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RESEARCH ARTICLE

water ENVIRONMENT RESEARCH

A technological and economic analysis of vacuum-assisted fermentation to augment enhanced biological phosphorus removal at full-scale

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[Correction added on 12 May 2025, after first online publication: The copyright has been changed.]

Abstract

IntensiCarb® represents a vacuum-driven intensification technology with applicability in fermentation or anaerobic digestion. Implementation of this technology in fermentation facilitates a 50% reduction in process volume while concurrently enhancing the yield of volatile fatty acids (VFAs) for advantageous utilization such as carbon source for enhanced biological phosphorus removal (EBPR). An analysis was conducted to assess the process performance and life-cycle costs of IntensiCarb in comparison to chemical addition and conventional fermentation methodologies for Total Phosphorus (TP) removal. Experimental outcomes pertaining to vacuum-assisted process intensification informed the process performance analysis. Greenhouse gas (GHG) emissions were estimated and compared between alternatives. The findings of the assessment indicate that IntensiCarb is a competitive option among alternatives that reduce effluent TP through EBPR. TP removal via $FeCl_3$ chemical addition was the most economically advantageous alternative based on current economics and estimated life-cycle cost.

Practitioner points:

- Novel vacuum-assisted fermentation, IntensiCarb®, was evaluated against other alternatives for process and cost comparisons in enhanced biological phosphorus removal facilities.
- IntensiCarb® has a similar life-cycle cost with conventional fermentation and MicroC® 2000 alternatives for achieving lower effluent total phosphorus.
- The evaluation suggests this technology may be economically viable at full-scale for enhanced biological phosphorus removal facilities.

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All authors are members of the Water Environment Federation (WEF).

- $FeCl_3$ addition and IntensiCarb had the highest greenhouse gas emission estimates relative to conventional fermentation or using MicroC® 2000.

KEYWORDS
enhanced biological phosphorus removal, fermentation, intensification, life-cycle cost

INTRODUCTION

The process of fermentation plays a pivotal role in wastewater treatment by enzymatically degrading volatile solids into desirable soluble components, particularly volatile fatty acids (VFA) (Hao & Wang, 2015). The fermentation process can be initiated with the utilization of primary sludge (PS) alone or a combination of PS and waste-activated sludge (WAS), yielding a substance known as fermentate (Garcia-Aguirre et al., 2019). Subsequent to fermentation, a solids separation procedure is conventionally applied to segregate residual sludge from the soluble by-products, exemplified by VFA, within the fermentate. The advantageous application of these soluble products extends to biological nutrient removal (BNR) treatment processes, notably for enhanced biological phosphorus removal (EBPR) (Nittami et al., 2017) or facilitating denitrification (Bahreini et al., 2021; Lim et al., 2008; Liu et al., 2018a). The remaining fermentate and sludge constituents undergo conveyance to anaerobic digestion for conclusive stabilization, culminating in the eventual generation of biogas (Nakashimada et al., 2008).

Fermentation presents process advantages by generating readily biodegradable substrates for BNR, thereby obviating the need for supplementary chemical additives. Nevertheless, the fermentation process may entail heightened odorous emissions and necessitate additional spatial allocation for the installation of new tanks. Utilities confronted with spatial constraints might exhibit reluctance toward incorporating fermentation into their operations, especially when confronted with competing priorities (Liu et al., 2018b). As an alternative, IntensiCarb®, an innovative space-efficient technology employing vacuum-driven process intensification, offers the advantages of fermentation while concurrently optimizing spatial utilization (Haroun et al., 2022).

IntensiCarb is a low-vacuum-based intensification technology, imparting process advantages in both fermentation and anaerobic digestion applications. The application of a vacuum to the fermenter facilitates the decoupling of Hydraulic Retention Time (HRT) and Solid Retention Time (SRT), resulting in a process that is amenable to deployment across a diverse spectrum of facilities, including municipalities and agricultural enterprises. Moreover, this technology holds the potential to enable the efficient separation and recovery of valuable resources like ammonia and VFA. Concurrently, it enhances sludge thickening, representing a noteworthy advantage for downstream solids treatment processes. Integration of this technology into the fermentation process results in a significant reduction in process volume while concurrently enhancing the yield of VFA for utilization by BNR processes.

The derivation of VFA from fermentation has merit for facilities requiring supplemental carbon to meet nutrient removal targets in their liquid treatment processes. This extracted VFA resource becomes instrumental for facilities aiming to enhance EBPR performance, consequently lowering effluent total phosphorus (TP) concentrations.

A comprehensive analysis was conducted to assess various aspects, including process performance, capital investment, life-cycle cost, and greenhouse gas (GHG) emissions, pertaining to the utilization of IntensiCarb in comparison with chemical addition and traditional fermentation alternatives for TP removal. The process benefits of IntensiCarb have been demonstrated at bench-scale, but an assessment of economic viability had not been completed until this study. This assessment utilized influent wastewater parameters from the Metropolitan Water Reclamation District (MWRD) Hanover Park Water Reclamation Plant (WRP) and Calumet WRP to base process calculations on real-world wastewater treatment experiences. Employing a SUMO process modeling software developed by Dynamita, VFA yield determination relied on combined sludge characteristics modeled after the Hanover Park WRP (Greeley and Hansen, 2022) and was further calibrated to align with yields obtained from laboratory-scale tests (Okoye et al., 2022).

The life-cycle cost evaluation was conducted with four alternatives:

- Ferric chloride ($FeCl_3$) addition
- Supplemental carbon addition to support EBPR
- Conventional fermentation (CF) for VFA production to support EBPR
- IntensiCarb® fermentation (ICF) for VFA production to support EBPR

The analytical framework outlined herein encompasses three principal objectives: firstly, to juxtapose the process performance disparities between conventional and intensified fermentation; secondly, to scrutinize the capital costs associated with both conventional and intensified fermentation; and thirdly, to evaluate the life-cycle cost of intensified fermentation versus alternative methods for the provision of sufficient supplemental carbon needed for EBPR to meet target effluent TP concentrations at a full-scale operational level. The analysis was expanded to encompass TP removal via FeCl3 addition and quantified differences in GHG emissions across all alternatives.

MATERIALS AND METHODS

Fermentation bench-scale study data

Bench-scale laboratory experiments were conducted to assess the impact of intensifying fermentation through vacuum evaporation employing the IntensiCarb technology (Okoye et al., 2022). Sludge was anaerobically mixed at 60°C under an internal absolute pressure of 150 millibar. The vacuum pressure was specifically calibrated for evaporation, aiming to decouple SRT and HRT within each vessel. Each experiment maintained a controlled 3-day SRT, while three distinct HRTs were subjected to testing: 1.5, 2.25, and 3 days. The intensification factor, denoting the ratio of HRT to SRT, was defined and tested at values of 1, 1.3, and 2. An intensification factor of 1 represented CF, while factors exceeding one denoted intensified fermentation processes. Sample collection from each experiment involved filtration through a 0.45 μm filter, and analyses, conducted in triplicate, encompassed the determination of soluble chemical oxygen demand (sCOD) and VFA using Hach methods. Further details are comprehensively documented by Okoye et al. (2022).

The results of the bench-scale study indicate that elevating the intensification factor correlates with improved Volatile Solids (VS) removal efficiency, enhanced recovery of ammonia in the condensate, and increased production of VFA. The application of vacuum-assisted intensification exhibits a favorable impact compared to conventional fermentation, particularly in the context of generating additional VFA. Comparable sCOD and VFA yields were observed between intensification factors of 1 and 1.3, but a noteworthy increase in yields was evident at an intensification factor of 2. Consequently, an intensification factor of 2 was deemed optimal for subsequent process modeling endeavors. In summary, the study underscores the advantages associated with higher intensification factors, emphasizing their positive influence on key parameters such as VS removal efficiency, ammonia recovery, and VFA production in the context of wastewater treatment processes.

Fermentation Alternatives Modeling.

The process alternatives concerning the implementation of ICF as an alternative to CF were conducted using a model that was developed in SUMO (Dynamita, France), specifically to mimic the physical and biological processes of ICF (Figure 1). An ICF model unit was developed to include a reactor volume with headspace and allowed for evaporation of water vapor and gas transfer through the liquid/gas phases of VFA and ammonia. The SUMO2 model was used to govern biological and chemical reactions. The model was calibrated using experimental data obtained from the Western University (WU) laboratories bench scale studies (Okoye et al., 2022). The work from WU demonstrated improvement in hydrolysis and fermentation when a vacuum was applied to extract water vapor and gases while concentrating the fermenter content. In this study, an intensification factor of 2 (i.e., half the fermenter reactor volume compared to CF) was simulated.

The model encompasses key components, namely a fermentation tank, a condensation unit (exclusive to ICF), and a solids separation unit. Its utility extended to the prediction of VFA yield (expressed as grams COD per grams volatile suspended solids [VSS]) and the total VFA loading available as a supplementary carbon source for EBPR, quantified in g-COD/d. The modeling approach incorporated combined PS and WAS characteristics derived from historical average operational data sourced from Hanover Park WRP. To enhance the model's accuracy, VFA stripping mechanisms were introduced, and the modeling encompassed both liquid and vapor phases for ammonia and VFA in the case of ICF.

Process calculations

The outcomes of process modeling, specifically focusing on VFA yield and the anticipated fermentate flow available for EBPR, were integrated into a Microsoft Excel-based model. This model served as a tool to project key parameters such as secondary effluent TP levels (expressed in mg-P/L), biogas production (in cubic feet per minute [cfm]), and biosolids production (in tons per day).

The comparative analysis of each alternative involved the utilization of influent wastewater characteristics sourced from two operational wastewater treatment facilities managed by the MWRD of Greater Chicago: Hanover Park WRP and Calumet WRP. Historical wastewater characteristics, process performance, and operating cost estimates for each facility were obtained from process model calibration technical memorandums and phosphorus removal feasibility studies (Black and Veatch, 2016; Greeley and Hansen, 2018, 2022).

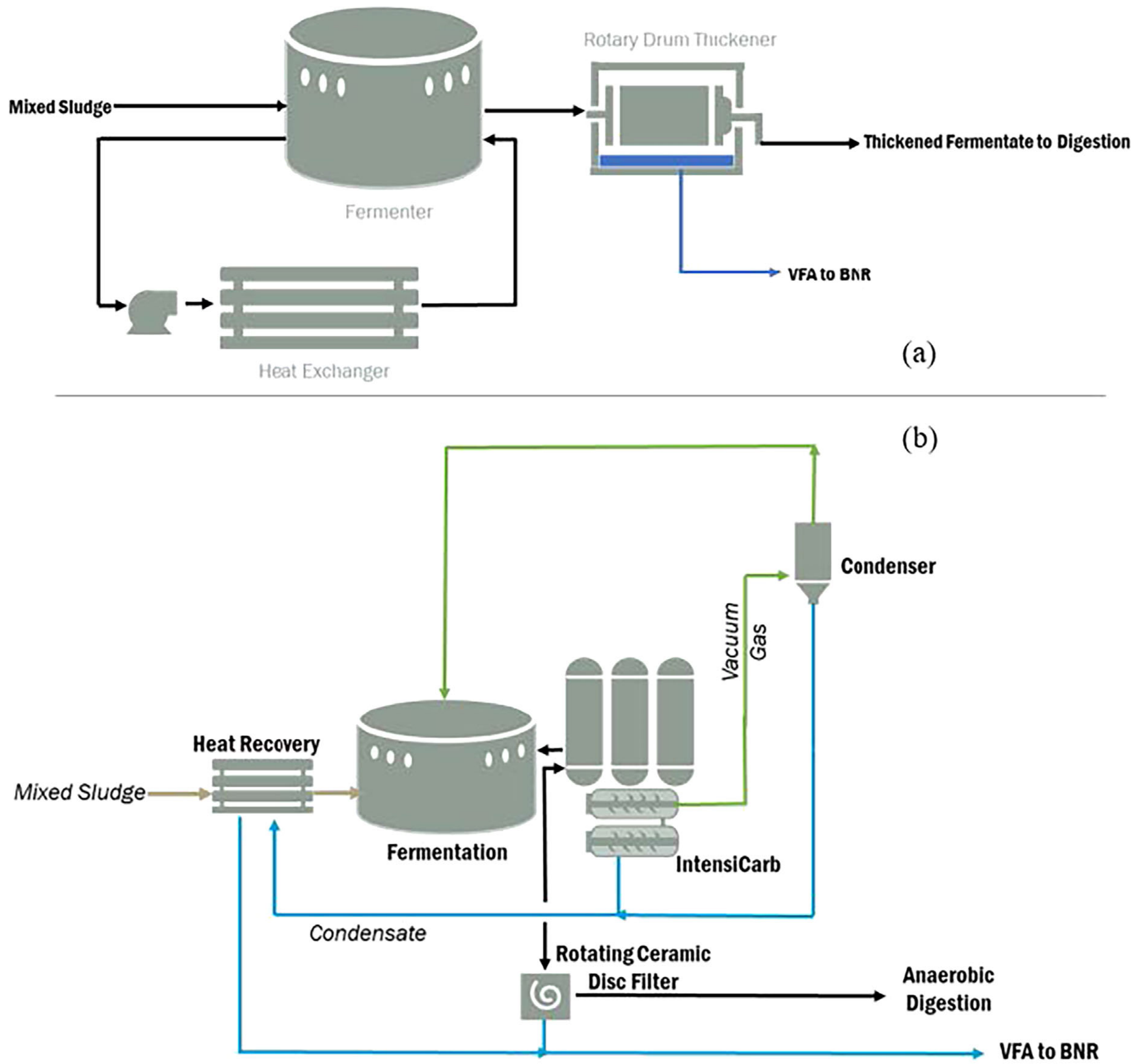

FIGURE 1 (a) Generalized process flow diagram for conventional sludge fermentation; (b) simplified process flow for IntensiCarb® for fermentation.

- Influent TSS concentration = 120 to 142 mg/l
- Influent VSS/TSS ratio = 0.75 to 0.88 g/g
- Influent COD:TP ratio = 45 to 80 g/g
- Influent COD:TKN ratio = 11.5 to 11.8 g/g
- Influent soluble unbiodegradable COD = 0.09 to 0.11 g/g-Total COD
- Influent particulate unbiodegradable COD = 0.22 to 0.30 g/g-Total COD
- Influent readily biodegradable COD = 0.13 to 0.20 g/g-Total COD

Other process performance parameters were assumed based on typical values, and these assumptions were validated by comparing process calculations to Hanover Park WRP and Calumet WRP annual average performance while using influent wastewater strength and fractions presented above.

- Primary TSS removal = 55%
- WAS volatile solids reduction = 30%
- PS volatile solids reduction = 60%

Liquid treatment process yields, decay rates, VSS production, and cell nutrient content were assumed based on typical values presented in Metcalf and Eddy (2014).

Capital and life-cycle costs approach

Capital cost estimates were prepared for each alternative, integrating vendor budgetary estimates obtained from PRAB for vacuum technology and condensation collection tanks into construction cost estimates for full-scale ICF. Estimates were considered Class V for all alternatives based on the level of project definition, characterized by an accuracy range of +100%/−50% (Bredehoeft et al., 2020). In the process of scaling up the Class V capital costs derived from a 38 megaliter per day (MLD) facility to accommodate larger capacities of 190 and 950 MLD, a scaling factor of 0.85 was applied using Equation 1. This scaling factor was predicated on the assumption that certain economies of scale would be realized as the facility capacity increased.

$$\text{Cost B} = \text{Cost A}^{*}\ (\text{Capacity B}/\text{Capacity A})\hat{}(\text{n}) \quad (1)$$

Where, n = scaling factor.

Capacity = fermentation tank design sludge flowrate, m^3/d.

The life-cycle cost assessment was a 20-year net present value (NPV) comparison, wherein all unit costs were reflective of values pertinent to the greater Chicago area or derived from the MWRD phosphorus feasibility studies.

Unit costs, operating, and life-cycle evaluation assumptions

The evaluation used specific unit cost assumptions integral to the analysis, encompassing the following parameters: the blended energy unit cost was established at $0.07 per kilowatt hour (kWh); the heating (natural gas) unit cost was fixed at $4.50 per million British thermal units (MMBTU) or approximately $0.0043 per megajoule (MJ); the unit cost of supplemental carbon (MicroC®-2000) was determined to be $1.13 per liter; the unit cost of $FeCl_3$ (40% weight by volume) was set at $468 per dry ton; labor unit cost amounted to $100 per full-time equivalent (FTE); and hauling cost was estimated at $99 per wet ton, assuming 40% total solids in biosolids. These unit cost values align with and draw from treatment cost data reported in lump sum values within the Phosphorus Feasibility Studies for Hanover Park and Calumet WRP conducted by (Greeley and Hansen, 2019, 2022), respectively. MicroC-2000 (MicroC) was assumed as the supplemental carbon source in alignment with phosphorus removal feasibility studies previously conducted for the two MWRD operational wastewater treatment facilities that were the basis of this evaluation. A unit cost of $4.29 per gallon for MicroC was derived from budgetary proposals furnished by Environmental Operating Solutions, Inc. (EOSI). MicroC was assumed to have 1,100,000 mg/l COD according to data sheets provided by EOSI.

The evaluation incorporated specific operating assumptions integral to the analysis, including Replacement and Rehabilitation (R&R) costs set at 2% of major mechanical equipment costs; labor allocations of 0.1 FTE for $FeCl_3$ and MicroC addition, 0.25 FTE for CF, and 0.5 FTE for ICF; a heat loss assumption of 3% for digestion and fermentation processes; and a 0.5% annual population and flow growth rate. The operating assumptions were developed based on a comparison of planning level assumptions developed by Brown and Caldwell's solids and energy practice subject matter experts, and labor assumptions were developed using PRAB recommendations on vacuum equipment requirements. Assumptions for energy and chemical utilization were developed based on typical ranges and averages documented in Metcalf and Eddy (2014), including a tank mixing energy of 8 kW per 1000 cubic meters was assumed for fermentation alternatives; pumping energy calculations assumed 15 m of head and a pump efficiency of 70% for sludge with a density of 1.03; $FeCl_3$ was assumed to be required at a rate of 5.24 g per gram of phosphorus (P) removed. Specific Chemical Oxygen Demand (COD) utilization assumptions were also based on Metcalf and Eddy (2014), and were assumed to vary depending on the carbon source: 20 g of influent readily biodegradable COD (rbCOD) per gram P removed, 20 g of MicroC COD per gram P removed, and 10 g of VFA as COD per gram P removed. An energy requirement for vacuum technology was assumed at 13 W-hours per liter of distillate produced. The vacuum technology energy requirements were based on PRAB equipment cut sheets assuming plant hot water is utilized and fermenter sludge is heated independently. The evaluation incorporated specific escalation and life-cycle assumptions essential to the analysis. A 1.7% annual escalation for Energy, Labor, Capital Cost, $FeCl_3$, and R&R costs was assumed to align with recent MWRD studies (Greeley and Hansen, 2019, 2022). A 2.6% annual escalation was applied to the unit cost of supplemental carbon due to observations of more rapidly increasing MicroC costs from increased customer demand. The supplemental carbon escalation rate assumption was calculated based on the Engineering News-Record Construction Cost Index for Chicago from 2019 to 2022. To reduce future cash flows and expenses, a nominal discount rate of 2.9% was employed based on the annual interest rate documented in previous, recent studies from MWRD (Greeley and Hansen, 2019). Greenhouse Gas Emissions Assumptions.

This section delineates the GHG emission assumptions, encompassing items categorized within Scope 1, 2, and 3 emissions, along with the corresponding emission factors. Process performance outcomes derived from process calculations were used to estimate GHG emissions. Scope 1 emissions comprise methane emissions from fugitive biogas, emissions from biogas combustion excluding biogenic carbon dioxide (with the assumption that all products are predominantly biogenic, including MicroC), and emissions arising from biosolids processing operations, which include polymer utilization and heating. Additionally, carbon sequestration by tree-covered land owned by the MWRD was accounted for as a Scope 1 sink. Scope 2 emissions were associated with purchased electricity use, while Scope 3 emissions encompassed the impact of purchased goods ($FeCl_3$ and MicroC). Emission factors, expressed in carbon dioxide equivalents (CO2e), were specified for various sources, including electricity (316 kg CO2e/MWh based on MWRD power supply), natural gas (53.1 kg CO2e/mmBTU), diesel fuel (10.2 kg CO2e/gal per EPA, 2018), polymer (1.6 kg CO2e per kg polymer used per the Biosolids Emissions Assessment Model [BEAM]), $FeCl_3$ (0.45 kg CO2e per kg $FeCl_3$ per Greeley and Hansen, 2019), MicroC (0.13 kg CO2e per gallon MicroC based on an assumed 10:1 biomass to methanol blending for the biodiesel product that MicroC is derived from), and biosolids carbon sequestration credit (−0.15 kg CO2e per dry lb biosolids land applied). Chemical and biosolids delivery distances were assumed to be 100 miles per trip with a 6000 gal tank size per delivery for MicroC and 17 wet tons per load capacity for biosolids haulers (based on the capacity of a tandem axle truck from the Illinois Depart of Transportation Size and Weight Laws). Diesel fuel efficiency was assumed to be 6.4 miles per gallon for these deliveries. These factors provide a quantitative basis for evaluating the environmental footprint associated with each alternative.

RESULTS

Process modeling and calculations

Fermentation modeling results predicted a higher VFA yield with ICF relative to CF. This was consistent with the trend observed in the bench-scale study data. Table 1 presents the model results. As shown, ICF increased the production of VFA and rbCOD. This results in a greater loading of VFA and rbCOD available for utilization in the BNR treatment process.

In Figure 2, process calculations illustrate anticipated alterations in PS VSS loading to digestion, WAS VSS loading to digestion, biogas production, and biosolids production across four scenarios: $FeCl_3$ addition, MicroC addition, CF, and ICF. These predictions are based on the average of influent wastewater characteristics summarized in process calculations materials and methods.

The introduction of $FeCl_3$ does not contribute to VSS loading in the digestion process. However, it does lead to a 7% increase in biosolids production attributed to the formation of hydroxide sludge. This increase in biosolids production elevates hauling costs, a crucial factor in economic analysis. It is important to note that $FeCl_3$ addition is predicated on ideal dosing assumptions, and the actual rise in sludge yield may exceed the predicted value. Therefore, refining such estimates becomes imperative in site-specific evaluations, especially when considering the sizing of solids handling processes and planning for end-use applications.

MicroC addition increases the WAS VSS loading to digestion by 16%, biogas production by 4%, and biosolids production by 7%. This could be a potential increase in revenue if biogas is utilized for resource recovery but also a potential increase in biosolids hauling costs.

Fermentation alternatives convert VSS that would otherwise be utilized by digestion into VFA that is redirected to the BNR for EBPR. There is maximum availability of VFA based on the SUMO fermentation model, and any additional supplement carbon requirement is assumed to be satisfied by MicroC addition. Therefore, PS VSS loading is predicted to decrease for both CF and ICF scenarios by about 24% to 29%, and a decrease in WAS VSS loading of 1% to 4% is predicted based on biomass yield from fermentate VFA and biodegradable COD. Biogas production decreases by about 19% to 22%. Biosolids are predicted to decrease based on process calculations; however, no decrease in biosolids will be assumed for the economic analysis for a more conservative life-cycle cost for fermentation scenarios.

TABLE 1 Fermentation model results.

Parameter	Unit	CF	ICF
SRT	Days	3	3
HRT	Hours	72	36
Intensification factor	--	1	2
VFA yield	%	10	15
Increase in VFA direction to BNR	%	--	67
Increase in rbCOD+VFA direction to BNR	%	--	26

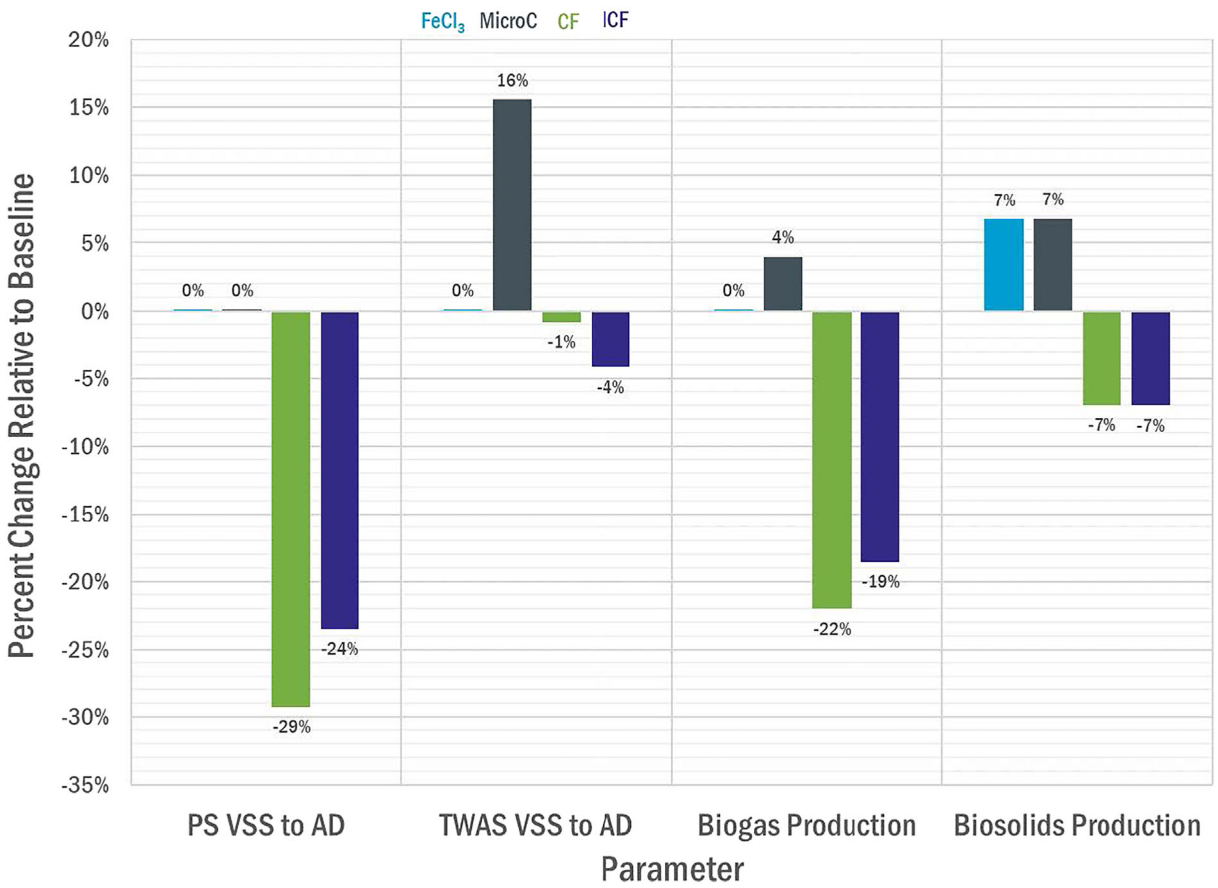

FIGURE 2 Process calculation predictions.

Capital and life-cycle costs

The comprehensive capital outlay for a CF facility varied within the range of \$9.3 to \$143 million, while for ICF, the total capital costs spanned from \$9.4 to \$166 million. Despite the higher estimated capital costs associated with ICF, the potential advantages of a smaller fermentation tank volume and reduced footprint could prove beneficial for utilities constrained by site limitations. The elevated cost primarily stems from the incorporation of vacuum technology. Notably, the capital cost estimates for facilities employing chemical addition methods, namely $FeCl_3$ and MicroC, were presumed to be identical. Capital cost estimates are provided in Tables 2 and 3 for reference.

The outcomes of life-cycle cost assessments exhibited a notable sensitivity to variations in influent wastewater characteristics. To underscore this sensitivity, the fractions representative of Hanover Park and Calumet WRP influence were employed in the process calculations. The ensuing results, depicted in Figure 3, incorporate the influent wastewater characteristics specific to each WRP. Projections for Hanover Park WRP indicate an ample supply of influent carbon for successful EBPR down to a 1 mg/l TP level. Consequently, a target of 0.5 mg/l TP was employed to facilitate the comparative analysis of life-cycle costs, aligning with the wastewater characteristics inherent to Hanover Park WRP.

Irrespective of influent concentration levels, $FeCl_3$ consistently demonstrated the lowest life-cycle cost. However, it's imperative to note that the exclusive use of $FeCl_3$ imposes a higher impact on resource utilization and Scope 3 GHG emissions. The subsequent discussion focuses on the comparative analysis of EBPR alternatives, namely MicroC, CF, and ICF.

With reference to Hanover Park WRP, if EBPR can be effectively executed with minimal chemical addition, then a chemical addition alternative emerges as the most favorable option. Meanwhile, based on the characteristics of Calumet WRP, facilities constrained by carbon limitations may find value in contemplating either CF or ICF. The rationale behind this lies in the lower operating costs associated with these alternatives, thereby resulting in a reduced 20-year life-cycle cost. Notably, MicroC addition was deemed necessary for CF but not for ICF, owing to disparities in VFA yield. Consequently, this discrepancy contributed to a lower life-cycle cost for ICF when compared to CF.

TABLE 2 Capital cost estimates for fermentation alternatives.

	Small facility (38 MlD)		Medium facility (190 MlD)		Large facility (950 MLD)	
Parameter	**CF**	**ICF**	**CF**	**ICF**	**CF**	**ICF**
Tanks	2.21 mil	1.10 mil	6.80 mil	3.40 mil	21.0 mil	10.5 mil
Building	2.93 mil	2.93 mil	9.05 mil	9.05 mil	27.9 mil	27.9 mil
Equipment	0.96 mil	3.28 mil	2.97 mil	10.8 mil	9.15 mil	44.4 mil
Yard piping	0.96 mil	0.96 mil	2.97 mil	2.97 mil	9.15 mil	9.15 mil
Construction markups	1.66 mil	1.94 mil	5.12 mil	6.15 mil	15.8 mil	21.6 mil
Contingency	0.95 mil	1.12 mil	2.94 mil	3.53 mil	9.06 mil	12.4 mil
Total	9.28 mil	9.42 mil	36.4 mil	39.7 mil	143 mil	166 mil

TABLE 3 Capital cost estimates for chemical and carbon addition alternatives.

	38 MLD		190 MLD		950 MLD	
Parameter	**$FeCl_3$**	**MicroC**	**$FeCl_3$**	**MicroC**	**$FeCl_3$**	**MicroC**
Total	1.2 mil	1.2 mil	4.7 mil	4.7 mil	18.5 mil	18.5 mil

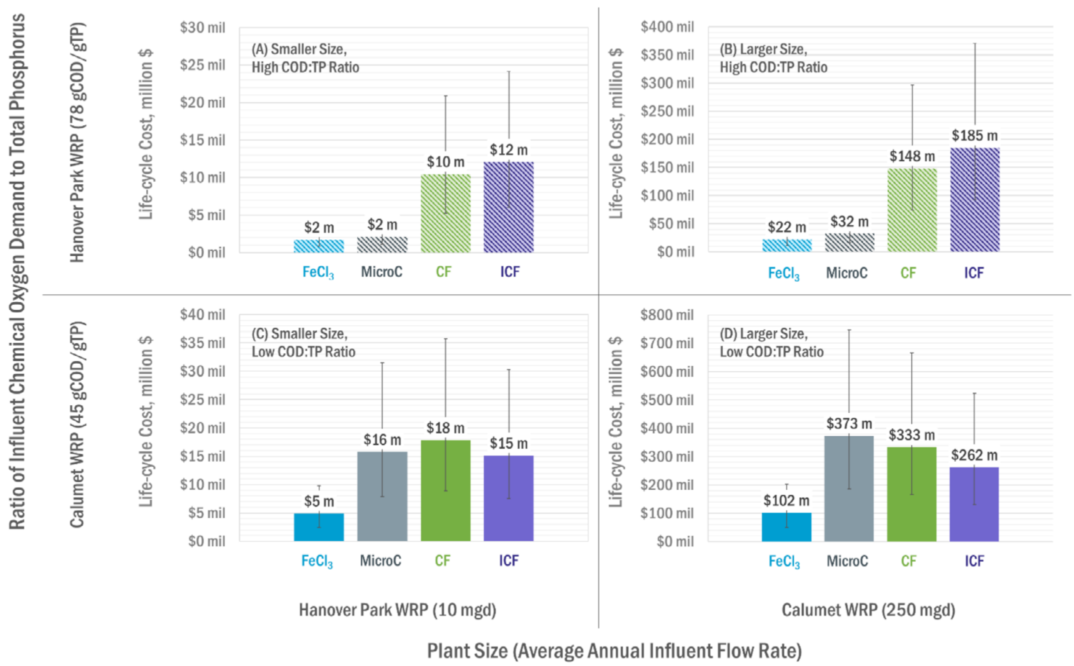

FIGURE 3 Comparison of life-cycle cost results for Hanover Park versus calumet (more carbon-limited) with −50% and +100% error bars.

Carbon footprint estimates

The GHG emission estimates for each alternative are visually depicted in Figure 4. $FeCl_3$ is notably associated with substantial emissions stemming from the production and transportation of the chemical. MicroC, on the other hand, is a byproduct derived from a biodiesel production process, wherein methanol is blended with biomass content at a ratio of 1:10. Despite the relatively low carbon content linked to MicroC production and the biosolids generated through its application, these emissions are counterbalanced by carbon sequestration resulting from the product's utilization on land owned by the MWRD. In contrast, fermentation alternatives redirect PS carbon from anaerobic digestion for utilization in the secondary treatment process, contributing to increased emissions due to the power requirements for operating mixing and pumping equipment. The ICF

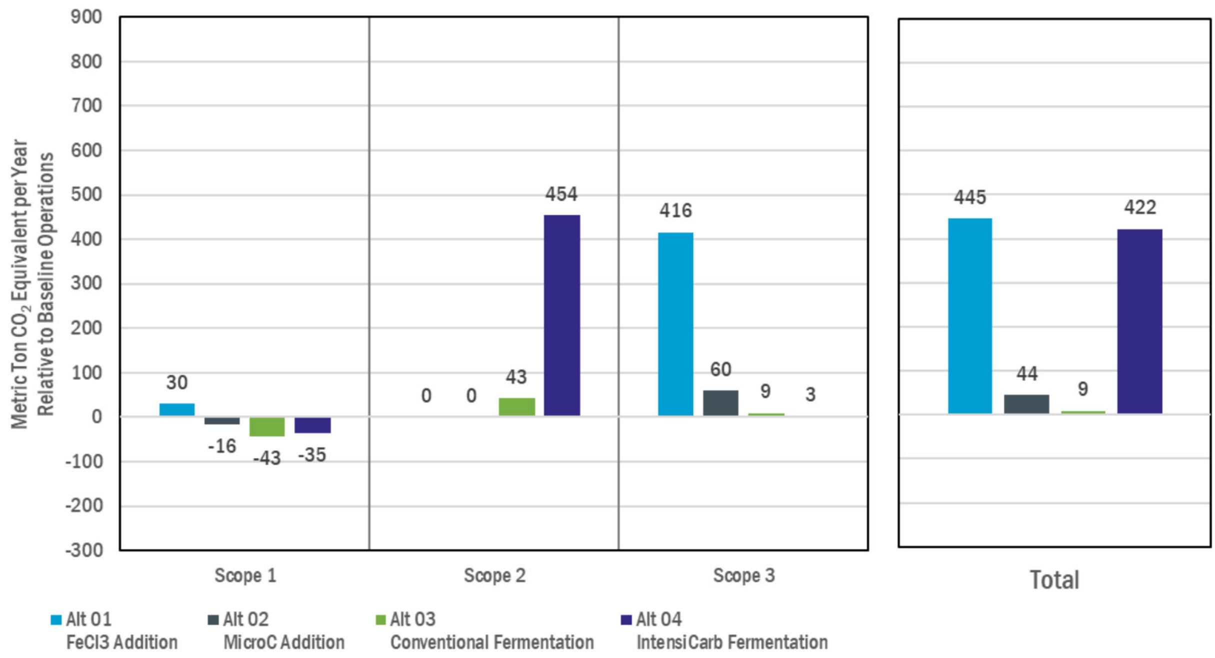

FIGURE 4 GHG emissions comparison by scope.

process, in particular, exhibits a significantly higher energy demand attributable to the operation of vacuum equipment.

In summary, the GHG emission estimates underscore $FeCl_3$ and ICF as the alternatives with the highest emissions. $FeCl_3$ is characterized by a substantial Scope 3 emission, primarily associated with the CO2e emissions linked to the manufacturing of the product. Conversely, ICF has the second highest emissions and a notably high Scope 2 emission, attributed to the electrical demand necessitated by the operation of vacuum technology.

DISCUSSION

Sensitivity analysis

The outcomes of the economic analysis indicate that IntensiCarb exhibits a life-cycle cost comparable to conventional fermentation. Smaller facilities, particularly those with a capacity below 38 MLD, might incur a lower life-cycle cost with chemical addition, while larger facilities, exceeding 190 MLD, could benefit from fermentation alternatives. Notably, the assumptions in this analysis point towards a lower NPV for $FeCl_3$ addition when compared to other EBPR alternatives, namely MicroC or fermentation.

Figure 5 illustrates the variations in capital cost estimates when utilizing scaling factors of 0.7, 0.85, or 1.0. Noteworthy discrepancies in capital costs emerge based on the chosen scaling factor, with the costs referenced from recent construction estimates pertaining to a facility categorized as near the "small" facility size.

In Figure 6, the difference between employing a 0.85 and 1.0 scaling factor on life-cycle costs is depicted. The higher scaling factor would lead to an elevated present value cost for fermentation alternatives within the 190 to 950 MLD facility size range. However, the Net Present Value costs for fermentation alternatives remain comparable to those associated with MicroC.

Considerations for chemical addition versus in-situ fermentation

The primary objectives of this study were to compare ICF against more traditional means of supplementing rbCOD for an EBPR process to achieve low effluent TP limits. In completing this study, the life-cycle cost associated with non-biological means of removing TP (i.e., $FeCl_3$ addition) was a necessary alternative for comparison to provide complete context on the cost to treat facilities with effluent TP limits. The results demonstrate significantly lower life-cycle costs for the $FeCl_3$ alternative. The lower life-cycle costs were primarily influenced by significantly lower capital construction costs. The assumed unit price for chemical deliveries, based on current market conditions, resulted in lower life-cycle costs over a 20-year duration. This study does not speculate on future price changes in unit pricing for $FeCl_3$. If regulatory pressures

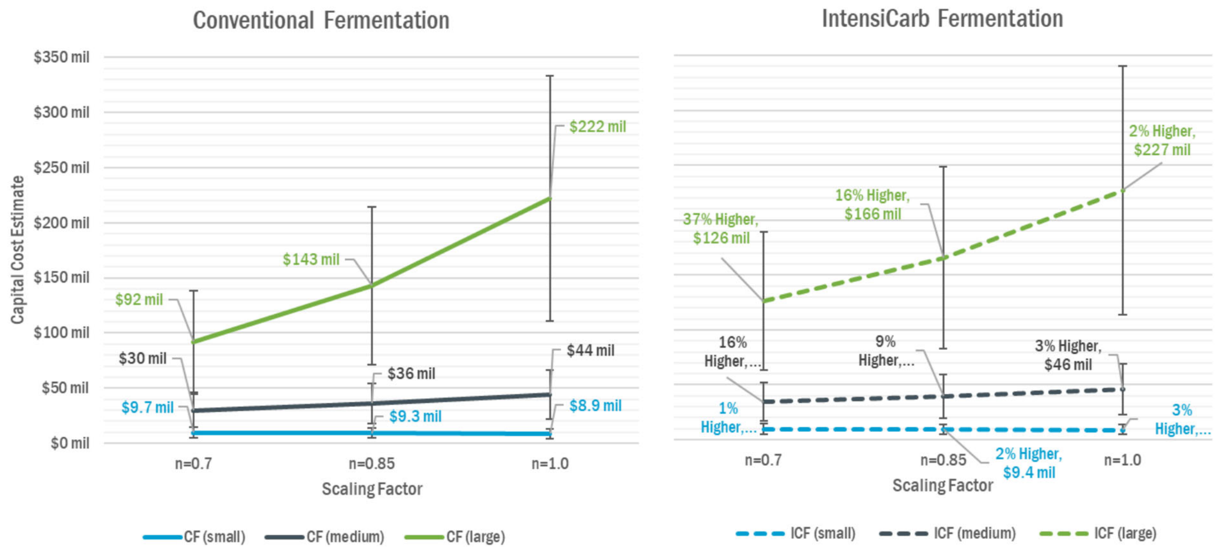

FIGURE 5 Comparison of capital cost estimates based on selected scaling factor.

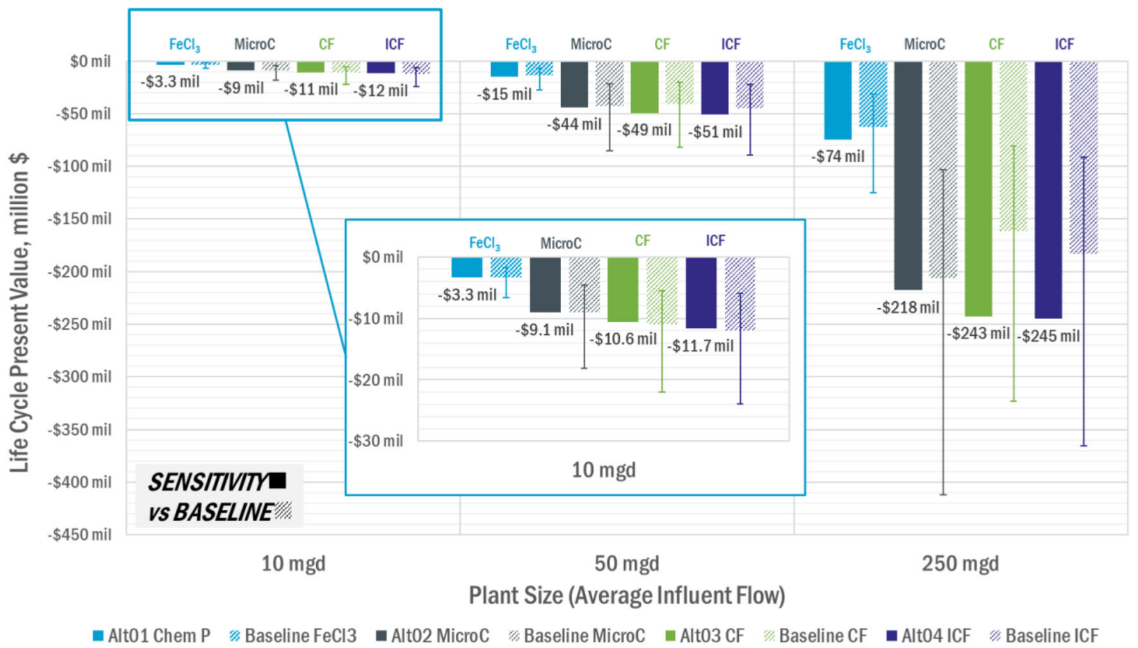

FIGURE 6 Life cycle cost comparison. * The baseline values represent capital cost estimates with a 0.85 scaling factor, and sensitivity shows results with a 1.0 scaling factor.

increase demand for $FeCl_3$, the annual operating costs may be subject to significant increases. In this scenario, the life-cycle cost comparison would be more comparable between CF and ICF alternatives and the $FeCl_3$ alternative. Non-economic considerations may also be considered by municipalities, such as the impact of chemical deliveries on the facility and surrounding communities. These types of considerations were not included in the scope of this study.

In addition to having the lowest life-cycle cost, the $FeCl_3$ alternative's GHG emissions were higher but similar to ICF. However, the $FeCl_3$ alternative's GHG emissions were primarily due to Scope 3 emissions, which a facility has limited to no control over. In contrast, ICF's GHG emissions were primarily due to Scope 2 emissions from the energy demand to operate the vacuum technology. Facilities have more control over minimizing these impacts by generating energy on-site via biogas utilization. Facilities with electricity produced via renewable means would also have lower GHG estimates than those calculated in this study, which were based on energy production specific to MWRD.

Fermentation alternatives provide utilities with greater control over operating cost stability but are assumed to have higher capital costs. Facilities with existing tankage that is readily convertible to a fermentation process (e.g., unused gravity settling or thickening tanks) may have lower life-cycle costs for CF and ICF than presented in the results of this study. Compared to CF, the ICF process, which requires a covered tank for the vacuum, may be desirable due to the inherent capture of odors via the vacuum. The ICF process also uses half the volume, which may allow for the reuse of existing tankage to be feasible compared to a CF process.

This discussion demonstrates unexplored considerations for site-specific or market-driven conditions that could make ICF economically beneficial. The capital cost and life-cycle cost similarities between CF and ICF are therefore supportive of further evaluation of ICF for projects that require EBPR enhancement to meet lower TP limits. The lowest effluent TP limit evaluated in this study was 0.5 mg/l TP, but even lower limits could result in the CF and ICF appearing more favorable based on the trend demonstrated in Figure 3.

CONCLUSION

This study compared IntensiCarb, a fermentation process intensification alternative, against conventional fermentation, supplemental carbon addition, and ferric chloride addition to evaluate economic feasibility and GHG emissions. The results suggest ICF may be an attractive alternative for facilities with low effluent TP requirements, limited footprint, and a low influent COD to TP ratio. Key findings include:

- Modeling results determined ICF increases available VFAs for EBPR by 67% and rbCOD plus VFAs for EBPR by 26% compared to CF. These results are at half the HRT of CF with an identical SRT.
- ICF results in a significant reduction in the required process volume for fermentation with an estimated increase in capital costs by 1.5% to 15% when compared to CF.
- Lifecycle cost comparisons between ICF and CF were similar but dependent on the influent carbon-to-phosphorus ratio.
    - A COD/TP ratio of 78 g/g resulted in 20% to 25% higher lifecycle cost for ICF versus CF.
    - A COD/TP ratio of 45 g/g resulted in 17% to 21% lower lifecycle cost for ICF versus CF.
    - This trend was more pronounced when comparing ICF and supplemental carbon with MicroC.
- $FeCl_3$ and ICF alternatives had the highest GHG emission estimates compared to supplemental carbon addition with MicroC and CF alternatives. FeCl3 GHG emissions were primarily Scope 3, and ICF emissions were primarily Scope 2 from electrical demand to maintain the vacuum technology.

Future work and recommendations.

The subsequent steps in the Techno-Economic Analysis could involve the incorporation of additional factors into the economic assessment. This encompasses a comprehensive examination of the market value of fermentation products, an evaluation of diverse fermentation flow schemes, and the consideration of various design criteria, such as utilizing PS exclusively or employing smaller tanks. Furthermore, an economic model could be deployed to compare the conventional and vacuum-based intensification technologies, specifically IntensiCarb, for both digestion and nitrogen removal applications. It is noteworthy that, in comparison to conventional fermentation, IntensiCarb exhibits distinctions in microbial communities, including an enrichment of acid-forming bacteria. Additionally, it demonstrates a lower viscosity, facilitating smoother operation and flow, along with higher solid destruction efficiency. Furthermore, IntensiCarb generates increased ammonia, a characteristic that could be strategically removed and recovered through the application of vacuum technology. While these differences did not directly impact the cost analysis, they may prove advantageous in various other applications within the wastewater treatment context.

AUTHOR CONTRIBUTIONS

Maxwell Armenta: Writing – original draft; writing – review and editing; investigation; formal analysis; software; data curation. **Farokh Laqa Kakar:** Writing – original draft; project administration; conceptualization; investigation; supervision; resources; funding acquisition. **Ahmed Al-Omari:** Investigation; conceptualization; validation; software. **Christopher Muller:** Writing – review and editing; supervision; conceptualization; methodology; validation; investigation. **George Nakhla:** Conceptualization; writing – review and editing; supervision; validation; investigation; funding acquisition; project administration; resources. **Domenico Santoro:** Conceptualization; supervision; investigation; funding acquisition; resources. **Katherine Bell:** Conceptualization; project administration; supervision; funding acquisition; validation; resources.

CONFLICT OF INTEREST STATEMENT

The authors declare no competing financial interest.

DATA AVAILABILITY STATEMENT

The data that support the findings of this study are available from the corresponding author upon reasonable request.

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