NAS7.010277
NASA-JPL
SSIC No. 9661

FIELD PILOT TESTING OF A DYNAMIC SUSPENDED BED REACTOR FOR REMOVAL OF PERCHLORATE IN GROUNDWATER AT JPL

Prepared for:

Naval Facilities Engineering Service Center

Prepared by:

Foster Wheeler Environmental Corporation

FIELD PILOT TESTING OF A DYNAMIC SUSPENDED BED REACTOR FOR REMOVAL OF PERCHLORATE IN GROUNDWATER AT JPL

Prepared for:

NAVAL FACILITIES ENGINEERING SERVICE CENTER
1100 23rd Avenue
Port Hueneme, CA 93043

Prepared under Contract:
N47408-01-C-7202

Prepared by:

FOSTER WHEELER ENVIRONMENTAL CORPORATION
1940 E. Deere Ave., Suite 200
Santa Ana, CA 92705

June 30, 2003

TABLE OF CONTENTS

PAGE

LIST OF TABLES iii

LIST OF FIGURES iv

ABBREVIATIONS AND ACRONYMNS v

1.0 INTRODUCTION 1-1
1.1 SCOPE OF REPORT 1-2
1.2 ORGANIZATION OF REPORT 1-3

2.0 BACKGROUND 2-1
2.1 PREVIOUS BENCH TESTING 2-1
2.2 INITIAL FIELD TESTING (PHASE I) AT JPL 2-1
2.2.1 Startup 2-2
2.2.2 Operation 2-3
2.2.3 Summary of Results and Conclusions 2-4

3.0 PHASE II (DSBR) TESTING 3-1
3.1 OBJECTIVE/DESIGN 3-1
3.1.1 DSBRs Versus other Bioreactors 3-2
3.2 MICROBIOLOGY/STARTUP 3-3
3.3 OPERATION 3-4
3.3.1 Test 1 3-4
3.3.2 Re-circulation Period 3-6
3.3.3 System and Experimental Modifications 3-6
3.3.4 Test 2 3-7
3.3.5 Comparison with Objectives/Deviations from Scope, Phase II 3-10
3.3.6 Additional Substrate Utilization Testing 3-11
3.4 SUMMARY OF RESULTS AND CONCLUSIONS, PHASE II 3-15
3.4.1 Startup 3-15
3.4.2 Forward Flow Testing 3-16

4.0 SUMMARY OF OBJECTIVES/CONCLUSIONS, PHASE I AND II TESTING 4-1

5.0 SCALE-UP COSTING ESTIMATES 5-1
5.1 DSBR SCALE-UP PARAMETERS 5-1
5.2 DSBR COSTS – NON-POTABLE WATER APPLICATIONS 5-2
5.2.1 DSBR Capital Costs 5-2
5.2.2 DSBR O&M Costs 5-2
5.2.3 DSBR Cost Estimate Summary (Non-potable Applications) 5-3
5.3 DSBR COSTS – POTABLE WATER APPLICATIONS 5-4
5.4 LIFE CYCLE COST - DSBR 5-5
5.5 DSBR FOOTPRINT 5-5

TABLE OF CONTENTS
(Continued)

PAGE

5.6 COMPARISON WITH OTHER CLO_4^- TREATMENT SYSTEMS 5-6
5.6.1 Other Biological Systems 5-6
5.6.2 IE Systems 5-8
5.6.3 NF Systems 5-9
5.6.4 Cost Comparison Summary 5-10

APPENDICES

Appendix A Phase I Field Notes

Appendix B Phase I Parameters and ClO_4^- Probe Calibration Curves

Appendix C Phase I Analytical Reports and Chain-of-Custody Records for Groundwater Samples

Appendix D Phase II Field Notes

Appendix E Phase II Forward Flow Parameters

Appendix F Phase II Analytical Results Summary and Analytical Reports and Chain-of-Custody Records for Groundwater Samples

Appendix G Results and Chain-of-Custody Forms for Substrate Utilization Study and Theoretical Acetate Requirement Calculations

Appendix H Life Cycle Cost Estimates for DSBR, CSTR, FBR, IE, and NF Systems

LIST OF TABLES

Table 1 Phase I Startup: ClO4- Probe Calibration Standards and System Readings

Table 2 Phase I Forward Flow Test Analytical Results

Table 3 Phase II Startup Parameters

Table 4 Summary of Key Operational Parameters

Table 5 Summary of Key Results

LIST OF FIGURES

Figure 1 Perchlorate Bioreactor Process Flow Diagram (Phase I)

Figure 2 Phase I Start-Up Results

Figure 3 Perchlorate Bioreactor Process Flow Diagram (Phase II)

Figure 4 Phase II Dissolved Oxygen Readings

Figure 5 pH and Temperature Tracking – Phase II/Test 1

Figure 6 ClO_4^-, NO_3^- and Acetate Removal – Phase II/Test 1

Figure 7 pH and Temperature Tracking –Phase II/Test 2

Figure 8 ClO_4^-, NO_3^- and Acetate Removal – Phase II/Test 2

Figure 9 Test 2 - Influent [ClO_4^-] Versus Percent Removal per Reactor

Figure 10 Test 2 - Pressure Drops through Bioreactors

Figure 11 October 18, 2002, Sampling: Water Quality Parameters Analytical Results

Figure 12 ClO_4^-, NO_3^- and DO Removal Tracking

Figure 13 Use of Acetate by the JPL Isolates and Perc1ace and Disappearance of Available Electron Acceptors.

Figure 14 Use of Acetate and Methanol at Various Ratios by the JPL Isolates and Disappearance of Available Electron Acceptors

Figure 15 500-gpm DSBR Process Flow Diagram

Figure 16 500-gpm DSBR – Equipment Layout

ABBREVIATIONS AND ACRONYMNS

BOD	biochemical oxygen demand
^{o}C	degrees Celsius
µg/L	micrograms per liter
µL	microliter
$CaCO_3$	calcium carbonate
CEM	Center for Environmental Microbiology
Cl^-	chloride
ClO_4^-	perchlorate
CSTR	continuously stirred tank reactor
DHS	Department of Health Services
DSBR	dynamic suspended bed reactor
DO	dissolved oxygen
EPA	U. S. Environmental Protection Agency
FBR	fluidized bed reactor
$FeCl_3$	ferric chloride
ft/min	feet per minute
FWENC	Foster Wheeler Environmental Corporation
g/L	grams per liter
gpm	gallons per minute
gpm/ft^2	gallons per minute per square foot
hp	horsepower
HPO_4-P	hydrogen phosphate
IE	ion exchange
JPL	Jet Propulsion Laboratory
kgal	one thousand gallons
K_2HPO_4	di-potassium hydrogen phosphate
KH_2PO_4	potassium di-hydrogen phosphate
kW	kilowatts
lb	pounds
lb/d	pounds per day

ABBREVIATIONS AND ACRONYMS
(Continued)

lb/ft^3	pounds per cubic foot
L	liter
LPGAC	liquid phase granular activated carbon
mA	milliamps
mg	milligrams
mg/L	milligrams per liter
$MgSO_4$	magnesium sulfate
mL	milliliter
mL/g	milliliters per gram
mL/min	milliliters per minute
mM	millimolar
m^2/g	square meters per gram
MSM	mineral salts medium
mV	millivolts
N/A	not applicable
$NaClO_4$	sodium perchlorate
NaOH	sodium hydroxide
ND	not detected
NF	nanofiltration
NH_4-N	ammonium nitrogen
$(NH_4)_2SO_4$	diammonium sulfate
NO_3^-	nitrate
NPDES	National Pollutant Discharge Elimination System
OD	optical density
O&M	operation and maintenance
O_2	oxygen
PBR	packed bed reactor
ppm	parts per million
psi	pounds per square inch
RO	reverse osmosis

ABBREVIATIONS AND ACRONYMS
(Continued)

RWQCB	Regional Water Quality Control Board
SD	standard deviation
SERDP	Strategic Environmental Research and Development Program
TDS	total dissolved solids
TOC	total organic carbon
UV	ultraviolet
VOC	volatile organic compound

1.0 INTRODUCTION

A number of sites in the United States have groundwater impacted with perchlorate (ClO_4^-) at concentrations ranging from low microgram per liter (µg/L) levels to tens or hundreds of milligrams per liter (mg/L) and higher. Several techniques have been used for removal of ClO_4^- from groundwater, including physicochemical and biological methods. Conventional physicochemical technologies such as ion exchange (IE) and reverse osmosis (RO) are relatively expensive, particularly for high concentration applications, due to operation and maintenance (O&M) costs for regenerating resins for IE or for disposal of rejectates for RO. The higher costs are compounded by the fact that both IE and RO do not destroy ClO_4^-, rather, they concentrate it into waste streams that must be addressed via secondary treatment (destruction) or are disposed of by other means. Biological treatment, on the other hand, completely destroys ClO_4^-, converting it to chloride (Cl^-) ions, and oxygen (O_2), which are innocuous.

Several bioreactor configurations have been developed at various scales to treat ClO_4^--impacted water, including fluidized bed reactors (FBRs), continuously stirred tank reactors (CSTRs), and packed bed reactors (PBRs). Although application of bioreactors for ClO_4^- treatment is relatively recent, several systems currently are in use. These include CSTRs and FBRs. The CSTR has been used to treat high levels (tens or hundreds of mg/L) of ClO_4^- in relatively concentrated waste streams at low flow rates. FBR technology thus far has proven to be effective for treating ClO_4^--contaminated groundwater. However, FBRs are considered to be more difficult and costly to operate than CSTRs or PBRs, and it has yet to be proven that FBR technology is the most favorable biological approach. On the other hand, PBRs are simple to operate and are potentially more efficient than FBRs in removing ClO_4^- from water. However, PBRs may be prone to clogging, which may diminish their effectiveness.

In the spring of 2001, a PBR was field tested by Foster Wheeler Environmental Corporation (FWENC) at the Jet Propulsion Laboratory (JPL) and successfully reduced ClO_4^- concentrations in groundwater from 0.42 mg/L to less than 4 µg/L[1] at a flow rate of 2 gallons per minute (gpm). The specifics of the field PBR testing (Phase I) are discussed in Section 2.0. Based upon performance of the PBR and results of testing, FWENC developed and tested a dynamic suspended bed reactor (DSBR) at JPL in the summer and fall of 2002. The DSBR was designed to provide improved flow characteristics [by using different media (for bacterial attachment) with less surface area and greater pore volume and allowing for limited bed flux and expansion in the reactors] than those of the PBR tested previously, while retaining the process simplicity and low operating costs of the PBR. The media used in the Phase I testing (Celite R-633) were a diatomaceous earth product with sand-like consistency (30/50 mesh) and a miniscule pore

[1] 4 µg/L was the reporting limit for EPA Method 314 for the analytical laboratory for this testing (MWH Laboratories, Pasadena, California).

volume of 1.5 milliliters per gram (mL/g) and very high surface area [1.3 square meters per gram (m^2/g)]. Based on this pore volume and a bed density of 24 pounds per cubic foot (lb/ft^3), and assuming the R-633 completely filled the PBR vessels used in the Phase I test, the void volume was only about 56 percent for these reactors. In contrast, the media proposed for one treatment train for the DSBR (i.e. Phase II) testing were Hydroxyl-PAC® high-density polyethylene cylinders, which were about ¾-inch diameter and ½-inch high and had a pore volume of about 90 percent based on manufacturer's estimates. In addition, these media would not be filled to the top of the DSBR reactor; instead, they would be added until the reactor was about two-thirds full to allow for media flux. Section 3.0 presents further details on the media specified for the Phase II (DSBR) testing.

In the Phase II testing, it was realized that influent concentrations of ClO_4^- had increased by an order of magnitude (from about 0.4 mg/L during Phase I to more than 6 mg/L at the beginning of Phase II), as a result of which flow rates were reduced to 1 gpm (per treatment train) during the test. The DSBR was capable of reducing ClO_4^- from more than 6 mg/Lto less than 4 μg/L[2].

1.1 SCOPE OF REPORT

This report summarizes results of:

- Preliminary bench testing, performed in 1999, of a PBR for reducing ClO_4^- in JPL groundwater, performed at flow rates ranging from 10 to 100 milliliters per minute (mL/min)
- Phase I field testing of a PBR at JPL conducted in the spring of 2001, performed using three PBRs in parallel, each with a different combination of inocula/substrate, at a total flow rate of 6 gpm
- Phase II field testing of a DSBR at JPL conducted from August to October 2002, performed using two parallel trains of DSBRs, each with a different packing material, at flow rates ranging from 1 to 3 gpm per train

The report also presents life cycle cost estimates for 500-, 1,000- and 2,000-gpm DSBRs for treatment of groundwater with similar characteristics to groundwater treated in the Phase II testing at JPL [i.e., 10 mg/L of influent ClO_4^-, 50 mg/L of influent nitrate (NO_3^-)]. For comparison, it also presents life cycle costs for 500-, 1,000-, and 2,000-gpm CSTRs, FBRs, nanofiltration (NF), and IE systems. It should be noted that the preliminary bench testing noted above is not part of this contract, but was done as part of the Feasibility Study for JPL. Results are included here, as they form the basis for the Phase I testing.

---

[2] 4 μg/L was MWH Laboratories' reporting limit for the Phase II ClO_4^- analyses.

1.2 ORGANIZATION OF REPORT

The remainder of this report is organized as follows:

- **Section 2.0 – Background:** presents the test parameters/objectives and summarizes the results and conclusions of the PBR bench test and Phase I PBR pilot test (startup and forward flow) at JPL.
- **Section 3.0 – Phase II (DSBR) Testing:** presents the objectives and design parameters for the tests and summarizes the results of the startup period, discusses system modifications implemented during the test, and presents results and conclusions of the forward flow testing. Also presents the results of substrate utilization studies performed following the Phase II testing for use in preparing DSBR scale-up costs.
- **Section 4.0 – Summary of Objectives/Conclusions, Phase I and II Testing:** summarizes the objectives, results (comparison with objectives), and conclusions/ recommendations of the Phase I/II testing at JPL.
- **Section 5.0 – Scale-up Costing Estimates:** presents scale-up parameters and capital, O&M and life cycle cost estimates for 500-, 1,000-, and 2000-gpm DSBRs, CSTRs, FBRs, NF, and IE systems for ClO_4^- removal in non-potable and potable applications.

2.0 BACKGROUND

2.1 PREVIOUS BENCH TESTING

In 1999, FWENC developed, designed, and constructed a PBR and conducted limited bench-scale testing to demonstrate process feasibility for treatment of JPL groundwater and to estimate initial scale-up parameters. This was done in support of the Feasibility Study being conducted by FWENC for JPL. Work was conducted in conjunction with Dr. W.T. Frankenberger, Jr., of the Center for Environmental Microbiology (CEM). Dr. Frankenberger is a professor of soil microbiology and biochemistry at the University of California at Riverside. A proprietary ClO_4^--reducing bacterium (perc1ace), previously isolated and characterized by Dr. Frankenberger's group, was used as the inoculum. Celite R-635 pellets, which are 0.25-inch diameter and 0.5-inch long) were used as the medium for bacterial attachment. Based upon Dr. Frankenberger's preliminary work, FWENC and CEM conducted bench-scale experiments at flow rates ranging from 10 to 100 mL/min (0.00264 to 0.0264 gpm), which corresponded to reactor residence times of 2.0 to 0.2 hours, respectively. Other initial operating parameters, including optimal carbon source (acetate) and concentration [0.5 grams per liter (g/L)], and optimal temperature [28 degrees Celsius (^oC)] also were based upon previous work conducted in Dr. Frankenberger's laboratory.

Data from these tests suggested that the PBR/perc1ace system was capable of reducing low concentrations of ClO_4^- (approximately 800 μg/L) in groundwater to non-detectable levels (less than 4 μg/L) with reactor residence times of approximately 0.3 hours. Data also indicated that influent acetate concentrations of less than 500 mg/L and potentially less than 250 mg/L yielded non-detectable effluent ClO_4^- levels. Moreover, low acetate concentrations (in many cases less than 50 mg/L) were present in the effluent. Other important findings were that NO_3^- also was removed in this system; pH remained relatively constant during the process; reduction of sulfate was not observed; and the need for addition of nitrogen and phosphorus as nutrients appeared to be minimal.

2.2 INITIAL FIELD TESTING (PHASE I) AT JPL

In March and April 2001, FWENC constructed a pilot scale (maximum 6 gpm – 3 trains of 2 gpm each) PBR system at JPL and conducted testing with the overall scope of demonstrating proof-of-concept for the PBR to remove ClO_4^- from JPL groundwater at a much larger scale. Within this overall scope, major objectives were to assess three sources of inoculum with respect to:

1) Startup efficiency (time to achieve functional biomass within the reactors)
2) Process efficiency (ability to achieve non-detectable effluent ClO_4^- levels)

Three test PBRs were set up in parallel (see Figure 1), each associated with a different ClO_4^--reducing inoculum/carbon source. The rationale for this approach was to identify whether proprietary microorganisms might be required for this system to operate optimally, which could potentially lead to increased expense. Inocula/carbon source combinations that were tested included perc1ace/acetate (Reactor 1 or R1), food waste/compost/ethanol (Reactor 2 or R2), and a combination of two ClO_4^--reducing enrichment cultures isolated from the JPL aquifer in previous experiments [conducted by Dr. Paul Hatzinger of Envirogen under Strategic Environmental Research and Development Program (SERDP) Project Number CU1163]/acetate (Reactor 3 or R3). The perc1ace and food waste bacteria were grown in mineral salts medium (MSM) at CEM in 10-liter batches using acetate in the presence of ClO_4^-.[3]

The system also included secondary treatments (downstream of the PBRs) consisting of an aerobic bioreactor (to remove residual organic carbon, which is added as a substrate), particulate filters (to remove sloughed biomass and suspended solids – to protect downstream processes), liquid phase granular activated carbon (LPGAC) [to remove volatile organic compounds (VOCs) that also are present in the influent], and IE (to remove residual ClO_4^-). Figure 1 shows these treatment processes. The secondary treatment processes were implemented to ensure that the final effluent met National Pollutant Discharge Elimination System (NPDES) permit requirements for these constituents prior to discharge. Originally, it was conceived that the effluent would be discharged to a storm drain system following verification of analytical results. However, in accordance with Regional Water Quality Control Board (RWQCB) direction, the effluent eventually was disposed off site, mainly due to high native sulfate and Cl^- levels.

Due to the experimental nature of the system, initial design and experimental parameters for the bioreactors were determined based upon previous bench-scale testing. The startup test was estimated to require approximately 2 weeks, and it was expected that the process efficiency test would run for 40 days. Field notes for the Phase I test are included in Appendix A.

2.2.1 Startup

Startup, which consisted of inoculating reactors with the respective bacteria and circulating a relatively rich basal salts medium through the reactors, occurred much more rapidly than expected. Based upon ClO_4^- disappearance data collected using a ClO_4^--specific probe (see Table 1, three right columns), ClO_4^--reducing populations were established in two of the three reactors (perc1ace and the JPL isolates) within just a few days. The third reactor (inoculated with food waste/compost) took slightly longer to colonize (approximately 1 week). Regression curves and "best fit" equations were generated for each day during startup based on readings for the ClO_4^- standards [1, 10, 25, 50, 75, and 100 parts per million (ppm) ClO_4^-]. The ClO_4^- probe readings, in millivolts (mV), varied slightly each day based upon temperature. These

[3] Acetate and ClO_4^- concentrations in MSM were 500 mg/L (as acetate ion) and 100 mg/L, respectively.

curves/equations are presented in Appendix B. The curves generally show a good correlation between the standard ClO_4^- concentrations and measured (actual) readings. Regression equations were used to convert probe readings in mV (see columns with headings "EFF R1," "EFF R2," and "EFF R3") into equivalent ClO_4^- concentrations shown in the three right columns in Table 1. Startup ClO_4^- concentrations are shown graphically in Figure 2. Note that these data represent only snapshots on each date and need to be interpreted in the context of what constituents (electron donor or ClO_4^-) were added to each reactor and when. A summary of details relevant to interpretation of the ClO_4^- probe readings is provided in Table 1.

Following the rapid startup period, the Navy was informed that the system was ready for the operation phase earlier than had been expected. The Navy then informed FWENC that there was a problem with the effluent discharge agreement and that further negotiations were required before discharge into the storm drain would be allowed and that discharge into the sewer was being explored. Therefore, the system remained in the re-circulation phase for several weeks. During this extended re-circulation period, substrate was fed into the reactors periodically, and disappearance of ClO_4^- was monitored (as shown in Table 1 and Figure 2) to maintain and verify the health of the bacterial cultures. The extended startup period contributed to proliferation of bacterial growth, and this, in conjunction with the small size of the particles comprising the packing medium (Celite R-633), resulted in increased pressure drops through the reactors, eventually causing one reactor (inoculated with food waste/compost and fed with ethanol) to plug. When plugging started to occur, the startup medium was diluted 1 to 3 in an effort to limit bacterial growth rates.

Throughout the startup period, efforts were made to adjust pH and temperature (i.e. maintain between about 7 and 8, and around 25°C, respectively). Without heating, water temperatures in the startup tanks generally were below the desired range (typically below 20°C). Heaters were more efficient than expected and generally were operated only while field personnel were on site. After about 1 week of startup re-circulation, the pH was high (above 8) in Reactors 1 and 3 (perc1ace and JPL isolates, respectively) and low in Reactor 2 (between 5 and 6) and was adjusted accordingly. Conductivity also was monitored during the startup period. Appendix B contains records of the pH, temperature and conductivity readings and pH/temperature adjustments. Appendix B also shows pressure drops for each of the reactors. Variation was between about 10 and 30 pounds per square inch (psi).

2.2.2 Operation

When forward flow was finally initiated for the remaining two reactors, R1 and R3 (24 days after startup commenced on March 26, 2001), the influent concentration of ClO_4^- was measured after 2 days of operation at 420 µg/L, and effluent ClO_4^- concentrations were non-detect (less than 4 µg/L), and 26 µg/L, respectively, for reactors inoculated with JPL isolates and perc1ace (at a

residence time of about 0.3 hours[4]). Over the next 6 days, further reactor plugging occurred. R1 was shut down after 4 days, and a substantial ClO_4^- breakthrough was observed (110 to 240 µg/L in effluent with influent ClO_4^- concentrations of approximately 420 µg/L). The ClO_4^- breakthrough likely was attributable to channeling and reduced residence time in the reactors due to plugging.

Table 2 summarizes the forward flow analytical results for ClO_4^- and the other constituents measured [acetate, total organic carbon (TOC), and NO_3^-] during the test. As shown in Table 2, NO_3^- concentrations increased on the first day of testing (likely due to sampling or laboratory error, as it could not be process-related) and decreased somewhat on the other dates. Although acetate and TOC concentrations were somewhat variable, influent versus effluent levels generally showed a slight decrease (expected, as substrate is consumed in the breakdown of ClO_4^-). A slight breakthrough of acetate/TOC also was observed. Appendix C contains copies of analytical reports and chain-of-custody records for groundwater samples collected during this test.

During the forward flow testing (April 18 through 26), pH and temperature were monitored continually and influent equalization tank water was heated occasionally to bring temperatures up to the desired range (influent groundwater temperature was around 20°C versus the desired 25°C).

2.2.3 Summary of Results and Conclusions

Results and conclusions from the Phase I field testing are as follows.

The three sources of inoculum tested, perc1ace, food waste/compost, and JPL enrichment cultures, all reduced ClO_4^- in the startup medium. The reactor inoculated with food waste/compost and fed ethanol experienced a proliferation of growth during startup (observed as biological material on the surface of the attachment medium), increasing the propensity for plugging, while ClO_4^--reducing ability actually was less efficient than that of the other two inocula. This may have been due to the presence of large numbers of non-ClO_4^--reducing organisms. Performance of the perc1ace culture and the JPL enrichment cultures was similar during startup/operation in terms of ClO_4^- reduction; however, plugging was less evident in the reactor containing JPL enrichment cultures.

---

[4] This was the target residence time, based on the bed volume with Celite R-633 and assuming a 2-gpm flow. Actual flow averaged about 0.5 gpm per reactor, based on flow totalizer readings. Residence time also was affected (reduced) by plugging.

Proof-of-concept was demonstrated for the PBR system at the field scale, as the reactor containing JPL isolates produced effluent with non-detectable ClO_4^- levels, despite significant overgrowth resulting from the extended startup period. In addition, the reactor containing perc1ace also demonstrated significant ClO_4^- reducing activity, reducing concentrations from 420 µg/L to 26 µg/L (96 percent removal). Successful operation of the PBR appeared to be related to reactor design (with packing material flow characteristics of primary importance), and the source of inoculum appeared to be of less importance than was originally hypothesized. Temperature and pH, which were monitored and adjusted frequently to keep within a relatively narrow range (especially during startup), did not appear to have a significant impact on performance of the bioreactors. Breakthrough was observed several days after forward flow testing began and was attributed to flow channeling due to media plugging.

3.0 PHASE II (DSBR) TESTING

3.1 OBJECTIVE/DESIGN

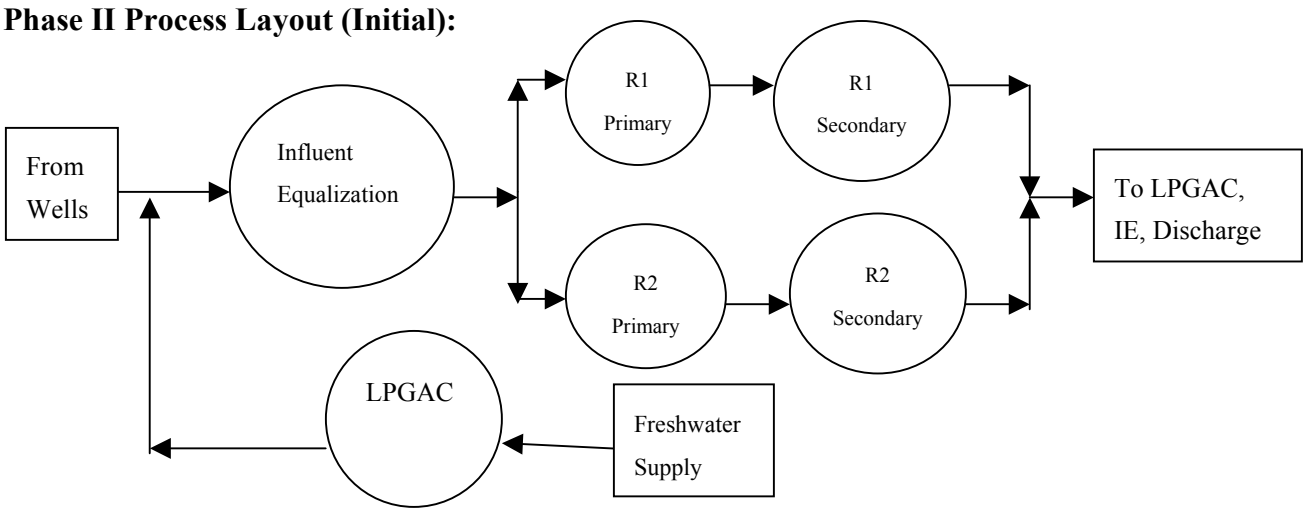
Based upon the results of Phase I testing, the following objectives were established for Phase II work:

1) Assess two different packing materials for their process efficiency in terms of ClO_4^- removal and ability to minimize plugging (Test 1).
2) Evaluate the ability of the system to treat very low ClO_4^- levels (approximately 50 to 100 μg/L), which could be encountered in potential applications (Test 2).

Both Tests 1 and 2 were expected to last for approximately 2 weeks.

To achieve the above objectives, the Phase I treatment process was redesigned with two parallel trains of reactors, Reactor 1 (or R1) and Reactor 2 (or R2), as shown in the diagram below. R1 vessels were filled with Hydroxyl-PAC high-density polyethylene media, and R2 vessels were filled with polyethylene dishwashing sponge scrubbers impregnated with Celite R-635 pellets[5]. Both of these media were specifically selected to minimize the plugging observed in Phase I. Also, a freshwater feed system was devised and installed for dilution of incoming groundwater (per Objective #2). This system consisted of a 500-gallon polyethylene tank, 500-pound LPGAC vessel, a supply pump, and water source connections. Although the equipment was installed, this system was not used for dilution, as explained in Section 3.3.

Phase II Process Layout (Initial):



[5] The sponges impregnated with Celite R-635 pellets were used in place of R-635 pellets alone (per the July 30, 2001, proposal) based on the desire to test a low-cost, "low-tech" medium with properties similar to a proprietary, proven media.

As in the Phase I test, substrate (i.e. acetate – ethanol would not be used based on the Phase I test results) feed was set up independently for each treatment train; whereas, the nutrient addition was installed with a single pump for both processes. Also, in-line mixers were installed in place of the previously used mixing tanks to streamline the process and eliminate the cycling of flow into the bioreactors. Figure 3 shows a detailed process flow diagram for the Phase II ClO_4^- bioreactor system, depicting all of the Phase II design elements. These improvements were installed by August 2001. At that time, the project was put on hold due to questions about effluent discharge limitations.

In April 2002, the Navy directed FWENC to resume testing, and to dispose of all effluent generated during the Phase II testing off site because some of the salt levels that occur naturally in the groundwater exceeded RWQCB discharge limits. The Navy further required that each tank load of effluent be analyzed prior to off-site disposal to confirm it was non-hazardous. Based on these requirements, three additional 20,000-gallon staging tanks were mobilized to the site in August 2002, prior to beginning the test. Use of the secondary treatment processes (IE, LPGAC, aerobic bioreactor) was also modified because of the off-site disposal requirement. Specifically, IE and LPGAC were retained (because VOCs and ClO_4^- are potentially constituents of concern), and the aerobic bioreactor was removed from the process [because effluent biochemical oxygen demand (BOD) was not a concern in off-site disposal].

3.1.1 DSBRs Versus other Bioreactors

Unlike the Phase I test, where Celite R-633 media completely filled the bioreactor vessels, for Phase II, both types of media used in the testing (the Hydroxyl-PAC and sponges impregnated with Celite R-633) were added to the reactors until each vessel was about two-thirds full. This design change, coupled with characteristics of the media described above (lower relative surface area and greater pore volume), was expected to diminish plugging associated with the Phase I bioreactors. Because the media bed would no longer be subject to static conditions (as in a PBR; in fact, it would be suspended within the reactor and therefore subject to flux), the Phase II bioreactor was renamed a DSBR to better reflect the actual process. It was also envisioned that the dynamic bed properties would make backwashing easier.

The DSBR is distinguished from CSTRs and FBRs as follows.

CSTRs have about the same residence time as DSBRs, but have a reduced packing density (about 20 to 30 percent) of their sponge-like media[6], which allows for complete mixing of the media and contacting with the incoming wastewater. Loadings [gallons per minute per square foot (gpm/ft^2)] are also generally higher than in the DSBR, due to recommended vessel configurations. A FBR, on the other hand, has generally the highest loading rates of biological

---

[6] Specifically, the CSTR offered by Ecomat was considered for this comparison. Other CSTRs may be available that do not use any media, or use media other than sponges.

systems, and the shortest residence times. Higher loading rates are needed to achieve (i.e., approximate) fluidization velocities. FBRs also use a granular medium for bacterial attachment, such as activated carbon or sand. To achieve the same flow rate through the granular media as the plastic media used in the DSBR/CSTR, higher pump discharge pressures are required, because pressure losses are expected to be higher through granular material. Hence, it is expected that FBR power usage would be higher than the DSBR/CSTR.

3.2 MICROBIOLOGY/STARTUP

Based upon the results of Phase I testing that showed similar ClO_4^- removal efficiencies for the JPL native bacteria and perc1ace, it was not envisioned that proprietary microbes would be required; hence, JPL isolates initially were selected for further testing. The JPL isolates (combined) were grown in MSM at CEM in 10-liter batches using acetate in the presence of ClO_4^-.[7] On August 15, 2002, the cultures were dispensed into approximately 200 gallons of JPL tap water that had been added to each pair of reactor vessels and startup (inocula) tanks (Figure 3)[8]. In addition, chemicals were added to formulate a relatively rich MSM, and sodium perchlorate ($NaClO_4$) and sodium acetate also were added at final concentrations of 100 milligrams of ClO_4^- per liter and 500 milligrams of acetate per liter, respectively. The field notes for Phase II test are presented in Appendix D.

The basic strategy for startup was to assess ClO_4^- removal at the outset to verify that the cultures were reducing ClO_4^- as expected. Oxygen was then introduced using a small pump to build biomass within the reactors[9], as it was not practical to add ClO_4^- continually to the medium. Periodic ClO_4^- additions and analyses (using a ClO_4^--specific probe) were conducted to verify viability/ClO_4^--reducing activity. Based on ClO_4^- concentrations calculated from ClO_4^--specific probe readings (Table 3), significant ClO_4^- removal from the startup solution was observed in less than 24 hours for both startup reactor systems. The ClO_4^- concentrations shown in Table 3 were calculated from the initial (laboratory) regression curve for the standards (regression curves were not calculated daily as in Test 1 – the Test 1 data showed that the reliability of the initial calibration for daily use, and relative disappearance data were satisfactory). At the end of the day on August 16 (ClO_4^- reduction had been observed during the day), each startup tank was aerated to sustain the cultures. The startup tanks were aerated for a few days, after which, the ClO_4^--reducing ability of the system was tested again – and confirmed – from August 22 through 23, 2002.

---

[7] Acetate and ClO_4^- concentrations in MSM were 500 mg/L (as acetate ion) and 100 mg/L, respectively.

[8] The tap water was passed through a 500-pound LPGAC vessel to remove chlorine prior to filling the tanks. The freshwater feed setup was used in this manner for making up all chemical feed solutions for the test.

[9] The JPL isolates are facultative and can thus use oxygen as a terminal electron acceptor in respiration. This was shown to have no adverse effects on ClO_4^- reduction by these bacteria and can be considered an asset as it eliminates the need to maintain strict anaerobic conditions in managing startup activities.

During the startup period, water quality parameters (pH, temperature, and conductivity) were monitored, and substrate (acetate) was added to ensure that the system was not deficient in electron donor concentrations. As shown in Table 3, the pH varied from about 6.8 to 7.8 during startup, and generally increased as the media were colonized. The temperature varied from about 21°C to 31°C. Temperature and pH were not adjusted daily, based upon the results of the Phase I testing, which had shown that these parameters did not vary significantly and were of minor importance compared with maintaining a consistent and appropriate residence time.

3.3 OPERATION

In the Phase II proposal, it was established that Test 1 and Test 2 would address the two objectives described above: 1) to assess two different packing materials for process efficiency in terms of ClO_4^- removal and ability to minimize plugging, and 2) to evaluate the ability of the system to treat very low ClO_4^- levels. However, about a week or so after operation commenced, it was evident that: 1) the influent ClO_4^- levels were an order of magnitude higher than anticipated and as such, would render dilution to the low levels proposed for Test 2 impractical; and 2) process modifications would be needed to enhance the system's capability to treat these higher ClO_4^- concentrations (reactor residence time, for instance, was based on the much lower influent concentrations encountered in the Phase I testing). Hence, Test 1 was performed up until the time that the effluent staging tanks were full; at which time, the treatment system was put into re-circulation mode. Process enhancements were then made in preparation for Test 2, which was re-scoped to evaluate additional residence time in light of the increased ClO_4^- concentrations and to test the performance of perc1ace in conjunction with Hydroxyl-PAC media.

3.3.1 Test 1

On August 26, 2002, groundwater was introduced into the influent equalization tank (Figure 3) and pumped through each reactor train initially at 2.5 gpm. The substrate (sodium acetate) was added for a final concentration of 300 mg/L (as acetate) initially (higher than the design level of 100 mg/L[10] in order to gradually reduce from 500 mg/L in the startup solution). Nutrients (di-ammonium phosphate) were added for a final concentration of 1.0 and 0.9 mg/L as hydrogen phosphate (HPO_4-P) and ammonium nitrogen (NH_4-N), respectively. Reactor influent, intermediate, and effluent water quality parameters [pH, temperature, conductivity, turbidity, and dissolved oxygen (DO)] were checked at least once daily using a Horiba U10 Water Quality Checker (the Horiba U10 malfunctioned and was replaced with a YSI Model 51B instrument for DO readings for the latter part of the test). Appendix E provides a summary of these readings. A reduction in DO from influent levels of 5 to 7 mg/L to around 1 mg/L or lower indicated that

[10] The objectives of this relatively short-term test did not include acetate feed optimization. Based on the objective of demonstrating process feasibility at the field scale, excess acetate was added to ensure that it was not a limitation in the reactors during the test. Appendix G shows the stoichiometry for the reaction between ClO_4^-, O_2, and NO_3^- with acetate and the theoretical acetate requirement calculation.

reducing conditions (appropriate for ClO_4^- reduction) were established in the reactors[11] (Figure 4). Figure 5 depicts pH versus time and shows that 1) influent pH was relatively stable between 7 and 7.5, 2) pH increased from influent to effluent for both reactors, and 3) effluent pH increased slightly with time but remained within the optimal metabolic range for the bacteria (6.5 to 8.5). Figure 5 also shows temperature versus time and shows that the water temperature ranged from about 19 to 24°C during the test[12]. As with the startup period, pH and temperature were not adjusted, based on the results of Phase I testing. Samples were collected for laboratory analysis for NO_3^- and ClO_4^- (combined influent, R1 and R2 intermediate and final effluents) and acetate (R1 and R2 influent, intermediate, and final effluent) twice a week (four times total) during this test.

Table 4 shows a summary of important operational parameters for Test 1, which was completed on September 9. The table shows that despite efforts to equalize flows through both reactors, the average flow rate for R1 was about 2.1 gpm versus about 2.8 gpm for R2. This implies that there were slightly better flow characteristics associated with the sponge media versus the Hydroxyl-PAC media. The table also shows that based on these flow rates, the average residence time ranged from 36 to 48 minutes for R2 and R1, respectively.

Table 5 summarizes the results of Test 1. Figure 6 shows ClO_4^- and NO_3^- removal efficiencies for both R1 and R2. The figures show that R1 (with Hydroxyl-PAC media) performed slightly better than R2 (with sponges) for ClO_4^- and NO_3^- removal[13]. Although ClO_4^- removal performance was mediocre, but not unexpected as the bacteria were still acclimating during this period and residence times were short given the high-influent ClO_4^- concentrations, both reactors showed capability for NO_3^- removal to non-detectable levels (less than laboratory reporting limits of 0.44 to 0.88 mg/L) from about 50 mg/L influent NO_3^-. Figure 6 also shows acetate addition and removal through the treatment process[14]. The figure shows:

1) Acetate feed concentrations were not consistent between R1 and R2, except on the August 28, 2002, sampling, 2 days after operation commenced.
2) In one instance (August 28, 2002, sampling), higher acetate concentrations were measured in the effluent relative to those measured in the influent (for both reactors).

---

[11] DO readings for September 3 to 4 (days 9 to 10), shown in Appendix E, should be disregarded due to a malfunctioning DO probe. Starting on September 5, a new DO probe, a YSI Model 51B, was used.

[12] These temperatures are based on samples collected from various sample ports in the treatment system, after the water had been retained in the influent (equalization) tank and subjected to solar heating. Groundwater temperatures at JPL are typically in the 16 to 20°C range, based on past groundwater monitoring results.

[13] The final sample result for R1 effluent from September 6, 2002, is shown as reported by the laboratory (not detected <0.004 mg/L, laboratory reporting limit). The result was 1.5 mg/L before the laboratory had to reanalyze the sample due to a quality control error. Although the sample was preserved, ClO_4^- reduction may have occurred in the sample in the laboratory due to the extended reaction time between analyses (if bacteria were still present and viable in the sample).

[14] Samples were not analyzed for acetate on the Day 4 sampling.

3) Acetate sometimes was present in the effluent (always for R1).

Based upon these results, acetate consumption could not be correlated with each reactor's performance. Interpretation of the acetate results was also complicated by laboratory equipment problems, which resulted in apparent data inconsistencies. Influent data appeared unreliable – although this could be due in part to inconsistent feed concentrations.

Test 1 was concluded after approximately 21,900 gallons had been treated in R1 and 33,300 gallons in R2. It must be emphasized that due to conditions beyond FWENC's control, the Test 1 forward flow duration was only 14 days. This normally would be considered part of the startup phase in the general sense of bioreactor operation (which can last up to 30 days). During this period, the biological system typically goes through an acclimation period, and performance is generally is not optimal.

Following Test 1, both reactor trains were put into a re-circulation mode, meaning the effluent was not discharged, but rather recycled back through the reactors for re-treatment. Re-circulation lasted for several weeks while treated effluent from Test 1, which had filled the staging tanks, was disposed off site.

3.3.2 Re-circulation Period

Profiling and disposal of the effluent generated from August 26 through September 9, 2002, was required before the testing could be continued. During the period of September 10 through October 1, 2002, groundwater was re-circulated through each reactor system, and substrate and nutrients were added. A few hundred gallons of fresh groundwater were pumped through the reactors daily to increase the influent DO concentration, and the influent and effluent equalization tanks were also aerated to ensure the presence of an electron acceptor for culture maintenance[15]. Groundwater parameters were checked and recorded (note: the Horiba U10 was replaced with a YSI Model 3500 for pH, temperature, and conductivity readings; the YSI 3500 could not measure turbidity; DO was still measured with the YSI Model 51B) during this re-circulation period (see Appendix E), and analytical samples (for ClO_4^-, NO_3^-, and acetate) were collected on a few occasions (see Appendix F). ClO_4^- was not detected at any point in the system during re-circulation.

3.3.3 System and Experimental Modifications

During the re-circulation/waste disposal period, adjustments were made to better equip the system to treat the much higher than anticipated influent ClO_4^- levels. Results of Test 1 showed that R1, filled with the Hydroxyl PAC media, had a slightly better ClO_4^- removal performance

[15] In lieu of the pump used in Test 1, a ½ horsepower air compressor was connected to piping with tiny holes to diffuse air into each tank. One of the 20,000-gallon staging tanks was used as the effluent equalization tank for the re-circulation period.

than R2, filled with the Celite R-635-impregnated sponge scrubbers. Both of these media displayed performances in the same range, but were significantly better than the media used in Phase I. As mentioned previously (see Subsection 3.1.1), this improvement was hypothesized as being a result of the better flow characteristics. It was therefore concluded that, provided flow characteristics were such that plugging was reduced (or eliminated), the type of media used in this test did not impact reactor performance. Based on this reasoning, the following changes were made to the treatment system/approach:

1) The (Celite-impregnated) sponge scrubber media were replaced with Hydroxyl-PAC media, and perc1ace was grown up and inoculated in R2. R1 was left unchanged. This change would allow comparison of perc1ace performance with JPL isolates, under similar (and, in comparison with Phase I, potentially optimal) flow conditions.
2) The flow rate was reduced for subsequent testing to about 1 gpm per reactor train to increase the residence time to about 100 minutes per train (two reactors) from 36 to 48 minutes.
3) A polishing reactor filled with the Hydroxyl-PAC media (with a 25-minute residence time at 2 gpm) was added in series with both reactor trains (Figure 3).
4) A second influent pump was added so that a separate pump could feed each reactor train – this was expected to improve the flow balancing between the two reactors and mitigate the flow rate variations observed in Test 1.

On September 24, 2002, perc1ace, MSM, sodium acetate, and $NaClO_4$ were re-circulated through R2 in the same manner as described previously with the JPL isolates. ClO_4^- disappearance from the startup media (from 100 mg/L to less than 1 mg/L) was achieved in less than 24 hours with the perc1ace (Appendix E). The R2 startup tank was then aerated and water was re-circulated to build up biomass in R2 prior to startup. While R2 was inoculated with perc1ace, R1 continued in the same re-circulation mode with influent and effluent tanks described above.

3.3.4 Test 2

Test 2 was performed from October 1 through 18 with both R1 and R2 operating in parallel, with the system and experimental modifications described above. Table 4 shows a summary of key operational parameters for Test 2. The overall flow rate averaged about 1 gpm for R1 and 0.9 gpm for R2 throughout the test. There were some daily flow fluctuations, though less than those observed in Test 1, as shown in Appendix E. Based on these flow rates, the average residence time increased to about 100 minutes for Reactor 1 and 111 minutes for R2. Pressure drops were also recorded through each reactor system throughout the test (Appendix E). Water quality parameters were measured with a YSI Model 3500 (for pH, temperature, and conductivity) and a YSI Model 51B (for DO) (Appendix E). DO readings again validated reducing conditions in the system (Figure 4). Figure 7 shows pH versus time for Test 2. It shows that 1) pH increased from influent to effluent, but did not change significantly for each sampling

location during the test, and 2) pH remained between 6.5 and 9.0 during the entire test, within the tolerance range for the bacteria. Also noteworthy is that the pH for R1 did not change significantly during the 3-week re-circulation period (compare end of Figure 5 with beginning of Figure 7). Figure 7 also shows temperature versus time and shows that the water temperature ranged from about 17°C to 25°C throughout the test. Again, pH and temperature were not adjusted during this test.

Analytical samples were collected for laboratory analysis for ClO_4^- and NO_3^- (combined influent, R1 and R2 intermediate and effluent, and final effluent – a new sampling point for the effluent of the polishing reactor) and acetate (R1 and R2 influent, R1 and R2 intermediate and effluent, and final effluent). Per the Navy's direction, Test 2 was concluded after 18 days with approximately 24,000 gallons treated in R1 and 22,600 gallons in R2.

For Test 2, Figure 8 shows that R2 (inoculated with perc1ace) generally had more efficient ClO_4^- removal compared to R1 (JPL isolates), and effluent ClO_4^- concentration for this reactor was twice non-detect (less than 4 µg/L, the method reporting limit for the laboratory) during the testing (after 3 days and 13 days; see also Appendix F). R2 achieved removal to 0.42 mg/L. ClO_4^- concentrations after the final polishing reactor ranged from 0.4 mg/L after 1 day of testing to less than 4 µg/L after 13 days (levels were 77 µg/L or below from Days 7 to 17)[16]. Thus, generally after 1 week of operation, after the final effluent reactor (total residence time approximately 125 minutes), ClO_4^- concentrations were reduced to low µg/L levels. Figure 9 shows ClO_4^- removal efficiency as a function of influent ClO_4^- concentrations for each reactor vessel – primary and secondary for R1 and R2 and final effluent (polishing reactor). Figure 9 shows that ClO_4^- removal rates were relatively high (about 80 percent) at both low and high concentrations and generally best at the lowest concentrations (i.e., less than 3 mg/L, where removal usually exceeded 80 percent), illustrating the flexibility of the DSBR. Thus, to some extent, the original objective of testing the DSBR at lower concentrations (see Section 3.1) was met.

Table 5 and Figure 8 show that NO_3^- removal efficiency was very high, and consistent for the 17-day test for both reactors. Aside from Day 1, both reactors reduced NO_3^- to non-detect levels (0.44 to 0.88 mg/L, the laboratory reporting limit) throughout the entire test period.

---

[16] Removal efficiencies shown in Table 5 are based on samples collected from R1 and R2 effluent, not final effluent.

Figure 8 shows acetate addition/removal for Test 2 and with similar trends to Test 1. Due to difficulties experienced by the laboratory in measuring acetate (because of various equipment problems) and resulting apparent data inconsistencies, implications of the acetate addition/removal trends for Test 2 are not clear. Based on performance of both reactor systems in the test, it does not appear that acetate was limiting in the system and that feed concentrations could be reduced to minimize breakthrough. Acetate optimization was not in the scope of the testing, but it is strongly recommended for future work as it would have a significant effect on O&M costs (see Subsections 3.3.6 and 5.2.2).

Figure 10 shows the pressure drops through each bioreactor system for the duration of Test 2. Pressure drops were not recorded during Test 1. Figure 10 shows that pressure losses increased slightly with time (about 3 psi for R1 and 6 psi for R2) and were slightly higher for R1 (likely because R1 had been on-line since August 26, 2002, 34 days longer than R2). As the consistency of pressure readings show, neither reactor system showed evidence of significant plugging that was associated with the Phase I media (Celite R-633)/PBR system. This likely can be attributed to the fact that, unlike the PBR in Phase I, the bed was not static (the media were suspended, and in flux). Also, inlet pressures were relatively low (less than 20 psi) compared with a typical FBR (about 40 to 45 psi).

Figure 11 shows the analytical results for several general water quality parameters (carbonate and bicarbonate alkalinity, sulfate, Cl^-) from two sampling events conducted on the final date of testing. Noteworthy are the increase in Cl^- concentrations from influent to final effluent indicating Cl^- production from the reduction ClO_4^-, and the stability of the sulfate concentrations, indicating that no sulfate reduction occurred. Absence of sulfate reduction is consistent with FWENC's prior work (bench and field) using the PBR to treat JPL groundwater. Reduction of sulfate would generate hydrogen sulfide, indicating that the redox potential was lower than required for ClO_4^- reduction to occur.

Figure 12 shows a schematic of the Phase II, Test 2 treatment process; flow rates; and influent, intermediate, and effluent ClO_4^-, NO_3^-, and DO concentrations for four data points during Test 2: one at the beginning of the test prior to the bacteria being acclimated (October 2, 2002, sampling) and the other three during the final week of testing (during optimal performance, as defined for this relatively short test). In looking at these figures, the following conclusions can be made:

1) DO reduction occurred mainly in the primary vessels. DO reduction was efficient enough such that ClO_4^- reduction that likely occurred downstream of the oxygen-reducing zone was still substantial (generally about 80 percent at higher ClO_4^- concentrations).
2) ClO_4^- removal rates ranged from about 70 to 80 percent in the primary and secondary vessels during the peak performance period and were highest in R2-secondary (80 to 90 plus percent) and in the final polishing reactor (generally above 90 percent). The latter treated water with the lowest influent ClO_4^- concentrations. The ClO_4^- removal

efficiency was about 70 to 80 percent regardless of the influent ClO_4^- concentration (1.4 to 10.6 mg/L).

3) NO_3^- removal rates exceeded 90 percent for a 50-minute residence time, and removal to non-detectable effluent concentrations was complete within 100 minutes.

4) Residence times required for ClO_4^- removal to 4 µg/L ranged from about 111 to 125 minutes. Low µg/L effluent ClO_4^- levels were attained consistently at 125 minutes or about 2 hours. Because of the short duration of the tests, it is possible that building and stabilization of the biomass may have been incomplete, and reactor efficiency thus may be greater than indicated in these tests.

5) The schematic for October 18, 2002, shows increases in Cl^- concentration corresponding to reduction of ClO_4^- during the treatment process. The increase in Cl^- across R1 (10 mg/L) is greater than the theoretical increase from the stoichiometry shown in Appendix G (i.e., the decrease in [ClO_4^-] by 7.9 mg/L across R2-primary should result in an increase in Cl^- concentration of approximately 3 mg/L), likely due to analytical variation.

3.3.5 Comparison with Objectives/Deviations from Scope, Phase II

Objectives for the Phase II testing of the ClO_4^- bioreactor were 1) to assess performance of two different packing materials for ClO_4^- removal and ability to minimize plugging and 2) to evaluate the ability of the system to treat very low ClO_4^- levels (e.g., 50 to 100 µg/L). For the first objective, Phase II testing of the DSBR showed that both media tested (the proprietary Hydroxyl-PAC and sponges impregnated with Celite R-635) performed adequately in terms of not plugging – although the sponge media seemed to have more favorable flow characteristics – and showed promise for efficient ClO_4^- removal based on the results of Test 1. In Test 2, the Hydroxyl-PAC media (when colonized with perc1ace) satisfied the performance objective for ClO_4^- removal, achieving non-detectable effluent levels with 111 minutes of residence time and showing favorable flow capabilities based on only slight increases in pressure drops over the course of the 18-day test. Thus, based on these results, either medium appears to be satisfactory in meeting performance requirements.

Because the unexpectedly high-influent ClO_4^- concentrations rendered dilution to very low levels unfeasible, reactor performance in treating low ClO_4^- concentrations was evaluated based on removal efficiencies through the final effluent polishing reactor in Test 2. As described above, the DSBR performed optimally at these low ClO_4^- concentrations. In instances where the effluent to the polishing reactor was below 0.5 mg/L[17] (the lowest was approximately 0.27 mg/L), ClO_4^- concentrations were reduced by 79 to 99 percent.

---

[17] Since samples were not collected from the combined effluent from R1 and R2, and since flow rates were similar, the effluent concentrations of each were averaged to generate a representative "influent" concentration to the polishing reactor.

3.3.6 Additional Substrate Utilization Testing

The next step following completion of the field testing described in the preceding subsections was to prepare cost estimates for scaling up of the DSBR. Due to the limited time that the system was actually operated in the forward flow mode, the cost estimates were expected to provide an initial approximation of what would be required to construct and operate a DSBR at JPL. During preparation of the cost estimate, it became evident that substrate utilization has a significant influence on operation costs. Per Navy direction, assessment of substrate (acetate) utilization was beyond the original scope of the testing, and acetate was provided well in excess of theoretical requirements so it would not limit evaluation of ClO_4^- removal, which was the chief objective. However, influent and effluent acetate concentrations were measured during forward flow operation, and it was expected that the difference in the respective measurements would provide an indication of acetate consumption under operating conditions. However, as discussed above in Subsections 3.3.1 and 3.3.4, it was not possible to accurately estimate acetate requirements.

Accordingly, it was agreed that additional laboratory microcosm tests would be conducted to obtain preliminary data to support the costing effort to the extent possible, rather than rely solely on calculations based on published chemical equations. Two experiments were therefore conducted, the objectives of which were as follows:

1) Experiment 1: Obtain an experimental estimate of acetate requirements of the JPL isolates and perc1ace for reduction of various concentrations of ClO_4^- in the presence of representative O_2 and NO_3^- levels relative to operation of the DSBR.

2) Experiment 2: Obtain an experimental assessment of acetate and methanol uptake by the JPL isolates for various concentrations of ClO_4^- in the presence of representative O_2 and NO_3^- levels for the DSBR, and to assess ClO_4^- reducing efficiency of various acetate and methanol mixtures.

The second experiment was based on the premise that methanol is an alternate substrate which is known to support ClO_4^- reduction in certain bacteria, and is less expensive than acetate salts. No data were available regarding methanol utilization by the JPL isolates. It was hypothesized that methanol could potentially be used to supplement or replace acetate, and thus reduce the acetate requirement and associated costs without sacrificing ClO_4^--reducing capacity. Importantly, it is recognized that these supplemental experiments were not expected to mimic the actual treatment process, but rather to provide experimental evidence to support preliminary evaluation of the stated objectives.

3.3.6.1 Materials and Methods

The experiments were conducted in sealed 500-milliliter (mL) Erlenmeyer flasks set up with MSM. The MSM contained acetate, methanol, or acetate plus methanol at various ratios as the sole carbon and energy source(s), and oxygen, nitrate, and ClO_4^- as the sole electron acceptors.

The medium was inoculated with standardized inoculum containing either the JPL isolates and/or perc1ace, and electron donors and acceptors were tracked analytically. Experimental details are given below.

Experimental Design

The experimental design for the additional substrate utilization studies is given below.

Treatments Experiment 1	Bacterial Isolate	$[ClO_4^-]$ (mg/L)	$[O_2]$ (mg/L)	$[NO_3^-]$ (mg/L as NO_3^-)	Acetate (mg/L)	Methanol (mg/L)
1	JPL	10	8	50	500	N/A
2	JPL	1	8	50	500	N/A
3	JPL	0.1	8	50	500	N/A
4	perc1ace	10	8	50	500	N/A
5	perc1ace	1	8	50	500	N/A
6	perc1ace	0.1	8	50	500	N/A
Experiment 2						
7	JPL(1)	10(2)	8	50	150	0
8	JPL	10	8	50	100	25
9	JPL	10	8	50	75	50
10	JPL	10	8	50	50	100
11	JPL	10	8	50	25	150
12	JPL	10	8	50	0	200

Notes:

(1) Only the JPL isolates were used in order to assess a greater range of acetate/methanol ratios as relevant to the JPL test.

(2) Only one concentration was used, based on the best trends observed in Experiment 1.

N/A – Not Applicable

Inoculum Preparation

JPL enrichments and perc1ace were used as specified above. The cultures were pre-grown in FTW mineral salts medium consisting of the following (in g/L): di-potassium hydrogen phosphate (K_2HPO_4), 0.225; potassium di-hydrogen phosphate (KH_2PO_4), 0.225; ammonium sulfate $(NH_4)_2SO_4$, 0.225; magnesium sulfate ($MgSO_4$) x $7H_2O$, 0.05; calcium carbonate ($CaCO_3$), 0.005; ferric chloride ($FeCl_2$) x $4H_2O$, 0.005, acetate, 1.0 (as sodium acetate), and 1 mL of trace elements solution. ClO_4^- was added at the required concentration (depending on the experiment) as $NaClO_4$. The medium was autoclaved (121°C, 15 minutes). Cultures were pre-grown to an optical density at 600 nanometers (OD_{600}) of 0.35. A total of 750 mL of each culture was centrifuged and washed with 40 mL of sterile FTW medium. Cells were resuspended in 15 mL of FTW medium.

Batch Studies of Electron Donor Uptake and Electron Acceptor Reduction

Medium: FTW medium was used in all experiments. Batch experiments were carried out in 500-mL Erlenmeyer flasks containing 500 mL of FTW medium, to which, ClO_4^-, NO_3^-, acetate, and methanol and were added. Three ClO_4^- concentrations (0.1, 1, and 10 mg/L) were used in Experiment 1, and based on results, a concentration of 10 mg/L was used in Experiment 2. NO_3^- was added to yield a final concentration of 50 mg/L in Experiments 1 and 2. Acetate was added in excess to yield a final concentration of 500 mg/L in Experiment 1, and acetate/methanol ratios were varied in Experiment 2. The medium was saturated with O_2 using an aquarium pump and with stirring.

Inoculation, Incubation, and Sampling: Each flask of batch culture medium was spiked with 2 mL of the inoculum. Flasks were sealed with rubber stoppers and incubated at an ambient temperature (approximately 22°C). Fifteen-mL samples were collected at time T_0, T_4, T_8, T_{12}, T_{16}, T_{20}, and T_{24} hours. Samples were removed as rapidly as possible, and headspaces were flushed with nitrogen gas during sampling to minimize O_2 diffusion into the medium.

Analysis of DO: DO levels were monitored *in vitro* using an Orion DO meter before samples were collected.

Analysis of ClO_4^-, Acetate, and NO_3^-: Samples were filtered and analyzed using a Dionex ion-chromatography system (Dionex, Sunnyvale, California), equipped with a GP50 gradient pump, an AS40 automated sampler, 740-microliter (µL) injection loop, an ionPac AS11 column, and an ED40 conductivity detector. For the ClO_4^- analyses, the eluent was 100 millimolar (mM) sodium hydroxide (NaOH) at 1/mL. An ASRS-II (4 mm) suppressor, operated at 300 milliamps (mA) was used to suppress the eluent, using water as the regenerant. Acetate and NO_3^- were analyzed using the same ion-chromatograph except that the eluent was 21 mM NaOH and the suppressor was operated at 100 mA. For Experiment 2, methanol analysis was carried out using U.S. Environmental Protection Agency (EPA) Method 8015B, performed by a subcontracted laboratory (Centrum Analytical Inc., Riverside, California).

3.3.6.2 Results and Discussion

Results of the above analyses from Experiments 1 and 2 are tabulated in Appendix G, which also contains analytical reports for the subcontracted analysis (methanol) for Experiment 2.

The results of Experiment 1 are depicted graphically in Figure 13 for the treatments receiving 1 and 10 milligrams (mg) ClO_4^-/liter. ClO_4^- was not detectable (less than 4 µg/L) in the treatment receiving 0.1 mg/L at 8 hours, and the data are not shown. The data for the treatments receiving 1 and 10 mg ClO_4^-/L show that oxygen depletion occurred within the first 4 hours, and that NO_3^- depletion was complete at 20 hours. ClO_4^- reduction occurred concurrently with the latter stages of NO_3^- depletion, and all electron acceptors were generally non-detectable at 20 hours, with the

exception of the perc1ace treatment at 10 mg ClO_4^-/L, which still had detectable perchlorate at the conclusion of the experiment.

Acetate consumption was evident over the 24-hour period and was reasonably linear. The acetate consumption required to bring about complete reduction of all available electron acceptors is thus approximated by calculating the mean amount of acetate consumed over 20 and 24 hours for the JPL and perclace isolates (respectively) receiving 10 mg/L ClO_4^- and over 20 hours for both isolate treatments receiving 1 mg/L ClO_4^-. The calculations are as follows.

1) Estimated acetate consumption required to bring about a complete reduction of all available electron acceptors by JPL isolates and perc1ace at a starting ClO_4^- concentration of 10 mg/L:

JPL	Initial [Ac] Rep 1 (mg/L)	420.7	**Perc1ace**	Initial [Ac] Rep 1 (mg/L)	422.4
	Final [Ac], 20 hr, Rep 1 (mg/L)	326.6		Final [Ac], 24 hr, Rep 1 (mg/L)	243.9
	Ac Consumption Rep 1 (mg/L)	94.1		Ac Consumption Rep 1 (mg/L)	178.5
	Initial [Ac] Rep 2 (mg/L)	441.4		Initial [Ac] Rep 2 (mg/L)	437.5
	Final [Ac], 20 hr, Rep 2 (mg/L)	351.7		Final [Ac], 24 hr, Rep 2 (mg/L)	295.3
	Ac Consumption Rep 2 (mg/L)	89.7		Ac Consumption Rep 2 (mg/L)	142.2
	Mean Ac Consumption (mg/L)	**91.9**		**Mean Ac Consumption (mg/L)**	**160.4**
	SD (for Rep 1 and Rep 2)	**3.0**		**SD (for Rep1 and Rep2)**	**25.7**

2) Estimated acetate consumption required to bring about complete reduction of all available electron acceptors by JPL isolates and perc1ace at a starting [ClO_4^-] of 1 mg/L:

JPL	Initial [Ac] Rep 1 (mg/L)	449.7	**Perc1ace**	Initial [Ac] Rep 1 (mg/L)	446.9
	Final [Ac], 20 hr, Rep 1 (mg/L)	286.3		Final [Ac], 20 hr, Rep 1 (mg/L)	420.1
	Ac Consumption Rep 1 (mg/L)	163.4		Ac Consumption Rep 1 (mg/L)	26.8
	Initial [Ac] Rep 2 (mg/L)	317.1		Initial [Ac] Rep 2 (mg/L)	495.1
	Final [Ac], 20 hr, Rep 2 (mg/L)	314.4		Final [Ac], 20 hr, Rep 2 (mg/L)	358.4
	Ac Consumption Rep 2 (mg/L)	2.4		Ac Consumption Rep 2 (mg/L)	136.7
	Mean Ac Consumption (mg/L)	**82.4**		**Mean Ac Consumption (mg/L)**	**81.8**
	SD (for Rep 1 and Rep 2)	**113.8**		**SD (for Rep 1 and Rep 2)**	**77.7**

Notes:

Ac – acetate
SD – standard deviation

Despite some variability in the data, these results suggest than in general, the acetate consumption required to bring about complete reduction of all available electron acceptors at levels relevant to the JPL scenario is in the range of 100 mg/L. The theoretical requirements for the same electron acceptor scenarios, based on published chemical equations, is 43 mg acetate/L

(refer to Appendix G). Allowing for 20 percent inefficiency due to cell-building processes, the theoretical requirement may be estimated at 52 mg acetate/L. The value obtained here is roughly twice what may be expected based on theoretical analysis; however, as acknowledged, the microcosm experiment was not necessarily intended to directly simulate in-process conditions. Nevertheless, the values agree fairly well, and the study provides reasonable experimental evidence for using 100 mg/L as the acetate feed rate in preliminary cost evaluation.

Results of Experiment 2 are shown in Figure 14. In this experiment, all of the electron acceptors were reduced to varying extents; however in general, greater reduction was observed in treatments receiving greater amounts of acetate. Acetate consumption was measured to varying degrees as well, and was generally correlated with electron donor removal. Methanol consumption, however, did not appear to be significant. However, there is information to the contrary, suggesting that significant ClO_4^- reduction by perc1ace growing on methanol has been reported (Peter Hall, EcoMat, Inc., personal communication, June 6, 2003). As noted above, methanol uptake was only assessed for the JPL isolates, in order to assess a greater range of acetate/methanol ratios as relevant to the JPL test. Acetate uptake also was generally observed to be in the range of 30 to 100 mg/L (for varying degrees of electron acceptor removal), lending further support to the measurements made in Experiment 1.

3.3.6.3 Conclusions

Conclusions from this study are as follows:

1) Experimental evidence (Experiment 1) was obtained suggesting that acetate consumption for operation is in the range of 100 mg/L, which will be used in the preliminary cost analysis provided in this report. Importantly, this is likely a worst-case scenario, as slightly lower acetate consumption was noted in Experiment 2. Thus, it may be determined in future long-term testing that less acetate is actually optimal.

2) Methanol does not appear to be an efficient electron donor for the JPL isolates. It is not clear whether acetate/methanol blending would provide costing benefits if perc1ace were used in the reactor. However, based on successful use of methanol with perc1ace by others, use of methanol will be considered in scale-up cost estimates for the DSBR. In addition, it may be possible to test other substrates such as various agricultural or industrial waste products (brewery or winery wastes, which contain yeast extracts or acetic acid, or dairy wastes, which contain whey or lactic acid).

3.4 SUMMARY OF RESULTS AND CONCLUSIONS, PHASE II

3.4.1 Startup

Each of the inocula tested – the JPL isolates and perc1ace (prior to Test 2) – effectively reduced ClO_4^- in the startup medium, and a relatively small amount of time (approximately 1 week) was

required for initial colonization of the media. For both inocula, data from the ClO_4^- selective probe indicated that reduction commenced immediately, and removal of approximately 50 to100 milligrams of ClO_4^-per liter was achieved within 24 hours after seeding. Additionally, start-up was equally efficient for both the Hydroxyl-PAC and sponge media (prior to Test 1).

After ClO_4^--reducing conditions were established, the startup bacterial cultures were maintained by adding oxygen (via aeration) as the terminal electron acceptor for respiration, after which, the bacteria were re-acclimated to ClO_4^-. When aeration was terminated and ClO_4^- was re-added to the system (for JPL isolates – ClO_4^- concentrations were not re-measured in the perc1ace startup media after returning to anaerobic mode prior to startup), reduction of ClO_4^- commenced rapidly (with complete reduction within 24 hours), indicating that reductase(s) responsible for ClO_4^- reduction are essentially constitutive (requiring little or no induction period), and the bacteria tested were not adversely affected by exposure to O_2.

3.4.2 Forward Flow Testing

Test 1 showed that both packing materials (the Hydroxyl-PAC and sponges impregnated with Celite R-635) showed promise as media for ClO_4^- reduction, but the reactor with Hydroxyl-PAC media had slightly better overall performance. Reactors with both media showed the ability to reduce NO_3^- to non-detectable levels. ClO_4^- removal efficiency was not optimal in Test 1, as the bacteria were likely still becoming established and acclimating (only 14 days into the test – typical startup periods are 30 days) and residence times were short given the unexpectedly high-influent ClO_4^- concentrations. Initial residence times were based on the much lower levels encountered in previous testing. Temperature and pH were not varied during the test and did not appear to have an impact on ClO_4^- removal efficiency.

Based on the results of Test 1, several process/experimental modifications were made during a several-week re-circulation period between tests to better equip the DSBR system to treat the unexpectedly high influent ClO_4^- concentrations. The residence time was increased for subsequent testing by reducing the flow rate. A second feed pump was installed, and R2 was inoculated with perc1ace to test the performance of a different source of inoculum. Finally, a polishing reactor was added to the process in series with the other vessels.

Test 2 showed that R2, inoculated with perc1ace, performed better than R1 (with the JPL isolates) for ClO_4^- removal and produced an effluent with non-detectable ClO_4^- (less than 4 μg/L) at about 111 minutes residence time. Following treatment in the polishing reactor (25 minutes additional residence time), ClO_4^- removal to low μg/L levels was achieved during the final 10 days of the test. NO_3^- was reduced to non-detectable levels by both reactors throughout the test. Again, during Test 2, variations in pH and temperature were minimal and were not believed to be significant to the ClO_4^- removal efficiency of either reactor. Pressure drops through the reactors were measured throughout the test and did not change significantly. This can be

attributed to the nature of the plastic media (having a relatively high porosity) and the fact that the bed was not static.

Results of Phase II testing, with the modifications made between tests, indicate that the DSBR is capable of treating ClO_4^- from several mg/L to low µg/L levels after a few weeks of acclimation and that it is highly efficient for NO_3^- removal. Results also show that the DSBR is capable of efficiently treating relatively high (above 10 mg/L) and low ClO_4^- (below 3 mg/L) concentrations, with removal efficiency generally independent of influent ClO_4^- concentrations. That is, removal rates were in the 70 to 80 percent plus range for influent concentrations from 1.5 up to 10.5 mg/L. Although the reactor/media combination that performed best was the Hydroxyl- PAC/perc1ace (in Test 2), the sponge media used in Test 1 and the JPL isolates also showed promise for ClO_4^- removal. In a longer term test, it is expected that sponge media (or any media with similar properties)/ JPL aquifer inoculum (or any native aquifer inoculum) would perform similarly to the Hydroxyl-PAC/perc1ace.

4.0 SUMMARY OF OBJECTIVES/CONCLUSIONS, PHASE I AND II TESTING

	PHASE I	PHASE II
Objectives	Assess three sources of inocula with respect to: • Startup efficiency (time to achieve functional biomass in the reactor). • Process efficiency (ability to achieve non-detectable effluent ClO_4^- levels).	Assess two different packing materials for their process efficiency in terms of ClO_4^- removal and ability to minimize plugging (Test 1). Evaluate the ability of the system to treat very low ClO_4^- levels (approximately 50 to 100 µg/L), which could be encountered in potential applications (Test 2).
Results – Comparison with Objectives	• All three sources of inocula reduced ClO_4^- in the startup media; startup occurred more rapidly than expected (within 1 week). • Reactor with JPL isolates produced effluent with non-detectable ClO_4^- levels, then had breakthrough due to plugging. Reactor with perc1ace produced an effluent with 26 µg/L of ClO_4^-. Reactor with food waste was not tested due to excessive plugging.	Both media tested (the proprietary Hydroxyl-PAC and sponges impregnated with Celite R-635) appeared to be satisfactory in meeting performance objectives. Reactor performance in treating low ClO_4^- concentrations was evaluated based on removal efficiencies through the final effluent polishing reactor in Test 2 – this reactor performed optimally in reducing ClO_4^- at low (below 0.5 mg/L) influent concentrations (80 to 99 percent removal).
Conclusions/ Recommendations	Process efficiency appeared to be related primarily to the flow characteristics of the medium as opposed to the source of inoculum. Temperature/pH appeared to have little effect on process efficiency. Recommended testing alternative packing materials with a reduced propensity for plugging.	ClO_4^- removal efficiency was in the 70 to 80 percent plus range per reactor, regardless of influent ClO_4^- concentration. Although the combination of Hydroxyl-PAC media and perc1ace performed most efficiently for this short-term test, it did not appear that performance would be either media- or inoculum- specific in a longer term test. Methanol has shown promise as an electron donor for specific ClO_4^- reducers such as per1ace. Since for full-scale operations significant cost savings can be expected with use of alternative substrates to acetate (such as methanol), additional bench and field substrate uptake testing is recommended; using per1ace with methanol and other substrates.

5.0 SCALE-UP COSTING ESTIMATES

5.1 DSBR SCALE-UP PARAMETERS

Based on the DSBR Phase II field testing, Test 2 (Subsection 3.3.4), an average ClO_4^- removal rate of approximately 80 percent per reactor was achieved with a residence time of about 50 minutes. For water of JPL quality[18], an equation can be formulated to determine the number of DSBR vessels required to reduce a given influent ClO_4^- concentration to a particular discharge level. This equation is as follows:

$$[ClO_4^-](effluent) = [ClO_4^-](influent)\text{^}.2^N$$

Where: N = number of reactors

For instance, for $[ClO_4^-]$(influent) = 10 mg/L, and $[ClO_4^-]$(effluent) = 0.004 mg/L (4 parts per billion)

- N = log(0.004/10)/log(.2) = 4.86 → 5

The size of the reactor vessel would depend on the influent flow rate, and be based on the residence time required to achieve the desired effluent quality. Thus, using the results of the Phase II testing, a residence time of 100 minutes was assumed for scale-up purposes, treating groundwater with 10 mg/L influent ClO_4^- to non-detectable (less than 4 µg/L) levels. Hence, five 10,000-gallon reactors, each with a residence time of 20 minutes, would be required for a flow rate of 500 gpm under these conditions. Figure 13 shows a process schematic for a 500-gpm DSBR and shows the basic equipment required (influent tank, pumps, bioreactor vessels, substrate and nutrient tanks, and so forth) and possible optional equipment [e.g. bag filter, ultraviolet disinfection (UV) system, and IE vessels] needed for potable water applications. Figure 13 shows the DSBR vessels configured in series-parallel. Flow would be split through two reactors initially (250-gpm each) and then three reactors (167-gpm each). It is expected that the first two reactors would remove most of the competing electron acceptors and some ClO_4^-, and the final three reactors would provide polishing to treat to non-detectable (less than 4 µg/L) ClO_4^- concentrations.

---

[18] In terms of concentrations of competing electron acceptors, i.e., 50 mg/L of NO_3^-, 8 mg/L of DO.

5.2 DSBR COSTS – NON-POTABLE WATER APPLICATIONS

5.2.1 DSBR Capital Costs

Cost estimates for 500, 1,000, and 2,000-gpm DSBRs are shown in Appendix H. Costs are presented for both non-potable and potable water applications. Capital costs for non-potable treatment scenarios include the basic DSBR equipment described above, as well as the media required to fill the bioreactor vessels, integral piping and valves, controls and instrumentation, and other appurtenances (such as an influent flow meter and DO meters). The costs do not include equipment to bring water to the system (e.g., well pumps and conveyance piping) or away from the system. Equipment costs for 1,000 and 2,000-gpm DSBRs were estimated from the 500-gpm system costs by multiplying various items by scaling factors of 1.5 to 2, depending on whether costs were expected to increase proportionally to the flow rate. For example, the cost of the bioreactor vessels for a 1,000-gpm DSBR was assumed to be double that of a 500-gpm system, since double the residence time and hence twice the treatment volume would need to be installed. Similar logic was applied to other items such as flow meters, media, etc. In cases where economies of scale could result in savings, such as with feed tanks, a scaling factor of 1.5 was used. The same scaling factors were applied to the 1,000-gpm system to estimate costs for the 2,000-gpm DSBR. All the scaling factors used are listed in Appendix H. In addition to equipment costs, the total capital costs also include construction costs (equal to 25 percent of equipment cost), engineering/legal costs (15 percent of equipment and construction costs), and a contingency fee (10 percent of equipment plus construction plus engineering/legal costs), as shown in the Appendix H cost estimate summary page. These costs are presented as a lump sum and amortized ($/year) over 20 years at 7 percent interest.

5.2.2 DSBR O&M Costs

DSBR (non-potable) annual O&M costs are shown in Appendix H for 500-, 1,000- and 2,000-gpm systems. These costs include chemicals added to the process water [i.e., substrate such as sodium acetate and nutrients (added as di-ammonium phosphate)], power consumption (for electrical equipment such as pumps, mixers, etc.), analytical costs for monitoring system performance and complying with discharge requirements, operator training/labor, and equipment maintenance. O&M costs also include costs for replacing the media in the reactors (if needed due to excessive biomass accumulation), and disposal of used media and any chemical feed stocks (in case of cleaning or maintenance required to feed tanks). The following are assumptions used in this estimate:

- Acetate and nutrient requirements are based on final concentrations used in Phase II testing, described in Subsection 3.3.1. Methanol requirements (alternate substrate – see below) are based on a vendor estimate (EcoMat, Inc., email, January 10, 2003)

- Media for all reactors are changed every 5 years. This is a conservative assumption, as the media can likely be cleaned, and reused. Chemical feed solution, equal to capacity of feed tanks, is disposed annually.
- Operating labor will include a technician for 16 hours per week.
- For performance monitoring, samples are collected from the influent and combined effluent from the first pair and final three reactors, respectively on a twice-weekly basis for the first month and monthly thereafter, and analyzed for ClO_4^-, NO_3^-, acetate, Cl^-, and SO_4^{2-}.
- For discharge monitoring, samples are collected monthly for compliance with NPDES requirements. Appendix H shows specific analyses expected to be required.
- Maintenance costs are estimated as 5 percent of the (lump sum) capital cost per year.

O&M costs for the 1,000- and 2,000-gpm DSBRs were computed in a similar fashion to the capital costs. As shown in Appendix H, for the 1,000-gpm system, a scaling factor of 2 (from 500-gpm costs) was used to estimate costs for electricity, chemical usage, and media replacement/disposal since these costs would be expected to be proportional to the flow rate, whereas a factor of 1.25 was applied to labor. Analytical costs were assumed to increase by 2 for the performance monitoring analyses (since with the added treatment train of vessels, there would be double the number of sampling points), but stay the same for the discharge monitoring element (since there would still be one discharge point). Maintenance costs were still calculated as 5 percent of the (lump sum) capital cost. Similar logic was applied for computing the O&M costs for the 2,000-gpm DSBR from the 1,000-gpm system costs.

DSBR O&M costs shown in Appendix H are also broken down according to whether acetate or an alternate substrate such as methanol is used during treatment. As described in Subsection 3.3.6, although the DSBR Phase II testing and subsequent acetate/methanol utilization testing have confirmed success only with acetate used in combination with the JPL isolates, methanol has shown promise as a substrate when used in combination with perc1ace by others. As shown in Appendix H, methanol is considerably less expensive on a unit cost basis [$0.15/pound (lb) versus $0.50/lb] than acetate and a smaller quantity [200 pounds per day (lb/d) versus more than 800 lb/d at 500 gpm] is required for treatment. This disparity in costs underscores the importance of identifying and testing potentially more cost-effective alternative substrates to acetate. It should also be noted that acetate consumption used in this estimate is based in part on 50 mg/L of influent NO_3^-, which, as Appendix G shows, has the greatest impact on the acetate feed requirement. While this is reasonable for the JPL site, NO_3^- concentrations would likely be lower at other sites.

5.2.3 DSBR Cost Estimate Summary (Non-potable Applications)

The following table summarizes capital and O&M costs for DSBRs for the 500-, 1,000-, and 2,000-gpm for non-potable uses, with acetate or methanol used as the substrate:

Non-potable Applications	Lump Sum Capital Cost ($)	Amortized Capital Cost – ($/year) [1]	O&M Cost – Using Acetate ($/year)	O&M Cost – Using Methanol ($/year)	Total Cost ($/acre foot) – Acetate	Total Cost ($/acre foot) – Methanol
500-gpm	506,494	47,809	355,595	203,776	500	311
1,000-gpm	978,556	92,369	641,080	335,438	455	265
2,000-gpm	1,905,463	179,862	1,221,078	631,408	434	246

[1] Using amortization factor of 10.6 (20 years at 7% interest)

This table illustrates the significant impact of substrate costs on the overall cost associated with DSBR operation. The total cost is about 1/3 less when using methanol for all flow scenarios considered. It is also apparent that costs decrease with an increased flow rate, mainly due to economies of scale associated with labor and certain equipment costs.

5.3 DSBR COSTS – POTABLE WATER APPLICATIONS

Equipment required for potable water applications shown on Figure 15 include an automatic sequencing duplex bag filter system to remove sloughed biomass and reduce turbidity to required levels, ultraviolet disinfection unit, and six U. S. Filter Magflex IE vessels in series-parallel. Costs for this equipment are shown in Appendix H. Similar equipment would be required for CSTR/FBR to meet standards for potable water. Scale-up costs for 1,000-gpm and 2,000-gpm systems were approximated as follows:

- IE vessels are rated for 200-gpm each, so six vessels are required at 500 gpm (3 pairs of vessels), ten are required at 1,000 gpm (5 pairs), and twenty are required at 2,000-gpm (10 pairs); 60 ft^3 of resin is required per vessel
- For pump and controls/instrumentation – assumed 1.5 scaling factor due to economies of scale
- For bag filter system – Basic design (1,200-gpm) quoted by vendor (John Bush, Rosedale Products, Inc., personal communication, May 29, 2003) is most economical for 500 and 1,000-gpm; the 2,000-gpm system requires minor retrofits, at 15 percent additional cost
- For UV disinfection equipment, used vendor-provided costs for 500 and 1,000 gpm (Hydroxyl Inc., personal communication, January 9, 2003); based on these provided costs, estimated cost for 2,000-gpm unit

Operating costs associated with potable water equipment include electricity to operate the conveyance pumps and UV bulbs, filter replacement bags and associated disposal, resin replacement and disposal (replacement rate was based on usage rate of 0.75 cubic feet of resin per acre foot of water treated), and maintenance (calculated as 5 percent of capital cost for

additional equipment). Assumptions for filter bag replacement frequency and disposal are listed in Appendix H.

Operating costs for 1,000- and 2,000-gpm DSBRs were calculated using scaling factors proportional to flow rate for all of the above items except maintenance costs, which were still calculated at 5 percent of the capital cost for the additional equipment.

The following table summarizes DSBR capital and O&M costs for potable water applications:

Potable Applications	Lump Sum Capital Cost ($)	Amortized Capital Cost – ($/year) [1]	O&M Cost – Using Acetate ($/year)	O&M Cost – Using Methanol ($/year)	Total Cost ($/acre foot) – Acetate	Total Cost ($/acre foot) – Methanol
500-gpm	984,424	92,923	468,535	316,317	696	507
1,000-gpm	1,719,960	162,352	856,239	550,597	631	442
2,000-gpm	3,238,629	305,704	1,643,913	1,036,243	604	416

[1] Using amortization factor of 10.6 (20 years, 7 percent interest)

Total costs for potable water treatment are approximately 1/3 higher than for non-potable applications when acetate is used, and more than 50 percent higher when methanol is the substrate.

5.4 LIFE CYCLE COST - DSBR

Appendix H also shows 20-year life cycle costs for 500-gpm, 1,000-gpm, and 2,000-gpm DSBRs for potable and non-potable applications and for use of acetate and methanol as a substrate. These costs range from approximately $5 million for a 500-gpm DSBR/non-potable/methanol to approximately $28 million for a 2,000-gpm DSBR/potable/acetate. For all scenarios, life cycle costs increase nearly linearly due to the fact that most capital and O&M costs are expected to be proportional to flow rate, as described in detail above.

5.5 DSBR FOOTPRINT

Figure 16 shows an equipment layout for a 500-gpm DSBR. The figure shows the approximate area required for the basic equipment (for non-potable applications) and a control room with parking is about 75 feet by 110 feet (8,250 square feet). The optional equipment would require an additional approximately 25 feet by 50 feet. The area required for larger systems would be based on the additional space needed for additional trains of reactor vessels, larger or multiple feed tanks, etc. The required increase in area would not be linear (for basic equipment), as obviously the same control room/parking facilities would be used.

5.6 COMPARISON WITH OTHER CLO_4^- TREATMENT SYSTEMS

5.6.1 Other Biological Systems

5.6.1.1 CSTRs

CSTRs have been tested successfully on a field-pilot scale by a few vendors, including EcoMat, Inc. (Hayward, California). The total capital cost for a 500-gpm CSTR to treat 50 mg/L of NO_3^- and 10 mg/L of ClO_4^- is approximately $1.39 million (as a lump sum) for non-potable applications, and $1.83 million for potable applications, based on vendor estimates. This is significantly higher than the DSBR costs shown above ($506,494 and $984,424 for potable and non-potable, respectively) and is likely due to the reactor design/manufacturing costs (since reactor flow dynamics are critical to CSTR operation). Total capital costs for 1,000- and 2,000-gpm CSTRs for non-potable applications were based on vendor estimates (Peter Hall, EcoMat, Inc., e-mail, January 10, 2003). For potable applications, additional equipment required would be identical to the DSBR. Appendix H shows the lump sum and 20-year amortized capital costs [$/year and $/thousand gallons (kgal)] for 500-, 1,000- and 2,000-gpm CSTRs for potable and non-potable water applications.

Appendix H also shows O&M costs for the CSTR at 500-, 1,000-, and 2,000-gpm for non-potable and potable water applications. O&M costs consist of essentially the same elements as the DSBR. CSTR O&M costs (non-potable) were estimated as follows:

- Power usage - based on vendor estimates of 15 horsepower (hp)/reactor for pumping (Peter Hall, Ecomat, Inc., June 6, 2003), plus 10 kilowatts (kW) for appurtenances such as the DO meter
- Chemicals - methanol is standard substrate used (usage rates provided by EcoMat, Inc., email, January 10, 2003); nutrient usage identical to DSBR
- Labor - assumed identical to DSBR due to similar complexity
- Process sampling - based on the number of sampling points from vendor-provided vessel configurations and identical analyses as DSBR – except, obviously acetate is not measured. Discharge sampling costs are identical to DSBR.
- Maintenance costs are again assumed at 5 percent of the capital cost per year
- Media replacement
- Waste sampling/disposal (chemical feed stocks) – assumed identical to DSBR

O&M costs (non-potable) are similar for the CSTR and DSBR (using methanol) at 500-, 1,000- and 2000-gpm. CSTR media replacement costs were based on a vendor estimate of 10 percent of the total cost of the remaining operating costs annually, compared to the cost of replacing all media in the DSBR reactors every 5 years at a cost $14/cubic foot. O&M costs for potable water treatment are identical to DSBR, as equipment supplied would be the same.

Appendix H also shows the expected 20-year life cycle costs for 500-, 1,000-, and 2,000-gpm CSTRs, based on 50 mg/L of influent nitrate and 10 mg/L of influent ClO_4^-. These costs range from approximately $8 million for a 500-gpm CSTR for non-potable applications to approximately $27 million for a 2,000-gpm CSTR for potable water treatment. Costs are similar to DSBR-methanol for non-potable and potable applications, with DSBR costs being slightly lower at 500 and 1,000 gpm and slightly higher at 2,000 gpm.

5.6.1.2 FBRs

FBRs are currently in full-scale use at several sites for treatment of ClO_4^- in non-potable applications and are manufactured by a few companies including U. S. Filter. In these applications, FBRs have successfully reduced ClO_4^- to non-detectable (less than 4 µg/L) levels. The total capital cost for a 500-gpm FBR to treat 50 mg/L of NO_3^- and 10 mg/L of ClO_4^- is $791,151 (as lump sum) for non-potable applications and $1.17 million for potable applications, based on vendor estimates for reactors/pumps/controls/media and assuming similar equipment as the DSBR for equalization, interconnecting piping (assume cost is about 25% lower than DSBR due to fewer reactors), chemical mixing and feed, and flow/DO monitoring. These costs are higher than the DSBR costs shown above ($506,494 and $984,424, for non-potable and potable, respectively). Total capital costs for 1,000- and 2,000-gpm FBRs for non-potable applications were similarly estimated from vendor estimates and DSBR-specified equipment listed above for the 500-gpm system. For potable applications, additional equipment required would be identical to the DSBR/CSTR. Appendix H shows the lump sum and 20-year amortized capital costs ($/year and $/kgal) for 500-, 1,000- and 2,000-gpm FBRs for potable and non-potable water applications.

Appendix H also shows O&M costs for FBRs at 500-, 1,000-, and 2,000-gpm for non-potable and potable water applications. O&M costs consist of essentially the same elements as the DSBR/FBR. These costs (for non-potable applications) were estimated as follows:

- Power usage – assumed 25 percent higher than DSBR due to greater plant operational pressures
- Chemicals (methanol or acetate, nutrients) – identical usage as DSBR
- Labor – assumed 33 percent higher than DSBR/CSTR due to greater system complexity
- Performance sampling – based on the number of sampling points from vendor-provided vessel configurations and identical analyses as DSBR
- Discharge sampling analyses are identical to DSBR/CSTR
- Maintenance – again assumed at 5 percent of capital cost per year

Based on discussions with a FBR vendor (Bill Himebaugh, U. S. Filter, June 10, 2003) FBR media, such as activated carbon, are continuously regenerated and therefore, do not require

disposal. Waste sampling/disposal is thus based solely on costs for substrate and nutrient batch solutions in equal quantities to DSBR/CSTR.

O&M costs (non-potable) are lower than CSTR, but higher than DSBR (methanol) at 500 and 1,000 gpm. At 2000 gpm, FBR costs are slightly lower than DSBR.

Appendix H also shows the expected 20-year life cycle costs for a FBR for 500-, 1,000- and 2,000-gpm, based on 50 mg/L of influent nitrate and 10 mg/L of influent ClO_4^-. These costs range from approximately $6 million for a 500-gpm FBR (methanol) for non-potable applications to approximately $38 million for a 2,000-gpm FBR (acetate) for potable water treatment.

Subsection 5.6.4 below presents a cost comparison summary for DSBR, CSTR, and FBR capital and operating costs for 500-, 1,000-, and 2,000-gpm potable and non-potable scenarios.

5.6.2 IE Systems

IE is currently the system of choice at a number of ClO_4^--impacted sites in the southern California area. This is primarily due to the following reasons:

1) IE can treat ClO_4^- to levels below 4 μg/L and potentially below 1 μg/L.
2) In drinking water applications, IE does not have the stigma associated with biological systems and is approved by the California Department of Health Services (DHS) for drinking water use.
3) By virtue of IE being a physicochemical process, performance is stable throughout operation.

However, there are some important disadvantages to using IE, both in general, and specifically at JPL, as follows:

1) The cost of brine disposed from IE systems is prohibitive. This has prompted a number of leading IE providers to supply a one-time use resin, followed by burning to dispose spent resin.
2) The presence of other anions, such as NO_3^- (present at JPL at 50 mg/L) further increases costs, due to increased resin usage.

Appendix H shows capital, O&M and 20-year life cycle cost estimates for 500-, 1000-, and 2000-gpm IE systems. For a 500-gpm IE system for treating influent with 10 mg/L ClO_4^- and 50 mg/L NO_3^-, capital costs (lump sum) are expected to be about $378,000, which are significantly lower than all three biological systems in potable water treatment scenarios. For all flow scenarios, capital costs for IE are lower than the biological systems. IE capital costs are based on vendor quotes for treatment vessels and initial resin fill. Costs for interconnecting piping and appurtenances are estimated from experience with similar systems. IE capital costs include the same duplex bag filter system with automatic sequencing specified for potable water

treatment for the three biological systems (it is needed for pre-filtration to remove TDS). Scale-up factors for 1,000- and 2,000-gpm IE capital costs are shown in Appendix H.

IE operating costs are significantly higher than for the three biological systems for all flow scenarios, primarily due to the prohibitive cost for resin change-outs. Resin usage rates were based on vendor estimates of approximately 8.8 cubic feet of resin per acre foot of water for disposable styrenic resin (Purolite, Inc., email, March 10, 2003). Resin costs, from vendor estimates (Purolite, Inc, email, March 10, 2003), were assumed to be $100/cubic foot for refill plus disposal of spent resin. IE O&M costs also include power consumption (to operate process pumps, which would be similar to power usage for biological-potable water scenarios (less usage for UV disinfection unit), operating labor (estimated as one day per week), performance and discharge sampling [the former based on sampling weekly and analysis for anions and total dissolved solids (TDS) – as shown in Appendix H], and maintenance costs (again, assuming 5 percent of capital cost per year). Scale-up factors for 1,000- and 2,000-gpm IE O&M costs are shown in Appendix H.

Appendix H also shows the expected 20-year life cycle costs for IE for 500-, 1,000- and 2,000-gpm, based on 50 mg/L of influent nitrate and 10 mg/L of influent ClO_4^-. These costs range from approximately $17 million for a 500-gpm IE system to approximately $63 million for a 2,000-gpm system. These costs are significantly higher than the most expensive biological system using acetate (i.e., $38 million for 2,000-gpm DSBR).

5.6.3 NF Systems

NF appears to be a promising emerging technology for ClO_4^- removal. The main advantages of NF are relatively low (compared to IE) operating costs (including rejectate treatment/disposal, if required) and high-quality permeate. Potential disadvantages are a lack of proven reliability for ClO_4^- removal and the fact that the concentrated rejectate (comprising about 10 to 25 percent of the influent flow) may need to be treated/disposed. Rejectate management costs are somewhat difficult to quantify. Currently, rejectates from some applications are disposed in brine pipelines that dump into the ocean. However, it is understood that brine disposal may soon come under closer scrutiny from the RWQCB and/or the EPA. If treatment is required, biological treatment and electrolytic reduction have shown some promise for ClO_4^- removal.

Based on vendor estimates (Jim Stewart, Best Technologies, Inc., email, June 22, 2003), the total capital cost for a 500-gpm NF system (non-potable), including rejectate treatment (using DSBR), is approximately $1,099,000, which is significantly more expensive than all but the CSTR. Capital costs include NF modules (multiples of 160-gpm), membrane washing system, interconnecting piping, and air pressure tank. Equipment costs for rejectate treatment costs (for 500 gpm) are assumed to be equal to a DSBR treating the same flow as the NF systems. Although the flow rate will be about 12 percent for rejectate treatment (from vendor estimates, or about 60 gpm), ClO_4^-, NO_3^-, and other anion concentrations will be about 8 times higher, so the overall loading would be

equivalent to a 500-gpm DSBR treating 10 mg/L of ClO_4^-, 50 mg/L of NO_3^-, and so forth. Scale-up capital costs for 1,000- and 2,000-gpm NF units, with DSBR rejectate treatment, are shown in Appendix H. Costs for all NF capital equipment were vendor-provided.

For potable water treatment, additional equipment would be required to treat the DSBR rejectate effluent, namely a bag filtration system and UV disinfection unit, and associated piping, controls and pumps. Costs are shown in Appendix H, and are mostly based on scaling down costs from vendor estimates for similar equipment for 500 gpm (as specified for secondary treatment for DSBR, CSTR, and FBR).

O&M costs (non-potable) include electricity to pump water through NF modules (from vendor-provided estimate for pump hp), operating labor (assumed to be equal to DSBR/CSTR – mainly due to rejectate treatment management), maintenance (e.g., for membrane/pump replacement – not assumed as 5 percent of capital cost – from vendor estimate), performance (based on weekly influent/effluent analyses for ClO_4^- and NO_3^-) and discharge monitoring (assumed identical to all other treatments), and rejectate management. Rejectate treatment costs were calculated as follows (for 500-gpm):

- Costs for chemicals (acetate/nutrients), performance monitoring and media replacement/disposal assumed identical to DSBR (treating same flow rate as NF unit)
- Labor – assumed no additional required (captured in labor cost for NF system operation)
- Maintenance – assumed to be 5 percent of 60-gpm rejectate unit equipment cost
- Power consumption estimated as about 17.5 kW total for process pumps, mixers, and appurtenances

Scale-up O&M costs for 1,000- and 2,000-gpm NF systems are shown in Appendix H.

For potable water applications, additional O&M costs are associated with operation of a process pump, filter bag change-outs, and maintenance. Assumptions are detailed in Appendix H.

Appendix H also shows the expected 20-year life cycle costs for NF for 500-, 1,000- and 2,000-gpm for potable and non-potable applications, based on 50 mg/L of influent nitrate and 10 mg/L of influent ClO_4^-. These costs range from $10.7 million for a 500-gpm NF system (non-potable) to $38.3 million for a 2,000-gpm system (potable). These costs are higher than all biological systems except potable water scenarios where acetate is used as a substrate, but lower than IE.

5.6.4 Cost Comparison Summary

The tables below summarize the pertinent costs for the various ClO_4^- treatment technologies evaluated for 500-, 1,000-, and 2,000-gpm flow rates, and for potable and non-potable treatment scenarios. The table clearly shows that biological treatment is the most cost-effective for all

scenarios if an economical substrate such as methanol is used. It should be noted that these costs do not include pilot testing, which is advisable for establishing scale-up design parameters for each technology type.

5.6.4.1 Non-Potable Water Applications

Non-potable Applications	Lump Sum Capital Cost ($)	Amortized Capital Cost – ($/year)	O&M Cost – ($/year)	O&M Cost – ($/year)	Total Cost ($/acre foot)	Total Cost ($/acre foot)
500-gpm						
DSBR	506,494	47,809	355,595 (Acetate)	203,776 (Methanol)	500 (Acetate)	311 (Methanol)
CSTR	1,391,500	131,348	N/A	254,023 (Methanol)	N/A	416 (Methanol)
FBR	791,151	74,679	382,128 (Acetate)	233,980 (Methanol)	566 (Acetate)	383 (Methanol)
IE	378,012	35,682	807,709		1,046	
NF	1,099,720	103,806	433,651		666	
1,000-gpm						
DSBR	978,556	92,369	641,880 (Acetate)	335,438 (Methanol)	455	265
CSTR	1,916,096	180,866	N/A	371,385 (Methanol)	N/A	342
FBR	1,124,656	106,160	642,113 (Acetate)	351,641 (Methanol)	464	284
IE	621,982	58,711	1,539,359		991	
NF	2,071,734	195,557	777,902		604	
2,000-gpm						
DSBR	1,905,463	179,862	1,221,078 (Acetate)	631,408 (Methanol)	434	246
CSTR	2,407,295	227,232	N/A	568,675 (Methanol)	N/A	247
FBR	2,100,468	198,269	1,159,973 (Acetate)	589,673 (Methanol)	421	244
IE	1,172,312	110,658	3,030,659		974	
NF	4,139,763	390,764	1,435,788		566	

5.6.4.2 Potable Water Applications

Potable Applications	Lump Sum Capital Cost ($)	Amortized Capital Cost – ($/year)	O&M Cost – ($/year)	O&M Cost – ($/year)	Total Cost ($/acre foot)	Total Cost ($/acre foot)
500-gpm						
DSBR	984,424	92,923	468,535 (Acetate)	316,317 (Methanol)	696 (Acetate)	507 (Methanol)
CSTR	1,831,797	172,811	N/A	366,964 (Methanol)	N/A	669 (Methanol)
FBR	1,173,495	110,770	495,069 (Acetate)	346,920 (Methanol)	751 (Acetate)	567 (Methanol)
IE	378,012	35,682	826,169		1,069	
NF	1,175,974	110,941	453,544		700	
1,000-gpm						
DSBR	1,719,960	162,352	856,239 (Acetate)	550,597 (Methanol)	631 (Acetate)	442 (Methanol)
CSTR	2,607,969	246,035	N/A	586,544 (Methanol)	N/A	516 (Methanol)
FBR	1,717,780	162,146	857,272 (Acetate)	566,800 (Methanol)	632 (Acetate)	452 (Methanol)
IE	621,982	58,711	1,539,359		991	
NF	2,183,541	205,994	815,652		633	
2,000-gpm						
DSBR	3,238,629	305,704	1,643,913 (Acetate)	1,036,243 (Methanol)	604	416
CSTR	3,659,356	342,222	N/A	991,510 (Methanol)	N/A	414
FBR	3,167,001	298,942	1,582,808 (Acetate)	1,012,508 (Methanol)	583	407
IE	1,172,312	110,658	3,030,659		974	
NF	4,296,393	405,320	1,507,340		593	

TABLES

TABLE 1

PHASE I STARTUP: ClO_4^- PROBE CALIBRATION STANDARDS AND SYSTEM READINGS

		STD1	STD2	STD3	STD4	STD5	STD6							
		mg ClO_4^-/L									mg ClO_4^-/L			
DAY	Date	1	10	25	50	75	100	EFF R1	EFF R2	EFF R3	R1	R2	R3	Notes on Substrate (Acetate/Ethanol) and ClO_4^- Additions:
1	27-Mar-01	262	215	192	175	164	157	158	156	155	83.9	86.1	87.2	Added $NaClO_4^-$ to each start-up tank to bring concentration to 100 mg ClO_4^-/L
2	28-Mar-01	263	215	189	173	165	157	263	148	241	0.0	94.7	0.0	
3	29-Mar-01	264	214	193	176	163	155	298	151	294	0.0	90.3	0.0	Added $NaClO_4^-$ to R1/R3 start-up tanks to bring concentration to 100 mg ClO_4^-/L (after readings)
4	30-Mar-01	268	214	192	174	164	157	296	153	295	0.0	88.7	0.0	After readings, added Acetate to R1/R3 (to make 1 g Ac-/L)
5	Weekend - No Readings Taken													
6	Weekend - No Readings Taken													
7	2-Apr-01	268	212	191	173	163	157	295	216	296	0.0	19.6	0.0	Added $NaClO_4^-$ to R1/R3 to bring concentrations up to 100 mg ClO_4^-/L. Added 0.5 g/L ethanol (substrate) to R2 (and $NaClO_4^-$ after subsequent reading confirmed disapperance of ClO_4^-).
8	3-Apr-01	265	212	191	173	163	157	158	156	157	82.8	85.0	83.9	
9	4-Apr-01	265	212	191	173	163	157	156	293	154	85.0	0.0	87.3	After readings, added Acetate to R1/R3 (to make 1 g Ac-/L), $NaCl0_4^-$ to R2
10	5-Apr-01	263	212	191	173	163	159	213	280	185	21.2	0.0	53.2	After readings, added Acetate (to make 1 g Ac-/L) to R1/R3, Ethanol (0.5 g/L) to R2
11	6-Apr-01	268	212	191	173	163	157	286	285	282	0.0	0.0	0.0	After readings, added $NaClO_4^-$ to make 100 mg ClO_4^-/L to each start-up tank
12	Weekend - No Readings Taken													
13	Weekend - No Readings Taken													
14	9-Apr-01	267	212	191	173	163	157	179	288	175	59.7	0.0	64.1	After readings, added Acetate (to make 1 g Ac-/L) to R1/R3; R2 off-line
15	10-Apr-01	262	212	191	173	163	157	191	R2 Offline	208	45.8	R2 Offline	26.6	
16	11-Apr-01	264	212	191	173	163	155	193		210	43.5		24.9	Diluted contents of R1/R3 by 1:3, added $NaClO_4^-$ (for 10 mg ClO_4^-/L) and Acetate to each start-up tank
17	12-Apr-01	262	212	191	173	163	159	202		216	33.5		17.5	After readings, added Acetate (to make 1 g Ac-/L) to R1/R3
18	13-Apr-01	266	212	191	173	163	156	288		289	0.0		0.0	Put system in aerobic mode for weekend

Notes:
R1 - perc1ace
R2 - compost/food waste
R3 - JPL isolates

$NaClO_4^-$ sodium perchlorate
Ac- acetate
STD 1-6 ClO_4^- probe standard
Values shown under columns "STD1 - STD6" and "EFFR1 - EFFR3" are in millivolts (mV)
Values shown under columns "R1, R2, and R3" are in mg ClO_4^-/L

TABLE 2
PHASE I FORWARD FLOW TEST ANALYTICAL RESULTS

			INFLUENT						EFFLUENT							
Date	Day	Event	INF ClO_4^- µg/L	INF NO_3^- mg/L	Reactor 1 INF ACE mg/L	Reactor 1 INF TOC mg/L	Reactor 3 INF ACE mg/L	Reactor 3 INF TOC mg/L	Reactor 1 EFF ClO_4^- µg/L	Reactor 1 EFF NO_3^- mg/L	Reactor 1 EFF ACE mg/L	Reactor 1 EFF TOC mg/L	Reactor 3 EFF ClO_4^- µg/L	Reactor 3 EFF NO_3^- mg/L	Reactor 3 EFF ACE mg/L	Reactor 3 EFF TOC mg/L
20-Apr-01	1	1	420	3.1	5.1	23.9	90	38.4	26	20	ND (<1)	15.3	ND (<4)	12	13	24.5
24-Apr-01	5	2	410	20	69	22.4	240	76.3	160	6.6	NS	21.1	110	6.2	NS	60.8
25-Apr-01	6	3	420	15	NS	NS	ND (<1)	15.1	NS	NS	NS	NS	200	9.2	ND (<1)	9.5
26-Apr-01	7	4	410	18	NS	NS	36	11.2	NS	NS	NS	NS	240	11	22	7.2

Notes:

Reactor 3 was offline after April 24, 2001

NS	Not Sampled
ND	Not Detected (Below Reporting Limit)
mg/L	milligrams per liter
mg/L	micrograms per liter
INF	influent
EFF	effluent
ACE	acetate
TOC	total organic carbon

Nitrate results are reported as mg/L NO_3^-

TABLE 3

PHASE II STARTUP PARAMETERS

		REACTOR 1				REACTOR 2				
DATE	TIME	pH	Temp (^oC)	Cond (dS/m)	mg $Cl0_4^-$/L [1]	pH	Temp (^oC)	Cond (dS/m)	mg ClO_4^-/L [1]	Comments
8/15/2002	1:00pm	7.01	25.9	3.42	-	7.01	25	3.3	-	Inoculated R1 and R2 with JPL isolates @1115
8/15/2002	1:15pm	-	-	-	109.6	-	-	-	61	Initial measurement after addition of ClO_4^-
8/15/2002	3:45pm	6.94	29.2	3.77	75.7	7.05	27.6	3.5	71.3	
8/16/2002	7:50am	6.88	24.7	3.42	1.45	6.98	24.4	1.71	<1	Time reflects time samples were collected, not measured
8/16/2002	11:10am	-	-	-	47.7	-	-	-	33.6	Added 109.5 g $NaClO_4$ (to make 100 mg/L ClO_4^-) to each system after taking readings
8/16/2002	4:20pm	6.8	30.7	3.7	< 1	6.88	29.4	3.76	<1	Time reflects time samples were collected, not measured
8/16/2002	4:45pm									Added 109.5 g $NaClO_4^-$ to each system after taking readings
										Turned on aerator over weekend to add O_2 to system (both reactors)
8/19/2002	7:10am	7.61	23.2	1.82	<1	7.8	23.7	3.29	<1	Time reflects time samples were collected, not measured
										Added 2 x 109.5 g $NaClO_4^-$ to each system after taking readings
8/19/2002	9:35am	7.59	23.2	3.62	62.8	7.59	21.9	4.78	69.9	Time reflects time samples were collected, not measured
8/19/2002	4:15pm	7.41	25.8	2.06	<1	7.59	25.6	1.94	<1	Added 1.3 kg NaAc (hyd) to each reactor (to make 0.5 g/L Ac^-) and turned on aerator
8/20/2002	6:45am	7.43	23.8	3.72	-	7.61	23.7	3.52	-	Observed foaming in R2 start-up tank
8/20/2002	7:20am									Added 650g NaAc (hyd) to R1 (to make 0.25 g/L)
8/20/2002	7:40am									Added 650 g NaAc (hyd) to R2
8/20/2002	3:30pm	7.54	25.5	4.21	-	7.67	25.2	4.09	-	
8/21/2002	7:00am	7.67	21.7	3.96	-	7.82	22	3.82	-	Shut off recirculation for 1.5 hours (to build up biomass)
8/21/2002	4:30pm	7.76	26.5	4.39	-	7.89	25.3	4.11	-	
8/22/2002	8:45am									Added 650 g NaAc (hyd) to each reactor
8/22/2002	9:00am	7.76	21.2	3.9	-	7.84	21.2	3.7	-	Shut off recirculation for 1.5 hours (to build up biomass)
8/22/2002	4:15pm									Added 126 g $NaClO_4$ (to make 100 mg/L ClO_4^-) to each reactor; shut off aerator
8/22/2002	5:10pm	7.63	27.5	2.7	75	7.81	26.4	4.64	10	Time reflects time samples were collected, not measured
8/23/2002	7:45am	7.53	21.9	2.37	<1	7.65	21.6	4.2	<1	Time reflects time samples were collected, not measured
8/23/2002	9:15am-10:40am									Shut off recirculation; turned on aerator

Note: All readings taken from effluent of final reactor
R1 - Reactor 1
R2 - Reactor 2
[1] From ClO_4- probe, using standard regression curve
-Not measured
dS - decisiemens
NaAc - sodium acetate
$NaClO_4$ - sodium perchlorate
Ac^- - acetate ion
hyd - hydrated (form of acetate salt)
g - grams

TABLE 4
SUMMARY OF KEY OPERATIONAL PARAMETERS

	TEST 1: DURATION - 14 DAYS		TEST 2: DURATION - 18 DAYS	
	Reactor 1	Reactor 2	Reactor 1	Reactor 2
Flow (gpm)	2.1	2.8	1	0.9
Residence Time (avg.) (minutes)	48	36	100	111
Acetate Concentration in Final Flow (mg Ac-/L)	100-300	100-300	100	100
Pressure Drop (psi)	No Data	No Data	15-21	5-16
Media	Hydroxyl-PAC	Sponges w/ Celite	Hydroxyl-PAC	Hydroxyl-PAC
Bacteria	JPL Isolates	JPL Isolates	JPL Isolates	Perc1ace

Notes:
Each Reactor consisted of two pressure vessels in series
Ac^- - acetate ion

TABLE 5
SUMMARY OF KEY RESULTS

	TEST 1: DURATION - 14		TEST 2: DURATION - 18	
	Reactor 1	Reactor 2	Reactor 1	Reactor 2
Flow (gpm)	2.1	2.8	1	1
Influent Perchlorate Concentrations (range) (mg/L)	5.8-6.6		8.3-10.6	
Influent Nitrate Concentrations (range) (mg NO_3^-/L)	45-48.4		47-48	
Perchlorate Removal Efficiency (range) (%)	19.7-75.9[1]	22.7-75.7	39.6-96[2]	58-99.9
Nitrate Removal Efficiency (range) (%)	89.1-98[3]	69-98[3]	98[3]	89.1-98[3]

Notes:
Each Reactor train consisted of two pressure vessels in series
Removal efficiencies for Test 2 do not include final polishing reactor.
[1] Disregarding the sample result from September 6, which showed 99.9% removal (sample was re-run due to lab QC error -- see Appendix F).
[2] Range is 73-96 if first sampling event (1 day after start-up) is disregarded.
[3] 98% is based on removal to ND (<0.44 or 0.88 mg/L)

FIGURES

T-1
INFLUENT TANK
1,100 GAL POLY TANK
INLET/OUTLET: 2" FPT

P-1
INFLUENT PUMP
0.5HP CENTRIFUGAL
INLET/OUTLET: 0.75"FPT

FM-1,-2,-3
FLOWMETER
TURBINE TYPE
RANGE: 0-3 GPM
INLET/OUTLET: 1"FPT

T-2,-7,-11
SUBSTRATE
INJECTION TANK
150 GAL
OUTLET: 0.5"FPT

MP-1,-2,-3,-4
SUBSTRATE
METERING PUMPS
8 GPH
INLET/OUTLET: 0.5"FPT

T-3
NUTRIENT
INJECTION TANK
150 GAL
INLET/OUTLET: 0.5"FPT

T-5,-8,-12
FLASH MIXING TANK
55 GAL
INLET/OUTLET: 1"FPT

P-2,-3,-4
MIXER PUMP
0.5HP CENTRIFUGAL
INLET/OUTLET: 0.75"FPT

T-4,-9,-13
INOCULUM TANK
50 GAL
INLET/OUTLET: 1"FPT

T-6,-10,-14
BIOREACTOR TANK
55 GAL
INLET/OUTLET: 1.25"FPT

T-15
INTERMEDIATE TANK
165 GAL
INLET/OUTLET: 1"FPT

FM-4,-5,-6
FLOWMETER
BLOCK STYLE
RANGE: 1.2-12 GPH
INLET/OUTLET: 0.25"FPT

P-5
INTERMEDIATE PUMP
0.5 HP CENTRIFUGAL
INLET/OUTLET: 0.75"FPT

PF-1
PARTICULATE FILTER
25 MICRON
INLET/OUTLET: 1"FPT

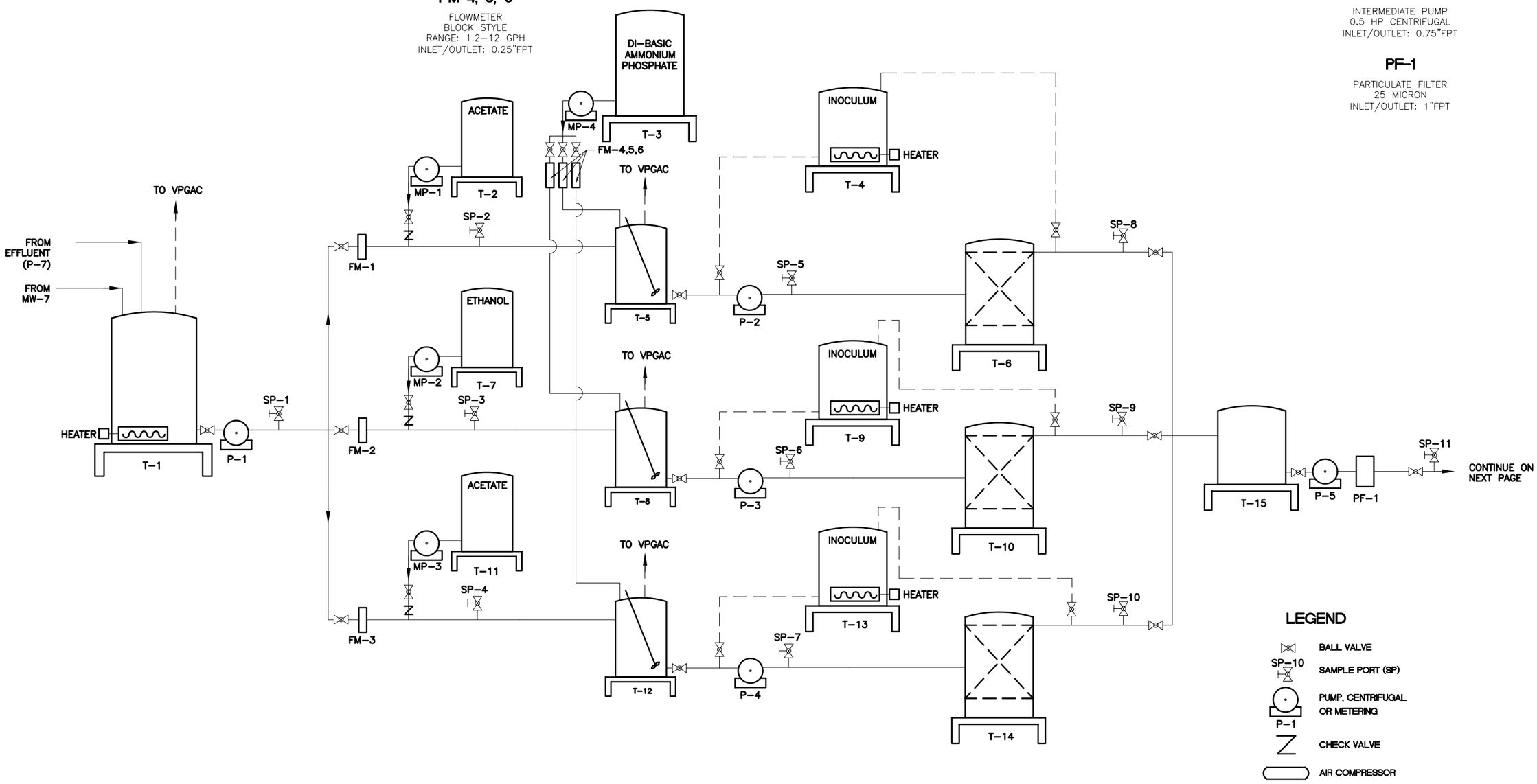


Figure 1 (Page 1 of 2)
PERCHLORATE BIOREACTOR PROCESS FLOW DIAGRAM (PHASE I)

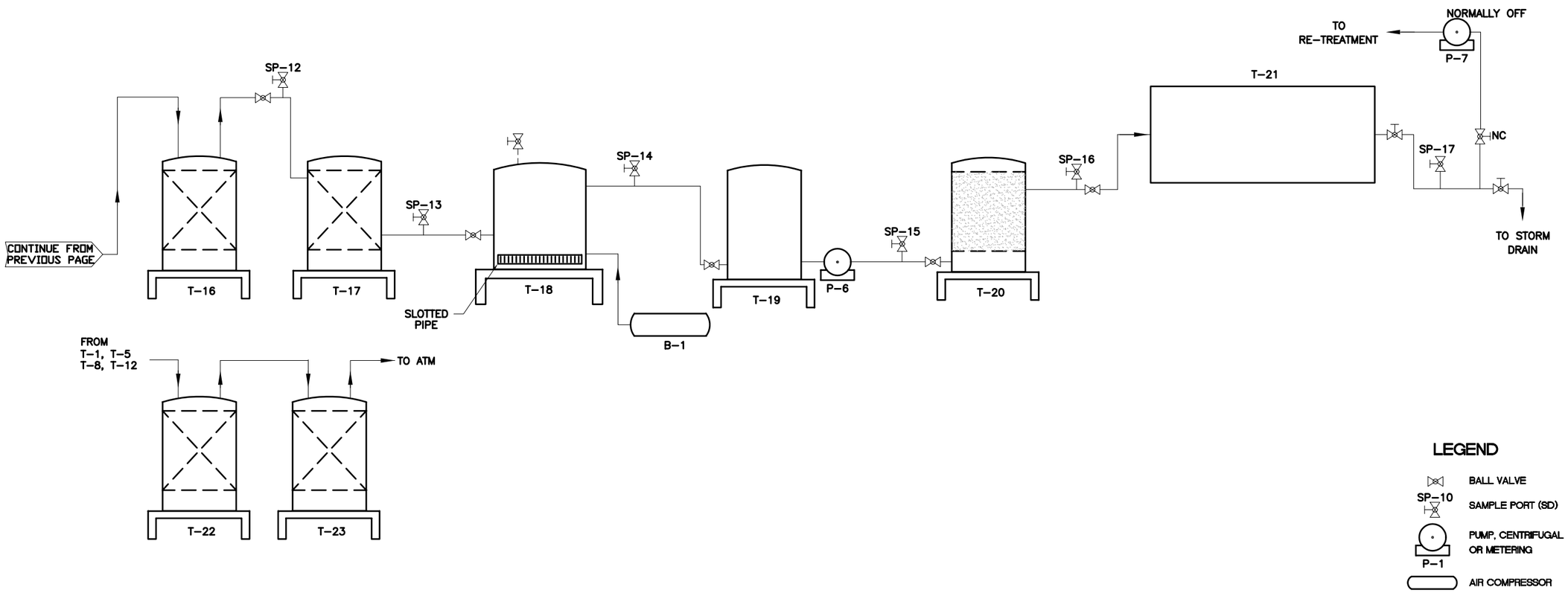
JPL

FOSTER WHEELER ENVIRONMENTAL CORPORATION

DESIGN BY: VH/DT | CHECKED BY: DT | APPROVED BY: VH | DATE: 01/15/03

I:\2423-NFESC\DWG\PILOT REPORT\FIG 1.DWG
PLOT/UPDATE: JUN 30 2003 16:17:39

T-16	T-17	T-18	B-1	T-19	P-6	T-20	T-21	P-7	T-22,-23
ION EXCHANGE SBA RESIN INLET/OUTLET: 1" FTP	GRANULAR ACTIVATED CARBON 1,000 LBS GAC INLET/OUTLET: 2"FPT	AEROBIC BIOTANK 55 GAL DRUM INLET/OUTLET: 1.25"FPT	AIR COMPRESSOR 4.5 CFM @ 90 PSI INLET/OUTLET: 0.25"FPT	EFFLUENT TANK 165 GAL INLET/OUTLET: 1"FPT	TRANSFER PUMP 0.5 HP CENTRIFUGAL INLET/OUTLET: 0.75"FPT	SAND FILTER 55 GAL INLET/OUTLET: 1.25"FPT	FRAC TANK 21,000 GAL INLET: From Open Top OUTLET: 4"FPT	TRANSFER PUMP 0.5 HP CENTRIFUGAL INLET/OUTLET: 0.75"FPT	VAPOR GAC 55 GAL DRUM 180-LB GAC INLET: 2" FPT OUTLET: 2" FPT



LEGEND

- BALL VALVE
- SP-10 — SAMPLE PORT (SD)
- P-1 — PUMP, CENTRIFUGAL OR METERING
- AIR COMPRESSOR

Figure 1 (Page 2 of 2)
PERCHLORATE BIOREACTOR PROCESS FLOW DIAGRAM (PHASE I)

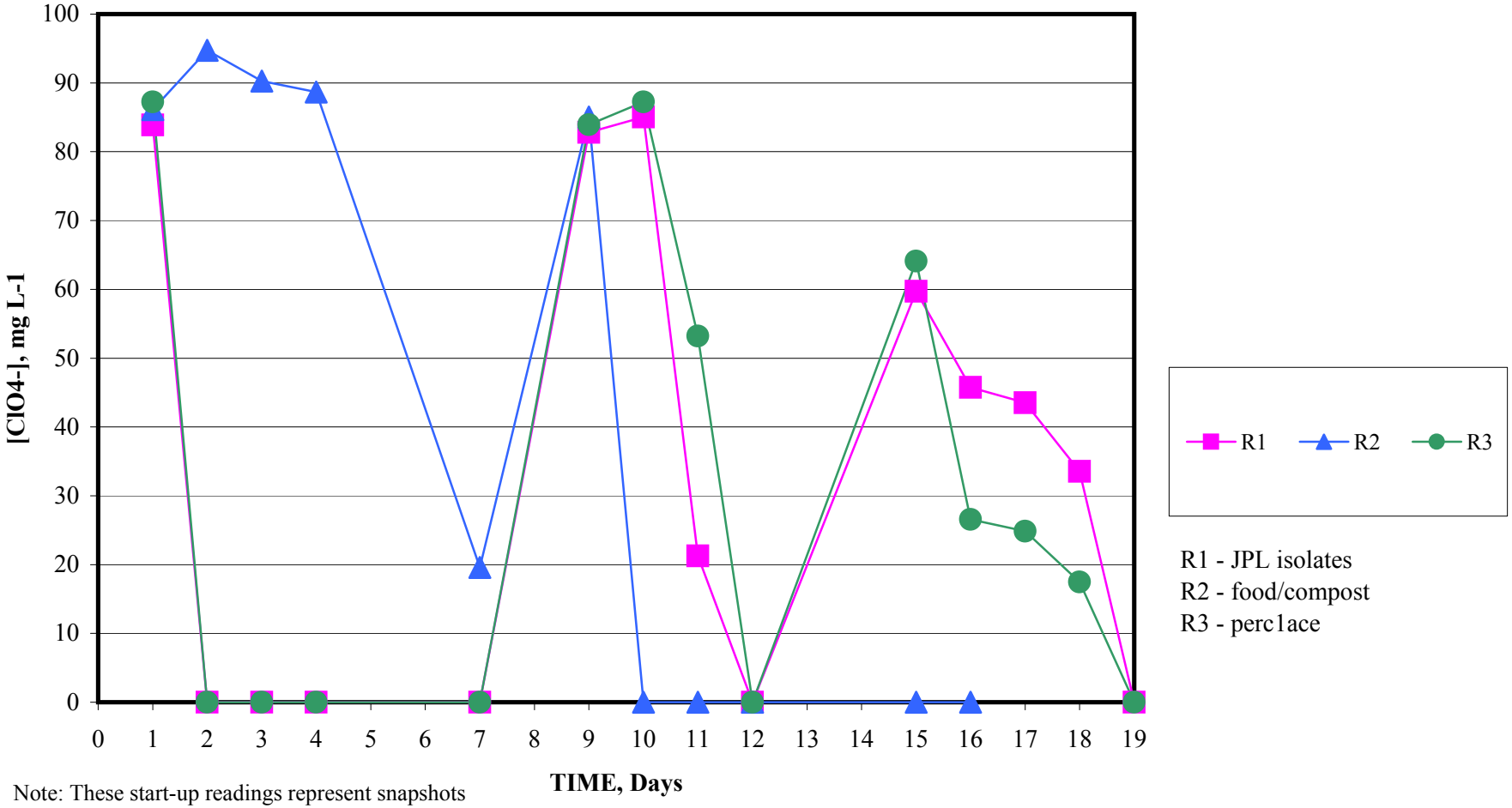
JPL

FOSTER WHEELER ENVIRONMENTAL CORPORATION

DESIGN BY: VH/DT | CHECKED BY: VH/DT | APPROVED BY: VH/ML | DATE: 12/19/02

I:\2423-NFESC\DWG\PILOT REPORT\FIG 1.DWG
PLOT/UPDATE: JUN 30 2003 16:17:39

FIGURE 2
PHASE I STARTUP RESULTS



Note: These start-up readings represent snapshots from each date . They should be interpreted in the context of whether or not substrates or ClO4- were added to each system and when (see Table 1).

T-1
INFLUENT TANK
1,100 GAL POLY TANK
(SAME AS PHASE I)

T-2, T-4
SUBSTRATE FEED
TANK – 150 GAL POLY TANK
(CONNECTED TO
500 GAL STAGING TANK, **T-3, T-5** RESPECTIVELY)

T-6
NUTRIENT FEED TANK,
2 X 150 GALLON
POLY TANKS

T-7A, B/T-8A, B
BIOREACTOR TANKS, 3 FT DIA
X 2 FT HIGH, STEEL

T-9, T-10
INOCULUM (START UP)
TANK, 50 GAL POLY
(SAME AS PHASE I)

T-11
BIOREACTOR "POLISHING" TANK,
3 FT DIA X 2 FT HIGH,
STEEL (TEST 2 ONLY)

T-12, T-15
INTERMEDIATE EQUALIZATION
TANK, 165 GAL POLY
(SAME AS PHASE I)

T-13
LIQUID PHASE GRANULAR
ACTIVATED CARBON VESSEL,
1000 LB GAC
(SAME AS PHASE I)

T-14
ION EXCHANGE UNIT W/
STRONG BASE ANION RESIN
(SAME AS PHASE I)

VPGAC
VAPOR PHASE GRANULAR
ACTIVATED CARBON
(SAME AS PHASE I)

B-1
AIR COMPRESSOR, .5 HP
(USED AS REQUIRED)

P-1, P-2
INFLUENT FEED PUMPS,
VARIABLE SPEED,
.5 HP

P-3, P-4
RECIRCULATION PUMPS,
CENTRIFUGAL, .5 HP

P-5, P-6
CENTRIFUGAL PUMPS,
.5 HP

P-7
RECIRCULATION/RE-TREATMENT
PUMP, .5 HP

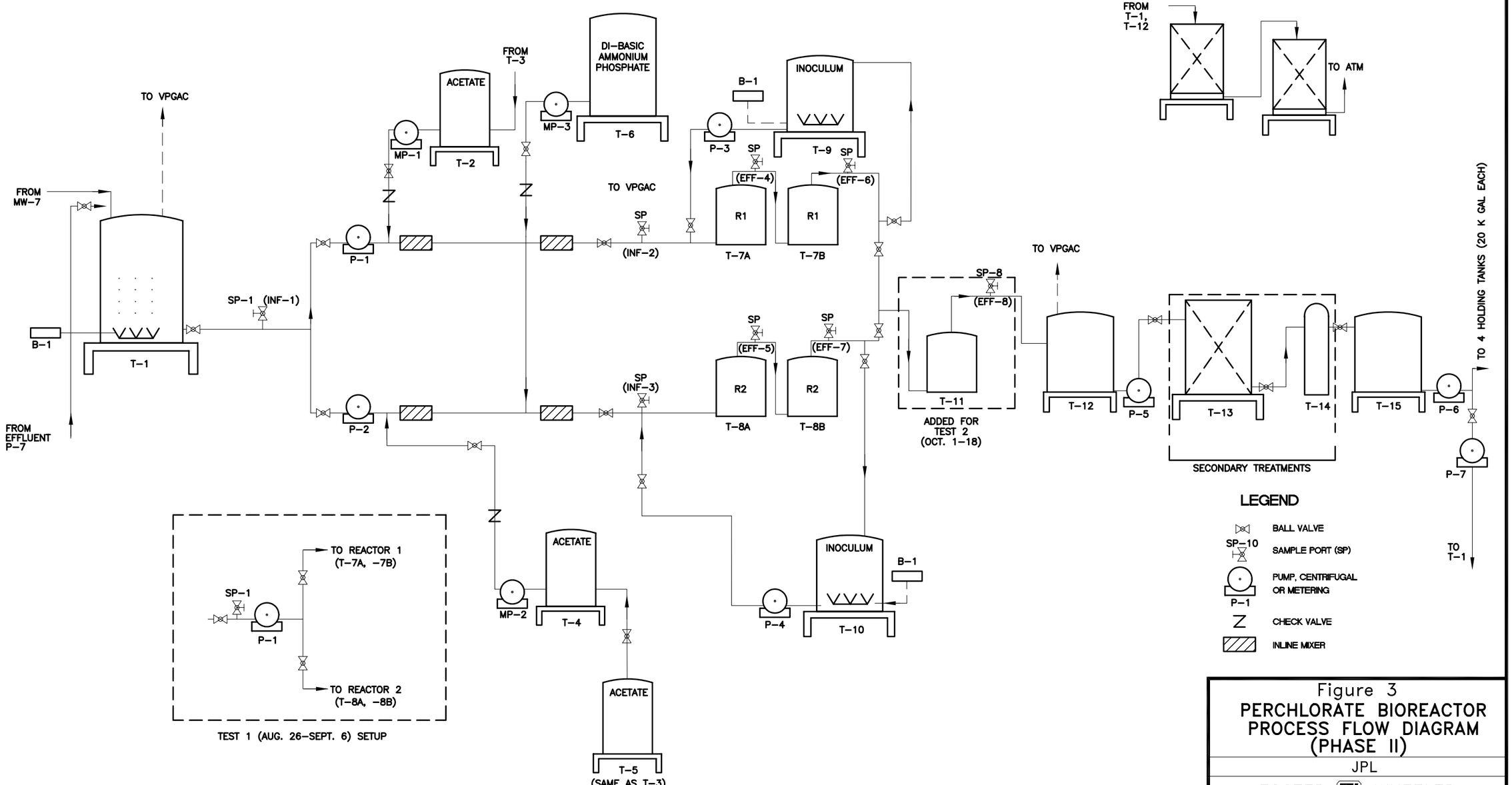


Figure 3
PERCHLORATE BIOREACTOR
PROCESS FLOW DIAGRAM
(PHASE II)

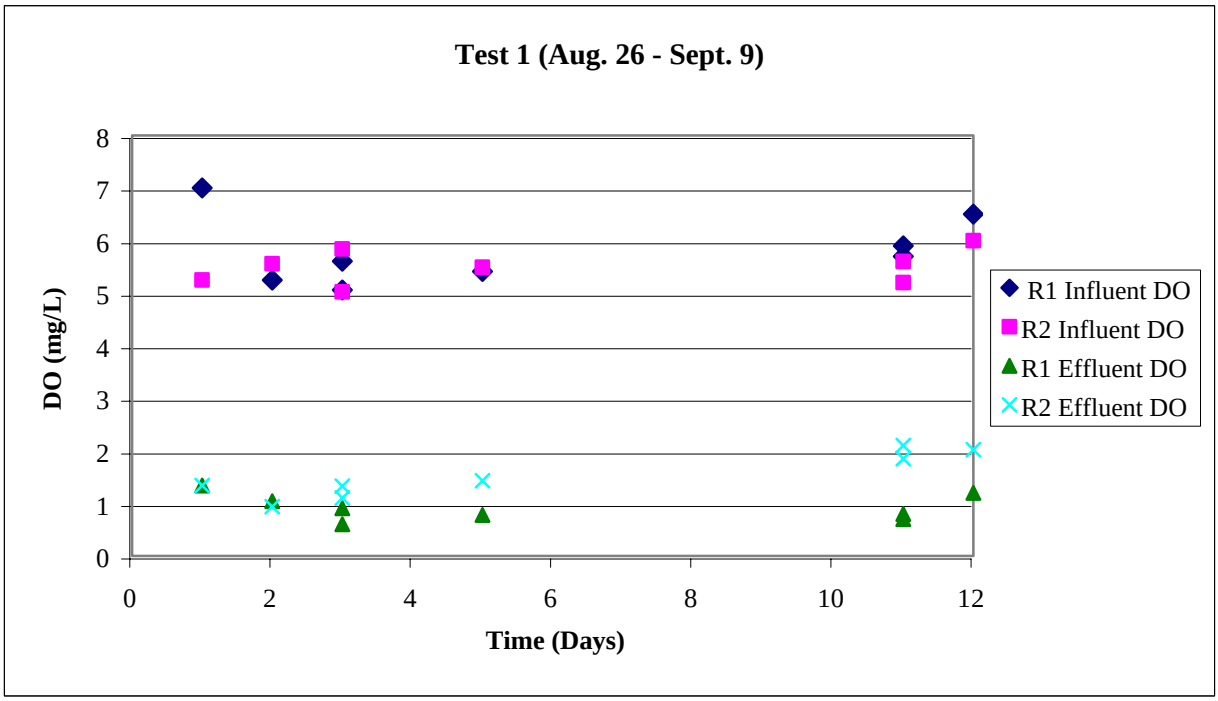
JPL

FOSTER WHEELER
ENVIRONMENTAL CORPORATION

DESIGN BY: VH/DT
CHECKED BY: DT
APPROVED BY: VH
DATE: 01/15/03

I:\2423-NFESC\DWG\PILOT REPORT\FIG 3.DWG
PLOT/UPDATE: JUN 30 2003 09:11:22

FIGURE 4
PHASE II DISSOLVED OXYGEN READINGS



Note:
Readings taken with Horiba U10 for first 5 days, YSI Model 51B for days 11-12. Horiba.
Readings from Days 9-10 not reported due to malfunctioning probe.

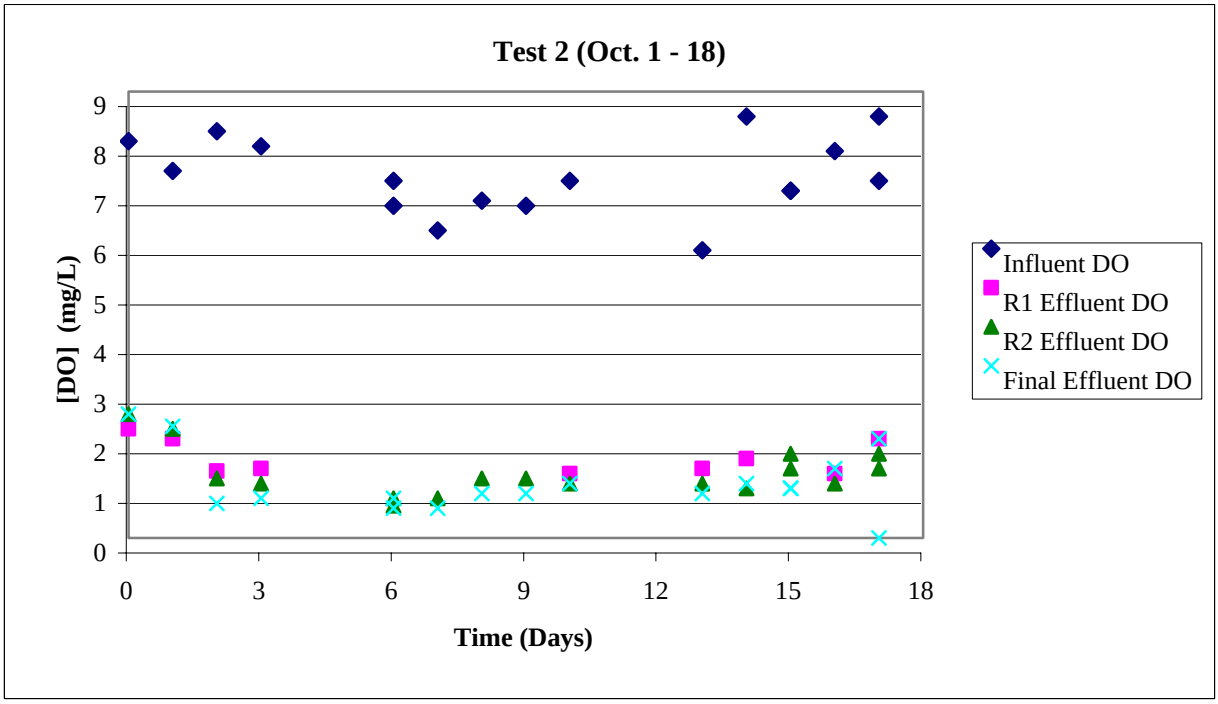
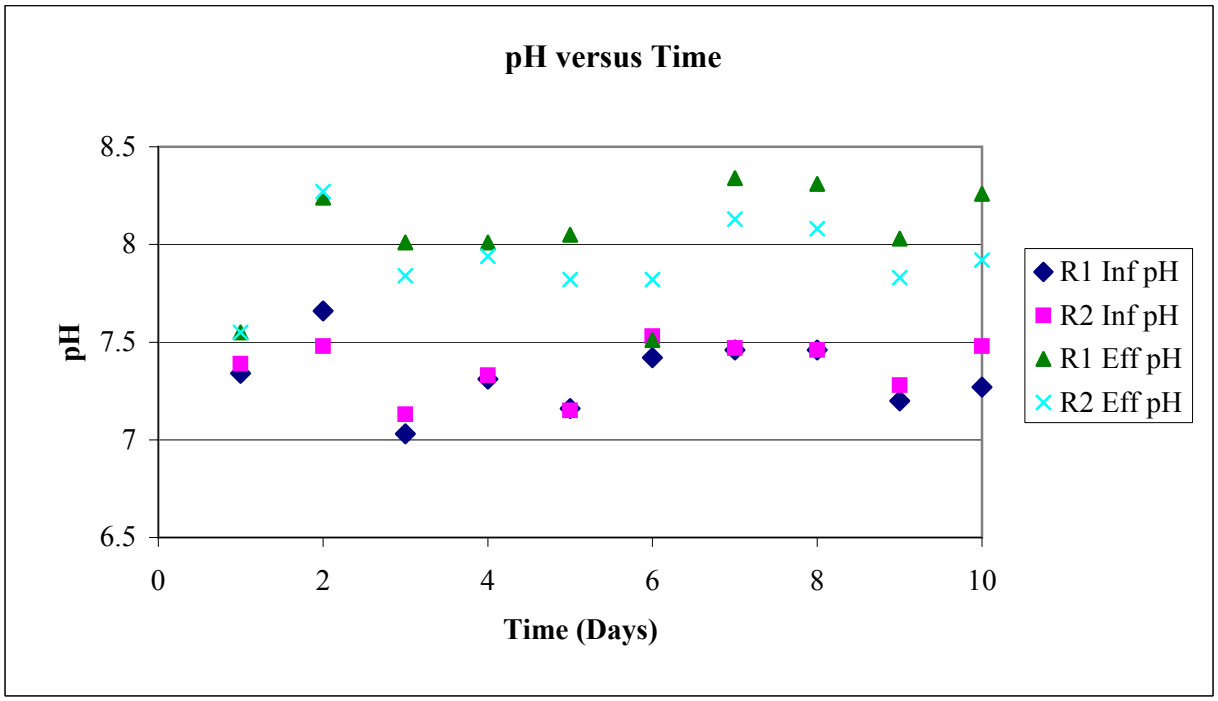


FIGURE 5
pH AND TEMPERATURE TRACKING - PHASE II/TEST 1



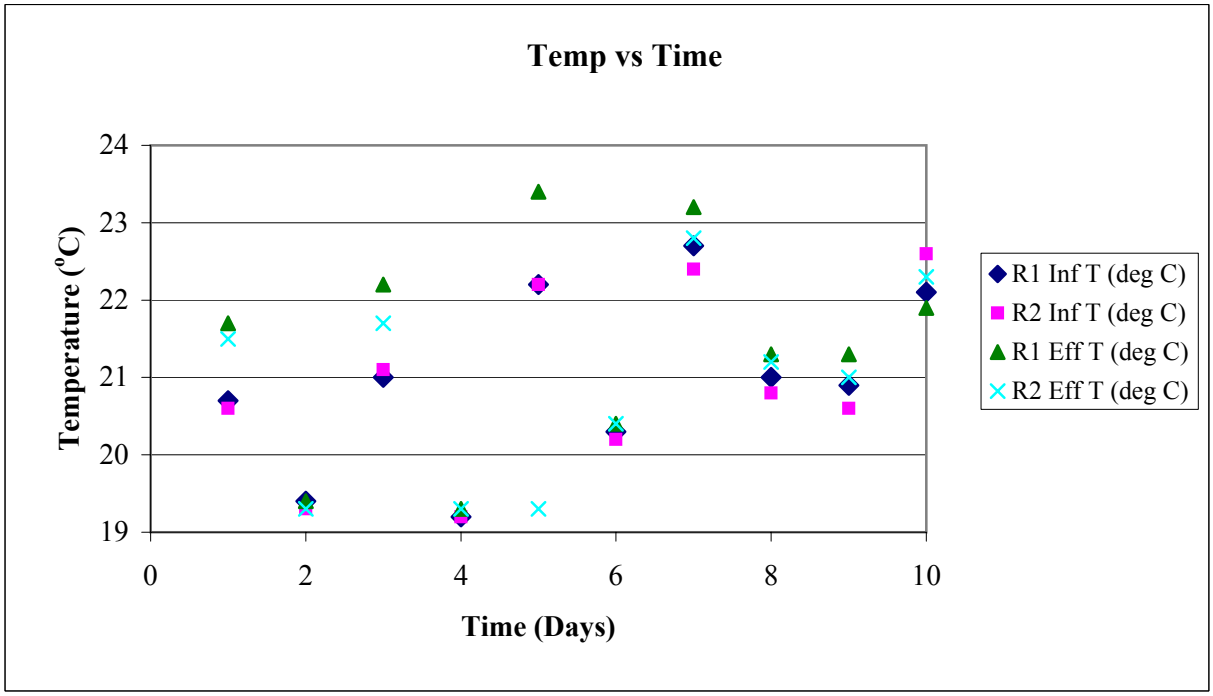
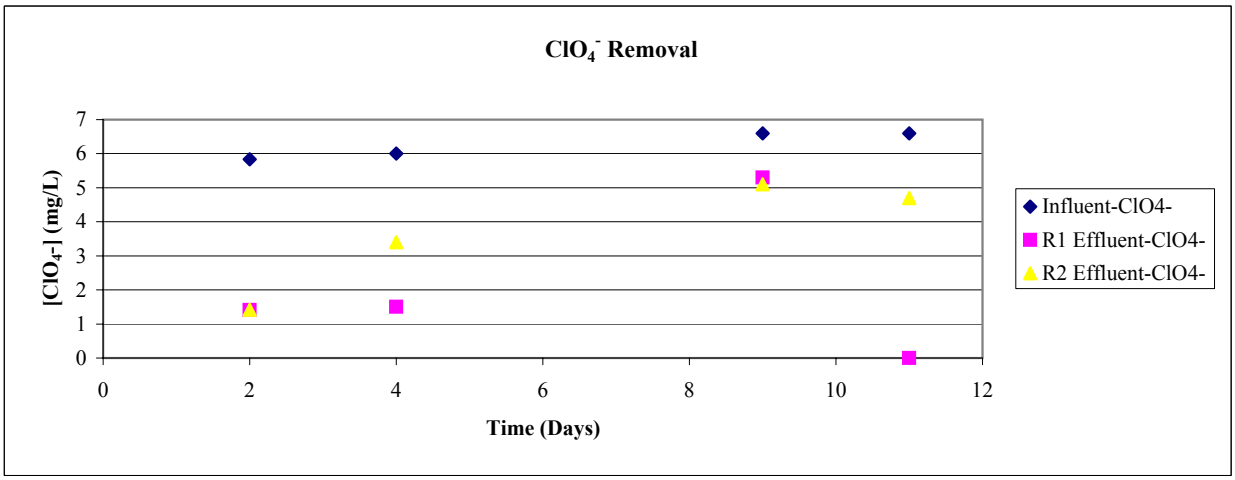
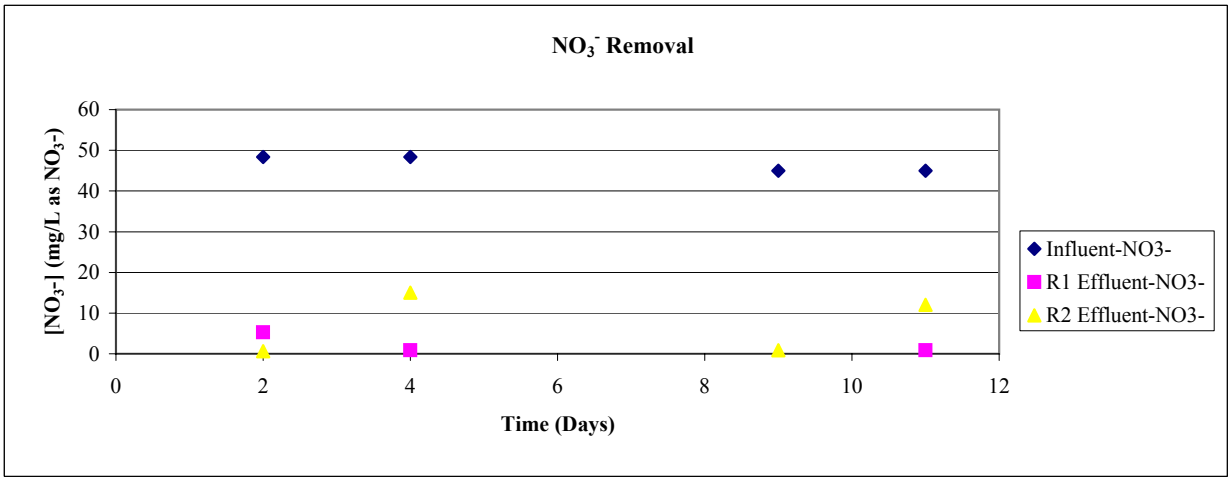
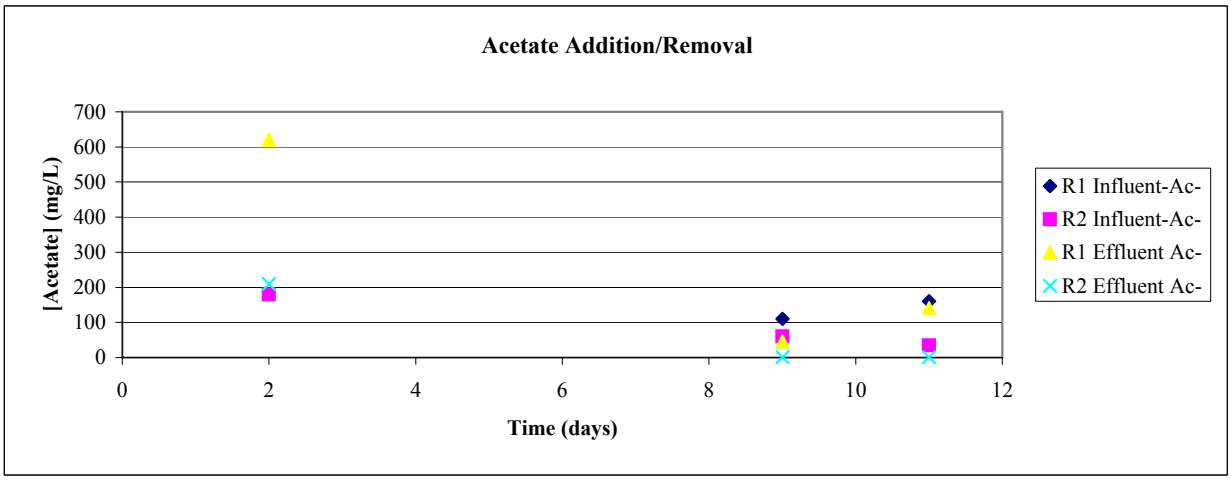


FIGURE 6
ClO_4^-, NO_3^- AND ACETATE REMOVAL, PHASE II/TEST 1

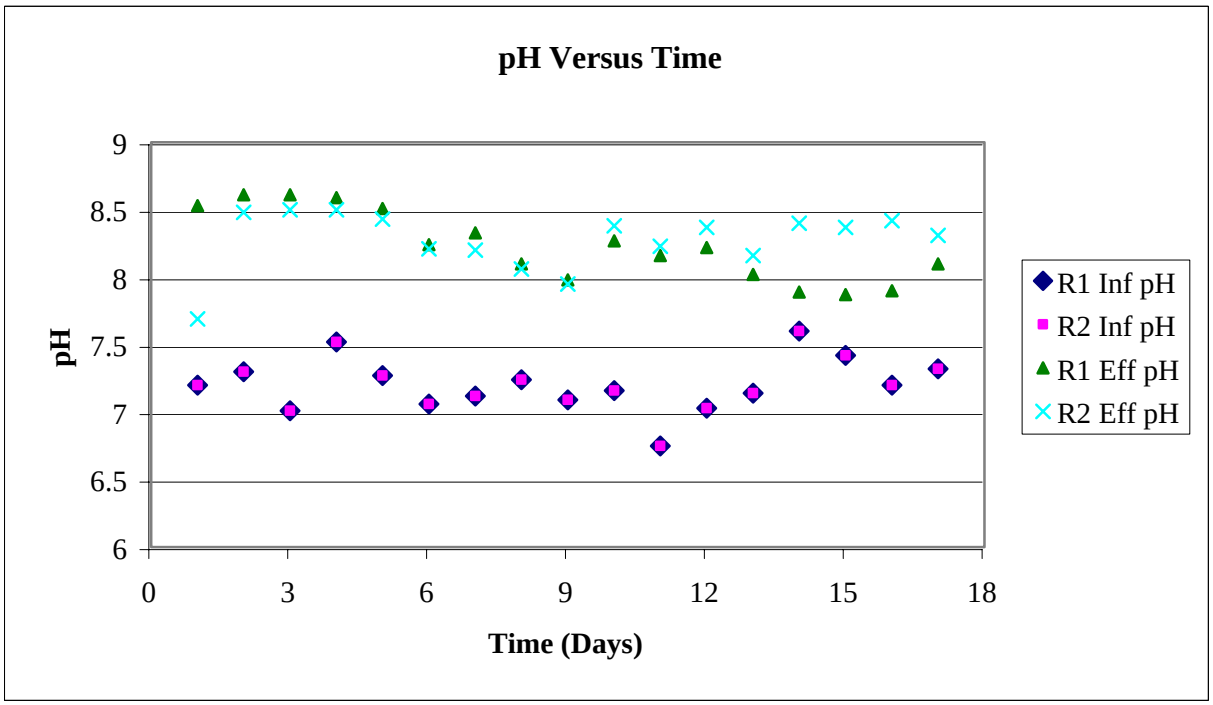






Note: Reactors 1 and 2 both inoculated with JPL isolates.
Avg. Residence Time: 48 minutes for Reactor 1 and 36 minutes for Reactor 2.
Where constituent was not detected, concentration was assumed equal to method reporting limit.

FIGURE 7
pH AND TEMPERATURE TRACKING - PHASE II/ TEST 2



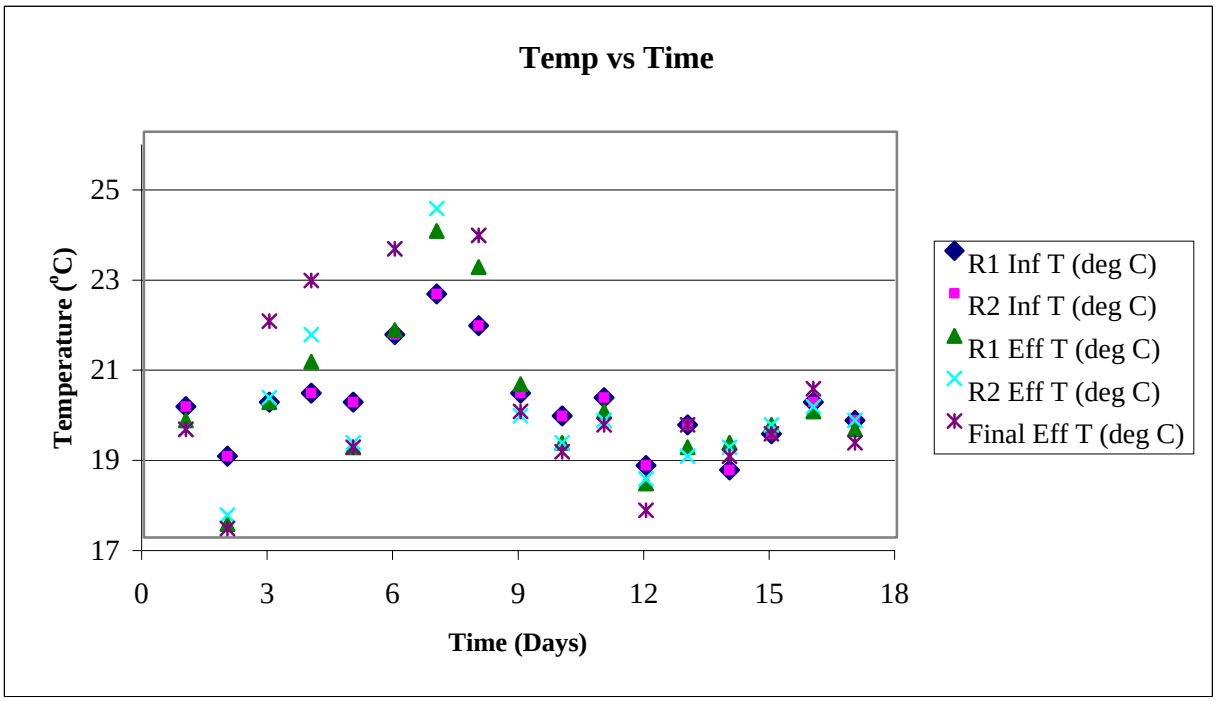
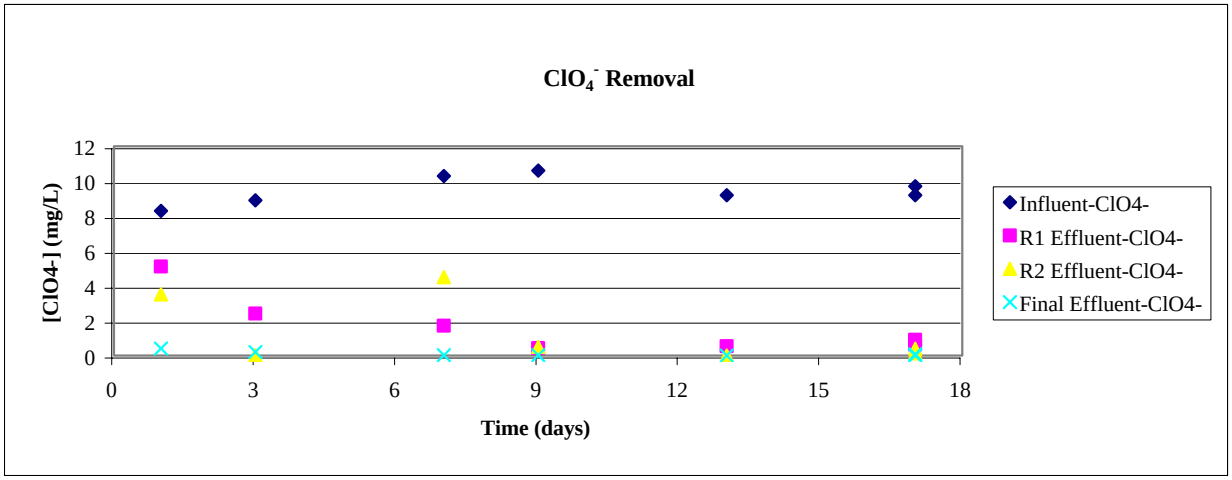
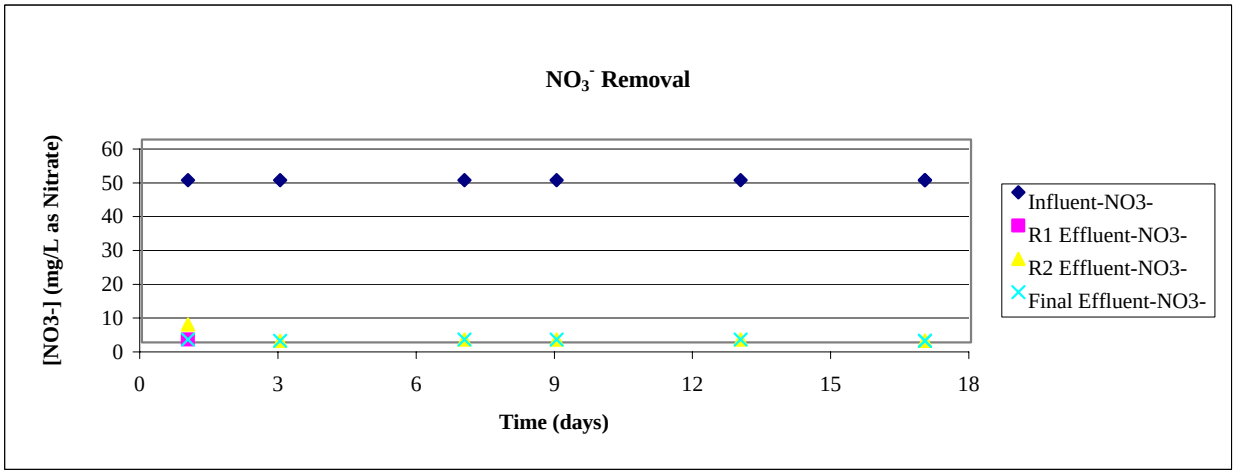
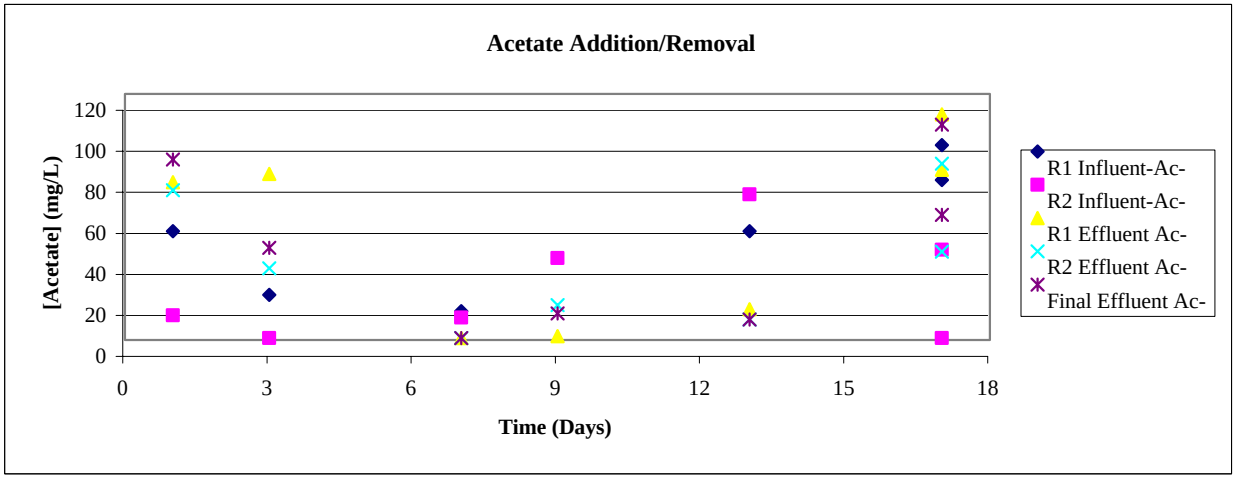


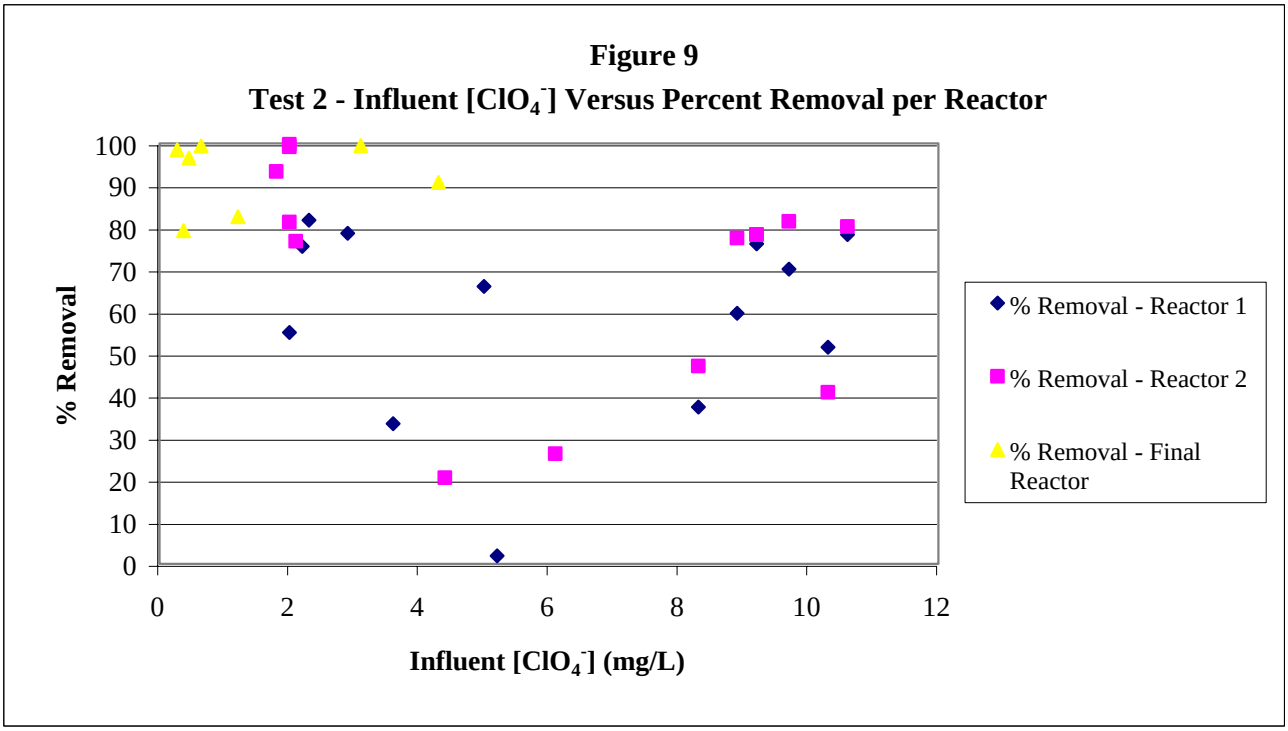
FIGURE 8
ClO_4^-, NO_3^-, AND ACETATE REMOVAL - PHASE II/TEST 2

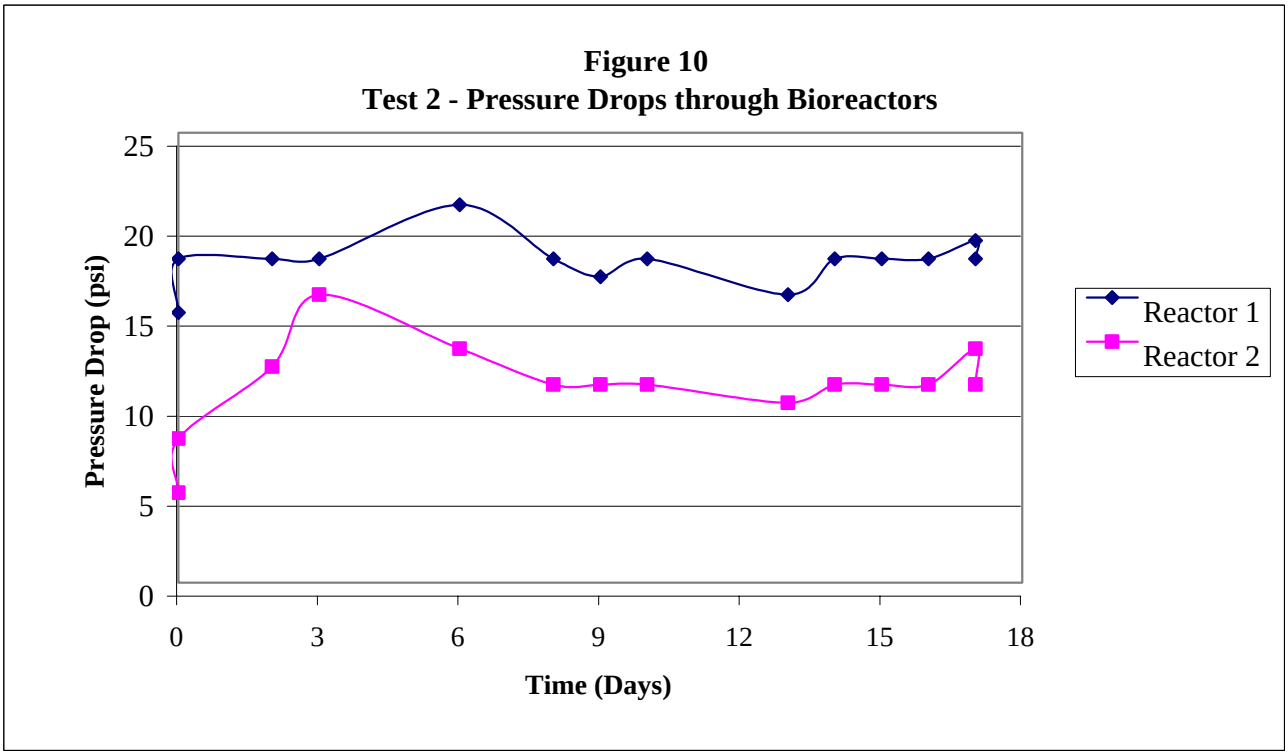






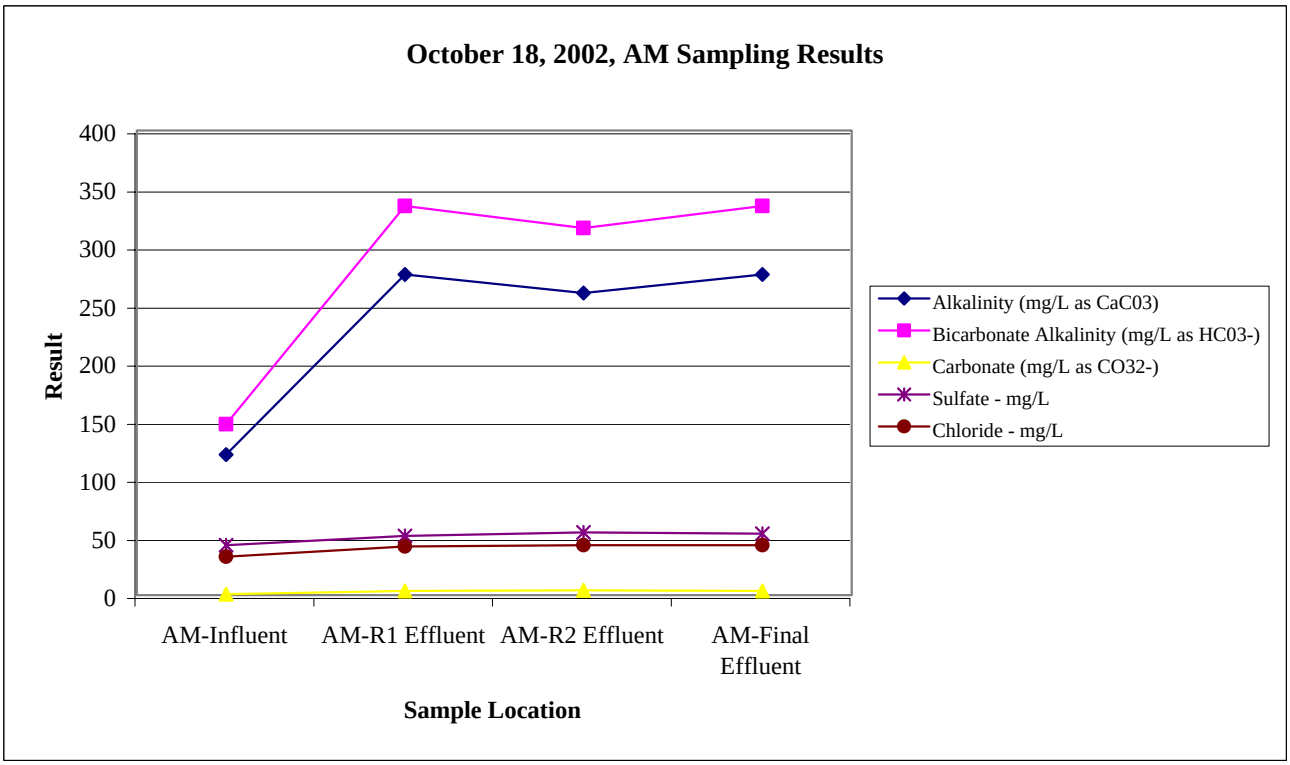
Note: Two data points shown for Day 18 represent two sampling events conducted on that date.
Reactor 1 was inoculated with JPL isolates, Reactor 2 was inoculated with perc1ace, Final Effluent was composite of Reactor 1 and Reactor 2 effluents after treatment in final reactor.
Avg. Residence Time: 100-111 minutes for both reactors (125-136 minutes for final effluent)

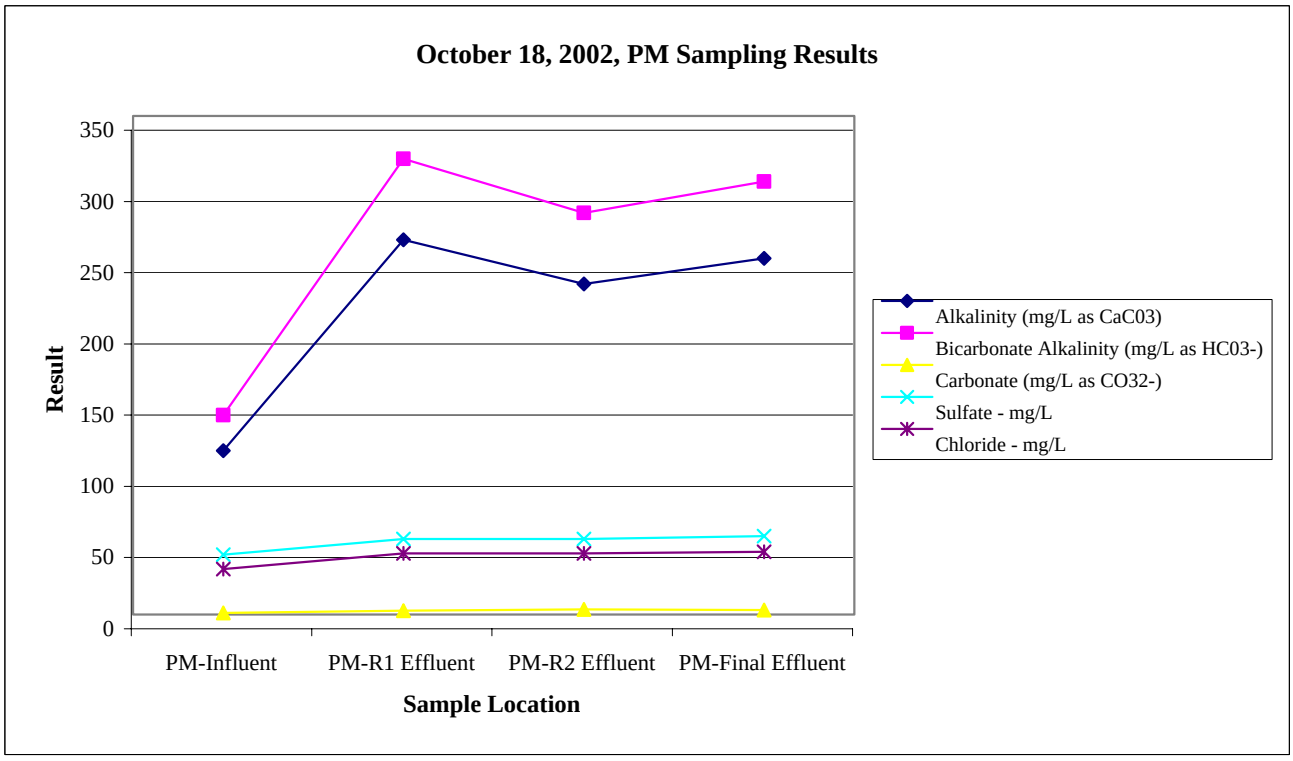




Note:
Assumes pressure at outlet of second vessel in treatment train was 0 psi (may have been 0-2 psi)

FIGURE 11
OCTOBER 18, 2002, SAMPLING:
WATER QUALITY PARAMETERS ANALYTICAL RESULTS

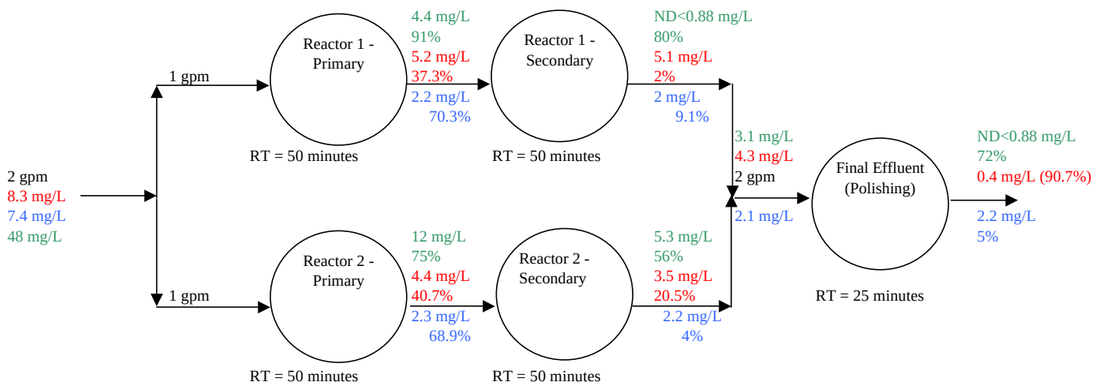




Note: Influent [ClO4-] = 9.7 mg/L

FIGURE 12
ClO_4^-, NO_3^- AND DO REMOVAL TRACKING

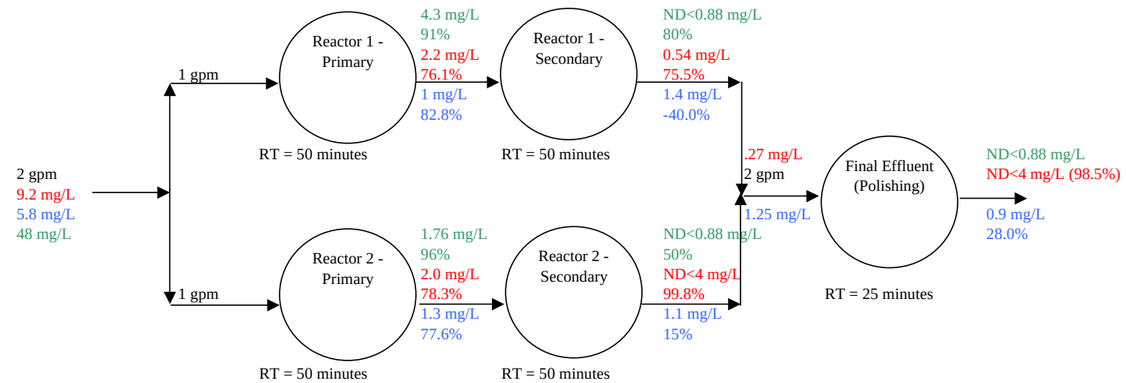
RESULTS FROM 10-2-2002 SAMPLING (PRE-ACCLIMATION)



Legend:

8.3 mg/L	[ClO_4^-]	37.3%	% removal (same for all species)
7.4 mg/L	[DO]		
48 mg/L	[NO_3^-]		
1 gpm	flowrate (approx.)		
RT	residence time (approx.)		

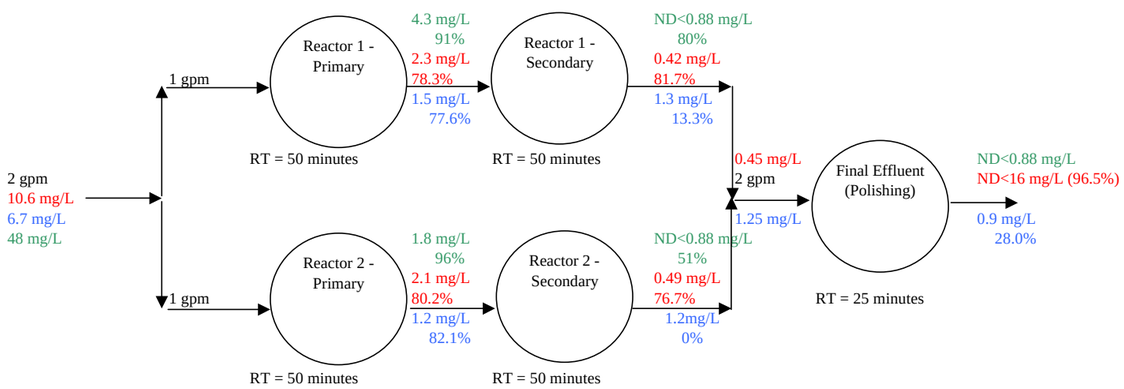
RESULTS FROM 10-14-2002 SAMPLING



Legend:

10.6 mg/L	[ClO_4^-]	76.1%	% removal (same for all species)
6.7 mg/L	[DO]		
48 mg/L	[NO_3^-]		
1 gpm	flow rate (approx.)		
RT	residence time (approx.)		

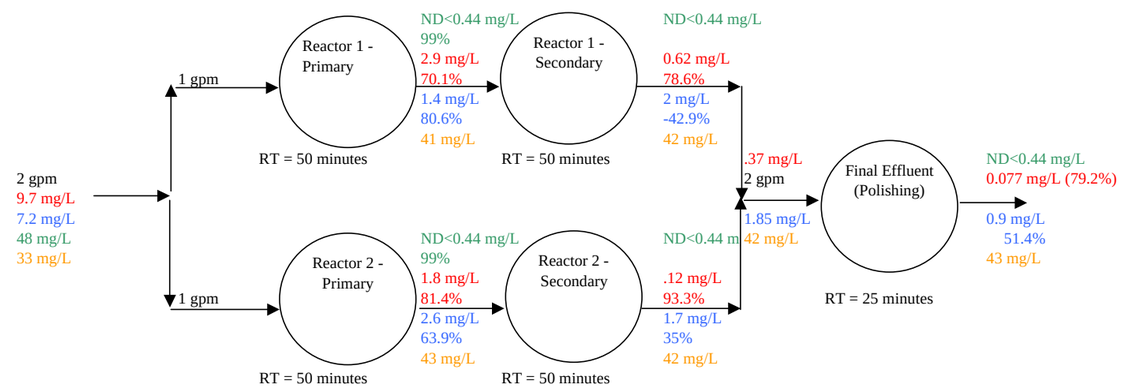
RESULTS FROM 10-10-02 SAMPLING



Legend:

10.6 mg/L	[ClO_4-]	78.3%	% removal (same for all species)
6.7 mg/L	[DO]		
48 mg/L	[NO_3^-]		
1 gpm	flow rate (approx.)		
RT	residence time (approx.)		

RESULTS FROM 10-18-02 SAMPLING #1



Legend:

10.6 mg/L	[ClO_4^-]	70.1%	% removal (same for all species)
6.7 mg/L	[DO]		
48 mg/L	[NO_3^-]		
33 mg/L	[Cl-]		
1 gpm	flow rate (approx.)		
RT	residence time (approx.)		

FIGURE 13
USE OF ACETATE BY THE JPL ISOLATES AND PERCLACE AND DISAPPEARANCE OF AVAILABLE ELECTRON ACCEPTOR

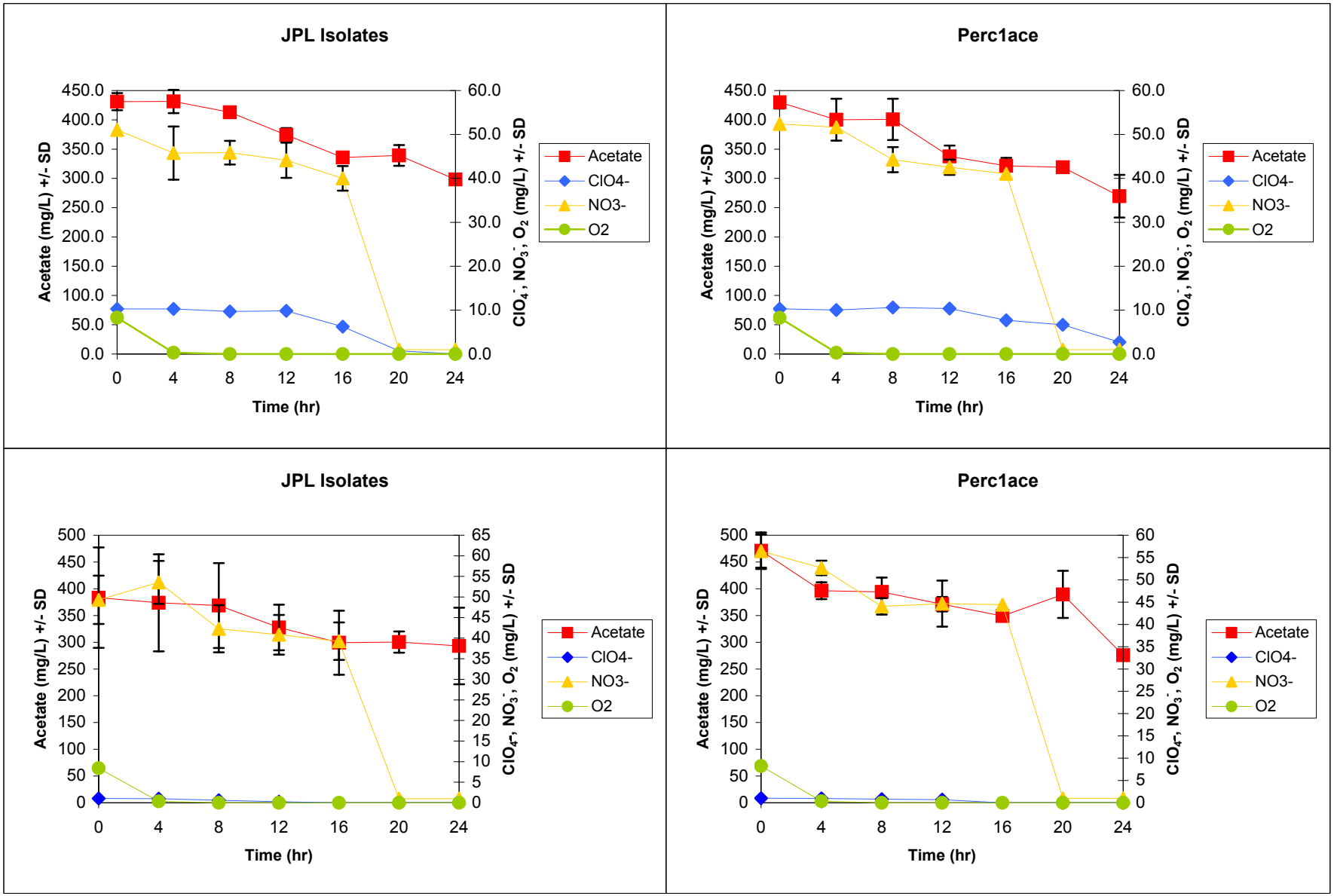
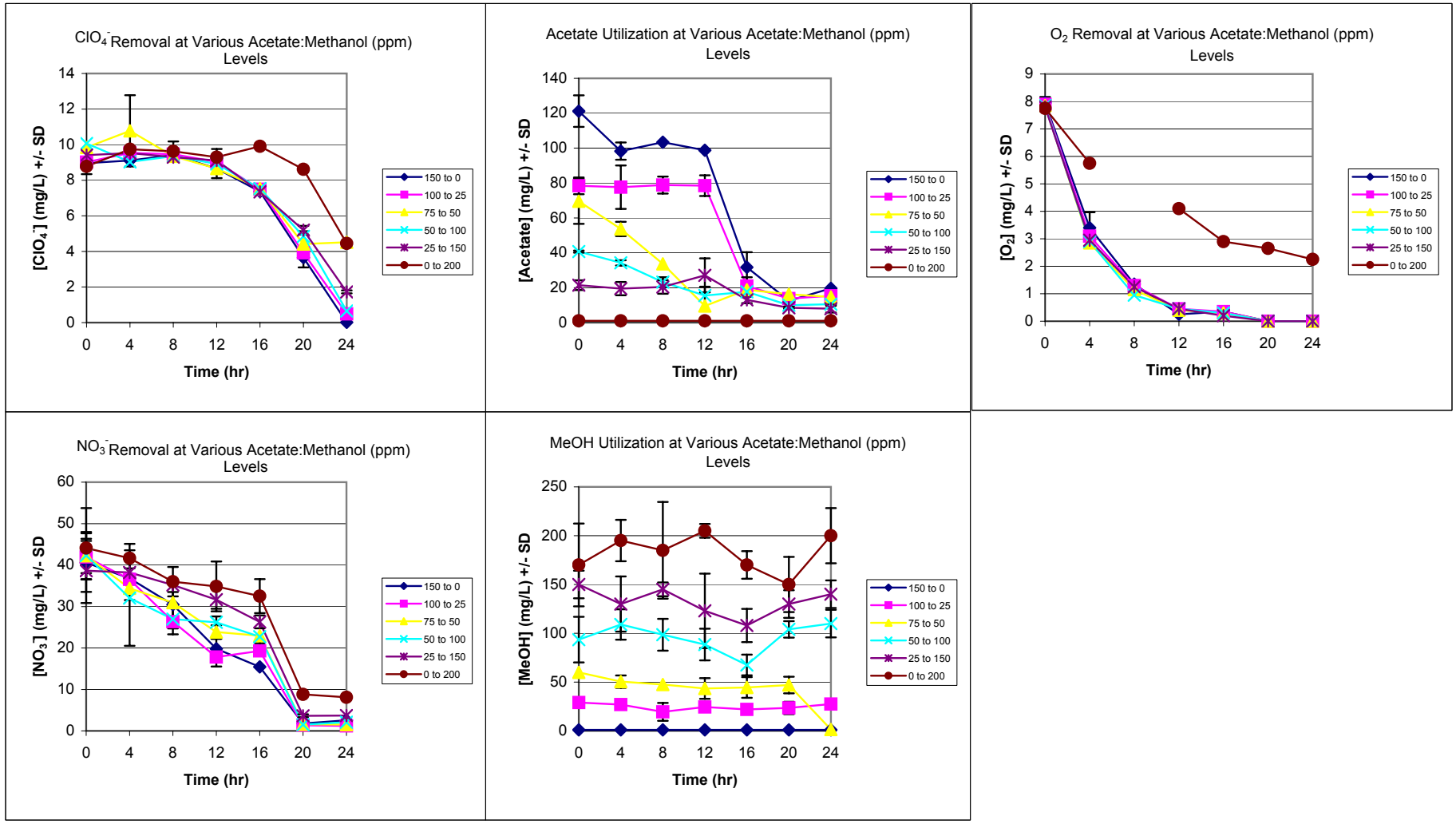


FIGURE 14
USE OF ACETATE AND METHANOL AT VARIOUS RATIOS BY THE JPL ISOLATES AND DISAPPEARANCE OF AVAILABLE ELECTRON ACCEPTORS



P-1
BIOREACTOR FEED PUMP, 500 GPM, 25 HP

T-1
INFLUENT EQUALIZATION TANK, 11,000 gal POLY TANK 2" INLET & OUTLET

FM-1
FLOWMETER, MAGNETIC– FOR CHEMICAL FEED PACING

P-2
BOOSTER PUMP, 500 GPM, 25 HP

T-2 a to e
BIOREACTOR VESSELS 10,000 gal STEEL TANKS, 4" INLET & OUTLET

T-3
SODIUM ACETATE (NaAc) FEED TANK, 11,000 gal POLY TANK, .5" OUTLET

MP-1
ACETATE METERING PUMPS, 1 GPM

M-1
ACETATE TANK MIXER, 2 hp

M-3, M-4
IN–LINE STATIC MIXER, 6"

T-4
DI–AMMONIUM PHOSPHATE $(NH_4)_2HPO_4$ (NUTRIENT) TANK 11,000 gal POLY TANK .5" OUTLET

M-2
NUTRIENT TANK MIXER, 2 hp

MP-2
NUTRIENT METERING PUMPS, 1 gpm

P-3
FILTER FEED PUMP, 500 GPM

T-5
DUPLEX BAG FILTER W/ AUTOMATIC SEQUENCING, 12 BAGS PER UNIT

T-6
UV DISINFECTION SYSTEM

T-7A-F
IE VESSELS (6)

LEGEND

- BALL VALVE
- SAMPLE PORT
- PUMP, CENTRIFUGAL OR METERING (P–1)
- MIXER
- IN LINE MIXER

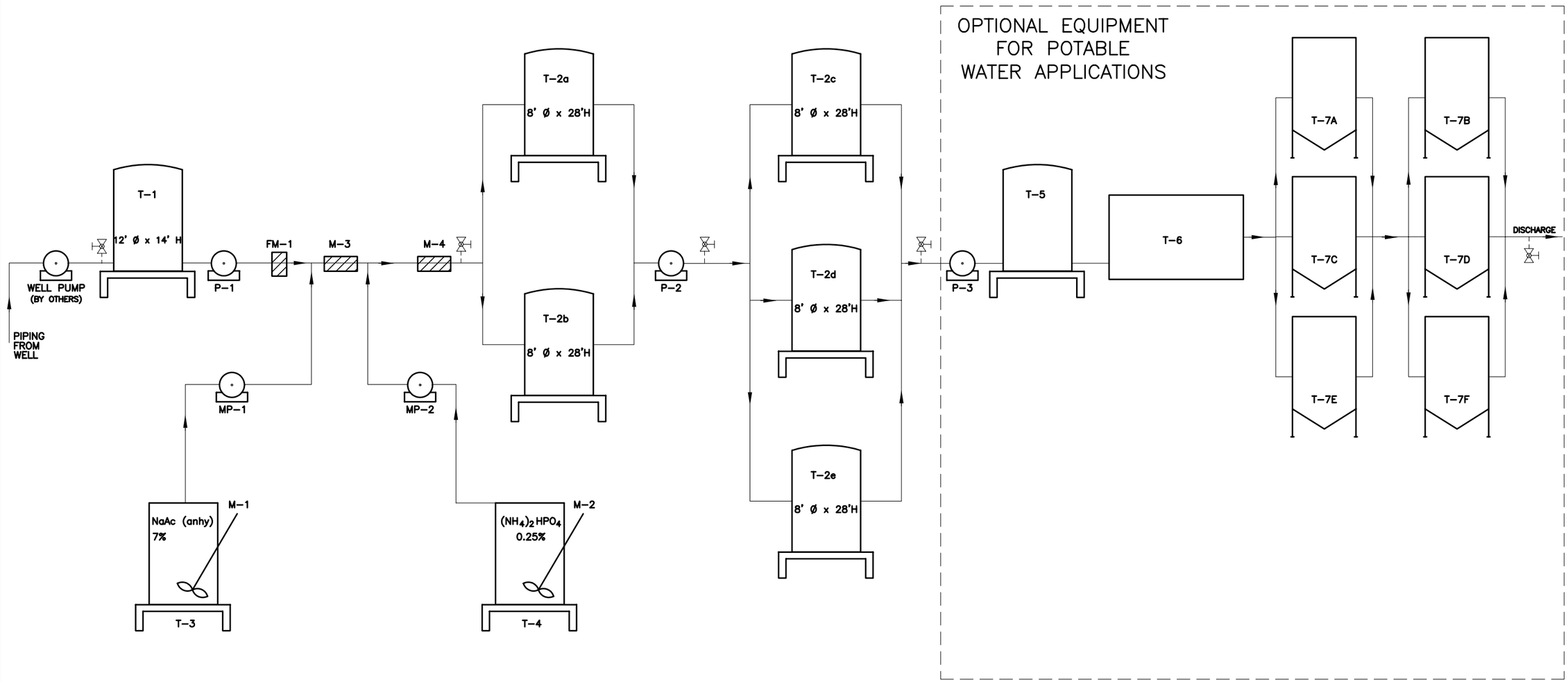


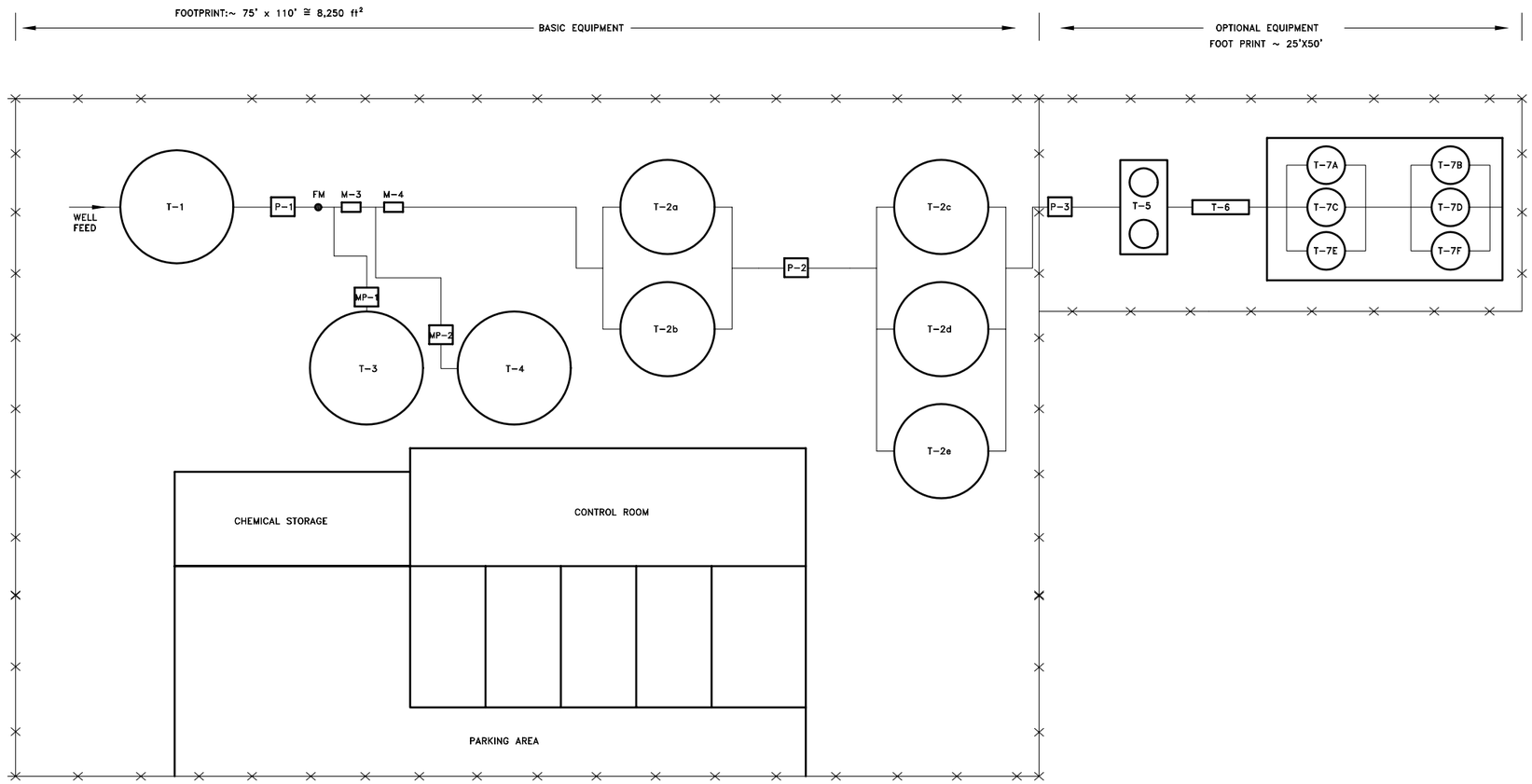
Figure 15
500–GPM DSBR
PROCESS FLOW DIAGRAM

JPL

FOSTER WHEELER
ENVIRONMENTAL CORPORATION

DESIGN BY: VH/DT | CHECKED BY: DT | APPROVED BY: VH | DATE: 01/15/03

I:\2423–NFESC\DWG\PILOT REPORT\FIG 15.DWG
PLOT/UPDATE: JUN 27 2003 15:48:54



LEGEND

T-10 REFER TO FIGURE 13 FOR SPECIAL EQUIPMENT DESIGNATION (T = TANK, P = PUMP, M = MIXER, FM = FLOW METER)

CONCRETE PADS FOR PUMPS

15 7.5 0 15 30

SCALE IN FEET

Figure 16
500-GPM DSBR
EQUIPMENT LAYOUT

JPL

FOSTER WHEELER
ENVIRONMENTAL CORPORATION

DESIGN BY: VH/DT | CHECKED BY: DT | APPROVED BY: VH | DATE: 01/15/03

I:\2423-NFESC\DWG\PILOT REPORT\FIG 16.DWG
PLOT/UPDATE: JUN 30 2003 09:09:18