

PROCESS FOR DEHYDRATING GLYCEROL TO ACROLEIN

The present invention relates to a process for manufacturing acrolein by dehydration of glycerol in the presence of molecular oxygen.

Acrolein is the simplest of the unsaturated aldehydes. It is also known as 2-propenal, acrylaldehyde or acrylic aldehyde. As a result of its structure, acrolein has high reactive power by virtue of the presence of its two reactive functions, which are capable of reacting individually or together. It is for this reason that acrolein finds many applications, especially as a synthetic intermediate. It is in particular a key intermediate for the synthesis of methionine, a synthetic protein used as an animal feed supplement, which has established itself as a substitute for fishmeal. Acrolein is a non-isolated synthetic intermediate of acrylic acid in the industrial production of acrylic acid by catalytic oxidation of propylene in the gas phase. The importance of the chemistry of acrylic acid and its derivatives is known. Acrolein also leads, via reaction with methyl vinyl ether followed by hydrolysis, to glutaraldehyde, which has many uses in leather tanning, as a biocidal agent in oil well drilling and during the processing of cutting oils, and as a chemical disinfectant and sterilizing agent for hospital equipment.

Acrolein is usually used as a synthetic intermediate of derivatives that are synthesized on the site of production to minimize the transportation of acrolein from the manufacturer to the client. The essential reason is linked to the toxicity of acrolein, which leads industrials to avoid the storage and transportation of this chemical product.

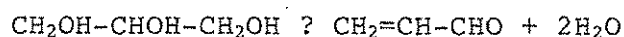
The most commonly used process for producing acrolein is based on the gas-phase catalytic oxidation reaction of propylene with atmospheric oxygen. The acrolein thus obtained may then be incorporated directly into an acrylic acid manufacturing process. When acrolein is used as starting material for the synthesis of methionine or for fine chemistry reactions, a purification section allows the removal of the reaction by-products, mainly carbon oxides, acrylic acid, acetic acid and acetaldehyde.

The production of acrolein is thus highly dependent on the propylene starting material obtained by steam cracking or catalytic cracking of petroleum fractions. This starting material, of fossil origin, furthermore contributes towards increasing the greenhouse effect. It thus appears necessary to have available an acrolein synthesis process that is not dependent on propylene as resource and that uses another starting material, which is preferably renewable. This process would be particularly advantageous for the synthesis of methionine, which might then be said to be "obtained from biomass". Specifically, during its use in animal feed, methionine is rapidly metabolized and the carbon dioxide expelled into the atmosphere contributes towards increasing the greenhouse effect. If acrolein is obtained from a renewable starting material, for example obtained from plant oil, the CO₂ emissions no longer enter into the process balance, since they compensate for the carbon dioxide used by the biomass for its growth; there is therefore no increase in the greenhouse effect. Such a process thus satisfies the criteria associated with the new concept of "green chemistry" within a more global context of durable development.

It has been known for a long time that glycerol can

lead to the production of acrolein. Glycerol (also known as glycerine) is derived from the methanolysis of plant oils at the same time as the methyl esters, which are themselves used especially as fuels or combustibles in diesel and domestic fuel oil. It is a natural product that has an "environmentally friendly" image, is available in large amount and may be stored and transported without difficulty. Many studies have been devoted to the financial upgrading of glycerol according to its degree of purity, and the dehydration of glycerol to acrolein is one of the routes envisaged.

The reaction involved for obtaining acrolein from glycerol is:



As a general rule, the hydration reaction is favoured at low temperatures, and the dehydration reaction is favoured at high temperatures. To obtain acrolein, it is thus necessary to use a sufficient temperature, and/or partial vacuum to shift the reaction. The reaction may be performed in the liquid phase or in the gas phase. This type of reaction is known to be catalysed by acids.

According to patent FR 695 931, acrolein is obtained by passing glycerol vapours at a sufficiently high temperature over salts of acids containing at least three acid functions, for instance phosphoric acid salts. The yields indicated are greater than 75% after fractional distillation.

In patent US 2 558 520, the dehydration reaction is performed in the gas/liquid phase in the presence of diatomaceous earths impregnated with phosphoric acid salts, suspended in an aromatic solvent. A degree of conversion of the glycerol into acrolein of 72.3% is obtained under these conditions.

The process described in patent application WO 99/05085 is based on a complex homogeneous catalysis, under a CO/H₂ atmosphere at a pressure of 20/40 bar and in the presence of a solvent such as an aqueous solution of sulfolane.

Chinese patent application CN 1 394 839 relates to a process for preparing 3-hydroxypropanaldehyde from glycerol. The acrolein produced as reaction intermediate is obtained by passing vaporized pure glycerol over a catalyst of potassium sulfate or magnesium sulfate type. The reaction yields are not given.

Patent US 5 387 720 describes a process for producing acrolein by dehydration of glycerol, in the liquid phase or in the gas phase over acidic solid catalysts defined by their Hammett acidity. The catalysts must have a Hammett acidity of less than +2 and preferably less than -3. These catalysts correspond, for example, to natural or synthetic siliceous materials, for instance mordenite, montmorillonite, acidic zeolites; supports, such as oxides or siliceous materials, for example alumina (Al₂O₃), titanium oxide (TiO₂), coated with mono-, di- or triacidic inorganic acids; oxides or mixed oxides such as gamma-alumina, the mixed oxide ZnO-Al₂O₃, or alternatively heteropolyacids. According to the said patent, an aqueous solution comprising from 10% to 40% of glycerol is used, and the process is performed at temperatures of between 180°C and 340°C in the liquid phase, and between 250°C and 340°C in the gas phase. According to the authors of the said patent, the gas-phase reaction is preferable since it enables a degree of conversion of the glycerol of close to 100% to be obtained, which leads to an aqueous acrolein solution containing side products. A proportion of about

10% of the glycerol is converted into hydroxypropanone, which is present as the major by-product in the acrolein solution. The acrolein is recovered and purified by fractional condensation or distillation. For a liquid-
5 phase reaction, a conversion limited to 15-25% is desired, to avoid excessive loss of selectivity. Patent US 5 426 249 describes the same gas-phase process for the dehydration of glycerol to acrolein, but followed by a hydration of the acrolein and a hydrogenation to lead to
10 1,2- and 1,3-propanediol.

The dehydration reaction of glycerol to acrolein is thus generally accompanied by side reactions leading to the formation of by-products such as hydroxypropanone, propanaldehyde, acetaldehyde, acetone, adducts of acrolein
15 with glycerol, glycerol polycondensation products, cyclic glycerol ethers, etc., but also phenol and polyaromatic compounds, which are the cause of the formation of coke on the catalyst. This results, firstly, in a reduction in the yield of and the selectivity towards acrolein, and
20 secondly in deactivation of the catalyst. The presence of by-products in the acrolein, such as hydroxypropanone or propanaldehyde, some of which are moreover difficult to isolate, necessitates separation and purification steps, which lead to high recovery costs for the purified
25 acrolein. Moreover, it is necessary to regenerate the catalyst very regularly in order to regain satisfactory catalytic activity.

The Applicant Company has sought to solve these problems and has found many advantages in using molecular
30 oxygen during the reaction for the dehydration of glycerol to acrolein. It has been observed, surprisingly, that supplying oxygen reduces the formation of aromatic compounds such as phenol, and of by-products originating

from a hydrogenation of dehydrated products, for instance propanaldehyde and acetone, but also from hydroxypropanone. The formation of coke on the catalyst is reduced. This results in inhibition of the deactivation of the catalyst and continuous regeneration of the catalyst. 5 Certain by-products are found to be present in markedly lower amount, which facilitates the subsequent purification steps.

10 One subject of the disclosure is thus a process for manufacturing acrolein by dehydration of glycerol in the presence of molecular oxygen.

In one aspect, there is provided a process for manufacturing acrolein by dehydration of glycerol in the presence of molecular oxygen, in an amount chosen so as 15 to be outside the flammability range at any point in the plant.

The molecular oxygen may be present in the form of air or in the form of a mixture of gases containing molecular oxygen. The amount of oxygen is chosen so as to be outside the flammability range at any point in the 20 plant. From Figure 4 of US patent application 2004/15012, the maximum oxygen content, in an acrolein/O₂/N₂ mixture is about 7% by volume in order to be entirely outside the flammability range. The oxygen content in the process 25 according to disclosure will generally be chosen so as not to exceed 7% relative to the mixture of gasses entering the reaction (mixture of glycerol/H₂/oxygen/inert gases). Preferably, the oxygen 30 content is less than 7% relative to the dry gas mixture leaving the reactor (mixture of acrolein/oxygen/inert gases).

The dehydration reaction is performed on acidic solid catalysts. The catalysts that are suitable are homogeneous or multi-phase materials, which are insoluble in the reaction medium, and which have a Hammett acidity, noted H_0 , of less than +2. As indicated in patent US 5 387 720, which refers to the article by K. Tanabe et al. in "Studies in Surface Science and Catalysis", Vol. 51, 1989, chap. 1 and 2, the Hammett acidity is determined by amine

titration using indicators or by adsorption of a base in the gas phase. The catalysts satisfying the acidity criterion H_0 less than +2 may be chosen from natural or synthetic siliceous materials or acidic zeolites; mineral supports, such as oxides, coated with mono-, di-, tri- or polyacidic inorganic acids; oxides or mixed oxides, or alternatively heteropolyacids.

The catalysts are advantageously chosen from zeolites, Nafion® composites (based on sulfonic acid of fluorinated polymers), chlorinated aluminas, phosphotungstic and/or silicotungstic acids and acid salts, and various solids of metal oxide type such as tantalum oxide Ta_2O_5 , niobium oxide Nb_2O_5 , alumina Al_2O_3 , titanium oxide TiO_2 , zirconia ZrO_2 , tin oxide SnO_2 , silica SiO_2 or silico-aluminate $SiO_2-Al_2O_3$, impregnated with acidic functions such as borate BO_3 , sulfate SO_4 , tungstate WO_3 , phosphate PO_4 , silicate SiO_2 or molybdate MoO_3 . According to the literature data, these catalysts all have a Hammett acidity H_0 of less than +2.

The preferred catalysts are sulfate zirconias, phosphate zirconias, tungsten zirconias, siliceous zirconias, sulfate titanium or tin oxides, and phosphate aluminas or silicas.

These catalysts all have a Hammett acidity H_0 of less than +2; the acidity H_0 may then vary within a wide range, up to values that may reach -20 in the scale of reference with the Hammett indicators. The table given on page 71 of the publication on acid-base catalysis (C. Marcilly) Vol. 1, published by Technip (ISBN No. 2-7108-0841-2) illustrates examples of solid catalysts in this acidity range. The content of molecular oxygen introduced into the reaction medium may depend on the nature of the catalyst used, its acidity and its capacity to form coke.

The reaction according to the invention may be performed in the gas phase or in the liquid phase, preferably in the gas phase. When the reaction is performed in the gas phase, various process technologies
5 may be used, i.e. a fixed-bed process, a fluidized-bed process or a circulating fluidized-bed process. In the last two processes, in a fixed bed or a fluidized bed, the regeneration of the catalyst may be separate from the reaction. It may take place ex situ, for example by
10 extraction of the catalyst and combustion in air or with a gaseous mixture containing molecular oxygen. In this case, the temperature and pressure at which the regeneration is performed do not need to be the same as those at which the reaction is performed. According to the process of the
15 invention, it may take place continuously in situ, at the same time as the reaction, given the presence of a small amount of molecular oxygen or of a gas containing molecular oxygen in the reactor. In this case, the regeneration is likened to an inhibition of deactivation
20 and takes place at the reaction temperature and pressure.

In the circulating fluidized-bed process, the catalyst circulates in two containers, a reactor and a regenerator. It is known that the dehydration reaction is endothermic, and energy must therefore be supplied to the
25 first container, whereas the regeneration consisting of the combustion of coke is exothermic, and heat must therefore be taken away from the second container. In the case of the circulating fluidized bed, the two systems may compensate each other: according to the process of the
30 invention, the regeneration of the catalyst under a flow of oxygen by combustion leads to heating of the catalyst and consequently supplies the energy required for the dehydration reaction when the heated catalyst returns into

the reactor. The residence time in each container depends on the rate of deactivation of the catalyst and on the level of coke formed on the catalyst. Specifically, a minimum level of coke is desirable in order to be able to
5 bring the solid to the correct temperature, and a maximum level of coke is necessary in order to prevent the solid from degrading by sintering during combustion.

The selection of the optimum process is made as a function of various criteria. The fixed-bed process has
10 the advantage of simplicity. The fluidized-bed processes make it possible to continuously discharge the spent catalyst and to permanently recharge fresh catalyst without stopping the production, with the possibility of being isothermic. The circulating fluidized-bed process
15 has the advantage of optimizing the reaction selectivity by permanently returning freshly regenerated catalyst into the reactor, while at the same time compensating for the energy exchange between the reactor and the regenerator.

According to one particular embodiment of the
20 invention, the process is performed in a reactor of the plate heat exchanger type. This reactor consists of plates forming between themselves circulation channels that can contain a catalyst. This technology has many advantages in terms of heat exchange, associated with high heat exchange
25 capacity. Thus, this type of reactor is particularly suitable for removing heat easily in the case of exothermic reactions, or for supplying heat in the start-up phases of reactions or in the case of endothermic reactions. More particularly, this reactor makes it
30 possible either to heat or to cool the catalyst. The heat exchange is particularly efficient with the circulation of a heat-exchange fluid in the system. The plates may be assembled in modules, which gives greater flexibility,

whether as regards the size of the reactor, its maintenance or the replacement of the catalyst. Systems that may be suitable for the process of the invention are, for example, the reactors described in documents EP 995 5 491 or EP 1 147 807, the content of which is incorporated by reference. These reactors are particularly suitable for the catalytic conversion of reaction media, specifically gaseous reaction media, such as those used in the present invention. The plate heat exchanger used for the 10 preparation of (meth)acrolein or (meth)acrylic acid via catalytic oxidation of C3 or C4 precursors, described in document US 2005/0020851, may also be suitable for the manufacture of acrolein via dehydration of glycerol, which is the subject of the present invention.

15 The dehydration of glycerol may also be performed in the liquid phase in a standard reactor for liquid-phase reaction on a solid catalyst, but also in a reactor of catalytic distillation type. Given the large difference between the boiling points of glycerol (280°C) and 20 acrolein (53°C), a liquid-phase process at a relatively low temperature that allows continuous distillation of the acrolein produced may also be envisaged. The reaction is permanently shifted, thus limiting the consecutive reactions on the acrolein in an equilibrium-shifted 25 continuous reactor.

The experimental conditions of the gas-phase reaction are preferably a temperature of between 250°C and 350°C and a pressure of between 1 and 5 bar. In the liquid phase, the reaction is preferably performed at a 30 temperature of between 150°C and 350°C and a pressure that may range from 3 to 70 bar. It has been observed that a lower temperature leads to a reduction in the degree of conversion of glycerol, but, at the same time, the

acrolein selectivity is increased. To avoid consecutive reactions and the formation of unwanted products, it is important to limit the residence time in the reactor; moreover, by increasing the residence time, it is also possible to have higher conversions. It is especially desirable to increase the contact time (residence time) of the reagents in the region of the catalyst in order to compensate for a decrease in the degree of conversion when a lower reaction temperature is used.

Glycerol is available inexpensively in the form of aqueous solutions. Advantageously, an aqueous glycerol solution with a concentration of between 10% and 50% and preferably between 15% and 30% by weight is used in the reactor. The concentration should not be too high, so as to avoid spurious reactions such as the formation of glycerol ethers or reactions between the acrolein produced and the glycerol. Moreover, the glycerol solution should not be too dilute on account of the energy cost involved in the evaporation of the aqueous glycerol solution. In any case, the concentration of the glycerol solution may be adjusted by recycling the water produced by the reaction. In order to reduce the glycerol transportation and storage costs, the reactor may be fed with concentrated solution of 40% to 100% by weight, dilution to the optimum content being performed by recycling some of the steam produced by the reaction and of the dilution water. Similarly, the recovery of heat at the reactor outlet may also allow the glycerol solution feeding the reactor to be vaporized.

Glycerol derived from the methanolysis of plant oils in basic medium may contain certain impurities such as sodium chloride or sulfate, non-glycerol organic matter, and methanol. The presence of sodium salts is in

particular detrimental to the catalytic dehydration reaction since these salts are capable of poisoning the acidic sites. A pretreatment of the glycerol by ion exchange may be envisaged.

5 Compared with the conventional process for preparing acrolein by selective oxidation of propylene, the acrolein produced according to the process of the invention may contain impurities of different nature or in different amount. According to the envisaged use, synthesis of
10 acrylic acid, synthesis of methionine or fine chemistry reactions, it may be envisaged to purify the acrolein according to the techniques known to those skilled in the art. More particularly, the by-products may be recovered and incinerated, thus producing vapour or energy. The
15 energetic upgrading of the by-products of the glycerol dehydration reaction furthermore makes it possible to greatly reduce the greenhouse-gas emissions of the process, compared with the conventional process, for which the CO₂ produced is derived from fossil carbon during the
20 incineration of the by-products.

The examples that follow illustrate the present invention without, however, limiting its scope.

Examples

25 In the examples, a tubular reactor consisting of a tube 85 cm long and with an inside diameter of 6 mm is used to perform the glycerol dehydration reaction in the gas phase at atmospheric pressure. This reactor is placed in a
30 heated chamber maintained at the reaction temperature, which is 300°C, unless otherwise indicated. The catalysts used are ground and/or pelletized to obtain particles of 0.5 to 1.0 mm. 10 ml of catalyst are loaded into the

reactor to form a catalytic bed 35 cm long. This bed is maintained at the reaction temperature for 5 to 10 minutes before introducing the reagents. The reactor is fed with an aqueous solution containing 20% by weight of glycerol at a mean feed flow rate of 12 ml/h, and with a flow rate of 0.8 l/h of molecular oxygen for the examples according to the invention. In this case, the O₂/vaporized glycerol/steam relative proportion is 6/4.5/89.5. The aqueous glycerol solution is vaporized in the heated chamber, and then passes over the catalyst. The calculated contact time is about 2.9 sec. The duration of a catalyst test is about 7 hours, which corresponds to about 80 ml of aqueous glycerol solution passed over the catalyst. After reaction, the products are condensed in a trap refrigerated with crushed ice.

Samples of the effluents are collected periodically. For each sample collection, the flow is interrupted and a gentle flow of nitrogen is passed through the reactor to purge it. The trap at the reactor outlet is then replaced, the nitrogen flow is stopped and the reactor is returned under a flow of reagent. The test is continued until appreciable deactivation of the catalyst is noted.

For each experiment, the total mass of products entering and leaving is measured, which allows a mass balance to be determined. Similarly, the products formed are analysed by chromatography. Two types of analysis are performed:

- an analysis by chromatography on a filled column (FFAP column 2 m*1/8") on a Carlo Erba chromatograph equipped with a TCD detector. The quantitative analysis is performed with an external standard (2-butanone);

- an analysis by chromatography on a capillary column

(FFAP column 50 m*0.25 mm) on an HP6890 chromatograph equipped with an FID detector with the same samples stored at -15°C.

The first method is particularly suitable for rapid
5 analysis of the products, and especially the yield of acrolein. The second method is used to have a more precise analysis of all the reaction by-products. Moreover, analyses by GC-MS or by chromatography after silylation were performed to confirm these results.

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The products thus quantified are the unreacted glycerol, the acrolein formed, and the by-products such as hydroxypropanone, acetaldehyde, propanaldehyde, acetone and phenol.

15 In the examples that follow, the glycerol conversion, the acrolein selectivity and the yields of the various products are defined as follows:

glycerol conversion (%) = $100 - \frac{\text{number of moles of glycerol remaining}}{\text{number of moles of glycerol introduced}} \times 100$;

20 acrolein yield (%) = $\frac{\text{number of moles of acrolein produced}}{\text{number of moles of glycerol introduced}} \times 100$;

acrolein selectivity (%) = $100 \times \frac{\text{number of moles of acrolein produced}}{\text{number of moles of glycerol reacted}}$.

25 The acetone or hydroxypropanone yield is calculated as for the acrolein yield:

acetaldehyde yield (%) = $\frac{2}{3} \times \frac{\text{number of moles of acetaldehyde produced}}{\text{number of moles of glycerol introduced}}$.

30 phenol yield (%) = $2 \times \frac{\text{number of moles of phenol produced}}{\text{number of moles of glycerol introduced}}$.

All the results are expressed as molar percentages relative to the glycerol introduced.

Example 1 (comparative and according to the invention)

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The catalyst used is a zeolite HZSM5 (Zeocat PZ - 2/54H 15% Aerosil - Ueticon). 10 ml, representing a mass of 6.41 g, were loaded into the reactor. The results are given in Table 1 below:

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Table 1

Cumulative glycerol introduced (g)	15	17	8	16	24
Addition of oxygen	no	no	yes	yes	yes
Glycerol conversion	79	64	96	88	83
Acrolein yield	39.1	24.3	40.6	40.9	37.5
Acrolein selectivity	49	38	42	46	45
Hydroxypropanone yield	5.6	3.7	1.5	2.2	2.4
Acetaldehyde yield	1.2	0.6	2.2	1.8	1.7
Propanaldehyde yield	1.5	0.0	1.6	0.9	0.8
Acetone yield	0.1	0.5	0.0	0.2	0.1
Phenol yield	1.0	0.0	0.0	0.0	0.0
Material balance (mass collected/mass introduced)	97.7	98.0	97.2	98.8	98.9
Quantified product balance (products assayed/glycerol introduced)	69.2	65.0	49.9	57.7	59.8

5 The addition of molecular oxygen makes it possible to maintain the glycerol conversion and the acrolein yield by inhibiting the deactivation of the catalyst and the formation of certain by-products.

Example 2 (according to the invention)

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In this Example 2, two types of catalyst (10 ml) are tested: a sulfate zirconia (90% ZrO_2 - 10% SO_4) from Daiichi Kigenso (supplier reference H1416) and a tungsten zirconia (90.7% ZrO_2 - 9.3% WO_3) from Daiichi Kigenso (supplier reference H1417). The first catalyst has a loss on ignition at 1000°C of 8.81% and a specific surface area

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of 54.3 m²/g (BET, 1 point). The second catalyst is characterized by a loss on ignition at 1000°C of 1.75% and a specific surface area of 47.4 m²/g (BET, 1 point). The results are given in Table 2 below:

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Table 2

Cumulative glycerol introduced (g)	9	18	27	21	33
Catalyst	Sulfate zirconia 16.5 g			Tungsten zirconia 17 g	
Glycerol conversion	100	100	100	100	100
Acrolein yield	42.3	53.8	52.5	54.9	53.0
Acrolein selectivity	42	54	52	55	53
Hydroxypropanone yield	0.0	0.0	0.0	0.0	0.0
Acetaldehyde yield	10.3	9.1	8.2	9.8	8.7
Propanaldehyde yield	4.9	3.7	4.0	2.1	1.4
Acetone yield	0.0	0.4	0.0	0.1	0.1
Phenol yield	0.0	0.0	0.3	0.0	0.0
Material balance (mass collected/mass introduced)	96.5	98.0	98.0	97.2	97.9
Quantified product balance(products assayed/glycerol introduced)	57.5	66.9	65.0	66.9	63.2

The formation of hydroxypropanone and of phenol is completely inhibited in the presence of molecular oxygen.

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Example 3 (comparative and according to the invention)

10 ml of zeolite H-beta from Valfor (CP 811BL-25)

representing a mass of 4.23 g were loaded into the reactor. For this example, the flow rate of molecular oxygen used is 0.34 l/h. The results are given in Table 3 below:

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Table 3

Cumulative glycerol introduced (g)	8	16	25	35	9	16	24	32
Addition of oxygen	no	No	no	no	yes	yes	yes	yes
Glycerol conversion	100	99	97	89	100	100	100	100
Acrolein yield	45.6	56.9	52.3	47.9	42.9	57.3	56.4	56.3
Acrolein selectivity	46	57	54	54	43	57	56	56
Hydroxypropanone yield	6.9	9.6	9.5	9.7	0.4	0.9	0.2	0.4
Acetaldehyde yield	5.1	5.2	5.1	4.8	6.6	7.2	6.8	6.1
Propanaldehyde yield	7.2	5.4	4.6	3.4	4.9	3.9	3.2	2.6
Acetone yield	0	0	0	0.2	0	0	0	0
Phenol yield	1.3	0.7	0.5	0.4	1.2	0.5	0.4	0.1
Material balance (mass collected/mass introduced)	95.6	-	98.6	98.9	97	99	98.9	99
Quantified product balance (products assayed/glycerol introduced)	66.2	78.7	74.8	77.2	56	69.8	67	65.5

The addition of molecular oxygen makes it possible to maintain the glycerol conversion and the acrolein yield, while at the same time reducing the formation of by-products.

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Example 4 (comparative and according to the invention)

A phosphate zirconia (91.5% ZrO_2 - 8.5% PO_4) from Daiichi Kigenso (Ref H1418) is used. This catalyst has a loss on
5 ignition at 1000°C of 4.23% and a specific surface area of 128.7 m^2/g . 10 ml of this catalyst, representing a mass of 12.7 g, are loaded into the reactor.

The results are given in Table 4 below.

Table 4

Cumulative glycerol introduced (g)	8	16	24	32	41	9.8	17.9	26.3	34.5	42
Addition of oxygen	0	0	0	0	0	0.34	0.34	0.34	0.34	0.82
Glycerol conversion	100	100	100	100	99	100	100	100	100	100
Acrolein yield	16.6	40.4	46.7	45.2	46.2	23.3	42.0	43.0	44.2	38.2
Acrolein selectivity	17	40	47	45	46	23	42	43	44	38
Hydroxypropanone yield	0.0	9.4	13.0	13.5	14.7	0.0	0.0	0.0	0.1	0.0
Acetaldehyde yield	6.9	6.3	5.0	4.7	4.3	12.2	11.7	9.9	7.7	10.6
Propanaldehyde yield	15.0	14.2	11.7	11.1	9.8	6.3	5.9	5.4	3.7	3.9
Acetone yield	0.0	0.0	0.0	0.0	0.0	2.0	0.6	0.0	0.2	0.2
Phenol yield	4.2	4.4	2.9	2.5	1.8	0.9	0.4	0.2	0.1	0.2
Material balance (mass collected/mass introduced)	95.0	98.2	95.2	97.7	97.6	-	97.8	97.9	98.8	98.6
Quantified product balance (products assayed/glycerol introduced)	42.7	74.6	79.3	77.0	77.4	44.8	60.7	58.5	56.0	53.1

The hydroxypropanone, propanaldehyde and phenol by-products are in markedly lower amount when the process is performed in the presence of oxygen.

Example 5 (comparative of the prior art)

10 ml of H₃PO₄/alpha-alumina catalyst prepared as described in patent US 5 387 720, representing a mass of 10 g, were
 5 loaded into the reactor. The catalyst was prepared in the following manner: 15.9 g of alpha-alumina from Ceramtec (Ref EO-19 - specific surface area 0.7 m²/g - mean pore diameter 2.5 μm - apparent porosity 65% - supplied in the form of rings and ground so as to retain only the
 10 particles of diameter 1-1.4 mm) were impregnated with 4 g of a 20% by weight phosphoric acid solution (prepared by addition of 16.25 ml of water and 5 g of 85% by weight phosphoric acid). The solid is then dried on a rotavapor at 80°C and used directly. The results are collated in
 15 Table 5.

Table 5

Cumulative glycerol introduced (g)	8	16	25	32
Glycerol conversion	91	69	42	17
Acrolein yield	54.5	32.2	20.6	3.8
Acrolein selectivity	60	46	49	23
Hydroxypropanone yield	12.3	9.3	6.5	2.1
Acetaldehyde yield	0.1	0.0	0.0	0.0
Propanaldehyde yield	0.3	0.2	0.1	0.0
Acetone yield	0.0	0.0	0.0	0.0
Phenol yield	1.0	0.1	0.1	0.0
Material balance (mass collected/mass introduced)	98.6	98.7	nd	98.9
Quantified product balance (products assayed/glycerol introduced)	77.6	72.6	84.9	89.4

Example 6 (comparative of the prior art)

10 ml of an H_3PO_4 /alpha alumina catalyst, representing a mass of 8.55 g, were loaded into the reactor. The catalyst was prepared in the same manner as for Example 5, but, after drying at 80°C , the solid was activated in air at 300°C for 3 hours in order to fix the phosphoric acid to the support.

10 The results are given in Table 6 below.

Table 6

Cumulative glycerol introduced (g)	8	16	24	32
Glycerol conversion	70	37	9	8
Acrolein yield	42.1	18.2	4.6	3.1
Acrolein selectivity	60	50	50	41
Hydroxypropanone yield	10.3	4.8	0.0	0.0
Acetaldehyde yield	0.0	0.0	0.0	0.0
Propanaldehyde yield	0.0	0.0	0.0	0.0
Acetone yield	0.0	0.0	0.0	0.1
Phenol yield	0.8	0.0	0.0	0.0
Material balance (mass collected/mass introduced)	98.5	98.9	98.0	99.0
Quantified product balance (products assayed/glycerol introduced)	83.2	86.2	95.5	95.6

15 Rapid deactivation of the catalyst is noted in these two comparative Examples 5 and 6.

CLAIMS

1. Process for manufacturing acrolein by dehydration of glycerol in the presence of molecular oxygen, in an amount
5 chosen so as to be outside the flammability range at any point in the plant.
2. Process according to Claim 1, characterized in that the molecular oxygen is in the form of air or in the form
10 of a mixture of gases containing molecular oxygen.
3. Process according to Claim 1 or 2, characterized in that the glycerol is in the form of an aqueous solution with a concentration of between 10% and 50% and preferably
15 between 15% and 30% by weight in the reactor.
4. Process according to one of Claims 1 to 3, characterized in that an acidic solid catalyst with a Hammett acidity H_0 of less than +2 is used.
20
5. Process according to Claim 4, characterized in that the catalyst is chosen from natural or synthetic siliceous materials or acidic zeolites; mineral supports, such as oxides, coated with mono-, di-, tri- or polyacidic
25 inorganic acids; oxides or mixed oxides, or alternatively heteropolyacids.
6. Process according to Claim 4 or 5, characterized in that the catalyst is chosen from zeolites, Nafion®
30 composites (based on sulfonic acid or fluorinated polymers), chlorinated aluminas, phosphotungstic and/or silicotungstic acids and acid salts, and various solids of metal oxide type such as tantalum oxide Ta_2O_5 , niobium

oxide Nb_2O_5 , alumina Al_2O_3 , titanium oxide TiO_2 , zirconia ZrO_2 , tin oxide SnO_2 , silica SiO_2 or silico-aluminate $\text{SiO}_2\text{-Al}_2\text{O}_3$, impregnated with acidic functions such as borate BO_3 , sulfate SO_4 , tungstate WO_3 , phosphate PO_4 , silicate
5 SiO_2 or molybdate MoO_3 .

7. Process according to one of Claims 4 to 6, characterized in that the catalyst is chosen from sulfate zirconias, phosphate zirconias, tungsten zirconias,
10 siliceous zirconias, sulfate titanium or tin oxides, and phosphate aluminas or silicas.

8. Process according to one of Claims 1 to 7, characterized in that the dehydration reaction is
15 performed in the gas phase.

9. Process according to Claim 8, characterized in that the reaction is performed in a fixed-bed reactor, a fluidized-bed reactor or a circulating fluidized-bed
20 reactor.

10. Process according to one of Claims 1 to 8, characterized in that it is performed in a plate heat exchanger.

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11. Process according to one of Claims 1 to 7, characterized in that the dehydration reaction is performed in the liquid phase.