



Simulation Coupled Preliminary Scale Up and Equipment Capacity Assessment of Corncob Based Lactic Acid Production

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ABSTRACT

The translation of lignocellulosic lactic acid processes from simulation to industrial development requires a consistent link between process flow rates, residence times and equipment capacity. This study presents a simulation coupled preliminary scale-up and equipment capacity assessment for an integrated corncob to lactic acid process comprising dilute acid pretreatment, enzymatic hydrolysis, fermentation, membrane filtration, acidification, purification and product storage. Reconciled steady state process outputs at a reference corncob throughput of $2,500 \text{ kg h}^{-1}$ were linked to residence time, working volume, membrane flux, tray spacing and storage duration relationships. An 80% vessel working-volume fraction and a 15% membrane-area allowance were applied. The calculated total requirements were 11.35 m^3 for pretreatment, $1,096.9 \text{ m}^3$ for enzymatic hydrolysis, 23.49 m^3 for fermentation, 169.34 m^2 of membrane area, 15.43 m^3 for acidification, a 13.20 m preliminary purification-column height and 49.26 m^3 of product storage. Nominal capacities of 12, 1,100, 25 and 20 m^3 were selected for the four process vessels, together with 170 m^2 membrane area, a 14 m column and a 50 m^3 storage tank. Hydrolysis accounted for approximately 95.6% of the calculated working volume of the residence time-controlled process vessels because of its 61-h residence time, identifying it as the principal volumetric bottleneck. The purification reboilers required $1,164 \text{ kW}$ and represented 43.5% of the total $2,678 \text{ kW}$ heating demand, identifying purification as the main thermal bottleneck. The estimates define process-scale capacity requirements and support pilot planning; they are not final mechanical specifications. Mixing, rheology, heat-transfer area, vessel internals, pressure design and vendor verification remain necessary before commercial implementation.

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INTRODUCTION

Lactic acid is a versatile platform chemical used in food, pharmaceutical, cosmetic and chemical applications and as the monomer for polylactic acid. Fermentative production offers optical selectivity, but conventional processes often depend on refined sugars or starch. Corncob is a non-food lignocellulosic residue containing cellulose and hemicellulose that can be converted into fermentable sugars through suitable pretreatment and hydrolysis. Its utilisation can

support agricultural-residue valorisation, although the recalcitrant lignocellulosic structure creates technical challenges that extend beyond fermentation performance (Abedi et al., 2020; Malacara-Becerra et al., 2022).

In Nigeria, converting locally available corncob into lactic acid could strengthen agricultural-residue utilisation and provide a technical basis for domestic biochemical production. Achieving this objective requires process-scale information that connects

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laboratory conversion performance with realistic equipment capacities, production throughput and utility requirements.

Published corncob studies have demonstrated lactic acid production through separate or simultaneous hydrolysis and fermentation. Rivas et al. (2004) mathematically interpreted simultaneous saccharification and fermentation of treated corncob, while Zhang et al. (2016) reported high-titre fermentation of pretreated corncob using *Bacillus coagulans*. More recently, Adoga et al. (2026b) statistically optimised dilute acid hydrolysis of corncob and established optimal conditions for maximum glucose yield, while Adoga et al. (2026a) developed a kinetic model for lactic acid fermentation from the hydrolysate. These findings provide a laboratory basis for process development; however, translating the established operating conditions and kinetic relationships into industrial equipment capacities remains unaddressed in those studies.

Scale-up of lignocellulosic conversion is not achieved by multiplying laboratory dimensions by a constant factor. Residence time, slurry rheology, mixing, enzyme distribution, and heat and mass transfer interact with equipment geometry and throughput. High-solids hydrolysis is especially challenging because increasing solids concentration reduces available free water, raises apparent viscosity and yield stress, and restricts convective transport (Modenbach and Nokes, 2013; Shiva et al., 2022). Consequently, preliminary capacity calculations can identify the magnitude and location of scale-dependent constraints, but they must be distinguished from detailed mechanical design.

Pilot and semi-industrial scale studies demonstrate the feasibility of larger lactic acid fermentation systems. Lu et al. (2012) used a 3,000 L fermenter coupled with ceramic microfiltration, while Fu et al. (2017) demonstrated D-lactate production in a 30 m³ fed-batch fermenter. Conceptual process studies have also connected laboratory fermentation results to separation design and larger production capacities (Daengpradab and Rattanaphanee, 2018). Nevertheless, limited published work links

an integrated corncob-to-lactic-acid simulation directly to preliminary capacity requirements for all major unit operations, including pretreatment, hydrolysis, fermentation, filtration, product recovery and storage.

The present study therefore builds on the hydrolysis optimisation and fermentation kinetics reported by Adoga et al. (2026b) and Adoga et al. (2026a) to develop a simulation-coupled preliminary scale-up and equipment capacity assessment for integrated corncob-based lactic acid production. Emphasis is placed on mass and energy balances, reactor sizing and throughput analysis. The objectives were to: (i) translate reconciled steady-state material flows and nominal residence times into installed equipment capacities; (ii) identify the dominant volumetric and thermal bottlenecks; and (iii) define the additional hydrodynamic, mechanical and vendor-based information required before detailed engineering. The methodology preserves the specified throughput and nominal residence times without claiming complete geometric, mixing, heat-transfer or mass-transfer similarity.

MATERIALS AND METHODS

Process boundary and simulation basis

The process boundary extended from dried corncob, process water and steam inputs to purified lactic acid, filtration cake, gypsum and off-gas outputs. The process sequence comprised feed preparation, dilute-acid pretreatment, enzymatic hydrolysis, fermentation with *Bacillus coagulans*, membrane filtration, calcium-lactate acidification, purification and product storage. A steady-state equation-oriented process model supplied the reconciled mass-flow rates, densities, temperatures, pressures and thermal duties used in the capacity calculations.

The reference throughput was 2,500 kg h⁻¹ of corncob. The simulated process produced 1,984 kg h⁻¹ of lactic acid in approximately 2,002 kg h⁻¹ of purified product containing 99.1 wt% lactic acid. Equipment sizing was performed on the basis of the principal process streams rather than the dry feed alone. The stream and design data used are summarised in Table 1.

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Table:1 Design Basis for Preliminary Equipment-Capacity Estimation

Parameter	Value	Unit	Basis
Corncob feed rate	2,500	kg h ⁻¹	Reference simulation
Pretreatment stream flow rate	14,912	kg h ⁻¹	Steady-state model
Pretreatment stream density	931	kg m ⁻³	Steady-state model
Pretreatment residence time	34	min	Process-design input
Hydrolysis inlet flow rate	14,875	kg h ⁻¹	Steady-state model
Hydrolysis inlet density	1,034	kg m ⁻³	Steady-state model
Hydrolysis residence time	61	h	Process-design input
Fermenter inlet flow rate	14,520	kg h ⁻¹	Steady-state model
Fermenter inlet density	1,067	kg m ⁻³	Steady-state model
Fermentation residence time	82.846	Hr.	Optimisation result
Filtrate flow rate	13.40	m ³ h ⁻¹	Steady-state model
Membrane design flux	91	L m ⁻² h ⁻¹	Design input
Acidification inlet flow rate	13,674	kg h ⁻¹	Steady-state model
Acidification inlet density	1,108	kg m ⁻³	Steady-state model
Purified-product flow rate	2,002	kg h ⁻¹	Reconciled mass balance
Purified-product density	1,219	kg m ⁻³	Steady-state model
Vessel working-volume fraction	0.80	–	Design assumption
Product-storage duration	24	h	Design assumption

Scale-up philosophy

Material and energy balances and capacity equations were evaluated on one consistent simulation basis. The approach preserved material throughput, nominal residence time and specified operating conditions. It did not impose geometric similarity or constant power input per unit volume because the information required to establish these criteria, particularly slurry rheology, impeller performance, mixing time and heat-transfer coefficients was not available. The resulting values are therefore process-level capacity estimates for screening and pilot planning.

Volumetric flow and vessel capacity

The required working volume of each residence-time-controlled vessel was calculated simultaneously with the process balances using:

$$V_w = Q_v \tau \quad (1)$$

$$V_T = \frac{V_w}{f_w} \quad (2)$$

where V_{working} is the required working volume, V_{total} is the total installed vessel capacity, τ is the specified residence time and f_w is the working-volume fraction, taken as 0.80.

Where the calculated capacity exceeded the selected practical capacity of a single vessel, the required number of parallel units was determined from:

$$n_v = \frac{V_t}{V_{\text{unit}}} \quad (3)$$

where V_{unit} is the selected nominal capacity of one vessel. The continuous capacity requirement was calculated in gProms, while rounding to a whole number of vessels was performed as an engineering post-processing step.

Membrane-area estimation

The membrane area was linked directly to the simulated filtrate flow rate and specified membrane flux:

$$A_{m,\text{design}} = 1.15 \left(\frac{Q_f}{J} \right) \quad (4)$$

where A_m is the required membrane area (m²), Q_f is the simulated filtrate flow rate (L/h), J is the design membrane flux (L/m² h), and 1.15 represents the 15% allowance for fouling, cleaning and operational flexibility.

Purification-column and storage estimates

For the specified column configuration, the active column height was estimated from:

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$$H_{active} = (N_s - 1)s_t \quad (5)$$

where H_{active} is the active column height, N_s is the specified number of equilibrium stages and s_t is the adopted tray spacing. The number of stages was treated as a column-design input rather than a simulation result.

Product storage capacity was linked to the simulated purified-product volumetric flow rate:

$$V_{storage} = \frac{Q_{product}t_{storage}}{f_w} \quad (6)$$

where $Q_{product}$ is the purified-product volumetric flow rate and $t_{storage}$ is the specified storage duration of 24 h.

Integration with Energy Requirements

Equipment-capacity calculations were evaluated concurrently with the unit-operation energy balances. Heating and cooling duties for pretreatment, hydrolysis, fermentation, acidification and purification were calculated from the corresponding material flows, thermal properties, operating temperatures and reaction-energy terms. Thus, changes in throughput simultaneously altered equipment-capacity and utility requirements, providing a consistent basis for scale-up, sustainability and techno-economic evaluation.

RESULTS AND DISCUSSION

Equipment-capacity requirements

The pretreatment stream volumetric flow was $16.02 \text{ m}^3 \text{ h}^{-1}$. At 34 min residence time, the working volume was 9.08 m^3 and the installed requirement was 11.35 m^3 ; a 12 m^3 nominal reactor was selected. The hydrolysis inlet flow was $14.39 \text{ m}^3 \text{ h}^{-1}$. Its 61-h residence time increased the working volume to 877.5 m^3 and the installed requirement to $1,096.9 \text{ m}^3$, for which a total nominal hydrolysis capacity of approximately $1,100 \text{ m}^3$ was selected.

The fermenter inlet flow was $13.61 \text{ m}^3 \text{ h}^{-1}$. The optimised fermentation residence time of 82.846 Hr, produced a working-volume requirement of 18.79 m^3 and a total installed requirement of 23.49 m^3 . A nominal 25 m^3 fermenter was therefore selected. The acidification inlet flow was $12.34 \text{ m}^3 \text{ h}^{-1}$. Because the simulation did not specify an acidification residence time, a preliminary 1 h value was adopted, yielding 12.34 m^3 working volume and 15.43 m^3 installed volume; a 20 m^3 nominal reactor was selected.

The filtration step processed $13,400 \text{ L h}^{-1}$ of filtrate. At $91 \text{ L m}^{-2} \text{ h}^{-1}$, the base membrane requirement was 147.25 m^2 . Application of the 15% allowance increased this to 169.34 m^2 , rounded to 170 m^2 . The purification column required 10.20 m of active tray height and 13.20 m including disengagement and sump allowances; a nominal height of 14 m was retained. The purified-product volumetric flow was $1.642 \text{ m}^3 \text{ h}^{-1}$, giving 39.41 m^3 working storage and 49.26 m^3 installed storage for 24 h; a 50 m^3 tank was selected.

Table 2: Calculated Requirements and Selected Nominal Equipment Capacities

Equipment	Working requirement	Installed/design requirement	Selected nominal size
Pretreatment reactor	9.08 m^3	11.35 m^3	12 m^3
Hydrolysis reactor system	877.5 m^3	$1,096.9 \text{ m}^3$	$1,100 \text{ m}^3$ total
Fermenter	18.79 m^3	23.49 m^3	25 m^3
Membrane-filtration unit	147.25 m^2	169.34 m^2 including allowance	170 m^2
Acidification reactor	12.34 m^3	15.43 m^3	20 m^3
Purification column	10.20 m active height	13.20 m estimated total height	14 m ; 18 stages
Product-storage tank	39.41 m^3	49.26 m^3	50 m^3

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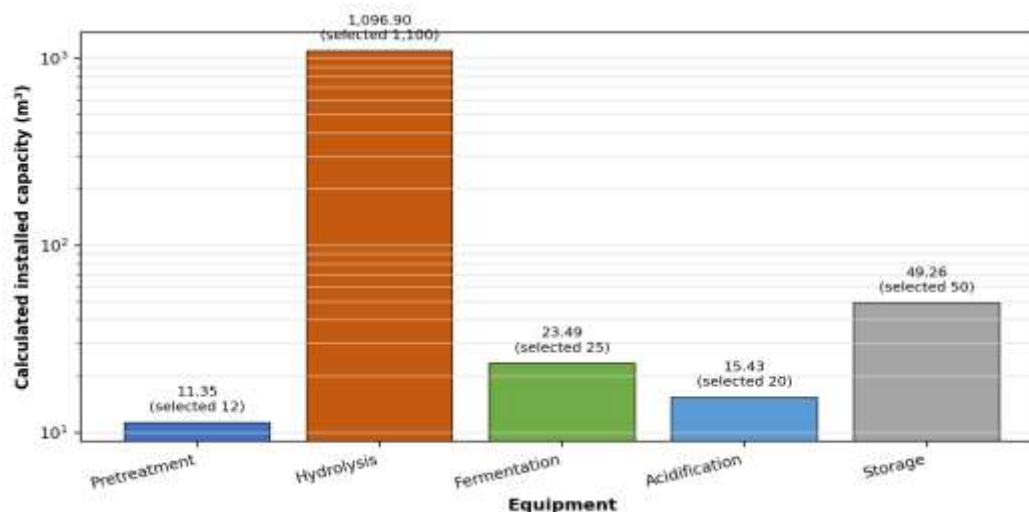


Figure 1: Calculated installed volumetric capacities and selected nominal equipment sizes.

A logarithmic vertical scale is used because the hydrolysis requirement is approximately two orders of magnitude larger than several other vessels.

Volumetric bottleneck

Hydrolysis represented approximately 95.6% of the combined calculated working volume of pretreatment, hydrolysis, fermentation and acidification vessels. Its installed capacity was about 46.7 times the fermenter requirement, even though their inlet volumetric flow rates were of similar magnitude. The difference was caused primarily by the 61-h hydrolysis residence time compared with 34 min pretreatment and 82.846-hour fermentation. Thus, residence time not simply feed throughput governed the principal volumetric constraint.

The calculated 1,100 m³ hydrolysis capacity should be interpreted as the total process requirement. A single vessel of this size would present substantial mixing, structural and operational challenges for a lignocellulosic slurry. Parallel reactors, staged hydrolysis, fed-batch solids addition or process intensification may provide more practical configurations. Selection among these alternatives requires rheological data and mixing-power calculations that were outside the present process-capacity assessment.

This finding is consistent with the broader lignocellulosic literature. Modenbach and Nokes (2013) identified water limitation, mixing difficulty, inadequate heat and mass transfer, and inhibitor accumulation as important high-solids hydrolysis constraints. Shiva et al. (2022) similarly showed that transport phenomena and rheology become increasingly important as solids loading rises. Malacara-Becerra et al. (2022) identified pretreatment and saccharification as major barriers to commercial-scale lactic acid production from corn residues.

Fermentation and filtration scale

The 25 m³ nominal fermenter was of the same order of magnitude as the 30 m³ semi-industrial fermenter reported by Fu et al. (2017). The comparison supports the physical plausibility of the nominal capacity but does not validate the present design directly. Fu et al. employed an engineered *Escherichia coli* strain in fed-batch glucose fermentation, whereas the present process used a continuous model basis, corn-cob-derived hydrolysate and *Bacillus coagulans*. Differences in organism, broth rheology, feeding strategy, gas evolution and heat release can materially alter agitation, cooling and control requirements.

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The 170 m² membrane estimate was based on a constant design flux and a fixed allowance. Pilot-scale lactic acid production coupled to ceramic microfiltration has been demonstrated in a 3,000 L system (Lu et al., 2012), confirming the relevance of membrane-assisted processing. Nevertheless, full-scale membrane design requires transmembrane pressure, module configuration, concentration polarisation, cleaning frequency, flux decline and membrane replacement to be evaluated over time.

Thermal bottleneck and integration opportunity

The positive heating duties were 742 kW for pretreatment, 318 kW for hydrolysis, 226 kW for fermentation, 228 kW for acidification and 1,164 kW for the purification reboiler, giving a total heating demand of 2,678 kW. The condenser required a separate cooling duty of 932 kW. At

1,984 kg h⁻¹ lactic acid production, the heating duties corresponded to approximately 4.86 MJ kg⁻¹ lactic acid.

Purification accounted for 43.5% of the total heating demand, followed by pretreatment at 27.7%. Purification was therefore the main thermal bottleneck, while hydrolysis was the main volumetric bottleneck. This distinction is important for process development: hydrolysis improvement should target residence time, enzyme accessibility and slurry handling, whereas purification improvement should target heat recovery, separation intensity and reduction of reboiler duty. Conceptual lactic acid separation studies have likewise shown that column configuration and operating variables strongly affect recovery and energy use (Daengpradab and Rattanaphanee, 2018).

Table 3 Thermal-Duty Distribution and Engineering Implication

Process unit	Duty	Share of heating demand	Primary design implication
Pretreatment	742 kW heating	27.70%	Steam supply, corrosion-resistant heat-transfer surfaces
Hydrolysis	318 kW heating	11.90%	Low-temperature heat supply and slurry mixing
Fermentation	226 kW heating	8.40%	Temperature control and biological heat effects
Acidification	228 kW heating	8.50%	Reaction heat and solids handling
Purification reboiler	1,164 kW heating	43.50%	Main target for heat integration
Purification condenser	932 kW cooling	Reported separately	Cooling-water and condenser-area requirement

Industrial design implications and limitations

The integrated calculations demonstrate the value of coupling equipment capacity to one reconciled process basis. Mass-flow rate, density, residence time and utility demand were not treated as unrelated spreadsheet inputs; each equipment requirement originated from the corresponding simulated process stream and operating condition. This improves internal consistency and allows future throughput or operating-condition changes to be propagated through material, energy and capacity calculations.

The estimates remain preliminary. Vessel shell thickness, material of construction, corrosion allowance, agitator type, shaft power, impeller diameter, heat-transfer area, pressure relief, nozzle arrangement, piping, control systems, access and maintainability were not designed. The assumed 1 h acidification residence time requires experimental verification. The purification column height was estimated from stage count and tray spacing, but the diameter cannot be specified until vapour and liquid traffic, flooding fraction, entrainment, weeping and pressure drop are calculated. Likewise, the

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constant membrane flux should be verified using long-duration fouling tests.

For pilot development, the hydrolysis system should receive first priority. Required measurements include slurry viscosity over conversion, yield stress, mixing time, torque, enzyme dispersion, particle settling, and heat-transfer coefficient. These data would support selection of an appropriate scale-up criterion, such as constant power per unit volume, constant impeller tip speed or constant mixing time. Pilot verification should then be used to revise the total capacity, determine the number of parallel reactors and quantify uncertainty before vendor quotation and detailed mechanical design.

CONCLUSIONS

A simulation-coupled methodology was used to translate the reconciled flows and residence times of an integrated corncob-to-lactic-acid process into preliminary equipment-capacity requirements. At a reference corncob throughput of $2,500 \text{ kg h}^{-1}$, the selected nominal requirements were a 12 m^3 pretreatment reactor, $1,100 \text{ m}^3$ total hydrolysis capacity, a 25 m^3 fermenter, 170 m^2 membrane area, a 20 m^3 acidification reactor, a 14 m purification column with 18 stages and a 50 m^3 product-storage tank.

The 61-h enzymatic-hydrolysis residence time made hydrolysis the dominant volumetric constraint, accounting for approximately 95.6% of the combined working volume of the principal reaction vessels. The 1,164-kW purification reboiler represented 43.5% of the 2,678-kW heating demand, making purification the main thermal constraint. These results separate the process into two priority areas: hydrolysis intensification and mixing for capacity reduction, and purification heat integration for energy reduction.

The reported values define the approximate scale and bottlenecks of the process; they are not final mechanical equipment specifications. Pilot-scale rheology, mixing, heat- and mass-transfer, membrane-fouling and column-hydraulic measurements, followed by vendor-based design, are required before commercial implementation.

The preliminary capacity assessment provides a reproducible engineering basis for evaluating local corncob conversion and planning subsequent pilot studies in Nigeria. Practical implementation will depend on feedstock logistics, experimental scale-up data, detailed mechanical design and verified equipment costs.

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