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Technologies and Costs for Removal of Arsenic from Drinking Water

TECHNOLOGIES AND COSTS FOR
REMOVAL OF ARSENIC FROM DRINKING WATER

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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
Fg/L	micrograms per liter

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 predominantle As(III) since 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, specifically activated alumina (AA);

#	Ion exchange (IX) processes, specifically anion exchange;

#	Membrane filtration, including reverse osmosis (RO), and electrodialysis reversal (EDR);

Alternative treatment processes specifcally greensand filtration; and

Point-of-use (POU) devices.

Many of these processes were evaluated to develop cost curves for the technologies. Discussions of the following technologies are included in this document for future consideration as viable processes for arsenic removal:

Iron oxide coated sand;
Nanofiltration;
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 $\digamma$g/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. The SDWA was later amended to require the final regualation to be issued by June 22, 2001.

1.3   DOCUMENT ORGANIZATION

This document contains the following chapters:

Chapter 1.0 Introduction - Provides an introduction to the arsenic statutory requirements, and defines technology categories. Also 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.

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 selected treatment options.

Chapter 6.0 References - Lists the literature cited in this document, as well as additional references which may be of interest to the reader.

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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 from drinking water. 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 of these processes may be technically feasible, their cost may be prohibitive.

Technologies discussed in this chapter 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 (C/F) 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 is 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.

Field Studies

The field operation of two coagulation filtration plants was studied by Battelle Memorial Institute, with funding from EPA (EPA, June 2000). One plant (Plant A) uses ozonation coupled with coagulation/filtration to treat up to 600 mgd. The other water system (Plant B) relies on coagulation, sedimentation, and filtration, and was designed to treat a much lower daily flow (62.5 mgd). Both plants demonstrated the ability to consistently reduce moderately high average influent arsenic concentrations (7.5 and 19.1 Fg/L) to less than 5 Fg/L (3.5 and 4.0 Fg/L) in finished water. Furthermore, it should be noted that these plants were not using optimal coagulant and/or polymer doses, and were not operated at the ideal pH for arsenic removal.

Based on the observation of the field operation of these systems, adsorption and coprecipitation of As(V) with iron and aluminum flocs appears to be the principal mechanism for arsenic removal at these plants. As part of this study, sludge samples were collected from both Plant A and Plant B and subjected to TCLP-testing. However, based on the results of the TCLP, these sludges would not be characterized as hazardous wastes.

Summary

Coagulation technology can successfully achieve As(V) removals greater than 90 percent. As noted in the field study discussed above, coagulation/filtration plants have demonstrated the capacity

to reduce arsenic levels below 5 ɧg/L. Moreover, if optimal operating practices are adopted, it is anticipated that effluent levels of less than 3 ɧg/L will be obtainable.

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. The presence of sulfates significantly decreases As(III) removal, but only 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 commonly used by facilities treating groundwater. The oxidation process used to remove iron and manganese leads to the formation of hydroxides that remove soluble arsenic by precipitation or adsorption reactions.

Arsenic removal during iron precipitation is 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 resulted in the removal of 83 percent of influent As(V) arsenic from a source with 22 µg/L As(V). Indeed, field studies of iron removal plants have indicated that this treatment can feasibly reach 3 ɧg/L (see discussion below). However, the removal efficiencies achieved by iron removal are not as high or as consistent as those realized by activated alumina (see section 2.3.1) or ion exchange (see section 2.4).

Note, however, that arsenic removal 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.

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

Substantial arsenic removal has been seen using greensand filtration (Subramanian, *et al.*, 1997). The active material in "greensand" is glauconite, a green, iron-rich, clay-like mineral that has ion exchange properties. Glauconite often occurs in nature as small pellets mixed with other sand particles, giving a green color to the sand. The glauconite sand is treated with $KMnO_4$ until the sand grains are 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, including 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 has 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.

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.

Field Studies

The field operation of two iron removal plants was studied by Battelle Memorial Institute, with funding from EPA (EPA, August 2000). One plant (Plant A) used an iron removal process followed by zeolite softening, while the second (Plant B) relied solely on an iron removal process. Both plants treated a similar volume of water (1.6 mgd and 1.4 mgd, respectively), however, the source water of the former contained an average concentration of 2,284 Fg/L of iron, while the source water of the latter was characterized by a much lower concentration (1,137 Fg/L). Since arsenic removal is heavily dependent on the amount of iron in the influent, it is not surprising that Plant B was not as effective in the removal of arsenic–while Plant A consistently reduced arsenic levels by 87 percent (from 20.3 Fg/L to 3.0 Fg/L), Plant B only reduced arsenic concentrations by 74 percent (from 48.5 Fg/L to 11.3 Fg/L). Based on the data collected during the study, and additional experience, the authors of the study concluded that the removal efficiency of Plant B could be improved by the addition of a coagulant such as ferric chloride.

As part of the same study, sludge samples were taken from Plant A and subjected to TCLP-testing. Based on the results of these tests, the wastes would not be characterized as hazardous, even under the strict hazardous waste regulations of California. Although sludge samples were not collected at Plant B during this study, an earlier test result would have resulted in the exceedance of California's limit on total arsenic for non-hazardous wastes.

2.2.3   Coagulation Assisted Microfiltration

Arsenic is removed effectively by 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 in a manner similar 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. This report is corroborated by pilot studies conducted by Clifford (1997), which found that coagulation assisted microfiltration could reduce arsenic levels below 2 Fg/L in waters with a pH of between 6 and 7, even when the influent concentration of Fe(III) is approximately 2.5 mg/L. These studies also found that the same level of arsenic removal could be achieved by this treatment process even if source water sulfate and silica levels were high. Further, coagulation assisted microfiltration can reduce arsenic levels to an even greater extent at a slightly lower pH (approximately 5.5).

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 for the reduction of disinfection byproduct (DBP) precursors for surface water systems which have sedimentation capabilities. The enhanced process 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.

Note that post-treatment pH adjustment may be required for corrosion control when the process is operated at a low pH.

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 pH 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 and 10 percent, whereas soluble As(V) removal at plants precipitating calcite and magnesium and/or ferric hydroxide was between 60 and 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 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.

Field Studies

Surveys of lime softening facilities by AWWARF (AWWARF 2000) show that the technology should be able to produce water with less than 3 µg/L–all five of the surveyed plants that operated at a pH of 10.2 or higher would be able to feasibly achieve these low arsenic levels. However, although lime softening processes are quite effective for arsenic removal when operated at high pH, lime softening facilities may not effectively remove arsenic to low levels when operated at or below pH 10.0. For example, a field study conducted by Battelle Memorial Institute, with funding from EPA, found that a 10 mgd lime softening plant only reduced average total arsenic concentrations of 32.0 µg/L by about 45 percent to an average of 16.6 µg/L in the finished water even though the treatment train included pre-softening oxidation and post-softening filtration (EPA, June 2000). None of the sludge samples collected as part of this study qualified as hazardous waste based on TCLP testing.

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 technology can be implemented by water systems to achieve greater than 90 percent removal of As(V). As discussed above, a survey of softening facilities by AWWARF found that those surveyed facilities operating at a pH of 10.2 or higher could feasibly reduce arsenic concentrations in the treated water below 3 µg/L.

Arsenic in the pentavalent arsenate form is more readily removed by this treatment process than the trivalent arsenite form. The optimum pH for As(V) removal by softening is approximately 10.5 and the optimum pH for removal of As(III) is approximately 11.0. As(III) removal appears to depend on the initial concentration of the contaminant. Initial findings indicated that As(V) removal was also a function of its concentration in the influent. However, more recent studies have shown that As(V) removal is actually independent of its 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 the 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 (BV). For the same level of As(V) removal, a column operating at pH 6.0 treated 13,391 BV and a column operating at a pH of 9.0 treated only 800 BV. 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)]:

$$OH^- > H_2AsO_4^- > Si(OH)_3O^- > F^- > HSeO_3^- > TOC > SO_4^{2-} > H_3AsO_3$$
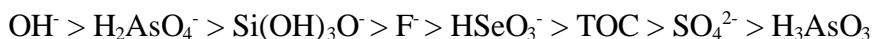
Several studies have illustrated the effects of this selectivity, particularly those associated with sulfate and chloride. Benjamin *et al.* (1998) found little effect produced by either sulfate or chloride. Increasing sulfate from 0 to 100 mg/L had only a small impact on the sorption of As(V). The presence 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.

Effect of Empty Bed Contact Time

The operation of AA beds, and in particular the EBCT, can also play a role in arsenic removal. EBCT represents the length 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.

Regeneration

Regeneration of AA beds is usually accomplished using a strong base solution, typically concentrated NaOH. Relatively few BV 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.

Field Studies

Several field studies have demonstrated that arsenic may be reduced to below 3 Fg/L using activated alumina filtration. For example, Stewart (1991) reports on the success of a small water utility in treating for arsenic. This utility, which serves 89 households in Bow, New Hampshire, relies on 2 wells for its source water. Each of the wells is characterized by a high level of arsenic contamination (mean arsenic concentrations of 62 Fg/L and 57 Fg/L). However, the utility was able to reduce arsenic levels below the current MCL of 50 Fg/L for 106 days while producing an average

treated flow of 13,730 gpd. In fact, the data shows that this system would have been able to reduce arsenic concentrations in treated water below 5 or 10 Fg/L.

In another study (Wang, 2000), the Battelle Memorial Institute, with funding from EPA, examined two activated alumina plants designed to treat for arsenic. The treatment trains for both of these plants consisted of two parallel sets of two tanks in series. The media in the lead tank of each set was replaced when it had reached its capacity for arsenic (approximately every 1.5 years). At this time, the second tank in the series was moved to the lead or 'roughing' position. Thus, in all cases, the second, or 'polishing,' tank contained virgin media. Challenged by arsenic concentrations ranging from 53 to 87 Fg/L (average concentration of 62 Fg/L), and 21 to 76 Fg/L (average concentration of 49 Fg/L), both systems reduced influent arsenic concentrations by more than 92 percent on average, consistently maintaining effluent arsenic concentrations below 5 Fg/L. Indeed, despite such high influent concentrations, both of these two systems were able to produce finished water with an arsenic concentration below 3 Fg/L for an extended period of time (40 and 10 weeks, respectively). Samples of spent media were collected from one of the treatment vessels and subjected to the TCLP test. In all cases, the spent media easily passed the TCLP test for toxicity. The results of these tests and their implications are discussed in more detail in chapter 4 of this document.

Media Fouling

Much like ion exchange resins, 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 BV 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

Field studies involving AA indicate that this technology can feasibly achieve arsenic removal to 3 Fg/L (see discussion above; Stewart 1991, Wang 2000). The operational experiences which have

been developed provide important information to be considered for AA processes; these are discussed here.

AA beds may be operated in series or parallel. Series operation increases removal and helps prevent 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. Systems operating activated alumina without regeneration will also benefit from series operation. Such an approach will provide greater utilization of the media before it is disposed.

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 prevent 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. For these reasons, it is recommended that systems utilize disposable activated alumina rather operate the process with regeneration.

Secondary Effects

AA processes may produce changes to the effluent water quality (EPA, 1994). When pre-treatment is 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 corrosion control to raise the pH would be necessary for those systems. Some systems, especially small systems, may choose to operate the process at the natural water pH. The two full-scale plants studied in the ORD project (Wang, 2000) were operated at natural pH values of 8 and 8.3. While this may not yield the optimal run length, it may be sufficient for smaller systems.

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 that can be treated before exhaustion varies with resin type and influent water quality. Typically from 300 to 60,000 BV can be treated before regeneration is required. In most cases, regeneration of the bed can be accomplished with only 1 to 5 BV of regenerant followed by 2 to 20 BV 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.

$$HCrO_4^- > CrO_4^{2-} > ClO_4^- > SeO_4^{2-} > SO_4^{2-} > NO_3^- > Br^- > (HPO_4^{2-}, HAsO_4^{2-}, SeO_3^{2-}, CO_3^{2-}) > CN^- >$$
$$NO_2^- > Cl^- > (H_2PO^{4-}, H_2AsO_4^-, HCO_3^-) > OH^- > CH_3COO^- > F^-$$
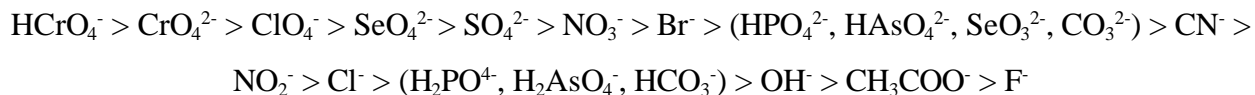
These resins have a relatively high affinity for arsenic in the arsenate form ($HAsO_4^{2-}$), however, previous studies have shown that high TDS and sulfate levels compete with arsenate and can reduce removal efficiency (AWWA, 1990). In general, ion exchange for arsenic removal is only applicable for low-TDS, low-sulfate source waters. 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 BV to exhaustion is limited by nitrate breakthrough. If the sulfate level is high (about 100 mg/L), however, the number of BV 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, resulting 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 above, 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 (following page). Data in Table 2-1 represent BV 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 because arsenic "leakage" often occurs 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 freshly-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).

TABLE 2-1

Typical IX Resins for Arsenic Removal

Resin	Type	Operating Parameters	BV 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 As(III) is
			    oxidized to As(V)]
				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

    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 (e.g., downward) and the regenerant is applied in the opposite direction (e.g., 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).

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, post-treatment corrosion control, 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 indicated by a 3-5 percent decrease in total BV to exhaustion over consecutive cycles, and by 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 BV 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 BV of 1.0 N NaCl or 5 BV of 0.25 N NaCl in "co-current flow" operation. Also, a rinsing cycle is required after regeneration; typically only a few BV 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. Spent regenerant 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, the spent solution must be treated and/or disposed.

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. The problem with this option is that the brine stream prior to chemical precipitation could be considered hazardous waste. As discussed in Chapter 4, brine streams with arsenic concentrations greater than 5 mg/L can be considered hazardous waste. Since arsenic will be even more concentrated in a brine reuse scenario, it is unlikely to be used unless the brine could be discharged to the POTW. The TDS content of the brine may restrict that option.

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 BV 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   Field Studies

Anion exchange processes have demonstrated the capacity to consistently reduce arsenic concentrations below 3 µg/L. The Battelle Memorial Institute, with funding from EPA, studied two ion exchange plants located in New England. (Wang, 2000) The treatment train of one of the plants consisted of a potassium permanganate greensand oxidizing filter followed by a mixed bed ion exchange system. This system was regenerated every 6 days and consistently reduced influent arsenic concentrations of 40 to 65 µg/L to below 5 µg/L. Indeed, this system was able to produce finished water with arsenic concentrations below 3 µg/L for approximately 30 weeks.

The treatment train of the second plant also incorporated pre-oxidation, consisting of a solid oxidizing media filter followed by an anion exchange system. However, although the influent arsenic concentrations facing this system (19 to 55 µg/L) were not as great as those handled by the other ion exchange system involved in this study, effluent arsenic concentrations for this plant were not maintained at a consistently low level, ranging from below 5 µg/L to more than 80 µg/L. Nonetheless, based on the breakthrough data collected during this study, the second system could feasibly achieve finished water levels of 5 µg/L, or even 3 µg/L, if it was regenerated with greater frequency (i.e., every 3 to 4 weeks).

This study also demonstrated that ion exchange has only a minimal impact on water pH when used to treat for arsenic. Following treatment, the average pH of one of the two plants was identical to the average influent pH (pH 7.5), while the average pH of the effluent from the second plant was only slightly below that of the influent (pH 7.3 and pH 8.3, respectively). Blending a treated portion with an untreated portion would reduce any impact on finished water pH.

2.4.12   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 must 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 while blocking the passage of others. The movement of constituents across a membrane requires a driving force (i.e. a potential difference between the two 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
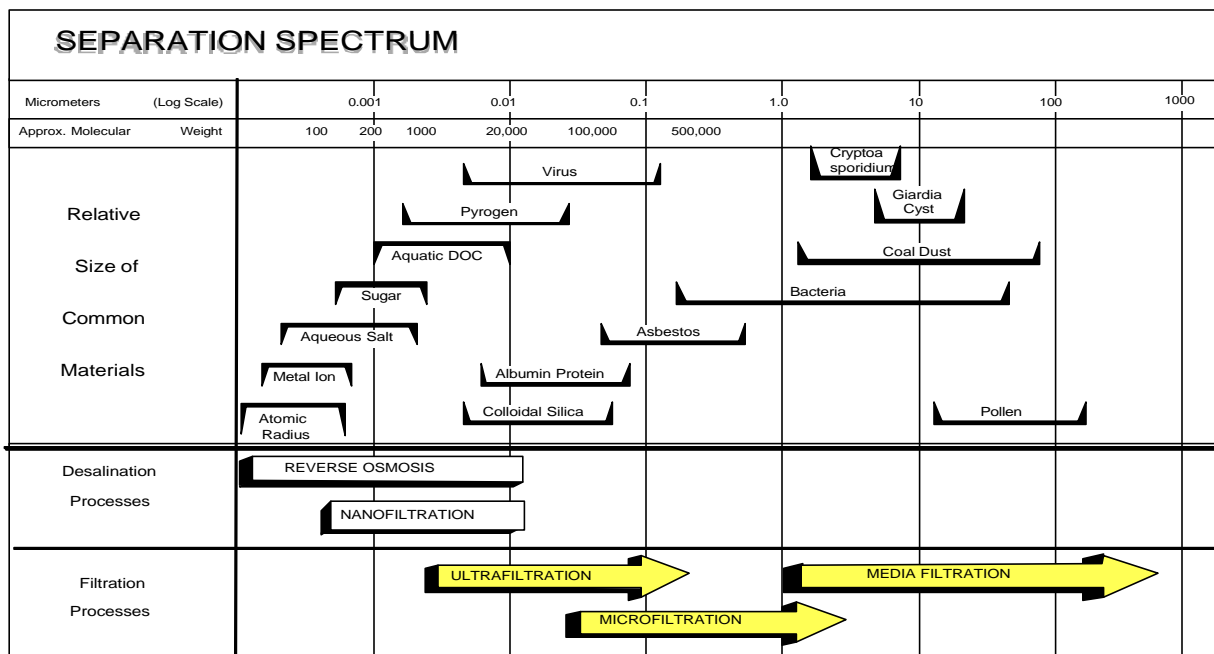
Figure 2-1 Pressure Driven Membrane Process Classification
(Westerhoff and Chowdhury, 1996)

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 on 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 (e.g., Ca, Mg), but not monovalent salts (e.g., 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 of 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 suggests that a negatively charged membrane could not keep high As (V) rejection rates for long durations without maintaining arsenic in the As(V) form. Additional 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.

Another study (Chang et al, 1994) revealed that the removal efficiency dropped significantly during pilot-scale tests where the process was operated at more realistic recoveries. When the membrane unit was operated at a recovery of 65%, the arsenic removal efficiency dropped to 65% and when the recovery was increased to 90%, the arsenic removal efficiency dropped down to 16%.

2.5.7　Reverse Osmosis

RO is the oldest membrane technology, traditionally used for the 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, and is very effective in removing dissolved constituents. 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 oxidation 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	C 3-5 gpd C 90% recovery C 20-100 psi	50%	Fox, 1989 Fox and Sorg, 1987
Eugene, OR	POU	C 3-5 gpd C 67% recovery C 195 psi	below MDL	Fox, 1989 Fox and Sorg, 1987
Fairbanks, AL	POU	C low-pressure (< 100 psi)	50%	Fox, 1989 Fox and Sorg, 1987
San Ysidro, NM	Pilot (hollow fiber, cellulose acetate)	C 50% recovery C pH adjustment to 6.3 C antiscalent addition	93-99%	Clifford and Lin, 1991
San Ysidro, NM	Pilot (hollow fiber, polyamide)	C 50% recovery C pH adjustment to 6.3 C antiscalent addition	99%	Clifford and Lin, 1991
San Ysidro, NM	POU	C 10-15% recovery	91%	Fox, 1989 Fox and Sorg, 1987
Tarrytown, NY	Pilot (FilmTec BW30, Hydranautics NCM1, Fluid Systems TCFL)	C 15 gfd C 10% recovery	below MDL	Malcolm Pirnie, 1992
Tarrytown, NY	POU	Not Available	86%	Rogers, 1989
Charlotte Harbor, FL	POU (several membrane types)	C 1000 gpd C 10-60% recovery	As(V) 96-99% As(III) 46-84%	Huxstep, 1987
Cincinnati, OH	POU	Not Available	As(III) 73%	Fox and Sorg, 1987
Hudson, NH	POU	Not Available	40%	USEPA, 1982

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.

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   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 affect 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.

2.6.2 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 capacities were obtained with final arsenic concentration of 0.050 mg/L. Arsenic removal was influenced by pH. Approximately

20 mg As per gram of iron was removed at pH 8, and 50 mg As per gram of iron was removed 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 per gram of 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 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.

A field study reported by Simms *et al.* (2000) confirms the efficacy of GFH for arsenic removal. Over the course of this study, a 5.3 mgd GFH plant located in the United Kingdom was found to reliably and consistently reduce average influent arsenic concentrations of 20 Fg/L to less than 10 Fg/L for 200,000 BV (over a year of operation) at an EBCT of 3 minutes. Despite insignificant headloss, routine backwashing was conducted on a monthly basis to maintain media condition and to reduce the possibility of bacterial growth. The backwash was not hazardous and could be recycled or disposed to a sanitary sewer. At the time of replacement, arsenic loading on the media was 2.3 percent. Leachate tests conducted on the spent media found that arsenic did not leach from the media.

The most significant weakness of this technology appears to be its cost. Currently, GFH media costs approximately $4,000 per ton. However, if a GFH bed can be used several times longer than an alumina bed, for example, it may prove to be the more cost effective technology. Indeed, the system profiled in the field study presented above tested AA as well as GFH and found that GFH was sufficiently more efficient that smaller adsorption vessels and less media could be used to achieve the same level of arsenic removal (reducing costs). In addition, unlike AA, GFH does not require pre-oxidation.

A treatment for leaching arsenic from the media to enable regeneration of GFH seems feasible, but it results in the generation of an alkaline solution with high levels of arsenate which requires further treatment to obtain a solid waste. Thus, direct disposal of spent GFH 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 can be recommended as an approved arsenic removal technology for drinking water.

2.6.5 Photo-Oxidation

Researchers at the Australian Nuclear Science and Technology Organisation (ANSTO) have found that in the presence of light and naturally occurring light-absorbing materials, the oxidation rate of As(III) by oxygen can be increased ten-thousandfold (Cooperative Research Centres for Waste Management and Pollution Control Limited, 1999). The oxidized arsenic, now As(V), can then be effectively removed by co-precipitation.

ANSTO evaluated both UV lamp reactors and sunlight-assisted-photo-oxidation using acidic, metal-bearing water from an abandoned gold, silver, and lead mine. Air sparging was required for sunlight-assisted oxidation due to the high initial As(III) concentration (12 mg/L). Tests demonstrated that near complete oxidation of As(III) can be achieved using the photochemical process. Analysis of process waters showed 97% of the arsenic in the process stream was present as As(V). Researchers also concluded that As(III) was preferentially oxidized in the presence of excess dissolved Fe(II) (22:1 iron to arsenic mole ratio). This is a contrast to conventional plants where dissolved Fe(II) represents an extra chemical oxidant demand which has to be satisfied during oxidation of As(III) (CRC-WMPC, 1999).

Photo-oxidation of the mine water followed by co-precipitation was able to reduce arsenic concentrations to as low as 17 Fg/L, which meets the current MCL for arsenic. Initial total arsenic concentrations were unknown, though the As(III) concentration was given as approximately 12 mg/L, which is considerably higher than typical raw water arsenic concentrations. ANSTO reported residuals from this process are environmentally stable and passed the Toxicity Characteristic Leaching Procedure (TCLP) test necessary to declare waste non-hazardous and suitable for landfill disposal.

Based on the removals achieved and residuals characteristics, it is expected that photo-oxidation followed by co-precipitation would be an effective arsenic removal technology. However, this technology is still largely experimental and should be further evaluated before recommendation 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 - chlorination;

Precipitative processes, including coagulation assisted microfiltration, enhanced coagulation/filtration, and enhanced lime softening;

Adsorption processes, specifically activated alumina;

Ion exchange processes, specifically anion exchange;

The next sections provide an overview of how the costs of each technology is estimated. The sections following include a brief description of the technology, design criteria, and capital and O&M cost curves for systems with design flows ranging from 0.01 to 430 mgd.

3.2   BASIS FOR COST ESTIMATES

3.2.1   Cost Modeling

Three models were used to develop costs: 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). The models were used for the following flow ranges; linear trends are used in the transition between models:

#	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

All three models rely on flows 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). Some processes (e.g., activated alumina and ion exchange) have slightly different ranges due to discrepancies between the models.

Activated alumina costs are not based on these models because they assume regeneration of the media and parallel operation of columns. Activated alumina costs in this document assume no regeneration (disposable media) and series operation of columns. Anion exchange costs are also not based on the Water or W/W models because the models are for nitrate removal with very high sulfate rather than arsenic removal in the sulfate ranges under consideration. The text summarizes the approach used for activated alumina and anion exchange costs. A more detailed description can be found in the Appendices.

3.2.2   Technology Design Panel Recommendations

Since the 1986 Safe Drinking Water Act (SDWA) reauthorization, EPA has relied mainly on the three cost models to estimate compliance costs for drinking water regulations. Following the reauthorization of the SDWA in 1996, EPA 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).

Total capital costs consist of three elements: process, construction, and engineering costs. Process costs include manufactured equipment, concrete, steel, electrical and instrumentation, and pipes and valves. Construction costs include sitework and excavation, subsurface considerations, standby power, contingencies, and interest during construction. Engineering costs include general contractor overhead and profit, engineering fees, and legal, fiscal, and administrative fees. Housing costs are specifically excluded from each of these cost category designations. Housing costs are

included in the estimates presented in this chapter, but are added to the total capital cost after application of the TDP cost approach.

To incorporate the TDP recommendations, the cost models are used to estimate total capital costs, and then steps are taken to adjust these costs. The TDP recommended that total capital cost estimates be based solely on process costs; therefore, in the first step, the process-related components of the models' estimated costs are generated using the capital cost breakdowns presented in Appendices A through C. In the second step, construction and engineering costs are then estimated using the factors in Table 3-1 to arrive at total capital costs that reflect the TDP recommendations.

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 excluding housing. Housing costs are added to the total cost.

Table 3-2 presents a sample capital cost breakdown for the VSS model for membrane equations. 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. The reports *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 to develop capital cost breakdown summaries 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 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 (ft²)						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	c
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	

*Housing costs are added to the total capital cost after application of the TDP cost approach

Table 3-4

Water Model Capital Cost Breakdown by Percentage for

Package Conventional Treatment (Coagulation/Filtration)

Cost Component	Filter Area (ft²)						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%

*Housing costs are added to the total capital cost after application of the TDP cost approach

Table 3-5

W/W Cost Model Capital Cost Breakdown for Sedimentation Basins

Cost Component	Area (A = ft²) 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 = ft²) 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 a 0.024 mgd (24 kgpd) plant the capital cost is:

   $$CAP = 0.86[24] + 41.1$$
   $$CAP = 61.74 \text{ or } \$61,740$$

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.

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

   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:

   $$\$69,250 \times 0.5697 = \$39,452$$

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

   $$\$39,452 \times 2.5 = \$98,629$$

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

3.2.3.2 Water Model

1. 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) = $376,138

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

   $376,138 x 2.5 = $940,345

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.

5. When housing costs are included for a unit process, they are added after multiplication of the process cost by the TDP total cost factor (2.5 in this example). Table 3-4 shows housing is 16.3 percent of the construction cost. The total capital cost is:

   ($692,066 x 0.1630) + $940,345 = $1,053,152

3.2.3.3 W/W Cost Model

1. The W/W Cost model output for a 1 mgd (1250 ft^2) 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

sources. The September 1998 index values used to develop 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
Building Energy Use	102.6 kWh/ft^2/yr

[1] Energy Information Administration.

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

The models present total capital costs and annual O&M costs. Annual O&M costs include the costs for materials, chemicals, power, and labor. Annual costs can be determined using the following equations:

Total annual cost (¢/kgal) =   Annualized Capital Cost (¢/kgal) + O&M Cost (¢/kgal)

Where:

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

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

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.

$$\text{Capital Recovery Factor} = \frac{(1 + i)^N}{((1 + i)^N - 1)/i}$$

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 in 1967 dollars. In 1986, the BLS conducted a comprehensive overhaul of the industrial price methodology resulting in a re-basing of all index information to a base year of 1982 (i.e., 1982=100). The BLS index information must be re-based to 1967 to use the indices in the models. A sample re-base calculation is presented below.

Sample Re-base Calculation:

Machinery =   1982 Base Factor / Re-base Factor = 1967 Base Factor

=   146.4 / 0.32895016 = **445.1**

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 flows were used for each of the flow regions, with the exception of the transition regions, where cost estimates are based on a linear regressions between the last data point of the previous region and the first data point of the following region. Table 3-11 presents the design and average flows, and cost models used in this process. Each model is used to estimate the cost for each flow. Regression analysis is then used to estimate an equation relating cost to flow for each region.

Table 3-11

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 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 methods. 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. Selection of treatments that reduce additional contaminants would generate additional benefits from the treatment technology.

Land

Land costs are not included as part of the construction and capital costs for installation of new facilities. Land costs and the amount of land needed will vary significantly from site to site. Technology selection and the footprint of the technology will also play a critical role in land needed for treatment. The November 1999 Technology and Cost Document presented ranges of costs based on the Technology Design Panel recommendations. Comments on the proposed rule provided very little documentation on land costs for compliance with arsenic. One comment from the Association

of California Water Agencies indicated that land costs could be as high as 5% of total annualized costs. This is based on land prices in California, so it represents an upper bound for the nation.

It is expected that systems with existing treatment will usually have adequate land available. Furthermore, small ground water systems are not expected to have land acquisition requirements. Intermediate to large ground water systems and large surface water systems with appreciable ground water use may need to acquire land. However, according to SDWIS, these systems comprise only about five percent of all community water systems in the United States.

Based on the above discussion, 5% of the capital costs was selected as the upper bound increase associated with land at ground water sites. This upper bound can be used in a sensitivity analysis to evaluate an upper bound impact of land on costs. Many systems may not need to purchase land for arsenic treatment.

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.

Permitting

The Very Small Systems Document, the Water Model and the W/W Model all include costs for legal, fiscal and administrative costs. These costs vary by model and system size. Legal, fiscal and administrative costs are included in the engineering cost factor. For the base case, permitting costs are covered by the legal, fiscal and administrative costs in the models.

An upper bound for sensitivity analysis was based on the Technology Design Panel recommendation of permitting costs as 3% of construction value (EPA, 1998). Participants in the TDP also recommended a floor of $2500 for systems serving less than 10,000 people. The upper bound that can be used to assess sensitivity of the costs to permitting is an additional 3% of the capital costs rather than 3% of the construction value. This approach will yield a higher cost for permitting.

Piloting

The Technology Design Panel stated that pilot testing needed to be part of EPA's high cost models or upper bound estimates (EPA, 1998). Members of the large system breakout group indicated that large systems will need to conduct pilot tests in all cases for treatment optimization. The costs in this document assume that pilot testing is covered by the existing capital costs due to the conservative assumptions used to develop these costs. The technology costs presented in this chapter do not assume optimized design and operation at each site. Systems which incur piloting costs will generally achieve more optimized, lower cost, technology applications.

An upper bound for piloting was also estimated. In other cost analyses, EPA has assumed that piloting costs would be approximately 3% of the total capital costs for a technology. The sensitivity analysis can use 3% of the capital costs as an upper bound.

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 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).

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

There are redundant items included in the cost equations for several of the technologies in Chapter 3. For example, redundant columns were included in the costs for anion exchange and activated alumina. Thus, redundancy is included for the major cost component in the design for these

two technologies. Redundancy for the lower cost components would be covered by the conservative assumptions used in deriving the capital costs.

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 for water treatment facilities in those States. Systems may incur additional costs if additional redundant items are required by the State.

3.4   COSTS FOR MULTIPLE REMOVAL PERCENTAGES

Capital and O&M cost estimates are presented for the maximum achievable removal in this document. Table 3-12 presents a removal technology matrix which identifies maximum removal percentages for the technologies for which costs have been estimated. Costs for facilities requiring less than the maximum removal to meet the arsenic MCL target, can be estimated using the blending approach discussed in Section 3.4.2.

3.4.1   Removal and Accessory Costs

Both the April 1999 and the November 1999 drafts of the Technology and Cost Document included costs for accessories in Appendix D. Accessory costs included 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. Inclusion of accessory costs was not based on a recommendation of the Technical Design Panel. A detailed examination of Section 3.11 of the November 1999 Technology and Cost Document revealed the apparent source of the accessory costs. Section 3.11 compared the costs in the November 1999 Technology and Cost Document with the costs from the 1993 Technology and Cost Document (Malcolm Pirnie, 1993a) and *Evaluation of Central Treatment Options as Small System Compliance Technologies* (SAIC, 1999). There is a statement in bold at the end of Section 3.11.1 that states that the 1993 estimates are greater than the 1999 estimates for many technologies. It further states that this is because the 1993 estimates included accessory costs; i.e., raw and finished water pumping and clearwell storage.

The 1993 Technology and Cost Document was examined to determine why the accessory costs were included for each technology. After examining the costs in this document, accessory costs, such as clearwell storage, were only included for one technology - coagulation/filtration. This is understandable because coagulation/filtration would not be installed solely for arsenic removal. This technology would likely be selected by surface water systems, especially since the surface water treatment rule (SWTR) had only promulgated only recently before the 1993 Technology and Cost Document was prepared. Surface water systems would need the clearwell storage for disinfection credit under the SWTR. The costs for lime softening in the 1993 Technology and Cost Document specifically state that clearwell storage is not included. Costs were also not included for anion exchange or activated alumina. **Therefore**, **cost estimates presented in this document do not include accessory capital and O&M.**

Table 3-12

Treatment Technology Maximum Achievable Removal Percentages

Treatment Technology	Maximum Percent Removal
Coagulation/Filtration	95
Enhanced Coagulation/Filtration[1]	95
Coagulation Assisted Microfiltration	90
Lime Softening (pH > 10.5)	90
Enhanced Lime Softening[1] (pH > 10.5)	90
Ion Exchange (sulfate < 50 mg/L)	95
Activated Alumina	95
Reverse Osmosis	>95
Greensand Filtration (20:1 iron:arsenic)	80
POU Activated Alumina	90
POU Ion Exchange	90

1 - Enhanced processes assume the existing plant can achieve 50 percent removal without modification. Process enhancements result in the balance to achieve the maximum removal. For example, an existing coagulation/filtration facility can achieve 50 percent removal. Process enhancements result in an additional 45 percent removal, for a total removal of 95 percent.

3.4.2   Use of Blending in Cost Estimates

Capital and O&M costs were estimated using the VSS, Water, and W/W Cost models. If raw water contaminant levels are sufficiently low, a utility may not need to achieve maximum removal to achieve a treatment goal. For example, assume a facility is considering installation of a coagulation filtration which can achieve 95 percent removal. If the raw water arsenic concentration is 20 $\digamma$g/L and the treatment objective is a finished water concentration of 10 $\digamma$g/L, the utility need only remove 50 percent of the arsenic in the raw water. In this scenario, the facility could treat a portion of the raw water and blend with untreated water and still achieve its treatment objective. The portion of the total process flow to be treated can be 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. Multiplying that fraction by the total plant flow will result in design and average operating flows that can be used to estimate capital and O&M costs for the treated portion of the flow, using the graphs and equations presented in this Section.

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 on 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. Potassium permanganate costs can be calculated for dosages of 10 mg/L. The *Very Small Systems Best Available Technology Cost Document* (Malcolm Pirnie, 1993) can be used for calculating costs for the flows below 1 mgd. For flows greater than 10 mgd the W/W Cost Model can be used to estimate the capital and O&M costs. Linear regressions can be used to estimate costs in the transition regions between the two models, i.e., 1 to 10 mgd. The following are some of the system design criteria that will affect costs when using the two models:

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 percent potassium permanganate solution.

The VSS document makes provisions for building (42.7 percent), fencing (49.4 percent), and road (33.8 percent) costs associated with potassium permanganate addition. It should be assumed that permanganate addition will be installed as part of a larger treatment process and that building, fence and road costs for the treatment facility would be adequate to accommodate permanganate addition.

O&M costs for very small systems can be calculated using equations in the very small systems cost document. Labor requirements should assume 1 hour per week of incremental labor.

For small system potassium permanganate addition, a dry chemical feed system capable of 1,000 pounds per day can be used.

The costs associated with potassium permanganate are in the costs of each of the treatment technologies considered below.

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 mg/L. Similar to potassium permanganate systems, the VSS Model was used for calculating costs for the flows below 1 mgd, and the W/W Cost Model was used to estimate the capital and O&M costs for flows greater than 10 mgd. Linear regressions were used to estimate costs in the transition regions between the models, i.e., 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 percent sodium hypochlorite feed stock.

> \#   The VSS Model makes provisions for building (52.2 percent), fencing (60.5 percent), and road (41.4 percent) costs associated with chlorine addition. Housing costs are only needed when the arsenic removal process is a centrally-managed point-of-use option. Otherwise, it is assumed that chlorination process would be housed in the same building as the larger central treatment process.

> \#   Capital costs were calculated for both with and without housing costs added.

> \#   Incremental labor requirements for O&M costs were assumed to be 1 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 3-1 and 3-2.

Figure 3-1
Pre-oxidation - 1.5 mg/L Chlorine
Capital Costs

$y = 15290x^{0.6554}$

$y = 6065.8x + 8494.2$

With Housing $y = 14560$

$y = 6875x + 401.97$

Without Housing $y = 7277$

Capital Costs ($)

Design Flow (mgd)
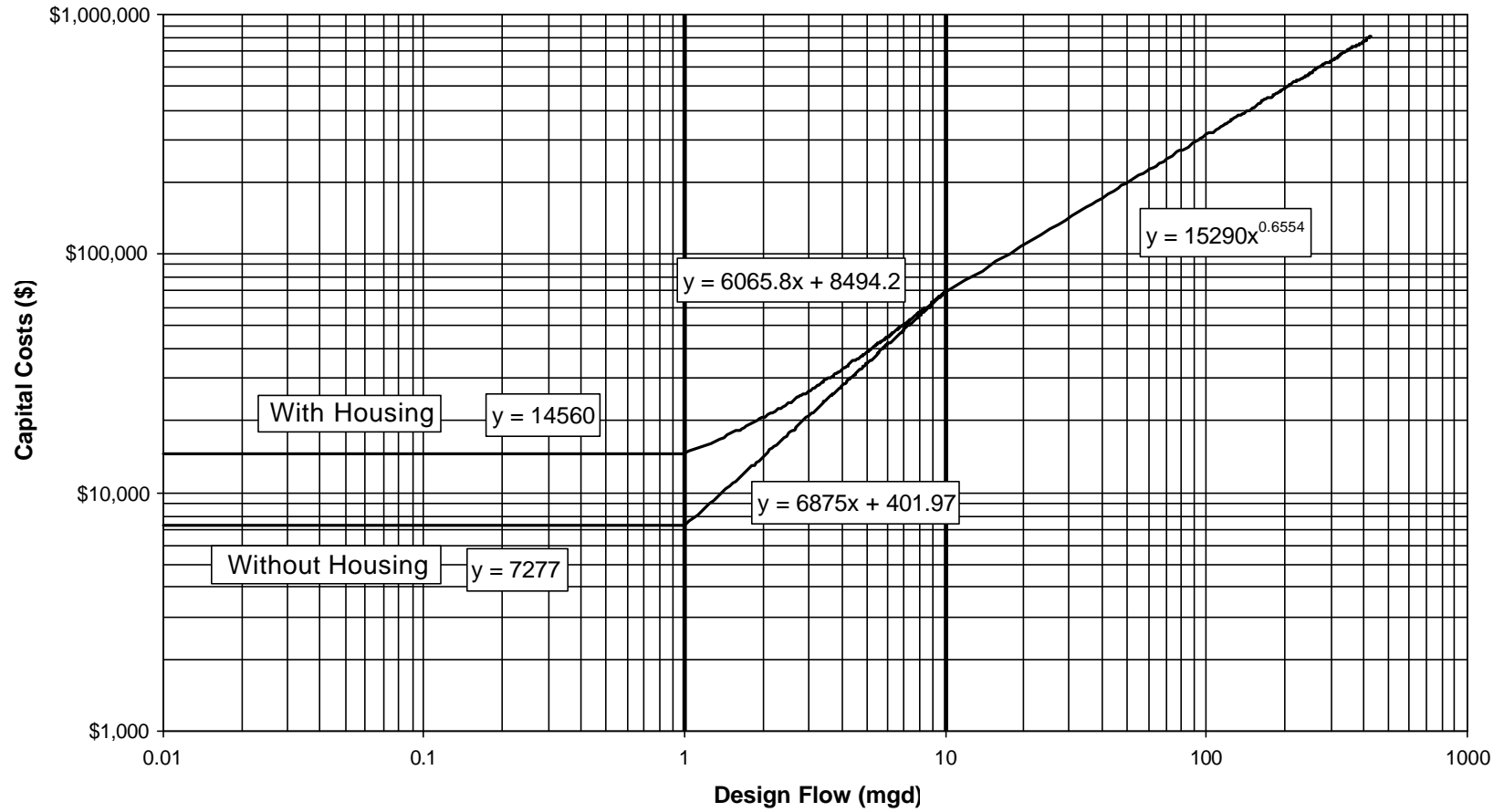
Figure 3-2
Pre-oxidation - 1.5 mg/L Chlorine
O&M Costs

$y = 4920.1x + 916.87$

$y = -0.3928x^2 + 823.35x + 19360$

$y = -1E-10x^2 + 4239.7x + 1161.8$
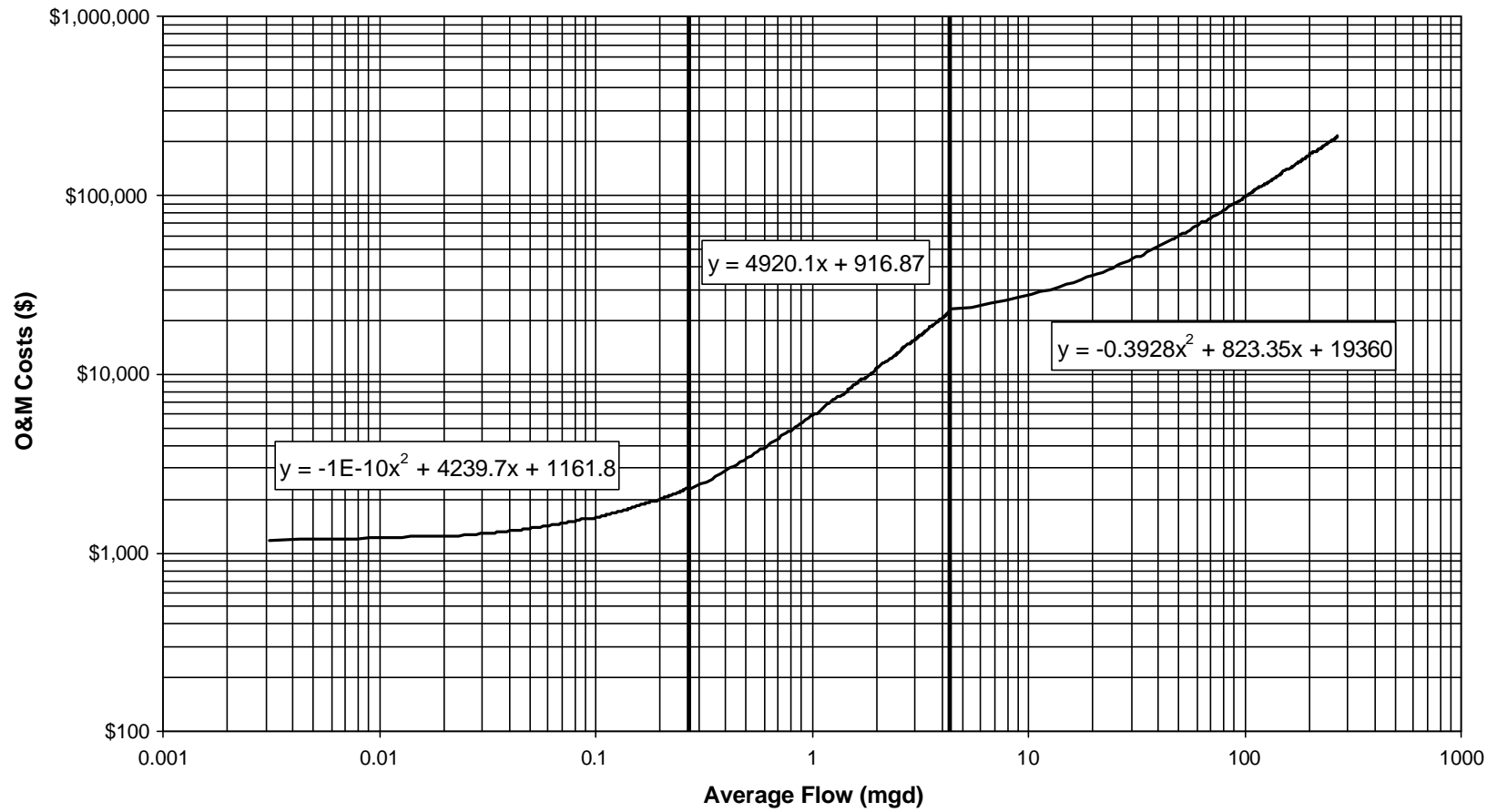
3.6   PRECIPITATIVE PROCESSES

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. C/F is unlikely to be installed solely for arsenic removal. Systems considering installation of this technology should design the process to operate in the optimal pH range if high removal efficiency is needed for compliance. 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. Costs are not presented for this option. Key design parameters that can be used with the cost models are presented for systems that wish to explore this option.

Very Small Systems (Less than 0.10 mgd)

Capital costs for very small systems can be 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 percent), fencing (2.1-7.5 percent) and road (1.2-4.7 percent) costs associated with each of the technologies presented. The following design criteria can be used to develop capital cost estimates for systems with a design flow of less than 0.10 mgd:

- #   Coagulant dosage, ferric chloride, 25 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 can be 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 can be 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; and
- Lime dose, 25 mg/L for pH adjustment.

Large Systems (Greater Than 1 mgd)

The W/W Cost model can be used to develop capital and O&M cost estimates for large C/F plants. The following design criteria can be 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;
- Rapid mix, 1 minute;
- Flocculation, 20 minutes;
- Sedimentation, 1000 gpd/ft^2 using rectangular tanks; and
- Dual media gravity filters, 5 gpm/ft^2.

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 is 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 (see 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 enhancement. It was also assumed that the only added O&M burden would result from power and materials costs, **no additional labor was assumed to be required.** Costs presented are for the enhancement only, and are in addition to any current annual debt incurred by the utility.

Small Systems (Less than 1 mgd)

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.

Large Systems (Greater Than 1 mgd)

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 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 3-3 and 3-4 present capital and O&M cost curves and equations for enhanced coagulation.

Figure 3-3
Enhanced Coagulation/Filtration
Capital Costs

$y = -6.7115x^2 + 6482.1x + 799902$

$y = 94000x - 75947$

$y = -3825.4x^2 + 14734x + 7144.5$
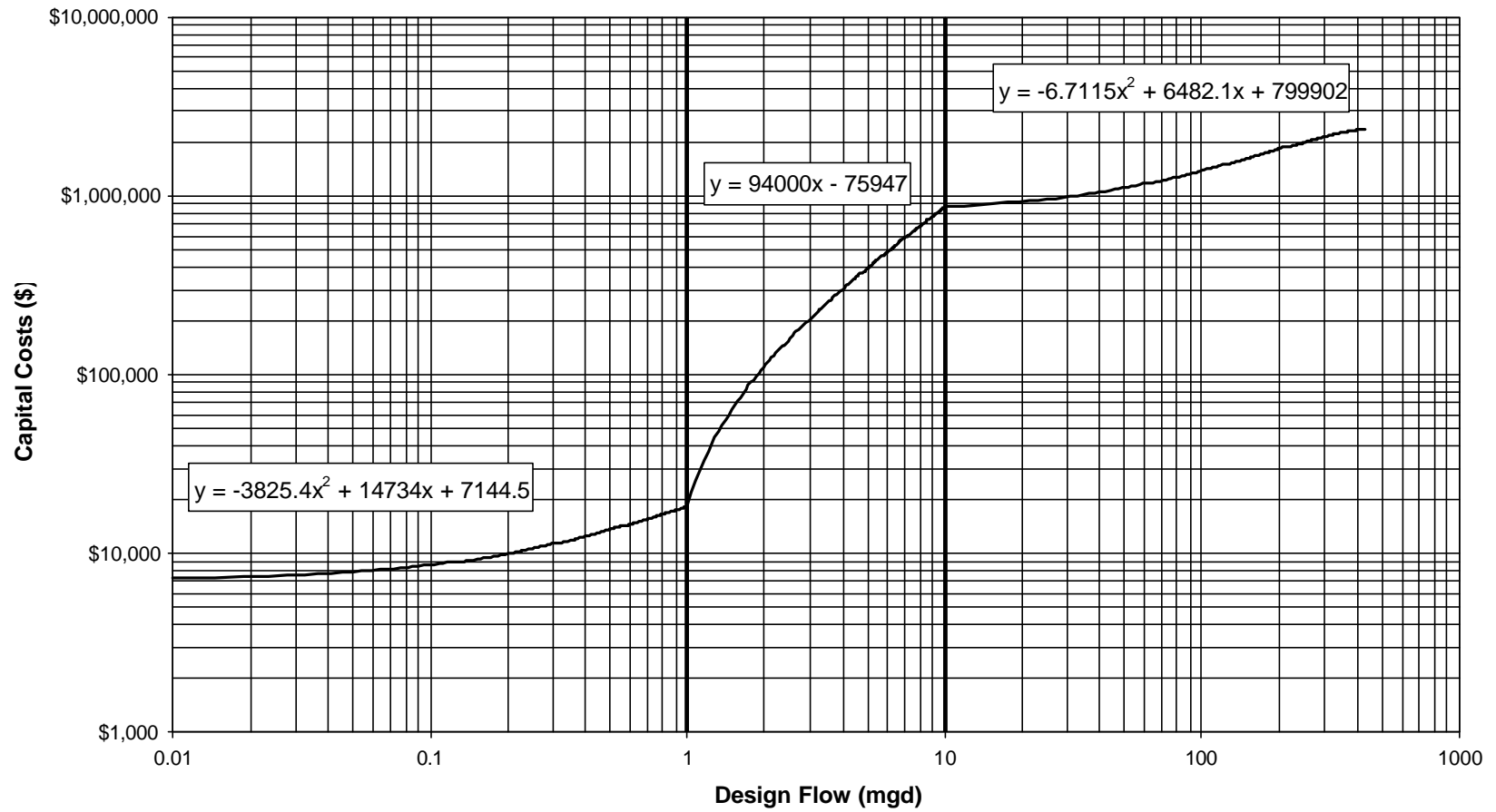
Figure 3-4
Enhanced Coagulation/Filtration
O&M Costs
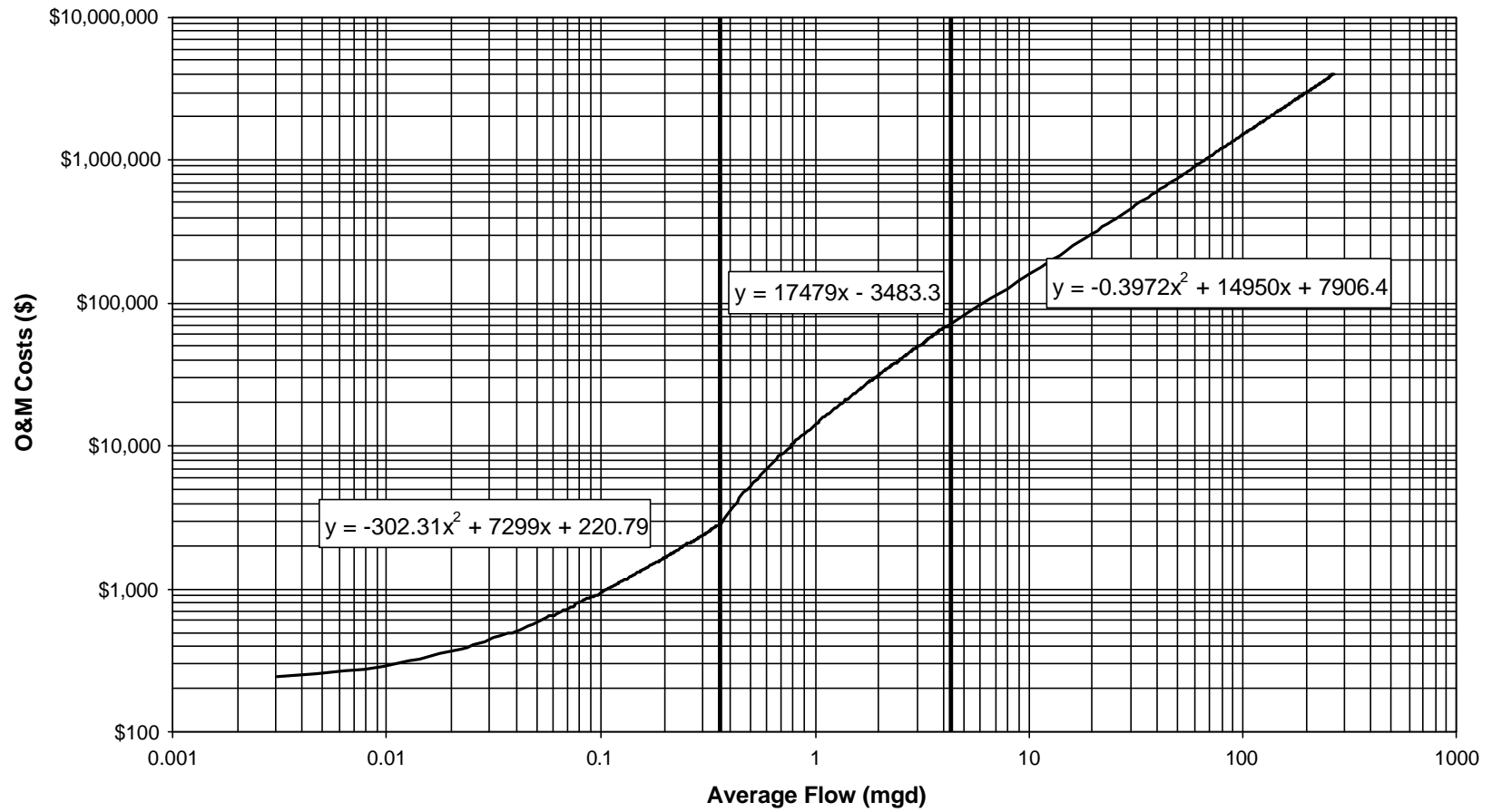
3.6.3 Direct Filtration

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 Section 3.5. Direct filtration is unlikely to be installed solely for arsenic removal. The technology may not be able to achieve high removals of arsenic. Design criteria are not presented for 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 coagulation, flocculation, sedimentation, and microfiltration. 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 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. The C/F design parameters given in Section 3.6.1 were used here as well, however no polymer addition was assumed due to the tendency of some membranes to foul when used with polymer. Very small system costs also include a 20 mg/L sodium hydroxide dose to adjust process pH. Microfilter specifications and cost estimates were developed based on vendor quotes and case studies. These costs were then added to the 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. 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, 25 mg/L;
- \# Sodium hydroxide dose, 20 mg/L; 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. The following design criteria were used to estimate capital and O&M costs for the coagulation portion of this process:

- \# Ferric chloride dose, 25 mg/L;
- \# Rapid mix, 1 minute;
- \# Flocculation, 20 minutes;
- \# Sedimentation, 1000 gpd/ft^2 in rectangular basins; and
- \# Standard microfilter specifications, provided by vendors.

Figures 3-5 and 3-6 present capital and O&M cost curves and equations for removal of arsenic by coagulation assisted microfiltration.

Figure 3-5
Coagulation Assisted Microfiltration
Capital Costs

$y = 1030810x + 1067733$

$y = -483591x^2 + 2308991x + 273143$

$y = 320x^2 + 921471x + 2129119$

$y = 2343199x + 228653$

$y = -11935465x^2 + 4880036x + 94324$

Capital Costs ($)

Design Flow (mgd)
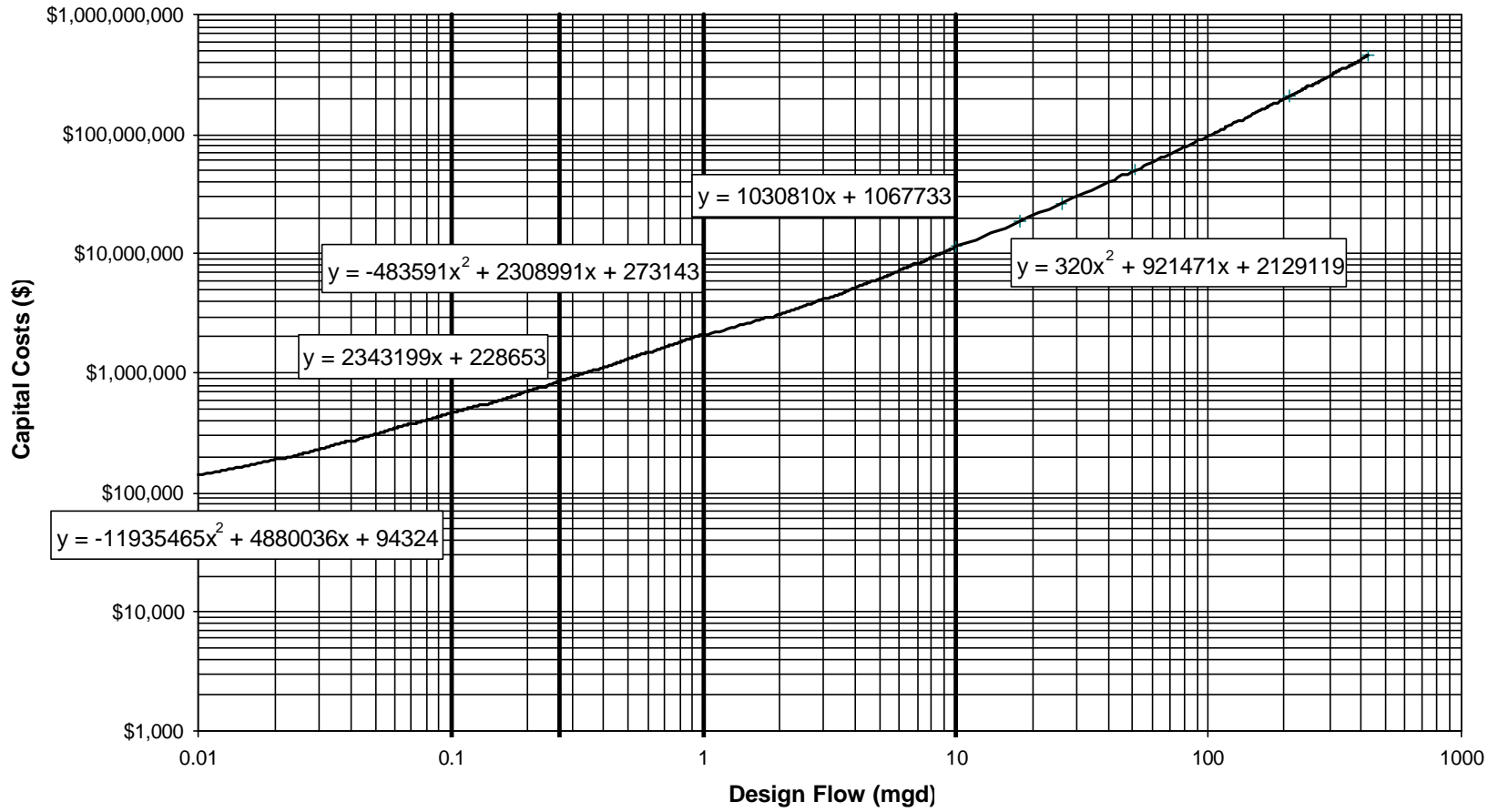
Figure 3-6
Coagulation Assisted Microfiltration
O&M Costs

y = 13311x + 51014

y = 80081x + 26977

y = 15236x + 42350

y = 136332x + 22139

y = 196829x + 20264

O&M Costs ($)

Average Flow (mgd)
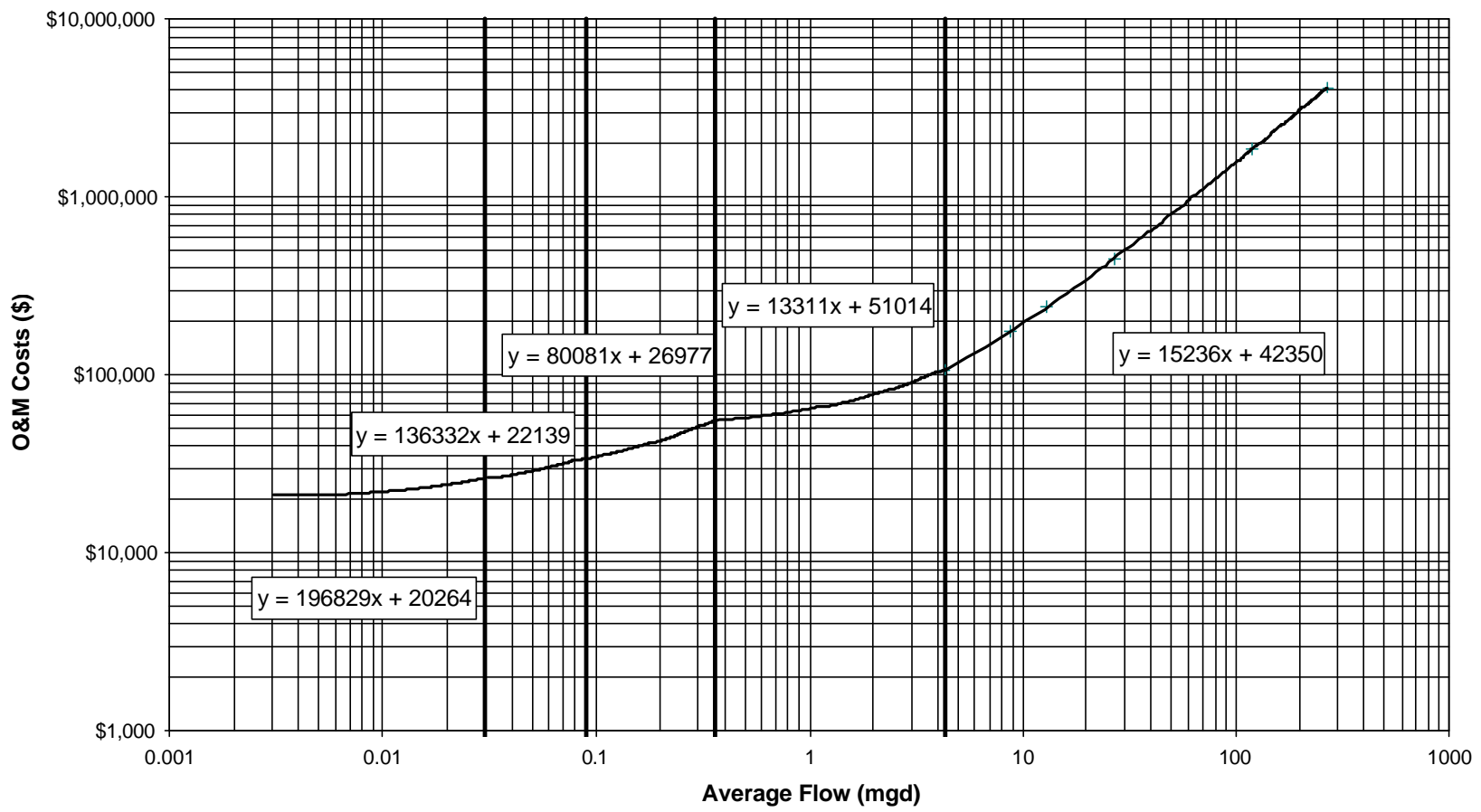
3.6.5 Lime Softening

Lime softening (LS) has been widely used for reducing hardness in large water treatment systems. The LS removal mechanism is discussed 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 Section 3.5.

Lime softening is unlikely to be installed solely for arsenic removal. Systems considering installation of this technology should design the process to operate in at a pH > 10.5 if high removal efficiency is needed for compliance. Costs are not presented for this option. Key design parameters that can be used with the cost models are presented for systems that wish to explore this option.

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 would have to be used to estimate capital and O&M costs for all LS treatment plants with less than 1 mgd capacity. The following design criteria can be used in the development of capital and O&M cost estimates:

- \# Package plant for all small systems;
- \# Lime dose, 250 mg/L; and
- \# Carbon dioxide (liquid), 35 mg/L for recarbonation.

Large Systems (Greater Than 1 mgd)

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

- \# Lime dose, 250 mg/L;
- \# Carbon dioxide (liquid), 35 mg/L for recarbonation;
- \# Rapid mix, 1 minute;
- \# Flocculation, 20 minutes;
- \# Sedimentation, 1500 gpd/ft^2 using circular tanks; and
- \# Dual media gravity filters, 5 gpm/ft^2.

3.6.6   Enhanced Lime Softening

Enhanced LS involves modifications to the typical LS treatment process to adjust the pH > 10.5 by 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 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. It was also assumed that the only added O&M burden would result from power and materials costs, **no additional labor was assumed to be required.** 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 were 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 3-7 and 3-8 present cost curves and equations for removal of arsenic by enhanced lime softening.

Figure 3-7
Enhanced Lime Softening
Capital Costs

$y = -7.539x^2 + 26610x + 969164$

$y = 133486x - 100351$

$y = -22974x^2 + 48211x + 7897.7$

Capital Costs ($)

Design Flow (mgd)
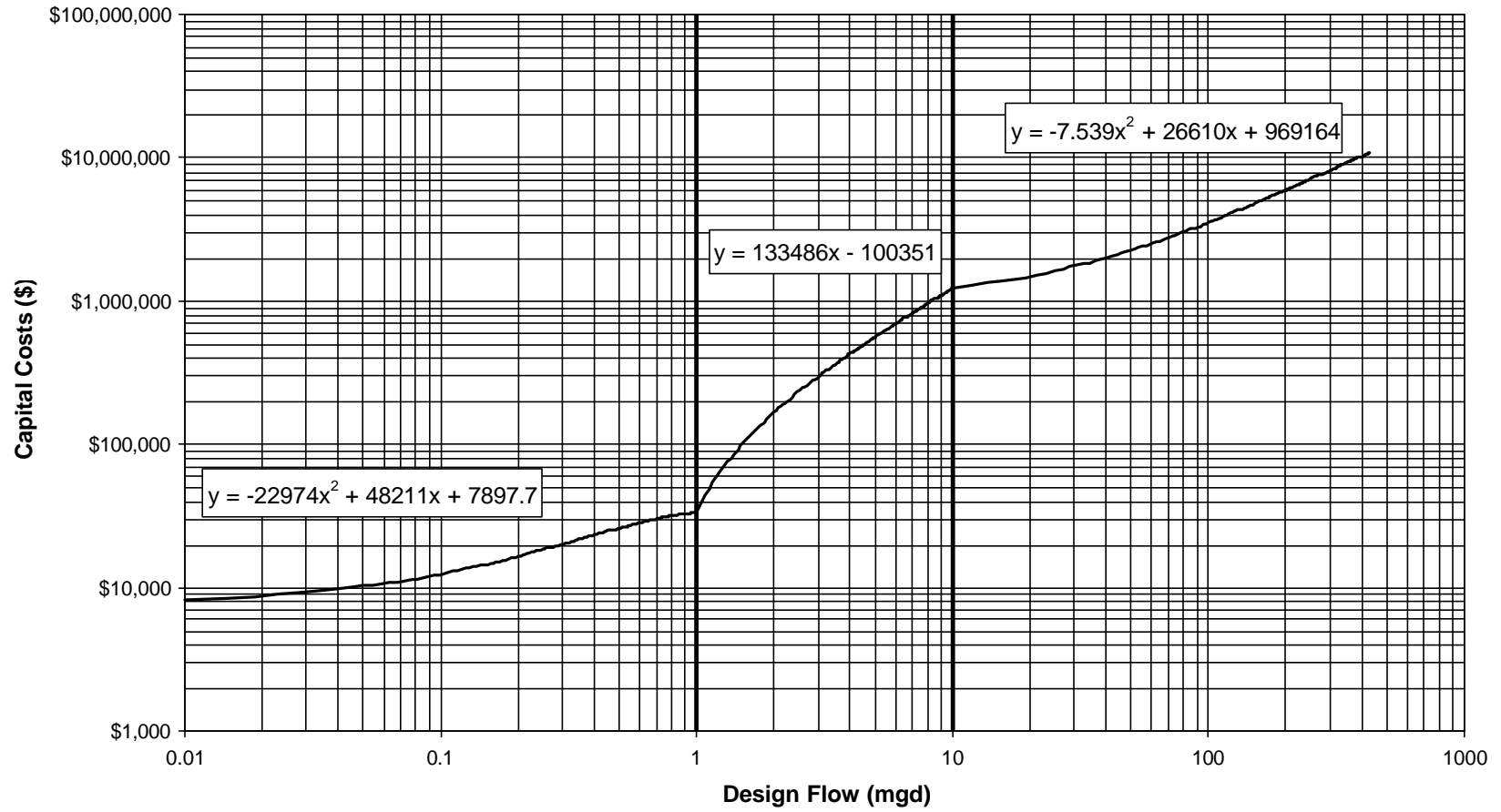
Figure 3-8
Enhanced Lime Softening
O&M Costs

Chart showing O&M Costs ($) versus Average Flow (mgd) on a log-log plot with the following fitted equations:

$y = 2144.7x^2 + 30525x + 319.67$

$y = 29145x + 1094.3$

$y = -0.4528x^2 + 26067x + 14956$
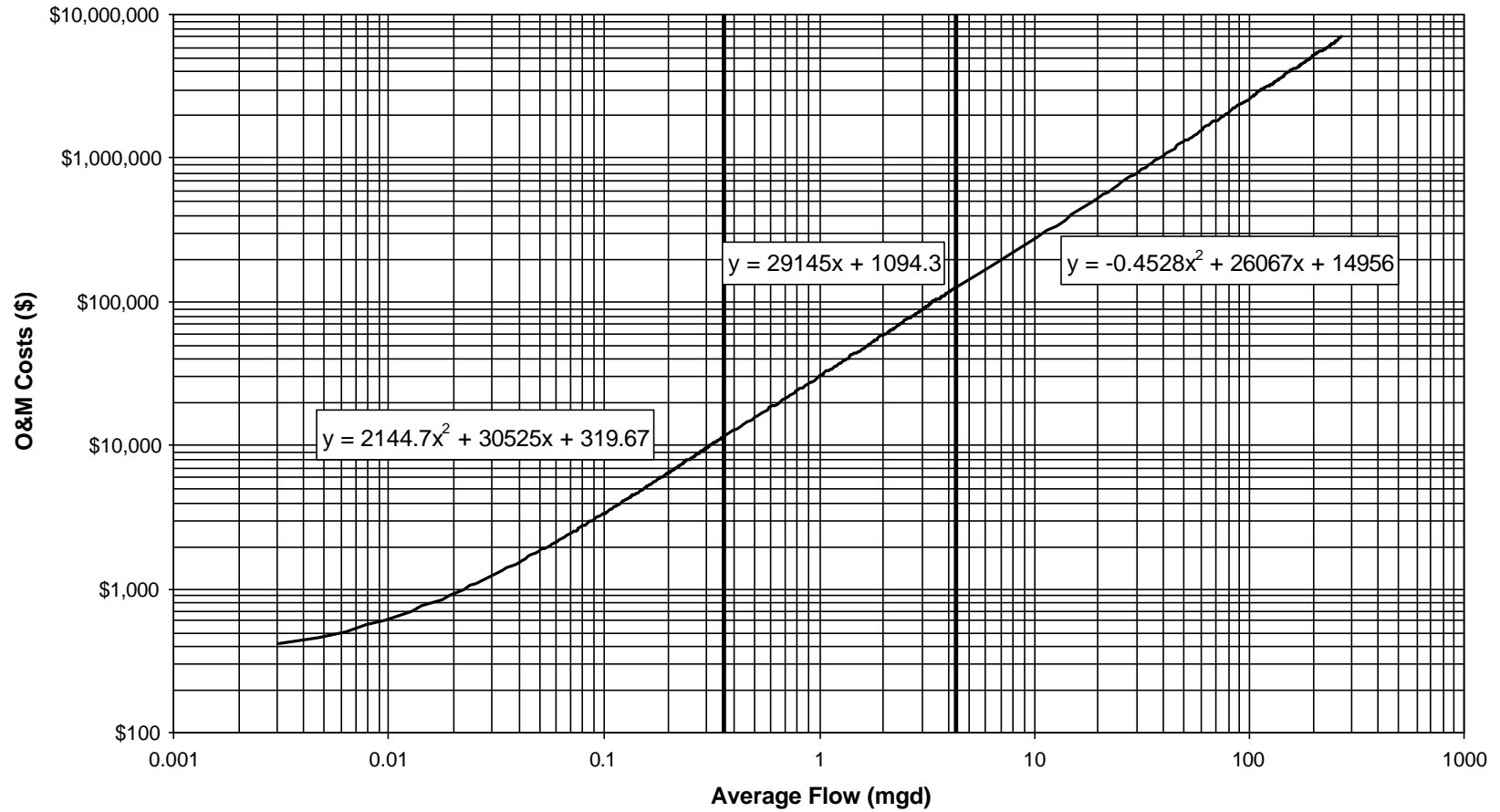
3.7   ADSORPTION PROCESSES

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. Disposable activated alumina (no regeneration) costs were developed using the following approach. Four sets of costs were developed - two for unadjusted pH and two where the pH has been adjusted to the optimal pH of 6. The costs for pH adjustment to optimal pH 6.0 are included as a separate module in these cost estimates since it is only used for two of the options.

The three models routinely used to estimate unit costs are all based on regeneration of the media, operation at optimal pH, and use of single columns. Most of the costs are based on fluoride removal rather than arsenic removal. It was determined that the existing cost models could not be used to estimate costs for disposable activated alumina. To estimate costs, it was assumed that the alumina would be arrayed in small columns operated in series rather than in one larger column. This will provide greater utilization of the media before disposal and is more consistent with the other designs using disposable activated alumina.

Design Assumptions: Capital Costs with Natural pH

The following assumptions were made to estimate the capital cost of activated alumina when the water is not adjusted for pH. The basis for these assumptions are provided in Appendix D.

1.   The pH will not be adjusted to operate the process at the optimal pH between 5.5 and 6.0. The activated alumina process will be operated at the natural pH of the system to simplify the process and avoid potential problems with lowering the pH. Two pH ranges will be evaluated 7.0 #pH < pH 8.0 and 8.0 # pH #8.3.

2.   The pH will not need to be adjusted after the activated alumina process. The activated alumina process should not adversely affect the finished water pH. Those systems with high natural water pH will not need a new pH adjustment process. Those systems with lower natural water pH values will continue to rely on the existing corrosion control process.

3. Empty Bed Contact Time (EBCT) is 5 minutes per column

4. The estimated cost of the activated alumina media is $0.82/lb for systems with a design flow greater than 0.1 mgd. The estimated cost of activated alumina media is $1.30/lb for systems with a design flow less than or equal to 0.1 mgd.

5. The density of the activated alumina media is assumed to be 47 lb/ft^3.

6. The bed depth assumption ranged from 3 to 6 feet, depending on the design flow. The maximum diameter per column was 12 feet. When the calculated media volume exceeded the maximum, the treatment media was divided into several beds to maintain a realistic bed diameter. Thus, larger designs include multiple activated alumina trains, each treating a portion of the total flow.

7. The vessel cost has been sized based on 50 percent bed expansion during backwash even though backwashing may not be necessary on a routine basis for smaller systems. The vessel volume was calculated as (1.5)(media volume). The vessel cost is given by the following equation:

$$Cost = 63.288 * (Vessel\ Volume\ in\ gallons) \wedge 0.679$$

8. The capital costs include a redundant column to allow the system to operate while the media is being replaced in the old roughing column.

9. The remaining components of the process equipment include concrete, pipe and valves and electrical and instrumentation. A cost factor of 14.43 percent of the manufactured equipment and activated alumina was used for these other construction cost components.

10. The capital costs have been estimated from the total process costs using a factor of 2.5 for small systems and 3.33 for large systems.

11. The capital costs estimated in the previous step do not include the cost of a building to house the process equipment and other add-on costs related to the site. Housing costs were assumed for all sites. Fence and road were only assumed for those systems that have no treatment. Table 3-13 shows the assumptions made for this analysis regarding the percentage of systems with no treatment in place. (See Appendix D for further information on these percentages.)

Table 3-13

Assumed Percentage of Ground Water Systems with No Treatment

Population Category	Percentage of GWS w/o any Treatment
25 - 100	43%
101 - 500	19%
501 - 1,000	16%
1,001 - 3,300	18%
3,301 - 10,000	18%
10,001 - 50,000	18%
50,001 - 100,000	18%
100,001 - 1 MIL	18%

12. The equation for process area, which is the key variable in estimating the costs for buildings, roads, and fences has been based on the design flow of the system. The equation used for process area is:

Process Area = 726.6 * Design Flow + 59

13. The building cost to house the activated alumina system is $45.38/sq ft for design flows at or below 0.65 mgd. The following equation is used for design flows between 0.65 and 1 mgd:

Building cost = 54.204 * Process Area + 14887

For design flows greater than 1 mgd, the building cost varied based on flow from $100 to $90 per square foot.

Design Assumptions: Capital Costs with pH Adjustment

For two options, the water pH is adjusted to the optimal pH of 6. The capital cost equation for the pH adjustment module, which includes post-treatment corrosion control is given by the following equation (Appendix D provides additional information):

Capital cost = 40517 * Design Flow + 31889

Design Assumptions: O&M Costs with and without pH Adjustment

There are three major components to the operation and maintenance (O&M) costs - replacement media, energy and labor.

1. The critical variable in calculating the replacement media costs is the run length - the number of bed volumes until the effluent concentration exceeds a specific concentration. The run length for the natural pH range of $7 \# pH < 8$ is 10,000 bed volumes (BV). The run length for the natural pH range of $8 \# pH \# 8.3$ is 5,200 BV. These run lengths are used to estimate the replacement of the activated alumina in half of the active columns.

2. The O&M costs were calculated at 2 run lengths when pH is adjusted to pH of 6. The two run lengths that were selected for the optimal pH of 6 are 15,400 and 23,100 BV.

3. Chemical costs also are incurred to adjust the pH. The following equation was used for chemical costs for pH adjustment:

Chemical cost (pH adjustment) = 81361* Average Flow + 151.7

4. The second major component in the O&M costs is incremental labor cost. Incremental labor is the labor associated with the additional maintenance that comes with a new process.

Adjustment in staffing or shifting of activities are not included in the incremental labor. The incremental labor for the activated alumina process without regeneration is one hour per week. In addition, time is also included for media replacement (at 5 percent breakthrough) depending on the volume of media being replaced. It was assumed that sixteen hours are required to replace the media for design flows less than 1 mgd and thirty-two hours are required for larger design flows. Incremental labor for the pH adjustment processes also is included in the two options operated at optimal pH. The incremental labor is an additional two hours per week.

5. The labor rate for small systems was $28/hour. The loaded labor rate for large systems was $52/hour.

6. The third component in the O&M costs is the energy costs for both the building and the process. Housing electrical energy use is based on an annual usage of 19.5 kwh/sq ft/yr, which includes lighting, ventilation, and heating. Process electrical energy was assumed to be 440 kwh/yr. Costs were based on a rate of $0.08/kwh.

   The capital and O&M cost curves for activated alumina are shown in figures 3-9 through 3-14.

Figure 3-9
Activated Alumina
Capital Costs

$y = 420286x + 10212$

$y = 515309x + 10214$

Capital Costs ($)

Design Flow (mgd)
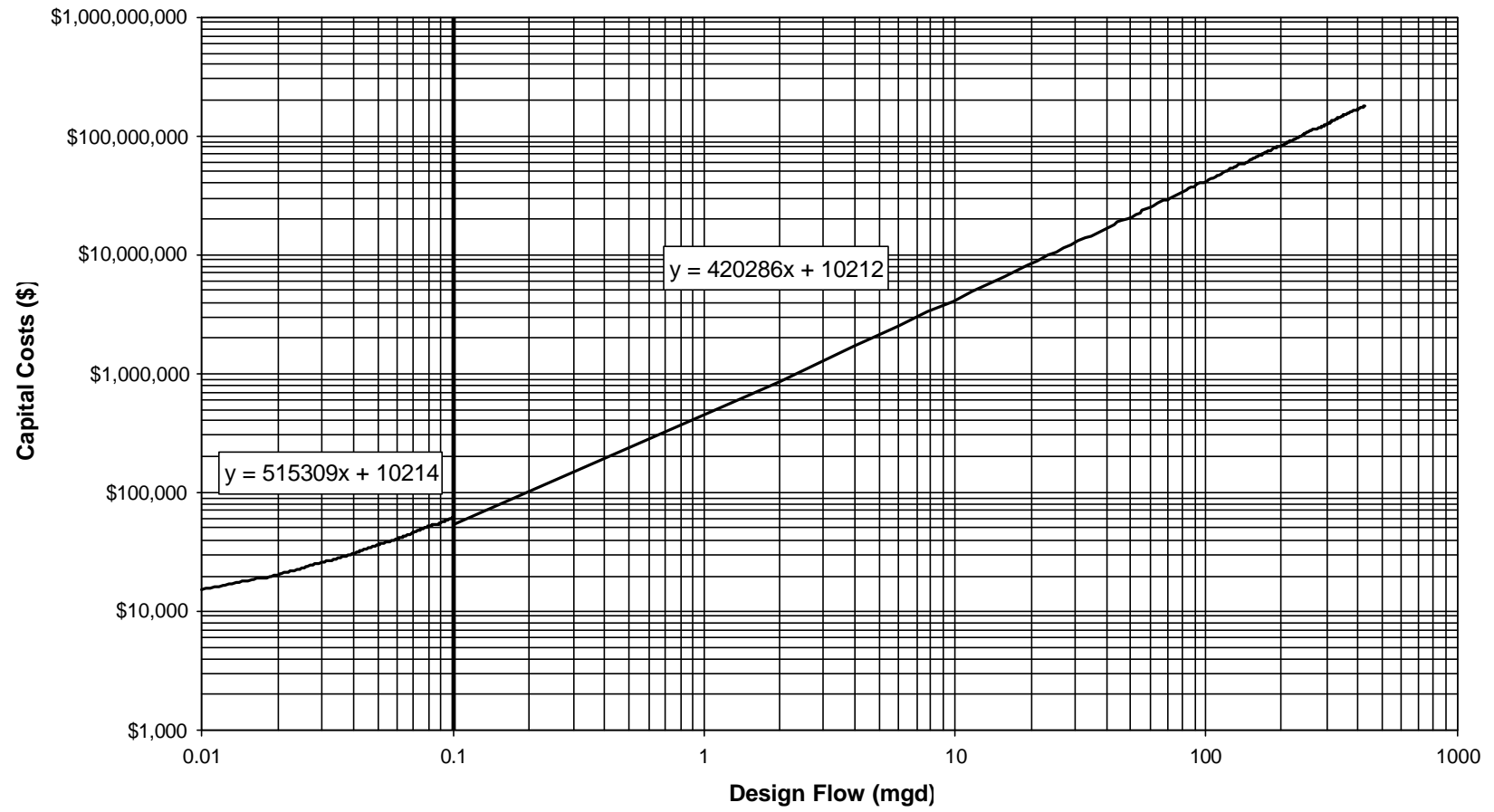
Figure 3-10
Activated Alumina (pH 7 - 8)
O&M Costs

$y = 190726x + 9798.2$

$y = 188890x + 4122.7$

O&M Costs ($)

Average Flow (mgd)
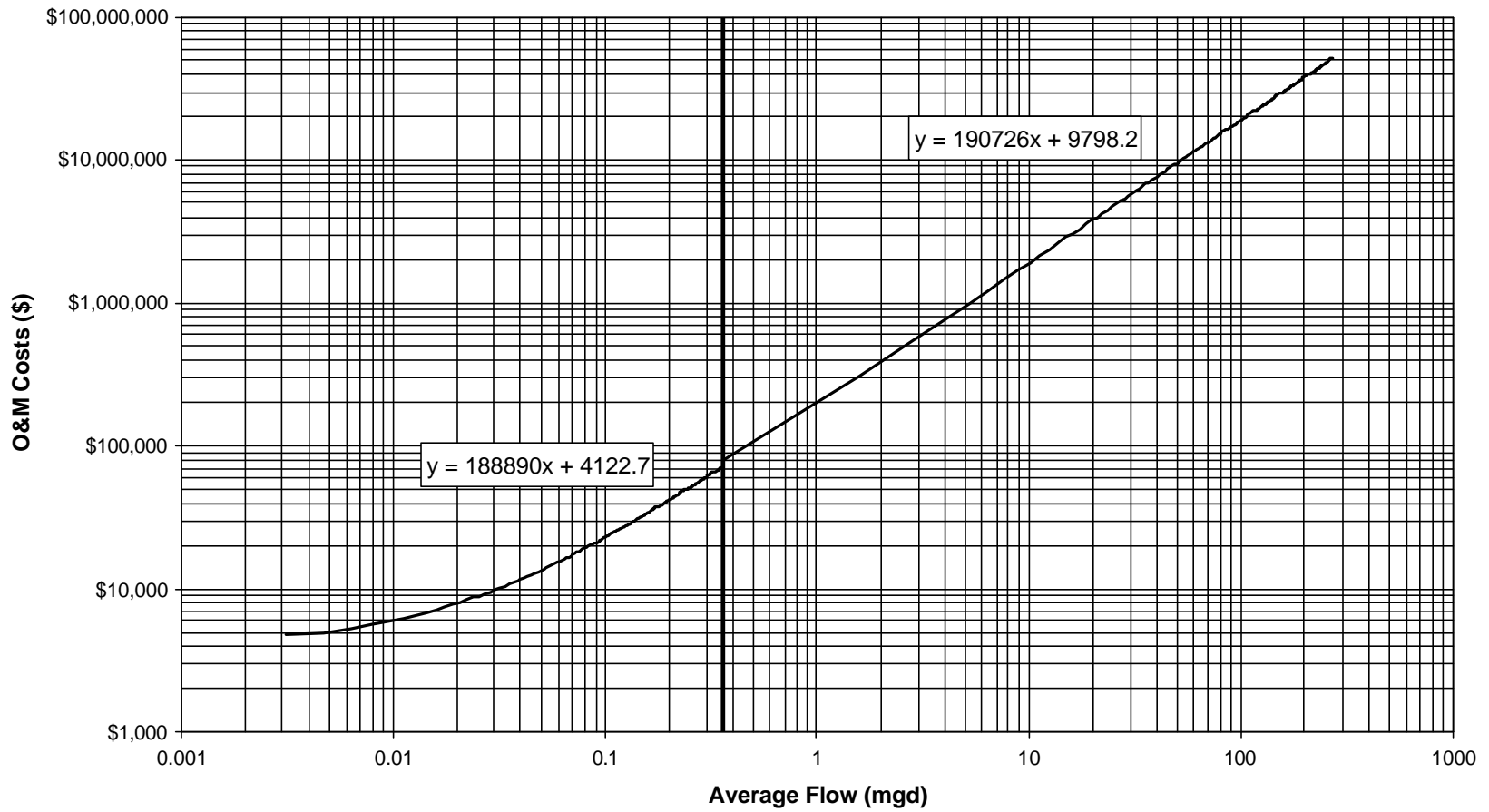
Figure 3-11
Activated Alumina (pH 8 - 8.3)
O&M Costs

$y = 364399x + 15765$

$y = 359695x + 6388.6$

O&M Costs (

Average Flow (mgd)
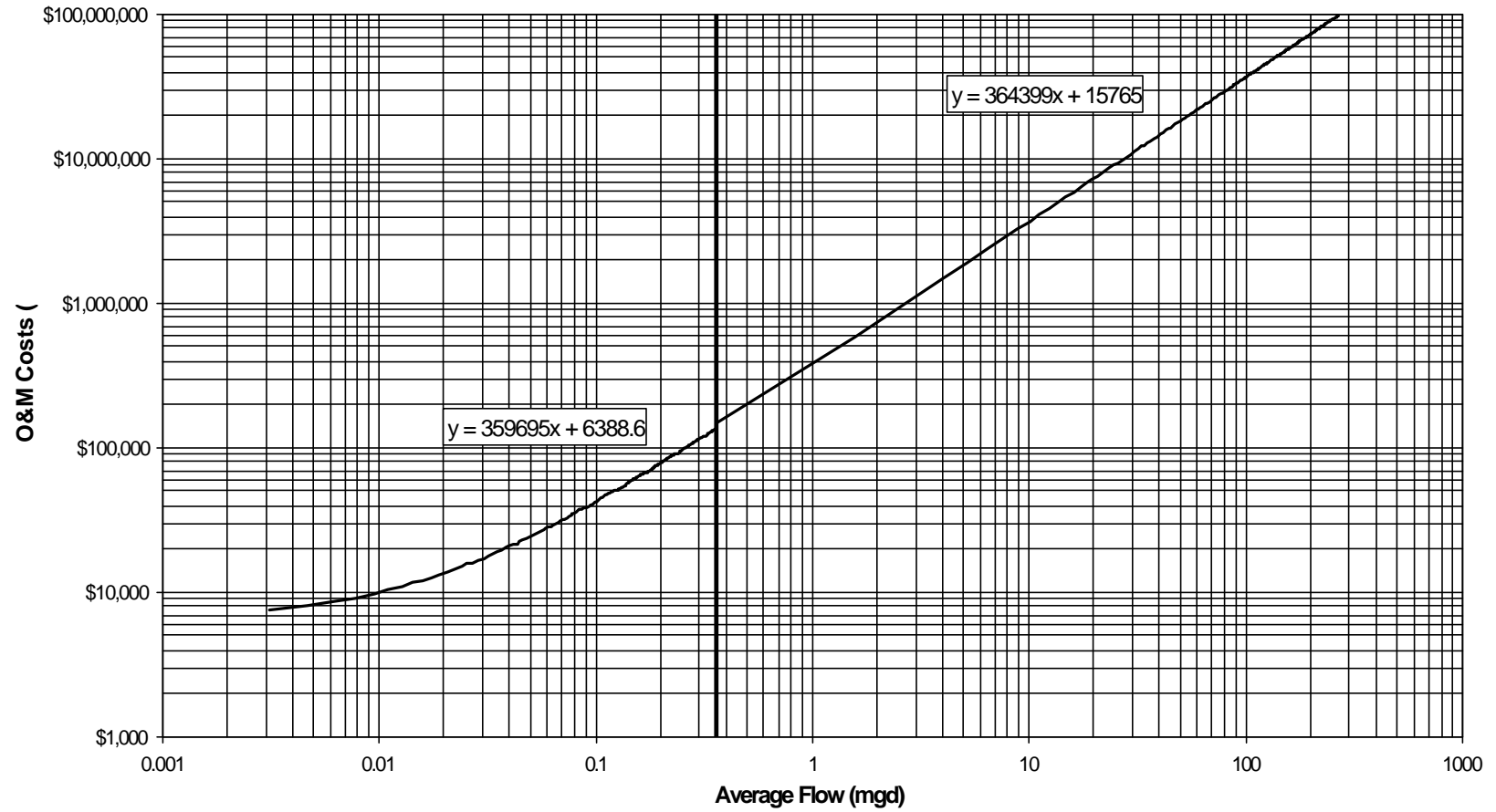
Figure 3-12
Activated Alumina (pH Adjusted to 6.0)
Capital Costs
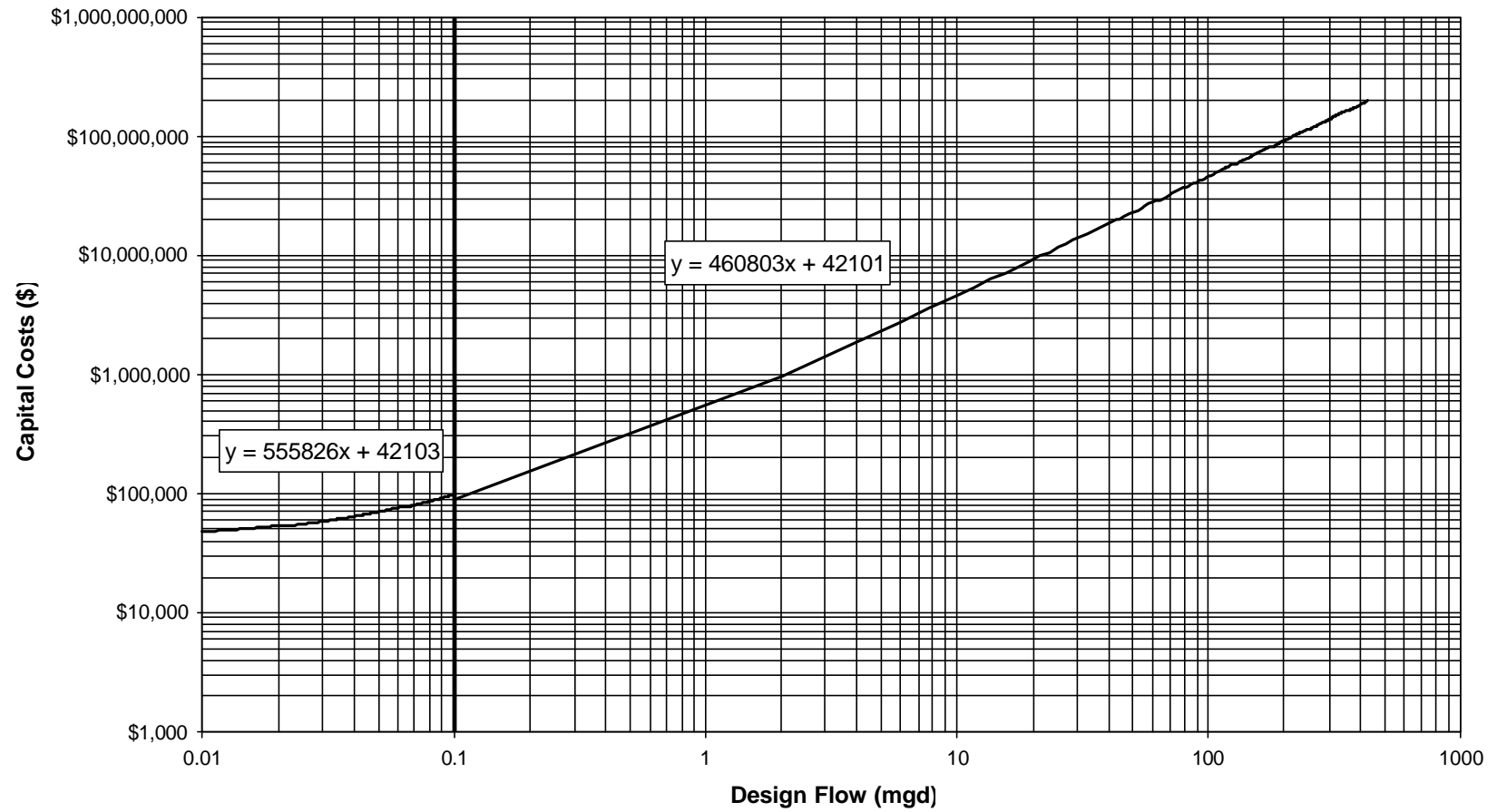
Figure 3-13
Activated Alumina (pH Adjusted to 6.0 - 23,100 BV)
O&M Costs
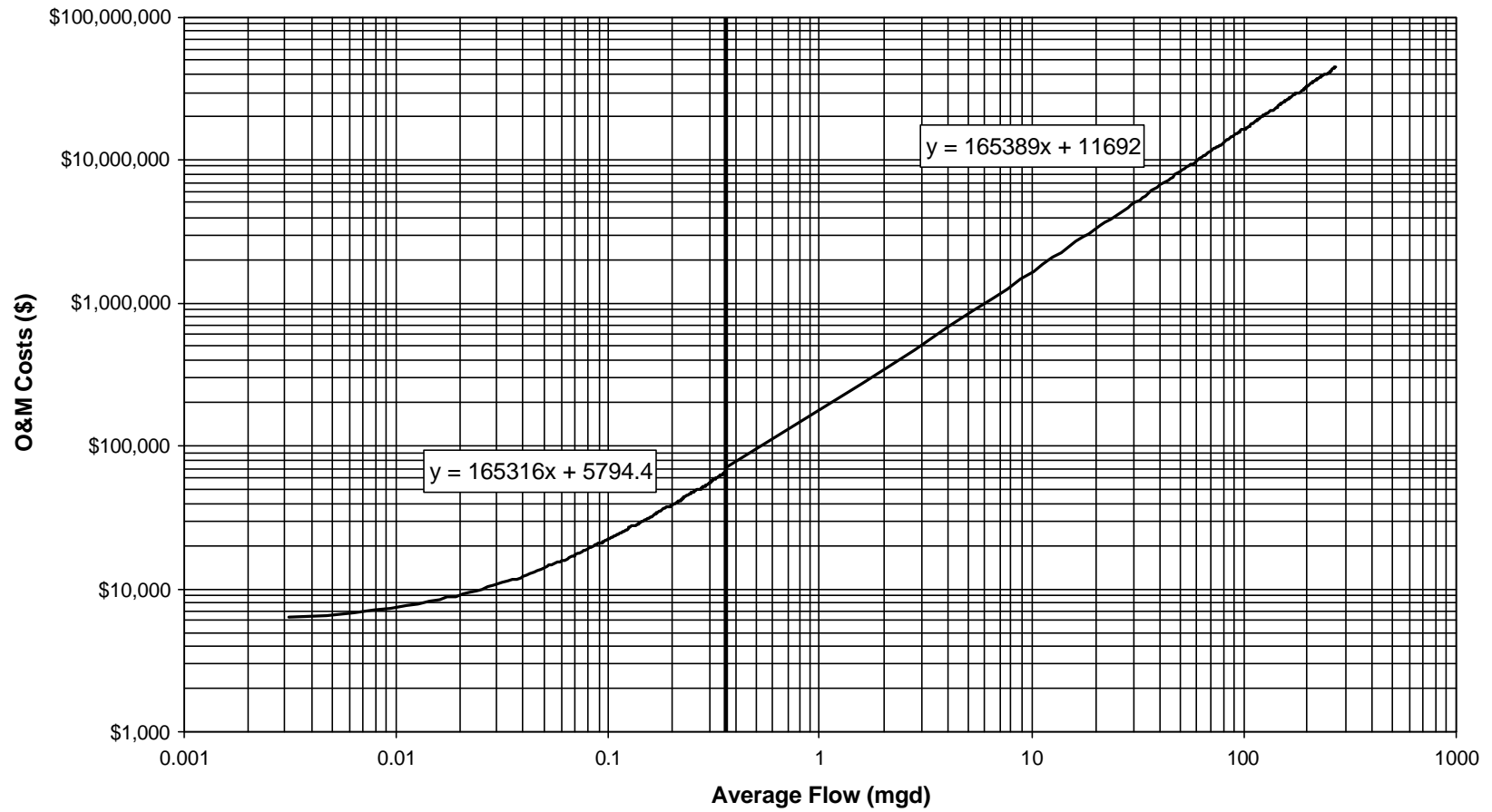
Figure 3-14
Activated Alumina (pH Adjusted to 6.0 - 15,400 BV)
O&M Costs

y = 206113x + 13092

y = 205368x + 6325.7

O&M Costs ($)

Average Flow (mgd)
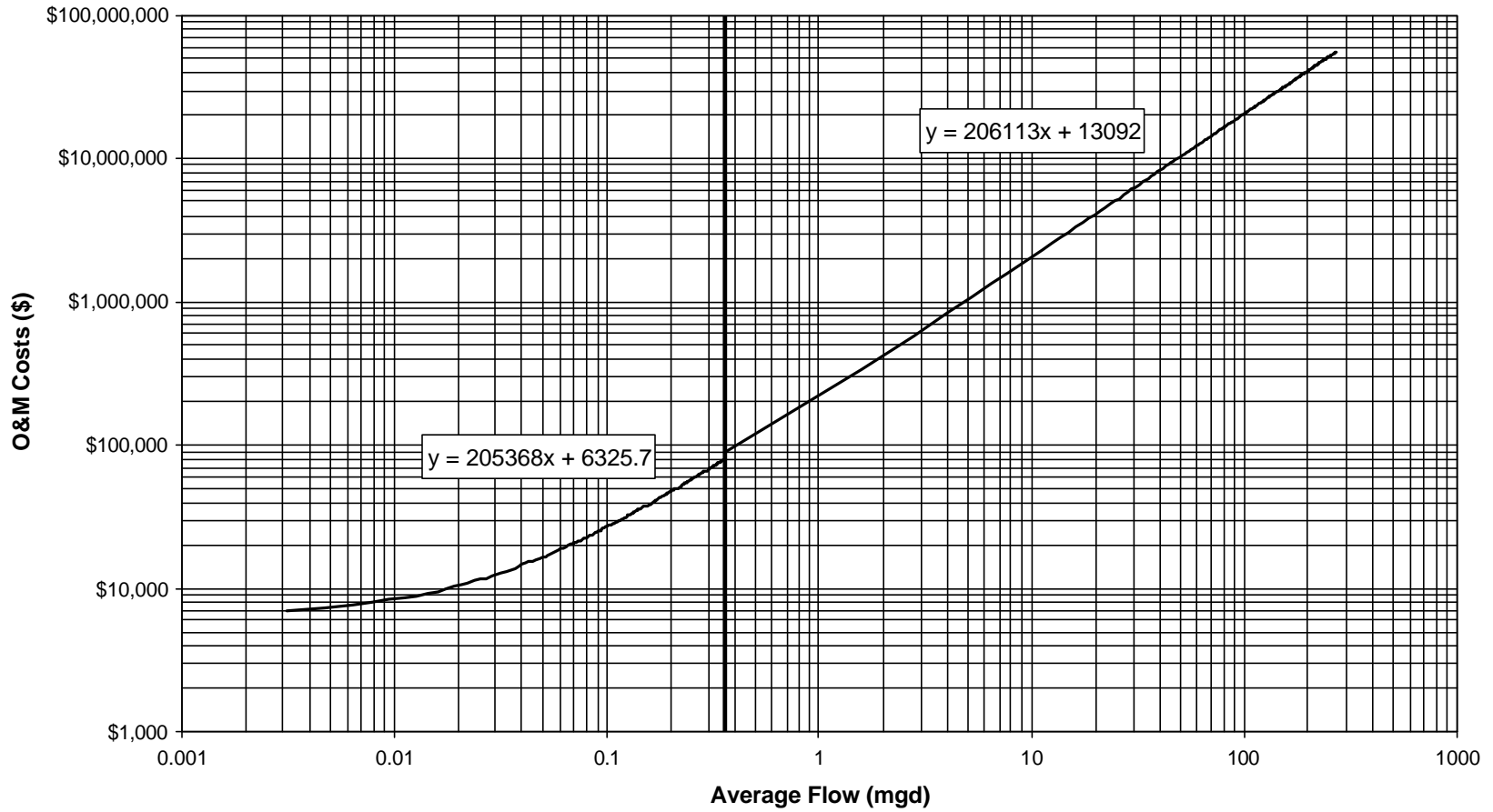
3.7.2 Granular Ferric Hydroxide

Granular ferric hydroxide is a technology that may allow for very long run length without the need for pH adjustment or pre-oxidation. It has been demonstrated for full-scale arsenic removal in England (Simms, et. al., 2000). It was not designated a BAT because of the lack of published data showing performance for a range of water qualities. Also, there is little published cost data, so costs are difficult to evaluate and are not presented.

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 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 Section 3.5. Neither pre-oxidation nor pre-filtration have been considered as part of the costs developed in this section. It is further assumed that corrosion control measures are not required because the pH is not significantly affected by the process, especially when blending can be utilized.

Sulfate concentrations in the influent water significantly affect the capacity of the IX resin with respect to the removal of arsenic. Clifford (1995) estimated bed volumes for 10 percent and 50 percent breakthrough of influent arsenic as a function of influent sulfate concentration. Figure 3-15 shows the bed volume and sulfate relationship estimated by Clifford (1995). Using this figure, the regeneration frequency for an IX column can be estimated for the entire range of sulfate concentration based on the upper bound concentration. This is a very conservative assumption and underestimates run length for systems with lower sulfate concentrations in the range. The run length until regeneration is the critical factor in estimating the O&M cost. Capital costs will be marginally affected by changes in bed volume to regeneration.

The costs in both the April 1999 Technology and Cost Document (*1*) and the November 1999 Technology and Cost Document (*2*) utilize models based on nitrate removal. The Water Model was one of the cost models used for small systems with design flows less than 1 mgd. Sulfate concentrations for this model were between 50 and 100 mg/L. The November 1999 Technology and Cost Document listed the design criteria for large systems (> 1 mgd) using the W/W model. The sulfate concentration was 80 mg/L and the nitrate concentration was 100 mg/L. It was not possible to modify the sulfate concentration in either of these two models. It was determined that the existing cost models could not be used to estimate costs for anion exchange options with sulfate at or below 50 mg/L.

Design Assumptions: Capital Costs

The following assumptions were made to estimate the capital cost of anion exchange. The basis for these assumptions are provided in Appendix E.

1. Empty Bed Contact Time (EBCT) = 2.5 minutes per column

2. The estimated cost of the anion exchange resin was $125/ft^3 for all system sizes.

3. The bed depth assumption ranged from 3 feet to 6 feet depending upon the design flow. The maximum diameter per column was 12 feet. When the calculated media volume exceeded the maximum, the treatment media was divided into several beds to maintain a realistic bed diameter. Thus, larger designs include multiple anion exchange columns, each treating a portion of the total flow.

4. The vessel cost has been sized based on 50% bed expansion during backwash. The vessel volume was calculated as (1.5)(media volume).

5. The vessel cost is based on the following equation:

    Cost = 63.288 * (Vessel Volume in gallons) ^ 0.679

6. The capital costs include a redundant column to allow the system to operate while the media is being regenerated in the other column.

7. The other equipment cost elements are the brine tank, brine pump and mixer. Costs for the brine tank were based on 30-day storage. Separate costs were prepared for the two sulfate ranges based on the number of regenerations per month.

8. The remaining components of the process equipment include concrete, pipe and valves and electrical and instrumentation. A cost factor of 25.02% of the manufactured equipment and activated alumina was used for these other construction cost components.

9. The capital costs have been estimated from the total process costs using a factor of 2.5 for small systems and 3.33 for large systems.

10. The capital costs estimated in the previous step do not include the cost of a building to house the process equipment and other add-on costs related to the site. Housing costs were assumed for all sites. Fence and road were only assumed for those systems that have no treatment. Table 3-13 (above) shows the assumptions made for this analysis regarding the percentage of systems with no treatment in place.

11. The equation for process area, which is the key variable in estimating the costs for buildings, roads, and fences has been based on the design flow of the system. The equation used for process area is:

Process Area = 118.7 * Design Flow + 118

12. The building cost to house the anion exchange system is $45.38/sq ft for design flows at or below 0.65 mgd. The following equation is used for design flows between 0.65 and 1 mgd:

Building cost = 54.204 * Process Area + 14887

For design flows greater than 1 mgd, the building cost varied based on flow from $100 to $90 per square foot.

Design Assumptions: Operation and Maintenance (O&M) Costs

There are major components to the operation and maintenance (O&M) costs are regeneration frequency, regeneration dose, and incremental labor. See Appendix E for more detail.

1. The critical variable in calculating the regeneration frequency is the run length - the number of bed volumes until the effluent concentration exceeds a specific concentration. The run length when sulfate is at or below 20 mg/L is 1500 bed volumes (BV).   The run length when sulfate is between 20 and 50 mg/L sulfate is 700 BV.

2. The regeneration dose is also a critical element for the O&M costs. The salt dose for regeneration was 10.2 lb/ft^3.

3. The second major component in the O&M costs is incremental labor cost. Incremental labor is the labor associated with the additional maintenance that comes with a new process. Adjustment in staffing or shifting of activities are not included in the incremental labor. The incremental labor for the anion exchange is one hour per week plus three hours per regeneration.

4. The labor rate for small systems was $28/hour. The loaded labor rate for large systems was $52/hour.

5. O&M Costs were calculated using the data on regeneration dose, regeneration frequency, labor hours per week and adjusted labor rates using the equation in the Very Small Systems Document.

The capital and O&M cost curves for anion exchange are shown in figures 3-16 through 3-19.

Figure 3-15
Bed Volumes to Arsenic Breakthrough as a Function
of Sulfate Concentration (Clifford, 1995)

Upper curve is based on BV to 50% breakthrough of arsenic. Lower curve is based on BV to 10% breakthrough (90% removal) of arsenic.

Both curves are based on trace levels of arsenic (< 1 mg/l) contamination in world average water containing 20 mg/l chloride, and 200 mg/l bicarbonate. Sulfate concentration is variable.

Bed Volumes to Arsenic Breakthrough

Sulfate Concentration (mg/l)

Figure 3-16
Anion Exchange (< 20 mg/L SO4)
Capital Costs

$y = -6278.7x^2 + 319169x + 36758$

$y = 344581x + 19546$

Capital Costs ($)

Design Flow (mgd)
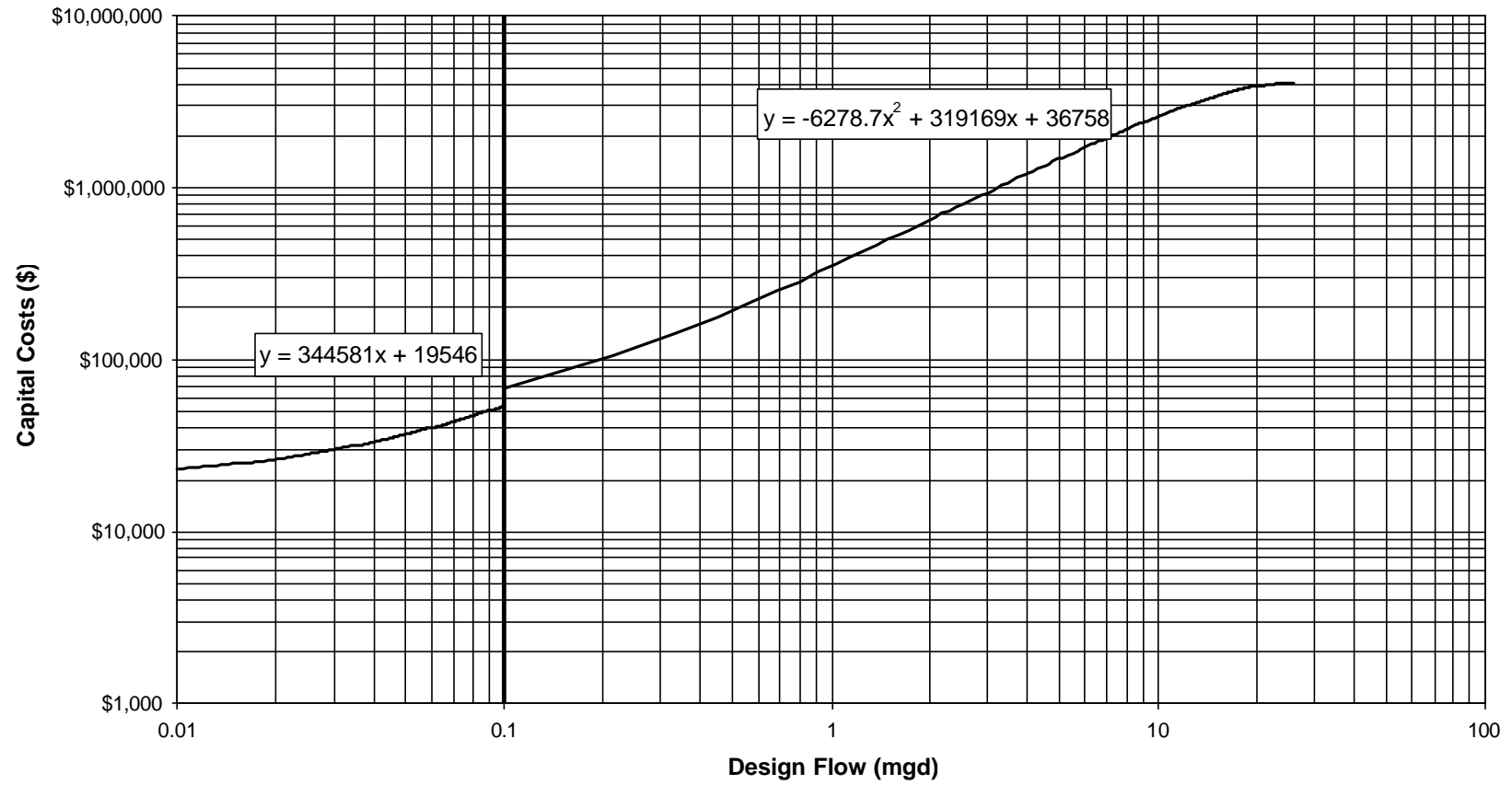
Figure 3-17
Anion Exchange (< 20 mg/L SO4)
O&M Costs

$y = -1695.5x^2 + 37350x + 16532$

$y = -67837x^2 + 77544x + 4997.4$

O&M Costs ($)

Average Flow (mgd)
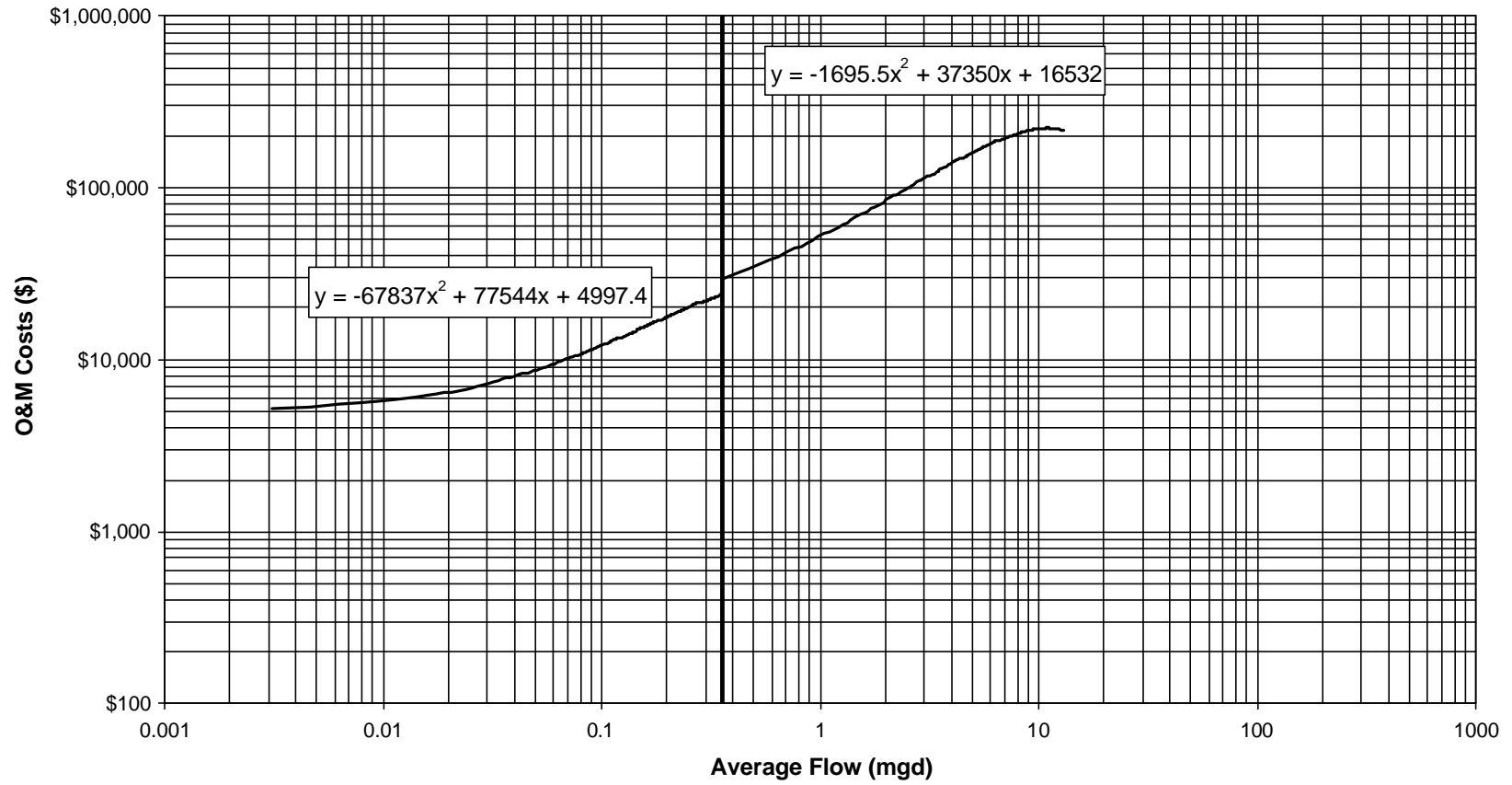
Figure 3-18
Anion Exchange (20-50 mg/L SO4)
Capital Costs

$y = -9756.1x^2 + 373847x + 73320$

$y = 454220x + 19546$

Capital Costs ($)

Design Flow (mgd)
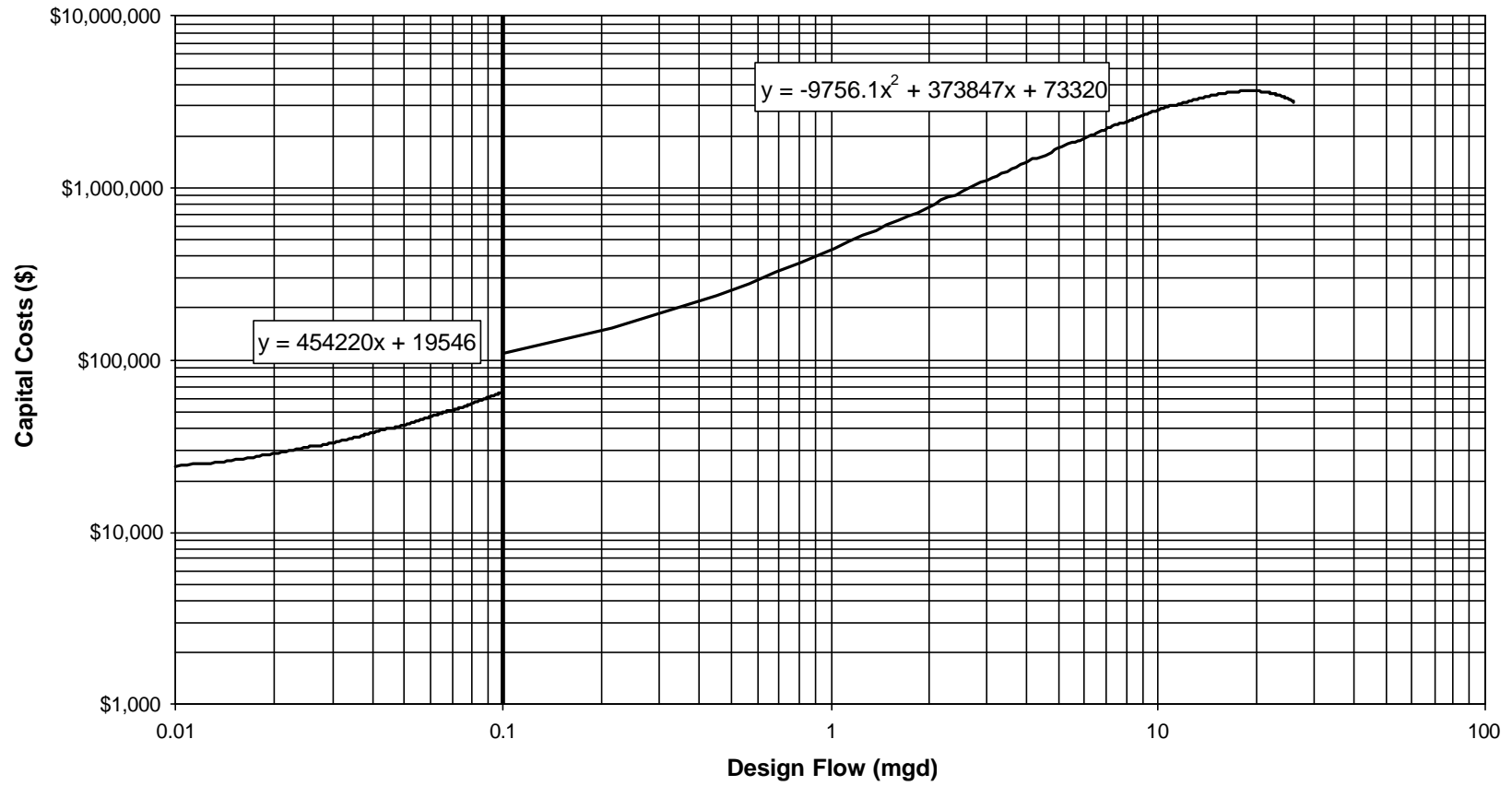
Figure 3-19
Anion Exchange (20-50 mg/L SO4)
O&M Costs

$y = -1843.1x^2 + 48269x + 24614$

$y = -82812x^2 + 94848x + 8449.9$

O&M Costs ($)

Average Flow (mgd)
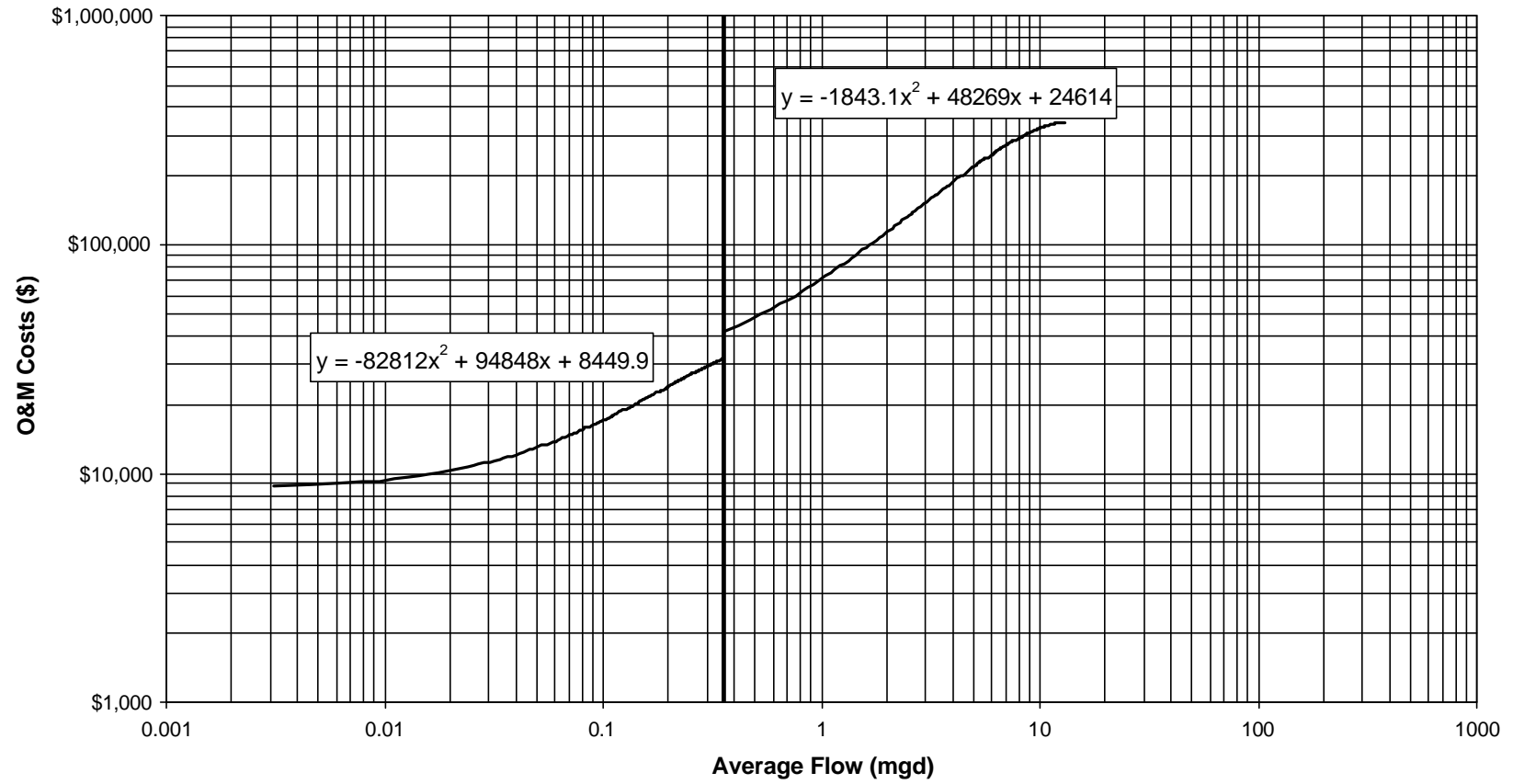
3.9   SEPARATION PROCESSES

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-mm prefiltration. Because MF is not an effective stand-alone technology for removal of arsenic, capital and O&M cost estimates for MF are not provided in this chapter.

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 200-mm prefiltration. UF has the benefit of being lower in both capital and O&M costs than high-pressure membrane processes. Because UF is not an effective stand-alone technology for removal of arsenic, capital and O&M cost estimates for UF are not provided in this chapter.

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. Due to decreased removal efficiency when operated at higher recoveries, NF is not yet demonstrated to be a reliable treatment for arsenic. Because NF is not an effective stand-alone technology for removal of arsenic, capital and O&M cost estimates for NF are not provided in this chapter.

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. Costs are not provided for RO because other options are more cost effective and have much smaller waste streams. RO may be cost effective if removal of other contaminants is needed and water quantity is not a concern.

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 can be reduced by 50 percent. The membrane replacement portion of the O&M costs can also be 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. The model also makes an assumption that recovery is 80 percent for systems of 1 to 10 mgd, and 85 percent for systems larger than 10 mgd.

3.10   GREENSAND FILTRATION

Greensand filtration is an oxidation filtration process that has demonstrated effectiveness for the removal of arsenic. The greensand filtration medium 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. 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 VSS model was used for estimating greensand filtration capital and O&M costs. Greensand filtration costs were not included in either the Water Model or the W/W Model. This technology is considered to be a small systems technology and as a result costs were not estimated for larger systems. This technology could be effectively operated in larger system sizes, but cost data are not readily available. The key parameter is the ratio of source water iron and arsenic. If high

removals are needed the ratio needs to be at least 20:1 iron to arsenic. Costs are based on the following design and operating criteria:

 #   Potassium permanganate feed, 10 mg/L;

 #   The filter medium is contained in a ferrosand continuous regeneration filter tank equipped with an underdrain;

 #   Filtration rate, 4 gpm/ft^2; and

 #   Backwash is sufficient for 40 percent bed expansion.

It is further assumed that corrosion control measures are not required because pH is not affected by the process.

Figures 3-20 and 3-21 present cost estimates for removal of arsenic by greensand filtration.

Figure 3-20
Greensand Filtration
Capital Costs

$y = 587584x^{0.838}$
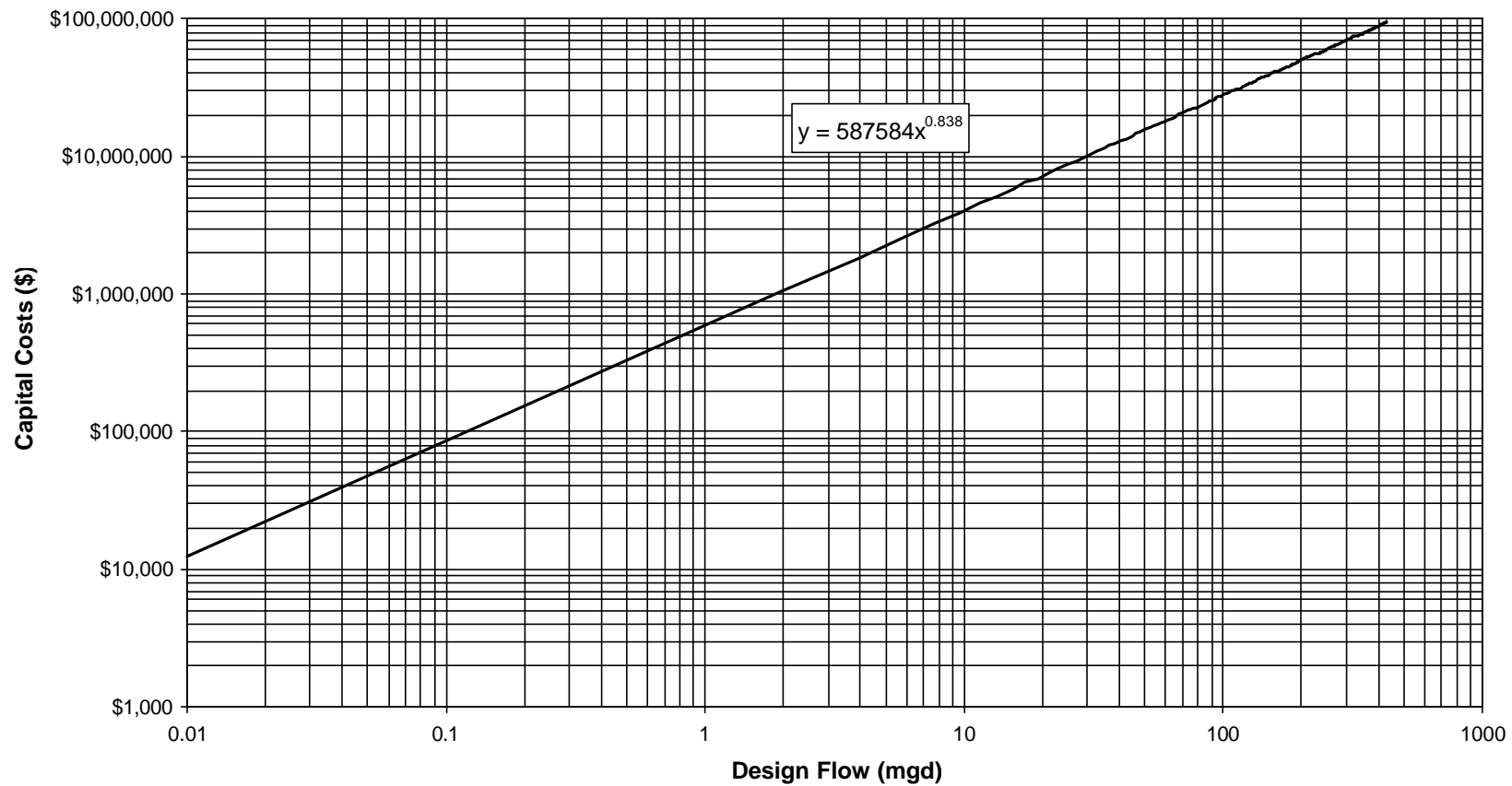
Figure 3-21
Greensand Filtration
O&M Costs

$y=0.0009x^2 + 58921x + 7392.9$
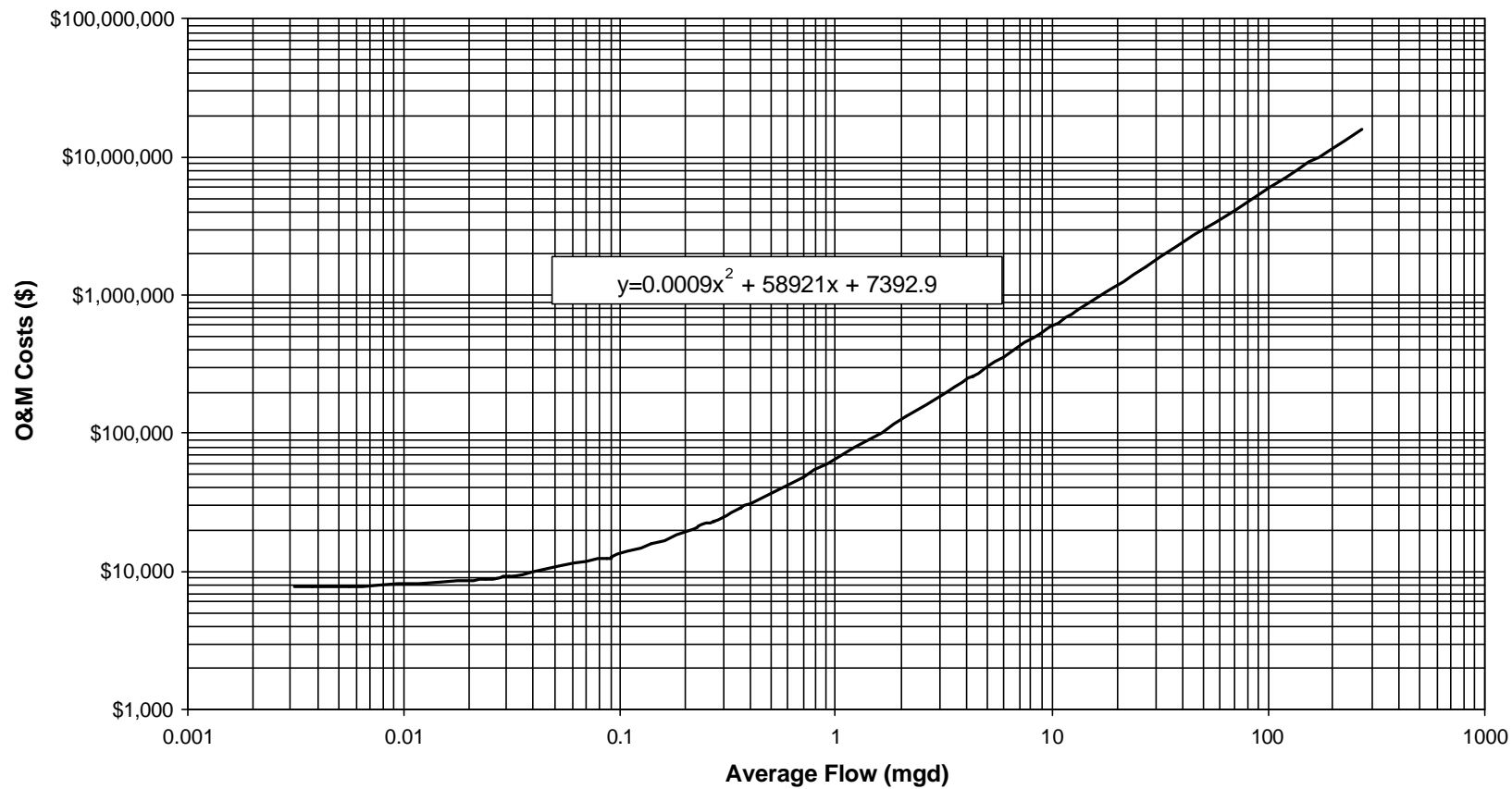
3.11   COMPARISON OF COSTS

The April 1999 and November 1999 drafts of the Technology and Cost Document compared capital and O&M cost estimates in this chapter were compared with actual cost data and cost data from other sources, including the 1993 draft Technology and Cost Document (Malcolm Pirnie, 1993). There are numerous problems with the approach. The first is that there is not a consistent set of design assumptions for a given technology.   For example, without knowing the EBCT used in a full-scale system, a comparison of capital costs is meaningless because the cost difference could be due to a larger media volume. The cost estimates in *Evaluation of Central Treatment Options as Small System Treatment Technologies* (SAIC, 1999) are not specific to arsenic removal. Differences could be due to different assumptions about run length or process design. The same problem would apply in making comparisons between the 1993 Technology and Cost Document and this final draft of the Technology and Cost Document. Activated alumina and anion exchange do not rely on the cost models used in that document because the models are inappropriate for arsenic removal costs. Therefore, this document will not present any comparison of costs.

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4.0   RESIDUALS HANDLING AND DISPOSAL ALTERNATIVES

4.1   INTRODUCTION

The implementation of any of the treatment technologies presented in Chapter 3 will result in the generation of solid and/or liquid waste streams containing elevated levels of arsenic. It is the purpose of this chapter to present the characteristics of the residuals generated by each of the treatment technologies and to discuss appropriate handling and disposal options.

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 those 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 quantity of residuals produced, which is dependent on 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 by a treatment process plays a significant role in determining the handling and disposal method that will be utilized. For example, many handling methods which are suitable for smaller systems are impractical for larger systems because of significant land requirements associated with particular methods. As a result, larger systems that process residuals on-site (as opposed to direct or indirect discharge) typically use mechanical methods to avoid large outlays for land purchase, preparation, and maintenance.

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 for large water systems. Transportation also can 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 greater maintenance than less complex methods.

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 first step in residuals processing, preceding mechanical dewatering processes, evaporation ponds, and/or storage lagoons.

Filter backwash streams are high volume, low solids slurries generated during the cleaning of granular filter media. Backwash volume depends on the number of filters used by a system and the frequency with which they are cleaned. Typical backwash 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 characterized by an average solids concentration of 0.5 to 2.0 percent (DPRA, 1993a).

When possible, backwash waters are recycled within the treatment process. In gravity thickening, backwash waters are fed to a tank where settling occurs naturally. The sludges resulting from gravity thickening are discharged and further treated for ultimate disposal, and the decant is either recycled or discharged to a surface water or to a publicly-owned treatment works (POTW). Gravity thickening reduces the quantity of water sent to waste 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 or a POTW, or may be treated by other mechanical or non-

mechanical dewatering processes. When backwash slurries cannot be recycled or discharged to a surface water or 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 and high O&M costs, compared to non-mechanical dewatering processes that can handle similar waste volumes, e.g., storage lagoons. Due to their high costs, mechanical dewatering processes are generally not suitable for application at very small water systems.

Filter presses have been used in industrial processes for decades, and their use has been increasing in the water treatment industry over the past several years. These devices have been successfully applied for the treatment of 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 may increase 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, but have high capital costs and are labor intensive (DPRA, 1993a). Capital and O&M costs are generally higher than those of comparable non-mechanical dewatering alternatives. As a result, pressure filtration is most applicable for 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 on an influent solids concentration of 1 to 10 percent. Centrifugation is a continuous process that requires very little time (8 to 12 minutes) to achieve optimal sludge solids concentration. Centrifuges have low land requirements and high capital costs. They are more labor intensive than non-mechanical alternatives, but less labor-intensive than filter presses. Again, due to the associated 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 (RO) and ion exchange (IX) (DPRA, 1993a). Ponds and drying beds are not generally suitable for dewatering alum and lime sludges.

The use of evaporation ponds and drying beds involve the discharge of brine waste to a pond for storage and dewatering via the natural process of evaporation. Typically, such ponds are designed to have large surface areas to maximize evaporation of residual water by the sun and wind. Pond size is determined by waste flow and storage capacity requirements.

As previously mentioned, evaporation ponds and drying beds are used primarily for brine wastes generated by RO and IX. Such processes produce large volumes of liquid wastes characterized by high levels of total dissolved solids (TDS), for which mechanical dewatering processes, such as filter presses, are impractical. Depending on the solids concentration of the brine waste stream, intermittent removal of solids may be required. For brines with a TDS content ranging from 15,000 to 35,000 mg/L, solids will accumulate in an evaporation pond at a rate of ½ to 1½ inches per year (DPRA, 1993a). When the depth of the solids in the pond reaches a predetermined level, discharge to the pond is halted and evaporation is allowed to continue until the solids concentration of the waste is suitable for disposal.

The use of evaporation ponds is extremely land intensive, requiring shallow basins with large surface areas. This can be an important consideration in densely populated regions. Reverse osmosis processes produce very large volumes of liquid wastes which increases land requirements and ultimately construction costs. As a result, evaporation ponds may not be suitable for large water systems utilizing RO. Evaporation ponds and drying beds have few O&M 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 be discharged to the same pond for several years (assuming adequate total capacity) 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, lagoons, like evaporation ponds, are land intensive (DPRA, 1993a). Storage lagoons are lined ponds designed to collect, retain, and dewater sludge for a predetermined period of time. Dewatering occurs through the evaporation and decanting of the supernatant. Lagoon size is determined based on the volume of sludge that must be handled and the desired storage time. As with evaporation ponds, when a lagoon reaches its design capacity, solids can be removed by means of heavy equipment and shipped for disposal.

Storage lagoons are best suited for dewatering lime softening (LS) process sludges, though they have been applied with some success to coagulation/filtration (C/F) process sludges. Storage lagoons can operate under a variety of sludge flows and solids concentrations, and chemical conditioning of alum sludges is not required (DPRA, 1993a). Typically, lime sludges are discharged to a lagoon with 3 percent solids, and are dewatered to 50 to 60 percent solids, whereas alum sludges are typically characterized by 1 percent solids when first discharged to a storage lagoon and can be dewatered to 7 to 15 percent solids (DPRA, 1993a). As evident from this data, alum sludges do not typically dewater well in storage lagoons. When the top layer of an alum 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 for acceptance at most landfills (DPRA, 1993a). Further, thickened alum sludges can be difficult to remove from lagoons, often requiring 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 assist in the dehydration of alum sludges.

4.3   DISPOSAL ALTERNATIVES

4.3.1   Direct Discharge

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

However, the discharge of liquid residuals to surface waters is subject to compliance with and permitting under the National Pollution Discharge Elimination System (NPDES). NPDES establishes limits for specific contaminants (including total solids and arsenic) based on a variety of factors, including ambient contaminant levels, low flow condition of the receiving water, and anticipated volume 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 Fg/L. If a water source is also used as a source of drinking water, the arsenic limit is reduced to 0.0175 Fg/L; although some States use the drinking water MCL for this purpose. These criteria are used by state regulatory agencies as a basis for the determination of discharge limits for arsenic depending on the classification of the receiving water.[1] 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 purchasing and installing the necessary 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 successfully to dispose of alum and lime sludges, as well as brine streams generated by systems using RO and IX (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 disposal option may require the installation of a conveyance system (i.e., piping) 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 LS sludges have also been successfully disposed of in this manner. However, the POTW receiving such wastes 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 may be classified as an industrial waste since they contain contaminants, namely arsenic, which may impact the quality of the sludges generated by the POTW. As a result, discharge to a POTW is only acceptable when arsenic concentrations fall within the Technically Based Local Limits (TBLL) established by the POTW under

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[1] Note that State and local limits may be more stringent than federal limits for specific contaminants, including arsenic.

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 municipal wastewater. Background levels are usually measured in domestic wastewater. The existing TBLL would be based on the untreated drinking water quality (higher arsenic). Removing arsenic from drinking water would change the background level in the municipal wastewater. The TBLL would likely be revised since the background level has been lowered. The revised TBLL would be used to determine if the brine stream from the drinking water process could be discharged to the POTW. One approach to evaluating if indirect discharge is an option is to perform a mass balance on arsenic based on the treated and untreated conditions. The critical factor in determining the arsenic increase at the POTW will be the volume of water lost between the drinking water treatment plant and the POTW. If there were no water lost between the drinking water treatment plant and the POTW, then there would be no arsenic increase to the POTW.

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 of the residual exceeds the Clean Sludge Limit of 41 mg/kg, the biosolids may be land applied, but the quantity will be limited to a total cumulative arsenic loading of 41 kg per hectare (36.6 lb/acre). As a result, most TBLLs are based on the Clean Sludge criterion (AWWARF, 1998). This source appears to overstate the impact of the arsenic concentration in sludge and the ability to land apply the biosolids. Biosolids with concentrations between 41 and 75 mg As/kg biosolids can be land applied as long as the arsenic accumulation is tracked. The lifetime accumulation is 41 kg As/hectare of land. Using the maximum arsenic concentration of 75 mg/kg biosolids, over 500,000 kg of biosolids would need to be applied to one hectare to exceed the limit, which is unlikely.

To illustrate the impact of Part 503 regulations on the development of an arsenic TBLL, consider a limit of 41 mg/kg for the land application of biosolids (AWWARF, 1998). 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 is located in an area that has a background arsenic concentration near 13 µg/L, it may not be permitted to discharge to the local POTW. There are several problems with the above example. The first is that the background concentration would be reduced because arsenic is being removed from drinking water. Therefore, the TBLL would need to be revised. The second is that 41 mg/kg is being used as the upper limit for land application when the actual upper limit is based on 75 mg As/kg biosolids since arsenic accumulation is unlikely to exceed the lifetime limit. The key variables to determine if arsenic in the POTW biosolids would exceed the land application limit are: existing background arsenic concentration, water loss between drinking water treatment plant and POTW, arsenic removal efficiency of the POTW, and the volume of biosolids generated by the POTW. Land application restrictions on the biosolids would only likely be an issue when background arsenic levels are high, water loss is 50% or higher between the drinking water treatment plant and the POTW, the POTW has a high arsenic removal efficiency and only generates a low volume of biosolids. Thus, arsenic accumulation in biosolids is unlikely to be as significant a restriction as total dissolved solids (TDS) increases due to the salt used for regeneration for anion exchange.

Anion exchange is regenerated using sodium chloride. Typical regenerations will use 10.2 lb/cubic feet of resin. Even though the brine stream is small in volume, the TDS concentration of the brine stream will be very high. As a result, the POTW discharge may increase the TDS content of the receiving waters. Areas of the country with high TDS naturally or limited quantities of water may find any increase in TDS to be unacceptable. In areas where small increases in TDS can be accommodated, the water loss between the drinking water treatment plant and the POTW and the size of the brine stream compared to the total volume of water treated at the POTW will affect the magnitude of the TDS increase.

As with direct discharge, 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. Other 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 placed on the local 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 on several variables, including soil and sludge chemistry and 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 up to 41 mg/kg can be land applied without tracking arsenic accumulation. When the concentration is in the range of 41 to 75 mg/kg, sludges can be land applied if arsenic accumulation is tracked and does not exceed 41 kg As/hectare. Due to the possibility of arsenic absorption by vegetation, application of sludges to non-food chain fields is preferred. Land application is also limited by the availability of land. In areas where grassland, farmland, or forested land is unavailable, transportation costs can significantly affect the overall cost effectiveness of this disposal option.

Land application can serve as a means of final disposal of LS, and to a lesser degree C/F, sludges. Lime sludges can have a beneficial impact on farmland by neutralizing soil pH, replacing the use of 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 governing the types of wastes that 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). Federal requirements include restrictions on location, operation and design criteria, groundwater 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 (typically determined by the Paint Filter Liquids Test, SW-846, Method 9095). 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 meeting even one of these criteria is considered hazardous. For drinking water 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 toxicity characteristic (TC) regulatory level for arsenic is 5 mg/L, which is 100 times the interim primary drinking water maximum contaminant limit (MCL) of 50 Fg/L. The TC regulatory levels for other inorganic contaminants were also based on 100 times their interim primary drinking water MCL. The MCLs for many of these inorganic contaminants have been revised since 1990 whereas the TC levels have not changed. Thus, lowering the arsenic MCL does not mean that the TC regulatory level will be lowered. A separate rulemaking would be required for the reduction of the TC level.

Many water treatment facilities currently dispose of their waste residuals in commercial or public 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 are generally less than those associated with the development of a sanitary landfill (DPRA, 1993a). Monofills control the type of waste that they accept more strictly and also limit potential future liabilities. However, it is unlikely that monofills will be utilized for arsenic treatment residuals.

4.3.5   Hazardous Waste Landfill Disposal

Water treatment residuals containing arsenic which fail the TCLP test for toxicity must be disposed of 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 the 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.

Hazardous waste 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 a treatment residual is 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, the generation 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. Due to the extremely high costs associated with this disposal method, it was assumed for the purpose of estimating costs for the Arsenic Rule that water systems would not select treatment processes that would require disposal of waste residuals as hazardous waste to meet the revised MCL.

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 pound of solids 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, typically characterized by 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 in previous sections of 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 (mgd).

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 previous sections of 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 mgd.

Selection of Handling and Disposal Options

As previously discussed, C/F blowdown and filter backwash are high volume waste streams characterized by low solids content. Gravity thickening may be used to increase the solids content of 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 a land-intensive disposal option, they may not be applicable for large water systems.

Disposal of C/F residuals containing arsenic is largely dependent on influent arsenic concentration, coagulant dose and suspended solids content. Disposal by direct discharge to a surface water is not typically appropriate.

Depending on the arsenic concentration of C/F sludges, land application may be a suitable disposal method. As discussed in section 4.3.2, total arsenic cannot exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with arsenic concentrations between 41 and 75 mg/kg may be land applied provided that the total loading does 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 of in a sanitary landfill. Otherwise, residuals must be disposed of in a hazardous waste landfill. A 1992 study (Bartley *et al.*, 1992) found that the waste sludges from C/F plants would pass the TCLP test. In this study, samples were taken from the waste sludges generated by four different water treatment plants and subjected to the TCLP test. One of these systems relied on C/F, and another used both C/F and lime softening processes. These two systems treated raw waters characterized by arsenic concentrations averaging 1.1 mg/L and less than 0.001 mg/L, respectively. The results of the TCLP tests conducted on the waste residuals from both systems were below 0.040 mg/L (range: 0.007 to 0.039 mg/L)–well below the current criterion for treatment as hazardous waste (5.0 mg/L).

A recent study (EPA, June 2000), examined the characteristics of the waste sludges generated by two C/F plants (Plant A and Plant B). The sludge from Plant A was generated from backwashing anthracite coal/pea gravel filters, while that from Plant B was generated as a result of sedimentation in primary and secondary clarifiers and from filter backwashing. Both sludges passed the TCLP test

for arsenic toxicity by a substantial margin (ranges: less than 0.050 to 0.106 mg/L and 0.058 to 0.160 mg/L, respectively) and would not have exceeded the strict TCLP limits established by California for nonhazardous waste. However, the sludge from Plant A would violate the soluble threshold limit concentrations (SLTCs) established by California for arsenic and copper. As a result, the sludge generated by this treatment plant would be considered a hazardous waste, requiring appropriate handling and disposal.

Tests conducted by the University of Colorado confirm the results found in these two studies, indicating 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 or exceeds another regulatory limit established for water treatment sludges by individual States.

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 only about 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 mgd.

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 section 4.2.2. As for 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 mgd.

Selection of Handling and Disposal Options

As previously discussed, enhanced coagulation blowdown is characterized by high volume and low solids content. Typical solids contents range from 0.5 to 2.0 percent, depending on coagulant type. Gravity thickening may be used to increase the solids content of enhanced 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 may not be applicable for large water systems since they are land intensive.

Disposal of enhanced coagulation arsenic residuals is largely dependent on influent arsenic concentration, coagulant dose, and suspended solids content. Disposal by direct discharge to a surface water is not typically appropriate.

Depending on the arsenic concentration of C/F sludges, land application may be a suitable method of disposal. Total arsenic should not exceed 41 mg/kg if sludges are to be land-applied without restrictions. Sludges with arsenic concentrations between 41 and 75 mg/kg may be land applied provided that the total loading does 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 of in a sanitary landfill. Otherwise, residuals must be disposed of in a hazardous waste landfill. Note that hazardous waste disposal is unlikely since tests conducted by the University of Colorado indicate that enhanced coagulation sludges will pass the TCLP test (AWWARF, 1998). Due to its high cost disposal to a hazardous waste landfill should only be relied on 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 used in this process. Sludge production is also affected by the suspended solids content of the raw water. Systems are unlikely to install direct filtration solely for arsenic removal. If this technology were installed, the disposal alternative discussed for coagulation/filtration would be appropriate for this technology.

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 $Fe(CH)_3$ precipitate.

The solids content of the backwash from coagulation assisted microfiltration processes were found to be less than 0.5 percent in one study (Clifford et al, 1997).

Selection of Handling and Disposal Options

The wastes from coagulation assisted microfiltration processes will consist of a very dilute slurry. Gravity thickening may be used as to increase the solids content of the sludge prior to the use of 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 means of handling the residuals generated by coagulation assisted microfiltration. Both require little oversight and maintenance, but are land intensive. As such, they may not be appropriate for large systems. A thorough comparison of handling options should be conducted to select the most cost effective method.

Land application may be a suitable disposal method for sludges from coagulation assisted microfiltration processes. As discussed in section 4.3.2, total arsenic cannot exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with arsenic concentrations between 41 and 75 mg/kg may be land applied provided that the total loading does not exceed 41 kg 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. However, tests conducted by Clifford (Clifford, 1997) and the University of Colorado (AWWARF, 1998) indicate that the sludges from this treatment process will pass the TCLP test for arsenic toxicity by a considerable margin, making it unlikely that hazardous waste disposal will be necessary. 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 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 the use of LS for the removal of non-carbonate hardness.

LS plants typically produce 1,000 to 8,000 pounds of solids per million gallons of water treated depending on the hardness of the influent water (AWWARF, 1998). Arsenic concentrations

of these sludges, however, are generally lower than those found in C/F sludges due to the increased volume of solids produced.

Selection of Handling and Disposal Options

LS blowdown is slightly more dense than C/F blowdown. Typical solids content ranges from 1.0 to 4.0 percent, depending on raw water hardness. Gravity thickening may be used to pretreat LS sludges prior to their 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.

Land application of LS treatment sludges is one possible disposal alternative. As discussed in section 4.3.2, total arsenic cannot exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with arsenic concentrations between 41 and 75 mg/kg may be land applied provided that the total loading does not exceed 41 kg per hectare.

LS sludges will require dewatering prior to landfill disposal. If the residuals pass the TCLP test they may be disposed of in a sanitary landfill. Otherwise, residuals must be disposed in a hazardous waste landfill. However, hazardous waste landfill disposal should only be used as a last resort if waste fails the TCLP test which is unlikely given the findings of available studies. Tests conducted by the University of Colorado indicate that LS sludges will pass the TCLP test (AWWARF, 1998). This finding is supported by the findings of Bartley, *et al.* (1992).

In the Bartley study, samples were taken from the waste sludges generated by four different water treatment plants and subjected to the TCLP test. Two of these systems relied on LS, and another used both C/F and LS processes. All three of these systems treated raw waters characterized by arsenic concentrations averaging less than 0.001 mg/L. The results of the TCLP tests conducted on the waste residuals from these systems ranged from 0.007 to 0.039 mg/L–significantly below the current criterion for treatment as hazardous waste (5.0 mg/L).

4.4.6   Enhanced Lime Softening

Enhanced LS is a modified LS process wherein lime dosage is increased. Residuals generated as a byproduct of this process are similar to those generated by a typical LS treatment process. The quantity of sludge produced is a function of water hardness. As with LS, the use of enhanced LS for

carbonate hardness removal produces approximately twice the amount of solids per pound of hardness removed as the use of this process for 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.

Selection of Handling and Disposal Options

Typical solids content of enhanced LS blowdown range from 1.0 to 4.0 percent, depending on 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.

Land application of treatment sludges is a possible disposal alternative. As discussed in section 4.3.2, total arsenic cannot exceed 41 mg/kg if sludges are to be applied with no restrictions. Sludges with arsenic concentrations between 41 and 75 mg/kg may be land applied provided that the total loading does not exceed 41 kg per hectare.

Sludges will require dewatering prior to landfill disposal. If the residuals pass the TCLP test they may be disposed of in a sanitary landfill. Otherwise, residuals must be disposed of in a hazardous waste landfill. However, as with LS sludges, such treatment is not likely to be necessary, and should only be used as a last result.

Tests conducted by the University of Colorado indicate that enhanced LS sludges will pass the TCLP test (AWWARF, 1998). Additional tests conducted by the University of Colorado confirmed that the treatment sludge from enhanced LS treatment will typically pass the TCLP (results: 0.0009 to 0.0284 mg/L; Amy, 1999). A case study presented by Sorg (2000) lends further strength to the expectation that the residuals from this treatment process may be treated as nonhazardous.

Sludge from the enhanced LS plant in a recent EPA study was collected, sent to a slurry tank, and then to a sludge lagoon (EPA, June 2000) . Approximately once every 2 years, or as required, the sludge lagoon was dewatered using an underdrain system. Arsenic was not detected in the leachate at any of the sampling locations used in the study (detection limit: 0.05 mg/L). As a result,

following dewatering, the dry sludge from this plant could be removed and applied to local farm fields as a beneficial amendment.

4.4.7   Ion Exchange

Ion exchange (IX) involves the use of a synthetic resin in the 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 chloride per equivalent of resin, i.e., 10.2 pounds of salt per cubic foot of resin. Regeneration requires approximately 2.8 bed volumes (BV) of brine and 1.2 BV displacement rinse. Therefore, 4 to 5 BV of waste are produced per regeneration cycle. (AWWARF, 1998). The rinse volume in another study was 5 BV (Clifford et al, 1997). Using this as an upper bound, the waste volume could be as high as 7.8 BV.

The other important factor is the arsenic concentration in the waste brine. An EPA study (EPA October 2000) which found that arsenic concentrations during brine regeneration ranged from 1.83 to 38.5 mg/L (average: 16.5 mg/L). The wastewater generated during the other three steps of the regeneration process (i.e., backwash, slow rinse, and fast rinse) evaluated by this study were characterized by lower arsenic concentrations (0.0594, 1.332, and 0.108 mg/L, respectively). Therefore, if these waste streams were combined prior to disposal, the overall arsenic concentration of the brine would be below that observed during the brine regeneration step.

Selection of Handling and Disposal Options

The November 1999 Technology and Cost Document listed three mechanisms to dispose of the brine stream used for regeneration. The options were: sanitary sewer, evaporation pond, and chemical precipitation. Many comments on the proposed rule were considered with waste streams being considered hazardous waste. Waste streams containing less than 0.5% solids are evaluated against the toxicity characteristic directly to determine if the waste is hazardous. Arsenic in the regeneration brine will likely exceed 5 mg/L for most systems with arsenic above 10 Fg/L and sulfate below 50 mg/L. Since the brine stream would likely be considered hazardous, the evaporation pond and the chemical precipitation options were eliminated as options for disposal of anion exchange wastes. Discharge to a sanitary sewer was retained because domestic sewage and any mixture of domestic sewage and other wastes that pass through a sewer system to a publicly-owned treatment works (POTW) for treatment is excluded from being considered solid waste (40 CFR 261.4).

Domestic sewage means untreated sanitary wastes that pass through a sewage system. Discharges meeting the above criteria are excluded from regulation as hazardous waste.

Although the brine stream may be deemed a hazardous waste, such characterization would not, in itself, prevent disposal to a POTW since wastes that passes through a sewer system to a POTW are exempt from RCRA regulation. Discharge to the sanitary sewer may be limited by TBLLs for arsenic or TDS. The TDS is the more likely limiting factor since the arsenic TBLL would need to be recalculated since the background arsenic will change. The only net increase in arsenic is related to the water loss from the drinking water treatment plant to the POTW. TDS is different. TDS is being added to the waste stream. Due to the potential for an increase in total dissolved solids, anion exchange would be favored in areas other than the southwest where the volume of brine is very small relative to the total volume of wastewater being treated at the POTW.

Although one small system was reported to have disposed of its waste brine to the facility's septic tank (EPA October 2000), such disposal is likely to be regulated by local and State regulations and therefore was not considered in the cost analysis for the final rule.

4.4.8 Activated Alumina

Activated alumina (AA) can be operated with or without pH optimization and regeneration. Systems optimizing pH and regenerating will produce a regenerant waste solution with a pH of approximately 12 and high in dissolved solids, aluminum, and arsenic (AWWARF, 1998). 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 (approximately 42 BV per regeneration cycle). Arsenic is strongly adsorbed to the media, so only about 50 - 70% of the adsorbed arsenic is removed even using a strong base.. The brine stream produced by the regeneration process then requires disposal. The November 1999 Technology and Cost Document listed discharge to a sanitary sewer as the disposal mechanism for the brines. It was noted that TBLLs for arsenic might restrict discharge of brine streams to the sanitary sewer. Since activated alumina run lengths are much longer than anion exchange, the arsenic concentrations in the brine stream would likely be much higher even with incomplete regeneration. Regeneration of activated alumina media is not recommended even for larger systems because disposal of the brine may be difficult, the regeneration process is incomplete which reduces subsequent run lengths, and for most systems it will be cheaper to replace the media rather than regenerate it. As a result, regeneration of AA was not considered for the final rule.

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

Residuals from systems operating on a 'throw away' basis, will not face the difficulties of those systems that regenerate since they need only dispose of spent media, rather than both concentrated brines and spent media. These systems can dispose of process residuals in a non-hazardous waste landfill. Spent alumina will have to pass the TCLP test before disposal at a sanitary landfill is possible. In a recent study (Wang *et al.*, 2000), TCLP tests were performed on samples of activated alumina taken from two different water systems. Both of these systems had similar set-ups: four tanks of activated alumina with two parallel sets of two tanks in series (a lead or 'roughing' tank followed by a 'polishing' tank). Each tank contained approximately 10 cubic feet of activated alumina. During operation, the flow rate through each parallel series was 5 gallons per minute. This resulted in a hydraulic loading rate of 1.59 gallons per minute per square foot and an empty bed contact time (EBCT) of 15 minutes. Over the course of the study, the influent concentration of arsenic ranged from 0.053 to 0.087 mg/L (average: 0.062 mg/L) and 0.021 to 0.076 mg/L (average: 0.049 mg/L) for the two systems. Both systems consistently removed greater than 92 percent of influent arsenic throughout the course of the study. The media from each of the roughing tanks was replaced about every 18 months. At the time of media replacement, samples of spent media from different parts of the roughing tanks (i.e., the top, middle, and bottom) were taken and subjected to the TCLP test. In all cases, the samples had TCLP test results of less than 0.05 mg/L, well below the statutory limit of 5.0 mg/L. Based on these findings and the fact that activated alumina binds most tightly to arsenic at the pH at which the TCLP is conducted, it is unlikely that the spent media will be classified as hazardous.

4.4.9 Nanofiltration

NF membranes are primarily used for softening, removing the larger, divalent ions associated with hardness (AWWARF, 1998). The recovery rate is dependent upon the source water quality. Typical recovery rates of this technology approach 85 percent. The reject stream from NF contains elevated levels of arsenic and other contaminants that are removed from the source water by the NF membranes. As discussed in Chapter 3, nanofiltration is not yet a proven reliable technology that can be operated at realistic recoveries. If data are generated to show it can be reliably operated at high recoveries, then the disposal options would be similar to those for reverse osmosis.

4.4.10 Reverse Osmosis

Reverse osmosis (RO) membranes will remove much smaller ions typically associated with TDS (AWWARF, 1998). The recovery rate for RO membranes is dependent on the source water quality but typically falls between 30 and 85 percent. The reject stream from this process contains elevated levels of arsenic and other contaminants. While reverse osmosis is an effective removal technology, it is unlikely to be selected solely for arsenic removal because other options are more cost effective for arsenic removal and do not reject a large volume of water. RO may be cost effective options if removal of other contaminants is needed and water quantity is not a concern.

The disposal technologies were presented in the November 1999 Technology and Cost Document. They were: direct discharge, indirect discharge, and chemical precipitation/non-hazardous landfill disposal. Since the recovery rate is typically less than 85% percent, reverse osmosis reject streams can be characterized as high volume with lower arsenic concentrations compared to the brine streams from anion exchange and activated alumina. Arsenic concentrations should not exceed the TC regulatory level since the maximum concentration effect is less than 6 to 1.

Selection of Handling and Disposal Options

Direct discharge of RO residuals to a receiving surface water is one possible disposal alternative. The large volume of the waste stream makes this alternative unlikely except for small systems (which produce a small waste volume) or systems located near the coast.

Indirect discharge to the POTW is also an option. Similar to anion exchange, the existing TBLLs would likely need to be modified because the background arsenic will change. The water loss between the drinking water plant and the POTW will be the critical factor in determining the arsenic increase to the POTW. The volume of the reject stream compared to the total volume at the POTW will also be a critical factor in determining if the POTW would accept this waste stream. Systems should check with the POTW before selecting this option.

Chemical precipitation of reverse osmosis streams is likely to be cost prohibitive. Reverse osmosis is already a fairly expensive treatment technology compared to other options and chemical precipitation of the large volume waste stream would significantly increase those costs. If direct discharge and indirect discharge are unavailable, systems should consider another treatment technology.

Costs are not presented for reverse osmosis waste disposal because it was not used to develop national costs and other options are more cost effective and produce much smaller waste streams.

4.5   RESIDUALS HANDLING AND DISPOSAL COSTS

Cost estimates were generated using the *Small Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993a), and the *Water System Byproducts Treatment and Disposal Cost Document* (DPRA, 1993b).

Costs are presented for four water treatment processes: (1) coagulation assisted microfiltration, (2) ion exchange, (3) activated alumina, and (4) greensand filtration. The disposal options assumed for each of these treatment processes are shown below:

- # Ion Exchange:
  - POTW (indirect) discharge

- # Coagulation Assisted Microfiltration:
  - Mechanical dewatering and nonhazardous landfill disposal
  - Non-mechanical dewatering and nonhazardous landfill disposal

- # Activated Alumina:
  - Nonhazardous landfill disposal (systems operating without regeneration)

- # Greensand Filtration:
  - POTW (indirect) discharge

The handling and disposal options presented represent those selected by EPA during the development of the arsenic rulemaking decision tree. These disposal options will be used by EPA during the Regulatory Impact Analysis. Costs for additional disposal options can be estimated using the DPRA disposal cost documents (DPRA, 1993a and 1993b). Capital and O&M cost estimates for the selected handling and disposal options are presented in Figures 4-1 through 4-14. **Capital cost estimates for nonhazardous landfill disposal are assumed to be zero.** As a result, capital cost estimates are not provided when landfill is the sole disposal technology under consideration (i.e., activated alumina systems choosing not to regenerate). Capital cost estimates are provided in instances where additional technologies are employed, for example, mechanical dewatering and nonhazardous landfill disposal. **Land costs are excluded from all capital cost estimates.**

Figure 4-1
Anion Exchange (< 20 mg/L SO4)
Waste Disposal Capital Costs

y = 21.47x + 5198

y = 5085

y = 3955
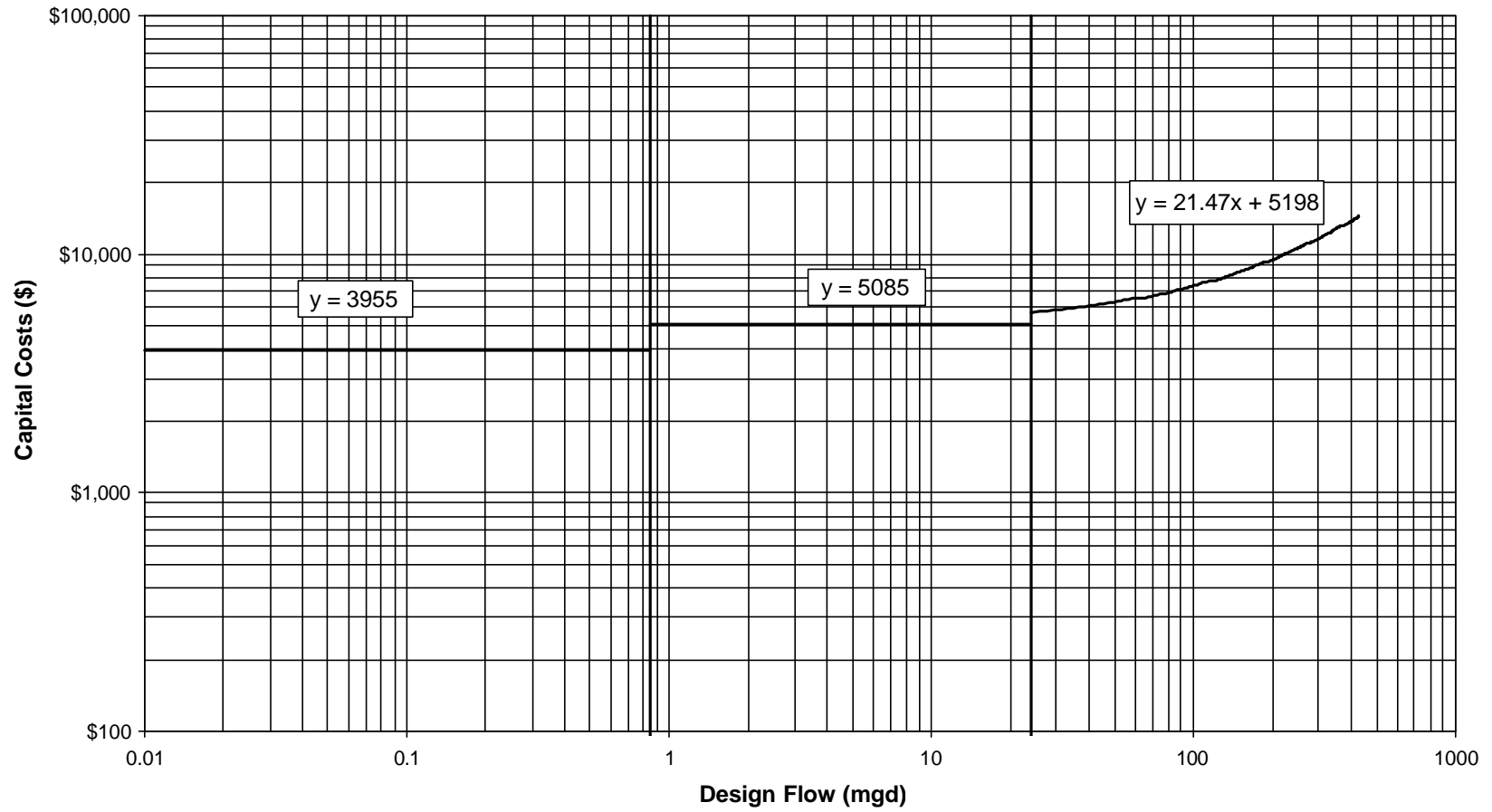
Figure 4-2
Anion Exchange (< 20 mg/L SO4)
Waste Disposal O&M Costs

y = 3429x + 375

O&M Costs ($)

Average Flow (mgd)
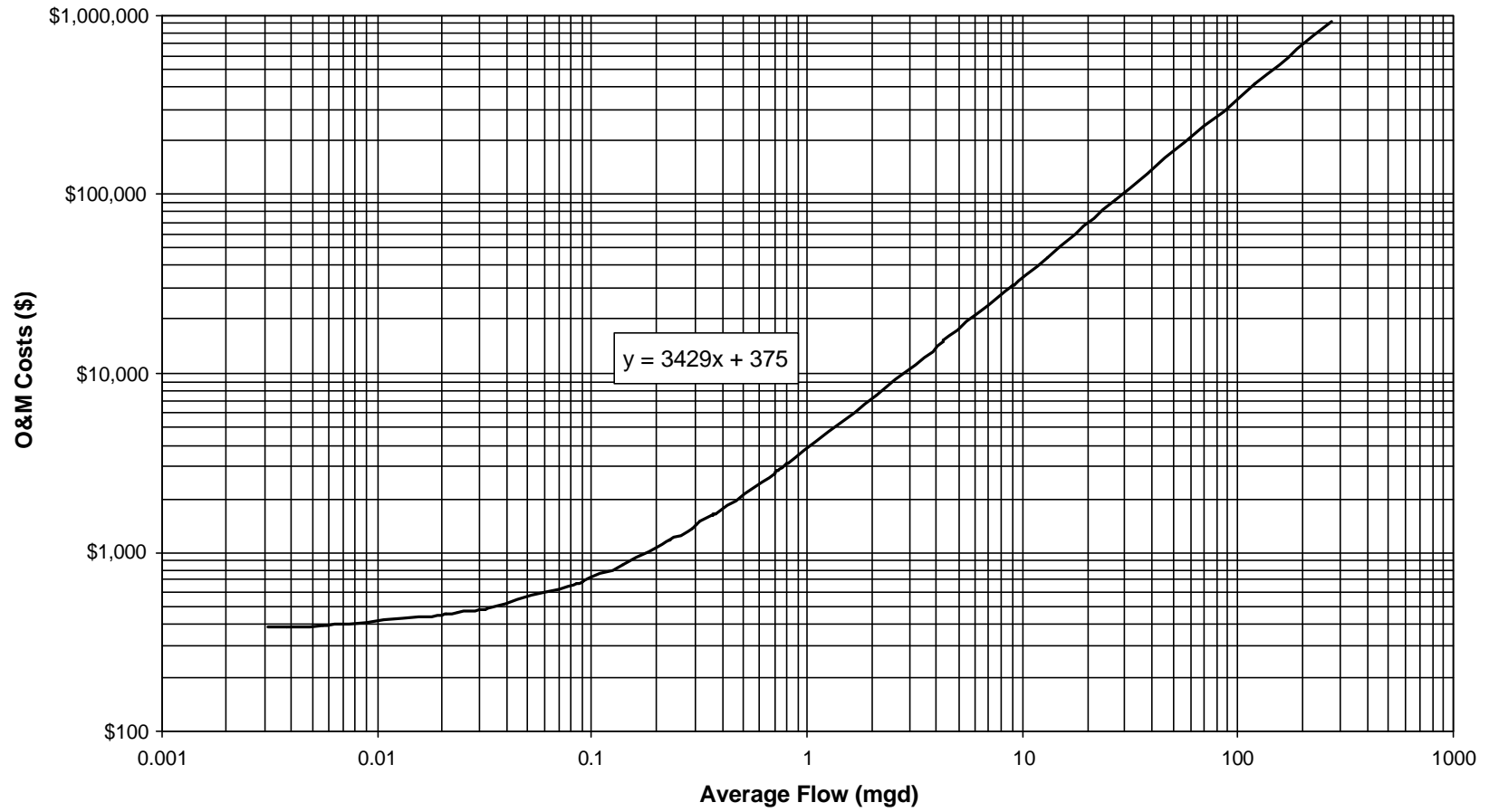
Figure 4-3
Anion Exchange (20-50 mg/L SO4)
Waste Disposal Capital Costs

y = 21.47x + 5198

y = 5085

y = 0x + 3955

Capital Costs ($)

Design Flow (mgd)
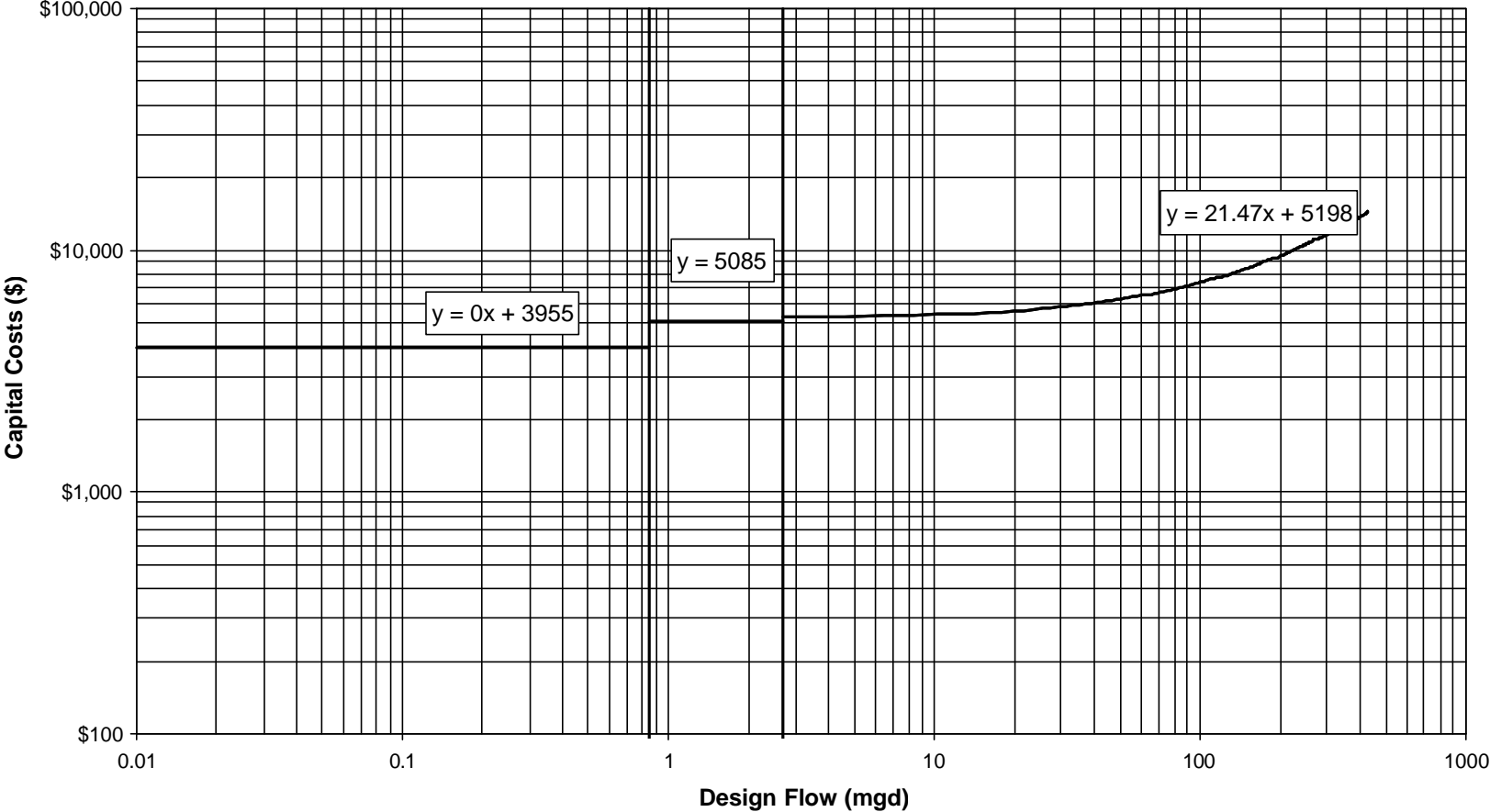
Figure 4-4
Anion Exchange (20-50 mg/L SO4)
Waste Disposal O&M Costs

$y = 7348x$
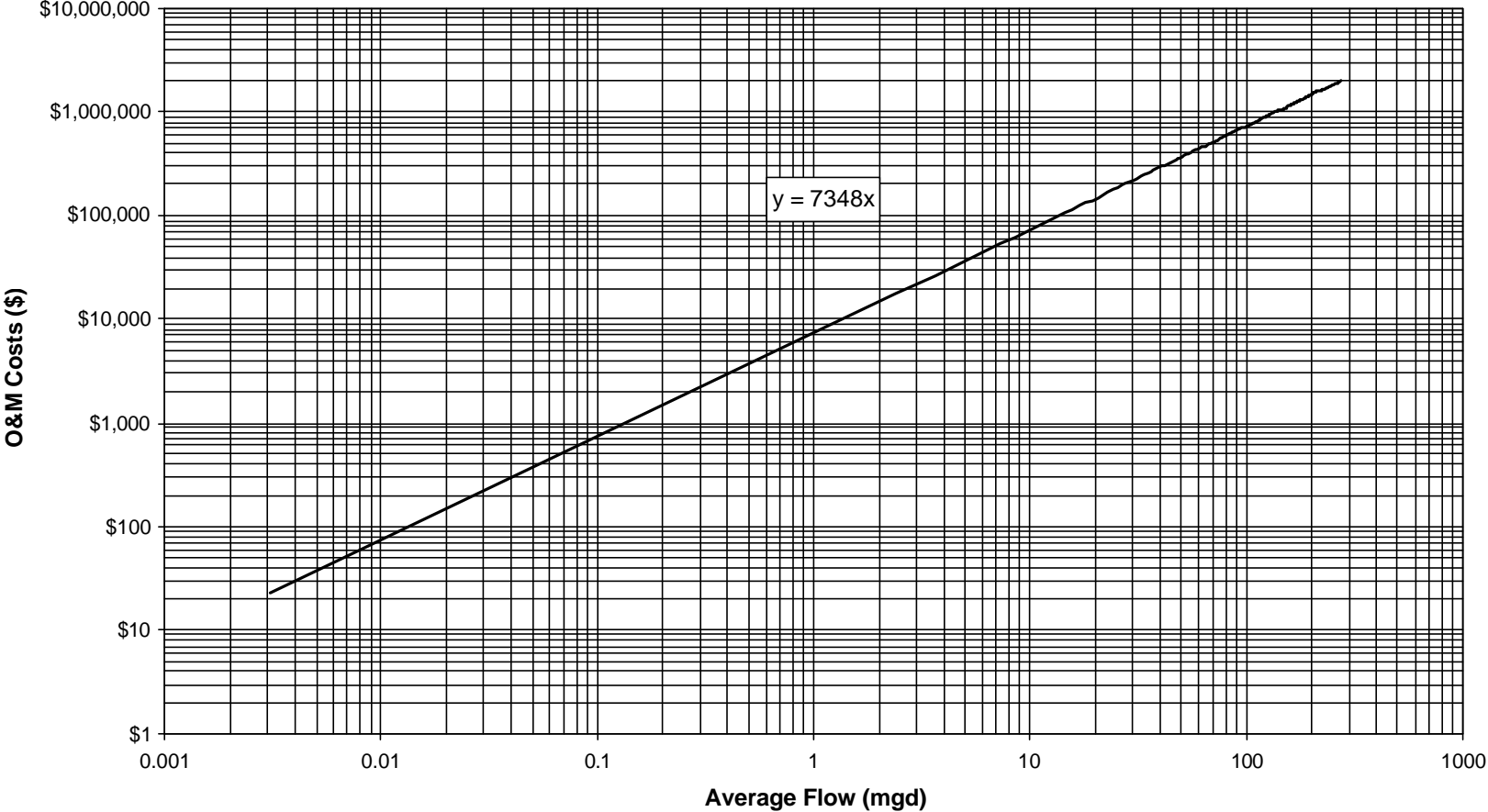
Figure 4-5
Coagulation Assisted Microfiltration (Mechanical)
Waste Disposal Capital Costs

$y = 211627x + 42043$

$y = -1644.1x^2 + 150402x + 157575$

$y = -692793x^2 + 455329x + 26748$

Capital Costs ($)

Design Flow (mgd)
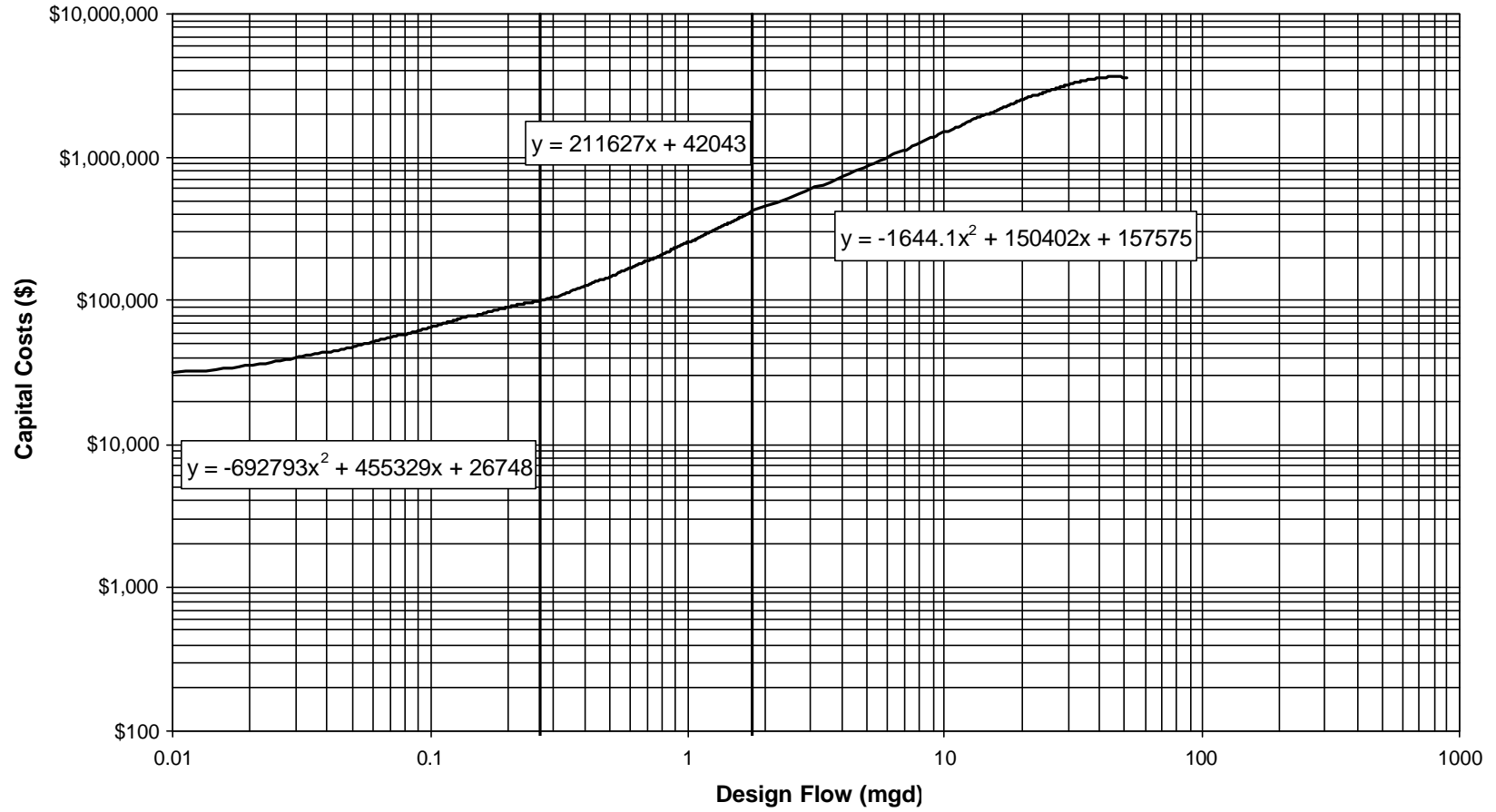
Figure 4-6
Coagulation Assisted Microfiltration (Mechanical)
Waste Disposal O&M Costs

$y = -359.66x^2 + 46936x + 21793$

$y = 25165x + 36857$

$y = -3476860x^2 + 684838x + 5839$
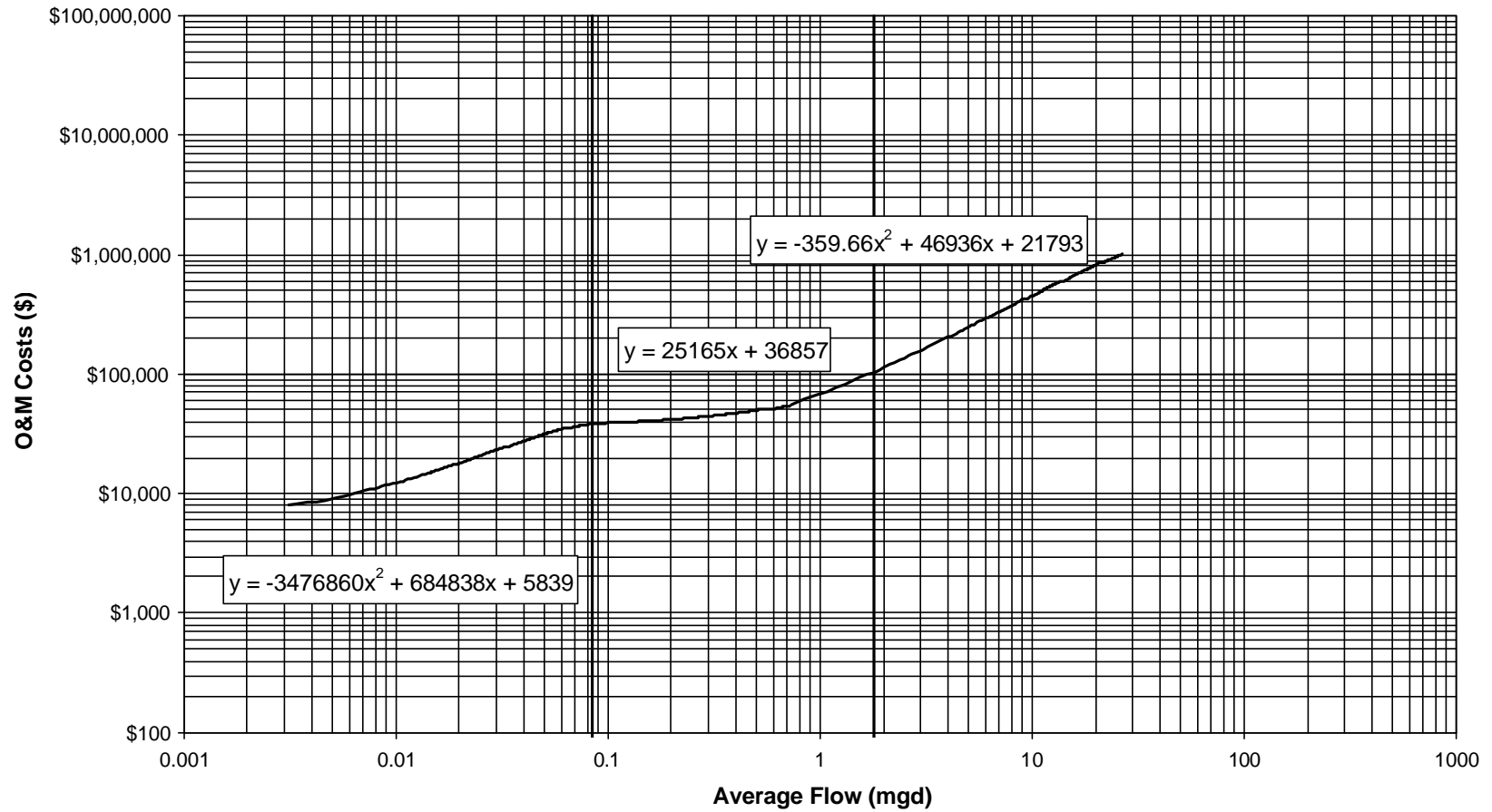
Figure 4-7
Coagulation Assisted Microfiltration (non-mechanical)
Waste Disposal Capital Costs

y = 1749352x + 108017

y = 1627970x + 326504

y = 3069360x - 790

Capital Costs ($)

Design Flow (mgd)
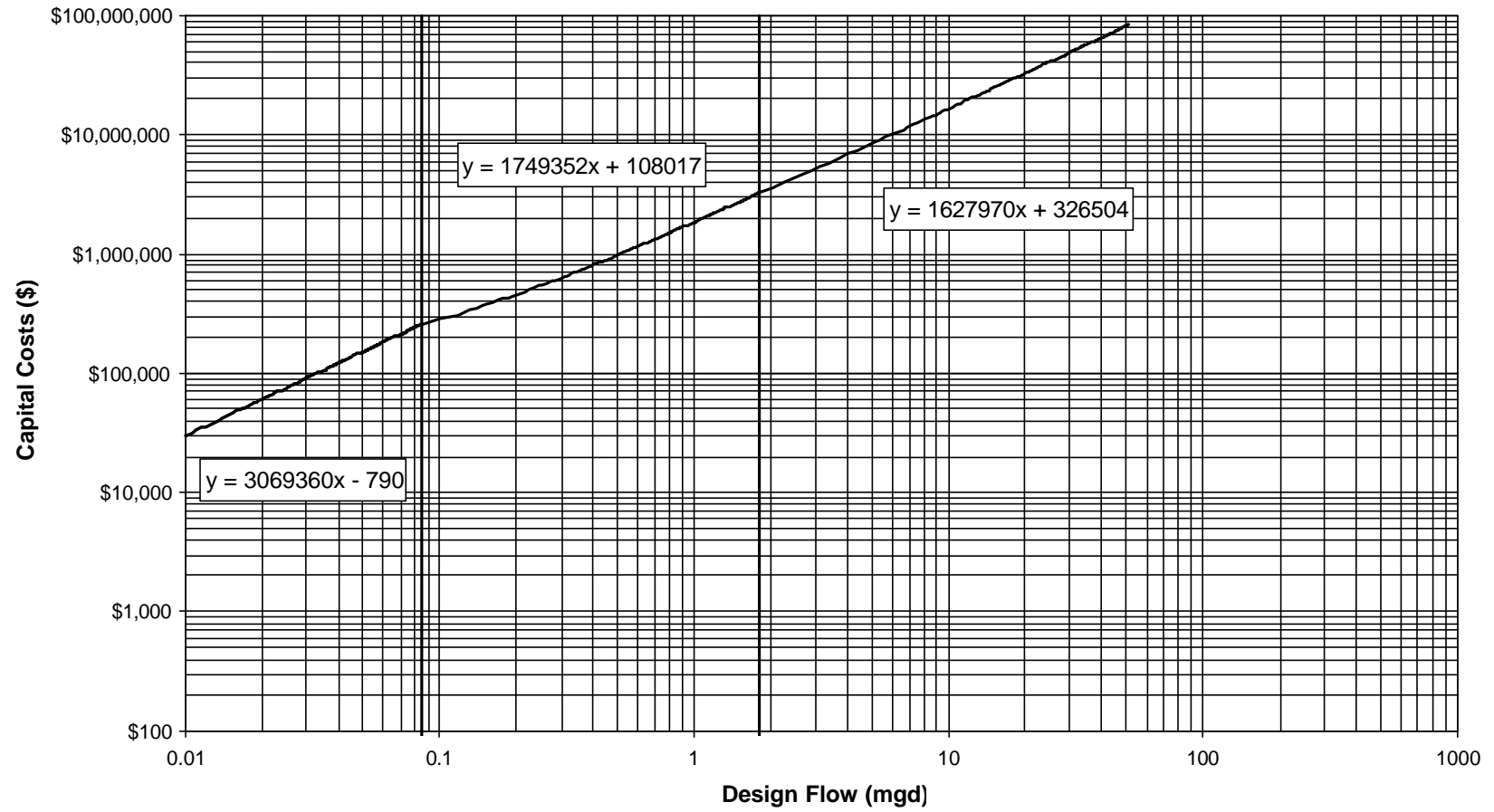
Figure 4-8
Coagulation Assisted Microfiltration (non-mechanical)
Waste Disposal O&M Costs

$y = 16.966x^2 + 60792x + 28760$

$y = 111819x - 6950$

$y = 18812x^2 + 4686x + 2124$

O&M Costs ($)

Average Flow (mgd)

4-30
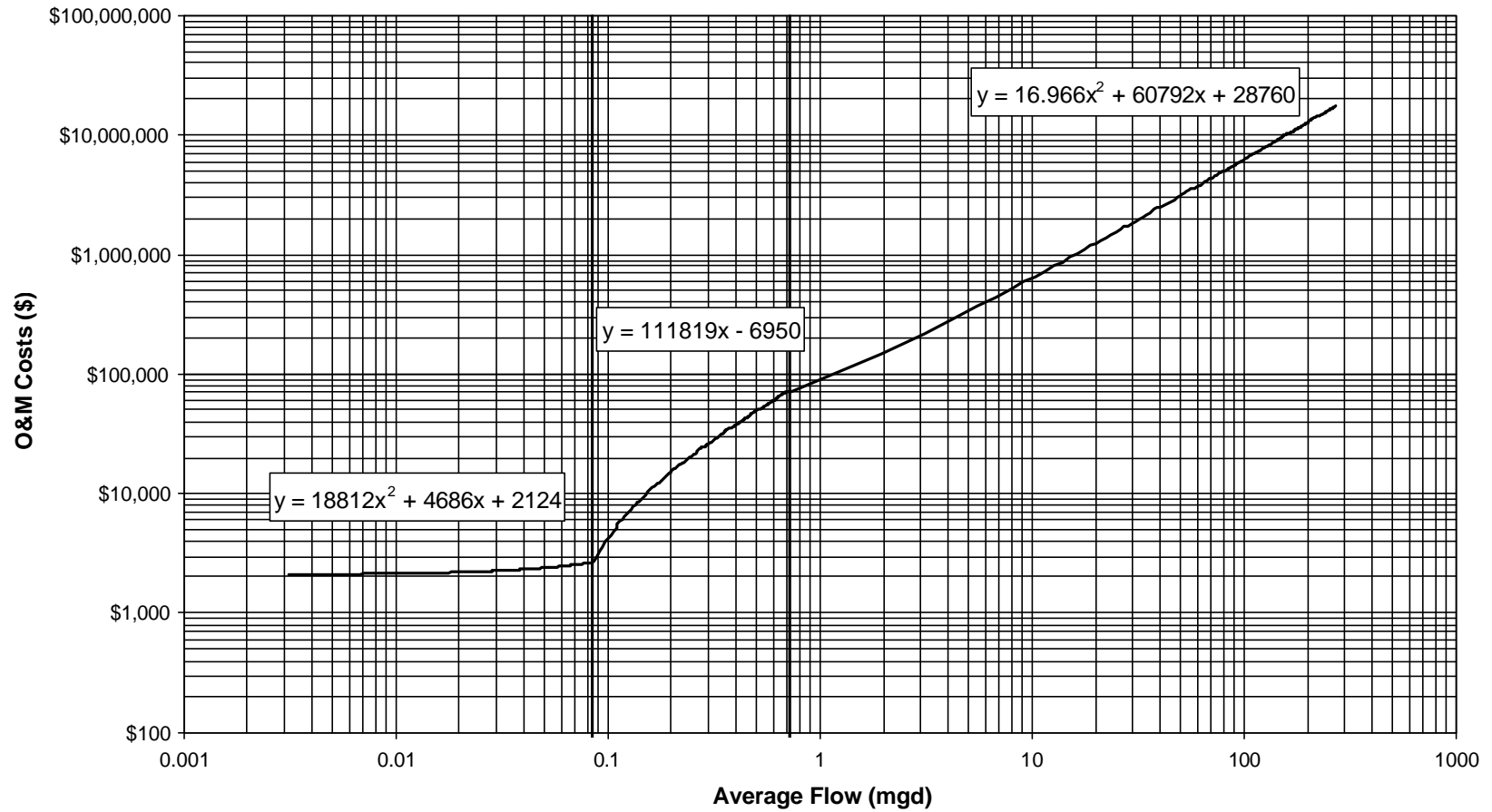
Figure 4-9
Activated Alumina (pH 7 - 8)
Waste Disposal O&M Costs
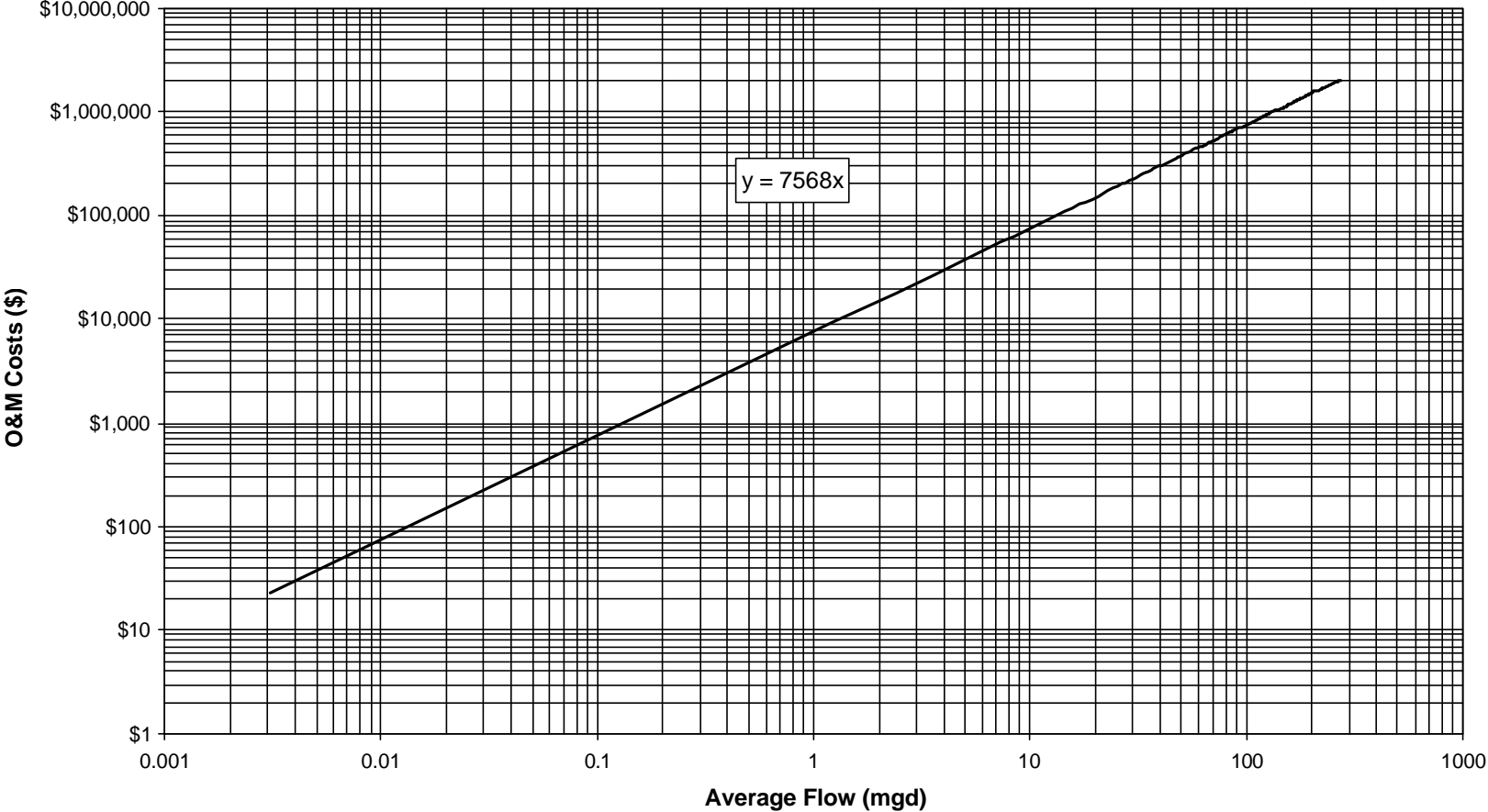
**Figure 4-10
Activated Alumina (pH 8 - 8.3)
Waste Disposal O&M Costs**

y = 14555x
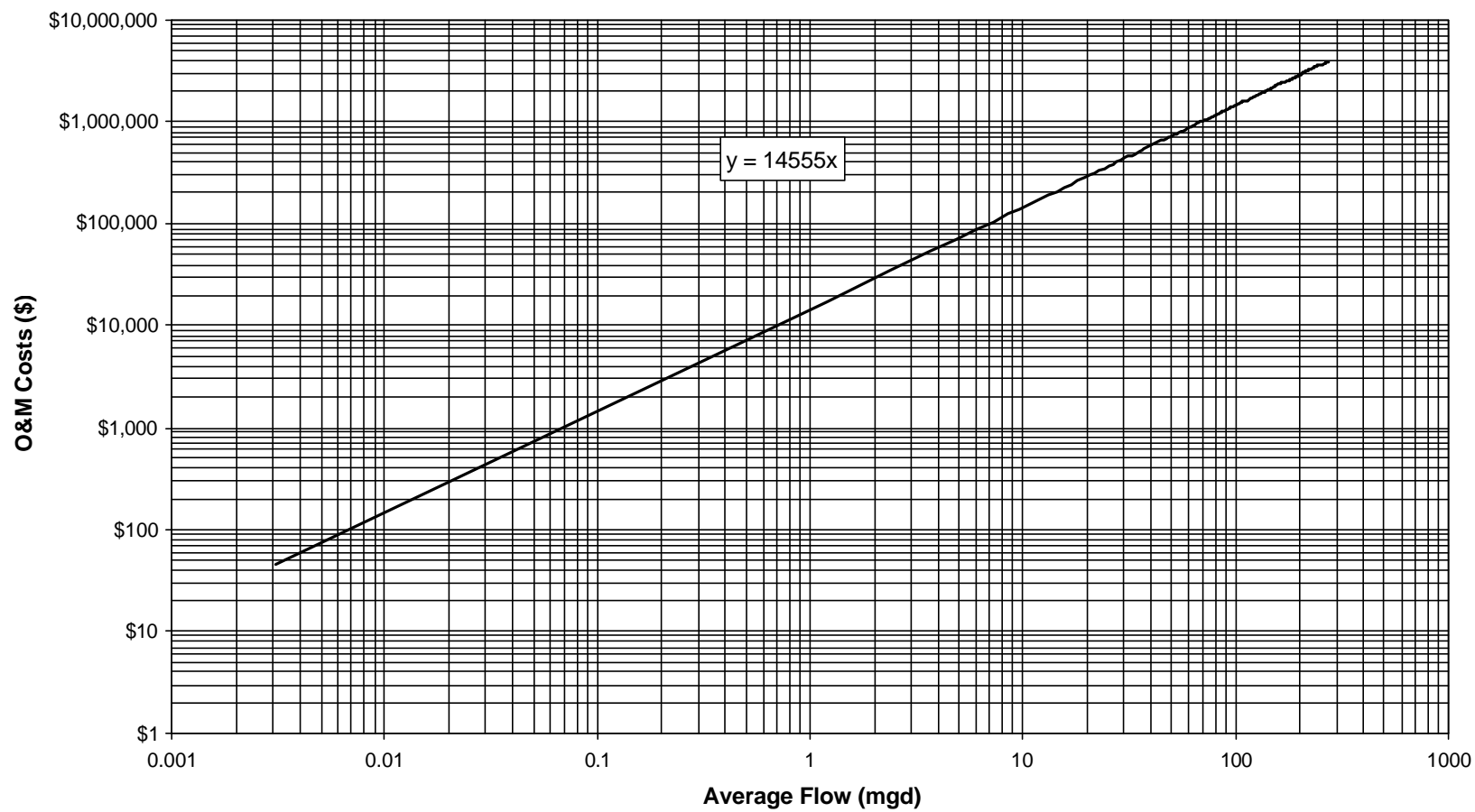
Figure 4-11
Activated Alumina (pH Adjusted to 6.0, 23,100 BV)
Waste Disposal O&M Costs

y = 3276x

(x-axis) **Average Flow (mgd)**
(y-axis) **O&M Costs ($)**
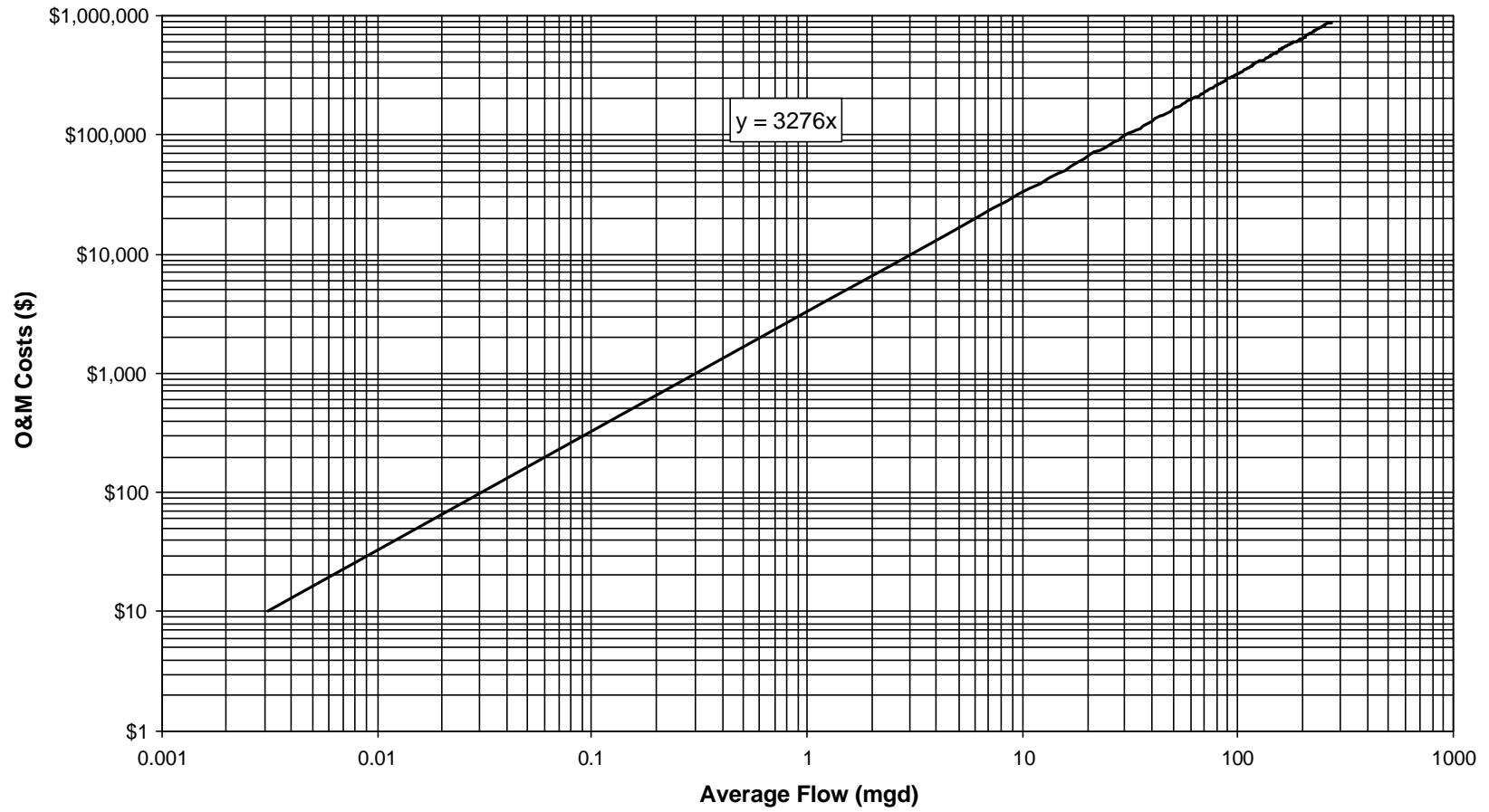
Figure 4-12
Activated Alumina (pH Adjusted to 6.0, 15,400 BV)
Waste Disposal O&M Costs

$y = 4915x$
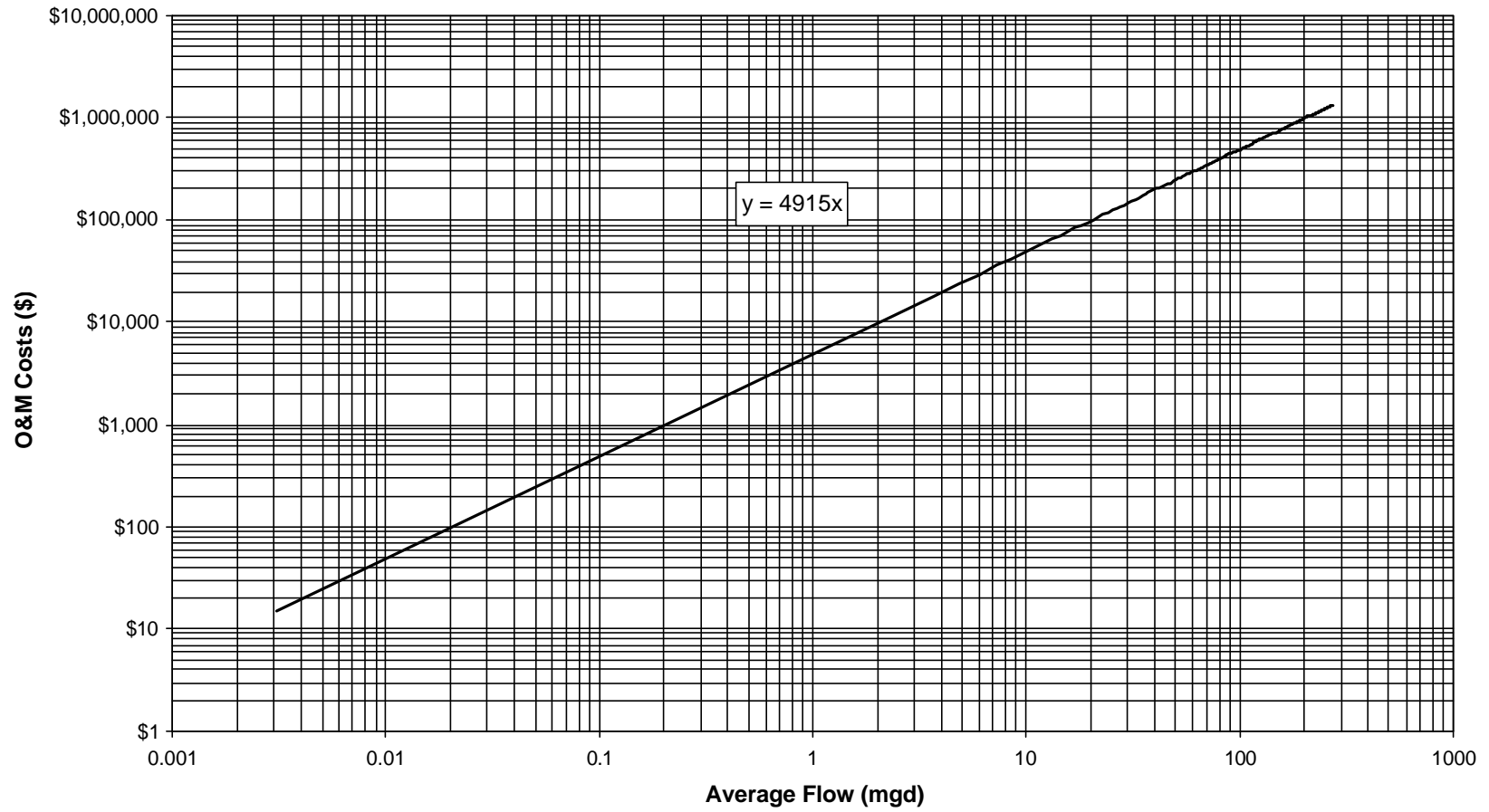
Figure 4-13
Greensand Filtration
Waste Disposal Capital Costs
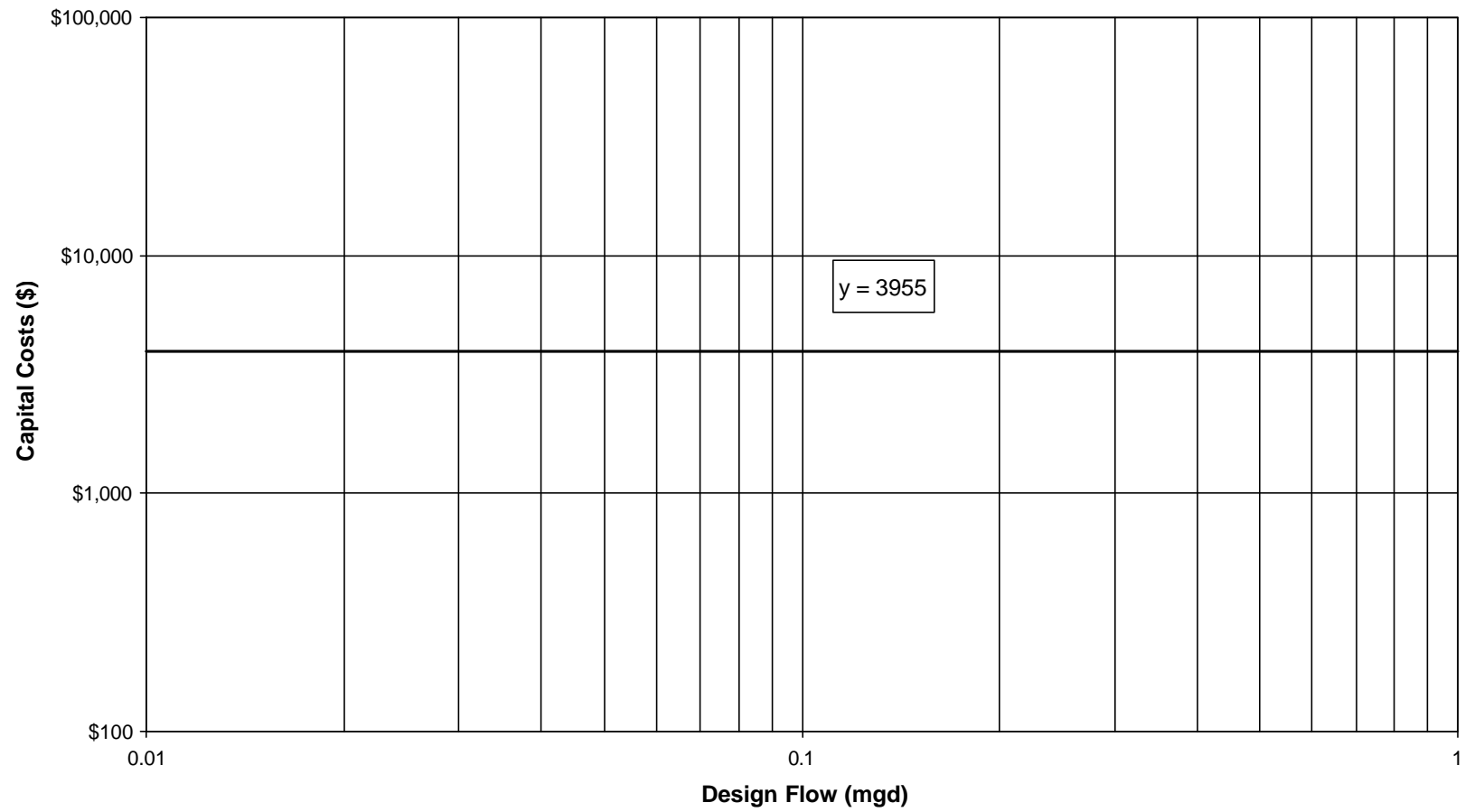
Figure 4-14
Greensand Filtration
Waste Disposal O&M Costs
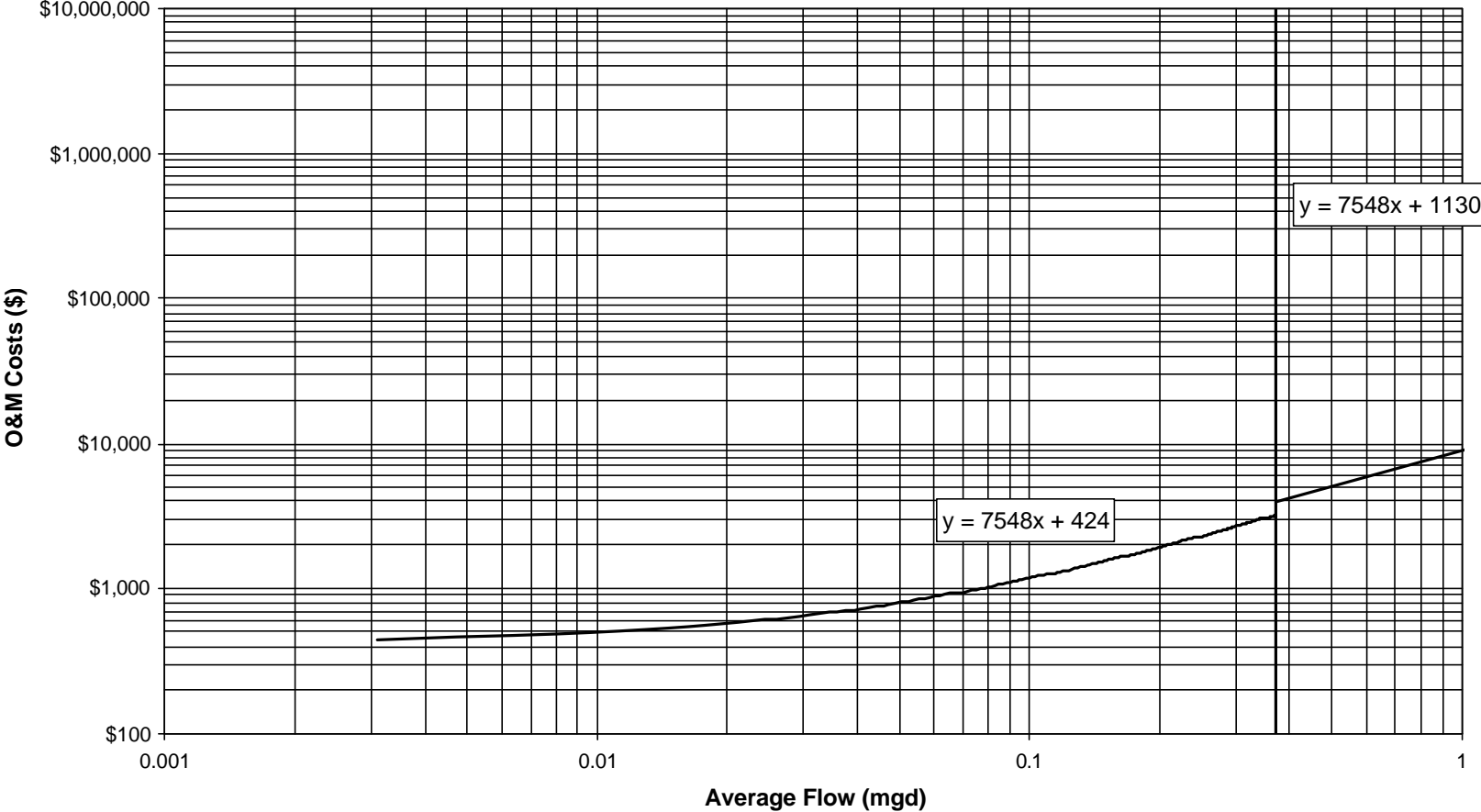
5.0   POINT-OF-ENTRY/POINT-OF-USE TREATMENT OPTIONS

5.1   INTRODUCTION

Centralized treatment is not always a feasible treatment option, for example, in areas where each home has a private well or where 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 500 or more, but less than 3,300; and
Population of 3,300 or more, 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 an effective means of removing arsenic from potable water (Fox and Sorg, 1987; Fox, 1989). 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 $HAsO_4^{-}$. 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.

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

chromatographic 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 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	14-225
Reverse Osmosis (high pressure)	50-80	684*	650**
Activated Alumina	NA	0-18,557**	0-1,226**
Ion Exchange	NA	0-20,935**	0-1,471**

* Based on EPA DWRD study conducted in Fairbanks, Alaska and Eugene, Oregon with arsenic concentrations ranging from 0.026 to 1.16 mg/L. (Fox, 1989)

** Taken off line before MCL exceeded.

NA = not applicable. Removal percentage does not adequately express unit performance for these technologies because run length is the determining factor when attempting to achieve a desired removal percentage.

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 on start-up, achieving 60 to 80 percent removal of arsenic. Over time, however, the arsenic removal efficiency decreased to 50 percent 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 percent removal, but with high influent arsenic concentrations, much higher removal efficiency is necessary to achieve the MCL. The high-pressure unit (196 psig) operated for 350 days and produced 684 gallons of treated water which met the MCL.

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) rather than resin. Over the course of this study, the AA beds effectively reduced arsenic levels; however, the media required pre-treatment to reduce the pH of the AA to 5.5 - 6.0 (the pH at which AA most effectively removes arsenic). Pre-treatment of the AA bed 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 longer than one year.

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.

During the study, observed arsenic concentrations of 0.068 mg/L to 0.02 mg/L were consistently reduced to less than the detectable limit (0.005 mg/L) by the POU RO units. Other contaminants were also effectively removed, including manganese (80 percent), iron (85 percent), and TDS (95 percent). Based on the manufacturer's literature, it appeared that the units were operating at an approximate recovery rate of 25 percent, 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 on 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 percent removal of arsenate by RO systems, and slightly less (75 percent) removal of arsenite. Field studies indicate that greater than 50 percent removal is possible, but data are inconclusive much beyond those levels.

5.4.1 Cost Estimates

The EPA document, *Cost Evaluation of Small System Compliance Options - Point-of Use and Point-of-Entry Treatment Units* (Cadmus Group, 1998), was used to estimate POE and POU RO treatment costs. Costs are presented in Figures 5-1 and 5-2 for total capital costs and annual O&M costs for POU RO treatment, respectively. The cost curves are based on 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 percent discount for 10 or more units, 15 percent 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.

POE RO is not considered a compliance technology because it could create corrosion control problems. Furthermore, water recovery would not be any higher than observed for central RO treatment, so concerns regarding water quantity that arise with central treatment would also apply to POE treatment. Therefore, POE RO treatment option is not considered in the final rule.

Figure 5-1
POU Reverse Osmosis
Total Capital Costs

$y = 864.66x^{0.9261}$

(X-axis: Households; Y-axis: Capital Costs ($))
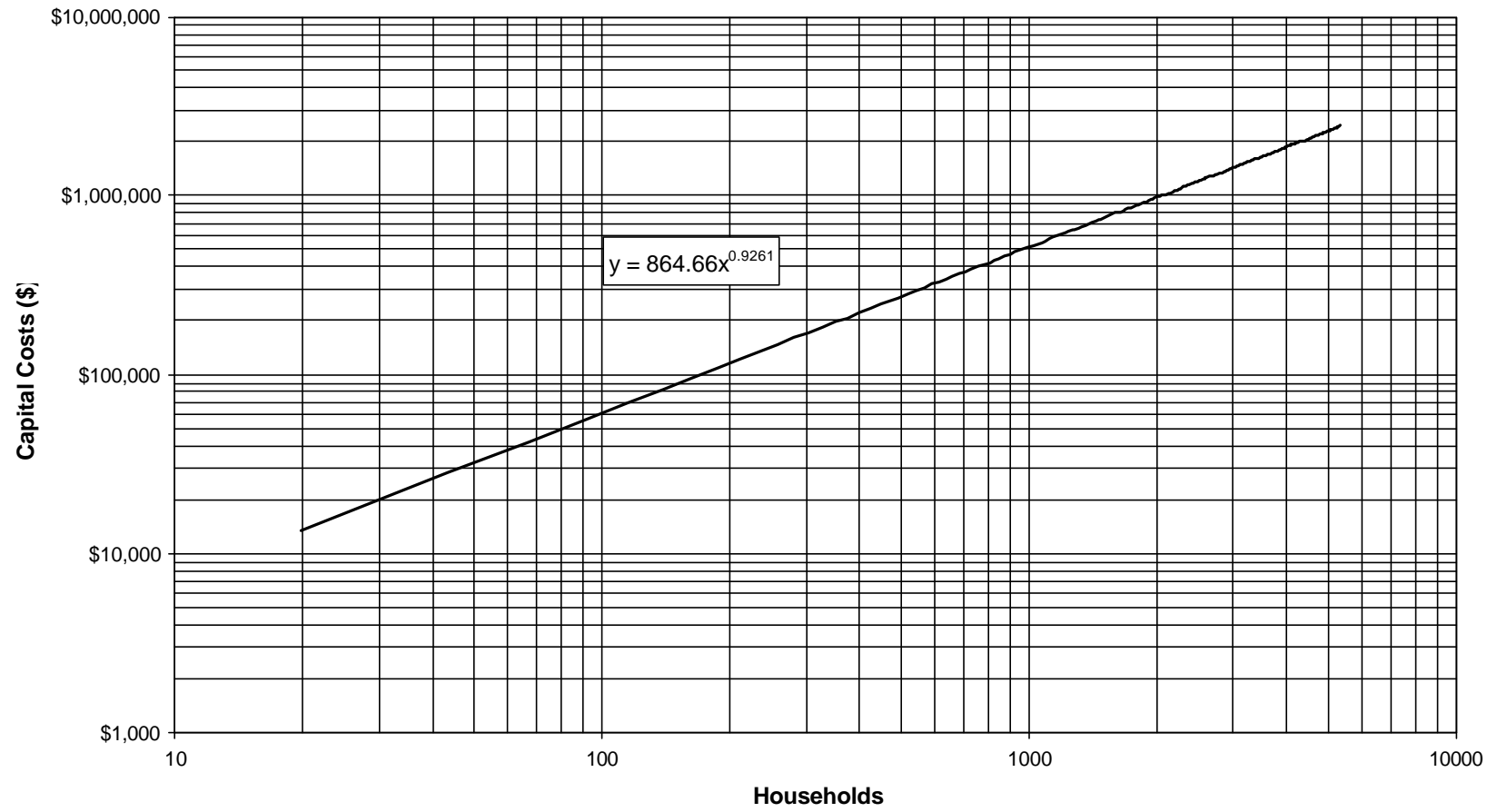
Figure 5-2
POU Reverse Osmosis
O&M Costs

$$y = 266.92x^{0.9439}$$

Y-axis: O&M Costs ($)
X-axis: Households
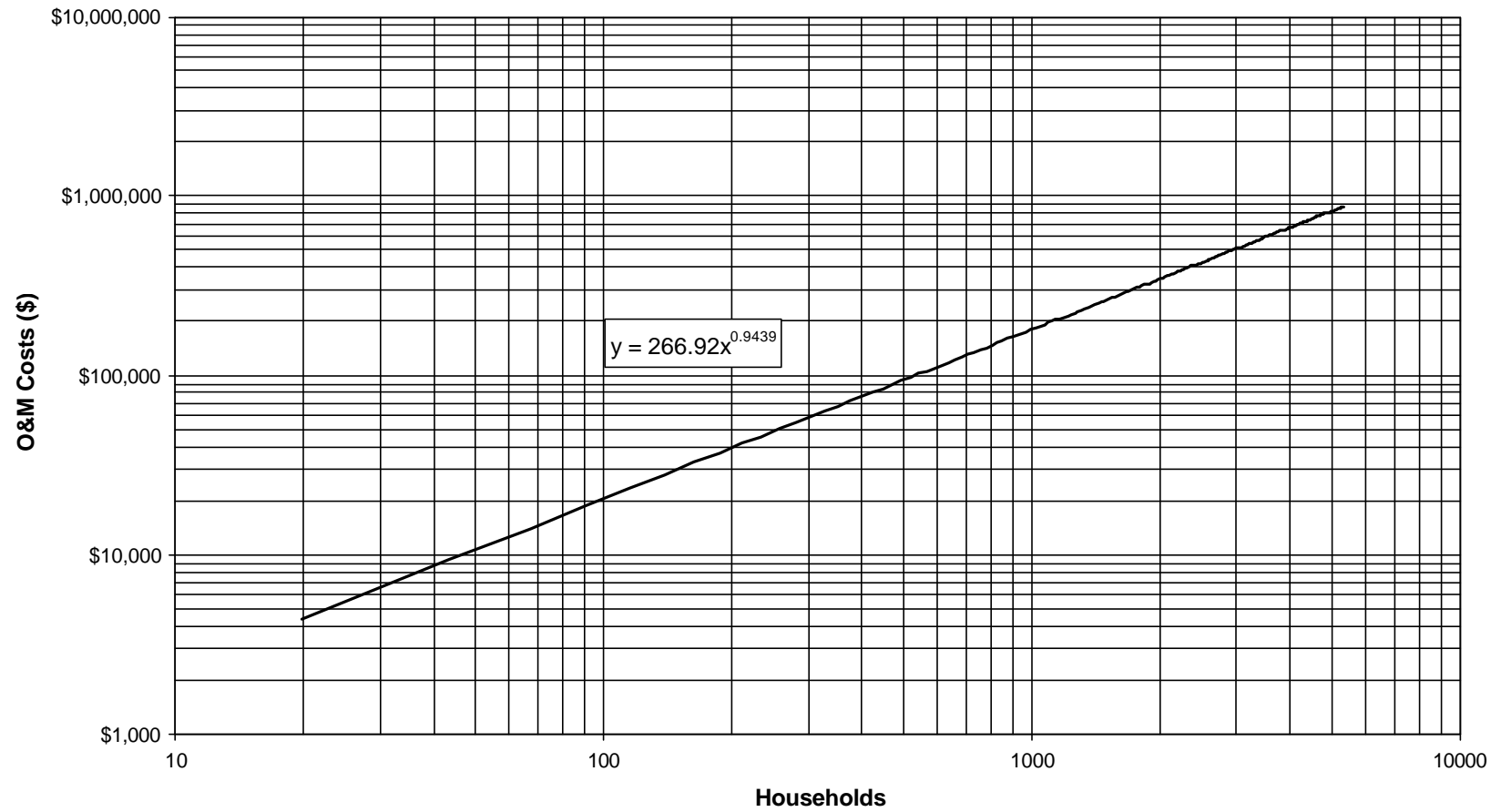
5.5   ION EXCHANGE

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 that this leads to better than average removal of arsenic. Further research is needed to explain this effect. 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 as the basis for estimates of POE and POU treatment costs. The following assumptions were made to estimate the cost of POE and POU IX:

- #   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 percent discount for 10 or more units, vendor retains 30 percent 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.

Due to concerns about chromatographic peaking and run lengths that typically would be less than 6 months, POU anion exchange is not listed as a compliance technology in the final rule (Kempic, 2000). POE IX also may present problems since the brine waste from the regeneration process is characterized by high TDS. In addition, POE run lengths were determined to be only a few months, so frequent regeneration would be required (Kempic, 2000). Therefore, POE anion exchange is also not listed as a compliance technology in the final rule.

5.6   ACTIVATED ALUMINA

Activated alumina (AA) can be used to remove inorganic contaminants, including arsenic, from source water. Contaminant removal occurs as contaminants are exchanged with the hydroxide ions on the alumina surface. Since AA is an amphoteric material, its properties are closely linked to the pH of the ambient environment. At high pH (i.e., pH above 8.2), AA will act as a cationic exchanger, while at lower pHs, it will perform as an anionic exchanger. However, it is important to note that AA is rarely used at high pH (>8.3) 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 with central treatment will need some type of pH adjustment to accommodate this requirement. For POE and POU systems this can not be accomplished because of practical reasons.

5.6.1   Cost Estimates

The Cadmus Group (1998) document was again used as the basis to develop estimates for POE and POU treatment costs. Costs are presented in Figures 5-3 for total capital costs and 5-4 for annual O&M costs. The cost curves are based on 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 percent discount for 10 or more units, vendor retains 30 percent 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.

Arsenic removal by AA is very sensitive to pH, and finished water pH typically will be higher than the optimal required for AA. A finished water pH for many systems would be in the range of pH 7 to pH 8. Using data on activated alumina run length and pH, it was determined that viable run lengths for POE AA were likely only when the finished water pH was at or below pH 7.5 (Kempic, 2000). Even in this pH range, the media may need to be replaced more frequently than once a year, which would make the option very expensive especially compared to the POU AA option. The run length data used for this analysis were from a site with very little competing ions (Simms and Azizian, 1997). Studies at other sites with higher levels of competing ions have much lower run lengths (Clifford et al, 1998). Based on the limited finished water pH range where POE AA might be effective and that the POU media needs replacing much less frequently due to lower water demand, POE AA has not been listed as a compliance technology. POE devices that utilize media that are less sensitive to pH adjustment may be listed as compliance technologies in the future once data on their performance is generated.

However, because the volume of water treated by POU AA is much smaller, POU units have typically have longer run lengths and require less frequent replacement than POE units. Therefore, the costs of POU AA treatment were considered in the final rule. The cost of replacement for POU AA is also much lower than POE AA.

Figure 5-3
POU Activated Alumina
Total Capital Costs

$y = 296.89x^{0.9257}$

Capital Costs ($) vs Households
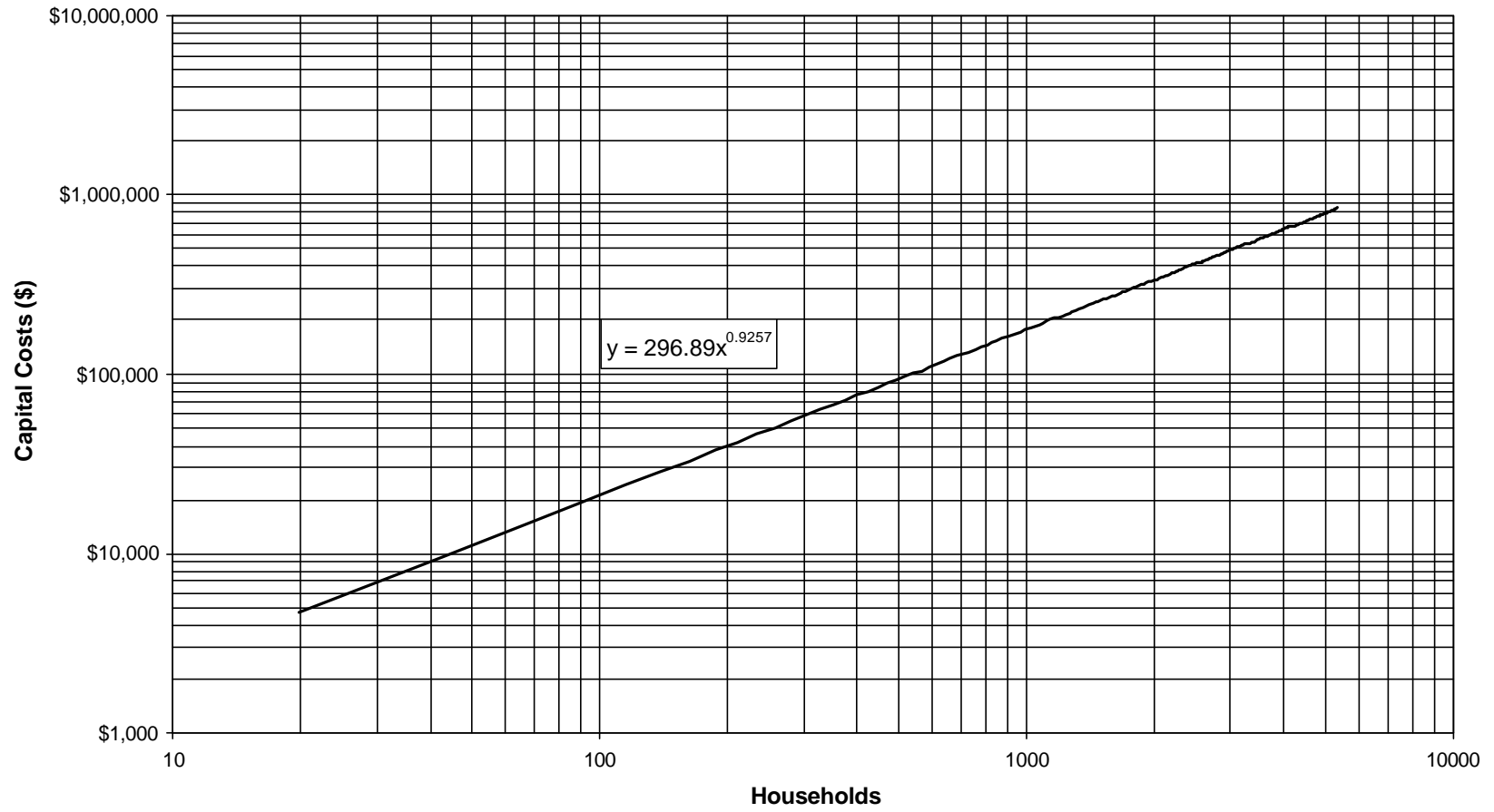
Figure 5-4
POU Activated Alumina
O&M Costs

$y = 412.61x^{0.9376}$

Axis labels: O&M Costs ($) (vertical), Households (horizontal)
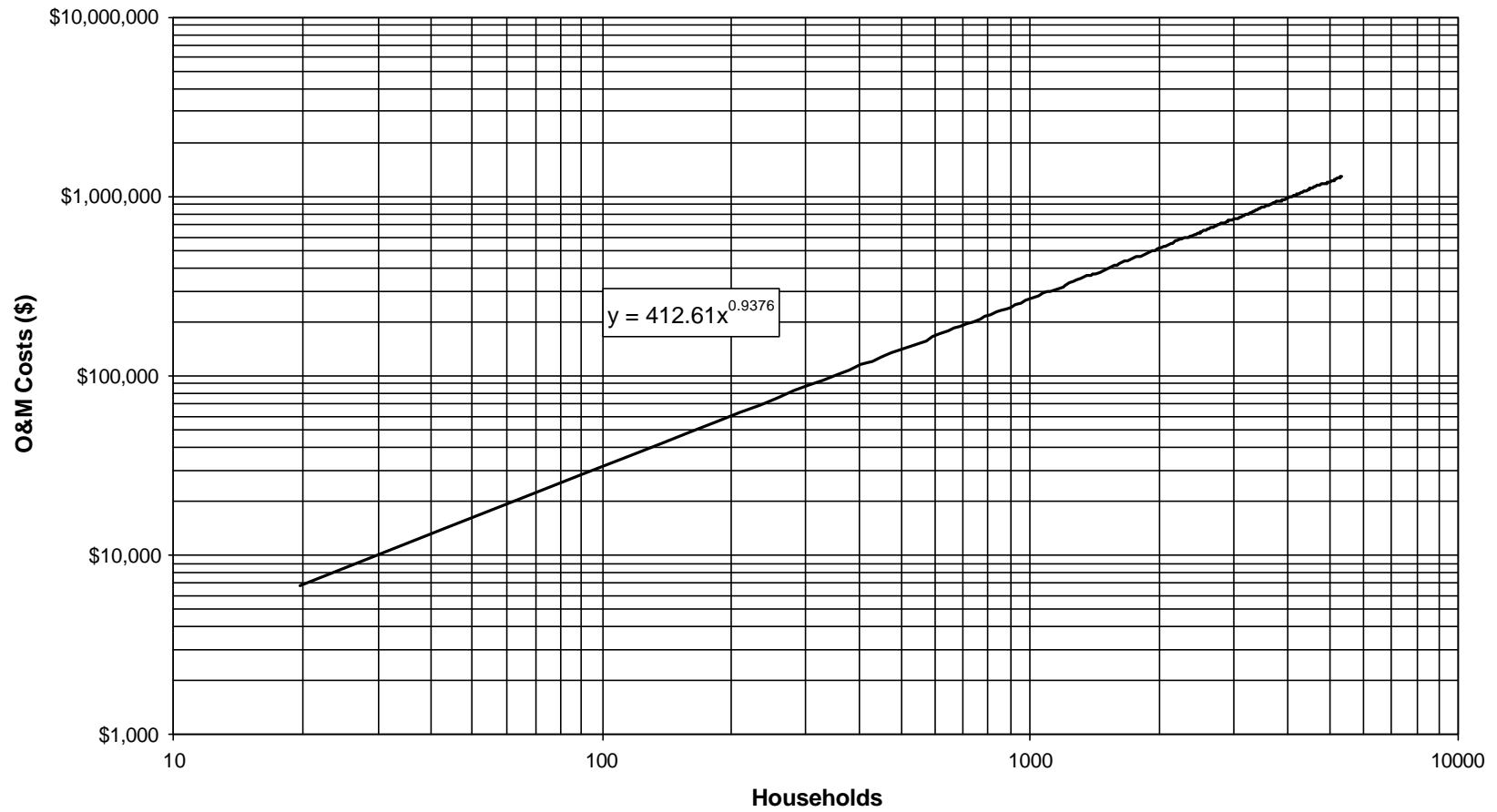
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Appendix A   Very Small Systems Capital Cost Breakdown Summaries

Table A1 - VSS Document Capital Cost Breakdown for Membrane Processes

Component	Capital Cost Factor	Percent of Total Capital Cost	Capital Cost Breakdown Category
Equipment	1.0000	56.97%	p
Installation	0.2500	14.24%	c
Sitework/Interface Piping	0.0750	4.27%	c
Standby Power	0.0625	3.56%	c
OH&P	0.1665	9.49%	e
Legal & Admin	0.0416	2.37%	e
Engineering	0.1596	9.09%	e
Contigencies	0.0000	0.00%	c
Total	1.7552	100.00%	

Table A2 - VSS Document Capital Cost Breakdown for Ion Exchange Processes

Component	Capital Cost Factor	Percent of Total Capital Cost	Capital Cost Breakdown Category
Equipment	1.0000	54.78%	p
Installation	0.3000	16.43%	c
Sitework/Interface Piping	0.0780	4.27%	c
Standby Power	0.0650	3.56%	c
OH&P	0.1732	9.49%	e
Legal & Admin	0.0433	2.37%	e
Engineering	0.1659	9.09%	e
Contigencies	0.0000	0.00%	c
Total	1.8254	100.00%	

Table A3 - VSS Document Capital Cost Breakdown for Chlorination

Component	Capital Cost Factor	Percent of Total Capital Cost	Capital Cost Breakdown Category
Equipment	1.0000	61.93%	p
Installation	0.1500	9.29%	c
Sitework/Interface Piping	0.0690	4.27%	c
Standby Power	0.0575	3.56%	c
OH&P	0.1532	9.49%	e
Legal & Admin	0.0383	2.37%	e
Engineering	0.1468	9.09%	e
Contigencies	0.0000	0.00%	c
Total	1.6148	100.00%	

Table A4 - VSS Document Capital Cost Breakdown for Potassium Permanganate Feed

Component	Capital Cost Factor	Percent of Total Capital Cost	Capital Cost Breakdown Category
Equipment	1.0000	64.74%	p
Installation	0.1000	6.47%	c
Sitework/Interface Piping	0.0660	4.27%	c
Standby Power	0.0550	3.56%	c
OH&P	0.1465	9.49%	e
Legal & Admin	0.0366	2.37%	e
Engineering	0.1404	9.09%	e
Contigencies	0.0000	0.00%	c
Total	1.5446	100.00%	

Table A5 - Typical VSS Document Capital Cost Breakdown

Component	Capital Cost Factor	Percent of Total Capital Cost	Capital Cost Breakdown Category
Equipment	1.0000	54.78%	p
Installation	0.3000	16.43%	c
Sitework/Interface Piping	0.0780	4.27%	c
Standby Power	0.0650	3.56%	c
OH&P	0.1732	9.49%	e
Legal & Admin	0.0433	2.37%	e
Engineering	0.1659	9.09%	e
Contigencies	0.0000	0.00%	c
Total	1.8254	100.00%	

Appendix B   Water Model Capital Cost Breakdown Summaries

Table B1.1 - Base Costs Obtained from the Water Model for Activated Alumina

Cost Component	Contactor Volume (ft3)							Capital Cost Category
	32	71	126	283	385	502	754	
Excavation & Sitework	$4,700	$4,700	$4,700	$4,700	$4,700	$4,700	$4,700	c
Manufactured Equipment	$12,800	$23,900	$39,100	$50,600	$64,500	$72,900	$101,000	p
Activated Alumina	$1,400	$3,100	$5,400	$11,900	$15,400	$19,600	$29,400	p
Concrete	$400	$1,200	$1,800	$2,000	$2,500	$3,200	$4,100	p
Labor	$1,200	$1,500	$2,000	$2,800	$3,300	$3,400	$4,200	c
Pipes and Valves	$5,200	$6,500	$6,500	$8,400	$12,800	$13,300	$20,100	p
Electrical	$6,400	$6,400	$6,400	$8,000	$8,000	$8,500	$9,600	p
Housing	$8,700	$14,400	$16,900	$17,900	$24,800	$34,400	$43,900	p
Subtotal	$40,800	$61,700	$82,800	$106,300	$136,000	$160,000	$217,000	
Contingencies	$6,100	$9,300	$12,400	$15,900	$20,400	$24,000	$32,600	c
Total	$46,900	$71,000	$95,200	$122,200	$156,400	$184,000	$249,600	

Table B1.2 - Water Model Base Construction Cost Analysis for Activated Alumina

Cost Component	Contactor Volume (ft3)							Average Percent
	32	71	126	283	385	502	754	
Excavation & Sitework	10.02%	6.62%	4.94%	3.85%	3.01%	2.55%	1.88%	4.70%
Manufactured Equipment	27.29%	33.66%	41.07%	41.41%	41.24%	39.62%	40.46%	37.82%
Activated Alumina	2.99%	4.37%	5.67%	9.74%	9.85%	10.65%	11.78%	7.86%
Concrete	0.85%	1.69%	1.89%	1.64%	1.60%	1.74%	1.64%	1.58%
Labor	2.56%	2.11%	2.10%	2.29%	2.11%	1.85%	1.68%	2.10%
Pipes and Valves	11.09%	9.15%	6.83%	6.87%	8.18%	7.23%	8.05%	8.20%
Electrical	13.65%	9.01%	6.72%	6.55%	5.12%	4.62%	3.85%	7.07%
Housing	18.55%	20.28%	17.75%	14.65%	15.86%	18.70%	17.59%	17.62%
Contingencies	13.01%	13.10%	13.03%	13.01%	13.04%	13.04%	13.06%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B2.1 - Base Costs Obtained from the Water Model for Anion Exchange

Cost Component	Resin Volume (ft3)						Capital Cost Category
	4	17	54	188	280	520	
Excavation & Sitework	$2,100	$2,100	$4,400	$4,400	$4,400	$5,300	c
Manufactured Equipment	$3,100	$8,600	$23,100	$64,100	$96,800	$164,800	p
Concrete	$300	$400	$5,500	$5,800	$6,000	$8,400	p
Steel	$0	$0	$7,800	$7,800	$7,800	$10,900	p
Labor	$400	$1,100	$12,100	$12,800	$12,900	$17,200	c
Pipes and Valves	$800	$800	$1,000	$2,600	$2,600	$3,100	p
Electrical	$3,100	$3,100	$3,100	$3,100	$3,100	$3,100	p
Housing	$5,600	$9,600	$11,100	$16,600	$19,200	$25,000	p
Subtotal	$15,400	$25,700	$68,100	$117,200	$152,800	$237,800	
Contingencies	$2,300	$3,900	$10,200	$17,600	$22,900	$35,700	c
Total	$17,700	$29,600	$78,300	$134,800	$175,700	$273,500	

Table B2.2 - Water Model Base Construction Cost Analysis for Anion Exchange

Cost Component	Resin Volume (ft3)						Average Percent
	4	17	54	188	280	520	
Excavation & Sitework	11.86%	7.09%	5.62%	3.26%	2.50%	1.94%	5.38%
Manufactured Equipment	17.51%	29.05%	29.50%	47.55%	55.09%	60.26%	39.83%
Concrete	1.69%	1.35%	7.02%	4.30%	3.41%	3.07%	3.48%
Steel	0.00%	0.00%	9.96%	5.79%	4.44%	3.99%	4.03%
Labor	2.26%	3.72%	15.45%	9.50%	7.34%	6.29%	7.43%
Pipes and Valves	4.52%	2.70%	1.28%	1.93%	1.48%	1.13%	2.17%
Electrical	17.51%	10.47%	3.96%	2.30%	1.76%	1.13%	6.19%
Housing	31.64%	32.43%	14.18%	12.31%	10.93%	9.14%	18.44%
Contingencies	12.99%	13.18%	13.03%	13.06%	13.03%	13.05%	13.06%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B3.1 - Base Costs Obtained from the Water Model for Basic Chemical Feed

Cost Component	Maximum Feed Rate (lb/day)							Capital Cost Category
	0.1-10	25	50	100	250	500	1000	
Dissolving Tank	$290	$430	$640	$910	$1,830	$2,200	$4,400	p
Mixer	$180	$200	$200	$240	$410	$620	$620	p
Metering Pump	$430	$700	$750	$1,230	$1,600	$1,670	$1,820	p
Pipes and Valves	$180	$180	$220	$220	$280	$280	$420	p
Labor	$180	$180	$240	$260	$300	$330	$400	c
Electrical	$80	$100	$150	$200	$250	$300	$400	p
Subtotal	$1,340	$1,790	$2,200	$3,060	$4,670	$5,400	$8,060	
Contingencies	$200	$270	$330	$460	$700	$810	$1,210	c
Total	$1,540	$2,060	$2,530	$3,520	$5,370	$6,210	$9,270	

Table B3.2 - Water Model Base Construction Cost Analysis for Basic Chemical Feed

Cost Component	Maximum Feed Rate (lb/day)							Average Percent
	0.1-10	25	50	100	250	500	1000	
Dissolving Tank	18.83%	20.87%	25.30%	25.85%	34.08%	35.43%	47.46%	29.69%
Mixer	11.69%	9.71%	7.91%	6.82%	7.64%	9.98%	6.69%	8.63%
Metering Pump	27.92%	33.98%	29.64%	34.94%	29.80%	26.89%	19.63%	28.97%
Pipes and Valves	11.69%	8.74%	8.70%	6.25%	5.21%	4.51%	4.53%	7.09%
Labor	11.69%	8.74%	9.49%	7.39%	5.59%	5.31%	4.31%	7.50%
Electrical	5.19%	4.85%	5.93%	5.68%	4.66%	4.83%	4.31%	5.07%
Contingencies	12.99%	13.11%	13.04%	13.07%	13.04%	13.04%	13.05%	13.05%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B4.1 - Base Costs Obtained from the Water Model for Chlorination

Cost Component	Cost	Capital Cost Category
Excavation & Sitework	$1,200	c
Manufactured Equipment	$2,700	p
Concrete	$300	p
Labor	$400	c
Pipes and Valves	$500	p
Electrical	$2,200	p
Housing	$7,800	p
Subtotal	$15,100	
Contingencies	$2,300	c
Total	$17,400	

Table B4.2 - Water Model Base Construction Cost Analysis for Chlorination

Cost Component	Cost	Average Percent
Excavation & Sitework	6.90%	6.90%
Manufactured Equipment	15.52%	15.52%
Concrete	1.72%	1.72%
Labor	2.30%	2.30%
Pipes and Valves	2.87%	2.87%
Electrical	12.64%	12.64%
Housing	44.83%	44.83%
Contingencies	13.22%	13.22%
Total	100.00%	100.00%

Table B5.1 - Base Costs Obtained from the Water Model for Underground Clearwell Storage

Cost Component	Design Capacity (gpd)					Capital Cost Category
	5,000	**10,000**	**50,000**	**100,000**	**500,000**	
Excavation & Sitework	$3,300	$5,700	$16,500	$25,300	$75,400	c
Concrete	$9,800	$16,500	$37,000	$64,000	$216,400	p
Steel	$300	$400	$500	$500	$600	p
Electrical	$2,600	$2,600	$2,600	$2,600	$2,600	p
Subtotal	$16,000	$25,200	$56,600	$92,400	$295,000	
Contingencies	$2,400	$3,800	$8,500	$13,900	$44,300	c
Total	$18,400	$29,000	$65,100	$106,300	$339,300	

Table B5.2 - Water Model Base Construction Cost Analysis for Underground Clearwell Storage

Cost Component	Design Capacity (gpd)					Average Percent
	5,000	**10,000**	**50,000**	**100,000**	**500,000**	
Excavation & Sitework	17.93%	19.66%	25.35%	23.80%	22.22%	21.79%
Concrete	53.26%	56.90%	56.84%	60.21%	63.78%	58.20%
Steel	1.63%	1.38%	0.77%	0.47%	0.18%	0.88%
Electrical	14.13%	8.97%	3.99%	2.45%	0.77%	6.06%
Contingencies	13.04%	13.10%	13.06%	13.08%	13.06%	13.07%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B6.1 - Base Costs Obtained from the Water Model for Package Conventional Treatment

Cost Component	Filter Area (ft2)						Capital Cost Category
	2	12	20	40	112	150	
Excavation & 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	c
Total	$82,100	$109,800	$133,700	$270,300	$392,600	$439,500	

Table B6.2 - Water Model Base Construction Cost Analysis for Package Conventional Treatment

Cost Component	Filter Area (ft2)						Average Percent
	2	12	20	40	112	150	
Excavation & 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 B7.1 - Base Costs Obtained from the Water Model for Ferric Chloride Feed

Cost Component	Maximum Feed Rate (lb/day)							Capital Cost Category
	1	10	25	50	100	250	750	
Storage Tank	$0	$0	$0	$0	$360	$780	$2,040	p
Wooden Stairway	$0	$0	$0	$0	$0	$300	$300	p
Metering Pump	$390	$390	$390	$390	$390	$1,090	$1,100	p
Pipes and Valves	$180	$180	$180	$180	$220	$280	$280	p
Labor	$120	$120	$130	$130	$210	$360	$410	c
Electrical	$80	$80	$80	$80	$100	$120	$120	p
Subtotal	$770	$770	$780	$780	$1,280	$2,930	$4,250	
Contingencies	$120	$120	$120	$120	$190	$440	$640	c
Total	$890	$890	$900	$900	$1,470	$3,370	$4,890	

Table B7.2 - Water Model Base Construction Cost Analysis for Ferric Chloride Feed

Cost Component	Maximum Feed Rate (lb/day)							Average Percent
	1	10	25	50	100	250	750	
Storage Tank	0.00%	0.00%	0.00%	0.00%	24.49%	23.15%	41.72%	12.76%
Wooden Stairway	0.00%	0.00%	0.00%	0.00%	0.00%	8.90%	6.13%	2.15%
Metering Pump	43.82%	43.82%	43.33%	43.33%	26.53%	32.34%	22.49%	36.53%
Pipes and Valves	20.22%	20.22%	20.00%	20.00%	14.97%	8.31%	5.73%	15.64%
Labor	13.48%	13.48%	14.44%	14.44%	14.29%	10.68%	8.38%	12.74%
Electrical	8.99%	8.99%	8.89%	8.89%	6.80%	3.56%	2.45%	6.94%
Contingencies	13.48%	13.48%	13.33%	13.33%	12.93%	13.06%	13.09%	13.24%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B8.1 - Base Costs Obtained from the Water Model for Package Lime Softening

Cost Component	Design Capacity (gpd)					Capital Cost Category
	15,000	**150,000**	**430,000**	**750,000**	**1,000,000**	
Excavation & Sitework	$3,500	$5,800	$6,700	$8,400	$9,800	c
Manufactured Equipment	$33,200	$49,800	$66,300	$86,200	$103,800	p
Concrete	$1,100	$2,500	$3,200	$5,900	$7,000	p
Labor	$14,000	$18,200	$28,000	$36,400	$43,800	c
Pipes and Valves	$5,200	$10,400	$14,100	$16,700	$45,900	p
Electrical	$8,500	$12,200	$17,000	$18,900	$26,700	p
Housing	$8,800	$16,400	$19,800	$30,000	$33,000	p
Subtotal	$74,300	$115,300	$155,100	$202,500	$270,000	
Contingencies	$11,100	$17,300	$23,300	$30,400	$40,500	c
Total	$85,400	$132,600	$178,400	$232,900	$310,500	

Table B8.2 - Water Model Base Construction Cost Analysis for Package Lime Softening

Cost Component	Design Capacity (gpd)					Average Percent
	15,000	**150,000**	**430,000**	**750,000**	**1,000,000**	
Excavation & Sitework	4.10%	4.37%	3.76%	3.61%	3.16%	3.80%
Manufactured Equipment	38.88%	37.56%	37.16%	37.01%	33.43%	36.81%
Concrete	1.29%	1.89%	1.79%	2.53%	2.25%	1.95%
Labor	16.39%	13.73%	15.70%	15.63%	14.11%	15.11%
Pipes and Valves	6.09%	7.84%	7.90%	7.17%	14.78%	8.76%
Electrical	9.95%	9.20%	9.53%	8.12%	8.60%	9.08%
Housing	10.30%	12.37%	11.10%	12.88%	10.63%	11.46%
Contingencies	13.00%	13.05%	13.06%	13.05%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B9.1 - Base Costs Obtained from the Water Model for Permanganate Feed

| Cost Component | Maximum Feed Rate (lb/day) | | | | | | Capital Cost Category |
	0.5-5	12.5	25	50	125	250	
Dissolving Tank	$290	$430	$640	$910	$1,830	$2,200	p
Mixer	$180	$200	$200	$240	$410	$620	p
Metering Pump	$430	$700	$750	$1,230	$1,600	$1,670	p
Pipes and Valves	$180	$180	$220	$220	$280	$280	p
Labor	$180	$180	$240	$260	$300	$330	c
Electrical	$80	$100	$150	$200	$250	$300	p
Subtotal	$1,340	$1,790	$2,200	$3,060	$4,670	$5,400	
Contingencies	$200	$270	$330	$460	$700	$810	c
Total	$1,540	$2,060	$2,530	$3,520	$5,370	$6,210	

Table B9.2 - Water Model Base Construction Cost Analysis for Permanganate Feed

| Cost Component | Maximum Feed Rate (lb/day) | | | | | | Average Percent |
	0.5-5	12.5	25	50	125	250	
Excavation & Sitework	18.83%	20.87%	25.30%	25.85%	34.08%	35.43%	26.73%
Manufactured Equipment	11.69%	9.71%	7.91%	6.82%	7.64%	9.98%	8.96%
Concrete	27.92%	33.98%	29.64%	34.94%	29.80%	26.89%	30.53%
Labor	11.69%	8.74%	8.70%	6.25%	5.21%	4.51%	7.52%
Pipes and Valves	11.69%	8.74%	9.49%	7.39%	5.59%	5.31%	8.03%
Electrical	5.19%	4.85%	5.93%	5.68%	4.66%	4.83%	5.19%
Contingencies	12.99%	13.11%	13.04%	13.07%	13.04%	13.04%	13.05%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B10.1 - Base Costs Obtained from the Water Model for Polymer Feed

| Cost Component | Maximum Feed Rate (lb/day) | | | | | Capital Cost Category |
	0.6	1	2.1	4.2	10.4	
Mixing Tank	$290	$430	$640	$910	$1,830	p
Mixer	$850	$850	$200	$1,050	$1,050	p
Metering Pump	$640	$700	$750	$1,230	$1,600	p
Pipes and Valves	$180	$180	$220	$220	$280	p
Labor	$180	$180	$240	$260	$300	c
Electrical	$80	$100	$150	$200	$250	p
Subtotal	$2,220	$2,440	$2,200	$3,870	$5,310	
Contingencies	$330	$370	$330	$580	$800	c
Total	$2,550	$2,810	$2,530	$4,450	$6,110	

Table B10.2 - Water Model Base Construction Cost Analysis for Polymer Feed

| Cost Component | Maximum Feed Rate (lb/day) | | | | | Average Percent |
	0.6	1	2.1	4.2	10.4	
Mixing Tank	11.37%	15.30%	25.30%	20.45%	29.95%	20.47%
Mixer	33.33%	30.25%	7.91%	23.60%	17.18%	22.45%
Metering Pump	25.10%	24.91%	29.64%	27.64%	26.19%	26.70%
Pipes and Valves	7.06%	6.41%	8.70%	4.94%	4.58%	6.34%
Labor	7.06%	6.41%	9.49%	5.84%	4.91%	6.74%
Electrical	3.14%	3.56%	5.93%	4.49%	4.09%	4.24%
Contingencies	12.94%	13.17%	13.04%	13.03%	13.09%	13.06%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B11.1 - Base Costs Obtained from the Water Model for Raw Water Pumping

Cost Component	Design Capacity (gpd)					Capital Cost Category
	28,800	144,000	504,000	720,000	1,008,000	
Excavation & Sitework	$11,700	$11,700	$12,300	$12,300	$12,800	c
Manufactured Equipment	$6,600	$7,800	$11,800	$12,600	$16,500	p
Concrete	$500	$500	$1,100	$1,100	$1,500	p
Labor	$3,700	$3,800	$5,800	$6,200	$8,500	c
Pipes and Valves	$1,500	$1,800	$2,700	$3,600	$4,500	p
Electrical	$800	$800	$1,400	$1,600	$2,100	p
Subtotal	$24,800	$26,400	$35,100	$37,400	$45,900	
Contingencies	$3,700	$4,000	$5,300	$5,600	$6,900	c
Total	$28,500	$30,400	$40,400	$43,000	$52,800	

Table B11.2 - Water Model Base Construction Cost Analysis for Raw Water Pumping

Cost Component	Design Capacity (gpd)					Average Percent
	28,800	144,000	504,000	720,000	1,008,000	
Excavation & Sitework	41.05%	38.49%	30.45%	28.60%	24.24%	32.57%
Manufactured Equipment	23.16%	25.66%	29.21%	29.30%	31.25%	27.72%
Concrete	1.75%	1.64%	2.72%	2.56%	2.84%	2.30%
Labor	12.98%	12.50%	14.36%	14.42%	16.10%	14.07%
Pipes and Valves	5.26%	5.92%	6.68%	8.37%	8.52%	6.95%
Electrical	2.81%	2.63%	3.47%	3.72%	3.98%	3.32%
Contingencies	12.98%	13.16%	13.12%	13.02%	13.07%	13.07%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B12.1 - Base Costs Obtained from the Water Model for Package Reverse Osmosis

| Cost Component | Plant Capacity (gpd) | | | | | | Capital Cost Category |
	2,500	10,000	50,000	100,000	500,000	1,000,000	
Manufactured Equipment	$20,300	$30,000	$69,600	$123,000	$454,800	$877,400	p
Labor	$800	$1,200	$1,500	$2,800	$7,500	$14,600	c
Electrical	$3,200	$4,600	$10,700	$18,700	$45,900	$62,100	p
Housing	$11,900	$13,900	$16,400	$18,500	$38,400	$52,500	p
Subtotal	$36,200	$49,700	$98,200	$163,000	$546,600	$1,006,600	
Contingencies	$5,400	$7,500	$14,700	$24,500	$82,000	$151,000	c
Total	$41,600	$57,200	$112,900	$187,500	$628,600	$1,157,600	

Table B12.2 - Water Model Base Construction Cost Analysis for Package Reverse Osmosis

| Cost Component | Plant Capacity (gpd) | | | | | | Average Percent |
	2,500	10,000	50,000	100,000	500,000	1,000,000	
Manufactured Equipment	48.80%	52.45%	61.65%	65.60%	72.35%	75.79%	62.77%
Labor	1.92%	2.10%	1.33%	1.49%	1.19%	1.26%	1.55%
Electrical	7.69%	8.04%	9.48%	9.97%	7.30%	5.36%	7.98%
Housing	28.61%	24.30%	14.53%	9.87%	6.11%	4.54%	14.66%
Contingencies	12.98%	13.11%	13.02%	13.07%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B13.1 - Base Costs Obtained from the Water Model for Sodium Hydroxide Feed

Cost Component	Maximum Feed Rate (lb/day)							Capital Cost Category
	0.8	4	8	42	83	417	834	
Storage and Feed Tanks	$60	$60	$90	$970	$2,040	$3,560	$6,940	p
Heating and Insulation	$0	$0	$0	$200	$410	$950	$1,620	p
Mixer	$0	$0	$0	$180	$240	$620	$640	p
Stairway	$0	$0	$0	$0	$0	$300	$600	p
Man. Transfer Pump	$100	$100	$100	$0	$0	$0	$0	p
Pipes and Valves	$310	$310	$310	$470	$470	$530	$790	p
Metering Pump	$390	$390	$390	$390	$410	$1,090	$1,100	p
Containment Wall	$120	$120	$150	$270	$400	$600	$880	p
Labor	$280	$280	$280	$420	$480	$650	$860	c
Electrical	$80	$80	$80	$100	$100	$120	$120	p
Subtotal	$1,340	$1,340	$1,400	$3,000	$4,550	$8,420	$13,550	
Contingencies	$200	$200	$210	$450	$680	$1,260	$2,030	c
Total	$1,540	$1,540	$1,610	$3,450	$5,230	$9,680	$15,580	

Table B13.2 - Water Model Base Construction Cost Analysis for Sodium Hydroxide Feed

Cost Component	Maximum Feed Rate (lb/day)							Average Percent
	0.8	4	8	42	83	417	834	
Storage and Feed Tanks	3.90%	3.90%	5.59%	28.12%	39.01%	36.78%	44.54%	23.12%
Heating and Insulation	0.00%	0.00%	0.00%	5.80%	7.84%	9.81%	10.40%	4.84%
Mixer	0.00%	0.00%	0.00%	5.22%	4.59%	6.40%	4.11%	2.90%
Stairway	0.00%	0.00%	0.00%	0.00%	0.00%	3.10%	3.85%	0.99%
Man. Transfer Pump	6.49%	6.49%	6.21%	0.00%	0.00%	0.00%	0.00%	2.74%
Pipes and Valves	20.13%	20.13%	19.25%	13.62%	8.99%	5.48%	5.07%	13.24%
Metering Pump	25.32%	25.32%	24.22%	11.30%	7.84%	11.26%	7.06%	16.05%
Containment Wall	7.79%	7.79%	9.32%	7.83%	7.65%	6.20%	5.65%	7.46%
Labor	18.18%	18.18%	17.39%	12.17%	9.18%	6.71%	5.52%	12.48%
Electrical	5.19%	5.19%	4.97%	2.90%	1.91%	1.24%	0.77%	3.17%
Contingencies	12.99%	12.99%	13.04%	13.04%	13.00%	13.02%	13.03%	13.02%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table B14.1 - Base Costs Obtained from the Water Model for Package Ultrafiltration

Cost Component	Membrane Area (ft2)						Capital Cost Category
	30	424	1,431	3,604	7,155	14,310	
Excavation & Sitework	$1,300	$2,400	$4,100	$5,700	$10,200	$14,900	c
Manufactured Equipment	$5,500	$25,300	$65,600	$129,800	$23,900	$415,100	p
Concrete	$1,800	$3,700	$5,800	$10,200	$16,700	$28,800	p
Labor	$1,100	$5,200	$13,500	$26,900	$49,500	$85,900	c
Pipes and Valves	$500	$1,100	$2,200	$3,800	$4,500	$6,200	p
Electrical	$1,500	$5,600	$13,300	$25,800	$48,100	$85,300	p
Housing	$7,800	$14,600	$21,700	$29,000	$40,800	$56,000	p
Subtotal	$19,500	$57,900	$126,200	$231,200	$193,700	$692,200	
Contingencies	$2,900	$8,700	$18,900	$34,700	$29,100	$103,800	c
Total	$22,400	$66,600	$145,100	$265,900	$222,800	$796,000	

Table B14.2 - Water Model Base Construction Cost Analysis for Package Ultrafiltration

Cost Component	Membrane Area (ft2)						Average Percent
	30	424	1,431	3,604	7,155	14,310	
Excavation & Sitework	5.80%	3.60%	2.83%	2.14%	4.58%	1.87%	3.47%
Manufactured Equipment	24.55%	37.99%	45.21%	48.82%	10.73%	52.15%	36.57%
Concrete	8.04%	5.56%	4.00%	3.84%	7.50%	3.62%	5.42%
Labor	4.91%	7.81%	9.30%	10.12%	22.22%	10.79%	10.86%
Pipes and Valves	2.23%	1.65%	1.52%	1.43%	2.02%	0.78%	1.60%
Electrical	6.70%	8.41%	9.17%	9.70%	21.59%	10.72%	11.05%
Housing	34.82%	21.92%	14.96%	10.91%	18.31%	7.04%	17.99%
Contingencies	12.95%	13.06%	13.03%	13.05%	13.06%	13.04%	13.03%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Appendix C W/W Cost Model Capital Cost Breakdown Summaries

Table C1.1 - Base Costs Obtained from the WATERCO$T Model for Activated Alumina

Cost Component	Plant Capacity (mgd)						Capital Cost Category
	0.7	2.0	6.8	27	54	135	
Manufactured Equipment	$26,760	$44,580	$138,330	$522,210	$1,031,270	$2,564,560	p
Activated Alumina	$8,300	$14,770	$83,080	$332,310	$664,610	$1,661,530	p
Labor	$10,280	$13,490	$48,010	$192,020	$384,060	$1,282,370	c
Pipes and Valves	$16,260	$19,320	$69,030	$273,210	$542,650	$1,368,060	p
Electrical	$10,050	$11,360	$22,300	$60,300	$119,030	$284,750	p
Housing	$6,960	$27,630	$62,120	$210,980	$374,840	$744,320	p
Contingencies	$11,790	$19,670	$63,430	$238,650	$467,470	$1,185,840	c
Total	$90,400	$150,820	$486,300	$1,829,680	$3,583,930	$9,091,430	

Table C1.2 - WATERCO$T Model Base Construction Cost Analysis for Activated Alumina

Cost Component	Plant Capacity (mgd)						Average Percent
	0.7	2.0	6.8	27	54	135	
Manufactured Equipment	29.60%	29.56%	28.45%	28.54%	28.77%	28.21%	28.86%
Activated Alumina	9.18%	9.79%	17.08%	18.16%	18.54%	18.28%	15.17%
Labor	11.37%	8.94%	9.87%	10.49%	10.72%	14.11%	10.92%
Pipes and Valves	17.99%	12.81%	14.19%	14.93%	15.14%	15.05%	15.02%
Electrical	11.12%	7.53%	4.59%	3.30%	3.32%	3.13%	5.50%
Housing	7.70%	18.32%	12.77%	11.53%	10.46%	8.19%	11.49%
Contingencies	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C2.1 - Base Costs Obtained from the WATERCO$T Model for Ammonia Feed Systems

Cost Component	Feed Capacity (lb/day)					Capital Cost Category
	250	500	1,000	2,500	5,000	
Manufactured Equipment	$13,260	$19,520	$30,450	$38,830	$59,200	p
Labor	$3,990	$5,680	$9,250	$10,620	$13,870	c
Pipes and Valves	$2,390	$3,520	$5,500	$7,000	$10,670	p
Electrical	$3,250	$3,770	$6,180	$8,480	$10,990	p
Housing	$4,500	$4,500	$4,500	$4,500	$6,430	p
Contingencies	$4,110	$5,550	$8,380	$10,410	$15,170	c
Total	$31,500	$42,540	$64,260	$79,840	$116,330	

Table C2.2 - WATERCO$T Model Base Construction Cost Analysis for Ammonia Feed Systems

Cost Component	Feed Capacity (lb/day)					Average Percent
	250	500	1,000	2,500	5,000	
Manufactured Equipment	42.10%	45.89%	47.39%	48.63%	50.89%	46.98%
Labor	12.67%	13.35%	14.39%	13.30%	11.92%	13.13%
Pipes and Valves	7.59%	8.27%	8.56%	8.77%	9.17%	8.47%
Electrical	10.32%	8.86%	9.62%	10.62%	9.45%	9.77%
Housing	14.29%	10.58%	7.00%	5.64%	5.53%	8.61%
Contingencies	13.05%	13.05%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C3.1 - Base Costs Obtained from the WATERCO$T Model for Backwash Water Pumping

| Cost Component | Pumping Capacity (mgd(gpm)) | | | | | Capital Cost Category |
	1,260 (1.8)	3,150 (4.5)	6,300 (9.1)	18,000 (25.9)	22,950 (33)	
Manufactured Equipment	$11,400	$14,600	$38,380	$76,780	$95,970	p
Labor	$3,050	$4,410	$4,880	$9,290	$12,440	c
Pipes and Valves	$9,780	$17,690	$17,690	$33,390	$44,780	p
Electrical	$13,350	$16,040	$16,740	$28,070	$33,250	p
Contingencies	$5,640	$7,910	$11,650	$22,130	$27,970	c
Total	$43,220	$60,650	$89,340	$169,660	$214,410	

Table C3.2 - WATERCO$T Model Base Construction Cost Analysis for Backwash Water Pumping

| Cost Component | Pumping Capacity (mgd(gpm)) | | | | | Average Percent |
	1,260 (1.8)	3,150 (4.5)	6,300 (9.1)	18,000 (25.9)	22,950 (33)	
Manufactured Equipment	26.38%	24.07%	42.96%	45.26%	44.76%	36.68%
Labor	7.06%	7.27%	5.46%	5.48%	5.80%	6.21%
Pipes and Valves	22.63%	29.17%	19.80%	19.68%	20.89%	22.43%
Electrical	30.89%	26.45%	18.74%	16.54%	15.51%	21.63%
Contingencies	13.05%	13.04%	13.04%	13.04%	13.05%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C4.1 - Base Costs Obtained from the WATERCO$T Model for Chemical Sludge Pumping

Cost Component	Capacity (gpm)						Capital Cost Category
	20	100	500	1,000	5,000	10,000	
Excavation & Sitework	$470	$600	$810	$970	$1,840	$2,220	c
Manufactured Equipment	$4,370	$6,230	$8,210	$10,390	$23,320	$38,440	p
Concrete	$1,500	$2,210	$3,220	$4,100	$9,270	$12,310	p
Steel	$1,510	$2,130	$3,120	$3,940	$8,640	$11,070	p
Labor	$5,280	$8,060	$12,880	$17,400	$47,850	$64,720	c
Pipes and Valves	$2,560	$4,570	$10,870	$18,190	$42,810	$79,060	p
Electrical	$6,290	$7,390	$7,880	$9,380	$10,380	$12,510	p
Housing	$5,880	$5,880	$5,880	$8,100	$8,100	$11,700	p
Contingencies	$4,180	$5,560	$7,930	$10,870	$22,830	$34,800	c
Total	$32,040	$42,630	$60,800	$83,340	$175,040	$266,830	

Table C4.2 - WATERCO$T Model Base Construction Cost Analysis for Chemical Sludge Pumping

Cost Component	Capacity (gpm)						Average Percent
	20	100	500	1,000	5,000	10,000	
Excavation & Sitework	1.47%	1.41%	1.33%	1.16%	1.05%	0.83%	1.21%
Manufactured Equipment	13.64%	14.61%	13.50%	12.47%	13.32%	14.41%	13.66%
Concrete	4.68%	5.18%	5.30%	4.92%	5.30%	4.61%	5.00%
Steel	4.71%	5.00%	5.13%	4.73%	4.94%	4.15%	4.78%
Labor	16.48%	18.91%	21.18%	20.88%	27.34%	24.26%	21.51%
Pipes and Valves	7.99%	10.72%	17.88%	21.83%	24.46%	29.63%	18.75%
Electrical	19.63%	17.34%	12.96%	11.26%	5.93%	4.69%	11.97%
Housing	18.35%	13.79%	9.67%	9.72%	4.63%	4.38%	10.09%
Contingencies	13.05%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C5.1 - Base Costs Obtained from the WATERCO$T Model for Chlorination

| Cost Component | Chlorine Feed Capacity (lb/day) | | | | | | Capital Cost Category |
	10	500	1,000	2,000	5,000	10,000	
Manufactured Equipment	$6,760	$21,630	$41,630	$65,950	$76,780	$114,360	p
Labor	$820	$2,610	$5,030	$7,960	$9,270	$13,810	c
Pipes and Valves	$540	$1,710	$3,300	$5,230	$6,080	$9,060	p
Electrical	$770	$2,450	$4,710	$7,460	$8,690	$12,940	p
Housing	$2,430	$18,360	$27,760	$46,550	$100,440	$186,490	p
Contingencies	$1,700	$7,010	$12,360	$19,970	$30,190	$50,500	c
Total	$13,020	$53,770	$94,790	$153,120	$231,450	$387,160	

Table C5.2 - WATERCO$T Model Base Construction Cost Analysis for Chlorination

| Cost Component | Chlorine Feed Capacity (lb/day) | | | | | | Average Percent |
	10	500	1,000	2,000	5,000	10,000	
Manufactured Equipment	51.92%	40.23%	43.92%	43.07%	33.17%	29.54%	40.31%
Labor	6.30%	4.85%	5.31%	5.20%	4.01%	3.57%	4.87%
Pipes and Valves	4.15%	3.18%	3.48%	3.42%	2.63%	2.34%	3.20%
Electrical	5.91%	4.56%	4.97%	4.87%	3.75%	3.34%	4.57%
Housing	18.66%	34.15%	29.29%	30.40%	43.40%	48.17%	34.01%
Contingencies	13.06%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C6.1 - Base Costs Obtained from the WATERCO\$T Model for Circular Clarifiers

Cost Component	Surface Area (SA=ft2) and Diameter (D=ft)							Capital Cost Category
	SA=707 D=30	SA=1,590 D=45	SA=5,027 D=80	SA=10,387 D=115	SA=15,393 D=140	SA=22,698 D=170	SA=31,416 D=200	
Excavation & Sitework	$1,530	$2,430	$4,900	$7,860	$10,280	$13,520	$17,130	c
Manufactured Equipment	$28,740	$34,410	$69,580	$97,180	$132,350	$189,060	$226,980	p
Concrete	$4,860	$7,710	$15,480	$24,800	$32,400	$42,560	$53,860	p
Steel	$14,160	$21,090	$67,240	$129,250	$188,720	$249,570	$335,140	p
Labor	$10,770	$16,180	$30,960	$46,980	$60,110	$77,640	$96,320	c
Pipes and Valves	$8,090	$8,420	$11,540	$15,660	$21,590	$26,590	$42,520	p
Electrical	$5,940	$5,940	$7,560	$8,270	$10,870	$12,370	$13,060	p
Contingencies	$11,110	$14,430	$31,090	$49,500	$68,450	$91,700	$117,750	c
Total	$85,200	$110,610	$238,350	$379,500	$524,770	$703,010	$902,760	

Table C6.2 - WATERCO\$T Model Base Construction Cost Analysis for Circular Clarifiers

Cost Component	Surface Area (SA=ft2) and Diameter (D=ft)							Average Percent
	SA=707 D=30	SA=1,590 D=45	SA=5,027 D=80	SA=10,387 D=115	SA=15,393 D=140	SA=22,698 D=170	SA=31,416 D=200	
Excavation & Sitework	1.80%	2.20%	2.06%	2.07%	1.96%	1.92%	1.90%	1.99%
Manufactured Equipment	33.73%	31.11%	29.19%	25.61%	25.22%	26.89%	25.14%	28.13%
Concrete	5.70%	6.97%	6.49%	6.53%	6.17%	6.05%	5.97%	6.27%
Steel	16.62%	19.07%	28.21%	34.06%	35.96%	35.50%	37.12%	29.51%
Labor	12.64%	14.63%	12.99%	12.38%	11.45%	11.04%	10.67%	12.26%
Pipes and Valves	9.50%	7.61%	4.84%	4.13%	4.11%	3.78%	4.71%	5.53%
Electrical	6.97%	5.37%	3.17%	2.18%	2.07%	1.76%	1.45%	3.28%
Contingencies	13.04%	13.05%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C7.1 - Base Costs Obtained from the WATERCO$T Model for Clearwell Storage

Cost Component	Capacity (gal)						Capital Cost Category
	10,000	50,000	100,000	500,000	1,000,000	7,500,000	
Excavation & Sitework	$140	$190	$410	$2,030	$19,440	$30,020	c
Concrete	$8,250	$14,430	$23,280	$66,330	$105,520	$622,500	p
Steel	$5,700	$9,240	$14,550	$32,670	$113,050	$350,700	p
Labor	$13,050	$21,480	$35,040	$84,090	$109,290	$394,160	p
Electrical	$1,270	$1,270	$6,010	$6,010	$9,800	$9,800	p
Contingencies	$4,260	$6,990	$11,890	$28,670	$53,570	$211,080	c
Total	$32,670	$53,600	$91,180	$219,800	$410,670	$1,618,260	

Table C7.2 - WATERCO$T Model Base Construction Cost Analysis for Clearwell Storage

Cost Component	Capacity (gal)						Average Percent
	10,000	50,000	100,000	500,000	1,000,000	7,500,000	
Excavation & Sitework	0.43%	0.35%	0.45%	0.92%	4.73%	1.86%	1.46%
Concrete	25.25%	26.92%	25.53%	30.18%	25.69%	38.47%	28.67%
Steel	17.45%	17.24%	15.96%	14.86%	27.53%	21.67%	19.12%
Labor	39.94%	40.07%	38.43%	38.26%	26.61%	24.36%	34.61%
Electrical	3.89%	2.37%	6.59%	2.73%	2.39%	0.61%	3.10%
Contingencies	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C8.1 - Base Costs Obtained from the WATERCO$T Model for Ferric Chloride Feed Systems*

Cost Component	Feed Capacity (lb/hr)				Capital Cost Category
	10.7	107	1,070	5,350	
Manufactured Equipment	$7,500	$13,100	$33,560	$160,940	p
Labor	$420	$1,130	$2,430	$12,160	c
Pipes and Valves	$2,000	$2,500	$3,000	$15,000	p
Electrical	$1,110	$2,260	$4,960	$19,000	p
Housing	$6,000	$13,300	$51,270	$174,590	p
Contingencies	$2,550	$4,840	$14,280	$57,250	c
Total	$19,580	$37,130	$109,500	$438,940	

*Numbers were unavailable for ferric chloride. However, numbers presented for ferrous sulfate and ferric   sulfate were identical.

It was assumed that these same relationships apply to ferric chloride

Table C8.2 - WATERCO$T Model Base Construction Cost Analysis for Ferric Chloride Feed Systems*

Cost Component	Feed Capacity (lb/hr)				Average Percent
	10.7	107	1,070	5,350	
Manufactured Equipment	38.30%	35.28%	30.65%	36.67%	35.22%
Labor	2.15%	3.04%	2.22%	2.77%	2.54%
Pipes and Valves	10.21%	6.73%	2.74%	3.42%	5.78%
Electrical	5.67%	6.09%	4.53%	4.33%	5.15%
Housing	30.64%	35.82%	46.82%	39.78%	38.27%
Contingencies	13.02%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

*Numbers were unavailable for ferric chloride. However, numbers presented for ferrous sulfate and ferric   sulfate were identical.

It was assumed that these same relationships apply to ferric chloride

Table C9.1 - Base Costs Obtained from the WATERCO$T Model for Finished Water Pumping

Cost Component	Plant Capacity (mgd)				Capital Cost Category
	1.5	15	150	300	
Manufactured Equipment	$15,410	$89,700	$567,600	$1,142,350	p
Labor	$3,880	$11,580	$80,400	$158,840	c
Pipes and Valves	$5,200	$16,570	$139,200	$270,100	p
Electrical	$7,180	$38,450	$210,490	$400,230	p
Contingencies	$4,750	$23,450	$149,650	$295,730	c
Total	$36,420	$179,750	$1,147,340	$2,267,250	

Table C9.2 - WATERCO$T Model Base Construction Cost Analysis for Finished Water Pumping

Cost Component	Plant Capacity (mgd)				Average Percent
	1.5	15	150	300	
Manufactured Equipment	42.31%	49.90%	49.47%	50.38%	48.02%
Labor	10.65%	6.44%	7.01%	7.01%	7.78%
Pipes and Valves	14.28%	9.22%	12.13%	11.91%	11.89%
Electrical	19.71%	21.39%	18.35%	17.65%	19.28%
Contingencies	13.04%	13.05%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C10.1 - Base Costs Obtained from the WATERCO\$T Model for Gravity Filtration

Cost Component	Total Filter Area (FA-ft2) and Plant Flow (Q=mgd)						Capital Cost Category
	FA=140 Q=1	FA=700 Q=5	FA=1,400 Q=10	FA=7,000 Q=50	FA=14,000 Q=100	FA=28,000 Q=200	
Excavation & Sitework	$1,950	$3,620	$5,520	$16,220	$25,590	$43,410	c
Manufactured Equipment	$26,360	$56,960	$78,300	$305,170	$529,360	$982,390	p
Concrete	$13,400	$27,040	$41,660	$95,490	$154,790	$275,570	p
Steel	$11,550	$19,960	$30,120	$73,530	$123,160	$209,960	p
Labor	$40,580	$88,490	$150,870	$356,380	$508,980	$1,000,670	c
Pipes and Valves	$20,580	$79,020	$127,340	$420,670	$590,150	$1,125,500	p
Electrical	$13,390	$38,410	$38,410	$99,140	$168,840	$265,310	p
Housing	$17,400	$40,480	$70,590	$291,940	$514,330	$968,520	p
Contingencies	$21,780	$53,100	$81,420	$248,780	$392,280	$730,700	c
Total	$166,990	$407,080	$624,230	$1,907,320	$3,007,480	$5,602,030	

Table C10.2 - WATERCO\$T Model Base Construction Cost Analysis for Gravity Filtration

Cost Component	Total Filter Area (FA-ft2) and Plant Flow (Q=mgd)						Average Percent
	FA=140 Q=1	FA=700 Q=5	FA=1,400 Q=10	FA=7,000 Q=50	FA=14,000 Q=100	FA=28,000 Q=200	
Excavation & Sitework	1.17%	0.89%	0.88%	0.85%	0.85%	0.77%	0.90%
Manufactured Equipment	15.79%	13.99%	12.54%	16.00%	17.60%	17.54%	15.58%
Concrete	8.02%	6.64%	6.67%	5.01%	5.15%	4.92%	6.07%
Steel	6.92%	4.90%	4.83%	3.86%	4.10%	3.75%	4.72%
Labor	24.30%	21.74%	24.17%	18.68%	16.92%	17.86%	20.61%
Pipes and Valves	12.32%	19.41%	20.40%	22.06%	19.62%	20.09%	18.98%
Electrical	8.02%	9.44%	6.15%	5.20%	5.61%	4.74%	6.53%
Housing	10.42%	9.94%	11.31%	15.31%	17.10%	17.29%	13.56%
Contingencies	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C11.1 - Base Costs Obtained from the WATERCO$T Model for Horizontal Paddle, G=50

Cost Component	Total Basin Volume (ft3)						Capital Cost Category
	1,800	10,000	25,000	100,000	500,000	1,000,000	
Excavation & Sitework	$470	$2,550	$4,290	$9,970	$40,080	$77,640	p
Manufactured Equipment	$12,140	$28,250	$35,410	$74,400	$220,800	$433,640	p
Concrete	$1,400	$7,610	$12,740	$29,770	$120,280	$232,960	p
Steel	$2,360	$12,550	$20,440	$46,500	$175,290	$339,510	p
Labor	$7,080	$20,220	$29,420	$75,460	$221,200	$439,770	c
Electrical	$6,980	$28,320	$28,320	$28,320	$141,610	$283,220	p
Contingencies	$4,560	$14,930	$19,590	$39,660	$137,890	$271,010	c
Total	$34,990	$114,430	$150,210	$304,080	$1,057,150	$2,077,750	

Table C11.2 - WATERCO$T Model Base Construction Cost Analysis for Horizontal Paddle, G=50

Cost Component	Total Basin Volume (ft3)						Average Percent
	1,800	10,000	25,000	100,000	500,000	1,000,000	
Excavation & Sitework	1.34%	2.23%	2.86%	3.28%	3.79%	3.74%	2.87%
Manufactured Equipment	34.70%	24.69%	23.57%	24.47%	20.89%	20.87%	24.86%
Concrete	4.00%	6.65%	8.48%	9.79%	11.38%	11.21%	8.59%
Steel	6.74%	10.97%	13.61%	15.29%	16.58%	16.34%	13.26%
Labor	20.23%	17.67%	19.59%	24.82%	20.92%	21.17%	20.73%
Electrical	19.95%	24.75%	18.85%	9.31%	13.40%	13.63%	16.65%
Contingencies	13.03%	13.05%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C12.1 - Base Costs Obtained from the WATERCO$T Model for Horizontal Paddle, G=80

| Cost Component | Total Basin Volume (ft3) | | | | | Capital Cost Category |
	1,800	10,000	25,000	100,000	500,000	
Excavation & Sitework	$470	$2,550	$4,290	$9,970	$40,080	c
Manufactured Equipment	$12,140	$34,210	$44,360	$115,770	$427,670	p
Concrete	$1,400	$7,610	$12,740	$29,770	$120,280	p
Steel	$2,360	$12,550	$20,440	$46,500	$175,290	p
Labor	$7,080	$22,190	$32,370	$90,170	$289,520	p
Electrical	$6,980	$28,320	$28,320	$28,320	$141,610	p
Contingencies	$4,560	$16,110	$21,380	$48,080	$179,170	c
Total	$34,990	$123,540	$163,900	$368,580	$1,373,620	

Table C12.2 - WATERCO$T Model Base Construction Cost Analysis for Horizontal Paddle, G=80

| Cost Component | Total Basin Volume (ft3) | | | | | Average Percent |
	1,800	10,000	25,000	100,000	500,000	
Excavation & Sitework	1.34%	2.06%	2.62%	2.70%	2.92%	2.33%
Manufactured Equipment	34.70%	27.69%	27.07%	31.41%	31.13%	30.40%
Concrete	4.00%	6.16%	7.77%	8.08%	8.76%	6.95%
Steel	6.74%	10.16%	12.47%	12.62%	12.76%	10.95%
Labor	20.23%	17.96%	19.75%	24.46%	21.08%	20.70%
Electrical	19.95%	22.92%	17.28%	7.68%	10.31%	15.63%
Contingencies	13.03%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C13.1 - Base Costs Obtained from the WATERCO$T Model for Hydraulic Surface Wash

Cost Component	Total Filter Area (ft2)						Capital Cost Category
	140	700	1,400	7,000	14,000	28,000	
Manufactured Equipment	$9,170	$12,050	$35,090	$82,010	$172,440	$401,200	p
Labor	$1,300	$2,770	$5,170	$14,710	$29,430	$66,600	c
Pipes and Valves	$2,570	$5,100	$7,020	$13,390	$32,290	$59,870	p
Electrical	$12,670	$17,920	$20,440	$37,900	$61,120	$92,360	p
Contingencies	$3,860	$5,680	$10,160	$22,200	$44,290	$93,000	c
Total	$29,570	$43,520	$77,880	$170,210	$339,570	$713,030	

Table C13.2 - WATERCO$T Model Base Construction Cost Analysis for Hydraulic Surface Wash

Cost Component	Total Filter Area (ft2)						Average Percent
	140	700	1,400	7,000	14,000	28,000	
Manufactured Equipment	31.01%	27.69%	45.06%	48.18%	50.78%	56.27%	43.16%
Labor	4.40%	6.36%	6.64%	8.64%	8.67%	9.34%	7.34%
Pipes and Valves	8.69%	11.72%	9.01%	7.87%	9.51%	8.40%	9.20%
Electrical	42.85%	41.18%	26.25%	22.27%	18.00%	12.95%	27.25%
Contingencies	13.05%	13.05%	13.05%	13.04%	13.04%	13.04%	13.05%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C14.1 - Base Costs Obtained from the WATERCO$T Model for In-Plant Pumping

Cost Component	Pumping Capacity (mgd)						Capital Cost Category
	1	5	10	50	100	200	
Excavation & Sitework	$100	$100	$130	$360	$600	$1,030	c
Manufactured Equipment	$6,300	$9,110	$14,780	$48,650	$83,400	$152,900	p
Concrete	$970	$970	$1,510	$4,770	$8,030	$14,090	p
Steel	$1,610	$1,610	$2,450	$7,630	$12,500	$21,330	p
Labor	$5,570	$10,410	$24,070	$63,330	$129,130	$331,030	c
Pipes and Valves	$5,090	$12,330	$16,300	$60,230	$114,200	$222,080	p
Electrical	$3,170	$4,930	$7,390	$25,760	$47,240	$89,360	p
Housing	$1,500	$1,500	$3,000	$14,520	$28,830	$58,080	p
Contingencies	$3,650	$6,140	$10,440	$33,790	$63,590	$133,490	c
Total	$27,960	$47,100	$80,070	$259,040	$487,520	$1,023,390	

Table C14.2 - WATERCO$T Model Base Construction Cost Analysis for In-Plant Pumping

Cost Component	Pumping Capacity (mgd)						Average Percent
	1	5	10	50	100	200	
Excavation & Sitework	0.36%	0.21%	0.16%	0.14%	0.12%	0.10%	0.18%
Manufactured Equipment	22.53%	19.34%	18.46%	18.78%	17.11%	14.94%	18.53%
Concrete	3.47%	2.06%	1.89%	1.84%	1.65%	1.38%	2.05%
Steel	5.76%	3.42%	3.06%	2.95%	2.56%	2.08%	3.31%
Labor	19.92%	22.10%	30.06%	24.45%	26.49%	32.35%	25.89%
Pipes and Valves	18.20%	26.18%	20.36%	23.25%	23.42%	21.70%	22.19%
Electrical	11.34%	10.47%	9.23%	9.94%	9.69%	8.73%	9.90%
Housing	5.36%	3.18%	3.75%	5.61%	5.91%	5.68%	4.92%
Contingencies	13.05%	13.04%	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C15.1 - Base Costs Obtained from the WATERCO$T Model for Ion Exchange

Cost Component	Plant Capacity (mgd)				Capital Cost Category
	1.1	3.7	6.1	12.3	
Excavation & Sitework	$740	$1,140	$1,470	$1,970	c
Manufactured Equipment	$39,960	$89,580	$137,770	$258,230	p
Media	$92,790	$313,160	$521,940	$1,043,880	p
Concrete	$2,410	$3,580	$4,750	$6,320	p
Steel	$3,830	$5,680	$7,530	$9,950	p
Labor	$17,420	$33,510	$61,460	$125,080	c
Pipes and Valves	$14,040	$38,780	$69,740	$139,480	p
Electrical	$27,700	$38,510	$60,820	$120,210	p
Housing	$21,920	$35,660	$57,440	$79,820	p
Contingencies	$33,120	$83,940	$138,440	$267,740	c
Total	$253,930	$643,540	$1,061,360	$2,052,680	

Table C15.2 - WATERCO$T Model Base Construction Cost Analysis for Ion Exchange

Cost Component	Plant Capacity (mgd)				Average Percent
	1.1	3.7	6.1	12.3	
Excavation & Sitework	0.29%	0.18%	0.14%	0.10%	0.18%
Manufactured Equipment	15.74%	13.92%	12.98%	12.58%	13.80%
Media	36.54%	48.66%	49.18%	50.85%	46.31%
Concrete	0.95%	0.56%	0.45%	0.31%	0.57%
Steel	1.51%	0.88%	0.71%	0.48%	0.90%
Labor	6.86%	5.21%	5.79%	6.09%	5.99%
Pipes and Valves	5.53%	6.03%	6.57%	6.80%	6.23%
Electrical	10.91%	5.98%	5.73%	5.86%	7.12%
Housing	8.63%	5.54%	5.41%	3.89%	5.87%
Contingencies	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C16.1 - Base Costs Obtained from the WATERCO$T Model for Lime Feed with Recalcination

Cost Component	Feed Capacity (lb/hr)		Capital Cost Category
	1,000	10,000	
Manufactured Equipment	$48,870	$80,660	p
Labor	$1,510	$3,060	c
Pipes and Valves	$3,120	$6,250	p
Electrical	$6,880	$12,320	p
Housing	$9,450	$26,250	p
Contingencies	$10,470	$19,280	c
Total	$80,300	$147,820	

Table C16.2 - WATERCO$T Model Base Construction Cost Analysis for Lime Feed with Recalcination

Cost Component	Feed Capacity (lb/hr)		Average Percent
	1,000	10,000	
Manufactured Equipment	60.86%	54.57%	57.71%
Labor	1.88%	2.07%	1.98%
Pipes and Valves	3.89%	4.23%	4.06%
Electrical	8.57%	8.33%	8.45%
Housing	11.77%	17.76%	14.76%
Contingencies	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%

Table C17.1 - Base Costs Obtained from the WATERCO$T Model for Permanganate Feed Systems

Cost Component	Feed Capacity (lb/day)				Capital Cost Category
	1	10	100	500	
Manufactured Equipment	$2,340	$2,600	$3,380	$5,220	p
Labor	$480	$480	$540	$770	c
Pipes and Valves	$970	$970	$970	$970	p
Electrical	$3,190	$3,190	$3,190	$3,190	p
Housing	$1,260	$1,580	$1,950	$2,940	p
Contingencies	$1,240	$1,320	$1,500	$1,960	c
Total	$9,480	$10,140	$11,530	$15,050	

Table C17.2 - WATERCO$T Model Base Construction Cost Analysis for Permanganate Feed Systems

Cost Component	Feed Capacity (lb/day)				Average Percent
	1	10	100	500	
Manufactured Equipment	24.68%	25.64%	29.31%	34.68%	28.58%
Labor	5.06%	4.73%	4.68%	5.12%	4.90%
Pipes and Valves	10.23%	9.57%	8.41%	6.45%	8.66%
Electrical	33.65%	31.46%	27.67%	21.20%	28.49%
Housing	13.29%	15.58%	16.91%	19.53%	16.33%
Contingencies	13.08%	13.02%	13.01%	13.02%	13.03%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C18.1 - Base Costs Obtained from the WATERCO$T Model for Polymer Feed Systems

Cost Component	Feed Capacity (lb/hr)				Capital Cost Category
	1	10	100	200	
Manufactured Equipment	$11,670	$11,670	$14,730	$18,970	p
Labor	$700	$700	$700	$760	c
Pipes and Valves	$280	$280	$280	$300	p
Electrical	$1,290	$1,290	$1,290	$1,290	p
Housing	$3,600	$3,600	$4,050	$4,500	p
Contingencies	$2,630	$2,630	$3,160	$3,870	c
Total	$20,170	$20,170	$24,210	$29,690	

Table C18.2 - WATERCO$T Model Base Construction Cost Analysis for Polymer Feed Systems

Cost Component	Feed Capacity (lb/hr)				Average Percent
	1	10	100	200	
Manufactured Equipment	57.86%	57.86%	60.84%	63.89%	60.11%
Labor	3.47%	3.47%	2.89%	2.56%	3.10%
Pipes and Valves	1.39%	1.39%	1.16%	1.01%	1.24%
Electrical	6.40%	6.40%	5.33%	4.34%	5.62%
Housing	17.85%	17.85%	16.73%	15.16%	16.90%
Contingencies	13.04%	13.04%	13.05%	13.03%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C19.1 - Base Costs Obtained from the WATERCO$T Model for Rapid Mix, G=900

Cost Component	Basin Volume (ft3)						Capital Cost Category
	100	500	1,000	5,000	10,000	20,000	
Excavation & Sitework	$220	$380	$490	$1,360	$2,720	$5,460	c
Manufactured Equipment	$4,310	$9,830	$14,760	$66,840	$133,670	$267,340	p
Concrete	$390	$870	$1,280	$3,610	$7,220	$14,450	p
Steel	$570	$1,350	$2,010	$5,600	$11,180	$22,360	p
Labor	$1,230	$2,300	$3,410	$13,140	$26,280	$52,550	c
Electrical	$6,980	$6,980	$7,180	$7,470	$8,760	$16,100	p
Contingencies	$2,060	$3,260	$4,370	$14,700	$28,470	$56,740	c
Total	$15,760	$24,970	$33,500	$112,720	$218,300	$435,000	

Table C19.2 - WATERCO$T Model Base Construction Cost Analysis for Rapid Mix, G=900

Cost Component	Basin Volume (ft3)						Average Percent
	100	500	1,000	5,000	10,000	20,000	
Excavation & Sitework	1.40%	1.52%	1.46%	1.21%	1.25%	1.26%	1.35%
Manufactured Equipment	27.35%	39.37%	44.06%	59.30%	61.23%	61.46%	48.79%
Concrete	2.47%	3.48%	3.82%	3.20%	3.31%	3.32%	3.27%
Steel	3.62%	5.41%	6.00%	4.97%	5.12%	5.14%	5.04%
Labor	7.80%	9.21%	10.18%	11.66%	12.04%	12.08%	10.50%
Electrical	44.29%	27.95%	21.43%	6.63%	4.01%	3.70%	18.00%
Contingencies	13.07%	13.06%	13.04%	13.04%	13.04%	13.04%	13.05%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C20.1 - Base Costs Obtained from the WATERCO$T Model for Recarbonation, Liquid Carbon Dioxide

Cost Component	Installed Capacity (lb/day)						Capital Cost Category
	380	750	1,500	3,750	7,500	15,000	
Manufactured Equipment	$27,000	$31,000	$35,250	$49,250	$73,000	$141,000	p
Labor	$7,650	$8,780	$12,170	$17,330	$28,990	$58,010	c
Pipes and Valves	$1,530	$2,340	$4,620	$8,710	$16,940	$37,540	p
Housing	$7,360	$7,360	$7,360	$7,360	$8,450	$8,900	p
Contingencies	$6,530	$7,420	$8,910	$12,400	$19,110	$36,820	c
Total	$50,070	$56,900	$68,310	$95,050	$146,490	$282,270	

Table C20.2 - WATERCO$T Model Base Construction Cost Analysis for Recarbonation, Liquid Carbon Dioxide

Cost Component	Installed Capacity (lb/day)						Average Percent
	380	750	1,500	3,750	7,500	15,000	
Manufactured Equipment	53.92%	54.48%	51.60%	51.81%	49.83%	49.95%	51.93%
Labor	15.28%	15.43%	17.82%	18.23%	19.79%	20.55%	17.85%
Pipes and Valves	3.06%	4.11%	6.76%	9.16%	11.56%	13.30%	7.99%
Housing	14.70%	12.93%	10.77%	7.74%	5.77%	3.15%	9.18%
Contingencies	13.04%	13.04%	13.04%	13.05%	13.05%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C21.1 - Base Costs Obtained from the WATERCO$T Model for Recarbonation Basins

Cost Component	Single Basin Volume (ft3)							Capital Cost Category
	770	1,375	2,750	5,630	8,800	17,600	35,200	
Excavation & Sitework	$520	$620	$980	$1,390	$1,790	$3,050	$5,570	c
Concrete	$1,380	$1,860	$2,820	$4,050	$5,190	$8,570	$15,320	p
Steel	$2,250	$3,010	$4,670	$6,560	$8,320	$13,960	$25,240	p
Labor	$2,830	$3,800	$5,730	$8,090	$10,240	$16,740	$29,730	c
Pipes and Valves	$90	$130	$250	$480	$680	$1,360	$3,360	p
Contingencies	$1,060	$1,410	$2,170	$3,090	$3,930	$6,550	$11,880	c
Total	$8,130	$10,830	$16,620	$23,660	$30,150	$50,230	$91,100	

Table C21.2 - WATERCO$T Model Base Construction Cost Analysis for Recarbonation Basins

Cost Component	Single Basin Volume (ft3)							Average Percent
	770	1,375	2,750	5,630	8,800	17,600	35,200	
Excavation & Sitework	6.40%	5.72%	5.90%	5.87%	5.94%	6.07%	6.11%	6.00%
Concrete	16.97%	17.17%	16.97%	17.12%	17.21%	17.06%	16.82%	17.05%
Steel	27.68%	27.79%	28.10%	27.73%	27.60%	27.79%	27.71%	27.77%
Labor	34.81%	35.09%	34.48%	34.19%	33.96%	33.33%	32.63%	34.07%
Pipes and Valves	1.11%	1.20%	1.50%	2.03%	2.26%	2.71%	3.69%	2.07%
Contingencies	13.04%	13.02%	13.06%	13.06%	13.03%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C22.1 - Base Costs Obtained from the WATERCO$T Model for Rectangular Clarifiers

| Cost Component | Area (A=ft2) and Length x Width (LW=ftxft) | | | | | | 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 & 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	$8,430	$12,820	$18,190	$22,070	p
Steel	$6,400	$13,110	$19,440	$32,620	$51,250	$69,680	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
Contingencies	$5,050	$7,970	$12,550	$18,460	$27,750	$34,860	c
Total	$38,710	$61,070	$96,230	$141,510	$212,750	$267,290	

Table C22.2 - WATERCO$T Model Base Construction Cost Analysis for Rectangular Clarifiers

| Cost Component | Area (A=ft2) and Length x Width (LW=ftxft) | | | | | | 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 & 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%	24.96%	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%

Table C23.1 - Base Costs Obtained from the WATERCO$T Model for Reverse Osmosis

| Cost Component | Plant Capacity (mgd) | | | | Capital Cost Category |
	1.0	10	100	200	
Manufactured Equipment	$474,210	$3,458,480	$29,174,260	$56,438,930	p
Labor	$70,420	$346,850	$2,312,340	$2,837,870	c
Electrical	$65,740	$486,270	$3,635,690	$6,947,480	p
Housing	$64,260	$462,650	$2,409,660	$4,176,740	p
Contingencies	$101,190	$713,140	$5,629,790	$10,560,150	c
Total	$775,820	$5,467,390	$43,161,740	$80,961,170	

Table C23.2 - WATERCO$T Model Base Construction Cost Analysis for Reverse Osmosis

| Cost Component | Plant Capacity (mgd) | | | | Average Percent |
	1.0	10	100	200	
Manufactured Equipment	61.12%	63.26%	67.59%	69.71%	65.42%
Labor	9.08%	6.34%	5.36%	3.51%	6.07%
Electrical	8.47%	8.89%	8.42%	8.58%	8.59%
Housing	8.28%	8.46%	5.58%	5.16%	6.87%
Contingencies	13.04%	13.04%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C24.1 - Base Costs Obtained from the WATERCO$T Model for Sodium Hydroxide Feed Systems

Cost Component	Feed Capacity (lb/day)				Capital Cost Category
	10	100	1,000	10000	
Manufactured Equipment	$6,440	$7,010	$5,720	$19,450	p
Labor	$640	$640	$790	$4,120	c
Pipes and Valves	$850	$850	$850	$850	p
Electrical	$3,190	$3,190	$3,190	$3,460	p
Housing	$1,010	$2,100	$8,400	$48,380	p
Contingencies	$1,820	$2,070	$2,840	$11,440	c
Total	$13,950	$15,860	$21,790	$87,700	

Table C24.2 - WATERCO$T Model Base Construction Cost Analysis for Sodium Hydroxide Feed Systems

Cost Component	Feed Capacity (lb/day)				Average Percent
	10	100	1,000	10,000	
Manufactured Equipment	46.16%	44.20%	26.25%	22.18%	34.70%
Labor	4.59%	4.04%	3.63%	4.70%	4.24%
Pipes and Valves	6.09%	5.36%	3.90%	0.97%	4.08%
Electrical	22.87%	20.11%	14.64%	3.95%	15.39%
Housing	7.24%	13.24%	38.55%	55.17%	28.55%
Contingencies	13.05%	13.05%	13.03%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C25.1 - Base Costs Obtained from the WATERCO$T Model for Sulfuric Acid Feed Systems

Cost Component	Feed Capacity (gpd)				Capital Cost Category
	10	100	1000	5000	
Manufactured Equipment	$1,560	$3,440	$12,400	$41,000	p
Labor	$640	$820	$2,840	$11,840	c
Pipes and Valves	$1,090	$1,090	$2,150	$2,150	p
Electrical	$1,670	$2,920	$2,920	$2,920	p
Housing	$2,520	$1,560	$1,560	$1,560	p
Contingencies	$1,120	$1,470	$3,280	$8,920	c
Total	$8,600	$11,300	$25,150	$68,390	

Table C25.2 - WATERCO$T Model Base Construction Cost Analysis for Sulfuric Acid Feed Systems

Cost Component	Feed Capacity (gpd)				Average Percent
	10	100	1000	5000	
Manufactured Equipment	18.14%	30.44%	49.30%	59.95%	39.46%
Labor	7.44%	7.26%	11.29%	17.31%	10.83%
Pipes and Valves	12.67%	9.65%	8.55%	3.14%	8.50%
Electrical	19.42%	25.84%	11.61%	4.27%	15.28%
Housing	29.30%	13.81%	6.20%	2.28%	12.90%
Contingencies	13.02%	13.01%	13.04%	13.04%	13.03%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

Table C26.1 - Base Costs Obtained from the WATERCO$T Model for Tube Settling Modules

Cost Component	Tube Module Area (ft2)					Capital Cost Category
	280	2,800	14,000	28,000	56,000	
Manufactured Equipment	$4,200	$31,000	$147,000	$282,000	$504,000	p
Steel	$2,000	$19,500	$95,000	$155,000	$300,000	p
Labor	$2,500	$11,200	$49,000	$95,000	$224,000	c
Contingencies	$1,300	$9,300	$43,700	$79,800	$154,200	c
Total	$10,000	$71,000	$334,700	$611,800	$1,182,200	

Table C26.2 - WATERCO$T Model Base Construction Cost Analysis for Tube Settling Modules

Cost Component	Tube Module Area (ft2)					Average Percent
	280	2,800	14,000	28,000	56,000	
Manufactured Equipment	42.00%	43.66%	43.92%	46.09%	42.63%	43.66%
Steel	20.00%	27.46%	28.38%	25.34%	25.38%	25.31%
Labor	25.00%	15.77%	14.64%	15.53%	18.95%	17.98%
Contingencies	13.00%	13.10%	13.06%	13.04%	13.04%	13.05%
Total	100.00%	100.00%	100.00%	100.00%	100.00%	100.00%

Table C27.1 - Base Costs Obtained from the WATERCO$T Model for Wash Water Surge Basins

Cost Component	Capacity (gal)				Capital Cost Category
	10,000	50,000	100,000	500,000	
Excavation & Sitework	$200	$520	$1,250	$4,400	c
Concrete	$11,560	$39,310	$71,480	$143,680	p
Steel	$7,990	$25,170	$44,680	$70,770	p
Labor	$18,270	$58,500	$107,590	$182,150	c
Pipes and Valves	$5,500	$7,500	$11,000	$16,000	p
Electrical	$1,300	$1,300	$6,000	$6,000	p
Contingencies	$6,720	$19,850	$36,300	$63,450	c
Total	$51,540	$152,150	$278,300	$486,450	

Table C27.2 - WATERCO$T Model Base Construction Cost Analysis for Wash Water Surge Basins

Cost Component	Capacity (gal)				Average Percent
	10,000	50,000	100,000	500,000	
Excavation & Sitework	0.39%	0.34%	0.45%	0.90%	0.52%
Concrete	22.43%	25.84%	25.68%	29.54%	25.87%
Steel	15.50%	16.54%	16.05%	14.55%	15.66%
Labor	35.45%	38.45%	38.66%	37.44%	37.50%
Pipes and Valves	10.67%	4.93%	3.95%	3.29%	5.71%
Electrical	2.52%	0.85%	2.16%	1.23%	1.69%
Contingencies	13.04%	13.05%	13.04%	13.04%	13.04%
Total	100.00%	100.00%	100.00%	100.00%	100.00%

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Appendix D Basis for Revised Activated Alumina Costs

APPENDIX D: BASIS FOR REVISED ACTIVATED ALUMINA COSTS

D.1   OVERVIEW

Four sets of costs were developed for disposable activated alumina (no regeneration) - two for unadjusted pH and two where the pH has been adjusted to the optimal pH of 6. The costs for pH adjustment to optimal pH 6.0 are included as a separate module in these cost estimates since it is only used for two of the options. The costs in the cost models routinely used to estimate unit costs are all based on regeneration of the media, operation at optimal pH, and use of single columns. Most of the costs are based on fluoride removal rather than arsenic removal. It was determined that the existing cost models could not be used to estimate costs for disposable activated alumina. A similar summary to this was prepared for the proposed rule for the disposable activated alumina costs. It was included in Appendix G of the November 1999 Technology and Cost Document (1). The main change between that design and the design used for the final rule is that smaller columns operated in series will be used instead of a single column. This will provide greater utilization of the media before disposal and is more consistent with the designs used by commenters in evaluating disposable activated alumina.

D.2   DESIGN ASSUMPTIONS AND BASIS

D.2.1   Capital Costs - Natural pH

1.   The pH will not be adjusted to operate the process at the optimal pH between 5.5 and 6.0. The activated alumina process will be operated at the natural pH of the system to simplify the process and avoid potential problems with lowering the pH. Two pH ranges will be evaluated 7.0 #pH < pH 8.0 and 8.0 # pH #8.3.

Basis. Comments from AWWA Arsenic Technical Workgroup Meeting on Treatment Technology. EPA expressed a concern about small systems adjusting pH downward and the AWWA Technical Workgroup agreed that there were significant risks if small systems were adjusting pH downward (the water could become very acidic if the treatment chemicals are overdosed). The AWWA Technical Workgroup believed that lowering the pH would require

much more oversight than most small systems would have available. In addition, the elimination of a chemical feed system will simplify the total treatment system oversight requirements. Since the run length is sensitive to pH, two ranges were developed for the natural pH option.

2. The pH will not need to be adjusted after the activated alumina process. The activated alumina process should not adversely affect the finished water pH. Those systems with high natural water pH will not need a new pH adjustment process. Those systems with lower natural water pH values will continue to rely on the existing corrosion control process.

   Basis. Post-treatment pH adjustment is not included in the White Rock Water Company case study*(2)*. The raw water pH of the two blended sources is approximately 7.6. The treated water is blended with untreated water in this case study. Post-treatment pH adjustment is also not included in the two full-scale case studies from New Hampshire in the ORD project (*3*). The average raw water pH values for these two systems are pH 8.1 and pH 8.3 (*3*) Samples were taken at the inlet and outlet to the activated alumina plants. Average pH and alkalinity values at the inlet and outlet to the two plants were approximately equal. Since it does not appear that activated alumina is adversely affecting the finished water pH, additional post-treatment corrosion control is not necessary.

3. Empty Bed Contact Time (EBCT) = 5 minutes per column

   Basis. Series operation with one column serving as a roughing column and the second column serving as the polishing column provides greater utilization of the activated alumina than a single column. Since the media is being operated on a disposable basis rather than being regenerated, greater utilization of the media is desired to reduce operation and maintenance (O&M) costs and minimize waste disposal. Since the volume of water treated is dependent upon the size of the bed, an EBCT of 5 minutes was selected based on a comparison with other designs. An EBCT of approximately 5 minutes was used for series operation with regeneration in the AwwaRF report (*4*). The Phoenix case study was the only one to use disposable media and the EBCT was 2.5 minutes (*3*). The EBCT in the White Rock Water

Company case study is 18 minutes (*2*). One full-scale site is based on a 7.5-minute EBCT and the other is based on a 4.3 minute-EBCT (*3*).

4. The estimated cost of the activated alumina media is \$0.82/lb for systems with a design flow greater than 0.1 mgd. The estimated cost of activated alumina media is \$1.30/lb for systems with a design flow less than or equal to 0.1 mgd.

    Basis. The five case studies with regeneration in the AwwaRF report use \$0.50/lb as the cost of activated alumina media (*3*). The Phoenix case study uses \$0.75/lb(*4*). Several manufacturers and distributors were contacted for quotes on activated alumina. A range of costs was obtained based on the size of the shipment. The lowest cost was based on a truckload of 20 tons. The truckload costs averaged \$0.68/lb. A higher cost of \$0.82/lb was assumed to account for higher costs that may be associated with supersack purchases at the lower end of the flow range. The highest cost of \$1.30/lb was for a 55 gallon barrel of 28x48 media. One of the full-scale plants uses Alcoa CPN 28x48 (fine) alumina and it cost \$1.98/lb when it was replaced in May 1999 (*3*). The other full-scale plant uses Alcoa DD-2 14x28 (coarse) alumina and it cost \$1.01/lb when it was replaced in December 1998 (*3*). The barrel cost was assumed for all very small systems with design flows at or below 0.1 mgd.

5. The density of the activated alumina media is assumed to be 47 lb/ft^3.

    Basis. The bulk density of Alcoa CPN activated alumina is 47 $/b/ft^3$ (*3*). The bulk density range for Alcoa DD-2 activated alumina is 39 to 52 lb/ft^3 (*3*). The average for the DD-2 media would be 45.5 lb/ft^3. The higher density was conservatively used to develop the costs.

6. The bed depth assumption ranged 3 feet to 6 feet depending upon the design flow. The maximum diameter per column was 12 feet. When the calculated media volume exceeded the maximum, the treatment media was divided into several beds to maintain a realistic bed diameter. Thus, larger designs include multiple activated alumina trains, each treating a portion of the total flow.

Basis. An old activated alumina design manual for fluoride removal recommends that the bed diameter be equal to or greater than the bed depth to prevent "wall effects" (*5*). This source also states that good practice dictates that the bed depth be a minimum of three feet and a maximum of six feet. The Water Model assumed a bed depth of 5 feet for all conceptual designs (*6*). Many of the conceptual designs in the cost estimating manual had bed depths that exceeded the diameter, so the recommendation in the design manual was not strictly applied to all designs. For design flows less than 50,000 gallons per day, the bed depth is greater than the bed diameter. The data from the three case studies where activated alumina is being operated without regeneration, the bed depth is greater than the bed diameter (*2, 3*).

7. The vessel cost has been sized based on 50% bed expansion during backwash even though backwashing may not be necessary on a routine basis for smaller systems. The vessel volume was calculated as (1.5)(media volume).

   Basis. The old design manual and the Water Model used 50% bed expansion to size the vessel (*5, 6*). Both of these designs are based on regeneration of the media rather than disposable operation. The bed expansion for the one case study where it could be calculated was 30% (*3*). The design approach used as the basis for the case studies in the AwwaRF report used 80% expansion (*7*). However, this design was also based on regeneration of the media rather than disposable media operation and was based on designs greater than 1 mgd. Since the expansion is only necessary for backwashing, 50% expansion was assumed for all sizes. Backwashing is not performed on a routine basis at any of the three small systems operating activated alumina without regeneration.

8. The vessel cost is based on the following equation:

   Cost = 63.288 * (Vessel Volume in gallons) ^ 0.679

   Basis. Quotes were obtained from two manufacturers for glass-lined carbon steel vessels with working pressure rating between 75 and 150 psi and meeting American Society of Mechanical Engineers (ASME) Code Section VIII requirements. Sizes ranging from 140 gallons to 6764 gallons were used to develop the equation. Vessel costs for sixteen different sizes were

regressed against their respective volumes to derive the equation. Costs from the AwwaRF report cannot be used for comparison because of an apparent error in the vessel costs. The vendor quote used for vessel costs is actually a vendor quote for a complete activated alumina system. The vendor quote presents budget prices for three design flows and different size vessels are used for each design flow. The vessel sizes are listed with the budget price and this is apparently the source of the mistake. The three full-scale systems using activated alumina on a disposable basis use fiberglass tanks instead of carbon steel tanks because the pressure drop is very small. Fiberglass tanks would be less expensive than the carbon steel vessels, so the vessel costs are conservative for very small systems.

9. The capital costs include a redundant column to allow the system to operate while the media is being replaced in the old roughing column.

    Basis. A redundant vessel has been included for each set of roughing and polishing columns. While the activated alumina media is being replaced in this column, the other two columns can still function as a roughing and polishing column. Activated alumina media is not included in the redundant vessel(s) as a capital cost to make it easier to estimate O&M costs. The Phoenix case study uses a redundant roughing and polishing column (*4*). This design did not seem to be the most efficient because one column would contain partially used activated alumina, but would not be in operation while the media in the other column was being replaced. The initial Albuquerque design did not contain a redundant column, but instead relied upon the old polishing column to treat all the water while the media was replaced in the old roughing column (*7*). The two full-scale sites in New Hampshire split the flow into two streams and had roughing and polishing columns treating half the water (3). Redundant columns were not included; however, one set of columns could treat all the water if there was a problem with the other set.

10. The remaining components of the process equipment include concrete, pipe and valves and electrical and instrumentation. A cost factor of 14.43% of the manufactured equipment and activated alumina was used for these other construction cost components.

Basis. The Water Model provided construction cost estimates for a variety of conceptual designs (*6*). Construction cost components included process costs such as manufactured equipment and media, concrete, pipes and valves, electrical and instrumentation as well as construction costs such as excavation and sitework. The conceptual designs are based on an EBCT of 7.5 minutes. Since this design is using a total EBCT of 10 minutes (two 5-min EBCTs in series), the activated alumina costs were adjusted to estimate the contribution of each component to the construction cost. The weighted percentage of each component was calculated for both EBCT values. The weighted percentages for concrete, pipe and valves and electrical and instrumentation were then expressed as a percentage of the manufactured equipment and activated alumina costs. This was done for the data with an EBCT of 10 minutes. Concrete was 3.17%, pipe and valves were 15.18%, and electrical and instrumentation were 11.12% of the equipment and media costs. However, the designs in the Water Model include pipes, valves, and instrumentation for the sulfuric acid and sodium hydroxide feed systems in the regeneration process. Those are not needed in this design. Therefore, the pipe and valves percentage has been reduced by a factor of 2. The electrical and instrumentation were reduced by a factor of 3 because pH sensors and alarms are not needed and the electrical needs are drastically reduced. The main electrical components of the activated alumina process in the Water Model are day tank mixers and pumps used for regeneration and the electrical immersion heater for the sodium hydroxide tank. None of these components are needed in this design. The adjusted cost factor is 14.43%. The other process costs were calculated by multiplying the adjusted factor by the sum of the manufactured equipment and activated alumina costs. The sum of the other process costs and the manufactured equipment and activated alumina costs is the total process costs. For the designs with pH adjustment, a separate module has been developed including pipes and valves for those chemical feed systems.

11. The capital costs have been estimated from the total process costs using a factor of 2.5 for small systems and 3.33 for large systems.

    Basis. The capital costs have been estimated from the process equipment costs using the ratio in the Guide for Implementing Phase I Water Treatment Cost Upgrades (*8*). The process costs were assumed to be 40% of the capital costs for small systems. The breakdown of capital

costs for small systems is as follows: 40% process costs, 40% construction costs, and 20% engineering costs. For large systems, the breakdown of capital cost is: 30% process costs, 40% construction, and 30% engineering.

12. The capital costs estimated in the previous step do not include the cost of a building to house the process equipment and other add-on costs related to the site. Housing costs were assumed for all sites. Fence and road were only assumed for those systems that have no treatment.

Basis. Data from the 1995 Community Water Systems Survey were used to identify the percentage of systems without treatment (*9*). The data from ground water systems is as follows:

Table D-1

Percentage of Ground Water Systems with No Treatment: Systems Serving # 3,300 People

Population Category			
25 - 100	101 - 500	501 - 1,000	1,001 - 3,300
43%	19%	16%	18%

Table D-2

Percentage of Ground Water Systems with No Treatment: Systems Serving > 3,300 People

Population Category			
3,301 - 10,000	10,001 - 50,000	50,001 - 100,000	100,001 - 1 MIL
13%	1%	11%	0%

Larger ground water systems rely on multiple entry points to supply ground water rather than one well. Thus, treatment may be present in the system, but only at one of the wells. Since the analysis is based on no treatment throughout the system, using the percentages for larger

systems could underestimate the need for roads and fences at some entry points. Therefore, the 18% without treatment from the 1,001 - 3,300 category was applied to the four larger system categories. The following table summarizes the percentage used for each of the size categories.

Table D-3

Assumed Percentage of Ground Water Systems with No Treatment

Population Category	Percentage of GWS w/o any Treatment
25 - 100	43%
101 - 500	19%
501 - 1,000	16%
1,001 - 3,300	18%
3,301 - 10,000	18%
10,001 - 50,000	18%
50,001 - 100,000	18%
100,001 - 1 MIL	18%

Thus, the cost of a road and fence was calculated and 43% of that total was included in the unit cost for activated alumina for systems serving between 25 and 100 people.

13. The equation for process area, which is the key variable in estimating the costs for buildings, roads, and fences has been based on the design flow of the system. The equation used for process area is:

Process Area = 726.6 * Design Flow + 59

Basis. The basis for the cost equations for buildings, fence and road are described in Section 4.0 of the Very Small Systems Document (*10*). The Very Small Systems Document provided costs as a percentage of the capital costs for design flows from 0.024 to 0.65 mgd. The process area was calculated from the building cost estimates by dividing by 40 (building costs

were based on $40/sq ft). The process areas for the design flows in the Very Small Systems Document were regressed and found to best fit a linear regression:

$$\text{Process Area (sq ft)} = 614.6 * \text{Design Flow} + 68$$

The process area for the conceptual designs in the Water Model (*6*) were also regressed and fitted to a linear regression. The conceptual designs covered a design flow range from 0.045 to 1.082 mgd. The linear regression from this source is:

$$\text{Process Area (sq ft)} = 726.6 * \text{Design Flow} + 59$$

The process area estimates using the regression of the older model data were selected to make cost estimates for buildings, road and fence. Since there would be a minimum building size, extrapolating a linear regression below the lowest data point does not create a significant problem with the intercept. These process area estimates were larger than the process area estimates from the Very Small Systems Document for design flows larger than 0.09 mgd. At the lower design flows, the process area estimates were slightly lower using the Water Model. The Water Model covered a broader range of design flows, so less extrapolation is required at higher design flows than would be the case if the Very Small Systems Document equation were used. Another factor that also adds to the conservative nature of the process area estimates is that the data are based on a design that includes regeneration. The space devoted to chemical feeders and storage tanks for regeneration of the media is unnecessary in this design. The process area estimates were compared with the minimum area needed for the vessels using the vessel diameter. The process area estimates are greater than three times the minimum area estimates for the vessels, including the redundant column.

14. The building cost to house the activated alumina system is $45.38/sq ft for design flows at or below 0.65 mgd. The following equation is used for design flows between 0.65 and 1 mgd:

$$\text{Building cost} = 54.204 * \text{Process Area} + 14887$$

For design flows greater than 1 mgd, the building cost varied based on flow from $100 to $90 per square foot.

Basis. The data from both the Very Small Systems Document and the Water Model were examined to develop housing costs for the activated alumina system. The housing rate listed in the Very Small Systems Document was $40/sq ft (*10*). This rate includes the following: basic storage building, foundation, electrical wiring, HVAC, and plumbing. The rate is for a pre-fabricated building installed on a concrete slab. The rate is based on data from pre-fabricated building manufacturers and the 1993 Means Building Construction Cost Data for storage buildings multiplied by a conversion factor for small-sized buildings. The housing costs in the Water Model vary based on size with rates ranging from $87/sq ft for the smallest conceptual design and $50/sq ft for a facility treating approximately 1 mgd (*6*). The only documentation on the costs in the Water Model states that in lieu of segregating building costs into several components, the housing category represents all material and labor costs associated with the building, including heating, ventilating, lighting, and normal convenience outlets. However, the fabrication and building type are unknown and, most critically, the sources used to derive the building costs are unknown. The costs in the Water Model were developed in 1983. The ENR Building Cost Index was used to update both sets of costs to September 1998. The housing rate from the Very Small Systems Document was selected based on better documentation and the use of more recent data for design flows up to 0.65 mgd. For design flows between 0.65 and 1 mgd, a linear regression of housing cost against process area was used to derive the equation listed above. For design flows greater than1 mgd, the building cost rate assumptions were the same as those used in the for Albuquerque Report (*7*) and for the case studies in the AwwaRF Report (*4*).

D.2.1 Capital Costs - pH Adjustment

15. For two options, pH adjustment to the optimal pH 6 is utilized. The capital cost equation for the pH adjustment module is:

Capital cost = 40517 * Design Flow + 31889

Basis. Capital cost components include piping, acid and caustic metering pumps, and storage tanks for the acid and caustic. For the metering pumps, pH controllers, pH sensors, and level sensors are included in the costs. Cost data for each of the components were obtained from manufacturers for design flows ranging from 0.1 to 3 mgd. Capital costs were estimated by multiplying the total process cost by 2.5 or 3.33 depending upon size(*8*). The Phoenix case study in the AwwaRF report provides the only realistic comparison for disposable activated alumina at optimal pH (*4*). The other case studies are based regeneration and would use acid and caustic for those purposes. A design flow of 0.2 mgd was evaluated in both the Phoenix case study and the process cost estimate used to derive the above equation. The process cost estimate of $16,200 is very similar to the process cost from the Phoenix case study of $16,750.

D.2.1   Operation and Maintenance Costs

There are three major components to the operation and maintenance (O&M) costs - replacement media, energy and labor.

16. The critical variable in calculating the replacement media costs is the run length - the number of bed volumes until the effluent concentration exceeds a specific concentration. The run length for the natural pH range of $7 \# pH < 8$ is 10,000 bed volumes (BV). The run length for the natural pH range of $8 \# pH \# 8.3$ is 5,200 BV. These run lengths are used to estimate the replacement of the activated alumina in half of the active columns.

   Basis. The full-scale systems in New Hampshire serve as the basis for the run lengths for the natural pH options (*3*). Plant C data were used for the range of $7 \# pH < 8$ and data from Plant D were used for the range of $8 \# pH \# 8.3$.

The source water at Plant C has the following critical characteristics:

Parameter	Average Source Water Results
Arsenic (total)	59 µg/L
pH	8.1
Alkalinity	84 mg/L CaCO$_3$
Fluoride	1.5 mg/L
Sulfate	26 mg/L

The roughing tank treated 10,050 BV of water before breakthrough (effluent = influent). The arsenic concentration in the effluent from the second tank was approximately 20 µg/L. Based on the sample results from September 2 and September 16, 1998, breakthrough above 3 µg/L (95% removal based on average inlet of 62.1 µg/L) in the effluent occurred somewhere within that time period. This converts to a run length of approximately 8500 BV. The roughing tanks were replaced with virgin activated alumina after treating 11,071 BV of water. The former polishing tanks then became the roughing tanks. The data indicate that the outlet arsenic concentration was still below 4 µg/L as the total run approached 16,000 BV. This supports a run length of 8500 BV per replacement of half of the media. Arsenic concentration was the only key influent parameter that varied to any extent during the testing.

Run length would be affected by pH and by the influent arsenic concentration. Since influent arsenic concentrations for most systems affected by the rule should be much lower than 59 µg/L and since arsenic removal is pH dependent, a run length of 10,000 BV was selected for the range of 7 ≤ pH < 8. This should be very conservative for systems with lower arsenic concentrations and pH values at the lower end of the range. The effluent pH and effluent concentrations of alkalinity, fluoride and sulfate indicated almost no change from their value in the influent. Thus, the existing corrosion control should be sufficient.

The source water at Plant D has the following critical characteristics:

Parameter	Average Source Water Results
Arsenic (total)	59.0 Ｆg/L
pH	8.3
Alkalinity	55.5 mg/L $CaCO_3$
Fluoride	1.13 mg/L
Sulfate	15 mg/L

The roughing tank treated 5,200 BV of water before breakthrough (effluent = influent). The arsenic concentration in the effluent from the second tank was less than 2 Ｆg/L as it was throughout the run. Arsenic concentration was the only key parameter that varied to any extent during the testing. The run length would be affected by pH and by the influent arsenic concentration. The pH was consistent and represents the upper bound on the range of applicability for this cost equation. A run length of 5,200 BV was used as the run length for the range of 8 # pH # 8.3. Since influent arsenic concentrations for most systems affected by the rule should be much lower than 59 Ｆg/L, this should be a very conservative run length for systems with lower arsenic concentrations and/or pH values at the lower end of the range. The effluent pH and effluent concentrations of alkalinity, fluoride and sulfate indicated almost no change from their value in the influent. Thus, the existing corrosion control should be sufficient.

The previous approach described in Appendix G of the November 1999 Technology and Cost Document (1) relied upon pilot-scale data from Severn-Trent (11). The study examined arsenic removal over a pH range from pH 6 to pH 7.5 in a single column. The three run lengths derived from that data ranged from 16,500 BV and pH 7 to 3,000 BV at pH 8 for 10% breakthrough in a single column. Actual run lengths to 50% breakthrough in that study for natural pH 7.5 ranged from 9,400 to 13,000 BV based on mesh size. The data from the two full-scale facilities (3) indicated that the roughing column could be operated almost complete breakthrough (effluent = influent) before concentrations in the effluent of the second column exceeded 5% of the influent concentration. This data also indicates that a natural pH run

length of 10,000 BV for 7 # pH < 8 may be conservative, especially for systems with lower pH.

One final factor that would increase the run length is intermittent operation at small systems. When a fixed bed process is operated intermittently, the sorbed ions can migrate deeper into the pore structure of the media, thereby exposing more external surface area to the ions in solution. This factor also supports the use of the uncorrected run lengths in developing the O&M costs for this process.

17. The two previous runs were based on operating at the natural pH without pH adjustment. Adjusting pH in larger systems is less likely to be an issue, so run lengths were also developed for operation at pH 6. Two run lengths were selected for the optimal pH of 6 - 15,400 and 23,100 BV.

Basis. The two run lengths for pH adjustment to pH 6 are based on pilot-scale data performed at Albuquerque (12). The Arsenic Treatment Evaluation Report (7) used the pilot-scale data to derive a run length estimate for series operation of activated alumina columns with regeneration. In the examination of this report, it was determined that there was an error in the derivation of the run length estimate (13). A revised estimate of run length was calculated using the same data. The procedure described in that following modified excerpt describes the procedure used to calculate the run length at pH 6 for disposable activated alumina.

A. The equation for arsenic effluent in the Arsenic Treatment Evaluation Report (7) was used to determine the adsorptive capacity of the lead column until arsenic breakthrough at 2 Fg/L. The equation was based on the data from Figure 14 of the Phase 3 Report (12). It was necessary to make one assumption to estimate the arsenic adsorbed onto the column. It was assumed that the influent concentration of 22.8 Fg/L was constant. From the Phase 3 Report, one column can treat 4300 BV before arsenic breaks through at 2 Fg/L. The amount of arsenic that passed through the column can be determined by estimating the area under the curve defined by the

equation $y = (2*10^{-8})x^2 + 0.0002x + 0.8394$. The integral of this equation was used to estimate the area under the curve. The integral equation is:

$z = (2*10^{-8})/3*x^3 + 0.0001x^2 + 0.8394x$.

For x = 4300 BV:

$As_{pass} = (2*10^{-8})/3*(4300)^3 + 0.0001*(4300)^2 + 0.8394*4300$
$As_{pass} = 5988 \ \mu g*BV/L$.

The arsenic adsorbed onto the activated alumina was calculated by subtracting the arsenic that passed through the column from the total arsenic in the influent.

$As_{ads} = As_{tot} - As_{pass} = (4300 \ BV)(22.8 \ \mu g/L) - 5988 \ \mu g*BV/L$
$As_{ads} = 98,040 - 5988 \ \mu g*BV/L = 92,052 \ \mu g*BV/L$

The arsenic adsorbed onto the activated alumina is often reported in milligrams arsenic per gram of activated alumina. The Phase 3 Report refers to this as As loading in Table 4. In the Phase 3 Report, a bed volume was 0.375 L. To calculate the arsenic loading, the density of Alcoa CPN is needed. The packed bulk density for Alcoa CPN is 47 lb/ft^3 or 750 kg/m^3. Arsenic loading is calculated as follows:

As loading = $\dfrac{[(92,052 \ \mu g*BV/L)(0.375 \ L/BV)(mg/1000 \ \mu g)]}{(0.375 \ L)(dm^3/L)(m^3/1000 \ dm^3)(750 \ kg/m^3)(1000 \ g/kg)}$

As loading = 0.12 mg As/g AA for breakthrough at 2 µg/L

Arsenic loading will vary based on the breakthrough concentration. For breakthrough at 10 µg/L, the run length is 16,500 BV. Using the integral equation yields the following:

$As_{pass} = 71,023 \ \mu g*BV/L$
$As_{ads} = (22.8 \ \mu g/L)(16,500 \ BV) - 71,023 \ \mu g*BV/L = 305,177 \ \mu g*BV/L$

The arsenic loaded onto the column is:

As loading = $\frac{[(305,177)(0.375)(1/1000)]}{(0.375)(1/1000)(750)(1000)}$

As loading = 0.41 mg As/g AA   {*Table 2 of Phase 3 Report lists 0.43 mg As/g AA*}

B. Using the bed volume size of 16,032 liters, yields the following adsorptive capacities:

As_{ads} = (92,052 Fg*BV/L)(16,032 L/BV)(1 gram/10^6 Fg) = 1476 grams

As_{tot} = (98,040)(16,032)(1/10^6) = 1572 grams

It appears that the adsorptive capacity of 1573 grams in the Arsenic Treatment Evaluation Report is based on total arsenic instead of the adsorbed arsenic.

C. The maximum run length will be based on an adsorptive capacity of 1476 grams in the second column. For this step of the analysis, it will be assumed that all of the arsenic passing through the first column will be adsorbed by the second column. This is a very conservative estimate because the second column will also have leakage through it and concentrations will rise until breakthrough is reached at 2 Fg/L. This is balanced by the fact that when the polishing column is moved into the roughing column position, some of its capacity will have already been used when it was the polishing column. Both of these factors will be addressed later. The integral equation in step 2 is again utilized to estimate the amount of arsenic passing through the first column. When this value is equal to 1476 grams, then the maximum run length can be determined.

For a run length of 18,500 BV:

As_{pass} = (2*10^{-8})/3*(18500)^3 + 0.0001*(18500)^2 + 0.8394*18500
As_{pass} = 1474 grams

The maximum run length is 18,500 bed volumes. The amount of arsenic adsorbed onto the first column is:

$$As_{ads} = (22.8 \ \digamma g/L)(18,500 \ BV)(16,032 \ L/BV)(g/10^6 \ \digamma g) - 1474 \ grams$$
$$As_{ads} = 5288 \ grams$$

Thus, the capacity of the roughing column before it needs to be regenerated is 5288 grams. This assumes that there is no arsenic already adsorbed onto the column. Under steady-state operation, the polishing column will become the roughing column and the redundant column with virgin media column becomes the polishing column. The media in the old roughing column is replaced with virgin media and becomes the new redundant column. The adsorbed arsenic on the former polishing column will decrease run length whereas arsenic leakage through the second column will increase run length.

D. The adsorptive capacity of the roughing column was 5288 grams based on the maximum run length of 18,500 bed volumes. When the "old" polishing column is shifted and becomes the "new" roughing column, it already has a maximum of 1474 grams adsorbed onto it. Therefore, the remaining capacity is considerably lower than 5288 grams. The absorptive capacity of the "new" roughing column was first estimated by subtracting the arsenic that passed through the first column. This estimate was called the minimum capacity of column 2, since the polishing column does not adsorb all of the arsenic that passed through the first column.

Min Adsorptive Capacity - Column 2 = 5288 - 1474 grams = 3814 grams.

E. As noted in the previous step, the minimum adsorptive capacity was estimated by assuming that all of the arsenic that passed through the first column was adsorbed onto the second column. This would not be the case. The second column would have leakage effects as well as increasing arsenic concentrations until the effluent concentration reached 2 $\digamma$g/L. Because the arsenic passing through the first column and reaching the second column is not constant, it is not possible to estimate a breakthrough curve. Therefore, a different approach was taken to better estimate the adsorptive capacity. Arsenic leakage from the single column study summarized in the Phase 3 Report (*12*) was examined to estimate leakage for the second

column. The Phase 3 Report noted that arsenic leakage was always greater from the once-regenerated alumina than the virgin alumina. For the 5-minute EBCT runs, an average arsenic leakage (up to 3200 BV) of 0.3 and 1.4 µg/L was observed for the virgin and once-regenerated Alcoa CPN alumina runs. For the Alcan alumina at an EBCT of 5 minutes, average arsenic leakage (up to 3200 BV) was 0.2 and 1.3 µg/L for the virgin and once-regenerated alumina runs. After 3200 BV, the normal breakthrough curve started, so the effluent arsenic was not solely due to leakage. Since the run length of 4300 BV and breakthrough equation were both based on the once-regenerated Alcoa CPN alumina, arsenic leakage was also estimated for this alumina. The average arsenic leakage was 1.4 µg/L   It was assumed that the second column would exhibit the same leakage characteristics as the first column. Thus, arsenic leakage for the second column was assumed constant at 1.4 µg/L. It was assumed that 1.4 µg/L would pass through both columns starting at the beginning of the run. The second column will probably remove some of the arsenic that is passing through the first column, so this approach likely overestimates the amount of arsenic leakage for the second column. However, since there was no way to estimate a breakthrough curve for the second column, it was not possible to estimate arsenic in the effluent due to reduced adsorptive capacity. The arsenic leakage estimate was used to account for all arsenic that would pass through the second column - from both leakage and reduced adsorptive capacity. Thus, it likely underestimates the amount of arsenic that passed through the second column into the finished water.   The arsenic leakage was estimated using the following equation:

$$As_{leakage} = (1.4 \text{ µg/L})(RL \text{ BV})(16{,}032 \text{ L/BV}) \quad \text{Where } RL = \text{run length in BV}$$

For a run length of 18,500 BV:

$$As_{leakage} = 415 \text{ grams}$$

F.   The available arsenic capacity of the second column is estimated by adding the arsenic leakage to the minimum arsenic capacity of the second column.

Available Arsenic Capacity of Column 2 = 3814 grams + 415 grams = 4229 grams

Since the available arsenic capacity of the "old" polishing column is lower than the maximum capacity, the steady-state run length must be lower than 18,500 bed volumes.

G. The run length was determined by an iterative approach. The integral equation was used to estimate the arsenic adsorbed onto the first column for the lower run lengths. The available arsenic capacity of the second column was also estimated using the arsenic leakage determined for the new run length. When the arsenic adsorbed on column 1 was equivalent to the available arsenic capacity of column 2, the steady-state run length was determined. The following will illustrate this process:

At 15,400 bed volumes:

Column 1 - arsenic adsorbed = 4651 grams
Column 1 - arsenic passed through = 978 grams
Column 2 - arsenic leakage = 346 grams
Column 2 - arsenic adsorbed = 632 grams
Column 2 - available arsenic capacity = 5288 - 632 = 4656 grams

The available arsenic capacity of column 2 is approximately equivalent to the amount of arsenic that can be adsorbed onto it when it becomes the roughing column (column 1) based on the maximum run length. This is the steady-state run length to meet a standard of 2 Fg/L at the effluent to the second column. After 15,400 bed volumes, the media in the roughing column is taken off-line and replaced. While this is occurring, the polishing column becomes the new roughing column and the redundant column becomes the new polishing column.

H. The Phase 3 Report (*12*) recommended that the run length be increased by up to 50% to account for improved performance when hydrochloric acid is used instead of sulfuric acid to adjust the pH. Based on the Phase 3 Report, low and high estimates of run length were made based on the use of hydrochloric acid for pH adjustment. For the low option, a 20% increase was assumed and a 50% increase was assumed for the high option. If hydrochloric acid is used for pH adjustment, the run length would range from 18,480 bed volumes to 23,100 bed volumes.

The lower bound run length at pH 6 was based on the calculated run length for Albuquerque water when the pH is adjusted using sulfuric acid (15,400). The upper bound run length at pH 6 was based on the higher end of the range when hydrochloric acid is used instead of sulfuric acid. There are several factors that could influence run length.

The first factor is that the run length for the single column is based on the data from once-regenerated activated alumina. It was not possible to determine the effect of regeneration because the influent concentration varied significantly between the two runs. Some studies have shown some degradation after regeneration. The use of virgin media produced a longer run length even though the influent concentration was much higher (38.8 μg/L). Effluent concentrations from virgin media did not exceed 2 μg/L (. 95% removal) until 7000 BV had been treated using either the Alcoa CPN and the Alcan media. This is nearly double the single unit run length of 4300 BV using the once regenerated media that serves as the basis for the two pH adjusted series operation options. Using the run lengths based on regenerated media is likely to be very conservative for costing disposable activated alumina options.

The second factor is that the leakage through each of the columns will be reduced. The data described above for virgin media was less than 0.4 μg/L. This may also extend the run length for series operation based on virgin media because less arsenic is loaded onto the second column due to leakage.

The third factor is that the data from the two full-scale facilities (*3*) indicated that the roughing column could be operated almost complete breakthrough (effluent = influent) before concentrations in the effluent of the second column exceeded 5% of the influent concentration. Complete breakthrough refers to the effluent of the first column (C_1) equaling the influent to the column (C_0). The finished water would be the effluent of the second column (C_2). The ratio of C_1/C_0 was derived for the Alcoa CPN data in Figure 14 of the Phase 3 Report (12).

C_1 . 15 μg/L at the end of the data collection (. 17,000 BV)

$C_0 = 38.8$ μg/L

$C_1/C_0 = 0.39$

If the first column can be operated until near complete breakthrough while maintaining arsenic concentrations at or below 2 Fg/L, then the run lengths of 15,400 BV and 23,100 BV are probably underestimated due to the conservative assumptions used in their derivation.

18. The other major chemical cost is the pH adjustment chemicals for the two designs where pH is adjusted to the optimal pH. The following equation was used for chemical costs for pH adjustment:

Chemical cost (pH adjustment) = 81361* Average Flow + 151.7

Basis. The pH is adjusted downward using hydrochloric acid. The chemical demand was calculated based on adjusting pH from pH 8 down to pH 6 with an alkalinity of 100 mg/L as $CaCO_3$. The pH is adjusted back to pH 8 using sodium hydroxide. Chemical costs were developed for average flows ranging from 0.05 to 1.5 mgd. The Phoenix case study presented chemical costs of $5742 for a flow of 0.2 mgd (*4*). The pH was adjusted from the pH 7.5 - 7.8 range down to pH 6 for those costs. The alkalinity was not listed for Phoenix. For an average flow of 0.15 mgd, the chemical costs were estimate to be $12,208. The chemical costs may be conservative for systems with lower pH and lower alkalinity. Systems with higher alkalinity may experience higher costs. It was not possible to compare with the other AwwaRF case studies because acid and base are used for regeneration as well as pH adjustment.

19. The second major component in the O&M costs is incremental labor cost. Incremental labor is the labor associated with the additional maintenance that comes with a new process. Adjustment in staffing or shifting of activities are not included in the incremental labor. The incremental labor for the activated alumina process without regeneration is one hour per week. In addition, time is also included for media replacement (at 5% breakthrough) depending upon the volume of media being replaced. Sixteen hours was assumed for design flows less than 1 mgd and thirty-two hours for larger design flows. Incremental labor for the pH adjustment processes is also included in the two options operated at optimal pH. The incremental labor is an additional two hours per week.

Basis. The labor requirements when operated at natural pH are minimal. The chemical feeders have been eliminated since the media is replaced rather than regenerated and the process is operated at the raw water pH. If backwashing every four months is necessary, then it could be done in the allotted time. Estimated total annual labor costs were provided for the Bow, NH system (*2*). These estimates were based on four hours per month for routine inspections and 16 hours/cycle for filter media replacement. The hours for activated alumina media replacement were doubled for the design flows greater than 1 mgd because of the larger volume of media to be replaced. When operated at the optimal pH, an additional two hours per week is used for the pH adjustment processes. Technical oversight is reduced due to the use of pH controllers, pH sensors, and level sensors in this design.]

20. The labor rate for small systems was $28/hour. The loaded labor rate for large systems was $52/hour.

    Basis. A rate of $28/hour is recommended as a loaded labor rate for small systems at the EPA Technology Design Workshop Summary included in Annex A of the Phase I Upgrade Document (*8*). A loaded labor rate of $52/hour was selected for large systems based on licensed operators. For this analysis, a large system was defined as a system with an average flow greater than 0.36 mgd (corresponds to a design flow of 1 mgd). This corresponds to populations lower than 3,300 people. The use of a large system rate may overestimate costs in systems serving between 2,700 and 10,000 people.

    The Design Workshop Summary (*8*) also noted that the Bureau of Labor Statistics (BLS) publishes monthly reports that include fully burdened labor costs on a regional basis. Other members of the Panel noted that some States have conducted operator surveys that include labor rates. The Panel recommended that EPA review and compare both the BLS and State material to derive representative rates. A preliminary analysis of such material indicates that the labor rates for both large and small systems may be very conservative and that actual labor rates may be much lower.]

21. The third component in the O&M costs is the energy costs for both the building and the process. Housing electrical energy use is based upon an annual usage of 19.5 kwh/sq ft/yr,

which includes lighting, ventilation, and heating. Process electrical energy was assumed to be 440 kwh/yr. Costs were based on a rate of $0.08/kwh.

Basis. The Water Model was used for the electrical usage estimates. The housing electrical use is taken directly from the O&M cost discussion on activated alumina. Lighting is required only when the operator is present, which was assumed to be average 3 hours per day. Lighting would likely be needed less frequently under the simplified design. The process electrical energy requirements were also taken from the Water Model. However, they were modified significantly since regeneration is not being performed in this design. Process electrical energy is required for day tank mixers and diaphragm pumps to feed from the day tanks. Mixers were assumed to operate one-half hour per day, with the diaphragm feed pumps operating 24 hours per day. The other major energy use component was the electrical immersion heater used to maintain the sodium hydroxide tanks above freezing (14.4 EC) and it was assumed to operate 4 hours per day and 180 days per year. Since all of these major energy use components are not included in the design, the process energy requirements in Table 67 of the Water Model were examined for the smallest design flow (0.045 mgd). The process energy requirements were assumed to be one-tenth of the listed energy requirements and assumed to be constant over the entire flow range. Power costs were consistently higher using this approach compared with the Phoenix data (*4*).

D.3   WASTE DISPOSAL COSTS

22.   The primary waste is the spent activated alumina. Rates for disposal at sanitary landfills are based on tons of material. A wet density is assumed to be 10% higher than the dry density to convert the volume of activated alumina into a mass for disposal.

Basis. The cost estimate for the White Rock Water Company was based on the mass of the media and did not assume a higher wet density (*2*). The Phoenix case study was the only one that investigated disposable activated alumina. The disposal costs are based on a 10% increase in the mass of the media (*4*). The wet density of the media was assumed to be 10% higher than the dry density.

23. The disposal cost for the spent activated alumina media is $60/ton.

   Basis. The disposal rate of $60/ton was used to estimate costs for the White Rock Water Company (*2*). The two water system byproducts documents (14, 15) contain the same cost equation for off-site nonhazardous landfill cost for various system sizes. The cost equation for disposal in a nonhazardous landfill from these two documents was also examined to estimate disposal costs. That cost equation had two variables: tons of sludge requiring disposal and transportation distance. The transportation distance could range from 5 to 50 miles. Using the upper bound of 50 miles, the equation simplified to a multiplier of $45.58/ton. The Phoenix case study used a disposal rate of $120/wet ton (*4*). The basis for this rate is not discussed. The $60/ton rate was selected for the cost analysis because it was based on a full-scale system that was disposing spent alumina. The volume of waste being generated is not significant, so even using $120/wet ton would not appreciably affect the waste disposal costs.

24. The only other potential waste stream is backwash water. Due to the infrequent need for backwashing and the low volume of waste, no costs were assigned for backwashing.

   Basis. Both the Very Small Systems Document (*10*) and the Water Model (*6*) provided a backwash rate of 8 - 9 gpm/sq ft and a backwash duration of 10 minutes. The backwash rate and duration are based on a design where the activated alumina media is being regenerated. Backwashing was not performed at the White Rock Water Company probably due to the frequent media replacement. Backwashing is not performed on a routine basis at the other two full-scale systems (*3*). At Plant C, backwashing was performed twice over a period from August 1997 to September 1999. When backwashing was performed on August 5, 1998, the backwash rate was 3 gpm and the duration was 5 minutes. At Plant D, backwashing was originally scheduled every four months, but this was discontinued after it was found to be unnecessary. Backwashing was not performed between May 1998 and September 1999. Backwashing costs are not included in the Phoenix case study, which is the only case study to evaluate disposable activated alumina (*4*). Routine backwashing is not essential to operation of these plants and if performed, would produces a very small waste stream. Backwash water

disposal costs, if needed, are covered by the conservative assumptions that have been used throughout the costing process.

D.3   REFERENCES

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Appendix E Basis for Revised Anion Exchange Costs

APPENDIX E: BASIS FOR REVISED ANION EXCHANGE COSTS

E.1   OVERVIEW

Two sets of costs were developed for anion exchange - one where the sulfate concentration is less than or equal to 20 mg/L and the other where sulfate is greater than 20 mg/L, but less than or equal to 50 mg/L. The costs in both the April 1999 Technology and Cost Document (*1*) and the November 1999 Technology and Cost Document (*2*) utilize models based on nitrate removal. The Water Model (*3*) was one of the cost models used for small systems with design flows less than 1 mgd. Sulfate concentrations for this model were between 50 and 100 mg/L. The November 1999 Technology and Cost Document listed the design criteria for large systems (> 1 mgd) using the W/W model. The sulfate concentration was 80 mg/L and the nitrate concentration was 100 mg/L. It was not possible to modify the sulfate concentration in either of these two models. It was determined that the existing cost models could not be used to estimate costs for anion exchange options with sulfate at or below 50 mg/L. A similar summary to this was prepared for the disposable activated alumina costs used in the proposed and final rules.

E.2   DESIGN ASSUMPTIONS AND BASIS

E.2.1   Capital Costs

1.   Empty Bed Contact Time (EBCT) = 2.5 minutes per column

Basis. The designs in the November 1999 Technology and Cost Document were based on an EBCT of 2.5 minutes. Case study costs in the AwwaRF report (*4*) were based on EBCTs of 1.5 minutes. The EBCTs for two full-scale anion exchange plants removing arsenic were 3.7 minutes for Plant A and 5.6 minutes for Plant B (*5*).

2. The estimated cost of the anion exchange resin was $125/ft^3$ for all system sizes.

    Basis. The November 1999 Technology and Cost Document (2) recommended this cost based on costs at existing ion exchange facilities and conversations with suppliers. The case study costs in the AwwaRF Document (4) use a range of $110/ft^3$ at 1 mgd to $90/ft^3$ at 30 mgd.

3. The bed depth assumption ranged from 3 feet to 6 feet depending upon the design flow. The maximum diameter per column was 12 feet. When the calculated media volume exceeded the maximum, the treatment media was divided into several beds to maintain a realistic bed diameter. Thus, larger designs include multiple anion exchange columns, each treating a portion of the total flow.

    Basis. The costs for the activated alumina used this approach (6).

4. The vessel cost has been sized based on 50% bed expansion during backwash. The vessel volume was calculated as (1.5)(media volume).

    Basis. The Water Model used 50% bed expansion to size the vessel (3).

5. The vessel cost is based on the following equation:

    Cost = 63.288 * (Vessel Volume in gallons) ^ 0.679

    Basis. Quotes were obtained from two manufacturers for glass-lined carbon steel vessels with working pressure rating between 75 and 150 psi and meeting American Society of Mechanical Engineers (ASME) Code Section VIII requirements. Sizes ranging from 140 gallons to 6764 gallons were used to develop the equation. Vessel costs for sixteen different sizes were regressed against their respective volumes to derive the equation.

6.   The capital costs include a redundant column to allow the system to operate while the media is being regenerated in the other column.

      Basis. A redundant vessel has been included for each anion exchange column. While the anion exchange column is being regenerated, the other column can produce drinking water. Redundant columns were not included at either of the two full-scale arsenic anion exchange plants (5).

7.   The other equipment cost elements are the brine tank, brine pump and mixer. Costs for the brine tank were based on 30-day storage. Separate costs were prepared for the two sulfate ranges based on the number of regenerations per month.

      Basis. Brine tank costs were based on the salt needed to perform regenerations every 1500 bed volumes and 700 bed volumes over a 30 day period. Cost data from RS Means "Environmental Remediation Cost Data" on plastic storage tanks were used to develop the brine tank costs (7). Data from this source were also used for the brine pump costs. A manufacturer cost estimate was used for the brine mixer.

8.   The remaining components of the process equipment include concrete, pipe and valves and electrical and instrumentation. A cost factor of 25.02% of the manufactured equipment and activated alumina was used for these other construction cost components.

      Basis. The Water Model provided construction cost estimates for a variety of conceptual designs (3). Construction cost components included process costs such as manufactured equipment and media, concrete, pipes and valves, electrical and instrumentation as well as construction costs such as excavation and sitework. The weighted percentages for concrete, pipe and valves and electrical and instrumentation were expressed as a percentage of the manufactured equipment costs. Resin costs were included in the manufactured equipment costs. Concrete was 7.32%, steel was 9.51%, pipe and valves were 3.02%, and electrical and instrumentation were 5.16% of the equipment costs. The total was 25.02%.

9. The capital costs have been estimated from the total process costs using a factor of 2.5 for small systems and 3.33 for large systems.

    Basis. The capital costs have been estimated from the process equipment costs using the ratio in the Guide for Implementing Phase I Water Treatment Cost Upgrades (*8*). The process costs were assumed to be 40% of the capital costs for small systems. The breakdown of capital costs for small systems is as follows: 40% process costs, 40% construction costs, and 20% engineering costs. For large systems, the breakdown of capital cost is: 30% process costs, 40% construction, and 30% engineering.

10. The capital costs estimated in the previous step do not include the cost of a building to house the process equipment and other add-on costs related to the site. Housing costs were assumed for all sites. Fence and road were only assumed for those systems that have no treatment.

    Basis. Data from the 1995 Community Water Systems Survey were used to identify the percentage of systems without treatment (*9*).   The data from ground water systems is as follows:

Table E-1.

Percentage of Ground Water Systems with No Treatment: Systems Serving # 3,300 People

Population Category			
25 - 100	101 - 500	501 - 1,000	1,001 - 3,300
43%	19%	16%	18%

Table E-2

Percentage of Ground Water Systems with No Treatment: Systems Serving > 3,300 People

Population Category			
3,301 - 10,000	10,001 - 50,000	50,001 - 100,000	100,001 - 1 MIL
13%	1%	11%	0%

Larger ground water systems rely on multiple entry points to supply ground water rather than one well. Thus, treatment may be present in the system, but only at one of the wells. Since the analysis is based on no treatment throughout the system, using the percentages for larger systems could underestimate the need for roads and fences at some entry points. Therefore, the 18% without treatment from the 1,001 - 3,300 category was applied to the four larger system categories. The following table summarizes the percentage used for each of the size categories.

Table E-3

Assumed Percentage of Ground Water Systems with No Treatment

Population Category	Percentage of GWS w/o any Treatment
25 - 100	43%
101 - 500	19%
501 - 1,000	16%
1,001 - 3,300	18%
3,301 - 10,000	18%
10,001 - 50,000	18%
50,001 - 100,000	18%
100,001 - 1 MIL	18%

Thus, the cost of a road and fence was calculated and 43% of that total was included in the unit cost for activated alumina for systems serving between 25 and 100 people.]

11. The equation for process area, which is the key variable in estimating the costs for buildings, roads, and fences has been based on the design flow of the system. The equation used for process area is:

Process Area = 118.7 * Design Flow + 118

Basis. The basis for the cost equations for buildings, fence and road are described in Section 4.0 of the Very Small Systems Document (*10*). The Very Small Systems Document provided costs as a percentage of the capital costs for design flows from 0.024 to 0.65 mgd. The process area was calculated from the building cost estimates by dividing by 40 (building costs were based on $40/sq ft). The process areas for the design flows in the Very Small Systems Document were regressed and found to best fit a linear regression:

Process Area (sq ft) = 118.7 * Design Flow + 118

The process area for the conceptual designs in the Water Model (*3*) were also regressed and fitted to a linear regression. The conceptual designs covered resin volumes from 4 to 520 cubic feet. The linear regression from this source is:

Process Area (sq ft) = 171.8 * Design Flow + 103

The process area estimates using the Very Small Systems Document equation were selected to make cost estimates for buildings, road and fence. The process area estimates using this equation were consistently higher than the Water Model estimates over the entire range of design flows under consideration. Since there would be a minimum building size, extrapolating a linear regression below the lowest data point does not create a significant problem with the intercept.

12. The building cost to house the activated alumina system is $45.38/sq ft for design flows at or below 0.65 mgd. The following equation is used for design flows between 0.65 and 1 mgd:

$$\text{Building cost} = 54.204 * \text{Process Area} + 14887$$

For design flows greater than 1 mgd, the building cost varied based on flow from $100 to $90 per square foot.

Basis. The data from both the Very Small Systems Document and the Water Model were examined to develop housing costs for the anion exchange system. The housing rate listed in the Very Small Systems Document was $40/sq ft (*10*). This rate includes the following: basic storage building, foundation, electrical wiring, HVAC, and plumbing. The rate is for a pre-fabricated building installed on a concrete slab. The rate is based on data from pre-fabricated building manufacturers and the 1993 Means Building Construction Cost Data for storage buildings multiplied by a conversion factor for small-sized buildings. The housing costs in the Water Model for activated alumina were used because it was easier to link to design flow and for consistency purposes with activated alumina costs (*3*). The costs in the Water Model were developed in 1983. The ENR Building Cost Index was used to update both sets of costs to September 1998. The housing rate from the Very Small Systems Document was selected based on better documentation and the use of more recent data for design flows up to 0.65 mgd. For design flows between 0.65 and 1 mgd, a linear regression of activated alumina housing cost against process area was used to derive the equation listed above. For design flows greater than1 mgd, the building cost rate assumptions were the same as those used for the case studies in the AwwaRF Report for activated alumina (*4*).

E.2.2 Operation and Maintenance Costs

There are major components to the operation and maintenance (O&M) costs are regeneration frequency, regeneration dose, and incremental labor.

13. The critical variable in calculating the regeneration frequency is the run length - the number of bed volumes until the effluent concentration exceeds a specific concentration. The run

length when sulfate is at or below 20 mg/L is 1500 bed volumes (BV). The run length when sulfate is between 20 and 50 mg/L sulfate is 700 BV.

Basis. The November 1999 Technology and Cost Document (*2*) utilized a breakthrough curve from a computer program based on equilibrium multicomponent chromatography theory with constant separation factors (*11*). The number of bed volumes for 10 percent and 50 percent breakthrough of influent arsenic were presented as a function of influent sulfate concentration. The 10 percent breakthrough curve would correspond to 90% removal of arsenic. This breakthrough curve was presented in Figure 3-37 of the November 1999 Technology and Cost Document (*2*). Run length for the entire range was based on the run length at the highest sulfate concentration. A run length of 1500 BV for sulfate concentrations up to 20 mg/L is based on the run length at 20 mg/L. A run length of 700 BV for sulfate concentrations in the range of 20 to 50 mg/L is based on the run length at 50 mg/L. This is a very conservative assumption. For example, a system with 25 mg/L would have a run length of 1300 BV, but is assumed to have a run length of 700 BV. The actual run length would be nearly double the run length being used to estimate costs. One other source of conservatism in these estimates. The breakthrough curve presented in the November 1999 Technology and Cost Document may underestimate actual run length. The run length observed in pilot testing at Albuquerque was 506 BV when regeneration was performed with 2.0 equivalent chloride per equivalent resin and downflow regeneration with 1 molar (M) salt solution (*12*). The sulfate level was 81.7 mg/L. The predicted run length would be approximately 420 BV using the graph in the November 1999 Technology and Cost Document (*2*).

The regeneration frequency can be calculated using the run length (number BV), the size of a BV (gallons), and the average daily flow. The regeneration frequency is reported in days. This can also be reported as number of regenerations per week.

14. The regeneration dose is also a critical element for the O&M costs. The salt dose for regeneration was 10.2 lb/ft^3.

Basis. Regeneration in the Albuquerque study (*12*) was achieved using a 1 M sodium chloride solution at a regeneration level of 2 eq chloride/eq resin (10.2 lbs NaCl/ft^3 resin).

15. The second major component in the O&M costs is incremental labor cost. Incremental labor is the labor associated with the additional maintenance that comes with a new process. Adjustment in staffing or shifting of activities are not included in the incremental labor. The incremental labor for the anion exchange is one hour per week plus three hours per regeneration.

   Basis. The Very Small Systems Document states that fours per day are required for resin regeneration, routine equipment maintenance and process oversight (10). This section covers both cation and anion exchange, so this estimate is likely based on daily or more frequent regeneration for cation exchange softening or nitrate removal. The Water Model (*3*) lists labor hours per year based on the frequency of regeneration. The number of hours varies based on resin volume. For daily regeneration, the range is between 200 and 300 hours per year - less than one hour per day. One hour per week for routine equipment maintenance and three hours per regeneration was assumed based on both sources.

16. The labor rate for small systems was $28/hour. The loaded labor rate for large systems was $52/hour.

   Basis. A rate of $28/hour is recommended as a loaded labor rate for small systems at the EPA Technology Design Workshop Summary included in Annex A of the Phase I Upgrade Document (*8*). A loaded labor rate of $52/hour was selected for large systems based on licensed operators. For this analysis, a large system was defined as a system with an average flow greater than 0.36 mgd (corresponds to a design flow of 1 mgd). This corresponds to populations lower than 3,300 people. The use of a large system rate may overestimate costs in systems serving between 2,700 and 10,000 people.

   The Design Workshop Summary (*8*) also noted that the Bureau of Labor Statistics (BLS) publishes monthly reports that include fully burdened labor costs on a regional basis. Other members of the Panel noted that some States have conducted operator surveys that include labor rates. The Panel recommended that EPA review and compare both the BLS and State material to derive representative rates. A preliminary analysis of such material indicates that

the labor rates for both large and small systems may be very conservative and that actual labor rates may be much lower.

17. O&M Costs were calculated using the data on regeneration dose, regeneration frequency, labor hours per week and adjusted labor rates using the equation in the Very Small Systems Document.

   Basis. The equation allows the critical variables to be used to estimate O&M costs. The equation produces O&M cost in cents per thousand gallons, so it must be converted to dollars per year. Run lengths of 700 BV and 1500 BV were included in the November 1999 Technology and Cost Document (*2*). The O&M costs produced by this costing procedure are consistently higher than those in the November 1999 Technology and Cost Document even at the higher flow rates. This is true for both run lengths. Even though the equation in the Very Small Systems Document listed a domain up to 0.27 mgd, it appears to function over the entire flow range.

E.3   WASTE DISPOSAL COSTS

18. The primary waste is the regenerant brine stream. Since arsenic in the brine stream would likely exceed the toxicity characteristic regulatory level of 5 mg/L, it could be considered hazardous waste. Any mixture of domestic sewage and other wastes that passes through a sewer system to a publicly-owned treatment works for treatment is excluded from being solid waste and thus can't be hazardous waste. Discharge to a POTW is the only waste disposal option being considered for anion exchange brine.

   Basis. The proposed rule listed three mechanisms to dispose of the brine stream used for regeneration.   The options were: sanitary sewer, evaporation pond, and chemical precipitation. Many comments on the proposed rule were concerned about waste streams being considered hazardous waste. Waste streams containing less than 0.5% solids are evaluated against the toxicity characteristic directly to determine if the waste is hazardous.

Arsenic in the regeneration brine will likely exceed 5 mg/L for most systems with arsenic above 15 Fg/L and sulfate below 50 mg/L. Since the brine stream would likely be considered hazardous, the evaporation pond and the chemical precipitation options were eliminated from the decision tree as options for disposal of anion exchange wastes. Discharge to a sanitary sewer was retained because domestic sewage and any mixture of domestic sewage and other wastes that pass through a sewer system to a publicly-owned treatment works (POTW) for treatment is excluded from being considered solid waste (40 CFR 261.4). Domestic sewage means untreated sanitary wastes that pass through a sewage system. Discharges meeting the above criteria are excluded from regulation as hazardous waste.

19. The volume of the waste stream is 7.7 bed volumes (BV).

    Basis: The volume of the salt solution used for regeneration is based on an equation in <u>Water Quality and Treatment</u> (*13*). The critical variables are the regenerant dose and the molarity of the salt solution. The brine volume is 2.7 BV. The displacement rinse volume can range from 2.5 to 5 BV (*12*). The upper bound of 5 BV was used to calculate waste volume for O&M costs. The total waste stream is 7.7 BV.

20. Capital and O&M costs were calculated using the two residuals management document for discharge to POTW.

    Basis: Equations in the two water system byproducts were used for the capital and O&M costs for discharge to POTW (*14*, *15*). Costs were updated to September 1998 for consistency with other cost equations.

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