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RESEARCH ARTICLE

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Optimization of a reactor for biostimulant production with emphasis on the economic aspects of hydrolysis

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Abstract

Biostimulants are increasingly used in agriculture to promote plant growth, improve stress tolerance, and support sustainable farming practices. One common method of production is chemical hydrolysis of protein-rich waste, such as tannery by-products, offering an economical and eco-friendly alternative to conventional raw materials. This approach aligns with the principles of the circular economy by transforming industrial waste into valuable agricultural inputs. This study investigates the first stage of biostimulant production, alkaline hydrolysis at elevated temperatures, by analyzing the interplay between reaction kinetics, conversion efficiency, and operational costs. Through experimental data and kinetic modelling using a first-order reaction and the Arrhenius equation, key parameters such as activation energy and frequency factor were established. A techno-economic evaluation was conducted to determine optimal operating conditions, emphasizing the importance of the cost ratio between the reactor and evaporator. The results show that higher conversion rates and shorter reaction times are achievable with optimized temperature and cost parameters. Pilot-scale validation (100 L) confirmed process scalability, and preliminary field trials on maize demonstrated effects comparable to those of commercial products. These insights contribute to the advancement of scalable and cost-effective hydrolysis methods for producing biostimulants from waste protein sources.

KEYWORDS

biostimulants, circular economy, hydrolysis, process optimization, protein

1 | INTRODUCTION

Biostimulants, also known as resistance inducers, represent a modern method of non-chemical protection for cultivated plants. These substances enhance the natural defence mechanisms of plants, increasing their resistance to diseases and pests.[1–5] Biostimulants are classified into several main groups, including humic substances, plant and seaweed extracts, and products containing amino acids such as protein hydrolyzates.[6,7]

Currently, most biostimulants based on protein hydrolyzates are produced from animal proteins obtained through acidic hydrolysis, while a smaller proportion consists of hydrolyzates from plant proteins, which are produced through enzymatic processes.[8] One significant category is protein hydrolyzates derived from by-products of the food industry.[9] Importantly, waste materials from the tannery industry constitute a valuable yet underexploited source of proteins, which can be effectively used in plant cultivation to enhance growth and resilience.[10–12] The processing of these by-products into protein hydrolyzates not only supports waste valorization but also offers a sustainable and effective alternative in biostimulant production. Collagen-based hydrolyzates have thus become an essential component in this group of biostimulants due to their natural origin and ability to support the plant's natural defence mechanisms.[13]

These protein hydrolyzates have attracted growing interest due to their beneficial effects on crop performance.[14] Rich in glycine, alanine, and proline, they stimulate plant growth by enhancing chlorophyll synthesis, chelating micronutrients, and improving plant resilience to abiotic stress.[15] Therefore, this ecological alternative to chemical agents contributes to plant protection and the sustainability of agricultural production.[7,16] Given the relatively high cost of commercially available biostimulants, it is advisable to optimize their production process, where collagen waste hydrolysis can be performed in either an alkaline or acidic environment.[10,17] However, due to the hydrolysis reactor walls' corrosion risk, the process is predominantly carried out under alkaline conditions.

This study investigates the valorization of collagen-rich wastes originating from the tannery and food industries through alkaline hydrolysis using potassium hydroxide (KOH). The process aims to produce protein-based hydrolyzates enriched with essential nutrients, particularly nitrogen and potassium, that can serve as sustainable raw materials for agricultural biostimulants or fertilizers. Phosphorus content is adjusted using phosphoric acid or calcium dihydrogen phosphate. Unlike previous studies that focused either on maximizing protein conversion without economic considerations or on basic techno-economic analysis,[17,18] this work integrates chemical kinetics with techno-economic modelling to optimize both technical performance and operational costs.

By combining waste valorization, nutrient recovery, and cost-optimized processing within a clear mathematical framework, this study addresses an important research gap and provides a practical approach for producing sustainable agricultural inputs from industrial waste. The process is designed to minimize operating costs while meeting key chemical and technical requirements, making it environmentally friendly and commercially viable. It also offers three major sustainability benefits: (i) waste diversion—converting disposal-bound material into value-added products, (ii) resource recovery—capturing protein for biostimulant activity and essential nutrients (N, K, P), and (iii) agricultural sustainability—reducing reliance on synthetic fertilizers.

2 | THEORY

2.1 | Determination of the objective function

The production process of collagen hydrolyzate occurs in two main steps. The first step is hydrolysis, which occurs in a reactor at elevated temperatures with intensive stirring of the reaction mixture. The second step is concentration, where the diluted aqueous solution (protein hydrolyzate) is adjusted to the desired commercial concentration. Operating costs include energy consumption for driving the stirrer, compensation for heat losses during the reaction, costs for concentrating the hydrolyzate, overhead and labour costs, and depreciation related to the technological equipment. This process enables the efficient production of collagen hydrolyzate that meets both quality and economic requirements, making it a sustainable product with added value.

These costs (Equation (1)) are given by the product of the stirrer electric motor's input power, specific electricity costs and operating time, plus heat losses associated with the specific cost of the heating medium, mainly water steam.

$$N_{\mathrm{RE}} = \left(M_{\mathrm{P}} \cdot K_{\mathrm{E}} + \varkappa \Delta t \mathrm{S} K_{\mathrm{P}}\right) \cdot \tau \tag{1}$$

Hydrolysis costs increase over time. The reactor produces a diluted solution of the hydrolyzate, and due to its application, it is necessary to include evaporation, which concentrates the diluted solution of the hydrolyzate to the desired concentration. The costs for achieving the desired concentration are determined by the product of the mass of the required evaporated water, the heat of vaporization, and the price of the heating steam.

$$\mathrm{No} = \mathrm{W} \Delta H_{\mathrm{V}} K_{\mathrm{P}} \tag{2}$$

Total balance:

$$h = P + W \tag{3}$$

Component balance:

$$h a_{H,h} = P a_{H,P} \tag{4}$$

Using the given Equations (3) and (4), W is calculated:

$$W = h\left(1 - \frac{a_{H,h}}{a_{H,P}}\right) \tag{5}$$

2.2 | Kinetics of hydrolysis

In this study, a first-order kinetic mechanism with respect to hydrolyzate concentration was adopted as a practical simplification of a complex heterogeneous chemical reaction process. During the alkaline hydrolysis of collagen-rich waste, several simultaneous phenomena occur, including peptide bond cleavage, diffusion of alkali into the solid matrix, and mass transfer of soluble products into the liquid phase. For industrial-scale applications, it is therefore essential to verify whether the apparent reaction rate is affected by transport limitations—particularly by the diffusion of alkali within the solid phase, which depends mainly on the particle size and thickness of the hydrolyzed material. If diffusion resistance were significant, additional disintegration or finer milling of the input proteinaceous waste would be required. Simplified kinetic models such as the first-order approach are widely accepted for process optimization and reactor design, provided their validity is theoretically or experimentally supported within the operating range.

In this study, the system is recognized as a heterogeneous reaction potentially influenced by diffusion limitations of the reacting species—specifically, the diffusion of KOH into the interior of the solid phase. The possible retarding effect of diffusion can be verified experimentally through the dependence of the reaction rate on the particle size or thickness of the hydrolyzed solid. Here, a theoretical approach based on the Fourier diffusion criterion (F_o) was adopted to estimate the maximum particle thickness at which diffusion resistance may be neglected:

$$F_O = \frac{D\tau}{b^2} \tag{6}$$

where D is the diffusion coefficient, τ is the diffusion time, and b is the half-thickness of the solid particle. When $F_o = 1$, diffusion effects become negligible and the process can be considered kinetically controlled. The diffusion coefficient of KOH was estimated using the Nernst equation[19] and reference data from Perry and Chilton.[20]

For the experiments, a laboratory cutting mill (Fritsch, Pulverisette 19, Germany) was used to achieve an average particle thickness of approximately $b = 0.1$ mm through mechanical milling of the solid feedstock. Based on this dimension and the calculated diffusion coefficient $D_{KOH} = 7.1 \times 10^{-9}\,m^2\ s^{-1}$, the characteristic diffusion time was estimated to be 1.4 s. Given that the total hydrolysis time exceeds 1 h, diffusion resistance can be safely neglected. Moreover, the actual particle thickness after milling is likely smaller than the assumed value, further confirming that the process is kinetically rather than diffusion controlled.

The assumption of first-order reaction kinetics concerning the volumetric concentration of the hydrolyzate.

$$B + A = H \tag{7}$$

$$\frac{dc_{H,h}}{d\tau} = k(c_E - c_{H,h}) \tag{8}$$

By integrating Equation (8), the following expression is obtained:

$$\int_0^{c_{H,h}} \frac{dc_{H,h}}{c_E - c_{H,h}} = k\tau \rightarrow -\ln\frac{c_E - c_{H,h}}{c_E} = k\tau \tag{9}$$

By introducing the degree of conversion,

$$x = \frac{c_{H,h}}{c_E}, \tag{10}$$

Equation (11) is obtained:

$$-\ln(1-x) = k\cdot\tau \tag{11}$$

For the concentration of alkali (KOH), the following applies:

$$-\frac{dc_A}{d\tau} = k\cdot c_A \tag{12}$$

Integration of Equation (12) results in:

$$-\ln\frac{c_A}{c_{AP}} = k\cdot\tau \tag{13}$$

From Equation (10), it follows:

$$hc_{H,h} = x\cdot c_E \tag{14}$$

In most cases, the densities of the waste protein and the hydrolyzate are approximately the same, and therefore, the following applies:

$$c_{H,h} = a_{H,h}\cdot\rho; c_E = a_E\cdot\rho \tag{15}$$

From this:

$$x = \frac{a_{H,h}}{a_E}; a_{H,h} = xa_E \tag{16}$$

Then, for the desired evaporated water volume W (Equation (3)), the following will apply:

$$W = h\left(1 - \frac{x a_{\mathrm{E}}}{a_{\mathrm{H,P}}}\right) \tag{17}$$

As solved, the objective function (N) is the sum of the costs for electricity consumption, heat losses from the hydrolysis reactor, and heat from the evaporator.

$$N = N_{RE} + N_0 \tag{18}$$

$$N = (M_P K_{\mathrm{E}} + \varkappa \Delta t S K_{\mathrm{P}})\tau + W \Delta H_{\mathrm{v}} K_{\mathrm{P}} \tag{19}$$

By substituting Equations (5), (12)–(14) into Equation (19), one obtains:

$$N = -\frac{(M_P \cdot K_{\mathrm{E}} + \varkappa \mathrm{S} \Delta t K_{\mathrm{P}})\ln(1-x)}{k} + h\left(1 - \frac{\mathrm{x} c_{\mathrm{E}}}{c_{\mathrm{H,P}}}\right)\Delta H_{\mathrm{V}} K_{\mathrm{p}} \tag{20}$$

Equation (20) can be rewritten as:

$$N = \ -\mathrm{A}\ln(1-\mathrm{x}) + \mathrm{B}(1-\mathrm{Dx}) \tag{21}$$

where

$$A = \frac{(M_P \cdot K_{\mathrm{E}} + \varkappa \mathrm{S} \Delta t K_{\mathrm{P}})}{k}; \tag{22a}$$

$$B = \Delta \mathrm{H}_V \mathrm{K}_{\mathrm{p}} h; \tag{22b}$$

$$D = \frac{c_E}{c_{H,P}} \tag{22c}$$

The recommended ratio of the reaction mixture's mass to the protein waste mass (N_{A}) is 5.[21] The average initial waste protein concentration (a_{BI}) is 80% based on the organic nitrogen content.

It holds that:

$$0.8 \cdot 1 = a_{BI} \cdot 5; a_{BI} = \frac{0.8}{5} = 0.16$$

Then, for the desired concentration of 30% protein in the final product, it can be expressed as:

$$a_{H,P} = 0.3; \frac{a_{BI}}{a_{H,P}} = \frac{0.16}{0.3} = 0.53 \sim 0.5 = D$$

$$N = -A ln(1-x) + B(1-0.5x) \tag{23}$$

The dependence of the objective function (N, Equation (23)) on the degree of conversion x is shown in Figure 1. It is related to the hydrolysis costs (A) and the concentration costs (B). The example refers to the specific case where $A = 1$€ and $B = 10$€ and demonstrates how the objective functions behave under these cost conditions.

For economic optimum, the following applies:

$$N^{'} = \frac{A}{1-x} - DB = 0 = \frac{dN}{dx} \tag{24}$$

Then,

$$x_{\mathrm{opt}} = 1 - \frac{A}{BD} = 1 - \frac{A}{0.5B} = 1 - \frac{1}{0.5 \cdot 10} = 0.8 \tag{25}$$

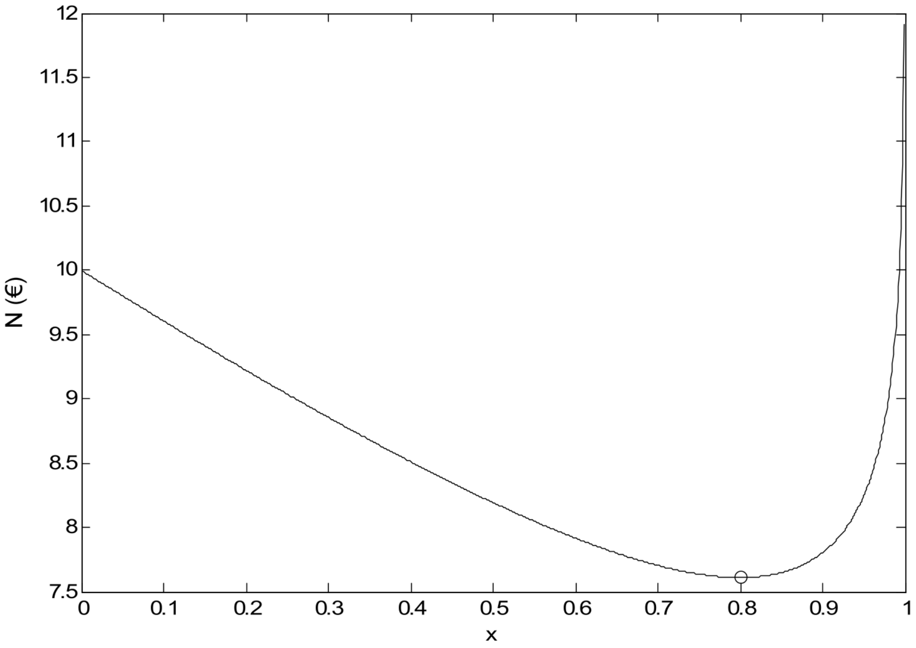

FIGURE 1 Dependence of the objective function on the degree of conversion.

The dependence of the objective function on different values of constant *B* is illustrated in Figure 2A. In contrast, the dependence of the optimal degree of conversion on various values of the constants *A* and *B* is shown in Figure 2B.

3 | EXPERIMENTAL

Collagen protein-rich tannery waste (non-chromium-tanned) was obtained from Devro s.r.o. (Czech Republic). Prior to the hydrolysis experiments, the material was characterized for its chemical composition, which was determined to be as follows (Table 1):

3.1 | Hydrolysis process

All laboratory experiments were conducted in triplicate at reaction temperatures ranging from 50 to 90°C. For each run, tannery waste (2.30 ± 0.02 g, weighed on an analytical balance with a precision of ±0.01 g) was

TABLE 1 Chemical characterization of tannery waste.

Parameter	Value ± SD
Dry matter (%)	79.7 ± 3.2
Ash content (%)[a]	1.1 ± 0.0
Nitrogen content (%)[a]	13.4 ± 0.2

[a] Results expressed in dry matter of the sample.

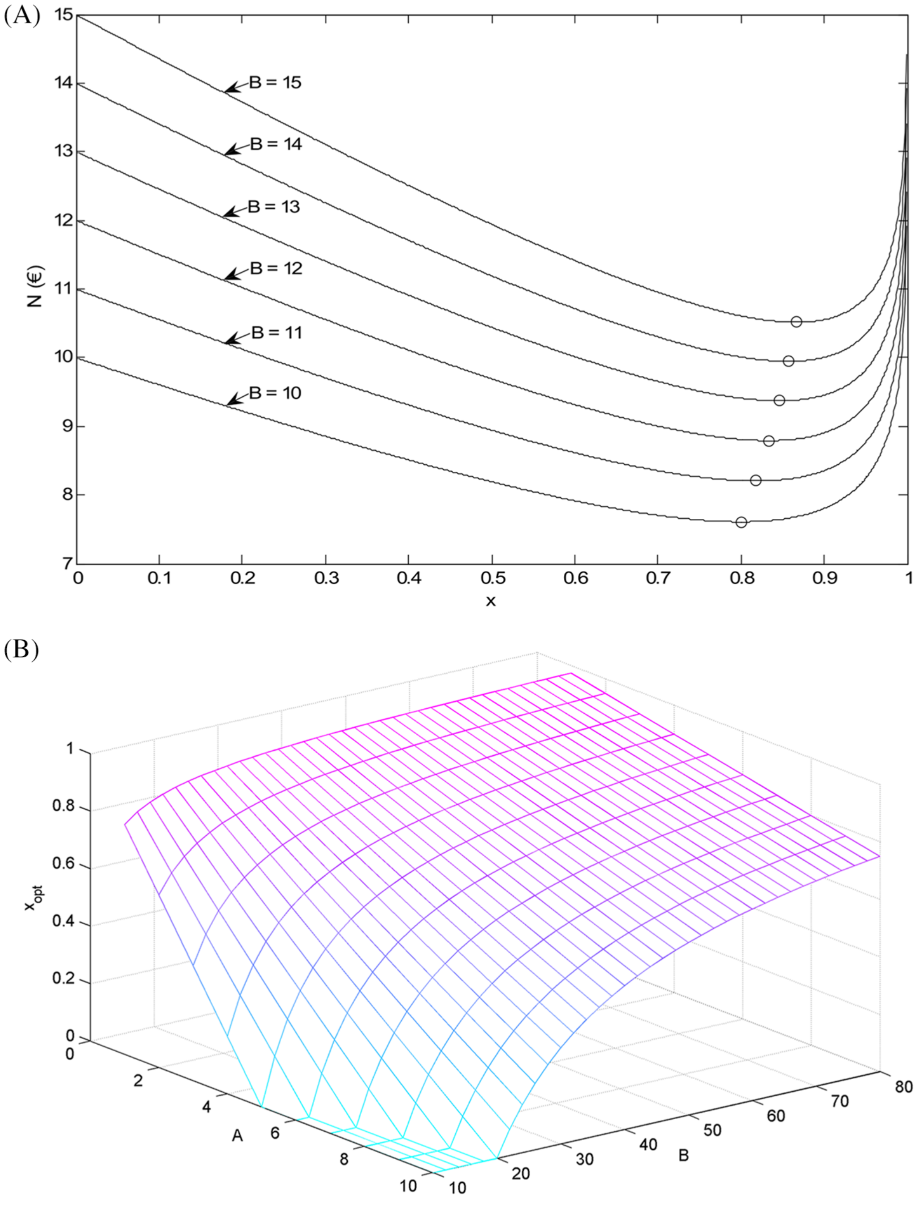

FIGURE 2 (A) Dependence of total costs on conversion *x* for different values of constant B (evaporator) at a fixed value of $A = 1$ (reactor); (B) The dependence of the optimal degree of conversion on different values of constants A and B.

dispersed in deionized water (100.0 ± 0.5 mL, volumetric flask) inside a 250 mL jacketed glass reactor equipped with a reflux condenser. Prior to processing, the waste samples were milled using a cutting mill (Fritsch Pulverisette 19, Germany) to ensure a uniform particle size. The reactor temperature was maintained with a heated magnetic stirrer (Heidolph MR Hei-Standard, Germany) controlled by a Hei-Con contact thermometer with an accuracy of ±1.0°C. After thermal equilibration for 15 min, 9.0 ± 0.1 mL of 10% KOH solution was added while stirring at 750 rpm throughout the entire reaction. The hydrolysis time varied depending on the pH change; once the pH value stabilized, the reaction was terminated. The minimum reaction time was 120 min. Throughout the hydrolysis, the pH of the reaction mixture was continuously monitored, and the recorded values were subsequently used to determine the reaction rate constants. To determine the errors, more precisely the mass deviations, we evaluated the overall mass balance.

For the pilot-scale experiment, 2.5 kg of milled tannery waste was mixed with 98 kg of 1% potassium hydroxide solution in a 100 L reactor equipped with direct steam heating. The reaction mixture was heated to boiling and maintained under these conditions for 1 h to ensure complete hydrolysis. The resulting hydrolyzate was then neutralized with phosphoric acid to pH 7.0. For the initial preliminary tests, the hydrolyzate intended for foliar application on maize was used without concentration by vacuum evaporation; instead, the freshly prepared hydrolyzate was diluted with water at a 1:5 ratio (v/v) and applied directly to the maize plants.

3.2 | Analytical analysis

The analytical methods for characterizing tannery waste and hydrolyzates included: determination of dry matter content gravimetrically by drying to constant weight at 105°C using a PrepAsh 340 Series automatic analyzer (Precisa Gravimetrics AG, Switzerland); measurement of the organic fraction (ash content determination) with the same instrument; analysis of nitrogen content using a Flash 2000 automatic analyzer (Thermo Scientific, USA) via the Dumas method; and determination of elemental content by X-ray fluorescence (XRF) spectroscopy using a SPECTRO iQ II spectrometer (SPECTRO CS, Czech Republic), evaluated by the fundamental parameters method. The pH was monitored continuously using a combination glass electrode (CHS FermPro, model CHS-5594-A0S7, Chromservis, Czech Republic) connected to a digital pH meter (Hanna Instruments pH 213, Microprocessor, Czech Republic) with a resolution of 0.01 pH units and an accuracy of ±0.01 pH. The electrode was calibrated using a two-point calibration (pH 7.01 and 10.01 buffer solutions) immediately before each experiment.

4 | RESULTS AND DISCUSSION

A series of experiments was carried out, and the results confirmed satisfactory reproducibility, with a measurement deviation of less than 10%. This deviation is common when using this type of feedstock—partly due to experimental inaccuracies caused by handling a heterogeneous reaction mixture, but primarily because the waste feedstock is naturally inconsistent, showing differences in particle thickness, grain size, and especially in its overall composition. The experiments aimed to determine the rate constant (k) using Equation (13). By plotting $\ln\frac{c_A}{c_{AP}}$ against time, a straight line is obtained, from the slope of which the numerical value of the rate constant can be determined (see Figure 3). Table 2 presents the measured and evaluated data at 50°C, and Figure 4 shows the progression pH at 50°C.

The remaining values for temperatures 60–90°C were obtained similarly and subsequently evaluated.

To determine the dependence of the hydrolysis rate, which is given by the value of the reaction rate constant, the Arrhenius Equation (26) was used:

$$k = A_F \cdot \mathrm{Exp}\left(-\frac{\Delta E}{RT}\right) \tag{26}$$

Linearization of Equation (26) yields:

$$\ln(k) = \ln(A_F) - \frac{\Delta E}{\mathrm{RT}} \tag{27}$$

A straight line is obtained by plotting the natural logarithm of the numerical value of k against the reciprocal of the absolute temperature, from whose parameters the frequency factor A_F and the activation energy E can be determined, as shown in Table 3. In the given equation, R is the universal gas constant $R = 8314\ \mathrm{J\ kmol^{-1} \cdot K^{-1}}$.

The frequency factor A_F from the previous graph is equal to 113.6 $\mathrm{min^{-1}}$, and the activation energy E is $2.38 \times 10^4\ \mathrm{J\ mol^{-1}}$.

To determine the minimum of the objective function, the numerical value of the rate (kinetic) constant (k) in the reactor constant (A) had to be known. Based on a techno-economic requirement, the optimal numerical value x_{opt} was estimated, and the rate constant was determined using Equations (22a), (22b), and (25).

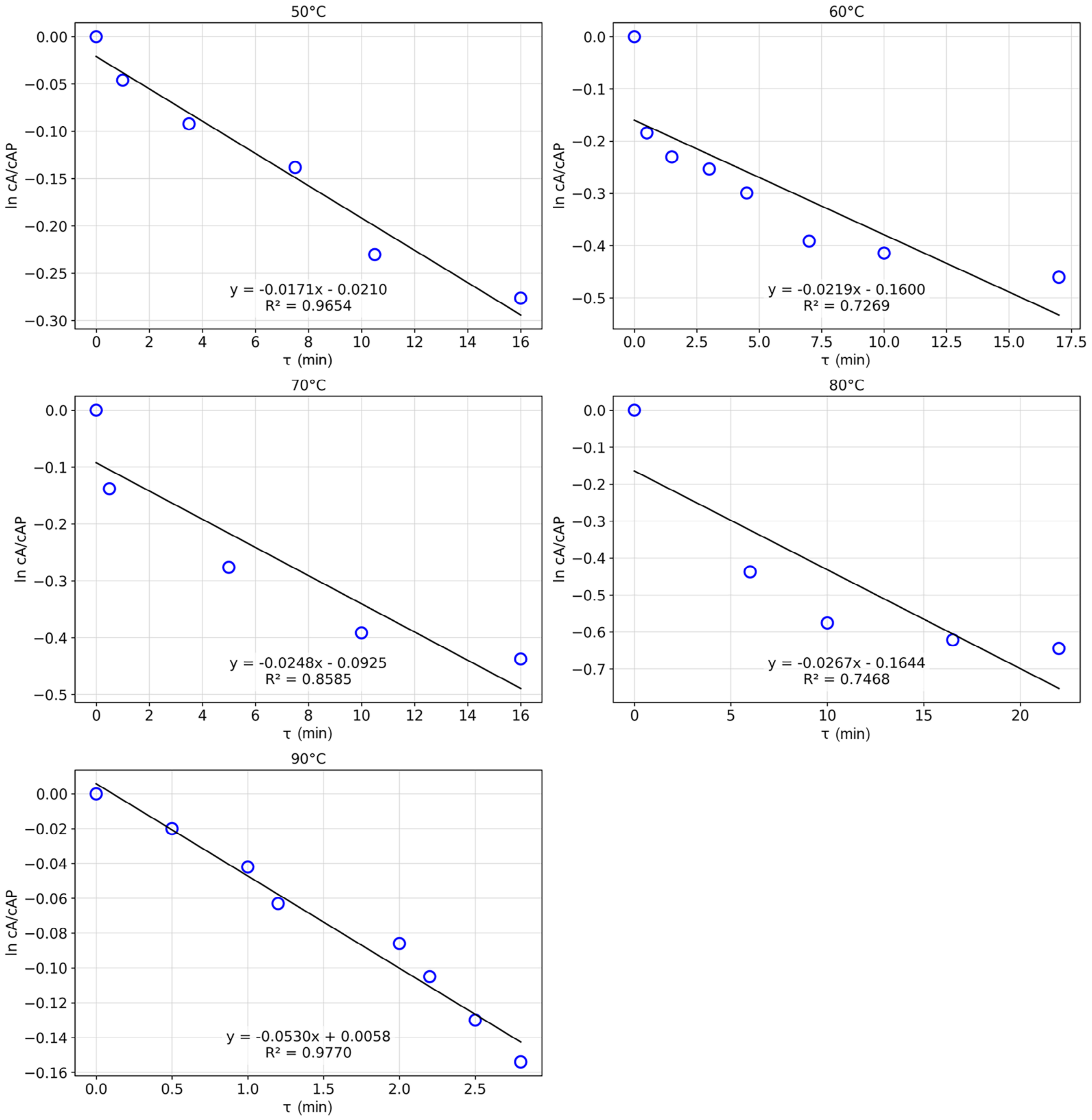

FIGURE 3 Dependencies of $\ln c_A/c_{AP}$ on τ at different temperatures.

$$k = \frac{M_P K_E + \kappa \Delta t S K_p}{\left(1 - x_{opt}\right) \cdot 0.5 - \Delta H_V \cdot K_p \cdot P} \tag{28}$$

The temperature of the hydrolysis reaction is determined from Equation (27). For demonstration, the following example is provided.

With the desired hydrolyzate production $p = 3000$ kg with mass fraction $a_{HP} = 0.3$, latent heat of water vaporization $\Delta H_w = 2.3$ MJ kg^{-1}, specific price of heating steam $K_P = 4 \times 10^{-8}$ € J^{-1}, power input of the hydrolysis reactor agitator $M_P = 10$ kW, specific price of electrical energy $K_E = 0.4$ € kWh^{-1}, heat transfer coefficient through the reactor wall $\kappa = 10$ W m^{-2} K^{-1}, heat exchange area of the reactor $S = 2$ m^2, temperature difference from reaction temperature to ambient temperature $\Delta t = 80$°C, the rate constant of hydrolysis was estimated from the required conversion of hydrolysis.

Example:

Degree of hydrolysis conversion $x_{opt} = 0.95$ during 1 h = 60 min.

TABLE 2 Measured and evaluated data at 50°C.

pH	pOH	τ (min)	c_A (mol L^{-1})	c_A/c_{AP} $c_{AP} = 0.0339$ (mol L^{-1})	Ln c_A/c_{AP}
12.53	1.47	0	0.0339	1	0
12.51	1.49	1	0.0324	0.955	−0.046
12.49	1.51	3.5	0.0309	0.912	−0.092
12.47	1.53	7.5	0.0295	0.871	−0.138
12.43	1.57	10.5	0.0269	0.794	−0.230
12.41	1.59	16	0.0257	0.759	−0.276
12.36	1.64	31	0.0229	0.676	−0.391
12.31	1.69	50	0.0204	0.603	−0.507
12.26	1.74	75	0.0182	0.537	−0.622
12.25	1.75	115	0.0178	0.525	−0.645
12.27	1.73	150	0.0186	0.550	−0.599
12.26	1.74	180	0.0182	0.537	−0.622
12.27	1.73	210	0.0186	0.550	−0.599

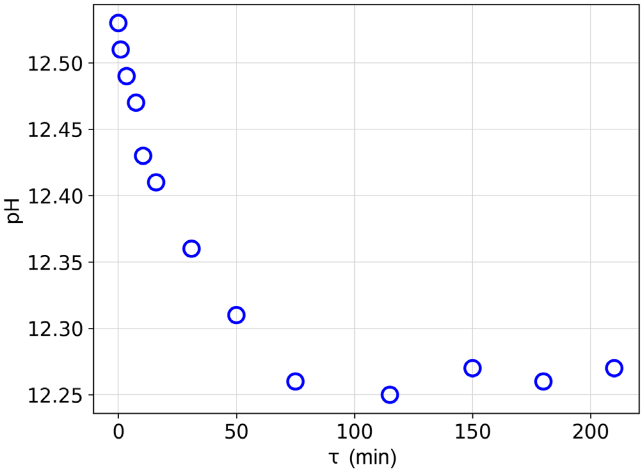

FIGURE 4 pH changes of the reaction mixture over time at 50°C.

TABLE 3 Temperature dependence of the rate constant.

t (°C)	k (min^{-1})	T (K)	$1/T$ (K^{-1})	Ln k
50	0.017	323	0.0031	−4.075
60	0.022	333	0.0030	−3.817
70	0.024	343	0.0029	−3.689
80	0.027	353	0.0028	−3.623
90	0.053	363	0.0028	−2.937

The constant is then calculated according to Equation (28).

$$k = 8.5 \cdot 10^{-4} s^{-1} = 0.051 \text{ min}^{-1}$$

The hydrolysis temperature is determined using Equation (27) and the values given in Figure 5.

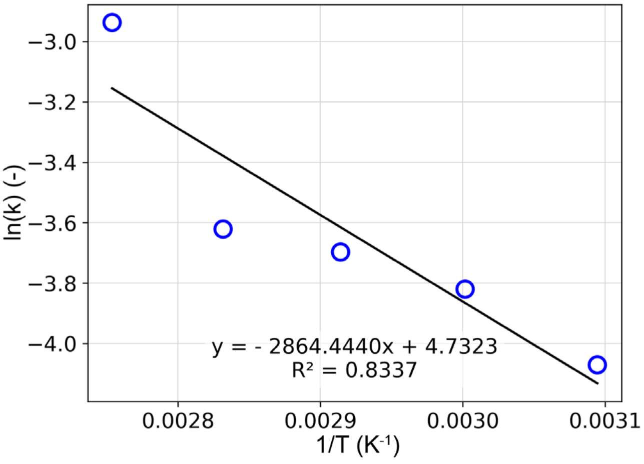

FIGURE 5 The obtained dependence of ln k on $1/T$.

$$T = -\frac{\Delta E}{R \cdot (\ln k - lnA_F)} = -\frac{2.38 \cdot 10^4}{8.314\,(-2.98 - 4.73)}$$
$$= 371\,K \sim 98^{\circ}\text{C}$$

Hydrolysis time (see Equation (11)):

$$\tau = -\frac{\ln 0.05}{0.051} \cong 59 \text{ min}$$

The desired optimal reaction temperature can be maintained using a commercially available controller. Currently, there is a wide range of controllers available on the market, differing in the number of regulated channels, supported types of input analogue signals, depending on the used sensor, output analogue or binary control signals for operating the actuator, and the configurability options of the controller. There is also the common option to select additional functions, such as alarms, automatic parameter tuning of the controller, and communication with a supervisory system via a selected standard industrial communication interface.

For the purpose of temperature control in a chemical reactor, a single-channel PID controller is fully sufficient, allowing for the connection of a temperature sensor with a measurement range of 0 –100°C and a current output of 4–20 mA. A relay binary output is required to control the actuator, enabling operation with 230 V alternating current. These required parameters are fully met by the DDR224 controller from Pixys Electronics, which is additionally equipped with universal software-configurable analog inputs that allow connection of thermocouples, PT, NI, PTC, NTC resistance thermometers, and standard analog voltage signals (0–10 V) and current signals (4–20 mA). The controller's parameter settings can be either automatic or manual. The controller is designed for installation on a DIN rail in an electrical distribution panel.[22]

The ratio of costs (excluding the unit prices of electricity and heat) for electricity and heat is determined by the required concentration of dry matter in the hydrolyzate. A higher concentration at the end of hydrolysis is associated with a lower consumption of heat for evaporation. Although the unit price of electricity is higher than that of heat, the share of evaporation costs is, in most cases involving a high degree of conversion, found to be higher. This is attributed to the considerable heat demand resulting from the high latent heat of evaporation. It is suggested that the use of bleed steam in multi-effect evaporators be considered, as is common in conventional chemical technology.

The rate constant of the hydrolytic reaction is regarded as a highly important parameter. The numerical value of this constant is influenced by the chemical mechanism of the hydrolysis reaction. In this study, a first-order mechanism with respect to alkali concentration, specifically potassium hydroxide, has been assumed. For industrial large-scale applications, the kinetic (maximum) reaction rate must be evaluated to determine whether it is limited by transport processes, particularly by the diffusion of the alkali. This issue is related to the particle size of the hydrolysed solid phase, particularly its thickness. If a significant diffusion-limiting effect is observed, disintegration of the input waste protein should be carried out. Simulation calculations based on a diffusion model (the second Fick's law) coupled with a chemical reaction are recommended. These calculations allow for an estimation of the extent to which such disintegration may influence the objective function.

As previously stated, the objective function consists of the costs associated with the hydrolysis reactor and the evaporator. Quantitatively, both cost components involve constants: A for the hydrolysis reactor and B for the evaporator. Both constants are considered time-independent. In particular, the physical parameters are not time-dependent. For the hydrolysis reactor, these parameters include the power input of the electric motor driving the stirrer for homogenizing the reaction mixture, the heat transfer coefficient, the temperature difference between the hydrolysis mixture and the ambient temperature, the heat transfer surface area, the unit price of electrical energy, and the unit price of heat, which accounts for thermal losses.

The time in Equation (1) is dependent on the rate of hydrolysis and is given by the expression $\ln(1 - x)/k = \tau$. The faster the reaction proceeds, the shorter the reaction time required. Reactor costs can be significantly reduced through the use of an enzymatic proteolytic catalyst; however, its application is limited by the range of reaction conditions, particularly the temperature and pH of the reaction mixture. For the evaporator, the parameters such as the latent heat of evaporation and the unit price of heat required to achieve the target commercial dry matter concentration are considered constant. In determining the optimal conversion, labour costs, overheads, and depreciation of investment, while influencing overall costs, can be considered negligible with respect to the hydrolysis time and thus do not affect the numerical value of the optimal conversion. In practical application, the production of the biostimulant in conjunction with economic viability is strongly dependent on the quantitative ratio between constants A and B. According to Equation (24), the optimal conversion (x_{opt}) exhibits a linear dependence on the ratio A/B. For the hydrolytic reaction to be economically viable, the inequality $A/(B{\cdot}0.5) < 0.5$ must be satisfied. An example is shown in Figure 1, where for the ratio $A/B = 1/10$, and with constants $A = 1$ € and $B = 10$ € (values that approximately correspond to the unit prices of electricity and heat, respectively), the optimal conversion is $x_{opt} = 0.8$. For higher A/B ratios, the optimal conversion decreases linearly, as illustrated in the following Table 4.

The higher the ratio of constants A/B, the lower the hydrolysis rate, that is, a lower value of the rate constant, and consequently, the longer the reaction time, until a point is reached where the operating costs of the reactor equal those of the evaporator, and the conversion becomes zero. To prevent this from occurring, the reaction temperature and, consequently, the pressure must be increased. An alternative approach involves selecting an appropriate catalyst.

The optimized reaction conditions established at laboratory scale were successfully validated in a pilot-scale operation, where 100 litres of proteinaceous hydrolyzate

A/B	0.5	0.45	0.4	0.3	0.15	0.10	0.05	0.025	0.01	0.0
x_{opt}	0	0.1	0.2	0.4	0.7	0.8	0.9	0.95	0.98	1.0

TABLE 4 Dependence of optimal conversion (x_{opt}) on the cost ratio A/B (reactor/evaporator).

were prepared under strictly controlled parameters, demonstrating efficient conversion without significant loss of efficiency. Comprehensive chemical analysis revealed a nitrogen content of 12.7% (w/w) in dry matter, corresponding to approximately 80% crude protein. The remaining fraction comprised inorganic components, particularly potassium and phosphorus compounds—macroelements critical for plant nutrition. The overall nutrient composition was closely aligned with that of conventional NPK fertilizers. However, the organic matrix of the hydrolyzate is expected to enhance nutrient bioavailability and promote a favourable soil microbiome balance. A preliminary field trial on maize using foliar application during early developmental stages showed positive effects on plant growth parameters, with performance comparable to conventional commercial products that are substantially more expensive.

5 | CONCLUSION

This study successfully integrated experimental kinetics and techno-economic modelling to optimize the alkaline hydrolysis of tannery waste under conditions relevant to industrial application. The process followed first-order kinetics with an activation energy of 2.38×10^4 J · mol^{-1} and a frequency factor of 113.6 min^{-1}, enabling accurate prediction of reaction rates across different temperatures. Techno-economic analysis identified optimal operating parameters based on conversion targets and the cost ratio between the hydrolysis reactor and evaporator (A/B), showing that lower A/B ratios favour higher conversions and shorter reaction times, improving cost efficiency. At 98°C, a 95% conversion was achieved with a rate constant of 0.051 min^{-1} and a 59-min reaction time. The optimized parameters were validated in a 100 L pilot-scale system, confirming process scalability and robustness. Preliminary field trials on maize indicated positive effects comparable to commercial foliar fertilizers, though broader multi-season validation is needed. Physical limitations such as inner diffusion were identified as potential rate-limiting factors; mechanical disintegration of input materials is recommended to enhance surface area and minimize diffusion resistance.

Despite these promising outcomes, several limitations remain. Laboratory-scale kinetic data may not fully capture industrial-scale phenomena such as mixing efficiency, heat transfer, and hydrolyzate rheology, and the simplified first-order model does not account for variability in waste composition. The economic assessment relied on estimated cost parameters without regional validation or life-cycle analysis. Future work should therefore focus on pilot-scale verification under realistic conditions, including rheological and energy analyses and comprehensive product characterization—particularly amino acid profiling, molecular weight distribution, and stability testing—to comply with biostimulant registration standards. In the longer term, efforts should aim at process intensification, enzymatic pre-treatment, continuous operation, and the expansion of feedstocks to other protein-rich wastes.

Overall, this work provides a comprehensive framework linking reaction kinetics, process optimization, and agronomic performance, demonstrating the potential of alkaline hydrolysis for valorizing tannery waste within a circular bioeconomy framework.

NOMENCLATURE

a_E	mass fraction of protein in the processed waste
$a_{H,h}$	mass fraction of the hydrolyzate in the dilute aqueous reaction blend (h) entering the evaporator in time
$a_{H,P}$	mass fraction of the hydrolyzate in the product P
A_F	frequency factor (min)
c	specific heat of the reaction mixture (J kg^{-1} K^{-1})
c_A	volumetric concentration of alkali in the reaction mixture (kg m^{-3})
c_{AP}	initial volumetric concentration of alkali in the reaction mixture (kg m^{-3})
c_E	volumetric equilibrium concentration of protein in the processed waste (kg m^{-3})
$c_{H,h}$	volumetric concentration of the hydrolyzate in the reaction mixture (kg · m^{-3})
$c_{H,P}$	volumetric concentration of the hydrolyzate in the product P (kg m^{-3})
E	activation energy (J mol^{-1})
h	mass of the reaction mixture (kg)
H	mass of hydrolyzate produced in time (τ) with concentration $a_{H,h}$ (kg)
k	rate constant of the hydrolysis reaction (s^{-1}, min^{-1})
K_E	specific price of electrical energy (€ kWh^{-1})
K_P	price of heating steam (€ GJ^{-1})
M_P	electric motor power consumption of the stirrer (W)

N	objective function (€)
N_o	cost of the evaporator (€)
N_{RE}	costs of protein waste hydrolysis (€)
P	mass of the desired product (kg)
S	heat transfer area of the reactor wall (m^2)
T	temperature (K)
W	mass of evaporated water (kg)
x	degree of conversion
ΔH_V	heat of vaporization ($J\ kg^{-1}$)
Δt	temperature difference between the temperature of the reaction mixture and the reactor surroundings (°C)

Greek letters

ρ	density ($kg \cdot m^{-3}$)
τ	reaction time (s, min)
τ_R	optimal reaction time (s)
κ	heat transfer coefficient of the reactor wall ($Wm^{-2}\ K^{-1}$)

AUTHOR CONTRIBUTIONS

Karel Kolomazník: Conceptualization; investigation; validation; methodology; supervision; writing – original draft; data curation; software. **Lubomír Šánek:** Writing – original draft; writing – review and editing; visualization; methodology; investigation; formal analysis; data curation. **Petr Dostálek:** Software; writing – original draft; data curation; methodology. **Clara Desroches:** Resources; writing – original draft; visualization; formal analysis.

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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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