Process Engineering Economic Evaluation of the Ethanol Extraction of Cottonseed: Preliminary Analysis

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This preliminary analysis was undertaken to determine if the operations being developed for the aqueous ethanol extraction of cottonseed oil are economical and whether further research of this process should be pursued. Results of the conversion of hypothetical hexane extraction plants to ethanol extraction, in the plant capacity range of 300–600 tons of cottonseed flakes/day and operating 150–350 days annually, show that two unconventional operations, namely, chill-separation of miscella exiting the extractor and reduction of oil in recycled ethanol by reverse osmosis, require less energy and are less expensive than conventional alternatives. However, additional work is needed to determine the overall efficiency of an alcohol process as compared to a conventional hexane process.

KEY WORDS: Alternate solvent, cost study, cottonseed, economics, ethanol, extraction.

Research has been initiated to develop an economical process to extract oil and to remove undesirable constituents, such as gossypol and aflatoxin, from cottonseed with a solvent from a renewable source. Benefits will include a safer solvent than hexane, less dependence on petroleum, increased value and new markets for the meal product (1), and a potential supply of gossypol (2). The studies so far have shown ethanol to be a potentially suitable solvent for removal of oil, and a process for that purpose is being developed (3–6). Research into the additional problems of gossypol and aflatoxin removal are also being pursued.

Engineering research into novel processes must first consider if these processes have the potential to compete economically with conventional ones before significant research effort to develop them has been expended. Early in the research stage of the ethanol extraction work it became apparent that the departures from conventional solvent extraction, needed to accomplish the objective, mandated an overall preliminary economic evaluation of the ethanol oil-extraction process. The first step of the economic evaluation was the development of a computer model that has been used to generate the material balances used in this study (7). The economic evaluation provides a means of viewing the process and the interactions of the various unit operations. For example, the economic evaluation of solvent extraction identified the need for in-depth study of marc desolventization because of higher residual solvent in the marc. Equally important, this study also provides a means to identify those unit operations that contribute significantly to the cost of the process. As a result, refinements of those high-cost unit operations are underway that will have the greatest impact on the economics. This research, along with the computer model, will allow a more accurate determination of performance and thereby enable a more precise, more economical selection of equipment.

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Conversion to aqueous ethanol extraction has been evaluated primarily to the extent that research has verified and supported its practicality. To a lesser extent, in the application of anticipated research results (*i.e.*, those not yet achieved but whose exploratory work shows promise for achievement), we have consulted with experts and conservatively have used results that are more easily achievable and that are only a fraction of what is potentially achievable.

This is an economic study of the conversion of hypothetical commercial cottonseed hexane extraction plants of three capacities to aqueous ethanol extraction. The novel ethanol extraction process described in this evaluation is based on previous work (4–6) and on more recent unpublished work.

PROCESS

Figure 1 is a flow sheet of the process. Cottonseed flakes of 0.15 mm (0.006 in.) to 0.20 mm (0.008 in.) thickness are dried to *ca.* 2% moisture on a continuous conveyor in a tunnel-type dryer by passing 82.2°C air downward at a velocity of 91.4 m/min (300 ft/min) through a 0.05 m (2 in.) depth of flakes for 5.2 min. Removal of moisture is necessary to avoid dilution of the ethanol, which significantly reduces its effectiveness as an extraction solvent. Recent modifications to the process suggest that a flake thickness of 0.25 mm (0.010 in.) and a moisture of 4% are sufficient for extraction. An analysis of the modified process will be reported in the near future. It should be pointed out that while we have not determined what the ultimate moisture of the meal is after desolventization, it would be possible to adjust its moisture content by addition of water.

From the dryer, the flakes are fed to a jacketed 10-stage countercurrent vertical loop-type extractor equipped with a 20-mesh bottom screen to retain the flakes. The flakes are extracted with lean miscella (99% aqueous ethanol, 1% oil by weight) near its boiling point (75°C) for 48 min to a residual lipids level in flakes of 0.8%. Temperature is important in the alcohol process because oil rapidly becomes less soluble in aqueous ethanol as the temperature of the mixture decreases from the boiling point of the ethanol. In the last extractor stage, the flakes are washed with pure aqueous ethanol recovered from desolventization of the marc. The solvent-to-flake ratio is 3.5:1, double to triple that for hexane extraction.

The miscella from the extraction operation, which is ethanol at 75°C saturated with cottonseed oil (*ca.* 13% by weight), is pumped to the chiller-separator. Tower cooling water and propylene glycol, the latter chilled by ammonia, are used, respectively, in a series of two thin-plate heat exchangers to lower the temperature of the miscella to 10°C, thereby reducing the solubility of the oil in the ethanol and improving separation. The chilled miscella is fed to a centrifuge and separated into two phases—a lean miscella phase containing 3% oil and an oil-gum emulsion phase.

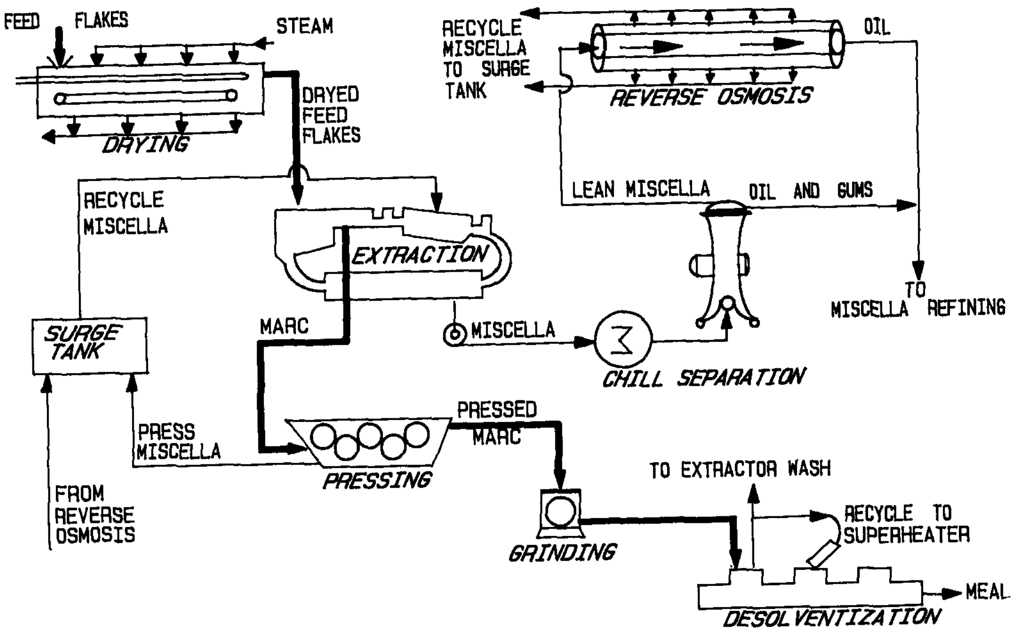

FIG. 1. Flow sheet for aqueous ethanol extraction of cottonseed.

Then the lean miscella is pumped to a reverse osmosis system where the oil content is reduced to less than 1%, thus making it possible for the lean miscella to be recycled to the extractor to achieve a residual oil level in final meal of 0.8%. The oil and gums containing 4% ethanol are sent to the miscella refining system. In refining, sodium hydroxide is added to the oil-gum emulsion at ambient conditions to remove free fatty acids, resulting in the formation of soapstock. This mixture is then centrifuged to separate the oil from the foots. The oil then goes to further refining. The foots are either returned to the marc before desolventization or sent to further processing for separation or detoxification of gossypol and aflatoxin. The separation of these toxic constituents from the foots, or preferably at an earlier stage of the process, is the subject of current research.

The marc from the extraction step, composed of 35% solids by weight, is pressed in a continuous belt-filter press operated at 75°C until it contains *ca.* 50% solids, which is equivalent to pressing out half of the ethanol carried out of the extractor. The pressed miscella is recycled to the surge tank.

The pressed marc from the filter press is ground and, along with the foots from miscella refining (which include 4% ethanol by weight) are fed to the desolventizer. Ethanol vapor is superheated to 149°C and 1 atm in a steam-operated heater. This superheated vapor is then used to vaporize the residual ethanol on the flakes, leaving the extracted flakes, gums, soapstock, and other foots as the meal. The meal, containing 41% protein by weight after equilibration, may be sold as ruminant feed. The ethanol vapor from the desolventizer is divided into two streams. Part of the ethanol vapor is condensed in a countercurrent condenser by the recycle miscella from the chiller-separator as it is recycled to the extractor wash. An additional condenser is necessary to condense completely the ethanol vapor to liquid at 75°C. The remainder of the ethanol vapor from the desolventizer is recycled through the superheater so that it can be used in the desolventizer again.

MATERIAL BALANCE

The material balance (Fig. 2) has been generated from experimental equilibrium data (7). The 73 kg of meal containing 43% protein and 2.7% moisture equilibrates to 76.g kg of meal containing 41% protein and 7.2% moisture.

PLANTS

Conversion was studied for hypothetical hexane extraction plants with daily capacities of 300, 420 and 600 tons of cottonseed flakes into the extractor, operating for 150, 200, 250, 300 and 350 days annually in Mississippi and Texas. The study was focused on four unit operations involved in conversion: The drying of flakes before extraction; the chill-separation of the miscella from the extractor for separation of oil and solvent, rather than using energy-intensive evaporation and stripping as is done in hexane extraction; reverse osmosis for reducing the oil content of the recycle ethanol; and the pressing-grinding of the marc from the extractor prior to desolventization.

Because this study shows that modification of conventional hexane desolventization equipment would be required, further experimental work needs to be done in this area to make a comprehensive economic evaluation possible. This is also the case for miscella refining.

In consideration of the long-term objective to reduce the gossypol and aflatoxin contents for production of edible cottonseed protein, materials of fabrication most generally used in conversion for surfaces in contact with

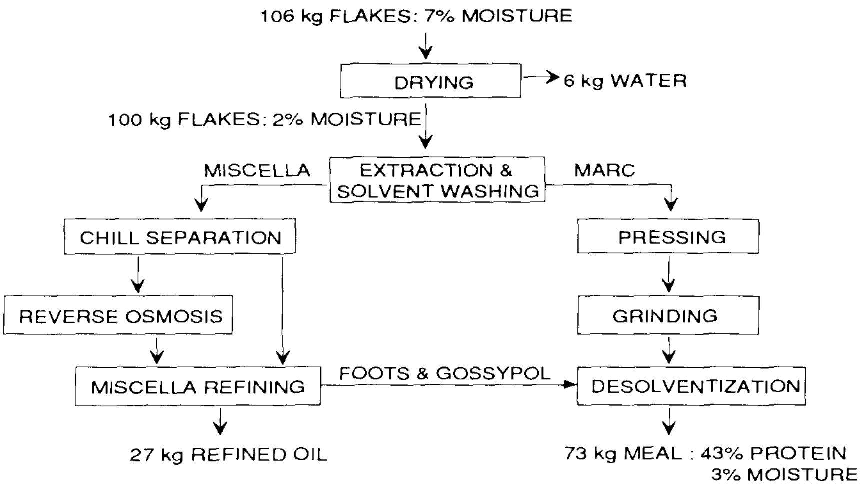

FIG. 2. Material balance: 100 kg cottonseed flakes into extractor.

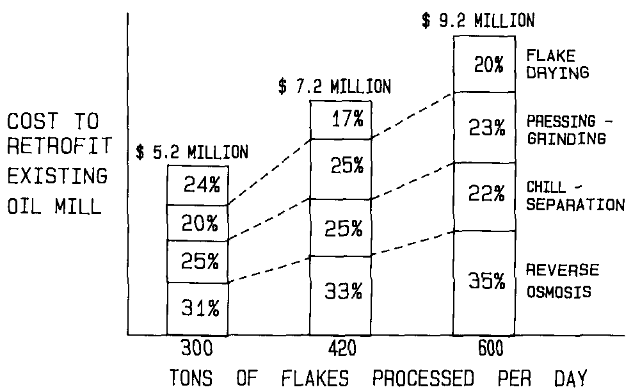

FIG. 3. Fixed capital investment for conversion to ethanol extraction.

the product include 304, 316 and 329 stainless steels and carbon steel with FDA-approved coatings.

All conversions include process equipment, piping, insulation, instrumentation and controls; service facilities, *e.g.*, a steam boiler and an ammonia refrigeration system; an enlargement of the process building; an addition to the boiler house; and an extension of yard improvements.

FIXED CAPITAL INVESTMENT

Fixed capital investment (Fig. 3) for the four unit operations is estimated to be $5.2 million for the 300 ton/day plant, $7.2 million for the 420 ton/day plant, and $9.2 million for the 600 ton/day plant. These estimates are based on the purchased costs of process equipment and service facilities quoted by equipment manufacturers.

For the plant sizes studied, the percentage of fixed capital investment attributable to reverse osmosis (Fig. 3) is consistently the highest of the four operations studied. A membrane flux of 0.53 m^3/m^2/day [15 gal/sq-ft/day (gfd)] was conservatively claimed in the evaluation, which is only half what might be achievable. Consequently, reverse osmosis costs could be only half those shown. Experts in membrane processing in the food industry have advised in a private communication that a flux of 0.53–1.1 m^3/m^2/day (15–30 gfd) is achieved in many industrial reverse osmosis applications. For either case, from current costs obtained from equipment manufacturers, this study shows that reverse osmosis is less expensive than evaporation or any other unit operation requiring the consumption of steam for vaporization of the ethanol.

MANUFACTURING COSTS AND GENERAL EXPENSES

Figures 4 and 5 show non-capital costs per ton vs. the weight of whole cottonseed for the three sizes of plants considered. In reading these Figures, note that 1 ton of whole seed corresponds to about 0.61 ton of flakes or meats. Additional manufacturing costs and general expenses (Fig. 4) resulting from conversion are indicated to be as much as $24.30/ton of cottonseed processed when operating the 300 ton/day plant for 150 days annually to as little as $12.15/ton when operating the 600 ton/day plant 350 days annually. General expenses are those items such as administrative costs, distribution and marketing costs and financing costs. Excluding general expenses, this corresponds to additional manufacturing costs (Fig. 5) of as much as $16.64/ton of cottonseed processed when operating the 300 ton/day plant 150 days annually to as little as $8.93/ton when operating the 600 ton/day plant 350 days annually.

An examination of chill-separation costs show that operation to be less expensive than evaporation. This has been made possible by using cooling tower water to cool the miscella out of the extractor from 75°C down to 43.3°C before chilling with propylene glycol, thereby reducing refrigeration requirements 38%. Manufacturing costs and general expenses for reverse osmosis are the most expensive of the four unit operations included in the cost study.

DISCUSSION

The extraction of cottonseed with ethanol requires several additional operations not needed in the conventional hex-

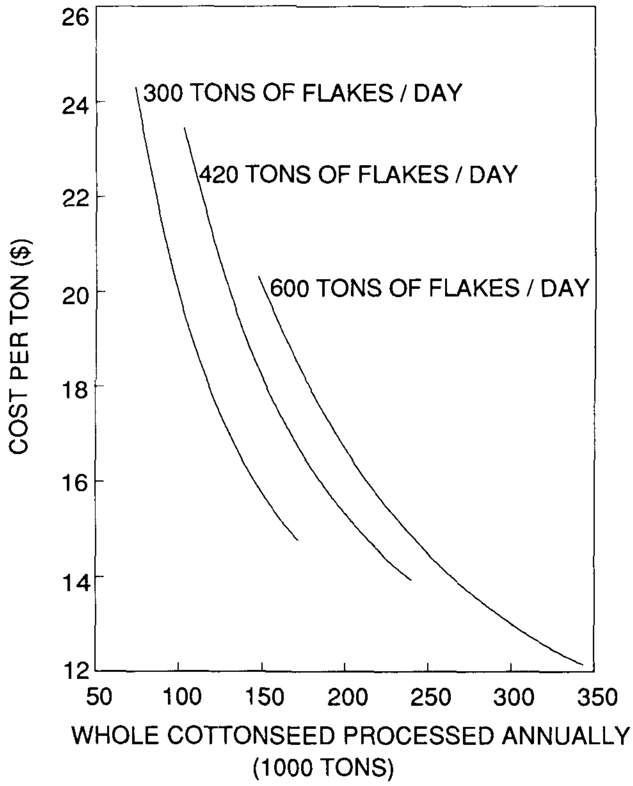

FIG. 4. Additional manufacturing costs plus general expenses after conversion to ethanol extraction.

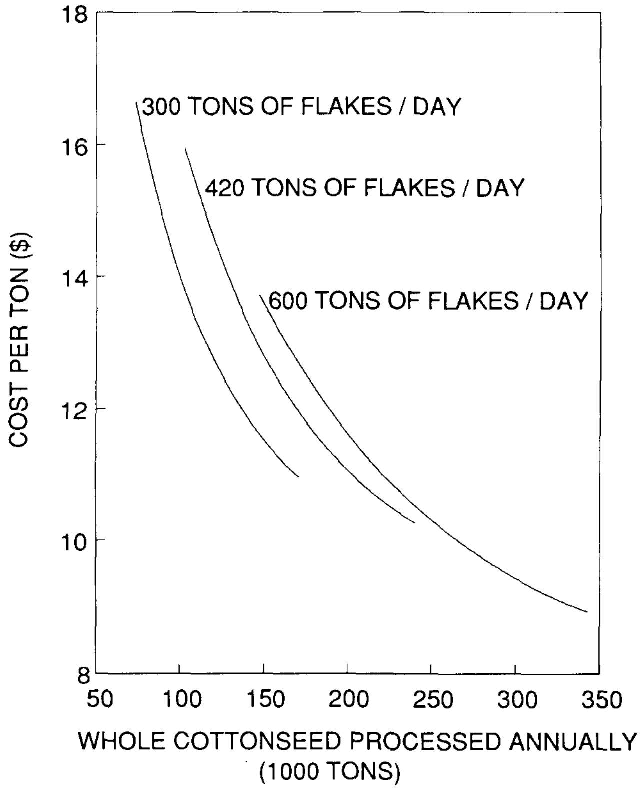

FIG. 5. Additional manufacturing costs after conversion to ethanol extraction.

ane operation. In this study we have considered the costs associated with these additional operations. Of the four operations considered, two of them, reverse osmosis and chill-separation, would replace conventional solvent recycle operations. Consequently, the additional costs associated with these new operations are mitigated by the removal of the conventional operations that they replace. However, the other two operations, flake drying and pressing-grinding, have no conventional counterpart. Because these last two operations make up 43% of the retrofit cost (Fig. 3), process modifications that avoid them are desirable. Extrusion of the seed before extraction has been found to dramatically increase drainage in hexane extractions. Incorporating this process in an ethanol extraction might do away with the need for pressing the flakes after extraction. The maximum increased cost of $24.30/ton for this alcohol extraction process, which also removes aflatoxin, could potentially be justified when it is considered that there is a penalty of as much as $35/ton on aflatoxin-contaminated seed (Johnson, LA., 1991, private communication).

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