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ORIGINAL PAPER

Techno-economic analysis of biomass to fuel conversion via the MixAlco process

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Abstract MixAlco is a robust process that converts biomass to fuels and chemicals. A key feature of the MixAlco process is the fermentation, which employs a mixed culture of acid-forming microorganisms to convert biomass components (carbohydrates, proteins, and fats) to carboxylate salts. Subsequently, these intermediate salts are chemically converted to hydrocarbon fuels (gasoline, jet fuel, and diesel). This work focuses on process synthesis, simulation, integration, and cost estimation of the MixAlco process. For the base-case capacity of 40 dry tonne feedstock per hour, the total capital investment is US $5.54/annual gallon of hydrocarbon fuels (US $3.79/annual gallon of ethanol equivalent), and the minimum selling price [with 10% return on investment (ROI), internal hydrogen production, and US $60/tonne biomass] is US $2.56/gal hydrocarbon, which is equivalent to US $1.75/gal ethanol. If plant capacity is increased to 400 tph, the minimum selling price of biomass-derived hydrocarbon fuels is US $1.76/gal hydrocarbon (US $1.20/gal ethanol equivalent), which can compete without subsidies with petroleum-derived hydrocarbons when crude oil sells for about US $65/bbl. At 40 tph, using the average tipping fee for municipal solid waste (US $45/dry tonne) and current price of external hydrogen (US $1/kg), the minimum selling price is only US $1.24/gal hydrocarbon (US $0.85/gal ethanol equivalent).

Keywords MixAlco · Techno-economic analysis · Simulation · Design · Biofuel · Cost estimation · Sensitivity analysis

Introduction

Technologies for production of renewable fuels from biomass are being developed to reduce dependence on imported petroleum, decrease greenhouse gas emissions, and improve national security. Many biochemical and thermochemical pathways have been proven technically; however, high production costs prevent many pathways from being economically viable without government subsidies. Figure 1 summarizes typical pathways to produce hydrocarbon fuels from biomass via alcohols:

- The thermochemical pathway gasifies biomass into syngas ($CO + H_2$), which is then catalytically converted into methanol or ethanol;
- The sugar crop-to-alcohol pathway is utilized in industrial production of bioethanol from sugarcane in Brazil;
- The starch crop-to-alcohol pathway uses amylase enzymes to produce intermediate sugars, as exemplified by the corn-to-ethanol industry in the USA;
- The lignocellulosic crop-to-alcohol pathway uses cellulase enzymes to produce intermediate sugars, which is in the process of being commercialized.

The latter three biochemical platforms require aseptic fermentation, which is expensive. Furthermore, because of its recalcitrance to enzymatic hydrolysis, lignocellulose

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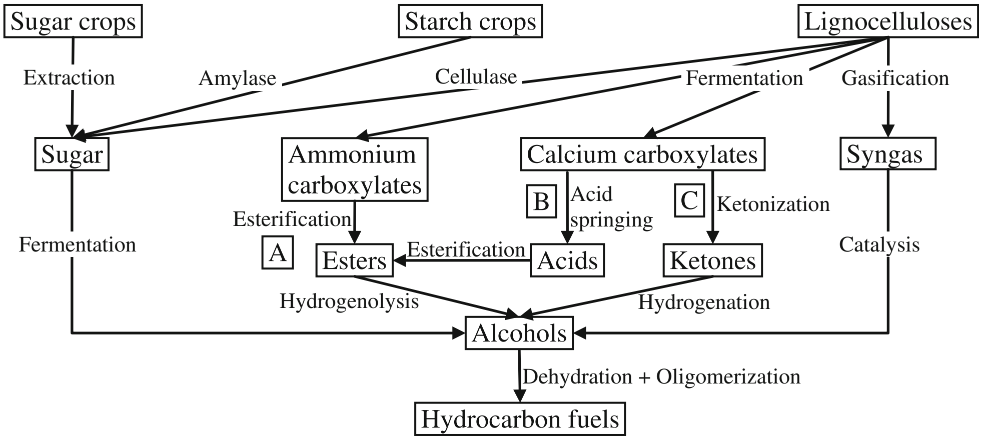


Fig. 1 Pathways for converting biomass to hydrocarbon fuels

requires high cellulase loadings. The MixAlco process produces fuels and chemicals from biomass without encountering those problems.

When producing hydrocarbon from alcohols, dehydration causes a mass loss of 30% (isopropanol), 39% (ethanol), and 56% (methanol). The theoretical loss of energy is 2% (isopropanol), 5% (ethanol), and 10% (methanol). Based on these examples, the pattern is clear: higher alcohols have greater retention of mass and energy when converted to hydrocarbons.

A key feature of the MixAlco process is the fermentation, which employs a mixed culture of acid-forming microorganisms to convert biomass components (carbohydrates, proteins, and fats) to carboxylate salts. The MixAlco process does not require aseptic conditions, which lowers capital costs and improves operability. The microorganisms produce their own enzymes—a type of consolidated bioprocessing—which reduces operating costs compared with traditional enzymatic pathways. Depending on the choice of buffer, the salts are ammonium carboxylates (buffered by NH_4HCO_3) or calcium carboxylates (buffered by $CaCO_3$). Via pathway A (Fig. 1), ammonium carboxylates are processed by esterification and hydrogenolysis, which produces a mixture of primary alcohols. Because there is almost no carbon loss from biomass to final products, this pathway has a high theoretical overall alcohol yield. Via pathway B (acid springing), the calcium carboxylates are converted to the corresponding carboxylic acids. From the acids, mixed alcohols are produced by esterification and hydrogenolysis. Via pathway C, calcium carboxylates are thermally converted into ketones, which are subsequently hydrogenated into a mixture of secondary alcohols. In this latter route (pathway C), the overall alcohol yield of is lower than that of the former (pathways A and B), but it does not demand as much hydrogen.

Previous work on MixAlco process economics includes a study by Holtzapple et al. [1], who estimated the economics of the calcium carboxylate platform (pathway C) using municipal solid waste or sugarcane bagasse as feedstocks. For the dewatering process, they employed extraction with amines and multi-effect evaporation. Lau et al. [2] evaluated production of ethanol from sweet sorghum via the acid springing platform (pathway B) using various scenarios of plant location, capacity, and incentives. Granda et al. [3] analyzed the process economics of the ammonium carboxylate pathway (pathway A) using municipal solid waste as feedstock in different scenarios of hydrogen sources and prices.

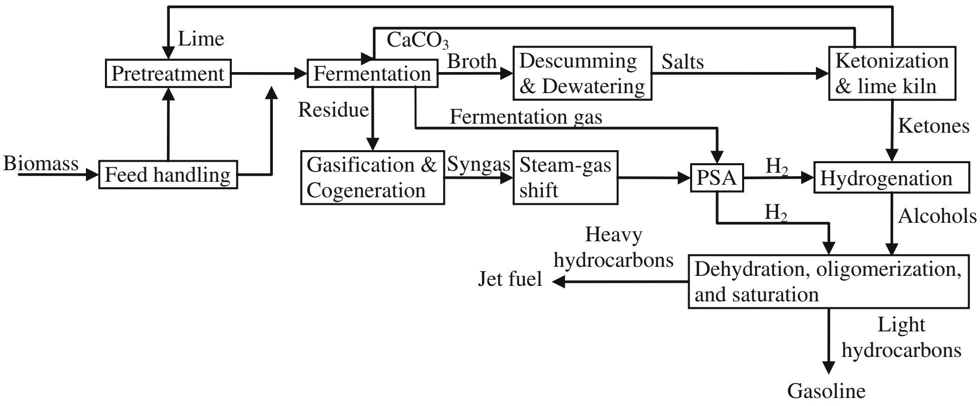
This work performs a techno-economic analysis for the MixAlco process based on the calcium carboxylate platform (pathway C) with vapor-compression dewatering. Compared with previous efforts, this work emphasizes process synthesis, integration, and analysis with simulation using Aspen Plus [4] and equipment cost estimation using Aspen Icarus Process Evaluator [5].

Process description

Figure 2 depicts biomass-to-hydrocarbon fuels conversion via the MixAlco process. To make hydrocarbon fuels, the MixAlco process uses the following steps: (1) pretreatment with lime, (2) fermentation with a mixed culture of acid-forming microorganism to obtain carboxylate salts, (3) dewatering using a high-efficiency vapor-compression evaporator, (4) thermal conversion of salts to ketones, (5) hydrogenation of the ketones to mixed alcohols, and (6) oligomerization of alcohols to hydrocarbons using zeolite catalysts.

If the lignocellulosic biomass has significant lignin content, it is pretreated with lime, which can be recycled using downstream processes. In the fermentation, a nutrient source (e.g., manure, sewage sludge) is mixed with the pretreated biomass. From the ketonization reactor, calcium carbonate is recycled to buffer the fermentation, which produces a broth of mixed calcium carboxylates. In the descumming and dewatering unit, the broth is concentrated

Fig. 2 Simplified process block diagram of the analyzed MixAlco process (pathway C)



to obtain solid salts, which are then thermally converted into ketones and calcium carbonate in the ketonization unit. Then, the ketones are hydrogenated into mixed alcohols. Potential sources of hydrogen include fermentation broth, gasified fermentation residue, reformed methane, and water electrolysis. To make hydrocarbons, the mixed alcohols are dehydrated and oligomerized to produce olefins, *n*-paraffins, iso-paraffins, and aromatics with boiling ranges of gasoline and kerosene, depending upon reaction conditions employed in the oligomerization reactor. The light fraction can be used for gasoline and the heavy fraction as jet fuel. To improve product quality, the olefins can be saturated.

In one process option, the undigested fermentation residues (about 20% of the biomass feed) is gasified and processed via cogeneration, steam-gas shift, and pressure-swing adsorption (PSA) units to generate steam, power, and hydrogen for the plant. Using pathway C, hydrogen from gasified biomass residue and fermentation gas are sufficient to supply the hydrogen needs of the plant.

In the MixAlco process, the fermentation broth contains 3–6% salts, which is concentrated using vapor-compression and crystallization units. The recovered distilled water is recycled to the fermentation and pretreatment. Water entering with the biomass feed is purged as distilled water, which can be sold as a byproduct.

Pretreatment and fermentation

In principle, any chemical or physical pretreatment method can be used in the MixAlco process; however, lime pretreatment is the best choice in terms of mass integration, reactor design, and operation. Recycled lime from the lime kiln is assumed to contribute 85% of lime demand in the pretreatment, with the 15% make-up lime purchased from external vendors. Using lime, the pretreatment can be performed in simple inexpensive pile reactors [3]. Pile pretreatment integrates with fermentation piles in a round-robin system, in which biomass solids are held in the same pile for both pretreatment and fermentation. Although the residence time of pretreatment (6 weeks) and fermentation (8 weeks) is long, the round-robin system results in steady flow rate and product concentration in the broth.

Using a mixed culture of microorganisms is the key feature of the MixAlco process. These microorganisms not only digest carbohydrates but also proteins and fats in biomass. For food wastes, this advantage gives a significantly higher overall yield than other common fermentation methods. The MixAlco process does not require sterilization or external enzymes, which reduces capital and operating costs. Consequently, reactors do not require stainless steel, and can be constructed using low-cost materials (e.g., concrete or plastic) that support piles or submerged fermentation.

Granda et al. [3] provide detailed information on the design of the pile reactors and operation of the round-robin system. Holtzapple et al. [1] discuss the design of submerged fermentation ponds.

Dewatering

In the produced fermentation broth, the calcium carboxylate concentration is 6% weight. Other components (e.g., dissolved carbon dioxide, microorganisms, undigested biomass, and other unknowns) are impurities and must be removed along with water. To purify the carboxylate salts, the broth is degasified by stripping, descummed using flocculant, evaporated using vapor compression, and crystallized (Fig. 3).

The key to vaporizing water efficiently is the novel design and optimized operating conditions of the vapor-compression unit. Figure 4 shows a simplified process flow diagram of the vapor-compression unit. At high pressure (8 bar), the descummed broth recovers heat from product streams (which contain condensed water), and is then evenly split into many stages. The figure shows three stages, but there can be tens of stages in practical plants.

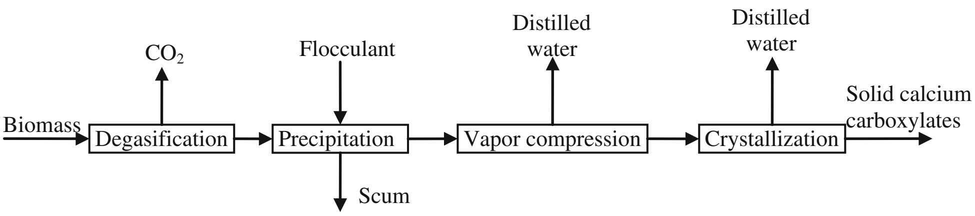


Fig. 3 Simplified process block of the descumming and dewatering unit

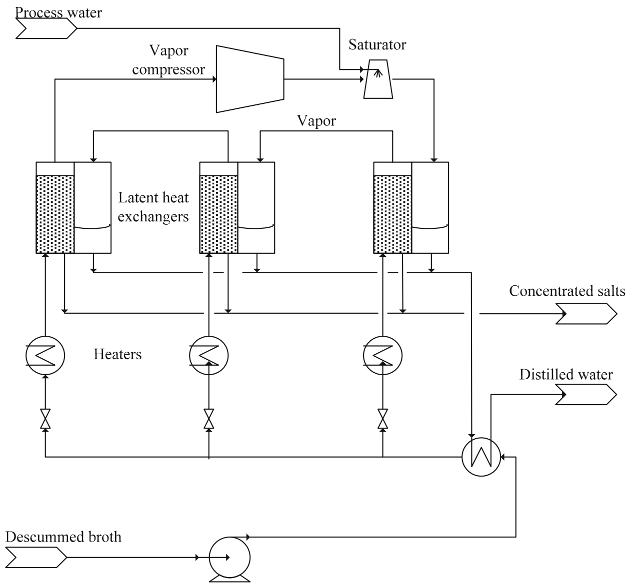


Fig. 4 A parallel configuration of multi-effect vapor-compression evaporator

In this work, six stages were chosen. In every stage, the inlet streams pass through a valve and a heater to be adjusted to saturation conditions. After that, they enter latent heat exchangers and use heat from the condensing vapor of an adjacent stage to vaporize water from the fermentation broth. The vapor from the first stage is compressed to a higher temperature and pressure and is then saturated so it readily condenses and transfers heat to the last stage. In Fig. 4, the pressure profile of the latent heat exchangers increases from left to right. With this profile, vapor from the right adjacent stage has higher temperature than the salt solution in the left adjacent stage; hence, heat transfer occurs. Using copper plates with a hydrophobic coating to promote dropwise condensation, an extremely high overall heat transfer coefficient of 170,350 W/(m^2 K) [30,000 Btu/(h ft^2 F)] was achieved at laboratory scale [6]. This allows a very small temperature approach of only 0.19 K (0.35°F) while maintaining a high heat flux. As a result, the compression ratio of the compressor is small, which saves both associated capital and operating costs. The net energy consumption of this vapor compression unit is only 5.5 kWh per 1,000 gallons of water vaporized, equal to 0.2% of the latent heat vaporization of the same amount of water.

Ketonization

The solid salts are sent to a dryer to remove residual moisture (Fig. 5). At high temperatures (430°C) in the ketonization reactor, calcium carboxylates are thermally converted into ketones and calcium carbonate as in the following representative equation:

$$\mathrm{R{-}COO\;Ca\;COO{-}R'} \rightarrow \mathrm{R{-}CO{-}R'} + \mathrm{CaCO_3},$$

where R and R′ represent hydrocarbon groups. The reactor is kept under vacuum (30 mmHg), which reduces residence time to avoid decomposition of the produced ketones. The ketone vapor is quickly removed from the reactor, quenched, and condensed. Part of the calcium carbonate is directly recycled to the fermentors, and the remaining portion is converted into quick lime (CaO) in a kiln. The quick lime is recycled to the pretreatment reactors. Unlike conventional lime kilns that are fed coarse limestone, this kiln processes fine calcium carbonate powder; thus, some processing steps (grinding, drying) are not needed.

Ketone hydrogenation

In this conversion step (Fig. 6), the ketone carbonyl groups react with hydrogen to form alcohol groups in an exothermic reaction:

$$\mathrm{R{-}CO{-}R'} + \mathrm{H_2} \rightarrow \mathrm{R{-}CH(OH){-}R'}.$$

The reaction is performed at high pressure (55 bar) and at isothermal (130°C) condition. The optimal design was found to be three continually stirred tank reactors (CSTR) in series [7]. In each CSTR, liquid ketones, solid Raney nickel catalyst, and hydrogen bubbles are well mixed. The heat of reaction is recovered by a pump-around system. Hydrogen is fed to every CSTR in 20% excess to maximize ketone conversion. The net

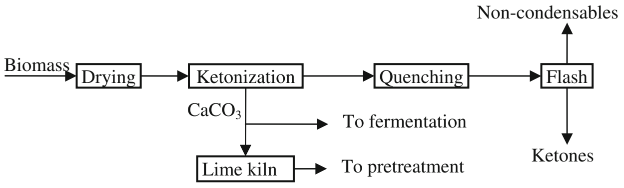


Fig. 5 Simplified block diagram of the ketonization and lime kiln unit

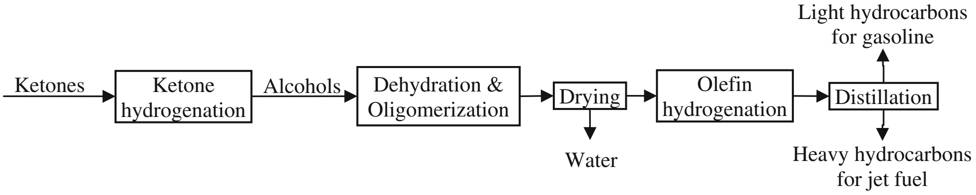


Fig. 6 Simplified block diagram of ketone hydrogenation, dehydration & oligomerization, and olefin hydrogenation units

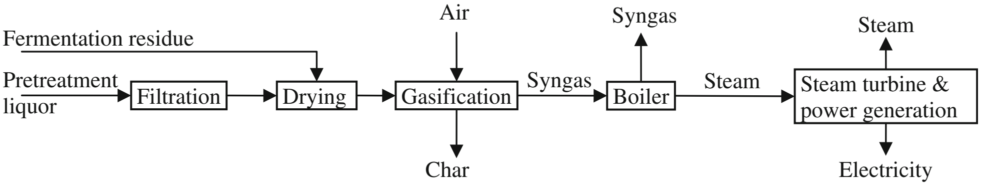


Fig. 7 Simplified block diagram of gasification and cogeneration unit

demand of hydrogen is 0.0225 kg H_2/kg mixed alcohol [25.0 standard cubic foot (SCF) per gallon of mixed alcohols] or 0.00687 kg H_2/kg dry ash-free biomass (1.30 SCF per dry ash-free pound of biomass). The produced mixture of secondary alcohols can be directly used as a transportation oxygenated fuel such as bioethanol, but it has higher energy content (net heating values are 34.6 and 26.8 MJ/kg, respectively [4]).

Dehydration and oligomerization

The mixed alcohols are further processed to produce hydrocarbon fuels (Fig. 6). Using H-ZSM-5 catalyst in a reactor at 300°C and 3 bar, the alcohols are dehydrated as follows:

$$\text{R-CH(OH)-CH}_2\text{-R}'' \rightarrow \text{R-CH=CH-R}'' + H_2O.$$

In the same reactor, the produced olefins are oligomerized as shown in the following simplified reaction:

$$C_mH_{2m} + C_nH_{2n} \rightarrow C_{m+n}H_{2m+2n}.$$

Depending upon the specific reaction conditions (time, pressure, temperature), the products are very complex and include olefins, *n*-paraffins, iso-paraffins, aromatics, and cyclics. Water in the products is removed in a drying unit using a salt filter.

Olefin hydrogenation

To improve fuel quality, the olefins can be hydrogenated to make corresponding paraffins (Fig. 6). Similar to the design of the ketone hydrogenation, this conversion unit employs CSTRs in series with Raney nickel catalyst. The carbon double bond (C=C) is saturated to stabilize the hydrocarbon product as follows:

$$C_xH_{2x} + H_2 \rightarrow C_xH_{2x+2}.$$

In this step, the net demand of hydrogen is 0.0139 kg H_2/kg hydrocarbon fuels (15.4 SCF per gallon of hydrocarbon fuel) or 0.0034 kg H_2/kg dry ash-free biomass (0.64 SCF per dry ash-free pound of biomass). Out of the reactor, the mixed hydrocarbons are distilled into C_{8-} and C_{9+} fractions. The light fraction and heavy components can be used as blending components for gasoline and jet fuel, respectively.

Hydrogen source

Hydrogen is required to produce alcohols and saturate hydrocarbons. Two scenarios of hydrogen sources were analyzed.

In the first scenario, hydrogen is produced by gasifying undigested biomass from the fermentors. The investment includes not only gasification but also cogeneration, steam-gas shift, and pressure-swing adsorption (PSA, next section) to supply hydrogen, steam, and power for the plant. Figure 7 shows the gasification and cogeneration processes in this scenario. First, the small amount of biomass suspended in the pretreatment liquor is recovered by filtration. Then, this filtered biomass is mixed with fermentation residue and dried in a rotary dryer using flue-gas heat. After drying, the moisture content in the biomass is about 10%. In the next step, the dried biomass is gasified to generate syngas and byproduct char. Energy from the hot syngas is used to make high-pressure steam, which is expanded in a steam turbine to generate power.

Figure 8 shows the gasifier, a fluidized bed followed by two cyclones that effectively remove particulates (char, ash) from the syngas [8]. Atmospheric-pressure air, but no steam, is introduced into the gasifier, which requires only a simple control system and low-cost cyclones. The capital cost of this gasifier is only 25% of commercial pressurized

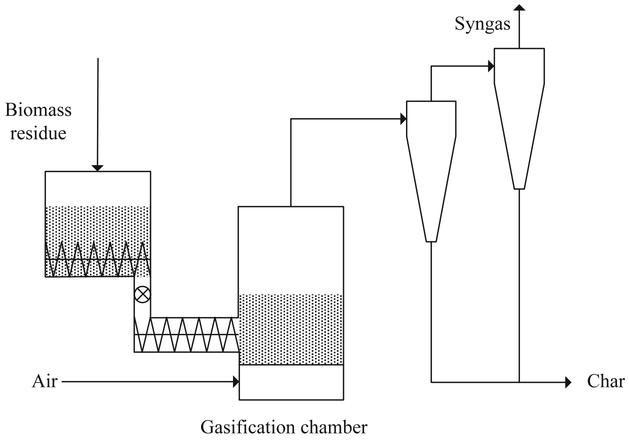


Fig. 8 Schematic of the atmospheric biomass gasifier

gasifiers that use steam [8]. There are multiple uses for the char recovered from the cyclones: (1) add to soil to sequester carbon and improve fertility, (2) sell to coal-fired power plants to provide "green" fuels, and (3) burn at the plant to recover alkaline ash that can be used to replace lime in the pretreatment.

In the second scenario, hydrogen is not produced in the plant but is purchased from external sources, such as pipelines or oil refineries. In this scenario, the gasification and cogeneration unit is retained to use biomass residues to generate steam and power, but an additional combustion chamber is installed after the cyclones to burn the syngas completely and produce more steam and power.

Steam-gas shift and pressure-swing adsorption

This stage is only needed when hydrogen is produced in the plant (as described in scenario 1). More hydrogen is made using the shift reaction between steam and carbon monoxide:

$$H_2O + CO \rightarrow H_2 + CO_2.$$

Because of the compositional characteristics of the syngas, a one-stage shift is sufficient for high conversion and low residence time. The hydrogen-rich syngas, along with fermentation gas, is passed through molecular sieve beds in the pressure-swing adsorption unit, which purifies hydrogen. Both of these technologies are well developed.

Cost estimation basis

For most sections in the plant, a detailed process flow diagram was developed using Aspen Plus software. Mass and energy balances around each piece of equipment were calculated. For equipment not readily modeled by Aspen Plus (e.g., pretreatment and fermentation reactors, gasifier, and pressure-swing adsorption), information from experts, literature, and vendors were used [3, 8–11].

Next, equipment was sized for key parameters (e.g., compressor power, heat exchanger area, distillation column diameter and height) which is necessary to estimate costs. Some equipment (e.g., crystallizer, drum dryer, clarifier) could not be sized by Aspen Plus with the available data, so reliable heuristics and assumption were used in these cases [3, 9, 12, 13].

Table 1 summarizes the key process performance parameters used in the mass and energy calculations. The long residence time of pretreatment and fermentation requires large piles, which have a volume of 160,000 m^3 each. When biomass is converted to hydrocarbon fuels, carbon is lost from the fermentation (carbon dioxide, fermentation residue) and ketonization units (calcium carbonate). Another yield loss (about 20%) occurs when oxygen is removed from alcohol molecules in the dehydration unit.

After sizes were specified, purchase equipment costs were estimated by using Aspen Icarus Process Evaluator [5]. Subsequently, other relevant costs (e.g., installation, instrumentation, site preparation) were estimated as factors of the purchase equipment cost. For pretreatment, fermentation, and support units (wastewater treatment, storage, and utilities), this work employs the modified factor method of NREL, which is suitable for aqueous-based processes [9]. This modification has a fivefold higher contingency factor than the original NREL method. Lang factors were applied for other units, which are similar to chemical and petrochemical processes [13]. If package quotes were available from the literature or vendors, those quotes were used instead of estimating the costs from individual pieces of equipment. Scaling factors for estimating equipment costs at various capacities are referred to NREL [9], which in turn took most of the scaling factors from Walas [12].

Project economics were analyzed using the assumptions in Table 2. Detailed cash flows over 20 years were constructed with the depreciation method and period recommended by NREL [9]. No subsidies or incentives were applied.

Table 3 lists key parameters for the base case. Sorghum is an energy crop that has been well studied and developed at Texas A&M University. Chicken manure supplies the nutrients for the fermentation. They are used in a sorghum-to-manure ratio of 80:20, which provides both an energy source and a nutrient source. In the base case, hydrogen is recovered in the plant from fermentation gas and shifted syngas. The recovered hydrogen is sufficient to meet the demands of ketone and olefin hydrogenation.

Table 1 Key process performance parameters

Unit	Parameter	Value
Pretreatment	Make-up lime demand	0.023 g CaO/g biomass
	Total lime loading	0.15 g CaO/g biomass
	Reaction time	6 weeks
Fermentation	Conversion	0.8 g digested/g volatile solids fed
	Selectivity	0.65 g carboxylic acids/g volatile solids digested
	Reaction time	8 weeks
Dewatering	Carboxylate recovery	95%
Ketonization	Conversion	99.5%
	Yield	0.583 g ketones/g carboxylic acids
Ketone hydrogenation	Conversion	98.4%
Dehydration and dimerization	Light hydrocarbon yield	0.6 g light hydrocarbon/g alcohols
	Heavy hydrocarbon yield	0.2 g heavy hydrocarbon/g alcohols
Olefin hydrogenation	Conversion	98.4%
Gasification and cogeneration	Gasification temperature	760°C
	Solid-to-air ratio	0.625
Steam-gas shift and PSA	Steam-gas shift temperature	254°C
	Hydrogen recovery	95%
The whole plant	Gasoline yield	62 gallons/dry ash-free tonne biomass
	Jet fuel yield	19 gallons/dry ash-free tonne biomass

Table 2 Assumptions for project economic evaluation

Parameters	Value
Plant life	20 years
US dollar	November 2009
Depreciation model	200% DDB for 7 years
Financing	100% equity
Construction period	1.5 years
Start-up time	6 months
Income tax rate	39% of profits
Subsidy	No
Operating season	8,000 h per year
Working capital investment	10% FCI

DDB double-decline balance

Results and discussion

For the base-case capacity, the fixed capital investment (FCI) is US $131 MM. Figure 9 shows how the capital is distributed in the plant. The steam-gas shift and PSA units contribute most to the FCI, followed by dewatering, gasification and cogeneration, ketonization, and olefin hydrogenation. This indicates that producing hydrogen from fermentation residue is expensive. Thanks to the low-cost pile design, pretreatment and fermentation require the least investment.

For the base case, the minimum selling price was estimated to be US $2.56 per gallon hydrocarbon, including a discount rate of 10%. For simplification in the sensitivity analysis, it was assumed that jet fuel and gasoline were sold at the same price. Table 4 shows the components of this selling price. The biomass feedstock is the highest cost component. This implies that using low-cost waste feedstocks will significantly reduce the product cost.

Sensitivity analysis

Figure 10 shows the sensitivity of the minimum selling price to the following key parameters: discount rate (or return on investment (ROI)), overall yields, fixed capital investment, external hydrogen prices, and key raw material prices. If investors expect an ROI of 20% or 25%, the product fuel must be sold for US $3.31 or US $3.74/gal hydrocarbon, respectively. For an increase of FCI by 50%, the product selling price is US $2.98/gal hydrocarbon. A decrease of overall yields by 10% results in selling prices of US $2.85/gal hydrocarbon. If the lime kiln is not employed and pretreatment is fully fed with purchased fresh lime, the selling price increases by US $0.09/gal hydrocarbon. In the worst-case scenario of manure and lime prices, the selling price is not affected much. Unlike the main biomass (sorghum), free manure does not reduce the hydrocarbon selling price significantly.

The sensitivity analysis shown in Fig. 10 reveals that the base case is not the best case. If external hydrogen is

Table 3 Inputs for the base case

Parameters	Value
Discount rate	10%
Hydrogen source	Produced in the plant, neutral net balance
Delivered price of forage sorghum	US $60/dry tonne
Delivered price of chicken manure	US $10/dry tonne
Lime (CaO) price	US $70/tonne
Plant capacity	32 dry tonne/h of forage sorghum containing 8.1% ash 8 dry tonne/h of chicken manure containing 35.4% ash

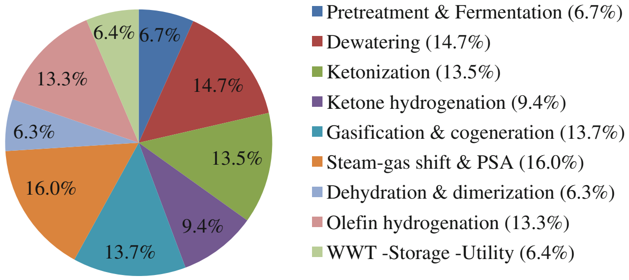


Fig. 9 Fixed capital investment for the base case. *WWT* Wastewater treatment

Table 4 Cost components of the minimum product selling price

Cost component	US $/gallon of product	Contribution (%)
Biomass	0.617	24.1
Chemicals and waste disposal	0.110	4.3
Utility	0.284	11.1
Labor	0.177	6.9
Maintenance	0.406	15.9
Operating overhead	0.082	3.2
Average capital depreciation	0.448	17.5
Average income tax	0.166	6.5
Average return on investment (ROI)	0.270	10.5
Total	2.560	100

available for US $1 or US $2/kg, buying hydrogen and burning the fermentation residue only for steam and power will lower the product cost. However, if the external hydrogen price is expensive (e.g., US $5/kg), then it is better to produce hydrogen from fermentation residues. Figure 11 shows that the break-even point is US $2.95/kg of external hydrogen. In Fig. 11, the relationship between hydrogen price and natural gas price is taken from Granda et al. [3].

Table 4 shows that feedstock costs have high impacts on product price. For the base-case plant, Figs. 12 and 13 depict the effect of feedstock price on the hydrocarbon selling prices. This sensitivity analysis was done for energy sorghum. For other feedstocks with the same volatile solids content as sorghum (91.9% weight), the results are identical. Some feedstocks (e.g., municipal solid waste, food waste) may come with tipping fees (which are described by negative prices in Figs. 12 and 13), whereas sorghum does not. If such feedstock is delivered to the gate of the MixAlco plant with a US average tipping fee of US $45/tonne and the plant uses internal hydrogen, the minimum product selling price is only US $1.57/gal hydrocarbon at the base-case capacity (Fig. 12). In a reasonable scenario where biomass is available with a tipping fee of US $45/tonne and external hydrogen is supplied at US $1/kg (the current price based on natural gas costing US $5.21/GJ or US $5.50/MMBtu), the minimum selling price of hydrocarbon fuels is US $1.24/gal hydrocarbon, as shown in Fig. 10 and 13a.

The base-case plant capacity (40 dry tonne/hour) is actually a small biorefinery. A large biorefinery can process up to 400 dry tonne biomass per hour. Figure 14 shows the hydrocarbon selling prices and fixed capital investment with respect to capacity for a plant at various prices of biomass. The FCI curve is represented by a function of capacity to a power of 0.72. At capacities of 300 tph or more and biomass cost of US $60/tonne, the minimum fuel selling price approaches US $1.75/gal hydrocarbon. Petroleum refineries sell gasoline and jet fuel at this price level when crude oil price is about US $65 per barrel, as shown in Fig. 15.

Conclusions

MixAlco is a robust process that converts biomass into oxygenated fuels (e.g., alcohol) or hydrocarbons (gasoline, jet fuel) for transportation. The process does not need external hydrogen, but it is more economic if external hydrogen is available for less than US $2.95/kg, which occurs when natural gas is below US $17.1/GJ (US $18.1/MMBtu).

The MixAlco process has the following advantages: no requirement for sterility, no external enzymes, low capital cost, and cost-effective dewatering. In particular, the minimum selling prices of hydrocarbon fuels can be around US $1.57/gal (US $1.07/gal equivalent ethanol) if

Fig. 10 Sensitivity analysis of the key factors on minimum product selling price

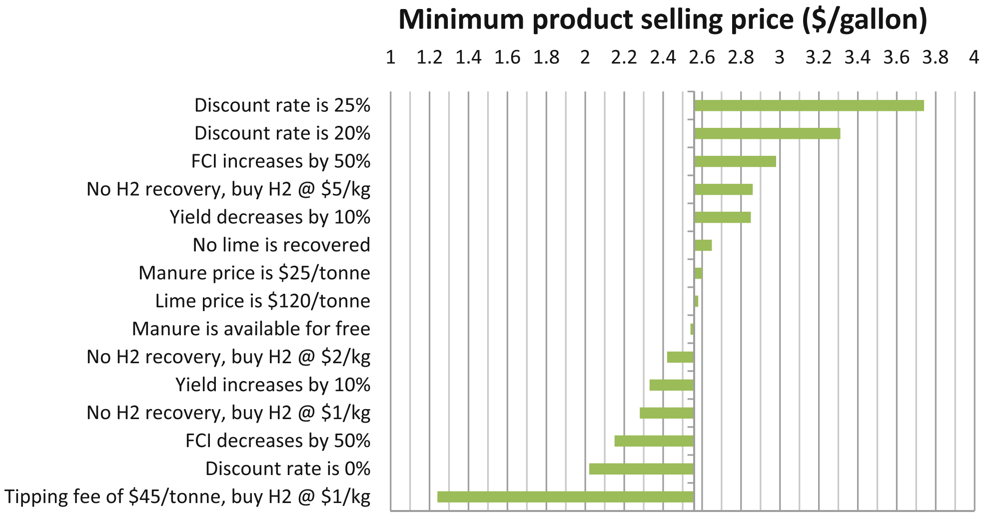


Fig. 11 Plot of minimum selling prices versus external hydrogen prices for the case of no hydrogen production

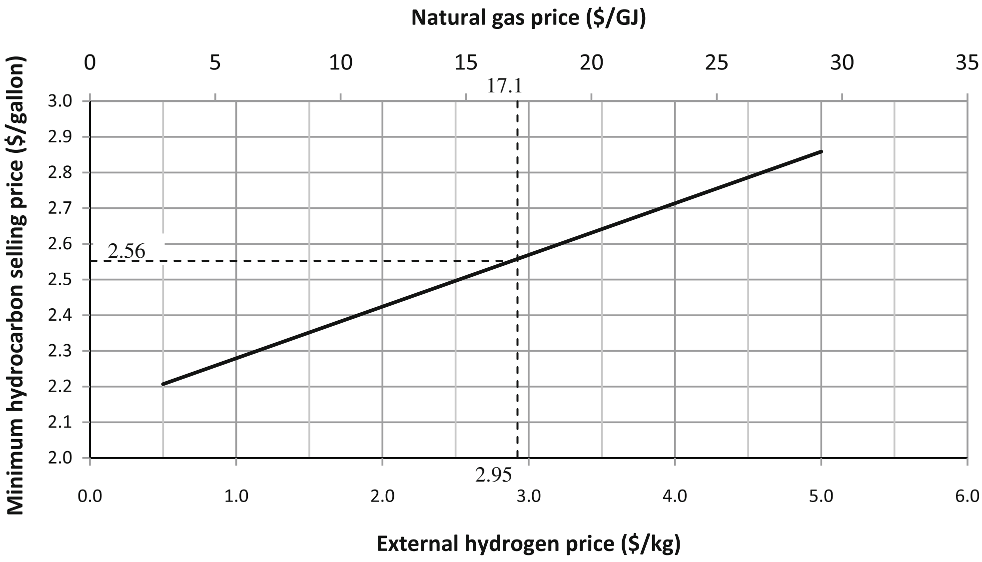
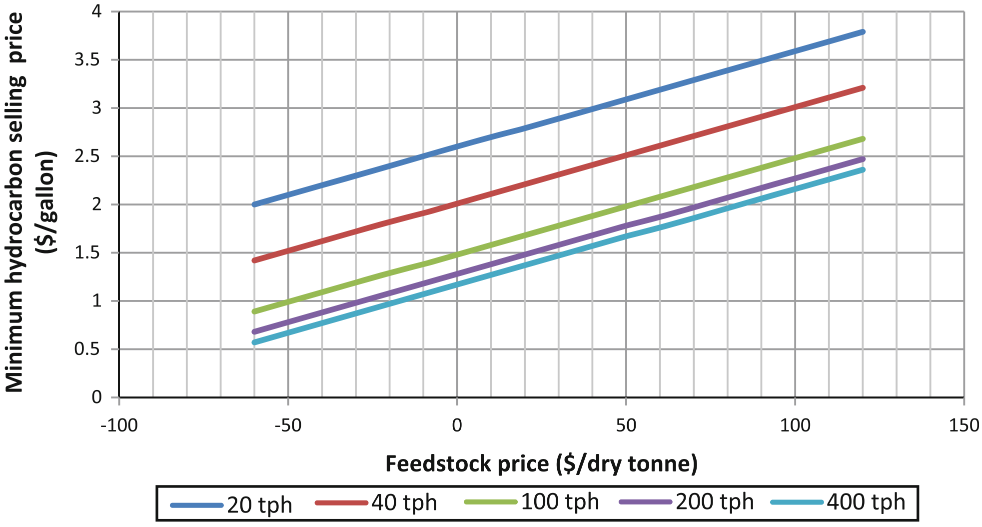


Fig. 12 Minimum product selling prices versus prices of biomass feedstocks using hydrogen generated from fermentation residues in various plant capacities



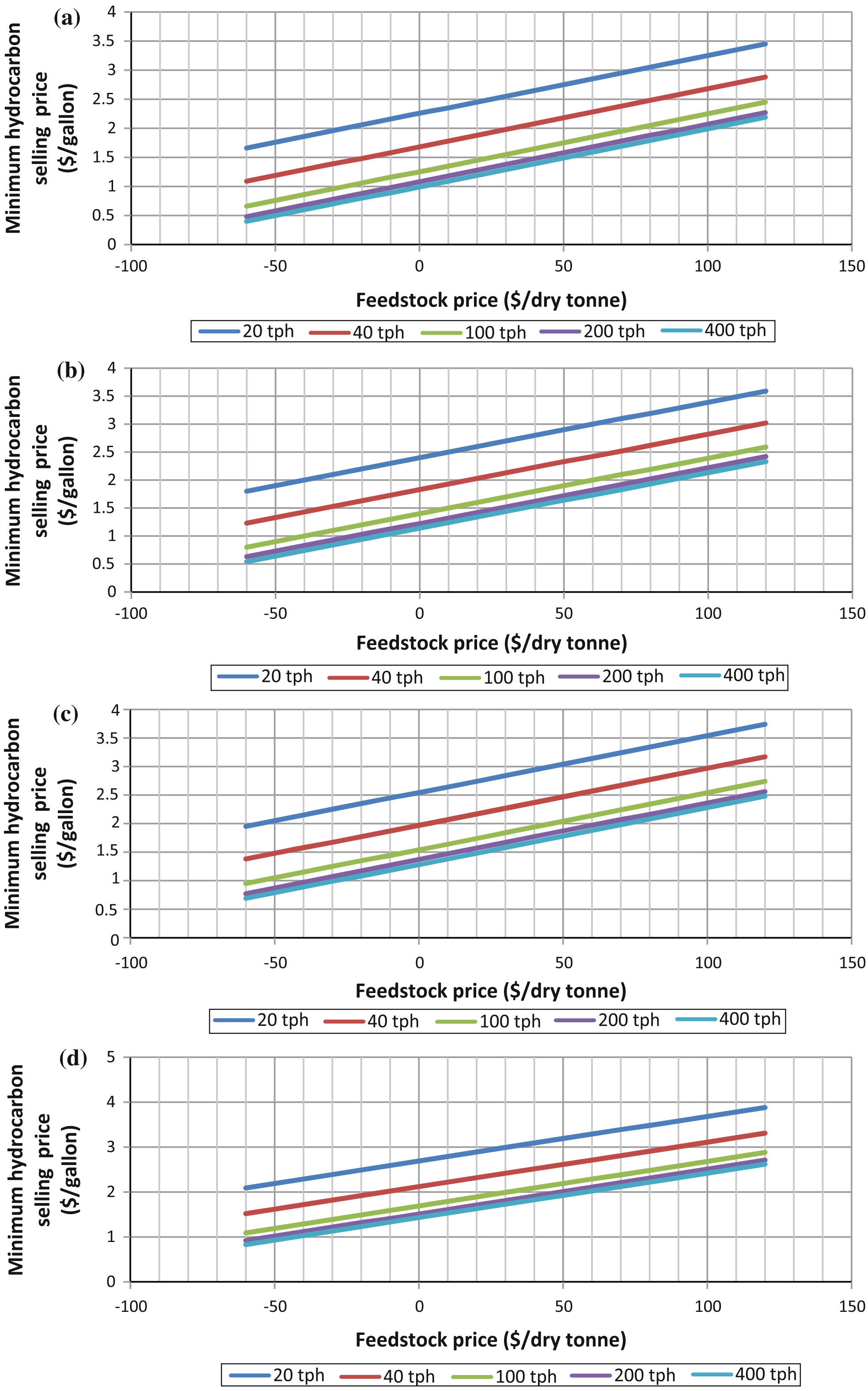


Fig. 13 Minimum product selling prices versus prices of biomass feedstocks in various plant capacities, using external hydrogen available for: **a** US $1/kg, **b** US $2/kg, **c** US $3/kg, and **d** US $4/kg

municipal solid waste is available at the US average tipping fee of US $45/dry tonne (40 tonne/h plant, with internal hydrogen production), or US $1.75/gal (US $1.20/gal equivalent ethanol) if the plant capacity is about 400 dry tonne per hour (US $60/tonne feedstock, with internal hydrogen production).

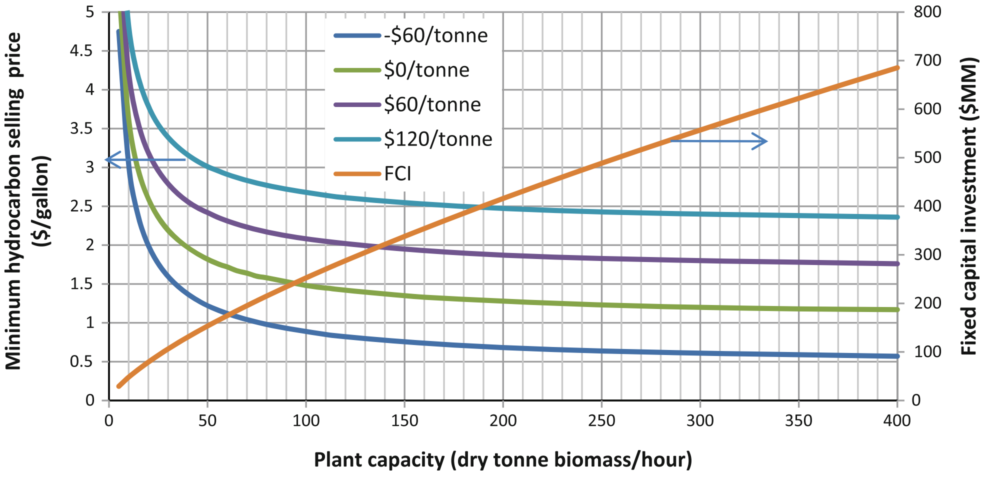


Fig. 14 Minimum selling prices and fixed capital investment versus plant capacities using biomass at various prices (sorghum) and US $10/dry tonne (manure). Hydrogen is produced by gasification

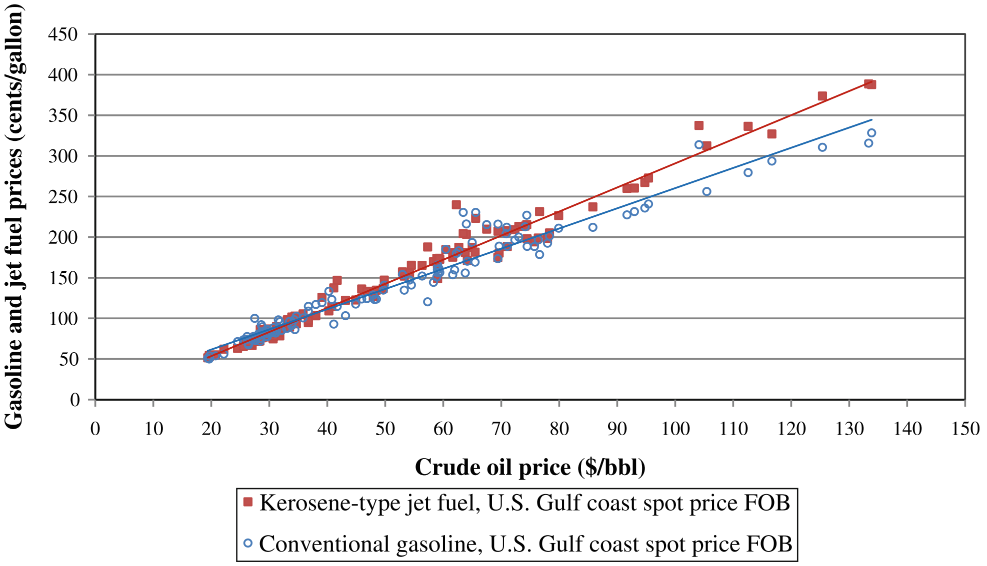


Fig. 15 Historical prices of crude oil, gasoline, and jet fuel (monthly data from February 2000 to February 2010) [14]. *FOB* Free on board

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