

PROCESS FOR THE PRODUCTION OF FATTY ACID ESTERS

The present invention relates to a process for the production of fatty acid esters. More particularly, it relates to a process for the production of detergent fatty acid esters. In a further arrangement the present invention relates to a process for the production of fatty alcohols.

Fatty alcohols, or higher alcohols as they are sometimes designated, are monohydric aliphatic alcohols containing six or more carbon atoms which are derived either from natural sources or are synthesised from petroleum feedstocks. They are often classified by their market usage. As the primary end use of primary alcohols containing between about 6 and about 11 carbon atoms is the production of plasticiser esters, such alcohols are often termed plasticiser alcohols. For higher alcohols containing, for example, from about 11 up to about 20 carbon atoms, the major use is the production of synthetic detergents, hence such alcohols are often termed detergent alcohols. The distinction between plasticiser alcohols and detergent alcohols is somewhat arbitrary and detergent alcohols may have from 10 carbon atoms. In addition, there is some production of phthalate esters from a C₁₃ "oxo" alcohol and also some production of, for example, non-ionic surfactants from C₈ to C₁₀ alcohols.

Although there are some natural products which contain esters which can be hydrogenated to produce alcohols in the plasticiser range, these are more usually produced synthetically from petroleum feedstocks by, for example, the so-called "oxo" process, a process which is also termed oxonation or hydroformylation. Detergent range alcohols, on the other hand, are typically produced by hydrogenation of low molecular alkyl esters of fatty acids. Such esters can be produced by transesterification of natural triglycerides or by esterification of the fatty acids obtained by hydrolysis of the triglycerides. Examples of triglycerides which can be used as raw materials include natural oils, such as coconut oil, rape seed oil, and palm oils, and animal fats such as lard, tallow, and fish oils. As such natural raw materials contain mixtures of triglycerides, the alcohol products obtained upon hydrogenation are also mixtures of n-

alkanols of differing molecular weight. Such mixtures of alkanols are acceptable for production of detergents without prior separation of the alkanols one from another.

One process for carrying out the esterification is described in US 5536856 the contents of which are incorporated herein by reference. In this process, the esterification of the fatty acid is carried out in a column reactor having a plurality of esterification trays, each having a predetermined liquid hold-up and containing a charge of a solid esterification catalyst. The fatty acid flows down the column reactor against an upflowing lower alkyl alcohol vapour stream such as methanol. The supplied alcohol is relatively dry and water of esterification is removed from the top of the column in the vapour stream. The product ester is recovered from the sump of the reactor. As the liquid flows down the trays it encounters progressively drier alcohol which drives the ester reaction towards 100% conversion. The ester may then be fed to a polishing reactor operated under liquid phase conditions. Once produced, these esters can be utilised as biodiesel or they can be hydrogenated to the desired alcohols.

An alternative process for the production of fatty alcohols is described in US5157168. In this process the fatty alcohols are produced by hydrogenation of lower alkyl esters, particularly methyl esters, of fatty acids derived from natural triglycerides, under conditions which minimize formation of by-product alkanes and ethers. The hydrogenation is followed by refining of any resulting ester-containing hydrogenation product to produce a substantially ester free fatty alcohol.

Whatever process is used for the esterification an excess of alkanol is generally used. Since water is produced during the esterification reaction and it is desirable to separate the excess alkanol from the water generated in the esterification.

One prior art process for achieving this separation is illustrated schematically in Figure 1. Fatty acid is fed in line 1 to the esterification reactor 2. In the reactor 2 the downflowing acid is reacted with upflowing alkanol which has been added to the reactor 2 in line 6. The ester product is removed from the reactor 2 in line 3.

The excess alkanol removes most of the water present as a gaseous overhead stream 4. This is then fed to an alkanol recovery distillation column 5 where the water and alkanol are separated. Reflux is returned to the distillation column 5 in line 9 and lights are purged in line 8. Alkanol is removed in line 6 and returned to the esterification reactor 2. Fresh alkanol is added in line 7.

Waste water is removed in line 10. This waste water stream will always also contain some or all of residual alkanol, alkyl ester and fatty acid. The bulk of the fatty acid and methyl esters can be recovered in a simple separation as they are immiscible with water. However, even after this separation, the water stream will still contain residual organic content that requires the stream to be treated by a waste effluent treatment unit.

This process does suffer from various disadvantages and drawbacks. First, useful feedstocks and products such as alkanol, fatty acids and esters are lost to the process thereby reducing the economics of the system. However, the main problem is that water treatment processes are required to reduce the chemical oxygen demand (COD) of the effluent stream.

COD is commonly used to indirectly measure the amount of organic compounds in water. The basis for the COD test is that nearly all organic compounds can be fully oxidized to carbon dioxide with a strong oxidizing agent under acidic conditions. The International Organization for Standardization describes a standard method for measuring COD in ISO 6060. Many governments impose regulations regarding the maximum COD allowed in wastewater before it can be returned to the environment. As environmental concerns increase, the requirement to reduce the COD further makes treatment plants for effluent streams increasingly costly to construct and operate.

The growth in demand for naturally derived detergent range alcohols such as that derived from coconut or palm kernel has driven a desire to provide an improved process for manufacturing the alcohols and for handling the effluents produced.

The problem associated with the prior art process such as that illustrated in Figure 1 has been solved by the surprising finding that when the wastewater stream is subjected to a second distillation step, the heavy fatty acid and alkyl ester components are removed as part of the overhead stream giving a waste water stream having a lower COD.

Thus according to the present invention there is provided a process for the production of fatty acid esters comprising:

- (a) subjecting a fatty acid or fatty acid mixture to esterification with a lower alkanol in an esterification reactor maintained under esterification conditions to form a stream comprising the corresponding lower alkyl ester or esters and recovering same;
- (b) recovering an overhead stream from the esterification reactor comprising unreacted lower alkanol, water, and a heavy organic content comprising one or more of lower alkyl ester and/or acids;
- (c) subjecting the recovered overhead stream to distillation in a first distillation column operating under first distillation conditions to separate a portion of the lower alkanol from a stream comprising the water, the remaining lower alkanol and the heavy organic content;
- (d) recovering the stream comprising the water, the remaining lower alkanol and the heavy organic content; and
- (e) subjecting the stream comprising the water, the remaining lower alkanol and the heavy organic content to distillation in a second distillation column operating under second distillation conditions to provide an overhead stream comprising lower alkanol and at least a major portion of the heavy organic content and a water stream having a reduced heavy organic and alkanol content.

Without wishing to be bound by any theory it is believed that an azeotrope may be formed between the alkanol and the heavy organic content in the second distillation zone thereby allowing at least a major portion of the heavy organic content to be separated from the water. This is surprising in view of the high boiling points of the components of the heavy organic content.

The presence of the second distillation column and the separation that occurs therein, provides various advantages. One benefit is that there is an optimal recovery of alkanol. Another benefit is the greater recovery of fatty material. More importantly the removal of the COD value from the waste water effluent is optimised.

The waste water recovered may be of sufficient purity to enable the water to be recycled. Where the process of the present invention is carried out on the same site where the raw oil or fat is split, some or all of the waste water may be recycled to the reactor in which the splitting occurs.

The term "lower alkyl" means C₁- to C₄-alkyl, including methyl, ethyl, n-propyl, isopropyl, n-butyl, iso-butyl and sec-butyl. The preferred lower alkyl radical is methyl. Similarly the term "lower alkanol" means C₁- to C₄-alkanols, including methanol, ethanol, n-propanol, iso-propanol, n-butanol, iso-butanol, and sec-butanol. Methanol is the preferred lower alkanol.

The term "fatty acids" means linear saturated, unsaturated or polyunsaturated aliphatic acids, such as linear alkyl, alkenyl, or hydroxyalkenyl carboxylic acids containing from about 6 to about 26 carbon atoms, preferably about 10 to about 20 carbon atoms. Examples of such fatty acids are decanoic acid (capric acid), dodecanoic acid (lauric acid), tetradecanoic acid (myristic acid), pentadecanoic acid, hexadecanoic acid (palmitic acid), heptadecanoic acid (margaric acid), octadecanoic acid (stearic acid or isostearic acid), octadecenoic acids (oleic acid, linoleic acid or linolenic acid), hydroxyoctadecenoic acid (ricinoleic acid), eicosanoic acid (arachidic acid) and docosanoic acid (behenic acid). Mixtures of fatty acids are of especial importance as raw materials from which the lower alkyl fatty acid esters used as starting material in the hydrogenation step are prepared. Such mixtures of acids can be obtained by hydrolysis of naturally occurring triglycerides such as coconut oil, rape seed oil, palm oils, tallow, lard and fish oils. If desired, such mixtures of acids can be subjected to distillation to remove lower boiling acids having a lower boiling point than a chosen temperature and thus produce a "topped" mixture of acids, or to remove higher boiling acids having a boiling point higher than a second chosen temperature and thus produce

a "tailed" mixture of acids, or to remove both lower and higher boiling acids and thus produce a "topped and tailed" mixture of acids.

In a preferred process of the present invention esterification of the fatty acid or fatty acid mixture with the lower alkanol (e.g. methanol) is effected by a procedure in which the fatty acid or fatty acid mixture and lower alkanol are passed in countercurrent flow through a column reactor provided with a plurality of esterification trays mounted one above another, each adapted to hold a predetermined liquid volume and a charge of solid esterification catalyst thereon, liquid downcomer means associated with each esterification tray adapted to allow liquid phase to pass down the column reactor from that esterification tray but to retain solid esterification catalyst thereon, and vapour upcomer means associated with each esterification tray adapted to allow vapour to enter that esterification tray from below and to agitate the mixture of liquid and solid esterification catalyst on that tray, in which the fatty acid or fatty acid mixture is supplied in liquid phase to the uppermost one of said plurality of esterification trays whilst the lower alkanol is supplied in vapour form beneath the lowermost one of said plurality of esterification trays, in which vapour comprising lower alkanol and water of esterification is recovered from an upper part of the column reactor, and in which a lower alkyl fatty acid ester or ester mixture is recovered from a lower part of the column reactor.

In such a procedure the water content of the lower alkanol vapour supplied to the column reactor should be less than about 5 mole % and the number of esterification trays and the reaction conditions should be selected so that the stream of lower alkyl fatty acid ester or esters has a low acid content of less than about 1 mole %, calculated on a lower alkanol free basis, and an ester content, also expressed on an alkanol free basis, of at least about 99 mole %.

Any suitable reaction conditions can be used for the esterification in the esterification zone. The esterification conditions used in the column reactor will normally include use of elevated temperatures up to about 160°C, for example a temperature in the range of from about 80°C to about 140°C, preferably in the range of from about 100°C to about

125°C. Such operating temperatures will be determined by factors such as the thermal stability of the esterification catalyst, the kinetics of the esterification reaction and the vapour temperature of the lower alkanol fed to the base of the column reactor at the relevant inlet pressure. Typical operating pressures at the vapour inlet of the column reactor range from about 0.1 bara to about 25 bara. A liquid hourly space velocity through the column reactor in the range of from about 0.1 hr⁻¹ to about 10 hr⁻¹, typically from about 0.2 hr⁻¹ to about 2 hr⁻¹, may be used.

The column reactor has a plurality of esterification trays. Although two or three trays may suffice in some cases, it will typically be necessary to provide at least about 5 up to about 20 or more esterification trays in the column reactor. Typically each esterification tray is designed to provide a residence time for liquid on each tray of from about 1 minute up to about 120 minutes, preferably from about 5 minutes to about 60 minutes.

The fatty acid or fatty acid mixture is supplied in liquid form to an upper part of the column reactor or in admixture with lower alkanol, in solution in recycled ester product, or in solution in an inert solvent or diluent therefor. It is possible to pre-react the lower alkanol and the fatty acid or fatty acid mixture prior to introduction to the column reactor. The resulting reaction mixture contains a mixture of lower alkyl fatty acid ester or ester mixture, water, and lower alkanol.

The solid esterification catalyst may be a granular ion exchange resin containing SO₃H and/or COOH groups. Macroreticular resins of this type are preferred. Examples of suitable resins are those sold under the trade marks "Amberlyst", "Dowex", "Dow" and "Purolite", such as Amberlyst 13, Amberlyst 66, Dow C351 and Purolite C150.

Different solid esterification catalysts may be used on different trays of the column reactor. Moreover different concentrations of solid esterification catalyst can be used on different trays.

The charge of solid particulate or granular esterification catalyst on each tray is typically sufficient to provide a catalyst liquid ratio on that tray corresponding to a resin

concentration of at least about 0.2% w/v for example, a resin concentration in the range of from about 2% w/v to about 20% w/v, preferably 5% w/v to 10% w/v, calculated as dry resin. Sufficient catalyst should be used to enable equilibrium or near equilibrium conditions to be established on the tray within the selected residence time at the relevant operating conditions. On the other hand not so much catalyst should be used on each tray that it becomes difficult to maintain the catalyst in suspension in the liquid on the tray by the agitation produced by the upflowing vapour entering the tray from below. For a typical resin catalyst a resin concentration in the range of from about 2% v/v to about 20% v/v, preferably 5% v/v to 10% v/v may be used.

The particle size of the catalyst should be large enough to facilitate retention of the catalyst on each tray by means of a screen or similar device. However, as larger catalyst particle sizes are more difficult to maintain in suspension and have lower geometrical surface area per gram, it is expedient to use not too large a catalyst particle size. A suitable catalyst particle size is in the range of from about 0.1 mm to about 5 mm.

In an alternative arrangement, the reaction may be autocatalysed.

One or more wash trays may be provided above the esterification trays in order to prevent loss of product, solvent and/or reagents from the column reactor.

The process of the invention utilises the vaporous stream of the lower alkanol to carry away water of esterification produced in the esterification reactor but without carrying with it significant quantities of the fatty acid or acids or of the lower alkyl fatty acid ester or esters. The carryover of fatty acid or acids or of the lower alkyl fatty acid ester or esters may be greater with a shorter alkyl chain length of the fatty acid or acids. The overhead stream may include lower alkanol dialkyl ether. In general the overhead stream recovered from the esterification will include: from about 5 to about 40 wt% water, preferably from about 15 to about 30 wt%; from about 50 to about 90 wt% alkanol, preferably from about 60 to about 85 wt% alkanol, from about 0.5 to about 20 wt% dialkyl ether, preferably from about 1 to about 10 wt% dialkyl ether; from about

0.01 wt% to about 1 wt% of fatty acid; and from about 0.01 wt% to about 1 wt% of fatty ester.

The overhead stream is passed to a first distillation column where a portion of the alkanol is separated. The separated alkanol portion is preferably recycled to the esterification reactor. The lower the water content of the lower alkanol vapour that is supplied to the esterification column, the further towards 100% conversion to ester the esterification equilibrium reaction can be driven and the lower the residual acidity of the ester containing product recovered from the bottom of the column reactor will be. However, a balance may often have to be struck between the cost of providing, for example, a substantially dry lower alkanol for vaporisation into the column reactor, on the one hand, and the cost of providing and operating any additional downstream processing facilities that may be required to upgrade the ester product to the required quality if a less dry alkanol is used. This will vary from lower alkanol to lower alkanol and will depend upon the interaction between water and lower alkanol (e.g. azeotrope formation) and its effect upon alkanol/water separation. As discussed above, where an alkanol/water azeotrope is present it may be necessary to provide an additional column to split the azeotrope by the use of, for example, pressure swing distillation or a membrane system in order to attain an alkanol stream of suitable dryness. In any case, the water content of the lower alkanol vapour supplied to the reactor is less than about 5 mole%, and even more preferably is less than about 1 mole%.

The first distillation column may be of any suitable configuration. In one arrangement the distillation column will operate at a pressure of from about 0.1 to about 5 bara. The top temperature may be in the region of about 40° to about 110°C and may be about 70 to about 80°C. The reboiler temperature may be in the region of about 80° to about 150°C and may be about 110 to about 120°C. In one arrangement the distillation column may be fabricated from stainless steel. It will generally have a plurality of trays or a structured packing. A shell and tube condenser may be used for the overheads. The overhead stream will comprise non-condensable dialkylether, nitrogen and lower alkanol.

Generally the amount of lower alkanol separated in the first distillation column for recycle to the esterification reaction will be in the region of about 95 wt% to about 99.8 wt%.

The lower alkanol may be taken from any suitable point in the first distillation column. In one arrangement, it may be taken as a side draw.

The column will generally include a reflux. A purge of lights, which may be mostly dialkyl ether, may also be taken.

A stream will be recovered from at or towards the bottom of the first distillation column. This stream will contain water, the remaining lower alkanol and the heavy organic content. The amount of lower alkanol present in this stream will depend on the amount of lower alkanol removed for return to the esterification zone.

Where a lower alkanol/water azeotrope is present, the stream may be subjected to a process to break the azeotrope. This process may be the application of pressure swing distillation or by application of membrane to remove any residual water.

This stream can then be passed to the second distillation column. The second distillation column may be of any suitable configuration. It may be of the same or different configuration to the first distillation column.

In one arrangement, the second distillation column may be operated to have a top temperature of about 40°C to about 110°C and may be about 60°C to about 100°C and a reboiler temperature of about 80°C to about 150°C and may be about 90°C to about 120°C.

The first and second distillation columns will generally be separate columns. However, it is possible that these may be configured as separate zones in the same shell.

The lower alkanol and heavy organic content separated from the water will be removed from the second distillation column in the overhead stream. The water will be removed from the bottom of the second distillation column. The overhead stream may be returned to the esterification reactor.

It will be understood that it is desirable that the amount of alkanol and heavy content be reduced as much as possible however, the water received may still contain some alkanol and/or heavy organic content. Thus the reference to the water having a reduced heavy organic content means that the water stream contains less alkanol and/or heavy organic content than the stream entering the second distillation column. In one arrangement it will be reduced from about 0.55 wt% total organics to about 0.06 wt% or less.

It will be understood that references to an overhead stream means a stream taken from the top of the column or from an upper area thereof. Similarly the reference to a stream being taken from the bottom of the column means a stream taken from the base of the column or from a lower area thereof.

Steam may be supplied to the second distillation column to strip the alkanol and fatty components. This can be freshly added stream or stream generated from within the process.

The water stream having the reduced heavy organic and alkanol may be recycled to a fat splitting section.

The present invention will now be described by way of example with reference to the accompanying figures in which:

- Figure 1 is a schematic representation of a process according to the prior art;
- Figure 2 is a schematic representation of the process of the present invention; and
- Figure 3 is a schematic representation of the process in a second distillation column in which stream stripping is used.

It will be understood by those skilled in the art that the drawings are diagrammatic and that further items of equipment such as reflux drums, pumps, vacuum pumps, temperature sensors, pressure sensors, pressure relief valves, control valves, flow controllers, level controllers, holding tanks, storage tanks, and the like may be required in a commercial plant. The provision of such ancillary items of equipment forms no part of the present invention and is in accordance with conventional chemical engineering practice.

A simplified version of a flow scheme of the present invention is illustrated schematically in Figure 2. A fatty acid or mixture of fatty acid is supplied in line 11 before being subjected to esterification in the reactor 12. The acid is reacted with upflowing alkanol added to the esterification reacts in lines 16 and 17. The product of the esterification reaction is withdrawn in line 13. The product of the esterification may be subjected to any subsequent treatment, polishing and the like.

The excess alkanol removes most of the water present as a gaseous overhead stream 14. This is then fed to a first distillation column 15 where a proportion of the alkanol is separated and removed from the first distillation column 15 in line 16 and returned to the esterification reactor 12. Fresh alkanol is added in line 17. Reflux from the top of the first distillation column 15 is returned in line 19 and lights are purged in line 18.

The water stream, which comprises water, remaining alkanol and the heavy organic content is removed from the bottom of the first distillation column 15 and passed in line 20 to the second distillation column 21. In this column at least a major portion of the remaining alkanol and heavy organic content is removed as overheads in line 22 and the water, which has a reduced content of alkanol and heavy organics, is removed in line 23. In an alternative arrangement, the alkanol and heavy organic content removed as overheads in line 22 may be returned to the esterification reactor 12.

An alternative arrangement for the second distillation column 21 is illustrated in Figure 3. The bottom stream from the first distillation column is fed to the second distillation

column 21 via line 20. Steam is added in line 25 to assist the stripping of the remaining alkanol and heavy organic content which leaves the second distillation column in line 23. A reflux system 24 will generally be located at the top of the column 21. The water having a reduce content of alkanol and heavy organics is removed in line 23.

The present invention will now be described with reference to the following examples.

Example 1

The experiment was carried out using a laboratory scale 1 inch glass distillation column. The feed point was located in the middle of the column with four pieces of structured packing located above and below. A three litre three necked round bottomed flask heated via an isomantle was used as the reboiler, with the temperature of the isomantle controlled by the skin temperature of the flask. Two thermocouples were attached to the top and bottom surface sections of the column and the column was insulated to minimise any heat losses. A vapour travelling up the column was condensed and then collected as a liquid in the reflux drum. Off take and reflux return rates were then adjusted to maintain a liquid level.

2013 g of a feed comprising 96.85 wt% water, 3.00 wt% methanol, 0.10 wt% methyl laurate and 0.05 wt% lauric acid was charged to the round bottomed flask and then heated to 100°C to generate vapour traffic. The reflux back to the column was set at approximately ten times that of the overheads take off. Ten overhead fractions ranging between 20 and 25g were taken.

The results are set out in Table 1.

Table 1

Time	Temperatures (°C)			Reflux Ratio	Comments	Analysis (wt%)			Acid by Titre
	Skin	Pot	OH's			MeOH	C12 Ester	Water	
0	199	97	63	10.0	Starting to take off				
25	195	98	63	12.0	Fraction 1	96.584	0.087	3.329	

130	195	100	63	10.0	Fraction 2	99.970		0.030	0.061
230	195	100	64	10.0	Fraction 3	99.977		0.023	0.064
420	195	100	96	8.0	Fraction 4	62.740	0.271	36.974	0.036
455	200	100	99	9.0	Fraction 5	15.584	0.021	84.395	
550	200	100	99	9.0	Fraction 6	3.175	0.006	96.819	
715	200	99	99	11.0	Fraction 7	0.530	0.019	99.451	
815	200	99	99	11.0	Fraction 8	0.231	0.046	99.723	
935	200	99	99	10.0	Fraction 9	0.228	0.009	99.762	
1055	200	99	99	10.0	Fraction 10	0.172	0.021	99.807	

Example 2

Example 1 was repeated, but pot samples were also removed. The final pot sample was also analysed for chemical oxygen demand (COD).

For the methanol water test work the overheads stream was analysed for water using a Karl Fischer Coulometric Analyser. The methanol content in the bottoms stream was determined by Gas Chromatography (GC), using flame ionisation detection and a constant injection volume. For the methanol, methyl laurate and water test work all streams were analysed by GC for methanol and methyl laurate and also by titration for lauric acid using 0.01 molar potassium hydroxide solution and phenolphthalein indicator.

The results are set out in Table 2.

Table 2

Time	Temperatures (°C)			Reflux Ratio	Comments	Overheads Analysis (wt%)			
	Skin	Pot	OH's	Ratio		MeOH	C12 Ester	Water	Acid by Titre
0	200	98	63	10.0	Starting to take off				
40	200	98	63	10.0	Fraction 1	89.701	1.532	7.956	2.312
90	200	99	66	10.0	Fraction 2	99.617	0.050	0.233	0.100
195	200	99	96	10.0	Fraction 3	51.926	0.154	47.852	0.000
260	200	100	98	12.0	Fraction 4	7.864	0.014	92.063	0.000
345	200	100	99	9.0	Fraction 5	1.925	0.001	98.050	
400	200	100	99	10.0	Fraction 6	0.404	0.000	99.596	
465	200	100	99	14.0	Fraction 7	0.637	0.011	99.351	

540	200	100	99	11.0	Fraction 8	0.313	0.019	99.570	
610	200	100	99	9.0	Fraction 9	0.014		99.986	
750	200	100	99	9.0	Fraction 10	0.006		99.994	
Time		Pot Analysis (wt%)							
0		MeOH	C12 Ester	Water					
40		1.736	0.007	98.24					
90		0.893		99.08					
195		0.07	0.003	99.91					
260		0.011		99.99					
345		0.001		100					
400				99.99					
465				99.98					
540				100					
610				99.99					
750				100					
COD of final pot contents		13		mgL-1 O2					

Example 3

A COD model was used to predict the reduction of cod values for the present invention.

The results are illustrated in Table 3.

Table 3

Decanter feed composition:						Feed	molesO2/	O2 req.	
		wt%				kmol/h	mol feed	kmol/h	mol wt
MeOH		0.025		CH3OH		0.0088	1.5	0.0132	32.042
C10 acid		0.0125		C10H20O2		0.0008	14	0.0114	172.27
C12 acid		0.205		C12H24O2		0.0115	17	0.1961	200.324
C14 acid		0.025		C14H28O2		0.0012	20	0.0247	228.37
C16 acid		0.005		C16H32O2		0.0002	23	0.0051	256.43
C18 acid		0.0025		C18H36O2		0.0001	26	0.0026	284.48
C10 ester		0.0125		C11H22O2		0.0008	15.5	0.0117	186.297
C12 ester		0.205		C13H26O2		0.0108	18.5	0.1994	214.351
C14 ester		0.025		C15H30O2		0.0012	21.5	0.0250	242.405
C16 ester		0.005		C17H34O2		0.0002	24.5	0.0051	270.459
C18 ester		0.0025		C19H38O2		0.0001	27.5	0.0026	298.514
water		99.475		H2O		62.2271			18.016
								0.4968	
Mass flow =		1127	kg/h						
Density =		950	kg/m3						
Vol flow =		1.18632	m3/h						
		1186.32	l/h						
Total O2 required =			0.4968	kmol/h					
			15.8981	kg/h					

	COD =		13401.2	mg/l					
Assumed Decanter water product composition									
		wt%				kmol/h	mol feed	kmol/h	mol wt
MeOH		0.025		CH ₃ OH		0.0088	1.5	0.0132	32.042
C10 acid		0.0125		C10H ₂₀ O ₂		0.0008	14	0.0114	172.27
C12 acid		0.03		C12H ₂₄ O ₂		0.0017	17	0.0287	200.324
C14 acid		0		C14H ₂₈ O ₂		0.0000	20	0.0000	228.37
C16 acid		0		C16H ₃₂ O ₂		0.0000	23	0.0000	256.43
C18 acid		0		C18H ₃₆ O ₂		0.0000	26	0.0000	284.48
C10 ester		0.0125		C11H ₂₂ O ₂		0.0008	15.5	0.0117	186.297
C12 ester		0.03		C13H ₂₆ O ₂		0.0016	18.5	0.0292	214.351
C14 ester		0		C15H ₃₀ O ₂		0.0000	21.5	0.0000	242.405
C16 ester		0		C17H ₃₄ O ₂		0.0000	24.5	0.0000	270.459
C18 ester		0		C19H ₃₈ O ₂		0.0000	27.5	0.0000	298.514
water		99.89		H ₂ O		62.4867			18.016
								0.0942	
Total O ₂ required =			0.0942	kmol/h					
			3.01541	kg/h					
	COD =		2541.83	mg/l					

The following Table 4 compared the measured results with the predicted results confirming that the prediction is reasonable.

Table 4

%MeOH Measured analytically	Measured COD	Predicted COD (MeOH only)	Predicted COD (MeOH + 0.03% C12 Acid + 0.12% C12 Ester)
0.02	4150	285	4207
1.09	23300	15512	18435
1.02	11490	14516	18438
0.35	13680	4981	8903
0.72	16900	10247	14169
2.17	34800	30862	34804

It can be seen from the above examples that the overhead include high boiling C₁₂ ester which has a boiling point of 261°C and C₁₂ acid which has a boiling point of 299°C even though the pot temperature in the batch distillation does not go above 100°C. Thus, as

distillation proceeds, not only is the alkanol removed but the organic levels are reduced to low COD values which are acceptable in wastewater.

The following Tables 5 to 8 exemplify the use of alkanols other than methanol and show the removal of a partial of the heavy organics as well as the alkanol from the water.

Table 5

Summary of ethanol, water, ethyl laurate and lauric acid distillation

Fraction	Overheads temperature / °C	Overheads mass / g	Analysis / wt%			
			EtOH	Ethyl Laurate	Lauric Acid	Water
Overheads						
1	79	26.6	56.537		0.149	38.813
2	95	21.54	61.953		0.110	36.809
3	96	28.55	44.145	0.230	0.103	53.538
4	97	24.95	10.645		0.091	88.984
5	97	33.72	0.857		0.077	98.697
6	97	37.75	0.209		0.123	97.889
7	97	20.19	0.093		0.084	99.705
8	97	127.96	0.038		0.073	99.351
9	97	120.71			0.070	99.831
10	97	72			0.072	99.680
11	97	50.45			0.063	99.800
Pot sample						
Feed			3.020	0.100	0.050	96.830

Table 6

Summary of ethanol, water, ethyl palmitate and palmitic acid distillation

Fraction	Overheads temperature / °C	Overheads mass / g	Analysis / wt%			
			EtOH	Ethyl Palmitate	Palmitic Acid	Water
Overheads						
1	75	34.76	57.960		0.223	40.057

2	96	24.99	53.940		0.020	46.592
3	96	17.45	42.086		0.017	57.182
4	97	33.58	6.276		0.009	93.815
5	97	58.91	0.323		0.024	99.588
6	97	34.21	0.137			99.823
7	97	124.2	0.048			99.918
8	97	58.4	0.022			99.936
9	97	45.7	0.013			99.942
10	97	26.2	0.012			99.945
Pot sample						
Feed			3.000	0.120	0.050	96.830

Table 7

Summary of butanol, water, butyl laurate and lauric acid distillation

Fraction	Overheads temperature / °C	Overheads mass / g	Analysis / wt%			
			1-BuOH	Butyl Laurate	Lauric Acid	Water
Overheads						
1 - top layer	90	41.08	62.877		0.384	35.889
1 - bottom layer			7.563		0.016	92.221
2 - top layer	90	15.29	62.868		0.025	36.997
2 - bottom layer			6.906		0.008	93.017
3 - top layer	89	20.36	64.748		0.025	35.153
3 - bottom layer			6.915		0.005	93.033
4 - top layer	97	21.49	62.331	0.128	0.020	36.657
4 - bottom layer			6.477		0.006	93.490
5	97	23.49	1.437		0.017	98.504
6	97	64.5	0.112			99.874
7	97	120.5	0.039			99.950
8	97	83.4	0.016	0.015		99.958
9	97	44.5	0.009	0.139		99.653
10	97	37.67	0.009	0.335		99.584
11	97	68.1	0.009	0.006		99.968
12	97	40.5	0.011	0.006		99.953
Pot Sample						
Feed			3.000	0.100	0.050	96.840

Table 8

Summary of propanol, water, propyl palmitate and palmitic acid distillation

Fraction	Overheads temperature	Overheads mass	Analysis / wt%			
	/ °C	/ g	1-Propanol	Propyl Palmitate	Palmitic Acid	Water
Overheads						
1	84	16.04	58.347	0.032	0.490	38.618
2	84	15.56	71.245	0.013	0.087	28.122
3	84	22.56	73.676	0.006		26.255
4	84	20.56	67.009			32.932
5	96	17.89	71.396		0.041	27.889
6	96	37.33	2.295		0.037	97.624
7	96	74.61	0.183		0.033	99.784
8	97	24.86	0.112		0.043	99.845
9	97	57.97	0.066	0.232	0.090	98.391
10	97	41.4	0.037		0.033	99.896
11	96	239.8	0.066	0.113	0.018	99.433
12	96	58.6	0.016		0.010	99.956
13	96	36.07	0.015	0.006	0.016	99.931
14	96	28.1	0.017	0.005	0.017	99.926
Pot Sample						
Feed			2.96	1.830	0.700	94.510

Claims

1. A process for the production of fatty acid esters comprising:
 - (a) subjecting a fatty acid or fatty acid mixture to esterification with a lower alkanol in an esterification reactor maintained under esterification conditions to form a stream comprising the corresponding lower alkyl ester or esters and recovering same;
 - (b) recovering an overhead stream from the esterification reactor comprising unreacted lower alkanol, water, and a heavy organic content comprising one or more of lower alkyl ester and/or acids;
 - (c) subjecting the recovered overhead stream to distillation in a first distillation column operating under first distillation conditions to separate a portion of the lower alkanol from a stream comprising the water, the remaining lower alkanol and the heavy organic content;
 - (d) recovering the stream comprising the water, the remaining lower alkanol and the heavy organic content; and
 - (e) subjecting the stream comprising the water, the remaining lower alkanol and the heavy organic content to distillation in a second distillation column operating under second distillation conditions to provide an overhead stream comprising lower alkanol and at least a major portion of the heavy organic content and a water stream having a reduced heavy organic and alkanol content.
2. A process according to Claim 1 wherein the esterification of the fatty acid or fatty acid mixture with the lower alkanol is effected in a column reactor in which the fatty acid or fatty acid mixture and lower alkanol are passed in countercurrent flow.
3. A process according to Claim 1 wherein the water content of the lower alkanol supplied in the esterification reaction is less than 5 mole % and the reaction conditions are selected so that the stream of lower alkyl fatty acid ester or esters has a low acid content of less than 1 mole %, calculated on a lower alkanol free

basis, and an ester content, also expressed on an alkanol free basis, of at least 99 mole %.

4. A process according to any one of Claims 1 to 3 wherein the esterification reaction is carried out in a temperature up to 160°C, or a temperature in the range of from 80°C to 140°C, or a temperature of from 100°C to 125°C.
5. A process according to any one of Claims 1 to 4 wherein the esterification reaction is carried out a pressure from 0.1 bara (10 kPa) to 25 bara (2500 kPa).
6. A process according to any one of Claims 1 to 5 wherein the esterification is carried out with a liquid hourly space velocity through the column reactor in the range of from 0.1 hr⁻¹ to 10 hr⁻¹, or from 0.2 hr⁻¹ to 2 hr⁻¹.
7. A process according to any one of Claims 1 to 6 wherein the esterification is carried out in the presence of a solid esterification catalyst which is a granular ion exchange resin containing SO₃H and/or COOH groups.
8. A process according to any one of Claims 1 to 6 wherein the esterification is autocatalysed.
9. A process according to any one of Claims 1 to 8 wherein the first distillation column is operated at a pressure of from 0.1bara (10 kPa) to 5 bara (500kPa).
10. A process according to any one of Claims 1 to 9 wherein the first distillation column has a top temperature in the region of 40 to 110°C.
11. A process according to any one of Claims 1 to 10 wherein the first distillation column has a reboiler temperature in the region of 80 to 150°C.
12. A process according to any one of Claims 1 to 11 additionally comprising a step of breaking any lower alkanol/water azeotrope present after step (c).

13. A process according to Claim 12 wherein the azeotrope is broken by the application of pressure swing distillation or a membrane to remove any residual water.
14. A process according to any one of Claims 1 to 13 wherein the second distillation column has a top temperature of from 40°C to 110°C.
15. A process according to any one of Claims 1 to 14 wherein the second distillation column has a reboiler temperature of 80°C to 150°C.
16. A process according to any one of Claims 1 to 15 wherein steam is added to the second distillation column.
17. A process according to any one of Claims 1 to 16 wherein the water stream having the reduced heavy organic and alkanol content is recycled to a fat splitting section.