

TREATMENT OF INDUSTRIAL EFFLUENT
FOR ARSENIC, MERCURY, AND SILVER

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Summary

The enclosed study involved designing a wastewater treatment facility for a power plant in Redding, Ca. The objective was to develop a design that would meet the Environmental Protection Agency's (EPA) discharge to surface water (DSW) requirements for heavy metals. An Additional design objective included obtaining an operational cost for the facility of \$200,000 per annum.

Currently, the power plant receives 5 mgd of cooling water from a local sewer facility, of which 2 mgd are lost to lost blowdown. The used cooling water (effluent) is sent back to the same sewer plant to be treated for heavy metals and discharged to the Sacramento River. The power plant is now required to treat its own effluent for high concentrations of mercury, arsenic, and silver before either discharging the effluent to the river itself, or recycling the effluent to the sewer treatment facility.

The metal concentrations in the industrial effluent are 500 ppb of arsenic, 200 ppt of mercury, and 500 ppb of silver. Arsenic and silver must be reduced to concentrations below 100 ppb, and mercury must be reduced to a concentration below 10 ppt. The scope of this study included designing a treatment facility, addressing the appropriate land issues involved (e.g. zoning, purchasing, leasing), and researching National Pollutant Discharge Elimination System (NPDES) permitting.

This report includes details of the design and operation of a WWTP on purchased land. NPDES permitting is considered for direct discharge of treated water to the river. The design of this WWTP plant involves a level of uncertainty

involving its performance and calculations. The largest area of uncertainty in the design involves the performance of the removal technologies for mercury and silver. The removal mechanism chosen for arsenic is well documented and supported in literature for both lab and full scale operations. The proposed mechanism for mercury removal involves the reduction and evaporation of mercury. The silver removal method involves the precipitation of silver chloride. Both methods are theoretically sound but need testing at the pilot plant scale.

The sources of error in the design performance will most likely come from the models and calculations used. Much of these calculations were carried out assuming ideal conditions and equipment performance. This error will most likely affect the removal efficiencies for each metal.

The proposed alternative reduces the concentrations of arsenic, silver, and mercury to 81 ppb, 36 ppb, and 10 ppt respectively. The total capital investment for this design was \$13.6 million. The annual operating cost was \$2.6 million, and the investors' rate of return was negative. The proposed plant never pays back; therefore; the construction of this plant is not recommended.

The main driver of high costs is the capital cost of the equipment. The large flowrate of water and required residence times necessitates large equipment sizes, despite the fact that only small quantities of contaminants are being removed. The design may become profitable if it is combined with treatment for other contaminants, or if the fine increases in size.

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1.0 Introduction

1.1 Overall Goal

The design presented in this report is for a wastewater treatment plant (WWTP) in Redding, CA. This plant treats used cooling water for heavy metals from a natural gas power plant in an adjacent lot. The wastewater treatment plant must be designed to handle three million gallons per day and cost less than \$200,000 per annum to operate. The designed treatment facility must reduce mercury from 200 ppt to 10 ppt, and arsenic and silver from 500 ppb to 100 ppb, in order to comply with the discharge to surface water standards (DSW) (see attached problem statement in Appendix D). In addition to designing a WWTP, other points addressed include National Pollutant Discharge Elimination System (NPDES) permitting and land acquisition.

1.2 Current Situation

A power plant is discharging water with high levels of mercury, silver, and arsenic to the local sewer treatment plant. The agreement under which the discharge has been occurring is now expired. In order to continue operating, the power plant must treat its wastewater for metals and reach the DSW requirements or pay up to \$1 million in fines and litigation annually. The plant is required by an Administrative Consent Order to meet these DSW standards by December 31, 2011.

Treatment of water on this scale often involves the generation and filtering of solids by some means. Common filtration methods are physical filtration and biofiltration. The solids may also be agglomerated by some method, and the

larger particles may be allowed to settle. Membrane filtration is unusual for treatment of wastewater streams to the quality proposed in this design, due to high costs per unit of water. The Roger Road WWTP in Tucson, AZ treats approximately 35 mgd of wastewater at a yearly cost of 24.8 cents per gallon treated per day.

1.3 Project Premise and Assumptions

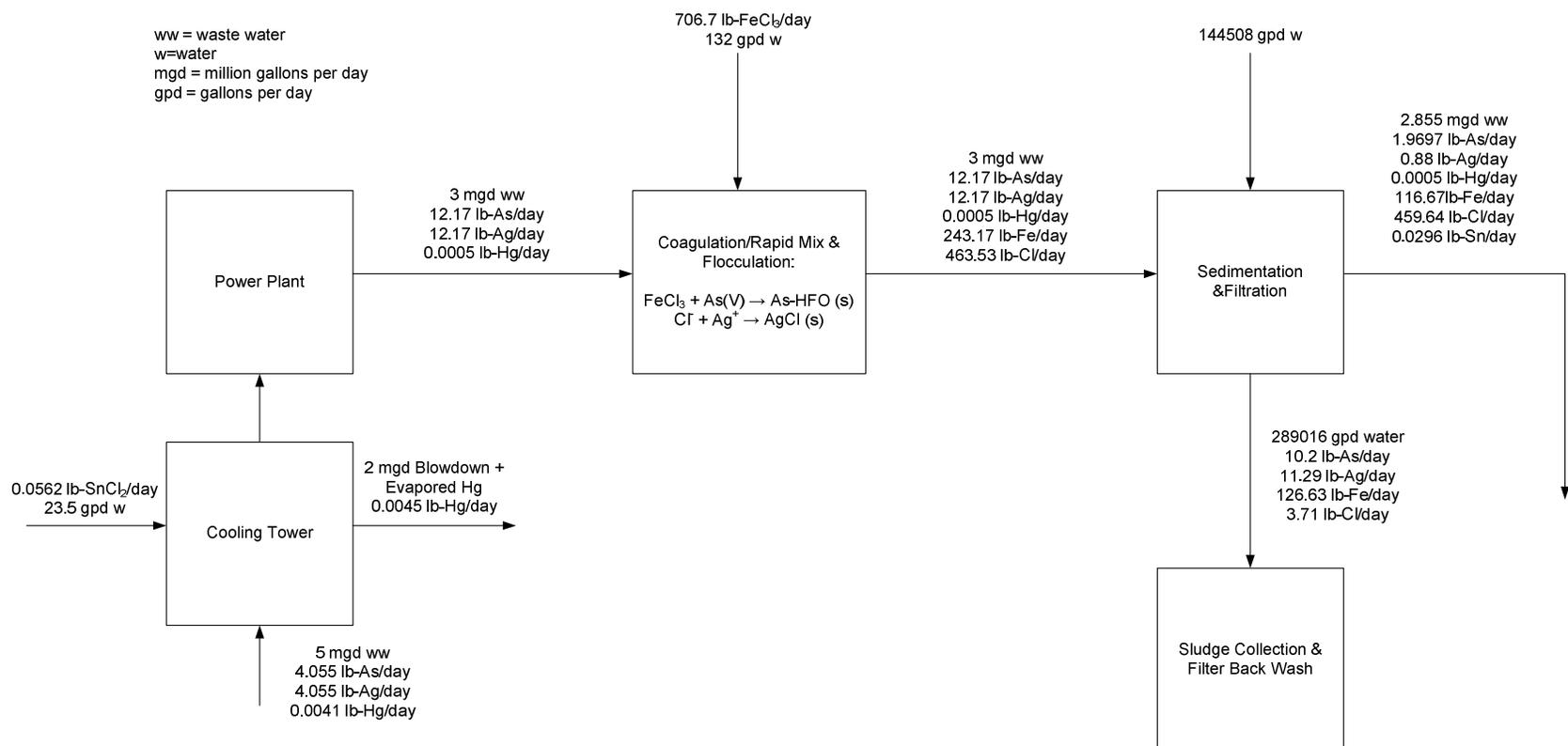
The initial proposal was to design a wastewater treatment plant on land adjacent to the power plant. The adjacent land is suitably zoned in order to construct and operate a WWTP. The treatment plant takes the power plant effluent water and treats it to meet DSW standards, after which it is discharged to the Sacramento River.

The Sacramento River is dammed near Redding for power production purposes. The river has a flowrate around 3.2 million gallons per minute, by conservative estimates. The discharges from the proposed plant will have a small effect on the river, the details of which are contained in the report.

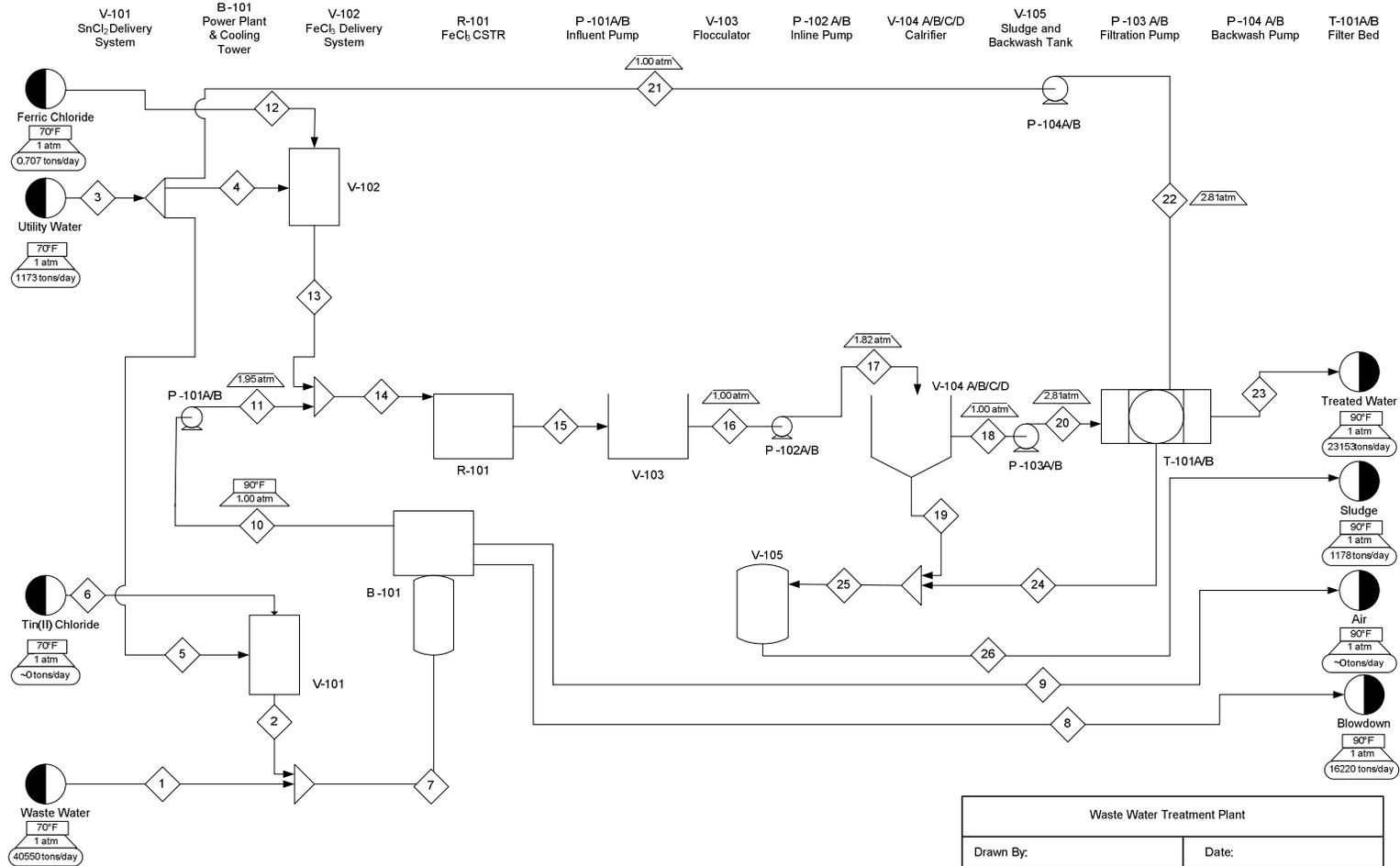
Complex cooling water flows in the power plant itself were not important for the design and they were ignored. The quality of the water that enters the process is unknown and is assumed to have no significant effect on process separation schemes.

2.0 Process Description, Rationale, and Optimization

2.1 Block Flow Diagram



2.2 Process Flow Diagram



Waste Water Treatment Plant	
Drawn By:	Date:
Checked By:	Date:
Approved By:	Date:
Drawing No.	Revision No.

2.3 Equipment Tables

Pumps	P-101A/B	P-102A/B	P-103A/B	P-104A/B
Type	Centrifugal	Centrifugal	Centrifugal	Centrifugal
Components	Influent	Flocculated Wastewater	Clarified Water	Backwash Water
Flow (gal/min)	2080	2080	2080	1204
Fluid Density (lb/gal)	8.6	8.6	8.6	8.6
Pressure Increase (psi)	14	12	26.6	26.6
Efficiency	0.88	0.88	0.89	0.88
Shaft Power (hP)	20.7	17.7	42.8	23.7
Power (hP)	23.5	20.1	48.1	26.9
MOC*	SS	SS	SS	SS

Vessels	V-101	V-102	V-103
Purpose	SnCl ₂ Storage	FeCl ₃ Storage	Flocculation Tank
Components	SnCl ₂ (aq)	FeCl ₃ (aq)	WW Slurry
Volume (gal)	23	100	62500
Area (ft ²)	0.95	-	840
Pressure (psig)	0	0	0
Temperature (°F)	70	70	90
Residence Time (min)	-	-	30
MOC*	Plastic	Fiberglass	Concrete

Vessels, Cont.	V-104A/B/C/D	V-105
Purpose	Clarifying and Settling Tank	Sludge and Backwash Tank
Components	WW Slurry	Sludge
Volume (gal)	250000	500000
Area (ft ²)	3000	-
Pressure (psig)	0	0
Temperature (°F)	90	90
Residence Time (min)	480	-
MOC*	Concrete	SS

Towers	T-101A/B
Type	Filter Bed
Components	Clarified WW
Volume (gal)	17200
Area (ft ²)	600
Pressure drop (psi)	2.59
Temperature (°F)	90
MOC*	SS

Reactors	R-101
Type	Coagulation Unit
Components	WW FeCl ₃
Volume (gal)	2100
Pressure (psig)	2
Temperature (°F)	90
Resonance Time (min)	1
MOC*	SS

*SS indicates stainless steel

2.4 Stream Tables

Stream Number	1	2	3	4	5	6	7	8	9
Temperature (°F)	70	70	70	70	70	70	70	-	-
Pressure (atm)	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Vapor Fraction	0	0	0	0	0	0	0	0	1
Mass flow (lb/day)	40600000	191	1170000	1070	191	0.0562	40600000	16200000	-
Component mass flow (lb/day)	-	-	-	-	-	-	-	-	-
Water	40600000	191	1170000	1070	191	0	40600000	16200000	-
Arsenic	4.1	0	0	0	0	0	4.1	-	0.00475
Mercury	0.00406	0	0	0	0	0	0.00406	-	0
Silver	4.06	0	0	0	0	0	4.06	-	0
Iron	0	0	0	0	0	0	0	-	0
Tin	0	0.0296	0	0	0	0.0296	0.0296	-	0
Chloride	0	0.0177	0	0	0	0.0177	0.0177	-	0

Stream #	11	12	13	14	15	16	17	18
Temperature (°F)	90	90	90	90	90	90	90	90
Pressure (atm)	1.95	1.00	1.00	1.95	1.95	1.00	1.82	1.00
Vapor frac	0	0	0	0	0	0	0	0
Mass flow (lb/day)	24300000	700	2000	24300000	24300000	24300000	24300000	24300000
Component mass flow (lb/day)	-	-	-	-	-	-	-	-
Water	24300000	0	1000	24300000	24300000	24300000	24300000	24300000
Arsenic	12.2	0	0	12.2	12.2	12.2	12.2	4.6
Mercury	0.00025	0.00025	0	0.00025	0.00025	0.00025	0.00025	0.00025
Silver	12.2	0	0	12.2	12.2	12.2	12.2	2.1
Iron	0	243	243	243	243	243	243	150
Tin	0.0296	0.0296	0.0000	0.0296	0.0296	0.0296	0.0296	0.0296
Chloride	0.0177	463	463	463	463	463	463	460

Stream #	19	20	21	22	23	24	25	26
Temperature (°F)	90	90	90	90	90	90	90	90
Pressure (atm)	1.00	2.81	1.00	2.81	1.00	1.00	1.00	1.00
Vapor frac	0	0	0	0	0	0	0	0
Mass flow (lb/day)	6000	24300000	1200000	1200000	23200000	2300000	1200000	1200000
Component mass flow (lb/day)	-	-	-	-	-	-	-	-
Water	6000	24300000	1200000	1200000	23200000	2300000	2300000	2300000
Arsenic	7.5	4.6	0	0	2.0	2.7	10.2	10.2
Mercury	0	0.00025	0	0	0.00025	0	0	0
Silver	10.1	2.1	0	0	0.9	1.2	11.3	11.3
Iron	94	150	0	0	117	33	127	127
Tin	0	0.0296	0	0	0.0296	0	0	0
Chloride	3.32	460	0	0	460	0.392	3.71	3.71

2.5 Utility Table and Raw Materials

Utility/Raw Material	Usage		Cost	per	Reference	Total/year
Water	145,000	gal/day	\$0.005	gal	(Warren, 2005)	\$264,806
Power	1,733	kW-hr/day	\$0.07	kW-hr	(Kelly, 2008)	\$44,308
FeCl ₃	129.4	ton/year	\$280.00	ton	(Fedrigo, 2008)	\$36,232
SnCl ₂ ·2H ₂ O	20.5	lb/year	\$29.58	lb	(Science Stuff, 2008)	\$606
Total Annual Costs						\$345,953

2.6 Written Description of Process

Throughout this section, refer to the equipment tables (section 5.3) and PFD (section 5.2) as needed for process and design information. As shown by the PFD, the process begins by removing mercury during the cooling tower pretreatment in the power plant. Stannous chloride dihydrate is added at a rate of 0.0562 lb/day to the water stream from V-101, a mixing tank, before entering the cooling tower. Flow in the cooling tower and power plant (B-101) is complicated and not known in detail; however, the tin reduces ionic mercury to elemental mercury during flow through the cooling tower and the reactor (Looney, 2003). The driving force for this is the equilibrium between tin/mercury valence states, as shown in the paper by Looney et al.

While the water is in the cooling tower, it will be contacted with air. Since elemental mercury is volatile, it will evaporate with the water in the cooling tower. Overall, 0.0051 lb of mercury per day enters the process and 0.0048 lb of this are evaporated per day. The sources of mercury are the input water stream and the

sulfuric acid pretreatment. The output water stream has mercury in a concentration lower than 10 ppt. The concentration in the output air stream is under 6.23 ppb (see Appendix A). Total emissions for the year are under 1000 g, well under the 3200 g/day limit required by the Sacramento Metropolitan Air Quality Management District (Sacramento Metropolitan Air Quality Management District, 1978).

The removal of arsenic (As) and silver ion (Ag^+) from wastewater begins by mixing iron salt in the form of ferric chloride (FeCl_3) with water in a mixing tank (V-102). This FeCl_3 solution is then fed to the power plant effluent stream by a gravity driven delivery system.

After the introduction of dissolved FeCl_3 to the wastewater, a rapid mix coagulation unit (R-101) is used to mix and contact the coagulant and arsenic, as well as silver ion and chlorine, for one minute. After the coagulation unit, the effluent stream enters the flocculation unit (V-103) where it is slowly mixed for 30 minutes, forming larger particle aggregates. The aggregates leaving the flocculation unit are then settled out in clarifiers (V-104A/B/C/D) for eight hours. After the clarifiers, the effluent stream is filtered. Filtration is the last separation step for arsenic and silver chloride.

For filtration, two modules (T-101A/B) run in parallel on an alternating basis. This is done so that while one filter is backwashed, the other continues treatment. The filters are composed of 20 in. of anthracite filter media, followed by 8 in. of silica sand, and 18 in. of gravel (Scott, 1995). The run time before the backwash and rinse cycles is five hrs. Using this design for an arsenic removal

treatment train, it is estimated that 83 % removal of arsenic can be achieved, meeting the requirement of 80% removal.

The sludge from the clarifier and the backwash are sent to the same storage tank. The composition of this waste is 4.34 ppm arsenic and 4.81 ppm silver. This is not classified as hazardous waste because of the mixing with the backwash. To be considered for hazardous waste classification, the waste must be solid (“Hazardous Waste Identification,” 2008). The waste product will periodically be pumped to drying ponds and then taken to a landfill.

2.7 Rationale for Process Choice

The treatment strategy for mercury has been chosen because of the constraints on this process. The two primary constraints that governed this choice were the desired low operating cost and the low initial and final concentrations of mercury. Valuing low operating costs eliminates more expensive methods such as membrane filtration or adsorption. However, the chosen treatment requires the addition of tin chloride salts, introducing these to the environment.

The treatment method has been shown experimentally to be effective in removing mercury to the desired concentration from initial concentrations comparable to those considered in this process (Looney, 2003). No other treatments of reasonable cost appear to be effective.

There is an ethical issue with regard to the transfer of mercury from water to air. The dilemma is that, rather than eliminating the release, the method of release is merely switched to avoid regulations. However, this method of release

may lead to mitigation of environmental effects. It is difficult to predict the long-term effects relating to the accumulation of mercury in the air, assuming there is accumulation. It is possible that sufficient atmospheric dispersion exists to displace the mercury, preventing regional accumulation. This issue is further addressed in the future work section (6.2)

The removal mechanism for arsenic treatment was chosen because it has been long been recognized as an effective method (Hering, 1996). Other treatment technologies that were investigated, such as solid sorbents, were excluded due to their high capital costs and high operational costs with respect to materials. Additionally, the efficiencies and removal abilities of this technology are well documented in literature, both for lab and full scale experiments. Studies by Janet Hering have shown that arsenic removal efficiencies of over 90% can be achieved by this method (Hering, 1996). The current proposed WWTP only needs to obtain an 80% reduction in the concentration of As.

The removal of arsenic (As) and silver ion (Ag^+) from wastewater involves the processes of adsorption, coagulation, flocculation, sedimentation, and filtration. In this process, iron salt in the form of ferric chloride (FeCl_3) is added to the wastewater in liquid form at a rate of 1777.17 lb/day and with a mass concentration of about 40%. When FeCl_3 dissolves in water, it forms a hydrous ferric oxide (HFO) precipitate (Hering, 1996). The As (V) species present in the water complexes with this HFO by adsorbing onto available sites (Boccelli, 2005). A Langmuir type adsorption isotherm can be used to determine the extent

of adsorption for this process (Wilkie, 1996). The silver present forms ionic bonds to the chloride freed by the formation of HFO from ferric chloride.

After the introduction of dissolved FeCl_3 to the wastewater, a rapid mix coagulation unit is used to mix and contact the coagulant and arsenic as well as silver ion and chlorine. The residence time for coagulation is one minute. The complexes formed in the rapid mix unit are then sent to a flocculation unit where they are slowly mixed for 30 minutes. This slow mixing will cause the newly formed As-HFO complexes and silver chloride to aggregate forming large particles called floc. The aggregates leaving the flocculation unit are then settled out in a clarifier.

The settling time for the clarifier is determined by the settling rate. The settling rate of a particle depends on its diameter and density, and is governed by Stoke's Law (Masters, 2004). The sizing of the clarifier is then determined based on the particle's settling rate. A large clarifier volume requirement, resulting from the settling rate, of 1,000,000 million gallons is required.

Silver chloride precipitation was chosen as the silver removal technology for several reasons. First, no additional chemicals were needed as all of the chloride ion needed is generated from the addition and dissolution of ferric chloride. The extremely low solubility of silver chloride allows for effective separation ("Butler," 1964). Additionally all the required equipment is already needed for arsenic removal, therefore there is no additional capital cost.

The last separation step for arsenic and silver chloride removal is filtration. This step is necessary because the process obtains only 70% removal of arsenic

through sedimentation (Scott, 1995). With effective filtration, arsenic removal percentages greater than 90% can be achieved (Scott, 1995). The removal of silver by sedimentation is 83%; with the addition of filtration 93% of the initial silver is removed. The filtration system to be used for this WWTP is modeled after the filtration system used by the Henry J. Mills plant in Riverside, CA. Its filter module design can handle the current flowrate of 3 mgd. The filtration run time for each module is 5 hrs and the time for backwashing is 30 minutes. The rinse time needed before beginning filtration again is 30 minutes as well. Fresh water is used for rinse to minimize the reloading of the filter with contaminants.. This results in an effective filtration rate of 2.55 gpm/ft².

The choice of FeCl₃ over an aluminum coagulant is due to the observed result that aluminum is less effective in removing arsenic than iron (Scott, 1995). The choice of FeCl₃ is also due in part to its low cost of \$0.14/lb (Fedrigo, 2008).

3.0 Equipment Description, Rationale, and Optimization

The materials of construction for the following process equipment were chosen based on their compatibility with the quality of water they will handle. This compatibility was determined using a given compatibility sheet (CHEE 442, 2008). The materials of construction for the WWTP consist of stainless steel, fiberglass, and concrete, and the criteria for compatibility was that of brackish water. Refer to the PFD and detailed calculations appendix (Appendix A) as needed, throughout this section.

There are two pieces of equipment involved in the treatment for mercury. The first is the mixing tank, V-101. The mixing tank is sized by arbitrarily choosing a flowrate out of the tank (0.016 gallons/min or 1 mL/s), chosen to be comparable to the flowrate of sulfuric acid used currently by the power plant. For one day of operation, the required volume is approximately 23 gallons. Because aqueous chlorides have the potential to be corrosive to carbon steel, plastic is chosen as the material of construction.

The other piece of equipment is the preexisting cooling tower, B-101. Since this already exists, no design choices were made for it. However, the process efficacy must be justified for the cooling tower. That is, it must be shown that the cooling tower will effectively evaporate the mercury from the mercury-water solution. The calculations justifying this are included in the detailed calculations in Appendix A. Mainly, an evaporation rate that corresponds to a 1 °F temperature drop in the tower would be sufficient to evaporate the mercury

(see calculations). Since the temperature drop is actually about 12 °F (Kelly, 2008), the cooling tower should effectively remove the mercury.

The treatment units specifically involved in the removal of arsenic are the chemical delivery system (V-102), coagulation unit (R-102), flocculation unit (V-103), clarifiers (V-104A/B/C/D), and filter beds (T-101A/B). Each unit was designed to achieve a final arsenic reduction of about 84% by mass. Calculations show that designing equipment for a specific separation of arsenic yielded a higher separation of silver. The optimization and design of the unit operations was based on observed values in literature, and removal calculations (see Appendix A). The optimization for the arsenic treatment train involved minimizing the amount of treatment chemicals needed; in the case of arsenic, the amount of FeCl_3 needed was minimized.

The chemical delivery system's purpose is to introduce FeCl_3 to the effluent stream. The contact time between the FeCl_3 and the utility water in this unit is small, just a few seconds. The mixing is done with a small, fast propeller. For this mixer ideal CSTR conditions are assumed. The validity of this assumption should be verified and is considered under future work. The volume of the mixing tank is 100 gallons. The material of construction is fiberglass.

The coagulation unit's purpose is to contact FeCl_3 with the effluent stream. In this unit operation, rapid mixing causes the arsenic to adsorb to the hydrous ferric oxide molecule (HFO) while silver forms ionic bonds to chloride. The size of the coagulation unit is based on the residence time needed to mix the effluent with the chemical stream sufficiently. Typical residence times for coagulation are

less than one min (Masters, 2004). For coagulation, the residence time is set at one minute. Using this residence time, the coagulation unit was designed to have a volume of about 2,100 gallons. This unit also includes a propeller for mixing, which delivers 277 ft-lb_f/s of work to the fluid for mixing. The material of construction is stainless steel.

The flocculation unit's purpose is to contact the As-HFO species with each other to form larger aggregates called floc. Silver chloride also forms flocs by the same mechanism. The residence time chosen for this unit is also based on literature, and has a set value of 30 minutes (Hering, 1996). The flocculation unit is designed to have a volume of about 62,500 gallons. The dimensions of the rectangular unit are 10 ft high, 100 ft long, and about 8.4 ft wide. The slow mixing is delivered by paddles. The paddles deliver 62.9 ft-lb_f/s of work to the fluid. The material of construction is concrete.

The sedimentation operation is used to settle out the larger aggregates formed in the flocculation operation. There are four sedimentation tanks used to settle out the floc (aggregates). Each tank has residence time of about 8 hours; this is based on the settling velocity calculated for a 10 μm diameter particle of As-HFO. The volume of each tank is about 250,000 gallons. The material of construction is stainless steel. From experience, the volume for these sedimentation tanks is not unusual. The Roger Road treatment plant in Tucson, AZ, has primary and secondary clarifiers with volumes greater than 500,000 gallons.

The filtration unit's purpose is to remove the smaller aggregates that were not removed by sedimentation. The filter media choice was based on a filter design for the Henry J. Mills Filtration Plant (Scott, 1995). The filter internals include 20 in. of anthracite media, 8 in of silica sand, and 18 in. of gravel. The filtration unit is operated for five hrs before the flow is switched to the second module and backwashing is done on the first module. Backwashing and rinsing of the filter are each 30 minutes long. The filtration rate for the filter is 3.46 gpm/ft², which requires a roughly 600 ft² filter bed. The effective filtration rate, which accounts for water used in backwash and rinsing, is 2.55 gpm/ft². The filtration rate during backwashing and rinsing is 2 gpm/ft². The material of construction for the filter (frame) is stainless steel.

4.0 Safety and Environment

4.1 Safety Issues

The objective of this section is to identify and address safety concerns related to employees and general practices (e.g. alarms). Detailed process safety concerning equipment is addressed at the end of this section in the form of process hazard assessments (PHA). Since the plant is not operating in the regime of high temperatures or pressures, and because none of the process equipment is extremely dangerous, more of a focus is given to employee safety and operational protocol. PPE for workers handling specialty chemicals or hazardous/toxic material will be addressed. Nine personal hazard assessments (PHA) were completed for different pieces of equipment are included at the end of this section.

The industrial effluent being treated contains metals, which, at high concentrations, can be hazardous to human health (e.g. arsenic) (Choong, 2007). Additionally, physical dangers related to equipment are inherent of processing facilities such as a WWTP. Due to these dangers, it is both appropriate and necessary to discuss the general safety protocol of plant employees.

For plant workers that will routinely be working with equipment, specific work attire will be required. This dress code will also be required of anyone that will be out and around the plant and equipment, regardless of how often. The attire for all employees is as follows: protective overall suit, steel-toed work boots, and protective eye wear (e.g. safety glasses). Employees are required to

wear these at all times, with no exceptions. Additional PPE can include hearing protection for those working in a loud area (e.g. pump station), and latex or other types of gloves for those working with solids, liquids, or other materials (Norzagaray, 2008). All the previous equipment will be provided by the company, with steel-toed boots being provided once per year if needed.

Some employees will work with specialty treatment chemicals, or handle toxic wastes produced by the plant. Ultimately, the PPE to be worn will be determined by the MSDS for each relevant chemical (Norzagaray, 2008). The treatment chemicals used by the plant include tin (II) chloride and iron (III) chloride. There is only one waste stream generated in the plant. This waste stream is comprised of adsorbed arsenic, iron hydroxide precipitate, and silver chloride precipitate. Working in the plant could result in exposure to the following chemicals: chloride, iron, silver, tin, arsenic, and mercury. The necessary attire for workers handling the chemicals in each situation is outline below.

For the following safety guidelines, refer to MSDS sheets as needed. Five MSDS sheets were used for the development of the following safety protocols. The chemical MSDS sheets used include arsenic trioxide, tin chloride (stannous chloride), mercury, ferric chloride, and silver chloride (“MSDSProvider,” 2008).

Workers that will be handling or encountering treatment chemicals will be exposed to either stannous chloride or ferric chloride. Those workers handling stannous chloride, in addition to the general safety attire, will be required to wear the following additional PPE: appropriate government approved respirator, chemical-resistant gloves, and face shield. It should also be noted that following

handling these materials workers should thoroughly wash their hands and contaminated clothing. Contaminated shoes should be discarded (“MSDSProvider,” 2008).

Workers handling ferric chloride, in addition to the general safety attire, will be required to wear the following additional PPE: rubber gloves and a Mine Safety and Health Administration approved respirator. As with all other chemicals, workers should avoid any contact. The MSDS suggests a fume hood for handling. Workers should work in a well-ventilated area, and again, wash hands thoroughly after handling (“MSDSProvider,” 2008).

For workers handling the sludge waste stream or sampling, in addition to the general safety attire, will be required to wear the following additional PPE: approved respirator, chemical-compatible gloves, and a face shield. Again, hygiene should be stressed; workers should avoid any contact and following handling workers should thoroughly wash hands and contaminated clothing (“MSDSProvider,” 2008).

Alarm and evacuation protocols are necessary for plant operation. For every building that makes up the plant, appropriate alarms will be placed and checked routinely. Appropriateness is determined by whether or not an employee is at risk to exposure from a certain chemical or hazard. If they are at risk, then an alarm will be necessary. Possible events that would call for an alarm include a gas leak, overflow, equipment failure, and chemical spill. In the event that such an alarm is sounded or an event occurs, specific evacuation plans should be implemented. These evacuation plans might involve the entire

facility to consolidate to one area of the plant, after safely securing their station. Further development of such alarm and evacuation systems will come with a final design and layout of the facility.

It should also be noted that PPE, alarm systems, and evacuation plans, outlined above would fall under OSHA regulations (29 CFR 1910) (OSHA, 2008). These regulations detail and qualify each type of PPE, alarm, and evacuation plan to be used. These regulations should be consulted before any final decisions are made with respect to worker safety.

Other hazards involved in this process that are of concern are those relating to the environment. In the event that the plant sees a failure, there may be risks posed to the surrounding environment. At the worst, if the plant fails in some major aspect the biggest risk posed by the plant is the leaking or discharging of contaminants to the surrounding environment. Other hazards might involve contacting treatment chemicals with the surrounding environment.

Discharges to the environment could involve liquids or solids. Solid discharges would come from the plant's residual sludge, and liquid discharge would come from the plant effluent. Further development of the environmental hazards posed by the plant can be found in the Environmental Impact Statement.

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: V-101	System: Wastewater treatment for heavy metals
Method: What-if	Type: Tank/Vessel	Design Intent: Produce a well-mixed aqueous solution of stannous chloride dihydrate		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		23 gallons, 0 psig		

No.: 1	Description: V-101 SnCl ₂ mixing tank				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Mixing system breaks	Loss of electricity	Manually stir the tank as necessary	None-manual mixing is adequate	Make sure a stirrer for manual stirring is available.
1.2	Foreign substance enters tank	Mishandling of chemicals	1. Empty tank into sludge tank 2. Refill tank with water/SnCl ₂	Ensure the mixing tank is closed when not being refilled.	Add a cover to the mixing tank
1.3	Tank leaks	Poor workmanship Accidental puncture	1. Empty tank 2. Patch the leak 3. Refill	Inspect tank for workmanship Keep tank in a place where it is not prone to puncture	Purchase a backup tank to prevent downtime during repair

Company: University of Arizona- Tucson Date	Plant:	Site: Redding, CA	Unit: V-102	System: Wastewater treatment for heavy metals
Method: What-if	Type: Tank/Vessel	Design Intent: Produce a well-mixed aqueous solution of ferric chloride dihydrate		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		210 gallons, 0 psig		

No.: 1	Description: V-102 FeCl ₃ Mixing Tank				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	The mixing tanks overflows?	Controls (e. g. Pump/Valve) malfunction or failure.	Emergency shutdown of unit.	Secondary containment tank for overflow. Appropriate PPE for responding personal.	1-Identify Cause 2-Rengineer/Replace 3-Resume Operation
1.2	FeCl ₃ is spilled while being fed to the mixer?	Personal Error/Accidental.	Physically contain chemical spill.	Physical containment barriers. Appropriate PPE for responding personal.	1-Identify causal action 2-remove chemical from work area 3-Resume Operation
1.3	The utility water feed fails?	Pump malfunction. Utility company malfunction.	Immediately stop of FeCl ₃ feed.	-	1-Fix malfunction 2-Resume Operation
1.4	A employee falls into unit?	Personal/Accidental Error	Emergency shutdown of entire process and immediate draining of unit.	Railing around open tanks.	1-Emergency attention given to employee 2-Identify Cause 2-Rengineer/Replace 3-Resume Operation (entire process currently at stop)
1.5	Mixer/paddle stops rotating	Pump malfunction. Utility company malfunction.	Immediately stop of FeCl ₃ feed and run everything else in batch mode	-	1-Fix malfunction 2-Resume Operation

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: P-101A/B	System: Wastewater treatment for heavy metals
Method: What-if	Type: Centrifugal Pump	Design Intent: Pump wastewater across R-101 and V-103 (CSTR and flocculator)		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		14 psi pressure increase, 2,080 gpm		

No.: 1	Description: P-101A/B – Pre-reactor pump				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	The pump fails?	Equipment failure. Controls (e.g. sensors) failure.	Switch to backup pump.	Pumps isolated in contained area.	1-Identify failure 2-Rengineer/ Replace 3-Resume Operation.
1.2	The pump explodes?	Controls (e.g. sensors) failure.	Emergency shutdown.	Pumps isolated in contained area.	1-Account for personnel 2- Seek necessary medical attention, if any. 3-Identify failure 4-Rengineer/ Replace 5-Resume Operation.

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: V-103	System: Wastewater treatment for heavy metals
Method: What-if	Type: Stirred vessel	Design Intent: Aggregate solids into larger particles		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		62,500 gallons, 0 psig		

No.: 1	Description: V-103 Coagulation and flocculation tank				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Foreign substance enters flocculator	Mishandling of chemicals	Allow substance to flow through and be clarified, filtered, or stored with sludge	None-existing safeguards adequate	
1.2	Flocculator leaks	Poor workmanship Accidental puncture Earthquake	1. Halt flow to flocculator 2. Patch the flocculator	Inspect tank for workmanship Build tank to withstand earthquake forces	Install spill containment system
1.3	Paddle system stops	Loss of electricity Machine failure	1. Determine root cause 2. Repair paddles	Multiple paddles installed so process is continued	Install backup power in case of electricity loss

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: V-104A/B/C/D	System: Wastewater treatment for heavy metals
Method: What-if	Type: Clarifier	Design Intent: Settle small particulates to form arsenic and silver sludge		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		270,000 gallons, 0 psig		

No.: 1	Description: V-104A/B/C/D Clarifiers				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	The tanks cracks, develops a hole, leaks	Metal fatigue, improper maintenance	Shut down and drain clarifier, push load to remaining clarifiers reducing residence time	Perform regular maintenance and spot check every 6 months for sign of fatigue or corrosion	1-Redesign clarifiers to handle loss of 1 clarifier
1.2	The clarifier overfills	Controls (e.g. sensors) failure	Shutdown overflowing clarifier	Design clarifiers to handle slight overflow. Place in depression so natural landscape will allow overflow to drain back into clarifier	1-Redesign clarifiers to handle excess wastewater
1.3	Personnel fall into clarifier	Operator error	Emergency shutdown of flow in and out of clarifier and shutdown scraper	Ensure railings and fences are maintained around open tanks. Place emergency shutdown controls outside tank for first responders and at several locations inside the tank for victim to shut down operation. Ensure flotation devices are located around open tanks	
1.4	Rainwater floods clarifiers	See 1.2	See 1.2	See 1.2	See 1.2

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: T-101A/B	System: Wastewater treatment for heavy metals
Method: What-if	Type: Triple media bed filter	Design Intent: Remove small floc not removed by sedimentation		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		600 ft ² , 2.59 psi pressure drop		

No.: 1	Description: T-101A/B Filter				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Breakthrough occurs in the filter?	Controls Failure (e.g. measurement devices). Media failure.	Switch to second filtration module.	Operating personnel supplied with appropriate PPE.	1-Identify failure 2-Engineer/ Replace. 3-Resume Operation.
1.2	Excessive head loss occurs over the filter?	Controls (e.g. sensors) failure.	Switch to second filtration module.	Operating personnel supplied with appropriate PPE.	1-Identify failure 2-Engineer/ Replace. 3-Resume Operation.
1.3	Excessive head loss causes the filter pump to explode or fail?	Controls (e.g. sensors) failure.	Switch to second filtration module and switch to backup filtration pump. If explosion occurs, emergency shutdown procedures will follow.	Pumps are kept in an enclosed area, isolated from personnel. Operating personnel supplied with appropriate PPE.	1-Account for personnel 2- Seek necessary medical attention, if any. 3-Identify failure 4-Engineer/ Replace 5-Resume Operation.

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: V-105	System: Wastewater treatment for heavy metals
Method: What-if	Type: Tank/Vessel	Design Intent: Hold sludge until it is taken off-site		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		500,000 gallons, 0 psig		

No.: 1	Description: V-105 Sludge storage tank				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Foreign substance enters tank	Mishandling of chemicals	Store substance and empty with rest of materials	None-existing safeguards adequate	
1.2	Tank leaks	Poor workmanship Accidental puncture Earthquake	Patch the tank	Inspect tank for workmanship Build tank to withstand earthquake forces	Install spill containment system
1.3	Tank overflows	Removal schedule is not adhered to Flow from power plant increases	1. Cut off flow to the tank 2. Remove existing sludge 3. Collect contaminated soil for treatment	Emergency shutoff system for overflow	See above

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: All	System: Wastewater treatment for heavy metals
Method: What-if	Type: Utility Power Loss	Design Intent: Provide utility power to plant and process equipment		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler			1,733 kW-hr per day	

No.: 1	Description: Parasitic power from supported power plant				
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Power is lost to plant	Problem with supported power plant	Switch to backup generators to run cooling water over turbine until it is cool	Maintain sufficient generator power and fuel to supply cooling water to power plant	
1.2	All lights are lost	See 1.1	Technicians use flashlights to resume work in the dark	Ensure enough batteries and flashlights for all technicians at all times.	

Company: University of Arizona- Tucson	Plant:	Site: Redding, CA	Unit: V-101,V-102, T-101	System: Wastewater treatment for heavy metals
Method: What-if	Type: Utility Water Loss	Design Intent: Provide utility water to process equipment		
Team Members: Geoffrey Steward, Benjamin Mills, and David Hubler		145,000 gallons per day		

No.: 1		Description: Utility water			
Item	What if...?	Root Causes/Related Questions	Responses	Safeguards	Action Items
1.1	Utility water is lost to V-101 or V-102	Problem with municipal water distribution system	Plant can continue operating until the chemical delivery systems are empty.	Retain excess water in tanks to allow for operation during water outage.	1-Redesign chemical delivery systems to include a 24 hour supply of utility water
1.2	Utility water is lost to T-101	See 1.1	Plant can continue to operate until filter bed is caked, then switch to back-up unit until it is caked. Both units can then be run as decreased efficiency until municipal water supply is restored.	Maintain back-up filter bed in clean condition. This allows plant to operate at peak efficiency for 10 hours without municipal water supply.	

4.2 Environmental Impact Statement

For this environmental impact statement (EIS), the releases to the environment prior to building the proposed design will be compared with the releases made under the proposed design. The global warming potentials (GWPs), ozone depletion potential (ODPs), human toxicity and environmental toxicity will be examined for all chemicals emitted by both processes.

An inventory of releases has been calculated for the two different designs. The basis for this inventory is 3 mgd of wastewater treated. Under current operation, all releases are in the water stream and consist of the three contaminants: arsenic, mercury, and silver. The proposed design releases some of the mercury in an air stream, as well as arsenic, mercury, silver, tin, iron, and chloride to the water stream. Table 1 below summarizes the emissions for the current design and the proposed design.

Evaluating these two designs to discern which is preferred will require weightings of the different substances released and the evaluation is therefore somewhat subjective. In lieu of this weighting and evaluation, the environmental and human effects of the substances are provided in the following.

None of the chemicals released before or after the plant, besides water and mercury, are volatile. There is no data on, or reason to believe that any of these chemicals, other than water, which is currently evaporated in the cooling tower, have an ODP or GWP. The incremental ODP or GWP due to water evaporation in this process is small.

Arsenic is well known as a toxic compound for humans. It targets enzyme reactions and affects organ systems, and can have severe long- or short-term effects in humans (ATSDR, 2008). Long-term, it is associated with cancers, and can cause vascular problems. It is associated with gangrene in the extremities. Acute exposure can cause “capillary leak” and result in shock.

Table 1: Summary of yearly releases and concentrations for a flowrate of 3 mgd.

	Current operation		Proposed operation	
	Total emissions	Concentration	Total emissions	Concentration
Arsenic	4440	500 ppb	719	81 ppb
Mercury (to water)	1.82	200 ppt	0.09	10 ppt
Mercury (to air)	-	-	1.73	<6.2 ppb (<7.2 µg/dscm)
Silver	4440	500 ppb	321	36 ppb
Tin	-	-	10.8	1.3 ppb
Iron	-	-	42600	4.8 ppm
Chloride	-	-	168000	18.8 ppm

Environmentally, arsenic can accumulate in fish or shellfish (“ToxFAQs,” 2008). Many arsenic compounds are water-soluble and so they can contaminate groundwater supplies. Arsenic generally ends up in soil or sediment. Arsenic is not substantially absorbed by plants (Peryea, 2001). The concentration of arsenic in soil may be orders of magnitude higher than the concentration in nearby plants.

The current mercury releases, as seen in Table 1, are all to a local river at a 1.82 lb/year. Under the proposed design 0.09 lbs/yr will be released the water while the remaining 1.73 lbs/yr are released as a gas. This will result in lower concentrations of mercury emitted to the environment.

Mercury has many toxic effects in humans and other living animals. It causes several types of tumors, and is a possible carcinogen (“ToxFAQs for Mercury,” 1999). Long-term exposure to high levels of mercury is known to cause tremors, shyness, irritability, changes in vision or hearing, and memory problems. Short term exposure to high levels of mercury vapors cause lung damage, nausea, vomiting, diarrhea, increase in blood pressure or heart rate, skin rash and eye irritation.

Large-scale environmental effects of mercury are largely unknown (U.S. Geological Survey, 2000). It is found that mercury bioaccumulates in most wildlife. Older and larger fish generally have the highest concentrations of mercury (“ToxFAQs for Mercury,” 1999). Mercury has been found to affect the reproduction of some species of fish and bird adversely. The proposed design lowers the concentration of mercury to the acceptable regulatory limit of 10 ppt. The long-term effects of mercury release to the environment for this plant will not be known in detail unless a study is done on this particular site, accounting for wind and flow patterns, as well as the rate of dispersion through the atmosphere and the ground.

The highest concentration of mercury vapor is at the exit of the cooling tower at $7.2 \mu\text{g}/\text{m}^3$, which is well below the OSHA limits of $150 \mu\text{g}/\text{m}^3$ (“ToxFAQs for Mercury,” 1999). With the plant in operation, the concentration and yearly mercury released to the river will be nearly 95% lower. This leads to the conclusion that the proposed design produces a favorable outcome.

Long-term exposure to silver either by ingestion or by inhalation causes argyria, a condition that turns skin blue-grey. The condition results from exposure to silver in amounts of about 1-25g over a lifetime ("Silver (CASRN 7440-22-4)," 1996). The EPA has provided a maximum reference dosage of 0.014 mg per kilogram of body weight per day. This has an oral retention factor of 4% built in, such that as long as one consumes less silver than this argyria will not set in. This means that a normal 70 kg person could drink 2 L of our discharged water daily before the WWTP is installed and 28 L daily with the proposed design without experiencing argyria. Conditions that are more serious are very rare and studies on them are inconclusive ("Silver (CASRN 7440-22-4)," 1996).

In the environment, silver has been found to bioaccumulate in fish and will adversely affect fish health (Hogstrand, 1998). The acute toxicity limits are 97 ppb for a 7-day LD50. The chronic toxicity occurs in the range of 0.1 to 5 ppb. The proposed design removes the acute threat to aquatic life but not the chronic threat. This situation is partially ameliorated by the fact that the effluent water will be diluted when it enters the river. Conservative estimates of the Sacramento River flowrate indicate that 3.2 million gallons per minute flow through the river naturally, on average, around Redding, CA (Larsen, 2006). The effluent is flowing at approximately 2 thousand gallons per minute. Thus, the discharge is diluted by a factor of 1600. The silver concentration in the river will be 22 ppt as a result of this process, well below the chronic toxicity limit.

Silver chloride will be present as result of its settling out through precipitation. Silver chloride is known to be an irritant and exposure should be minimized (“MSDSProvider,” 2008). When contact is made between silver chloride and the body, it should be immediately rinsed. There are no known ecological effects for silver chloride (“MSDSProvider,” 2008).

With the operation of this plant, stannous chloride will be introduced as part of the mercury removal process. Stannous chloride produces a detectable odor at 3.5 ppm and becomes an irritant at 15 ppm (Ceramic Materials, 2003). At concentrations of 50 ppm short exposures are dangerous and 1000 ppm may be fatal even if exposure is brief. Known hazards to the environment for stannous chloride include an LC50 of 0.9 mg/L for rainbow trout (“MSDSProvider,” 2008). The proposed design releases water with a concentration of 2×10^{-3} mg/L of stannous chloride.

Ferric chloride is used and output in large quantities in this process. There are some human toxicity effects related to this chemical (Baker, 2006); inhalation or ingestion are especially bad as ferric chloride is corrosive and will damage tissue. Strong concentrations can cause iron poisoning.

There are also some potential environmental effects of ferric chloride. The MSDS also provides lethal concentrations for some fish (Baker, 2006). These concentrations vary depending on the development of the fish, but 4 mg/L is reported as a 24-hour lethal concentration for striped bass larvae. The proposed design releases water with a concentration of 13.5 mg/L of ferric chloride. Dilution by the river results in a concentration of ferric chloride of approximately

8.4 ppb for this process. Other information on this compound is limited, but in water, it can clearly be unhealthy for some animal species.

The chlorine concentration of 18.4 mg/L in the output water stream will increase the salinity of the water. The concentration in the river resulting from this process will be 11.5 ppb. Chlorine adversely affects respiration of humans and animals (Lenntech, 2008); however, chlorine is not likely to accumulate in plants or animals. Chlorine is harmful to organisms living in contaminated soil or water.

Aside from the environmental effects, it is also important to recognize the legislation that will affect this process. Per the problem statement, the facility will be fined \$1 million/year if they continue under current operation. However, the proposed design must also be in regulatory compliance. Also in the problem statement, the required concentrations of contaminants are 100 ppb arsenic, 10 ppt mercury, and 100 ppb silver. The proposed design accomplishes these goals.

However, the goals are accomplished by adding other chemicals to the waste streams. Releases of these chemicals should be checked against legislation. The EPA does not have established drinking water Maximum Contaminant Levels for tin, chloride, and iron in drinking water ("Drinking Water Contaminants," 2008). The EPA maintains secondary (non-enforceable) levels of 250 mg/L chloride and 0.3 mg/L iron. The effluent concentration of chloride is well below this threshold, and the concentration of iron is below this after dilution

by the river. Although it is not legally necessary to meet drinking water standards for this plant, performance can be judged against these standards.

Additionally, mercury is transferred from the water stream to an air stream. This release should also remain below established regulatory limits. Rule 903 of the Sacramento Metropolitan Air Quality Management District states that air emissions of mercury are limited to the order of thousands of grams per day for sludge incineration (Sacramento Metropolitan Air Quality Management District, 1978). The proposed design releases less than 1000 grams per year, so it complies with the spirit of these regulations, although the exact requirements may be different for this type of emission.

The concentration of mercury emissions is also an issue. The New Jersey Department of Environmental Protection limits the concentration of mercury release in air to 28 $\mu\text{g}/\text{dscm}$ (New Jersey Department of Environmental Protection, 2006). The proposed process releases will be under 7.2 $\mu\text{g}/\text{dscm}$ (see Appendix A), so this aspect also complies with regulations.

5.0 Economic Analysis

A detailed economic analysis for this WWTP has been performed and is discussed below. The many factors that affect the cost of the plant were determined and then formulae detailed by Seider were used in the analysis (Warren, 2005). The plant will be built in Redding, California. The plant is somewhere between the cost of an integrated facility and grass roots construction. The plant will be built on an adjacent property to the existing power plant, but it will maintain separate facilities.

The annual costs and benefits of the plant are comprised from feedstocks needed for the separation process, utilities, operating expenses and the benefit of not paying a contamination fine. The raw materials, including utilities, needed for separation cost nearly \$350,000 per annum. A benefit of \$1 million is achieved by operating the plant instead of paying fines. A breakdown of the raw materials and benefit of the plant can be seen in Table 2.

The equipment costs have been calculated at \$7.4 million using the methods in Seider (Seider, 2005) and are summarized in Table 5 in the detailed economic calculations appendix (Appendix C). Appendix B contains all equations relating to the costs of equipment.

The equipment costs were used along with many other factors as outlined in Sieder to calculate a total capital investment of \$13.6 million (see Table 6). Among the factors detailed in Table 6 is the cost of site preparation. The lowest value of grass roots plant, 10% of total depreciable capital, has been used because the WWTP will be built on an adjacent but separate lot. This requires

the building of many but not all support facilities required for a grass roots operation. The final cost of plant start-up is approximated at \$940,000. Contractor and contingency fees have been estimated as 18% of the direct permanent investment according the methods outlined by Seider et al. The total fee amounts to just over \$1.2 million.

The operating costs have been totaled in Table 7 at \$2.6 million per year, well over the target operating costs of \$200,000.

Table 2: Raw materials and sales

Utility/Raw Material	Usage		Cost	per	Reference	Total/year
Water	145000	gal/day	\$0.005	gal	(Warren, 2005)	\$264,806
Power	1,733	kW-hr/day	\$0.07	kW-hr	(Kelly, 2008)	\$44,308
FeCl ₃	129	ton/year	\$280.00	ton	(Fedrigon, 2008)	\$36,120
SnCl ₂ ·2H ₂ O	20.5	lb/year	\$29.58	lb	(Science Stuff, 2008)	\$606
Total Annual Costs						\$345,841
Product	Production per year		Selling Price		Reference	Annual Benefit
Water	1	fine	\$1,000,000		Problem Statement	\$1,000,000
Total Annual Sales						\$1,000,000

Table 8 in Appendix C illustrates the cash flow worksheet for the project. In order to calculate cumulative present value an interest rate of only 10% was assumed for this project due to the public works nature of the project. With a total of about \$13.6 million in capital, two-year startup period, and \$1 million in benefits from avoiding fines, the project will never be profitable.

Economic hazards are an important factor in considering whether to invest in a project. Addressing the economic hazards pertaining to a certain design

project can illuminate the risks involved with continuing the project. The hazards of this design are the market fluctuations in chemical costs and utility costs.

The market for treatment chemicals could change depending on supply and demand, and could drive the prices up or down. Utilities, also susceptible to supply and demand forces, could have fluctuating prices as well. Quantifying the volatilities of the chemical and utility prices requires a more detailed economic analysis and market research. Owing to the large capital cost, no decrease in chemical or utility cost will make this project profitable.

A second analysis was performed to determine what fine would be necessary to make the project a prudent investment. This cash flow worksheet is contained in Table 9. From the analysis, when the fines for discharging contaminants exceed \$4.5 million, building and operating the WWTP will be preferable to paying the fines.

6.0 Conclusions and Recommendations

6.1 Conclusions

The proposed design is not economically feasible, and for this reason, it is not recommended that this project be constructed as designed. The fine imposed by the Administrative Consent Order is not sufficiently large to justify the proposed treatment method, is not a sufficiently large justification. This wastewater treatment plant design becomes economically viable when the total economic incentive from selling silver-rich sludge and avoidance of fines is \$4.5 million.

Other possible solutions to the heavy metal problem remain to be investigated. These solutions include paying the total fine for all metals of \$1,000,000 per year. For a 420 MW power plant, this amounts to an increase in cost of \$0.00027 per kW-hr produced. It is possible that a deal be made with the current municipal WWTP to accept a donation of equipment to treat the power plant effluent. The donation of may be more economically favorable than constructing the proposed design, due to the lower operating costs and potential tax benefits involved.

Finally, the mercury treatment process may be implemented independently of treatment for the other contaminants. Since this occurs independently from the removal of other contaminants as part of the existing process, it could be implemented without needing to construct the large equipment necessary for treatment of silver and arsenic. The mercury treatment

is cheap, but introduces tin chloride to the cooling water stream. The treatment plan for mercury alone may be justified if a lower fine can be negotiated.

6.2 Future Work

Following completion of this report, some future work remains to be done. If the choice is made to continue exploring the treatment design proposed in this report, some testing and verification must be done. The two proposed removal methods must be tested for the quality of their treatment. A representative sample of the water stream to be treated should be collected from the power plant.

The mercury removal testing should be applied before testing arsenic and silver removal, since that is the order they are treated in the process. Stannous chloride dihydrate should be added to the water, reducing mercuric ion (Hg^{1+}) to neutral mercury (Hg^0). This reaction should be checked for its specificity. If the stannous chloride also reduces silver or arsenic to their elemental forms, its addition may cause problems later on in the process. Following addition of stannous chloride, the wastewater sample should be sparged for 30 minutes with a 10:1 ratio for volume of air to volume of water. Then, test for final mercury content to see if it is below the desired concentration (10 ppt).

Future investigations on the practicality of silver ion removal by silver chloride precipitation are necessary. Bench scale tests should be conducted; upon bench scale verification, a pilot plant should be built. Once the problems, if any, with silver precipitation are addressed the WWTP could be built.

If the process is chosen, consideration will have to be given to testing during operation. Testing the process output will be necessary to ensure compliance with regulations. For example, a maximum flowrate for the discharge will allowed by the permit (or, alternatively, a deal with the sewer plant). Testing must be done to monitor this flowrate and ensure it remains, on average, below the permitted limit. Other aspects to be monitored will include concentrations of all three of the contaminants of concern: arsenic, sliver, and mercury.

The above testing will monitor short-term averages of the releases. It will also be necessary to monitor long-term contamination of the area, including the output stream and surrounding land for the releases. Although release of mercury to the air is well below established standards, accumulation on the ground in the surrounding area should be checked periodically. It is important to satisfy ethical requirements that the process mitigates environmental impact from mercury, rather than shifting the impact by media transfer. Soil near the output should also be monitored for accumulation of the metals, as well as other chemicals added in the process, such as chloride ion. In the event that accumulation exists at alarming levels, additional equipment should be designed and added. For example, a mercury stripper could be implemented in an effort to capture emissions, preventing dangerous levels of accumulation in the surrounding area.

One final consideration will be a potential donation to the existing sewer plant. Such a deal would have to be worked out individually, but it has the potential of cutting operational costs by using the existing wastewater treatment

personnel, and donating the equipment or required capital costs. An arrangement like this may have tax benefits as well.

Removal of arsenic by coagulation, flocculation, sedimentation, and filtration is well documented within literature and experimentally proven to be effective. The method is supported quite well by the citations and sources used within this report, as illustrated in the detailed calculations appendix (Appendix A). However, to ensure the investment the performance of this process and the equipment must be verified. It is recommended that future work on this project include planning for the construction and operation of a pilot plant, in an effort to obtain the real performance of the process and equipment.

Due to the limitations of this design project, there remain unanswered questions regarding the full design and operation of the proposed plant. One such issue involves the management of the wastewater treatment plant's residual sludge. This sludge is high in content for heavy metals. Several options exist for sludge management ranging from land application and landfill disposal to hazardous waste treatment by a third party. Ultimately, future work should include a more detailed characterization of the plant's residual sludge, possible via a pilot plant, and leaching tests of the sludge to determine the correct management option.

There is also a possibility of earning significant income from the heavy metal-laden sludge by selling the silver it contains. At current silver prices, if pure silver could be generated more than 1 million dollars is available to recapture. This money could be made either by building a silver reclamation unit

in the WWTP or by finding a precious metal recovery business capable of handling high concentrations of arsenic. An economic analysis on the cost of building and operating the reclamation unit against selling the silver-laden sludge at a discounted price would have to be performed in order to determine which route would be more successful.

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8.0 Appendices

A. Detailed Calculations

As Removal Calculations

Coagulation (R-101):

Coagulation residence times are typically less than one min (Masters, 2004). For coagulation, the residence time is set at one min, similar to residence times as reported by other experiments (Hering, 1996). With the residence time specified, the size (V) of the coagulation unit can be calculated.

$$V = \theta \times F = 60 \text{ s} \times \left[\frac{3,000,000 \frac{\text{gal}}{\text{D}}}{24 \frac{\text{hr}}{\text{day}} \times 60 \frac{\text{min}}{\text{hr}} \times 60 \frac{\text{s}}{\text{min}}} \right] \times \frac{0.1337 \text{ ft}^3}{\text{gal}} = 278.5 \text{ ft}^3 \quad (\text{A-1})$$

Perfect coagulation would result in 100 % of the As(V) present adsorbing onto the hydrous ferric oxides (HFO). To determine the extent of As(V)-HFO complexation, the following model given by the American Water Works Association and Hering (Amy, 2000; Hering, 1996) will be used. In the below equation, $[Fe]mM$ represent the concentration of iron that is in the form of a HFO.

$$\% \text{ As Adsorbed} = 100 \times \frac{K \times [Fe]mM}{1 + K \times [Fe]mM} \quad (\text{A-2})$$

Through experimental procedures, K has been determined by best-fit analysis to be 81 mM^{-1} (Amy, 2000).

It is estimated that it takes about 1 mg/L of iron to remove 50 µg/L of arsenic (EPA, 2004). This comes out to be a 20:1 mass ratio of iron to arsenic and about a 60:1 mass ratio of FeCl_3 to arsenic. The iron salt that will be used to produce the HFOs is ferric chloride (FeCl_3). To determine the amount of ferric chloride (FeCl_3) that needs to be added, the mass flow of arsenic into the rapid

mix unit is multiplied by twenty and then divided by the weight fraction of iron in $FeCl_3$.

$$\begin{aligned} Mass\ FeCl_3 &= \left[5.518 \frac{kg-As}{day} \times 20 \frac{Kg-Fe}{Kg-As} \times \frac{1}{0.3443} \frac{Kg-FeCl_3}{Kg-Fe} \times \frac{2.2046\ lb}{kg} \right] \\ &= 706.65 \frac{lb-FeCl_3}{day} \end{aligned} \quad (A-3)$$

The next step is to determine the amount of arsenic that complexes with the HFOs. The stoichiometric conversion for $FeCl_3 \rightarrow HFO$ is about 0.55 mg/L HFO for every 1 mg/L of $FeCl_3$ added (Hering, 1996). This factor will be represented by β . Using the model specified earlier, the iron concentration is calculated as follows.

$$[Fe]mM = \frac{\frac{58.09\ g-FeCl_3}{1\ g-As} \times \beta \times 5,518 \frac{g-As}{D} \times \frac{0.3443\ g-Fe}{1\ g-FeCl_3} \times \frac{1\ mole-Fe}{55.85\ g}}{3.785412 \frac{L}{gallon} \times 3,000,000 \frac{gallons}{day} \times 0.001 \frac{mole}{mmole}} = 0.0957\ mM \quad (A-4)$$

For a complete mass balance in the end, the amount of iron removed by both sedimentation and filtration, along with the amount discharged to the river, needs to be determined. To do this the amount of arsenic that forms a complex with the HFOs is calculated and a ratio of arsenic to iron is established.

$$\% As\ Adsorbed = 100 \times \frac{81\ mM^{-1} \times [0.0957]mM}{1 + 81\ mM^{-1} \times [0.0957]mM} = 88.57\ \% \quad (A-5)$$

$$Removable\ Arsenic = \frac{88.57\ \%}{100} \times 5.518 \frac{kg}{day} \times \frac{2.2046\ lb}{kg} = 10.7745 \frac{lb}{day} \quad (A-6)$$

$$\frac{Arsenic}{Fe} = \frac{10.7745 \frac{lb}{day}}{0.55 \frac{\frac{mg-HFO}{L}}{\frac{mg-FeCl_3}{L}} \times 706.65 \frac{lb-FeCl_3}{day} \times \frac{0.3443\ Kg-Fe}{1\ Kg-FeCl_3}} = 0.0805 \quad (A-7)$$

For every pound of arsenic removed, approximately 12.42 lbs of iron is also removed. For chlorine, the mass concentration can be calculated, since

removal of chlorine will only come about from backwash in the filter and the water lost in sedimentation.

$$C_{Cl} = \frac{706.65 \text{ lb-FeCl}_3 \times (1-0.3443) \frac{\text{Kg-Cl}}{\text{Kg-FeCl}_3} \times \frac{1 \text{ kg}}{2.2046 \text{ lb}} \times \frac{10^9 \mu\text{g}}{\text{kg}}}{3,000,000 \frac{\text{gallons}}{\text{day}} \times 3.67866 \frac{\text{kg}}{\text{gal}}} = 19 \text{ ppm} \quad (\text{A-8})$$

Since no arsenic is removed in the coagulation unit, a mass balance on the unit shows no net loss of arsenic

$$0 = \text{Accumulation} = \text{in} - \text{out} \quad (\text{A-9})$$

$$0 = 12.165 \frac{\text{lb}_{\text{As},10}}{\text{day}} - 12.165 \frac{\text{lb}_{\text{As},13}}{\text{day}} \quad (\text{A-10})$$

The mixing intensity for the coagulation unit can be approximated by scaling down reported values for larger full size conventional plants. For a design of 54.5 mgd, the mixing (G) intensity for a coagulation unit is 1,000-2,000 s^{-1} (Scott, 1995). For this rapid mix unit, G is calculated as follows:

$$G = \frac{1}{6} \times 1,500 \text{ s}^{-1} = 250 \text{ s}^{-1} \quad (\text{A-11})$$

The power requirement can be calculated with the following equation

$$G = \sqrt{\frac{P}{\mu \times V}} \quad (\text{A-12})$$

The viscosity at 80°F is

$$\mu(80^\circ\text{F}) = 5.1140 \times 10^{-4} \frac{\text{lb}_m}{\text{ft-s}}; \text{ (NIST, 2008)} \quad (\text{A-13})$$

giving a power requirement of

$$\begin{aligned} P &= G^2 \times \mu \times V_{floc} = 250^2 \text{ s}^{-2} \times 5.1140 \times 10^{-4} \frac{\text{lb}_m}{\text{ft-s}} \times 278.5 \text{ ft}^3 \times \frac{\text{lb}_f \text{-s}^2}{32.17 (\text{lb}_m \text{-ft})} \\ &= 277 \frac{\text{ft-lb}_f}{\text{s}} \end{aligned} \quad (\text{A-14})$$

Flocculation (V-103):

To characterize the flocculation of As-HFO aggregates, a number of parameters must be known. First, a kinetic equation that models the rate of aggregation must be found. Following this, number concentrations for the inlet and outlet must be known, along with information on the particle size distribution (PSD). The collision efficiency (α), power, tank volume, and fluid properties must also be specified.

A typical PSD for flocculation with FeCl_3 is given by Han (Han, 2002). According to this PSD, about 60% of the total particle volume is made up of about 10 μm diameter particles. Therefore, the sedimentation basin will be designed for particles with a 10- μm diameter. For the sake of easy calculation, it is assumed that the flocculation kinetics are first order and is represented by the following model given by Masters (Masters, 2004):

$$\frac{N_o}{N} = 1 + k \times \frac{V_b}{Q} \quad (\text{A-15})$$

where N_o is the influent number concentration, representing the particles of coagulated As-HFO complex, and N is the effluent number concentration. V_b the volume of the flocculation unit, Q is equal to the flowrate (F), and k is the first order rate constant defined as follows:

$$k = \frac{\alpha \times 4 \times \Omega \times G}{\pi} \quad (\text{A-16})$$

where α is the collision efficiency factor, Ω is the floc volume, and G is the mixing intensity. The floc volume can be defined as follows:

$$\Omega = \frac{\pi \times d_p^3 \times N_o}{6} \quad (\text{A-17})$$

where d_p is the particle diameter for the influent particles.

Using the model explained above, with one more assumption and a couple specifications based off experiment, a size for the flocculation unit can be determined.

This last assumption is that $N_o/N = 100$. With this assumption, the following equation, obtained by the principle of volume conservation, can be used to determine the influent particle diameter.

$$\frac{N_o \times d_p^3}{N \times d_{p,effluent}^3} = 100 \frac{d_p^3}{(10 \mu m)^3} \quad (A-18)$$

Then, d_p is 2.15 μm .

The problem now is that the system is underspecified. N_o must be known to calculate our required mixing intensity. However, instead of guessing a value for N_o randomly, educated guesses will be made, by using a scaled down full size plant mixing intensity and experimental residence time as reported by Scott and Hering (Scott, 1995). In the full size scale experiment done by Scott, the flocculation unit had a mixing intensity of 123 s^{-1} (Scott, 1995). In the paper "Arsenic Removal by Ferric Chloride," Hering uses a slow mix time of 30 min (Hering, 1996). The 123 s^{-1} mixing intensity can be scaled down by a factor of six, like the coagulation unit, giving a flocculation mixing intensity of

$$G = \frac{1}{6} \times 123 \text{ s}^{-1} = 20.5 \text{ s}^{-1} \quad (A-19)$$

The slow mix residence time can now be used to calculate the volume of our flocculation unit and use the mixing intensity to determine the power input to the flocculation paddles by the following equations. Additionally, flocculation unit can be specified to be 10 ft high.

$$Volume = \theta_{floc} \times F = 30 \text{ min} \times \left[\frac{3,000,000 \frac{\text{gal}}{\text{day}}}{24 \frac{\text{hr}}{\text{day}} \times 60 \frac{\text{min}}{\text{hr}}} \right] \times \frac{0.1337 \text{ ft}^3}{\text{gal}} = 8356 \text{ ft}^3 \quad (\text{A-20})$$

$$Base \text{ Area} = \frac{38356 \text{ ft}^3}{10 \text{ ft}} = 835.6 \text{ ft}^2 \quad (\text{A-21})$$

$$G = \sqrt{\frac{P}{\mu \times V_{floc}}} \quad (\text{A-22})$$

The viscosity at 80°F is

$$\mu(80^\circ F) = 5.7609 \times 10^{-4} \frac{\text{lb}_m}{\text{ft-s}}; \text{ (NIST, 2008)} \quad (\text{A-23})$$

which yields a calculated P of

$$P = 62.9 \frac{\text{ft-lb}_f}{\text{s}} \quad (\text{A-24})$$

Sedimentation / Clarifier (V-104):

To determine the residence time and the size for our sedimentation tank the settling velocity for the floc particles needs to be calculated. The settling velocity is calculated using Stoke's Law.

$$v_s = \frac{g \times (\rho_p - \rho) \times d_p^2}{18 \times \mu} \quad (\text{A-25})$$

where g is the acceleration of gravity, ρ_p is the settling particle density, ρ is the density of water, d_p and is the settling particle diameter. There are several surface reactions that can take place forming different As-HFO complexes; one such example is the complex FeH_2AsO_4 (Wilkie, 1996). To estimate the density of this complex, it is assumed that its density is that of iron (III) arsenate (FeAsO_4). The density of FeAsO_4 is 2.79 g/cm^3 (Au-Yeung, 1984). Using this information, and the fact that most of the particles have a diameter of $10 \mu\text{m}$, v_s can be calculated.

$$v_s = \frac{9.807 \frac{m}{s^2} \times (3010 \frac{kg}{m^3} - 996.61 \frac{kg}{m^3}) \times (10 \times 10^{-6} m)^2}{18 \times 8.5731 \times 10^{-4} \frac{kg}{m \cdot s}} = 1.28 \times 10^{-4} \frac{m}{s} \quad (A-26)$$

$$v_s = 4.2 \times 10^{-4} \frac{ft}{s} \quad (A-27)$$

For the sedimentation tank to be effective, enough time for settling must be provided. This time can be determined by setting $v_s \geq v_o$, where v_o is the critical velocity. The critical velocity is defined as the minimum settling velocity that a particle can have and still be guaranteed to be removed in the settling basin. It is given by the following equation (Masters, 2004):

$$v_o = h/\theta_o \quad (A-28)$$

The last specification needed is the height of the sedimentation tank. For this calculation, the height is set at 12 ft, giving a residence time in the sedimentation tank of

$$\theta_o = \frac{12 \text{ ft}}{4.2 \times 10^{-4} \frac{ft}{s}} \cong 28,592.6 \text{ s} \sim 8 \text{ hrs} \quad (A-29)$$

To calculate the volume of the sedimentation tank θ_o is multiplied by the flowrate F .

$$V = \theta_o \times F = 8 \text{ hrs} \times \frac{1 D}{24 \text{ hr}} \times 3,000,000 \frac{\text{gal}}{D} = 1,000,000 \text{ gal} \quad (A-30)$$

Because of the large volume required for settling, the flow will be split and sent to four clarifiers. Each clarifier will have a volume of 250,000 gallons (33,425 ft³), resulting in a diameter of about 60 ft.

A study on the removal of arsenic was carried out for an existing full-scale (106 mgd) conventional drinking water plant by Scott. The results of this paper showed that for coagulant dosages of 10 mg/L, up to 78% of the arsenic was removed in the sedimentation basin alone (Scott, 1995). The sedimentation

basin above was designed to remove 10 μm diameter particles, which account for 60% of the total volume and is the mean of the Gaussian PSD. Using this knowledge, a removal of 70% of complexed arsenic will be assumed in our sedimentation basin as a safe estimate. The remaining removal needed will be accomplished by filtration.

A mass balance on arsenic for the sedimentation unit yields the following:

$$\textit{Accumulation} = \textit{in} - \textit{out} \quad (\text{A-31})$$

$$\textit{As}, 17 = \textit{As}, 18 + \textit{As}, 19 \quad (\text{A-32})$$

$$\textit{As}, \textit{complexed} = \textit{As}, \textit{Total} \times 0.8857 = 10.7745 \frac{\textit{lbAs}}{\textit{day}} \quad (\text{A-33})$$

$$12.165 \frac{\textit{lbAs}, 15}{\textit{day}} = (0.7 \times 10.7745) \frac{\textit{lbAs}, 17}{\textit{day}} + x \frac{\textit{lbAs}, 16}{\textit{day}} \quad (\text{A-34})$$

$$x \equiv \textit{As remaining} = 4.6228 \frac{\textit{lbAs}, 16}{\textit{day}} \quad (\text{A-35})$$

where the remaining arsenic is the arsenic that continues onto the filtration unit.

Filtration (T-101):

In the same study carried out by Scott, the filter setup used for additional removal of arsenic was 20 in. of anthracite filter media, followed by 8 in of silica sand, and 18 in. of gravel. This setup was designed for a capacity of 54.5 mgd, with actual filtration rates ranging from 2.87-6.48 gpm/ft². Backwashing the filters occurred at a head loss of 6.0 ft, and filter aid accompanied arsenic removal. The removal of arsenic by the filter alone for coagulant dosages of 10 mg/L was 18% of the total. This means the removal efficiency of the filter is about 79% for the remaining arsenic, not removed in the sedimentation basin (Scott, 1995).

The dosage for this experiment is about 29 mg/L. Using the above full scale results, it is expected that for a higher dosage the same results, at least, would be achieved. The study by Scott shows that increasing the coagulant dosage, increases the percent removal (Scott, 1995). Because this process has a much higher coagulant dosage, it is assumed that the absence of filter aid will not significantly affect removal efficiencies.

This design capacity for this project's filter(s) will be 3 mgd. The same filtration rate of 3.46 gpm/ft² will be used as the plant in Scott's study.

A mass balance on arsenic for the filtration unit(s) yields the following equations

$$\text{Accumulation} = \text{in} - \text{out} \quad (\text{A-36})$$

$$\text{As accumulation} = \text{As},20 + (1 - \varepsilon_{\text{Filter}}) \text{As},23 \quad (\text{A-37})$$

where $\varepsilon_{\text{Filter}}$ is the removal efficiency specified above.

Note that an accumulation term is used here because filtration is a cyclic process. Remember, the removal efficiency refers the complexed arsenic only.

$$\begin{aligned} \text{As accumulation in filter} &= 4.6228 \frac{\text{lbAs},18}{\text{day}} - [(1 - 0.79) \times 3.23235 + (4.6228 - \\ &3.23235)] \frac{\text{lbAs},20}{\text{day}} = 2.55356 \frac{\text{lb As}}{\text{day}} \end{aligned} \quad (\text{A-38})$$

To determine how much arsenic is removed totally by filtration, the amount removed per day with the backwash needs to be calculated and included. Before that is calculated, the arsenic concentration in the effluent stream should be determined.

$$M_{\text{As},20} = 4.6228 \frac{\text{lbAs},18}{\text{day}} - 2.55356 \frac{\text{lbAs},\text{acc}}{\text{day}} = 2.06924 \frac{\text{lbAs},20}{\text{day}} \quad (\text{A-39})$$

$$C_{As,20} = \frac{(2.0693 \frac{lb,As}{day} \times (453.6 \times 10^6) \frac{\mu g}{lb})}{3,000,000 \frac{gal}{day} \times 3.67866 \frac{kg}{gal}} \cong 85 \text{ ppb} \quad (\text{A-40})$$

To meet DSW standards, an 80% reduction in the arsenic concentration must be obtained. According to these calculations, our reduction target is achieved.

$$\% C_{As, \text{reduction}} = \frac{500 \text{ ppb} - 85 \text{ ppb}}{500 \text{ ppb}} \times 100\% = 83\% \quad (\text{A-41})$$

In addition to the removal of arsenic obtained by filtration, the backwashing time, backwashing volume, and filter capacity need to be specified. For this calculation, it is assumed that the filter has a maximum solid load (MSL) of 120 grams FeCl₃/square foot of filter bed. This assumption is based off the MSL for the filter with a mono media bed at Paradise Valley (Boysen, 2008).

Using this information, the time until the maximum capacity load is reached can be calculated.

$$\begin{aligned} \text{Time} &= \frac{120 \text{ g-FeCl}_3}{\text{ft}^2 - \text{filter bed}} \times \frac{2083.34 \text{ gpm}}{3.46 \frac{\text{gpm}}{\text{ft}^2}} \times \left[\frac{706.65 \frac{\text{lb-FeCl}_3}{D} \times \frac{\text{kg}}{2.2046 \text{ lb}} \times 1000 \frac{\text{g}}{\text{kg}}}{24 \frac{\text{hr}}{D} \times 60 \frac{\text{min}}{\text{hr}}} \right]^{-1} \\ &= 346 \text{ min} \cong 5.8 \text{ hrs} \end{aligned} \quad (\text{A-42})$$

To be safe, the filtration time until backwash will be set at 5 hrs. Now with the filter time specified, the backwash and rinse time can also be chosen. The backwash and rinse time are set to be 30 minutes. Additionally, the filtration rate for the backwash and rinse cycles will be 2 gpm/ft². Now the effective filtration rate can be calculated from the above information.

$$r_{ef} = \frac{(V_f - V_b - V_r)}{A_f \times t_c}; (\text{Masters, 2004}) \quad (\text{A-43})$$

where V_f is the volume of filtered water per cycles, V_b is the volume of water used for backwash, V_r is the volume of water used for rinsing the filter, A_f is the filter area, and t_c is the duration of time for a complete filter cycle.

$$r_{ef} = \frac{\left((5 \text{ hr} \times 60 \frac{\text{min}}{\text{hr}} \times 2083.33 \text{ gpm}) - (30 \text{ min} + 30 \text{ min}) \times 2 \frac{\text{gpm}}{\text{ft}^2} \times \frac{2083.33 \text{ gpm}}{3.46 \frac{\text{gpm}}{\text{ft}^2}} \right)}{\frac{2083.33 \text{ gpm}}{3.46 \frac{\text{gpm}}{\text{ft}^2}} \times (5 \text{ hr} \times 60 \frac{\text{min}}{\text{hr}} + 30 \text{ min} + 30 \text{ min})}$$

$$= 2.55 \frac{\text{gpm}}{\text{ft}^2} \quad (\text{A-44})$$

The total arsenic removal is the sum of the amount removed via sedimentation, filtration, and backwash, and is calculated in the following manner:

$$V_b = \left((30 \text{ min}) \times 2 \frac{\text{gpm}}{\text{ft}^2} \times \frac{2083.33 \text{ gpm}}{3.46 \frac{\text{gpm}}{\text{ft}^2}} \times 4 \frac{\text{cycles}}{\text{day}} \right) = 144,508 \frac{\text{gal}}{\text{day}} \quad (\text{A-45})$$

$$\text{Mass As removed by } V_b = 144,508 \frac{\text{gal}}{\text{day}} \times 3.67866 \frac{\text{kg}}{\text{gal}} \times \frac{2.2046 \text{ lb}}{\text{kg}}$$

$$\times (85 \times 10^{-9}) \frac{\text{lb-As}}{\text{lb}} = 0.1 \text{ lb - As} \quad (\text{A-46})$$

$$\text{Total As removal} = \frac{\text{lbAs}}{\text{day}} (\text{sedimentation} + \text{filtration} + \text{backwash}) \quad (\text{A-47})$$

$$\text{Total As removal} = 7.54215 \frac{\text{lbAs,17}}{\text{day}} + (2.55356 + 0.1) \frac{\text{lbAs,21}}{\text{day}}$$

$$= 10.1957 \frac{\text{lb As}}{\text{day}} \quad (\text{A-48})$$

$$\% \text{ As removed} = \frac{10.1957 \frac{\text{lbAs}}{\text{day}}}{12.165 \frac{\text{lbAs}}{\text{day}}} \times 100\% = 84\% \quad (\text{A-49})$$

Sludge Tank (V-105):

The backwash from the filter is mixed with the sludge from the clarifier. The sludge needs to be characterized to determine if can be considered hazard waste. It is assumed that the water content in the sludge is 98%. Then the sludge is characterized by its arsenic concentration and is calculated as follows:

$$C_{As,waste} = \frac{(V_b \times C_{As,b} + m_{As, filter} + m_{As, sludge})}{(2 \times V_b \times \rho + m_{solids,b}) + \frac{m_{solids}}{0.02} \frac{m_{solids}}{m_{total}}} \quad (A-50)$$

Here ρ is taken to be the density of water: 3.67866 kg/gal. All the other values are on a daily period.

$$C_{As,b} = C_{As,out} \cong 85 \text{ ppb} \quad (A-51)$$

$$(V_b \times C_{As,b} + m_{As, filter} + m_{As, sludge}) = \text{Numerator} \quad (A-52)$$

$$\text{Numerator} = \left(144,508 \text{ gal} \times 3.67866 \frac{\text{kg}}{\text{gal}} \times \frac{(85 \times 10^{-9}) \text{ kg}_{As}}{\text{kg}} + (2.55356 + 7.54215) \frac{\text{lb}_{As}}{\text{D}} \times \frac{\text{kg}}{2.2046 \text{ lb}} \right) \quad (A-53)$$

$$(2 \times V_b \times \rho + m_{solids,b}) + \frac{m_{solids}}{0.02} \frac{m_{solids}}{m_{total}} = \text{Denominator} \quad (A-54)$$

$$\text{Denominator} = 2 * 144,508 \text{ gal} \times 3.67866 \frac{\text{kg}}{\text{gal}} + 32.9575 \frac{\text{lb}_{Fe}}{\text{D}} + 0.3915 \frac{\text{lb}_{Cl}}{\text{D}} + 1.19 \frac{\text{lb}_{Ag}}{\text{D}} + 2.6532 \frac{\text{lb}_{As}}{\text{D}} + \frac{\left(7.54215 \frac{\text{lb}_{As}}{\text{D}} + 10.0951 \frac{\text{lb}_{Ag}}{\text{D}} + 93.6745 \frac{\text{lb}_{Fe}}{\text{D}} + 3.3229 \frac{\text{lb}_{Cl}}{\text{D}} \right) \times \frac{\text{kg}}{2.2046 \text{ lb}}}{0.02 \frac{m_{As}}{m_{total}}} \quad (A-55)$$

$$C_{As,waste} = \frac{\text{Numerator}}{\text{Denominator}} \quad (A-56)$$

$$C_{As,waste} = 4.34 \text{ ppm} \quad (A-57)$$

This is not a solid waste, and therefore is not considered a hazardous waste ("Hazardous Waste Identification," 2008). The storage tank needs to

contain the waste before it is discharged. The waste storage tank size will be 500,000 gallons.

Hg Removal Calculations

The problem statement indicates that mercury in the 3 mgd wastewater stream must be removed so that the output concentration is 10 ppt. The initial concentration of mercury is 200 ppt, meaning that out of approximately 0.0051 pounds, 0.0048 pounds will be removed per day.

The removal is accomplished through a two-step process that is presented in “Removal of Mercury from Low-Concentration Aqueous Streams Using Chemical Reduction and Air Stripping” (Looney, 2003). First, stannous chloride dihydrate ($\text{SnCl}_2 \cdot 2\text{H}_2\text{O}$) is added to the water stream. This reacts preferentially with the mercury, reducing all ionic mercury (Hg^{2+}) to neutral mercury (Hg^0). Following this, the water stream is contacted with air to strip mercury from the water. This can be done by sparging; in this solution, it is proposed that this may be accomplished in the already existing cooling tower.

Chemical Needs:

The only reagent needed in this process is the stannous chloride dihydrate. Looney, et al, suggest that a 5 to 25 times stoichiometric ratio of stannous chloride dihydrate to mercury is sufficient to reduce the mercury (Looney, 2003). For the purposes of this design, the ratio is chosen to be 10.

Thus, approximately 25.5 grams of $\text{SnCl}_2 \cdot 2\text{H}_2\text{O}$ will be needed:

$$\frac{2.27 \text{ grams Hg}}{\text{day}} \times \frac{1 \text{ mol Hg}}{200.59 \text{ g Hg}} \times \frac{10 \text{ mol SnCl}_2 \cdot 2\text{H}_2\text{O}}{\text{mol Hg}} \times \frac{225.616 \text{ g SnCl}_2 \cdot 2\text{H}_2\text{O}}{\text{mol SnCl}_2 \cdot 2\text{H}_2\text{O}} = \frac{25.5 \text{ grams SnCl}_2 \cdot 2\text{H}_2\text{O}}{\text{day}}$$

(A-58)

This results in a total need of 9.32 kg of stannous chloride dihydrate per year or 0.0562 lb stannous chloride per day.

Mixing Tank:

A concentrated aqueous solution of stannous chloride dihydrate will be mixed in a tank prior to diluting into the water stream. Making the design choice that 1 ml/s is pumped out of the mixing tank, the tank will be about 8.64 L in order to hold a day's worth of solution. To this, the 25.5 grams of stannous chloride dihydrate will be added.

Assuming the tank is cylindrical and that the height to diameter ratio is 3 (Turton, 2003),

$$V = \frac{\pi D^2 h}{4} = 86400 \text{ cm}^3 \quad (\text{A-59})$$

$$D^3 = 36669 \text{ cm}^3. \quad (\text{A-60})$$

The results of this are a 33.3 cm diameter and a 99.7 cm height for the tank.

Evaporation:

The water-mercury-stannous chloride mixture must then be contacted with air to strip the mercury out. It is proposed that this occurs in the cooling tower of the power plant. The data from Looney, et al, for a residence time of 60 minutes, indicate that a volume of gas to volume of water ratio of 20 should be sufficient for evaporation of 95% of the mercury (Looney, 2003). So, a volume of 55,000 gallons/minute of air is chosen as a basis for calculation. In reality, the actual flowrate must be much larger to achieve 500-600 gpm evaporation.

At the operating temperature of 90 °F, the maximum water evaporated into this air is 1.89 gallons/minute. This is approximately 0.1% of the water being

cooled. The blowdown factsheet from the North Carolina Department of Environment and Natural Resources indicates that about 1.2% of the water should be evaporated to achieve a 10 °F temperature drop (“Water Efficiency,” 2008). The needed evaporation is only one-tenth of that; thus, even a 1 °F temperature drop should indicate that enough water is evaporated to ensure the mercury is removed. Tim Kelly indicates that the temperature change across the tower is approximately 12 °F, and this is sufficient.

One concern with this solution is that mercury is emitted through an air stream. The concentration of mercury in this air stream should be kept low enough to meet standards. Under this design, the mercury concentration in the air is as follows:

$$\frac{1.73 \text{ lb Hg}}{\text{year}} = \frac{785 \text{ grams Hg}}{\text{year}} \quad (\text{A-61})$$

$$\begin{aligned} \text{moles air} &= \frac{1 \text{ atm} \times \frac{55000 \text{ gallons}}{\text{minute}} \times \frac{3.785 \text{ L}}{\text{gallon}} \times \frac{525960 \text{ minutes}}{\text{year}}}{0.08206 \frac{\text{L} \cdot \text{atm}}{\text{mol} \cdot \text{K}} \times 305.4 \text{ K}} \quad (\text{A-62}) \\ &= \frac{4.370 \times 10^9 \text{ moles air}}{\text{year}} \end{aligned}$$

$$\begin{aligned} \text{mass air} = \text{moles air} \times \text{molar mass} &= \frac{4.370 \times 10^9 \text{ moles air}}{\text{year}} \times \frac{28.84 \text{ g}}{\text{mole}} \quad (\text{A-63}) \\ &= \frac{1.260 \times 10^{11} \text{ g air}}{\text{year}} \end{aligned}$$

$$\frac{785 \text{ grams Hg}}{1.260 \times 10^{11} \text{ g air}} = \frac{6.23 \times 10^{-9} \text{ g Hg}}{\text{g air}} = 6.23 \text{ ppb} \quad (\text{A-64})$$

Ag Removal Calculations

The reaction:



has been proposed to reduce silver concentration from 500 ppb to 100 ppb.

The K_{sp} for this reaction is 1.8×10^{-10} (Butler, 1964). Using the calculated $[Cl^-]$ from Geoff, 0.5 mmol by addition of $FeCl_3$ the saturation conditions for $[Ag^+]$ and hence precipitation point can be calculated.

$$K_{sp} = 1.8 \times 10^{-10} = [Ag^+][Cl^-] \quad (A-66)$$

Rearranging this equation,

$$[Ag^+] = \frac{K_{sp}}{[Cl^-]} = \frac{1.8 \times 10^{-10} M^2}{0.0005 M} = 3.6 \times 10^{-7} M = 38 ppb \quad (A-67)$$

This number is significantly lower than the required 100 ppb. Therefore, process calculations can proceed.

In order to determine mixing and settling time, some data on flocculation kinetics is needed. With floc sizes at 0.9 microns in diameter, 4 hours is necessary to settle 90% of the Ag^+ in stagnant water with diffusion controlled kinetics (Insley, 1967).

In our situation, the $AgCl$ is flowed into a rapid mix tank followed by a slow mix flocculation tank. These two effects will increase floc size as well as shorten the time needed for the crystals to grow.

In the absence of any other data, it is assumed that the diameter of silver flocs increase by an order of magnitude, thus the diameter is 9 microns.

With this estimation, a settling speed calculation can be applied to ensure $AgCl$ will settle in an appropriate time. The settling speed equation is given by

$$v_s = \frac{g \times (\rho_p - \rho) \times d_p^2}{18 \times \mu} \quad (A-68)$$

where g is the acceleration of gravity, ρ_p is the settling particle density, ρ is the density of water, d_p and is the settling particle diameter.

The viscosity at $80^\circ F$ is

$$\mu(80^\circ F) = 5.7609 \times 10^{-4} \frac{lb_m}{ft-s}; \text{ (NIST, 2008)} \quad (\text{A-69})$$

The density of silver chloride is given as 5560 kg/m^3 (“MSDSProvider,” 2008). Inputting all the numbers, the settling velocity is

$$v_s = \frac{9.807 \frac{m}{s^2} \times \left(5560 \frac{kg}{m^3} - 996.61 \frac{kg}{m^3}\right) \times (9 \times 10^{-6} m)^2}{18 \times 8.5731 \times 10^{-4} \frac{kg}{m-s}} = 2.4 \times 10^{-4} \frac{m}{s} = 7.9 \times 10^{-4} \frac{ft}{s} \quad (\text{A-70})$$

For the sedimentation tank to be effective, enough time for settling must be provided. This time can be determined by setting $v_s \geq v_o$, where v_o is the critical velocity. The critical velocity is defined as the minimum settling velocity that a particle can have and still be guaranteed to be removed in the settling basin. It is given by the following equation (Masters, 2004):

$$v_o = h/\theta_o \quad (\text{A-71})$$

The last specification needed is the height of the sedimentation tank. For this calculation, the height is set at 12 ft, giving a residence time in the sedimentation tank of

$$\theta_o = \frac{12 \text{ ft}}{7.9 \times 10^{-4} \frac{ft}{s}} \cong 15189 \text{ s} = 4.2 \text{ hr} \quad (\text{A-72})$$

This settling time is less than that require by the HFO complex and thus no further modifications are needed for the process.

All calculations confirm that it should be possible to precipitate out 462 ppb, but for an added safety measure, an efficiency of 90% of the theoretically

possible value is used. This results in 415 ppb settled out and a remaining concentration of 85 ppb.

B. Equipment Costing

Coagulation/Rapid Mix Unit (R-101)

The cost of the Coagulation unit will include the vessel itself and the mixing propeller. The equation for a closed vessel agitator is

$$C_p (\text{Propeller}) = 2,600S^{0.17} \quad (\text{B-1})$$

where S is the size factor with a range of 1-8 Hp. Since our power requirement is less than one Hp, S will be 1.

$$C_p (\text{Propeller}) = 2,600 \times (1)^{0.17} = \$2600 \text{ (CE Index 394)} \quad (\text{B-2})$$

For the rapid mix vessel the following equation for a cone roofed process vessel is used

$$C_p (\text{Coagulator}) = 14V^{0.51} \quad (\text{B-3})$$

where V is the volume in gals. The coagulator has a volume of about 2,100 gallons.

$$C_p (\text{Coagulator}) = 14 \times 2,100^{0.51} = \$692.6 \quad (\text{B-4})$$

To find the total installed cost, multiply the total purchase cost by a bare module costing factor taken from Seider (2005)

$$C_p (\text{Propeller}) + C_p (\text{Coagulator}) = C_p = \$3,292.6 \quad (\text{B-5})$$

$$C_{BM} (\text{Propeller}) = C_p \times F_{BM} = \$3,292.6 \times 3.05 = \$10,042.3 \quad (\text{B-6})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$13,356.3 \quad (\text{B-7})$$

Chemical Delivery System (V-102)

The cost for the chemical delivery system for ferric chloride will include the tank, mixing propeller, and the first year's supply of ferric chloride will not be a need for a pump since it is a gravity fed system. The cost for a 1 Hp propeller is

$$C_p(\text{Propeller}) = 2,600S^{0.17} = 2,600 \times 1^{0.17} = \$2,600 \quad (\text{B-8})$$

The cost for the 100 gal closed roof vessel made of stainless steel is

$$C_p(\text{tank}) = 210V^{0.51} \times F_M = 210 \times 100^{0.51} \times 2 = \$4,398 \quad (\text{B-9})$$

The cost of ferric chloride is \$0.14/lb and includes delivery (Fedrigon, 2008).

$$C_p(\text{FeCl}_3) = 365 \frac{\text{Days}}{\text{year}} \times 706.65 \frac{\text{lb}}{\text{Day}} \times 1 \frac{\text{year}}{\text{startup}} \times 0.14 \frac{\$}{\text{lb}} = \$36,109.8 \quad (\text{B-10})$$

To find the total installed cost, multiply the total purchase cost by a bare module costing factor taken from Seider (2005)

$$C_p(\text{Propeller}) + C_p(\text{tank}) + C_p(\text{FeCl}_3) = C_p = \$43,107.7 \quad (\text{B-11})$$

$$C_{BM}(\text{Propeller}) = C_p \times F_{BM} = \$40,908.8 \times 3.4 = \$146,566 \quad (\text{B-12})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$194,933 \quad (\text{B-13})$$

Flocculation Unit (V-103)

The cost of the Flocculation unit will include the vessel itself and the mixing propeller. The equation for an open vessel agitator is

$$C_p(\text{Propeller}) = 1,810S^{0.17} \quad (\text{B-14})$$

where again S is the size factor with a range of 1-8 Hp. Since our power requirement is less than one Hp, S will be 1.

$$C_p(\text{Propeller}) = 1,810 \times (1)^{0.17} = \$1,810 \quad (\text{B-15})$$

For the slow mix vessel the following equation for a concrete clarifier is used

$$C_p (\text{Flocculation}) = 1,700A^{0.58} \quad (\text{B-16})$$

where A is the base area in ft². The flocculation vessel has a base area of about 835.6 ft².

$$C_p (\text{Flocculation}) = 1700 \times 835.6^{0.58} = \$84,179.82 \quad (\text{B-17})$$

The equation above used to calculate the cost of the flocculation unit is only valid is the area is about 8,000 ft². However, it is assumed here that this gives a good cost estimate.

To find the total installed cost, the total purchase cost is multiplied by a bare module costing factor taken from Seider (2005):

$$C_p (\text{Propeller}) + C_p (\text{Flocculation}) = C_p = \$85,989.8 \quad (\text{B-18})$$

$$C_{BM} (\text{Propeller}) = C_p \times F_{BM} = \$85,989.8 \times 2.45 = \$210,675 \quad (\text{B-19})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$280,198 \quad (\text{B-20})$$

Sedimentation Basin/Clarifier (V-104 A/B/C/D)

The cost for a clarifier made out of concrete is given by

$$C_p (\text{Clarifier}) = 1,700A^{0.58} \quad (\text{B-21})$$

where A is the settling area in ft². Each clarifier has a settling area of about 2,827 ft².

$$C_p (\text{Clarifier}) = 1,700 \times 2,827^{0.58} = \$240,979.6 \quad (\text{B-22})$$

Therefore, the total cost for all four clarifiers is

$$C_p = 4 \times C_p (\text{Clarifier}) = \$682,775 \quad (\text{B-23})$$

To find the total installed cost, multiply to total purchase cost by a bare module costing factor taken from Seider (2005):

$$C_{BM} (\text{Clarifier}) = C_p \times F_{BM} = \$682,775 \times 2.45 = \$1,672,799.8 \quad (\text{B-24})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$2,224,823.7 \quad (\text{B-25})$$

Waste Storage Tank (V-105)

For a stainless steel waste storage tank the cost is

$$C_p (\text{Waste Tank}) = 210 \times 500,000^{0.51} \times F_M = 2 \times \$169,314 = \$338,628 \quad (\text{B-26})$$

$$C_{BM} (\text{Waste Tank}) = C_p \times F_M \times F_{BM} = \$338,628 \times 2.45 = \$829,638 \quad (\text{B-27})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$1,103,418.5 \quad (\text{B-28})$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$1,103,418.5 \quad (\text{B-29})$$

Filter (T-101A/B)

The cost of the filter will include the module and the initial cost for internals. The internals of the filter are anthracite coal, silica sand, and support gravel. The stainless steel filter module cost is based on its area (A). Since there are two modules, the final C_p should be doubled:

$$C_p (Filter) = 3,800 \times A^{0.52} \times F_M + V \times \frac{30}{ft^3 - Anthracite} + V \times \frac{38}{ft^3 - Silica Sand} + V \times \frac{35}{ft^3 - Gravel} \quad (B-30)$$

$$A = \frac{2083.33 \frac{gpm}{ft^2}}{3.46 \frac{gpm}{ft^2}} = 602.12 ft^2 \quad (B-31)$$

$$Anthracite = \frac{20 \frac{in}}{12 \frac{in}}}{ft} \times A = 1003.53 ft^2 \quad (B-32)$$

$$Silica Sand = \frac{8 \frac{in}}{12 \frac{in}}}{ft} \times A = 401.41 ft^2 \quad (B-33)$$

$$Support Gravel = \frac{18 \frac{in}}{12 \frac{in}}}{ft} \times A = 903.18 ft^2 \quad (B-34)$$

$$C_p (Filter) = (3,800 \times 602.12^{0.52} \times 2 + 1003.53 \times 30 + 401.41 \times 38 + 903.18 \times 35) \times 2 \text{ modules} \\ = \$577,857 \quad (B-35)$$

$$C_{BM} (Filter) = C_p \times F_{BM} = \$577,857 \times 3.05 = \$1,762,462.5 \quad (B-36)$$

To bring the costs up to their current value, multiply by 1.33, which is the ratio of the present cost index to the CE 394 index.

$$C_{BM} \times 1.33 = \$2,344,075 \quad (B-37)$$

Pumps:

The purchase cost of a pump depends on parameters that include pressure drop and flowrate. Pressure drop was an assumed quantity based on heuristic 38 in Seider et al (Warren, 2005). It was assumed for pump 101A/B that it would be providing power for 200 ft of piping, and would pump the water across one control valve, for a total pressure increase of 14 psi. Then, the theoretical horsepower (THp) is calculated using heuristic 39 by the equation

$$\text{THp} = (\text{gpm})(\text{Pressure increase, psi})/1714 . \quad (\text{B-38})$$

For pump 101A/B, the flowrate is 2080 gpm, for a THp of 17.0 hp.

Then, the size factor for a pump is given by equation 16.13 in Seider et al (Warren, 2005):

$$S = Q(H)^{0.5} \quad (\text{B-39})$$

where S is the size factor, Q is the flowrate in gallons per minute, and H is the pump head in feet. For this pressure drop and flowrate with water as the fluid, the size factor is 11,804. The base cost (C_B) for the pump is based on the size factor, from equation 16.14:

$$C_B = \exp(9.2951 - 0.6019[\ln(S)] + 0.0519[\ln(S)]^2), \quad (\text{B-40})$$

which gives a base cost of \$3.693 for the size factor.

Then, the final purchase cost, C_P , require two factors, one for pump type and one for materials. For a stainless steel pump, the material factor, F_M , is 2.00. For a flowrate of 2080 gpm, the necessary centrifugal pump has a factor, F_T , of 2.00. Then, C_P is given by

$$C_P = F_T F_M C_B, \quad (\text{B-41})$$

which is \$14,772 for this pump. This number is then multiplied by 2, since a spare pump is being purchased, by the bare module factor, F_{BM} , and by a conversion from 2000 dollars to 2008 dollars (Lozowski, 2008):

$$C_{BM} = 2F_{BM} C_P \frac{I}{I_{base}} \quad (\text{B-42})$$

where I and I_{base} are cost index values for the two years. For 2000 and 2008,

$$\frac{I}{I_{base}} \approx 1.33. \quad (B-43)$$

The bare module factor for the pump is 3.4, from Table 16.14 in Seider (Warren, 2005). Putting these together results in a bare module cost of \$133,831 for pump 101A/B.

Corresponding values for all pumps are summarized in Table 3 below.

Table 3: Costing parameters and costs for pumps

Pump	Flowrate (gpm)	Pressure increase (psi)	Theoretical Horsepower	Size factor	C _B	C _P	C _{BM}
P-101A/B	2080	14	17.0	11804	\$3,693	\$14,773	\$133,831
P-102A/B	2080	12	14.6	10928	\$3,590	\$14,361	\$130,095
P-103A/B	2080	26.588	32.3	17757	\$4,335	\$17,341	\$157,097
P-104A/B	1204	26.588	18.7	9418	\$3,405	\$13,621	\$123,393

Pressure increase for pump 102A/B includes 1 control valve and 100 ft of pipe.

Pumps 103A/B and 104A/B both include 2 control valves, 100 ft of pipe, and the filter bed.

Each pump also needs a motor, the cost of which is based on the power required by the pump, given by equation 16.16 in Seider:

$$P_C = \frac{THp}{\eta_P \eta_M} \quad (B-44)$$

where P_C is the power consumption and η_P and η_M are efficiency factors. The efficiency from the volumetric flowrate, η_P , is based on Q , the volumetric flowrate in gallons per minute, and is given by equation 16.17:

$$\eta_P = -0.316 + 0.24015 (\ln Q) - 0.01199 (\ln Q)^2, \quad (B-45)$$

which is 0.82 for 2080 gpm. The other efficiency, η_M , is a function of pump brake horsepower, and is given by equation 16.18:

$$\eta_M = 0.80 + 0.0319 (\ln P_B) - 0.00182 (\ln P_B)^2, \quad (\text{B-46})$$

where P_B is the brake horsepower, given by THp/η_P . For pump 101A/B, η_M is 0.88. Combining these factors gives the result $P_C = 23.5$ hp.

The base motor cost is a function of P_C , given by equation 16.19:

$$C_B = \exp(5.4866 + 0.13141[\ln P_C] + 0.053255[\ln P_C]^2 + 0.028628[\ln P_C]^3 - 0.0035549[\ln P_C]^4) \quad (\text{B-47})$$

which results in a base cost of \$1,078. To get a final cost, the base cost is multiplied by 2 (for two pumps); a type factor, F_T , of 0.9, corresponding to a 1,800 rpm motor and open, drip-proof enclosure; the bare module factor, F_{BM} , of 3.4; and the cost index adjustment of 1.33:

$$C_{BM} = 2F_T F_{BM} C_B \frac{I}{I_{base}}. \quad (\text{B-48})$$

The resulting price for the motors on pump 101A/B is \$8,788.

Table 4 below summarizes the above pump motor parameters and costs for the other three pumps.

Table 4: Costs and cost parameters for pump motors

Pump	η_P	η_M	P_C (hp)	C_B	C_{BM}
Pump 101A/B	0.82	0.88	23.5	\$1,940	\$8,788
Pump 102A/B	0.82	0.88	20.1	\$1,696	\$7,683
Pump 103A/B	0.82	0.89	48.1	\$3,818	\$17,295
Pump 104A/B	0.79	0.88	26.9	\$2,189	\$9,915

C. Detailed Economic Calculations

Estimation factors for the information contained in the following charts is taken from Seider, et al (Seider, 2005), unless otherwise noted. Estimated land cost is taken from Loopnet, property ID 15478980 (Loopnet.com, 2008).

Table 5: Equipment Costing Table

Equipment	Size factor		C _B	Factors	C _P	F _{BM}	C _{BM}	Dollar conversion	Final Cost
P-101A/B	11804		\$3,693	8	\$29,546	3.4	\$100,456	1.332	\$133,831
P-102A/B	10928		\$3,590	8	\$28,721	3.4	\$97,652	1.332	\$130,095
P-103A/B	17757		\$4,335	8	\$34,682	3.4	\$117,920	1.332	\$157,097
P-104A/B	9418		\$3,405	8	\$27,242	3.4	\$92,621	1.332	\$123,393
P-101A/B motor	23.5	hP	\$1,078	1.8	\$1,940	3.4	\$6,596	1.332	\$8,788
P-102A/B motor	20.1	hP	\$942	1.8	\$1,696	3.4	\$5,767	1.332	\$7,683
P-103A/B motor	48.1	hP	\$2,121	1.8	\$3,818	3.4	\$12,982	1.332	\$17,295
P-104A/B motor	26.9	hP	\$1,216	1.8	\$2,189	3.4	\$7,442	1.332	\$9,915
V-101	23.0	gal							Too small to cost
V-102	100.0	gal			\$43,108	3.4	\$146,566	1.332	\$184,909
V-103	835.6	ft ²			\$85,990	2.45	\$210,675	1.332	\$280,198
V-104A/B/C/D	2827	ft ²		4	\$682,775	2.45	\$1,672,800	1.332	\$2,228,560
V-105	500000	gal			\$338,629	2.45	\$829,641	1.332	\$1,105,275
R-101	100	gal			\$3,293	3.05	\$10,042	1.332	\$13,379
T-101A/B	602	ft ²		2	\$577,857	3.05	\$1,762,464	1.332	\$2,348,013
R-101 Propeller	1	hP			\$2,600	3.3	\$8,580	1.332	\$11,431
									\$6,759,861

Table 6: Total Capital Investment

Step in Plant Construction			
Spares	\$675,986		
C _{PM}	\$6,759,861		
C _{TBM}		\$7,435,847	
Cost of Site Preparation		\$743,585	
Total direct permanent investment			\$8,179,432
Cost of Contingency and Contractor's Fee			\$1,226,915
Total Depreciable Capital, C _{TDC}			\$9,406,347
Cost of Land (2 acres)			\$63,400
Cost of Royalties			\$188,127
Cost of Startup			\$940,635
Total Permanent Investment			\$10,598,508
Working Capital, C _{WC}			\$283,848
Capital Investment			\$10,882,357
West Coast Building Cost			\$2,720,589
Total Capital Investment, C _{TCI}			\$13,602,946

Table 7: Cost of Operation

Expense		
Utilities		\$345,841
Operations		
Direct Wages and Benefits (DW&B)		
Hours / Year Shift	2000	
Operators / Shift	1	
Shifts	5	
Direct Wages and Benefits (DW&B)		\$300,000
Direct Salaries and Benefits		\$45,000
Operating Supplies and Services		\$2,700
Technical Assistance		\$52,000
Control Laboratory		\$57,000
Maintenance		
Wages and Benefits		
Fluid Handling		\$329,222
Solid-Fluid Handling		\$423,286
Saleries and benefits		\$75,000
Materials and Services		\$300,000
Maintenance Overhead		\$15,000
Operating Overhead		
General Plant Overhead		\$90,987
Mechanical Department Services		\$30,756
Employee Relations Department		\$1,282
Business Services		\$12,815
Property Tax and insurance		\$470,317
Cost of Operation		\$2,551,206
General Expenses		
Selling Expense		\$10,000
Administrative Expense		\$20,000
Management Insentive		\$12,500
Direct Research		\$10,000
Total Operational Cost		\$2,603,706

Table 8. Cash Flow Worksheet

Year	Fraction of Total Capital Investment	Cost of Working Capital	Depreciation	Cost Excluding Depreciation	Sales	Net Earnings	Cash Flow	Cumulative Present Value
2010	-\$6,801,473	-\$283,848					-\$7,085,321	-\$6,801,473
2011	-\$6,801,473		-\$340,074	\$2,603,706		-\$2,943,780	-\$9,405,179	-\$16,206,652
2012			-\$646,140	\$2,603,706	\$1,000,000	-\$2,249,846	-\$1,603,706	-\$17,664,566
2013			-\$581,526	\$2,603,706	\$1,000,000	-\$2,185,232	-\$1,603,706	-\$18,989,943
2014			-\$523,713	\$2,603,706	\$1,000,000	-\$2,127,419	-\$1,603,706	-\$20,194,831
2015			-\$471,342	\$2,603,706	\$1,000,000	-\$2,075,048	-\$1,603,706	-\$21,290,184
2016			-\$423,732	\$2,603,706	\$1,000,000	-\$2,027,438	-\$1,603,706	-\$22,285,959
2017			-\$401,287	\$2,603,706	\$1,000,000	-\$2,004,993	-\$1,603,706	-\$23,191,209
2018			-\$401,287	\$2,603,706	\$1,000,000	-\$2,004,993	-\$1,603,706	-\$24,014,164
2019			-\$401,967	\$2,603,706	\$1,000,000	-\$2,005,673	-\$1,603,706	-\$24,762,304
2020			-\$401,287	\$2,603,706	\$1,000,000	-\$2,004,993	-\$1,603,706	-\$25,442,432
2021			-\$401,967	\$2,603,706	\$1,000,000	-\$2,005,673	-\$1,603,706	-\$26,060,730
2022			-\$401,287	\$2,603,706	\$1,000,000	-\$2,004,993	-\$1,603,706	-\$26,622,820
2023			-\$401,967	\$2,603,706	\$1,000,000	-\$2,005,673	-\$1,603,706	-\$27,133,810
2024			-\$401,287	\$2,603,706	\$1,000,000	-\$2,004,993	-\$1,603,706	-\$27,598,346
2025			-\$401,967	\$2,603,706	\$1,000,000	-\$2,005,673	-\$1,603,706	-\$28,020,652
2026			-\$200,643	\$2,603,706	\$1,000,000	-\$1,804,349	-\$1,603,706	-\$28,404,567
2027			\$0	\$2,603,706	\$1,000,000	-\$1,603,706	-\$1,603,706	-\$28,753,580
2028			\$0	\$2,603,706	\$1,000,000	-\$1,603,706	-\$1,603,706	-\$29,070,864
2029			\$0	\$2,603,706	\$1,000,000	-\$1,603,706	-\$1,603,706	-\$29,359,305
2030		\$283,848	\$0	\$2,603,706	\$1,000,000	-\$1,603,706	-\$1,319,858	-\$29,575,112

Table 9: Cash Flow Worksheet with \$4.5 Million Fine

Year	Fraction of Total Capital Investment	Cost of Working Capital	Depreciation	Cost Excluding Depreciation	Sales	Net Earnings	Cash Flow	Cumulative Present Value at 10% Interest
2010	-\$6,801,473	-\$283,848					-\$7,085,321	-\$6,801,473
2011	-\$6,801,473		-\$340,074	\$2,603,706		-\$2,943,780	-\$9,405,179	-\$16,206,652
2012			-\$646,140	\$2,603,706	\$4,535,612	\$1,285,766	\$1,931,906	-\$14,450,373
2013			-\$581,526	\$2,603,706	\$4,535,612	\$1,350,380	\$1,931,906	-\$12,853,757
2014			-\$523,713	\$2,603,706	\$4,535,612	\$1,408,193	\$1,931,906	-\$11,402,287
2015			-\$471,342	\$2,603,706	\$4,535,612	\$1,460,564	\$1,931,906	-\$10,082,769
2016			-\$423,732	\$2,603,706	\$4,535,612	\$1,508,174	\$1,931,906	-\$8,883,208
2017			-\$401,287	\$2,603,706	\$4,535,612	\$1,530,619	\$1,931,906	-\$7,792,697
2018			-\$401,287	\$2,603,706	\$4,535,612	\$1,530,619	\$1,931,906	-\$6,801,324
2019			-\$401,967	\$2,603,706	\$4,535,612	\$1,529,939	\$1,931,906	-\$5,900,075
2020			-\$401,287	\$2,603,706	\$4,535,612	\$1,530,619	\$1,931,906	-\$5,080,758
2021			-\$401,967	\$2,603,706	\$4,535,612	\$1,529,939	\$1,931,906	-\$4,335,925
2022			-\$401,287	\$2,603,706	\$4,535,612	\$1,530,619	\$1,931,906	-\$3,658,804
2023			-\$401,967	\$2,603,706	\$4,535,612	\$1,529,939	\$1,931,906	-\$3,043,239
2024			-\$401,287	\$2,603,706	\$4,535,612	\$1,530,619	\$1,931,906	-\$2,483,634
2025			-\$401,967	\$2,603,706	\$4,535,612	\$1,529,939	\$1,931,906	-\$1,974,903
2026			-\$200,643	\$2,603,706	\$4,535,612	\$1,731,263	\$1,931,906	-\$1,512,420
2027			\$0	\$2,603,706	\$4,535,612	\$1,931,906	\$1,931,906	-\$1,091,981
2028			\$0	\$2,603,706	\$4,535,612	\$1,931,906	\$1,931,906	-\$709,764
2029			\$0	\$2,603,706	\$4,535,612	\$1,931,906	\$1,931,906	-\$362,294
2030		\$283,848	\$0	\$2,603,706	\$4,535,612	\$1,931,906	\$2,215,755	\$0

13.4 Supporting documentation

Problem statement

Design a 3 mgd wastewater treatment facility for a private company to treat cooling waters used in their power plant. The wastewater treatment plant could be favorably sited on unlimited CI-2 zoned privately owned real estate adjacent to the power plant abutting a major river in Redding, CA on the Sacramento River at competitively negotiated lease rates. The power plant, located on a CI-3 zoned twenty acre parcel, burns natural gas in a combustion turbine producing 420 MW on average year round.

County owned sewer plant reclaimed waters are fed to the power plant at 70 °F. All initial effluent fed to the sewer plant are transported via area groundwater sources. (Assume sewer plant is 10 mgd facility.) Unfortunately, Sacramento River groundwaters are laden with silver, mercury and arsenic. Previous power plant owners had a ten year agreement allowing discharge waters being returned to the sewer plant at 90 °F in more concentrated heavy metal levels. The expiration of the agreement and county and state authority's refusal to continue allowance of concentrated levels of heavy metals has led to an Administrative Consent Order requiring the power plant to clean up its discharge waters to meet state and federal discharge to surface water (DSW) levels before December 31, 2011.

Incoming denitrified waters to the power plant contain 0.1 ppm silver, 100 ppt mercury and 0.1 ppm arsenic. Each metal cycles up in the power plant's water cooling system. Ag cycles up to 0.5 ppm; Hg cycles up to 200 ppt; and As cycles up to 0.5 ppm in the cooling tower.

Blowdown from the cooling tower is 1 mgd, therefore, 2 mgd is lost to ambient air flow.

At the power plant, 96% sulfuric acid is fed to the cooling tower at 0.1 gpm to control pH. Unfortunately the sulfuric acid contains 0.7 ppm mercury. Present sources of sulfuric acid, based on transportation costs, are from northern California or Canada, none of which is mercury free.

State and federal environmental DSW regulations require waters containing silver being reduced to 100 ppb, mercury being reduced to 10 ppt and arsenic being reduced to 100 ppb.

The waste water treatment plant can be designed with mechanical, biological or chemical separation techniques.

Operating costs of the designed wastewater treatment plant cannot exceed \$200 K annually. Construction costs will be part of the decision to build the wastewater treatment plant or continue doing business under an ACO with litigation and potential fines exceeding \$1 million annually.

Presently, the power plant discharges waters to the local sewer plant since they maintain the NPDES permit allowing adjacent river (major ocean tributary) discharge. It may be worthwhile for the power plant to acquire an NPDES permit, and discharge directly into the same river located one half mile to the northwest.

Determine: (1) to what extent it is in the private owner's interest to acquire an NPDES permit; (2) if supplemental or replacement power plant cooling tower pretreatment is favorable; (3) whether private donation of secondary treatment for county owned sewer plant including engineering design, construction and startup is favorable; (4) appropriate ROIs; and (5) optimally design a wastewater treatment plant system for the private owners.