

# **Continuous Backwash Filters: An Attractive Technology for Phosphorus Removal and Recovery, and Nitrate Reduction**

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## **ABSTRACT**

Continuous backwash filters (CBFs) have been widely applied for suspended solids and nutrient removal. A more recent development is the use of a reactive media, namely hydrous ferric oxide coated silica sand in CBFs, to achieve effluent total phosphorus (TP) levels less than 0.1 mg/l. The results from case studies involving the use of a commercial embodiment of the reactive filtration technology, demonstrate the ability to achieve an effluent TP level of 10 µg/l in the treatment of municipal wastewater secondary effluent. The reactive filtration technology appears to be well suited as the means to achieve phosphorus removal and recovery in a new flowsheet designed to achieve energy sustainability, and water and nutrient recovery.

A mathematical model was developed for the reactive filtration, phosphorus removal technology. The observed results from case studies were in general agreement with the model predictions. A model was also developed for application of a tertiary treatment CBF system for nitrate removal. Observed results derived from various locations where the denitrification CBF system was in operation generally agreed with the model values.

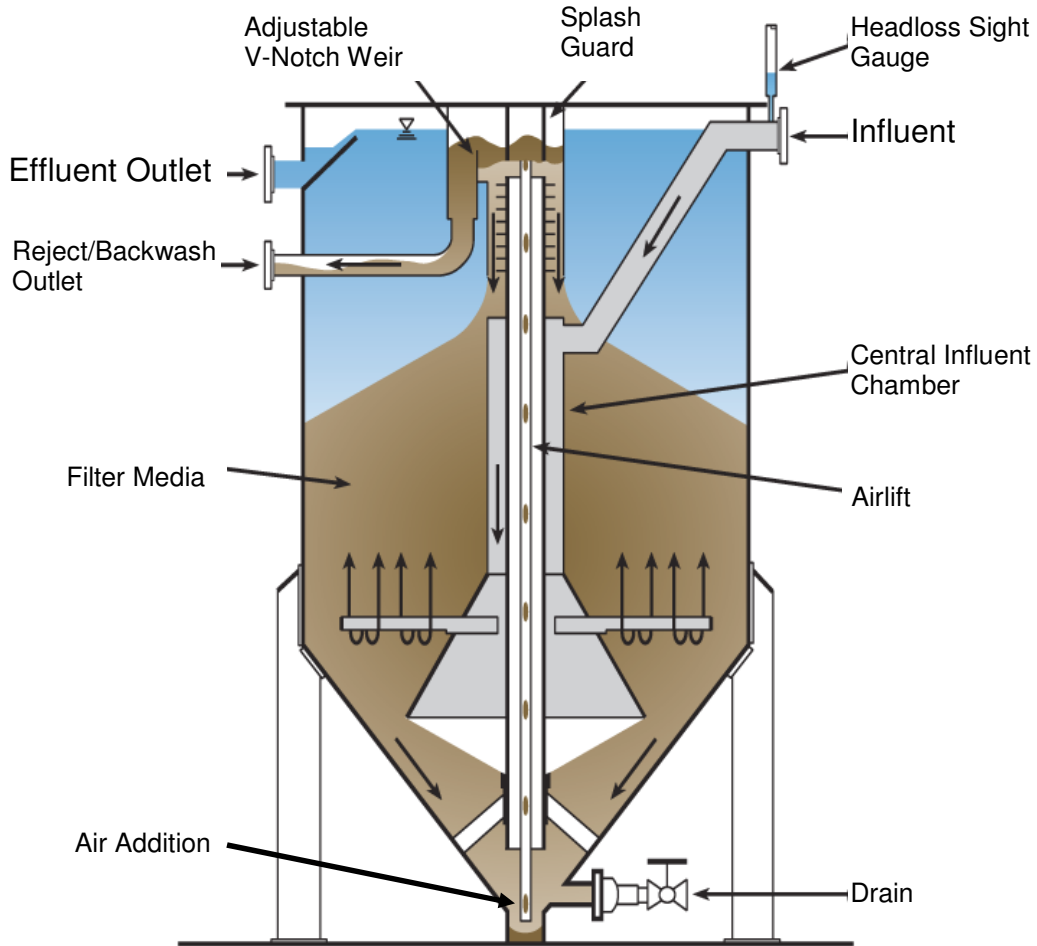
**KEYWORDS:** Nutrient removal, continuous backwash filters, mathematical modeling.

## **INTRODUCTION**

Continuous backwash filters (CBFs) have been applied for suspended solids and turbidity removal in North America since the mid-1980's (Weinschrott and Tchobanoglous, 1986). Since the early 1990s, research and full scale applications of the CBF technology for nitrate reduction through biological denitrification from municipal and industrial wastewaters have been reported (Koopman et al., 1990; STOWA, 1993; Cantor et al., 1995). These applications involved the use of inert, non-reactive sand as the bed media. More recently, the technology has been applied or proposed for phosphorus removal (Lancaster and Madden, 2008), combined phosphorus removal and nitrate reduction (Schauer et al., 2006), ammonia oxidation through biological nitrification, and the biological reduction of perchlorate and selenium from contaminated groundwaters or process wastewaters. The bed media used in certain of these applications is reactive, in the form of hydrous ferric oxide (HFO) coated silica sand (Newcombe et al., 2008a) or granular activated carbon (GAC) (Sutton, 2006).

In most commercial system embodiments of the CBF technology, the influent wastewater (i.e., typically municipal wastewater secondary effluent) is routed to the bottom of the filter and distributed across the bed media. Suppliers of the technology to North America include, Blue Water Technologies, Hayden, ID (Blue Water), Siemens Water Technologies and Parkson. Blue Water's embodiment of the CBF technology, referred to as the Centra-flo Filter, is depicted in Figure 1. In the system, the wastewater influent flows upward through the filter media which is

**Figure 1 - Centra-flo Filter Embodiment of the CBF Technology**



moving downward countercurrent to the influent flow. The treated wastewater or effluent is removed at the top of the filter. The rate of downward movement of the bed media or the bed turnover rate, is controlled by the rate of media airlift which occurs through the introduction of air into a pipe located in the center of the filter. The media is scoured as it passes upward through the airlift pipe. At the top of the filter, the liquid slurry of scoured media and other particulates (i.e., material removed from the media surface or trapped in the pores of the bed) is discharged to a baffled chamber or washbox. The washbox is designed hydraulically to allow for media washing by a portion of the filter effluent and gravity separation of the media. The

settled media falls through the washbox and is returned to the top of the filter bed while the reject or backwash flows out of the filter (Figure 1).

Conventional downflow filters for denitrification have been utilized in North America at the commercial scale for over 25 years (Chen, 1980). Unlike CBFs, the bed media in these filters does not move or is static during feed processing. As feed solids are trapped in the media pores and biomass growth occurs in the pores or as a film on the media, the pressure drop increases across the filter necessitating intermittent media backwash. The principles governing operation of these static bed filters (SBFs) and the method for control of solids and biomass build-up, make it difficult to operate the reactors under equilibrium conditions. In CBF reactors, the ability to control the bed turnover rate on a continuous basis by adjustment of the rate of media airlift (i.e., air addition rate), translates to a correlation with the filter solids retention time (SRT) and the media biofilm thickness, providing a means for optimizing the biomass growth and substrate removal conditions. Furthermore, this biomass inventory control strategy allows the reactor to operate under stable hydraulic conditions (i.e., prevents feed short circuiting or bypass flow). The use of continuous backwash and the ability to adjust the reactor biomass solids inventory simplifies operation and maximizes the volumetric efficiency of the filter, respectively.

Phosphorus removal in CBF systems occurs by chemical and physical mechanisms. In systems in which the filter bed is composed of an inert, non-reactive sand media, phosphorus removal occurs through the capture of precipitated, single (e.g., ferric phosphate) or mixed cation phosphates (e.g., calcium or aluminum phosphates, hydroxyphosphates). In order to meet very low effluent total phosphorus (TP) requirements (e.g., less than 0.1 mg/l) utilizing CBF systems designed around the precipitation-filtration removal mechanisms, the addition of a high concentration of metal salts (e.g., minimum Al to P molar ratio in the range from 2 to 8) is required, as dictated by chemical precipitation equilibrium. In addition, a small concentration (e.g., 0.5 mg/l) of a costly polymer solution (e.g., \$5.50 per kg) is typically required to minimize the release of phosphorus containing particulates. A significant reduction in metal salt addition can be achieved in CBF systems when the media is reactive versus inert, in the form of HFO coated silica sand media. In this case, phosphorus removal occurs by the process referred to as reactive filtration. Chemical phosphorus removal using HFO occurs by two fundamental mechanisms, adsorption of phosphate onto preformed HFO and co-precipitation of phosphate into the HFO structure (Newcombe et al., 2008b; Smith et al., 2008). Coupling these mechanisms with physical filtration to achieve reactive filtration using the CBF technology, embodied commercially by Blue Water as the Blue PRO system, minimizes the iron (e.g., ferric chloride) addition requirements while achieving a residual phosphate concentration less than that predicted according to chemical phosphate removal dissociation and solubility principles. When the CBF is located downstream of an existing suspended growth biological system (e.g., conventional activated sludge system), the role of the adsorption mechanism in achieving phosphorus removal is maximized and the iron addition requirements minimized, when a portion or all of the CBF backwash (i.e., filter backwash) containing HFO coated particulates is recycled to or in front of the biological reactor.

The CBF reactive filtration technology can be operated as a combined phosphorus removal and advanced oxidation process (AOP) by the addition of a small amount of ozone to the filter influent. Blue Water's commercial embodiment of the CBF technology with Fe and ozone

addition, is referred to as the Blue CAT system. The combined CBF reactive filtration-AOP technology represents an attractive physical-chemical technology to achieve phosphorus removal and advanced oxidation, with the potential for phosphorus recovery. There is a growing interest in developing flowsheets which are designed with the objective of treating wastewater not as a waste but as a resource (e.g., phosphorus recovery) while reducing the impact of wastewater treatment on greenhouse gas (GHG) emissions. The Blue CAT system appears to be an ideal technology for achieving phosphorus removal and recovery as a unit process in a recently developed municipal wastewater treatment flowsheet which achieves these goals.

The purpose of this paper is to derive information from pilot or full scale CBFs operated to achieve phosphorus removal through reactive filtration or to achieve nitrate reduction, and to use the information to develop design and performance relationships with the aid of computer modeling and simulation. Information will be derived from projects completed at treatment plants in Marlborough, MA, Sunrise, FL, Bellefonte, PA, and Grangeville and Hayden, ID. In addition, information will be presented on the application of the combined, CBF reactive filtration – AOP technology for achieving phosphorus removal and recovery.

## **FILTERS FOR PHOSPHORUS REMOVAL OR NITRATE REDUCTION**

To achieve a low effluent TP (e.g., 1.0 mg/l), enhanced biological phosphorus removal (EBPR) is widely applied in the U.S. for the treatment of municipal wastewater, particularly when the wastewater flow is large (e.g., greater than 18,750 m<sup>3</sup>/day). It is generally acknowledged effluent TP levels between 0.1 and 1.0 mg/l can be consistently achieved by combining EBPR with conventional tertiary sand, multi-media (e.g., sand plus anthracite) or cloth filtration, to ensure particulate P is reduced to levels ideally below 0.2 mg/l. To achieve this effluent particulate P level, the corresponding total suspended solids (TSS) concentration will normally need to be less than 5 mg/l. In SBFs and CBFs designed around the precipitation-filtration removal mechanisms utilizing non-reactive bed media to achieve very low effluent TP levels, defined previously as less than 0.1 mg/l, the addition of a high mass of metal salts is required. To consistently meet this effluent TP level while attempting to minimize chemical requirements (i.e., metal salts, polymer solutions, pH control chemicals), the SBF and CBF system suppliers have introduced multi-stage systems such as Parkson Corporation's DynaSand D2 filtration system (i.e., two-stage CBF) and Siemens Water Technologies Trident HS process system (i.e., two-stage clarification plus SBF stage). This same level of phosphorus removal performance can be achieved through the reactive filtration process in single or two-stage CBFs utilizing HFO coated silica sand media, with a significant reduction in chemical requirements (Lancaster and Madden, 2008; Benisch et al., 2007; Leaf and Johnson, 2006). This effective decrease in solids load to the filter will allow operation at a higher hydraulic loading rate or HLR (i.e., volume of influent applied per reactor cross-sectional per unit time) and potentially decrease the filter volume requirement.

It should be noted it is very difficult to achieve ultra-low effluent TP levels (e.g., less than 0.01 mg/l) in SBFs and CBFs, and in other physical-chemical systems (e.g., Cambridge Water Technology's CoMag system, Kruger's ACTIFLO system) due to the presence of non-reactive,

recalcitrant phosphorus compounds in the influent at concentrations typically exceeding 0.01 mg/l (Lancaster and Madden, 2008; Benisch et al., 2007).

Full scale application of deep bed filters for denitrification of nitrified municipal wastewater began in the U.S. in the early 1970s (U.S. EPA, 1975) with installation of fine media (i.e., sand, anthracite coal) based SBFs. CBFs began to be applied for denitrification some years later, as previously referenced. Since that time, the results from pilot studies combined with information from full scale installations, have provided a basis for design of the filters to meet an effluent total nitrogen (TN) requirement as low as 3 mg/l or less. In the denitrification application, typically SBFs and CBFs are sized on the basis of a specified HLR and a nitrate-N volumetric loading rate or VLR (i.e., mass of N applied per volume per unit time). The reactor volume term is normally defined as the reactor volume occupied by the media. The filter hydraulic retention time (HRT) is also normally calculated on this basis (i.e., empty bed HRT). Selecting the appropriate HLR and VLR values in design of SBFs and CBFs to meet a certain mean daily effluent TN value, requires consideration of a number of factors including the following;

- 1) the influent concentration characteristics (i.e.,  $\text{NO}_3\text{-N}$ ,  $\text{NO}_2\text{-N}$ , dissolved oxygen (DO), TSS), the influent mean daily flow rate, and the diurnal flow and concentration variability,
- 2) the influent temperature and pH,
- 3) the mean daily effluent  $\text{NO}_x\text{-N}$  (i.e.,  $\text{NO}_3\text{-N}$  plus  $\text{NO}_2\text{-N}$ ), TP and TSS requirements, and
- 4) the specific external carbon source to be added (e.g., methanol or alternative source such as MicroC G from Environmental Operating Solutions or Unicarb DN from Univar USA) and the mass addition rate.

In addition, consideration needs to be given to the design and operating characteristics of the specific SBF or CBF technology (e.g., Blue Water Centra-flo Filter versus Siemens ASTRASAND filter or Parkson DynaSand Filter) in selecting the HLR and VLR values. For example, the sand characteristics (e.g., particle size, roundness), sand bed depth and the bed turnover rate may affect the values of HLR and VLR that are appropriate to achieve the performance requirements. It is not surprising a significant range of HLR and VLR values have been reported by researchers and practitioners as appropriate to obtain a specific filter performance level, given the discussion above. The reported values imply CBFs can be operated at a mean daily HLR of up to approximately 10 m/h whereas the comparable value for SBFs would be 7 m/h. These maximum values are feasible only under certain ideal conditions such as near the optimal temperature for growth of denitrifiers (i.e., 20° to 30° C) and low influent or less stringent effluent  $\text{NO}_x\text{-N}$  requirements (e.g., influent  $\text{NO}_x\text{-N}$  5 mg/l, effluent  $\text{NO}_x\text{-N}$  2 mg/l). Reported values imply CBFs and SBFs can be operated at a mean daily nitrate-N VLR of up to approximately 2 kg/m<sup>3</sup>·day under these ideal conditions. The ability to control the bed media turnover rate in CBFs does provide the opportunity to maximize the nitrate volumetric removal rate by maximizing the mass of active denitrifiers in the filter, as previously discussed. The accumulation of influent TSS and the production of TSS due to biomass growth provide filtration (i.e., resulting in a low effluent TSS) and  $\text{NO}_x$  reduction. The accumulation and production of TSS result in a pressure drop across the bed media. The pressure drop can be controlled in-part by adjusting the bed turnover rate (i.e., adjusting the rate of media airlift through adjusting the air addition rate) beyond that which occurs automatically as a result of

changes in the HLR. The correlation between pressure drop and the biomass inventory, and the ability to control the pressure drop, translates to a control strategy which when implemented represents a means of maximizing the active biomass inventory. This control strategy can translate to more consistent performance under varying hourly influent conditions (Kramer et al., 1999). It is worthwhile noting this control strategy, characteristic to CBFs, has the potential to be further exploited since there is likely a relationship between the level of media scouring that occurs in the airlift pipe, the depth of the media bed, the media characteristics, the HLR and the media biofilm thickness. If the biomass inventory and the media biofilm thickness could be controlled, conditions optimal for maximizing the substrate (i.e., NO<sub>x</sub>-N and oxygen) removal rate could be achieved. Controlling the media biofilm thickness may be particularly important when a more complex organic carbon source (e.g., MicroC G) versus methanol is used to promote substrate reduction as in this case biofilm control has been reported to be more problematic (deBarbadillo et al., 2008).

It has been reported that CBFs designed to achieve significant denitrification (i.e., 5 mg/l NO<sub>x</sub>-N removal or greater), did not perform as well as SBFs when the goal was also to achieve a very low effluent P concentration (Husband and Becker, 2007). The authors speculated the continuous backwashing of the media resulted in a lower mass of attached biomass growth in the CBF versus that present in a parallel operating SBF. In this application when the orthophosphate concentration is not sufficient to support growth of the biomass required to achieve this level of denitrification (deBarbadillo et al., 2006), a significant amount of the required phosphorus must be derived from the particulate fraction present in the filter. Sources of particulate phosphorus include that derived from the feed solids that are trapped in the filter pores, or derived from biomass produced in the filter and present as suspended growth trapped in the pores or as a biofilm. Particulate P in the feed solids could be associated with biomass or an inorganic precipitate released from an upstream unit process (e.g., bioreactor). It is likely at a certain CBF bed turnover rate, the mass of particulate P present in the filter will be reduced below the critical level to support significant denitrification. On this basis, it is likely there is an optimal bed turnover rate specific to this application and when operated at this condition, the CBF would perform equal or better than the SBF. If the performance requirement is a very low effluent P level and significant denitrification, a CBF designed to achieve P removal through reactive filtration using HFO coated sand media versus through precipitation- filtration, is favored due to the formation of a lower mass of chemical solids (i.e., lower chemical mass addition rate). Formation of chemical solids effectively represents a solids load to the filter requiring more filter capacity or a greater bed turnover rate. This assertion assumes the phosphorus required to support denitrification can be derived from solids present in the feed and captured, or from biomass formed in the filter.

A qualitative comparison of the CBF technology versus SBFs for phosphorus removal or nitrate reduction is presented in Table 1. The comparison assumes the CBF technology relies on the reactive filtration process using HFO coated sand media in the phosphorus removal application. Although the Table 1 comparison is specific to phosphorus removal or nitrate reduction, the discussion above implies an advantage for the CBF reactive filtration technology when the need is to achieve a very low effluent P concentration and significant denitrification, in the same filter.

**Table 1 - Comparison of CBFs and SBFs for Phosphorus Removal or Nitrate Reduction<sup>1</sup>**

Comparative Factor	Continuous Backwash Filters	Conventional Downflow Static Bed Filters
Space or Footprint Requirements	<ul style="list-style-type: none"> <li>• CBF design minimizes ancillary vessel/equipment requirements.</li> <li>• Use of reactive filtration process to achieve very low effluent TP reduces solids load to filter associated with chemical addition, translating to smaller filter volume requirements and allowing operation at higher HLR.</li> <li>• Mean daily HLR of up to 10 m/h for N reduction application minimizes filter footprint.</li> </ul>	<ul style="list-style-type: none"> <li>• Need for off-line backwash and related vessels/equipment increases footprint.</li> <li>• Reactive filtration process not applicable necessitating larger chemical addition, increasing reactor size requirements or need for multiple stages.</li> <li>• Lower HLR for N reduction translates to greater space requirements.</li> </ul>
Volumetric Efficiency	<ul style="list-style-type: none"> <li>• Automatic bed media turnover rate control strategy allows performance to be maintained under varying hourly influent loading conditions.</li> <li>• Use of reactive filtration to achieve very low effluent total P, reduces chemical addition rate allowing operation at higher VLR.</li> <li>• In denite application, mean daily nitrate-N volumetric loading of up to 2 kg/m<sup>3</sup> · day.</li> </ul>	<ul style="list-style-type: none"> <li>• Varying hourly influent loading conditions less easily handled.</li> <li>• Reactive filtration process not applicable. Greater chemical addition rate translates to operation at lower VLR.</li> <li>• In denite application, can operate at comparable mean daily volumetric loading.</li> </ul>
Performance Stability	<ul style="list-style-type: none"> <li>• Tolerant to adverse influent conditions due to ability to automatically control bed media turnover rate without interrupting operation.</li> </ul>	<ul style="list-style-type: none"> <li>• Adverse influent conditions can lead to shorter feed cycle/more frequent backwashing required, requiring closer operator attention to maintain performance.</li> </ul>
Mechanical Simplicity, Maintenance Requirements	<ul style="list-style-type: none"> <li>• Continuous filter operation with no internal moving parts reduces maintenance needs.</li> </ul>	<ul style="list-style-type: none"> <li>• Feed and backwash cyclic operation translates to more complex mechanical design, and potential for more equipment and vessel maintenance.</li> </ul>
Solids Generation and Reuse	<ul style="list-style-type: none"> <li>• Use of reactive filtration reduces chemical requirements and associated solids generation in P removal application. Recovered solids represent potential fertilizer product.</li> </ul>	<ul style="list-style-type: none"> <li>• Greater chemical requirements result in more solids generation. Solids not as attractive for reuse.</li> </ul>
Equipment and Construction Costs	<ul style="list-style-type: none"> <li>• Need for internal airlift limits individual filter cross-sectional area, typically requiring supply of multiple filter units potentially translating to higher capital costs at larger wastewater flows.</li> </ul>	<ul style="list-style-type: none"> <li>• At larger wastewater flows fewer individual filter units required potentially reducing capital costs.</li> </ul>
Operating and Mechanical Maintenance Costs	<ul style="list-style-type: none"> <li>• CBF design involves less equipment reducing mechanical maintenance.</li> <li>• CBF design and automatic bed media turnover rate control strategy, simplify operating requirements particularly under adverse influent conditions.</li> <li>• Use of reactive filtration to achieve very low effluent total P, reduces chemical requirements and associated operating cost.</li> </ul>	<ul style="list-style-type: none"> <li>• More equipment to maintain increasing costs, although ensures media clogging does not become an issue and no need for airlift piping maintenance.</li> <li>• Feed and backwash cyclic operation increases operating manpower requirements.</li> <li>• Need for greater chemical addition to achieve very low effluent total P, increases operating cost.</li> </ul>

<sup>1</sup> The comparison is based on achieving P removal in CBFs to effluent levels less than 0.1 mg/l TP, through the reactive filtration process using HFO coated sand media.

This advantage assumes biomass growth in the filter does not prevent accomplishing phosphorus removal through the reactive filtration process. Even though the CBF design and operating characteristics translate to advantages versus SBFs when the technology is applied for phosphorus removal or nitrate reduction, the advantages may not be as important or translate to the most economical filter choice under all circumstances. Control of the solids accumulated or produced in CBFs through adjustment of the bed media turnover rate may not be relevant in certain applications (e.g., less than 5 mg/l NO<sub>x</sub>-N removal required). Hydraulic considerations with respect to the design of the CBF internal airlift piping, limit the design cross-sectional area of the filter which may translate to higher equipment and construction costs versus SBFs in the treatment of larger wastewater flow rates.

It is worthwhile noting the fluidized bed reactor (FBR) configuration represents the ideal fixed-film reactor when control of the solids produced in the reactors as a biofilm is important (i.e., biofilm thickness control). FBRs can be operated at much higher VLRs than CBFs or SBFs (Sutton, 2006) when they are designed using silica sand or GAC as the bed media and with a means to control the media biofilm thickness, such as the FBR system offered by Basin Water, Rancho Cucamonga, CA. An issue with FBRs is that physical filtration does not occur in the reactor and therefore the TSS concentration in the effluent will be equal or greater than the feed TSS concentration.

## **DESIGN OF REACTIVE FILTRATION BASED CBFs FOR PHOSPHORUS REMOVAL**

The fundamentals governing phosphorus removal through the reactive filtration process in CBFs utilizing HFO coated silica sand as the filter bed media, are detailed elsewhere (Newcombe et al., 2008a; Newcombe et al., 2008b). In the following section, data developed from operation of Blue Water’s commercial embodiment of the CBF reactive filtration process for phosphorus removal (i.e., Blue PRO system), is used to develop a steady-state model for the reactive filtration process. The model is used in the section entitled, CBF Case Studies in order to compare the observed effluent phosphorus results with the model predicted values.

### **Reactive Filtration Process Modeling**

A reactive filtration model was developed using data provided by Blue Water from operation of two stage Blue PRO CBF pilot systems at municipal wastewater treatment plants in Hayden and Grangeville, ID. Details regarding the specific data used for model development are provided elsewhere (Schraa, 2009). Achieving near complete phosphorus removal in the treatment of municipal wastewater secondary effluents, requires a high degree of TSS removal. Therefore, the reactive filtration model includes a TSS removal component. The TSS data required for model development was determined from turbidity measurements using the following equations (Tchobanogolous, 2003).

$$TSS_{inf} = 2.35NTU_{inf} \quad (1)$$

$$TSS_{p1} = 1.45NTU_{p1} \quad (2)$$

$$TSS_{p2} = 1.45NTU_{p2} \quad (3)$$

where;

- $TSS_{inf}$  = influent TSS concentration (mg/l),
- $TSS_{p1}$  = first stage effluent TSS concentration (mg/l),
- $TSS_{p2}$  = second stage effluent TSS concentration (mg/l),
- $NTU_{inf}$  = influent turbidity (NTU),
- $NTU_{p1}$  = first stage effluent turbidity (NTU), and
- $NTU_{p2}$  = second stage effluent turbidity (NTU).

The reactive filtration model is a steady-state model comprised of mass balances on the chemical oxygen demand (COD), inorganic suspended solids, metal precipitates, nitrogen, and phosphorus water quality variables used in the ASM3 (Henze et al., 2000) biological model with metal precipitates and phosphorus variables having been added to the model. A steady-state modeling approach was taken as the data were recorded daily and the dynamics of the reactive filtration process are expected to occur on a smaller time scale justifying a steady-state model. An empirical expression was used within the mass balances to predict the filtration of particulates and a steady-state version of the ASM2d chemical precipitation model was used to model the adsorption and co-precipitation of dissolved phosphorus.

The TSS removal or filtration component of the model uses an exponential rise function and a linear expression to calculate the concentration-based TSS removal efficiency as follows, given the influent TSS loading rate to the CBF and the polymer addition rate.

$$E_{TSS} = E_{TSS,u} \left( 1 - e^{\left( \frac{-SLR}{\tau} \right)} \right) + \frac{\lambda Q_{polymer}}{A} \quad (4)$$

where;

- $E_{TSS}$  = concentration based TSS removal efficiency due to filtration (-),
- $E_{TSS,u}$  = maximum or ultimate concentration based TSS removal efficiency (-),
- $SLR$  = solids loading rate (g TSS/(m<sup>2</sup>·h)),
- $\tau$  = solids loading rate at 63% of  $E_{TSS,u}$  (g TSS/(m<sup>2</sup>·h)),
- $\lambda$  = factor for impact of polymer addition rate per cross-sectional area ((h·m<sup>2</sup>)/l),
- $Q_{polymer}$  = polymer addition rate (l/h), and
- $A$  = cross-sectional area of filter (m<sup>2</sup>).

Further details concerning selection, development and verification of this filtration component of the reactive filtration model are provided elsewhere (Schraa, 2009) including the data base used to estimate the model parameters and confidence limits for the parameters. The reactive filtration model was implemented in GPS-X and the filtration component model parameters were

estimated using a nonlinear least squares objective function and the optimization algorithm found in GPS-X (Hydromantis, 2006). The estimated values for the parameters  $E_{TSS,u}$ ,  $\tau$  and  $\lambda$  were respectively, 0.764, 6.99 g/m<sup>2</sup> · h and 1.04 h · m<sup>2</sup>/l.

The reactive filtration model also includes an adsorption and co-precipitation component. The adsorption/co-precipitation component of the model is based on the chemical precipitation model found in ASM2d (Henze et al., 2000). This is a kinetic model based on dynamic mass balances and rate expressions for the precipitation and redissolution of ferric phosphate as follows. The alkalinity switching function has been removed for simplicity.

### Mass balance on soluble phosphorus

$$\frac{ds_{OP,eff}}{dt} = \frac{(Q_{inf} s_{op,inf} - Q_{eff} s_{op,eff} - Q_{rej} s_{op,rej})}{V} - k_{pre} s_{op,rej} x_{MeOH,rej} + k_{red} x_{MeP,rej} \quad (5)$$

### Mass balance on metal hydroxide

$$\frac{dx_{MeOH,eff}}{dt} = \frac{(Q_{inf} x_{MeOH,inf} - Q_{eff} x_{MeOH,eff} - Q_{rej} x_{MeOH,rej})}{V} - 3.45 k_{pre} s_{op,rej} x_{MeOH,rej} + 3.45 k_{red} x_{MeP,rej} \quad (6)$$

### Mass balance on metal phosphate

$$\frac{dx_{MeP,eff}}{dt} = \frac{(Q_{inf} x_{MeP,inf} - Q_{eff} x_{MeP,eff} - Q_{rej} x_{MeP,rej})}{V} + 4.87 k_{pre} s_{op,rej} x_{MeOH,rej} - 4.87 k_{red} x_{MeP,rej} \quad (7)$$

where,

- $Q_{inf}$  = influent flow rate (m<sup>3</sup>/day),
- $Q_{eff}$  = effluent flow rate (m<sup>3</sup>/day),
- $Q_{rej}$  = reject stream flow rate (m<sup>3</sup>/day),
- $V$  = volume of reactive filter (m<sup>3</sup>),
- $s_{op,inf}$  = concentration of orthophosphate-P in influent (mg P/l),
- $s_{op,eff}$  = concentration of orthophosphate-P in effluent (mg P/l),
- $s_{op,rej}$  = concentration of orthophosphate-P in reject stream (mg P/l),
- $x_{MeOH,inf}$  = concentration of ferric hydroxide in influent (mg/l),
- $x_{MeOH,eff}$  = concentration of ferric hydroxide in effluent (mg/l),
- $x_{MeOH,rej}$  = concentration of ferric hydroxide in reject stream (mg/l),
- $x_{MeP,inf}$  = concentration of ferric phosphate in influent (mg/l),
- $x_{MeP,eff}$  = concentration of ferric phosphate in effluent (mg/l),
- $x_{MeP,rej}$  = concentration of ferric phosphate in reject stream (mg/l),
- 3.45 = ratio of molecular mass of ferric hydroxide to molecular mass of phosphorus (g MeOH/g P),
- 4.87 = ratio of molecular mass of ferric phosphate to molecular mass of phosphorus (g MeP/g P),

$k_{pre}$  = rate constant for MeP precipitation ( $m^3/(gMeOH \cdot day)$ ),  
 $k_{red}$  = rate constant for MeP redissolution (1/day), and  
 $t$  = time (day).

The adsorption and co-precipitation model parameters were estimated using the optimizer in GPS-X and a nonlinear least squares objective function. The estimated parameters are stated in Table 2 along with their 95 percent confidence regions and the parameter correlation matrix. The confidence limits indicate that both the parameters are significant and the correlation matrix does not show any significant correlation between the parameters. The kinetic rate parameters in the model are very large indicating that the adsorption and co-precipitation processes occur very rapidly in the reactive filter. As a result, a steady-state version of the dynamic adsorption/co-precipitation model can be used. The model can be transformed into a steady-state model by setting the derivative terms to zero.

A plot (Figure 2) of the modeled values versus the measured values for the effluent orthophosphate-P concentration indicates the modeled results do not show any noticeable trends relative to the measured values. The coefficient of determination for the regression is 0.81 indicating that 81 percent of the variation is explained by the model.

**Table 2 - Parameter Summary for Orthophosphate-P Adsorption and Co-precipitation Model**

Parameter	Estimated Value	95 Percent Confidence Limits	Approximate Correlation Matrix
$k_{pre}$ , $m^3/g \cdot day$	$2.28 \times 10^9$	$\pm 2.00 \times 10^6$	1.00 -0.69
$k_{red}$ , 1/day	$2.58 \times 10^8$	$\pm 2.04 \times 10^5$	-0.69 1.0

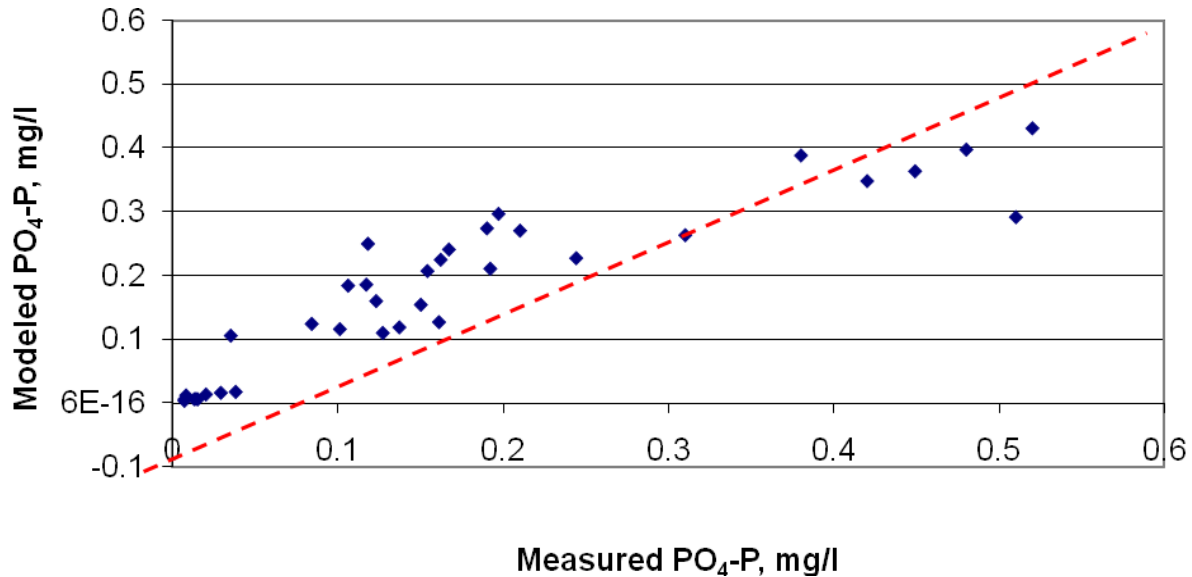
Further details concerning selection of the adsorption/co-precipitation model component and its verification plus information concerning the complete reactive filtration model is provided elsewhere (Schraa, 2009).

### CBF Case Studies

A number of pilot plant studies have been completed involving operation of Blue Water's, CBF Blue PRO system. The results from four of these studies follow.

**Marlborough, MA.** A Blue Water two stage, CBF pilot system was operated at the Westerly wastewater treatment plant in Marlborough, MA in the spring of 2007. The objective of the pilot study was to achieve an effluent TP concentration of less than 0.1 mg/l. The pilot system was operated for a period of approximately two weeks following a start-up phase during which equilibrium operation was established. The pilot system treated secondary effluent from a single stage activated sludge system designed to achieve seasonal nitrification, which during the pilot study utilized alum addition to the primary clarifiers to achieve an effluent TP concentration of less than 1 mg/l. The CBF pilot system was operated in the Blue PRO mode (i.e., without ozone

**Figure 2 - Modeled Versus Measured CBF Effluent Orthophosphate Values**



addition) during most of the study. Each pilot filter was composed of 0.65 m<sup>2</sup> of filtration area with the HFO coated sand media occupying a volume of 0.991 m<sup>3</sup>. The filter backwash or reject flow from each filter was routed to drain. The conditions during operation of the CBF pilot system in the Blue PRO mode and the observed performance results are stated respectively, in Tables 3 and 4. The influent and effluent values were based on analyses of daily composite samples generated from combining grab samples taken hourly over a period from two to four hours. Details regarding the analytical protocols are discussed elsewhere (Lancaster and Madden, 2008).

The results (Table 4) demonstrate the ability to achieve an effluent TP of less than 0.1 mg/l from a single stage Blue PRO CBF with the addition of a relatively small amount of ferric chloride. The presence of non-reactive, recalcitrant phosphorus compounds in the influent did have a negative impact on the performance of the filters. It should be noted if the Blue PRO CBF technology was applied at the full scale level at the Westerly plant, the following design changes and operating or performance impacts, would be anticipated.

- 1) The full scale CBF backwash flow rate as a percent of the influent flow rate would be reduced from approximately 13 to 7 percent due to full versus pilot scale-up effects.
- 2) The CBF backwash flow would be routed to the existing activated sludge system. This practice would reduce the quantity of chemicals used at the plant to achieve P removal versus that anticipated if the backwash was routed to drain, while meeting an effluent TP concentration of less than 0.1 mg/l utilizing a one versus two stage CBF Blue PRO system.

The adsorption/co-precipitation model component of the complete reactive filtration model was used to predict the effluent phosphorus values. The predicted results are in reasonable agreement with the observed values (Table 4).

**Table 3 – CBF Pilot System Operating Conditions: Marlborough, MA<sup>1</sup>**

Parameter	Value
<b>First Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	102
Backwash Flow, l/min	9 to 15
Air Flow, l/h	708
Fe Addition, mg/l	8 to 11
HLR, m/h	9.4
HRT, min	9.7
Temperature, °C	14
Influent pH	7.5 to 7.9
Effluent pH	7.4 to 7.6
<b>Second Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	89
Backwash Flow, l/min	9 to 15
Air Flow, l/h	708
Fe Addition, mg/l	6 to 9
HLR, m/h	8.2
HRT, min	11.1
Temperature, °C	14
Effluent pH	7.3 to 7.5

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode. Airflow stated at standard conditions. HRT calculated on empty bed basis. Fe added as ferric chloride.

<sup>2</sup> The influent flow rate values are set points. The actual flow rate did not vary significantly around each value.

**Sunrise, FL.** A Blue Water CBF pilot system was operated at the Southwest wastewater treatment plant in Sunrise, FL from early September through mid-December 2007. The objective of the pilot study was to meet an effluent TP concentration limit of 10 µg/l. Initially, a single stage CBF treated effluent from a pilot membrane bioreactor (MBR) system designed to achieve biological nutrient removal. Treating an influent containing a TP concentration of approximately 30 to 80 µg/l, little phosphorus removal was achieved across the CBF system. The poor performance was attributed to variable MBR operation and performance, water chemistry issues and other uncontrollable factors. Starting in early November, a two stage CBF was operated treating secondary effluent from the existing full scale oxidation ditch, activated sludge system. Each pilot filter was comprised of 0.28 m<sup>2</sup> of filtration area with the HFO coated

**Table 4 – CBF Pilot System Performance Results: Marlborough, MA<sup>1</sup>**

Parameter	Value
Influent, mg/l	
TP	0.321 (0.037)
Soluble TP	0.191 (0.060)
PO <sub>4</sub> -P	0.114 (0.025)
Alkalinity as CaCO <sub>3</sub>	165 (14)
TSS	< 5
First Stage Effluent, mg/l	
TP Observed	0.083 (0.022)
TP Predicted	0.110
Soluble TP Observed	0.060 (0.023)
Soluble TP Predicted	0.082
PO <sub>4</sub> -P Observed	< 0.005
PO <sub>4</sub> -P Predicted	0.003
Alkalinity as CaCO <sub>3</sub>	153 (17)
TSS	< 5
Second Stage Effluent, mg/l	
TP Observed	0.057 (0.020)
TP Predicted	0.092
Soluble TP Observed	0.052 (0.019)
Soluble TP Predicted	0.078
PO <sub>4</sub> -P Observed	< 0.005
PO <sub>4</sub> -P Predicted	< 0.001
Alkalinity as CaCO <sub>3</sub>	141 (17)
TSS	< 5

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode. Bracketed values represent the standard deviation.

sand media occupying a volume of 0.425 m<sup>3</sup>. The filter backwash flow from each filter was routed to drain. The CBF pilot system was operated in the Blue PRO mode and in the Blue CAT mode (i.e., ozone addition to the second stage, or to the first and second stage). The results discussed herein are those from operation in the Blue PRO mode. A performance optimization operating period and subsequent equilibrium operation, led to the conclusion that to minimize the effluent TP from the two stage system, it was necessary to add Fe and a small amount of polymer

to the influent to each filter. These chemicals were added in the form of a ferric chemical mix designated CB-9100F from CCI Chemical Corporation, Vernon, CA. The CBF system operating conditions and performance results achieved during the equilibrium period are stated respectively, in Tables 5 and 6. The phosphorus values reported were based on analysis of daily

**Table 5 – CBF Pilot System Operating Conditions: Sunrise, FL<sup>1</sup>**

Parameter	Value
<b>First Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	42
Backwash Flow, l/min	11
Air Flow, l/h	368
Fe Addition, mg/l	< 10
HLR, m/h	8.9
HRT, min	10.1
Temperature, °C	24
Influent pH	7.1 to 7.5
Effluent pH	6.9 to 7.0
<b>Second Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	30
Backwash Flow, l/min	11
Air Flow, l/h	368
Fe Addition, mg/l	< 10
HLR, m/h	6.5
HRT, min	14.2
Temperature, °C	24
Effluent pH	6.5 to 6.8

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode under equilibrium conditions (12/17 through 12/19/07). Airflow stated at standard conditions. HRT calculated on empty bed basis. Fe added as chemical mix also containing polymer.

<sup>2</sup> The influent flow rate values are set points. The actual flow rate did not vary significantly around each value.

**Grangeville, ID.** A Blue Water two stage, CBF pilot system was operated at the Grangeville wastewater treatment plant in Grangeville, ID in the summer of 2008. The objective of the pilot study was to meet an effluent TP concentration limit of 0.067 mg/l. The pilot system was operated for a period of approximately two weeks following a start-up phase during which

**Table 6 – CBF Pilot System Performance Results: Sunrise, FL<sup>1</sup>**

Parameter	Value
Influent, mg/l	
TP	2.32 (0.042)
PO <sub>4</sub> -P	2.0 (0.1)
Alkalinity as CaCO <sub>3</sub>	73 (3)
TSS	< 5
First Stage Effluent, mg/l	
TP Observed	0.047 (0.035)
TP Predicted	0.120
PO <sub>4</sub> -P Observed	0.038 (0.028)
PO <sub>4</sub> -P Predicted	0.091
Alkalinity as CaCO <sub>3</sub>	58 (6)
TSS	< 5
Second Stage Effluent, mg/l	
TP Observed	< 0.0044
TP Predicted	0.019
PO <sub>4</sub> -P Observed	< 0.0026
PO <sub>4</sub> -P Predicted	0.0027
Alkalinity as CaCO <sub>3</sub>	49 (5)
TSS	< 5

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode. Bracketed values represent the standard deviation. TSS values estimated from correlation with turbidity.

equilibrium operation was established. The pilot system treated secondary effluent from an extended aeration, activated sludge system. The CBF pilot system was operated in the Blue PRO mode throughout the study with ferric chloride addition to each stage, and with polymer addition to either the first or second stage during the second week of equilibrium operation. The polymer was added in the form of a quaternary amine polymer solution, designated RO-291 from CCI Chemical Company. Each pilot filter was composed of 0.28 m<sup>2</sup> of filtration area with the HFO coated sand media occupying a volume of 0.425 m<sup>3</sup>. The filter backwash flow from each filter was routed to drain. The CBF system operating conditions and observed performance results achieved during the equilibrium period are stated respectively, in Tables 7 and 8. Data derived on three consecutive days over the period are excluded as during those days, a mechanical issue prevented filter backwashing. The analytical results (Table 8) were based on the analysis of two

**Table 7 – CBF Pilot System Operating Conditions: Grangeville, ID<sup>1</sup>**

Parameter	Value
<b>First Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	45
Backwash Flow, l/min	8 to 14
Air Flow, l/h	360
Fe Addition, mg/l	10 to 12
HLR, m/h	9.6
HRT, min	9.4
Temperature, °C	20
Influent pH	7.5 to 7.9
Effluent pH	6.8 to 7.0
<b>Second Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	34
Backwash Flow, l/min	8 to 11
Air Flow, l/h	382
Fe Addition, mg/l	9
HLR, m/h	7.3
HRT, min	12.5
Temperature, °C	20

<sup>1</sup> Values stated are means unless otherwise indicated. Airflow stated at standard conditions. HRT calculated on empty bed basis. Fe added as ferric chloride. Polymer added during second week of equilibrium operation either to first or second stage filter.

<sup>2</sup> The influent flow rate values are set points. The actual flow rate did not vary significantly around each value.

daily grab samples (i.e., samples taken AM and PM). EPA methods 365.1 and 365.4 were followed to determine respectively, total and orthophosphate values. Hach instruments or procedures were used to determine the other analytical values reported.

The observed results demonstrate the ability to consistently achieve an effluent TP level of less than 0.1 mg/l from a two stage Blue PRO CBF system with a relatively small amount of Fe addition, despite treating an influent containing greater than 4 mg/l TP. It is important to note this same level of performance was achieved during the first week of equilibrium operation when polymer was not added to the CBF system. The predicted results from the adsorption/co-precipitation model are in close agreement with the observed values (Table 8).

**Table 8 – CBF Pilot System Performance Results: Grangeville, ID<sup>1</sup>**

Parameter	Value
Influent, mg/l	
TP	4.34 (0.27)
PO <sub>4</sub> -P	4.16 (0.32)
Alkalinity as CaCO <sub>3</sub>	89 (6)
TSS	< 15
First Stage Effluent, mg/l	
TP Observed	0.37 (0.12)
TP Predicted	0.32
PO <sub>4</sub> -P Observed	0.27 (0.13)
PO <sub>4</sub> -P Predicted	0.28
Alkalinity as CaCO <sub>3</sub>	64 (5)
TSS	< 5
Second Stage Effluent, mg/l	
TP Observed	0.041 (0.013)
TP Predicted	0.020
PO <sub>4</sub> -P Observed	0.014 (0.008)
PO <sub>4</sub> -P Predicted	0.009
Alkalinity as CaCO <sub>3</sub>	41 (5)
TSS	< 5

<sup>1</sup> Values stated are means unless otherwise indicated. Bracketed values represent the standard deviation. TSS values estimated from correlation with turbidity.

**Hayden, ID.** Since mid-2005, Blue Water has been operating CBFs at their research facility located at the Hayden Regional Wastewater Treatment Plant in Hayden, ID. When operating in the Blue PRO or Blue CAT mode, the CBFs have treated secondary effluent from the Hayden plant which is an activated sludge system configured as an oxidation ditch. The Blue PRO system at Hayden consists of two, CBF filters operated in series. Each filter is composed of 4.64 m<sup>2</sup> of filtration area with the HFO coated sand media occupying a volume of 7.08 m<sup>3</sup>. The filter backwash flow from each filter is routed to the Hayden plant oxidation ditch. Results from operation of the two stage CBF system in the Blue PRO mode in late 2005 through early 2006 have been discussed elsewhere (Newcombe et al., 2008a; Leaf and Johnson, 2006). The results presented here are based on equilibrium operation during the month of January 2006 beginning on 1/3 through 1/31/06. The conditions during operation of the CBF system and the observed performance results during this period are stated respectively in Tables 9 and 10. Details

**Table 9 – CBF Pilot System Operating Conditions: Hayden, ID**

Parameter	Value
<b>First Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	662
Backwash Flow <sup>2</sup> , l/min	57
Air Flow, m <sup>3</sup> /h	5.10
Fe Addition, mg/l	15
HLR, m/h	8.5
HRT, min	10.7
Temperature, °C	12
Influent pH	6.6 to 7.6
Effluent pH	6.6 to 7.5
<b>Second Stage Filter</b>	
Influent Flow <sup>2</sup> , l/min	606
Backwash Flow <sup>2</sup> , l/min	57
Air Flow, m <sup>3</sup> /h	3.96
Fe Addition, mg/l	10
HLR, m/h	7.8
HRT, min	11.7
Temperature, °C	12
Effluent pH	6.0 to 7.3

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode. Airflow stated at standard conditions. HRT calculated on empty bed basis. Fe added as ferric chloride.

<sup>2</sup> The influent and backwash flow rate values are set points. The actual flow rate did not vary significantly around each value.

regarding the sampling and analytical protocols are discussed elsewhere (Newcombe et al., 2008a). The observed results (Table 10) demonstrate the ability to consistently achieve an effluent TP of less than 0.1 mg/l from a single stage Blue PRO CBF with the addition of a relatively small amount of ferric chloride. There is reasonable agreement between the observed results and those predicted from the adsorption/co-precipitation model (Table 10).

It is important to note the practice of returning the backwash flow to the Hayden plant oxidation ditch had a significant impact on the TP concentration in the secondary effluent or influent to the first stage Blue PRO filter. Prior to introducing the backwash flow, the secondary effluent TP

**Table 10 – CBF Pilot System Performance Results: Hayden, ID<sup>1</sup>**

Parameter	Value
Influent, mg/l	
TP	0.643 (0.0341)
PO <sub>4</sub> -P	0.419 (0.286)
Alkalinity as CaCO <sub>3</sub>	123 (29)
TSS	< 5
First Stage Effluent, mg/l	
TP Observed	0.062 (0.029)
TP Predicted	0.062
PO <sub>4</sub> -P Observed	< 0.004
PO <sub>4</sub> -P Predicted	0.008
Alkalinity as CaCO <sub>3</sub>	84 (34)
TSS	< 5
Second Stage Effluent, mg/l	
TP Observed	0.009 (0.004)
TP Predicted	0.023
PO <sub>4</sub> -P Observed	< 0.005
PO <sub>4</sub> -P Predicted	< 0.001
Alkalinity as CaCO <sub>3</sub>	66 (30)
TSS	< 5

<sup>1</sup> Values stated are means unless otherwise indicated during operation in Blue PRO mode. Bracketed values represent the standard deviation.

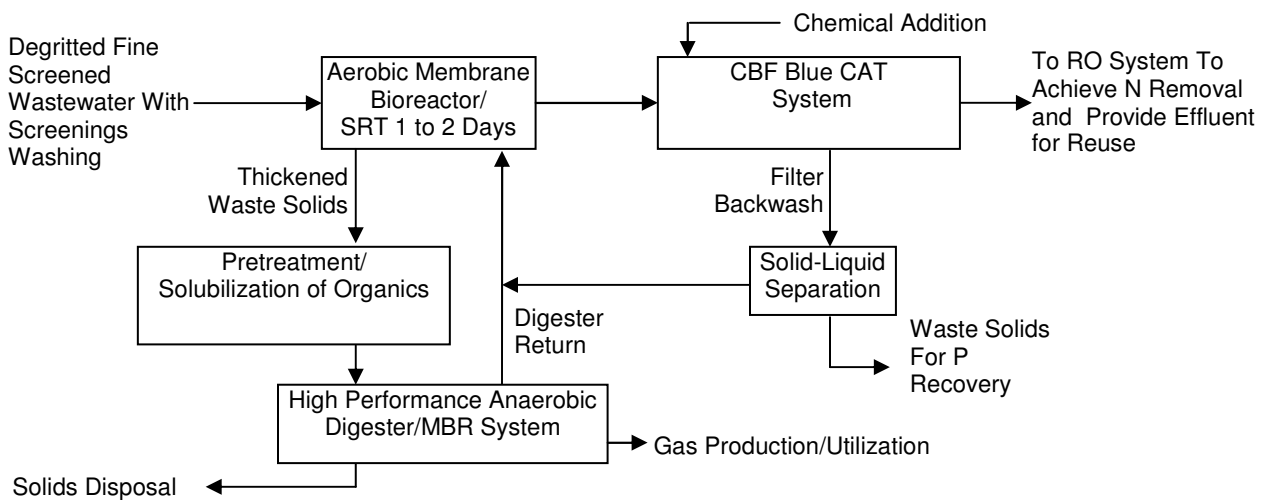
<sup>2</sup> The detection limit (i.e., 0.002 mg/l) was used for values determined as non-detect in completing calculations therefore actual observed value is less than value stated.

was approximately 4 mg/l. After equilibrium was established and accounting for the effects of dilution (i.e., total backwash flow approximately 5 percent of Hayden plant flow), this practice reduced the TP concentration in the secondary effluent by at least 50 percent. On this basis, it can be shown the mass of TP removed across the complete system (i.e., oxidation ditch plus Blue PRO CBFs) as a result of Fe addition to the filters was more than six times greater than that calculated based on the performance of the filters alone.

## CBF Reactive Filtration – AOP Technology Application

Blue Water’s combined CBF reactive filtration – AOP technology referred to as the Blue CAT system, represents an attractive physical-chemical technology for achieving phosphorus removal and recovery as a unit process in a recently developed municipal wastewater treatment flowsheet. This “new flowsheet” achieves energy sustainability, water and nutrient recovery and minimizes the production of residual solids (i.e., biomass) and the release of GHGs. Basin Water holds the proprietary rights to the new flowsheet. One embodiment of the new flowsheet is depicted schematically in Figure 3.

**Figure 3 - Simplified Schematic Representation of the New Flowsheet to Achieve Complete Treatment of Municipal Wastewater with Resource Recovery**



The new flowsheet derives energy from the wastewater by first shunting a large fraction of the organic carbon measured as COD, to a particulate or solids slurry form and then ultimately treating the solids via anaerobic digestion. The aerobic membrane bioreactor (MBR) configuration plays a key role in the new flowsheet in achieving the organic carbon shunt. Use of the MBR and uniquely designed physical-chemical process systems allows for the production of water for reuse, and the capture of phosphorus and nitrogen from the wastewater. Treating the wastewater organics anaerobically with methane/energy recovery, and avoiding biological nitrogen removal and thus the potential for nitrous oxide production, minimizes the release of GHGs. In the near term, the new flowsheet likely will have a greater appeal internationally in countries where wastewater reuse and the green revolution have gained more prominence than in North America. It is likely for economic reasons, the new flowsheet will prove to be particularly attractive at this time for treating smaller municipal wastewater flows (e.g., less than 18,925 m<sup>3</sup>/day), perhaps in the form of a decentralized treatment plant (i.e., satellite plant, water-mining concept).

An objective of the new flowsheet is nutrient removal, and the opportunity for phosphorus and nitrogen recovery. Evaluations to-date imply the Blue CAT system represents an attractive

choice for the phosphorus removal step. The CBF reactive filtration-AOP technology achieves phosphorus removal with a relatively small addition of Fe and oxidation of any residual organics not removed in the upstream low SRT based MBR system (Figure 3) by the addition of a small amount of ozone (i.e., 1 to 3 mg/l). The reactive filtration mechanism minimizes the Fe requirement. The fact the feed to the Blue CAT step contains no suspended solids is likely to further minimize chemical addition requirements and the size of the filter. The mechanisms responsible for phosphorus removal in this application of the reactive filtration process, are as follows;

- phosphorus adsorption onto the chemically coated sand grains,
- sand filtration of the chemically precipitated phosphorus solids formed at the point of chemical addition, and
- MBR system membrane separation of all precipitated or sorbed phosphorus in the return stream from the solid-liquid separator, removed from the sand filter as a result of filter backwashing (Figure 3).

A large fraction of the phosphorus contained in the degrittied and fine screened wastewater feed to the new flowsheet is captured as waste solids from the solid-liquid separator receiving the backwash from the Blue CAT system (Figure 3). This statement is based on results from characterization of the backwash stream by Blue Water at their Hayden, ID research facility. In the context of the new flowsheet, the results (Blue Water, 2006) imply the effective concentration of Fe added to the aerobic MBR will be approximately 2 mg/l. On this basis, only a small mass of phosphorus is expected to be removed across the MBR system as ferric phosphate. Therefore, only a small mass of phosphorus will be present in the solids wasted from the anaerobic digester. This statement assumes near complete release of the phosphorus contained in the digester biomass into the return stream (Figure 3) as a result of designing the new flowsheet coupled pretreatment and anaerobic digestion steps to achieve approximately 80 percent COD reduction commensurate with a high degree of biomass solubilization. The underflow waste solids from the solid-liquid separation step receiving the backwash from the Fe based reactive filtration Blue CAT system, have the potential to represent a slow release, high P containing fertilizer product following dewatering and drying. The fact Fe is found to strongly fix P, and P release from the solids matrix could be controlled by adjustment of the soil pH (Silveria et al., 2006), support this conjecture.

## **DESIGN OF CBFs FOR NITRATE REDUCTION**

When CBFs are designed and operated to achieve biological nitrate reduction they are normally considered biofilm reactors and therefore a biofilm model should be capable of predicting their performance. It is possible to determine the characteristics of the biofilm (e.g., thickness, density) in the FBR and certain other moving bed reactor, fixed-film configurations, affording mechanistic modeling of the reactions occurring in these reactors. Defining the biofilm characteristics in CBFs and conventional static bed fixed-film reactors (e.g., trickling filters) is more difficult, explaining why normally empirical measures are used to design the reactors and explain their performance. Specific to the application of CBFs for nitrate reduction, measures such as the nitrate VLR are used for design purposes, as discussed previously. In the following

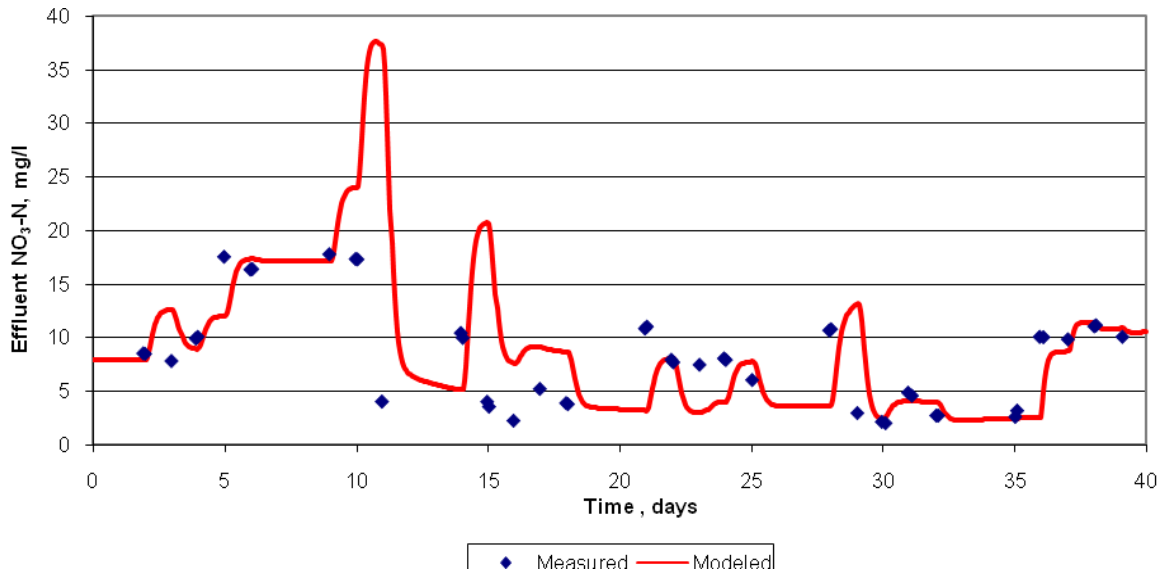
section, data developed from operation of Blue Water’s commercial embodiment of the CBF technology for nitrate reduction (i.e., Blue NITE system) was used to develop a model for the system. The validity of the model is discussed in this section and the subsequent section entitled, CBF Denitrification Performance at Other Locations.

### CBF Nitrate Reduction Modeling

A model for the CBF Blue NITE system was developed in GPS-X as a new process model using the existing mechanistic, dynamic, 1-D trickling filter biofilm model (Hydromantis, 2006) in GPS-X (with air transfer from the atmosphere removed) and the TSS removal component of the reactive filtration process model, discussed previously. The biofilm model includes equations for diffusion across the boundary layer between the bulk liquid and the biofilm, diffusion and biological growth in the biofilm, and attachment and detachment of solids to and from the biofilm. A continuous reject/backwash flow stream is incorporated into the model. The ASM3 biological model (Henze et al., 2000) was used to represent the biological processes occurring in the CBF. The data used for calibration of the model was derived from a Blue NITE CBF pilot system. Details regarding the calibration procedure and the specific input data used, are provided elsewhere (Schraa, 2009). The calibration procedure involved only the adjustment of the TSS removal model parameters. The default parameters within the biofilm model including the ASM3 model parameters, were not adjusted.

A plot (Figure 4) of the modeled versus measured effluent nitrate-N during the period of model calibration indicates the model reasonably depicts the performance of the CBF given the level of error expected with field pilot plant measurements and certain deficiencies in the data base (Schraa, 2009).

**Figure 4 – Modeled Versus Measured CBF Effluent Nitrate Values**



## CBF Denitrification Performance at Other Locations

Since mid-2000, Blue Water has installed a number of pilot or full scale CBF Blue NITE systems. Information obtained from Blue Water from operation of five of these systems is summarized in Table 11. The predicted effluent NO<sub>x</sub>-N values are those derived from the Blue NITE CBF system model. The following should be noted with respect to applying the model to determine the predicted values at the various sites.

- At the Hayden, ID location where methanol was the carbon source, the ASM3 stoichiometric and kinetic parameters were adjusted to be consistent with the stoichiometric and kinetic data for methylotrophs as established by Dold et al. (2007). This includes using a lower yield and a lower maximum specific growth rate for the heterotrophs. The anoxic reduction factor was set to one as it assumed that the filter is primarily anoxic so that most of the heterotrophic bacteria present would be capable of denitrification. The storage yields were set to one as the storage process is assumed to not be utilized by methylotrophic bacteria. The methanol dosing rate used was 4.5 g methanol as COD per g NO<sub>3</sub>-N.
- At the other locations where Unicarb DN or MicroC were the carbon sources, the default ASM3 stoichiometric and kinetic parameters were used recognizing the carbon sources are made up of various carbon compounds. The anoxic reduction factor was set to one as it was assumed that the filter is primarily anoxic so that most of the heterotrophic bacteria present would be capable of denitrification. The storage yields were set to one as it was assumed that storage is not an important phenomenon in a tertiary filter as alternating feast and famine conditions are not present. At the Bellefonte location, a dosing rate of 8 g Unicarb DN as COD per g NO<sub>3</sub>-N was used. At the locations where MicroC was used, a dosing rate of 6.45 g MicroC as COD per g NO<sub>3</sub>-N was used (Cherchi et al., 2008).

Analysis of the operating and performance results presented in Table 11 reveals the following.

- The CBFs operated at nitrate-N VLRs between approximately 0.5 and 2.2 kg/m<sup>3</sup>·day. The lowest rate corresponded to operation of pilot filters at the Hayden, ID and Bellefonte, PA locations. At the Hayden site, the CBF was operated to achieve both phosphorus removal and nitrate reduction (Table 11, footnote 2). At the Bellefonte site, it is expected the nitrate removal rate was limited by the amount of external carbon added to the CBF (Table 11, footnote 4). These factors may account for operation at VLRs well below the anticipated maximum value, as stated previously.
- The model predicted effluent NO<sub>x</sub>-N results are in reasonable agreement with the observed values.

## SUMMARY

CBFs have been widely applied for suspended solids and phosphorus removal, and nitrate reduction. A more recent development related to the phosphorus removal application is the use of a reactive media, namely HFO coated silica sand in CBFs, to achieve effluent TP levels less

than 0.1 mg/l. With the use of this media, phosphorus removal occurs by adsorption of phosphate onto preformed HFO and co-precipitation of phosphate into the HFO structure.

**Table 11 - Operating Conditions and Observed Versus Model Predicted Performance Results For Blue Water Denitrification CBF Systems<sup>1</sup>**

Operating or Performance Measure	Value at Pilot or Full Scale CBF Location				
	Hayden, ID <sup>2</sup>	Bellefonte, PA	Westford Valley, MA	Lynfield, MA	Hanover, MA
<b>Operating Conditions</b>					
Influent Flow, l/min	190	64	76	38	132
Backwash Flow, l/min	19	6	7	4	13
Bed Depth, m	1.52	1.72	1.02	1.02	1.02
Filtration Area, m <sup>2</sup>	4.6	1.1	1.8	1.1	3.5
Carbon Source	Methanol <sup>3</sup>	Unicarb DN <sup>4</sup>	Micro C <sup>3</sup>	Micro C <sup>3</sup>	Micro C <sup>3</sup>
Temperature, °C	11-15	10-14	10-16	10-16	10-16
<b>Influent Characteristics, mg/l</b>					
NO <sub>3</sub> -N	13	10.5	37.5	37.5	40
NO <sub>2</sub> -N	< 2	< 1	< 1	< 1	< 1
DO	3	5.5	< 2	< 2	< 2
TSS	< 5	10	15	15	15
<b>Effluent Results, mg/l</b>					
Observed NO <sub>x</sub> -N	≤ 1.5	≤ 3.5	≤ 2.0	≤ 2.0	≤ 2.0
Predicted NO <sub>x</sub> -N	0.55	0.18	1.1	0.21	0.59

<sup>1</sup> Observed performance results stated should be considered mean values based on operation at equilibrium under operating conditions stated. Information provided by Blue Water.

<sup>2</sup> CBF operated to achieve P removal and nitrate reduction. Ferric chloride added to promote P removal at 15 mg/l Fe.

<sup>3</sup> Carbon source added in excess of stoichiometric value (i.e., COD basis) necessary to reduce influent DO and NO<sub>x</sub>-N concentrations.

<sup>4</sup> Unicarb DN addition rate may have limited NO<sub>x</sub>-N removal rate.

Referred to as reactive filtration, the process has been embodied commercially by Blue Water as the CBF Blue PRO system. The results from a number of pilot plant case studies demonstrate the ability to achieve an effluent TP concentration as low as 10 µg/l utilizing a two stage Blue PRO system in the treatment of municipal wastewater secondary effluents. The CBF reactive filtration technology can be operated as a combined phosphorus removal and AOP by the addition of a small amount of ozone to the filter influent. The combined CBF reactive filtration-AOP technology represents an attractive physical-chemical technology to achieve advanced

oxidation, and phosphorus removal and recovery in a new flowsheet designed to achieve energy sustainability, and water and nutrient recovery.

A mathematical model was developed for the reactive filtration phosphorus removal technology consisting of a TSS removal component combined with a component to represent the adsorption and co-precipitation processes. The adsorption/co-precipitation model component was used to predict the effluent phosphorus values observed in various Blue PRO pilot plant case studies. The model predictions and observed results were generally in agreement.

The design of CBFs for nitrate reduction typically relies on specification of the value of an empirical parameter such as the nitrate VLR. Data developed from Blue Water's Blue NITE system, a commercial embodiment of the CBF technology for nitrate reduction, was used to develop a model for the system. Observed results derived from various locations where the denitrification CBF system was in operation, generally agreed with the model values.

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## CREDITS

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