be acceptable treatment alternatives. POE and POU systems offer ease of installation, simplify operation and maintenance, and generally have lower capital costs. These systems may also reduce engineering, legal and other fees typically associated with centralized treatment options. Use of POE and POU systems does not reduce the need for a well-maintained water distribution system. In fact, increased monitoring may be necessary to ensure that the treatment units are operating properly.

Home water treatment can consist of either whole-house or single faucet treatment. Whole-house, or POE treatment is necessary when exposure to the contaminant by modes other than consumption is a concern. POU treatment is preferred when treated water is needed only for drinking and cooking purposes. POU treatment usually involves single-tap treatment.

Section 1412(b)(4)(E) of the 1996 Safe Drinking Water Act (SDWA) Amendments requires the EPA to issue a list of technologies that achieve compliance with Maximum Contaminant Levels (MCLs) established under the act. This list must contain technologies for each NPDWR and for each of the small public water systems categories listed below:

- Population of more than 50, but less than 500;
- Population of more than 500, but less than 3,300; and
- Population of more than 3,300, but less than 10,000.

The SDWA identifies POE and POU treatment units as potentially affordable technologies, but stipulates that POE and POU treatment systems "shall be owned, controlled and maintained by the public water system, or by a person under contract with the public water system to ensure proper operation and compliance with the maximum contaminant level or treatment technique and equipped with mechanical warnings to ensure that customers are automatically notified of operational problems."

Research has shown that POE and POU devices can be effective means of removing arsenic from potable water. Water systems with high influent arsenic concentrations, i.e., greater than 1 mg/L, may have difficulty meeting MCLs much lower than the 10 to 20 μg/L level. As a result, influent arsenic concentration and other source water characteristics must be considered when evaluating POE and POU devices for arsenic removal. Reverse osmosis, activated alumina and ion exchange are three treatment techniques that have been evaluated and shown to be effective. This document looks at the removals achieved by each of these three treatment techniques, and presents total costs for each treatment option.

REGIONALIZATION

Regionalization involves purchasing and transferring water from one community or water source to another. In effect, regionalization expands the region served by a water distribution system. There are a number of factors which can influence the decision to implement regionalization, including water availability, water quality, geography and economic factors.

Accordingly, community water systems faced with installation of treatment facilities to address arsenic contamination issues may opt for regionalization. This document presents the costs associated with regionalization. Costs can be largely variable, with many site specific considerations. The estimates presented are for typical installation and do not include costs associated with site specific construction conditions, such as rugged terrain, severe elevation changes and land costs.

1.0 INTRODUCTION

1.1 OVERVIEW

Arsenic (As) is a naturally occurring element present in food, water, and air. Known for centuries to be an effective poison, some animal studies suggest that arsenic may be an essential nutrient at low concentrations. Non-malignant skin alterations, such as keratosis and hypo- and hyper-pigmentation, have been linked to arsenic ingestion, and skin cancers have developed in some patients. Additional studies indicate that arsenic ingestion may result in internal malignancies, including cancers of the kidney, bladder, liver, lung, and other organs. Vascular system effects have also been observed, including peripheral vascular disease, which in its most severe form, results in gangrene or Blackfoot Disease. Other potential effects include neurologic impairment (Lomaquahu and Smith, 1998).

The primary route of exposure to arsenic for humans is ingestion. Exposure via inhalation is considered minimal, though there are regions where elevated levels of airborne arsenic occur periodically (Hering and Chiu, 1998). Arsenic occurs in two primary forms; organic and inorganic. Organic species of arsenic are predominantly found in foodstuffs, such as shellfish, and include such forms as monomethyl arsenic acid (MMAA), dimethyl arsenic acid (DMAA), and arseno-sugars. Inorganic arsenic occurs in two valence states, arsenite (As III) and arsenate (As V). As(III) species consist primarily of arsenious acid (H_3AsO_3) in natural waters. As(V) species consist primarily of $H_2AsO_4^-$ and $HAsO_4^{2-}$ in natural waters (Clifford and Lin, 1995). Most natural waters contain the more toxic inorganic forms of arsenic. Natural groundwaters contain the more toxic As(III) (among the inorganic species) as reducing conditions prevail. In natural surface waters As(V) is the dominant species. Arsenic removal technologies for drinking water include:

- Precipitative processes, including coagulation/filtration (C/F), direct filtration, coagulation assisted microfiltration, enhanced coagulation, lime softening (LS), and enhanced lime softening;

- Adsorption processes, including activated alumina (AA), and iron oxide coated sand (IOCS);

- Ion exchange (IX) processes, specifically anion exchange;
- Membrane filtration, including microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), reverse osmosis (RO), and electrodialysis reversal (EDR);
- Alternative treatment processes, including biological processes, granular ferric hydroxide, sulfur-modified iron and iron filings, and greensand filtration; and
- Point-of-entry (POE) and point-of-use (POU) devices.

Many of these processes were evaluated to develop cost curves for the technologies. The above list also includes some experimental technologies which could not be costed at this time. Discussions of the following technologies are included in this document for future consideration as viable processes for arsenic removal:

- Iron oxide coated sand;
- Greensand filtration;
- Iron filings and sulfur-modified iron; and
- Granular ferric hydroxide.

1.2 STATUTORY REQUIREMENTS

In 1976 EPA issued a National Interim Primary Drinking Water Regulation (NIPWDR) for arsenic at 50 parts per billion (ppb or μg/L). Under the 1986 amendments to the Safe Drinking Water Act (SDWA), Congress directed EPA to publish Maximum Contaminant Level Goals (MCLGs) and promulgate National Primary Drinking Water Regulations (NPDWRs) for 83 contaminants, including arsenic. When EPA missed the statutory deadline for promulgating an arsenic regulation, a citizens' group filed suit to compel EPA to do so; EPA entered into a consent decree to issue the regulation. The EPA Office of Ground Water and Drinking Water (OGWDW) held internal workgroup meetings throughout 1994, addressing risk assessment, treatment, analytical methods, arsenic occurrence, exposure, costs, implementation issues, and regulatory options before deciding in early 1995 to defer the regulation to better characterize health effects and treatment technology.

With the reauthorization of the SDWA on August 6, 1996, Congress added section 1412(b)(12)(A) to the act. This addition specifies in part, that EPA propose a NPDWR for arsenic by January 1, 2000 and issue a final regulation by January 1, 2001.

1.3 DOCUMENT ORGANIZATION

This document contains the following chapters:

Chapter 1.0 Introduction - Provides an introduction to the arsenic statutory requirements, including BAT and variance technology requirements, as well as defines technology categories and presents the organization of the document.

Chapter 2.0 Arsenic Removal Technologies - Presents discussions on available arsenic removal technologies, removal efficiencies, factors affecting arsenic removal and associated pilot- and full-scale studies.

Chapter 3.0 Technology Costs - Presents capital and O&M costs, for each of the removal technologies in graphical format. This chapter also contains a comparison of the cost estimates presented in this document to actual capital and O&M costs obtained during an EPA survey of small water systems.

Chapter 4.0 Residuals Handling and Disposal Alternatives - Presents capital and O&M cost equations for a variety of residuals handling and disposal alternatives.

Chapter 5.0 Point-of-Entry/Point-of-Use Treatment Options - Evaluates a number of POE and POU treatment options effective for arsenic removal, as well as presents capital and O&M costs in graphical form for each of the treatment options.

Chapter 6.0 Regionalization - Presents estimates for regionalization as opposed to centralized treatment.

Chapter 7.0 References - Lists the literature cited in this document, as well as additional references which may be of interest to the reader.

2.0 ARSENIC REMOVAL TECHNOLOGIES

2.1 INTRODUCTION

Arsenic removal technologies are discussed in this chapter. Some of these technologies are traditional treatment processes which have been tailored to improve removal of arsenic. Several treatment techniques discussed here are at the experimental stage with regard to arsenic removal, and some have not been demonstrated at full scale. Although some processes may be technically feasible, cost may be prohibitive.

Technologies discussed in this section are grouped into four broad categories: precipitative processes, adsorption processes, ion exchange processes, and separation (membrane) processes. Each category is discussed here, with at least one treatment technology described in each category.

2.2 PRECIPITATIVE PROCESSES

2.2.1 Coagulation/Filtration

Coagulation/filtration is a treatment process by which the physical or chemical properties of dissolved colloidal or suspended matter are altered such that agglomeration is enhanced to an extent that the resulting particles will settle out of solution by gravity or will be removed by filtration. Coagulants change surface charge properties of solids to allow agglomeration and/or enmeshment of particles into a flocculated precipitate. In either case, the final products are larger particles, or floc, which more readily filter or settle under the influence of gravity.

The coagulation/filtration process has traditionally been used to remove solids from drinking water supplies. However, the process is not restricted to the removal of particles. Coagulants render some dissolved species (e.g., natural organic matter (NOM), inorganics and hydrophobic synthetic organic compounds (SOCs)) insoluble and the metal hydroxide particles produced by the addition of metal salt coagulants (typically aluminum sulfate, ferric chloride or ferric sulfate) can adsorb other dissolved species. Major components of a basic coagulation/filtration facility include chemical feed systems, mixing equipment, basins for rapid mix, flocculation, settling, filter media, sludge handling equipment, and filter backwash facilities. Settling may not be necessary in situations where the influent particle concentration is very low. Treatment plants without settling are known as direct filtration plants.

As(III) removal during coagulation with alum, ferric chloride, and ferric sulfate has been shown to be less efficient than As(V) under comparable conditions (Hering, et al., 1996; Edwards, 1994; Shen, 1973; Gulledge and O'Conner, 1973; Sorg and Logsdon, 1978). If only As(III) is present, consideration should be given to oxidation prior to coagulation to convert As(III) to As(V) species.

Effect of Coagulant Type

Batch studies were conducted at the University of Illinois to demonstrate the removal of As(V) by coagulation, sedimentation, and filtration **(**Gulledge and O'Conner, 1973). Raw water was spiked to obtain an initial concentration of 0.05 mg/L As(V); alum or ferric sulfate were used as coagulants at varying dosages. The pH was varied between 5.0 and 8.0, which is higher than the optimum pH range of 5.0 to 7.0 for alum coagulation, but within the optimum pH range for ferric sulfate coagulation. The results of these studies demonstrate that ferric sulfate coagulation within the optimum pH range achieved better removals than alum coagulation over a larger coagulant dosage range. Over 90 percent of As(V) was removed with alum coagulation but only at dosages greater than 30 mg/L. With ferric sulfate coagulation, over 95 percent of the As(V) was removed within the pH range of 5.0 to 7.5 for dosages between 10 and 50 mg/L.

Logsdon et al. (1974) showed that at an influent concentration of 0.3 mg/L, removals ranged from 40 to 60 percent with ferric sulfate coagulation, compared to 5 to 15 percent with alum coagulation. Higher As(III) removals were achieved in the pH range of 5.0 to 8.5 for ferric sulfate and 5.0 to 7.0 for alum. When As(III) was oxidized with 2 mg/L of chlorine, removals increased for both alum and ferric sulfate within the same pH range, but ferric sulfate still achieved higher removals. Over 95 percent of the oxidized As(III) was removed with ferric sulfate coagulation, and between 83 and 90 percent was removed with alum coagulation.

Scott, et al. (1995) conducted a full-scale study at the Metropolitan Water District of Southern California (MWDSC) to determine arsenic removals using alum and ferric chloride. The average concentration of arsenic in the source water was 2.1 µg/L. When the source water was treated with 3 to 10 mg/L of ferric chloride, arsenic removal was 81 to 96 percent. When the source water was treated with 6, 10, or 20 mg/L of alum, arsenic removal was 23 to 71 percent.

McNeill and Edwards (1997a) reported that solubility and stability of the metal hydroxide flocs play an important role in arsenic removal. When ferric coagulants are added, most of the ferric ends up as ferric hydroxide. In alum coagulation, however, a significant portion of the added aluminum remains as soluble complexes. Because only particulate metal hydroxides can mediate arsenic removal, alum plants must carefully consider aluminum solubility when arsenic removal is required. Aluminum complexes can pass through filters and decrease overall arsenic removal.

Effect of Coagulant Dosage

In general, higher removal efficiencies can be achieved with increased coagulant dosages (Cheng, et al., 1994; Edwards, 1994; Gulledge and O'Conner, 1973). Hering et al. (1996) demonstrated in coagulation experiments with ferric chloride at pH 7.0 that both As(III) and As(V) removal were dependent on coagulant dosage. "Complete" removal of As(V) was observed for coagulant dosages above 5 mg/L ferric chloride. "Complete" removal of As(III) was not observed under the range of conditions examined.

Predictions based on existing data and the use of a diffuse-layer model indicated that As(III) removals by coagulation were primarily controlled by coagulant dosage, whereas the converse was true for As(V) (Edwards, 1994). A database compiled by Edwards (1994) containing much previously published work on arsenic coagulation indicated that, at all dosages greater than 20 mg/L as ferric chloride or 40 mg/L as alum, greater than 90 percent removal of As(V) was always achieved. At lower coagulant dosages there was considerable scatter in the data attributed to poor particle removal, high initial As(V) concentrations, and possible interferences from other anions in the different waters tested.

Effect of Coagulation pH

Sorg and Logsdon (1978) demonstrated that arsenic removal with alum coagulation is most effective at pH 5 to 7 and ferric coagulation are most effective at pH 5 to 8. As discussed earlier, Edwards (1994) summarized that at significant coagulant dosages As(V) removal was similar for both alum and ferric coagulants at pH 7.6 or lower. At pH values greater than 7.6, however, the average removals were 87 percent for 10 mg/L ferric chloride and only 67 percent for 20 mg/L alum.

Analyzing previously collected research data for As(III) removal by iron and aluminum coagulation, Edwards (1994) demonstrated that removal of As (III) is much higher during iron coagulation when compared with that of alum. Furthermore, As(III) removal by adsorption onto aluminum hydroxides decreases markedly above pH 8.0.

Hering et al. (1996) observed the opposite effect. In coagulation experiments with ferric chloride over the pH range of 4 to 9, pH did not appear to influence the As(V) removal. However, strong pH dependence was observed for As(III) in coagulation experiments with ferric chloride, with a minimum in removal efficiency at pH 6.0.

Effect of Initial As(III)/As(V) Concentration

Logsdon et al. (1974) conducted several jar tests on spiked well water to analyze the initial concentration and form of arsenic, and determine the type of coagulant most effective in arsenic removal. The study found the initial arsenic concentration to have a significant effect on removals. For initial As(V) concentrations between 0.1 and 1.0 mg/L, a dosage of 30 mg/L of either alum or ferric sulfate in the optimum pH range removed over 95 percent As(V). Above an initial concentration of 1.0 mg/L, removals decrease with increasing concentrations. For concentrations of As(III) greater than 0.1 mg/L, neither alum nor ferric sulfate dosed at 30 mg/L could remove As(III) to concentrations below 0.05 mg/L. In both cases, higher coagulant dosages (60 to 100 mg/L) resulted in higher removals.

Hering et al. (1996) demonstrated in coagulation experiments, with ferric chloride dose of 4.9 mg/L at pH 7.0 and varied initial arsenic concentration from 2 to 100 µg/L, that both As(III) and As(V) removal was independent of initial concentration. Cheng et al. (1994) showed that As(V) removal was independent of initial concentration when treated with 20 mg/L of alum and 30 mg/L of ferric chloride while varying the initial As(V) concentration from 2.2 to 128 µg/L.

Effect of Co-occurring Inorganic Solutes

Co-occurring inorganic solutes, such as sulfate and calcium, may compete for surface binding sites onto oxide surfaces and influence the adsorption of trace contaminants, such as arsenic. Hering et al. (1996) investigated the effects of sulfate and calcium on the efficiency of As(III) and As(V) removal during coagulation with 4.9 mg/L of ferric chloride. The results indicated that at pH below 7.0, As(III) removal was significantly decreased in the presence of sulfate. However, only a slight decrease in As(V) was observed. At higher pH, removal of As(V) was increased in the presence of calcium.

Optimization Hierarchy for Coagulation/Filtration Facilities

McNeill and Edwards (1997a) developed a simple model for predicting As(V) concentration during coagulation with alum or ferric salts. Using inputs of aluminum hydroxide formed, ferric hydroxide present in the influent, ferric hydroxide formed, and a single sorption constant, the model predicted As(V) removal to within 13% for the 25 utility sampling events in this study. The authors suggested an optimization hierarchy strategy for coagulation/filtration facilities which are unable to meet arsenic removal requirements with their existing treatment scheme. If any As(III) is present in the raw water, the most cost-effective method of improving removal is to convert poorly sorbed As(III) to As(V). Thereafter, for facilities practicing alum coagulation, it is critical to minimize residual soluble aluminum to enhance the formation of aluminum hydroxide solids which mediate the As(V) removal. Jar testing should be performed to identify pH and coagulant dosage that might be altered to reduce aluminum residuals. The final option is to increase the coagulant dosage or to consider changing the coagulant type.

Summary

Coagulation is a successful technology for achieving As(V) removals greater than 90 percent. Arsenic in the pentavalant arsenate form is more readily removed than the trivalent arsenite form. At pH 7.6 or lower iron and aluminum coagulants are of equal effectiveness in removing As(V). However, iron coagulants are advantageous if pH is above 7.6, if soluble coagulant metal residuals are problematic, or if As(III) is present in the raw water. In general, higher arsenic removal efficiencies are achieved with increased coagulant dosages. The effectiveness of iron coagulants in removing As(III) diminishes at pH 6.0. Recent studies have shown that arsenic removal is independent of initial concentration. This contradicts initial findings which indicate that arsenic removals decrease with increasing initial concentrations. Presence of sulfate significantly decreases

As(III) removal, whereas sulfate slightly affects As(V) removal. At pH higher than 7.0, removal of As(V) increases in the presence of calcium.

2.2.2 Iron/Manganese Oxidation

Iron/Manganese (Fe/Mn) oxidation is dominant in facilities treating groundwater. Oxidation to remove iron and manganese leads to formation of hydroxides that remove soluble arsenic by precipitation or adsorption reactions.

Arsenic removal during iron precipitation is expected to be fairly efficient (Edwards, 1994). Removal of 2 mg/L of iron achieved a 92.5 percent removal of As(V) from a 10 µg/L As(V) initial concentration by adsorption alone. Even removal of 1 mg/L of iron is capable of adsorbing 83 percent of a 22 µg/L As(V) influent concentration. However, removal of arsenic during manganese precipitation is relatively ineffective when compared to iron even when removal by both adsorption and coprecipitation are considered. For instance, precipitation of 3 mg/L manganese removed only 69 percent of As(V) of a 12.5 µg/L As(V) influent concentration.

Effect of Co-occurring Inorganic Solutes

McNeill and Edwards (1995) demonstrated that a Fe/Mn facility with 400 mg/L sulfate and 5.2 µg/L arsenic in the raw water attained 83 percent removal of arsenic. Results from two other Fe/Mn facilities with 10 mg/L sulfate in the raw water showed 87 and 93 percent arsenic removals. This analysis suggests that sulfate interferes only slightly with sorption of arsenic onto ferric iron precipitates.

2.2.3 Coagulation Assisted Microfiltration

Arsenic is removed effectively during the coagulation process, as described in section 2.2.1. Microfiltration is used as a membrane separation process to remove particulates, turbidity, and microorganisms. In coagulation assisted microfiltration technology, microfiltration is used similarly to a conventional gravity filter. The advantages of microfiltration over conventional filtration are outlined below (Muilenberg, 1997):

- more effective microorganism barrier during coagulation process upsets;
- smaller floc sizes can be removed (smaller amounts of coagulants are required); and
- increased total plant capacity.

Vickers et al. (1997) reported that microfiltration exhibited excellent arsenic removal capability. Addition of a coagulant did not significantly affect the membrane cleaning interval, although the solids level to the membrane system increased substantially. With an iron and manganese removal system, it is critical that all of the iron and manganese be fully oxidized before they reach the membrane to prevent fouling (Muilenberg, 1997).

2.2.4 Enhanced Coagulation

The Disinfectant/Disinfection Byproduct (D/DBP) Rule requires the use of enhanced coagulation treatment technique for the reduction of disinfection byproduct (DBP) precursors for surface water systems which have sedimentation capabilities. This treatment technique involves modifications to the existing coagulation process such as increasing the coagulant dosage, reducing the pH, or both.

Cheng et al. (1994) conducted bench, pilot, and demonstration scale studies to examine As(V) removals during enhanced coagulation. The enhanced coagulation conditions in these studies included increase of alum and ferric chloride coagulant dosage from 10 to 30 mg/L, decrease of pH from 7 to 5.5, or both. Results from these studies indicated the following:

- Greater than 90 percent As(V) removal can be achieved under enhanced coagulation conditions. As(V) removals greater than 90 percent were easily attained under all conditions when ferric chloride was used.

- Enhanced coagulation using ferric salts is more effective for arsenic removal than enhanced coagulation using alum. With an influent arsenic concentration of 5 µg/L, ferric chloride achieved 96 percent As(V) removal with a dosage of 10 mg/L and no acid addition. When alum was used, 90 percent As(V) removal could not be achieved without reducing the pH.

- Lowering pH during enhanced coagulation improved arsenic removal by alum coagulation. With ferric coagulation pH does not have a significant effect between 5.5 and 7.0.

2.2.5 Lime Softening

Hardness is predominantly caused by calcium and magnesium compounds in solution. Lime softening (LS) removes this hardness by creating a shift in the carbonate equilibrium. The addition of lime to water raises the pH. Bicarbonate is converted to carbonate as the pH increases, and as a result, calcium is precipitated as calcium carbonate. Soda ash (sodium carbonate) is added if insufficient bicarbonate is present in the water to remove hardness to the desired level. Softening for calcium removal is typically accomplished at a pH range of 9 to 9.5. For magnesium removal, excess lime is added beyond the point of calcium carbonate precipitation. Magnesium hydroxide precipitates at pH levels greater than 10.5. Neutralization is required if the pH of the softened water is excessively high (above 9.5) for potable use. The most common form of pH adjustment in softening plants is recarbonation with carbon dioxide.

LS has been widely used in the U.S. for reducing hardness in large water treatment systems. LS, excess lime treatment, split lime treatment, and lime-soda softening are all common in municipal water systems. All of these treatment methods are effective in reducing arsenic. As(III) or As(V) removal by LS is pH dependent. Oxidation of As(III) to As(V) prior to LS treatment will increase removal efficiencies if As(III) is the predominant form. Considerable amounts of sludge are produced in a LS system and its disposal is expensive. Large capacity systems may find it economically feasible to install recalcination equipment to recover and reuse the lime sludge and reduce disposal problems. Construction of a new LS plant for the removal of arsenic would not generally be recommended unless hardness must also be reduced.

Effect of Initial As(V)/As(III) Concentration

McNeill and Edwards (1997b) showed that the percentage of As(V) removal by calcium carbonate and magnesium hydroxide is constant regardless of the initial As(V) concentration. At pH 10.5-12, As(V) removal was 23±4 percent for removal by calcium carbonate over the range of As(V) concentrations of 5-75 µg/L. At pH 11, As(V) removal was 37±5 percent for removal by magnesium hydroxide over the range of As(V) concentrations of 5-160 µg/L.

These results differ from those of Logsdon et al. (1974) who found that arsenic removal was dependent on the initial arsenic concentration. In the optimum pH range, As(V) or oxidized As(III) was reduced to 0.05 mg/L when the initial concentration was 0.35 mg/L or lower, while As(III) was reduced to 0.05 mg/L when the initial concentration was less than 0.1 mg/L.

McNeill and Edwards (1997b) also found that As(V) removal by manganese hydroxide solids is sensitive to As(V) initial concentrations. At pH of 10.5, there was about 80 percent removal in the system with 75 µg/L of As(V) versus about 30 percent of removal in the 150 µg/L As(V) solution.

Effect of Arsenic Oxidation State

As(V) was generally more effectively removed by LS than As(III). Sorg and Logsdon (1978) conducted several LS pilot studies for the removal of both As(III) and As(V). Two of the tests were performed at pHs 9.5 and 11.3. At a pH of 11.3, 99 percent of an initial As(V) concentration of 0.58 mg/L was removed, whereas only 71 percent of an initial As(III) concentration of 0.34 mg/L was removed. At a pH of 9.5, 53 percent of an initial As(V) concentration of 0.42 mg/L was removed, whereas only 24 percent of an initial As(III) concentration of 0.24 mg/L was removed.

Effect of pH

The optimum pH for As(V) removal by LS is approximately 10.5, and the optimum pH for As(III) removal is approximately 11 (Logsdon, et al., 1974; Sorg and Logsdon, 1978). Logsdon, et al. (1974) studied the effectiveness of excess LS on the removal of arsenic in jar tests. The test water was a well water that contained 300 mg/L hardness as $CaCO_3$ spiked with 0.4 mg/L As(V). The pH varied between 8.5 and 11.5. At pH 10.5 and above, nearly 100 percent arsenic removal was obtained. Below the optimum pH, the removals decreased with decreasing pH. When the water was spiked with As (III), removals were only around 75 percent in the optimum pH range. Below the optimum pH range, removals sharply decreased to less than 20 percent. Removals of oxidized As(III), however, were almost identical to removals of As(V).

Effect of Type of Precipitative Solids Formed

Arsenate removal during softening is controlled by formation of three solids including calcium carbonate, magnesium hydroxide, and ferric hydroxide. Calcium carbonate and magnesium hydroxide are produced from reactions which remove hardness from water after addition of lime, caustic soda, and soda ash. Ferric hydroxide can be formed by precipitation of iron naturally present in treatment plant influent or by addition of iron coagulant during softening.

A survey of full-scale plants by McNeill and Edwards (1995) indicated that soluble As(V) removal is mediated primarily by sorption to magnesium and/or ferric hydroxide solids during water softening operations. At softening facilities precipitating only calcite, soluble As(V) removal was between 0 to 10 percent, whereas soluble As(V) at plants precipitating calcite and magnesium and/or ferric hydroxide was between 60 to 95 percent.

McNeill and Edwards (1997b) performed bench-scale studies to investigate the role of iron addition in optimizing the As(V) removal. At pH 9 without any iron addition, only a small amount of As(V) was removed. However, adding increasing amounts of iron at this pH improved As(V) removal, with 82 percent of the As(V) removed at an iron dose of 9 mg/L. At pH 9.7, a 38 percent As(V) removal without iron addition was observed, versus 63±8.4 percent removal for iron dosages between 0.25 and 9 mg/L.

Effect of Other Constituents

The competitive effects of sulfate and carbonate for surface binding sites onto magnesium hydroxide surfaces and the influence on the adsorption of arsenic was examined by McNeill and Edwards (1998). These effects were investigated in experiments with preformed magnesium hydroxide by adding 20 mg/L Mg^{+2} and raising the pH to 12 after spiking the source water with 20 mg/L of As(V). Samples were collected as pH was incrementally lowered at ten minute intervals.

At pH 11 and above, no appreciable sulfate or carbonate interference was observed compared to the control case. However, at pH 10 to 10.5, the system with carbonate exhibited significantly lower As(V) removal (78 percent versus 96 percent in the control and sulfate systems) and nearly twice as much of the magnesium was measured as soluble (6.3 versus 3.3 mg/L). These results suggest that carbonate is somehow increasing the concentration of Mg^{+2}, leaving less solid available for As(V) sorption.

McNeill and Edwards (1997b) investigated the interference of orthophosphate on the As(V) removal by softening. Softening of raw water containing 15 µg/L As(V) at pH 12 indicated greater than 95 percent As(V) removal. After spiking raw water with 32 µg/L orthophosphate, As(V) removal was slightly lower at intermediate pH values. Because the amount of calcium and magnesium removed during softening with and without orthophosphate was nearly equal, it seems that orthophosphate interferes with arsenic removal by competing for sorption sites.

Optimization Hierarchy for Softening Facilities

McNeill and Edwards (1997b) developed a simple model for predicting As(V) during softening. Using inputs of calcium carbonate, magnesium and ferric hydroxide solid concentrations formed during softening, the model can predict percentage As(V) removal.

McNeill and Edwards (1997b) suggested an optimization hierarchy strategy for softening facilities which are unable to meet arsenic removal requirements with their existing treatment scheme similar to optimization of coagulation hierarchy. If As(III) is present, the most cost-effective method of improving arsenic removal is preoxidation of As(III) to As(V), since As(V) is more readily removed by precipitation of calcium carbonate, and magnesium and ferric hydroxide. For facilities that are currently precipitating only calcium carbonate, addition of iron can dramatically improve arsenic removal. A final option is to raise the softening pH in order to precipitate magnesium hydroxide which strongly sorbs As(V). These removal trends should be quantitatively confirmed with jar testing for optimizing arsenic removal.

Summary

Softening is a successful technology for achieving greater than 90 percent As(V) removals. Arsenic in the pentavalent arsenate form is more readily removed than the trivalent arsenite form. The optimum pH for As(V) removal by softening is approximately 10.5 and the optimum pH of As(III) is approximately 11.0. Recent studies have shown that As(V) removal is independent of initial concentration. This contradicts initial findings which indicate that As(V) removal is a function of initial concentration. As(III) removal appears to depend on initial concentration. Facilities precipitating only calcium carbonate observed lower As(V) removals when compared to facilities precipitating calcium carbonate, and magnesium and ferric hydroxide. Addition of iron improves As(V) removal. Presence of sulfate and carbonate in the raw water does not interfere with As(V) removal at pH 11. As(V) removal, however, is reduced in the presence of carbonate at pH 10 to 10.5 and presence of orthophosphate at pH less than 12.0.

2.3 ADSORPTIVE PROCESSES

2.3.1 Activated Alumina

Activated Alumina (AA) is a physical/chemical process by which ions in the feed water are sorbed to the oxidized AA surface. AA is considered an adsorption process, although the chemical reactions involved are actually an exchange of ions (AWWA, 1990). Activated alumina is prepared through dehydration of $Al(OH)_3$ at high temperatures, and consists of amorphous and gamma alumina oxide (Clifford and Lin, 1995). AA is used in packed beds to remove contaminants such as fluoride, arsenic, selenium, silica, and NOM. Feed water is continuously passed through the bed to remove contaminants. The contaminant ions are exchanged with the surface hydroxides on the alumina. When adsorption sites on the AA surface become filled, the bed must be regenerated. Regeneration is accomplished through a sequence of rinsing with regenerant, flushing with water, and neutralizing with acid. The regenerant is a strong base, typically sodium hydroxide; the neutralizer is a strong acid, typically sulfuric acid.

Many studies have shown that AA is an effective treatment technique for arsenic removal. Factors such as pH, arsenic oxidation state, competing ions, empty bed contact time (EBCT), and regeneration have significant effects on the removals achieved with AA. Other factors include spent regenerant disposal, alumina disposal, and secondary water quality.

Effect of pH

pH may have significant effects on arsenic removal with AA. A pH of 8.2 is significant because it is the “zero point charge” for AA. Below this pH, AA has a net positive charge resulting in a preference for adsorption of anions, including arsenic (AWWA, 1990). Acidic pH levels are generally considered optimum for arsenic removal with AA, however, some studies have presented conflicting effects of pH.

Several researchers have shown optimum pH for arsenic removal to be in the range of 5.5 to 6.0 for tests conducted on synthetic waters (Singer and Clifford, 1981; Rosenblum and Clifford, 1984). Others have also found improved performance at lower pH levels. Simms and Azizian (1997) found that incrementally lowering the pH from 7.5 to 6.0 increased the number of bed volumes which could be treated by 2 to 12 times. Hathaway and Rubel (1987) reported that the performance of AA for As(V) removal deteriorates as the pH increases from 6.0 to 9.0. Operating at an As(V) removal of 50 percent and at a pH of 5.5, a column treated 15,500 bed volumes (BVs). For the same level of As(V) removal, a column operating at pH 6.0 treated 13,391 BVs and a column operating at a pH of 9.0 treated only 800 BVs. Column studies conducted by Clifford and Lin (1985) also showed this trend. For a target arsenic effluent concentration of 0.05 mg/L, a column operating at a pH of 6.0 treated 8,760 BVs of water, but at pH of 7.3 the column treated only 1,944 BVs. In contrast to these results, Benjamin et al. (1998) found almost no dependence on pH level. The authors conducted isotherm and column studies with AA to investigate the removals of As(V) at pH 5.5, 7.0, and 8.5. Results indicated increasing pH from 5.5 to 8.5 had almost no effect on sorption of As(V) on AA.

Effect of Arsenic Oxidation State

Like nearly all other treatment technologies, the oxidation state of arsenic plays a large role in its removal; As(V) is much more easily adsorbed than As(III). Frank et al. (1986) conducted two column runs at pH 6. The influent in one run was 0.1 mg/L As(V) and in the other 0.1 mg/L As(III). The column treating water containing As(V) treated about 23,400 BVs before the effluent levels reached 0.05 mg/L. The other column showed a breakthrough of As(III) almost immediately and treated only 300 BVs before 0.05 mg/L was reached in the effluent. Benjamin et al. (1998) found that adsorption of As(V) was much faster than adsorption of As(III). The authors also showed that sorption onto AA was relatively rapid during the first few hours of exposure and slower thereafter. The ratio of As(V) adsorption densities at 2 and 24 hours was approximately 88 percent, whereas the ratio of As(III) adsorption densities was approximately 60 percent.

Effect of Competing Ions

Like ion exchange processes, AA exhibits preference for some ions. Interestingly, AA tends to have increased preference for ions which ion exchange does not. AA, however tends to be specific for arsenic and is not as greatly affected by competing ions (AWWA, 1990). As is indicated by the general selectivity sequence shown below (Clifford and Lin, 1995), AA preferentially adsorbs $H_2AsO_4^-$ [As(V)] over H_3AsO_3 [As(III)]:

$$OH^- > H_2AsO_4^- > Si(OH)_3O^- > F^- > HSeO_3^- > TOC > SO_4^{2-} > H_3AsO_3$$

Several studies have illustrated the effects of this selectivity, particularly those associated with sulfate and chloride. Benjamin et al. (1998) found little effect produced by either sulfate or chloride. Increasing sulfate from 0 to 100 mg/L had only a small impact on the sorption of As(V). The presence of chloride also did not affect As(V) removal. The addition of organics, however, had a much greater effect. The addition of 4 mg/L DOC reduced As(V) sorption onto AA by about 50 percent.

Clifford and Lin (1986) found significant effects of sulfate and total dissolved solids (TDS) on adsorption. They found the addition of 360 mg/L of sulfate and almost 1,000 mg/L TDS decreased the sorption of As(V) onto AA by approximately 50 percent compared to sorption from deionized water. Rosenblum et al. (1984) also reported that sulfate and chloride significantly reduced arsenic removal in AA systems. Arsenic removal in a water containing approximately 530 mg/L of chloride was 16 percent less than that achieved in a deionized water, and the presence of 720 mg/L of sulfate resulted in more than 50 percent less arsenic removal than that achieved in deionized water.

Simms and Azizian (1997) reported competition with silicate. In this particular study, the AA media became saturated with silicate much more quickly than with arsenic. No de-sorption of silicate was observed after saturation.

<u>Effect of Empty Bed Contact Time</u>

The operation of AA beds, and in particular the EBCT, can also play a role in arsenic removal. EBCT represents the lenght of time in which the feed water is in contact with the AA medium. Benjamin et al. (1998) conducted AA column tests using arsenic-spiked water from Lake Washington. All the column tests were run by adjusting the feed solution to pH 7. Sampling ports at various points in the system allowed EBCTs ranging from 2.5 to 15 minutes to be tested. Low arsenic concentrations (i.e. <5 µg/L) were achieved for more than 2,000 hours of operation. Comparing EBCTs, the data show that adsorption increased slightly with increasing EBCT.

<u>Regeneration</u>

Regeneration of AA beds is usually accomplished using a strong base solution, typically concentrated NaOH. Relatively few BVs of regenerant are needed. After regeneration with strong base, the AA medium must be neutralized using strong acid; typically two percent sulfuric acid. Arsenic is more difficult to remove during regeneration than other ions such as fluoride (Clifford and Lin, 1995). Because of this, slightly higher base concentrations are used; typically 4 percent NaOH. Even at this increased concentration , however, not all arsenic may be eluted. Clifford and Lin (1986) found only 50 to 70 percent of arsenic was removed from the AA columns during regeneration. Other researchers have also documented the difficult regeneration of AA for arsenic. Regeneration tests conducted by Benjamin et al. (1998) indicated that exposure of the AA medium to 0.1 N NaCl or 0.2 N NaOH did not regenerate the AA to a significant extent. Arsenic recovery was limited and in most cases was less than 50 percent of the sorbed arsenic. Higher recoveries have been reported, however. Hathaway and Rubel (1987) found that 80 percent of the adsorbed arsenic was eluted using 1.0 to 1.25 M NaOH solution. Simms and Azizian (1997) found that up to 85% of the capacity of an AA bed could be recovered using NaOH.

Regeneration also affects successive bed life and efficiency. Bed life is shortened and adsorption efficiency is decreased by regeneration. Benjamin et al. (1998) found that arsenic breakthrough patterns from the AA columns using regenerated media were qualitatively similar to those using fresh media, but the removal efficiency declined slightly after each of two regenerations.

Clifford (1986) demonstrated that regeneration has a clearly negative effect on the adsorption capacity of activated alumina. The unrecovered As(V) and changes in the AA surface induced by the regeneration process may cause the length of the adsorption runs to decrease by 10 to 15 percent after each regeneration.

Resin Fouling

Much like ion exchange resin, AA media may be fouled. Fouling reduces the number of adsorption sites thus decreasing removal effectiveness.

Hydraulic considerations should also be given. During treatment, AA media may become clogged with suspended solids present in the feed water. This can result in increased headloss across the bed. If the headloss buildup is significant, the media must be backwashed to removed the solids. Simms and Azizian (1997) found that headloss buildup across the bed after 75,000 BVs treated was minimal for a groundwater with 2 mg/L suspended solids and which was not pre-filtered.

In addition to suspended solids, Clifford and Lin (1995) note that silica and mica are particularly problematic foulants. In a study performed in Hanford, California, mica fouling was found to be a significant problem (Clifford and Lin, 1986).

Operational Considerations

Experience with AA processes is limited and full-scale applications are virtually non-existent. Therefore a large amount of information still needs to be obtained. The operational experience which has been developed, however, provides important information to be considered for AA process; these are discussed here.

AA beds may be operated in series or parallel. Series operation increases removal and helps prevents leakage, but limits throughput (leakage simply refers to elevated levels of arsenic in the effluent). Parallel operation on the other hand increases throughput, but does not improve effluent quality (AWWA, 1990). When operated in series, a "merry-go-round" configuration is often used. This configuration uses three beds: two in production and one in regeneration mode at a given time. When exchange capacity of the first bed in series is exhausted, the first bed is removed from service to be regenerated. The second bed in series then becomes the first and a fresh regenerated bed is brought on-line to become the second. This allows the maximum exchange capacity of beds to be used and prevents leakage since a fresh bed is always last in line. This also helps minimize regeneration frequency.

Degradation of AA media must also be considered. Alumina tends to dissolve over successive cycles due to the strong base/strong acid cycling during regeneration. As a result of this, alumina beds may become "cemented" if close care is not given (EPA, 1994). Backwashing the AA media may help in preventing cementation. Another important consideration is operator involvement. Strong acid and strong base are handled on a frequent basis and can present a safety hazard. An operator must be capable of handling these chemicals and must have a good understanding of pre-treatment, post-treatment, and regeneration practices if the process is to be operated efficiently. This presents a problem particularly for small systems.

Disposal Issues

Disposal of both spent regenerant and spent media is an important issue with arsenic removal using AA. Spent regenerant can contain high levels of arsenic. Simms and Azizian (1997) documented 20 to 40 mg/L of arsenic in spent regenerant liquid. Although little work has been done in this area, it has been speculated that the spent AA media would pass toxicity tests and could be landfilled. It is doubtful if spent regenerant could be discharge directly to the a sanitary sewer. There is evidence, though, that spent regenerant may be effectively treated prior to disposal (AWWA, 1990). This is possible because, during regeneration and acidification, enough aluminum dissolves to make precipitation of $Al(OH)_3$ a potential treatment. Arsenic is removed through its coprecipitation with the solid aluminum hydroxide. Arsenic is removed via the aluminum sludge which can be subsequently dried and landfilled if toxicity limits are not exceeded.

Although the possibility of regenerant reuse exists, it may not be feasible for arsenic removal. Direct reuse would probably not be possible due to the strong affinity of AA for arsenic. In other words, arsenic in the reused regenerant may actually be added to the column during regeneration. Spent regenerant, however, may be treated prior to reuse. By precipitating the arsenic from the regenerant, reuse may be possible assuming the regenerant solution was replenished and remained concentrated enough to replenish the AA bed.

Secondary Effects

AA processes will produce changes to the effluent water quality (EPA, 1994). Because pre-treatment is typically used to reduce the pH to low levels (less than 6.0) to optimize the process, the effluent pH will be less than typically desired in the distribution system. For this reason, post-treatment to raise the pH would be necessary. Another important effect of AA is the increased TDS levels of the effluent. AA processes tend to increase TDS levels which may have important implications for some utilities, such as corrosion issues or water quality issues.

2.3.2 Iron Oxide Coated Sand

Iron oxide coated sand (IOCS) is a rare process which has shown some tendency for arsenic removal. IOCS consists of sand grains coated with ferric hydroxide which are used in fixed bed reactors to remove various dissolved metal species. The metal ions are exchanged with the surface hydroxides on the IOCS. IOCS exhibits selectivity in the adsorption and exchange of ions present in the water. Like other processes, when the bed is exhausted it must be regenerated by a sequence of operations consisting of rinsing with regenerant, flushing with water, and neutralizing with strong acid. Sodium hydroxide is the most common regenerant and sulfuric acid the most common neutralizer.

Several studies have shown that IOCS is effective for arsenic removal. Factors such as pH, arsenic oxidation state, competing ions, EBCT, and regeneration have significant effects on the removals achieved with IOCS.

Effect of pH

pH appears to have an effect on arsenic adsorption by IOCS. Benjamin et al. (1998) conducted isotherm and column studies with IOCS to investigate the removals of As(V) at various pH levels. Results indicated that increasing the pH from 5.5 to 8.5 decreased the sorption of As(V) by approximately 30 percent.

Effect of Arsenic Oxidation State

As with other processes, the oxidation state of arsenic plays a role in its removal: As(V) appears to be more easily removed than As(III). Benjamin et al. (1998) showed that As(V) sorption onto IOCS was much more rapid than As(III) sorption during the first few hours of exposure and slower thereafter. The ratio of As(V) adsorption densities at 2 and 24 hours was approximately 60 percent, whereas the ratio of As(III) adsorption densities was only about 50 percent.

Effect of Competing Ions

Concentrations of competing ions will be an important consideration for arsenic removal with IOCS. Benjamin et al. (1998) evaluated the effect of sulfate and chloride on IOCS arsenic adsorption. They found that increasing sulfate from 0 to 100 mg/L had only slight impact on the sorption of As(V), and the presence of chloride did not appear to affect As(V) removal. Organic matter, however, did appear to present some competition for arsenic. The addition of 4 mg/L DOC reduced As(V) sorption by about 50 percent.

Effect of Empty Bed Contact Time

The EBCT can effect the arsenic removal efficiency of IOCS. Benjamin et al. (1998) conducted continuous flow IOCS column tests using arsenic-spiked water from Lake Washington. All tests were run by adjusting the feed solution to pH 7. Sampling ports at various points in the system allowed EBCTs ranging from 2.5 to 15 minutes to be tested. Low arsenic concentrations (i.e. <5 µg/L) were achieved for more than 2,000 hours of operation. Adsorption seemed to increase slightly with increasing EBCT. Based on the adsorption density at complete breakthrough, the initial capacity of the IOCS for either As(V) or As(III) was between 175 and 200 µg As/mL of media.

Regeneration

Regeneration of IOCS is performed in a similar fashion to that performed with activated alumina processes. Regeneration is accomplished using a strong base, typically NaOH, and subsequent neutralization is accomplished using strong acid, typically H_2SO_4. Regeneration tests conducted by Benjamin et al. (1998) indicated that exposure of the IOCS medium to 0.1 N NaCl or 0.2 N NaOH did not regenerate IOCS to a significant extent. Arsenic recovery was limited and in most cases was less than 50 percent of the sorbed arsenic. The arsenic breakthrough patterns from the IOCS columns using regenerated media were qualitatively similar to those using fresh media, but the removal efficiency declined slightly after each of two regeneration steps.

2.4 ION EXCHANGE

2.4.1 Introduction

Ion exchange (IX) is a physical/chemical process by which an ion on the solid phase is exchanged for an ion in the feed water. This solid phase is typically a synthetic resin which has been chosen to preferentially adsorb the particular contaminant of concern. To accomplish this exchange of ions, feed water is continuously passed through a bed of ion exchange resin beads in a downflow or upflow mode until the resin is exhausted. Exhaustion occurs when all sites on the resin beads have been filled by contaminant ions. At this point, the bed is regenerated by rinsing the IX column with a regenerant - a concentrated solution of ions initially exchanged from the resin. The number of bed volumes (BVs) that can be treated before exhaustion varies with resin type and influent water quality.

Typically from 300 to 60,000 BVs can be treated before regeneration is required. In most cases, regeneration of the bed can be accomplished with only 1 to 5 BVs of regenerant followed by 2 to 20 BVs of rinse water.

Important considerations in the applicability of the IX process for removal of a contaminant include water quality parameters such as pH, competing ions, resin type, alkalinity, and influent arsenic concentration. Other factors include the affinity of the resin for the contaminant, spent regenerant and resin disposal requirements, secondary water quality effects, and design operating parameters.

2.4.2 Effect of pH

The chloride-arsenate exchange chemical reaction typically occurs in the range of pH 8 to 9 when using chloride-form, strong-base resins (Clifford and Lin, 1995). IX removals with strong-base resins, though, is typically not sensitive to pH in the range of pH 6.5 to 9.0 (Clifford, et al., 1998). Outside of this range, however, arsenic removal decreases quickly. Groundwaters which are naturally contaminated with arsenic typically exhibit fairly high pH giving IX a slight advantage for these types of source water. Adjustment of pH prior to IX for arsenic removal is generally not necessary.

2.4.3 Effect of Competing Ions

Competition from background ions for IX sites can greatly affect the efficiency, as well as the economics, of IX systems. The level of these background contaminants may determine the applicability of IX at a particular site. Typically, strong-base anion exchange resins are used in arsenic removal. Strong-base anion resins tend to be more effective over a larger range of pH than weak-base resins. The order of exchange for most strong-base resins is given below, with the adsorption preference being greatest for the constituents on the far left.

$$HCrO_4^- > CrO_4^{2-} > ClO_4^- > SeO_4^{2-} > SO_4^{2-} > NO_3^- > Br^- > (HPO_4^{2-}, HAsO_4^{2-}, SeO_3^{2-}, CO_3^{2-}) > CN^-$$
$$> NO_2^- > Cl^- > (H_2PO^{4-}, H_2AsO_4^-, HCO_3^-) > OH^- > CH_3COO^- > F^-$$

These resins have a relatively high affinity for arsenic in the arsenate form ($HAsO_4^{2-}$), however, previous studies have shown that high TDS and sulfate levels compete with arsenate and can reduce removal efficiency (AWWA, 1990). In general, ion exchange for arsenic removal is only applicable for low-TDS, low-sulfate source waters. Source waters with TDS levels above 500 mg/L and sulfate levels above 25 mg/L are not recommended. Previous studies have confirmed this generalization; the low-sulfate/low-TDS source water in a Hanford, CA study proved to be amenable to IX treatment whereas the high-sulfate/high-TDS source water in a San Ysidro, NM study proved to be impractical for IX treatment (Clifford and Lin, 1986; Clifford and Lin, 1995).

If nitrate removal is being performed concurrent with arsenic removal, sulfate level can also be an important factor in arsenic removal. Clifford and others (1998) have shown that when sulfate levels are low (about 40 mg/L), the number of BVs to exhaustion is limited by nitrate breakthrough. If the sulfate level is high (about 100 mg/L), however, the number of BVs to exhaustion is limited by arsenic breakthrough. In other words, sulfate competes with both nitrate and arsenic, but competes more aggressively with arsenic than nitrate.

The presence of iron, Fe(III), in feed water can also affect arsenic removal. When Fe(III) is present, arsenic may form complexes with iron. These complexes are not removed by IX resins and therefore arsenic is not removed. Utilities with source waters high in Fe(III) may need to address this issue for IX use or evaluate other treatment techniques for arsenic removal (Clifford, et al., 1998).

When an ion is preferred over arsenate, higher arsenic levels in the product water than exist in the feed water can be produced. If a resin prefers sulfate over arsenate, for example, sulfate ions may displace previously sorbed arsenate ions and result in levels of arsenic in the effluent which are greater than the arsenic level in the influent. This is often referred to as chromatographic peaking. As a result, the bed must be monitored and regenerated well in advance of the onset of this peaking. Clifford and Lin (1995) recommend operating the bed to a known BV setpoint to avoid peaking.

2.4.4 Resin Type

As stated earlier, strong-base resins are typically used in IX arsenic removal. These resins, however, tend to prefer some ions, sulfate and chloride in particular, over arsenate. As mentioned in Section 2.4.2, this can result in chromatographic peaking if beds are not monitored adequately. Recent studies have also found that sulfate-selective resins tend to be superior to nitrate-selective resins for arsenic removal (Clifford, et al., 1998). Future research, however, may produce monovalent-selective resins which will be arsenate-selective and may eliminate non-arsenic ion competition (EPA, 1994).

Many resins are available for arsenic removal. Some of the commercially available resins which have been used in relevant IX studies are summarized in Table 2-1. Data in Table 2-1 represent BVs to exhaustion using virgin IX resins. It should be noted, however, that the capacity of the bed may decrease slightly over time. Choice of resin will ultimately be site-specific, making preliminary studies a necessity to determine optimum resin type.

2.4.5 Process Configuration

Properly configuring IX columns can improve arsenic removal and help minimize regeneration frequency. This is due to the fact that arsenic "leakage" will often occur in IX columns. In some situations, series operation or implementation of certain operating methods may be needed to achieve low arsenic levels.

Series operation, also known as "merry-go-round" operation, uses three beds: two in production and one in regeneration mode at a given time. When exchange capacity of the first bed in series is exhausted, the first bed is removed from service to be regenerated. The second bed in series then becomes the first and a fresh regenerated bed is brought on-line to become the second. This allows the maximum exchange capacity of beds to be used and prevents leakage since a fresh bed is always last in line. This also helps minimize regeneration frequency (EPA, 1995).

Another approach for minimizing effluent levels is to operate IX columns in "counter-current flow" operation. In this mode, feed water is applied in one direction (i.e. downward) and the regenerant is applied in the opposite direction (i.e. upward). This minimizes leakage from the column. Typically columns are designed for "co-current flow" operation where the feed water and regenerant are applied in the same direction. Co-current operation increases chances for leakage, however, since regeneration in this mode concentrates the contaminant on the effluent end of the IX column. Using the "counter-current flow" method also minimizes regenerant requirements, i.e. volume and concentration (EPA, 1995).

TABLE 2-1

Typical IX Resins for Arsenic Removal

Resin	Type	Operating Parameters	BVs to Exhaustion	Reference
Dowex 11	2	Bed Volume: 0.8 cu ft Flowrate: 1 gpm EBCT: 5.6 min Depth: 2.5 - 5 ft Sulfate/As Ratio: 60:1 TDS/As Ratio: 2500:1	4,200	Clifford and Lin (1986)
Ionac ASB-2	2	Bed Volume: 0.8 cu ft Flowrate: 1 gpm EBCT: 5.6 min Depth: 2.5 - 5 ft Sulfate/As Ratio: 60:1 TDS/As Ratio: 2500:1	4,940	
Dowex SBR-1	1	Bed Volume: 1.0 cu ft Flowrate: 1 gpm EBCT: 7.5 min Depth: 3.8 ft Sulfate/As Ratio: NR TDS/As Ratio: NR	2,800	Hathaway and Rubel (1987) Fox (1989)
Ionac ASB-1 and Dowex 11	1	Bed Volume: 0.014 cu ft Flowrate: NR EBCT: NR Depth: NR Sulfate/As Ratio: 420:1 TDS/As Ratio: 9200:1	• 200 • 400-500 (projected if oxidation to As(V) is preformed)	Clifford and Lin (1985)
• A-300E (bench-scale) • A-300E (full-scale)	2	Bed Volume: 0.0018 cu ft Flowrate: 0.035 gpm EBCT: NR Depth: 1.33 ft Sulfate/As Ratio: 300:1 TDS/As Ratio: NR	• 1,340-1,640 • 5,000-7,000	Malcolm Pirnie (1992)

NR = Not Reported

2.4.6 Secondary Effects

Chloride-form resins are often used in arsenic removal. Chloride ions are displaced from the column as contaminants (arsenic) are sorbed onto the column. As a result, the potential exists for increases in the chloride concentration of the product water. Increases in chlorides can greatly increase the corrosivity of the product water. Chlorides increase the corrosion potential of iron and as a result increase the potential for red water problems (EPA, 1995). Corrosion problems are worsened when high chloride levels are intermittent. In situations where chlorides pose a problem, demineralization, blending, or alternate treatment techniques may be required.

Also, effluent pH may be lowered as a result of IX treatment. pH of the product water may be less than 7 at the beginning of a cycle. Again, decreases in pH may increase the corrosivity of the effluent. In some situations, pH restabilization may be necessary to prevent disturbances in the distribution system.

2.4.7 Resin Fouling

IX resin beads may be fouled if appropriate pretreatment is not practiced. Generally, fouling of IX resins is caused by scaling of minerals (i.e. Ca) or by particulates in the feed stream. Iron precipitates have also been known to cause resin fouling (Malcolm Pirnie, 1993a). If scaling is a problem, chemical addition may be needed to lower the scale-forming potential of the feed water. If suspended solids are found in the feed stream, multi-media filtration ahead of IX columns may be necessary. A previous study performed in Hanford, California found that IX resin was significantly fouled by mica present in the source water. This was determined by observing a 3-5 percent decrease in total BVs to exhaustion over consecutive cycles and by observing a black coating on the exhausted resin. Most, but not all, of the black coating could be removed from the resin beads during the NaCl regeneration cycle (Clifford and Lin, 1986).

2.4.8 Regeneration

With chloride-form resins, concentrated NaCl solution is typically used as the regenerant. Only a few number of BVs of regenerant are usually required to replenish the resin, depending on the solution strength. Arsenic elutes readily from IX columns, regardless of resin type, mainly because it is a divalent ion and as such is subject to selectivity reversal in high ionic strength (> 1M) solution (Clifford and Lin, 1995). Clifford and Lin also found that dilute regenerants tend to be more efficient than concentrated regenerants in terms of the ratio of regenerant equivalents to resin equivalents. For example, they found that two resins (Dowex-11 and Ionac ASB-2) could be regenerated equivalently using either 2 BVs of 1.0 N NaCl or 5 BVs of 0.25 N NaCl in “co-current flow” operation. Also, a rinsing cycle is required after regeneration; typically only a few BVs are required for rinsing as well.

2.4.9 Regenerant Reuse and Treatment

Spent regenerant is produced during IX bed regeneration. Typically this spent regenerant will have high concentrations of arsenic and other sorbed contaminants. Spent regenerant must be treated and/or disposed of appropriately. This can be an expensive part of the IX process and must be given thorough consideration. Spent brine can be disposed of either directly to a surface water source, or indirectly to a sanitary sewer, depending on contaminant levels. Spent solution can also be treated and disposed of as described at the end of this subsection. Spent regenerant, however, may be reused many times. Clifford and others (1998) estimate that regenerants may be used 25 times or more before treatment and disposal are required. Regenerants do not need treatment prior to reuse, except to replenish the chloride concentration to maintain a 1 M solution. Once the contaminant concentration becomes too high in the regenerant, however, the spent solution must be treated and/or disposed.

Treatment of spent regenerant is accomplished in a number of ways. First, the spent regenerant can be dewatered in some fashion. Common methods of dewatering IX residuals include mechanical dewatering, drying beds, gravity thickeners, and lagoon dewatering. The solids generated by these processes would need to be tested for toxicity and disposed accordingly. If determined to be non-toxic according to disposal regulations, the dried solids could be landfilled. Waste liquid generated by these drying processes could be either directly discharged to a surface water source or indirectly discharged to the sanitary sewer, depending on contaminant levels (Malcolm Pirnie, 1996). Second, spent brine can be treated by precipitation. Clifford and Lin (1995) have shown that arsenic levels can be substantially reduced using iron and aluminum coagulants as well as lime. Much greater than the stoichiometric amounts (up to 20 times as much), however, are needed in actual practice to reduce arsenic to low levels. In addition, pH adjustment may be necessary to ensure optimum coagulation conditions. Reductions from 90 mg As(V)/L to less than 1.5 mg As(V)/L have been seen using iron and aluminum metal salts (Clifford and Lin, 1995). Both coagulant types seem to work well, however, iron precipitates tend to settle better due to their weight. Dried sludge from brine reduced to 1.5 mg As(V)/L using precipitation passed an EP toxicity test with only 1.5 mg/L As(V) in the leachate. In this situation, dried sludge could have been disposed of in a landfill. Reuse of decontaminated regenerant has yet to be evaluated, however, the possibility of reuse does exist.

2.4.10 EBCT

A few studies have been performed to test the effect of EBCT on IX performance. Clifford and Lin (1986) reduced EBCT from 5 to 1.4 in a Hanford, CA study and found no significant reduction in arsenic removal performance. In a recent AWWARF study, four IX columns were run with EBCTs varying between 2.5 and 15 minutes. Data from this study show that the shorter the EBCT, the more BVs can be treated before breakthrough. The disadvantage to shorter EBCT, however, is increased regeneration frequency. Based on these data, shorter EBCTs may be preferred to reduce capital costs (AWWARF, 1998).

2.4.11 Typical Design Parameters

Through extensive research, Clifford and others (1998) assembled typical operating parameters and suggested options for ion exchange processes. Although many design parameters should be tailored to the specific treatment situation, Table 2-2 gives typical values and options.

TABLE 2-2

Typical Operating Parameters and Options for IX

1.5 minute EBCT (15 gpm/ft^2 at 3 ft/day)
0.5 - 1.0 M NaCl (1-2 eq Cl$^-$/eq resin)
Operate the column to a fixed BV endpoint (to prevent leakage)
Regenerant Surface Loading Velocity should be greater than 2 cm/min
Regenerant may be used 25 times or more (with Cl$^-$ concentration of 1 M maintained)
Ferric coagulant should be used for $Fe(OH)_3$•As from regenerant waste

2.5 MEMBRANE PROCESSES

2.5.1 Introduction

Membranes are a selective barrier, allowing some constituents to pass and blocking the passage of others. The movement of constituents across a membrane requires a driving force (i.e. a potential difference between sides of the membrane). Membrane processes are often classified by the type of driving force, including pressure, concentration, electrical potential, and temperature. The processes discussed here include only pressure-driven and electrical potential-driven types.

Pressure-driven membrane processes are often classified by pore size into four categories: microfiltration (MF), ultrafiltration (UF), nanofiltration (NF), and reverse osmosis (RO). Typical pore size classification ranges are given in Figure 2-1. High-pressure processes (i.e., NF and RO) have a relatively small pore size compared to low-pressure processes (i.e., MF and UF). Typical pressure ranges for these processes are given in Table 2-3. NF and RO primarily remove constituents through chemical diffusion (Aptel and Buckley, 1996). MF and UF primarily remove constituents through physical sieving. An advantage of high-pressure processes is that they tend to remove a broader range of constituents than low-pressure processes. However, the drawback to broader removal is the increase in energy required for high-pressure processes.

TABLE 2-3

Typical Pressure Ranges for Membrane Processes

Membrane Process	Pressure Range
MF	5 - 45 psi
UF	7 - 100 psi
NF	50 - 150 psi
RO	100 - 150 psi

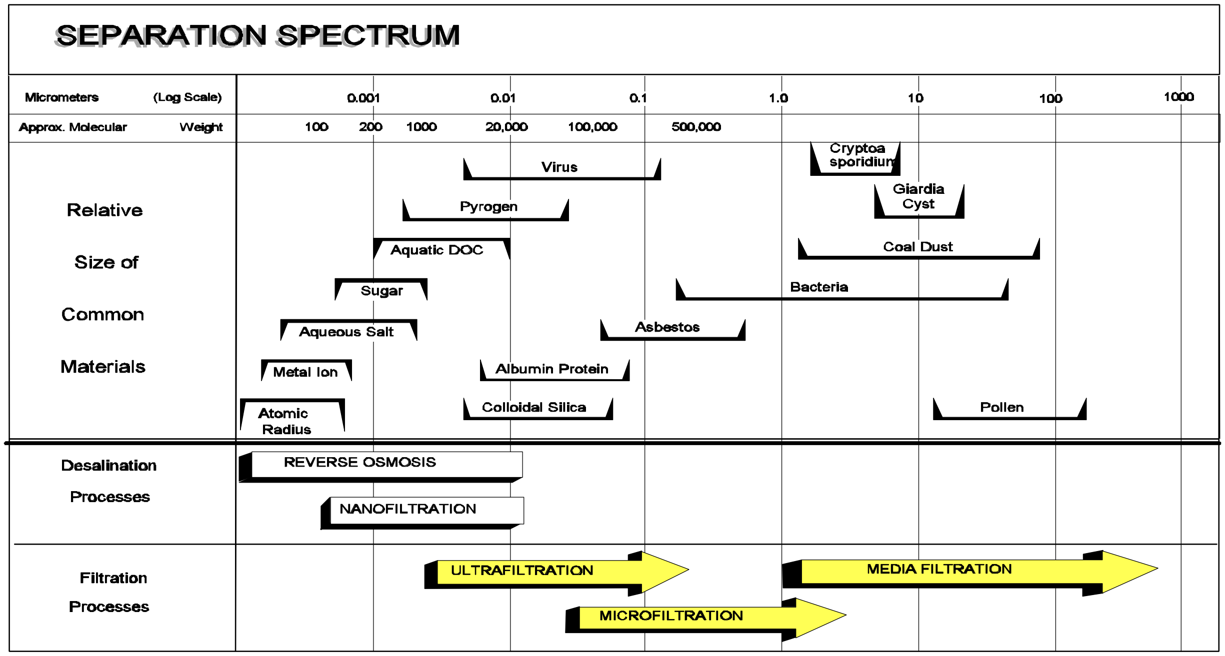

Figure 2-1 Pressure Driven Membrane Process Classification (Westerhoff and Chowdhury, 1996)

Electrical potential-driven membrane processes can also be used for arsenic removal. These processes include, for the purposes of this document, only electrodialysis reversal (EDR). In terms of achievable contaminant removal, EDR is comparable to RO. The separation process used in EDR, however, is ion exchange (Aptel and Buckley, 1996). EDR is discussed further in Section 2.5.8.

2.5.2 Important Factors for Membrane Performance

Commercial pressure-driven membranes are available in many types of material and in various configurations. The chemistry of the membrane material, in particular surface charge and hydrophobicity, play an important role in rejection characteristics since membranes can also remove contaminants through adsorption. Membrane configuration and molecular weight cut-off (MWCO), i.e. pore size, also influence rejection properties, as well as operational properties, to a great extent. These options must be chosen appropriately depending on source water characteristics and removal requirements.

Source water quality is also important in the selection of a membrane process. Water quality can have significant effects on membrane operation and rejection. Water temperature is very important to all membrane processes. Lower water temperatures will decrease the flux at any given pressure. To compensate, additional membrane area and/or higher feed pressures must be provided to maintain equivalent production at lower temperatures. Depending on source water quality, pretreatment is often necessary, particularly with the high-pressure processes. The small pore size of NF and RO membranes makes them more prone to fouling than UF or MF membranes. The application of NF and RO for surface water treatment is generally not accomplished without extensive pretreatment for particle removal and possibly pretreatment for dissolved constituents. The rejection of scale-causing ions, such as calcium, can lead to precipitation on the membrane surface. Organic compounds and metal compounds, such as iron and manganese, can promote fouling as well. Precipitation can result in irreversible fouling and must be avoided by appropriate pretreatment, including addition of anti-scaling chemical and/or acid to the feed water.

The percentage of product water that can be produced from the feed water is known as the *recovery*. Recovery for MF and UF is typically higher than recovery for RO and NF. The recovery is limited by the characteristics of the feed water and membrane properties. Typical recoveries for membrane processes are given in Table 2-4.

TABLE 2-4

Typical Recovery for Membrane Processes

Membrane Process	Recovery
MF	to 99%
UF	to 95%
NF	to 85%
RO	30-85%

2.5.3 Arsenic Removal with Membrane Processes

Membrane processes can remove arsenic through filtration, electric repulsion, and adsorption of arsenic-bearing compounds. If particulate arsenic compounds are larger than a given membrane pore size, they will be rejected due to size exclusion. Size, however, is only one factor which influences rejection. Studies have shown that some membranes can reject arsenic compounds which are one to two orders of magnitude smaller than the membrane pore size, indicating removal mechanisms other than just physical straining (AWWARF, 1998). Shape and chemical characteristics of arsenic compounds play important roles in arsenic rejection. Membranes may also remove arsenic compounds through repulsion by or adsorption on the membrane surface. These depend on the chemical characteristics, particularly charge and hydrophobicity, of both the membrane material and the feed water constituents. Many studies have been performed which evaluated various membrane processes for arsenic removal. These processes and corresponding research are discussed in the remainder of this section.

2.5.4 Microfiltration

Microfiltration's viability as a technique for arsenic removal is highly dependent on the size distribution of arsenic-bearing particles in the source water. MF pore size is too large to substantially remove dissolved or colloidal arsenic. Although MF can remove particulate forms of arsenic, this alone does not make the process efficient for arsenic removal unless a large percentage of arsenic is found in this form. Arsenic found in groundwater is typically less than 10 percent particulate while arsenic found in surface waters can vary from 0 percent to as much as 70 percent particulate (AWWARF, 1998; McNeill and Edwards, 1997). Unfortunately, the percentage of particulate arsenic does not seem to be related to specific water types. In a recent study, AWWARF (1998) did not find arsenic size distribution to correlate with turbidity or organic content, indicating that arsenic size distribution was specific to individual waters.

To increase removal efficiency in source waters with a low percentage of particulate arsenic content, MF can be combined with coagulation processes. Coagulation assisted microfiltration for arsenic removal is discussed in Section 2.2.3. For utilities using MF alone for particulate arsenic removal, removal would primarily depend on the influent arsenic concentration and percentage of particulate arsenic since the MF rejection mechanism is mechanical sieving. Therefore, the effectiveness of MF arsenic rejection is a function of pore size. Variation in MF performance is due to pore size distribution.

2.5.5 Ultrafiltration

Ultrafiltration processes are generally capable of removing some colloidal and particulate constituents, based upon the above discussion on particulate arsenic occurrence. Considering this, UF alone, like MF, may not be a viable technique for arsenic removal for groundwaters, however, UF may be appropriate for surface waters with high colloidal and particulate arsenic concentrations.

Recent research has found that electric repulsion of UF may play an important role in arsenic rejection and increase rejection beyond that achievable with only pore size-dependent sieving. AWWARF (1998) performed bench-scale tests on two low-MWCO UF membranes. Single element testing was performed on Desal GM and FV UF membranes for a spiked, deionized water. Flat sheet testing was also performed on Desal GM, FV, and PM UF membranes for spiked, deionized water. Since the samples were spiked, no particulate or colloidal arsenic was present. Results of this study are given in Table 2-5.

TABLE 2-5

As(V) and As(III) Removal by UF Membranes

Membrane Type	MWCO	Membrane Charge	Arsenic Species	pH	Total Arsenic Rejection (%)
Single Element					
GM2540F	8,000	(-)	V	6.9	63
GM2540F	8,000	(-)	V	2.0	8
GM2540F	8,000	(-)	III	7.2	<1
GM2540F	8,000	(-)	III	10.8	53
FV2540F	10,000	None	V	6.9	3
FV2540F	10,000	None	III	6.8	5
Flat Sheet					
GM	8,000	(-)	V		52
FV	10,000	None	V		NA
PW	10,000	None	V		5

NA: Not Available

For the negatively charged GM2540F membrane, As(V) rejection was high at neutral pH but very low at acidic pH. On the other hand, with the same membrane, As(III) rejection was high at basic pH and negligible at neutral pH. The uncharged FV2540F membrane showed poor rejection of both As(V) and As(III) at neutral pH. High rejection rates were seen even though the MWCO of the membranes were two orders of magnitude larger than the arsenic compounds (AWWARF, 1998). The authors theorize that the high rejection rates seen were due to electrostatic interaction between the negatively charged membrane surface and the arsenic ions. This will be pH dependent since the anionic As(V) and the nonionic As(III) will be charged (protonated/deprotonated) at different pH levels. In effect, membrane charge and pH may play an important role in arsenic rejection. In fact, the authors found that electrostatic repulsion becomes increasingly important moving from RO to NF to UF, while size exclusion becomes increasingly important moving from UF to NF to RO. The flat sheet testing produced rejection rates comparable, and slightly conservative, to the single element rejection rate. As with single element testing, the negatively charged membrane proved more effective for arsenic rejection than the neutral charged membrane.

AWWARF (1998) also performed UF pilot-scale tests. Single element pilot tests were performed on two groundwaters, one with a DOC level of 11 mg/L and one with a DOC level of 1 mg/L, and a spiked, finished surface water. Arsenic removal results from these tests are shown in Table 2-6.

TABLE 2-6

Arsenic Removal by UF at Pilot-Scale

Membrane	MWCO	Water Type	As Species	As Rejection
Desal GM2540F	8,000	High DOC GW	Total As	70%
Desal GM2540F	8,000	Low DOC GW	Total As	30%
		Finished SW	V	47%
			III	10%

As seen in Table 2-6, arsenic removal varied with DOC levels, being much higher in the high DOC groundwater (70%) than in the low DOC groundwater (30%). The authors postulated that this difference was due to a reduction in electrostatic forces caused by adsorption of NOM to the membrane surface. Adsorption of NOM would reduce the surface charge of the membrane and would, in effect, increase the repulsion towards negatively charged arsenic compounds. Increases in the apparent size of the arsenic molecules through "bridging" with humic substances was ruled out since a concurrent increase in UV_{254} removal was not seen. In contrast to DOC levels, changes in flux and recovery did not seem to impact the arsenic rejection rate. As shown in Table 2-6, testing on the finished surface water showed fairly effective removal of As(V), but unimpressive As(III) removal. Considering the MWCO, however, these removals were expected.

2.5.6 Nanofiltration

Nanofiltration membranes are capable of removing significant portions of the dissolved arsenic compounds in natural waters due to their small pore size. NF will primarily remove divalent ions (i.e. Ca, Mg), but not monovalent salts (i.e. Na, Cl). Through size exclusion, NF can remove both dissolved As(V) and As(III). This makes NF a reliable arsenic removal process for groundwater which contains up to 90% dissolved arsenic (AWWARF, 1998). The small pore size, however, makes NF membranes more prone to fouling than UF or MF membranes. The application of NF for surface water treatment is typically not accomplished without extensive pretreatment for particle removal and possibly pretreatment for dissolved constituents to prevent fouling.

Several NF studies for have been undertaken, and the results show that NF processes are effective for the removal of arsenic. Removal however depends on operating parameters, membrane properties, and arsenic speciation. AWWARF (1998) performed NF bench-scale studies for arsenic removal on spiked deionized water and on a lake water. Single element and flat sheet testing were performed on a negatively charged NF membrane for a lake water and a spiked, deionized water. Results are shown in Table 2-7.

As seen in Table 2-7, As(III) removal was low at only 12 percent. However, As(V) rejection for the negatively charged membrane was high at 89 and 85 percent for the lake water and deionized water, respectively. Flat sheet testing produced a comparable As(V) rejection 90 percent.

TABLE 2-7

As(V) and As(III) Removal by NF Membranes

Membrane Type	MWCO	Membrane Charge	Water Type	Species	pH	Total Arsenic Rejection (%)
Single Element						
NF 45-2540	300	(-)	DI	V	6.7	85
NF 45-2540	300	(-)	Lake	V	6.9	89
NF 45-2540	300	(-)	DI	III	6.9	12
Flat Sheet						
NF 45-2540	300	(-)	DI	V	NA	90

NA: Not Available

AWWARF also performed several single element and array NF pilot-scale tests. Two of these tests were conducted on groundwaters, one high in DOC (11 mg/L) and one low in DOC (1 mg/L). Another test was performed on spiked, high-DOC groundwater. One other test was performed on spiked, finished surface water. These tests are summarized in Table 2-8.

As shown in Table 2-8, during the single element tests on the groundwaters the membranes demonstrated substantial arsenic removal. Removal in the low DOC water, however, was only 60 percent compared to over 80 percent in the high DOC water. As discussed in Section 2.5.5, this was presumably due to changes in electrostatic repulsion at the membrane surface through NOM adsorption. As in the UF pilot study, NF arsenic rejection rate did not seem to be affected by changes in flux or recovery.

Single element tests performed on the spiked, finished surface water showed substantial As(V) rejection (>95 percent). As(III) rejection, however, was reduced with an average for all three membranes of only 40 percent. The authors point out that these results attest to the influence of diffusion and electrostatic repulsion on As(III) removal. As(III) is small and can more easily diffuse through very small NF pores. As(III) is also not as repulsed by surface charge as As(V). Combining NF with an oxidizing process to convert As(III) to As(V) would probably be the most effective option for its removal.

TABLE 2-8

Arsenic Removal with NF at Pilot-Scale

Membrane	MWCO	Water Type	Charge	As Species	As Rejection
Single Element					
Accumem	400	High DOC GW	(-)	Total As	80%
		Low DOC GW	(-)	Total As	60%
NF 1	NA	Finished SW	NA	V	>95%
				III	52%
NF 2	NA	Finished SW	NA	V	>95%
				III	20%
NF 3	NA	Finished SW	NA	V	>95%
				III	30%
Array					
Accumem	400	High DOC GW	(-)	Total As	75% (initial) 3-16% (final)

NA: Not Available

The array test results, as shown in Table 2-8, were somewhat surprising. Arsenic rejection rate declined over time. Rejection at the beginning of the test was approximately 75 percent but proceeded to decline to 11 percent by day 60. Rejection stayed between 3 percent and 16 percent for the remainder of the 80-day period. This was surprising given the fact that the membrane showed high arsenic rejection in single-element tests. Samples taken throughout the array indicate that a speciation change from As(V) to As(III) was taking place within the filter. Since As(III) is more difficult to remove than As(V), overall arsenic removal dropped. This decrease in rejection over time presents doubt that a negatively charged membrane could keep high As (V) rejection rates for long durations without maintaining arsenic in the As(V) form. It is evident that more long-term testing is needed to verify these results for other membranes and situations. If speciation changes are influential for arsenic removal, keeping the membrane surface in an oxidized state may be an option.

A NF pilot-scale study to determine arsenic removals with NF membranes was conducted in Tarrytown, NY (Malcolm Pirnie 1992). Two NF membranes were tested: 1) NF70 manufactured by

Dow Chemical Company (FilmTec) and 2) TFCS manufactured by UOP Fluid Systems. The NF membranes were operated at a flux varying between 17 and 21 gfd and at a recovery of 15 percent. Feed water conductivity varied from 460 to 950 uS, pH ranged from 7.7 to 8.3, and feed water arsenic ranged from 0.038 - 0.154 mg/L. A second feed solution was mixed that had approximately twice the TDS and arsenic levels as found in the original test solution to simulate arsenic rejections by the last element in an NF membrane system operating at 50 percent recovery. Arsenic rejection was very high with only one of eight permeate samples from the NF membranes exceeding the detection limit with a level of 0.0025 mg/L, corresponding to 95% rejection.

2.5.7 Reverse Osmosis

RO is the oldest membrane technology which has traditionally been used for desalination of brackish water and sea water. RO produces nearly pure water by maintaining a pressure gradient across the membrane greater than the osmotic pressure of the feed water. Osmotic pressure becomes great in RO systems compared to other membrane processes due to the concentration of salts on the feed side of the membrane. The majority of the feed water passes through the membrane, however, the rest is discharged along with the rejected salts as a concentrated stream. Discharge concentrate can be substantial, between 10 and 50 percent of the influent flow, depending on influent water quality and membrane properties.

RO performance is adversely affected by the presence of turbidity, iron, manganese, silica, scale-producing compounds and other constituents. Like NF, RO requires extensive pretreatment for particle removal and often pretreatment for dissolved constituents. RO often requires pretreatment even for high quality source waters. RO has sometimes been used as a polishing step for already treated drinking water. Pretreatment can make RO processes costly. Treated waters from RO systems typically have extremely high quality, however, and blending of treated water and raw water can be used to produce a finished water of acceptable quality. This may reduce cost to some extent.

RO is an effective arsenic removal technology proven through several bench- and pilot-scale studies. RO is very effective for removing dissolved constituents including dissolved arsenic. Since the arsenic found in groundwater is typically 80 to 90 percent dissolved, RO is a suitable technology for arsenic removal in groundwater. Several previous RO bench-scale and pilot-scale studies for arsenic removal are summarized in Table 2-9. These studies indicate that RO can be an effective process for arsenic removal, however, membrane type and operating conditions will affect removal and must be chosen appropriately. As with other processes, RO removes As(V) to a greater degree than As(III), so maintaining oxidization conditions may be important to the process.

AWWARF (1998) performed bench- and pilot-scale RO testing. Short-term, single element testing and flat sheet testing were performed for a DK2540F RO membrane manufactured by DESAL on a lake water and on spiked deionized water. Results from this testing are shown in Table 2-10.

TABLE 2-9

Summary of Arsenic Removal with RO

Location	Type	Operating Parameters	As Removal	Reference
Eugene, OR	POU	• 3-5 gpd • 90% recovery • 20-100 psi	50%	Fox, 1989 Fox and Sorg, 1987
Eugene, OR	POU	• 3-5 gpd • 67% recovery • 195 psi	below MDL	Fox, 1989 Fox and Sorg, 1987
Fairbanks, AL	POU	• low-pressure (< 100 psi)	50%	Fox, 1989 Fox and Sorg, 1987
San Ysidro, NM	Pilot (hollow fiber, cellulose acetate)	• 50% recovery • pH adjustment to 6.3 • antiscalent addition	93-99%	Clifford and Lin, 1991
San Ysidro, NM	Pilot (hollow fiber, polyamide)	• 50% recovery • pH adjustment to 6.3 • antiscalent addition	99%	Clifford and Lin, 1991
San Ysidro, NM	POU	• 10-15% recovery	91%	Fox, 1989 Fox and Sorg, 1987
Tarrytown, NY	Pilot (FilmTec BW30, Hydranautics NCM1, Fluid Systems TCFL)	• 15 gfd • 10% recovery	below MDL	Malcolm Pirnie, 1992
Tarrytown, NY	POU	NA	86%	Rogers, 1989
Charlotte Harbor, FL	POU (several membrane types)	• 1000 gpd • 10-60% recovery	As(V) 96-99% As(III) 46-84%	Huxstep, 1987
Cincinnati, OH	POU	NA	As(III) 73%	Fox and Sorg, 1987
Hudson, NH	POU	NA	40%	USEPA, 1982

NA: Not Available

TABLE 2-10

Arsenic Removal with RO at Bench-Scale

Membrane	MWCO	Source Water	Speciation	pH	As Rejection
Single Element					
DK2540F	180	Deionized	V	6.8	96%
DK2540F	180	Lake Water	V	6.9	96%
DK2540F	180	Deionized	III	6.8	5%
DK2540F	180	Lake Water	III	6.8	5%
Flat Sheet					
DK2540F	180	Deionized	V		88%

These results indicate very high rejection for As(V) but very low rejection for As(III) at neutral pH. Again, this points to the fact that oxidation conditions would be desirable and that surface charge/electrostatic repulsion probably plays a role in arsenic rejection. Also, flat sheet testing produced a rejection rate comparable, and slightly conservative, to the single element rejection rate.

Several RO pilot-scale tests were also performed (AWWARF, 1998). Two tests were performed on high- and low-DOC groundwaters. Another set of tests was performed on spiked, finished surface water. The results from these pilot tests are summarized in Table 2-11.

Table 2-11 shows substantial rejection for both the low- and high-DOC waters. Rejection was only slightly higher with the high DOC water. As with UF and NF, flux and recovery changes did not seem to affect arsenic rejection. Results for the four membranes tested on spiked finished water also showed substantial removal. For all membranes during this test, As(V) exceeded 95 percent, however, As(III) rejection averaged only 74 percent.

TABLE 2-11

Arsenic Removal with RO at Pilot-Scale

Membrane	Water Type	Charge	As Species	As Rejection
Single Element				
TFCL-HR	High DOC GW	(-)	Total As	>90%
	Low DOC GW	(-)	Total As	>80%
RO 1	Finished SW	NA	V	>95%
			III	60%
RO 2	Finished SW	NA	V	>95%
			III	75%
RO 3	Finished SW	NA	V	>95%
			III	68%
RO 4	Finished SW	NA	V	>95%
			III	85%

NA: Not Available

Overall, RO is capable of achieving finished water arsenic concentrations below 0.002 mg/L when arsenic is present as As(V). As(III) rejection is not as significant, however, conversion to As(V) can be achieved with pre-oxidation.

2.5.8 Electrodialysis Reversal

Electrodialysis (ED) is a process in which ions are transferred through membranes that are selectively permeable towards cations or anions under the influence of direct electric current. The separation mechanism is actually an ion exchange process. The ions travel from a lesser to a higher concentrated solution. In this process, the membranes are arranged in an array or stack placed between opposite electrodes, with alternating cation and anion exchange membranes. The mobility of the cations or anions is restricted to the direction of the attracting electrodes, and this results in alternating sets of compartments containing water with low and high concentrations of the ions. The electrodialysis reversal (EDR) process is an ED process with periodic reversal of the direction of travel of the ions caused by reversing the polarity of the electrodes. The advantage of polarity reversal is the decreased potential for fouling of the membranes, which also minimizes the pretreatment requirements of this process.

EDR is designed specifically for each application based on the desired quantity and quality of product water. Equipment at an EDR plant, besides the stack itself, includes feedwater pumps, recycle pumps, valving, stream switching, product water diversion, pressure regulation, and electrode stream control. EDR systems are fully automated and require little operator attention, with the exception of data collection and routine maintenance. Routine maintenance consists of changing cartridge filters, calibrating and maintaining instruments, replacing membranes, maintaining pumps and valves, and replacing electrodes. EDR systems are also attractive since they do not require chemical addition (EPA, 1994). EDR systems, however, are typically more expensive than NF and RO systems (EPA, 1994). EDR systems are often used in treating brackish water to make it suitable for drinking. In terms of effluent water quality, EDR has been compared to RO (AWWARF, 1996; Robinson, et al., 1998). EDR processes have also been applied in the industry for wastewater recovery.

EDR can achieve high removals of TDS from water and typically operates at a recovery of 70 to 80 percent (Kempic, 1994a). Very few studies have been conducted to exclusively evaluate this process for the removal of arsenic. One of the studies was conducted using EDR to treat water from San Ysidro, New Mexico, which was a site for several other arsenic removal studies (Clifford and Lin, 1985). Studies by a leading manufacturer of EDR equipment also provide data on arsenic removal (Ionics Inc., 1989-1990). These are discussed below.

In the San Ysidro EDR study, a recovery of 85 percent was achieved by using an internal brine recycle system. Pretreatment for the unit consisted of a standard 10-micron cartridge filter and a granular activated carbon (GAC) column that were part of the system provided by the manufacturer. The unit was tested for two different waters, a city water that contained a mixture of As(III) and As(V), and a groundwater that contained mostly As(III). The well water contained 0.188 mg/L of arsenic. The groundwater was nearly all As(III). Arsenic removals by EDR were low, at only 28 percent, and the effluent concentrations were high at 0.136 mg/L.

The city water quality is shown in Table 2-12. The unit was run for 5 days with a recovery of 81 percent. The overall removal of arsenic was estimated at 73 percent. Approximately 60 percent of the As(III) was removed, which was higher than expected, based on the assumption that 90 percent of As(V) was removed.

Another mobile unit equipped with RO, ED, and EDR systems, along with the necessary pretreatment and post-treatment equipment, was used to test waters from eight New Mexico communities (New Mexico State University, 1979). In one of the studies conducted at Bluewater, New Mexico, EDR brought the level of arsenic in the treated water down to 0.003 mg/L from the influent level of 0.021 mg/L. This corresponds to a removal of approximately 86 percent. The feed water to the EDR unit was drawn from a point before chlorination of the community water supply. The test flow rate was 4.8 gpm, and 80 percent recovery was obtained. Raw water quality for the community water is shown in Table 2-13.

TABLE 2-12

Influent Water Quality for San Ysidro EDR Study

Parameter	Concentration (mg/L)
pH	7.1 (units)
TDS	810
As(total)	0.085
Fluoride	2.4
Sulfate	36
Bicarbonate	552
Chloride	142

TABLE 2-13

Raw Water Quality for Bluewater EDR Study

Parameter	Concentration (mg/L)
pH (units)	7.1
TDS	908
Na	78
Sulfate	398
Silica	16
Chloride	52

In another study, process water from in-situ mining was treated using a 30,000-gpd EDR unit (Garling,1981). The unit removed about 59 percent of the 0.022 mg/L arsenic in the feedwater operating at a recovery of approximately 81 percent.

2.6 ALTERNATIVE TECHNOLOGIES

2.6.1 Oxidation Filtration

Oxidation filtration technologies may be effective arsenic removal technologies. Research of oxidation filtration technologies has primarily focused on greensand filtration. As a result, this discussion focuses on the effectiveness of greensand filtration as an arsenic removal technololgy.

Substantial arsenic removal has been seen using greensand filtration (Subramanian, et al., 1997). Greensand is a zeolite-type glauconite mineral which is produced by treating glauconite sand with $KMnO_4$ until the granular material (sand) is coated with a layer of manganese oxides, particularly manganese dioxide. The principle behind this arsenic removal treatment is multi-faceted and includes oxidation, ion exchange, and adsorption. Arsenic compounds displace species from the manganese oxide (presumably OH^- and H_2O), becoming bound to the greensand surface - in effect an exchange of ions. The oxidative nature of the manganese surface converts As(III) to As(V) and As(V) is adsorbed to the surface. As a result of the transfer of electrons and adsorption of As(V), reduced manganese (MnII) is released from the surface.

The effectiveness of greensand filtration for arsenic removal is dependent on the influent water quality. Subramanian et al. (1997) showed a strong correlation between influent Fe (II) concentration and arsenic percent removal. Removal increased from 41 percent to more than 80 percent as the Fe/As ratio increased from 0 to 20 when treating a tap water with a spiked As(III) concentration of 200 mg/L. The tap water contained 366 mg/L sulfate and 321 mg/L TDS; neither constituent seemed to affect arsenic removal. The authors also point out that the influent Mn(IV) concentration may play an important role. Divalent ions, such as calcium, can also compete with arsenic for adsorption sites. Water quality would need to be carefully evaluated for applicability for treatment using greensand. Other researchers have also reported substantial arsenic removal using this technology. Other researchers (Lauf and Magyar) have both reported arsenic removals of greater than 90 percent for treatment of groundwater (Subramanian, et al., 1997).

As with other treatment media, greensand must be regenerated when its oxidative and adsorptive capacity have been exhausted. Greensand filters are regenerated using a solution of excess potassium permanganate ($KMnO_4$). Like other treatment media, the regeneration frequency will depend on the influent water quality in terms of constituents which will degrade the filter capacity. Regenerant disposal for greensand filtration has not been addressed in previous research.

2.6.2 Sulfur-Modified Iron

A patented Sulfur-Modified Iron (SMI) process for arsenic removal has recently been developed (Hydrometrics, 1997 and 1998). The process consists of three components: 1) finely-divided metallic iron, 2) powdered elemental sulfur, or other sulfur compounds, and 3) an oxidizing agent. The powdered iron, powdered sulfur, and the oxidizing agent (H_2O_2 in preliminary tests) are thoroughly mixed and then added to the water to be treated. The oxidizing agent serves to convert As(III) to As(V). The solution is then mixed and settled.

Using the SMI process on several water types, high adsorptive were capacities obtained with final arsenic concentration of 0.050 mg/L. Arsenic removal was influenced by pH. Approximately 20 mg As removed/g iron at pH 8, and 50 mg As removed/g iron at pH 7. Arsenic removal seems to be very dependent on the iron to arsenic ratio.

Packed bed column tests demonstrated significant arsenic removal at residence times of 5 to 15 minutes. Significant removal of both arsenate and arsenite was measured. The highest adsorption capacity measured was 11 mg As removed/g iron. Flow distribution problems were evident, as several columns became partially plugged and better arsenic removal was observed with reduced flow rates.

Spent media from the column tests were classified as nonhazardous waste. Projected operating costs for SMI, when the process is operated below a pH of 8, are much lower than alternative arsenic removal technologies such as ferric chloride addition, reverse osmosis, and activated alumina. Cost savings would increase proportionally with increased flow rates and increased arsenic concentrations.

Possible treatment systems using SMI include continuous stirred tank reactors, packed bed reactors, fluidized bed reactors, and passive in situ reactors. Packed bed and fluidized bed reactors appear to be the most promising for successful arsenic removal in pilot-scale and full-scale treatment systems based on present knowledge of the SMI process.

2.6.3 Granular Ferric Hydroxide

A new removal technique for arsenate, which has recently been developed at the Technical University of Berlin (Germany), Department of Water Quality Control, is adsorption on a granular ferric hydroxide (GFH) in fixed bed reactors. This technique combines the advantages of the coagulation-filtration process, efficiency and small residual mass, with the fixed bed adsorption on activated alumina, and simple processing.

Driehaus et al. (1998) reported that the application of GFH in test adsorbers showed a high treatment capacity of 30,000 to 40,000 bed volumes with an effluent arsenate concentration never exceeding 10 µg/L. The typical residual mass was in the range of 5-25 g/m^3 treated water. The residue was a solid with an arsenate content of 1-10 g/kg. Table 2-14 summarizes the data of the adsorption tests.

Table 2-14

Adsorption Tests on GFH

	Units	Test 1	Test 2	Test 3	Test 4
Raw Water Parameters					
pH		7.8	7.8	8.2	7.6
Arsenate Concentration	µg/L	100-800	21	16	15-20
Phosphate Concentration	µg/L	0.70	0.22	0.15	0.30
Conductivity	µS/cm	780	480	200	460
Adsorption Capacity for Arsenate	g/kg	8.5	4.5	3.2	N/D
Adsorber					
Bed Height	m	0.24	0.16	0.15	0.82
Filter Rate	m/h	6-10	7.6	5.7	15
Treatment Capacity	BV	34,000	37,000	32,000	85,000
Maximum Effluent Concentration	µg/L	10	10	10	7
Arsenate Content of GFH	g/kg	8.5	1.4	0.8	1.7
Mass of Spent GFH (dry weight)	g/m^3	20.5	12	18	8.6

N/D: not determined

The competition of sulfate on arsenate adsorption was not very strong. Phosphate, however, competed strongly with arsenate, which reduced arsenate removal with GFH. Arsenate adsorption decreases with pH, which is typical for anion adsorption. At high pH values GFH out-performs alumina. Below a pH of 7.6 the performance is comparable. The most significant weakness, however, appears to be cost. Currently, GFH media costs approximately $4,000 per ton. The effect on total O&M costs should be evaluated. It is possible that if a GFH bed can be used several times longer than an alumina bed, for example, the overall effect may be minimal.

A treatment for leaching arsenic and the regeneration of GFH seems possible, but it leads to an alkaline solution with arsenate and requires a further treatment to obtain a solid waste. Thus, a direct deposition of spent GFH as hazardous waste should be favored.

2.6.4 Iron Filings

Iron filings and sand may be used to reduce inorganic arsenic species to iron co-precipitates, mixed precipitates and, in conjunction with sulfates, to arsenopyrites. This type of process is essentially a filter technology, much like greensand filtration, wherein the source water is filtered through a bed of sand and iron filings. Unlike some technologies, ion exchange for example, sulfate is actually introduced in this process to encourage arsenopyrite precipitation.

This arsenic removal method was originally developed as a batch arsenic remediation technology. It appears to be quite effective in this use. Bench-scale tests indicate an average removal efficiency of 81% with much higher removals at lower influent concentrations. This method was tested to arsenic levels of 20,000 ppb, and at 2000 ppb consistently reduced arsenic levels to less than 50 ppb (the current MCL). While it is quite effective in this capacity, its use as a drinking water treatment technology appears to be limited. In batch tests a residence time of approximately seven days was required to reach the desired arsenic removal. In flowing conditions, even though removals averaged 81% and reached greater than 95% at 2000 ppb arsenic, there is no indication that this technology can reduce arsenic levels below approximately 25 ppb, and there are no data to indicate how the technology performs at normal source water arsenic levels. This technology needs to be further evaluated before it should be recommended as an approved arsenic removal technology for drinking water.

3.0 TECHNOLOGY COSTS

3.1 INTRODUCTION

This chapter presents estimated capital and operations and maintenance (O&M) expenditures for the following arsenic removal technologies and unit processes:

- Pre-oxidation technologies, including chlorination and potassium permanganate;
- Precipitative processes, including coagulation/filtration, direct filtration, coagulation assisted microfiltration, enhanced coagulation, lime softening and enhanced lime softening;
- Adsorption processes, including activated alumina;
- Ion exchange processes, specifically anion exchange;
- Separation processes, including ultrafiltration, microfiltration, nanofiltration and reverse osmosis.

Each section includes a brief technology description, design criteria, and capital and O&M cost curves for systems ranging from 0.01 to 430 mgd.

3.2 BASIS FOR COST ESTIMATES

3.2.1 Cost Modeling

Three cost models were used in cost development: the *Very Small Systems Best Available Technology Cost Document* (Malcolm Pirnie, 1993), hereafter referred to as the VSS model; the Water Model (Culp/Wesner/Culp, 1984); and the W/W Cost Model (Culp/Wesner/Culp, 1994). Curve fitting analysis was conducted on the modeled cost estimates including the utilization of transition flow regions to provide better estimates within the breakpoints between models.

The following flow ranges have been generally established for each model and transition flow region:

- VSS - 0.015 to 0.100 mgd
- Transition 1 - 0.100 to 0.270 mgd
- Water Model - 0.27 to 1.00 mgd
- Transition 2 - 1 to 10 mgd
- W/W Cost Model - 10 to 200 mgd

Some processes (i.e., activated alumina and ion exchange) have slightly different ranges due to discrepancies between the models. Membrane processes have different ranges since these costs were generally developed without use of the models. All three models require flow to calculate capital and operation and maintenance (O&M) costs. In addition, the Water and W/W Cost models require several user-specified variables to generate direct capital cost. These additional user inputs include design factors, cost indices (Table 3-7), and other various unit costs (Tables 3-8 and 3-9).

3.2.2 Technology Design Panel Recommendations

Since the 1986 Safe Drinking Water Act (SDWA) reauthorization, EPA has relied mainly on the previously mentioned unit cost models to estimate compliance costs for drinking water regulations. Following the reauthorization of the SDWA in 1996, EPA has critically evaluated its tools for estimating the costs and benefits of drinking water regulations. As part of this evaluation, EPA solicited technical input from national drinking water experts at the Denver Technology Workshop (which was sponsored by EPA and held November 6 and 7, 1997) to improve the quality of its compliance cost estimating process for various drinking water treatment technologies. The Technology Design Panel (TDP) formed at the workshop for this purpose recommended several modifications to existing cost models to improve the accuracy of EPA's compliance cost estimates.

The TDP developed guidelines for estimating capital costs using the three cost models. The guidelines are discussed in greater detail in *Guide for Implementing Phase I Water Treatment Upgrade* (EPA, 1998a) and *Water Treatment Costs Development (Phase I): Road Map to Cost Comparisons* (EPA, 1998b).

Total capital costs consist of three elements: process, construction, and engineering costs. Process costs include manufactured equipment, concrete, steel, electrical and instrumentation, pipes and valves, and housing costs. Construction costs include sitework and excavation, subsurface considerations, standby power, land, contingencies, and interest during construction. Engineering costs include general contractor overhead and profit, engineering fees, and legal, fiscal, and administrative fees (including permitting).

The TDP recommended that total capital cost estimates be generated based solely upon process costs. That is, the models can be used to estimate total capital costs, but process costs are then generated using the capital cost breakdowns presented in Appendices A through C, and applying an appropriate factor for construction and engineering costs. These factors are based upon system size and are presented in Table 3-1.

Table 3-1

TDP Capital Cost Factors

System Size	Process Cost Factor (Percent of Total)	Construction Cost Factor (Percent of Total)	Engineering Cost Factor (Percent of Total)	Total Cost Factor[1] (Percent of Total)
Very Small	1.00 (40%)	1.00 (40%)	0.50 (20%)	2.50 (100%)
Small	1.00 (40%)	1.00 (40%)	0.50 (20%)	2.50 (100%)
Large	1.00 (30%)	1.33 (40%)	1.00 (30%)	3.33 (100%)

1 - This factor can be multiplied by the process cost to obtain the total capital cost.

Table 3-2 presents a sample capital cost breakdown for the VSS model membrane equations. The table also lists the capital costs assumptions associated with the VSS model. Capital cost breakdowns for all technologies costed using the VSS model are presented in Appendix A.

The Water and W/W Cost assumptions for capital cost components vary by design and average flow. Supporting documentation was used to develop capital cost breakdown summaries for the Water and W/W Cost models. *Estimation of Small System Water Treatment Costs* (Culp/Wesner/Culp, 1984) and *Estimating Treatment Costs, Volume 2: Cost Curves Applicable to 1 to 200 mgd Treatment Plants* (Culp/Wesner/Culp, 1979) were used for the Water and W/W Cost models, respectively. These documents present the design assumptions used in developing the cost models, as well as associated costs. The percent of total cost for each component cost was calculated for each design condition. These percentages were averaged to arrive at a universal capital cost breakdown which could be applied for developing the Phase I capital costs. Tables 3-3 through 3-6 demonstrate the methodology described here.

Table 3-2

VSS Capital Cost Breakdown for Membrane Processes (Including Microfiltration and Ultrafiltration)

Cost Component	Model Assumption	Cost Factor	Percent of Total Capital	Capital Cost Category
Manufactured Equipment	100%	1.000	56.97%	p
Installation	25%	0.2500	14.24%	c
Sitework and Interface Piping	6%	0.0750	4.27%	c
Standby Power	5%	0.0625	3.56%	c
General Contractor Overhead & Profit	12%	0.1665	9.49%	e
Legal, Fiscal and Administrative Fees	3%	0.0416	2.37%	e
Engineering	10%	0.1596	9.09%	e
Miscellaneous and Contingencies	0%	0.000	0.00%	c
TOTAL		1.7552	100.00%	

p = process, c = construction, e = engineering

Output from the Water and W/W Cost models includes construction costs and additional capital costs, which together make up the total capital cost. Additional capital costs include sitework and interface piping, standby power, overhead and profit, engineering, legal, fiscal, and administrative fees. There are no process costs associated with the additional capital costs. As a result, cost breakdowns need only consider the construction cost output from these two models. Tables 3-4 and 3-6 present sample capital cost breakdowns for the Water and W/W Cost models, respectively. Capital cost breakdowns for each technology and unit process are presented in Appendices A through C for the VSS, Water, and W/W Cost models, respectively

Table 3-3

Water Model Capital Cost Breakdown for Package Conventional Treatment (Coagulation/Filtration)

Cost Component	Filter Area (ft^2)						Capital Cost Category
	2	12	20	40	112	150	
Excavation and Sitework	$3,500	$3,500	$4,700	$5,800	$7,000	$9,300	c
Manufactured Equipment	$31,000	$44,900	$53,500	$111,300	$176,600	$190,500	p
Concrete	$1,000	$1,000	$1,500	$4,500	$5,700	$6,800	p
Labor	$9,900	$14,700	$17,500	$36,400	$57,800	$62,400	c
Pipes and Valves	$4,200	$8,300	$10,400	$20,900	$29,200	$41,700	p
Electrical	$3,200	$4,500	$5,300	$11,100	$17,600	$19,000	p
Housing	$18,600	$18,600	$23,400	$45,000	$47,500	$52,500	p
Subtotal	$71,400	$95,500	$116,300	$235,000	$341,400	$382,200	
Contingencies	$10,700	$14,300	$17,400	$35,300	$51,200	$57,300	e
Total	$82,100	$109,800	$133,700	$270,300	$392,600	$439,500	

Table 3-4

Water Model Capital Cost Breakdown by Percentage for Package Conventional Treatment (Coagulation/Filtration)

Cost Component	Filter Area (ft^2)						Average Percent
	2	12	20	40	112	150	
Excavation and Sitework	4.26%	3.19%	3.52%	2.15%	1.78%	2.12%	2.84%
Manufactured Equipment	37.76%	40.89%	40.01%	41.18%	44.98%	43.34%	41.36%
Concrete	1.22%	0.91%	1.12%	1.66%	1.45%	1.55%	1.32%
Labor	12.06%	13.39%	13.09%	13.47%	14.72%	14.20%	13.49%
Pipes and Valves	5.12%	7.56%	7.78%	7.73%	7.44%	9.49%	7.52%
Electrical	3.90%	4.10%	3.96%	4.11%	4.48%	4.32%	4.15%
Housing	22.66%	16.94%	17.50%	16.65%	12.10%	11.95%	16.30%
Contingencies	13.03%	13.02%	13.01%	13.06%	13.04%	13.04%	13.03%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table 3-5

W/W Cost Model Capital Cost Breakdown for Sedimentation Basins

Cost Component	Area (A = ft^2) and Length x Width (LW = ft x ft)						Capital Cost Category
	A=240 LW = 30x8	A=600 LW=60x10	A=1260 LW=90x14	A=2240 LW=140x16	A=3600 LW=200x18	A=4800 LW=240x20	
Excavation and Sitework	$1,060	$2,000	$3,060	$4,680	$6,670	$8,090	c
Manufactured Equipment	$8,540	$12,080	$24,470	$32,020	$53,110	$63,440	p
Concrete	$2,970	$5,490	$84,430	$12,820	$19,190	$22,070	p
Steel	$6,400	$13,110	$19,440	$32,620	$51,250	$39,680	p
Labor	$6,220	$11,260	$17,320	$26,390	$37,570	$45,300	c
Pipes and Valves	$6,960	$7,400	$9,100	$12,500	$16,100	$21,450	p
Electrical	$1,510	$1,760	$1,860	$2,020	$2,110	$2,400	p
Subtotal	$33,660	$53,100	$83,680	$123,050	$190,000	$232,430	
Contingencies	$5,050	$7,970	$12,550	$18,460	$27,750	$34,860	e
Total	$38,710	$61,070	$96,230	$141,510	$212,750	$267,290	

Table 3-6

W/W Cost Model Capital Cost Breakdown by Percentage for Sedimentation Basins

Cost Component	Area (A = ft^2) and Length x Width (LW = ft x ft)						Average Percent
	A=240 LW = 30x8	A=600 LW=60x10	A=1260 LW=90x14	A=2240 LW=140x16	A=3600 LW=200x18	A=4800 LW=240x20	
Excavation and Sitework	2.74%	3.27%	3.18%	3.31%	3.14%	3.03%	3.11%
Manufactured Equipment	22.06%	19.78%	25.43%	22.63%	27.96%	23.73%	23.10%
Concrete	7.67%	8.99%	8.76%	9.06%	8.55%	8.26%	8.55%
Steel	16.53%	21.47%	20.20%	23.05%	24.09%	26.07%	21.90%
Labor	16.07%	18.44%	18.00%	18.65%	17.66%	16.95%	17.63%
Pipes and Valves	17.98%	12.12%	9.46%	8.83%	7.57%	8.02%	10.66%
Electrical	3.90%	2.88%	1.93%	1.43%	0.99%	0.90%	2.01%
Contingencies	13.05%	13.05%	13.04%	13.05%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

3.2.3 Implementing TDP Recommended Costing Upgrades

The capital cost breakdowns presented above and in the appendices of this document can be used to estimate the modified capital cost, i.e., the capital cost estimate developed using the TDP recommendations. The following sections briefly demonstrate how the capital cost breakdowns are applied, and modified capital cost estimates are generated.

3.2.3.1 VSS Model

1. The VSS model presents capital and O&M costs as functions of design and average flow, respectively. Accordingly, the capital cost equation for package microfiltration units is:

   CAP = 0.86[DES] + 41.1

   Where:
   CAP = Total Capital Cost, $1,000s
   DES = Design Treated Flow, kgpd

2. Thus, for an 0.024 mgd (24 kgpd) plant the capital cost is:

   CAP = 0.86[24] + 41.1
   CAP = 61.74 or $61,740

3. The VSS model equations produce estimates in 1993 dollars. To escalate to September 1998, multiply the equation-generated capital cost by the ratio of the Engineering News Record (ENR) Building Cost Index for September 1998 to the 1993 index value.

   $61,740 x (3375/3009) = $69,250

   The escalated capital cost for a 0.024 mgd package microfiltration plant is $69,250.

4. Using the capital cost breakdown in Table 3-2, the total process cost is:

   $69,250 x 0.5697 = $39,452

5. The modified capital cost can then be calculated using the total cost factor presented in Table 3-1.

   $39,452 x 2.5 = $98,629

   Thus, the modified capital cost is $98,629.

3.2.3.2 Water Model

1. Assume the Water model output for a 0.27 mgd (270,000 gpd) package conventional treatment (coagulation/flocculation/filtration) plant is $692,066 (escalated to 1998 dollars).

2. Using the capital cost breakdown in Table 3-4, the total process cost is:

   $692,066 x (0.4136 + 0.0132 + 0.0752 + 0.0415 + 0.1630) = $488,945

3. The modified capital cost can then be calculated using the total cost factor presented in Table 3-1.

   $488,945 x 2.5 = $1,222,362

4. This approach must be applied to each unit process (e.g., backwash pumping) separately, then totaled for the entire treatment process to estimate the modified capital cost.

3.2.3.3 W/W Cost Model

1. Assume the W/W Cost model output for a 1 mgd (1250 sq ft.) rectangular sedimentation basin is $416,574 (escalated to 1998 dollars).

2. Using the capital cost breakdown in Table 3-6, the total process cost is:

   $416,574 x (0.2311 + 0.0855 + 0.2190 + 0.1066 + 0.0201) = $275,897

3. The modified capital cost can then be calculated using the total cost factor presented in Table 3-1.

   $275,897 x 3.33 = $918,737.

4. This approach must be applied to each unit process separately (e.g., acid feed), then totaled for the entire treatment process to estimate the modified capital cost.

3.2.4 Cost Indices and Unit Costs

Both the Water Model and the W/W Cost Model require a number of standard indices and various unit costs from the Bureau of Labor Statistics, the Engineering News Record, and other referenced sources. The values used in conjunction with the development of cost estimates are reported in Tables 3-7 through Table 3-9.

Table 3-7

Costs Indices Used in the Water and W/W Cost Models

Description	Index Reference	Numerical Value[1]
Concrete Ingredients and Related Products	BLS 132	448.8
Electrical Machinery and Products	BLS 117	281.8
General Purpose Machinery and Equipment	BLS 114	445.1
Metals and Metal Products (Steel)	BLS 1017	405.1
Miscellaneous General Purpose Equipment (Pipes &Valves)	BLS 1149	521.5
PPI Finished Goods Index	BLS 3000	364.0
ENR Building Cost Index		3375.31
ENR Skilled Labor		5317.36
ENR Materials Prices		2189.24

(1) BLS numerical values were re-based to 1967 base year (see Section 3.2.5)

Table 3-8

Unit and General Cost Assumptions

Electricity[1]	$0.08/kWh
Diesel Fuel[1]	$1.25/gallon
Natural Gas[1]	$0.006/scf
Labor[2]	Large systems: $40/hr Small systems: $28/hr Very small systems: 8 hrs/wk
Building Energy Use	102.6 kWh/sq ft/yr
Housing Costs	$125/sq ft

[1] Energy Information Administration.

[2] Technical Design Panel (EPA, 1998a)

TABLE 3-9

Chemical Costs

Chemical	Cost	Units
Alum, Dry Stock	\$300	per ton
Carbon Dioxide, Liquid	\$340	per ton
Chlorine, 1 ton cylinder	\$350	per ton
Chlorine, 150 lb cylinder	\$400	per ton
Chlorine, Bulk	\$280	per ton
Ferric Chloride	\$350	per ton
Hexametaphosphate	\$1276	per ton
Lime, Quick Lime	\$95	per ton
Phosphoric Acid	\$300	per ton
Polymer	\$2.25	per lb
Potassium Permanganate	\$2700	per ton
Soda Ash	\$400	per ton
Sodium Hypochlorite, 12%	\$1100	per ton
Sodium Chloride	\$99	per ton
Sodium Hydroxide, 50% solution	\$371	per ton
Sulfuric Acid	\$116	per ton

This document presents total capital costs and annual O&M costs. Annual O&M costs include the costs for materials, chemicals, power and labor. Annualized costs can be determined using the following equations:

$$\text{Total annual cost (¢/kgal)} = \text{Annualized Capital Cost (¢/kgal)} + \text{O\&M Cost (¢/kgal)}$$

Where:

$$\text{Annualized Capital Cost} = \frac{\text{Capital Cost (\$) * Amortization Factor * 100 ¢/\$}}{\text{Average Daily Flow (mgd)*(1000 kgal/mgal)*365 days/year}}$$

$$\text{O\&M Cost (¢/kgal)} = \frac{\text{Annual O\&M (\$) * 100 (¢/\$)}}{\text{Average Daily Flow (mgd)*1000 kgal/mgal*365 days/year}}$$

Amortization, or capital recovery, factors for interest rates of 3, 7, and 10 percent for 20 years are reported in Table 3-10. Alternative capital recovery factors can be calculated using the formula presented below.

Capital Recovery Factor = $i(1 + i)^N / (1 + i)^N - 1$

Where:
i = interest rate
N = number of years

Table 3-10

Amortization Factors

Interest Rate (%)	Amortization Period	Amortization Factor
3	20	0.0672157
7	20	0.0943929
10	20	0.1174596

3.2.5 Re-Basing Bureau of Labor Statistics Cost Indices

The Water Model and W/W Cost Model uses BLS cost index information based to 1967. In 1986, the BLS conducted a comprehensive overhaul of the industrial price methodology resulting in a re-basing of all index information to a 1982=100 base year. This requires a re-basing of BLS index information to 1967 prior to use in the models for the development of cost estimates. Table 3-11 provides the re-base factors. A sample re-base calculation is presented below.

Sample Rebase Calculation:

Machinery = 1982 Base Factor / Rebase Factor = 1967 Base Factor

= 147.8 / 0.32895016 = **449.3**

Table 3-11

Bureau of Labor Statistics Rebase Information

BLS Series:	Index Reference	1982=100 Number	Re-base Factor [1]	1967=100 Number	Date
Machinery	BLS 114	147.8	0.32895016	449.3	9/98
Concrete	BLS 132	148.8	0.32261652	461.2	9/98
Steel	BLS 1017	113.3	0.28608856	396.0	9/98
Pipes & Valves	BLS 1149	162.2	0.30909034	524.8	9/98
Electrical	BLS 117	120.8	0.43185069	279.7	9/98
PPI Finish Goods Index	BLS 3000	130.6	0.35633299	366.5	9/98

(1) Provided by the BLS

3.2.6 Flows Used in the Development of Costs

Flow categories were developed to provide adequate characterization of costs across each of the flow regions presented in Section 3.2.1. A minimum of four data points were generated for each of the flow regions, with the exception of the transition regions, where cost estimates are based upon a linear regressions between the last data point of the previous region and the first data point of the following region. Table 3-12 presents the design and average flows, and cost models used in this process.

Table 3-12

Flows Used in the Cost Estimation Process

Design Flow (mgd)	Average Flow (mgd)	Cost Model
0.010	0.0031	VSS
0.024	0.0056	VSS
0.087	0.024	VSS
0.10	0.031	VSS
0.27	0.086	Water
0.45	0.14	Water
0.65	0.23	Water
0.83	0.30	Water
1.0	0.36	Water
1.8	0.7	W/W Cost
4.8	2.1	W/W Cost
10	4.5	W/W Cost
11	5	W/W Cost
18	8.8	W/W Cost
26	13	W/W Cost
51	27	W/W Cost
210	120	W/W Cost
430	270	W/W Cost

Shaded rows represent data used in the estimation of costs with the transition regions.

3.3 COSTS FOR MULTIPLE REMOVAL PERCENTAGES

Capital and O&M cost estimates are presented for multiple removal percentages in this document; specifically, 30, 50, 80 and 95 percent removal. There were some instances where one or more of these specified removals could not be achieved for a given contaminant. In those instances, costs were estimated for the maximum achievable removal and the specified removals which could be achieved. Section 3.5 presents a contaminant and applicable removal technology matrix which identifies applicable technologies for each of the contaminants discussed in Chapter 2, the technologies which have been demonstrated as effective for removal of those contaminants, and removal percentages for which costs have been estimated.

3.3.1 Removal and Accessory Costs

Costs for each of the removal technologies presented in this document can be separated into two categories: removal and accessory. Accessory costs include raw and finished water pumping, and clearwell storage. Removal costs include any process item directly associated with the removal of a particular contaminant, e.g., the ion exchange bed in ion exchange processes.

Accessory costs are independent of the desired removal percentage. For example, a one mgd treatment plant must still pump one million gallons of raw water into the plant, pump one million gallons of finished water, and have adequate storage (10% of daily production). Conversely, removal costs are dependent upon the desired removal. If contaminant levels are such that the plant need only remove 30 percent of the contaminant to reach the treatment goal, then the treatment process can be scaled to treat a portion of the flow. The treated flow is then blended with the untreated portion prior to distribution. Section 3.3.2 discusses the blending approach used in the development of cost estimates.

Cost estimates presented in this document do not include accessory capital and O&M. Cost curves and equations for accessory costs (i.e., raw and finished water pumping, and clearwell storage) are presented in Appendix D.

3.3.2 Use of Blending in Cost Estimates

Capital and O&M costs were estimated using the VSS, Water and W/W Cost models. As previously discussed, it may not be necessary to treat the total process flow. If contaminant levels are sufficiently low, a utility may only need to remove 30 percent of the contaminant to achieve a treatment goal. Removal cost curves and equations for each technology are presented in Appendix D. The portion of the total process flow to be treated is calculated using the following equation:

$$Q_{treated} = \frac{Q_{total}}{[((C_{max} - C_{desired})/C_{desired}) + 1]}$$

Where:

$Q_{treated}$ = Treated portion of the total process flow, mgd

Q_{total} = Total daily process flow, mgd

C_{max} = Maximum achievable removal efficiency, %

$C_{desired}$ = Desired removal efficiency, %

If 1 is substituted for the total daily flow (Q_{total}) in the above equation, the treated portion of the flow ($Q_{treated}$) is expressed as a fraction of the total flow. Accessory capital and O&M costs (based on the total daily flow) and removal capital and O&M costs (based on a treated portion of the flow) were totaled to estimate capital and O&M costs for each contaminant, technology and removal efficiency combination.

3.4 ADDITIONAL CAPITAL COSTS

The cost models discussed in the previous sections are good tools for estimating capital and O&M costs associated with various drinking water treatment technologies. There are additional capital costs, however, which the models do not account for and may be a very real expense for public water utilities. The need for additional capital costs can be affected by a number of factors, including: contaminants present, quality of the source water, land availability, retrofit of existing plants, permitting requirements, piloting issues, waste disposal issues, building or housing needs, and redundancy. Tables with additional capital cost estimates for each technology discussed in this document are presented in Appendix E.

Contaminants

Arsenic is typically present in drinking water in one of two oxidation states, As(III) or As(V). As(V) is more effectively removed by each of the removal technologies discussed in this document. However, As(III) can be easily oxidized to As(V) using chlorination, potassium permanganate or other method. Groundwaters typically contain As(III), while As(V) is more commonly found in surface waters.

The presence of additional contaminants, for example, inorganics (sulfate, aluminum, manganese), pathogenic contaminants (*Giardia*, *Cryptosporidium*), or organic contaminants (trihalomethanes, haloacetic acids), can raise additional treatment concerns and result in decreased process performance. Changes in coagulant dosage or type, sedimentation time, or membrane efficiency are just a few of the concerns that may arise. Presence of pathogens can result in a need for disinfection of finished water.

Water Source

For the purpose of developing cost estimates, it is assumed that the source water is from a clean, consistent, single source. Facilities combining plant influent from multiple sources can affect source water quality, which may in turn affect the removal efficiency of the treatment technologies discussed. Ion exchange, for example, may not be an effective removal technology for source waters with high influent sulfate levels.

Costs may vary for ground and surface water systems as well. Treatment technologies susceptible to fouling by suspended solids (i.e., surface water systems) may require additional pretreatment (pre-filtration). Costs for these items are included in the additional capital cost tables presented in Appendix E.

Land

Land requirements were calculated based upon TDP recommendations (EPA, 1997) and engineering judgement. Appendix E presents two scenarios for land costs. The low cost scenario assumes land costs to be $1,000 per acre for small systems (i.e., less than 1 mgd) and $10,000 per acre for large systems. All land costs are $100,000 per acre for the high cost scenario (SAIC, 1998).

Retrofitting

All costs presented in this document are for new construction, with the exception of the enhanced coagulation and enhanced lime softening processes. All processes contained in the cost models include pipes and valves, electrical and instrumentation, and other costs associated with retrofitting. It was assumed that the costs included are sufficient for the retrofit of existing coagulation/filtration and softening plants. As a result, costs for retrofitting are excluded from Appendix E.

Permitting

Permitting costs follow the recommendations of the TDP as presented in the *Technology Design Conference Information Package* (EPA, 1997). A technology specific summary of low and high cost permitting scenarios is presented in Table 3-13. The number of permits required can vary by location, depending upon State and Local regulations, as well as technology. Some technologies may require permitting for storage tanks used for process chemicals, while others may necessitate NPDES permits, if the disposal option for process residuals happens to be discharge to a nearby surface water.

Piloting

Piloting costs are neglected in this document and are not included in Appendix E.

Waste Disposal

The characteristics of arsenic containing waste streams is presented in Chapter 4. Appropriate handling and disposal methods are discussed for residuals generated by each each treatment process for which capital and O&M cost estimates are provided. Cost equations for disposal by each of these methods are presented in *Small Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993a) and *Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993b).

Table 3-13

Permitting Scenarios

Permit Type	CF^2	ECF^2	DF^2	CMF^2	LS^2	ELS^2	AA^2	IX^2	UF^2	NF^2	RO^2
Possible Permits for All Technologies^3											
Land Development	H	H	H	H	H	H	H	H	H	H	H
Stormwater Managment	H	H	H	H	H	H	H	H	H	H	H
Soil Erosion & Sediment Control	H	H	H	H	H	H	H	H	H	H	H
Building	H	H	H	H	H	H	H	H	H	H	H
Potable Water	B	B	B	B	B	B	B	B	B	B	B
Technology Specific Permits^3											
Sludge Disposal	B	B	B	B	B	B			B	B	B
Air Quality											
NPDES	B	B	B	B			B	B	B	B	B
UIC											H
Site Dependent Permits^3											
UST/AST Registration											
Stormwater NPDES	H	H	H	H	H	H	H	H	H	H	H
SPCC Plan	B	B	B	B	B	B		B			B
Highway Occupancy	H	H	H	H	H	H	H	H	H	H	H
Rodent & Insect Control	H	H	H	H	H	H	H	H	H	H	H
EA/EIS	H	H	H	H	H	H	H	H	H	H	H
Building Occupancy	H	H	H	H	H	H	H	H	H	H	H
Wetlands	H	H	H	H	H	H	H	H	H	H	H

1 Based upon *Technical Design Conference Information Package* (EPA, 1997)

2 CF - Coagulation/Filtration, ECF - Enhanced Coagulation, DF - Direct Filtration, CMF - Coagulation Assisted Microfiltration, LS - Lime Softening, ELS - Enhanced Lime Softening, AA - Activated Alumina, IX - Ion Exchange, UF - Ultrafiltration, NF - Nanofiltration, RO - Reverse Osmosis

3 T - Typical Cost Scenario, H - High Cost Scenario, B - Both Typical and High Cost Scenarios

Storage/Building

All of the cost models used in preparing the technologies and costs document include costs for housing of equipment. It is assumed that the costs included in the model output is sufficient. As a result, additional building costs are not included. It is also assumed for all scenarios that source water production is consistent, and storage for source water is not provided.

Redundancy

The cost models include standby pumps for some of the unit processes used in generating the cost estimates presented in this document, e.g., raw and finished water pumping. Further, it is good design practice to include additional filtration structures and sedimentation basins to allow continued operation during maintenance of one or more of the structures. Backup pumps are not included for chemical feed systems. As a result, there may be some additional capital costs associated with redundancy for these items. *Recommended Standards for Water Works* (Great Lakes Upper Mississippi River Board of State Public Health and Environmental Managers, 1997), often referred to as the Ten State Standards, presents a comprehensive discussion of redundancy and recommended redundant items. The Ten State Standards were used for presenting costs for redundant items in Appendix E.

3.5 PRE-OXIDATION PROCESSES

Inorganic arsenic occurs in two primary valence states, arsenite (As III) and arsenate (As V). Surface waters more typically contain As(V), while As(III) is the dominant species found in ground waters. Each of the treatment technologies presented in this document remove As(V) more readily than As(III). As a result, pre-oxidation may be necessary depending upon source water conditions.

Potassium permanganate addition and chlorination are two oxidation technologies that have been evaluated and deemed effective for the conversion of arsenite to arsenate. Chlorination may cause disinfection by-product (DBP) formation in source waters with high TOC concentrations. Further, chlorination may cause fouling in some membrane processes. Source water characteristics should be thoroughly evaluated when considering pre-oxidation technologies. Additional oxidation technologies, such as ozonation and hydrogen peroxide, may be effective, but need further evaluation.

3.5.1 Potassium Permanganate

Potassium permanganate can be used as a pre-oxidation technology for conversion of As(III) to As(V). Potassium permanganate is more expensive than chlorination; $2700 per ton compared with $350 to $400 per ton. However, unlike chlorination, potassium permanganate is not known to form measurable DBPs and does not foul membranes. Raw water and downstream process considerations should be made when selecting a pre-oxidation technology. For this document potassium permanganate costs were calculated for dosages of 1.5, 3.0 and 5.0 mg/L. The *Very Small Systems Best Available Technology Cost Document* (Malcolm Pirnie, 1993) was used for calculating costs for the flows below 0.10 mgd. Capital and O&M costs for flows between 0.27 and 1 mgd were calculated using the Water Model. For flows greater than 10 mgd the W/W Cost Model was used to estimate the capital and O&M costs. Linear regressions were used to estimate costs in the transition regions between the models, i.e., 0.10 to 0.27 mgd, and 1 to 10 mgd. The following are some highlights of the system design used at the time of cost estimation:

- For very small systems, the potassium permanganate feed system is equipped with a metering pump, solution tank with mixer, pipes and valves, and instrumentation and controls. The system utilizes a 3% potassium permanganate solution.
- The VSS document makes provisions for building (42.7%), fencing (49.4%), and road (33.8%) costs associated with potassium permanganate addition.
- O&M costs for very small systems were calculated using equations in the very small systems cost document. Labor requirements were assumed to be 8 hours per week.
- For small system potassium permanganate addition, a dry chemical feed system capable of 1,000 pounds per day was used.

Figures 3-1 through 3-6 present capital and O&M cost curves and equations for potassium permanganate addition.

3.5.2 Chlorination

As previously stated, chlorination can cause DBP formation in source waters with high TOC concentrations. Chlorination has also been shown to cause fouling in some membrane processes. As a result, source water characteristics and downstream process needs should be thoroughly evaluated when considering chlorination as an oxidation technology. Capital and O&M costs were developed for cylinder and tank feed chlorination systems at dosages of 1.5, 3.0, and 5.0 mg/L. Similar to potassium permanganate systems, the VSS Model was used for calculating costs for the flows below 0.10 mgd. Capital and O&M costs for flows between 0.27 and 1 mgd were calculated using the Water Model. For flows greater than 10 mgd the W/W Cost Model was used to estimate the capital and O&M costs. Linear regressions were used to estimate costs in the transition regions between the models, i.e., 0.10 to 0.27 mgd, and 1 to 10 mgd. The following are some highlights of the system design used at the time of cost estimation:

- For very small systems, chlorination is accomplished with a hypochlorite feed system capable of providing dosages to 10 mg/L as chlorine. The system is equipped with a 150 gallon storage tank and utilizes a 15% sodium hypochlorite feed stock.
- The VSS Model makes provisions for building (52.2%), fencing (60.5%), and road (41.4%) costs associated with chlorine addition.
- O&M costs for very small systems were calculated using equations in the very small systems cost document. Labor requirements were assumed to be 8 hours per week.
- For small systems, cylinder feed chlorination system capital and O&M costs were estimated.

It should be noted that some systems currently using chlorine for disinfection may be able to modify existing chlorine feed systems to utilize chlorine as a preoxidant with significant capital cost savings. Capital and O&M cost curves and equations are presented in Figures 3-7 through 3-12.

3.6 PRECIPITATIVE PROCESSES

3.6.1 Coagulation/Filtration

Coagulation/filtration (C/F) is a treatment process that alters the physical or chemical properties of colloidal or suspended solids, enhancing agglomeration, and allowing these solids to settle out of solution by gravity or be removed by filtration. The C/F removal mechanism is discussed in greater detail in Chapter 2. A typical C/F process includes coagulant addition, which may be followed by polymer addition to aid agglomeration, flocculation, sedimentation and filtration.

C/F is widely used as a treatment for removing suspended solids from surface water supplies. Most ground waters are low in turbidity and do not require this type of treatment. Source waters containing high As(III) concentrations may opt for oxidation as a pre-treatment for C/F. Pre-oxidation options are presented in Section 3.5.

Very Small Systems (Less than 0.10 mgd)

Capital costs for very small systems were developed using the VSS model. The design parameter most affecting capital cost is the filtration rate. It affects the size of the filter structure and volume of filter media, the most cost intensive process in a C/F plant. The VSS model also makes provisions for building (14.9-28.1%), fencing (2.1-7.5%) and road (1.2-4.7%) costs associated with each of the technologies presented. The following design criteria were used to develop capital cost estimates for systems with a design flow of less than 0.10 mgd:

- Coagulant dosage, alum or ferric chloride, 30 mg/L;
- Polymer dosage, 0.4 mg/L; and
- Filtration rate, 2.5 gpm/ft^2.

O&M costs are most affected by chemical costs associated with coagulant and polymer dosages. As a result, the very small systems O&M cost estimates were escalated using the BLS Chemical and Allied Products Index. Labor requirements were estimated at 8 hours per week.

Small Systems (Less than 1 mgd)

The Water Model was used to estimate capital and O&M costs for small C/F treatment plants. The following design criteria were used in developing capital and O&M cost estimates:

- Package plant for all small systems, filtration rate 5 gpm/ft^2;
- Ferric chloride dose, 25 mg/L;
- Polymer dose, 2 mg/L;
- Lime dose, 25 mg/L for pH adjustment; and
- Waste flow, 1% of treated flow at 1% solids.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to develop capital and O&M cost estimates for large C/F plants. The following design criteria were used to estimate capital and O&M costs:

- Ferric chloride dose, 25 mg/L;
- Polymer dose, 2 mg/L;
- Lime dose, 25 mg/L for pH adjustment;
- Waste flow, 1% of treated flow at 1% solids;
- Rapid mix, 1 minute;
- Flocculation, 20 minutes;
- Sedimentation, 2.5 hours using rectangular tanks; and
- Dual media gravity filters, 5 gpm/ft^2.

Figures 3-13 through 3-20 present capital and O&M cost curves and equations for removal of arsenic by C/F.

3.6.2 Enhanced Coagulation

Enhanced coagulation involves modifications to the typical C/F process such as increasing the coagulant dosage, reducing the pH, or both. The process in nearly identical to that of conventional C/F with those two exceptions. Source waters with high influent As(III) concentrations may require pre-oxidation for conversion of arsenite to arsenate. Pre-oxidation technologies are discussed in more detail in Section 3.5.

For the purpose of estimating costs, it was assumed that a typical C/F treatment plant could remove 50 percent of the influent arsenic prior to modification, i.e., enhancement. Costs presented are for the enhancement only, and are in addition to any current annual debt incurred by the utility.

<u>Small Systems (Less than 1 mgd)</u>

The VSS Model makes no appropriations for estimating enhanced coagulation capital and O&M costs. As a result, the Water Model was used to estimate capital and O&M costs for all enhanced coagulation treatment plants with a capacity of less than 1 mgd. The following design criteria were used in developing capital and O&M cost estimates:

Additional ferric chloride dose, 10 mg/L;
Additional feed system for increased ferric chloride dose;
Additional lime dose, 10 mg/L for pH adjustment;and
Additional feed system for increased lime dose.

<u>Large Systems (Greater than 1 mgd)</u>

The W/W Cost model was used to estimate capital and O&M costs for large enhanced coagulation plants. The following design criteria were used to estimate capital and O&M costs:

Additional ferric chloride dose, 10 mg/L;
Additional feed system for increased ferric chloride dose;
Additional lime dose, 10 mg/L for pH adjustment; and
Additional feed system for increased lime dose.

Figures 3-21 through 3-24 present capital and O&M cost curves and equations for enhanced coagulation.

3.6.3 Direct Filtration

Direct filtration is a modified C/F treatment process utilized for source waters with low influent suspended solids concentrations. Because of the low solids content, settling is not required and coagulation is followed immediately by filtration. Direct filtration includes all of the typical C/F process elements with the exception of flocculation and sedimentation. Source waters with high influent As(III) concentrations may require pre-oxidation for conversion of arsenite to arsenate. Pre-oxidation technologies are discussed in more detail in Section 3.5.

Very Small Systems (Less than 0.10 mgd)

Capital costs for very small systems were developed using the VSS Model. The design parameter most affecting capital cost is the filtration rate. This parameter affects the size of the filter structure and volume of filter media, the most cost intensive processes within a direct filtration plant. The VSS Model also makes provisions for building (23.5-19.5%), fencing (3.2 - 7.5 %) and road (2.0 - 5.0 %) costs associated with each of the technologies presented. The following design criteria were used to develop capital cost estimates for very small direct filtration systems:

Coagulant dosage, alum or ferric chloride, 20 mg/L;
Polymer dosage, 1.0 mg/L; and
Filtration rate, 5.0 gpm/ft^2.

O&M costs are most affected by chemical costs associated with coagulant and polymer dosages. As a result, the very small systems O&M cost estimates were escalated using the BLS Chemical and Allied Products Index. Note that labor requirements were estimated at 8 hours per week.

Small Systems (Less than 1 mgd)

The Water Model was used to estimate capital and O&M costs for small direct filtration plants. The following design criteria were used in developing capital and O&M cost estimates:

Ferric chloride dose, 10 mg/L;
Polymer dose, 1 mg/L;
Lime dose, 10 mg/L for pH adjustment; and
Waste flow, 1% of treated flow at 1% solids.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to develop capital and O&M cost estimates for large direct filtration plants. The following design criteria were used to estimate capital and O&M costs:

Ferric chloride dose, 10 mg/L;
Polymer dose, 1 mg/L;
Lime dose, 10 mg/L for pH adjustment;
Waste flow, 1% of treated flow at 1% solids;
Rapid mix, 1 minute;
Flocculation, 20 minutes; and
Dual media gravity filters, 5 gpm/ft^2.

Figures 3-25 through 3-32 present capital and O&M cost curves and equations for removal of arsenic by direct filtration.

3.6.4 Coagulation Assisted Microfiltration

Coagulation assisted microfiltration is another modified C/F process wherein microfiltration is used in place of a conventional gravity filter. The process includes all of the conventional C/F process elements. However, the microfilter replaces the gravity filter. Coagulation assisted microfiltration is capable of removing smaller particle floc which results in decreased coagulant dosage and increased plant capacity. Source waters with high influent As(III) concentrations may require pre-oxidation for conversion of arsenite to arsenate. Pre-oxidation technologies are discussed in more detail in Section 3.5.

Very Small Systems (Less than 0.10 mgd)

Capital and O&M costs for very small systems for the coagulation portion of this process were developed using the VSS Model. C/F design parameters given in Section 3.3.1 were used here as well. Microfilter specifications and cost estimates were developed based upon vendor quotes and case studies. These costs were then added to C/F cost estimates.

Small Systems (Less than 1 mgd)

The Water Model was combined with vendor data and case studies to estimate capital and O&M costs for coagulation assisted microfiltration treatment plants. Note the coagulant dose is less than that of conventional coagulation/filtration treatment. This is to prevent fouling of the microfilter. The following design criteria were used in developing capital and O&M cost estimates for the coagulation portion of this process:

- Package plant for all small systems, filtration rate 5 gpm/ft^2;
- Ferric chloride dose, 10 mg/L;
- Waste flow, 1% of treated flow at 1% solids; and
- Standard microfilter specifications, provided by vendors.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to develop capital and O&M cost estimates for large coagulation assisted microfiltration plants. Note the coagulant dose is less than that of conventional coagulation/filtration. This is to prevent fouling of the microfilter. The following design criteria were used to estimate capital and O&M costs for the coagulation portion of this process:

- Ferric chloride dose, 10 mg/L;
- Waste flow, 1% of treated flow at 1% solids;
- Rapid mix, 1 minute;
- Flocculation, 20 minutes;
- Sedimentation, 2.5 hours using rectangular tanks; and
- Standard microfilter specifications, provided by vendors.

Figures 3-33 through 3-40 present capital and O&M cost curves and equations for removal of arsenic by coagulation assisted microfiltration.

3.6.5 Lime Softening

Lime softening (LS) has been widely used for reducing hardness in large water treatment systems. The LS removal mechanism is presented in Chapter 2. As(III) or As(V) removal by LS is largely pH dependent, and pre-oxidation of arsenite to arsenate will significantly improve arsenic removal efficiencies. Pre-oxidation technologies are discussed in more detail in Section 3.5.

Considerable amounts of sludge are produced by the LS process. Large systems may find it economically feasible to install recalcination equipment to recover and reuse the process sludge and reduce disposal costs.

Small Systems (Less than 1 mgd)

The VSS Model provides no estimation methods for LS treatment. Therefore, the Water Model was used to estimate capital and O&M costs for all LS treatment plants with less than 1 mgd capacity. The following design criteria used in the development of capital and O&M cost estimates:

- Package plant for all small systems;
- Lime dose, 250 mg/L;
- Carbon dioxide (liquid), 35 mg/L for recarbonation; and
- Waste flow, 2% of total flow at 1% solids.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to develop capital and O&M cost estimates for large LS plants. The following design criteria were used to estimate capital and O&M costs:

- Lime dose, 250 mg/L;
- Carbon dioxide (liquid), 35 mg/L for recarbonation;
- Waste flow, 2% of total flow at 1% solids;
- Rapid mix, 1 minute;
- Flocculation, 20 minutes;
- Sedimentation, 1000 gpd/ft^2 using circular tanks; and
- Dual media gravity filters, 5 gpm/ft^2.

Figures 3-41 and 3-48 present capital and O&M cost curves and equations for lime softening.

3.6.6 Enhanced Lime Softening

Enhanced LS involves modifications to the typical LS treatment process in the form of increased lime dosage and possibly increased soda ash dosage. This may result in the need for pH adjustment of treated water via recarbonation. Source waters with high influent As(III) concentrations may require pre-oxidation for conversion of arsenite to arsenate. Pre-oxidation technologies are discussed in more detail in Section 3.5.

For the purpose of estimating costs, it was assumed that an existing LS plant could achieve 50 percent removal of arsenic from the source water prior to modification, i.e., enhancement. Therefore, costs presented are associated with the enhancement only, and are in addition to current annual debt incurred by the utility.

Small Systems (Less than 1 mgd)

The Water Model was used to estimate capital and O&M costs for small enhanced LS treatment plants. The following design criteria used in the development of capital and O&M cost estimates:

Additional lime dose, 50 mg/L;
Chemical feed system for increased lime dose;
Additional carbon dioxide (liquid), 35 mg/L for recarbonation; and
Chemical feed system for increased carbon dioxide dose.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to develop capital and O&M cost estimates for large enhanced LS plants. The following design criteria were used to estimate capital and O&M costs:

Additional lime dose, 50 mg/L;
Chemical feed system for increased lime dose;
Additional carbon dioxide (liquid), 35 mg/L for recarbonation; and
Chemical feed system for increased carbon dioxide dose.

Figures 3-49 through 3-52 present cost curves and equations for removal of arsenic by enhanced lime softening.

3.7 ADSORPTION PROCESSES

3.7.1 Activated Alumina

Activated Alumina (AA) is a physical/chemical process by which ions in the feed water are sorbed to the oxidized AA surface. Feed water is continuously passed through the bed to remove contaminants. The contaminant ions are exchanged with the surface hydroxides on the alumina. When all adsorption sites on the AA surface are filled, the bed must be regenerated. Regeneration is accomplished through a sequence of rinsing with regenerant, flushing with water, and neutralizing with acid. The regenerant is a strong base, typically sodium hydroxide; the neutralizer is a strong acid, typically sulfuric acid.

The adsorption of arsenic on activated alumina is highly dependent on the influent arsenic concentration. At the present time the literature does not contain a standard relationship between the influent arsenic concentration and the number of bed volumes which can be treated with AA. However, a general relationship between the two is given in Table 3-14.

Table 3-14

Regeneration Frequency vs. Influent Arsenic Concentration for Activated Alumina

Influent Arsenic Level (µg/L)	No. of Bed Volumes Treated
5 to 15	56,800
15 to 35	33,600
greater than 35	22,400

These relationships were used in a previous arsenic removal cost estimation document (*National Compliance Assessment and Costs for Regulations of Arsenic in Drinking Water*, 1997). They were calculated using equilibrium modeling which assumed an average raw water quality matrix. A recent paper presenting operating data from an activated alumina plant in the United Kingdom agreed with the data given in the above table (Simms and Azizian, 1998). The water source contained approximately 22 µg/L of arsenic and the number of bed volumes treated as a function of pH is given in Table 3-15.

Table 3-15

Influent pH vs. Regeneration for Activated Alumina

Influent pH	Bed Volumes Treated to 10 µg/L limit
7.5	9,000
7	20,000
6.5	57,500
6	110,500

As shown by these results, the number of bed volumes is highly dependent on pH. Due to insufficient information available in the literature, the cost equations given in this document do not take influent pH into account. However, when estimating the cost for a particular system, the influent arsenic concentration and the influent pH should be taken into account when possible.

Very Small Systems (Less than 0.10 mgd)

Activated alumina capital and O&M costs for very small systems were developed based on equations given in the VSS Model. Since an assumed bed volume (BV) to breakthrough was not given, a BV of 2000 was assumed for the O&M costs since this was approximately the BV assumed by the W/W Cost model. O&M costs for other BVs were calculated in proportion to the assumed 2000 BV cost by using the proportional O&M costs for each regeneration frequency in the small systems model. The small systems model contains six AA processes, each with a different regeneration frequency. The proportions of these O&M costs for different regeneration frequencies were used to determine the O&M costs for the very small systems. For acid and base feed process costs, the Water Model was used since there were no appropriate very small systems equations.

The very small systems cost equation is based on design flow and EBCT. EBCT affects the volume of media which affects the size and number of AA units. The VSS Model also makes provisions for building (18.9 - 47.1 %), fencing (6.8 - 44.3 %) and road (4.3 - 30.0 %) costs.

Some very small systems may not be capable of optimizing process pH or providing the necessary maintenance for regeneration. Such systems will experience decreased operating efficiency, and be required to utilize alternative regeneration options, e.g., off-site regeneration. For those

systems with influent arsenic levels greater than 25 μg/L, O&M costs are best represented by those in Figure 3-55 (2000 BV). O&M costs for systems with **influent arsenic levels less than 25 μg/L are best represented by Figure 3-57 (10000 BV).**

The following design parameters were to develop capital and O&M cost estimates:

- EBCT of 15 minutes;
- Sulfuric acid feed, 70 mg/L;
- Caustic feed, 50 mg/L;
- Regenerant dose of 0.3 lb NaOH/kgal

Small Systems (Less than 1 mgd)

The Water Model was used to estimate capital and O&M costs for small treatment plants. O&M costs were determined in the same manner as described above for the very small systems. The design criteria used in development of costs are:

- EBCT of 15 minutes;
- Sulfuric acid feed, 70 mg/L;
- Caustic feed, 50 mg/L;
- Regenerant dose of 0.3 lb NaOH/kgal

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to determine the activated alumina capital and O&M costs for large systems. The following design criteria were assumed for determination of AA costs:

- Sulfuric acid feed, 70 mg/L;
- Caustic feed, 50 mg/L;
- Operation at pH 5.5;
- Low influent suspended solids;
- 10-foot deep beds;
- 100 psi working pressure;
- 80 percent bed expansion during backwash;
- Regeneration storage facilities sized for 30-day requirement;
- NaOH in solid phase for plants less than 10 mgd;
- NaOH in 50% solution for plants greater than 10 mgd;
- Media replacement of 10% per year.

The W/W Cost model assumes number of bed volumes to breakthrough is 1600. The O&M costs for other BVs is calculated in proportion to the assumed BV of 1600. The W/W Cost model was also used to calculate capital and O&M costs associated with acid and base feed. Figures 3-53 through 3-59 present capital and O&M cost curves and equations for AA.

3.8 ION EXCHANGE PROCESSES

3.8.1 Anion Exchange

Ion exchange (IX) is a physical/chemical treatment process in which an ion on the solid phase (IX resin) is exchanged for an ion in the feed water, thereby removing contaminants from the feed water. The IX removal mechanism is discussed in greater detail in Chapter 2. Ion exchange resin can be fouled by suspended and dissolved contaminants in the feed water. If the feed water contains suspended solids the IX process will need to be preceded by a pretreatment process, typically multi-media filtration. Also, source waters high in As(III) concentration may require pre-oxidation for conversion of arsenite to arsenate. Pre-oxidation is discussed in more detail in Section 3.5. Neither pre-oxidation or pre-filtration have been considered as part of the costs developed in this section.

Sulfate concentrations in the influent water significantly effect the capacity of the IX resin with respect to the removal of arsenic. Clifford (1993) estimated bed volumes for 10 percent and 50 percent breakthrough of influent arsenic as a function of influent sulfate concentration. Figure 3-60 shows the bed volume and sulfate relationship estimated by Clifford (1993). Figures 3-61 through 3-64 were developed using the relationship shown in Figure 3-60. Using these figures, the regeneration frequency for an IX column can be estimated if the influent arsenic, sulfate and target effluent arsenic concentrations are known. Straight line fits of the data points derived from Figure 3-60 are also shown on figures 3-61 through 3-64. Once the BV is known, the corresponding equation may be used to estimate the O&M cost. Capital costs will not be affected by changes in bed volume to regeneration.

Very Small Systems (Less than 0.10 mgd)

EPA developed a modified capital cost estimation method based upon the VSS Model and the Water Model (Kempic, 1994d). This method was used to develop the very small systems capital costs presented in this document. The ion exchange plant was assumed to be comprised of the following components:

- Pressure ion-exchange unit;
- Brine dilution pump;
- Brine pump;
- Salt storage tank;
- Pipes and valves;
- Instrumentation and controls.

The referenced EPA cost approach assigns percentages to each of the construction cost components (i.e., sitework, equipment, concrete, steel, labor and installation, pipes and valves, electrical, and housing) and escalates each component using the appropriate BLS cost index. The following design parameters were used in the development of the very small systems capital costs:

- EBCT of 2.5 minutes;
- Regenerant dose of 15 lb/ft^3 of resin;
- Regenerant frequency of once per day.

O&M costs for ion exchange process were based on the O&M cost equation given in the VSS Model. The use of these equations was confirmed in the revised EPA cost method presented above (Kempic, 1994d). The document assumed 576 bed volumes were treated per day. Based on this number of bed volumes, regeneration frequency was calculated for each assumed number of bed volumes to breakthrough. The regeneration frequency was used as an input to the O&M cost equation. The assumed labor requirement was 10 hours per week.

Small Systems (Less than 1 mgd)

The Water Model was used to estimate capital and O&M costs for small treatment plants. For the O&M costs, the Water Model assumes daily regeneration of the IX bed. However, the number of bed volumes treated per day is not given for this model. Therefore, a bed volume to breakthrough of 160 was assumed since this was the same number assumed in the W/W Cost program for large systems. The O&M costs for each bed volume to breakthrough were then calculated in proportion to the assumed regeneration frequency of 160 bed volumes.

Large Systems (Greater than 1 mgd)

The W/W Cost model was used to determine the ion exchange capital and O&M costs for large systems. The following design criteria were assumed for determination of IX costs:

- Six-foot deep IX beds;
- 100 psi working pressure;
- Nitrate = 100 mg/L;
- Sulfate = 80 mg/L;
- Other anions = 120 mg/L;
- Nitrate capacity = 7 kilograins/ft^3 resin;
- Regenerant requirement = 15 lb NaCl/ft^3;
- Regeneration time = 54 minutes;
- Backwashing time = 10 minutes;
- Rinsing time = 24 minutes;
- 25% resin replacement per year.

In the W/W Cost model, the number of bed volumes to breakthrough is assumed to be 160. The O&M costs for other BVs is calculated in proportion to the assumed regeneration frequency of 160 BVs.

Figures 3-65 through 3-70 present capital and O&M cost curves and equations for ion exchange. **All costs include redundant exchange beds**, i.e., beds for use during maintenance of another that keep the system on-line. Table 3-16 presents the number of IX beds included in the cost estimates provided.

Table 3-16

Number of IX Beds Included in Cost Estimates[1]

Plant Capacity (mgd)	Number of IX Beds
< 1.2	2
1.2 - 3.9	3
3.9 - 6.5	5
6.5 - 13	10
> 13	10

1- Culp/Wesner/Culp (1979, 1984)

3.9 SEPARATION PROCESSES

3.9.1 Microfiltration

Microfiltration is a low-pressure membrane process which has only a marginal ability to remove arsenic due to its relatively large pore size in comparison to other membrane processes. MF removes contaminants from a feed stream primarily through sieving. Typically, MF does not require pretreatment beyond approximately 500-µm prefiltration.

As previously stated, MF is not an effective stand alone technology for removal of arsenic. Due to this fact, capital and O&M cost estimates for removal of arsenic by MF are not provided in this chapter. Appendix E does contain estimated costs for MF treatment.

The Water and W/W Cost models used in determination of other process costs in this document do not include MF. Although the VSS Model did include MF, the equations were only used as a final check for reasonableness of costs for systems with a capacity less than 0.10 mgd. Since MF is a relatively new technology, very little cost information is contained in the literature. However, the MF cost information which currently exists provided the basis for the estimates contained in Appendix E. This information consisted of cost surveys from existing plants and vendor quotes. Vendor quotes were obtained from Memtec, a manufacturer of package MF systems. Cost survey information was obtained from several sources. The primary source was a survey of 21 MF plants conducted as part of a 1996 AWWARF study.

The largest MF plant which provided cost information had a capacity of 20 mgd. For this reason, the economies-of-scale which exist beyond a capacity of 20 mgd could not be accurately estimated. Therefore, no economies-of-scale were assumed beyond the boundary condition of 20 mgd for either capital or O&M costs.

3.9.2 Ultrafiltration

Ultrafiltration is a low-pressure membrane process which removes contaminants from a feed stream primarily through sieving. Typically, UF does not require pretreatment beyond approximately 500-µm prefiltration. UF has the benefit of being lower in both capital and O&M costs than high-pressure membrane processes.

Although the W/W Cost and VSS models included UF, only the W/W Cost model was used for UF cost calculations. The very small systems cost equations were used only as a backcheck against final UF costs. The W/W Cost estimation was valid to a capacity of 1 mgd. Since the time when the W/W Cost model was assembled, however, UF membrane module costs have decreased by approximately 30 percent. For this reason, the membrane module portion of the capital costs were reduced by 30 percent to account for this. Also, the membrane replacement portion of the O&M costs were reduced by 30 percent to account for this as well. Actual plant cost information was also used for UF cost estimation. Since UF is a new technology, however, very little cost information is contained in the literature. The UF cost information which currently exists was used in calculation of the UF capital and O&M costs. This information consisted of cost surveys from existing plants. Cost survey information was obtained from several sources. The primary source was a survey of 7 UF plants conducted as part of an AWWARF study in 1996.

The largest UF plant which provided cost information had a capacity of 28 mgd. For this reason, the economies-of-scale which exist beyond a capacity of 28 mgd could not be accurately estimated. Therefore, no economies-of-scale were assumed beyond the boundary condition of 28 mgd for either capital or O&M costs.

Figures 3-71 through 3-76 present capital and O&M cost curves and equations for UF.

3.9.3 Nanofiltration

Nanofiltration is a high-pressure membrane process capable of significant arsenic removal. NF removes contaminants from a feed stream primarily through a combination of diffusion and sieving mechanisms. Typically, NF requires pretreatment to remove suspended solids and other foulants from the feed stream. NF has greater arsenic removal capabilities than low-pressure membrane processes, however, capital and O&M costs for NF are usually greater than equivalent costs for low-pressure processes.

The W/W Cost and Water models used in the determination of other process costs in this document did not include NF. Although the VSS Model did include NF, the cost equations were used only as a backcheck against final NF costs. Since NF is a relatively new technology, very little cost information is contained in the literature. However, the NF cost information which currently exists provided the basis for the NF cost calculations given in this document. This information consisted of cost surveys from existing plants and vendor quotes. Cost survey information was obtained from a cost survey performed by Bergman (1996). A summary of how the cost estimates were derived from this cost survey data is given for large and small systems below.

Small Systems (Less than 1 mgd)

Capital costs for small systems, i.e. less than 1 mgd, were determined by contacting vendors to obtain membrane cost estimates for small systems. The cost information that was obtained consisted of costs for complete treatment systems. This was done in 1992 during the regulatory negotiations of the DBP Rule as part of the work conducted by the Technologies Working Group (TWG). The TWG concluded that these costs were representative of small system NF costs. For this document, the 1992 costs were updated to 1998 dollars using the BCI.

Large Systems (Greater than 1 mgd)

Capital and O&M cost estimates for large systems, i.e. greater than 1 mgd, were developed from NF cost data presented in a NF plant survey conducted by Bergman (1996). Costs presented in this survey were escalated to 1997 using the 1997 and 1995 ENR Building Cost Indices. Capital and O&M costs presented here were derived from cost data submitted by existing plants. The Bergman survey, however, presented capital cost and O&M cost data obtained between 1988 and

1996. It is recognized that spiral-wound membrane modules, which include the majority of NF membranes, have decreased in cost significantly in recent years. Based on vendor information, costs for spiral-wound membrane modules have been reduced by approximately 50 percent over the past five years. For this reason, costs for membrane modules presented in the above reference obtained between 1988 and 1995 were reduced by 50 percent. The reduced cost items included new membrane capital costs and O&M membrane replacement costs.

The costs given in the Bergman survey consist solely of nanofiltration costs from Florida plants. The source waters treated by these plants are warm, resulting in higher membrane flux values than potential flux values for lower temperature waters of comparable quality. As a result, the costs presented in these references may not be representative of costs for all areas of the country. For this reason, the costs were adjusted to equivalent costs at 20 degrees Celsius. This was accomplished by assuming a temperature of 25 degrees Celsius for the Florida plants and adjusting the membrane capital costs and O&M membrane replacement costs to account for the additional membrane area that would be required at a lower temperature. The temperature correction equation for permeate flux $J_T/J_{25}=1.03^{(T-25)}$ was used for these calculations (Wiesner and Aptel, 1996).

Best-fit curves were generated for capital and O&M costs. Capital costs were separated into membrane module costs and facility costs. Using the Bergman survey data, the average facility cost was found to be between two and three times the cost for membrane equipment. However, capital and operational costs for clearwells and high service pumping will not be required in a retrofit situation. Subtracting the capital costs for these two components results in a factor of 1.5 to 2.0 for facility costs when compared to membrane system costs. For this reason, facility costs used in determining the best-fit equation were calculated by multiplying the membrane cost for each plant by two (a conservative estimate). It should be noted that the largest plant surveyed was 14 mgd. Since at the present time very few facilities above this capacity exist, there is no way to accurately judge the economies-of-scale that may be seen beyond this point. For this reason, it was conservatively assumed that no economies-of-scale would exist beyond 14 mgd.

Figures 3-77 through 3-84 present capital and O&M cost curves and equations for NF.

3.9.4 Reverse Osmosis

Reverse Osmosis is a high-pressure membrane process which removes dissolved contaminants from a feed stream primarily through diffusion rather than physical straining. RO requires a high quality feed stream and often requires substantial pretreatment to remove suspended solids and other foulants. RO also often requires pH adjustment after the membrane process and may require the addition of an anti-scalant before the membrane process. For the purpose of this analysis, costs were not provided for a substantial pre-treatment system, other than the anti-scalant system. RO has the benefit of greater arsenic removal compared to low-pressure membrane processes, but is typically associated with higher capital and O&M costs.

Both the VSS Model and the W/W Cost Model included cost estimation for RO. Since the W/W Cost Model was assembled, however, RO spiral-wound membrane module costs have decreased by approximately 50 percent. For this reason, the membrane module portion of the capital costs were reduced by 50 percent. The membrane replacement portion of the O&M costs were also reduced by 50 percent to account for reductions in membrane costs. The W/W Cost Model for RO was only valid up to a capacity of 200 mgd. For this reason, no economies-of-scale were assumed for plants with a capacity larger than the boundary condition of 200 mgd. The model also makes an assumption that recovery is 80% for systems 1 to 10 mgd, and 85% for systems larger than 10 mgd. Costs were adjusted to reflect a recovery of 75%.

Figures 3-85 through 3-92 present capital and O&M cost estimates for RO.

3.10 GREENSAND FILTRATION

Greensand filtration is an oxidation filtration process that has demonstrated effectiveness for the removal of arsenic. Greensand is a zeolite-type glauconite material which is produces by treating glauconite sand with $KMnO_4$ until the granular material (sand) is coated with a layer of manganese oxides, particularly manganese dioxide. Arsenic compounds displace species from the manganese oxide (presumably OH^- and H_2O), becoming bound to the greensand surface - in effect an exchange of ions. The oxidative nature of the manganese surface converts As(III) to As(V), and As(V) is adsorbed to the surface.

The cost models make no appropriation for estimating oxidation filtration technology costs. As a result, the cost estimates for greensand filtration are based upon vendor supplied data, and were originally published in *Technologies and Costs for Removal of Radionuclides from Potable Water Supplies* (Malcolm Pirnie, 1992; ICI, 1999). Capital and O&M estimates presented in that document were escalated to 1998 dollars using the ENR Building Cost Index, and the BLS Chemical and Allied Products Index, respectively. Costs are based upon the following design and operating criteria:

- Chlorine feed and potassium permanganate feed systems are provided;
- Filter media is contained in a ferrosand continuous regeneration filter tank equipped with an underdrain; and
- Backwash is sufficient for 40 percent bed expansion.

Figures 3-93 through 3-100 present cost estimates for removal of arsenic by greensand filtration.

3.11 COMPARISON OF COSTS

Capital and O&M cost estimates in this chapter were compared with actual cost data presented in *Evaluation of Full-Scale Treatment Technologies at Small Drinking Water Systems* (ICF and ISSI, 1998) and estimates found in *Evaluation of Central Treatment Options as Small System Treatment Technologies* (SAIC, 1999). It was found that the estimates presented in this document are reasonable. Capital cost estimates were routinely conservative, but followed the general trends seen in actual data. O&M estimates typically represented an approximate average of the actual costs. Actual data was not available for all technologies, and as a result, comparisons are not presented for some of the technologies discussed in this document.

Figures 3-101 through 3-116 provide graphical representations of comparisons between the cost estimates in this document (labeled "ICI, 1999" on the figures) and actual cost data ("Actual Data" on the figures). Also, where applicable, cost data are included from: 1) the 1993 T&C document ("EPA, 1993"); and 2) the EPA report, *Evaluation of Central Treatment Options as Small System Treatment Technologies* ("SAIC, 1999") prepared by SAIC (1999). The ICI 1999 curves are based on the cost equations presented in this chapter. For the purpose of comparison, it was assumed that 100 percent of the process flow is treated.

3.11.1 Capital Cost Comparison

Capital cost estimates were routinely conservative and followed the general trends seen in actual data. This trend is noted in the comparison of capital cost data for coagulation/filtration. The majority of actual data points fall below the SAIC 1999 and ICI 1999 curves, indicating a conservative estimate projected by both SAIC and ICI. Data reported by EPA in 1993 for the coagulation/filtration technology tends to conform with the ICI 1999 data.

The majority of actual data projected for direct filtration also fall below the SAIC 1999 and ICI 1999 curves, indicating a conservative estimate. No direct filtration data was developed by EPA in 1993.

One actual data point was available for lime softening capital cost comparison. Although there is a sparsity of actual data, the similarity of the SAIC and ICI curves suggests that the curves are a realistic representation of actual capital costs.

Three actual data points were available for the activated alumina capital cost comparison. The capital cost curves developed by SAIC and ICI are the approximate average of the "Actual Data," and the three data points fall within a reasonable distance of the curves.

Ion exchange capital cost curves developed by SAIC and ICI are similar and are the approximate average of the "Actual Data," and the EPA 1993 data.

Very little actual data exists on nanofiltration and ultrafiltration. Comparisons of ICI 1999 and SAIC 1999 capital costs show similar expenditures indicating realistic estimates.

The reverse osmosis comparison of capital costs shows that the majority of actual data points fall below the SAIC 1999 and ICI 1999 curves, indicating a conservative estimate projected by both SAIC and ICI.

Two actual data point were available for greensand filtration capital costs. Both data points were below the ICI 1999 curve, indicating a conservative estimate.

In many cases EPA 1993 estimates fall below the ICI 1999 estimates. This is due to the fact that 1993 estimates included accessory costs, i.e., raw and finished water pumping and clearwell storage.

3.11.2 O&M Cost Comparison

O&M estimates typically represent an approximate average of the actual costs among the technologies discussed. The O&M cost curve developed in this document for coagulation/filtration shows that trend. The majority of actual data points fall evenly on each side of the ICI 1999 curve. The SAIC 1999 curve falls below the actual data and tends to be more liberal.

One actual data point for direct filtration O&M cost falls between the SAIC and ICI 1999 curves, indicating a realistic representation of the O&M cost curves for the direct filtration technology.

The SAIC and ICI 1999 curves for O&M costs for lime softening are the approximate average of the actual data points plotted.

Two actual data points were available for the activated alumina O&M cost comparison. The O&M cost curves developed by SAIC and ICI project a conservative estimate when compared to the actual data.

Ion exchange O&M cost curves developed by SAIC and ICI are the approximate average of the actual data.

Again, little actual data was available on nanofiltration and ultrafiltration O&M costs. The SAIC 1999 curve for nanofiltration costs was considerably lower than the ICI 1999 curve, which followed the trend of 1993 data. Ultrafiltration estimates were very similar above 0.1 mgd, with the SAIC estimate being much more conservative below that production level.

The comparison of reverse osmosis O&M cost curves indicate that the curves are an approximate average of the actual data points. The ICI 1999 curve is a little more conservative than the SAIC 1999 curve.

The ICI 1999 cost curve developed for greensand filtration O&M indicates a projected conservative estimate. The actual data points fall below, but close to, the ICI curve.

Estimates presented in this chapter were, in many cases, less expensive than estimates presented in the 1993 T&C document. This is due to the fact that the 1993 document included accessory O&M for many of the processes.

4.0 RESIDUALS HANDLING AND DISPOSAL ALTERNATIVES

4.1 INTRODUCTION

Each of the treatment technologies presented in Chapter 3 will produce residuals, either solid or liquid streams, containing elevated levels of arsenic. It is the purpose of this chapter to present the characteristics of arsenic waste generated by each of the treatment technologies and discuss appropriate handling and disposal options. Costs for residuals handling and disposal are not presented in this chapter, however, references which contain appropriate cost information are noted.

4.1.1 Factors Affecting Residuals Handling and Disposal Costs

There are a number of factors which can influence residuals handling and disposal costs. This discussion is concerned with factors affecting capital cost, as well as factors affecting operations and maintenance (O&M) costs. Capital costs include equipment, construction, installation, contractor overhead and profit, administrative and legal fees, land, and other miscellaneous costs. The primary factor affecting capital cost is the amount of residuals produced, which is dependent upon the design capacity of the water treatment plant and the treatment process utilized (e.g., coagulation/filtration vs. lime softening).

The amount of waste generated plays a significant role in determining the handling and disposal method to be utilized. Many handling methods which are suitable for smaller systems are impractical for larger systems because of the significant land requirements. For larger systems that process residuals on-site (as opposed to direct or indirect discharge), mechanical methods are typically used because of the limited land requirements.

Operations and maintenance costs include labor, transportation, process materials and chemicals, and maintenance. Many handling and disposal methods require extensive oversight which can be a burden on small water systems. Generally, labor intensive technologies are more suitable to large water systems. Transportation can also play a significant role in determining appropriate handling and disposal options. If off-site disposal requires extensive transportation, alternative disposal methods should be evaluated. Complex handling and disposal methods usually require more maintenance.

4.l.2 Methods for Estimating Residuals Handling and Disposal Costs

Residuals handling and disposal costs can be difficult to estimate. There are a number of factors which affect capital and O&M costs, and disposal costs can be largely regional. EPA has published two manuals for estimating residuals handling and disposal costs; Small Water System Byproducts Treatment and Disposal Cost Document (DPRA, 1993a), and Water System Byproducts Treatment and Disposal Cost Document (DPRA, 1993b). Both present a variety of handling and disposal options, applications and limitations of those technologies, and capital and O&M cost equations.

4.2 RESIDUALS HANDLING OPTIONS

4.2.1 Gravity Thickening

Gravity thickening increases the solids content of filter backwash, sedimentation basins and treatment process sludges. It is generally used as a pre-treatment for mechanical dewatering processes, evaporation ponds and storage lagoons.

Filter backwash streams are high volume, low solids slurries generated during the cleaning of granular filter media. Backwash volume depends upon the number of filters and cleaning frequency. Typical volumes range from 0.5 to 5 percent of the processed water flow with larger plants creating less backwash per million gallons produced than small systems due to increased plant efficiency (DPRA, 1993a). Backwash waters have an average solids concentration of 0.8 percent, compared to coagulation sludges which are typically 0.5 to 2.0 percent (DPRA, 1993a).

When possible, backwash waters are recycled to the treatment process. In gravity thickening, backwash waters are fed to a tank where settling occurs naturally. Sludges are discharged and further treated for ultimate disposal, and the decant is either recycled or discharged to a surface water or POTW. Gravity thickening reduces the quantity of water lost due to backwashing, as well as the total quantity of sludge generated (DPRA, 1993a). When recycling is not feasible, backwash waters may be discharged to a surface water, publicly-owned treatment works (POTW), or be treated by other mechanical or non-mechanical dewatering processes. When backwash slurries cannot be recycled or discharged to a surface water of POTW, they must be treated and disposed.

4.2.2 Mechanical Dewatering

Mechanical dewatering processes include centrifuges, vacuum-assisted dewatering beds, belt filter presses, and plate and frame filter presses (DPRA, 1993a). Such processes generally have high capital, as well as high O&M costs, compared to similar capacity non-mechanical dewatering processes, e.g., storage lagoons. Due to the high costs such processes are generally not suitable for application at very small water systems.

Filter presses have been used in industrial processes for years, and their use has been increasing in the water treatment industry over the past several years. These devices have been successfully applied to both lime and alum sludges. Prior to pressure filtration, alum sludges may require the addition of lime to lower the resistance of the sludge to filtration. This is generally done by adjusting the pH to approximately 11. Pre-conditioning also increases the sludge volume by as much as 20 to 30 percent. Lime sludges can attain final solids concentrations of 40 to 70 percent, while alum sludges may reach 35 to 50 percent total solids. Filter presses require little land, have high capital costs, and are labor intensive (DPRA, 1993a). Capital and O&M costs are generally higher than comparable non-mechanical dewatering alternatives. As a result, pressure filtration is most applicable to larger water systems.

Centrifuges have also been used in the water industry for years. They are capable of producing alum sludges with final solids concentrations of 15 to 30 percent and lime sludges with 65 to 70 percent total solids, based upon an influent solids concentration of 1 to 10 percent. Centrifugation is a continuous process requiring minimal time (8 to 12 minutes) to achieve the optimal sludge solids concentration. Centrifuges have low land requirements and high capital costs. They are more labor intensive than non-mechanical alternatives, but less intensive than filter presses. Again, due to the capital and O&M requirements centrifuges are more suitable for larger water systems.

4.2.3 Evaporation Ponds and Drying Beds

Evaporation ponds and drying beds are non-mechanical dewatering technologies wherein favorable climatic conditions are used to dewater waste brines generated by treatment processes such as reverse osmosis and ion exchange (DPRA, 1993a). Brine waste is discharged to a pond for storage and evaporation. Ponds and drying beds are not generally suitable for alum and lime sludges.

Typically, such ponds are designed with large surface areas to allow the sun and wind to effectively evaporate residual water. Size is determined by waste flow and storage capacity requirements.

Evaporation ponds and drying beds are used primary for brine wastes generated by reverse osmosis and ion exchange processes. Such processes produce large volumes of high TDS waste streams and make mechanical dewatering processes, such as filter presses, impractical. Depending upon the solids concentration of the brine waste stream, intermittent removal of solids may be required. For brines with a total dissolved solid (TDS) content ranging from 15,000 to 35,000 mg/L, solids will accumulate in the pond at a rate of ½ to 1½ inches per year (DPRA, 1993a). When the depth of the solids reaches a predetermined level, flow to the pond is halted and evaporation continues until the solids concentration is suitable for disposal.

Evaporation is an extremely land intensive handling option requiring shallow basins with large surface areas. This can be an important consideration in densely populated regions. Reverse osmosis produces a very large volume reject stream which increases the land requirement and ultimately construction costs. As a result, evaporation ponds may not be suitable for large water systems utilizing reverse osmosis. Evaporation ponds and drying beds have few operations and maintenance requirements, but are only feasible in regions with favorable climatic conditions, i.e., high temperatures, low humidity and low precipitation (DPRA, 1993a). Waste streams with low TDS concentrations can allow a pond to operate for several years before solids accumulation warrants removal.

4.2.4 Storage Lagoons

Lagoons are the most common, and often least expensive, method to thicken or dewater treatment sludges; however, they are land intensive (DPRA, 1993a). Lagoons are lined ponds designed to collect and dewater sludge for a predetermined period of time. Dewatering occurs by evaporation and decanting of the supernatant. Lagoon size is determined by the volume of sludge produced and the storage time desired. As with evaporation ponds, when a lagoon reaches the design capacity solids can be removed with heavy equipment and shipped for disposal.

Storage lagoons are best suited for dewatering lime softening process sludges, though they have been applied with some success to coagulation/filtration process sludges. They can operate under a variety of sludge flows and solids concentrations, and do not require chemical conditioning of alum sludges (DPRA, 1993a). Typically, lime sludges enter the lagoon at three percent solids, and can be dewatered to 50 to 60 percent solids, whereas alum sludges enter at one percent solids and can be dewatered to 7 to 15 percent solids (DPRA, 1993a). Alum sludges do not typically dewater well in storage lagoons. When the top layer of sludge is allowed to dry, it hardens, sealing moisture in the layers below. Even after several years, alum sludges may require additional dewatering to achieve the 20 percent solids content required at most landfills (DPRA, 1993a). Further, thickened alum sludges can be difficult to remove from lagoons, and often require dredging or vacuum pumping by knowledgeable operators.

As previously stated, lagooning is a land intensive process with limited applicability in densely populated areas, or areas with limited land availability. Such areas need to compare the cost of regular lagoon cleaning and disposal with land acquisition costs. Lagoons are best suited for areas with favorable climatic conditions, i.e., high temperatures, low humidity and low precipitation. In northern climates, winter freezing can dehydrate alum sludges.

4.3 DISPOSAL ALTERNATIVES

4.3.1 Direct Discharge

Direct discharge to a surface water is a common method of disposal for water treatment byproducts. No pretreatment or concentration of the byproduct stream is necessary prior to discharge, and the receiving water dilutes the waste concentration and gradually incorporates the sludge or brine (DPRA, 1993a).

Discharge of liquid residuals containing arsenic to a surface water will be subject to compliance with the National Pollution Discharge Elimination System (NPDES). NPDES establishes limits based upon a variety of factors, including ambient contaminant levels, low flow condition of the receiving water, and design flow of the proposed discharge. Most NPDES limits for solids discharge are around 30 mg/L (AWWARF, 1998).

EPA has established water quality criteria under authority of the Clean Water Act. For waters used for fish consumption the ambient water quality criterion for arsenic was set at 0.14 μg/L. If the water source is used for drinking as well, that limit is reduced to 0.0175 μg/L. These critera will be used by state regulatory agencies to determine discharge limitations for arsenic depending upon the classification of the receiving water. The allowable discharge is therefore affected by the ability of the receiving water to assimilate the arsenic without exceeding the water quality criteria.

The primary cost associated with direct discharge is that of the piping. Accommodations must be made for washout ports to prevent clogging because of sedimentation in pipelines. Valving is necessary to control waste flow in the event of pipe bursts, and pipe must be laid at a sufficient depth to prevent freezing in winter months. Direct discharge requires little oversight, and operator experience and maintenance requirements are minimal. This method has been used to successfully dispose of alum and lime sludges, as well as brine streams generated at reverse osmosis and ion exchange water plants (DPRA, 1993a).

4.3.2 Indirect Discharge

In some cases, water treatment process sludges, slurries and brines may be discharged to a POTW. This most often occurs when the treatment plant and POTW are under the same management authority. This may require addition of a conveyance system to access the sanitary sewer if an adequate system is not already in place (DPRA, 1993a).

Indirect discharge is a commonly used method of disposal for filter backwash and brine waste streams. Coagulation/filtration and lime softening sludges have also been successfully disposed of in this manner. However, the POTW must be able to handle the increased hydraulic and solids loading. The capacity of the sewer system must also be considered when selecting indirect discharge as a disposal option.

The residuals generated from an arsenic treatment process will be classified as an industrial waste since it contains contaminants, namely arsenic, which may impact the POTW. As a result, discharge to a POTW is only acceptable when arsenic concentrations fall within the established Technically Based Local Limits (TBLL) of the current Industrial Pretreatment Program (AWWARF, 1998). The Industrial Pretreatment Program serves to prevent NPDES violations, as well as unacceptable accumulation of contaminants in POTW sludges and biosolids. TBLLs are individually determined for each POTW, and take into account background levels of contamination in the municipal wastewater. TBLLs for arsenic will typically be limited by the contamination of biosolids rather than effluent limitations or process inhibition (AWWARF, 1998).

40 CFR 503 specifies the allowable limits for arsenic concentration in biosolids as a function

of disposal method. POTWs utilizing land application are subject to the Land Disposal Limit, Land Application Ceiling Limit, and Land Application Clean Sludge Limit which are 73 mg/kg, 75 mg/kg and 41 mg/kg, respectively. If the arsenic concentration exceeds the Clean Sludge criteria, land application is limited to 41 kg per hectare (36.6 lb/acre). As a result, most TBLLs are based upon the Clean Sludge criterion. The typical POTW removal efficiency for arsenic is approximately 45 percent. Assuming biosolids production is around 1,200 pounds per million gallons of water treated, the maximum allowable headworks loading will be around 0.109 pounds of arsenic per million gallons of wastewater treated. This equates to a total (municipal and industrial) influent concentration of around 13 μg/L (AWWARF, 1998). As a result, if a water system has a background arsenic concentration near 13 μg/L, it may not be possible to discharge to the sanitary sewer.

The primary cost associated with indirect discharge is that of the piping. Accommodations must also be made for washout ports to prevent clogging because of sedimentation in pipelines. Valving is necessary to control waste flow in the event of pipe bursts, and pipe must be laid at a sufficient depth to prevent freezing in winter months. Additional costs associated with indirect discharge may include lift stations, additional piping for access to the sewer system, or other surcharges to accommodate the increased demands on the POTW.

4.3.3 Dewatered Sludge Land Application

Dewatered sludge can be disposed by spreading the material over an approved land surface. Application is dependent upon several variables, including soil and sludge chemistry or the crop planted in the application field. Dewatered sludges are typically stored on site until they are transported for application. Monitoring of soils, run off from land application, and potentially affected water sources is advisable to protect open land that may become cropland and to protect local water quality (DPRA, 1993a).

As discussed in the previous section, land application of water treatment residuals containing arsenic is limited to 41 mg/kg. If these concentrations cannot be achieved application of sludges is limited to 41 kg arsenic per hectare. Due to the possibility of arsenic absorption by vegetation, non-food chain fields are preferred for application. Land application is also limited by the availability of land. In areas where grassland, farmland or forested land is unavailable, transportation can significantly affect the cost effectiveness of this disposal option.

Land application can be a means of final disposal of lime softening, and to a lesser degree coagulation/filtration, sludges. Lime sludges can be used in farmland to neutralize soil pH in place of other commercial products. Alum sludges offer no benefit to soil chemistry and are generally used as fill material.

4.3.4 Sanitary Landfill Disposal

Two forms of sanitary landfill are commonly used for disposal of water treatment byproducts: monofills and commercial nonhazardous waste landfills (DPRA, 1993a). Monofills only accept one type of waste, for example, fly ash or water treatment sludges. Commercial nonhazardous waste landfills accept a variety of commercial and industrial wastes.

Sanitary landfills are regulated by both state and federal regulations. States have guidelines on what types of waste can be landfilled, and determine construction and operation criteria. In many cases, state requirements are more stringent than the federal regulations promulgated under the Resource Conservation and Recovery Act (RCRA). The federal requirements include restrictions on location, operation and design criteria, ground water monitoring requirements, corrective action requirements, closure and post-closure requirements, and financial assurance.

Landfill disposal requires that residuals be in a solid form and contain no free liquids. Sanitary landfill disposal also requires that sludges meet specific criteria that determine if a waste is hazardous. 40 CFR 261 establishes four characteristics of hazardous waste: flammability, corrosivity, reactivity and toxicity. A waste must meet only one of the criteria to be considered hazardous. With treatment residuals containing arsenic, toxicity is the primary characteristic of concern.

EPA has established an analytical method, the Toxicity Characteristic Leaching Procedure (TCLP), to measure the toxicity of a waste. The current TCLP limit for arsenic is 5 mg/L, which is 100 times the current MCL of 50 μg/L. If the MCL is lowered in the future, the TCLP value will be lowered accordingly. For example, if the MCL were lowered to 20 μg/L or 2 μg/L, the TCLP would be lowered to 2.0 mg/L or 200 μg/L, respectively. As a result, water treatment residuals containing arsenic may meet current sanitary landfill disposal criteria, but may not under a future regulatory framework.

Many water treatment facilities currently dispose of their waste in commercial or public-owned landfills (DPRA, 1993a). In some parts of the country, decreasing landfill availability, rising costs and increasing regulations are making landfill disposal more expensive. As a result, the benefits of monofills are being discussed within the industry. Costs associated with development of monofills is generally less than that of a sanitary landfill (DPRA, 1993a). Monofills control the type of waste disposed more strictly and limit the potential future liabilities, as well.

4.3.5 Hazardous Waste Landfill Disposal

Water treatment residuals containing arsenic which fail the TCLP test for toxicity must be disposed in a designated and licensed hazardous waste landfill. Hazardous waste landfills are regulated by the federal government under authority of RCRA or by individual states who have received authorization under RCRA. Hazardous waste landfills are required to be permitted in accordance with 40 CFR 270 which specifies landfill construction and operation criteria, and are designed to isolate hazardous contaminants from the environment.

The primary limitation affecting disposal of arsenic containing residuals in a hazardous waste landfill is the presence of free liquids. If any water treatment sludge contains free liquids, usually determined by the Paint Filter Liquids Test (SW-846, Method 9095), it is not suitable for landfilling. Sludges containing free liquids must be stabilized or treated by another method to remove free liquids prior to disposal in a hazardous waste landfill.

These facilities have extensive monitoring and operational requirements which cause the cost of this method of disposal to be much greater than that of a typical sanitary landfill (AWWARF, 1998). If the residuals are determined to be hazardous, transportation to the landfill must be manifested and the owner may never be free of responsibility for that waste. As a result, production of a hazardous arsenic residual should be avoided if at all possible. Hazardous waste landfill disposal is the most expensive disposal alternative discussed in this document, and should be used only after all other disposal options have been exhausted.

4.4 RESIDUALS CHARACTERISTICS

4.4.1 Coagulation/Filtration

Coagulation/filtration (C/F) residual production is a function of coagulant type and suspended solids content. For alum coagulation, approximately 0.26 pound of solids are produced for every pound of alum added. For ferric coagulation, approximately 0.54 pounds of solid are produced for each pound of ferric chloride added (AWWARF, 1998). Sludge production is also affected by the suspended solids content of the raw water.

Sludges removed from C/F sedimentation basins are high in water content and typically have a solids content of less than 1.0 percent (AWWARF, 1998). As a result, such sludges are usually discharged to a sanitary sewer or dewatered by one of the methods discussed earlier in this chapter. Discharge to sanitary sewers is generally only an option for treatment plants with an average flow of less than 10 million gallons per day.

Filter backwash is a high volume liquid waste stream with a solids content generally less than 1.0 percent. Typical volumes range from 1.0 to 2.0 percent of the treated flow. Backwash streams are typically discharged to a sanitary sewer or processed using one of the mechanical methods discussed in this chapter. As with sedimentation sludges, discharge of filter backwash streams to a sanitary sewer is generally only an option for treatment plants with an average flow of less than 10 million gallons per day.

The concentration of arsenic in the sedimentation sludges is a function of the amount of coagulant used and the removal efficiency of the process. Using the design criteria specified in Chapter 3, consider a treatment plant that utilizes ferric chloride at a dose of 25 mg/L. Neglecting the impact of suspended solids, approximately 110 pounds of solids are produced for each million gallons of water treated. The sedimentation sludges, or blowdown, will have a solids concentration of 1.0 percent and be produced at a rate of 1,400 gallons per million gallons of water treated. Assuming 40 μg/L of arsenic is removed from the influent during the treatment proces, the result is 0.33 pounds of arsenic per million gallons of water treated. The resulting residuals arsenic concentration is 28.0 mg/L, or 3,000 mg/kg on a dry weight basis. Filter backwash will be produced at a rate of 10,000 gallons per million gallons of water produced, and will have an approximate solids concentration of 1 percent (10,000 mg/L).

Selection of Handling and Disposal Options

C/F blowdown and filter backwash are high volume, low solids content waste streams. Gravity thickening may be used as a pretreatment for C/F sludges and backwash prior to handling by other mechanical or non-mechanical dewatering processes. Filter presses are capable of attaining final solids contents in the range of 35 to 50 percent, while scroll centrifuges may achieve final solids contents of 15 to 30 percent. Evaporation ponds and storage lagoons may be suitable for smaller treatment plants, but because they are land intensive may not be applicable for large water systems.

Disposal of C/F arsenic residuals is largely dependent upon influent arsenic concentration, coagulant dose and suspended solids content. Disposal by direct discharge to a surface water is not likely. In the example presented earlier in this section, the typical solids concentration of 1.0 percent, or 10,000 mg/L, far exceeds the usual NPDES limit of around 30 mg/L.

The blowdown arsenic concentration in the above example is 28,000 μg/L. Most Industrial Protection Programs have TBLLs in the range of 50 μg/L to 1,000 μg/L (AWWARF, 1998). As a result, it is unlikely that indirect discharge to a POTW is an acceptable disposal method for C/F sludges.

Depending upon the arsenic concentration of C/F sludges, land application may be a suitable disposal method. Total arsenic should not exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with higher arsenic concentration may be land applied providing the total loading does not exceed 41 kg per hectare. In the previous example, the arsenic concentration is 2,930 mg/kg on a dry weight basis. As a result, this sludge may be land applied, but loading must not exceed 41 kg per hectare.

All C/F sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that most C/F sludges will pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.2 Enhanced Coagulation

Enhanced coagulation is a modified C/F process that includes increased coagulant dosage, reduction in process pH, or both. As a result, enhanced coagulation process residuals are nearly identical to typical C/F residuals. The exception is increased solids production as a result of the increased coagulant dosage.

Sludges removed from enhanced coagulation sedimentation basins are high in water content and typically have a solids content of approximately 1.0 percent (AWWARF, 1998). As a result, such sludges are usually discharged to a sanitary sewer or dewatered by one of the methods discussed earlier in this chapter. Discharge to sanitary sewers is generally only an option for treatment plants with an average flow of less than 10 million gallons per day.

Filter backwash is a high volume liquid waste stream with a solids content generally less than 1.0 percent. Typical volumes range from 1.0 to 2.0 percent of the treated flow. Backwash streams are typically discharged to a sanitary sewer or processed using one of the mechanical methods discussed in this chapter. As with sedimentation sludges, discharge of filter backwash streams to a sanitary sewer is generally only an option for treatment plants with an average flow of less than 10 million gallons per day.

The concentration of arsenic in the treatment process residuals is a function of the amount of coagulant used and the removal efficiency of the process. Using the design criteria specified in Chapter 3, consider a treatment plant that utilizes ferric chloride at a dose of 35 mg/L. Neglecting the impact of suspended solids, approximately 160 pounds of solids are produced for each million gallons of water treated. The sedimentation sludges, or blowdown, will have a solids concentration of 1.0 percent and be produced at a rate of 1,900 gallons per million gallons of water treated. Assuming 40 μg/L of arsenic is removed from the influent during the treatment process, the result is 0.33 pounds of arsenic per million gallons of water treated. The resulting residuals arsenic concentration is 21 mg/L, or 2,060 mg/kg on a dry weight basis. Filter backwash will be produced at a rate of 10,000 gallons per million gallons of water produced, and will have an approximate solids concentration of 1 percent (10,000 mg/L).

Selection of Handling and Disposal Options

Enhanced coagulation blowdown is a high volume, low solids content waste stream. Typical solids contents range from 0.5 to 2.0 percent, depending upon the coagulant type. Gravity thickening may be used as a pretreatment for C/F sludges prior to handling by other mechanical or non-mechanical dewatering processes. Filter presses are capable of attaining final sludge solids contents in the range of 35 to 50 percent, while scroll centrifuges may achieve final solids contents of 15 to 30 percent. Evaporation ponds and storage lagoons may be suitable for smaller treatment plants, but because they are land intensive may not be applicable for large water systems.

Disposal of enhanced coagulation arsenic residuals is largely dependent upon influent arsenic concentration, coagulant dose, and suspended solids content. Disposal by direct discharge to a surface water is not likely. In the example presented earlier in this section, the solids concentration is 1.0 percent, or 10,000 mg/L. This is a typical solids content for C/F sludges, and far exceeds the usual NPDES limit of around 30 mg/L.

The blowdown arsenic concentration in the above example is 21,000 μg/L. Most Industrial Protection Programs have TBLLs in the range of 50 μg/L to 1,000 μg/L (AWWARF, 1998). As a result, it is unlikely that indirect discharge to a POTW is an acceptable disposal method for C/F sludges.

Depending upon the arsenic concentration of C/F sludges, land application may be a suitable disposal method. Total arsenic should not exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with higher arsenic concentration may be land applied providing the total loading does not exceed 41 kg per hectare. In the previous example, the arsenic concentration is 2,060 mg/kg on a dry weight basis. As a result, this sludge may be land applied, but loading must not exceed 41 kg per hectare.

All enhanced coagulation sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that enhanced coagulation sludges will pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.3 Direct Filtration

Direct filtration is a modified C/F process that lacks the sedimentation unit process. Accordingly, direct filtration residuals are the result of filter backwash, and typically have lower TDS concentrations than a typical C/F process. This is due to the reduced coagulant dose. Sludge production is also affected by the suspended solids content of the raw water.

Backwash from direct filtration plants is high in water content and typically has a solids content of less than 1.0 percent. As a result, such sludges are usually discharged to a sanitary sewer or dewatered by one of the methods discussed earlier in this chapter. Discharge to sanitary sewers is generally only an option for treatment plants with an average daily flow of less than 10 million gallons per day.

The concentration of arsenic in the treatment process residuals is a function of the amount of coagulant used and the removal efficiency of the process. Using the design criteria specified in Chapter 3, consider a treatment plant that utilizes ferric chloride at a dose of 10 mg/L. Neglecting the impact of suspended solids, approximately 45 pounds of solids are produced for each million gallons of water treated. The filter backwash will have a solids concentration of 0.0005 percent and be produced at a rate of 20,000 gallons per million gallons of water treated. Assuming 40 μg/L of arsenic is removed from the influent during the treatment process, the result is 0.33 pounds of arsenic per million gallons of water treated. The backwash residuals arsenic concentration is approximately 2.0 mg/L, or 7,333 mg/kg on a dry weight basis.

Selection of Handling and Disposal Options

Direct filtration backwash is a high volume, low solids content waste stream. Gravity thickening may be used as a pretreatment for sludges prior to handling by other mechanical or non-mechanical dewatering processes. Filter presses are capable of attaining final sludge solids contents in the range of 35 to 50 percent, while scroll centrifuges may achieve final solids contents of 15 to 30 percent. Evaporation ponds and storage lagoons may be suitable for smaller treatment plants, but because they are land intensive may not be applicable for large water systems.

Disposal of direct filtration arsenic residuals is largely dependent upon influent arsenic concentration, coagulant dose and suspended solids content. Disposal by direct discharge to a surface water may be possible. In the example presented earlier in this section, the typical solids concentration is 0.0005 percent, or 5 mg/L, which meets the usual NPDES limit of around 30 mg/L. However, the above example neglects suspended solids and is based upon a backwash volume of 20,000 gallons. Higher influent suspended solids and/or smaller backwash volumes will impact the ability of a treatment facility to dispose of direct filtration residuals by direct discharge.

The backwash arsenic concentration in the above example is 2,000 μg/L. Most Industrial Protection Programs have TBLLs in the range of 50 μg/L to 1,000 μg/L (AWWARF, 1998). As a result, it is unlikely that indirect discharge to a POTW is an acceptable disposal method for C/F sludges.

Depending upon the arsenic concentration of direct filtration sludges, land application may be a suitable disposal method. Total arsenic should not exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with higher arsenic concentration may be land applied providing the total loading does not exceed 41 kg per hectare. In the previous example, the arsenic concentration is 7,333 mg/kg on a dry weight basis. As a result, this sludge may be land applied, but loading must not exceed 41 kg per hectare.

All sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that direct filtration sludges will pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.4 Coagulation Assisted Microfiltration

Coagulation assisted microfiltration is a modified C/F process wherein the flocculation/sedimentation and filtration unit processes are replaced by microfiltration. Residuals generated by this process consist of a filter backwash stream containing a dilute $Fe(OH)_3$ precipitate concentration. Based upon the design criteria in Chapter 3, assume a ferric chloride dose of 10 mg/L, a recovery rate of 95 percent, and that each pound of ferric chloride added produces 0.54 pounds of precipitate. Under these conditions, 52,600 gallons of backwash containing 45 pounds of precipitate at a concentration of 103 mg/L will be produced for every million gallons of water produced. Assuming an arsenic removal of 40 μg/L, the backwash will have an approximate arsenic concentration of 744 μg/L, or 7,333 mg/kg on a dry weight basis.

<u>Selection of Handling and Disposal Options</u>

Residuals from coagulation assisted microfiltration processes will be a very dilute slurry with a solids concentration of approximately 0.01% (103 mg/L). Gravity thickening may be used as a pre-treatment for other mechanical or non-mechanical dewatering options. Filter presses and centrifuges are appropriate methods of residuals handling. However, these methods are capital intensive and may not be appropriate for extremely large systems. Evaporation ponds and storage lagoons are also appropriate for coagulation assisted microfiltration residuals handling. Both require little oversight and maintenance, but are land intensive. As such, these may not be appropriate for large systems. A thorough comparison of handling options should be conducted to select the most cost effective method.

Direct discharge of coagulation assisted microfiltration residuals is not a likely disposal option. In the example above, the solids content of the residuals stream is 103 mg/L. This is greater than the typical NPDES limit of 30 mg/L. As a result, direct discharge is an unlikely disposal option for coagulation assisted microfiltration residuals.

The effluent arsenic concentration of 744 μg/L estimated for coagulation assisted microfiltration facilities is within the range of TBLLs (50 μg/L to 1,000 μg/L). Therefore, discharge to a sanitary sewer will be an acceptable method of disposal for coagulation assisted microfiltration residuals containing arsenic. The Industrial Protection Program should be checked prior to disposal to verify that the arsenic concentration of the residuals does not exceed the TBLL.

Land application may be a suitable disposal method for coagulation assisted microfiltration sludges. The predicted arsenic concentration of 7,333 mg/kg is much higher than the Clean Sludge criteria of 41 mg/kg, as well as the Land Application Ceiling limit of 75 mg/kg. Coagulation assisted microfiltration sludges may be land applied provided the total loading does not exceed 41 kg of arsenic per hectare.

All coagulation assisted microfiltration sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that sludges should pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.5 Lime Softening

The quantity of residuals produced at lime softening (LS) facilities is typically much greater than the quantity produced by C/F plants (AWWARF, 1998). The quantity of sludges produced is a function of water hardness. LS for carbonate hardness removal produces approximately twice the amount of solids per pound of hardness removed than non-carbonate hardness removal.

LS plants typically produce 1,000 to 8,000 pounds of solid per million gallons of water treated depending upon the hardness of the water (AWWARF, 1998). Arsenic concentrations, however, are generally lower than C/F sludges due to the increased volume of solids produced.

Using the design criteria in Chapter 3, assume a treatment plant generates 20,000 gallons of blowdown per million gallons of treated water. A solids concentration of 1.0 percent will result in 1,665 pounds of solids per million gallons of treated water. If 40 μg/L of arsenic are removed in the process, 0.33 pounds of arsenic are produced per million gallons treated. This equates to 2.0 mg/L in the blowdown, or 200 mg/kg on a dry weight basis.

Selection of Handling and Disposal Options

LS blowdown is slightly denser than C/F blowdown. Typical solids contents range from 1.0 to 4.0 percent, depending upon the raw water hardness. Gravity thickening may be used as a pretreatment for LS sludges prior to handling by other mechanical or non-mechanical dewatering processes. Filter presses are capable of attaining final LS sludge solids contents in the range of 40 to 70 percent, while scroll centrifuges may achieve final solids contents of 65 to 70 percent. Evaporation ponds and storage lagoons may be suitable for smaller treatment plants, but because they are land intensive may not be applicable for large water systems.

Direct discharge of LS sludges to a surface water is not a likely disposal alternative. LS sludges are typically 1.0 to 4.0 percent solids (10,000 to 40,000 mg/L) and will exceed the 30 mg/L limit in most NPDES permits. Further, discharge to a sanitary sewer is not appropriate. In the example presented earlier in this section, the arsenic concentration of the blowdown is approximately 2.0 mg/L which exceeds the typical TBLL (50 to 1,000 μg/L) of an Industrial Protection Program.

Land application of LS treatment sludges is one possible disposal alternative. Based upon the above example, it appears that LS sludges will exceed the Clean Sludge criteria for arsenic of 41 mg/kg. Application would therefore be limited to 41 kg of arsenic per hectare.

LS sludges will require dewatering prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that LS sludges will pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.6 Enhanced Lime Softening

Enhanced LS is a modified LS process wherein lime dosage is increased. Residuals produced are similar to those of a typical LS treatment process. The quantity of sludges produced is a function of water hardness. Enhanced LS for carbonate hardness removal produces approximately twice the amount of solids per pound of hardness removed than non-carbonate hardness removal.

Enhanced LS plants typically produce 1,000 to 8,000 pounds of solid per million gallons of water treated depending upon the hardness of the water (AWWARF, 1998). Arsenic concentrations, however, are generally lower than C/F sludges due to the increased volume of solids produced.

Using the design criteria in Chapter 3, assume a treatment plant generates 20,000 gallons of blowdown per million gallons of treated water. A solids concentration of 1.0 percent will result in 1,665 pounds of solids per million gallons of treated water. If 40 μg/L of arsenic are removed in the process, 0.33 pounds of arsenic are produced per million gallons treated. This equates to 2.0 mg/L in the blowdown, or 200 mg/kg on a dry weight basis.

Selection of Handling and Disposal Options

Typical enhanced LS blowdown solids contents range from 1.0 to 4.0 percent, depending upon the raw water hardness. Gravity thickening may be used as a pretreatment for sludges prior to handling by other mechanical or non-mechanical dewatering processes. Filter presses are capable of attaining final sludge solids contents in the range of 40 to 70 percent, while scroll centrifuges may achieve final solids contents of 65 to 70 percent. Evaporation ponds and storage lagoons may be suitable for smaller treatment plants, but because they are land intensive may not be applicable for large water systems.

Direct discharge of enhanced LS sludges to a surface water is not a likely disposal alternative. LS sludges are typically 1.0 to 4.0 percent solids (10,000 to 40,000 mg/L) and will exceed the 30 mg/L limit in most NPDES permits. Further, discharge to a sanitary sewer is not appropriate. In the example presented earlier in this section, the arsenic concentration of the blowdown is approximately 2.0 mg/L which exceeds the typical TBLL (50 to 1,000 μg/L) of an Industrial Protection Program.

Land application of treatment sludges is one possible disposal alternative. Based upon the above example, it appears that enhanced LS sludges will exceed the Clean Sludge criteria for arsenic of 41 mg/kg. Application would therefore be limited to 41 kg of arsenic per hectare. As with C/F sludges, once this loading is achieved it may never be used again for disposal purposes.

Sludges will require dewatering prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Tests conducted by the University of Colorado indicate that enhanced LS sludges will pass the TCLP test (AWWARF, 1998). Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.7 Ion Exchange

Ion exchange (IX) uses a synthetic resin in a chloride form for arsenic removal. With time, the efficiency of the resin is reduced as exchange sites are depleted. The IX resin can be regenerated using a NaCl solution. The regenerant is added at a rate of approximately 2 equivalents per equivalent of resin, i.e., 15 pounds of salt per cubic foot of resin. Regeneration requires approximately 2.8 BV of brine and 1.2 BV displacement rinse. Therefore, 4 to 5 BV of waste are produced per regeneration cycle. (AWWARF, 1998).

Arsenic removal is severely affected by sulfate concentration of the source water. Assuming the sulfate concentration limits the run length to 1,000 BV before regeneration is required, at an influent arsenic concentration of 40 μg/L, approximately 1,132 mg of arsenic can be removed per cubic foot of resin. Regeneration will produce a brine waste solution with an arsenic concentration of around 10 mg/L and a brine concentration of about 20,000 mg/L (AWWARF, 1998). Arsenic can be precipitated from the brine stream using ferric chloride. The resulting precipitate has an arsenic concentration of approximately 14,250 mg/kg on a dry weight basis.

<u>Selection of Handling and Disposal Options</u>

Evaporation ponds, drying beds, and storage lagoons are often used for brine waste stream handling. In regions with favorable climatic conditions, evaporation ponds may be the preferred handling option. Evaporation ponds are land intensive and require shallow basins with large surface areas. As such, ponds may not be suitable to areas where available land is scarce or acquisition costs cannot be justified for construction of a pond. If construction is feasible, operations and maintenance are minimal and make this a primary candidate for IX residuals handling.

Direct discharge of IX residuals to a receiving surface water is an unlikely disposal alternative.

Assume the brine waste stream generated has an arsenic concentration of 10 mg/L. Now, assume the state regulatory agency has set an ambient arsenic water quality standard at 20.5 μg/L for fish consumption. IX will produce approximately 4,000 gallons of brine waste for each million gallons of treated water. Based upon this scenario, a dilution factor of around 487 is needed for the receiving water to assimilate the arsenic from the brine stream. In other words, a one million gallon per day treatment plant would require a dilution flow of 3.1 cfs to assimilate the 4,000 gallons of waste brine generated (AWWARF, 1998). Furthermore, to be discharged directly to a receiving water the brine stream must pass the whole effluent toxicity (WET) test. It is unlikely that the waste brine would pass the WET test due to the high arsenic concentration and salt content (AWWARF, 1998). Therefore, even if a stream had sufficient flow for dilution of the IX residuals, it is unlikely discharge would be allowed.

The 10 mg/L concentration of the brine waste stream far exceeds the typical TBLL of 50 to 1,000 μg/L. Therefore, it is unlikely that IX residuals could be discharged to a sanitary sewer for treatment at a POTW. If the solids content of the waste brine were increased to 5 percent by ferric chloride precipitation, the arsenic concentration is increased to 712 mg/L (AWWARF, 1998). Again, this exceeds typical TBLL values and would not likely be a candidate for discharge to a sanitary sewer.

Land application of IX brine streams is an unlikely disposal alternative. The high salinity of the residual stream would result in a significant increase in the salt content of the receiving soil. This salinity build-up would make plant growth virtually impossible. Therefore, an alternative technology should be selected for disposal or IX residuals. If land application is considered appropriate, arsenic loading may not exceed 41 kg per hectare. Once land has been used for arsenic residuals disposal it may never be used for that purpose again.

IX residuals may be disposed at a sanitary or hazardous waste landfill, but will require extensive dewatering. The solids content can be increased to approximately 5 percent by ferric chloride precipitation. The precipitate will require dewatering by one of the methods presented in this chapter prior to disposal. The precipitate will likely pass the TCLP test and be a candidate for disposal at a sanitary landfill (AWWARF, 1998). However, should the residuals fail the TCLP test, disposal at a hazardous waste landfill will be required.

4.4.8 Activated Alumina

Activated alumina (AA) will produce a regenerant waste solution with a pH of approximately 12 and high in dissolved solids, aluminum, and arsenic (AWWARF, 1998). Arsenic can be removed from the solution by aluminum hydroxide precipitation and reducing the regenerant pH. Regeneration of AA is accomplished using 15 to 25 bed volumes (BV) of 2N NaOH, 7 BV of rinse, and 15 BV of 2N H_2SO_4 for neutralization. Thus, the total volume of waste produced in approximately 42 BV per regeneration cycle.

Assume that an AA system treats 10,000 BV of water before it reaches exhaustion, and that approximately 40 μg/L of arsenic are being removed from the source water. The resulting waste solution will have an approximate arsenic concentration of 9.52 mg/L. Ferric chloride precipitation of the residual arsenic would produce a residual with an arsenic concentration of around 14,250 mg/kg on a dry weight basis (AWWARF, 1998).

Selection of Handling and Disposal Options

Evaporation ponds, drying beds and storage lagoons are often used for brine waste stream handling. In regions with favorable climatic conditions, evaporation ponds may be the preferred handling option. Evaporation ponds are land intensive and require shallow basins with large surface areas. As such, ponds may not be suitable to areas where available land is scarce or acquisition costs cannot be justified for construction of a pond. If construction is feasible, operations and maintenance are minimal and make this a primary candidate for AA residuals handling.

Direct discharge of AA residuals to a receiving surface water is an unlikely disposal alternative. Assume the brine waste stream generated has an arsenic concentration of 9.52 mg/L. Now, assume the state regulatory agency has set an ambient arsenic water quality standard at 20.5 μg/L for fish consumption. AA will produce approximately 4,200 gallons of brine waste for each million gallons of treated water. Based upon this scenario, a dilution factor of around 463 is needed for the receiving water to assimilate the arsenic from the brine stream. In other words, a one million gallon per day treatment plant would require a dilution flow of 3.0 cfs to assimilate the 4,200 gallons of waste brine generated (AWWARF, 1998). Furthermore, to be discharged directly to a receiving water the brine stream must pass the WET test. It is unlikely that the waste brine would pass the WET test due to the high arsenic concentration (AWWARF, 1998). Therefore, even if a stream had sufficient flow for dilution of the AA residuals, it is unlikely discharge would be allowed.

The 9.52 mg/L concentration of the brine waste stream far exceeds the typical TBLL of 50 to 1,000 μg/L. Therefore, it is unlikely that AA residuals could be discharged to a sanitary sewer for treatment at a POTW. If the solids content of the waste brine were increased to 5 percent by ferric chloride precipitation, the arsenic concentration is increased to 712 mg/L (AWWARF, 1998). Again, this exceeds typical TBLL values and would not likely be a candidate for discharge to a sanitary sewer.

Land application of AA residuals may be inappropriate. The aluminum content of the waste solution may combine with phosphorus in the soil matrix and prevent uptake of the phosphorus by vegetation. If land application is considered appropriate, arsenic loading may not exceed 41 kg per hectare. Once land has been used for arsenic residuals disposal it may never be used for that purpose again.

AA residuals may be disposed at a sanitary or hazardous waste landfill, but will require extensive dewatering. The solids content can be increased to approximately 5 percent by aluminum hydroxide precipitation. The precipitate will require dewatering by one of the methods presented in this chapter prior to disposal. Tests conducted at the University of Colorado indicate the precipitate will likely pass the TCLP test and be a candidate for disposal at a sanitary landfill (AWWARF, 1998). However, should the residuals fail the TCLP test, disposal at a hazardous waste landfill will be required.

4.4.9 Microfiltration

Microfiltration (MF) pore size is too large to remove dissolved or colloidal arsenic, but is capable of removing particulate arsenic. Considering particulate arsenic concentrations in groundwater are generally less than 10 percent, and surface water particulate arsenic concentrations vary from 0 to 70 percent, MF alone may not be a viable removal technology. Recovery rates are much higher than RO and NF processes, with typical recovery approaching 99. The reject stream contains elevated levels of arsenic and other contaminants that are removed from the source water by the MF membranes.

As an example, assume that a MF system has a recovery rate of 99 percent, arsenic rejection by the membrane is 20 percent, and the feed arsenic concentration is 50 μg/L. The reject will have an arsenic concentration of approximately 980 μg/L. This system will also produce 10,100 gallons of reject for every million gallons of treated water (AWWARF, 1998). Due to the large volume of reject water, it may not be feasible to implement ferric chloride precipitation to remove the arsenic from the reject stream.

Selection of Handling and Disposal Options

MF generates high TDS waste streams. Evaporation ponds and drying beds have been used successfully for similar waste streams in the past. Both are land intensive handling options and may not be suitable for large MF facilities, or for water systems where land acquisition costs make construction infeasible. If determined to be appropriate, evaporation ponds and drying beds require little oversight and maintenance is minimal. Depending upon TDS concentrations in the brine stream, other mechanical and non-mechanical dewatering devices may also be applicable.

Direct discharge of MF brine streams to a surface water is not a likely disposal alternative. MF brine streams typically contain greater than 100 mg/L total solids, and will exceed the 30 mg/L limit in most NPDES permits. Discharge to a sanitary sewer may be appropriate. In the example presented earlier in this section, the arsenic concentration of the blowdown is approximately 980 μg/L which is within the upper bounds of the typical TBLL (50 to 1,000 μg/L) of an Industrial Protection Program.

Land application may be a suitable disposal method for MF brine streams. MF brine streams may be land applied providing the total loading does not exceed 41 kg of arsenic per hectare.

All MF sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.10 Ultrafiltration

Ultrafiltration (UF) membranes are primarily used for removal of colloidal and particulate contaminants. Considering arsenic found in groundwater is typically less than 10 percent particulate, and surface waters contain 0 to 70 percent particulate arsenic, UF may not be a viable arsenic removal technology. The recovery rate for UF is generally higher than that of RO or NF, with typical recovery approaching 95%. The reject stream contains elevated levels of arsenic and other contaminants that are removed from the source water by the UF membranes.

As an example, assume that a membrane system has a recovery rate of 95 percent, arsenic rejection by the membrane is 70 percent, and the feed arsenic concentration is 50 μg/L. The reject will have an arsenic concentration of approximately 660 μg/L. This system will also produce 52,600 gallons of reject for every million gallons of treated water (AWWARF, 1998). Due to the large volume of reject water, it may not be feasible to implement ferric chloride precipitation to remove the arsenic from the reject stream.

Selection of Handling and Disposal Options

UF generates high volume, high TDS waste streams. Evaporation ponds and drying beds have been used successfully for similar waste streams in the past. Both are land intensive handling options and may not be suitable for large UF facilities, or for water systems where land acquisition costs make construction infeasible. If determined to be appropriate, evaporation ponds and drying beds require little oversight and maintenance is minimal. Depending upon TDS concentrations in the brine stream, other mechanical and non-mechanical dewatering devices may also be applicable.

Direct discharge of UF brine streams to a surface water may be likely, providing TDS concentrations do not exceed the 30 mg/L limit in most NPDES permits. Discharge to a sanitary sewer may also be an appropriate method of disposal. In the example presented earlier in this section, the arsenic concentration of the brine stream is approximately 660 μg/L which is in the upper bounds of the typical TBLL (50 to 1,000 μg/L) of an Industrial Protection Program.

Land application is a possible candidate for disposal of UF arsenic containing residuals. The reject water would have to be applied at a rate matching the evapotransportation requirements of the cover crop grown and is limited to 41 kg of arsenic per hectare.

All UF sludges must be dewatered prior to landfill disposal. If the residuals pass the TCLP test they may be disposed in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. Hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test.

4.4.11 Nanofiltration

NF membranes are primarily used for softening, removing the larger, divalent ions associated with hardness (AWWARF, 1998). The recovery rate, or percent of water passing through the membrane rather than being rejected by the membrane, is dependent upon the source water quality. Typical recovery rates approach 85% percent. The reject stream contains elevated levels of arsenic and other contaminants that are removed from the source water by the NF membranes.

Based upon the design criteria in Chapter 3, assume that a NF system has a recovery rate of 85 percent. Arsenic rejection by the membrane is 85 percent and the feed arsenic concentration is 50 μg/L. The reject will have an arsenic concentration of approximately 225 μg/L. This system will also produce 176,000 gallons of reject for every million gallons of treated water (AWWARF, 1998). Due to the large volume of reject water, it may not be feasible to implement ferric chloride precipitation to remove the arsenic from the reject stream.

<u>Selection of Handling and Disposal Options</u>

NF generates high volume, high TDS waste streams. Evaporation ponds and drying beds have been used successfully for similar waste streams in the past. Both are land intensive handling options and may not be suitable for large facilities, or for water systems where land acquisition costs make construction infeasible. If determined to be appropriate, evaporation ponds and drying beds require little oversight and maintenance is minimal. Because of the high volume of NF waste streams, other mechanical and non-mechanical dewatering devices are generally not applicable.

Direct discharge of NF residuals to a receiving surface water is one possible disposal alternative. For example, assume the brine waste stream generated has an arsenic concentration of 225 μg/L, and the state regulatory agency has set an ambient arsenic water quality standard at 20.5 μg/L for fish consumption. NF will produce approximately 176,000 gallons of brine waste for each million gallons of treated water. Based upon this scenario, a dilution factor of around 12 is needed for the receiving water to assimilate the arsenic from the brine stream. In other words, a one million gallon per day treatment plant would require a dilution flow of 3.1 cfs to assimilate the 176,000 gallons of waste brine generated. Furthermore, to be discharged directly to a receiving water the brine stream must pass the WET test. Due to the low arsenic concentration in the NF process effluent, it is unlikely that the waste brine would fail the WET test (AWWARF, 1998). Therefore, direct discharge to a receiving water may be a possible disposal alternative for NF treatment facilities.

The effluent arsenic concentration of 225 μg/L estimated for the NF process is well within the bounds of the range of TBLLs (50 μg/L to 1,000 μg/L). Therefore, in most cases discharge to a sanitary sewer will be an acceptable method of disposal for NF residuals containing arsenic. The appropriate Industrial Protection Program should be checked prior to disposal to verify that the arsenic concentration of the residuals does not exceed the TBLL.

NF treatment produces high volume liquid waste streams. As a result, land application is an unlikely candidate for disposal of arsenic containing residuals. The reject water would have to be applied at a rate matching the evapotransportation requirements of the cover crop grown and is limited to 41 kg of arsenic per hectare. Therefore, land application is deemed inappropriate for NF residuals disposal (AWWARF, 1998).

The liquid waste produced by NF treatment can not be sent to a sanitary or hazardous waste landfill. The high free liquids content makes dewatering uneconomical. The arsenic concentration is well below TCLP limits and makes hazardous waste landfill disposal unnecessary. Landfill disposal of NF residuals is inappropriate (AWWARF, 1998).

4.4.12 Reverse Osmosis

RO membranes will remove much smaller ions typically associated with TDS (AWWARF, 1998). The recovery rate, or percent of water passing through the membrane rather than being rejected by the membrane, is dependent upon the source water quality. Typical recovery rates vary from 30 to 85 percent. The reject stream contains elevated levels of arsenic and other contaminants that are removed from the source water by the RO membranes.

Based upon the design criteria in Chapter 3, assume that a RO membrane system has a recovery rate of 85 percent. Arsenic rejection by the membrane is 95 percent and the feed arsenic concentration is 50 μg/L. The reject will have an arsenic concentration of approximately 265 μg/L. This system will also produce 176,000 gallons of reject for every million gallons of treated water (AWWARF, 1998). Due to the large volume of reject water, it may not be feasible to implement ferric chloride precipitation to remove the arsenic from the reject stream.

Selection of Handling and Disposal Options

RO generates high volume, high TDS waste streams. Evaporation ponds and drying beds have been used successfully for similar waste streams in the past. Both are land intensive handling options and may not be suitable for large RO facilities, or for water systems where land acquisition costs make construction infeasible. If determined to be appropriate, evaporation ponds and drying beds require little oversight and maintenance is minimal. Because of the high volume of RO waste streams, other mechanical and non-mechanical dewatering devices are generally not applicable.

Direct discharge of RO residuals to a receiving surface water is one possible disposal alternative. Assume the brine waste stream generated has an arsenic concentration of 265 μg/L. Now, assume the state regulatory agency has set an ambient arsenic water quality standard at 20.5 μg/L for fish consumption. RO will produce approximately 176,000 gallons of brine waste for each million gallons of treated water. Based upon this scenario, a dilution factor of around 12 is needed for the receiving water to assimilate the arsenic from the brine stream. In other words, a one million gallon per day treatment plant would require a dilution flow of 3 cfs to assimilate the 176,000 gallons of waste brine generated. Furthermore, to be discharged directly to a receiving water the brine stream must pass the WET test. Due to the low arsenic concentration in the RO process effluent, it is unlikely that the waste brine would fail the WET test (AWWARF, 1998). Therefore, direct discharge to a receiving water may be a possible disposal alternative for RO treatment facilities.

The effluent arsenic concentration of 265 μg/L estimated for the RO process is well within the bounds of the range of TBLLs (50 μg/L to 1,000 μg/L). Therefore, in most cases discharge to a sanitary sewer will be an acceptable method of disposal for RO residuals containing arsenic. The appropriate Industrial Protection Program should be checked prior to disposal to verify that the arsenic concentration of the residuals does not exceed the TBLL.

RO treatment produces high volume liquid waste streams. As a result, land application is an unlikely candidate for disposal of arsenic containing residuals. The reject water would have to be applied at a rate matching the evapotransportation requirements of the cover crop grown and is limited to 41 kg of arsenic per hectare. Therefore, land application is deemed inappropriate for RO residuals disposal (AWWARF, 1998).

The liquid waste produced by RO treatment can not be sent to a sanitary or hazardous waste landfill. The high free liquids content makes dewatering uneconomical. The arsenic concentration is well below TCLP limits which makes hazardous waste landfill disposal unnecessary. Landfill disposal of RO residuals is inappropriate (AWWARF, 1998).

4.5 SUMMARY

Characteristics of arsenic containing residuals were presented for the following treatment technologies evaluated within this document: coagulation/filtration; direct filtration; coagulation assisted microfiltration; enhanced coagulation; lime softening; enhanced lime softening; ion exchange; activated alumina; reverse osmosis; nanofiltration; ultrafiltration; and microfiltration. Table 4-1 summarizes the residuals characteristics for each of these treatment processes.

This chapter also evaluated handling and disposal options for water treatment residuals containing arsenic. Specifically, the following handling options were presented: gravity thickening; mechanical dewatering, including filter presses and centrifuges; evaporation ponds and drying beds; and storage lagoons. Moreover, the following disposal options were evaluated: direct discharge to surface water; indirect discharge - discharge to sanitary sewer for treatment at POTW; land application; sanitary landfill disposal; and hazardous waste landfill disposal. Table 4-2 summarizes the applicability of each of the handling and disposal options presented in this chapter to the treatment technologies presented throughout this document.

TABLE 4-1

Summary of Residuals Characteristics*

Treatment Technology	Waste Type	Waste per Million Gallons Water Produced	Solids Production per Million Gallons Water Produced	Arsenic Concentration
Coagulation/Filtration	Blowdown	1,400 gallons	110 pounds	28.0 mg/L 2,930 mg/kg
	Backwash	10,000 gallons	10,000 mg/L	2,930 mg/kg
Enhanced Coagulation	Blowdown	1,900 gallons	160 pounds	21.0 mg/L 2,060 mg/kg
	Backwash	10,000 gallons	10,000 mg/L	2,060 mg/kg
Direct Filtration	Backwash	20,000 gallons	45 pounds	2.0 mg/L 7,333 mg/kg
Coagulation Assisted Microfiltration	Brine	52,600 gallons	45 pounds 103 mg/L	744 μg/L 7,333 mg/kg
Lime Softening	Blowdown	20,000 gallons	1,665 pounds 1.0%	2.0 mg/L 200 mg/kg
Enhanced Lime Softening	Blowdown	20,000 gallons	1,665 pounds 1.0%	2.0 mg/L 200 mg/kg
Ion Exchange	Brine	--	--	10.0 mg/L 14,250 mg/kg
Activated Alumina	Brine	--	--	9.5 mg/L 14,250 mg/kg
Microfiltration	Brine	10,100 gallons	--	980 μg/L
Ultrafiltration	Brine	52,600 gallons	--	660 μg/L
Nanofiltration	Brine	176,000 gallons	--	225 μg/L
Reverse Osmosis	Brine	176,000 gallons	--	260 μg/L

* Based upon design criteria established in Chapter 3 of this document. For general residuals characteristics, see the appropriate section of this chapter.

TABLE 4-2

Summary of Arsenic Residuals Handling and Disposal Options

Treatment Technology	Handling Options				Disposal Options				
	GT	**MD**	**EP**	**SL**	**DD**	**ID**	**LA**	**SD**	**HD**
Coagulation/Filtration	!	!	"	"	×	■	"	"	□
Enhanced Coagulation	!	!	"	"	×	■	"	"	□
Direct Filtration	!	!	"	"	■	■	"	"	□
Coagulation Assisted Microfiltration	!	!	"	"	×	"	"	"	□
Lime Softening	!	!	"	"	×	■	"	"	□
Enhanced Lime Softening	!	!	"	"	×	■	"	"	□
Ion Exchange	■	■	"	"	×	■	"	"	□
Activated Alumina	■	■	"	"	×	■	"	"	□
Microfiltration	!	■	"	"	■	"	"	"	□
Ultrafiltration	!	■	"	"	"	"	■	■	□
Nanofiltration	■	■	"	"	"	"	×	■	×
Reverse Osmosis	■	■	"	"	"	"	×	■	×

GT = gravity thickening, MD = mechanical dewatering, EP = evaporation ponds and drying beds, SL = storage lagoons,
DD = direct discharge, ID = indirect discharge, LA = land application, SD = sanitary landfill disposal, HD = hazardous landfill disposal

! - Yes
" - Yes, with limitations. Disposal may depend upon NPDES limits, TBLLs or presence of free liquids.
■ - Not likely
□ - Only as a last resort
× - Not a feasible option

5.0 POINT-OF-ENTRY/POINT-OF-USE TREATMENT OPTIONS

5.1 INTRODUCTION

Centralized treatment is not always a feasible treatment option, for example, in areas where each home has a private well or centralized treatment is cost prohibitive. In these instances, point-of-entry (POE) and point-of-use (POU) treatment options may be acceptable treatment alternatives. POE and POU systems offer ease of installation, simplify operation and maintenance, and generally have lower capital costs (Fox, 1989). These systems may also reduce engineering, legal and other fees typically associated with centralized treatment options. Use of POE and POU systems does not reduce the need for a well-maintained water distribution system. In fact, increased monitoring may be necessary to ensure that the treatment units are operating properly.

Home water treatment can consist of either whole-house or single faucet treatment. Whole-house, or POE treatment is necessary when exposure to the contaminant by modes other than consumption is a concern. POU treatment is preferred when treated water is needed only for drinking and cooking purposes. POU treatment usually involves single-tap treatment.

Section 1412(b)(4)(E) of the 1996 Safe Drinking Water Act (SDWA) Amendments requires the EPA to issue a list of technologies that achieve compliance with MCLs established under the act. This list must contain technologies for each NPDWR and for each of the small public water systems categories listed below:

Population of more than 50, but less than 500;
Population of more than 500, but less than 3,300; and
Population of more than 3,300, but less than 10,000.

The SDWA identifies POE and POU treatment units as potentially affordable technologies, but stipulates that POE and POU treatment systems "shall be owned, controlled and maintained by the public water system, or by a person under contract with the public water system to ensure proper operation and compliance with the maximum contaminant level or treatment technique and equipped with mechanical warnings to ensure that customers are automatically notified of operational problems."

Research has shown that POE and POU devices can be effective means of removing arsenic from potable water (Fox and Sorg, 1987; Fox, 1989). Water systems with high influent arsenic concentrations, i.e., greater than 1 mg/L, may have difficulty meeting MCLs much lower than the 10 to 20 μg/L level. As a result, influent arsenic concentration and other source water characteristics must be considered when evaluating POE and POU devices for arsenic removal. To be effective these devices should work with minimal attention and be relatively inexpensive for the user. Reverse osmosis, activated alumina, and ion exchange are three treatment techniques that have been evaluated and shown to be effective. This chapter looks at the removals achieved by each of these three treatment techniques, and presents total costs for each treatment option.

5.2 VARIABLES AFFECTING REMOVAL EFFICIENCY

5.2.1 Speciation

Arsenic speciation is critical to the removal efficiency of every technology presented in this document. As previously discussed, inorganic arsenic occurs in two primary forms, arsenite (AsIII) and arsenate (AsV). Arsenite is removed less efficiently because it predominantly occurs in the uncharged (H_3AsO_3) state in source waters with a pH of less than 9.0. The dominant arsenate forms are anionic species, $H_2AsO_4^{2-}$ and $HAsO_4^{-}$.

Arsenic removal is dependent upon water chemistry and arsenic speciation. As a result, identification of the ionic form of arsenic is necessary for selection and design of a removal process. All technologies discussed in this document remove arsenate more effectively than arsenite. Therefore, if arsenite is the predominant species present, oxidation to arsenate may be required to achieve the desired removal.

5.2.2 pH

As previously stated, pH plays a significant role in determining the removal efficiency of a particular technology. Most processes are relatively unaffected by pH in the range of 6.5 to 9.0. However, activated alumina studies have shown the optimum pH for arsenic removal to be between 5.5 and 6.0, and reverse osmosis processes may require pH adjustment to prevent precipitation of salts on the membrane surface.

Source water pH can be adjusted by addition of dilute acid. The addition of acid, however, may need pH adjustment in treated water to contain corrosion. This can lead to increase in capital, as well as operations and maintenance costs. In POE and POU systems there is an added concern of monitoring to insure that pH levels in treated water are safe.

5.2.3 Co-occurrence

Co-occurrence of inorganic contaminants, such as sulfate and silica, as well as suspended solids, can cause interference with arsenic removal. Sulfate is preferentially adsorbed relative to arsenic by ion exchange processes. This preference can result in another phenomenon known as peaking, which occurs when arsenic is displaced on the resins by the sulfate causing effluent concentrations in excess of the influent levels.

A slight decrease in activated alumina performance has been seen in waters with high sulfate concentration, however, the effect is not as great as in ion exchange processes. At higher treatment pH levels silica may also be preferred relative to arsenic.

5.3 POE/POU DEVICE CASE STUDIES

Several field studies conducted to evaluate the effectiveness of POE and POU treatment units for arsenic removal indicate that POE and POU systems can be effective alternatives to centralized treatment options. These studies conducted evaluated reverse osmosis (RO), activated alumina (AA), and ion exchange (IX) processes.

The following sections present the results of two of these studies. Table 5-1 summarizes source water quality and influent arsenic concentrations. Table 5-2 summarizes the arsenic removals achieved by each of the technologies evaluated.

TABLE 5-1

Source Water Summary - Point-of-Use Case Studies

Source Water Characteristic	Concentration[1]
Alkalinity	56 - 206
Arsenic	<0.005 - 1.16
Calcium	8.9 - 22
Chloride	<10
Fluoride	0.6 - 5.2
Iron	<0.1 - 2.5
Magnesium	5.3 - 10.6
Manganese	<0.6
pH	7.4 - 8.3
Silica	NA
Sodium	4.4 - 62
Sulfate	<15
Total Dissolved Solids	<1,500
Total Hardness	109 - 547
Turbidity	0.24 - 0.48

[1] All concentrations are given in mg/L, except turbidity (NTU) and pH units. Note measurements for each parameter were not taken at each test site.

TABLE 5-2

Observed Arsenic Removal by Technology for POE and POU Units

Treatment Option	Arsenic Removal (%)	Quantity of Water Treated When MCL Exceeded* (gal)	Time On-Line Until MCL Exceeded* (days)
Reverse Osmosis (low pressure)	50-86	50-332	35-225
Reverse Osmosis (high pressure)	50-80	684	350
Activated Alumina	NA	0-15,427	0-1,226
Ion Exchange	NA	0-16,254	0-1,471

* Based upon EPA DWRD study conducted in Fairbanks, Alaska and Eugene, Oregon only.

5.3.1 Case Study 1: Fairbanks, Alaska and Eugene, Oregon

The EPA Drinking Water Research Division (DWRD) conducted field POU studies in Fairbanks, Alaska and Eugene, Oregon (Fox and Sorg, 1987; Fox, 1989). Pilot systems were installed in two homes in each community, and each system consisted of an activated alumina bed, ion exchange bed and reverse osmosis system. Influent arsenic concentrations ranged from 0.05 to 1.16 mg/L and was believed to be naturally occurring.

The RO systems at each of the four locations performed well upon start-up, achieving 60 to 80% removal of arsenic. Over time, however, the arsenic removal efficiency decreased to 50% or less. At the initial removal efficiencies, effluent arsenic levels met the current MCL of 50 ppb, but over time failed to sufficiently reduce the levels to below the MCL.

Low-pressure RO units (40 - 60 psig) consistently achieved greater than 50% removal, but with high influent arsenic concentrations much higher efficiency is necessary to achieve the MCL. The high-pressure units (196 psig) operated for 350 days and produced 684 gallons of treated water which met the MCL. The DWRD discontinued the study after 350 days, but the homeowner continued to utilize the high-pressure RO unit.

The IX beds evaluated were 1 cubic foot in size and were filled with a strong base anion exchange resin (Dowex-SBR). The IX beds effectively reduced arsenic levels to below the MCL, but required pre-treatment to ensure effective removal. This involved regeneration and chemical treatment of the resin to the chlorine form.

The AA beds were identical to the IX beds with the exception that they were filled with granular activated alumina (Alcoa-F1). The AA beds used effectively reduced arsenic levels, however required pre-treatment to reduce the pH to 5.5 - 6.0. Regeneration involved passing a sodium hydroxide solution through the tank, rinsing with clean water and then treating with dilute sulfuric acid. Improperly treated alumina performed poorly initially (30 to 40 percent removal), and performance significantly deteriorated over time (5 to 20 percent removal). Proper pre-treatment, however, allowed for efficient operation periods of nearly one year or greater.

5.3.2 Case Study 2: San Ysidro, New Mexico

A field study was also conducted in San Ysidro, NM to evaluate the effectiveness of POU RO units. This work is documented in several sources (Thomson and O'Grady, 1998; Fox, 1989; Fox and Sorg, 1987; and Clifford and Lin, 1985). San Ysidro source water is from an infiltration gallery under the local river banks and contains 5.2 mg/L fluoride and 0.23 mg/L arsenic. The water is also high in other inorganic contaminants, including iron (2.5 mg/L), manganese (0.6 mg/L) and total dissolved solids (1,500 mg/L). San Ysidro is a small community with limited financial resources, and central treatment was not a viable treatment option. San Ysidro applied to the DWRD for a cooperative agreement to evaluate POU RO treatment for the entire village. The project was funded in August 1995, and seventy-three units were initially purchased and installed in homes, restaurants, gas stations and municipal buildings.

Arsenic concentrations in the source water were consistently reduced from 0.068 mg/L to 0.02 mg/L to less than the detectable limit (0.005 mg/L). Other contaminants were also effectively removed, including manganese (80 percent), iron (85 percent), and TDS (95 percent). Based upon the manufacturer's literature, it appeared that the units were operating at an approximate recovery rate of 25%, i.e., for every 100 gallons of influent, 25 gallons of treated water are produced.

The water supply for San Ysidro is chlorinated at the wellhead. As a result, a carbon pre-filter was installed with each unit to remove residual chlorine and particulates to prevent membrane fouling. A carbon post-filter was installed for polishing treated water. Since the conclusion of the study, the village has assumed ownership of the units and is now responsible for their maintenance.

5.4 REVERSE OSMOSIS

Reverse osmosis (RO) is a separation process that utilizes a membrane system to reject compounds based upon molecular properties. Water molecules pass through the membrane, but most contaminants including arsenic are rejected by the membranes. While a portion of the feed water passes through the membrane, the rest is discharged with the rejected contaminants in a concentrated stream. Membrane performance can be adversely affected by the presence of turbidity, iron, manganese, scale-producing compounds and other contaminants. A detailed discussion of the RO removal mechanism is presented in Chapter 2.

POU RO systems can be operated at both high (approximately 200 psig) and low (40 - 60 psig) pressures. High pressure RO devices typically operate at a product-to-reject water ratio of 1 to 3 (Fox, 1989), and require a booster pump to achieve the desired operating pressure. Low pressure RO devices are less efficient and operate with a product-to-reject water ratio of about 1 to 10 (Fox, 1989). This can be a significant deterrent to RO treatment in dry regions or regions with frequent water shortages.

Manufacturer and laboratory data suggest greater than 95% removal of arsenate by RO systems, and slightly less (75%) removal of arsenite. Field studies indicate that greater than 50% removal is possible, but data are inconclusive much beyond those levels. Accordingly, water systems with high influent arsenic concentrations, i.e., greater than 1.0 mg/L, may consider other POE and POU treatment options.

5.4.1 Cost Estimates

The EPA document <u>Cost Evaluation of Small System Compliance Options - Point-of Use and Point-of-Entry Treatment Units</u> (Cadmus Group, 1998) was used to estimate POE and POU RO treatment costs. Costs are presented in Figure 5-1, and are based upon the following assumptions:

- Average household - 3 individuals, 1 gallon each per day, 1,095 gallons per year;
- Annual treatment - 1,095 gallons (POU), 109,500 gallons (POE);
- Minimally skilled labor - $14.50 per hour (population less than 3,300 individuals);
- Skilled labor - $28.00 per hour (population greater than 3,300 individuals);
- Life of unit - 5 years (POU), 10 years (POE);
- Duration of cost study - 10 years (therefore, two POU devices per household);
- Cost of water meter and automatic shut-off valve included;
- No shipping and handling costs required;
- Volume discount schedule - retail for single unit, 10% discount for 10 or more units, 15% discount on more than 100 units;
- Installation time - 1 hour unskilled labor (POU), 3 hours, skilled labor (POE);
- O&M costs include maintenance, replacement of pre-filters and membrane cartridges, laboratory sampling and analysis, and administrative costs.

5.5 ION EXCHANGE

Ion exchange (IX) has been used effectively for the removal of arsenic. There are two types of IX systems, anionic and cationic. IX for arsenic removal is typically anionic since arsenic present in natural waters is generally anionic. IX is a process wherein an ion in the solid phase, e.g., a synthetic resin, is exchanged for an ion in the source water. To accomplish this exchange, the source water is passed through the IX bed in either a downflow or upflow mode until the resin is exhausted. Exhaustion occurs when unacceptable levels of the contaminant are observed in the bed effluent. In POE and POU systems water often sits in the beds for extended periods of time because the required flow is not constant. It is possible this leads to better than average removal of arsenic. Further research is needed to explain this phenomenon. The IX removal mechanism is discussed in greater detail in Chapter 2.

5.5.1 Cost Estimates

The Cadmus Group (1998) document was again used to estimate POE and POU treatment costs. Costs are presented in Figure 5-2, and are based upon the following assumptions:

- Average household - 3 individuals, 1 gallon each per day, 1,095 gallons per year;
- Annual treatment - 1,095 gallons (POU), 109,500 gallons (POE);
- Minimally skilled labor - $14.50 per hour (population less than 3,300 individuals);
- Skilled labor - $28.00 per hour (population greater than 3,300 individuals);
- Life of unit - 5 years (POU), 10 years (POE);
- Duration of cost study - 10 years (therefore, two POU devices per household);
- Cost of water meter and automatic shut-off valve included;
- No shipping and handling costs required;
- Volume discount schedule - retail for single unit, 10% discount for 10 or more units, vendor retains 30% profit on more than 100 units;
- Installation time - 1 hour unskilled labor (POU), 3 hours, skilled labor (POE);
- O&M costs include maintenance, replacement of pre-filters and resin cartridges, laboratory sampling and analysis, and administrative costs.

5.6 ACTIVATED ALUMINA

Activated alumina (AA) can be used in packed beds to remove inorganic contaminants, such as arsenic, selenium, etc., from source water. In the AA process, contaminants are exchanged with the hydroxide ions on the alumina surface. Depending upon pH, AA can act as either an anionic (pH greater than 8.2) or cationic (pH less than 8.2) exchange process, but is rarely used as a cationic process for water treatment.

Arsenic removal by AA has been shown to be most effective near pH of 5.5 to 6.0. Most water systems will need some type of pH adjustment to accommodate this requirement. For POE and POU systems this can be accomplished by treating the AA bed with dilute sulfuric acid. This minimizes the possibility of unsafe treated water due to acidic pH, as well as the likelihood that additional pH adjustment would be necessary to raise the pH after treatment.

5.6.1 Cost Estimates

The Cadmus Group (1998) document was again used to estimate POE and POU treatment costs. Costs are presented in Figure 5-3, and are based upon the following assumptions:

- Average household - 3 individuals, 1 gallon each per day, 1,095 gallons per year;
- Annual treatment - 1,095 gallons (POU), 109,500 gallons (POE);
- Minimally skilled labor - $14.50 per hour (population less than 3,300 individuals);
- Skilled labor - $28.00 per hour (population greater than 3,300 individuals);
- Life of unit - 5 years (POU), 10 years (POE);
- Duration of cost study - 10 years (therefore, two POU devices per household);
- Cost of water meter and automatic shut-off valve included;
- No shipping and handling costs required;
- Volume discount schedule - retail for single unit, 10% discount for 10 or more units, vendor retains 30% profit on more than 100 units;
- Installation time - 1 hour unskilled labor (POU), 3 hours, skilled labor (POE);
- O&M costs include maintenance, replacement of pre-filters and resin cartridges, laboratory sampling and analysis, and administrative costs.

6.0 REGIONALIZATION

6.1 BACKGROUND

The term regionalization is used to define the process of purchasing and transporting water from one community to another. In effect, regionalization expands the region served by a water distribution system. Numerous economic, geographic, and operational factors can influence the decision to implement regionalization, including: 1) the availability of water; 2) water quality; 3) geography; and 4) economic factors.

Thriving communities that rapidly expand can easily outgrow their water source and find themselves faced with water shortage problems. To alleviate this problem, communities may decide to purchase water from other available sources in the region or neighboring communities. Water quality also plays a role in the decision making process. If a community's source water is contaminated, it may be cheaper for the community to purchase water from another rather that treat its own water source. In some cases, contaminated water cannot be sufficiently treated and a community may be faced with a choice to establish a new water source or to purchase water from a neighboring community.

Economic factors affecting the decision for regionalization include design, materials, construction, land, labor, and operational costs. Design costs include the engineering fees paid for the design of the regionalization system. Material costs include piping, fittings, gaskets, bends, valves, booster stations, pumps, and cathodic protection, among others. Construction costs include the costs associated with equipment rental and operation, excavation, backfilling, compaction, and landscaping. Land costs include the land required for the placement of the piping and booster stations, and the land required for the pipeline right-of-way. Labor costs include equipment operators, laborers, superintendents, and site engineer. Operational costs include energy costs, replacement parts, calibration, retrofitting, and operator costs.

The geographic location of a community will greatly affect the economic feasibility of regionalization. The distance from the water source will affect construction and equipment costs, and hilly or mountainous terrain can add significant design and construction costs. In addition, obtaining right-of-way for pipelines and booster stations may be a significant factor in the decision making process.

Additional factors include the lack of availabile water sources or change in the source availability due to increased draw down of groundwater, droughts affecting reservoirs, and other man-made or natural changes to the water source. Increased per capita water use can increase the demand for a larger water source or a new one, which also affects the decision process when considering regionalization. Political issues associated with natural drainage boundaries, the desire to avoid dependence on a single water source, and the reliability of the water source supply can also effect the decision making for regionalization.

6.2 COST ESTIMATES

Estimated costs for regionalization of drinking water were developed based upon three construction cost estimating sources: 1) the 1995 Environmental Cost Handling Options and Solutions (ECHOS); 2) the 1994 National Construction Estimator (NCE); and 3) the 1997 $^{\text{RS}}$Means - Site Work and Landscape Cost Data, 16th edition. The data collected for the cost estimates are included in Appendix F. Table 6-1 compares the cost of regionalization using reinforced concrete pipe, HDPE pipe, and ductile iron pipe as the conveyance conduit material. The following assumptions were made for the purpose of estimating regionalization costs:

- A 92" wide by 120" deep trench was excavated for the placement of the conveyance conduit. The width of the trench allows hand compaction around the pipe, the depth is an average depth.
- Type of soil was not taken into consideration, and no-rock excavation was assumed.
- 12" of fine gravel and sand were used to underlay the pipe.
- 3-48 magnesium anodes, at a spacing of 5 per mile, are assumed for the ductile iron pipe cathodic protection.
- The costs developed do not include costs associated with fittings, bends, gaskets, tees, etc. These costs may vary greatly depending upon the topography of the site.

- Air valves were assumed at 2 valves per mile. The location of air valves is also dependent on the topography of the site; valves are usually located at the high points.

- One booster station with a 100 GPM, 150' Head, and a 10 HP centrifugal pump, is assumed every two miles along the pipeline. The spacing and size of booster stations are site dependent.

- The cost of land purchase is not included.

- No escalation factors were used.

- The cost estimates do not include design costs, contractor profit and/or additional costs.

Table 6-1

Regionalization Cost Estimates

Pipe Construction	Pipe Diameter (inches)	1995 ECHOS ($/mile)	1994 NCE ($/mile)	1997 ^{RS}MEANS ($/mile)
Reinforced Concrete - Class 3	12	$331,399	$206,707	$286,128
	15	$341,906	$236,117	$301,176
	24	$402,098	$315,950	$374,568
HDPE	12	NA	$290,659	$329,688
	16	NA	$376,195	$356,088
	24	NA	$597,955	$501,288
Ductile Iron	12	$409,706	$389,188	$428,956
	16	$456,170	$497,639	$518,716
	24	$664,202	$698,543	$685,036

NA = Not Available

7.0 REFERENCES

Amy, G.L. and P. Brandhuber (1998). “Arsenic Rejection by Nanofiltration and Ultrafiltration Membranes: Arsenic Forms vs. Membrane Properties,” AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Aptel, P. and C.A. Buckley (1996). “Categories of Membrane Operations,” Chapter in Water Treatment Membrane Processes, McGraw Hill, 1996. American Water Works Research Foundation, Lyonnaise des Eaux, Water Research Commision of South Africa, 1996.

AWWA (1990). Water Quality and Treatment - A Handbook of Community Water Systems, McGraw-Hill Publishing Company, New York.

AWWA (1997). Arsenic Regulation, USEPA, Office of Ground Water and Drinking Water, October 1997.

AWWARF (1996) - Water Treatment Membrane Processes, McGraw-Hill Publishing Company, New York.

AWWARF (1998). Arsenic Treatability Options and Evaluation of Residuals Management Issues, Amy, G.L., M. Edwards, M. Benjamin, K. Carlson, J. Chwirka, P. Brandhuber, L. McNeill and F. Vagliasindi, Draft Report, April 1998.

Bellack, E. (1971). “Arsenic Removal from Potable Water,” *J. AWWA*, 63:7:454-458.

Benjamin, M.M., R.S. Sletten, R.P. Bailey and T. Bennett (1996). “Sorption and Filtration of Metals Using Iron-Oxide Coated Sand,” *Water Research*, 30:11:2609.

Benjamin, M. M., R.S. Sletten, R.P. Bailey and T. Bennett (1998). “Sorption of Arsenic by Various Adsorbents,” AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Cadmus Group, Inc. (1998). Cost Evaluation of Small System Compliance Options - Point-of-Use and Point-of-Entry Treatment Units, Prepared for USEPA Office of Ground Water and Drinking Water.

CH2M Hill (1998). Telephone conversation with Joseph Chwirka, CH2M Hill, Albuquerque, New Mexico.

Chang, S.D., W.D. Bellamy and H. Ruiz (1994). “Removal of Arsenic by Enhanced Coagulation and Membrane Technology,” Proceedings AWWA Annual Conference, New York.

Chen, H., M. Frey, D. Clifford, L.S. McNeill and M. Edwards (1998). “Raw Water Quality and Arsenic Treatment,” Submitted for Publication in *J. AWWA.*

Chen, H., and M. Edwards (1996). "Speciation of As in US Waters," Poster Session AWWA Annual Conference, Toronto.

Chen, S.L., S. R. Dzeng and M. Yang (1994). "Arsenic Species in Groundwaters of the Blackfoot Disease Area, Taiwan," *Environmental Science and Technology,* 28:5:877.

Cheng, R.C., S. Liang, H-C Wang and J. Beuhler (1994). "Enhanced Coagulation for Arsenic Removal," *J. AWWA,* 9:79-90.

Clifford, D., L. Ceber and S. Chow (1983). "As (III) and As (V) Separation by Chloride-Form Ion Exchange Resins," Proceedings AWWA WQTC, Norfolk, Virginia.

Clifford, D. and C.C. Lin (1985). Arsenic (III) and Arsenic (V) Removal From Drinking Water in San Ysidro, New Mexico, University of Houston, Houston, Texas.

Clifford, D. and C.C. Lin (1986). Arsenic Removal From Groundwater in Hanford, California - A Preliminary Report, University of Houston, Department of Civil/Environmental Engineering. Cited in: AWWA (1990). Water Quality and Treatment - A Handbook of Community Water Systems, McGraw-Hill Publishing Company, New York.

Clifford, D. (1994). "Computer Prediction of Arsenic Ion Exhange," *J. AWWA,* 86:4:10.

Clifford, D. and C.C. Lin (1995). "Ion Exchange, Activated Alumina, and Membrane Processes for Arsenic Removal from Groundwater," Proceedings of the 45th Annual Environmental Engineering Conference, University of Kansas, February 1995.

Clifford, D., G. Ghurye and A. Tripp (1998). "Arsenic Removal by Ion Exchange With and Without Brine Reuse," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Culp/Wesner/Culp (1979). Estimating Water Treatment Costs, Volume 2: Cost Curves Applicable to 1 to 200 mgd Treatment Plants, CWC Engineering, San Clemente, California.

Culp/Wesner/Culp (1984). Estimation of Small System Water Treatment Costs, CWC Engineering Software (for USEPA), San Clemente, California.

Culp/Wesner/Culp (1994). WATERCO$T Model - A Computer Program for Estimating Water and Wastewater Treatment Costs (Version 2.0), CWC Engineering Software, San Clemente, California.

Davis, M.K., K.D. Reich and M.W. Tikkanen (1994). "Nationwide and California Arsenic Occurrence Studies," Published in Arsenic Exposure and Health, Chapman and Hall Publishing Company, pp. 31-40.

DPRA, Inc. (1993a). Small Water System Byproducts Treatment and Disposal Cost Document, Prepared for USEPA Office of Ground Water and Drinking Water.

DPRA, Inc. (1993b). Water System Byproducts Treatment and Disposal Cost Document, Prepared for USEPA Office of Ground Water and Drinking Water.

Driehaus, W., M. Jekel and U. Hildebrandt (1998). "Granular Ferric Hydroxide - A New Adsorbent for the Removal of Arsenic from Natural Water," *J. Water SRT - Aqua*, 47:1:30-35.

Edwards, M.A. (1994). "Chemistry of Arsenic Removal During Coagulation and Fe-Mn Oxidation," *J. AWWA,* September 1994, pp. 64-77.

Edwards, M., S. Patel, L. McNeill, H. Chen, M. Frey, A.D. Eaton, C.R. Antweiler and E. Taylor II (1998). "Considerations in Arsenic Analysis and Speciation," *J. AWWA*, 90:3:103.

Elson, C.M., D.H. Davies, and E.R. Hayes (1980). "Removal of Arsenic from Contaminated Drinking Water by Chitosan/Chitin Mixture," *Water Research,* 14:1307.

EPA (1989). SAB Review of Arsenic Issues Relating to the Phase II of Proposed Regulations from the ODW, EPA SAB-EHC-89-038.

EPA (1992). Review of ORD's Arsenic Research Recommendations, EPA SAB-DWC-92-018.

EPA (1994). Review of the Draft Drinking Water Criteria Document on Inorganic Arsenic, EPA SAB-DWC-94-004.

EPA (1995). An SAB Report: Review of Issues Related to the Regulation of Arsenic in Drinking Water, EPA SAB-DWC-95-015.

EPA (1997). Annotated Outline for Technology and Cost Document in Draft Regulatory Analysis and Standards Support Documentation, September 8, 1997.

EPA (1997). Technology Design Conference Information Package, USEPA Office of Ground Water and Drinking Water.

EPA (1998a). Guide for Implementing Phase I Water Treatment Upgrade, USEPA Office of Ground Water and Drinking Water.

EPA (1998b). Water Treatment Costs Development (Phase I): Road Map to Cost Comparisons, USEPA Office of Ground Water and Drinking Water.

Ficklin, W.H. (1983). "Separation of As (III) and As (V) in Ground Waters by Ion-Exchange," *Talanta,* 30:5:371.

Fox, K.R. and T.J. Sorg (1987). "Controlling Arsenic, Fluoride, and Uranium by Point-of-Use Treatment," *J. AWWA,* 10:81-84.

Fox, K.R. (1989). "Field Experience with Point-of-Use Treatment Systems for Arsenic Removal," *J. AWWA*, 2:94-101.

Fox, K.R. (1994). "Point-of-Use Treatment for Arsenic Removal," USEPA Risk Reduction Engineering Laboratory 20th Annual Symposium, March 15-17, 1994, pp. 52-54.

Frank, P. and D. Clifford (1990). Project Summary: Arsenic (III) Oxidation and Removal from Drinking Water, EPA/600/S-2-86/021.

Frey, M.M. and M. Edwards (1997). "Surveying Arsenic Occurrence," *J. AWWA,* 89:3:105.

Frey, M.M., M. Edwards and G. Amy (1997). National Compliance Assessment and Costs for the Regulation of Arsenic in Drinking Water, AWWA, Washington.

Frey, M., D.M. Owen, Z.K. Chowdhury, R.S. Raucher and M. Edwards (1998). "Cost to Utilities of a Lower MCL for Arsenic," *J. AWWA*, 90:3:89.

Fuller, C.C., J.A. Davis, G.W. Zellwegger and K.E. Goddard (1988). "Coupled Chemical, Biological, and Physical Processes in Whitewood Creek, South Dakota: Evaluation of the Controls of Dissolved Arsenic," Proceedings of the Technical Meeting, Arizona, US Geological Survey Toxic Substances Hydrology Program, September 26-30, 1988.

Garling, R.A. (1981). "Evaluation of Electrodialysis for Process Water Treatment for In-Situ Mining," Presented at Fifth Annual Uranium Seminar, Albuquerque, NM.

Ghurye, G., D. Clifford and J. Tong (1998). "Iron Coagulation/Filtration for Arsenic Removal," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Gibbs, J.W. and L.P. Scanlan (1998). "Arsenic Removal in the 1990's: Full Scale Experience from Park City, Utah," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998

Great Lakes Upper Mississippi River Board of State Public Health and Environmental Managers (1997). Recommended Standards for Water Works, Health Education Services, Albany, NY, 1997.

Gulledge, J.H. and J.T. O'Connor (1973). "Removal of Arsenic (V) From Water by Adsorption on Aluminum and Ferric Hydroxides," *J. AWWA,* 8:548-552.

Gupta, S.K. and K.Y. Chen (1978). "Arsenic Removal by Adsorption," *Journal of WPCF,* 3:3:493.

Harper, T.R. and N.W. Kingham (1992). "Removal of Arsenic from Wastewater Using Chemical Precipitation Methods," *Water Environment Research,* 64:3:200.

Hathaway, S.W. and F. Rubel, Jr. (1987). "Removing Arsenic From Drinking Water," *J. AWWA*, 79:8:61-65.

Hecht, P.M. U.G. Kelkar and D.J. Hiltebrand (1994). "Enhanced Coagulation: Impacts on Residual Management Programs," Proceedings AWWA Annual Conference, New York.

Hering, J.G. and M. Elimelech (1996). Arsenic Removal by Enhanced Coagulation and Membrane Processes, AWWA, Denver.

Hering, J.G., P. Chen, J.A. Wilkie, M. Elimelech, and S. Liang (1996). "Arsenic Removal by Ferric Chloride," *J. AWWA*, 88:4:155-167.

Hering, J.G., P-Y. Chen, J.A. Wilkie and M. Elimelech (1997). "Arsenic Removal From Drinking Water During Coagulation," *J. Env. Engineering,* 123:8:800-807.

Hering, J.G., and V.Q. Chiu (1998). "The Chemistry of Arsenic: Treatment and Implications of Arsenic Speciation and Occurrence," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Hund, R., E. Lomaquahu, A. Eaton and X.C. Le (1998). "Detection and Preservation of Arsenic Species," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Hydrometrics, Inc. (1997). Summary Report on the Sulfur-Modified Iron (SMI) Process, Hydrometrics, Inc., Helena, Montana.

Hydrometrics, Inc. (1998). Second Interim Report on the Sulfur-Modified Iron (SMI) Process for Arsenic Removal, Hydrometrics, Inc., Helena, Montana.

ICF, Incorporated and ISSI, Incorporated (1998). Evaluation of Full-Scale Treatment Technologies at Small Drinking Water Systems, Prepared for USEPA Office of Ground Water and Drinking Water.

Indiana-American Water Company, Inc. (1998). Personal communication with Jeff Robinson, Indiana-American Water Company, Inc., Greenwood, Indiana.

International Consultants, Inc. (1998). National Level Affordability Criteria Under the 1996 Amendments to the Safe Drinking Water Act, Prepared for USEPA Office of Ground Water and Drinking Water.

Ionics, Inc. (1989-90). Comments on the USEPA National Primary and Secondary Drinking Water Regulations - Proposed Rule, Federal Register Vol. 54, No. 97. Submitted to USEPA, 1989; Addendum to the Comments, 1990.

Kempic, J.B. (1994a). "Arsenic Removal Technologies: An Evaluation of Cost and Performance," Published in: Arsenic: Exposure and Health Effects, Chapman and Hall Publishing Company, pp. 393-405.

Kempic, J.B. (1994b). Arsenic: Best Available Technology - Presentation to the Science Advisory Board, USEPA, Drinking Water Standards Division, Office of Ground Water and Drinking Water, August 1994.

Kempic, J.B. (1994c). Arsenic: Regulatory Impact Analysis Decision Tree - Presentation to the Science Advisory Board, USEPA, Drinking Water Standards Division, Office of Ground Water and Drinking Water, August 1994.

Kempic, J.B. (1994d). Basis for Revised Anion Exchange Costs, USEPA, Drinking Water Standards Division, Office of Ground Water and Drinking Water, December 1994.

Lauf, G.F. (1998). "Arsenic Removal by Oxidation with Potassium Permanganate in a Conventional Iron and Manganese Removal Process," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Le, X.C., and M. Ma (1998). "Determination of μg/L Levels of Arsenic Species in Water and Human Urine," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Logsdon, G.S., T.J. Sorg, and J.M. Symons (1974). "Removal of Heavy Metals by Conventional Treatment," Proc. 16th Water Quality Conference - Trace Metals in Water Supplies: Occurrence, Significance, and Control, University Bulletin No. 71, U. of Illinois.

Lomaquahu, E.S., and A.H. Smith (1998). "Feasibility of New Epidemiology Studies on Arsenic Exposures at Low Levels," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Malcolm Pirnie, Inc. (1991). RIA/EIA Assessment, AWWA Government Affairs, Washington D.C.

Malcolm Pirnie, Inc. (1992). Arsenic Removal Pilot-Scale Studies, Prepared for USEPA Office of Ground Water and Drinking Water.

Malcolm Pirnie, Inc. (1993a). Treatment and Occurrence of Arsenic in Potable Water Supplies, USEPA, September 1993.

Malcolm Pirnie, Inc. (1993b). Very Small Systems Best Available Technology Cost Document, USEPA, Office of Ground Water and Drinking Water, Drinking Water Technology Branch.

Malcolm Pirnie, Inc. (1994). Baseline Assumptions and Estimated Unit Process Costs for Arsenic Removal, Technical Memorandum No. I.

Malcolm Pirnie, Inc. (1996). Treatment and Residual Costs for Arsenic Removal, AWWA/University of Colorado, Technical Memorandum.

McMullin, M., R. Berry and E. Winchester (1998). "Low-Cost Direct Filtration for Removal of Arsenic From Groundwater," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

McNeill, L.S. and M.A. Edwards (1995). "Soluble Arsenic Removal at Water Treatment Plants," *J. AWWA,* April 1995, pp. 105-113.

McNeill, L.S. and M. Edwards (1996), "Predicting Arsenate Removal in the Presence of Metal Oxides," Proceedings AWWA Annual Conference, Toronto.

McNeill, L.S. and M. Edwards (1997a), "Arsenic Removal During Precipitative Softening," *ASCE Journal of Environmental Engineering,* 125:5:453.

McNeill, L.S. and M. Edwards (1997b). "Predicting Arsenic Removal During Metal Hydroxide Precipitation," *J. AWWA,* 89:1:75.

McNeill, L.S. and M. Edwards (1998). "Arsenic Removal Via Softening and Conventional Treatment", AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Mortazavi, S., K. Volchek, A. Tremblay, F.H. Tezel and H. Whittaker (1995). "An Advanced Process for the Removal of Arsenic From Effluents," Proceedings of the Twelfth Technical Seminar of Chemical Spills, June 12-13, 1995, Edmonton, Alberta, Canada, pp. 21-37.

Muilenberg, T. (1997). "Microfiltration Basics: Theory and Practice," Proceedings Membrane Technology Conference, February 23-26, 1997, New Orleans, LA.

New Mexico State University (1979). Water Treatment for Small Public Supplies - Report of Operation: Cuba, Carrizozo, La luz, San Jon, San Ysidro, Bluewater, Moriarty, Hagerman, La Cruces, NM.

Nikolaidis, N.P., J. Lackovic and G. Dobbs (1998). "Arsenic Remediation Technology - AsRT," Environmental Research Institute Technical Publication.

Papadimas, S.P., Z.K. Chowdhury and M. Frey (1997). "Development of Treatment and Residual Cost Curves for Arsenic Removal," AWWA Annual Conference Proceedings, Atlanta, GA, June 1997.

Pickering, K., and M.R. Wiesner (1993). "Cost Model for Low-Pressure Membrane Filtration," *J. Env. Eng.,* 119:5:772.

Pontius, F.W. (1998). "New Horizons in Federal Regulation," *J. AWWA*, 90:3:38-50.

Raucher, R.S. (1994). Estimating the Cost of Compliance With Drinking Water Standards: A User's Guide, AWWA, Washington.

Reiber, S., G. Dostal, L. Onnen, J. McCafferty, W. Sund, and L. Andreasen (1997). Final Report: Groundwater Disinfection and Arsenic Concentration in the Fremont Distribution System, Report to USEPA, November 1997.

Robinson, J. (1998). "Arsenic Removal - Case Study of Full Scale Coagulation/Filtration Plant," AWWA Inorganic Contaminants Workshop, San Antonio, TX, February 23-24, 1998.

Rosenblum, E.R. and D.A. Clifford (1984). The Equilibrium Arsenic Capacity of Activated Alumina, PB 84/10 527, NTIS, Springfield.

SAIC, HDR (1994). Summary of Arsenic Treatment Workshop, USEPA, Office of Ground Water and Drinking Water, January 1994.

SAIC (1998). Technolologies and Costs for the Removal of Radon from Drinking Water, Draft Report, Prepared for USEPA Office of Ground Water and Drinking Water.

SAIC (1999). Evaluation of Central Treatment Options as Small System Treatment Technologies - Technology Cost Estimates, Prepared for USEPA Office of Ground Water and Drinking Water.

Scott, K.N, et al. (1994). "Arsenic Removal Using Elevated Dosages of Ferric Chloride and Alum at a Full-Scale Conventional Treatment Plant," Proceedings AWWA Annual Conference, San Francisco, California.

Scott, K., J. Green, H.D. Do and S.J. McLean (1995). "Arsenic Removal by Coagulation," *J. AWWA*, 4:114-126.

Sethi, S. and M.R. Wiesner (1995). "Performance and Cost Modeling of Ultrafiltration,"*J. Env. Eng.*, 121:12:883.

Shen, Y.S. (1973). "Study of Arsenic Removal from Drinking Water," *J. AWWA*, 8:543-548.

Simms, J. and F. Azizian (1997). "Pilot Plant Trials on the Removal of Arsenic from Potable Water Using Activated Alumina," Proceedings AWWA Water Quality Technology Conference, November 9-12, 1997.

Sorg, T.J. and G.S. Logsdon (1978). "Treatment Technology to Meet the Interim Primary Drinking Water Regulations for Inorganics: Part 2," *J. AWWA*, 7:379-392.

Sorg, T.J. (1990). “Methods of Removing Drinking Water Contaminants and Their Limitations: Inorganics and Radionuclides,” USEPA Drinking Water Research Division, Cincinnati, Ohio, PB91-162792.

Stewart, H.T. and K.J. Kessler (1991). “Evaluation of Arsenic Removal by Activated Alumina Filtration at a Small Community Public Water Supply,” *Journal of New England Water Works Association,* 105:179-199.

Subramanian, K.S., T. Viraraghavan, T. Phommavong and S. Tanjore (1997). “Manganese Greensand for Removal of Arsenic in Drinking Water,” *Water Quality Research Journal Canada,* 32:3:551-561.

Thomson, B.M. and M. O’Grady (1998). Evaluation of Point-of-Use Water Treatment Systems, San Ysidro, New Mexico, Final Report to USEPA, February 1998.

Vickers, J.C., A. Braghetta, and R.A. Hawkins (1997). “Bench Scale Evaluation of Microfilitration for Removal of Particles and Natural Organic Matter,” Proceedings Membrane Technology Conference, February 23-26, 1997, New Orleans, LA.

Wageman, R. (1978). “Some Theoretical Aspects of Stability and Solubility of Inorganic Arsenic in the Freshwater Environment,” *Water Research,* 12:139.

WMA, Inc. (1994). Assembly of Compliance Decision Tree for Arsenic, Correspondence with Ms. Heather Shank-Givens, EPA, Office of Ground Water and Drinking Water, November 1994.

WMA, Inc. (1995). Arsenic: Impact Summaries (EPA Contract 68-C3-0368), Correspondence with Ms. Heather Shank-Givens, EPA, Office of Ground Water and Drinking Water, January 1995.

Waypa, J.J., M. Elimelech and J.G. Hering (1997). “Arsenic Removal by RO and NF Membranes,” *J. AWWA,* 89:10:102.

Westerhoff, G. and Z.K. Chowdhury (1996). “Water Treatment Systems,” Chapter in Water Resources Handbook, L.M. Mays ed., McGraw Hill, New York, 1996.

Wiesner, M.R., S. Sethi, J. Hackney, J. Jacangelo and J.M. Laine (1994). “A Comparison of Cost Estimates for Membrane Filtration and Conventional Treatment,” *J. AWWA,* 86:12:33-41.

Wilkie, J.A. and J.G. Hering (1998). “Rapid Oxidation of Geothermal Arsenic (III) in Streamwaters of the Eastern Sierra Nevada,” *Environmental Science and Technology,* 32:5:657-662.

APPENDIX A

VERY SMALL SYSTEMS
CAPITAL COST BREAKDOWN SUMMARIES

APPENDIX B

WATER MODEL
CAPITAL COST BREAKDOWN SUMMARIES

APPENDIX C

W/W COST MODEL
CAPITAL COST BREAKDOWN SUMMARIES

APPENDIX D

COST EQUATIONS AND CURVE FITS
FOR REMOVAL AND ACCESSORY COSTS

APPENDIX E

ADDITIONAL CAPITAL COSTS

APPENDIX F

REGIONALIZATION COST