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(54) **SUPERCRITICAL FLUID PHASE SYNTHESIS OF METHYLENE LACTONES USING NOVEL CATALYST FIELD OF INVENTION**

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This patent is subject to a terminal disclaimer.

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(52) **U.S. Cl.** ..... **549/295**; 549/326

(58) **Field of Classification Search** ..... 549/295, 549/326

See application file for complete search history.

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(57) **ABSTRACT**

Process for converting certain lactones to their alpha-methylene substituted forms that not only exhibits high initial activity (conversion), but also provides high reactor productivity (mass of product per mass of catalyst per unit of time) and sustained maintenance of a high level of activity and productivity with time on stream.

**9 Claims, No Drawings**

**SUPERCRITICAL FLUID PHASE SYNTHESIS  
OF METHYLENE LACTONES USING NOVEL  
CATALYST FIELD OF INVENTION**

CROSS-REFERENCE TO RELATED  
APPLICATION

This application claims priority under 35 U.S.C. §119 from U.S. Provisional Application Ser. No. 60/591,565, filed Jul. 27, 2004.

FIELD OF INVENTION

The invention pertains to a method of producing unsubstituted and substituted alpha-methylene lactones by a supercritical or near-critical fluid phase reaction of starting lactones with formaldehyde in the presence of a novel catalyst that not only exhibits high initial activity (conversion), but also provides high reactor productivity (mass of product per mass of catalyst per unit of time) and sustained maintenance of a high level of activity and productivity with time on stream.

BACKGROUND

Alpha-methylene-gamma-butyrolactone and methyl alpha-methylene-gamma-butyrolactone are useful monomers in the preparation of both homopolymers and copolymers. In addition, the alpha-methylene-gamma-butyrolactone group is an important structural feature of many sesquiterpenes of biological importance.

U.S. 2003-0166949 A1 describes a method for converting certain starting lactones to alpha-methylenelactones in a supercritical fluid (SCF) phase using a heterogeneous so-called basic catalyst that can be selected from the Group I, Group II, and Lanthanide Group oxides, hydroxides, carbonates, hydrogen carbonates, silicates, oxalates, carboxy-

lates, acetates and phosphates, and mixtures thereof, any of which may be supported or unsupported. The basic catalyst may include additives and promoters to enhance catalyst efficiency. The method involves a reaction between the starting lactone and formaldehyde and may be carried out in a batch or continuous mode. The process can be run in either a single homogeneous phase over the catalyst, or the reactants and SCF may be in two different phases over the catalyst. The temperature of the reaction can range from about 70° C. to about 400° C., with a preferred range of about 100° C. to about 350° C. A more preferred range is about 200° C. to about 350° C. Pressure ranges are those required to achieve the supercritical or near-critical state under a given set of reaction conditions. The pressure of the reaction can range from about 5 to about 60 MPa, with a preferred range of about 15 to about 40 MPa.

One important measure of reactor performance that is a strong function of the catalyst activity is termed the "reactor productivity," which can be expressed as the mass of product produced per mass of catalyst per unit of time. High and sustained reactor productivity is desired for a manufacturing process to improve economic viability. In addition, high catalyst activity (reactant conversion) is desirable, in general, to minimize additional processing required to separate the product from unconverted reactants. Although this performance parameter was not explicitly shown for the examples of the method, the reactor productivity can be estimated from the data provided. Table 1 shows the variation in both catalyst activity (GVL conversion) and reactor productivity for the various examples conducted in continuous reactors. These data illustrate that problems inherent in the method include maximum sustained reactor productivities on the order of only 0.7 gram MeMBL/gram catalyst-hour and an undesirable decrease in activity with time on stream.

TABLE 1

Reactor Productivity Estimates for Reference US 2003-0166949A1					
US 2003-0166949A1 Example No.	Catalyst Composition	WHSV (g GVL/g cat.-h)	TOS (h)	GVL Conversion (%)	Calculated Reactor Productivity* (g MeMBL/g cat.-h)
15	K <sub>2</sub> CO <sub>3</sub>	0.22	—	77.1	0.19
16	K <sub>2</sub> CO <sub>3</sub>	0.13	0.84	32.9	0.05
			1.09	34.2	0.05
			1.93	33.3	0.05
			2.46	38.6	0.06
			3.00	40.2	0.06
17	20% K/SiO <sub>2</sub>	0.7	—	64.9	0.51
18	20%	0.7	—	78.4	0.61
19	20%	0.7	—	88.8	0.70
20	20%	0.7	—	69.2	0.54
21	20% Ca/	0.7	—	8.0	0.06
22	20% Ba/	0.7	—	31.5	0.25
23	20% CSiO <sub>2</sub>	0.4	1.25	89.9	0.40
			1.75	88.5	0.40
			2.25	86.2	0.39
			2.75	83.8	0.38
			3.00	82.3	0.37
24	20%	0.7	1.45	94.9	0.74
			1.95	90.3	0.71
			2.45	88.7	0.70
			2.95	86.0	0.67
			3.20	84.9	0.67

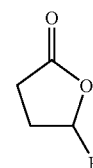
TABLE 1-continued

Reactor Productivity Estimates for Reference US 2003-0166949A1								
US 2003- 0166949A1 Example No.	Catalyst Composition	WHSV (g GVL/g cat.-h)	TOS (h)	GVL Conversion (%)	Calculated Reactor Productivity* (g MeMBL/ g cat.-h)			
25	20%	0.7	1.30	94.1	0.74			
			1.80	92.8	0.73			
			2.30	93.5	0.73			
			2.80	94.6	0.74			
			3.30	95.3	0.75			
			4.10	94.9	0.74			
26	20%	2.5	0.80	72.8	2.04			
			1.00	53.8	1.51			
			1.50	37.1	1.04			
			2.00	29.9	0.84			
			2.50	27.2	0.76			
			3.00	25.9	0.72			
			3.50	24.8	0.69			
			4.00	25.2	0.71			
			5.00	26.3	0.74			
			5.50	26.0	0.73			
27	20%	0.7	1.20	30.9	0.24			
			2.00	30.6	0.24			
			2.50	27.2	0.21			
			3.00	26.0	0.20			
			3.50	23.9	0.19			
			4.00	22.1	0.17			
28	20%	2.5	0.90	17.9	0.50			
			1.20	5.1	0.14			
			1.60	0.8	0.02			
			1.90	0.6	0.02			
			2.40	0.3	0.01			
29	15%	0.7	1.9	95.4	0.75			
			2.4	93.8	0.74			
			2.9	90.7	0.71			
			3.9	89.8	0.70			
			1.90	62.4	0.49			
	(1 <sup>st</sup> Regen.)	0.7	0.7	2.50	57.5	0.45		
				3.00	55.2	0.43		
				(2 <sup>nd</sup> Regen.)	0.7	1.80	63.4	0.50
						2.50	55.7	0.44
						3.00	54.1	0.42
30	15%	1.7	—	4.8	0.09			
31	20%	1.1	—	28.7	0.35			

\*Productivity estimated as product of WHSV, GVL Conversion, and Ratio of Molecular Weights of MeMBL/GVL (112.1/100.12) assuming 100% Selectivity to MeMBL.

It would be advantageous, therefore, to have a lactone conversion process that not only exhibits high initial activity (conversion), but also provides high reactor productivity (mass of product per mass of catalyst per unit of time) and sustained maintenance of a high level of activity and productivity with time on stream.

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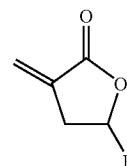
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## SUMMARY OF THE INVENTION

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This need is met by the present invention, which, in its first aspect, is a process for preparing a reaction product comprising an alpha-methylene lactone of the Formula II, said process comprising combining a lactone of the Formula I with formaldehyde derived from a formaldehyde source and a solvent to produce a reaction mixture;

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II

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wherein R is selected from the group consisting of hydrogen, methyl, ethyl, and straight or branched C<sub>3</sub>-C<sub>5</sub> alkyl;

at a temperature and pressure sufficient to cause the reaction mixture to exist as a supercritical or near-critical fluid, said temperature being sufficient to cause the formation of said alpha-methylene lactone of Formula II; said reaction mixture being in the presence of a catalyst;

said catalyst being made by (or obtainable by) a process comprising:

(a) contacting (i) porous silica, optionally containing at least one first element selected from the group consisting of aluminum, zirconium and titanium, said silica having a pore volume of at least 0.4 cc/g of porous silica attributable to pores having a pore diameter between 65 and 3200 Angstroms, with (ii) a solution comprising a solvent and an organic compound of at least one second element selected from the group consisting of sodium, potassium, rubidium, cesium, and barium;

(b) drying the product of step (a) to remove at least a portion of said solvent;

(c) heating the product of step (b) to a temperature in the range of 350° C. to 550° C. to produce a catalyst precursor;

(d) flushing at a preselected flow rate an oxygen-containing gas over said catalyst precursor either during step (c), or after step (c) while the temperature is still in the range of 350° C. to 550° C. to produce a catalyst candidate comprising the at least one second element in an amount from about 0.1% to about 40% by weight of the combined weight of said porous silica and the second element, or the combined weight of the porous silica containing the at least one first element and the second element;

(e) determining by porosimetry whether said catalyst candidate has a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms; and

(f) if said catalyst candidate does not have a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms, repeating, optionally more than once, steps (a) through (e) using in step (d) flow rates successively greater than said preselected flow rate until the catalyst candidate has a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms.

In its second aspect, the present invention is the catalyst made by the process recited above.

The use of such a catalyst in the conversion of lactones of the Formula I to those of Formula II leads not only to high initial activity (conversion), but also provides high reactor productivity (mass of product per mass of catalyst per unit of time) and sustained maintenance of a high level of activity and productivity with time on stream.

#### DETAILED DESCRIPTION OF THE INVENTION

The following terms generally are abbreviated as follows:

“alpha-methylene-gamma-butyrolactone” is abbreviated MBL;

“gamma-butyrolactone” is abbreviated GBL;

“gamma-valerolactone” is abbreviated GVL;

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“alpha-methylene-gamma-valerolactone” is abbreviated MVL;

“gamma-methyl alpha methylene gamma butyrolactone” is abbreviated MeMBL;

“time on stream” is sometimes abbreviated TOS;

“cubic centimeters” is abbreviated as cc or cm<sup>3</sup>;

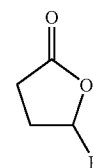
“mass spectroscopy” is abbreviated MS;

“gas chromatography” is abbreviated GC;

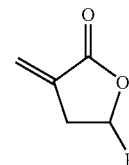
“supercritical fluid” is abbreviated SCF; and

“weight hour space velocity” is abbreviated WHSV.

The process of the present invention concerns a supercritical or near-critical fluid phase methylenation of lactones of Formula I to yield alpha-methylene lactones of Formula II.

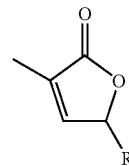


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II

Specifically, a lactone of Formula I is reacted with formaldehyde to give a reaction product comprising alpha methylene lactones of Formula II. The substituent -R group is selected from the group consisting of hydrogen, methyl, ethyl, and straight or branched C<sub>3</sub>-C<sub>5</sub> alkyl. Also produced is an internal isomer of the lactone of Formula III, represented by Formula III, below.



III

In a preferred embodiment the lactone of Formula I is gamma-butyrolactone (R is H) and the alpha-methylene lactone of Formula II is alpha-methylene-gamma-butyrolactone. In a most preferred embodiment the lactone of Formula I is gamma-valerolactone (R is CH<sub>3</sub>) and the alpha-methylene lactone of Formula II is alpha-methylene-gamma-valerolactone.

The process of the present invention is conducted at reaction conditions to achieve a supercritical or near-critical fluid state. The temperature is in the range of from about 70° C. to about 400° C. A temperature in the range of from about 100° C. to about 350° C. is preferred. A temperature in the range of from about 200° C. to about 350° C. is most preferred. The pressure is in the range of from about 5 MPa to about 60 MPa, with a preferred range of from about 15 MPa to about 40 MPa. The catalyst contact time can be

selected to achieve desired yields and selectivities. Contact time can be manipulated by increasing or decreasing flow rates over the catalyst.

The lactones of Formula I, formaldehyde, and the solvent can be in a homogeneous supercritical fluid phase. Alternatively the lactones of Formula I, formaldehyde, and solvent may be in two different phases (one supercritical) over the solid catalyst.

The formaldehyde may be supplied to the reaction in the form of an aqueous solution (formalin), anhydrous formaldehyde, formaldehyde hemiacetal, a low molecular weight polyformaldehyde (paraformaldehyde), or formaldehyde trimer (trioxane). The use of paraformaldehyde, trioxane, or anhydrous formaldehyde is preferred since this reduces the need to remove water from the process. Hemiacetals work effectively, but require separate steps to release the formaldehyde from the alcohol and to recover and recycle the alcohol.

The catalyst used in the present invention comprises silicon and oxygen that form a matrix (or support) for a catalytic element. The catalytic element is at least one element selected from the group consisting of potassium, rubidium, and cesium. Rubidium and cesium are preferred. The catalytic element is deposited on or dispersed within the matrix by contacting the matrix with a solution of an organic compound of at least one of these elements. Organic compounds do not include carbonates. The matrix comprising silicon and oxygen can optionally comprise at least one compound comprising an element selected from the group consisting of titanium, aluminum and zirconium. The catalytic element should constitute from about 0.1% to about 40% by weight of the combined weight of the starting silica and the added catalytic element (as opposed to the entire compound of which the element is a part).

The catalyst must be porous and have a pore-size distribution such that those pores having a diameter between 65 and 3200 Angstroms provide a pore volume of at least about 0.3 cubic centimeters per gram of the catalyst.

In some cases, reaction conditions may result in a decrease of catalyst efficiency. In these situations it may be useful to periodically reactivate the catalyst. For example, contacting the present catalysts, when activity drops below an acceptable level, with oxygen at elevated temperatures has the effect of reactivating the catalyst. Contact temperatures with oxygen may range from about 225° C. to about 500° C., with temperatures of about 250° C. to about 425° C. being preferred.

Selectivities and yields of product may be influenced by the total contact time with the catalyst. As stated previously, yields and selectivities may be increased by adjusting reactant flow rates.

The present method exploits several advantages of using a supercritical fluid (SCF) as the reaction solvent. SCFs are attractive media for conducting chemical transformations, primarily because the solvent and transport properties of a single solution, including the density, can be varied appreciably and continuously with relatively minor changes in temperature or pressure. The density variation in a SCF also influences the chemical potential of solutes and thus reaction rates and equilibrium constants. Thus, the solvent environment can be optimized for a specific reaction application by tuning the various density-dependent fluid properties. For a discussion of advantages and applications of supercritical fluid media for chemistry and catalysis, see Hutchenson, K. W., "Organic Chemical Reactions and Catalysis in Supercritical Fluid Media," in *Supercritical Fluid Technology in*

*Materials Science and Engineering*, Y.-P. Sun (ed.), Marcel Dekker: New York (2002), pp. 87-187.

A fluid is in the SCF state when the system temperature and pressure exceed the corresponding critical point values defined by the critical temperature ( $T_c$ ) and pressure ( $P_c$ ). For pure substances, the critical temperature and pressure are the highest at which vapor and liquid phases can coexist. Above the critical temperature, a liquid does not form for a pure substance, regardless of the applied pressure. Similarly, the critical pressure and critical molar volume are defined at this critical temperature corresponding to the state at which the vapor and liquid phases merge. Similarly, although more complex for multicomponent mixtures, the mixture critical state is identified as the condition at which the properties of coexisting vapor and liquid phases become indistinguishable. For a discussion of supercritical fluids, see *Kirk-Othmer Encyc. of Chem. Technology*, 4<sup>th</sup> Ed., Vol. 23, pg. 452-477.

In addition to typical factors such as chemical inertness, cost, toxicity, etc., the critical temperature must be considered when selecting a potential solvent for conducting chemical transformations in the SCF regime. For practical applications, thermal and catalytic chemical reactions can only be conducted in a relatively narrow temperature range. Lower temperatures result in unacceptable reaction rates, and higher temperatures can result in significant selectivity and yield losses, as well as catalyst deactivation. To obtain practical solvent densities and the corresponding density-dependent properties, this temperature optimization must be balanced against a general desire to operate in the vicinity of the mixture critical point of the reaction system to fully exploit the potential advantages afforded by SCF operation. The phase behavior of the reaction mixture, which is strongly influenced by the solvent critical temperature, is fundamentally important in defining this operating window, so one must select a solvent to provide the desired phase behavior. The phase behavior of SCF systems can also be manipulated to control the number and composition of coexisting phases, thus controlling both reaction effects, as well as the separation of products or homogeneous catalysts from the reaction mixture.

In practice, a number of desirable properties characteristic of the SCF state are also realized in the expanded liquid region that exists at temperatures and pressures slightly below this critical point. Hence, for the purposes of this invention, the term "supercritical fluid" also includes such "near-critical fluids," where the fluid is either at or below the critical temperature and the properties begin to approach those of a supercritical fluid. For the purposes of this invention, a "near-critical fluid" is considered to exist at those conditions where the fluid is at temperatures from about 75% of the critical temperature to about 100% of the critical temperature, and at pressures from about 25% of the critical pressure to about 100% of the critical pressure.

One can visually observe the phase behavior of the reaction mixture by conducting the reaction in a vessel equipped with a transparent window, or by simulating the reaction mixture with a solution of similar concentration in such a vessel. Systematic determination of the phase boundaries of the reaction mixture can be determined by standard techniques using such a vessel that is also equipped with a means of varying the vessel volume at fixed composition and temperature. The vessel is loaded with the various components at the specified composition of the reaction mixture, heated to the reaction temperature, then the solution pressure is varied by changing the vessel volume until a phase transition is visually observed. After measuring the

phase boundary of a solution of interest over the range of anticipated compositions, one can define the operating conditions necessary to achieve the supercritical or near-critical state for conducting the desired reaction.

Any suitable SCF solvent may be used in the process of this invention, including, but not limited to, carbon dioxide, nitrous oxide, sulfur hexafluoride, fluoromethane, trifluoromethane, tetrafluoromethane, ethane, ethylene, propane, propanol, isopropanol, propylene, butane, butanol, isobutane, isobutene, pentane, hexane, cyclohexane, benzene, toluene, o-xylene, water, and mixtures thereof, provided that it is inert to all reagents and products. Preferred SCF solvents include carbon dioxide or a C<sub>1</sub>-C<sub>6</sub> alkane, optionally substituted with Cl, F or Br. More preferred is where the supercritical fluids are carbon dioxide, trifluoromethane, pentane, or propane.

Separation and/or purification of the desired products, including MBL or MeMBL, from unreacted starting lactone and/or reaction byproducts may be performed by processes known in the art. A particularly suitable method to recover the desired product is to polymerize MBL in GBL solution, or MeMBL in GVL solution, using standard free-radical polymerization, isolate the polymer by precipitation, and then thermally depolymerize back to MBL or MeMBL, as the case may be, by heating under vacuum. Separation of the MeMBL from the internal isomer of Formula III can be accomplished by the polymerization of MeMBL. An appropriate polymerization technique is taught in U.S. Pat. No. 6,723,790. Finally, MBL can be separated from GBL by melt crystallization. Another effective method is liquid-liquid extraction.

Non-limiting reactors suitable for the process of the instant invention include tubular, fluidized bed, fixed bed, trickle bed, transport bed, and stirred tank reactors. The process can be run in either batch or continuous mode as described, for example, in H. Scott Fogler, *Elements of Chemical Reaction Engineering*, 2<sup>nd</sup> Edition, Prentice-Hall Inc, CA, 1992. The process can also be run in either a single homogeneous phase over the solid catalyst, or the reactants and SCF may be in two different phases over the solid catalyst.

The reaction can be carried out by passing solutions of the formaldehyde and lactone over the catalyst at elevated temperatures and pressures.

The catalysts of the present invention can be made by (or are obtainable by) a method as follows.

Porous silica powder, such as the material sold by Grace Davison, Inc. (Columbia, Md.) with a pore volume of at least 0.4 cc/g attributable to pores having a diameter between 65 and 3200 Angstroms is used as a catalyst matrix. Porosity preferably is determined by mercury porosimetry. Preferably the porous silica contains compounds of aluminum, titanium and/or zirconium. These latter silicas are preferred because of their hydrothermal stability.

A suitable way of choosing appropriate starting porous silica is to eliminate first those silicas that have a mean pore diameter less than 65 Angstroms and porosity less than 0.4 cc/g attributable to pores having diameters between 65 and 3200 Angstroms. Next, a porous silica manufacturer that describes its products in terms of mean pore diameter can be consulted to see if the manufacturer has the underlying data from which the mean pore diameter was calculated. If so, the manufacturer may be able to specify which, if any, of its products have pore volumes of greater than 0.4 Angstroms attributable to pores having diameters between 65 and 3200 Angstroms. These materials preferably should be indepen-

dently tested by mercury porosimetry to determine that they meet the porosity characteristics required by this invention.

Organic compounds such as the carboxylates, such as acetate, propionate, butyrate, and 2-ethylhexanoate of a catalytic element selected from the group consisting of potassium, rubidium, and cesium is dissolved in aqueous or non-aqueous solvent and contacted with the porous silica. Organic compounds do not include carbonates of the aforesaid catalytic elements. Organic compounds containing acetates are preferred. Other organic anions such as acetylacetonates can be used. One convenient method for introducing the catalytic element into the porous silica is to dissolve a suitable weight of the organic compound of the catalytic element in just enough solvent to equal the volume of the pores of the selected amount of the porous silica. The amount of organic compound should be chosen to provide to the silica from 0.1 wt % to 40 wt % of the element relative to the combined weight of the porous silica plus the element (as opposed to the compound of which the element is a part). The resulting material is allowed to dry, preferably in a nitrogen environment for an extended time. The purpose of the drying is to remove at least a portion of the solvent in which the organic compound is dissolved.

Organic compounds such as the alkoxides can also be used. Organic alkoxides of an element selected from the group consisting of potassium, rubidium, and cesium can contain from one to 20 carbon atoms and preferably 1 to 5 carbon atoms in the alkoxide group. The organic alkoxide should be soluble in the solvent. Most alkoxides can be dissolved in non-aqueous solutions such as ethanol, propanol, or isopropyl alcohol. Subsequent methods for introducing the element and drying are the same.

The dried material is then heated (for example in an alumina boat placed in a tube furnace) at an ambient temperature of 350° C. to 550° C. (The temperature of the catalyst material may be somewhat higher because of exothermic reactions taking place on the material.) A temperature between 450° C. and 550° C. is preferred. Either during the heating or subsequent to it, but at the same temperature, the material is flushed with an oxygen-containing gas (e.g. air), which is believed to burn off organic residues formed during the heating step. In a tube furnace, an airflow rate of at least 110 cc/min in a 3 cm diameter tube furnace, which corresponds to a linear velocity of 15.6 cm/min was found to be acceptable. Use of sufficiently high airflow rates is important to produce a high surface area material. In a tube furnace, the material can be heated at a rate of 5° C./min to 120° C., and can be exposed to this temperature for 4 hours. It can be heated subsequently at a rate of 5° C./min to approximately 450° C. and held at this temperature for 16 hours. Other equipment can be used to perform the heating step. Such equipment includes fluidized bed and rotary calcination equipment.

Heating can be accomplished in air or in a combination of an inert gas such as nitrogen, argon, or krypton for parts of the cycle, followed by air. An initial drying step at 120° C. in nitrogen, another inert gas, or air is preferred for a period of 30 minutes to 24 hours. Following this drying step, the catalyst can be heated in air or nitrogen to a temperature of 350° C. to 550° C. For acetate precursors, 450° C. to 550° C. is required. Heating times can range from 30 minutes to 48 hours. The final heating step preferably is performed in air for at least 30 minutes.

Once the catalyst is made as described above, it can be tested to confirm that it has a pore volume of at least about 0.3 cc/g of catalyst attributable to pores having a diameter between about 65 and 3200 Angstroms. (The reason that the

porosity of the starting porous silica decreases after treatment is believed to be attributable to the inclusion of the catalytic metal into the silica.)

Pore volume may be obtained by a variety of techniques, but preferred techniques are mercury (Hg) and nitrogen porosimetric techniques, with Hg porosimetry being most preferred.

Mercury porosimetry data can be obtained at 414 MPa using, for example, a Micromeritics 1 Model 9420 AutoPore III Instrument (Micromeritics Inc., One Micromeritics Drive, Norcross Ga. 30093-1877). This technique permits one to measure the pore volume and size by forcing mercury to penetrate inside the open porosity. Mercury is used because it behaves as a non-wetting liquid with a large number of materials.

Mercury is forced to enter into the pores by applying a controlled increasing pressure. As a sample holder is filled with mercury under vacuum conditions, mercury surrounds the sample without entering the pores due to the very low residual pressure. During the test, the pressure is increased, and the volume of mercury penetrated is detected by means of a capacitive system. The decreasing volume of mercury in the sample holder represents the pore volume. The penetration pressure is directly related to the pore access size by a well-known mathematical model, expressed by the Washburn equation:

$$D = -4\gamma \cos(\theta) / P_c$$

Where:

$\gamma$  is surface tension of pure mercury (480 dyne/cm);  
 $\theta$  is contact angle between mercury and the solid (average value 140° C.);  
 $P_c$  is mercury penetration equilibrated pressure; and  
 $D$  is pore diameter.

The distribution of pore size, as well as the total porosity, bulk and apparent density and the specific pore volume can be obtained by the relationship between the pressure necessary for penetration (the pore dimension) and the volume of penetrated mercury (pore volume). At each pressure, therefore, a differential volume of Hg can be calculated which occupies the pores of the solid; this represents the additional Hg volume which is intruded as a consequence of a pressure increase. At lower applied pressure, larger pores are filled with Hg. With increasing applied pressure, smaller pores are occupied up to the smallest pore diameter reasonably measurable by this technique, 65 Angstroms. In this way, a distribution of pore sizes as a function of applied pressure can be obtained.

Pores are assumed to be of a cylindrical shape, as is standard for this technique.

Sample compressibility correction is calculated post priori by determining the volume of samples and pores not yet intruded as a function of applied pressure. The resulting relationship is used to correct the raw intrusion data for sample compression effects.

From the Washburn equation it is clear that the pore size range that can be investigated by mercury porosimetry is directly related to the pressure range.

Alternatively, nitrogen porosimetry may be used. Dinitrogen adsorption/desorption measurements can be performed at 77.3° K. using, for example, Micromeritics ASAP model 2400/2405 porosimeters. Samples can be degassed at 150° C. overnight prior to data collection. Pore volume distributions can be determined using a 27 point desorption isotherm and can be analyzed using the BJH method described in E. P. Barret, L. G. Joyner and P. P. Halenda, *J. Amer. Chem. Soc.*, 73, 373(1951).

In this invention, pores having diameters of 65 Angstroms up to 3200 Angstroms should contribute at least 0.3 cc/g pore volume to the final catalyst. Above 3200 Angstroms, inter-particle pores and void spaces are measured, and are not important for this invention.

If the catalyst possesses the correct porosity, it may be used in the lactone conversion method. If not, it may be necessary to repeat the catalyst synthesis using higher oxygen-containing gas flow rates than were used initially. The process may have to be repeated several times with successively higher gas flow rates until a catalyst with the correct porosity is finally obtained.

#### COMPARATIVE EXAMPLE

Comparative Catalyst 1 ("CC1") (20% by Weight of Rb, on Davicat® Si 1251; Small Pore Volume Silica)

About 20 g of SiO<sub>2</sub> powder, Davicat® Si 1251 (Grace Davison, Inc., Columbia, Md.) having a pore volume of 0.36 cm<sup>3</sup>/g was used as support material to make the catalyst support. 8.45 g of rubidium acetate (Aldrich Chemical Co.) was dissolved in enough water to match to pore volume of the Si 1251 support; that is, enough solution was added to approximately fill the pores of the support. The amount of water used was approximately 7.2 cm<sup>3</sup>. The material was allowed to dry for at least 12 hours in a nitrogen environment. Approximately 10 cm<sup>3</sup> of the material was loaded into an alumina boat and heated in a tube furnace. The internal diameter of the tube furnace was 10 cm. The airflow rate was greater than 1220 cm<sup>3</sup>/min, which corresponds to a linear velocity of greater than 15.6 cm/min. Use of this higher airflow is important to produce a high surface area material. The material was heated at a rate of 5° C./min to 120° C.; this temperature (120° C.) was maintained for four hours. It was subsequently heated at a rate of 5° C./min to approximately 450° C. (as measured by a thermocouple placed approximately 0.5 cm over the catalyst bed) and was held at this temperature for 16 hours.

The catalyst produced as described above has a porosity of less than 0.01 cc/g, and, therefore, does not have the porosity characteristics required by the present invention, i.e., greater than 0.3 cc/g attributable to pores having diameters between 65 and 3200 Angstroms.

#### EXAMPLES OF THE INVENTION

Catalyst 1 (20% by Weight of Rb on Davicate® Si 1700, SiO<sub>2</sub> Support)

The procedure outlined in Example 1 was used with a Davicat® Si 1700 support (Grace Davison, Inc., Columbia Md.) instead of Si 1251. Davicate® Si 1700 is a SiO<sub>2</sub> powder, and has a greater pore volume than Davicat® Si 1251 (approximately 1.57 cc/g). Therefore, 31.4 cm<sup>3</sup> of water was used, instead of 7.2 cm<sup>3</sup> as in Example 1.

Catalyst 2 (20% by Weight of Rb, on Davicat® Si 1405, SiO<sub>2</sub> Support)

The procedure outlined in Example 1 was used with a Davicat® Si 1405 support (Grace Davison, Inc., Columbia Md.) instead of Davicat® Si 1251. Davicat® Si 1405 is a SiO<sub>2</sub> powder which has a greater pore volume than Si 1251 (approximately 1.10 cc/g). Therefore, 22.0 cm<sup>3</sup> of water was used, instead of 7.2 cm<sup>3</sup> as in Example 1. The same heating protocol was used as described in example 1, except that a 3 cm diameter tube furnace was used instead of a 10 cm diameter tube furnace. In this tube furnace, an airflow rate of

at least 110 cc/min was used, which corresponds to a linear velocity of 15.6 cm/sec. The material was heated at a rate of 5° C./min to 120° C. The temperature was held at 120° C. for four hours. The material was subsequently heated at a rate of 5° C./min to approximately 450° C. (as measured by a thermocouple placed approximately 0.5 cm over the catalyst bed) and was held at this temperature for 16 hours and allowed to cool to room temperature in air.

Catalyst 3 (20% by Weight of Rb, on Davicat® Si 1415, SiO<sub>2</sub> Support)

The procedure outlined in Example 1 was used with a Davicat® Si 1415 support (Grace Davison, Inc., Columbia Md.) instead of Davicat® Si 1251. Davicat® Si 1415 has a greater pore volume than Davicat® Si 1251 (approximately 0.85 cc/g). Therefore, 17.0 cm<sup>3</sup> of water was used, instead of 7.2 cm<sup>3</sup> as in Example 1.

The same heating protocol was used as described in example 1, except that a 3 cm diameter tube furnace was used instead of a 10 cm diameter tube furnace. In this tube furnace, an airflow rate of at least 110 cc/min was used, which corresponds to a linear velocity of 15.6 cm/sec. The material was heated at a rate of 5° C./min to 120° C. The temperature was held at 120° C. for four hours. The material was subsequently heated at a rate of 5° C./min to approximately 450° C. (as measured by a thermocouple placed approximately 0.5 cm over the catalyst bed) and was held at this temperature for 16 hours and allowed to cool to room temperature in air.

Catalyst 4 (20% by Weight of Rb, on Davicat® SiZr 4101, Zr—SiO<sub>2</sub> Support)

The same procedure as described in example 1 was used, except for the following differences: 20 g of Davicat® SiZr 4101 support was used, and 8.45 g of rubidium acetate was dissolved in 19.60 g of water and impregnated into the support.

Catalyst 5 (20% by Weight of Rb, on Davicat® SiTi 4151, Ti—SiO<sub>2</sub> Support)

The same procedure as described in example 1 was used, except for the following differences: 20 g of Davicat® SiTi 4151 support was used, and 8.45 g of rubidium acetate was dissolved in 20.0 g of water and impregnated into the support.

Catalyst 6 (20% by Weight of Cs, on Davicat® Si 1700, SiO<sub>2</sub> Support)

The same procedure as described in example 1 was used, except for the following differences: 20 g of Davicat® Si 1700 support was used, and 7.22 g of cesium acetate was dissolved in 29.0 g of water and impregnated into the support.

Catalyst 7: (20% by Weight of Cs, on Davicat® Si 1405, SiO<sub>2</sub> Support)

The same procedure as described in example 1 was used, except for the following differences: 20 g of Davicat® Si 1405 support was used, and 7.22 g of cesium acetate was dissolved in 19.0 g of water and impregnated into the support.

Catalyst 8: (20% by Weight of Cs, on Davicat® SiZr 4101, Zr—SiO<sub>2</sub> Support)

About 20 g of SiO<sub>2</sub> powder, Davicate® Si Zr 4101 (Grace Davison, Inc., Columbia, Md.) having a pore volume of 0.98 cm<sup>3</sup>/g was used as support material to make the catalyst support. 7.22 g of cesium acetate (Aldrich Chemical Co.) was dissolved in enough water to match the pore volume of the Si Zr 4101 support; that is, enough solution was added

to approximately fill the pores of the support. The amount of water used was approximately 19 cm<sup>3</sup>. The material was allowed to dry for at least 12 hours in a nitrogen environment. Approximately 10 cm<sup>3</sup> of the material was loaded into an alumina boat and heated in a tube furnace. The internal diameter of the tube furnace was 10 cm. The airflow rate was greater than 1220 cm<sup>3</sup>/min, which corresponds to a linear velocity of greater than 15.6 cm/min. Use of this higher airflow is important to produce a high surface area material. The material was heated at a rate of 5° C./min to 120° C.; this temperature (120° C.) was maintained for four hours. It was subsequently heated at a rate of 5° C./min to approximately 450° C. (as measured by a thermocouple placed approximately 0.5 cm over the catalyst bed) and was held at this temperature for 16 hours.

Catalyst 9: (10.7% by Weight of K, on Davicate® SiTi 4151, Ti—SiO<sub>2</sub> Support)

About 20 g of SiO<sub>2</sub> powder, Davicat® SiTi 4151 (Grace Davison, Inc., Columbia, Md.) having a pore volume of 1.0 cm<sup>3</sup>/g was used as support material to make the catalyst support. 5.74 g of potassium acetate (Aldrich Chemical Co.) was dissolved in enough water to match to pore volume of the SiTi 4151 support; that is, enough solution was added to approximately fill the pores of the support. The amount of water used was approximately 20 cm<sup>3</sup>. The material was allowed to dry for at least 12 hours in a nitrogen environment. Approximately 10 cm<sup>3</sup> of the material was loaded into an alumina boat and heated in a tube furnace. The internal diameter of the tube furnace was 10 cm. The airflow rate was greater than 1220 cm<sup>3</sup>/min, which corresponds to a linear velocity of greater than 15.6 cm/min. Use of this higher airflow is important to produce a high surface area material. The material was heated at a rate of 5° C./min to 120° C.; this temperature (120° C.) was maintained for four hours. It was subsequently heated at a rate of 5° C./min to approximately 450° C. (as measured by a thermocouple placed approximately 0.5 cm over the catalyst bed) and was held at this temperature for 16 hours.

Catalyst 10: (20% by Weight of Rb, on Davicat® Si 1700, SiO<sub>2</sub> Support)

The same procedure as described in catalyst 1 was used.

TABLE 2

Mercury Porosimetry Data of Catalysts		
Cat. No.	Composition wt % = (wt element/wt Support + wt element) × 100	Pore Volume (cc/g) from pores having diameter of 65 Å or greater
CC1	20% Rb/Si 1251	<0.01
1	20% Rb/Si 1700	1.04
2	20% Rb/Si 1405	0.908
3	20% Rb/Si 1415	0.771
4	20% Rb/SiZr 4101	0.313
5	20% Rb/SiTi 4151	0.732
6	20% Cs/Si 1700	0.817
7	20% Cs/Si 1405	0.889
8	20% Cs/SiZr 4101	0.378
9	10.3% K/SiTi 4151	0.716
10	20% Rb/Si 1700	1.04

#### EXAMPLES OF SUPERCRITICAL FLUID PHASE REACTION

The following examples were conducted in a continuous fixed bed reactor consisting of a 0.25-inch o.d.x0.049-inch



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wall×4.5-inch long 316 stainless steel tube packed with granular or powder catalyst, except as noted. The reactor was heated by cartridge-type electrical heaters mounted in an aluminum block enclosing the reactor. The lactone was combined with ethanol hemiacetal as the formaldehyde precursor and metered to the reactor as a liquid feed with a syringe pump. The ethanol hemiacetal was prepared by refluxing a 50 mol % paraformaldehyde solution in ethanol for four hours at 95° C., followed by cooling to room temperature and filtration. The solvent was metered as a condensed liquid with a second positive-displacement pump, and the two streams were heated and combined prior to entering the reactor. Liquid-phase reactor effluent samples were collected downstream in an ice bath after venting the solvent, and reaction products were quantified by gas chromatography. The reactor pressure was controlled by a back-pressure regulator located downstream of the reactor.

## Comparative Example 1

The reactor was charged with 1.24 g of 20% Rb/Davicat® Si 1251 catalyst (Comparative Catalyst CC1). The reactant feed solution consisted of 52.3 wt % GVL, with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.17 g gamma-valerolactone/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 7.5 mol % in the reactor feed. The reactor was operated at temperatures of 245° C. and 300° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 0.52 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.04 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.77	14.7	96.4	0.19
1.25	5.2	55.9	0.04
2.10	3.1	37.1	0.02

## Example 1

The reactor was charged with 0.48 g of 20% Rb/Davicat® Si 1700 catalyst (Catalyst 1). The reactant feed solution consisted of 52.3 mol % GVL, with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.17 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 3.0 mol % in the reactor feed. The reactor was operated at a temperature of 245° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 23.0 mol GVL converted/mol of catalyst, the overall reactor productivity was 1.17 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

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Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.67	95.6	96.5	1.21
1.33	84.2	97.7	1.08
2.00	79.7	97.8	1.02
3.00	78.0	97.7	1.00
4.00	77.8	97.5	0.99
5.00	80.0	96.9	1.02

## Example 2

The reactor was charged with 0.57 g of 20% Rb/Davicat® Si 1405 catalyst (Catalyst 2). The reactant feed solution consisted of 52.3 wt % GVL, with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.17 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 3.6 mol % in the reactor feed. The reactor was operated at a temperature of 245° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 23.2 mol GVL converted/mol of catalyst, the overall reactor productivity was 1.16 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.58	89.4	97.0	1.14
1.33	79.5	97.8	1.02
2.08	77.6	96.7	0.98
3.08	78.3	97.3	1.00
4.08	80.3	96.7	1.02
5.08	82.8	96.0	1.04

## Example 3

The reactor was charged with 0.61 g of 20% Rb/Davicat® Si 1415 catalyst (Catalyst 3). The reactant feed solution consisted of 48.3 wt % GVL, 7.7 wt % p-xylene as an internal standard, and the balance was made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.26 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 3.9 mol % in the reactor feed. The reactor was operated at a temperature of 250° C. and a pressure of about 20.1 MPa. The cumulative turnover number for the reaction was 10.2 mol GVL converted/mol of catalyst, the overall reactor productivity was 1.28 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

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Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.50	91.7	96.5	1.28
1.00	93.0	96.2	1.29
1.50	93.8	95.9	1.30
2.00	94.1	95.5	1.30

## Example 4

The reactor was charged with 0.79 g of 20% Rb/Davicat® SiZr 4101 catalyst (Catalyst 4). The reactant feed solution consisted of 48.3 wt % GVL, 7.7 wt % p-xylene as an internal standard, and the balance was made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.16 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 3.9 mol % in the reactor feed. The reactor was operated at a temperature of 251° C. and a pressure of about 20.0 MPa. The cumulative turnover number for the reaction was 7.8 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.99 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (gMeMBL/g catalyst-h)
0.50	95.0	96.6	1.03
1.00	92.0	97.5	1.00
1.50	91.1	96.7	0.99
2.00	91.1	96.5	0.98

## Example 5

The reactor was charged with 0.80 g of 20% Rb/Davicat® SiTi 4151 catalyst (Catalyst 5). The reactant feed solution consisted of 52.3 wt % GVL with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.17 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 4.4 mol % in the reactor feed. The reactor was operated at a temperature of 246° C. and a pressure of about 23.6 MPa. The cumulative turnover number for the reaction was 21.5 mol GVL converted/mol of catalyst, the overall reactor productivity was 1.08 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

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Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.58	87.2	97.5	1.12
1.08	74.6	98.3	0.96
1.58	71.5	97.8	0.92
2.08	69.8	97.6	0.89
3.08	69.0	97.7	0.88
4.08	73.7	96.8	0.94
5.08	76.2	95.1	0.95

## Example 6

The reactor was charged with 0.61 g of 20% Cs/Davicat® Si 1700 catalyst (Catalyst 6). The reactant feed solution consisted of 50.9 wt % GVL, 2.6 wt % diphenyl ether as an internal standard, and the balance was made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.32 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 3.9 mol % in the reactor feed. The reactor was operated at a temperature of 251° C. and a pressure of about 20.0 MPa. The cumulative turnover number for the reaction was 29.6 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.97 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.52	84.5	97.0	1.21
1.02	76.0	97.5	1.10
1.52	72.6	97.4	1.05
2.02	68.6	97.3	0.99
3.02	64.3	97.2	0.93
4.02	60.4	97.2	0.87
5.02	57.8	97.1	0.83

## Example 7

The reactor was charged with 0.63 g of 20% Cs/Davicat® Si 1405 catalyst (Catalyst 7). The reactant feed solution consisted of 52.3 wt % GVL with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 0.75 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 2.6 mol % in the reactor feed. The reactor was operated at a temperature of 246° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 20.5 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.66 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.55	94.2	94.8	0.75
1.13	84.7	97.0	0.69
1.63	79.4	97.5	0.65
2.07	78.9	97.5	0.65
3.07	78.9	97.0	0.64
4.07	78.6	96.5	0.64
5.07	81.4	96.0	0.66

## Example 8

The reactor was charged with 0.80 g of 20% Cs/Davicat® SiZr 4101 catalyst (Catalyst 8). The reactant feed solution consisted of 50.9 wt % GVL, 2.6 wt % diphenyl ether as an internal standard, and the balance was made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.30 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 5.0 mol % in the reactor feed. The reactor was operated at a temperature of 250° C. and a pressure of about 20.0 MPa. The cumulative turnover number for the reaction was 30.3 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.97 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.50	88.4	92.1	1.19
1.00	80.9	93.8	1.10
1.50	75.6	94.8	1.04
2.00	71.5	95.2	0.99
3.00	67.5	95.0	0.93
4.00	64.6	95.0	0.89
5.00	61.9	95.2	0.86

## Example 9

The reactor was charged with 0.67 g of 10.3% K/Davicat® SiTi 4151 catalyst (Catalyst 9). The reactant feed solution consisted of 52.3 wt % GVL with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 1.32 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 4.7 mol % in the reactor feed. The reactor was operated at a temperature of 245° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 21.0 mol GVL converted/mol of catalyst, the overall reactor productivity was 1.19 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
0.50	91.1	95.7	1.29
1.00	81.2	97.1	1.16
1.50	75.9	97.3	1.09
2.03	74.3	97.2	1.07
3.02	70.3	97.2	1.01
4.03	65.2	96.9	0.93
5.03	69.3	96.1	0.98

## Example 10

This example was conducted in a continuous fixed bed reactor consisting of a 0.50-inch o.d.×0.065-inch wall×11.5-inch long 316 stainless steel tube packed with catalyst. The reactor was heated by electrical band heaters mounted around an aluminum block enclosing the reactor. The reactor was charged with 3.98 g of 20% Rb/Davicat® Si 1700 catalyst (Catalyst 10). The reactant feed solution consisted of 52.3 wt % GVL with the balance made up with the ethanol hemiacetal solution. This solution resulted in a 1.2:1 ratio of formaldehyde to GVL in the reactor feed, which was metered at a rate resulting in a weight hour space velocity (WHSV) in the reactor of 0.94 g GVL/(g catalyst-h). CO<sub>2</sub> was used as the SCF solvent phase, and the flow rate was metered independently to give a final total organic concentration of 5.0 mol % in the reactor feed. The reactor was operated at a temperature of 245° C. and a pressure of about 23.5 MPa. The cumulative turnover number for the reaction was 30.4 mol GVL converted/mol of catalyst, the overall reactor productivity was 0.84 g MeMBL/g catalyst-h, and the corresponding reaction profile showing conversion of GVL to MeMBL is summarized below:

Time on Stream (h)	GVL Conversion (%)	MeMBL Selectivity (%)	Reactor Productivity (g MeMBL/g catalyst-h)
1.37	90.0	94.2	0.89
2.12	85.0	94.5	0.84
2.87	85.0	94.0	0.84
3.87	85.2	94.4	0.85
4.87	85.5	93.4	0.84
5.87	85.7	92.1	0.83
6.87	84.7	91.4	0.81
7.87	83.8	90.7	0.80
8.87	81.9	90.0	0.77
9.87	78.5	88.1	0.73

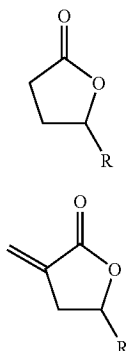
The data of Examples 1–10 show that while most catalysts exhibit a decline in GVL conversion and reactor productivity with time on stream, all demonstrate significantly higher and more sustained activity than that of comparative catalyst 1 (insufficient pore volume). In addition, these examples show more sustained activity with high GVL conversion levels and higher reactor productivities with time on stream than the examples from U.S. 2003-0166949 A1 that are illustrated in Table 1.

What is claimed is:

1. A process for preparing a reaction product comprising an alpha-methylene lactone of the Formula II, said process comprising combining a lactone of the Formula I with

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formaldehyde derived from a formaldehyde source and a solvent to produce a reaction mixture,



wherein R is selected from the group consisting of hydrogen, methyl, ethyl, and straight or branched C<sub>3</sub>-C<sub>5</sub> alkyl;

at a temperature and pressure sufficient to cause the reaction mixture to exist as a supercritical or near-critical fluid, said temperature being sufficient to cause the formation of said alpha-methylene lactone of Formula II;

said reaction mixture being in the presence of a catalyst; said catalyst made by a process comprising:

- (a) contacting (i) porous silica, optionally containing a first element, said first element being one, two or three members selected from the group consisting of aluminum, zirconium and titanium, said silica having a pore volume of at least 0.4 cc/g of porous silica attributable to pores having a diameter between 65 and 3200 Angstroms, with (ii) a solution comprising a solvent and an organic compound of a second element, said second element being one, two or three members selected from the group consisting of potassium, rubidium, and cesium;
- (b) drying the product of step (a) to remove at least a portion of said solvent;
- (c) heating the product of step (b) to a temperature in the range of 350° C. to 550° C. to produce a catalyst precursor;
- (d) flushing at a preselected flow rate an oxygen-containing gas over said catalyst precursor either during step (c), or after step (c) while the temperature

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is still in the range of 350° C. to 550° C. to produce a catalyst candidate comprising the second element in an amount from about 0.1% to about 40% by weight of the combined weight of said porous silica and the second element, or the combined weight of the porous silica containing the first element and the second element;

- (e) determining by porosimetry whether said catalyst candidate has a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms; and
- (f) if said catalyst candidate does not have a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms, repeating, optionally more than once, steps (a) through (e) using in step (d) flow rates successively greater than said preselected flow rate until the catalyst candidate has a pore volume of at least about 0.3 cubic centimeters per gram of catalyst attributable to pores having a diameter between 65 and 3200 Angstroms.

2. The process of claim 1 wherein the porosimetry is mercury porosimetry.

3. The process of claim 1 wherein R is hydrogen or methyl.

4. The Process of claim 1 wherein the solvent is carbon dioxide or a C1-C6 alkane, optionally substituted with Cl, F, or Br.

5. The process of claim 4 wherein the solvent is carbon dioxide, pentane, trifluoromethane, or propane.

6. The process according to claim 1 wherein the formaldehyde source is selected from the group consisting of trioxane, anhydrous formaldehyde, formalin, formaldehyde oligomer, formaldehyde cyclic oligomer, formaldehyde acetal, formaldehyde hemiacetal, and formaldehyde polymer.

7. The process according to claim 6 wherein the formaldehyde source is in the form of formalin, trioxane, formaldehyde hemiacetal or paraformaldehyde.

8. The process of claim 1 wherein the temperature is in the range of from about 70° C. to about 400° C. and the pressure is in the range of from about 5 MPa to about 60 MPa.

9. The process of claim 1 further comprising separating said alpha methylene lactone from said reaction product.

\* \* \* \* \*