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CARBOXYLIC ACIDS, MANUFACTURE

Carboxylic acids having 6–24 carbon atoms are commonly known as fatty acids. Shorter-chain acids, such as formic, acetic, and propionic acid, are not classified as fatty acids and are produced synthetically from petroleum sources (see Acetic acid; Formic acid and derivatives; Oxo process). Fatty acids are produced primarily from natural fats and oils through a series of unit operations. Clay bleaching and acid washing are sometimes also included with the above operations in the manufacture of fatty acids for the removal of impurities prior to subsequent processing.

The composition of common fats and oils are found in Table 1. The most predominant feedstocks for the manufacture of fatty acids are tallow and grease, coconut oil, palm oil, palm kernel oil, soybean oil, rapeseed oil, and cottonseed oil. Another large source of fatty acids comes from the distillation of crude tall oil obtained as a by-product from the Kraft pulping process (see Tall oil; Carboxylic acids, fatty acids from tall oil).

1. Manufacture of Fatty Acids from Natural Fats and Oils

There are essentially four steps or unit operations in the manufacture of fatty acids from natural fats and oils: (*1*) batch alkaline hydrolysis or continuous high pressure hydrolysis; (*2*) separation of the fatty acids usually by a continuous solvent crystallization process or by the hydrophilization process; (*3*) hydrogenation, which converts unsaturated fatty acids to saturated fatty acids; and (*4*) distillation, which separates components by their boiling points or vapor pressures. A good review of the production of fatty acids has been given (1).

1.1. Hydrolysis

Saponification or hydrolysis involves converting the fat or oil (a triglyceride) to a fatty acid and glycerol. This can be done in a number of ways including Twitchell splitting, autoclave batch splitting, continuous high pressure splitting, and enzymatic splitting. The feed should be of quality commensurate with the quality of fatty acid desired. Also, the feedstock generally is given a pretreatment such as clay bleaching or acid washing prior to splitting that allows hydrolysis to take place more efficiently and to give a higher quality product in some cases (2).

Twitchell splitting is an acid catalyzed hydrolysis that uses sulfuric acid as the catalyst along with surface-active agents such as petroleum sulfonates or sulfonated oleic and naphthenic acids (Twitchell reagent). The splitting is carried out in acid-resistant vessels using about 60% fat or oil, 40% water, 0.5% sulfuric acid, and 1% petroleum sulfonate. The advantages of Twitchell splitting are its simple process and equipment required compared with the other methods. The disadvantages are darker colored products that usually contain traces of sulfur-containing materials, disposal of the aqueous acid layers, inefficient use of heat, and long cycle times. Twitchell splitting is used on a relatively small scale (3, 4).

Table 1. Fatty Acid Composition (%) and Significant Properties of Important Fats and Oils

Name	Chain length	Double bonds	Cotton oil	Coconut oil	Palm kernel oil	Corn oil	Palm oil	Castor oil	Rapeseed oil (low erucic)	Rapeseed oil (high erucic)	Soybean oil	Sunflower oil	Herring oil	Sardine oil	Tallow	Tall oil
caproic acid	C_6	0		0–1	[a]											
caprylic acid	C_8	0		5–10	3–6											
capric acid	C_{10}	0		5–10	3–5											
lauric acid	C_{12}	0		43–53	40–52	0–1							[a]			
myristic acid	C_{14}	0	0–2	15–21	14–18		0–2		[a]	[a]	[a]	[a]	5–10	4–6	1–6	
palmitic acid	C_{16}	0	17–29	7–11	6–10	8–19	30–48	2–3	3–6	0–5	7–12	3–10	11–16	9–11	20–37	1–3
stearic acid	C_{18}	0	1–4	2–4	1–4	0–4	3–6	2–3	0–3	0–3	2–6	1–10	0–3	1–3	6–40	0–1
arachidic acid	C_{20}	0	0–1				0–1		0–2	0–2	0–3	0–1			[a]	0–1
behenic acid	C_{22}	0	[a]						[a]	0–2	[a]	0–1				
palmitoleic acid	C_{16}	0–2							[a]	[a]	[a]	0–1	5–12	10–15	1–9	
oleic acid	C_{18}	1	13–44	6–8	9–16	19–50	38–44	4–9	50–66	9–25	20–30	20–40	8–15	15–25	20–50	9–16
gadoleic acid	C_{20}	1	[a]						0–5	5–15	0–1	[a]				
erucic acid	C_{22}	1							0–5	30–60		[a]				
ricinoleic acid	C_{18}	1						80–87								
linoleic acid	C_{18}	2	33–58	1–3	1–3	34–62	9–12	2–7	18–30	11–25	48–58	50–70	2–4	3–8	0–5	20–32
linolenic acid	C_{18}	3				0–2			6–14	5–12	4–10	0–1	0–2	1–3	0–3	
unsaturated fatty acids	C_{20}	2–6							[a]	[a]			20–30	15–30	[a]	[a]
unsaturated fatty acids	C_{22}	2–6							[a]	0–2			10–28	15–20		
rosin acids																23–37
								Properties								
iodine value, g I_2/100 g			96–112	8–12	14–23	103–128	44–54	81–91	105–120	91–108	120–140	120–140	120–145	170–193	35–55	
saponification value, mg KOH/g			190–198	250–264	245–255	188–193	194–206	174–186	185–198	170–185	190–195	186–194	178–194	189–193	190–200	130–170

[a] Trace.

Autoclave batch splitting is generally carried out at 1.03–3.45 MPa (150–500 psig). By using certain metal oxides, such as zinc oxide, as catalysts, lower pressures can be used. In a typical batch the amount of water is about 30–60% of the fat weight. Headspace air is removed with steam to minimize oxidation and the autoclave is heated to the desired temperature and pressure. After 6–10 h a split of about 92% is obtained at the lower pressure whereas at higher pressures splits as high as 95% have been realized. After the desired degree of split is obtained the lower aqueous glycerol layer is separated from the upper fatty acid layer. If metal catalysts have been used the fatty acids are treated with sulfuric acid to decompose the soaps, and finally, any residual mineral acid is removed from the fatty acids with a hot water wash (5).

Continuous high pressure splitting was developed by Colgate-Emery and by Procter & Gamble (6, 7). Temperatures of 240–270°C are preferred, giving pressures of 4.8–5.2 MPa (700–750 psig). The splitting is carried out in a cylindrical-shaped tower, 18.3–24.4 m high and 0.51–1.22 m in diameter of 316 L stainless steel or 316 L cladding on carbon steel. Splitting coconut oil requires better corrosion resistance because of the shorter-chain fatty acids present. Corrosion-resistant linings such as Carpenter 20 Cb or Incoloy 825 can be utilized. The tower is operated with countercurrent flow, with water being introduced into the top part of the tower and fat at the bottom of the tower. The tower contains disengaging zones where fatty acids are collected at the top of the tower and aqueous glycerol (sweet water) at the bottom (Fig. 1). Heat is conserved by the use of heat exchangers that cool the existing fatty acid while heating the incoming water (steam); the exiting sweet water is cooled by the entering fat, which in turn is heated in the exchanger. Make-up heat is applied to the center of the tower (the hydrolysis section containing the continuous fat phase) using internal steam coils, electric heating, or direct superheated steam. The pressure in the tower is controlled by a backpressure valve in the fatty acid discharge line, whereas the fat/sweet water interface is controlled by the rate of sweet water discharged. About 98–98.5% split is usually obtained. The sweet water contains 10–15% glycerol and is purified in a series of steps involving removal of any dissolved salts, fat, and oil impurities, and then concentration by evaporation of water and/or distillation (8). In some continuous high pressure splitting units, approximately 0.05% of ZnO catalyst is used to speed the reaction rates and raise conversion to $>99.0\%$.

Enzymatic fat splitting was developed as a means to minimize energy costs and where a high order of specificity of splitting is desired. Enzymatic fat splitting is not currently carried out commercially in the United States but it is utilized in Japan where energy costs are high. Further details on enzymatic fat splitting are available (9).

1.2. Separation Techniques

Current methods for separating fatty acids are by solvent crystallization or by the hydrophilization process. Other methods that have been used in the past, or perhaps could be used in the future, are panning and pressing, solvent extraction, supercritical fluid extraction,the use of metal salts in assisting in separation, separations using urea complexes, and adsorption/desorption.

Panning and pressing is no longer used because of high labor costs, but it gave us the terms single-pressed, double-pressed, and triple-pressed stearic acid. These terms are widely used to denote the quality of stearic acid. These commercial “stearic” acids are actually mixtures of palmitic acid and stearic acid.

There are two commercial solvent crystallization processes. The Emersol Process, patented in 1942 by Emery Industries, uses methanol as solvent; and the Armour-Texaco Process, patented in 1948, uses acetone as solvent. The fatty acids to be separated are dissolved in the solvent and cooled, usually in a double-pipe chiller. Internal scrapers rotating at low rpm remove the crystals from the chilled surface. The slurry is then separated by means of a rotary vacuum filter. The filter cake is sprayed with cold solvent to remove free liquid acids, and the solvents are removed by flash evaporation and steam stripping and recovered for reuse (10).

When tallow fatty acids are the feed, stearic acid (actually 60/40 C16/C18) and oleic acids are the products. Solvent separation is also used to separate stearic acid from isostearic acid when hydrogenated monomer is the feed, and oleic acid from linoleic acid when using tall oil fatty acids as feed.

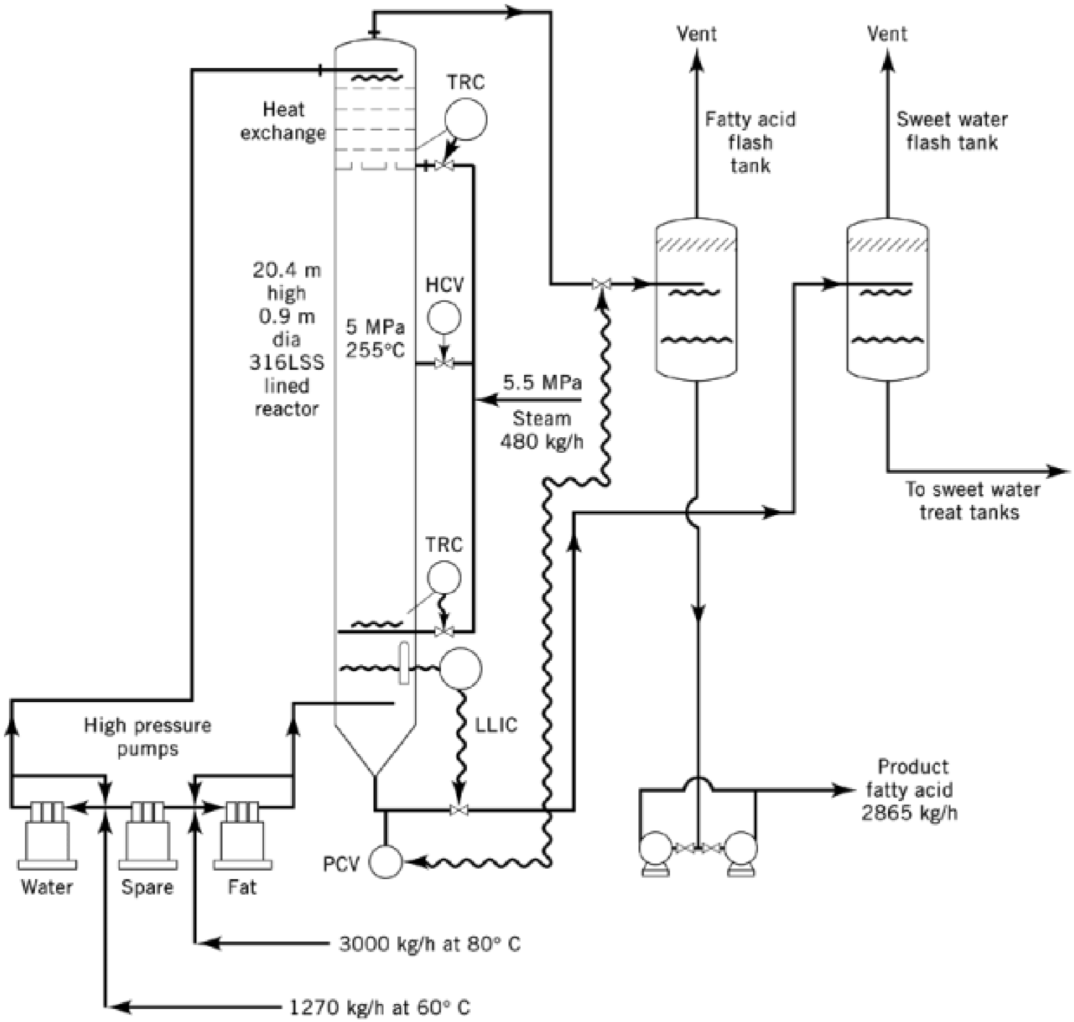

Fig. 1. Fat splitter. TRC, temperature recorder controller; LLIC, liquid level indicator controller; PCV, pressure control valve; and HCV, heat control valve. To convert MPa to psi, multiply by 145.

Several processes have been evaluated in attempts to lower the cost of the low temperature refrigeration required in the above solvent separation systems. The use of adiabatic cooling under vacuum to a temperature below the equilibrium temperature for crystallization has been described in which the resulting supercooled solution was transferred to a crystallization vessel where nucleation started and crystallization was reported to be complete within a few minutes. Spheroid-shaped crystals were obtained, which were more easily separated and washed than the usual crystals produced from other solvent processes (11). Two processes have been described in which crystallization occurs at a higher temperature. One uses methyl formate containing 5–10% water as solvent, where crystallization occurs about 15°C higher than when using methanol or acetone as solvent (12). The other process uses 2-nitropropane as solvent and gives a good separation of oleic acid from linoleic acid at -15°C instead of -30°C when using methanol or acetone as solvents (13).

To avoid the hazards and costs of using solvents, the Hydrophilization Process was developed by Henkel in Germany. The steps in the process include (*1*) cooling to obtain a slurry; (*2*) addition of aqueous solution of a wetting agent; (*3*) high speed agitation to form a dispersion of the solid fatty acid in the liquid; (*4*) addition of an electrolyte to stabilize the dispersion; and (*5*) separation of the solid fatty acid from the liquid fatty acid by means of a centrifuge. The quality of the products are generally not as good as when solvent separation is used (14–17).

Liquid–liquid extraction can be used to obtain high purity linoleic acid from safflower fatty acids or linoleic acid from linseed fatty acids using furfural and hexane as solvents (18). High purity linoleic acid has been obtained from sunflower fatty acids using a dimethylformamide and hexane solvent system (19).

Supercritical fluid extraction (SFE) has been investigated on oleochemical separations. This method uses a gas in the supercritical state, which generally means working at pressures of 8.3–48.5 MPa (1200–7000 psi), and at relatively low temperatures. Initial costs for commercial equipment can be quite high and will probably limit the use of SFE to higher priced specialty products (20). Methyl esters of Menhaden oil have been separated using supercritical CO_2 into docosa-hexaenoic acid [6217-54-5] (DHA) and eicosapentaenoic acid [10417-94-4] (EPA). These omega-3 fatty acids have been found to be important dietary factors, beneficial in reducing the development of atherosclerotic lesions. Accordingly, there is great interest in obtaining the polyunsaturated fatty acids present in fish oils in a more concentrated form (21).

Adsorption processes have recently been described to separate fatty acids into high purity products. Lauric acid was separated from myristic acid using crystalline silica as the adsorbent and was desorbed using a ketone such as acetone or methyl ethyl ketone (22). Another system, using cross-linked polystyrene as the adsorbent, separated a mixture of palmitic and stearic acids (23). Separation of saturated fatty acid, such as palmitic/stearic acid from an unsaturated fatty oleic acid, using a molecular sieve plus crystalline clay as the adsorbent, has been described (24). The desorbent was acetone. The separation of oleic acid from linoleic acid has been described using as an adsorbent either cross-linked polystyrene, or a molecular sieve-silicate as the adsorbent with a variety of desorbents listed (25). Separation of fatty acids from rosin acids can be accomplished using molecular sieves that have been modified with silicalite and a phosphorus-modified alumina. The preferred desorbents were methyl ethyl ketone–acetic acid or short-chain acids or esters with less than six carbon atoms (26). The separation of fatty acids from unsaponifiables has been carried out using a molecular sieve, comprising a crystalline silica with a silica to alumina ratio of at least 12, as the adsorbent. The desorbent was acetone (27).

1.3. Hydrogenation

In the manufacture of fatty acids hydrogenation is used to saturate the ethylenic linkage to produce a more saturated acid. Generally, the feedstocks for hydrogenation are (*1*) the solid fractions from crystallizations where the iodine value (IV) is 4–15 and the desired IV after hydrogenation may be as low as 0.5; (*2*) tallow fatty acids to give a rubber-grade stearic acid having an IV of about 5–12; (*3*) coconut, palm, and palm kernel fatty acids to give hydrogenated products, which are then fractionated to give saturated fatty acids of six- to sixteen-carbon atoms; (*4*) fish fatty acid, such as menhaden, and high erucic rapeseed fatty acids to give arachidic (C-20) and behenic (C-22) acids; and (*5*) castor oil fatty acids (ricinoleic) to give 12-hydroxystearic acid. These and other feedstocks can be partially hydrogenated to give products having iodine values of about 20–80 that have found use in specific areas of application. Monomer acids, by-products from the dimerization of unsaturated fatty acids (primarily tall oil fatty acids), are hydrogenated to give a mixture of primarily stearic and isostearic acid, which are then separated from solvent by crystallization (28).

Hydrogenations can be carried out in batch reactors, in continuous slurry reactors, or in fixed-bed reactors. The material of construction is usually 316 L stainless steel because of its better corrosion resistance to fatty acids. The hydrogenation reaction is exothermic and provisions must be made for the effective removal or control of the heat; a reduction of one IV per g of C_{18} fatty acid releases 7.1 J (1.7 cal), which raises the temperature 1.58°C. This heat of hydrogenation is used to raise the temperature of the fatty acid to the desired reaction temperature and is maintained with cooling water to control the reaction.

The size of a typical batch reactor is 18–23 t. Some reactors have a hydrogen recycle system, whereas in others the hydrogen is internally recycled by the use of a hollow agitator shaft that allows hydrogen to be drawn from the head space to below the impeller blades where it mixes with the liquid fatty acids. This recycle of the hydrogen gives better contact with the liquid and effective mixing does not have to depend solely on agitation. It is also an advantage to vent some of the hydrogen periodically, which allows removal of inert material and possibly catalyst poisons (29).

Continuous slurry reactors are generally either of one of two designs. One type uses a reactor loop, generally known as a Buss loop design; the other is a co-current hydrogen/fatty acid/catalyst system mainly marketed by Lurgi. Continuous slurry reactors are more popular in Europe, Asia, and South America than in the United States.

Fixed-bed reactors have been described in detail and their advantages and disadvantages listed (29). It is reported that only one manufacturer uses fixed-bed hydrogenation for fatty acids (29).

Dry reduced nickel catalyst protected by fat is the most common catalyst for the hydrogenation of fatty acids. The composition of this type of catalyst is about 25% nickel, 25% inert carrier, and 50% solid fat. Manufacturers of this catalyst include Calsicat (Mallinckrodt), Harshaw (Engelhard), United Catalysts (Süd Chemie), and Unichema. Other catalysts that still have some place in fatty acid hydrogenation are so-called wet reduced nickel catalysts (formate catalysts), Raney nickel catalysts, and precious metal catalysts, primarily palladium on carbon. The spent nickel catalysts are usually sent to a broker who sells them for recovery of nickel value. Spent palladium catalysts are usually returned to the catalyst supplier for credit of palladium value.

The most important reaction variables in the hydrogenation of fatty acids are temperature, pressure, agitation, catalyst loading, and catalyst addition. Temperature is normally in the range of 150 to 210°C. Below 150°C the nickel catalyst is not activated sufficiently; however, above 210°C some degradation may occur, which could inactivate the catalyst thus slowing the reaction rate. Pressures of industrial hydrogenations are about 2.08–3.47 MPa (300–500 psig). Pressures below 2.08 MPa require longer times to reach the same IV if the hydrogenation is carried out at 3.47 MPa. Pressure higher than 3.47 MPa has little or no effect on reaction rate. Increasing the agitation in the reactor increases the reaction rate. Determination of the optimum level of agitation is dependent on the configuration inside each specific reactor. Because the reaction is mass and diffusion limited, increasing the speed of the agitator increases the dispersion of catalyst in the fatty acid until it reaches a maximum that corresponds to a maximum reaction rate, assuming everything else is constant. Proper agitation also helps to keep the solid catalyst in suspension, to help maintain temperature control, and to bring the hydrogen in contact with the catalyst/fatty acid. Increasing the catalyst loading increases the reaction rate. However, when extremely large amounts of catalysts are used it may cause a rapid decrease in hydrogen concentration in the fatty acid resulting in a dehydrogenation reaction. The optimum time of catalyst addition is near the reaction temperature and immediately thereafter hydrogen is introduced. Prolonged exposure of a nickel catalyst to hot fatty acids leads to catalyst deactivation (29).

The quality of the feedstock is important since it affects not only the product quality but the rate of hydrogenation. Some of the impurities that affect the rate are sulfur, phosphorus, halides, polyethylene, and moisture. Impurities are usually removed by clay treatment or by distillation (30).

1.4. Distillation

Most fatty acids are distilled to produce high quality products having excellent color and a low level of impurities. Distillation removes odor bodies and low boiling unsaponifiable material in a light ends or heads fraction, and higher boiling material such as polymerized material, triglycerides, color bodies, and heavy decomposition products are removed as a bottoms or pitch fraction. The middle fractions sometimes can be used as is, or they

Table 2. Boiling Points of Fatty Acids,[a] °C

Pressure, kPa[b]	Caproic acid	Caprylic acid	Capric acid	Lauric acid	Myristic acid	Palmitic acid	Stearic acid	Oleic acid	Linoleic acid
0.133 kPa	61.7	87.5	110.3	130.2	149.2	167.4	183.6	177.6	178.5
0.267 kPa	71.9	97.9	121.1	141.8	161.1	179.0	195.9	189.5	190.1
0.533 kPa	82.8	109.1	132.7	154.1	173.9	192.2	209.2	202.6	202.8
1.07 kPa	94.6	121.3	145.5	167.4	187.6	206.1	224.1	217.0	216.9
2.13 kPa	107.3	134.6	159.4	181.8	202.4	221.5	240.0	232.9	232.6
4.27 kPa	120.8	149.2	174.6	197.4	218.3	238.4	257.1	250.6	250.0
8.53 kPa	136.0	165.3	191.3	214.6	236.3	257.1	276.8	270.3	269.7
17.1 kPa	152.5	183.3	209.8	234.3	257.3	278.7	299.7	292.5	291.9
34.1 kPa	171.5	203.0	230.6	256.6	281.5	303.6	324.8	317.7	317.2
68.3 kPa	192.5	225.6	254.9	282.5	309.0	332.6	355.2	346.5	346.5
101.3 kPa	205.8	239.7	270.0	298.9	326.2	351.5	376.1	364.9	365.2

[a] Ref. 31.

[b] To convert kPa to mm Hg, multiply by 7.5

can be fractionated (separated) into relatively pure materials such as lauric, myristic, palmitic, and stearic acids.

Because fatty acids, and especially unsaturated fatty acids, have limited stability when subjected to high temperatures, most distillations are carried out in continuous distillation columns as opposed to batch-type distillations. Almost all distillations are carried out under vacuum and sometimes with the injection of steam to further reduce the temperature at which the fatty acid will distill. Boiling points of various fatty acids at different pressures are shown in Table 2.

The crude fatty acid feed is usually preheated and degassed to remove air and water. If a vapor feed is desired it is then flashed in a vaporizer and fed to the distillation column as a vapor or vapor–liquid. A continuous distillation using one column will generally produce a satisfactory product if a suitable feedstock is used. If the feedstock contains large amounts of low boilers, color, and odor bodies, it is then preferable to use a two-column system. The low boilers and odor are removed as a heads fraction in the first column; the remaining material is removed at the bottom and then fed to a second column where a residue cut is taken off the bottom and the desired fatty acid fraction is removed as a side stream. It is possible to use a one-column system if the side stream is taken in the stripping section below the feed and taken as a vapor and condensed. Doing this will eliminate any high boilers in the side steam.

Most distillation columns in the past used sieve trays or bubble cap trays where the pressure drop could be substantial (1–2 mm/tray) or about 30–50 mm in a column of 20–30 trays. Because of this pressure drop the distillation temperature must be higher and this higher temperature causes some decomposition of the fatty acids. A packing called structured packing has low pressure drop per theoretical plate thus making it ideal for use in the distillation of heat-sensitive fatty acids. It is available throughout the world as either Flexipac or Mellapac (31). Wiped-film evaporators can be used in the distillation of heat-sensitive material because the contact time with a hot surface is extremely short. They have also been used in depitching crude tall oil (31).

Fatty acids are corrosive at high temperatures and selection of materials of construction for distillation systems is critical. Stainless steels with various contents of molybdenum have proved satisfactory. For example, 316 L has 2% Mo and is satisfactory for service up to 260°C; 317 L has 3% Mo and can be used satisfactorily up to 285°C, whereas 904 L can be used up to 310°C (31).

2. Synthetic Routes to Fatty Acids from Petroleum

These synthetic processes have been reviewed in detail (32).

2.1. Catalytic Oxidation for Straight-Chain Paraffinic Hydrocarbons

Synthetic fatty acids (SFA) are produced by Eastern European countries, Russia, and China using a manganese-catalyzed oxidation of selected paraffinic streams. The technology is based on German developments that were in use during World War II. The production volume in 1984 was estimated to be about 5.5×10^5 t/yr. The oxidation is highly exothermic and is carried out at about 105–125°C, mostly in continuous equipment.

2.2. Oxidation of Straight-Chain 1-Olefins

Oxidation of α-olefins has been thoroughly studied using ozone, peracids, nitric acid, chromic acid, and others.

2.3. Carboxylation/Oxidation of Straight-Chain 1-Olefins

Selective carboxylation of α-olefins to predominately straight-chain aldehydes is realized through specific catalyst systems and by careful control of reaction conditions. The aldehyde produced is then air-oxidized to the acid using a Mn catalyst. Heptanoic acid [111-14-8] and pelargonic acid [112-05-0] are produced commercially in this manner.

2.4. Carboxylation of Straight-Chain 1-Olefins

Carboxylation is the selective addition of CO and water to an olefin to give either a straight-chain or branched-chain acid. The use of specific catalysts and reaction conditions has given a straight-chain/branched-chain isomer ratio as high as 98:2. In spite of a one-step method to predominately straight-chain isomers of fatty acids, no commercialization using this route has yet occurred in the United States.

2.5. Oxidation of Straight-Chain Alcohols

Two methods have been developed. One uses an air oxidation catalyzed by a metal, eg, copper, platinum, etc, whereas the other is a caustic oxidation. Generally, however, fatty alcohols are priced higher on the world market than their corresponding fatty acids and, consequently, these conversions are uneconomical.

2.6. Branched-Chain Carboxylic Acids

Branched-chain acids such as 2-methylbutyric, 3-methylbutyric, isooctanoic, and isononanoic acids are produced by the oxo reaction, giving first the corresponding aldehyde, which is then oxidized to the acid. 2-Ethylhexanoic acid is produced by the aldol route from butyaldehyde in three steps: aldol condensation; hydrogenation of the carbon–carbon double bond; and oxidation of the branched-chain saturated aldehyde to 2-ethylhexanoic acid (see Carboxylic Acids, branched-chain acids).

2.7. Highly Branched Acids

These acids, called neoacids, are produced from highly branched olefins, carbon monoxide, and an acid catalyst such as sulfuric acid, hydrogen fluoride, or boron trifluoride. 2,2,2-Trimethylacetic acid (pivalic acid) is made from isobutylene and neodecanoic acid is produced from propylene trimer (see Carboxylic Acids, trialkylacetic acids).

3. Environmental Aspects

Environmental regulation in the oleochemical industry addresses pollution of air, surface, and groundwater, along with land pollution and solid waste disposal. This is administered by the Environmental Protection Agency (EPA) on the national level, an equivalent agency on the state level, and sometimes local agencies also deal with various aspects of pollution abatement.

In-plant controls are perhaps the best approach to eliminate waste generation and pollution problems, and many times good payback exists on recovery of products lost because of poor process controls. If the production department is responsible for the generation and in-plant control of wastes, this will help ensure that initial standards for water use and process loss are reasonable and that they are maintained (33).

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