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Economic evaluation of ion-exchange processes for nutrient removal and recovery from municipal wastewater

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Ion exchange (IEX) processes are a promising alternative to remove and recover nutrients from municipal wastewater. To assess the feasibility and viability of IEX processes for full-scale application, this study aimed at providing an evaluation of performance and economics on upscaling these processes for two different configurations in a 10,000 population equivalent wastewater treatment plant (WWTP) and compared them with a traditional biological nutrient removal (BNR) plant. The IEX processes were designed based on existing pilot-scale data, and after aerobic or anaerobic carbon removal stages. The nutrients were recovered from spent regenerants in the form of $(NH_4)_2SO_4$ and hydroxyapatite $Ca_5(PO_4)_3(OH)$, allowing regenerant reuse. The 40-year whole life cost (WLC) of IEX coupled with traditional activated sludge processes was estimated to be ~£7.4 M, and WLC of IEX coupled with anaerobic membrane process was estimated to be £6.1 M, which was, respectively, 17% and 27% less than the traditional BNR based WWTP. Furthermore, ~98 tonnes of $(NH_4)_2SO_4$ and 3.4 tonnes of $Ca_3(PO_4)_2$ could be recovered annually. The benefits of lower costs, reduction in greenhouse gas emissions and nutrient recovery aligned with circular economy, illustrated that IEX processes are attractive for nutrient removal and recovery from municipal wastewater.

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INTRODUCTION

Excess nutrients released from wastewater treatment plants (WWTPs) can lead to eutrophication with detrimental effects to the aquatic environment. Consequently, existing legislation provides stringent control of WWTPs discharges, limiting to 1 mg NH_4-N/L and as low as 0.5 mg PO_4-P/L[1]. On the other hand, nutrients are essential to all life forms, and should be recovered, particularly phosphorus considering the fast depletion of natural reserves[2]. Nutrient recovery can also enable alternative revenue streams and is aligned with the delivery of the circular economy[3].

Although activated sludge process variations, such as biological nutrient removal (BNR), predominate in WWTPs nowadays[4], ion exchange (IEX) processes are rarely applied in municipal wastewater treatment. The main reasons are limited media selectivity, bed clogging, and costly regenerations. Nevertheless, recent advances are tackling these limitations, and IEX processes are being considered for nutrient removal and recovery from municipal wastewater. More specifically, recent studies demonstrated that mesolite, a synthetically produced zeolite, has a high capacity for ammonia (NH_4-N) adsorption with reported values of 4.6 and 4.9 meq/g, compared with 2.0 meq/g for Clinoptilolite[5]. The outstanding performance of synthetic zeolites is attributed to the increased consistency Si:Al ratios of 1:1, defining a high cation exchange capacity[5]. A comparison experiment with different empty bed contact times (EBCT) suggested 10 min to be a balance point between performance and costs for ammonia removal with mesolite[6]. Furthermore, the tests showed that regenerant can be re-used for consecutive cycles before clean-up, and to reach breakthrough point of 1 mg NH_4-N/L, the process could be operated for 1000 bed volumes (BV). Nevertheless, this is dependent on ammonia concentrations and 300–400 BV between regenerations is expected if 10–20 mg NH_4-N/L is in the influent[7]. The study also revealed that potassium chloride regenerant was more effective than sodium chloride with the efficiencies ranged between 72 and 94 and 64–68%, respectively. In addition, potassium chloride was less consumed as it still remained effective down to 0.1 M, at the same regeneration efficiency as of 1.0 M sodium chloride[6]. Five BV KCl solution (10%) was found effective to regenerate the mesolite at higher PH value. The pH can be adjusted to 12 by adding 2% NaOH solution.

As for phosphorus removal by IEX, one of the most promising media is hybrid ion exchange resins (HAIX) with ferric oxide nanoparticles embedded within the base structure of the resin. The removal of phosphorus with HAIX is believed to be more efficient than chemical precipitation or biological methods, especially when removing down to very low level (<0.1 mg PO_4-P/L)[8]. The high capacity of HAIX results in extended operating cycles. For instance, the bed life of HAIX, treating a wastewater with an initial phosphorous concentration of 0.26 mg/L, demonstrated an order of magnitude greater (from 200 to 2000 BVs) than that of a conventional ion exchanger, IRA-900[9]. A realistic expectation of operational capacity at full scale was estimated at 4.9–6.2 mg P/g media with an EBCT of 5 min but capacity was reduced to 2.5 and 3.7 mg P/g media after multiple runs. A 10 BV NaOH solution (2–4%) was found effective to regenerate the HAIX and the regenerant could be re-used 10 times before requiring clean-up[9].

Although these ion-exchange processes look promising, the frequent regenerations with new chemical solution every time and high costs (typically £65/ton) for disposal of spent regenerant as hazardous waste were found non-economical in the long-term run[10].

Recovery of the regenerants is critical to ensure economic feasibility. Recent studies indicated that a promising solution is to clean-up and reuse the spent regenerant whilst recovering the nutrients at the same time. The clean-up of KCl regenerant and ammonia recovery in the form of ammonium sulphate can be achieved using commercially available modules of hollow fibre membrane contactor (HFMC) with sulphuric acid[6]. The phosphorus can be recovered in the form of calcium phosphate

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(hydroxyapatite) by adding hydrated lime to the spent NaOH regenerant leading to instant formation of hydroxyapatite $Ca_5(PO_4)_3(OH)$ that can be separated from the liquid by filtration[7]. Media replacement was estimated after 600 regeneration cycles, although this also needs further verification at a demonstration or full scale[7].

Besides technology, other challenges also remain to hinder the nutrient recovery. These include a lack of viable commercialisation pathway, and limited understanding of the recovered products markets[11]. This study aims at providing an evaluation of performance and economics of IEX processes as tertiary treatment process, for both ammonia and phosphorus removal and recovery, after traditional activated sludge process (aimed at carbon removal) or anaerobic treatment with a combined upflow anaerobic sludge blanket and membrane reactor (anMBR) compared with traditional BNR combined with iron dosing for a 10,000 population equivalent (PE) municipal WWTP.

RESULTS

Design, mass and energy balances

The designed processes and mass balance are shown in Fig. 1. The results demonstrate that all the flowsheets investigated, BNR + iron dosing, ASP + IEX and AnMBR + IEX processes could produce the desired effluent quality of COD < 20 mg/L, NH_4-N < 1 mg/L and PO_4-P < 0.5 mg/L. Nevertheless, process stability was different. More specifically, the BNR process was sensitive to low temperature, shock loads and dependent on carbon availability and struggled to reach very high nutrient removal rates (>95%) required to meet stringent nutrient discharge limits (<1 mg NH_4-N/L, <0.5 mg PO_4-P/L). Fundamentally, the reasons for instability were owing to competition on carbon sources uptake, especially for low-strength wastewater. According to influent characteristics and effluent quality requirement, COD removal rate must be above 93%, and phosphate removal rate must be above 85%, which could not be realised simultaneously with ammonium removal by BNR only. Therefore, supplementary processes were designed for further removal, including ferric dosing and a tertiary filter. The consumption of ferric salt was estimated to be 6438 kg/year, which was double theoretically calculated demand to ensure the removal efficiency. Reported optimum weight ratios for iron, have varied from <1 to as high as 7.5. The US EPA (USEPA, 1987) suggests that a weight ratio between 2 and 3 of Fe:P is needed to achieve 95 percent removal of phosphorus from municipal wastewater[12]. Regarding energy consumption, aeration consumed the most of energy required for the BNR + iron dosing up as anticipated[13] (Fig. 2).

In the ASP + IEX flowsheet, the conventional activated sludge process (ASP) was effective and capable of removing 90% BOD and COD[14,15] (Fig. 1). The IEX process was effective and capable of removing ammonia and phosphate to the required low level or even very low level such as <0.1 mg PO_4-P/L. There was additional benefit of 40–50% COD removal by HAIX media, which was believed to be associated with base resin of the HAIX media[8]. Others have demonstrated that IEX processes can be impacted by changes in temperature and initial pollutant concentrations (e.g., diurnal shock loads) owing to standard chemical kinetics and mass transfer limitations[16]. Nevertheless, the fact regeneration can be triggered dependent on effluent quality, makes IEX less sensitive to low temperature and diurnal shock loads, compared with biological processes. In relation to energy consumption, aeration consumed the majority of total electricity needs, representing 53%, followed 38% by drum filter (Fig. 2). The HFMC was able to clean-up the KCl brine and by adding sulphuric acid, ammonium sulphate $(NH_4)_2SO_4$ was recovered with annual production of 98 tonnes. The cloth filter was capable of cleaning up NaOH brine and by adding hydrated lime, hydroxyapatite $Ca_5(PO_4)_3(OH)$ was recovered with annual production of 3.4 tonnes.

The UASB-configured AnMBR provided suitable treatment (Fig. 1). The optimised intermittent dead-end biogas sparging regime resulted in low energy demands and low operational costs[17]. The COD removal efficiency in the AnMBR was 87%, but the further 6% removal, to reach the required 93% in total, could be supplemented by IEX, as HAIX media was found capable of removing COD by 40–50%[9]. The specific energy consumption clearly showed the benefits of biogas production as most of the energy consumed was offset by production in the AnMBR with a total energy production of 0.12 kWh/m^3 (does not consider the other processes consumption) (Fig. 2).

Overall, the IEX processes were more reliable to meet low level or even very low level of nutrients discharge limits compared with BNR. Importantly, nutrient recovery could be achieved during regenerant cleaning up, allowing for >85% nutrient recovery (Tables 7 and 8). Furthermore, IEX processes are generally independent of temperature, resilient to shock loads and consumes relatively low electricity[6,8,9]. However, high concentration of suspended solids and competing ions (SO_4^{2-}, Ca^{2+}, NO_3^- etc.) in the feed can potentially reduce performance efficiency[8]. Column experiments indicated that adsorption capacity was reduced by up to 50% when high level suspended solids were presented[6]. To minimise the impacts of suspended solids on the process performance, pre-treatment with filtration, or similar, is normally required. This increases processes complexity and costs. AnMBR coupling with IEX have more benefits compared with ASP + IEX. Besides energy recovery, a crucial point is that AnMBR provides solid free effluent feeding into IEX, which improve adsorption efficiency, secure stability of IEX process. The influence of SO_4 in the influent at the level of 50 mg/L to IEX was considered to be limited[9]. A test with synthetic solution showed that SO_4 reduced the capacity of HAIX from 114 mg P/g to 101 mg P/g[9] (at initial P and SO_4 concentrations 100 mg/L) but this reduction in capacity did not continue when the sulphate concentration was >100 mg/L. This was linked with the sulphate competing for both sorption sites and ferric nanoparticles embedded in HAIX structure[18]. Presence of high concentration of SO_4 (>100 mg/L) may impact the AnMBR process operation and methane yield through the carbon utilisation by sulphate reducing bacteria, but the effect can be mitigated by precipitation of sulphate salt produced from added chemical (Fe^{2+}) and optimising solid retention time and specific gas demands[19]. On the other side, Fe^{2+} needs to be balanced with other potential negative impacts downstream, including loss of recoverable phosphate in the IEX regenerant as well as adding potential further competing ions reducing capacity, specifically for the cation IEX process.

Economic evaluation

The CAPEX estimated for the BNR + iron dosing process was £3.94 M (Fig. 3). The largest contribution to the CAPEX was the construction of the BNR concrete tanks accounting for 60%, followed by 17% for tertiary filter, and 15% for installation of aeration systems (Fig. 3). The total operational expenditure (OPEX) was estimated to be £316 k/year. The maintenance costs constituted the majority of the total OPEX, representing 31%. Sludge and labour costs make up 25% and 17% of the annual OPEX, followed by 17% for energy consumption (Fig. 3). Total whole life cost (WLC) was estimated at £8.4 M.

The CAPEX estimated for the ASP + IEX process was £3.48 (Fig. 3). Installation of IEX (Fig. 3) accounted for the majority of total CAPEX representing 44%, followed by 33% for construction of conventional ASP concrete tanks. Total OPEX was estimated at £282k/year. Operation of ASP accounted for 79% of total OPEX, including electricity consumption for aeration, sludge disposal,

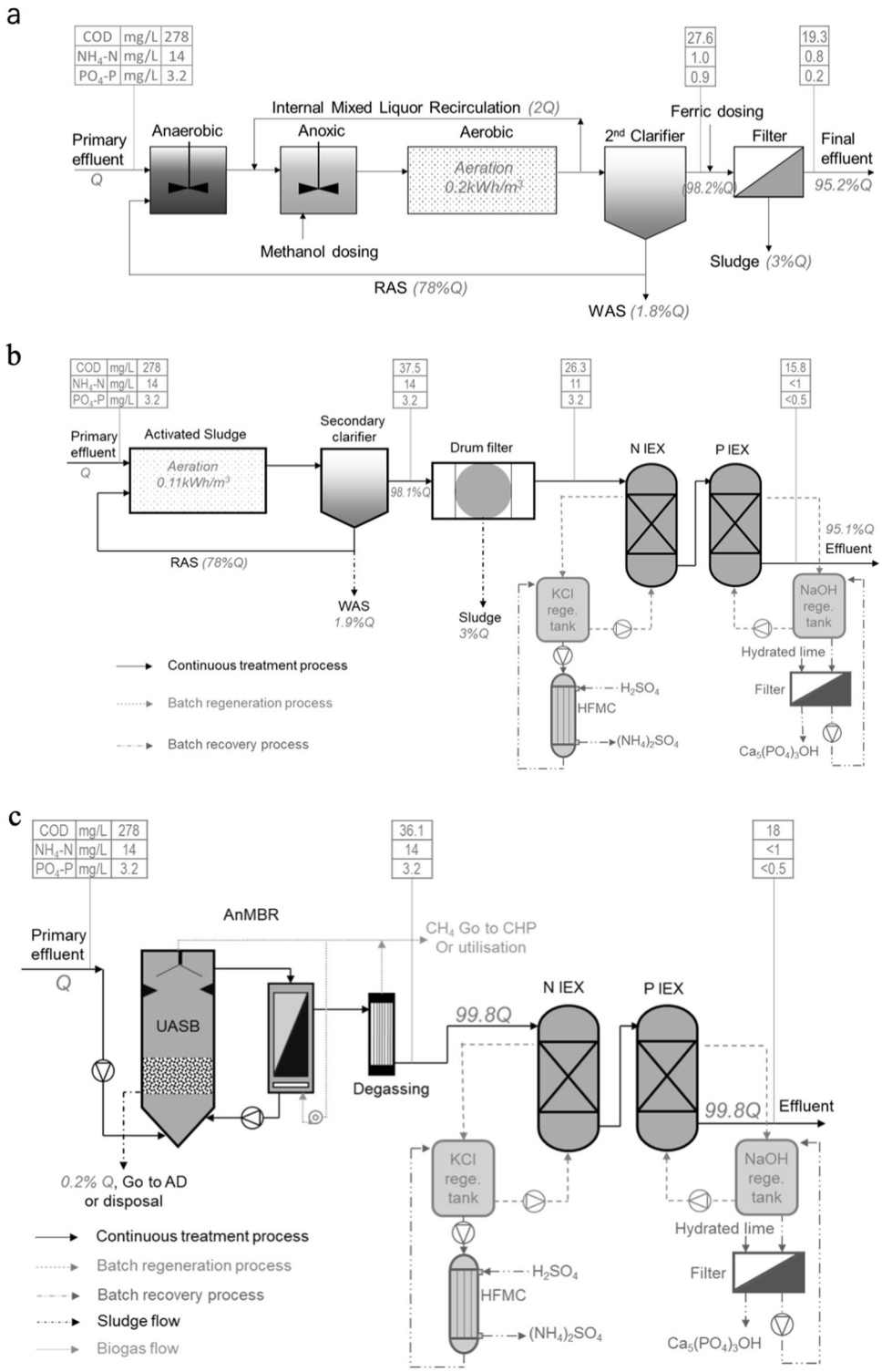

Fig. 1 Designed processes, mass and energy balances. a BNR (A^2O) with iron dosing and tertiary filtration; **b** ASP + IEX with clean-up and reuse of regenerant; and **c** AnMBR + IEX with clean-up and reuse of regenerant.

and related maintenance costs (Fig. 3). The WLC was estimated at £7.4 M.

The CAPEX estimated for the AnMBR+IEX process was £3.6 M. The purchase of membrane and construction of UASB reactor accounted for the majority of total CAPEX (Fig. 2). The total OPEX was estimated to be £177k/year accounting with the energy consumption of biogas sparging and recirculation pumps (Fig. 3). The WLC was estimated at £6.1 M.

Comparing the three flowsheets, Fig. 4 shows costs by year as TOTEX, by process and costs by process combination in the designed business scenario. Comparatively, IEX consumed much less electricity. More importantly, regenerant clean-up and reuse substantially lowered down the costs compared with which regenerant was used only once. For N IEX column, reuse of regenerant reduced the total OPEX by 50%, and recovery decreased further 15–65% in comparison with using new

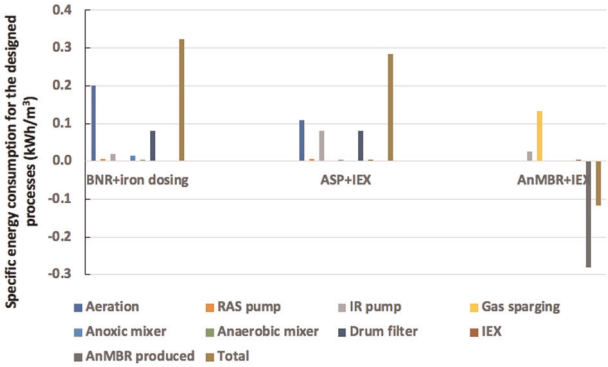

Fig. 2 Specific energy consumption for the designed processes.

regenerant every time. For P IEX column, HAIX media costs were expected to be 30–50% lower when using local products for scaled up treatment units. The results also indicated that WLC of ASP + IEX and AnMBR + IEX are 10.5% and 19.8% less than that of BNR + iron dosing, respectively.

When very high removal efficiency is required, such as >95% or even >98%, the costs of biological processes are dramatically increased. An estimation revealed that the operational costs were doubled to reach effluent discharge from <1 mg P/L to 0.1 mg P/L[20]. Instead, IEX OPEX is less sensitive to stringent nutrient discharge control compared with BNR, especially if the regenerants can be managed in such a way that allow for multiple use and recovery, as here described. Studies also illustrated the operational preference of having high capacity over high durability. For example, when mesolite capacity had a fourfold

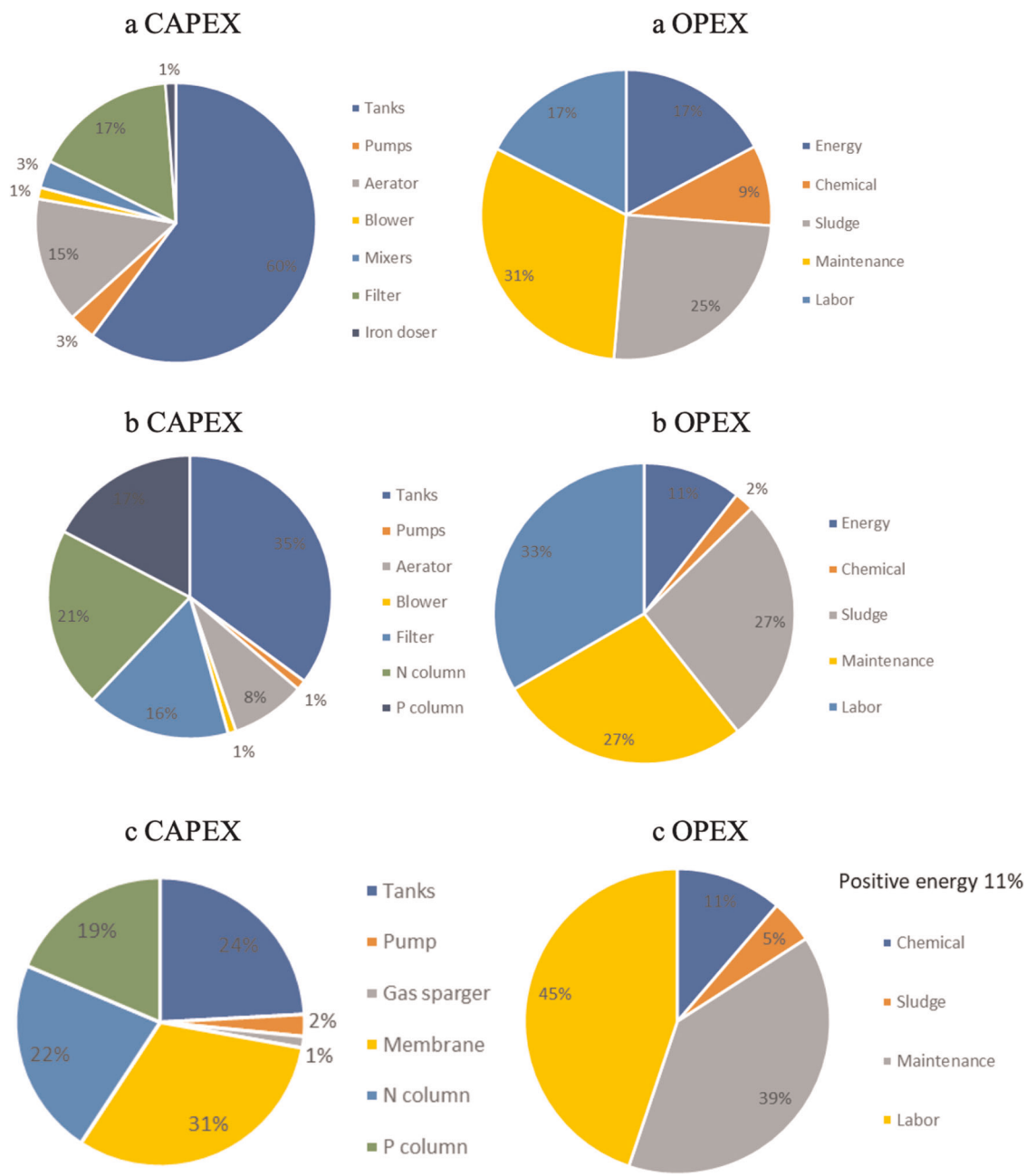

Fig. 3 Processes CAPEX breakdown by segments and processes OPEX breakdown by classification for **a** BNR (A^2O) with iron dosing and tertiary filtration; **b** conventional ASP + IEX with clean-up and reuse of regenerant; and **c** AnMBR + IEX with clean-up and reuse of regenerant.

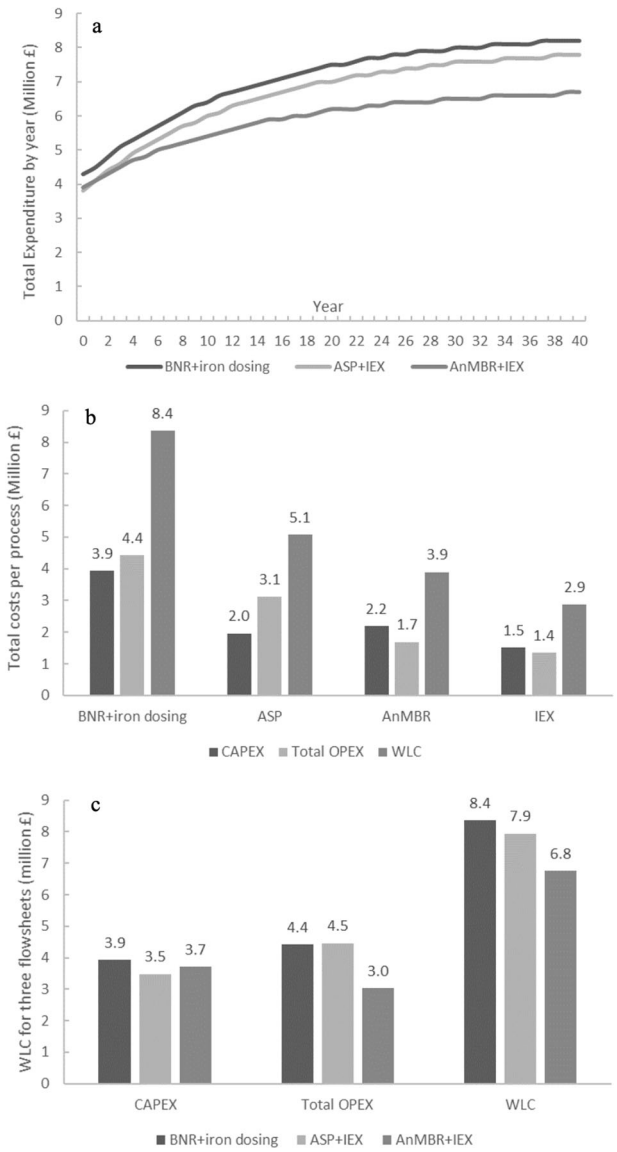

Fig. 4 Costs comparison for the three processes investigated. a Total expenditure (TOTEX) over the 40 years period; **b** total expenditure per process; and **c** whole life costs for the three flowsheets.

increase, the yearly OPEX was reduced by ~50%. However, when media life had been prolonged 10 times, the yearly OPEX was reduced by only 10%[6].

Direct greenhouse emissions

Biological treatment processes release CO_2 from the oxidation of organic matter, nitrous oxide (N_2O) from nitrification/denitrification and methane (CH_4) from anaerobic methanogenesis, which contributes to greenhouse gas (GHG) emissions[21]. Energy consumptions from grid or combined heat and power lead to indirect GHG emission. The estimated GHG emissions for the studied processes BNR + iron dosing, ASP + IEX and AnMBR + IEX were 936 CO_{2eq} t/y, 507 CO_{2eq} t/y and 171 CO_{2eq} t/y, respectively, based on the factors described in Table 1 and assuming emissions from other fugitive emissions, such as methane, to be 5%[22].

Compared with BNR, ASP has less GHG emissions owing to lower electricity consumption and no evident N_2O release as there is no nitrification and denitrification reactions. Meanwhile, AnMBR achieved minimal environmental impact because of nearly complete biogas recovery and significant reduction in the energy consumption. IEX consumes very low electricity, with no evident direct GHG emission in the process. The recovered nutrients from the brine also offer a route for resources recycling. IEX coupling with AnMBR considerably reduces direct greenhouse emissions and is aligned with circular economy for sustainable development through both energy and nutrients recovery.

Table 1. Conversions between greenhouses gases, electricity and CO_2 emissions.

Inventory	Parameters	Value	Unit	Reference
Potential	N_2O	310	kg CO_{2eq}/kg	29
	CH_4	23	kg CO_{2eq}/kg	29
Process	CO_2 from oxidation	1.375	kg CO_{2eq}/kg $BOD_{removed}$	33
	N_2O from nitrification	0.035	kg N_2O-N/kg $N_{removed}$	33
Electricity	Grid	0.523	kg CO_{2eq}/kWh	29
	CHP	0.295	kg CO_{2eq}/kWh	29

DISCUSSION

IEX processes provide a route to capture the nutrients from wastewater into concentrated brine and offered the possibility for subsequent nutrients extraction processes. "Resources factory" has become a buzzword in recent decades for describing how WWTPs should be designed and operated in the near future. This notion aims at more sustainable pattern for development by maximising recovery, reuse and recycling. A review analysed liquid, gas and solid nitrogen recovery with all the state of art technologies and compared the costs with that of industrial nitrogen synthesis from air[23]. The result indicated that N recovery from wastewater is neither currently economical nor urgently needed. It is estimated that the costs of IEX-based approach is relatively low at £1.98 ± 0.23/kg N in the case of NH_4NO_3 recovery[23] compared with other processes, and the operational costs of $(NH_4)_2SO_4$ recovery in this study was around £2.35/kg NH_4-N, but it still cannot compete with industrial synthesis approach (£0.28/kg N). A recommended recovery method with minimal technical input is for agricultural irrigation or selling it as liquid or aqua ammonia fertiliser. Notwithstanding, there is a valid argument if considering industrial nitrogen synthesis process (Haber–Bosch approach) consume 1–2% of global power production and emit 4–8 tons of CO_2 per ton of fertiliser produced[24]. Although nitrogen only recovery process is feasible, they are not extensively implemented so far[11]. This may change in the future with less availability and rising costs of natural gas that industrial process uses.

On the other side, phosphorus is non-renewable element mined from rock phosphate, and ultimately flows into oceans after usage. Considering the limited reserves of rock phosphate, uneven distribution in different countries and increased prices in recent years[25], phosphorus recovery from wastewater becomes increasingly imperative as wastewater is one of the richest sources of phosphorus containing 250,000 tons per year, averagely[25]. Furthermore, the form of recovered phosphorus can be selected based on maturity of the extractive technology, cost of the process and resale value of the product. For example, vivianite is recommended as it has highest value and added iron can simultaneously improve the process performance[26]. In the studied scenario, the recovered nutrients products were ammonia sulphate and hydroxyapatite and their properties are described in Table 2.

The marketability of the recovered nutrient products depends on product purity and quality, legislation, application industry,

Table 2. Properties of recovered nutrient products.

Name	Ammonium sulphate	Hydroxyapatite
Formula	$(NH_4)_2SO_4$	$Ca_5(PO_4)_3(OH)$
Content (TN-TP-P_2O_5-K-Mg-Ca-Fe-S) (%)	21-0-0-0-0-0-0-24	0-19-42-0-0-40-0-0
Physical form	Dry, solid or liquid	Dry, solid, 1–3 mm
Extracted by	Membrane stripping and evaporation	Chemical precipitation, filtration
Purity	High purity	Superior purity
Productivity	98 tons/year	3.4 tons/year
Comparable products	Commercial ammonium sulphate; ammonia aqua	Triple superphosphate ($CaH_4P_2O_8$)
Use in agriculture	Main: agricultural and ornamental crop fertiliser; agricultural spray adjuvant for water soluble insecticide, herbicides and fungicides.	Main: agricultural and ornamental crop fertiliser, turf, nursery garden fertiliser
Use in industry	Niche: used in refractory or flame-retardant materials; ingredient for many vaccines; food additive (acidity regulator in flours and breads); dyeing agent; leather deliming agent; used in battery production, lanthanide extraction; protein purification.	Niche: ingredient for pigment; poultry food additives; dental filler; plastic surgery, metal implant and screw coating; orthopaedic bone grafting or adhesive agent; "bone-building" supplements; sorbent for heavy metal contained in flue gas.
Reference	[34]	[11]

entry points, comparable exiting product sale price and the other aspects. A study using energy-dispersive X-ray spectroscopy analysis suggested high purity of the recovered ammonia sulphate by HFMC from spent regenerant, with only N, O and S elements detected and in proportion close to ideal products[27]. Studies also suggested superior purity of the recovered hydroxyapatite compared with alternative sources or other recovered phosphorus products[28].

To make the recovered nutrient a marketable product, consistent quality of the recovered nutrients must be reached, such as solid content >95%, dust <1% by weight, 2–3 mm of size for agricultural products, 1–1.25 mm of granular size for golf course turf[11]. Specifications for product to enter a specific market must be met. As fertiliser, the products should have no pathogen or heavy metals, and nutrient content and release rate must reach required level. It is to note that recovered nutrient products from wastewater as fertiliser will not likely be able to be marketed with "certified organic". There are potentially four entry points for recovered nutrients products as fertiliser[11]. The recovered products can be sold directly to end-user, wholesaler, broker and dealer or through wholesaler, broker and dealer network to end-user.

The fertiliser market is the focus for the recovered ammonia sulphate and hydroxyapatite. However, there are also alternative uses in other industries. The prices of commercial ammonia sulphate, calcium phosphate and hydroxyapatite products on the markets as fertiliser and chemical were shown in Table 3.

The prices vary widely because of different quality and purity in different grade, such as fertiliser grade, food grade, or pharmaceutical grade. Unless the recovered products resale price is at the same level with these comparable existing products, it is hard to gain the markets. If all the recovered products are sold as fertiliser, the revenue is only £12,834/year, representing 13.2% of annual IEX OPEX, i.e., after drying the HAP or further concentrate the N-liquid. But if all the recovered products, particularly hydroxyapatite, are sold as chemical used for health care, the annual revenue can be £450,285/year, representing fourfold annual IEX OPEX, because the unit price of hydroxyapatite as agent is ~400 times of calcium phosphate used as fertiliser (Table 4). However, considering potential risk and social acceptance, it is still currently impracticable to enter into food and pharmaceutical markets for the recovered products from wastewater.

In conclusion, the ion-exchange processes using mesolite and HAIX present high adsorption capacity for ammonia and phosphorus removal as secondary and tertiary stage of municipal wastewater treatment. Furthermore, nutrients can be recovered from the regenerants and these can be re-used multiple times. Compared with BNR, the benefits of ion-exchange processes include technical reliability, low costs, low GHG emission. Furthermore, nutrient recovery illustrated that IEX is an attractive alternative for nutrient removal and recovery from municipal wastewater, when compared with BNR. Especially, the coupling of AnMBR and ion-exchange processes allows for both energy and nutrient recovery, which is aligned with the circular economy vision.

The ion-exchange processes can be applied to small size or middle scale municipal wastewater plants when very quality effluent is required. For application in large WWTPs, the ion exchange columns can be divided into small units, allowing for simple operation. Overall, IEX processes are attractive for nutrient removal and recovery from municipal wastewater.

Table 3. Prices of recovered nutrient products[22].

Prices	Uses	Unit	Lower	Higher	Average
$(NH_4)_2SO_4$	Fertiliser	£/ton	50	450	120
$CaPO_4$	Fertiliser	£/ton	160	500	300
$(NH_4)_2SO_4$	Reagent	£/ton	400	720	550
$Ca_5(PO_4)_3OH$	Reagent	£/kg	8	312	120

Table 4. Annual revenue of recovered nutrient products.

Item	Industry	Quantities	Price	Revenue
$(NH_4)_2SO_4$	Fertiliser	98.7 t/y	£120 /t	£ 11,844/y
$CaPO_4$	Fertiliser	3.4 t/y	£300 /t	£ 990/y
Total revenue as fertiliser		13.2% of annual IEX OPEX		£ 12,834/y
$(NH_4)_2SO_4$	reagent	98.7 t/y	£550/t	£54,285/y
$Ca_5(PO_4)_3OH$	reagent	3.4 t/y	£120/kg	£396,000/y
Total revenue as reagent		466% of annual IEX OPEX		£450,285/y

METHODS

Process design

Three different flowsheets were designed: 1: BNR + iron dosing, 2: ASP + IEX and 3: upflow anaerobic sludge bioreactor coupled with submerged membrane configuration (AnMBR) + IEX (Fig. 1), for a 10,000 PE municipal WWTP treating settled wastewater to achieve the discharge limits summarised in Table 5. The flow assumed was three dry weather flow (DWF, i.e., the wastewater flow in a sewer system during periods of dry weather with minimum infiltration) at 540 L/(PE.day) corresponding to 5400 m^3/d.

Anaerobic-Anoxic-Oxic (A^2O) process was selected as typical BNR owing to its simplicity and widespread application, and it was designed based on the "sludge retention time (SRT)" method[14]. The influent was assessed to be suitable for using A^2O process and the design was based on the computation approach described in the Supplemental information S1. The process was assumed under typical UK conditions with an average wastewater temperature of 14 °C. The key parameter designs are summarised Supplemental information S1.

The combined ASP + IEX flowsheet considered a conventional ASP targeted at BOD removal and it was designed based on the "SRT" method[14]. The key parameter designs are summarised in Supplemental information S1. To further remove suspended solids from the ASP effluent, and avoid frequent clogging of IEX bed, high head losses, and inefficient operation owing to resin binding caused by residual organics associated with suspended solids, a drum filter was designed with loading capacity of 4500 L/m^2 h (key parameters designs were summarised in Supplemental information S1).

The IEX processes design consisted of two columns, regeneration units, clean-up and nutrient recovery systems for ammonia and phosphorus, respectively[9,29]. The design was based on the computation approach described in Table 6 and design parameters are summarised in Tables 7 and 8.

In the AnMBR process, biogas sparging is crucial to control membrane fouling. However, this consumes significant amounts of the energy, which normally accounts for two-third of the total energy demands of AnMBR process. The design was optimised by utilising intermittent regime and innovative gas sparging devices, so that energy demand of dead-end gas sparging could reach as low as 0.13 kWh per m^3 permeate, whereas specific gas demand can be reached as low as 0.5 m^3/m^2 h with sustained membrane permeability[17].

This significantly reduced capital costs because of obviation of blower purchase and less membrane area invested, as well as operational costs owing to less energy consumed[17]. Data from a 3-year's run AnMBR process for treatment of municipal wastewater at an average temperature of 14 °C supported the feasibility of the designed AnMBR. To recover dissolved methane in AnMBR effluent, a degassing membrane contactor was designed. A study showed that it was effective by a highest recorded recovery of dissolved methane from AnMBR effluent was up to 95%[30]. This makes possible to turn nearly all the methane produced into energy/electricity. The case illustrated that COD removal rate could maintain at 87 ± 1% and methane yield between 0.18 and 0.23 Nm^3 CH_4/kg $COD_{removed}$ and membrane flux at 10–13 L/m^2 [31]. The key parameter designs were summarised in Tables 9 and 10.

Table 5. Influent characterisation and effluent discharge limits for a 10,000 PE WWTP.

Influent			Effluent discharge limits
Parameter	Load g/(PE.day)	Concentration (mg/L)	Concentration (mg/L)
BOD	60	111	N/A
COD	150	278	20
TSS	80	148	N/A
TN	12	22	N/A
NH_4-N	7.5	14	1
TP	2.5	4.6	N/A
PO_4-P	1.75	3.2	0.5
SO_4	N/A	50	N/A

N/A not available or described in the effluent discharge limits.

Table 6. Computation approach for IEX process design.

Step description
1) Select and obtain media characterisation data.
2) Select optimal empty bed contact time (EBCT).
3) Determine ion exchange column volume and configuration.
4) Select regenerant concentration and EBCT, determine brine tank volume and configuration.
5) Select brine clean-up frequency, determine nutrient concentration in brine.
6) Determine recovery configuration and chemical dosage and production.

Costs estimation

The coefficients of cost curve adopted are listed in Table 11, in which *S* denotes size parameter, *a* denotes constant, *b* denotes multiplier, and *n* denotes exponent. The purchased equipment cost on a US Gulf Coast basis

Table 7. Design parameters for fixed bed IEX process for ammonia removal.

Parameter	Unit	Value	Note/reference
N column adsorption			
Media		Mesolite	6
EBCT	min	10	6
Bed volume	m^3	37.5	Calculated
No. of vessels		7	One spare
Vessel diameter	m	3	Optimal from 0.5 to 3 m
Bed depth	m	0.88	<1 m, avoid head loss
Media capacity	BV	350	Reach breakthrough 1 mg/L
N column regeneration			
KCl concentration	%	10	6
NaOH concentration	%	2	35
Regenerant EBCT	min	60	6
Regenerant required	BV	5	29
No. of cycles	/year	148	Calculated
Brine tank volume	m^3	50	Assumed
No. of tanks		4	Calculated
N spent regenerant clean-up			
Frequency	Cycles	3	6
NH_4 concentration	mg/L	2917	Calculated
Membrane	Liqui-cell 4 × 28	Liqui-cell	
Membrane area	m^2	20	Calculated
Flow rate	m^3/h	5	Assumed
Pressure drop	bar	0.41	Calculated
Brine volume	m^3	187.5	Calculated
Time to clean-up	Day	7	Calculated
NaOH required	ton/clean-up	0.0075	Adjust PH from 7 to 11
H_2SO_4 required	ton/clean-up	1.5	H_2SO_4:NH_4 = 4.5:1
$(NH_4)_2SO_4$ formed	ton/clean-up	2.01	H_2SO_4:$(NH_4)_2SO_4$ = 1:5

Table 8. Design parameters for fixed bed IEX for phosphorus removal.

Parameter	Unit	Value	Note/reference
P column adsorption			
Media		HAIX	[9]
EBCT	min	5	[9]
Bed volume	m^3	18.75	Calculated
No. of vessels		5	One spare
Vessel diameter	m	3	Optimal from 0.5–3 m
Bed depth	m	0.88	Calculated
Media life	cycle	600	[9]
P column regeneration			
NaOH concentration	%	4	[10]
Regeneration EBCT	Minutes	20	[10]
Regenerant volume	BV	10	[9]
Regenerant reuse	BV	1000	[36]
One cycle time	day	1.2	Calculated
No. of cycles/year	/year	297	Calculated
P spent regenerant clean-up			
Clean-up frequency	cycles	10	[9]
No. of clean-up	/year	29.7	Calculated
Time to clean-up	day	12.3	Calculated
P concentration in brine	mg/L	1134	Calculated
Calcium required		2	Ca:P ratio
$Ca_3(PO_4)_2$ precipitated	Tonne/ clean-up	0.12	$3Ca(OH)_2 + 2H_3PO_4 = 6H_2O + Ca_3(PO_4)_2$

Table 9. Computation approach for AnMBR process design.

Step	Description
1)	Obtain influent characteristic data and effluent discharge limits.
2)	Select appropriate organic load rate and optimal upflow velocity.
3)	Determine the volumes of UASB, calculate COD removal and sludge production.
4)	Calculated the gas and methane yield and electricity production.
5)	Select average flux, SGD_m, $MLSS_{mt}$ and $Q_{recirculation}$ based on the pilot tests.
6)	Calculated membrane surface area, biogas demands and electricity consumption.
7)	Prepare the design summary table.

(2010) C_e can be estimated by Eq. 1.

$$C_e = a + b \times S^n \quad (1)$$

The unit costs of other items are listed in Table 12. The cost of each item was obtained by multiplying the unit cost and quantities, which came from design assumptions or calculation results. To ensure the accuracy of the attained results, cross-checking and verification were made through different sources or methods, where possible. All the cost values were converted to British Pounds (£) by applying exchange rate when the source data were in other currencies. Historical cost data were converted to present value using the Chemical Engineering Plant Cost Index by Eq. 2.

$$Cost\ in\ year\ A = Cost\ in\ year\ B \times \frac{Cost\ index\ in\ year\ A}{Cost\ index\ in\ year\ B} \quad (2)$$

The location factor *LF* were applied for the cost data from US Gulf Coast (USGC) for application in other countries by Eq. 3.

$$Cost\ of\ plant\ in\ location\ A = Cost\ in\ plant\ in\ USGC \times LF_A \quad (3)$$

where LF_A = location factor for location *A* relative to USGC basis.

Table 10. Design parameters for AnMBR.

Parameter	Unit	Value	Note/reference
UASB			
Temperature	°C	14	Assumed
Upflow velocity	m/h	0.8	Range 0.8–0.9[17]
Organic load rate	kg COD/m^3.d	1.6	Assumed[17]
SRT	day	155	Calculated
HRT	hour	6	Calculated
Volume	m^3	1350	Calculated
Gas production	m^3/d	738	Calculated
Energy production	kWh/d	4928	Calculated
Submerged membrane			
Flux	L/m^2 h	10	[33]
SGD	m^3/m^2 h	0.5	[17]
Energy demand	kWh/m^3	0.13	[17]
Chemical cleaning	/year	0.5	[34]

Table 11. Cost curves for major equipment[32].

Item	Unit	*S*	S_{lower}	S_{higher}	*a*	*b*	*n*
Tank	£/m^3	Capacity	10	4000	5800	1600	0.7
Pump	£/(L/s)	flow	0.2	126	8000	240	0.9
Mixer	£/kW	Power	5	75	17000	1130	1.05
Blower	£/(m^3/h)	flow	200	5000	4450	57	0.8
Filter	m^3	Capacity	0.4	1.4	128,000	89,000	0.5
Drum filter	£/m^2	area	10	180	−73,000	93,000	0.3

Note: *S* denotes parameter size, *a* denotes constant, *b* denotes multiplier, *n* denotes exponent.

Table 12. Unit costs of major items and chemicals.

Item	Unit	Unit cost	Note/reference
N contactor	£/m	20,105	12,794–27,417[6]
P vessel	£/m	4350	2175–6525[9]
Membrane	£/m^2	10	HFMC from Alibaba
Mesolite	£/L	1.64	[6]
HAIX	£/L	13.5	[9]
Sand	£/ton	175	[9]
Screen	£/m^2	16321	[9]
Bag filter	£/bag	7.52	[9]
NaCl	£/ton	62.53	[6]
NaOH	£/ton	469	[6]
H_2SO4	£/ton	70	[6]
$FeCl_3$	£/ton	290	[9]
Lime	£/ton	600	[9]
Methanol	£/ton	600	[37]
Electricity	£/kWh	0.085	[6]
Sludge disposal	£/kg	0.279	[38]
Labour	£/m h	26.5	[6]

Location factors for international locations have a strong function of currency exchange rates and hence fluctuate with time[32] and this can be estimated by Eq. 4.

$$LF\ in\ the\ year\ A = LF\ in\ the\ year\ B \times \frac{Exchange\ rate\ in\ the\ year\ A}{Exchange\ rate\ in\ the\ year\ B} \quad (4)$$

Location factors for United Kingdom is 1.02 in 2003[32].

Capital expenditure

Estimation of capital expenditure (CAPEX) was based on "the factorial method"[21]. Each major equipment component of the process was listed and calculated individually based on the designed specifications, and whole CAPEX was subsequently obtained by multiplying a Lang factor of 4.74 to the sum of the major purchased equipment items (Eq. 5).

$$C = F\left(\sum C_e\right) \quad (5)$$

Where, C is total plant ISBL (inside battery limits) capital cost, $\sum C_e$ is the total delivered cost of all the major equipment items and F is 4.74, an Lang factor for fluids processing plant[32].

OPEX

Based on unit price and quantities of the items in the process, OPEX was calculated as the sum of: i) energy costs; ii); chemical costs; iii) sludge treatment or disposal costs; iv) maintenance costs (assumed as 2.5% of CAPEX)[32]; and v) labour costs.

WLC calculation

The WLC consists of the initial CAPEX and sum of the OPEX for the life span of the process. The life span was assumed as 40 years, and the present value conversion of the OPEX in the future years were performed at the discount rate of 7%, which could be estimated by Eq. 6[6].

$$WLC(whole\ life\ cost) = Initial\ CAPEX + (Annual\ OPEX \times 14) \quad (6)$$

DATA AVAILABILITY

Data can be made available upon request.

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AUTHOR CONTRIBUTIONS

X.H. completed all the key calculations for the manuscript and draughted the manuscript. S.G. advised on the manuscript data collection and design and provided minor comments on the manuscript. B.J. acted as an advisor on the project and provided minor comments on the manuscript. A.S. is the project principal investigator, having provided significant input on the data collection interpretation, writing of the manuscript, and competed all the revisions.

COMPETING INTERESTS

The authors declare no competing interests.

ADDITIONAL INFORMATION

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