

INDUSTRIAL PROCESS FOR PRODUCTION OF HIGH-PURITY DIARYL CARBONATE

Technical Field

5 The present invention relates to an industrial process for the production of a high-purity diaryl carbonate. More particularly, the present invention relates to a process for industrially producing, from a cyclic carbonate and an aromatic monohydroxy compound, stably for a prolonged period of time a large amount of a high-purity diaryl carbonate
10 required for producing a high-quality high-performance aromatic polycarbonate.

Background Art

 As a process for producing a diaryl carbonate, a process of
15 reacting an aromatic monohydroxy compound with phosgene has been known from long ago, and has also been the subject of a variety of studies in recent years. However, this process has the problem of using phosgene, and in addition chlorinated impurities that are difficult to separate out are present in the diaryl carbonate produced using this
20 process, and hence the diaryl carbonate cannot be used as a starting material for the production of an aromatic polycarbonate as is. The reason for this is that such chlorinated impurities markedly inhibit the polymerization reaction in the transesterification method for producing the aromatic polycarbonate which is carried out in the presence of an
25 extremely small amount of a basic catalyst; for example, even if such chlorinated impurities are present in an amount of only 1 ppm, the

polymerization hardly proceeds at all. To make the diaryl carbonate capable of being used as the starting material of a transesterification method aromatic polycarbonate, a troublesome multi-stage separation / purification process involving thorough washing with a dilute aqueous alkaline solution and hot water, oil / water separation, distillation and so on is thus required. Furthermore, the yield decreases due to hydrolysis loss and distillation loss during this separation / purification process. There are thus many problems in carrying out this process economically on an industrial scale.

10 On the other hand, a process for producing aromatic carbonates through transesterification reactions between the dialkyl carbonate and the aromatic monohydroxy compound is also known. However, such transesterification reactions are all equilibrium reactions. The equilibrium is biased extremely toward the original system and the reaction rate is slow, and hence there have been many difficulties in producing aromatic carbonates industrially in large amounts using such a process.

Several proposals have been made to improve on the above difficulties, but most of these have related to development of a catalyst to increase the reaction rate. Many metal compounds have been proposed as catalysts for this type of the transesterification reaction. However, the problem of the disadvantageous equilibrium cannot be resolved merely by developing the catalysts, and hence there are very many issues to be resolved including the reaction system in order to provide a process for industrial production aiming for mass production.

Attempts have also been made to devise a reaction system so as

to shift the equilibrium toward the product system as much as possible, and thus improve the aromatic carbonate yield. For example, for the reaction between dimethyl carbonate and phenol, there have been proposed a method in which methanol produced as a by-product is
5 distilled off by azeotropy together with an azeotrope-forming agent (see Patent Document 1: Japanese Patent Application Laid-Open No. 54-48732 (corresponding to West German Patent Application Laid-Open No. 736063, and U.S. Patent No. 4252737)), and a method in which the methanol produced as a by-product is removed by being adsorbed onto a
10 molecular sieve (see Patent Document 2: Japanese Patent Application Laid-Open No. 58-185536 (corresponding to U.S. Patent No. 410464)). Moreover, a method has also been proposed in which, using an apparatus in which a distillation column is provided on top of a reactor, an alcohol produced as a by-product in the reaction is separated off from
15 the reaction mixture, and at the same time unreacted starting material that evaporates is separated off by distillation (see Patent Document 3: Japanese Patent Application Laid-Open No. 56-123948 (corresponding to U.S. Patent No. 4182726)).

However, these reaction systems are basically batch system or
20 switchover system. This is because there is a limitation on how much the reaction rate can be improved through catalyst development for such the transesterification reaction, and hence the reaction rate is still slow, and thus it has been thought that a batch system is preferable to a continuous system. Of these, a continuous stirred tank reactor (CSTR)
25 system in which a distillation column is provided on top of a reactor has been proposed as a continuous system, but there are problems such as

the reaction rate being slow, and the gas-liquid interface in the reactor being small based on the volume of the liquid. It is thus not possible to make the reaction ratio high. Accordingly, it is difficult to attain the object of producing the aromatic carbonate continuously in large amounts stably for a prolonged period of time by means of the above methods, and many issues remain to be resolved before economical industrial implementation is possible.

The present inventors have developed reactive distillation methods in which such a transesterification reaction is carried out in a continuous multi-stage distillation column simultaneously with separation by distillation, and have been the first in the world to disclose that such a reactive distillation system is useful for such a transesterification reaction, for example a reactive distillation method in which the dialkyl carbonate and the aromatic hydroxy compound are continuously fed into a multi-stage distillation column, and reaction is carried out continuously inside the column in which a catalyst is present, while continuously withdrawing by distillation a low boiling point component containing an alcohol produced as a by-product, and continuously withdrawing a component containing a produced alkyl aryl carbonate from a lower portion of the column (see Patent Document 4: Japanese Patent Application Laid-Open No. 3-291257), a reactive distillation method in which an alkyl aryl carbonate is continuously fed into the multi-stage distillation column, and reaction is carried out continuously inside the column in which a catalyst is present, while continuously withdrawing by distillation a low boiling point component containing a dialkyl carbonate produced as a by-product, and continuously withdrawing a component

containing a produced diaryl carbonate from a lower portion of the column (see Patent Document 5: Japanese Patent Application Laid-Open No. 4-9358), a reactive distillation method in which these reactions are carried out using two continuous multi-stage distillation columns, and
5 hence a diaryl carbonate is produced continuously while efficiently recycling a dialkyl carbonate produced as a by-product (see Patent Document 6: Japanese Patent Application Laid-Open No. 4-211038 (corresponding to WO 91/09832, European Patent No. 0461274, and U.S. Patent No. 5210268)), and a reactive distillation method in which a
10 dialkyl carbonate and an aromatic hydroxy compound or the like are continuously fed into the multi-stage distillation column, and a liquid that flows down through the column is withdrawn from a side outlet provided at an intermediate stage and / or a lowermost stage of the distillation column, and is introduced into a reactor provided outside the distillation
15 column so as to bring about reaction, and is then introduced back in through a circulating inlet provided at a stage above the stage where the outlet is provided, whereby reaction is carried out in both the reactor and the distillation column (see Patent Document 7: Japanese Patent Application Laid-Open No. 4-235951).

20 These reactive distillation methods proposed by the present inventors are the first to enable aromatic carbonates to be produced continuously and efficiently, and processes in which the diaryl carbonate is produced from the dialkyl carbonate using two continuous multi-stage distillation columns based on the above disclosures have been proposed
25 thereafter (see, for example, Patent Document 8: Japanese Patent Application Laid-Open No. 6-157424 (corresponding to European Patent

No. 0582931, and U.S. Patent No. 5334742), Patent Document 9:

Japanese Patent Application Laid-Open No. 6-184058 (corresponding to

European Patent No. 0582930, and U.S. Patent No. 5344954), Patent

Document 10: Japanese Patent Application Laid-Open No. 9-40616,

5 Patent Document 11: Japanese Patent Application Laid-Open No.

9-59225, Patent Document 12: Japanese Patent Application Laid-Open

No. 9-176094, Patent Document 13: WO 00/18720 (corresponding to U.S.

Patent No. 6093842), Patent Document 14: Japanese Patent Application

Laid-Open No. 2001-64235).

10 Among reactive distillation systems, the present applicants have
further proposed, as a method that enables highly pure aromatic
carbonates to be produced stably for a prolonged period of time without a
large amount of a catalyst being required, a method in which high boiling
point material containing a catalyst component is reacted with an active
15 substance and then separated off, and the catalyst component is
recycled (see Patent Document 15: WO 97/11049 (corresponding to
European Patent No. 0855384, and U.S. Patent No. 5872275)), and a
method carried out while keeping the weight ratio of a polyhydric
aromatic hydroxy compound in the reaction system to a catalyst metal at
20 not more than 2.0 (see Patent Document 16: Japanese Patent Application
Laid-Open No. 11-92429 (corresponding to European Patent No. 1016648,
and U.S. Patent No. 6262210)). Furthermore, the present inventors
have proposed a method in which 70 to 99 % by weight of phenol
produced as a by-product in a polymerization process is used as a
25 starting material, and diphenyl carbonate is produced using a reactive
distillation method; this diphenyl carbonate can be used as a starting

material for polymerization to produce an aromatic polycarbonate (see Patent Document 17: Japanese Patent Application Laid-open No. 9-255772 (corresponding to European Patent No. 0892001, and U.S. Patent No. 5747609)).

- 5 However, in all of these prior art documents in which the production of aromatic carbonates using a reactive distillation method is proposed, there is no disclosure whatsoever of a specific process or apparatus enabling mass production on an industrial scale (e.g. 1 ton / hr), nor is there any description suggesting such a process or apparatus.
- 10 For example, the descriptions regarding heights (H_1 and H_2 : cm), diameters (D_1 and D_2 : cm), numbers of stages (n_1 and n_2), and starting material feeding rates (Q_1 and Q_2 : kg / hr) for two reactive distillation columns disclosed for producing mainly diphenyl carbonate (DPC) from dimethyl carbonate and phenol are as summarized in Table 1.

15 Table 1

TABLE 1

H_1	D_1	n_1	Q_1	H_2	D_2	n_2	Q_2	PATENT DOCUMENTS
600	25	20	66	600	25	20	23	6
350	2.8	-	0.2	305	5~10	15+ PACKINGS	0.6	9
500	5	50	0.6	400	8	50	0.6	10
100	4	-	1.4	200	4	-	0.8	11
300	5	40	1.5	-	5	25	0.7	12
1200	20	40	86	600	25	20	31	15 16
600	-	20	66	600	-	20	22	17

In other words, the biggest pairs of continuous multi-stage

distillation columns used when carrying out this reaction using the reactive distillation system are those disclosed by the present applicants in Patent Document 15 (WO 97/11049 (corresponding to European Patent No. 0855384, and U.S. Patent No. 5872275)) and Patent Document 16 (Japanese Patent Application Laid-Open No. 11-92429 (corresponding to European Patent No. 1016648, and U.S. Patent No. 6262210)). As can be seen from Table 1, the maximum values of the various conditions for the continuous multi-stage distillation columns disclosed for the above reaction are $H_1 = 1200$ cm, $H_2 = 600$ cm, $D_1 = 20$ cm, $D_2 = 25$ cm, $n_1 = n_2 = 50$ (the only conditions: Patent Document 10 (Japanese Patent Application Laid-Open No. 9-40616)), $Q_1 = 86$ kg / hr, and $Q_2 = 31$ kg / hr, and the amount of diphenyl carbonate produced has not exceeded approximately 6.7 kg / hr, which is not an amount produced on an industrial scale.

The dialkyl carbonate used in a step (II) of the present invention must be produced on an industrial scale, and furthermore must not contain a halogen. The only process in which such a dialkyl carbonate is industrially produced in a large amount as the starting material for the aromatic polycarbonate is an oxidative carbonylation process in which methanol is reacted with carbon monoxide and oxygen to produce dimethyl carbonate and water. However, with this oxidative carbonylation process (see, for example, Patent Document 18: WO 03/016257), the reaction must be carried out in a slurry state using a large amount of CuCl-HCl as a catalyst, and hence there is a problem of the corrosivity being very high in the reaction system and a separation / purification system. Moreover, in this process, the carbon monoxide is

prone to being oxidized into carbon dioxide, and hence there is a problem that the selectivity based on the carbon monoxide is low at approximately 80%.

On the other hand, several processes for the production of a dialkyl carbonate and a diol from the reaction between a cyclic carbonate and an aliphatic monohydric alcohol have been proposed. With this reaction, the dialkyl carbonate can be produced without using a halogen, and hence this is a preferable process. As reaction systems, four systems have been proposed. These four reaction systems are used in a process for the production of dimethyl carbonate and ethylene glycol from ethylene carbonate and methanol, which is the most typical reaction example, and are (1) a completely batch reaction system, (2) a batch reaction system using a reaction vessel having a distillation column provided on top thereof, (3) a flowing liquid reaction system using a tubular reactor, and (4) a reactive distillation system first disclosed by the present inventors (see, for example, Patent Document 19: Japanese Patent Application Laid-Open No. 4-198141, Patent Document 20: Japanese Patent Application Laid-Open No. 9-194435, Patent Document 21: WO99/64382 (corresponding to European Patent No. 1086940, and U.S. Patent No. 6346638), Patent Document 22: WO00/51954 (corresponding to European Patent No. 1174406, and U.S. Patent No. 6479689), Patent Document 23: Japanese Patent Application Laid-open No. 5-213830 (corresponding to European Patent No. 0530615, and U.S. Patent No. 5231212), Patent Document 24: Japanese Patent Application Laid-Open No. 6-9507 (corresponding to European Patent No. 0569812, and U.S. Patent No. 5359118), Patent Document 25: Japanese Patent

Application Laid-Open No. 2003-119168 (corresponding to
WO03/006418), Patent Document 26: Japanese Patent Application
Laid-Open No. 2003-300936, Patent Document 27: Japanese Patent
Application Laid-Open No. 2003-342209). However, there have been
5 problems with these systems as follows.

In the case of (1) and (3), the upper limit of the cyclic carbonate
conversion is determined by the composition put in and the temperature,
and hence the reaction cannot be carried out to completion, and thus the
conversion is low. Moreover, in the case of (2), to make the cyclic
10 carbonate conversion high, the produced dialkyl carbonate must be
distilled off using a very large amount of the aliphatic monohydric alcohol,
and a long reaction time is required. In the case of (4), the reaction can
be made to proceed with a higher conversion than with (1), (2) or (3).
However, processes of (4) proposed hitherto have related to producing
15 the dialkyl carbonate and the diol either in small amounts or for a short
period of time, and have not related to carrying out the production on an
industrial scale stably for a prolonged period of time. That is, these
processes have not attained the object of producing the dialkyl carbonate
continuously in a large amount (e.g. not less than 2 ton / hr) stably for a
20 prolonged period of time (e.g. not less than 1000 hours, preferably not
less than 3000 hours, more preferably not less than 5000 hours). For
example, the maximum values of the height (H: cm), diameter (D: cm),
and number of stages (n) of the reactive distillation column, the amount
produced P (kg / hr) of dimethyl carbonate, and the continuous
25 production time T (hr) in examples disclosed for the production of
dimethyl carbonate (DMC) and ethylene glycol (EG) from ethylene

carbonate and methanol are as in Table 2.

Table 2

TABLE 2

PATENT DOCUMENT	H : cm	D : cm	NO. STAGES : n	P : kg/hr	T : hr
19	100	2	30	0.106	400
20	160	5	40	0.743	NOTE 5
21	200	4	PACKING COLUMN (Dixon)	0.932	NOTE 5
22	NOTE 1	5	60	0.275	NOTE 5
23	250	3	PACKING COLUMN (Raschig)	0.392	NOTE 5
24	NOTE 2	NOTE 2	NOTE 2	0.532	NOTE 5
25	NOTE 3	NOTE 3	42	NOTE 4	NOTE 5
26	NOTE 3	NOTE 3	30	3750	NOTE 5
27	200	15	PACKING COLUMN (BX)	0.313	NOTE 5

NOTE 1 : OLDERSHAW DISTILLATION COLUMN.

NOTE 2 : NO DESCRIPTION WHATSOEVER DEFINING DISTILLATION COLUMN.

NOTE 3 : ONLY DESCRIPTION DEFINING DISTILLATION COLUMN IS NUMBER OF STAGES.

NOTE 4 : NO DESCRIPTION WHATSOEVER OF PRODUCED AMOUNT.

NOTE 5 : NO DESCRIPTION WHATSOEVER REGARDING STABLE PRODUCTION FOR PROLONGED PERIOD OF TIME.

In Patent Document 26 (Japanese Patent Application Laid-Open

5 No. 2003-300936), it is stated at paragraph 0060 that "The present example uses the same process flow as for the preferred mode shown in FIG. 1 described above, and was carried out with the object of operating a commercial scale apparatus for producing dimethyl carbonate and ethylene glycol through transesterification by a catalytic conversion

10 reaction between ethylene carbonate and methanol. Note that the following numerical values in the present example can be adequately used in the operation of an actual apparatus", and as that example it is stated that 3750 kg / hr of dimethyl carbonate was specifically produced. The scale described in that example corresponds to an annual production

15 of over 30,000 tons, and hence this implies that operation of the world's

largest scale commercial plant using this process had been carried out at the time of the filing of the patent application for Patent Document 26 (Japanese Patent Application Laid-Open No. 2003-300936) (April 9, 2002),. However, even at the time of filing the present application, there is not the above fact at all. Moreover, in the example of Patent Document 26, exactly the same value as the theoretically calculated value is stated for the amount of dimethyl carbonate produced, but the yield for ethylene glycol is approximately 85.6%, and the selectivity is approximately 88.4%, and hence it cannot really be said that a high yield and high selectivity have been attained. In particular, the low selectivity indicates that this process has a fatal drawback as an industrial production process. (Note also that Patent Document 26 (Japanese Patent Application Laid-Open No. 2003-300936) was deemed to have been withdrawn on July 26, 2005 due to examination not having been requested).

With the reactive distillation method, there are very many causes of fluctuation such as composition variation due to reaction and composition variation due to distillation in the distillation column, and temperature variation and pressure variation in the column, and hence continuing stable operation for a prolonged period of time is accompanied by many difficulties, and in particular these difficulties are further increased in the case of handling large amounts. To continue mass production of the dialkyl carbonate and the diol using the reactive distillation method stably for a prolonged period of time while maintaining high yields and high selectivities for the dialkyl carbonate and the diol, the reactive distillation apparatus must be cleverly devised. However,

the only description of continuous stable production for a prolonged period of time with the reactive distillation method proposed hitherto has been 400 hours in Patent Document 19 (Japanese Patent Application Laid-Open No. 4-198141).

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Disclosure of Invention

Problems to be Solved by the Invention

It is an object of the present invention to provide a specific process that enables a high-purity diaryl carbonate required for
10 producing a high-quality high-performance aromatic polycarbonate to be produced industrially in a large amount (e.g. not less than 1 ton / hr) stably for a prolonged period of time (e.g. not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours) from a cyclic carbonate and an aromatic monohydroxy compound.

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Means for Solving the Problems

The present inventors have carried out studies aimed at discovering a specific process enabling the above object to be attained, and as a result have reached to the present invention. That is, according to the first aspect of the present invention, there are provides:

- 20 1. an industrial process for the production of a high-purity diaryl carbonate in which a high-purity diaryl carbonate is continuously produced from a cyclic carbonate and an aromatic monohydroxy compound, the process comprising the steps of:

25 (I) continuously producing a dialkyl carbonate and a diol through a reactive distillation system of continuously feeding the cyclic carbonate and an aliphatic monohydric alcohol into a continuous multi-stage

distillation column T_0 in which a catalyst is present, carrying out reaction and distillation simultaneously in said column, continuously withdrawing a low boiling point reaction mixture containing a produced dialkyl carbonate from an upper portion of the column in a gaseous form, and
5 continuously withdrawing a high boiling point reaction mixture containing the diol from a lower portion of the column in a liquid form;

(II) continuously producing a diaryl carbonate by taking said dialkyl carbonate and an aromatic monohydroxy compound as a starting material, continuously feeding the starting material into a first continuous
10 multi-stage distillation column in which a catalyst is present, carrying out reaction and distillation simultaneously in said first column, continuously withdrawing a first column low boiling point reaction mixture containing a produced alcohol from an upper portion of said first column in a gaseous form, continuously withdrawing a first column high boiling point reaction
15 mixture containing a produced alkyl aryl carbonate from a lower portion of said first column in a liquid form, continuously feeding said first column high boiling point reaction mixture into a second continuous multi-stage distillation column in which a catalyst is present, carrying out reaction and distillation simultaneously in said second column,
20 continuously withdrawing a second column low boiling point reaction mixture containing a produced dialkyl carbonate from an upper portion of said second column in a gaseous form, continuously withdrawing a second column high boiling point reaction mixture containing a produced diaryl carbonate from a lower portion of said second column in a liquid
25 form, and continuously feeding the second column low boiling point reaction mixture containing the dialkyl carbonate into the first continuous

multi-stage distillation column; and

(III) purifying said diaryl carbonate so as to obtain the high-purity diaryl carbonate;

wherein:

- 5 (a) said continuous multi-stage distillation column T_0 comprises a structure having a cylindrical trunk portion having a length L_0 (cm) and an inside diameter D_0 (cm) and having an internal having number of stages n_0 thereinside, and comprises a gas outlet having an inside diameter d_{01} (cm) at a top of the column or in the upper portion of the
- 10 column near to the top, a liquid outlet having an inside diameter d_{02} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one first inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one
- 15 second inlet provided in the middle portion and/or the lower portion of the column above the liquid outlet, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} respectively satisfy the following formulae (1) to (6),

$$2100 \leq L_0 \leq 8000 \quad (1),$$

$$180 \leq D_0 \leq 2000 \quad (2),$$

$$4 \leq L_0 / D_0 \leq 40 \quad (3),$$

20 $10 \leq n_0 \leq 120 \quad (4),$

$$3 \leq D_0 / d_{01} \leq 20 \quad (5), \text{ and}$$

$$5 \leq D_0 / d_{02} \leq 30 \quad (6);$$

- (b) said first continuous multi-stage distillation column comprises a structure having a cylindrical trunk portion having a length L_1 (cm) and
- 25 an inside diameter D_1 (cm), and having an internal with number of stages n_1 thereinside, and comprises a gas outlet having an inside diameter d_{11}

(cm) at a top of the column or in the upper portion of the column near to the top, a liquid outlet having an inside diameter d_{12} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one third inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one fourth inlet provided in the middle portion and/or the lower portion of the column above the liquid outlet, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} respectively satisfy the following formulae (7) to (12),

$$1500 \leq L_1 \leq 8000 \quad (7),$$

$$100 \leq D_1 \leq 2000 \quad (8),$$

$$2 \leq L_1 / D_1 \leq 40 \quad (9),$$

$$20 \leq n_1 \leq 120 \quad (10),$$

$$5 \leq D_1 / d_{11} \leq 30 \quad (11), \text{ and}$$

$$3 \leq D_1 / d_{12} \leq 20 \quad (12);$$

(c) said second continuous multi-stage distillation column comprises a structure having a cylindrical trunk portion having a length L_2 (cm) and an inside diameter D_2 (cm), and having an internal with number of stages n_2 thereinside, and comprises a gas outlet having an inside diameter d_{21} (cm) at a top of the column or in the upper portion of the column near to the top, a liquid outlet having an inside diameter d_{22} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one fifth inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one sixth inlet provided in the central portion and/or the lower portion of the column above the liquid outlet, wherein L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} respectively satisfy the following formulae (13) to (18),

$$1500 \leq L_2 \leq 8000 \quad (13),$$

$$100 \leq D_2 \leq 2000 \quad (14),$$

$$2 \leq L_2 / D_2 \leq 40 \quad (15),$$

$$10 \leq n_2 \leq 80 \quad (16),$$

$$5 \quad 2 \leq D_2 / d_{21} \leq 15 \quad (17), \text{ and}$$

$$5 \leq D_2 / d_{22} \leq 30 \quad (18),$$

2. the process according to item 1, wherein not less than 1 ton / hr of the high-purity diaryl carbonate is produced,

3. the process according to item 1 or 2, wherein said d_{01} and said d_{02}
 10 for said continuous multi-stage distillation column T_0 used in the step (I) satisfy the formula (19)

$$1 \leq d_{01} / d_{02} \leq 5 \quad (19),$$

4. the process according to any one of items 1 to 3, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} for said continuous multi-stage distillation
 15 column T_0 satisfy respectively $2300 \leq L_0 \leq 6000$, $200 \leq D_0 \leq 1000$, $5 \leq L_0 / D_0 \leq 30$, $30 \leq n_0 \leq 100$, $4 \leq D_0 / d_{01} \leq 15$, and $7 \leq D_0 / d_{02} \leq 25$,

5. the process according to any one of items 1 to 4, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} for said continuous multi-stage distillation
 column T_0 satisfy respectively $2500 \leq L_0 \leq 5000$, $210 \leq D_0 \leq 800$, $7 \leq L_0 / D_0 \leq 20$, $40 \leq n_0 \leq 90$, $5 \leq D_0 / d_{01} \leq 13$, and $9 \leq D_0 / d_{02} \leq 20$,
 20

6. the process according to any one of items 1 to 5, wherein said continuous multi-stage distillation column T_0 is a distillation column having a tray and/or a packing as said internal,

7. the process according to item 6, wherein said continuous multi-stage
 25 distillation column T_0 is a plate type distillation column having the tray as said internal,

8. the process according to item 6 or 7, wherein said tray in said continuous multi-stage distillation column T_0 is a sieve tray having a sieve portion and a downcomer portion,
9. the process according to item 8, wherein said sieve tray in said continuous multi-stage distillation column T_0 has 100 to 1000 holes / m^2 in said sieve portion thereof,
10. the process according to item 8 or 9, wherein a cross-sectional area per hole of said sieve tray in said continuous multi-stage distillation column T_0 is in a range of from 0.5 to 5 cm^2 ,
11. the process according to any one of items 8 to 10, wherein an aperture ratio of said sieve tray in said continuous multi-stage distillation column T_0 is in a range of from 1.5 to 15%,
12. the process according to any one of items 1 to 11, wherein said d_{11} and said d_{12} for said first continuous multi-stage distillation column used in the step (II) satisfy the following formula (20), and said d_{21} and said d_{22} for said second continuous multi-stage distillation column used in the step (II) satisfy the following formula (21);
- $$1 \leq d_{12} / d_{11} \leq 5 \quad (20) \text{ and}$$
- $$1 \leq d_{21} / d_{22} \leq 6 \quad (21),$$
13. the process according to any one of items 1 to 12, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for said first continuous multi-stage distillation column used in the step (II) satisfy respectively $2000 \leq L_1 \leq 6000$, $150 \leq D_1 \leq 1000$, $3 \leq L_1 / D_1 \leq 30$, $30 \leq n_1 \leq 100$, $8 \leq D_1 / d_{11} \leq 25$, and $5 \leq D_1 / d_{12} \leq 18$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for said second continuous multi-stage distillation column used in the step (II) satisfy respectively $2000 \leq L_2 \leq 6000$, $150 \leq D_2 \leq 1000$, $3 \leq L_2 / D_2 \leq$

30, $15 \leq n_2 \leq 60$, $2.5 \leq D_2 / d_{21} \leq 12$, and $7 \leq D_2 / d_{22} \leq 25$,

14. the process according to any one of items 1 to 13, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for said first continuous multi-stage distillation column satisfy respectively $2500 \leq L_1 \leq 5000$, $200 \leq D_1 \leq 800$,

5 $5 \leq L_1 / D_1 \leq 15$, $40 \leq n_1 \leq 90$, $10 \leq D_1 / d_{11} \leq 25$, and $7 \leq D_1 / d_{12} \leq 15$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for said second continuous multi-stage distillation column satisfy respectively $2500 \leq L_2 \leq 5000$, $200 \leq D_2 \leq 800$, $5 \leq L_2 / D_2 \leq 15$, $20 \leq n_2 \leq 50$, $3 \leq D_2 / d_{21} \leq 10$, and $9 \leq D_2 / d_{22} \leq 20$,

10 15. the process according to any one of items 1 to 14, wherein each of said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is a distillation column having a tray and/or a packing as said internal,

15 16. the process according to item 15, wherein said first continuous multi-stage distillation column is a plate type distillation column having the tray as said internal, and said second continuous multi-stage distillation column is a distillation column having both the packing and the trays as said internal,

20 17. the process according to item 15 or 16, wherein each of said trays in said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is a sieve tray having a sieve portion and a downcomer portion,

25 18. the process according to item 17, wherein each of said sieve trays in said first continuous multi-stage distillation column and said second continuous multi-stage distillation column has 100 to 1000 holes / m^2 in said sieve portion,

19. the process according to item 17 or 18, wherein a cross-sectional area per hole of each of said sieve trays in said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is in a range of from 0.5 to 5 cm²,

5 20. the process according to item 15 or 16, wherein said second continuous multi-stage distillation column is a distillation column having, as said internal, the packing in the upper portion of the column, and the tray in the lower portion of the column,

10 21. the process according to any one of items 15 to 20, wherein said packing of said internal in said second continuous multi-stage distillation column is one or a plurality of sets of structured packings,

22. the process according to item 21, wherein said structured packings in said second continuous multi-stage distillation column are of at least one type selected from the group consisting of Mellapak, Gempak,

15 Techno-pack, Flexipac, a Sulzer packing, a Goodroll packing, and Glitschgrid,

23. the process according to any of items 1 to 22, wherein the diaryl carbonate purification step (III) is carried out by distillation.

20 In addition, according to the second aspect of the present invention, there are provided:

24. a high-purity diaryl carbonate produced in an amount of not less than 1 ton / hr, which is produced by the process according to any one of items 1 to 23,

25 25. the high-purity diaryl carbonate according to item 24, which has a halogen content of not more than 0.1 ppm,

26. the high-purity diaryl carbonate according to item 25, which has a

halogen content of not more than 1 ppb.

Advantageous Effects of the Invention

It has been discovered that by implementing the process
5 according to the present invention, a high-purity diaryl carbonate
required for producing a high-quality high-performance aromatic
polycarbonate can be produced on an industrial scale of not less than 1
ton / hr from the cyclic carbonate and the aromatic monohydroxy
compound. Moreover, it has been discovered that such a high-purity
10 diaryl carbonate can be produced stably for a prolonged period of time of,
for example, not less than 2000 hours, preferably not less than 3000
hours, more preferably not less than 5000 hours. The present invention
thus provides a process that achieves excellent effects as an industrial
process for the production of the high-purity diaryl carbonate.

15

Brief Description of Drawings

FIG. 1 is an examples of schematic drawing of a continuous
reactive distillation column T_0 preferable for carrying out the present
invention, the distillation column having internals comprising sieve trays
20 provided inside a trunk portion thereof;

FIG. 2 is an example of schematic drawing of a first continuous
reactive distillation column preferable for carrying out the present
invention, the distillation column having internals provided in a trunk
portion thereof;

25

FIG. 3 is an example of schematic drawing of a second continuous
reactive distillation column preferable for carrying out the present

invention, the distillation column having provided, in a trunk portion thereof, internals comprising structured packings in an upper portion and sieve trays in a lower portion; and

FIG. 4 is an example of schematic view of an apparatus comprising the first continuous reactive distillation column and the second continuous reactive distillation column linked together preferable for carrying out the present invention.

Description of Reference Numerals

(In Figure 1)

1: gas outlet, 2: liquid outlet, 3-a to 3-e: inlet, 4-a to 4-b: inlet, 5: end plate, 6: internal, 7: trunk portion, 10: continuous multi-stage distillation column, L: length (cm) of trunk portion, D: inside diameter(cm) of trunk portion, d_1 : inside diameter of gas outlet, d_2 : inside diameter (cm) of gas outlet

(In Figures 2, 3, and 4)

1: gas outlet, 2: liquid outlet, 3: inlet, 4: inlet, 5: end plate, L_1 and L_2 : length (cm) of trunk portion, D_1 and D_2 : inside diameter (cm) of trunk portion, d_{11} and d_{21} : inside diameter (cm) of gas outlet (cm), d_{12} and d_{22} : inside diameter (cm) of liquid outlet,

6: internal, 7: trunk portion, 10: continuous multi-stage distillation column, L: length (cm) of trunk portion, D: inside diameter(cm) of trunk portion, d_1 : inside diameter of gas outlet, d_2 : inside diameter (cm) of gas outlet, 101: first continuous multi-stage distillation column, 201: second continuous multi-stage distillation column, 11, 12, and 21:

inlet, 13 and 23: column top gas outlet, 14, 24, 18, and 28: heat exchanger, 17 and 27: column bottom liquid outlet, 16 and 26: column top component outlet, 31: column bottom component outlet of the second continuous multi-stage distillation column.

5

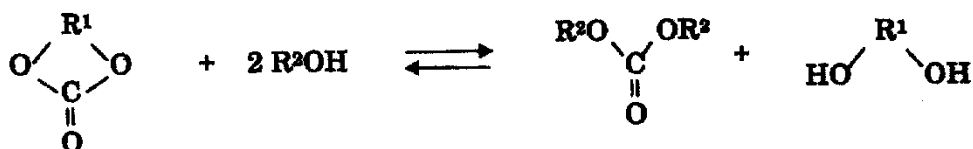
Best Mode for Carrying Out the Invention

Following is a detailed description of the present invention.

In the present invention, first, a step (I) of continuously producing a dialkyl carbonate and a diol on an industrial scale from a cyclic carbonate and an aliphatic monohydric alcohol is carried out.

10

The reaction of step (I) is a reversible transesterification reaction represented by the following formula:



15

wherein R^1 represents a bivalent group $-(\text{CH}_2)_m-$ (m is an integer from 2 to 6), one or more of the hydrogens thereof being optionally substituted with an alkyl group or aryl group having 1 to 10 carbon atoms. Moreover, R^2 represents a monovalent aliphatic group having 1 to 12 carbon atoms, one or more of the hydrogens thereof being optionally substituted with an alkyl group or aryl group having 1 to 10 carbon atoms.

20

Examples of the cyclic carbonate include an alkylene carbonate such as ethylene carbonate or propylene carbonate, or 1,3-dioxacyclohexa-2-one, 1,3-dioxacyclohepta-2-one, or the like, ethylene carbonate or propylene carbonate being more preferably used

due to ease of procurement and so on, and ethylene carbonate being particularly preferably used.

Moreover, as the aliphatic monohydric alcohol, one having a lower boiling point than the diol produced is used. Although possibly varying depending on the type of the cyclic carbonate used, examples of the aliphatic monohydric alcohol include methanol, ethanol, propanol (isomers), allyl alcohol, butanol (isomers), 3-buten-1-ol, amyl alcohol (isomers), hexyl alcohol (isomers), heptyl alcohol (isomers), octyl alcohol (isomers), nonyl alcohol (isomers), decyl alcohol (isomers), undecyl alcohol (isomers), dodecyl alcohol (isomers), cyclopentanol, cyclohexanol, cycloheptanol, cyclooctanol, methylcyclopentanol (isomers), ethylcyclopentanol (isomers), methylcyclohexanol (isomers), ethylcyclohexanol (isomers), dimethylcyclohexanol (isomers), diethylcyclohexanol (isomers), phenylcyclohexanol (isomers), benzyl alcohol, phenethyl alcohol (isomers), phenylpropanol (isomers), and so on. Furthermore, these aliphatic monohydric alcohols may be substituted with substituents such as halogens, lower alkoxy groups, cyano groups, alkoxycarbonyl groups, aryloxy carbonyl groups, acyloxy groups, and nitro groups.

Of such aliphatic monohydric alcohols, ones preferably used are alcohols having 1 to 6 carbon atoms, more preferably alcohols having 1 to 4 carbon atoms, i.e. methanol, ethanol, propanol (isomers), and butanol (isomers). In the case of using ethylene carbonate or propylene carbonate as the cyclic carbonate, preferable aliphatic monohydric alcohols are methanol and ethanol, methanol being particularly preferable.

When carrying out the reactive distillation of step (I), the method of making a catalyst be present in the reactive distillation column may be any method, but in the case, for example, of a homogeneous catalyst that dissolves in the reaction liquid under the reaction conditions, the catalyst
5 can be made to be present in a liquid phase in the reactive distillation column by feeding the catalyst into the reactive distillation column continuously, or in the case of a heterogeneous catalyst that does not dissolve in the reaction liquid under the reaction conditions, the catalyst can be made to be present in the reaction system by disposing the
10 catalyst as a solid in the reactive distillation column; these methods may also be used in combination.

In the case that a homogeneous catalyst is continuously fed into the reactive distillation column, the homogeneous catalyst may be fed in together with the cyclic carbonate and / or the aliphatic monohydric
15 alcohol, or may be fed in at a different position to the starting materials. The reaction actually proceeds in the distillation column in a region below the position at which the catalyst is fed in, and hence it is preferable to feed the catalyst into a region between the top of the column and the position(s) at which the starting materials are fed in.
20 The catalyst must be present in at least 5 stages, preferably at least 7 stages, more preferably at least 10 stages.

Moreover, in the case of using a heterogeneous solid catalyst, the catalyst must be present in at least 5 stages, preferably at least 7 stages, more preferably at least 10 stages. A solid catalyst that also has an
25 effect as a packing in the distillation column may be used.

Examples of the catalyst used in step (I) include:

alkali metals and alkaline earth metals such as lithium, sodium, potassium, rubidium, cesium, magnesium, calcium, strontium, and barium;

basic compounds of alkali metals and alkaline earth metals such as hydrides, hydroxides, alkoxides, aryloxides, and amides;

basic compounds of alkali metals and alkaline earth metals such as carbonates, bicarbonates, and organic acid salts;

tertiary amines such as triethylamine, tributylamine, trihexylamine, and benzyldiethylamine;

nitrogen-containing heteroaromatic compounds such as N-alkylpyrroles, N-alkylindoles, oxazoles, N-alkylimidazoles, N-alkylpyrazoles, oxadiazoles, pyridine, alkylpyridines, quinoline, alkylquinolines, isoquinoline, alkylisoquinolines, acridine, alkylacridines, phenanthroline, alkylphenanthrolines, pyrimidine, alkylpyrimidines, pyrazine, alkylpyrazines, triazines, and alkyltriazines;

cyclic amidines such as diazobicycloundecene (DBU) and diazobicyclononene (DBN);

thallium compounds such as thallium oxide, thallium halides, thallium hydroxide, thallium carbonate, thallium nitrate, thallium sulfate, and thallium organic acid salts;

tin compounds such as tributylmethoxytin, tributylethoxytin, dibutyldimethoxytin, diethyldiethoxytin, dibutyldiethoxytin, dibutylphenoxytin, diphenylmethoxytin, dibutyltin acetate, tributyltin chloride, and tin 2-ethylhexanoate;

zinc compounds such as dimethoxyzinc, diethoxyzinc, ethylenedioxyzinc, and dibutoxyzinc;

aluminum compounds such as aluminum trimethoxide, aluminum triisopropoxide, and aluminum tributoxide;

titanium compounds such as tetramethoxytitanium, tetraethoxytitanium, tetrabutoxytitanium, dichlorodimethoxytitanium, 5 tetraisopropoxytitanium, titanium acetate, and titanium acetylacetonate;

phosphorus compounds such as trimethylphosphine, triethylphosphine, tributylphosphine, triphenylphosphine, tributylmethylphosphonium halides, trioctylbutylphosphonium halides, and triphenylmethylphosphonium halides;

10 zirconium compounds such as zirconium halides, zirconium acetylacetonate, zirconium alkoxides, and zirconium acetate;

lead and lead-containing compounds, for example lead oxides such as PbO , PbO_2 , and Pb_3O_4 ;

lead sulfides such as PbS , Pb_2S_3 , and PbS_2 ;

15 lead hydroxides such as $\text{Pb}(\text{OH})_2$, $\text{Pb}_3\text{O}_2(\text{OH})_2$, $\text{Pb}_2[\text{PbO}_2(\text{OH})_2]$, and $\text{Pb}_2\text{O}(\text{OH})_2$; plumbites such as Na_2PbO_2 , K_2PbO_2 , NaHPbO_2 , and KHPbO_2 ;

plumbates such as Na_2PbO_3 , $\text{Na}_2\text{H}_2\text{PbO}_4$, K_2PbO_3 , $\text{K}_2[\text{Pb}(\text{OH})_6]$, K_4PbO_4 , Ca_2PbO_4 , and CaPbO_3 ;

20 lead carbonates and basic salts thereof such as PbCO_3 and $2\text{PbCO}_3 \cdot \text{Pb}(\text{OH})_2$;

alkoxylead compounds and aryloxylead compounds such as $\text{Pb}(\text{OCH}_3)_2$, $(\text{CH}_3\text{O})\text{Pb}(\text{OPh})$, and $\text{Pb}(\text{OPh})_2$;

lead salts of organic acids, and carbonates and basic salts thereof, 25 such as $\text{Pb}(\text{OCOCH}_3)_2$, $\text{Pb}(\text{OCOCH}_3)_4$, and $\text{Pb}(\text{OCOCH}_3)_2 \cdot \text{PbO} \cdot 3\text{H}_2\text{O}$;

organolead compounds such as Bu_4Pb , Ph_4Pb , Bu_3PbCl , Ph_3PbBr , Ph_3Pb

(or Ph_6Pb_2), Bu_3PbOH , and Ph_2PbO (wherein Bu represents a butyl group, and Ph represents a phenyl group);

lead alloys such as Pb-Na, Pb-Ca, Pb-Ba, Pb-Sn, and Pb-Sb;

lead minerals such as galena and zinc blende; and

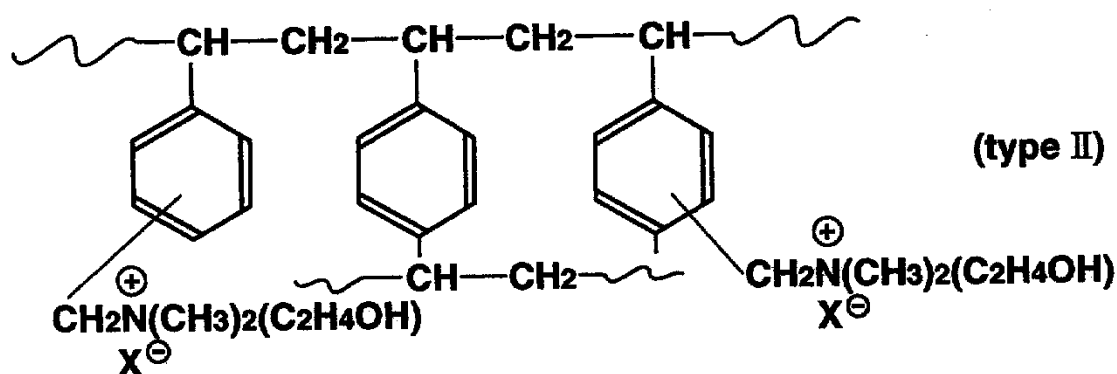
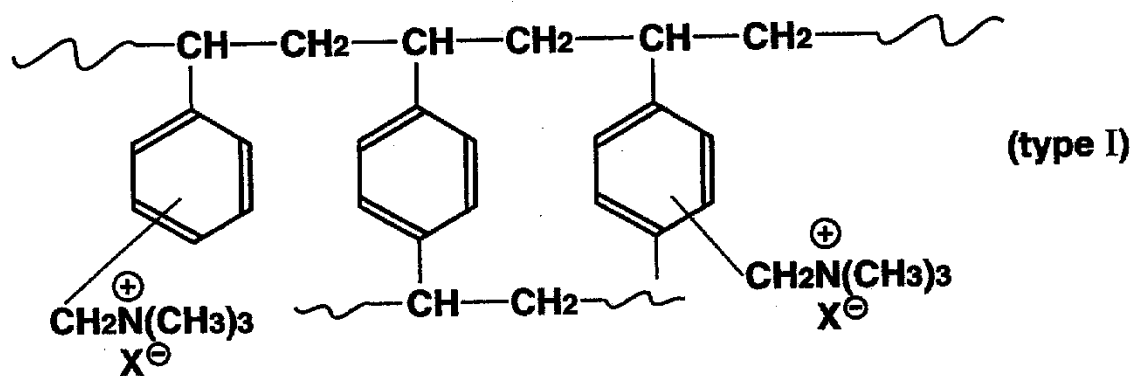
5 hydrates of such lead compounds.

In the case that the compound used dissolves in the starting material of the reaction, the reaction mixture, a reaction by-product or the like, the compound can be used as a homogeneous catalyst, whereas in the case that the compound does not dissolve, the compound can be
10 used as a solid catalyst. Furthermore, it is also preferable to use, as a homogeneous catalyst, a mixture obtained by dissolving a compound as above in a starting material of the reaction, the reaction mixture, a reaction by-product or the like in advance, or by reacting to bring about dissolution.

15 Furthermore, ion exchangers such as anion exchange resins having tertiary amino groups, ion exchange resins having amide groups, ion exchange resins having at least one type of exchange groups selected from sulfonate groups, carboxylate groups and phosphate groups, and solid strongly basic anion exchangers having quaternary
20 ammonium groups as exchange groups; solid inorganic compounds such as silica, silica-alumina, silica-magnesia, aluminosilicates, gallium silicate, various zeolites, various metal-exchanged zeolites, and ammonium-exchanged zeolites, and so on can also be used as the catalyst.

25 As a solid catalyst, a particularly preferably used one is a solid strongly basic anion exchanger having quaternary ammonium groups as

exchange groups, examples thereof including a strongly basic anion exchange resin having quaternary ammonium groups as exchange groups, a cellulose strongly basic anion exchanger having quaternary ammonium groups as exchange groups, and an inorganic carrier supported type strongly basic anion exchanger having quaternary ammonium groups as exchange groups. As a strongly basic anion exchange resin having quaternary ammonium groups as exchange groups, for example a styrene type strongly basic anion exchange resin or the like can be preferably used. A styrene type strongly basic anion exchange resin is a strongly basic anion exchange resin having a copolymer of styrene and divinylbenzene as a parent material, and having quaternary ammonium groups (type I or type II) as exchange groups, and can be schematically represented, for example, by the following formula:



wherein X represents an anion; as X, generally at least one type of anion

selected from F^- , Cl^- , Br^- , I^- , HCO_3^- , CO_3^{2-} , $CH_3CO_2^-$, HCO_2^- , IO_3^- , BrO_3^- , and ClO_3^- is used, preferably at least one type of anion selected from Cl^- , Br^- , HCO_3^- , and CO_3^{2-} . Moreover, as the structure of the resin parent material, either a gel type one or a macroreticular (MR) type one can be used, the MR type being particularly preferable due to the organic solvent resistance being high.

An example of a cellulose strongly basic anion exchanger having quaternary ammonium groups as exchange groups is cellulose having $-OCH_2CH_2NR_3X$ exchange groups obtained by converting some or all of the $-OH$ groups in the cellulose into trialkylaminoethyl groups. Here, R represents an alkyl group; methyl, ethyl, propyl, butyl or the like is generally used, preferably methyl or ethyl. Moreover, X represents an anion as above.

An inorganic carrier supported type strongly basic anion exchanger having quaternary ammonium groups as exchange groups means an inorganic carrier that has had $-O(CH_2)_nNR_3X$ quaternary ammonium groups introduced thereto by modifying some or all of the $-OH$ surface hydroxyl groups of the inorganic carrier. Here, R and X are defined as above. n is generally an integer from 1 to 6, preferably $n = 2$. As the inorganic carrier, silica, alumina, silica-alumina, titania, a zeolite, or the like can be used, it being preferable to use silica, alumina, or silica-alumina, particularly preferably silica. Any method can be used as the method of modifying the surface hydroxyl groups of the inorganic carrier.

As the solid strongly basic anion exchanger having quaternary ammonium groups as exchange groups, a commercially available one

may be used. In this case, the anion exchanger may also be used as the transesterification catalyst after being subjected to ion exchange with a desired anionic species in advance as pretreatment.

Moreover, a solid catalyst comprising a macroreticular or gel-type organic polymer having bonded thereto heterocyclic groups each containing at least one nitrogen atom, or an inorganic carrier having bonded thereto heterocyclic groups each containing at least one nitrogen atom can also be preferably used as the transesterification catalyst. Furthermore, a solid catalyst in which some or all of these nitrogen-containing heterocyclic groups have been converted into a quaternary salt can be similarly used. Note that the solid catalyst such as an ion exchanger may also act as a packing.

The amount of the catalyst used in the step (I) varies depending on the type of the catalyst used, but in the case of continuously feeding in a homogeneous catalyst that dissolves in the reaction liquid under the reaction conditions, the amount used is generally in a range of from 0.0001 to 50 % by weight, preferably from 0.005 to 20 % by weight, more preferably from 0.01 to 10 % by weight, as a proportion of the total weight of the cyclic carbonate and the aliphatic monohydric alcohol fed in as the starting materials. Moreover, in the case of using the solid catalyst installed in the distillation column, the catalyst is preferably used in an amount in a range of from 0.01 to 75 vol%, more preferably from 0.05 to 60 vol%, still more preferably from 0.1 to 60 vol%, based on the empty column volume of the distillation column.

There are no particular limitations on the method of continuously feeding the starting material cyclic carbonate and aliphatic monohydric

alcohol into the continuous multi-stage distillation column T_0 constituting the reactive distillation column used in step (I); any feeding method may be used so long as the cyclic carbonate and the aliphatic monohydric alcohol can be made to contact the catalyst in a region of at least 5 stages, preferably at least 7 stages, more preferably at least 10 stages, of the distillation column. That is, the cyclic carbonate and the aliphatic monohydric alcohol can be continuously fed in from a required number of inlets in stages of the continuous multi-stage distillation column satisfying the conditions described earlier. Moreover, the cyclic carbonate and the aliphatic monohydric alcohol may be introduced into the same stage of the distillation column, or may be introduced into different stages to one another.

The starting material cyclic carbonate and aliphatic monohydric alcohol are fed continuously into the continuous multi-stage distillation column T_0 in a liquid form, in a gaseous form, or as a mixture of the liquid and the gas. Other than feeding the starting materials into the distillation column in this way, it is also preferable to additionally feed in a gaseous starting material intermittently or continuously from a lower portion of the distillation column. Moreover, another preferable method is one in which the cyclic carbonate is continuously fed in a liquid form or a gas / liquid mixed form into a stage of the distillation column above the stages in which the catalyst is present, and the aliphatic monohydric alcohol is continuously fed in a gaseous form and / or a liquid form into the lower portion of the distillation column. In this case, the cyclic carbonate may of course contain the aliphatic monohydric alcohol.

In step (I), the starting materials fed in may contain the product

dialkyl carbonate and / or diol. The content thereof is, for the dialkyl carbonate, generally in a range of from 0 to 40 % by weight, preferably from 0 to 30 % by weight, more preferably from 0 to 20 % by weight, in terms of the percentage by weight of the dialkyl carbonate in the aliphatic monohydric alcohol / dialkyl carbonate mixture, and is, for the diol, generally in a range of from 0 to 10 % by weight, preferably from 0 to 7 % by weight, more preferably from 0 to 5 % by weight, in terms of the percentage by weight of the diol in the cyclic carbonate / diol mixture.

When carrying out the reaction of step (I) industrially, besides fresh cyclic carbonate and / or aliphatic monohydric alcohol newly introduced into the reaction system, material having the cyclic carbonate and / or the aliphatic monohydric alcohol as a main component thereof recovered from this process and / or another process can also be preferably used for the starting materials. It is an excellent characteristic feature of the present invention that this is possible. An example of another process is step (II) in which a diaryl carbonate is produced from the dialkyl carbonate and the aromatic monohydroxyl compound, the aliphatic monohydric alcohol being by-produced in step (II) and recovered. The recovered by-produced aliphatic monohydric alcohol generally often contains the dialkyl carbonate, the aromatic monohydroxy compound, an alkyl aryl ether and so on, and may also contain small amounts of an alkyl aryl carbonate, the dialkyl carbonate and so on. The by-produced aliphatic monohydric alcohol may be used as is as the starting material in step (I), or may be used as the starting material in step (I) after the amount of contained material having a higher boiling point than that of the aliphatic monohydric alcohol has been

reduced through distillation or the like.

A cyclic carbonate preferably used in step (I) is one produced through reaction between, for example, an alkylene oxide such as ethylene oxide, propylene oxide or styrene oxide and carbon dioxide; a
5 cyclic carbonate containing small amounts of such compounds or the like may be used as the starting material in step (I).

In step (I), a ratio between amounts of the cyclic carbonate and the aliphatic monohydric alcohol fed into the reactive distillation column varies according to the type and amount of the transesterification
10 catalyst and the reaction conditions, but a molar ratio of the aliphatic monohydric alcohol to the cyclic carbonate fed in is generally in a range of from 0.01 to 1000 times. To increase the cyclic carbonate conversion, it is preferable to feed in the aliphatic monohydric alcohol in an excess of at least 2 times the number of mols of the cyclic carbonate, but if the
15 amount of the aliphatic monohydric alcohol used is too great, then it is necessary to make the apparatus larger. For such reasons, the molar ratio of the aliphatic monohydric alcohol to the cyclic carbonate is preferably in a range of from 2 to 20, more preferably from 3 to 15, yet more preferably from 5 to 12. Furthermore, if much unreacted cyclic
20 carbonate remains, then the unreacted cyclic carbonate may react with the product diol to by-produce oligomers such as a dimer or a trimer, and hence in industrial implementation, it is preferable to reduce the amount of unreacted cyclic carbonate remaining as much as possible. In the process of the present invention, even if the above molar ratio is not
25 more than 10, the cyclic carbonate conversion can be made to be not less than 98%, preferably not less than 99%, more preferably not less

than 99.9%. This is another characteristic feature of the present invention.

In step (I), preferably not less than approximately 0.4 ton / hr of the dialkyl carbonate is continuously produced; the minimum amount of the cyclic carbonate continuously fed in to achieve this is generally 0.44 P ton / hr, preferably 0.42 P ton / hr, more preferably 0.4 P ton / hr, based on the amount P (ton / hr) of the dialkyl carbonate to be produced. In a yet more preferable case, this amount can be made to be less than 0.39 P ton / hr.

The continuous multi-stage distillation column T_0 used in step (I) comprises a structure having a cylindrical trunk portion having a length L_0 (cm) and an inside diameter D_0 (cm) and having an internal with a number of stages n_0 thereinside, and further having a gas outlet having an inside diameter d_{01} (cm) at a top of the column or in an upper portion of the column near to the top, a liquid outlet having an inside diameter d_{02} (cm) at a bottom of the column or in a lower portion of the column near to the bottom, at least one first inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one second inlet provided in the central portion and/or the lower portion of the column above the liquid outlet, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} must satisfy respectively the following formulae (1) to (6):

$$2100 \leq L_0 \leq 8000 \quad (1)$$

$$180 \leq D_0 \leq 2000 \quad (2)$$

$$4 \leq L_0 / D_0 \leq 40 \quad (3)$$

$$10 \leq n_0 \leq 120 \quad (4)$$

$$3 \leq D_0 / d_{01} \leq 20 \quad (5), \text{ and}$$

$$5 \leq D_0 / d_{02} \leq 30 \quad (6).$$

Note that the term "the top of the column or the upper portion of the column near to the top" used in the present invention means the portion from the top of the column downward as far as approximately 0.25 L_0 , and the term "the bottom of the column or the lower portion of the column near to the bottom" means the portion from the bottom of the column upward as far as approximately 0.25 L_0 . (Likewise for the first and second continuous multi-stage distillation columns, described later, but with 0.25 L_1 and 0.25 L_2 respectively.)

It has been discovered that by using the continuous multi-stage distillation column T_0 that simultaneously satisfies the formulae (1), (2), (3), (4), (5) and (6), the dialkyl carbonate and the diol can be produced on an industrial scale of preferably not less than 0.4 ton / hr of the dialkyl carbonate and / or preferably not less than 0.26 ton / hr of the diol with a high conversion, high selectivity, and high productivity stably for a prolonged period of time of, for example, not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours, from the cyclic carbonate and the aliphatic monohydric alcohol. The reason why it has become possible to produce the dialkyl carbonate and the diol on an industrial scale with such excellent effects by implementing step (I) as described above is not clear, but this is supposed to be due to a composite effect brought about when the conditions of the formulae (1) to (6) are combined. Preferable ranges for the respective factors are described below.

If L_0 (cm) is less than 2100, then the conversion decreases and hence it is not possible to attain the desired production amount.

Moreover, to keep down the equipment cost while securing a reaction ratio enabling the desired production amount to be attained, L_0 must be made to be not more than 8000. A more preferable range for L_0 (cm) is $2300 \leq L_0 \leq 6000$, with $2500 \leq L_0 \leq 5000$ being yet more preferable.

5 If D_0 (cm) is less than 180, then it is not possible to attain the desired production amount. Moreover, to keep down the equipment cost while attaining the desired production amount, D_0 must be made to be not more than 2000. A more preferable range for D_0 (cm) is $200 \leq D_0 \leq 1000$, with $210 \leq D_0 \leq 800$ being yet more preferable.

10 If L_0 / D_0 is less than 4 or greater than 40, then stable operation becomes difficult. In particular, if L_0 / D_0 is greater than 40, then the pressure difference between the top and bottom of the column becomes too great, and hence prolonged stable operation becomes difficult. Moreover, it becomes necessary to increase the temperature in the lower
15 portion of the column, and hence side reactions become liable to occur, bringing about a decrease in the selectivity. A more preferable range for L_0 / D_0 is $5 \leq L_0 / D_0 \leq 30$, with $7 \leq L_0 / D_0 \leq 20$ being yet more preferable.

If n_0 is less than 10, then the conversion decreases and hence it is not possible to attain the desired production amount. Moreover, to
20 keep down the equipment cost while securing the conversion enabling the desired production amount to be attained, n_0 must be made to be not more than 120. Furthermore, if n_0 is greater than 120, then the pressure difference between the top and bottom of the column becomes too great, and hence prolonged stable operation becomes difficult. Moreover, it
25 becomes necessary to increase the temperature in the lower portion of the column, and hence side reactions become liable to occur, bringing

about a decrease in the selectivity. A more preferable range for n_0 is $30 \leq n_0 \leq 100$, with $40 \leq n_0 \leq 90$ being yet more preferable.

If D_0 / d_{01} is less than 3, then the equipment cost becomes high. Moreover, a large amount of a gaseous component is readily released to the outside of the system, and hence stable operation becomes difficult. If D_0 / d_{01} is greater than 20, then the gaseous component withdrawal amount becomes relatively low, and hence stable operation becomes difficult, and moreover a decrease in the conversion is brought about. A more preferable range for D_0 / d_{01} is $4 \leq D_0 / d_{01} \leq 15$, with $5 \leq D_0 / d_{01} \leq 13$ being yet more preferable.

If D_0 / d_{02} is less than 5, then the equipment cost becomes high. Moreover, the liquid withdrawal amount becomes relatively high, and hence stable operation becomes difficult. If D_0 / d_{02} is greater than 30, then the flow rate through the liquid outlet and piping becomes excessively fast, and hence erosion becomes liable to occur, bringing about corrosion of the apparatus. A more preferable range for D_0 / d_{02} is $7 \leq D_0 / d_{02} \leq 25$, with $9 \leq D_0 / d_{02} \leq 20$ being yet more preferable.

Furthermore, it has been found that it is further preferable for d_{01} and d_{02} for the continuous multi-stage distillation column T_0 used in step (I) to satisfy the following formula (7):

$$1 \leq d_{01} / d_{02} \leq 5 \quad (7).$$

The term "prolonged stable operation" referred to in step (I) means that operation can be carried out continuously in a steady state based on the operating conditions with no flooding, weeping, clogging of piping, or erosion for not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours, and predetermined amounts of

the dialkyl carbonate and the diol can be produced while maintaining a high conversion, high selectivity, and high productivity.

The "selectivity" for each of the dialkyl carbonate and the diol in step (I) is based on the cyclic carbonate reacted. In the present invention, a high selectivity of not less than 95% can generally be attained, preferably not less than 97%, more preferably not less than 99%. Moreover, the "conversion" in step (I) generally indicates the cyclic carbonate conversion, in the present invention it being possible to make the cyclic carbonate conversion be not less than 95%, preferably not less than 97%, more preferably not less than 99%, yet more preferably not less than 99.5%, still more preferably not less than 99.9%. It is one of the excellent characteristic features of step (I) that the high conversion can be maintained while maintaining high selectivity in this way.

The continuous multi-stage distillation column T_0 used in step (I) is preferably a distillation column having trays and/or packings as the internals. The term "internal" used in the present invention means the parts in the distillation column where gas and liquid are actually brought into contact with one another. Examples of the tray include a bubble-cap tray, a sieve tray, a valve tray, a counterflow tray, a Superfrac tray, a Maxfrac tray, or the like. Examples of the packing include an irregular packing such as a Raschig ring, a Lessing ring, a Pall ring, a Berl saddle, an Intalox saddle, a Dixon packing, a McMahon packing or Heli-Pak, or a structured packing such as Mellapak, Gempak, Techno-pack, Flexipac, a Sulzer packing, a Goodroll packing or Glitschgrid. A multi-stage distillation column having both a tray portion

and a portion packed with the packings can also be used. Furthermore, the term "number of stages n of the internals" used in the present invention means the number of trays in the case of the trays, and the theoretical number of stages in the case of the packings.

5 In the case of a multi-stage distillation column having both a tray portion and a portion packed with the packing, the number of stages n is thus the sum of the number of trays and the theoretical number of stages.

In step (I) in which the cyclic carbonate and the aliphatic
10 monohydric alcohol are reacted together, it has been discovered that a high conversion, high selectivity, and high productivity can be attained even if a plate type continuous multi-stage distillation column in which the internals comprise trays and/or packings having a predetermined number of stages and/or a packed column type continuous multi-stage
15 distillation column is used, but the plate type distillation column in which the internals are trays is preferable. Furthermore, it has been discovered that sieve trays each having a sieve portion and a downcomer portion are particularly good as the trays in terms of the relationship between performance and equipment cost. It has also been discovered
20 that each sieve tray preferably has 100 to 1000 holes / m^2 in the sieve portion. A more preferable number of holes is from 120 to 900 holes / m^2 , yet more preferably from 150 to 800 holes / m^2 .

Moreover, it has been discovered that the cross-sectional area per hole of each sieve tray is preferably in a range of from 0.5 to 5 cm^2 . A
25 more preferable cross-sectional area per hole is from 0.7 to 4 cm^2 , yet more preferably from 0.9 to 3 cm^2 . Furthermore, it has been discovered

that it is particularly preferable if each sieve tray has 100 to 1000 holes / m^2 in the sieve portion, and the cross-sectional area per hole is in a range of from 0.5 to 5 cm^2 .

Furthermore, it has been discovered that an aperture ratio of each
5 of the sieve trays is preferably in a range of from 1.5 to 15%. A more preferable aperture ratio is in a range of from 1.7 to 13%, yet more preferably from 1.9 to 11%. Here, the "aperture ratio" of the sieve tray indicates the ratio of the cross-sectional area of all of the holes (the total cross-sectional area of the holes) present in the sieve tray to the area of
10 the sieve portion containing the cross-sectional area of all of the holes. The area of the sieve portion and / or the total cross-sectional area of the holes may differ between the sieve trays, in which case the aperture ratio of each of the sieve trays is preferably within the above range. Furthermore, the number of holes in the sieve portion may be the same
15 for all of the sieve trays, or may differ. It has been shown that by adding the above conditions to the continuous multi-stage distillation column T_0 , the object for step (I) can be attained more easily.

When carrying out step (I), a dialkyl carbonate and a diol are continuously produced by continuously feeding the cyclic carbonate and
20 the aliphatic monohydric alcohol as the starting materials into the continuous multi-stage distillation column in which the catalyst is present, carrying out reaction and distillation simultaneously in the column, continuously withdrawing a low boiling point reaction mixture containing the produced dialkyl carbonate from an upper portion of the column in a
25 gaseous form, and continuously withdrawing a high boiling point reaction mixture containing the diol from a lower portion of the column in a liquid

form.

Moreover, in step (I), as the continuous feeding of the starting material cyclic carbonate and aliphatic monohydric alcohol into the continuous multi-stage distillation column T_0 , the cyclic carbonate and the aliphatic monohydric alcohol may be fed in as a starting material mixture or separately, in a liquid form and/or a gaseous form, from inlet(s) provided in one place or a plurality of places in the upper portion or the middle portion of the column below the gas outlet in the upper portion of the distillation column. A method in which the cyclic carbonate or the starting material containing a large amount of the cyclic carbonate is fed into the distillation column in a liquid form from inlet(s) in the upper portion or the middle portion of the distillation column, and the aliphatic monohydric alcohol or the starting material containing a large amount of the aliphatic monohydric alcohol is fed into the distillation column in a gaseous form from inlet(s) provided in the middle portion or the lower portion of the column above the liquid outlet in the lower portion of the distillation column is also preferable.

The reaction time for the transesterification reaction carried out in step (I) is considered to equate to the average residence time of the reaction liquid in the continuous multi-stage distillation column T_0 . The reaction time varies depending on the form of the internals in the distillation column and the number of stages, the amounts of the starting materials fed in, the type and amount of the catalyst, the reaction conditions, and so on. The reaction time is generally in a range of from 0.1 to 20 hours, preferably from 0.5 to 15 hours, more preferably from 1 to 10 hours.

The reaction temperature in step (I) varies depending on the type of the starting material compounds used, and the type and amount of the catalyst. The reaction temperature is generally in a range of from 30 to 300 °C. It is preferable to increase the reaction temperature so as to increase the reaction rate. However, if the reaction temperature is too high, then side reactions become liable to occur. The reaction temperature is thus preferably in a range of from 40 to 250 °C, more preferably from 50 to 200 °C, yet more preferably from 60 to 150 °C. In the present invention, the reactive distillation can be carried out with the column bottom temperature set to not more than 150 °C, preferably not more than 130 °C, more preferably not more than 110 °C, yet more preferably not more than 100 °C. An excellent characteristic feature of step (I) is that the high conversion, high selectivity, and high productivity can be attained even with such a low column bottom temperature.

Moreover, the reaction pressure varies depending on the type of the starting material compounds used and the composition therebetween, the reaction temperature, and so on. The reaction pressure may be any of a reduced pressure, normal pressure, or an applied pressure, and is generally in a range of from 1 to 2×10^7 Pa, preferably from 10^3 to 10^7 Pa, more preferably from 10^4 to 5×10^6 Pa.

The reflux ratio used for the continuous multi-stage distillation column T_0 in step (I) is generally in a range of from 0 to 10, preferably from 0.01 to 5, more preferably from 0.05 to 3.

The material constituting the continuous multi-stage distillation column T_0 used in step (I) is generally a metallic material such as carbon steel or stainless steel. In terms of the quality of the dialkyl carbonate,

diol and high-purity diaryl carbonate to be produced, stainless steel is preferable.

In the present invention, next, a step (II) of continuously producing a diaryl carbonate on an industrial scale from the dialkyl carbonate produced in step (I) and the aromatic monohydroxy compound is carried out.

The dialkyl carbonate used in step (II) is a compound represented by the following formula as shown in the following formula:



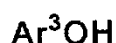
wherein R^2 is defined as before.

Examples of dialkyl carbonates having such an R^2 include dimethyl carbonate, diethyl carbonate, dipropyl carbonate (isomers), diallyl carbonate, dibutenyl carbonate (isomers), dibutyl carbonate (isomers), dipentyl carbonate (isomers), dihexyl carbonate (isomers), diheptyl carbonate (isomers), dioctyl carbonate (isomers), dinonyl carbonate (isomers), didecyl carbonate (isomers), dicyclopentyl carbonate, dicyclohexyl carbonate, dicycloheptyl carbonate, dibenzyl carbonate, diphenethyl carbonate (isomers), di(phenylpropyl) carbonate (isomers), di(phenylbutyl) carbonate (isomers), di(chlorobenzyl) carbonate (isomers), di(methoxybenzyl) carbonate (isomers), di(methoxymethyl) carbonate, di(methoxyethyl) carbonate (isomers), di(chloroethyl) carbonate (isomers) and di(cyanoethyl) carbonate (isomers).

Of these dialkyl carbonates, ones preferably used in the present invention are dialkyl carbonates in which R^2 is an alkyl group having not more than four carbon atoms and not containing a halogen atom. A

particularly preferable one is dimethyl carbonate. Moreover, of preferable dialkyl carbonates, particularly preferable ones are dialkyl carbonates produced in a state substantially not containing a halogen, for example ones produced from an alkylene carbonate substantially not containing a halogen and an alcohol substantially not containing a halogen.

The aromatic monohydroxy compound used in step (II) is a compound represented by the following formula. The type of the aromatic monohydroxy compound is not limited, so long as the hydroxyl group is directly bonded to the aromatic group:



wherein Ar^3 represents an aromatic group having 5 to 30 carbon atoms. Examples of aromatic monohydroxy compounds having such an Ar^3 include phenol; various alkylphenols such as cresol (isomers), xlenol (isomers), trimethylphenol (isomers), tetramethylphenol (isomers), ethylphenol (isomers), propylphenol (isomers), butylphenol (isomers), diethylphenol (isomers), methylethylphenol (isomers), methylpropylphenol (isomers), dipropylphenol (isomers), methylbutylphenol (isomers), pentylphenol (isomers), hexylphenol (isomers) and cyclohexylphenol (isomers); various alkoxyphenols such as methoxyphenol (isomers) and ethoxyphenol (isomers); arylalkylphenols such as phenylpropylphenol (isomers); naphthol (isomers) and various substituted naphthols; and heteroaromatic monohydroxy compounds such as hydroxypyridine (isomers), hydroxycoumarin (isomers) and hydroxyquinoline (isomers).

One of these aromatic monohydroxy compounds may be used, or

a mixture of a plurality may be used.

Of these aromatic monohydroxy compounds, ones preferably used in the present invention are aromatic monohydroxy compounds in which Ar^3 is an aromatic group having 6 to 10 carbon atoms. Phenol is particularly preferable. Moreover, of these aromatic monohydroxy compounds, ones substantially not containing a halogen are preferably used in the present invention.

The diaryl carbonate in the present invention is thus generally represented by the following formula:



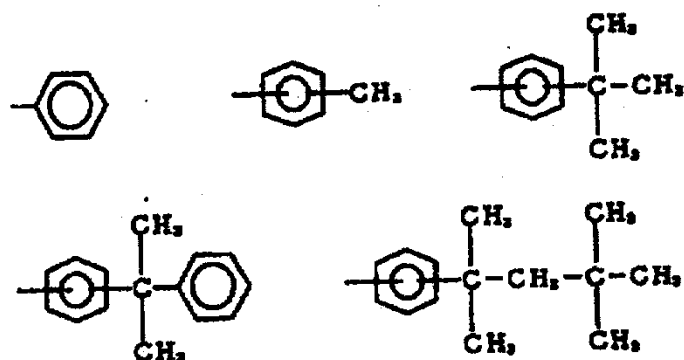
wherein each of Ar^3 and Ar^4 is a monovalent aromatic group.

Each of Ar^3 and Ar^4 represents a monovalent carbocyclic or heterocyclic aromatic group, and in each of Ar^3 and Ar^4 , one or more of the hydrogen atoms may be substituted with another substituent that does not have an adverse effect on the reaction, for example a halogen atom, an alkyl group having 1 to 10 carbon atoms, an alkoxy group having 1 to 10 carbon atoms, a phenyl group, a phenoxy group, a vinyl group, a cyano group, an ester group, an amide group, a nitro group, or the like. Ar^3 and Ar^4 may be the same as one another, or different.

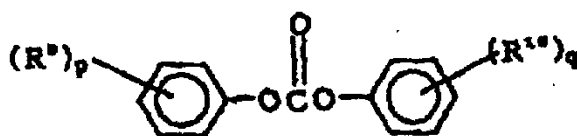
Typical examples of each of the monovalent aromatic groups Ar^3 and Ar^4 include a phenyl group, a naphthyl group, a biphenyl group, and a pyridyl group. Each of these may be substituted with one or a plurality of substituent(s) as above.

Preferable examples of each of Ar^3 and Ar^4 include those shown in

the following formula:



A particularly preferable diaryl carbonate is unsubstituted or substituted diphenyl carbonate represented by the following formula:



5

wherein each of R^9 and R^{10} independently represents a hydrogen atom, an alkyl group having 1 to 10 carbon atoms, an alkoxy group having 1 to 10 carbon atoms, a cycloalkyl group having 5 to 10 carbon atoms in the ring thereof, or a phenyl group, and each of p and q is an integer from 1 to 5; in the case that p is 2 or more, the R^9 's may be different from one another, and in the case that q is 2 or more, the R^{10} 's may be different from one another.

Among these diaryl carbonates, a symmetrical diaryl carbonate such as unsubstituted diphenyl carbonate or a lower alkyl group-substituted diphenyl carbonate such as ditolyl carbonate or di-*t*-butylphenyl carbonate is preferable, diphenyl carbonate, which has the simplest structure, being particularly preferable. One such diaryl carbonate may be used alone, or a plurality may be used in combination.

A molar ratio of the dialkyl carbonate to the aromatic monohydroxy compound used in the starting material in step (II) is preferably in a range of from 0.1 to 10. Outside this range, the amount of unreacted starting material remaining relative to a predetermined amount of the
5 desired diaryl carbonate produced becomes high, which is not efficient, and moreover much energy is required to recover the starting material. For such reasons, the above molar ratio is more preferably in a range of from 0.5 to 5, yet more preferably from 0.8 to 3, still more preferably from 1 to 2.

10 In the present invention, not less than 1 ton / hr of a high-purity diaryl carbonate is continuously produced, and to achieve this, the minimum amount of the aromatic monohydroxy compound fed in continuously in step (II) is generally 15'Q ton / hr, preferably 13 Q ton / hr, more preferably 10 Q ton / hr, based on the amount of the high-purity
15 diaryl carbonate (Q ton / hr) to be produced. In a more preferable case, this amount can be made to be less than 8 Q ton / hr.

The dialkyl carbonate and the aromatic monohydroxy compound used in the starting material in step (II) may each be of high purity, or may contain other compounds, for example may contain compounds or
20 reaction by-products produced in the first continuous multi-stage distillation column and / or the second continuous multi-stage distillation column. In the case of industrial implementation, for the starting material, besides fresh dialkyl carbonate and aromatic monohydroxy compound newly introduced into the reaction system, it is also preferable
25 to use dialkyl carbonate and / or aromatic monohydroxy compound recovered from the first continuous multi-stage distillation column and /

or the second continuous multi-stage distillation column, or recovered from another process. In the process of the present invention, column top components constituting a low boiling point reaction mixture from the second continuous multi-stage distillation column are fed into the first continuous multi-stage distillation column. Here, the second column low boiling point reaction mixture may be fed into the first continuous multi-stage distillation column as is, or after some of the components thereof have been separated out.

Accordingly, in the present invention, which is implemented industrially, it is preferable for the starting material fed into the first continuous multi-stage distillation column to contain any of an alcohol, an alkyl aryl carbonate, the diaryl carbonate, an alkyl aryl ether, and so on. The starting material further containing small amounts of high boiling point by-products such as Fries rearrangement products of the alkyl aryl carbonate and diaryl carbonate which are the products, derivatives thereof, and so on can also be preferably used. In the present invention, in the case, for example, of producing methyl phenyl carbonate and diphenyl carbonate using the starting material containing dimethyl carbonate as the dialkyl carbonate and phenol as the aromatic monohydroxy compound, it is preferable for this starting material to contain methanol, and methyl phenyl carbonate and diphenyl carbonate, which are the reaction products, and the starting material may further contain small amounts of anisole, phenyl salicylate and methyl salicylate, which are reaction by-products, and high boiling point by-products derived therefrom.

The high-purity diaryl carbonate of the present invention is

preferably used for producing an aromatic polycarbonate through polymerization with an aromatic dihydroxy compound, and in this polymerization step a large amount of the aromatic monohydroxy compound is by-produced and recovered. This by-produced aromatic monohydroxy compound is preferably used as the starting material in
5 step (II) of the present invention.

The diaryl carbonate produced in step (II) is obtained through transesterification between the dialkyl carbonate and the aromatic monohydroxy compound. Included under this transesterification are a
10 reaction in which one or both of the alkoxy groups of the dialkyl carbonate is / are exchanged with the aryloxy group of the aromatic monohydroxy compound and an alcohol is eliminated, and a reaction in which two molecules of the alkyl aryl carbonate produced are converted into the diaryl carbonate and the dialkyl carbonate through a
15 transesterification reaction therebetween, i.e. a disproportionation reaction. In step (II), in the first continuous multi-stage distillation column, the alkyl aryl carbonate is mainly obtained, and in the second continuous multi-stage distillation column, the diaryl carbonate and the dialkyl carbonate are mainly obtained through the disproportionation
20 reaction of the alkyl aryl carbonate. The diaryl carbonate obtained in step (II) does not contain a halogen at all, and hence is important as the starting material when industrially producing the aromatic polycarbonate in the present invention. The reason for this is that if a halogen is present in the starting material for the polymerization even in a small
25 amount of less than 1 ppm, then this may impede the polymerization reaction, thus impeding stable production of the aromatic polycarbonate,

or cause a deterioration of the properties of, or discoloration of, the aromatic polycarbonate produced.

As a catalyst used in the first continuous multi-stage distillation column and / or the second continuous multi-stage distillation column in
 5 step (II), for example, a metal-containing compound selected from the following compounds can be used:

<lead compounds>

lead oxides such as PbO , PbO_2 and Pb_3O_4 ;

lead sulfides such as PbS and Pb_2S ;

10 lead hydroxides such as $\text{Pb}(\text{OH})_2$ and $\text{Pb}_2\text{O}_2(\text{OH})_2$;

plumbites such as Na_2PbO_2 , K_2PbO_2 , NaHPbO_2 and KHPbO_2 ;

plumbates such as Na_2PbO_3 , $\text{Na}_2\text{H}_2\text{PbO}_4$, K_2PbO_3 , $\text{K}_2[\text{Pb}(\text{OH})_6]$,

K_4PbO_4 , Ca_2PbO_4 and CaPbO_3 ;

lead carbonates and basic salts thereof such as PbCO_3 and

15 $2\text{PbCO}_3 \cdot \text{Pb}(\text{OH})_2$;

lead salts of organic acids, and carbonates and basic salts thereof, such as $\text{Pb}(\text{OCOCH}_3)_2$, $\text{Pb}(\text{OCOCH}_3)_4$ and $\text{Pb}(\text{OCOCH}_3)_2 \cdot \text{PbO} \cdot 3\text{H}_2\text{O}$;

organolead compounds such as Bu_4Pb , Ph_4Pb , Bu_3PbCl , Ph_3PbBr , Ph_3Pb (or Ph_6Pb_2), Bu_3PbOH and Ph_3PbO (wherein Bu represents a butyl
 20 group, and Ph represents a phenyl group);

alkoxylead compounds and aryloxylead compounds such as $\text{Pb}(\text{OCH}_3)_2$, $(\text{CH}_3\text{O})\text{Pb}(\text{OPh})$ and $\text{Pb}(\text{OPh})_2$;

lead alloys such as Pb-Na, Pb-Ca, Pb-Ba, Pb-Sn and Pb-Sb;

lead minerals such as galena and zinc blende; and

25 hydrates of such lead compounds;

<copper family metal compounds>

salts and complexes of copper family metals such as CuCl , CuCl_2 , CuBr , CuBr_2 , CuI , CuI_2 , $\text{Cu}(\text{OAc})_2$, $\text{Cu}(\text{acac})_2$, copper oleate, Bu_2Cu , $(\text{CH}_3\text{O})_2\text{Cu}$, AgNO_3 , AgBr , silver picrate, $\text{AgC}_6\text{H}_6\text{ClO}_4$, $[\text{AuC}\equiv\text{C}-\text{C}(\text{CH}_3)_3]_n$ and $[\text{Cu}(\text{C}_7\text{H}_8)\text{Cl}]_4$ (wherein acac represents an acetylacetone chelate
 5 ligand);

<alkali metal complexes>

alkali metal complexes such as $\text{Li}(\text{acac})$ and $\text{LiN}(\text{C}_4\text{H}_9)_2$;

<zinc complexes>

zinc complexes such as $\text{Zn}(\text{acac})_2$;

10 <cadmium complexes>

cadmium complexes such as $\text{Cd}(\text{acac})_2$;

<iron family metal compounds>

complexes of iron family metals such as $\text{Fe}(\text{C}_{10}\text{H}_8)(\text{CO})_5$, $\text{Fe}(\text{CO})_5$, $\text{Fe}(\text{C}_4\text{H}_6)(\text{CO})_3$, $\text{Co}(\text{mesitylene})_2$, $(\text{PEt}_2\text{Ph})_2$, $\text{CoC}_5\text{F}_5(\text{CO})_7$, $\text{Ni}-\pi-\text{C}_5\text{H}_5\text{NO}$
 15 and ferrocene;

<zirconium complexes>

zirconium complexes such as $\text{Zr}(\text{acac})_4$ and zirconocene;

<lewis acid type compounds>

lewis acids and Lewis acid-forming transition metal compounds
 20 such as AlX_3 , TiX_3 , TiX_4 , VOX_3 , VX_5 , ZnX_2 , FeX_3 and SnX_4 (wherein X represents a halogen atom, an acetoxy group, an alkoxy group or an aryloxy group); and

<organo-tin compounds>

organo-tin compounds such as $(\text{CH}_3)_3\text{SnOCOCH}_3$,
 25 $(\text{C}_2\text{H}_5)_3\text{SnOCOC}_6\text{H}_5$, $\text{Bu}_3\text{SnOCOCH}_3$, $\text{Ph}_3\text{SnOCOCH}_3$, $\text{Bu}_2\text{Sn}(\text{OCOCH}_3)_2$,
 $\text{Bu}_2\text{Sn}(\text{OCOC}_{11}\text{H}_{23})_2$, $\text{Ph}_3\text{SnOCH}_3$, $(\text{C}_2\text{H}_5)_3\text{SnOPh}$, $\text{Bu}_2\text{Sn}(\text{OCH}_3)_2$,

$\text{Bu}_2\text{Sn}(\text{OC}_2\text{H}_5)_2$, $\text{Bu}_2\text{Sn}(\text{OPh})_2$, $\text{Ph}_2\text{Sn}(\text{OCH}_3)_2$, $(\text{C}_2\text{H}_5)_3\text{SnOH}$, Ph_3SnOH , Bu_2SnO , $(\text{C}_8\text{H}_{17})_2\text{SnO}$, Bu_2SnCl_2 and $\text{BuSnO}(\text{OH})$. Each of these catalysts may be a solid catalyst fixed inside the multi-stage distillation column, or may be a soluble catalyst that dissolves in the reaction system.

Each of these catalyst components may of course have been reacted with an organic compound present in the reaction system such as an aliphatic alcohol, an aromatic monohydroxy compound, the alkyl aryl carbonate, the diaryl carbonate or the dialkyl carbonate, or may have been subjected to heating treatment with the starting material or products prior to the reaction.

In the case of carrying out step (II) with a soluble catalyst that dissolves in the reaction system, the catalyst is preferably one having a high solubility in the reaction liquid under the reaction conditions. Examples of preferable catalysts in this sense include PbO , $\text{Pb}(\text{OH})_2$ and $\text{Pb}(\text{OPh})_2$; TiCl_4 , $\text{Ti}(\text{OMe})_4$, $(\text{MeO})\text{Ti}(\text{OPh})_3$, $(\text{MeO})_2\text{Ti}(\text{OPh})_2$, $(\text{MeO})_3\text{Ti}(\text{OPh})$ and $\text{Ti}(\text{OPh})_4$; SnCl_4 , $\text{Sn}(\text{OPh})_4$, Bu_2SnO and $\text{Bu}_2\text{Sn}(\text{OPh})_2$; FeCl_3 , $\text{Fe}(\text{OH})_3$ and $\text{Fe}(\text{OPh})_3$; and such catalysts that have been treated with phenol, the reaction liquid or the like. The catalyst used in the first continuous multi-stage distillation column and the catalyst used in the second continuous multi-stage distillation column may be the same as one another, or different.

The first continuous multi-stage distillation column used in step (II) comprises a structure having a cylindrical trunk portion having a length L_1 (cm) and an inside diameter D_1 (cm), and having internals with a number of stages n_1 thereinside, and has a gas outlet having an inside

diameter d_{11} (cm) at a top of the column or in an upper portion of the column near to the top, a liquid outlet having an inside diameter d_{12} (cm) at the bottom of the column or in a lower portion of the column near to the bottom, at least one third inlet provided in the upper portion and/or a central portion of the column below the gas outlet, and at least one fourth inlet provided in the middle portion and / or the lower portion of the column above the liquid outlet, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} must satisfy the following formulae (7) to (12) respectively:

$$1500 \leq L_1 \leq 8000 \quad (7),$$

$$100 \leq D_1 \leq 2000 \quad (8),$$

$$2 \leq L_1 / D_1 \leq 40 \quad (9),$$

$$20 \leq n_1 \leq 120 \quad (10),$$

$$5 \leq D_1 / d_{11} \leq 30 \quad (11), \text{ and}$$

$$3 \leq D_1 / d_{12} \leq 20 \quad (12).$$

Moreover, the second continuous multi-stage distillation column used in step (II) comprises a structure having a cylindrical trunk portion having a length L_2 (cm) and an inside diameter D_2 (cm), and having internals with a number of stages n_2 thereinside, and has a gas outlet having an inside diameter d_{21} (cm) at the top of the column or in an upper portion of the column near to the top, a liquid outlet having an inside diameter d_{22} (cm) at the bottom of the column or in a lower portion of the column near to the bottom, at least one fifth inlet provided in the upper portion and / or a middle portion of the column below the gas outlet, and at least one sixth inlet provided in the middle portion and / or the lower portion of the column above the liquid outlet, wherein L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} must satisfy the following formulae (13) to

(18) respectively:

$$1500 \leq L_2 \leq 8000 \quad (13),$$

$$100 \leq D_2 \leq 2000 \quad (14),$$

$$2 \leq L_2 / D_2 \leq 40 \quad (15),$$

$$5 \quad 10 \leq n_2 \leq 80 \quad (16),$$

$$2 \leq D_2 / d_{21} \leq 15 \quad (17), \text{ and}$$

$$5 \leq D_2 / d_{22} \leq 30 \quad (18).$$

It has been discovered that by using the first continuous multi-stage distillation column and the second continuous multi-stage distillation column that simultaneously satisfy all of the formulae (7) to (18), the diaryl carbonate can be produced from the dialkyl carbonate and the aromatic monohydroxy compound on an industrial scale of not less than 1 ton / hr with high selectivity and high productivity stably for a prolonged period of time, for example not less than 2000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours. The reason why it has become possible to produce the diaryl carbonate on an industrial scale with such excellent effects by implementing the process according to the present invention is not clear, but this is supposed to be due to a composite effect brought about when the conditions of the formulae (7) to (18) are combined. Preferable ranges for the respective factors for the continuous multi-stage distillation columns used in step (II) are described below.

If L_1 (cm) or L_2 (cm) is less than 1500, then the conversion decreases and hence it is not possible to attain the desired production amount. Moreover, to keep down the equipment cost while securing the conversion enabling the desired production amount to be attained, L_1 and

L_2 must each be made to be not more than 8000. More preferable ranges for L_1 (cm) and L_2 (cm) are respectively $2000 \leq L_1 \leq 6000$ and $2000 \leq L_2 \leq 6000$, with $2500 \leq L_1 \leq 5000$ and $2500 \leq L_2 \leq 5000$ being yet more preferable.

- 5 If D_1 (cm) or D_2 (cm) is less than 100, then it is not possible to attain the desired production amount. Moreover, to keep down the equipment cost while attaining the desired production amount, D_1 and D_2 must each be made to be not more than 2000. More preferable ranges for D_1 (cm) and D_2 (cm) are respectively $150 \leq D_1 \leq 1000$ and $150 \leq D_2 \leq$
 10 1000 , with $200 \leq D_1 \leq 800$ and $200 \leq D_2 \leq 800$ being yet more preferable.

For the first continuous multi-stage distillation column and the second continuous multi-stage distillation column, so long as D_1 and D_2 are within the above ranges, each of the columns may have the same inside diameter from the upper portion thereof to the lower portion
 15 thereof, or the inside diameter may differ for different portions. For example, for each of the continuous multi-stage distillation columns, the inside diameter of the upper portion of the column may be smaller than, or larger than, the inside diameter of the lower portion of the column.

If L_1 / D_1 or L_2 / D_2 is less than 2 or greater than 40, then stable
 20 operation becomes difficult. In particular, if L_1 / D_1 or L_2 / D_2 is greater than 40, then the pressure difference between the top and bottom of the column becomes too great, and hence prolonged stable operation becomes difficult. Moreover, it becomes necessary to increase the temperature in the lower portion of the column, and hence side reactions
 25 become liable to occur, bringing about a decrease in the selectivity.

More preferable ranges for L_1 / D_1 and L_2 / D_2 are respectively $3 \leq L_1 / D_1$

≤ 30 and $3 \leq L_2 / D_2 \leq 30$, with $5 \leq L_1 / D_1 \leq 15$ and $5 \leq L_2 / D_2 \leq 15$ being yet more preferable.

If n_1 is less than 20, then the conversion decreases and it is not possible to attain the desired production amount for the first continuous multi-stage distillation column. Moreover, to keep down the equipment cost while securing the conversion enabling the desired production amount to be attained, n_1 must be made to be not more than 120. Furthermore, if n_1 is greater than 120, then the pressure difference between the top and bottom of the column becomes too great, and hence prolonged stable operation of the first continuous multi-stage distillation column becomes difficult. Moreover, it becomes necessary to increase the temperature in the lower portion of the column, and hence side reactions become liable to occur, bringing about a decrease in the selectivity. A more preferable range for n_1 is $30 \leq n_1 \leq 100$, with $40 \leq n_1 \leq 90$ being yet more preferable.

Moreover, if n_2 is less than 10, then the conversion decreases and it is not possible to attain the desired production amount for the second continuous multi-stage distillation column. Moreover, to keep down the equipment cost while securing the conversion enabling the desired production amount to be attained, n_2 must be made to be not more than 80. Furthermore, if n_2 is greater than 80, then the pressure difference between the top and bottom of the column becomes too great, and hence prolonged stable operation of the second continuous multi-stage distillation column becomes difficult. Moreover, it becomes necessary to increase the temperature in the lower portion of the column, and hence side reactions become liable to occur, bringing about a decrease in the

selectivity. A more preferable range for n_2 is $15 \leq n_2 \leq 60$, with $20 \leq n_2 \leq 50$ being yet more preferable.

If D_1 / d_{11} is less than 5, then the equipment cost for the first continuous multi-stage distillation column becomes high. Moreover, large amounts of gaseous components are readily released to the outside of the system, and hence stable operation of the first continuous multi-stage distillation column becomes difficult. If D_1 / d_{11} is greater than 30, then the gaseous component withdrawal amount becomes relatively low, and hence stable operation becomes difficult, and moreover a decrease in the conversion is brought about. A more preferable range for D_1 / d_{11} is $8 \leq D_1 / d_{11} \leq 25$, with $10 \leq D_1 / d_{11} \leq 20$ being yet more preferable. Furthermore, if D_2 / d_{21} is less than 2, then the equipment cost for the second continuous multi-stage distillation column becomes high. Moreover, large amounts of gaseous components are readily released to the outside of the system, and hence stable operation of the second continuous multi-stage distillation column becomes difficult. If D_2 / d_{21} is greater than 15, then the gaseous component withdrawal amount becomes relatively low, and hence stable operation becomes difficult, and moreover a decrease in the conversion is brought about. A more preferable range for D_2 / d_{21} is $5 \leq D_2 / d_{21} \leq 12$, with $3 \leq D_2 / d_{21} \leq 10$ being yet more preferable.

If D_1 / d_{12} is less than 3, then the equipment cost for the first continuous multi-stage distillation column becomes high. Moreover, the liquid withdrawal amount becomes relatively high, and hence stable operation of the first continuous multi-stage distillation column becomes difficult. If D_1 / d_{12} is greater than 20, then the flow rate through the

liquid outlet and piping becomes excessively fast, and hence erosion becomes liable to occur, bringing about corrosion of the apparatus. A more preferable range for D_1 / d_{12} is $5 \leq D_1 / d_{12} \leq 18$, with $7 \leq D_1 / d_{12} \leq 15$ being yet more preferable. Furthermore, if D_2 / d_{22} is less than 5, then the equipment cost for the second continuous multi-stage distillation column becomes high. Moreover, the liquid withdrawal amount becomes relatively high, and hence stable operation of the second continuous multi-stage distillation column becomes difficult. If D_2 / d_{22} is greater than 30, then the flow rate through the liquid outlet and piping becomes excessively fast, and hence erosion becomes liable to occur, bringing about corrosion of the apparatus. A more preferable range for D_2 / d_{22} is $7 \leq D_2 / d_{22} \leq 25$, with $9 \leq D_2 / d_{22} \leq 20$ being yet more preferable.

Furthermore, it has been found that in step (II) it is further preferable for d_{11} and d_{12} to satisfy the following formula (20), and for d_{21} and d_{22} to satisfy the following formula (21):

$$1 \leq d_{12} / d_{11} \leq 5 \quad (20), \text{ and}$$

$$1 \leq d_{21} / d_{22} \leq 6 \quad (21).$$

The term "prolonged stable operation" for step (II) means that operation can be carried out continuously in a steady state based on the operating conditions with no flooding, weeping, clogging of piping, erosion or the like for not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours, and a predetermined amount of the diaryl carbonate can be produced while maintaining high selectivity.

A characteristic feature for step (II) is that the diaryl carbonate can be produced stably for a prolonged period of time with high

selectivity and with a high productivity of not less than 1 ton / hr, preferably not less than 2 ton / hr, more preferably not less than 3 ton / hr. Moreover, another characteristic feature for step (II) is that in the case that L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for the first continuous multi-stage distillation column satisfy respectively $2000 \leq L_1 \leq 6000$, $150 \leq D_1 \leq 1000$, $3 \leq L_1 / D_1 \leq 30$, $30 \leq n_1 \leq 100$, $8 \leq D_1 / d_{11} \leq 25$, and $5 \leq D_1 / d_{12} \leq 18$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for the second continuous multi-stage distillation column satisfy respectively $2000 \leq L_2 \leq 6000$, $150 \leq D_2 \leq 1000$, $3 \leq L_2 / D_2 \leq 30$, $15 \leq n_2 \leq 60$, $2.5 \leq D_2 / d_{21} \leq 12$, and $7 \leq D_2 / d_{22} \leq 25$, not less than 2 ton / hr, preferably not less than 2.5 ton / hr, more preferably not less than 3 ton / hr of the diaryl carbonate can be produced.

Furthermore, another characteristic feature for step (II) is that in the case that L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for the first continuous multi-stage distillation column satisfy respectively $2500 \leq L_1 \leq 5000$, $200 \leq D_1 \leq 800$, $5 \leq L_1 / D_1 \leq 15$, $40 \leq n_1 \leq 90$, $10 \leq D_1 / d_{11} \leq 25$, and $7 \leq D_1 / d_{12} \leq 15$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for the second continuous multi-stage distillation column satisfy respectively $2500 \leq L_2 \leq 5000$, $200 \leq D_2 \leq 800$, $5 \leq L_2 / D_2 \leq 10$, $20 \leq n_2 \leq 50$, $3 \leq D_2 / d_{21} \leq 10$, and $9 \leq D_2 / d_{22} \leq 20$, not less than 3 ton / hr, preferably not less than 3.5 ton / hr, more preferably not less than 4 ton / hr of the diaryl carbonate can be produced.

The "selectivity" for the diaryl carbonate in step (II) is based on the aromatic monohydroxy compound reacted. In step (II), a high selectivity of not less than 95% can generally be attained, preferably not less than 97%, more preferably not less than 98%.

Each of the first continuous multi-stage distillation column and the second continuous multi-stage distillation column used in step (II) is preferably a distillation column having trays and/or packings as the internals. The term "internals" used in the present invention means the parts in the distillation column where gas and liquid are actually brought into contact with one another. As the trays, ones as described in the section on step (I) are preferable. Moreover, the term "number of stages of the internals" has the same meaning as before.

In step (II), in the first continuous multi-stage distillation column, a reaction in which the alkyl aryl carbonate is produced from the dialkyl carbonate and the aromatic monohydroxy compound mainly occurs. This reaction has an extremely low equilibrium constant, and the reaction rate is slow, and hence it has been discovered that a plate type distillation column having trays as the internals is particularly preferable as the first continuous multi-stage distillation column used in the reactive distillation. Moreover, in the second continuous multi-stage distillation column, it is mainly a disproportionation reaction of the alkyl aryl carbonate that occurs. This reaction also has a low equilibrium constant, and a slow reaction rate. Even so, it has been discovered that a distillation column having both the packings and the trays as the internals is preferable as the second continuous multi-stage distillation column used in the reactive distillation. Furthermore, it has been discovered that a distillation column having the packings installed in the upper portion thereof and the trays installed in the lower portion thereof is preferable as the second continuous multi-stage distillation column. It has been discovered that the packings in the second continuous

multi-stage distillation column are preferably structured packings, and that of the structured packings, Mellapak is particularly preferable.

Furthermore, it has been discovered that, as each of the trays installed in the first continuous multi-stage distillation column and the
5 second continuous multi-stage distillation column, a sieve tray having a sieve portion and a downcomer portion is particularly good in terms of the relationship between performance and equipment cost. It has also been discovered that each sieve tray preferably has 100 to 1000 holes / m^2 in the sieve portion. A more preferable number of holes is from 120
10 to 900 holes/ m^2 , yet more preferably from 150 to 800 holes/ m^2 .

Moreover, it has been discovered that the cross-sectional area per hole of each sieve tray is preferably in a range of from 0.5 to 5 cm^2 . A more preferable cross-sectional area per hole is from 0.7 to 4 cm^2 , yet more preferably from 0.9 to 3 cm^2 . Furthermore, it has been discovered
15 that it is particularly preferable if each sieve tray has 100 to 1000 holes / m^2 in the sieve portion, and the cross-sectional area per hole is in a range of from 0.5 to 5 cm^2 . It has been shown that by adding the above conditions to the continuous multi-stage distillation columns, the object of the present invention can be attained more easily.

20 When carrying out step (II), the diaryl carbonate is continuously produced by continuously feeding the dialkyl carbonate and the aromatic monohydroxy compound constituting the starting material into the first continuous multi-stage distillation column in which the catalyst is present, carrying out reaction and distillation simultaneously in the first column,
25 continuously withdrawing a first column low boiling point reaction mixture containing a produced alcohol from an upper portion of the first column in

a gaseous form, continuously withdrawing a first column high boiling point reaction mixture containing a produced alkyl aryl carbonate from a lower portion of the first column in a liquid form, continuously feeding the first column high boiling point reaction mixture into the second continuous multi-stage distillation column in which the catalyst is present, carrying out reaction and distillation simultaneously in the second column, continuously withdrawing a second column low boiling point reaction mixture containing a produced dialkyl carbonate from an upper portion of the second column in a gaseous form, continuously withdrawing a second column high boiling point reaction mixture containing a produced diaryl carbonate from a lower portion of the second column in a liquid form, and continuously feeding the second column low boiling point reaction mixture containing the dialkyl carbonate into the first continuous multi-stage distillation column.

As mentioned earlier, the starting material may contain the alcohol, the alkyl aryl carbonate and the diaryl carbonate that are reaction products, and reaction by-products such as an alkyl aryl ether and high boiling point compounds. Considering the equipment and cost required for separation and purification in other processes, when actually implementing the present invention industrially, it is preferable for the starting material to contain small amounts of such compounds.

In step (II), when continuously feeding the dialkyl carbonate and the aromatic monohydroxy compound constituting the starting material into the first continuous multi-stage distillation column, this starting material may be fed into the first distillation column in a liquid form and / or a gaseous form from inlet(s) provided in one or a plurality of positions

in the upper portion or the middle portion of the first distillation column below the gas outlet in the upper portion of the first distillation column.

It is also preferable to feed the starting material containing a large proportion of the phenol into the first distillation column in a liquid form from an inlet provided in the upper portion of the first distillation column, and feed a starting material containing a large proportion of the dialkyl carbonate into the first distillation column in a gaseous form from an inlet provided in the lower portion of the first distillation column above the liquid outlet in the lower portion of the first distillation column.

Moreover, in step (II), the first column high boiling point reaction mixture containing the alkyl aryl carbonate continuously withdrawn from the lower portion of the first continuous multi-stage distillation column is continuously fed into the second continuous multi-stage distillation column. Here, the first column high boiling point reaction mixture is preferably fed into the second distillation column in a liquid form and/or a gaseous form from inlet(s) provided in one or a plurality of positions in the upper portion or the middle portion of the second distillation column below the gas outlet in the upper portion of the second distillation column.

Moreover, in the case of using, as the second distillation column, a distillation column having a packing portion in the upper portion thereof and a tray portion in the lower portion thereof, which is a preferable embodiment of the present invention, it is preferable for at least one position where an inlet is provided to be between the packing portion and the tray portion. Moreover, in the case that the packings comprise a plurality of sets of structured packings, it is preferable for an inlet to be installed in a space between the sets of structured packings.

Moreover, in step (II), it is also preferable to carry out a reflux operation of condensing the gaseous component withdrawn from the top of each of the first continuous multi-stage distillation column and the second continuous multi-stage distillation column, and then returning
5 some of this component into the upper portion of that distillation column. In this case, the reflux ratio for the first continuous multi-stage distillation column is in a range of from 0 to 10, preferably from 0 to 5, more preferably from 0 to 3. The reflux ratio for the second continuous multi-stage distillation column is in a range of from 0.01 to 10, preferably
10 from 0.08 to 5, more preferably from 0.1 to 2. For the first continuous multi-stage distillation column, not carrying out such a reflux operation (i.e. reflux ratio = 0) is also a preferable embodiment.

In step (II), the method of making the catalyst be present in the first continuous multi-stage distillation column may be any method. In
15 the case that the catalyst is a solid that is insoluble in the reaction liquid, it is preferable for the catalyst to be fixed inside the column by, for example, being installed on stages inside the first continuous multi-stage distillation column or being installed in the form of a packing. In the case of a catalyst that dissolves in the starting material or the reaction
20 liquid, it is preferable to feed the catalyst into the first distillation column from a position above the middle portion of the first distillation column. In this case, a catalyst solution obtained by dissolving the catalyst in the starting material or reaction liquid may be introduced into the column together with the starting material, or may be introduced into the column
25 from a different inlet to the starting material. The amount of the catalyst used in the first continuous multi-stage distillation column in the present

invention varies depending on the type of catalyst used, the types and proportions of the starting material compounds, and reaction conditions such as the reaction temperature and the reaction pressure. The amount of the catalyst is generally in a range of from 0.0001 to 30 % by weight, preferably from 0.0005 to 10 % by weight, more preferably from 0.001 to 1 % by weight, based on the total weight of the starting material.

Moreover, in step (II), the method of making the catalyst be present in the second continuous multi-stage distillation column may be any method. In the case that the catalyst is a solid that is insoluble in the reaction liquid, it is preferable for the catalyst to be fixed inside the column by, for example, being installed on stages inside the second continuous multi-stage distillation column or being installed in the form of a packing. In the case of a catalyst that dissolves in the starting material or the reaction liquid, it is preferable to feed the catalyst into the second distillation column from a position above the middle portion of the second distillation column. In this case, a catalyst solution obtained by dissolving the catalyst in the starting material or reaction liquid may be introduced into the column together with the starting material, or may be introduced into the column from a different inlet to the starting material.

The amount of the catalyst used in the second continuous multi-stage distillation column in the present invention varies depending on the type of catalyst used, the types and proportions of the starting material compounds, and reaction conditions such as the reaction temperature and the reaction pressure. The amount of the catalyst is generally in a range of from 0.0001 to 30 % by weight, preferably from 0.0005 to 10 % by weight, more preferably from 0.001 to 1 % by weight, based on the

total weight of the starting material.

In step (II), the catalyst used in the first continuous multi-stage distillation column and the catalyst used in the second continuous multi-stage distillation column may be the same or different, but are preferably the same. More preferably, the same catalyst is used in both columns, this catalyst being one that dissolves in the reaction liquid in both columns. In this case, the catalyst dissolved in the high boiling point reaction mixture in the first continuous multi-stage distillation column is generally withdrawn from the lower portion of the first distillation column together with the alkyl aryl carbonate and so on, and fed into the second continuous multi-stage distillation column as is; this is a preferable embodiment. If necessary, more of the catalyst can be newly added into the second continuous multi-stage distillation column.

The reaction times for the transesterification reactions carried out in step (II) are considered to equate to the average residence times of the reaction liquids in the first continuous multi-stage distillation column and the second continuous multi-stage distillation column. Each of these reaction times varies depending on the form of the internals in that distillation column and the number of stages, the amount of the starting material fed into the column, the type and amount of the catalyst, the reaction conditions, and so on. The reaction time in each of the first continuous multi-stage distillation column and the second continuous multi-stage distillation column is generally in a range of from 0.01 to 10 hours, preferably from 0.05 to 5 hours, more preferably from 0.1 to 3 hours.

The reaction temperature in the first continuous multi-stage

distillation column varies depending on the type of the starting material compounds used, and the type and amount of the catalyst. This reaction temperature is generally in a range of from 100 to 350°C. It is preferable to increase the reaction temperature so as to increase the reaction rate. However, if the reaction temperature is too high, then side reactions become liable to occur, for example production of by-products such as an alkyl aryl ether increases, which is undesirable. For this reason, the reaction temperature in the first continuous multi-stage distillation column is preferably in a range of from 130 to 280°C, more preferably from 150 to 260°C, yet more preferably from 180 to 250°C.

The reaction temperature in the second continuous multi-stage distillation column varies depending on the type of the starting material compounds used, and the type and amount of the catalyst. This reaction temperature is generally in a range of from 100 to 350°C. It is preferable to increase the reaction temperature so as to increase the reaction rate. However, if the reaction temperature is too high, then side reactions become liable to occur, for example production of by-products such as an alkyl aryl ether, and Fries rearrangement products of the starting material compounds and the produced alkyl aryl carbonate and diaryl carbonate, and derivatives thereof increases, which is undesirable. For this reason, the reaction temperature in the second continuous multi-stage distillation column is preferably in a range of from 130 to 280°C, more preferably from 150 to 260°C, yet more preferably from 180 to 250°C.

Moreover, the reaction pressure in the first continuous multi-stage

distillation column varies depending on the type of the starting material compounds used and the composition of the starting material, the reaction temperature, and so on. The first continuous multi-stage distillation column may be at any of a reduced pressure, normal pressure, 5 or an applied pressure. The pressure at the top of the column is generally in a range of from 0.1 to 2×10^7 Pa, preferably from 10^5 to 10^7 Pa, more preferably from 2×10^5 to 5×10^6 Pa.

The reaction pressure in the second continuous multi-stage distillation column varies depending on the type of the starting material 10 compounds used and the composition of the starting material, the reaction temperature, and so on. The second continuous multi-stage distillation column may be at any of a reduced pressure, normal pressure, or an applied pressure. The pressure at the top of the column is generally in a range of from 0.1 to 2×10^7 Pa, preferably from 10^3 to 10^6 15 Pa, more preferably from 5×10^3 to 10^5 Pa.

Furthermore, as the first continuous multi-stage distillation column in step (II), a plurality of distillation columns may be used. In this case, the distillation columns may be linked together in series, in parallel, or in a combination of series and parallel. Moreover, as the second 20 continuous multi-stage distillation column in step (II), a plurality of distillation columns may be used. In this case, the distillation columns may be linked together in series, in parallel, or in a combination of series and parallel.

The material constituting each of the first continuous multi-stage 25 distillation column and the second continuous multi-stage distillation column used in step (II) is generally a metallic material such as carbon

steel or stainless steel. In terms of the quality of the aromatic carbonates produced, stainless steel is preferable.

The second column high boiling point reaction mixture continuously withdrawn from the lower portion of the second continuous multi-stage distillation column in a liquid form in step (II) has the diaryl carbonate as a main component thereof, but generally also contains unreacted alkyl aryl carbonate, a small amount of unreacted starting material, a small amount of high boiling point by-products and so on, and moreover in the case that a homogeneous catalyst is used, also contains this catalyst component. It is thus necessary to carry out a purification step (III) for obtaining a high-purity diaryl carbonate from the second column high boiling point reaction mixture. In step (III), any process may be used, so long as the high-purity diaryl carbonate can be obtained from the second column high boiling point reaction mixture. Examples include distillation and / or recrystallization and so on. It has been discovered that, of these, in the present invention, it is particularly preferable to carry out step (III) using a distillation process.

Furthermore, in the present invention, it has been discovered that it is further preferable for step (III) to comprise a distillation separation process in which two distillation columns (a high boiling point matter separating column, and a diaryl carbonate purifying column having a side cut outlet) are used, separation is carried out continuously in the high boiling point matter separating column into a column top component having as a main component thereof the diaryl carbonate, unreacted alkyl aryl carbonate, and a small amount of unreacted starting material, and a column bottom component having as a main component thereof a small

amount of high boiling point by-products and so on and/or a catalyst component, the column top component from the high boiling point matter separating column is continuously fed into the diaryl carbonate purifying column, and separation is carried out continuously in the diaryl carbonate purifying column into three components, i.e. a column top component, a side cut component, and a column bottom component, the high-purity diaryl carbonate being obtained as the side cut component.

Furthermore, it is preferable for all or some of the column bottom component from the high boiling point matter separating column to be circulated back into the first continuous multi-stage distillation column and / or the second continuous multi-stage distillation column and reused as catalyst component in step (II). Moreover, the column top component from the diaryl carbonate purifying column generally contains a small amount of the diaryl carbonate, and hence a method in which, with the column top component as is or after carrying out other distillation separation, some or all of the column bottom component is returned into the high boiling point matter separating column and / or the diaryl carbonate purifying column, so as to be recovered as high-purity diaryl carbonate is also preferable.

In step (III), a high-purity diaryl carbonate of purity generally not less than 99.9%, preferably not less than 99.99%, is obtained. The content of high boiling point by-products therein is generally not more than 100 ppm, preferably not more than 50 ppm, more preferably not more than 10 ppm. Moreover, in the present invention, a starting material and catalyst not containing a halogen are generally used, and hence the halogen content of the high-purity diaryl carbonate obtained is

not more than 0.1 ppm, preferably not more than 10 ppb, more preferably not more than 1 ppb, which is out of the detection limit by an ion chromatography.

5 EXAMPLES

Following is a more detailed description of the present invention through Examples. However, the present invention is not limited to the following Examples.

The halogen content was measured using an ion chromatography
10 method.

Example 1:

(1) Step (I) of continuously producing dimethyl carbonate and ethylene glycol

15 <Continuous multi-stage distillation column T_0 >

A continuous multi-stage distillation column as shown in FIG. 1 having $L_0 = 3300$ cm, $D_0 = 300$ cm, $L_0 / D_0 = 11$, $n_0 = 60$, $D_0 / d_{01} = 7.5$, and $D_0 / d_{02} = 12$ was used. In this example, as the internals, sieve trays each having a cross-sectional area per hole in the sieve portion
20 thereof of approximately 1.3 cm^2 and a number of holes of approximately 180 to 320 / m^2 were used.

<Reactive distillation>

3.27 Ton / hr of ethylene carbonate in a liquid form was continuously introduced into the distillation column T_0 from an inlet (3-a)
25 provided at the 55th stage from the bottom. 3.238 Ton / hr of methanol in a gaseous form (containing 8.96 % by weight of dimethyl carbonate)

and 7.489 ton / hr of methanol in a liquid form (containing 6.66 % by weight of dimethyl carbonate) were respectively continuously introduced into the distillation column T_0 from inlets (3-b and 3-c) provided at the 31st stage from the bottom. The molar ratio of the starting materials introduced into the distillation column T_0 was methanol / ethylene carbonate = 8.36.

The catalyst used was obtained by adding 4.8 ton of ethylene glycol to 2.5 ton of KOH (48 % by weight aqueous solution), heating to approximately 130 °C, gradually reducing the pressure, and carrying out heat treatment for approximately 3 hours at approximately 1300 Pa, so as to produce a homogeneous solution. This catalyst solution was continuously introduced into the distillation column T_0 from an inlet (3-e) provided at the 54th stage from the bottom (K concentration: 0.1 % by weight based on ethylene carbonate fed in). Reactive distillation was carried out continuously under conditions of a column bottom temperature of 98 °C, a column top pressure of approximately 1.118×10^5 Pa, and a reflux ratio of 0.42.

It was possible to attain stable steady state operation after 4 hours. A low boiling point reaction mixture withdrawn from the top 1 of the column in a gaseous form was cooled using a heat exchanger and thus turned into a liquid. The liquid low boiling point reaction mixture, which was continuously withdrawn from the distillation column at 10.678 ton / hr, contained 4.129 ton / hr of dimethyl carbonate, and 6.549 ton / hr of methanol. A liquid continuously withdrawn from the bottom 2 of the column at 3.382 ton / hr contained 2.356 ton / hr of ethylene glycol, 1.014 ton / hr of methanol, and 4 kg / hr of unreacted ethylene carbonate.

Excluding the dimethyl carbonate contained in the starting material, the actual produced amount of dimethyl carbonate was 3.340 ton / hr, and excluding the ethylene glycol contained in the catalyst solution, the actual produced amount of ethylene glycol was 2.301 ton / hr. The
 5 ethylene carbonate conversion was 99.88%, the dimethyl carbonate selectivity was not less than 99.99%, and the ethylene glycol selectivity was not less than 99.99%.

Prolonged continuous operation was carried out under these conditions. After 500 hours, 2000 hours, 4000 hours, 5000 hours, and
 10 6000 hours, the actual produced amounts per hour were 3.340 ton, 3.340 ton, 3.340 ton, 3.340 ton, and 3.340 ton respectively for dimethyl carbonate, and 2.301 ton, 2.301 ton, 2.301 ton, 2.301 ton, and 2.301 ton respectively for ethylene glycol, the ethylene carbonate conversions were respectively 99.90%, 99.89%, 99.89%, 99.88%, and 99.88%, the dimethyl
 15 carbonate selectivities were respectively not less than 99.99%, not less than 99.99%, not less than 99.99%, not less than 99.99%, and not less than 99.99%, and the ethylene glycol selectivities were respectively not less than 99.99%, not less than 99.99%, not less than 99.99%, not less than 99.99%, and not less than 99.99%.

20 (2) Step (II) of continuously producing diphenyl carbonate
 <First continuous multi-stage distillation column 101>

A continuous multi-stage distillation column as shown in FIG. 2 having $L_1 = 3300$ cm, $D_1 = 500$ cm, $L_1 / D_1 = 6.6$, $n_1 = 80$, $D_1 / d_{11} = 17$, and $D_1 / d_{12} = 9$ was used. In this Example, sieve trays each having a
 25 cross-sectional area per hole of approximately 1.5 cm^2 and a number of holes of approximately $250 / \text{m}^2$ were used as the internals.

<Second continuous multi-stage distillation column 201>

A continuous multi-stage distillation column as shown in FIG. 3 having $L_2 = 3100$ cm, $D_2 = 500$ cm, $L_2 / D_2 = 6.2$, $n_2 = 30$, $D_2 / d_{21} = 3.85$, and $D_2 / d_{22} = 11.1$ was used. In this Example, as the internals, two sets of Mellapak (11 theoretical stages in total) were installed in the upper portion, and sieve trays each having a cross-sectional area per hole of approximately 1.3 cm^2 and a number of holes of approximately $250 / \text{m}^2$ were used in the lower portion.

<Reactive distillation>

Diphenyl carbonate was produced by carrying out reactive distillation using an apparatus in which the first continuous multi-stage distillation column 101 and the second continuous multi-stage distillation column 201 were connected together as shown in FIG. 4.

A starting material 1 containing phenol and dimethyl carbonate in a weight ratio of phenol / dimethyl carbonate = 1.9 was introduced continuously in a liquid form at a flow rate of 50 ton / hr from an upper inlet 11 of the first continuous multi-stage distillation column 101. On the other hand, a starting material 2 containing dimethyl carbonate and phenol in a weight ratio of dimethyl carbonate / phenol = 3.6 was introduced continuously in a gaseous form at a flow rate of 50 ton / hr from a lower inlet 12 of the first continuous multi-stage distillation column 101. The molar ratio for the starting materials introduced into the first continuous multi-stage distillation column 101 was dimethyl carbonate / phenol = 1.35. The starting materials substantially did not contain halogens (outside the detection limit for the ion chromatography, i.e. not more than 1 ppb). $\text{Pb}(\text{OPh})_2$ as a catalyst was introduced from the

upper inlet 11 of the first continuous multi-stage distillation column 101 such that a concentration thereof in the reaction liquid would be approximately 100 ppm. Reactive distillation was carried out continuously under conditions of a temperature at the bottom of the first continuous multi-stage distillation column 101 of 225 °C, a pressure at the top of the column of 7×10^5 Pa and a reflux ratio of 0. A first column low boiling point reaction mixture containing methanol, dimethyl carbonate, phenol and so on was continuously withdrawn in a gaseous form from the top 13 of the first column, was passed through a heat exchanger 14, and was withdrawn at a flow rate of 34 ton / hr from an outlet 16. On the other hand, a first column high boiling point reaction mixture containing methyl phenyl carbonate, dimethyl carbonate, phenol, diphenyl carbonate, the catalyst and so on was continuously withdrawn in a liquid form from the bottom 17 of the first column.

A stable steady state was attained after 24 hours. The first column high boiling point reaction mixture was thus then fed continuously as is into the second continuous multi-stage distillation column 201 at a flow rate of 66 ton / hr from a starting material inlet 21 installed between the Mellapak and the sieve trays. The liquid fed into the second continuous multi-stage distillation column 201 contained 18.2 % by weight of methyl phenyl carbonate and 0.8 % by weight of diphenyl carbonate. Reactive distillation was carried out continuously in the second continuous multi-stage distillation column 201 under conditions of a temperature at the bottom of the column of 210°C, a pressure at the top of the column of 3×10^4 Pa, and a reflux ratio of 0.3. It was possible to attain stable steady state operation after 24 hours. A second column

low boiling point reaction mixture containing 35 % by weight of dimethyl carbonate and 56 % by weight of phenol was continuously withdrawn from the top 23 of the second column, the flow rate at an outlet 26 being 55.6 ton / hr, and a second column high boiling point reaction mixture containing 38.4 % by weight of methyl phenyl carbonate and 55.6 % by weight of diphenyl carbonate was continuously withdrawn from the bottom 27 of the second column. The second column low boiling point reaction mixture was continuously fed into the first continuous multi-stage distillation column 101 from the inlet 11. At this time, the amounts of fresh dimethyl carbonate and phenol newly fed into the first continuous multi-stage distillation column 101 were adjusted so as to maintain the above compositions and amounts of the starting material 1 and the starting material 2, taking into consideration the composition and amount of the second column low boiling point reaction mixture. It was found that the amount of diphenyl carbonate produced per hour was 5.74 ton. The selectivity for the diphenyl carbonate based on the phenol reacted was 98%.

Prolonged continuous operation was carried out under these conditions. The produced amounts of diphenyl carbonate per hour after 500 hours, 2000 hours, 4000 hours, 5000 hours, and 6000 hours (excluding the diphenyl carbonate contained in the starting material) were 5.74 ton, 5.75 ton, 5.74 ton, 5.74 ton, and 5.75 ton respectively, and the selectivities were 98%, 98%, 98%, 98%, and 98% respectively, and hence the operation was very stable. Moreover, the aromatic carbonates produced substantially did not contain halogens (not more than 1 ppb).

(3) Step (III) of obtaining high-purity diphenyl carbonate

The second column high boiling point reaction mixture withdrawn from the bottom of the second continuous multi-stage distillation column was continuously fed into a high boiling point matter separating column (length 1700 cm, inside diameter 340 cm, 30 stages), and distillation was carried out continuously with a temperature at the bottom of the column of 206 °C, a pressure at the top of the column of 3800 Pa, and a reflux ratio of 0.6. A column top component continuously withdrawn from the top of the high boiling point material separating column was continuously fed as is into an inlet of a diaryl carbonate purifying column having a side cut outlet (length 2200 cm, inside diameter 280 cm, 12 stages above inlet, 18 stages between inlet and side cut outlet installed therebelow, 5 stages below side cut outlet). In the diaryl carbonate purifying column, distillation was carried out continuously with a temperature at the bottom of the column of 213 °C, a pressure at the top of the column of 5000 Pa, and a reflux ratio of 1.5. The purity of diphenyl carbonate continuously withdrawn from the side cut outlet was not less than 99.999%, and the halogen content was not more than 1 ppb.

20 Example 2:

(1) Step (I) of continuously producing dimethyl carbonate and ethylene glycol

Reactive distillation was carried out under the following conditions using the same continuous multi-stage distillation column as in Example

25 1.

2.61 Ton / hr of ethylene carbonate in a liquid form was

continuously introduced into the distillation column from the inlet (3-a) provided at the 55th stage from the bottom. 4.233 Ton / hr of methanol in a gaseous form (containing 2.41 % by weight of dimethyl carbonate) and 4.227 ton / hr of methanol in a liquid form (containing 1.46 % by weight of dimethyl carbonate) were respectively continuously introduced into the distillation column from the inlets (3-b and 3-c) provided at the 31st stage from the bottom. The molar ratio of the starting materials introduced into the distillation column was methanol / ethylene carbonate = 8.73. The catalyst was made to be the same as in Example 1, and was continuously fed into the distillation column. Reactive distillation was carried out continuously under conditions of a column bottom temperature of 93 °C, a column top pressure of approximately 1.046×10^5 Pa, and a reflux ratio of 0.48.

It was possible to attain stable steady state operation after 24 hours. A low boiling point reaction mixture withdrawn from the top 1 of the column in a gaseous form was cooled using a heat exchanger and thus turned into a liquid. The liquid low boiling point reaction mixture, which was continuously withdrawn from the distillation column at 8.17 ton / hr, contained 2.84 ton / hr of dimethyl carbonate, and 5.33 ton / hr of methanol. A liquid continuously withdrawn from the bottom 2 of the column at 2.937 ton / hr contained 1.865 ton / hr of ethylene glycol, 1.062 ton / hr of methanol, and 0.2 kg / hr of unreacted ethylene carbonate. Excluding the dimethyl carbonate contained in the starting material, the actual produced amount of dimethyl carbonate was 2.669 ton / hr, and excluding the ethylene glycol contained in the catalyst solution, the actual produced amount of ethylene glycol was 1.839 ton /

hr. The ethylene carbonate conversion was 99.99%, the dimethyl carbonate selectivity was not less than 99.99%, and the ethylene glycol selectivity was not less than 99.99%.

Prolonged continuous operation was carried out under these conditions. After 1000 hours, 2000 hours, 3000 hours, and 5000 hours, the actual produced amounts per hour were 2.669 ton, 2.669 ton, 2.669 ton, and 2.669 ton respectively for dimethyl carbonate, and 1.839 ton, 1.839 ton, 1.839 ton, and 1.839 ton respectively for ethylene glycol, the ethylene carbonate conversions were respectively 99.99%, 99.99%, 99.99%, and 99.99%, the dimethyl carbonate selectivities were respectively not less than 99.99%, not less than 99.99%, not less than 99.99%, and not less than 99.99%, and the ethylene glycol selectivities were respectively not less than 99.99%, not less than 99.99%, not less than 99.99%, and not less than 99.99%.

(2) Step (II) of continuously producing diphenyl carbonate

Reactive distillation was carried out under the following conditions using the same apparatus as in Example 1.

A starting material 1 containing phenol and dimethyl carbonate in a weight ratio of phenol / dimethyl carbonate = 1.1 was introduced continuously in a liquid form at a flow rate of 40 ton / hr from the upper inlet 11 of the first continuous multi-stage distillation column 101. On the other hand, a starting material 2 containing dimethyl carbonate and phenol in a weight ratio of dimethyl carbonate / phenol = 3.9 was introduced continuously in a gaseous form at a flow rate of 43 ton / hr from the lower inlet 12 of the first continuous multi-stage distillation column 101. The molar ratio for the starting materials introduced into

the first continuous multi-stage distillation column 101 was dimethyl carbonate / phenol = 1.87. The starting materials substantially did not contain halogens (outside the detection limit for the ion chromatography, i.e. not more than 1 ppb). Pb(OPh)_2 as a catalyst was introduced from the upper inlet 11 of the first continuous multi-stage distillation column 101 such that a concentration thereof in the reaction liquid would be approximately 250 ppm. Reactive distillation was carried out continuously under conditions of a temperature at the bottom of the first continuous multi-stage distillation column 101 of 235°C, and a pressure at the top of the column of 9×10^5 Pa and a reflux ratio of 0. A first column low boiling point reaction mixture containing methanol, dimethyl carbonate, phenol and so on was continuously withdrawn in a gaseous form from the top 13 of the first column, was passed through the heat exchanger 14, and was withdrawn at a flow rate of 43 ton / hr from the outlet 16. On the other hand, a first column high boiling point reaction mixture containing methyl phenyl carbonate, dimethyl carbonate, phenol, diphenyl carbonate, the catalyst and so on was continuously withdrawn in a liquid form from the bottom 17 of the first column.

A stable steady state was attained after 24 hours. The first column high boiling point reaction mixture was thus then fed continuously as is into the second continuous multi-stage distillation column 201 at a flow rate of 40 ton / hr from the starting material inlet 21 installed between the Mellapak and the sieve trays. The liquid fed into the second continuous multi-stage distillation column 201 contained 20.7 % by weight of methyl phenyl carbonate and 1.0 % by weight of diphenyl carbonate. Reactive distillation was carried out continuously in the

second continuous multi-stage distillation column 201 under conditions of a temperature at the bottom of the column of 205°C, a pressure at the top of the column of 2×10^4 Pa, and a reflux ratio of 0.5. It was possible to attain stable steady state operation after 24 hours. A second column low boiling point reaction mixture was continuously withdrawn from the top 23 of the second column, and a second column high boiling point reaction mixture containing 36.2 % by weight of methyl phenyl carbonate and 60.8 % by weight of diphenyl carbonate was continuously withdrawn from the bottom 27 of the second column. The second column low boiling point reaction mixture was continuously fed into the first continuous multi-stage distillation column 101 from the inlet 11. At this time, the amounts of fresh dimethyl carbonate and phenol newly fed into the first continuous multi-stage distillation column 101 were adjusted so as to maintain the above compositions and amounts of the starting material 1 and the starting material 2, taking into consideration the composition and amount of the second column low boiling point reaction mixture. It was found that the produced amount of diphenyl carbonate per hour was 4.03 ton. The selectivity for the diphenyl carbonate based on the phenol reacted was 97%.

Prolonged continuous operation was carried out under these conditions. The produced amounts of diphenyl carbonate per hour after 500 hours, 1000 hours, and 2000 hours were 4.03 ton, 4.03 ton, and 4.04 ton respectively, and the selectivities based on the reacted phenol were 97%, 97%, and 97% respectively, and hence the operation was very stable.

(3) Step (III) of obtaining high-purity diphenyl carbonate

This was carried out using the same process as in Example 1. The purity of diphenyl carbonate continuously withdrawn from the side cut outlet was not less than 99.999%, and the halogen content was not more than 1 ppb.

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Example 3:

(1) Step (I) of continuously producing dimethyl carbonate and ethylene glycol

A continuous multi-stage distillation column as shown in FIG. 1 having $L_0 = 3300$ cm, $D_0 = 300$ cm, $L_0 / D_0 = 11$, $n_0 = 60$, $D_0 / d_{01} = 7.5$, and $D_0 / d_{02} = 12$ was used. In this example, as the internals, sieve trays each having a cross-sectional area per hole in the sieve portion thereof of approximately 1.3 cm^2 and a number of holes of approximately 220 to 340 / m^2 were used.

15 3.773 Ton / hr of ethylene carbonate in a liquid form was continuously introduced into the distillation column from the inlet (3-a) provided at the 55th stage from the bottom. 3.736 Ton / hr of methanol in a gaseous form (containing 8.97 % by weight of dimethyl carbonate) and 8.641 ton / hr of methanol in a liquid form (containing 6.65 % by weight of dimethyl carbonate) were respectively continuously introduced into the distillation column from the inlets (3-b and 3-c) provided at the 31st stage from the bottom. The molar ratio of the starting materials introduced into the distillation column was methanol / ethylene carbonate = 8.73. The catalyst was made to be the same as in Example 1, and
20 was continuously fed into the distillation column. Reactive distillation was carried out continuously under conditions of a column bottom

temperature of 98 °C, a column top pressure of approximately 1.118×10^5 Pa, and a reflux ratio of 0.42.

It was possible to attain stable steady state operation after 24 hours. A low boiling point reaction mixture withdrawn from the top of the column in a gaseous form was cooled using a heat exchanger and thus turned into a liquid. The liquid low boiling point reaction mixture, which was continuously withdrawn from the distillation column at 12.32 ton / hr, contained 4.764 ton / hr of dimethyl carbonate, and 7.556 ton / hr of methanol. A liquid continuously withdrawn from the bottom of the column at 3.902 ton / hr contained 2.718 ton / hr of ethylene glycol, 1.17 ton / hr of methanol, and 4.6 kg / hr of unreacted ethylene carbonate. Excluding the dimethyl carbonate contained in the starting material, the actual produced amount of dimethyl carbonate was 3.854 ton / hr, and excluding the ethylene glycol contained in the catalyst solution, the actual produced amount of ethylene glycol was 2.655 ton / hr. The ethylene carbonate conversion was 99.88%, the dimethyl carbonate selectivity was not less than 99.99%, and the ethylene glycol selectivity was not less than 99.99%.

Prolonged continuous operation was carried out under these conditions. After 1000 hours, 2000 hours, 3000 hours, and 5000 hours, the actual produced amounts per hour were 3.854 ton, 3.854 ton, 3.854 ton, and 3.854 ton respectively for dimethyl carbonate, and 2.655 ton, 2.655 ton, 2.655 ton, and 2.655 ton respectively for ethylene glycol, the ethylene carbonate reaction ratios were respectively 99.99%, 99.99%, 99.99%, and 99.99%, the dimethyl carbonate selectivities were respectively not less than 99.99%, not less than 99.99%, not less than

99.99%, and not less than 99.99%, and the ethylene glycol selectivities were respectively not less than 99.99%, not less than 99.99%, not less than 99.99%, and not less than 99.99%.

(2) Step (II) of continuously producing diphenyl carbonate

5 Reactive distillation was carried out under the following conditions using the same apparatus as in Example 1 except that the cross-sectional area per hole of each of the sieve trays in the second continuous multi-stage distillation column 201 was made to be approximately 1.8 cm^2 .

10 A starting material 1 containing phenol and dimethyl carbonate in a weight ratio of phenol / dimethyl carbonate = 1.7 was introduced continuously in a liquid form at a flow rate of 86 ton / hr from the upper inlet 11 of the first continuous multi-stage distillation column 101. On the other hand, a starting material 2 containing dimethyl carbonate and
15 phenol in a weight ratio of dimethyl carbonate / phenol = 3.5 was introduced continuously in a gaseous form at a flow rate of 90 ton / hr from the lower inlet 12 of the first continuous multi-stage distillation column 101. The molar ratio for the starting materials introduced into the first continuous multi-stage distillation column 101 was dimethyl
20 carbonate / phenol = 1.44. The starting materials substantially did not contain halogens (outside the detection limit for the ion chromatography, i.e. not more than 1 ppb). $\text{Pb}(\text{OPh})_2$ as a catalyst was introduced from the upper inlet 11 of the first continuous multi-stage distillation column 101 such that a concentration thereof in the reaction liquid would be
25 approximately 150 ppm. Reactive distillation was carried out continuously under conditions of a temperature at the bottom of the first

continuous multi-stage distillation column 101 of 220°C, a pressure at the top of the column of 8×10^5 Pa and a reflux ratio of 0. A first column low boiling point reaction mixture containing methanol, dimethyl carbonate, phenol and so on was continuously withdrawn in a gaseous form from the top 13 of the first column, was passed through the heat exchanger 14, and was withdrawn at a flow rate of 82 ton / hr from the outlet 16. On the other hand, a first column high boiling point reaction mixture containing methyl phenyl carbonate, dimethyl carbonate, phenol, diphenyl carbonate, the catalyst and so on was continuously withdrawn in a liquid form from the bottom 17 of the first column.

A stable steady state was attained after 24 hours. The first column high boiling point reaction mixture was thus then fed continuously as is into the second continuous multi-stage distillation column 201 at a flow rate of 94 ton / hr from the starting material inlet 21 installed between the Mellapak and the sieve trays. The liquid fed into the second continuous multi-stage distillation column 201 contained 16.0 % by weight of methyl phenyl carbonate and 0.5 % by weight of diphenyl carbonate. Reactive distillation was carried out continuously in the second continuous multi-stage distillation column 201 under conditions of a temperature at the bottom of the column of 215°C, a pressure at the top of the column of 2.5×10^4 Pa, and a reflux ratio of 0.4. It was possible to attain stable steady state operation after 24 hours. A second column low boiling point reaction mixture was continuously withdrawn from the top 23 of the second column, and a second column high boiling point reaction mixture containing 35.5 % by weight of methyl phenyl carbonate and 59.5 % by weight of diphenyl carbonate was continuously

withdrawn from the bottom 27 of the second column. The second column low boiling point reaction mixture was continuously fed into the first continuous multi-stage distillation column 101 from the inlet 11. At this time, the amounts of fresh dimethyl carbonate and phenol newly fed into the first continuous multi-stage distillation column 101 were adjusted so as to maintain the above compositions and amounts of the starting material 1 and the starting material 2, taking into consideration the composition and amount of the second column low boiling point reaction mixture. It was found that the produced amount of diphenyl carbonate per hour was 7.28 ton. The selectivity for the diphenyl carbonate based on the phenol reacted was 98%.

Prolonged continuous operation was carried out under these conditions. The produced amounts of diphenyl carbonate per hour after 500 hours, 1000 hours, and 2000 hours were 7.28 ton, 7.29 ton, and 7.29 ton respectively, and the selectivities based on the reacted phenol were 98%, 98%, and 98% respectively, and hence the operation was very stable.

(3) Step (III) of obtaining high-purity diphenyl carbonate

This was carried out using the same process as in Example 1. The purity of diphenyl carbonate continuously withdrawn from the side cut outlet was not less than 99.999%, and the halogen content was not more than 1 ppb.

Industrial Applicability

It has been discovered that by implementing the process according to the present invention, a high-purity diaryl carbonate

required for producing a high-quality high-performance aromatic polycarbonate can be produced on an industrial scale of not less than 1 ton / hr from the cyclic carbonate and the aromatic monohydroxy compound. Moreover, it has been discovered that such a high-purity diaryl carbonate can be produced stably for a prolonged period of time of, for example, not less than 2000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours. The present invention thus provides a process that achieves excellent effects as an industrial process for the production of the high-purity diaryl carbonate.

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CLAIMS

We claim:

1. An industrial process for the production of a high-purity diaryl carbonate in which a high-purity diaryl carbonate is continuously produced from a cyclic carbonate and an aromatic monohydroxy compound, the process comprising the steps of:

(I) continuously producing a dialkyl carbonate and a diol through a reactive distillation system of continuously feeding the cyclic carbonate and an aliphatic monohydric alcohol into a continuous multi-stage distillation column T_0 in which a catalyst is present, carrying out reaction and distillation simultaneously in said column, continuously withdrawing a low boiling point reaction mixture containing a produced dialkyl carbonate from an upper portion of the column in a gaseous form, and continuously withdrawing a high boiling point reaction mixture containing the diol from a lower portion of the column in a liquid form;

(II) continuously producing a diaryl carbonate by taking said dialkyl carbonate and an aromatic monohydroxy compound as a starting material, continuously feeding the starting material into a first continuous multi-stage distillation column in which a catalyst is present, carrying out reaction and distillation simultaneously in said first column, continuously withdrawing a first column low boiling point reaction mixture containing a produced alcohol from an upper portion of said first column in a gaseous form, continuously withdrawing a first column high boiling point reaction mixture containing a produced alkyl aryl carbonate from a lower portion of said first column in a liquid form, continuously feeding said first column high boiling point reaction mixture into a second continuous

multi-stage distillation column in which a catalyst is present, carrying out reaction and distillation simultaneously in said second column, continuously withdrawing a second column low boiling point reaction mixture containing a produced dialkyl carbonate from an upper portion of said second column in a gaseous form, continuously withdrawing a second column high boiling point reaction mixture containing a produced diaryl carbonate from a lower portion of said second column in a liquid form, and continuously feeding the second column low boiling point reaction mixture containing the dialkyl carbonate into the first continuous multi-stage distillation column; and

(III) purifying said diaryl carbonate so as to obtain the high-purity diaryl carbonate;

wherein:

(a) said continuous multi-stage distillation column T_0 comprises a structure having a cylindrical trunk portion having a length L_0 (cm) and an inside diameter D_0 (cm) and having an internal having number of stages n_0 thereinside, and comprises a gas outlet having an inside diameter d_{01} (cm) at a top of the column or in the upper portion of the column near to the top, a liquid outlet having an inside diameter d_{02} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one first inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one second inlet provided in the middle portion and/or the lower portion of the column above the liquid outlet, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} respectively satisfy the following formulae (1) to (6),

$$2100 \leq L_0 \leq 8000 \quad (1),$$

$$180 \leq D_0 \leq 2000 \quad (2),$$

$$4 \leq L_0 / D_0 \leq 40 \quad (3),$$

$$10 \leq n_0 \leq 120 \quad (4),$$

$$3 \leq D_0 / d_{01} \leq 20 \quad (5), \text{ and}$$

$$5 \leq D_0 / d_{02} \leq 30 \quad (6);$$

(b) said first continuous multi-stage distillation column comprises a structure having a cylindrical trunk portion having a length L_1 (cm) and an inside diameter D_1 (cm), and having an internal with number of stages n_1 thereinside, and comprises a gas outlet having an inside diameter d_{11} (cm) at a top of the column or in the upper portion of the column near to the top, a liquid outlet having an inside diameter d_{12} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one third inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one fourth inlet provided in the middle portion and/or the lower portion of the column above the liquid outlet, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} respectively satisfy the following formulae (7) to (12),

$$1500 \leq L_1 \leq 8000 \quad (7),$$

$$100 \leq D_1 \leq 2000 \quad (8),$$

$$20 \leq L_1 / D_1 \leq 40 \quad (9),$$

$$20 \leq n_1 \leq 120 \quad (10),$$

$$5 \leq D_1 / d_{11} \leq 30 \quad (11), \text{ and}$$

$$3 \leq D_1 / d_{12} \leq 20 \quad (12);$$

(c) said second continuous multi-stage distillation column comprises a structure having a cylindrical trunk portion having a length L_2 (cm) and an inside diameter D_2 (cm), and having an internal with

number of stages n_2 thereinside, and comprises a gas outlet having an inside diameter d_{21} (cm) at a top of the column or in the upper portion of the column near to the top, a liquid outlet having an inside diameter d_{22} (cm) at a bottom of the column or in the lower portion of the column near to the bottom, at least one fifth inlet provided in the upper portion and/or a middle portion of the column below the gas outlet, and at least one sixth inlet provided in the central portion and/or the lower portion of the column above the liquid outlet, wherein L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} respectively satisfy the following formulae (13) to (18),

- 10 $1500 \leq L_2 \leq 8000$ (13),
 $100 \leq D_2 \leq 2000$ (14),
 $2 \leq L_2 / D_2 \leq 40$ (15),
 $10 \leq n_2 \leq 80$ (16),
 $2 \leq D_2 / d_{21} \leq 15$ (17), and
 15 $5 \leq D_2 / d_{22} \leq 30$ (18).

2. The process according to Claim 1, wherein not less than 1 ton / hr of the high-purity diaryl carbonate is produced.

- 20 3. The process according to Claim 1 or 2, wherein said d_{01} and said d_{02} for said continuous multi-stage distillation column T_0 used in the step (I) satisfy the formula (19)

$$1 \leq d_{01} / d_{02} \leq 5 \quad (19).$$

- 25 4. The process according to any one of Claims 1 to 3, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} for said continuous multi-stage

distillation column T_0 satisfy respectively $2300 \leq L_0 \leq 6000$, $200 \leq D_0 \leq 1000$, $5 \leq L_0 / D_0 \leq 30$, $30 \leq n_0 \leq 100$, $4 \leq D_0 / d_{01} \leq 15$, and $7 \leq D_0 / d_{02} \leq 25$.

5 5. The process according to any one of Claims 1 to 4, wherein L_0 , D_0 , L_0 / D_0 , n_0 , D_0 / d_{01} , and D_0 / d_{02} for said continuous multi-stage distillation column T_0 satisfy respectively $2500 \leq L_0 \leq 5000$, $210 \leq D_0 \leq 800$, $7 \leq L_0 / D_0 \leq 20$, $40 \leq n_0 \leq 90$, $5 \leq D_0 / d_{01} \leq 13$, and $9 \leq D_0 / d_{02} \leq 20$.

10

6. The process according to any one of Claims 1 to 5, wherein said continuous multi-stage distillation column T_0 is a distillation column having a tray and/or a packing as said internal.

15 7. The process according to Claim 6, wherein said continuous multi-stage distillation column T_0 is a plate type distillation column having the tray as said internal.

8. The process according to Claim 6 or 7, wherein said tray in
20 said continuous multi-stage distillation column T_0 is a sieve tray having a sieve portion and a downcomer portion.

9. The process according to Claim 8, wherein said sieve tray
in said continuous multi-stage distillation column T_0 has 100 to 1000
25 holes / m^2 in said sieve portion thereof.

10. The process according to Claim 8 or 9, wherein a cross-sectional area per hole of said sieve tray in said continuous multi-stage distillation column T_0 is in a range of from 0.5 to 5 cm².

5 11. The process according to any one of Claims 8 to 10, wherein an aperture ratio of said sieve tray in said continuous multi-stage distillation column T_0 is in a range of from 1.5 to 15%.

12. The process according to any one of Claims 1 to 11,
10 wherein said d_{11} and said d_{12} for said first continuous multi-stage distillation column used in the step (II) satisfy the following formula (20), and said d_{21} and said d_{22} for said second continuous multi-stage distillation column used in the step (II) satisfy the following formula (21);

$$1 \leq d_{12} / d_{11} \leq 5 \quad (20) \text{ and}$$

15 $1 \leq d_{21} / d_{22} \leq 6 \quad (21).$

13. The process according to any one of Claims 1 to 12, wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for said first continuous multi-stage distillation column used in the step (II) satisfy respectively
20 $2000 \leq L_1 \leq 6000$, $150 \leq D_1 \leq 1000$, $3 \leq L_1 / D_1 \leq 30$, $30 \leq n_1 \leq 100$, $8 \leq D_1 / d_{11} \leq 25$, and $5 \leq D_1 / d_{12} \leq 18$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for said second continuous multi-stage distillation column used in the step (II) satisfy respectively $2000 \leq L_2 \leq 6000$, $150 \leq D_2 \leq 1000$, $3 \leq L_2 / D_2 \leq 30$, $15 \leq n_2 \leq 60$, $2.5 \leq D_2 / d_{21} \leq 12$, and $7 \leq D_2 / d_{22} \leq 25$.

25

14. The process according to any one of Claims 1 to 13,

wherein L_1 , D_1 , L_1 / D_1 , n_1 , D_1 / d_{11} , and D_1 / d_{12} for said first continuous multi-stage distillation column satisfy respectively $2500 \leq L_1 \leq 5000$, $200 \leq D_1 \leq 800$, $5 \leq L_1 / D_1 \leq 15$, $40 \leq n_1 \leq 90$, $10 \leq D_1 / d_{11} \leq 25$, and $7 \leq D_1 / d_{12} \leq 15$, and L_2 , D_2 , L_2 / D_2 , n_2 , D_2 / d_{21} , and D_2 / d_{22} for said second continuous multi-stage distillation column satisfy respectively $2500 \leq L_2 \leq 5000$, $200 \leq D_2 \leq 800$, $5 \leq L_2 / D_2 \leq 15$, $20 \leq n_2 \leq 50$, $3 \leq D_2 / d_{21} \leq 10$, and $9 \leq D_2 / d_{22} \leq 20$.

15. The process according to any one of Claims 1 to 14,
wherein each of said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is a distillation column having a tray and/or a packing as said internal.

16. The process according to Claim 15, wherein said first continuous multi-stage distillation column is a plate type distillation column having the tray as said internal, and said second continuous multi-stage distillation column is a distillation column having both the packing and the trays as said internal.

17. The process according to Claim 15 or 16, wherein each of said trays in said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is a sieve tray having a sieve portion and a downcomer portion.

18. The process according to Claim 17, wherein each of said sieve trays in said first continuous multi-stage distillation column and

said second continuous multi-stage distillation column has 100 to 1000 holes / m² in said sieve portion.

19. The process according to Claim 17 or 18, wherein a
5 cross-sectional area per hole of each of said sieve trays in said first continuous multi-stage distillation column and said second continuous multi-stage distillation column is in a range of from 0.5 to 5 cm².

20. The process according to Claim 15 or 16, wherein said
10 second continuous multi-stage distillation column is a distillation column having, as said internal, the packing in the upper portion of the column, and the tray in the lower portion of the column.

21. The process according to any one of Claims 15 to 20,
15 wherein said packing of said internal in said second continuous multi-stage distillation column is one or a plurality of sets of structured packings.

22. The process according to Claim 21, wherein said structured
20 packings in said second continuous multi-stage distillation column are of at least one type selected from the group consisting of Mellapak, Gempak, Techno-pack, Flexipac, a Sulzer packing, a Goodroll packing, and Glitschgrid.

23. The process according to any of Claims 1 to 22, wherein
25 the diaryl carbonate purification step (III) is carried out by distillation.

24. A high-purity diaryl carbonate produced in an amount of not less than 1 ton / hr, which is produced by the process according to any one of Claims 1 to 23.

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25. The high-purity diaryl carbonate according to Claim 24, which has a halogen content of not more than 0.1 ppm.

26. The high-purity diaryl carbonate according to Claim 25,

10 which has a halogen content of not more than 1 ppb.