



Cost effectiveness of phosphorus removal processes in municipal wastewater treatment

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HIGHLIGHTS

- IFAS-EBPR is the most cost-effective system (\$42.25/lb-P removed; $P_{\text{eff}} = 0.82$ mg/L).
- MBR systems are the least cost-effective (\$60.89/lb-P removed; $P_{\text{eff}} = 1.02$ mg/L).
- Tertiary filtration (high investment) is cost effective due to low $P_{\text{eff}} = 0.05$ mg/L.
- Side stream struvite recovery is not helpful in meeting stringent effluent TP limits.
- Due to high capital and maintenance costs, MBR systems are not cost effective.

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ABSTRACT

Meeting stringent phosphorus (P) discharge standards remains one of the major challenges for wastewater utilities due to increased economic burdens associated with advanced (i.e., secondary, tertiary) treatment processes. In a trade-off between higher treatment cost and enhanced P removal, it is critical for the treatment plants to be able to select the most appropriate technology. To this end, established/emerging high performing P removal/recovery technologies (e.g., Modified University of Cape Town process, Bardenpho process, membrane bioreactors, IFAS-EBPR, struvite recovery, tertiary reactive media filtration) were identified and full-scale treatment plant designs were developed. Using advanced mathematical modeling techniques, six different treatment configurations were evaluated in terms of performance and cost effectiveness (\$/lb of P removed). Results show that the unit cost for P removal in different treatment alternatives range from \$42.22 to \$60.88 per lb of P removed. The MUCT BNR + tertiary reactive media filtration proved to be one of the most cost effective configurations (\$44.04/lb P removed) delivering an effluent with total P (TP) concentration of only 0.05 mg/L. Although struvite recovery resulted in significant reduction in biosolids P, the decrease in effluent TP was not sufficient to meet very stringent discharge standards.

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1. Introduction

The social, economic and environmental issues associated with uncontrolled phosphorus (P) release to surface waters have been recognized for decades (Chislock et al., 2013). Consequently, increasing regulatory impetus to reduce P release to the environment has resulted in tightening of discharge permit limits for municipal wastewater treatment plants (WWTPs). With this

continuing trend, there has been an ongoing need for investment (i.e., infrastructure, chemical, energy) to efficiently remove/manage P in domestic wastewater. With the world's increasing population and growing energy crisis, modern WWTPs are also faced with the new challenge of shifting to sustainable, and energy/ resource positive operations (Wang et al., 2017a). Currently, significant efforts are being expended to integrate low-energy consumption processes with resource (e.g., energy, nutrients) recovery in wastewater treatment layouts (Chen et al., 2015). To address the huge challenge of ensuring sustainable treatment and simultaneous maximum pollutant removal, careful assessment of treatment cost will be required.

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Nomenclature

WWTP	wastewater treatment plant
TP	total phosphorus
BNR	biological nutrient removal
AOB	ammonia oxidizing bacteria
GHG	greenhouse gas
POTW	publicly owned treatment works
EBPR	enhanced biological phosphorus removal
MBR	membrane bioreactor
WAS	waste activated sludge
SRT	solids retention time
HRT	hydraulic retention time
MUCT	Modified University of Cape Towne
IFAS	integrated fixed-film activated sludge
PAO	polyphosphate accumulating organism
TMDL	total maximum daily load

Several established and emerging P removal technologies have been practiced on full-scale systems. Among the established P removal technologies, enhanced biological phosphorus removal (EBPR) and chemical precipitation have the most widespread use (Pratt et al., 2012). However, biological treatment schemes use substantial amount of energy for aeration (Foley et al., 2010). These systems are also known to be inconsistent in achieving low P concentrations in the effluent (Oehmen et al., 2007; Pratt et al., 2012). Chemical P removal is generally an expensive treatment option and these systems are known to increase sludge production. Also, PO_4^{3-} forms a strong bond with Al^{3+} and Fe^{3+} metal ions which often reduces its plant availability in the biosolids (Tian et al., 2016).

Few emerging alternative P removal technologies implemented on a full-scale basis that show great promise include: (i) Membrane bioreactors (MBRs), (ii) Integrated Fixed-Film Activated Sludge Systems with Enhanced Biological Phosphorus removal (IFAS-EBPR), and (iii) Continuous flow, no backwash, upflow, deep-bed granular media filters (USEPA, 2013). MBRs have gained popularity over the last 10–15 years due to smaller footprint and ability to meet stringent discharge limits with high process reliability (Smith et al., 2015). However, membrane fouling remains a critical obstacle for widespread implementation of MBRs as it decreases the membrane lifespan and increases energy requirements for sludge recirculation. IFAS-EBPR is a hybrid process that enables separate solids retention time (SRT) control for slower growing nitrifiers and faster growing heterotrophs (i.e., denitrifiers or Polyphosphate Accumulating Organisms (PAOs)), which is a significant advantage over conventional biological processes (Bai et al., 2016). Continuous backwash, upflow, deep-bed granular media filter units combine co-precipitation and adsorption of P on a reactive media filter (USEPA, 2013). Full-scale implementations of this technology for tertiary treatment are meeting effluent total P (TP) limits of 0.05 mg/L (USEPA, 2007).

Over the past 20 years though, P found in wastewater streams has been recognized as a recoverable product rather than a pollutant (Desmidt et al., 2013). Several P recovery technologies have been developed that can produce high-grade P minerals, in the form of struvite or Ca phosphate, for use as fertilizer. Currently, fourteen full-scale struvite recovery systems are operational in North America (Ostara, 2017) with most of the commercially-available processes primarily implemented on the side streams.

A major focus of the Federal Water Pollution Control Act and its

amendments has been to improve surface water quality through control of point source P inputs from wastewater treatment plants to waterbodies. A review of USEPA's Discharge Monitoring Report (DMR) data shows that states like Florida, Michigan, New Hampshire, Massachusetts, Maryland, Delaware and Vermont have already set P permit limits for more than 40% of the publicly owned treatment works (POTWs). Among these states, Massachusetts (0.1 mg/L), New Hampshire (0.2 mg/L), Michigan (0.3 mg/L) and Maryland (0.3 mg/L) have the lowest TP permit limits for one or more POTWs (USEPA, 2017). Currently, most US treatment plants have a TP limit in the range of 0.5–1.5 mg/L. With the USEPA focusing towards allotment of total maximum daily loads (TMDL) for P management in river basins, it is evident that future discharge limits will be increasingly more stringent. Restricting P loads from dischargers will inevitably exert a significant financial burden on the municipalities. Greater emphasis will also be placed on recovering P from wastewater due to its worldwide dwindling high-quality supplies (Desmidt et al., 2013). To keep up with the future trends, it is very important for the treatment facilities to identify opportunities for upgrade and develop robust long-range economic plans.

To date, not much research has been conducted to compare the operational and economic advantages/disadvantages of established and emerging P removal systems. Several studies have examined established competing P removal configurations in terms of performance and cost effectiveness (Falk et al., 2013; Hao et al., 2001; Jiang et al., 2004, 2005; Ohio EPA, 2013; WSDOE, 2011; Zhang et al., 2009). Jiang et al. (2004, 2005) estimated the cost of P removal from municipal wastewater for facilities constructed *de novo* and adaptation of existing facilities. From the analysis, they concluded that for an effluent TP concentration between 0.5 and 2.0 mg/L, activated sludge (AS) process followed by alum addition is the most economical configuration. However, under a TP limit of 0.13 mg/L, three step Biological nutrient removal (BNR), chemical addition and tertiary filtration was found to be the most cost effective configuration for a large treatment facility (>10 MGD) by Jiang et al. (2004, 2005). WSDOE (2011) includes cost curves that can be used to estimate costs of certain treatment processes scaled to design capacity. However, most of these studies are limited in scope in terms of the number of alternative process configurations considered. While the established treatment technologies (e.g., AS, chemical precipitation) were the main focus, emerging treatment alternatives (e.g., tertiary filtration, fixed film systems, membrane processes) were ignored in most studies. Also, recycling and recovery of P from the sidestream was not considered, which is a significant limitation.

Therefore, the major goal of this study is to perform an advanced mathematical model based technical and economical evaluation of the high-performing P removal/recovery processes (secondary and tertiary) that have been implemented on full-scale systems. This goal encompasses the following objectives/components:

1. Technical and economic evaluation of established and emerging secondary P removal processes;
2. Technical and economic evaluation of implementing a side stream P recovery system in conjunction with a mainstream P removal process;
3. Evaluation of process modifications and incremental operational costs to achieve a specific nutrient removal goal (effluent $\text{TP} \leq 0.05$ mg/L) using tertiary treatment; and
4. Comparison of treatment technologies in terms of overall annual economic cost (total capital, operational and maintenance costs) and cost effectiveness (\$/lb of P removed).

2. Methodology

2.1. Treatment process scenarios and simulation

For the analysis, six full-scale treatment scenarios were considered. Each configuration consisted of preliminary and primary treatment, mainstream P removal, sludge line (thickening, dewatering and mesophilic anaerobic digestion of primary and waste activated sludge), and biogas recovery from the sludge. The six P removal/recovery processes considered are: (i) Modified University of Cape Towne (MUCT) process, (ii) five-stage Bardenpho Process, (iii) membrane bioreactors (MBRs), (iv) Integrated Fixed-Film Activated Sludge Systems with Enhanced Biological Phosphorus Removal (IFAS-EBPR), (v) struvite recovery by chemical precipitation, and (vi) tertiary media filtration. Relevant publications on nutrient removal systems by USEPA (Ohio EPA, 2013; USEPA, 2007) were reviewed to identify the above high performing P removal/recovery technologies.

The general layouts of the whole plant process configurations considered in this study are presented in Fig. 1. **Scenario 1 (S1)** represented MUCT process nutrient removal configuration (anaerobic, primary anoxic, secondary anoxic and aerobic). This system consists of a two-step anoxic process to avoid the intrusion of nitrate in the anaerobic zone, which results in enhanced P removal performance (WEF, 2008).

Scenario 2 (S2) comprised “advanced” nutrient removal using 5-stage (anaerobic, primary anoxic, primary aerobic, secondary anoxic, secondary aerobic) Bardenpho process configuration. The Bardenpho process has been implemented in many developed

countries for excellent nutrient (N & P) removal (Foley et al., 2010). **Scenario 3 (S3)** involved a MUCT process activated sludge treatment and a MBR. **Scenario 4 (S4)** represented an IFAS-EBPR process configuration with side-stream sludge treatment. The mainstream P removal configuration simulated in S1 (i.e., MUCT process) was coupled with side-stream P recovery from anaerobic digestate to come up with **Scenario 5 (S5)**. A similar full-scale treatment configuration has been successfully implemented at the local Nine Springs WWTP (Madison, WI). The sludge from anaerobic digestion was thickened and the supernatant containing elevated levels of soluble P entered the struvite precipitator. Magnesium chloride (MgCl_2) was added in the struvite precipitator based on a 1:1 molar ratio of the supernatant P feed. To maintain an operating pH of 8 in the reactor, sodium hydroxide (NaOH) was also added to the reactor. In **Scenario 6 (S6)**, a reactive media filtration step (continuous backwash, upflow, deep-bed granular media filter) was added to the process configuration considered in S1 to reduce the effluent TP level to 0.05 mg/L. The secondary system in S6 is intended to lower effluent TP to ≤ 1 mg/L using biological treatment. This step was followed by hydrous ferric oxide (HFO) coated sand filters for adsorptive P removal. The reject stream from the filtration system was recycled to the secondary system to allow for P uptake by the excess adsorptive capacity remaining in the HFO waste particulates. The chemically-bound P was removed with the wasted sludge. To regenerate adsorptive capacity of the HFO coated sand, ferric chloride (FeCl_3) was dosed before the filters (Newcombe et al., 2006a, 2006b).

All six scenarios were constructed using a widely-used process model simulating software BioWin v. 5.2 (EnviroSim Associates

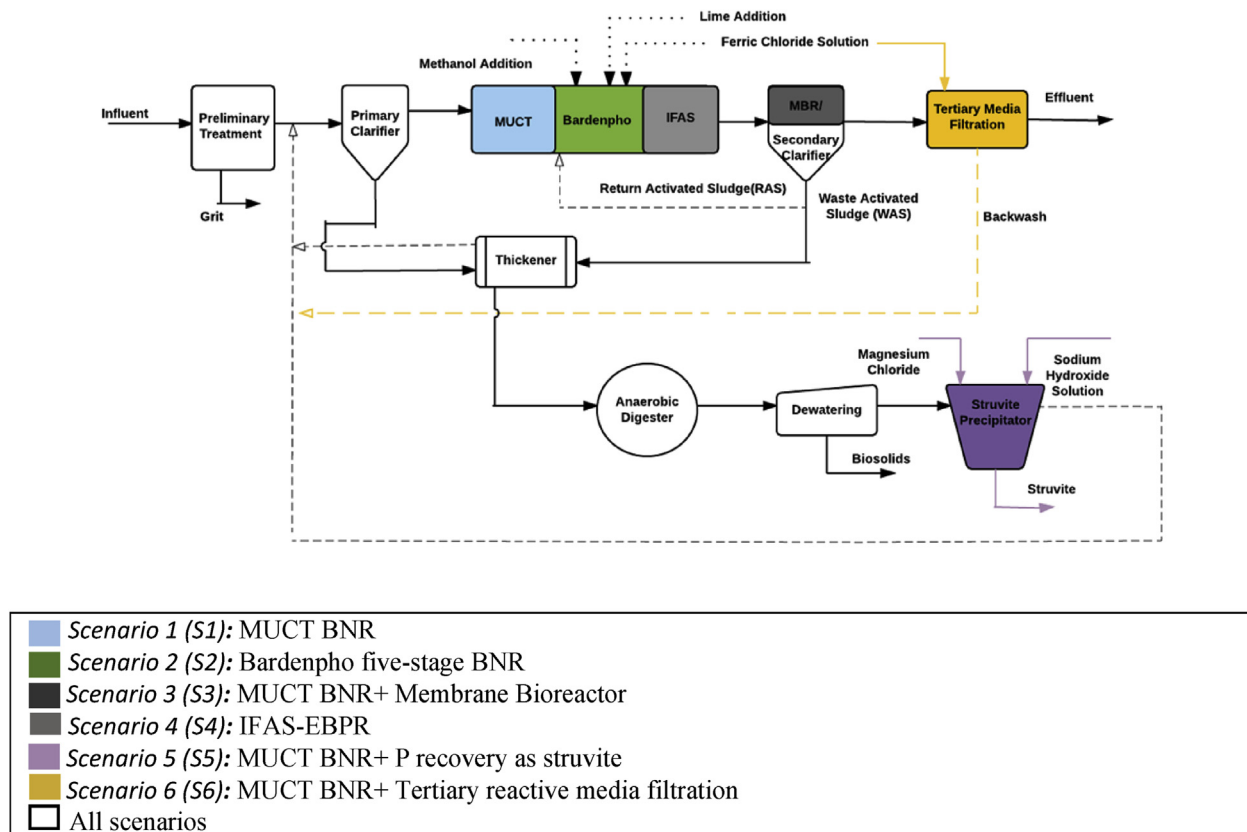
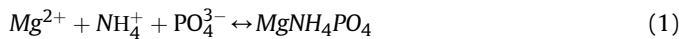


Fig. 1. General layout of process flow diagrams for six treatment scenarios simulated in this study. Preliminary, primary, and sludge treatments (thickening, dewatering and mesophilic anaerobic digestion) were common to all the treatment configurations. Secondary clarifier was replaced by MBR in S3. The ferric salt addition location was transferred from the anoxic reactor to tertiary media filter in S6.

Ltd., Canada). BioWin uses an integrated kinetic model and mass balance approach, incorporating pH/alkalinity and general Activated Sludge/Anaerobic Digestion Models (Envirosim, 2007). The most recent version of the software (v. 5.2) also enables users to obtain a plant-wide inventory of power demand and operating cost. Steady state simulations, with an effluent TP goal < 1 mg/L and Total Inorganic Nitrogen (TIN) < 8 mg/L were performed for S1 to S5. Biological P removal processes were combined with chemically-assisted precipitation of P to reach the effluent goal where necessary. For S6, a stringent target effluent TP limit 0.05 mg/L was considered. These treatment objectives were selected based on the generally accepted performance of established and emerging nutrient removal technologies for municipal WWTPs, as suggested in WSDOE (2011).

2.2. Calibration of P recovery module

To simulate struvite precipitation, BioWin's default chemical precipitation module (developed based on Musvoto et al., 2000) was calibrated and validated using literature data from full- and/or pilot-scale studies. The precipitation of struvite or magnesium ammonium phosphate (MAP) is expressed as:



The rate of crystallization was expressed as (Musvoto et al., 2000):

$$\frac{d}{dt} [\text{MgNH}_4\text{PO}_4] = R_{\text{precip}} \left[[\text{Mg}^{2+}]^{1/3} [\text{NH}_4^+]^{1/3} [\text{PO}_4^{3-}]^{1/3} - K_{\text{spMgNH}_4\text{PO}_4}^{1/3} \right]^3 \quad (2)$$

where:

Mg^{2+} , NH_4^+ and PO_4^{3-} correspond to soluble magnesium, ammonium and phosphate species
 MgNH_4PO_4 represents struvite or MAP, and
 $K_{\text{spMgNH}_4\text{PO}_4}$ is the struvite solubility product. In the above model, the equation is valid only if $[\text{Mg}^{2+}]^{1/3} [\text{NH}_4^+]^{1/3} [\text{PO}_4^{3-}]^{1/3} > K_{\text{spMgNH}_4\text{PO}_4}^{1/3}$.

Three independent sets of experimental data (Battistoni et al., 2001; Jaffer et al., 2002; Pastor et al., 2010) were used for calibration and three additional independent datasets (Pastor et al., 2008; Rahaman et al., 2008, 2014) were used for validation. The dataset for calibration was chosen such that it covers a wide range of pH (7.5–8.9) and Mg:P ratios (0.8–1.7). The module with the calibrated precipitation rate constant parameter (R_{precip}) minimized the Mean Squared Error (MSE) based on measurements of orthophosphate removal. The MSE values used for model calibration can be defined as:

$$\text{MSE} = \frac{1}{n} \sum_{i=1}^n (Y_{\text{exp}} - Y_{\text{model}})^2 \quad (3)$$

where,

Y_{exp} = Experimental data on orthophosphate removal (from literature),
 Y_{model} = Modeled orthophosphate removal values,
 n = Number of observations.

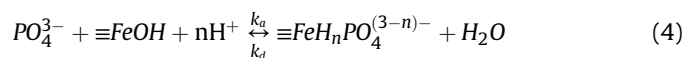
An iterative approach for model calibration was adopted until the MSE value could not be further minimized. Detailed calibration

approach for the struvite recovery module is included in Supplementary Information (S1).

2.3. Simulation of tertiary media filtration

To model adsorptive P removal in BioWin, a 'model builder reactor' element was used in conjunction with a solids removal unit. The adsorption process occurred in the model builder element. The solids removal unit was configured such that it selectively retained the media and recycled back to the system. BioWin's user defined variable UD4 was used to represent the adsorbed component. The reactive media filtration system was designed to consist of six Centra-flo moving bed sand filters (model CF-200, Blue Pro, Nexom.). Each filter has a filtration surface area of 18.6 m² (200 ft²) with a hydraulic loading rate of 578.7 gpm, which is consistent with the manufacturer's guidelines. To regenerate the media periodically, ferric salt was added prior to tertiary media filtration. The application rate was set to 15 mg Fe/L, which is consistent with the typical application rates in full-scale systems of similar influent and effluent P concentration (Newcombe et al., 2006b).

The kinetic model developed by Mao and Yue (2016) was adopted to simulate adsorption of phosphate by preformed HFO in BioWin. The surface complexation reaction was conceptualized as:



k_a and k_d are adsorption and desorption reaction rate constants
 PO_4^{3-} represents the dissolved phosphate species
 $\equiv\text{FeOH}$ represents the active surface sorption sites on the media and, $\equiv\text{FeH}_n\text{PO}_4^{(3-n)-}$ is representative of the surface complex that forms over the HFO media

The rate of adsorption was expressed by the following equation:

$$\frac{d}{dt} (P_{\text{ads}}) = k_a [\text{PO}_4^{3-}] [\equiv\text{FeOH}] [\text{H}^{n+}] - k_d [P_{\text{ads}}] \quad (5)$$

where, P_{ads} represents the concentration of adsorbed P and n represents the number of protons associated with each surface complex. Experimentally determined values for k_a , k_d and number of active sites were obtained from Mao and Yue (2016).

2.4. Influent characteristics and design parameters

Each treatment configuration was simulated under identical influent flowrate and characteristics to facilitate direct comparisons. It was assumed that the facility is a mid-sized U.S. treatment WWTP with a capacity of 5 MGD (average dry weather flow). To establish the wastewater load characteristics, the monthly average data for the time period of January 2010 to November 2015 were collected from the local Nine Springs Wastewater Treatment Plant (Madison, WI). The averaged data (Table 1) were introduced into BioWin. Influent COD, TKN and TP fractionation parameters were calculated from the collected data (S2.1 in Supplementary Information). The influent composition thus obtained was further compared to the typical municipal wastewater characteristics suggested by Tchobanoglous et al. (2013) to ensure all the concentrations were within the typical range.

Tank sizes and process parameters, such as, hydraulic retention time (HRT), SRT, were established according to the standards set forth in the WEF manual (WEF, 2008). Summary of design

Table 1

Influent operational data from Nine Springs Wastewater Treatment Plant, Madison, WI (N = 71).

Parameter	Mean	Min	Max	SD	Typical range ^a
Flow, million gallons per day (MGD)	38.0	32.0	48.4	3.9	–
Chemical Oxygen Demand (COD), mg/L	507	450	540	28.2	250–800
Biochemical oxygen demand (BOD ₅), mg/L	239.4	177.8	295.6	24.2	110–350
pH	7.5	7.3	9.6	0.3	7.0–8.0
Total Kjeldahl Nitrogen (TKN), mg/L	42.5	32.5	49.0	4.1	20–70
Total Phosphorus (TP), mg/L	5.7	4.6	6.7	0.5	4–12
Total Suspended Solids (TSS), mg/L	230.0	183.0	277.3	19.5	12–400
Calcium (Ca), mg/L	86.5	85.3	89.3	1.1	–
Magnesium (Mg), mg/L	45.6	44.8	46.8	0.6	–
NH ₃ -N, mg-N/L	26.8	20.1	32.7	3.1	20–75
Alkalinity, eqv/m ³	4.8	4.6	5.0	0.2	1–7

Min = Minimum, Max = Maximum, SD = Standard Deviation.

^a Tchobanoglous et al. (2013).

parameters for the treatment configurations are presented in S3 in Supplementary Information.

2.5. Cost effectiveness of process configurations

The cost-effectiveness of the configurations was determined using the following equation:

$$CE_K = \frac{TAC_K}{PE_K} \quad (6)$$

where,

CE = cost-effectiveness (\$/lb-P removed),
 TAC = total annualized cost (\$),
 PE = amount of P removed in pounds (lb-P), and
 k = process configuration.

The overall annualized expenses associated with a process configuration consist of capital investments, operation and maintenance costs, and was calculated as follows (Jiang et al., 2004):

$$TAC_K = (C_{Ca} * CRF) + C_{OM} \quad (7)$$

where,

C_{Ca} = total capital investment cost,
 C_{OM} = annual operation and maintenance cost, and
 CRF = capital recovery factor

Capital costs were annualized assuming a CRF of 8% and 20 year life-span of the WWTP. Cost of land, civil engineering, labor and construction were excluded because they are all highly dependent on location.

2.5.1. Capital costs

The capital cost for scenarios S1 and S2 was obtained from Ohio EPA (2013) and USEPA (2007), respectively. The aforementioned studies derived capital costs for EBPR systems based on case studies, “CAPDEnetworks” and literature sources for expansion and modifications over secondary treatment configurations. These estimates were updated to reflect conditions in 2017 using construction cost index from the Engineering News Records (ENR, 2017). For capital investment cost and membrane replacement reserve in scenario S3, pricing information of MBR was obtained on ZeeWeed 500 membrane from GE Power (GE Corporation, 2017). The moving bed biofilm reactors (MBBRs) for IFAS process in scenario S4 were designed to be filled with plastic biofilm carriers (500 m²/m³ protective surface area) at a biofilm-to-media filling

ratio of 33%. The pricing information for the media was obtained from SEWPCC (2008). For scenario S5, the investment costs for struvite recovery reactor was obtained using Water Environment Research Foundation (WERF) P recovery tool from Ostara Nutrient Recovery Technologies Inc. The cost for the sand filter in scenario S6 was obtained from Nexom Blue Pro (Nexom, 2017), as they have performed several successful full-scale implementations of this technology in municipal WWTPs (USEPA, 2013).

2.5.2. Operation and maintenance costs

The operation and maintenance costs consisted of four items: energy, chemical, sludge disposal and maintenance & insurance. For estimation of energy requirements, aeration in the oxic zone, liquid mixing in the anoxic and anaerobic zone, pumping of solids, and mixing and heating for anaerobic sludge digestion unit were considered. BioWin aeration model was used to determine the oxygen transfer rate. The electricity consumption for the aeration unit was subsequently calculated by assuming a standard oxygen transfer efficiency of 6.5% per meter of reactor depth, based on values typically stated by vendors of fine bubble aeration diffusers (Foley et al., 2010).

Electricity consumption for all pumps was estimated from individual flow rate and the assumed pumping head (identical for all scenarios). Hydraulic efficiency of pumps was estimated from standard curves with a motor efficiency of 95% (Wang et al., 2012a). The stirring power for liquid mixing in the anoxic and anaerobic zones was assumed to be 5 W/m³ liquor (Envirosim, 2007). The amount of heat (KWh) required per wet metric ton (1000 kg) of sludge during anaerobic digestion was calculated from the difference between the initial and desired temperatures multiplied by specific heat capacity of sludge with 6% solids content (1117 KWh/kg deg C) and the heat loss from the digester using available heat transfer coefficients (Wang et al., 2012b).

Chemical addition for the treatment configurations primarily consisted of FeCl₃ for P removal, lime addition for alkalinity correction, and methanol dosing for carbon source. For the wastewater entering the denitrification system, the methanol-to-nitrogen (nitrate) ratio was kept at 3:1 based on typical amount used in full-scale facilities. The degree of chemical use was acquired directly from the BioWin simulator. Unit costs for chemicals used for the study are listed in S4.2 in Supplementary Information.

Repair and maintenance costs associated with scenarios S1, S2 and S6 were calculated as 4% of the total capital expenditure (Jiang et al., 2004). To account for the mechanical cleaning requirements in S3, an additional 1.5% of capital cost associated with membranes was added to the maintenance cost considered for the MUCT process in S1 (SEWPCC, 2008). The amount of 50% citric acid and 12.5% sodium hypochlorite required for chemical cleaning of membranes

was estimated according to WSDOE (2011). The maintenance costs associated with Scenario S4 was calculated according to SEWPCC (2008). In Scenario S5, it was assumed that citric acid cleaning of the reactor will be required every four months, which is similar to the full-scale operation at the local Nine Springs WWTP. The cost due to taxes and insurance was calculated as 2% of the total capital expenditure in all the treatment scenarios (Jiang et al., 2004).

Methane production from anaerobic digestion is an economic benefit and was included in the operating cost calculation as a negative cost. The power and heat efficiency of the CHP technology was 33% and 35%, respectively (Wan et al., 2016). Since recovered struvite is expected to be land applied as fertilizer, it was assumed that synthetic fertilizer (mono-ammonium phosphate, MAP) will be replaced. The cost of recovered struvite product (Table S3) was added as a credit to that particular scenario.

3. Results and discussion

3.1. Effluent quality

The TP removal efficiencies in six different treatment scenarios ranged from 82.4% to 99.1% (Table 2). Among the scenarios, S3 and S6 were particularly efficient in solids removal attributable to the filtration mechanism of membrane bioreactors and the media filters to remove total suspended solids (TSS).

Scenarios S1 and S5 are directly comparable, since the only difference between the two is the addition of side-stream struvite recovery process in S5. Simulation results show that the inclusion of struvite recovery in S5 increased the TP removal by 6%, compared to S1. This is primarily due to the reduction of P concentration in the recycle stream after struvite recovery. Configuration S4 was able to achieve the target N removal goal at a shorter SRT of 7.5 days than the other scenarios where the SRT ranged from 10 to 12 days. All the

treatment scenarios needed lime addition for alkalinity correction in the nitrification process. Other than S4, all the configurations also required significant amount of methanol addition for heterotrophic denitrification process. In general, all of the treatment scenarios could achieve very low concentrations of $\text{NH}_4\text{-N}$ (≤ 1 mg/L) and nitrite (≤ 0.5 mg/L) in the effluent.

3.2. P mass balance

On a mass basis, between 0.9 and 17.6% of the influent TP was found in the effluent for the six treatment scenarios considered (Table 3). Other than S5, the remaining five scenarios removed 82 to 99 percent of influent P mass from wastewater that ended up in the biosolids. Importantly, implementation of the struvite P recovery system (S5) resulted in a substantial reduction (35.6%) in biosolids P content compared to the highly efficient BNR process MUCT (S1), which focuses only on P removal. Overall, for all treatment configurations, biosolids is the major outlet for P other than recovery as struvite. Safe and efficient application of biosolids is a critical component of nutrient management on a watershed basis. Biosolids normally supply similar amounts of plant-available P as N, but crops require only one-fifth to one-half as much P as N (Evanylo, 2009). It should be pointed out here that, harvesting P (as struvite) from the side-stream of a treatment process could have a secondary benefit for biosolids management by yielding a more agronomically-balanced fertilizer product.

3.3. Total cost & energy requirements

In this section, operational costs and energy for different treatment configurations are compared in terms of net energy requirements, aeration requirement, chemical and sludge disposal costs, and energy recovery. These parameters are expected to have

Table 2
Effluent quality for the six treatment scenarios evaluated in this study.

	S1 (MUCT)	S2 (Bardenpho)	S3 (MBR)	S4 (IFAS-EBPR)	S5 (P Recovery)	S6 (Tertiary Filtration)
Effluent						
Total BOD, mg/L	2.05	1.20	0.80	1.36	1.46	1.04
Total COD, mg/L	33.41	30.10	28.47	29.18	29.79	29.90
TSS, mg/L	2.87	1.24	<0.01	1.28	1.85	<0.01
Ammonia N, mg/L	0.62	0.06	0.08	0.35	0.31	0.62
$\text{NO}_3\text{-N}$, mg/L	7.04	5.29	6.99	7.36	7.15	7.05
$\text{NO}_2\text{-N}$	0.18	0.02	0.02	0.11	0.41	0.18
TiN, mg/L	7.84	5.37	7.09	7.82	7.87	7.85
TP, mg/L	0.90	0.95	1.02	0.82	0.58	0.05
TP Removal %	84.4	83.5	82.4	85.5	90.2	99.1
pH	6.99	7.04	7.10	7.02	7.02	6.95

Table 3
Phosphorus mass balance for the six treatment scenarios evaluated.

	Scenario	Location	Flow ML/d	Phosphorus Concentration mg/L	Phosphorus Load kg/day (% of influent P)
Inputs	All	Influent	18.93	5.73	108.45 (100%)
Outputs	S1	Effluent	18.79	0.93	17.46 (16%)
		Biosolids	0.14	617.39	90.99 (84%)
	S2	Effluent	18.77	0.95	17.90 (16.5%)
		Biosolids	0.16	562.15	90.54 (83.5%)
	S3	Effluent	18.79	1.02	19.10 (17.6%)
		Biosolids	0.17	583.34	89.35 (82.4%)
	S4	Effluent	19.26	0.82	15.74 (14.5%)
		Biosolids	0.16	585.03	92.70 (85.5%)
	S5	Effluent	18.35	0.58	10.60 (9.8%)
		Biosolids	0.22	234.92	52.44 (48.4%)
	S6	Struvite	0.36	125.18	44.84 (41.3%)
		Effluent	18.77	0.05	0.99 (0.9%)
		Biosolids	0.17	708.67	107.46 (99%)

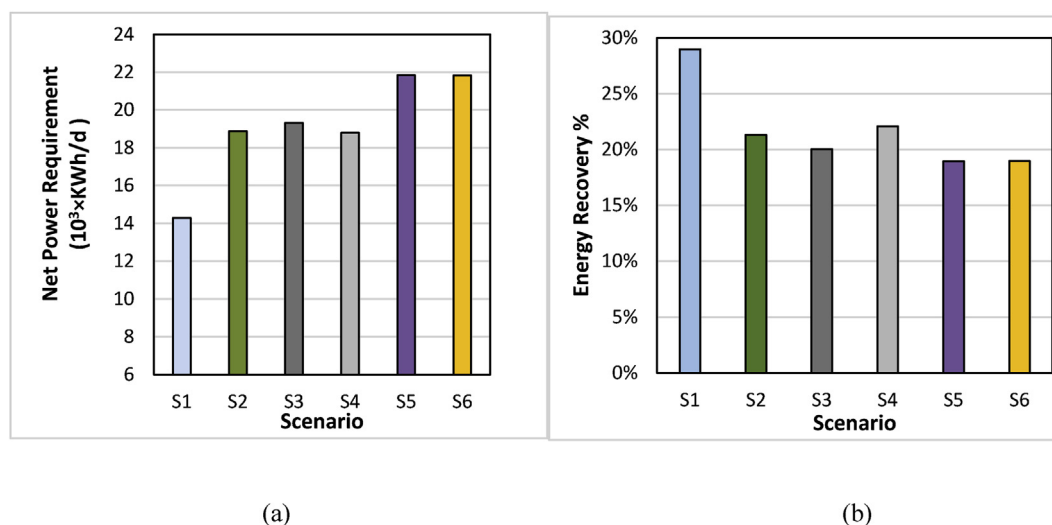


Fig. 2. (a) Net power requirements and (b) Percentage of energy recovered relative to net power requirements for different treatment scenarios.

a significant impact on the overall cost of the configurations.

The highest energy consuming scenarios were S5 and S6 (Fig. 2a). The five-stage Bardenpho configuration (S2) showed a distinct increase in energy demand compared to the MUCT process (S1). This increase was mainly due to the extended aeration and higher mixing requirements in the five-stage process. Simulation results show that the aeration energy requirement in S2 is 16% higher than S1 and about 34% higher than S5. In S5, struvite recovery resulted in the reduction of N and P content in the recycle stream, which could be linked to the 21% reduction in aeration requirement compared to S1. Although S3 was very efficient in removing organics and suspended solids, the aeration requirement was the highest of all the treatment scenarios evaluated due to the need for air scouring in addition to biological processes.

Fig. 2a also shows that there was no increase in net energy requirement for additional P removal in S6 (effluent TP = 0.05 mg/L) compared to S5 (effluent TP ≤ 1 mg/L). This is mainly due to high energy requirement to operate the struvite reactors in S5. The increased P removal in S6 to meet the stringent effluent limit resulted in a distinct increase in the net energy requirement compared to all the other scenarios except S5. The addition of MBR to the MUCT process in S3 required 1.4x higher energy input compared to the MUCT process alone in S1. It should be pointed out that scenario S4 resulted in a slightly higher (2%) P removal compared to S2 for similar energy input and significantly less methanol addition (Table 4).

The power produced by the CHP engine appears as 157–180 kW/h in different scenarios. Comparing the results from

S1 and S5, we can conclude that P recovery did not appear to have a negative impact on biogas generation (Table 4). Since the net power requirement of S5 and S6 are high, only about 18% energy recovery value could be obtained (Fig. 2b). The best scenario in the context of energy recovery was S1, which was likely due to the low net energy requirement of the treatment process and high biomass yield. The addition of MBR system in S3 to the MUCT process in S1 resulted in about 7% reduction in energy recovery (Table 4) primarily due to the reduction in total volatile solids loading to the anaerobic digester by about 11%.

Simulation results suggest that S1 and S3 were the most chemically intensive scenarios (Fig. 3a and Table 4). In S3, a significant amount of chemical cost was required for membrane cleaning using sodium hypochlorite and citric acid. Although the side-stream P recovery installation helped eliminate the ferric salt addition completely from the mainstream due to low P concentration in the recycle stream, the struvite recovery process (S5) required chemical addition in the precipitative reactor (Table 4). The Mg input cost associated with struvite precipitation was estimated to be \$145 per ton of struvite produced. The chemical consumption in S4 was relatively low compared to all the other scenarios, as the fixed film system helped achieve target P and N removal without requiring substantial amount of methanol and ferric salt addition (Table 4). The primary reason could be that the fixed film systems improve nitrification process by decoupling the growth rate of nitrifying population from PAOs and denitrifiers (Onnis-Hayden et al., 2011). The decoupling and separate SRT control allow for simultaneous optimization of N and P removal.

Table 4
Chemical consumption, energy requirement and energy recovery for the six treatment configurations.

	S1	S2	S3	S4	S5	S6
Chemical Requirement						
Ferric, kg/day	—	264	158	—	—	283
Methanol, kg/day	952	294	912	78.66	397	952
Lime Addition, kg/day	1274	448	1274	2119	1344	1272
Sludge Treatment						
Sludge Production, kg TSS/day	2459	2228	2507	2249	1594	2509
Sludge Production, kg TSS/kg COD _{removed}	0.274	0.247	0.276	0.249	0.176	0.278
Energy Usage						
Gross Power, kWhd ⁻¹	18,432	22,896	23,192	22,944	25,980	25,968
Energy Recovery, kWhd ⁻¹	4142	4023	3873	4152	4143	4143
Net Power Requirement, kWh/m ³	0.76	0.99	1.02	0.99	1.15	1.15

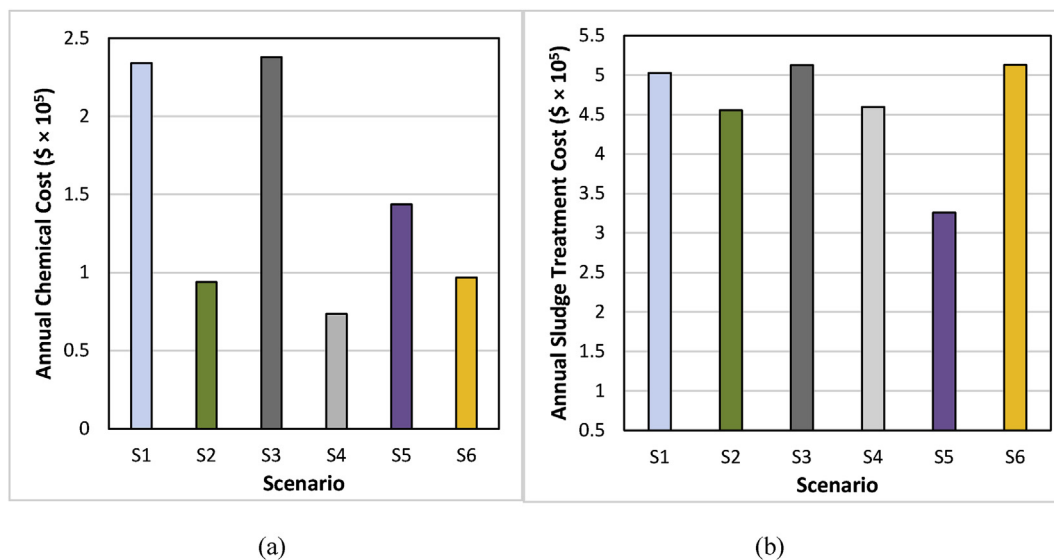


Fig. 3. (a) Annual chemical cost, and (b) annual sludge disposal cost for different treatment scenarios.

The sludge disposal cost was the highest for S6 as almost all the P (99%) was diverted to the biosolids fraction (Fig. 3b). Also, this scenario required ferric salt addition for media regeneration and was very efficient in TSS removal. There was a substantial reduction (37%) in sludge disposal cost (Fig. 3b) for S5 (MUCT process with sidestream struvite recovery) compared to S1 (MUCT process without sidestream struvite recovery). This reduction might be linked to N and P recovery as struvite and a significant reduction (~555 kg/day) in methanol addition requirement in S5 compared to S1 (Table 4). The struvite recovery process removed 44.84 kg/day of P and 20.40 kg/day of N with an annual struvite production of 142.2 tons. The sludge production in S3 was also 2%, 12.5% and 11% higher, respectively, compared to S1, S2 and S4. This is attributable to the improved capture of solids associated with the filtration process and the chemically intense nature of the treatment system.

3.4. Total annualized cost and cost effectiveness (\$/lb-P removed)

Overall, it was evident that scenarios S1 (\$42.25/lb-P removed), S4 (\$42.22/lb-P removed), S5 (\$44.60/lb-P removed) and S6 (\$44.04/lb-P removed) represented the most favorable options

from a cost effectiveness standpoint (\$/lb-P removed) (Fig. 4a). Although S2 appeared to be a very attractive option for enhanced N removal, the unit cost for P removal (\$46.01/lb-P removed) in this scenario was higher than most of the other scenarios (S1, S4, S5, S6). S4 was one of the most cost-effective configuration due to low chemical requirement and flexibility to adjust SRT without impacting the nitrification capacity of the system. Configuration S5 highlights the potential of WWTPs to recover P as struvite, which would otherwise get deposited and create blockage in the plumbing system. The total annualized cost of the P recovery configuration is substantially higher compared to S1 and S4 (Fig. 4b). This is because of high chemical cost associated with P recovery, energy inputs to operate the reactors, and the higher initial investment cost. Although the total annualized costs in S6 was about 20% higher than S1, it still appeared to be a cost-effective alternative as 15% higher P removal was obtained (Table 2). The membrane system (S3) did not seem to be a cost-effective solution for P removal. Membranes usually have a shorter economic life (10–12 years) and the operational and maintenance costs are higher. Our results indicate that maintenance costs (chemical cleaning of membrane and replacement costs) associated with S3 is

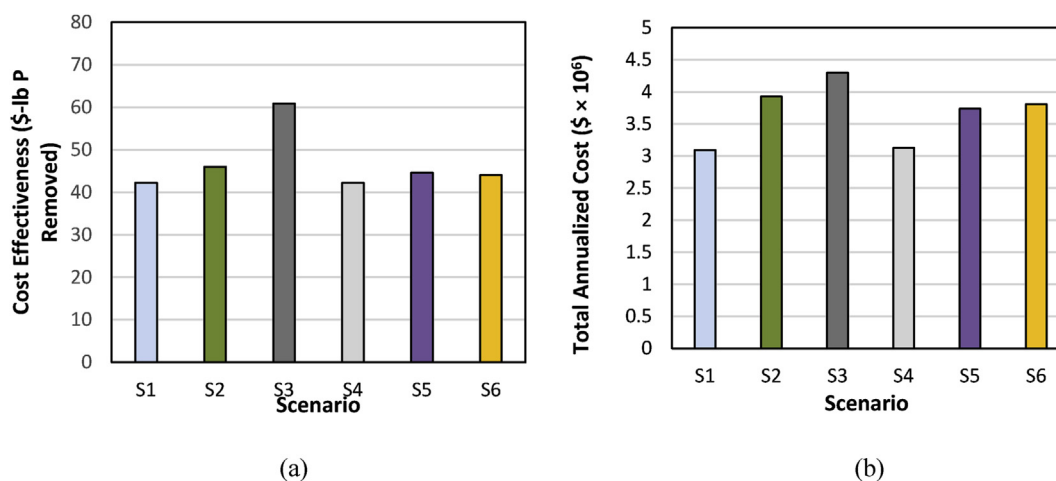


Fig. 4. (a) Cost effectiveness (\$/lb P removed), and (b) Total annualized cost for different treatment scenarios.

about 27% of the total annualized cost, which is the highest of all scenarios evaluated (Fig. 5).

A breakdown of operational and maintenance costs in Fig. 5 shows that electricity consumption accounted for the largest portion in each scenario. It was evident that, improved levels of P removal and effluent quality requires an increase in energy consumption for treatment. Chemical costs varied from 4 to 11 percent of the total operational costs. It should be noted that the characteristics of the wastewater or market situation might have a significant impact on chemical costs associated with P removal. The cost of chemicals required for wastewater treatment usually varies from year to year. A mean cost of each type of chemical based on historical trends was adopted in this study. Another important factor affecting the cost of P removal was sludge production. In all the treatment scenarios except S5, the cost of sludge disposal accounted for about 20% of the total operational cost (Fig. 5). The cost of sludge disposal may increase in the near future, as suitable landfill locations diminish. If sludge has to be transported to distant locations, or further treated due to regulatory requirements, cost of P removal might rise sharply. Under such circumstances, cost of P removal in scenarios S1 or S6 could be prohibitively high.

3.5. Comparisons with previous studies

As shown in Table 5, unit cost estimates for our treatment scenarios compared favorably with two other previously published studies (Jiang et al., 2005; WSDOE, 2011). Both these studies reported a decrease in unit costs for TP removal with increasing plant capacity. While we obtained higher unit cost estimates compared to Jiang et al. (2005), the differences can be attributed to the time elapsed (12 years), which would have an impact on electricity, chemical, and maintenance costs. However, our unit cost estimate for 5 MGD capacity is well within the range reported by WSDOE (2011) for 1 and 10 MGD plants. Also, for the 3-step BNR with chemical addition and tertiary filtration configuration, our estimate is lower than that obtained by Jiang et al. (2005). Possible reasons include consideration of labor costs and engineering fees by Jiang et al. (2005), which sharply increase for such advanced treatment technologies. These cost components were not considered in the current study.

Apart from the two studies mentioned above, some of the recent studies published within the last five years focus on not only meeting the discharge permit limits economically but also on the environmental sustainability of the treatment systems. Falk et al.

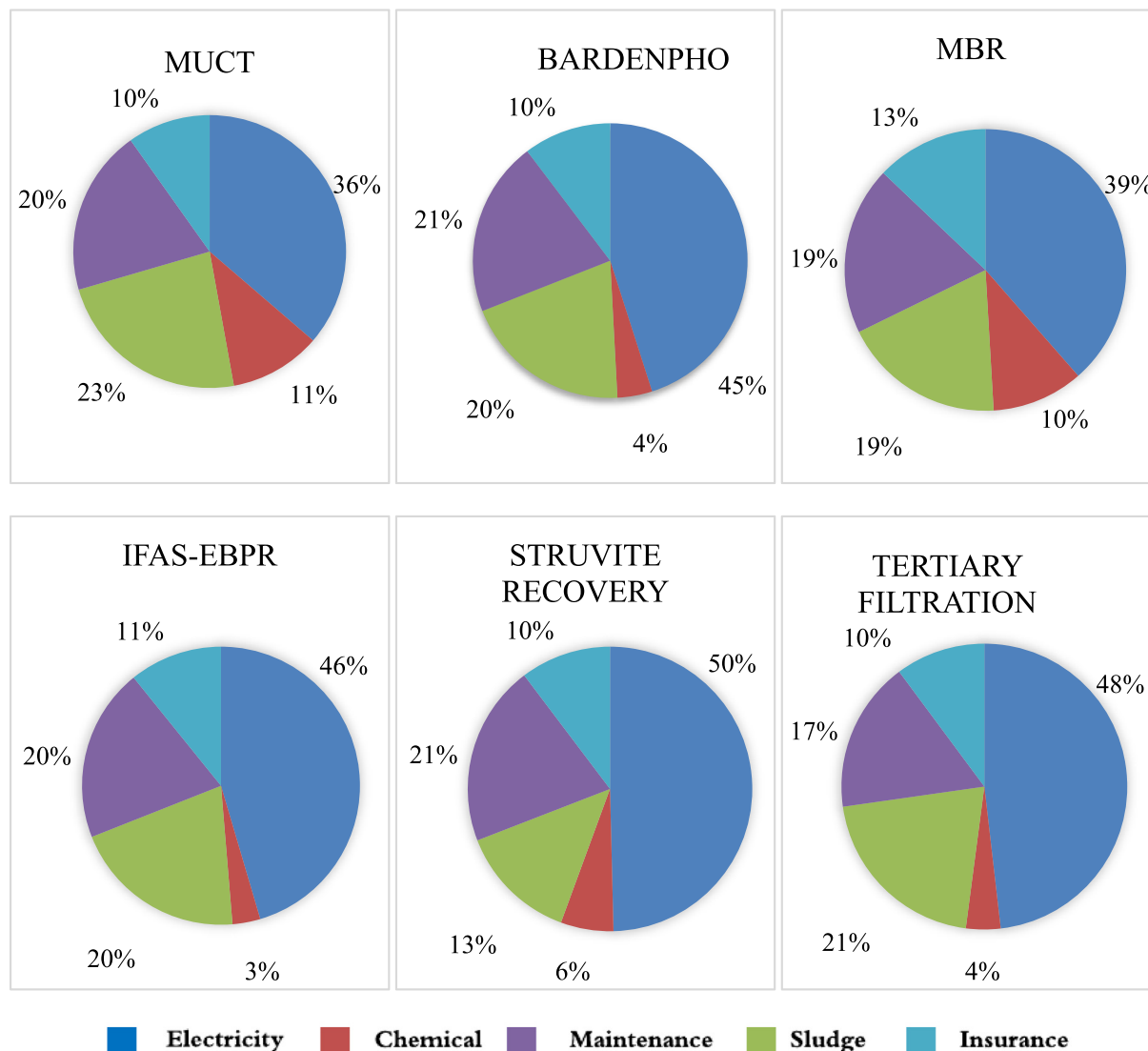


Fig. 5. Operational and maintenance cost breakdown for different treatment scenarios.

Table 5

Comparison of our cost estimates with those from other relevant studies.

Study	Treatment Process	Capacity (MGD)	Effluent TP (mg/L)	Unit Cost (\$/kg P removed)
WSDOE (2011)	3-step BNR + Chemical Addition	1	1	118.49
		10	1	52.82
Jiang et al. (2005)	3-step BNR + Chemical Addition	1	1	87
		10	1	37.7
	3 step BNR + Chemical Addition + Filtration	1	0.5	269
		10	0.5	89
This study	3-step BNR + Chemical Addition	5	1	92.95
	3 step BNR + Chemical Addition + Filtration	5	0.05	96.8

(2013) recommended a triple bottom line for fulfilling an effluent discharge objective that includes environmental, economical and social pillars. Site specific costs associated with land, construction and labor are included in some studies while one or all of the items are excluded in other studies. Some of the studies also ignore the maintenance costs associated with the treatment processes. Thus, the differences in methodology and components considered prevented us from performing a direct comparison with additional published studies.

3.6. Limitations of our study

The assumptions used in this study could be a source of certain limitations. Firstly, no consideration was given to land requirements in our economic calculations. The membrane systems are compact and require much less footprint compared to conventional activated sludge systems. Therefore, the utilities forced to meet very stringent discharge standards but incapable of further expansion due to land unavailability, could find these systems useful. Secondly, the simulation exercises ought to be performed for much longer periods, to cover both dry- and wet-weather conditions, as well as seasonal variations. Thirdly, meeting the target effluent P was the major focus while designing and simulating the treatment systems in this study. But in reality, the wastewater treatment systems are employed to remove numerous constituents and not just P. While exploring various strategic alternatives for P removal and/or recovery, we also had to consider the need to achieve an acceptable effluent composition in terms of TSS, COD, BOD and N.

4. Conclusions

We have presented a comprehensive operational and economical evaluation of six different wastewater treatment scenarios, covering a wide variety of competing P removal processes from secondary to advanced tertiary treatment and one side-stream P recovery process. The main conclusions from this study are as follows:

- The unit cost for P removal in different treatment alternatives ranges from \$42.22 to \$60.88 per lb of P removed. The MUCT BNR + tertiary reactive media filtration proved to be one of the most cost-effective configurations (\$44.97/lb P removed) producing effluent with a TP concentration of 0.05 mg/L. This treatment method had the second lowest unit P removal cost among the six scenarios evaluated.
- Incorporation of struvite precipitation in WWTPs recovers P in a solid form that is separate from biosolids and can be used as a fertilizer product. But, the accompanying reduction in effluent P due to struvite recovery is low (~6%). Therefore, side-stream P recovery process, while diverting away P loads from biosolids, would not help with meeting stringent TP discharge limits.

- A major operational expense during struvite precipitation arises from chemical addition requirement. The typical Mg concentration in municipal wastewater is very low for struvite precipitation. To reach the optimal ranges for struvite precipitation, pH levels had to be adjusted as well. Future research into alternatives to chemical inputs for pH adjustment and Mg is necessary for cost effective P recovery.
- In all the treatment scenarios, biosolids was the major outlet for P (48–99% of influent P). So land application of biosolids is extremely important for efficient P management at the watershed level. However, waste activated sludge from WWTP concentrates numerous contaminants, such as persistent organic pollutants and heavy metals (Zhao et al., 2017; Xu et al., 2017). A large portion of these contaminants is not degraded by the most common sludge treatment methods employed by the WWTPs, such as anaerobic digestion or fermentation (Wang et al., 2017b). The remaining pollutants in the biosolids have the potential to enter into the environment and cause risks to natural organisms and other biological processes. Thus, biosolids management in an environmentally safe manner needs to be ensured.

We anticipate use of this study results to determine the life cycle cost and environmental impact assessment of different P removal and recovery processes.

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Appendix A. Supplementary data

Supplementary data related to this article can be found at <https://doi.org/10.1016/j.chemosphere.2017.12.169>.

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