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Production of glucose-galactose syrup and milk minerals from Greek yogurt acid whey

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Greek yogurt acid whey (GAW) is a byproduct stream from Greek yogurt production with costly and environmentally unsustainable disposal methods. A process was developed and tested at the benchtop and pilot scale to produce multiple products from GAW. Pilot scale filtration and neutralization operations resulted in the precipitation of calcium phosphate. The large precipitate was harvested *via* size classification using a hydrocyclone and crushed. The fine precipitate was separated from a lactose-rich stream by membrane filtration and converted to a calcium-rich powder, referred to as milk minerals, in a pilot scale spray dryer. A benchtop process sequence, consisting of acid-catalyzed lactose hydrolysis, filtration, and evaporation, converted the concentrated aqueous lactose filtrate to a sweetener syrup consisting primarily of glucose and galactose. A techno-economic analysis, based on the experimental mass balance data, indicated that the process can be highly profitable (IRR 34.0%) at a GAW inlet of 907 metric tons per day and match the food industry average (IRR 6.25%) at a GAW inlet of 192 metric tons per day GAW. A life cycle assessment demonstrated that the proposed GGS production process had significantly fewer negative environmental impacts than either the production of high fructose corn syrup from corn or sucrose from sugarcane.

1. Introduction

The market share of Greek yogurt in the US yogurt market has risen from 1% to over 50% from 2007 to 2020.[1,2] As a result, Greek yogurt acid whey (GAW), the main byproduct stream of Greek yogurt production, has risen to approximately 1.8 million metric tons per year in the US[3] alone. The US dairy industry has struggled to deal with the large quantity of GAW byproduct. Due to GAW's low protein and high mineral contents, processing methods used for cheese whey are not economically feasible.[4] Thus, many US Greek yogurt producers pay for treatment of GAW in municipal wastewater plants.[5,6] However, the high biological and chemical oxygen demand (BOD and COD, respectively) of GAW can overwhelm the capacity of the wastewater treatment facilities.[7] Alternatively, limited amounts of GAW can be used by farmers, as a component of livestock feed or as a component of farmland fertilizer.[8] Unfortunately, the cow's digestive system can react negatively to excessive GAW,[9,10] and high land application rates can reduce farmland productivity over time due to the salt content and acidity of GAW.[11]

New approaches to GAW valorization have been developed in recent years. For example, GAW has been processed in anaerobic digesters to produce methane, which is burned to generate electricity. However, this route is often not commercially viable,[12] relying on government subsidies, and it can often only handle a fraction of total GAW production.[5] Research institutions, such as the Center for Dairy Research at the University of Wisconsin–Madison, are investigating the use of advanced filtration techniques to produce dairy ingredients and powders, but these technologies have not reached industrial scale.[3] Recent research from Cornell University has led to a start-up company called Capro-X which uses a bioreactor system to produce medium-chain carboxylic acids,[8,13] which can be used in a variety of high-value applications,[14,15] but this technology is only in the semi-pilot plant scale. In short, there remains a need to develop a technology with demonstrated potential economic viability which can produce high-value products from GAW at the industrial scale.

A GAW valorization option which has demonstrated potential economic viability is catalytic hydrolysis of the lactose component of GAW to produce a sweetener syrup called glucose-galactose syrup (GGS).[16,17] While lactose hydrolysis can be performed with either enzymatic or acid catalysts,[16–27] acid catalysts have been reported to be more economical than enzymes.[16,18] A techno-economic analysis conducted on this process showed that a 20% IRR was achievable if the GGS was

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sold for a price similar to the sweetener high fructose corn syrup. However, this approach had multiple drawbacks. The high concentration of salts in the GAW required a large amount of ion exchange resin treatment, resulting in high operating costs and mineral waste. The two catalytic methods examined were both flawed, requiring either the direct addition of concentrated sulfuric acid to the filtered GAW or the use of heterogeneous acid catalysts which were found to be economically unfeasible. Furthermore, ultrafiltration retentate was assumed to be a potential feedstock for the production of valuable whey protein concentrate (WPC), an assumption that would likely not hold given the problematically high acid and mineral content of the retentate unless diafiltration was employed.

A recent study from the University of Wisconsin–Madison has shown that milk minerals, a high value calcium supplement, can be produced from acid whey at higher yields than the incumbent production feedstock of sweet whey.[28] The milk minerals were produced by neutralizing nanofiltration concentrate of GAW UF permeate to precipitate calcium phosphate, followed by classification of the precipitate in a hydrocyclone. Large particles harvested by the hydrocyclone rapidly settled by gravity, allowing for decanting and drying to produce high purity milk minerals. The hydrocyclone overflow was treated with microfiltration to separate the precipitate fines from the soluble solids in the liquid. The precipitate was sent to a spray dryer where the final milk minerals product is made. Milk minerals products are currently produced from cheese whey permeate[29] and sold in dozens of countries[30] under names such as Capolac (Arla Foods) and TruCal (Glanbia Nutritionals). These products make up a part of the roughly 139 000 metric ton US nutritional calcium market.[31] It would be ideal to combine the GGS and milk production technologies to valorize GAW.

The objectives of this paper are to demonstrate, at the laboratory benchtop and pilot scale, a process to produce high value GGS and milk minerals products from GAW and to assess the process' economic potential. The drawbacks of the previous studies on acid-catalyzed lactose hydrolysis in GAW are addressed with novel steps in the process model. The milk minerals production process is incorporated with adjustments due to the different feed used. We collect data at both the laboratory and pilot plant scale. These data were used to develop a rigorous economic analysis, and we show how sensitive this analysis is to different process and economic parameters. In addition, we illustrate the feasibility of the proposed process by comparing it with high fructose corn syrup (HFCS) and sucrose from sugarcane (SS) production in terms of environmental impacts.

2. Materials and methods

2.1 Feedstocks

1140 liters of Greek yogurt acid whey (GAW) were obtained by from local Greek yogurt producer. The composition of the GAW feedstock is found in Stream 5 of Table 4. The Greek yogurt was produced with a ceramic ultrafiltration unit, meaning the protein and fat remaining in the UF GAW permeate was negligible.

2.2 Pilot plant filtration and spray drying of GAW

Fig. 1 shows all operations performed at UW-Madison, including the filtration and spray drying process sequence performed on the GAW at the pilot plant owned by the Center for Dairy Research (CDR) at UW-Madison. The GAW was first passed through a nanofiltration unit (NF), and the NF permeate was concentrated using reverse osmosis (RO). The NF retentate was heated to 65 °C and neutralized to pH 7 with potassium hydroxide in a heated tank to precipitate the calcium phosphate. When the pH reached 6, the NF retentate changed from clear and yellow to turbid and off-white as calcium phosphate particles were formed.

The neutralized NF retentate was then processed in a hydrocyclone for size classification of the precipitated calcium phosphate. Crowley *et al.* reported in a similar acid whey system that the stream exiting the bottom of the hydrocyclone, called hydrocyclone bottoms, contained 4.7 times more calcium than the feed. They calculated that the hydrocyclone bottoms contained 6% of the total calcium, but 45% of the calcium was not accounted for and was described as lost. The hydrocyclone bottoms also contained 1.3% of water and every other component.[28] The hydrocyclone bottoms were collected for characterization and decanted to purify the calcium phosphate. Crowley *et al.* also reported that an early batch of hydrocyclone bottoms was purified and spray dried into milk minerals that contained 29% calcium, but there were large abrasive precipitate that were problematic for the dryer feed pump. The stream exiting the top of the hydrocyclone, called the hydrocyclone overflow, was processed with microfiltration (MF) where the precipitate fines were separated from the permeate. The MF lactose-rich permeate was used as the feedstock to produce glucose-galactose syrup at the laboratory scale. The MF retentate underwent diafiltration with water to purify the precipitate. This purified precipitate was subsequently sent to a spray dryer to produce the final milk minerals product.

2.2.1 Nanofiltration. Two 9.65 cm diameter, 96.5 cm long nanofiltration (Synder, Vacaville, CA) membrane elements (NFX-2B-3838, 150–300 Da membrane pores, 0.79 mm diamond spacer, control bypass without tail) were placed in series in stainless steel tubing vessel. UF GAW permeate contained in a holding vessel was pumped through the filtration elements at 30 bar baseline pressure. The permeate flow was 10.7 lpm initially, and the permeate was collected in a separate vessel. The retentate was recirculated into the feed vessel until the permeate flux was nearly negligible. NF permeate was concentrated using two Hydranautics model Dairy reverse osmosis (RO) spiral filtration elements (size 9.65 cm dia. × 96.5 cm long sanitary spiral elements containing 0.76 mm spacers) in series to concentrate it from 1% solids to 8% solids. The RO permeate flow was 5.7 lpm initially, and the RO permeate had negligible total solids.

2.2.2 Neutralization and heating. NF retentate was heated to approximately 60 °C in a 1135 L stainless steel stirred tank with an impeller spinning at 18–20 rpm. 50% KOH was poured in the stirring tank, and the pH was quickly checked. This was

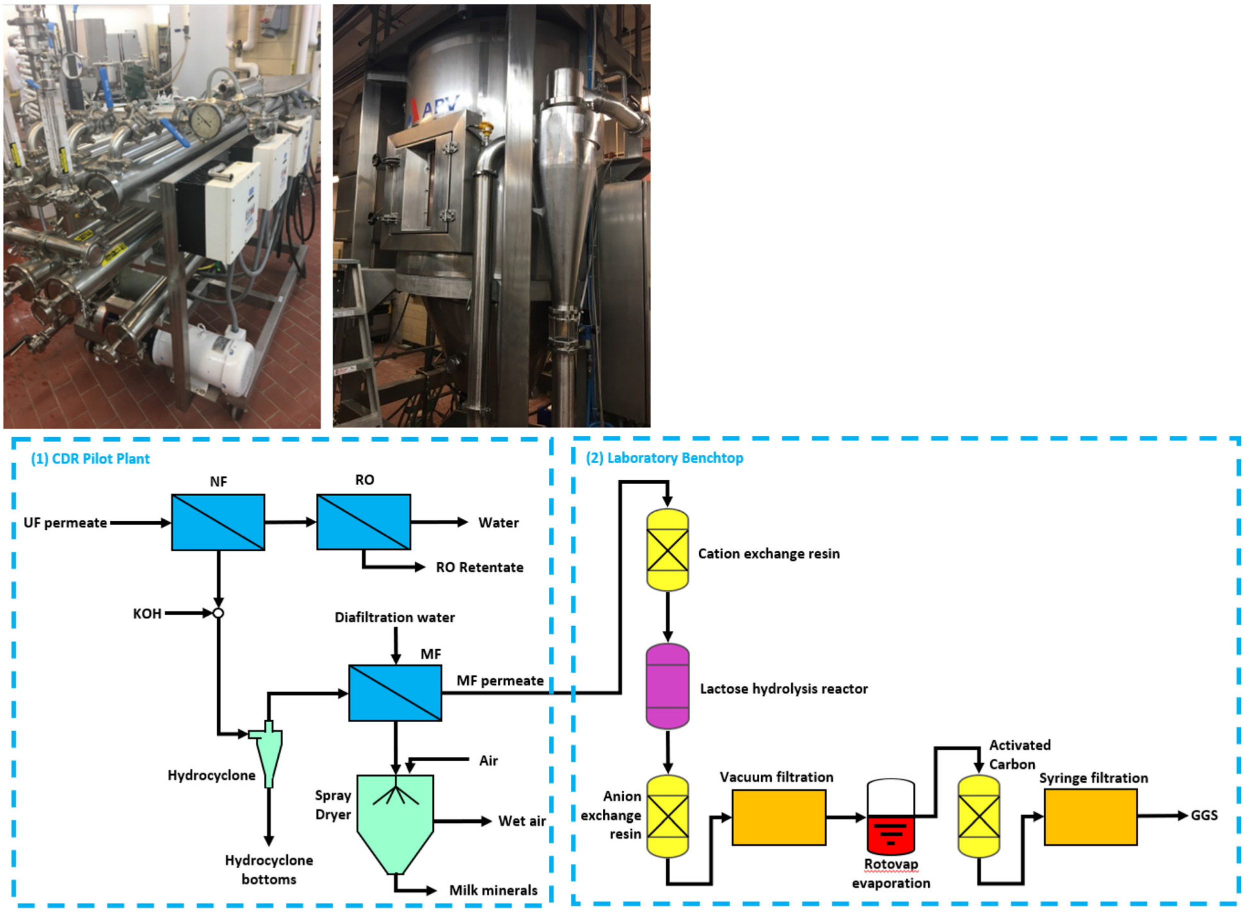

Fig. 1 (Bottom) Process flow diagram for the pilot-scale and laboratory-scale operations performed at UW-Madison to produce GGS and milk minerals from ultrafiltered GAW permeate. Section 1 (bottom left) shows operations performed at the Center for Dairy Research pilot plant at UW-Madison, and Section 2 (bottom right) shows operations performed at the laboratory benchtop-scale in the research lab at UW-Madison. (Top left) Pilot-scale nanofiltration and reverse osmosis system. (Top right) Pilot-scale spray dryer.

repeated until the pH reached 7.18. When the pH rose above 6, the GAW turned from clear to opaque due to the precipitation of calcium phosphate. The neutralized NF retentate was diluted with water to 15% total solids for optimal hydrocyclone performance.[28]

2.2.3 Hydrocyclone. A stainless steel hydrocyclone (VorSpin model AZ-VS-2a, Compatible Components Corporation, Houston, TX) described elsewhere[28] was utilized to classify the precipitate formed in the NF retentate neutralization step. The feed was delivered using 60 Hz to generate 2.39 bar pump with a FZX2100 model Fristam liquid ring. The inlet flowrate was 55.6 lpm and the bottoms flowrate was 2.3 lpm. The hydrocyclone bottoms was captured in a milk can, and the overflow was captured in a collection vessel. 14.7 kg of hydrocyclone bottoms were collected from an initial 355 liters of feed.

2.2.4 Microfiltration. One large-pore ultrafiltration membrane filter (Spira-cel DS UV200 3838G. 200 kDa pores, 9.65 cm diameter, 96.5 cm length, 2.0 mm parallel spacer, polyvinylidenefluoride membrane) from Microdyn-Nadir (Goleta, CA) was used to perform microfiltration on the hydrocyclone overflow. To prevent confusion with the ultrafiltration performed on the GAW by the GAW supplier, the large-pore ultrafiltration will be referred to as microfiltration (MF) in this study. The temperature of the feed was 60.6 °C, and the baseline pressure was maintained at 0.50 bar. The flowrate of permeate exiting the system was 4.81 lpm and was collected in 20 liter buckets. The retentate was collected in a milk can. Diafiltration at 4.35 lpm was performed on the MF retentate that remained in the system after the MF permeate was collected.

2.2.5 Spray dryer. The MF retentate was spray dried using a single-stage, natural gas, direct-fire, pilot-scale spray dryer (PSD 55, APV, Copenhagen, Denmark). The feed was heated to 71 °C prior to entering the atomizing nozzle (Spraying Systems Co., Glendale Heights, IL. SK Spraydry nozzle, 0.34 mm diameter orifice). The inlet air temperature was 188 °C, and the outlet air temperature was in the range of 82–99 °C. The atomization pressure varied widely due to issues the pump had in dealing with the precipitated calcium phosphate particles. When the atomization pressure dropped, striking the pump with a rubber mallet caused the pressure to rise again, possibly because the mallet strikes dislodged some precipitate inside the pump.

2.3 Laboratory-scale processing of MF permeate to produce GGS

The laboratory-scale process to produce GGS from MF permeate is shown in Fig. 1. The MF permeate is treated with a cation exchange resin to purify and acidify the whey. The acidified MF permeate then undergoes homogeneous acid-catalyzed lactose hydrolysis in a heated reactor. The hydrolyzed whey is neutralized and further purified by treatment with an anion exchange resin. Vacuum filtration ensures no resin particles remain in the feed. The neutralized, filtered reaction product is concentrated by evaporation using a laboratory-scale rotovap evaporator. Activated carbon is then used to decrease the coloration, followed by syringe filtration to removes any pieces of activated carbon, resulting in the final GGS product.

2.3.1 Ion exchange. MF permeate produced by the pilot-scale process at the Center for Dairy Research was treated with a cation exchange resin prior to entering the lactose hydrolysis reactor. About 300 mL of Amberlyst 15 cation exchange resin (DOW, Midlands, MI) was placed in a 3.18 cm diameter glass column with a filter. Water in an amount ten times the volume of the resin was flowed through the column to wet the resin, and the resin was never allowed to be dry. MF permeate was allowed to flow through the column at 1 mL min^{-1} (linear velocity 0.13 cm s^{-1}), a rate consistent with literature.[32] The first 60 mL of outflowing liquid were discarded as it was mostly water. After that, the MF permeate exiting the column was collected in 100 mL increments and tested for pH. Roughly 800 mL of whey exited the column at a pH of 1.1, after which the pH began to rise. Once the pH began to rise, no new whey was added to the column, and the column was allowed to drain with the stopcock fully open until the liquid was just above the level of the resin. The MF permeate that flowed through the column with the stopcock fully open was recycled by adding it to the next batch of MF permeate that passed through the cation exchange resin column after the column was regenerated.

When the pH of the MF permeate began to rise, the cation exchange column was regenerated before additional MF permeate was treated. Regeneration was done by adding concentrated food grade sulfuric acid (Sigma Aldrich) as per the literature,[32] after which water was flowed through the column at 1–2 mL min^{-1} until the pH of the water exiting the column rose above 3.

Treatment with the anion exchange resin Amberlite IRN-78 (DOW) was done in a beaker. The anion exchange resin was rinsed with water until no fish smell remained before use, as this fish smell can impart an undesirable fish flavor to the glucose-galactose syrup. The product of the lactose hydrolysis reaction was mixed with resin using a stir bar on a stir plate at a stir rate sufficient to cause there to be no stagnant resin until the pH reached 7. At this point, the beaker was removed from the stir plate and the whey was vacuum filtered to separate the resin from the whey. The vacuum filtration was performed quickly upon the pH reaching 7 because the pH will continue to rise if the whey remains in contact with the resin. The resin was regenerated by spinning the resin in a concentrated NaOH solution[32] followed by rinsing in water until the pH of the water was below 11.

2.3.2 Lactose hydrolysis. The lactose present in the acidified MF permeate was hydrolyzed in a benchtop flow reactor shown in Fig. 2. The 1.27 cm diameter stainless steel reactor

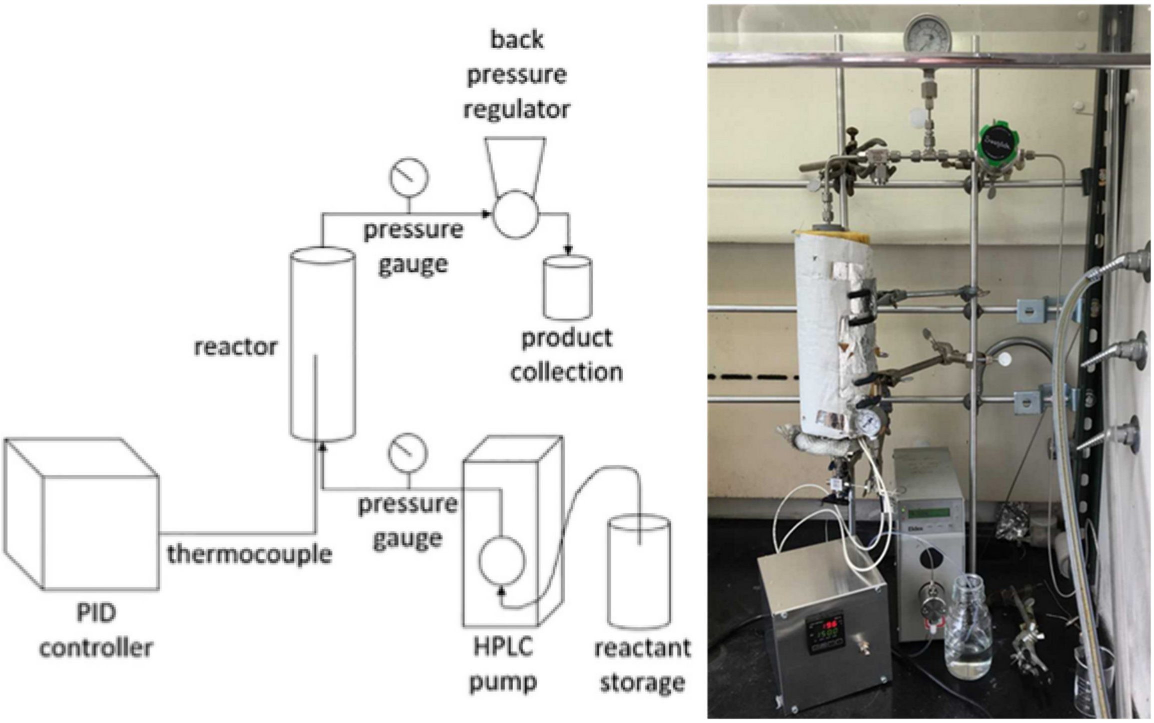

Fig. 2 Scheme (a) and photo (b) of lactose hydrolysis reactor system.

Table 1 Reaction conditions and product selectivities for acid-catalyzed lactose hydrolysis

Parameter	Value
Temperature (°C)	140
Flowrate (mL min^{-1})	2.4
WHSV (h^{-1})	3.7
Feed pH	1.1
Lactose conversion (%)	90
Monosaccharide selectivity (%)	99
Monosaccharide yield (%)	89

was heated by heating tape wrapped around an aluminum cylinder and surrounded by insulation (see Fig. 2). Temperature was maintained at 140 °C using a PID controller connected to a thermocouple and the heating tape. Pressure was controlled at 6 bar with a back pressure regulator. Reactant was introduced by an Eldex (Napa, CA) Optos 1SM model HPLC pump. The homogeneous acid catalyst for the lactose hydrolysis reaction was provided by treatment of the MF permeate with the cation exchange resin. The reaction conditions are summarized in Table 1. The flowrate of acidified MF permeate was adjusted to maintain the lactose conversion near 90%, with a flowrate of 2.4 mL min^{-1} being typical. This corresponds to a weight hourly space velocity (WHSV) of 3.7 h^{-1}. Slight differences in the flow rate needed to reach the desired lactose conversion were due to variations in salt concentrations among the GAW batches.

2.3.3 Characterization (HPLC, total solids, particle size distribution, ICP, *etc.*). The methods used to determine the composition of the Greek yogurt acid whey feed has been described in previous publications.[16] The concentration of glucose, galactose, lactose, HMF, citric acid, and lactic acid for the various filtered GAW streams were characterized by High Performance Liquid Chromatography (HPLC) using a Shimadzu (Kyoto, Japan) HPLC with Aminex (Bio-Rad, Hercules, CA) HPX-87H acid column at 30 °C with mobile phase of 0.6 mL min^{-1} of 5 mM sulfuric acid. Samples were filtered using a 0.2 micrometer membrane (VMR brand) and diluted to 10% of the original concentration with distilled water before injection in the HPLC. HMF was quantified at 290 nm using the UV detector, and all other compounds were quantified using the refractive index detector (RI). Total solids were quantified by weighing a sample before and after evaporation in a 110 °C oven for 4–24 hours.

The chemical composition of the acid whey samples was found using a Vista MPX Simultaneous inductively coupled plasma optical emission spectrometry ICP-OES (Varian Inc., Palo Alto, CA) with an Agilent (Santa Clara, CA) SPS 3 Autosampler. Prior to analysis with the ICP-OES, samples were filtered through a 20 mm syringe filter, acidified, and diluted as needed.

Particle size distribution of Ca precipitates was determined by laser light scattering using a Mastersizer 3000 (Malvern Instruments, Worchestershire, UK). Measurements were performed at an obscuration value of 10–15%. The particle size distribution was calculated from the light scattering pattern using Mie theory.[33] A refractive index of 1.63 and absorption of 0.01 were used. Deionized water was used as the dispersant.

2.3.4 Activated carbon and syringe filtration. The syrup produced by evaporation was treated with Hydrodarco 3000 activated carbon (Cabot Corporation, Alpharetta, GA). Activated carbon equaling 1% of the weight of the syrup was added to the syrup along with a stir bar. Using a clamp, the syrup with the activated carbon and stir bar was carefully lowered and centered in an oil bath at 60 °C. After 30 minutes, the vial was removed and the contents were filtered with a 0.45 μm PES (polyethersulfone) filter to produce the final glucose-galactose syrup.

2.3.5 Rotovap evaporation. The neutralized reaction product was evaporated at 0.175 bar using a Buchi (New Castle, DE) rotary evaporator (V-700 Vacuum Pump, V-850 Vacuum Controller, and R-210 Rotavapor) and an oil bath at 75 °C. A 1 L round bottom flask was filled with up to 500 mL of reaction product, connected to the instrument, and lowered into the oil bath. A separate cooler (Polystat R6L) flowed a 50 : 50 ethylene glycol : water mixture through the cooling tube at 10 °C to condense the water vapor. The evaporation was stopped when the syrup in the round bottom flask bubbled aggressively and threatened to move out of the flask towards the coolant rig, which generally occurred when ~83% of the initial mass of the reaction product had been evaporated.

2.4 Process modeling and economic analysis

On the basis of experimental data and literature results,[16,17,28,34–36] Aspen Plus (Aspen Technology Inc, Massachusetts, USA) process models were developed to estimate key mass and energy flows for the for the production of GGS from GAW. A discounted cash flow analysis was used to determine the internal return rate (IRR) of GGS under the set of financial assumptions shown in Table 4. The GAW feed rate of 907 metric tons per day corresponds to a large Greek yogurt manufacturing plant. The GAW credit is $18 per metric ton,[37] an estimate of the average price dairy plants pay for GAW removal. The capital costs, as well as variable and fixed operating costs, were then estimated for the calculated flow rates. The capital costs are based on quotes by engineering companies and Aspen simulation estimates, whereas the variable operating costs are estimated based on the raw materials prices given in Table 8. An extra 10% of the sum of installed equipment costs was added to account for storage and other outside battery limits costs.

2.5 Life cycle assessment

OpenLCA software, which is licensed under the Mozilla Public License (open source), was used to estimate the environmental impacts of the proposed GGS production from GAW. Agribalyse provided input information for agriculture and food materials. The life cycle impact assessment (LCIA) converted the resource consumptions and emissions of the life cycle inventory data into impact categories. In this regard, ReCiPe 2016 midpoint (H) LCIA method was applied to highlight three impact categories including global warming potential, water consumption, and ozone formation.[38] Additionally, we conducted an evaluation of HFCS[39] and SS[40] to compare the same environmental impacts to that of GGS production.

3. Results and discussion

3.1 Pilot plant operation

Pilot plant filtration and spray drying equipment was used to produce the calcium supplement milk minerals and a feedstock for production of the sweetener syrup glucose-galactose syrup (GGS) from Greek yogurt acid whey (GAW). The GAW, which was previously treated with ultrafiltration (UF), was treated with pilot-scale nanofiltration (NF) to concentrate the lactose, calcium and phosphate. The NF permeate was 1% total solids, 0.3% lactic acid, 0.4% salt, 0.1% lactose, and 0.2% galactose. (See Table 4 Streams 5 through Stream 22. The material balance for these streams was primarily created using data from the pilot plant characterization data, although the total solids are lower in the pilot plant experiments due to flux limitations for the NF.) While future research may find applications for NF permeate, it was considered a waste stream with a disposal cost equal to that of GAW ($18 per metric ton (ref. 37)). A reverse osmosis (RO) unit was used to reduce the volume of the waste by removing water from the stream. The RO unit produced water with 0% solids that could be reused as process water elsewhere in the system. The RO retentate had 8.2% solids, meaning the RO concentrated the waste NF permeate by a factor of 8.

The NF retentate was 17.8% total solids and the NF permeate was 1.0% total solids, corresponding to approximately a 5 × concentration factor. 98% of the lactose and 95% of the calcium was retained by the NF membrane. Crowley *et al.* found that a stream with 17% solids entering the hydrocyclone resulted in smaller particle sizes *versus* feeds with a lower total solids, so water was added to the NF retentate to reduce the total solids to 15%.[28] Then the NF retentate was heated to 65 °C and neutralized to pH 7.1 using potassium hydroxide to precipitate calcium phosphate. The heating and addition of base reduce the solubility of calcium, resulting in the precipitation of calcium phosphate.[41,42] When the pH approached 6, the NF retentate changed from clear and yellow to turbid and off-white as calcium phosphate particles were formed. The neutralized NF retentate was sent to a hydrocyclone to fractionate the calcium particles by size. As the abrasiveness of calcium phosphate particles is directly related to particle size,[28] removing the largest particles from the stream should reduce damage to downstream equipment. The literature suggested that 6% of the total calcium and 1.3% of the water containing soluble components exited in the hydrocyclone bottoms,[28] and the remainder exited as hydrocyclone overflow.

The hydrocyclone overflow was next treated with microfiltration (MF) to isolate the precipitate. While no precipitate was visible in the MF permeate, analysis of the stream using ICP-OES found that 1616 ppm of calcium remained in the solution, along with 654 ppm of sodium, 346 ppm of phosphorus, and 7297 ppm of potassium. The potassium concentration is higher than other elements because of the KOH added to precipitate calcium phosphate. The MF permeate was 13% total solids with 10.0% lactose, 1.1% ash, and 0.1% organic acids. The MF retentate remained in the MF unit and was processed with diafiltration to remove soluble components from the precipitate prior to spray drying. A portion of the resulting purified MF retentate was processed by the spray dryer to produce 0.64 pounds of milk minerals.

Fig. 3 shows the particle size distribution of the milk minerals produced by spray drying the MF retentate. The peak corresponds to a particle diameter of 8.2 µm, the mean particle diameter is 8.9 µm, and there are no particles greater than 30 µm. Characterization using ICP, HPLC, and total solids showed that the composition of the milk minerals was 15% lactose, 3% galactose, 62.7% inorganic salt, including 21.6% calcium, and 12.2% water (Table 2). When compared to a commercial product (TruCal made by Glanbia[36]), the milk minerals produced here had a higher moisture and sugar content, but lower salt content. The increased moisture content could be due to a delay between the production and characterization of the milk minerals in which some moisture

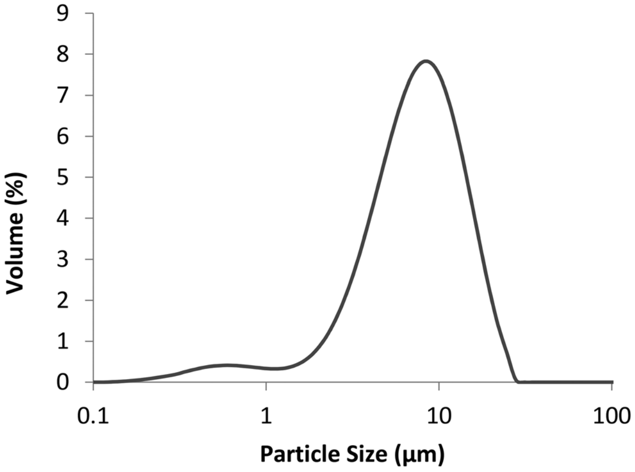

Fig. 3 Particle size distribution of the milk minerals produced by spray drying the MF retentate.

Table 2 Composition of milk minerals produced in the CDR pilot plant compared to TruCal,[36] a commercial milk minerals product produced by Glanbia. Units in weight %, except for particle size

	TruCal (wt%)	Milk minerals (wt%)
Total minerals	78	69.2
Inorganic mineral (ash)	>71	62.7
Calcium	>23.5	21.6
Phosphorus	>13	11.3
Organic mineral (citrate)	4	6.5
Sugars	7	17.8
Protein	<7	<0.5
Moisture	<6	12.2
Fat	<0.5	<0.5
Particle size	90% < 7 µm	90% < 16 µm

could have been absorbed by the milk minerals. Additional diafiltration of the precipitate during microfiltration could reduce the concentration of soluble components, thus increasing the concentration of calcium in the milk minerals.

Crowley *et al.* produced calcium phosphate particles large enough that gravity settling "promptly occurred" in the hydrocyclone bottoms,[28] allowing for decanting as a simple calcium phosphate purification method. Crowley also reported that increasing the total solids of the hydrocyclone feed from 5% to 10% could result in the precipitate particles colliding with each other inside the hydrocyclone, resulting in particle size reduction. In our pilot plant experiment, the hydrocyclone feed had a total solids of 15% and the hydrocyclone bottoms did not gravity settle within an hour, likely due to the lack of large particles in the feed or particle-to-particle collisions in the hydrocyclone. Instead of using hydrocyclones to fractionate calcium phosphate by particle size, the hydrocyclone could be used to slightly reduce downstream equipment damage by classifying (removing) the largest calcium phosphate particles.

3.2 Production of glucose-galactose syrup from MF permeate

The MF permeate was converted into glucose-galactose syrup (GGS) by the process shown in Fig. 1. First, the MF permeate passed through a cation exchange resin. The cations in the MF permeate were adsorbed by the resin which in turn release acidic protons into the solution, dropping the pH to around 1.1. The acidified MF permeate was then sent to a lactose hydrolysis reactor maintained at 140 °C. Here, the lactose undergoes lactose hydrolysis with the acid in the stream acting as a homogeneous acid catalyst. At a feed flowrate of around 2.4 mL min^{-1}, lactose is converted to the monosaccharides glucose and galactose at 90% conversion with a 99% monosaccharide selectivity (see Table 1). The reaction product was treated with anion exchange resin in a batch process. 94% of the anions in the stream, including organic acids, adsorbed to the resin resulting in the release of hydroxide ions. When the pH reached 6.5–7, the anion exchange resin treatment was stopped by removing the resin from the reaction product using vacuum filtration. This pH range was chosen because increases in pH indicate that anions and other non-sugar components are being adsorbed by the resin, which is desirable for the end product. However, the syrup should be prevented from becoming basic, since corn sweeteners are mildly acidic.[43,44]

The neutralized reaction product underwent evaporation at 75 °C using a rotary evaporator to remove water and increase the total solids. It was desirable to reach 71% total solids, the same as existing sweetener syrups such as high fructose corn syrup. However, the syrup in the evaporator bubbled violently at 45% total solids, forcing a stoppage of evaporation to avoid product loss. It is expected that commercial or pilot-scale evaporators could reach 71% total solids. The syrup, which is deep brown in color, underwent treatment with activated carbon in a batch process to reduce the coloration. The final glucose-galactose syrup (GGS) was obtained by using a syringe filter to remove the activated carbon.

Table 3 Comparison of GGS and HFCS-42. NPN refers to non-protein nitrogen, and HMF refers to 5-hydroxymethyl furfural

Sweetener	High fructose corn syrup 42	Glucose-galactose syrup
Composition (wt%)		
Water	29.0	29.0
Oligosaccharides	3.6	8.4
Monosaccharides	67.5	60.6
Organic acids	0.0	0.0
Ash	0.0	0.1
NPN	0.0	0.0
HMF	0.0	0.5
Humins/other	0.0	1.4
Relative sweetness	0.92	0.87

The composition of the GGS produced by this method was compared with the composition of high fructose corn syrup 42 (HFCS-42) in Table 3. The GGS produced had 6.9 wt% fewer monosaccharides than HFCS-42, 2 wt% more non-sugar components, and 4.8 wt% more oligosaccharides (the unreacted lactose). The relative sweetness for GGS was expected to be within 5% of HFCS-42 based on a study by Shah and Nickerson[45] despite the relative sweetness of the individual sugars in GGS being far less than in HFCS-42.[46] Shah and Nickerson explained that the components of GGS exhibited a synergistic sweetness effect which caused the relative sweetness of similar syrups to be higher than expected based on the individual sugars alone.[45] A higher level of lactose conversion would result in a syrup with a higher monosaccharide and lower oligosaccharide content. This could theoretically increase the sweetness but could also result in more off-flavors due to the increase in side reactions that would occur.

3.3 Process model

A process model was developed for production of GGS and milk minerals from GAW. For simplicity, this process model omits heat exchangers, storage vessels, and other equipment where no change in the material composition of streams occurs. Fig. 4 shows how the model is broken down into four sections that are separated by dashed-line boxes: (1) ultrafiltration, (2) nanofiltration and reverse osmosis, (3) milk minerals production, and (4) GGS production.

Prior to section 1 of the model, a milk inlet of 1359 metric tons per day (Stream 1) produced 453 tons per day of Greek yogurt (Stream 2) and 1000 metric tons per day of GAW (Stream 3). In the US there are at least two facilities of this size[47,48] and several others within an order of magnitude. The material balance was calculated using experimental pilot and laboratory data from Sections 3.1 and 3.2.

In Section 1 of the model, the GAW (Stream 3) passed through a UF unit where 100% of the macromolecules (protein, fat, bacteria, *etc.*) and 5% of other components remained in the UF retentate (Stream 4). The UF retentate was sold as animal feed since the protein and fat content provides

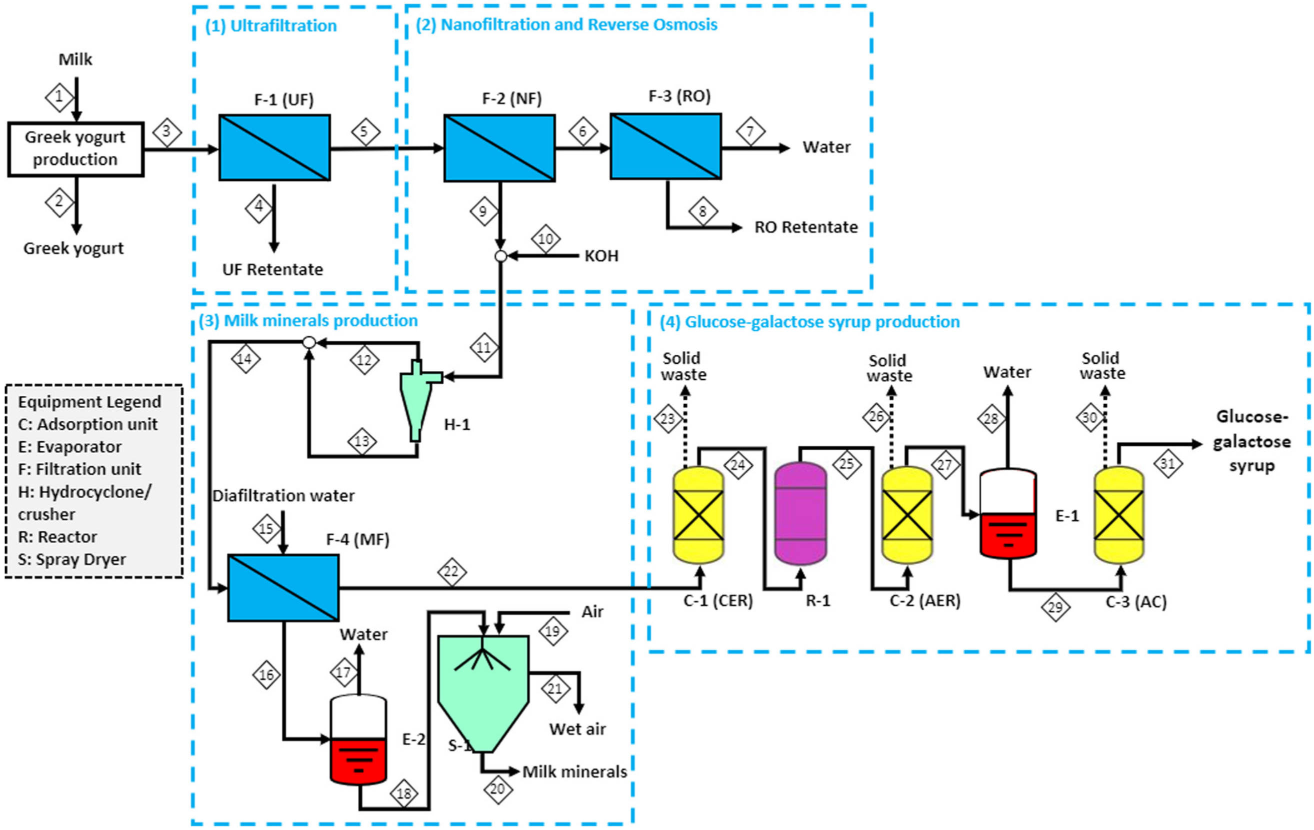

Fig. 4 Block flow diagram of the UW GAW process. For simplicity, this diagram omits heat exchangers, storage vessels, and other equipment.

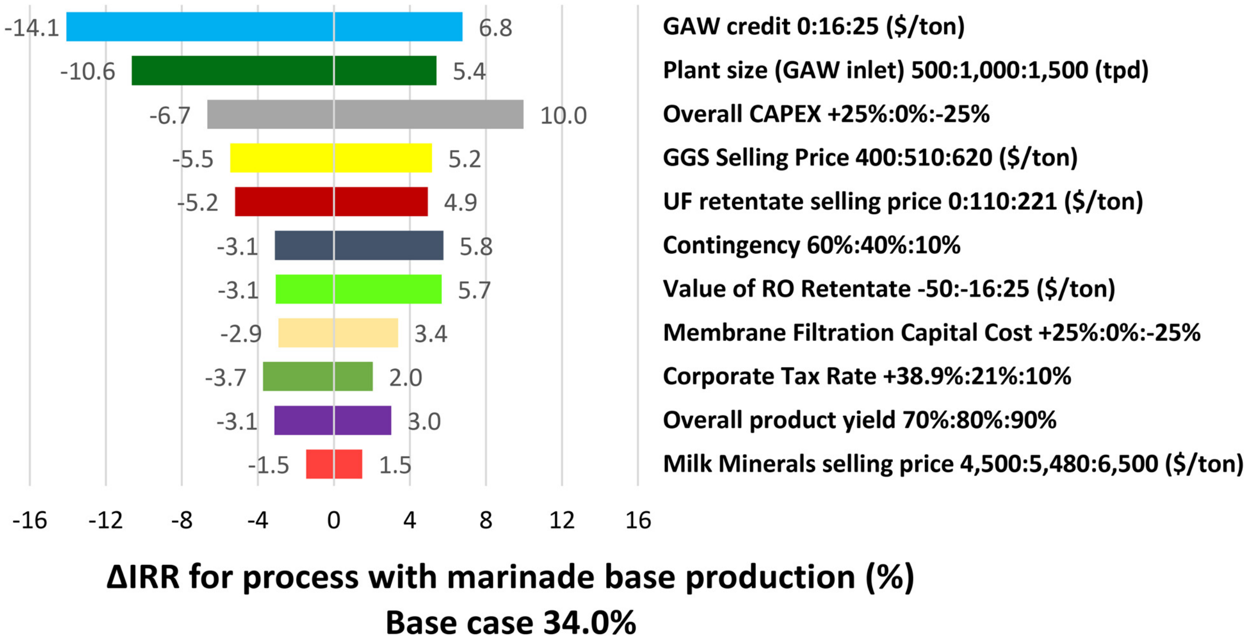

Fig. 5 Sensitivity analysis for the UW GAW Process. The center line represents the base case internal rate of return (IRR) of 34.0%. Bars extending to the right and left of center reflect an increased and decreased IRR, respectively. The values for each variable are shown to the right of the variable name as low IRR case : base case : high IRR case.

Table 4 Material balance for the process to produce glucose-galactose syrup and milk minerals from Greek yogurt acid whey. Units are metric tons per day. The stream numbers refer to the streams in Fig. 4. NPN refers to non-protein nitrogen, and HMF refers to 5-hydroxymethyl furfural

Stream	1	2	3	4	5	6	7	8	9	10	11
Total	1358.7	452.9	905.8	47.2	858.6	716.6	625.8	90.8	142.0	1.4	143.4
Water			851.2	42.6	808.7	709.3	625.8	83.5	99.4	1.0	100.4
Lactose			34.4	1.7	32.7	0.7	0.0	0.7	32.0	0.0	32.0
Galactose			7.8	0.4	7.4	1.4	0.0	1.4	6.0	0.0	6.0
Glucose			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Organic acids			2.5	0.1	2.4	1.9	0.0	1.9	0.5	0.0	0.5
Ash			7.2	0.4	6.9	2.8	0.0	2.8	4.1	0.3	4.4
Calcium (part of Ash)			1.1	0.1	1.0	0.1	0.0	0.1	1.0	0.0	1.0
HMF			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NPN			0.6	0.0	0.6	0.5	0.0	0.5	0.1	0.0	0.1
Humins			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Fat			1.1	1.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Protein			0.9	0.9	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Air			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

Stream	12	13	14	15	16	17	18	19	20	21	22
Total	141.5	1.9	143.4	85.3	20.6	16.3	4.4	131.2	1.5	134.1	208.0
Water	99.1	1.3	100.4	85.3	19.3	16.3	3.1	0.0	0.2	2.9	166.4
Lactose	31.5	0.4	32.0	0.0	0.2	0.0	0.2	0.0	0.2	0.0	31.7
Galactose	5.9	0.1	6.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	6.0
Glucose	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Organic acids	0.5	0.0	0.5	0.0	0.1	0.0	0.1	0.0	0.1	0.0	0.4
Ash	4.3	0.1	4.4	0.0	1.0	0.0	1.0	0.0	1.0	0.0	3.4
Calcium (part of Ash)	0.9	0.1	1.0	0.0	0.3	0.0	0.3	0.0	0.3	0.0	0.7
HMF	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NPN	0.1	0.0	0.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.1
Humins	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Fat	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Protein	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Air	0.0	0.0	0.0	0.0	0.0	0.0	0.0	131.2	0.0	131.2	0.0

Stream	23	24	25	26	27	28	29	30	31
Total	2.8	205.2	205.2	3.3	202.0	151.0	51.0	1.5	49.5
Water	0.0	166.4	165.3	0.0	165.3	151.0	14.4	0.0	14.4
Lactose	0.8	30.9	4.4	0.1	4.3	0.0	4.3	0.1	4.1
Galactose	0.2	5.8	18.8	0.5	18.3	0.0	18.3	0.6	17.8
Glucose	0.0	0.0	13.0	0.3	12.7	0.0	12.7	0.4	12.3
Organic acids	0.0	0.4	0.4	0.4	0.0	0.0	0.0	0.0	0.0
Ash	1.8	1.7	1.7	1.6	0.1	0.0	0.1	0.0	0.1
Calcium (part of Ash)	0.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HMF	0.0	0.0	0.6	0.3	0.3	0.0	0.3	0.1	0.2
NPN	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Humins	0.0	0.0	1.0	0.0	1.0	0.0	1.0	0.3	0.7
Fat	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Protein	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Air	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0

valuable nutrition to livestock.[49] Whether or not a UF filtration system is needed depends on the Greek yogurt production methods, since some of these methods retain nearly all the protein and fat present in the milk.

In Section 2 of the model, UF permeate (Stream 5) entered an NF system and was split into the NF retentate (Stream 6) and the NF permeate (Stream 9). The NF system was assumed to increase the total solids of the NF retentate (Stream 9) to 30%. The NF membranes retained 98% of the lactose, 81% of the galactose, and 84% of the multivalent cations. The non-protein nitrogen (NPN), monovalent cations, and organic acids were minimally retained by the NF membrane such that only 17% of the NPN, 20% of the monovalent cations and 18% of the organic acids were retained.

The NF permeate (Stream 6) may find future applications, but it was considered to be a waste stream in this model. The NF permeate passed through an RO unit where 88% of the water was removed and reused as process water (Stream 7). The model RO retentate (Stream 8), comprised of salt, organic acids, and galactose, was disposed of at the same disposal cost as GAW ($18 per metric ton). The RO reduced the volume of waste in the model from 717 metric tons per day to 91 metric tons per day. If the RO unit is removed from the economic analysis, the IRR dropped from 37.4% to 31.4% (see Section 3.4).

The NF treatment increased the concentration of calcium in the whey from 0.12% in the UF permeate to 0.6% in the NF retentate. The NF retentate was heated to 65 °C and neutralized to pH 7 with potassium hydroxide (Stream 10) to precipitate calcium phosphate. Potassium hydroxide was used as the neutralizing base instead of sodium hydroxide because sodium is consumed in excess by adults in the United States, whereas the same population has insufficient potassium consumption.[50]

Section 3 of the model encompasses particle size reduction treatment, microfiltration with extensive diafiltration, reverse osmosis, and spray drying. First, the neutralized NF retentate (Stream 11) passed to a hydrocyclone where the largest calcium phosphate particles were sent to the bottoms and may also experience size reduction, as discussed in section 3.1. The hydrocyclone bottoms would then undergo a separate crushing process to obtain particle sizes similar to the calcium phosphate particles in the hydrocyclone overflow. After the crushing process, the bottoms and overflow were recombined (Stream 14).[28]

Stream 14 passed to a microfiltration (MF) unit which retained 33% of the salt, including all the precipitate, 0.6% of the lactose, 0.1% of the galactose, and 4.5% of the organic acids. The model included the addition of 85.3 metric tons per day of diafiltration water (Stream 15) during the MF process to force the soluble components (sugars, NPN, organic acids, soluble salts) through to the permeate (Stream 22). The diafiltration water diluted the model MF permeate to 20% total solids from 30% total solids in Stream 14.

In the model, the MF retentate (Stream 16) passed through an RO unit where 86.4% of the water was removed and reused as process water (Stream 17). This increased the total solids of the RO retentate (Stream 18) to 30% from 5% in the MF retentate. Reducing the feed flowrate to the spray dryer using the RO unit reduced the cost of the spray dryer, a significant capital cost, by 82%. The spray drying process removed water to produce 1.5 metric tons per day of the milk minerals powder (Stream 20), modeled as having the same composition as the milk minerals produced in Section 3.1 (Table 2).

In Section 4 of the model, MF permeate (Stream 22) underwent ion exchange, homogeneous acid-catalyzed lactose hydrolysis, evaporation, and activated carbon filtration to produce GGS (Stream 31), similar to the process outlined in Fig. 1. First, the MF permeate passed through an anion exchange resin column. The model assumed that the industrial anion exchange resin and activated carbon columns prevented the resin or carbon from exiting the column, which eliminated the need for the vacuum filtration and syringe filtration steps from Fig. 1. The cation exchange resin removed half of the non-protein nitrogen and salt from the MF permeate as waste (Stream 23), but also 2.6% of the sugar. Because the cation exchange resin was in H^+ form, the ion exchange released protons into the MF permeate, acidifying the feed to a pH of 1.1.

The acidified MF permeate (Stream 24) underwent lactose hydrolysis in a flow reactor at the conditions listed in Table 1, except the monosaccharide yield is assumed to be 80% instead of 89% due to limitations of heat and mass transfer in large reactors. The reaction product (Stream 25) underwent anion exchange resin treatment where 94% of organic acids, 2.6% of sugars, 96% of anions, and 50% of HMF and NPN were removed as waste (Stream 26). The model set the WHSV such that the pH leaving the anion exchange resin does not exceed 7 as discussed previously. The model used an industrial scale evaporator like those used to produce corn syrup to increase the total solids to 71%[44] in Stream 29, unlike the 45% total solids obtained from the benchtop rotary evaporator in Section 3.2. An activated carbon column was used in the model to reduce the coloration of the syrup. The activated carbon was modeled to remove 7% of the organic acids, 4% of the ash, 27% of the HMF, 30% of the NPN, 28% of the humins, and 3% of the sugars as waste (Stream 30). The model predicted

Table 5 Assumptions of techno-economic analysis

Assumption	Value
GAW feed (metric tons per day)	907
GAW disposal cost ($ per metric ton)	18
Plant life (years)	30
Operating days/year	328.5
Depreciation schedule	7 years MACRS
Contingency (40%)	40
Tax rate (%)	21
GGS selling price ($ per metric ton)	562
UF retentate selling price ($ per metric ton)	121
Milk minerals selling price ($ per metric ton)	6041
Process water ($ per metric ton)	0.32

Table 6 Capital costs for UW GAW process

Row	Process area	Calculation method	Installed cost ($ million)
1	1. UF		2.0
2	2. NF and RO		2.8
3	3. Milk minerals production		2.6
4	4. GGS production		6.3
5	**ISBL**	= 1 + 2 + 3 + 4	**13.7**
6	Other OSBL (storage *etc.*)		0.7
7	Warehouse	4.0% of ISBL	0.5
8	Site development	9.0% of ISBL	1.2
9	Additional piping	4.5% of ISBL	0.6
10	**Total direct costs (TDC)**	= 5 + 6 + 7 + 8 + 9	**16.8**
11	Prorateable expenses	10.0% of TDC	1.7
12	Field expenses	10.0% of TDC	1.7
13	Home office & construction fee	20.0% of TDC	3.4
14	Project contingency	40.0% of TDC	6.7
15	Other costs (start-up, permits, *etc.*)	10.0% of TDC	1.7
16	**Total indirect costs**	= 11 + 12 + 13 + 14 + 15	**15.1**
17	**Fixed capital investment (FCI)**	= 10 + 16	**31.9**
18	Land & working capital	5.5% of FCI	1.8
19	**Total capital investment (TCI)**	= 17 + 18	**33.6**

that 49.5 metric tons per day of GGS would be produced with the same composition as was listed in Table 2.

We note that roughly 10% of the sugars are lost to non-GGS streams, and roughly 70% of the calcium was lost to non-milk minerals streams during the laboratory and pilot plant experiments. This represents an opportunity for further process improvement and optimization. It's know that a high percentage of the calcium phosphate precipitates when employing the routine described above.

3.4 Techno-economic analysis and life cycle assessment

A techno-economic analysis (TEA) was conducted to estimate the internal rate of return (IRR) generated by this technology, hereby called the UW GAW Process. The assumptions and fixed variables are listed in Table 5. Because the average cost of GAW disposal paid by Greek yogurt-producing companies is 18 $ per ton, a credit of $18 per ton for GAW disposal was assumed.[37] This further assumes that the GAW producer owns the GGS-producing plant and thus would not need to purchase GAW from itself, an assumption which is relaxed later in Fig. 5. The GGS selling price of $562 per metric ton was the same as the most recent price for HFCS-42.[51] The price for the UF retentate was estimated by multiplying the price of whey protein concentrate (WPC-34)[52] by the ratio of protein concentration in UF retentate *versus* WPC-34. Since this is a new technology, a high contingency of 40% was included in the assumptions.

The process from Fig. 4 and the material balance from Table 4 were input into the Aspen Plus model, and equipment and utility costs were estimated. While not shown in Fig. 4 or Table 4, equipment such as heat exchangers and storage vessels were included in the Aspen Plus model and incorporated in the TEA. Table 7 shows the capital expenditures for the process. The process areas used to calculate the ISBL (inside battery limits) costs were the sections of the process flow diagram in Fig. 4. The installed equipment costs for those areas were summed to generate an ISBL cost of $13.7 million. The Fixed Capital Investment (FCI) was calculated by summing the direct and indirect costs, and finally the land and working capital was added to the FCI to generate the total capital investment (TCI) of $33.6 million (Table 6).

The total revenue produced by each of the four product streams is shown in Table 8. The total revenue produced by the UW GAW Process was $19.3 million per year. The breakdown in the revenue streams was GGS production $9.2 million per year, GAW disposal credit $5.3 million per year, milk min-

Table 7 Revenue streams of the UW GAW process

Product	Stream number	Value ($ per metric ton)	Flow (metric tons per day)	Flow (metric tons per year)	Value ($ million per year)
GAW disposal credit	3	18	907.2	298 010	5.3
GGS (wet basis)	31	562	49.5	16 281	9.2
Milk minerals	20	6041	1.5	313	3.0
UF retentate (sold as animal feed)	4	121	47.3	15 535	1.9
Total revenue					**19.3**

Table 8 Operating costs for the UW GAW process

Raw material	Amount	Unit	Price	Unit	$ million per year
Sulfuric acid (98% food grade)	12	Metric tons per day	221	$ per metric ton	0.83
sodium hydroxide	5	Metric tons per day	386	$ per metric ton	0.59
Ion exchange resins					0.02
Filter replacements and CIP					2.07
RO retentate disposal	91	Metric tons per day	18	$ per metric ton	0.53
Process water	399	Metric tons per day	0.32	$ per metric ton	0.00
Low pressure steam	4859	kW	1.90×10^{-6}	$ per kJ	0.26
Cooling water	264	kW	2.12×10^{-7}	$ per kJ	0.00
Grid electricity	16	kW	0.0691	$ per kW per h	0.37
Total variable operating costs					**4.15**

Position	Salary	# Required	Total	$ million per year
Plant engineer	$ 91 283	1	$ 91 283	
shift operators	$ 47 333	15	$ 709 995	
Total salaries			$ 801 278	0.80
Labor burden (90%)			$ 491 112	0.49
Maintenance	3.0% of ISBL		$ 409 525	0.41
Property insur. & tax	0.7% of FCI		$ 222 406	0.22
Total fixed operating costs				**1.92**
Total operating costs				**6.07**

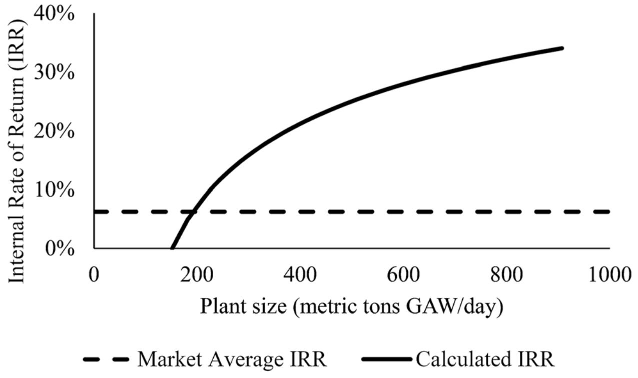

Fig. 6 The effect of Greek yogurt acid whey (GAW) inlet (metric tons per day) on the predicted internal rate of return (IRR). The dashed is at an IRR of 6.25%, the average for the US food sector.[54]

erals \$3.0 million per year, and UF retentate \$1.9 million per year. The operating costs are listed in Table 8. The variable operating costs, which included reactants, consumables, waste disposal, and utilities, totaled \$4.15 million per year, with the largest cost coming from filter replacements and CIP (clean in place) chemicals.[53] The utilities costs were generated by the economic analysis feature of Aspen Plus process simulation software.

The fixed operating costs, which included employee salaries, insurance, and maintenance, are also shown in Table 8 and totaled \$1.92 million per year. The number of required operators was estimated by assuming 4.8 operators per shift position. This provides a four-shift rotation and accounts for weekends, holidays, vacation, and overtime. For a fluids and solids batch process, at least three shift positions would be necessary.[54] Thus, the process is estimated to require 15 shift operators and one engineer. The maintenance and property insurance and tax were calculated as a percentage of the ISBL

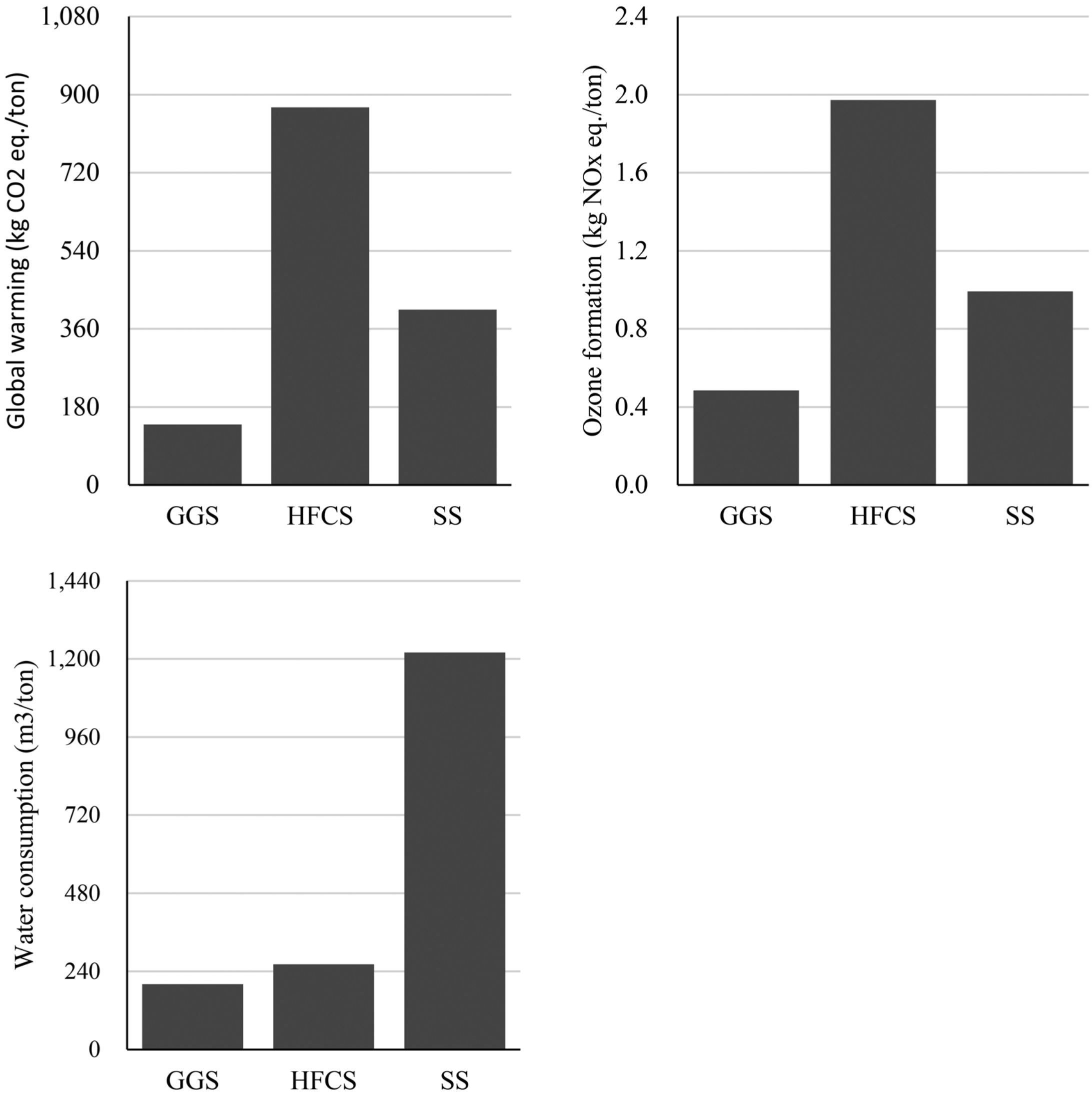

Fig. 7 The environmental impact of GGS, HFCS, and SS in terms of global warming, ozone formation, and water consumption categories.

costs and FCI, respectively. The total operating costs was \$5.82 million per year.

The IRR for the process was calculated using a discounted cash flow rate of return (DCFROR) analysis. The IRR was calculated to be 34.0%. For comparison, the most recent reported return on investment for General Mills, a large Greek yogurt producer, was 9.80%, and the average for the food products sector was 6.25%.[54] A sensitivity analysis was conducted to analyze the effect of individual variables on the IRR for the process (Fig. 5).

Of all the variable changes examined in this study, the one with the largest negative effect on the IRR was the GAW disposal credit. With a GAW credit of \$0 per metric ton, the IRR decreased from 34.0% to 19.9%, still more than double the 9.80% reported by General Mills and more than triple the food sector average. There are two main situations where the GAW credit could be \$0 per metric ton. First, feedstock transportation costs could eliminate any credits if the plant were constructed far enough away from the GAW-producing facility. Second, if the plant were not owned by the GAW producer, the price of the GAW would have to be negotiated. A more in-depth supply-chain analysis is beyond the scope of this paper but needed in the future.

While not shown in the figure, the IRR decreased to 29.3% if the RO unit which treats the NF permeate was removed, showing the value of waste reduction. Increasing the membrane filtration capital cost by 25% reduced the IRR by 2.9%, compared to a 6.7% reduction when the overall capital expenditure (CAPEX) was increased by the same amount. This shows the relative importance of filtration costs to the overall capital costs.

A sensitivity analysis was conducted on the plant size (Fig. 6). It was assumed that the capital expenditure scaled to a power of 0.6 and the operating costs and revenue scaled linearly. The IRR was 0% at 151 metric tons per day GAW inlet, showing the range of plant sizes a profit could theoretically be obtained. At 192 metric tons per day GAW inlet, the IRR rose above 6.25%, the average for the food sector.[54]

An LCA was performed to illustrate the environmental impact of GGS production from GAW in comparison to HFCS and SS production (Fig. 7). The production of GGS from GAW showed between 65.49 to 83.97 percent lower global warming potential (139.6 kg CO_2 eq. per ton) than from production of HFCS[39] from corn and the production of SS.[40] The ozone formation (0.5 kg NO_x eq. per ton), and water consumption (200.5 m^3 per ton) for GGS were also 23.25 to 83.57 percent lower than from HFCS and SS. Thus, the proposed process in this study for GGS production is the most environment-friendly among the three sweeteners. HFCS had the highest global warming potential (870.8 kg CO_2 eq. per ton) and ozone formation (2.0 kg NO_x eq. per ton),[39] while SS exhibited the highest water consumption (1220.0 m^3 per ton) among the scenarios.[40] Although the primary analysis may need to be further refined, the results so far indicate that the GGS production from GAW has a lower carbon footprint, ozone formation, and water consumption compared with alternatives.

4. Conclusion

A process was developed where milk minerals (CaP), a sweetener (GGS) and an animal feed were produced from GAW. Pilot plant filtration and spray drying equipment produced the calcium supplement milk minerals and a feedstock for production of a sweetener called glucose-galactose syrup. During the pilot plant operations, a particle size reduction was observed inside the hydrocyclone, but fractionation of the particles by size did not occur. The milk minerals produced during the pilot plant experiments had a calcium phosphate concentration similar to commercial milk minerals products. Laboratory-scale ion exchange and activated carbon filtration, lactose hydrolysis, and evaporation operations were conducted to produce glucose-galactose syrup. Data from these pilot and laboratory-scale experiments were input to a process model and TEA. For a GAW inlet of 907 metric tons per day, the after-tax net revenue is \$9.9 million per year and the capital expenditure is \$33.6 million. An IRR of 34.0% was calculated, for this process, which is far above the food and dairy industry average of 6.25%. LCA results showed that GGS production from GAW is more environment-friendly than HFCS from corn and SS from sugarcane in terms of global warming potential, ozone formation, and water consumption categories.

Conflicts of interest

There are no conflicts to declare.

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