

PROCESS FOR INDUSTRIALLY PRODUCING DIALKYL CARBONATE AND DIOL

Technical Field

The present invention relates to an industrial process for the production of a dialkyl carbonate and a diol with a high yield. More particularly, the present invention relates to a process for industrially producing large amounts of the dialkyl carbonate and the diol with the high yield stably for a prolonged period of time by a reactive distillation system of taking a cyclic carbonate and an aliphatic monohydric alcohol as starting materials, continuously feeding the starting materials into a continuous multi-stage distillation column in which a catalyst is present, and carrying out reaction and distillation simultaneously in the column.

Background Art

Several processes for the production of a dialkyl carbonate and a diol from a reaction between a cyclic carbonate and an aliphatic monohydric alcohol have been proposed, but most of these proposals relate to a catalyst. As reaction systems, four systems have been proposed hitherto. 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.

A first system is a completely batch reaction system in which ethylene carbonate, methanol and a catalyst are put into an autoclave, which is a batch reaction vessel, and reaction is carried out by holding

for a predetermined reaction time under an applied temperature at a reaction temperature above the boiling point of methanol (see, for example, Patent Document 1: U.S. Patent No. 3642858, Patent Document 2: Japanese Patent Application Laid-Open No. S54-48715 (corresponding to U.S. Patent No. 4181676), Patent Document 5: Japanese Patent Application Laid-Open No. S54-63023, Patent Document 6: Japanese Patent Application Laid-Open No. S54-148726).

A second system is a system that uses an apparatus in which a distillation column is provided on top of a reaction vessel; ethylene carbonate, methanol and a catalyst are put into the reaction vessel, and reaction is made to proceed by heating to a predetermined temperature. With this system, to make up for methanol distilled off through azeotropy with the produced dimethyl carbonate, methanol is added to the reaction vessel continuously or in batches, but in any event with this system the reaction proceeds only in the reaction vessel, which is batch type, in which the catalyst, ethylene carbonate and methanol are present. The reaction is thus batch type, the reaction being carried out using a large excess of methanol under reflux taking a long time in a range of from 3 to over 20 hours (see, for example, Patent Document 3: Japanese Patent Application Laid-Open No. S51-122025 (corresponding to U.S. Patent No. 4062884), Patent Document 4: Japanese Patent Application Laid-Open No. S54-48716 (corresponding to U.S. Patent No. 4307032), Patent Document 11: U.S. Patent No. 3803201).

A third system is a continuous reaction system in which a mixed solution of ethylene carbonate and methanol is continuously fed into a tubular reactor maintained at a predetermined reaction temperature, and

a reaction mixture containing unreacted ethylene carbonate and methanol and dimethyl carbonate and ethylene glycol which are produced is continuously withdrawn in a liquid form from an outlet on the other side. Either of two processes is used depending on the form of the catalyst used. That is, there are a process in which a homogeneous catalyst is used, and is passed through the tubular reactor together with the mixed solution of ethylene carbonate and methanol, and then after the reaction the catalyst is separated out from the reaction mixture (see, for example, Patent Document 7: Japanese Patent Application Laid-Open No. 63-41432 (corresponding to U.S. Patent No. 4661609), Patent Document 10: U.S. Patent No. 4734518), and a process in which a heterogeneous catalyst fixed inside the tubular reactor is used (see, for example, Patent Document 8: Japanese Patent Application Laid-Open No. S63-238043, Patent Document 9: Japanese Patent Application Laid-Open No. S64-31737 (corresponding to U.S. Patent No. 4691041)). The reaction producing dimethyl carbonate and ethylene glycol through reaction between ethylene carbonate and methanol is an equilibrium reaction, and hence with this continuous flow reaction system using a tubular reactor, it is impossible to make the ethylene carbonate conversion higher than the equilibrium conversion determined by the composition ratio put in and the reaction temperature. For example, according to Example 1 in Patent Document 7 (Japanese Patent Application Laid-Open No. S63-41432 (corresponding to U.S. Patent No. 4661609)), for a flow reaction at 130°C using a starting material put in with a molar ratio of methanol / ethylene carbonate = 4 / 1, the ethylene carbonate conversion is 25%. This means that a large amount of unreacted ethylene carbonate and

methanol remaining in the reaction mixture must be separated out, recovered, and recirculated back into the reactor, and in actual fact, with the process of Patent Document 9 (Japanese Patent Application Laid-Open No. S64-31737 (corresponding to U.S. Patent No. 4691041)), much equipment is used for such separation, purification, recovery, and recirculation.

A fourth system is a reactive distillation system first disclosed by the present inventors (see, for example, Patent Document 12: Japanese Patent Application Laid-Open No. H4-198141, Patent Document 13: Japanese Patent Application Laid-Open No. H4-230243, Patent Document 14: Japanese Patent Application Laid-Open No. H9-176061, Patent Document 15: Japanese Patent Application Laid-Open No. H9-183744, Patent Document 16: Japanese Patent Application Laid-Open No. H9-194435, Patent Document 17: International Publication No. WO97/23445 (corresponding to European Patent No. 0889025, U.S. Patent No. 5847189), Patent Document 18: International Publication No. WO99/64382 (corresponding to European Patent No. 1086940, U.S. Patent No. 6346638), Patent Document 19: International Publication No. WO00/51954 (corresponding to European Patent No. 1174406, U.S. Patent No. 6479689), Patent Document 20: Japanese Patent Application Laid-Open No. 2002-308804, Patent Document 21: Japanese Patent Application Laid-Open No. 2004-131394), that is a continuous production process in which ethylene carbonate and methanol are each continuously fed into a multi-stage distillation column, and reaction is carried out in the presence of a catalyst in a plurality of stages in the distillation column, while dimethyl carbonate and ethylene glycol which are

produced are separated off. Patent applications in which such a reactive distillation system is used have subsequently been filed by other companies (see, for example, Patent Document 22: Japanese Patent Application Laid-Open No. H5-213830 (corresponding to European Patent No. 0530615, U.S. Patent No. 5231212), Patent Document 23: Japanese Patent Application Laid-Open No. H6-9507 (corresponding to European Patent No. 0569812, U.S. Patent No. 5359118), Patent Document 24: Japanese Patent Application Laid-Open No. 2003-119168 (corresponding to International Publication No. WO03/006418), Patent Document 25: Japanese Patent Application Laid-Open No. 2003-300936, Patent Document 26: Japanese Patent Application Laid-Open No. 2003-342209).

In this way, the processes proposed hitherto for producing the dialkyl carbonates and the diols from the cyclic carbonate and the aliphatic monohydric alcohol are the four systems:

- (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.

However, there have been problems with these 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 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 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 a dialkyl carbonate continuously in a large amount (e.g. not less than 2 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).

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 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 1.

Table 1

TABLE 1

PATENT DOCUMENT	H : cm	D : cm	NO. STAGES : n	P : kg/hr	T : hr.
12	100	2	30	0.106	400
15	160	5	40	0.427	NOTE 5
16	160	5	40	0.473	NOTE 5
18	200	4	PACKING COLUMN (Dixon)	0.932	NOTE 5
19	NOTE 1	5	60	0.275	NOTE 5
20	NOTE 1	5	60	0.258	NOTE 5
21	NOTE 1	5	60	0.258	NOTE 5
22	250	3	PACKING COLUMN (Raschig)	0.392	NOTE 5
23	NOTE 2	NOTE 2	NOTE 2	0.532	NOTE 5
24	NOTE 3	NOTE 3	42	NOTE 4	NOTE 5
25	NOTE 3	NOTE 3	30	3750	NOTE 5
26	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.

Note that Patent Document 25 (Japanese Patent Application Laid-Open No. 2003-300936) (paragraph 0060) describes 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 reaction between ethylene carbonate and methanol. It should be noted 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 of 30,000 tons or more, and hence this implies

that at the time of the filing of the patent application for Patent Document 25 (Japanese Patent Application Laid-Open No. 2003-300936) (April 9, 2002), operation of the world's first large scale commercial plant using this process had been carried out. However, even at the time of filing the present invention application, there is not the fact at all. Moreover, in the example of Patent Document 25 (Japanese Patent Application Laid-Open No. 2003-300936), exactly the same value as the theoretically calculated value is stated for the produced amount of dimethyl carbonate, 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 25 (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 carbonates and the diols using the reactive distillation method stably for a prolonged period of time while maintaining high yields and high selectivities for the dialkyl carbonates

and the diols, 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 the 200 to 400 hours in Patent Document 12 (Japanese Patent Application Laid-Open No. H4-198141) and Patent Document 13 (Japanese Patent Application Laid-Open No. H4-230243).

Disclosure of Invention

Problems to be Solved by the Invention

It is an object of the present invention to provide, for the case of producing dialkyl carbonates and diols industrially in large amounts (e.g. not less than 2 ton / hr for the dialkyl carbonates, and not less than 1.3 ton / hr for the diols) through a reactive distillation system of taking a cyclic carbonate and an aliphatic monohydric alcohol as starting materials, continuously feeding the starting materials into a continuous multi-stage distillation column in which a homogeneous catalyst is present, and carrying out reaction and distillation simultaneously in the column, a specific process in which the dialkyl carbonates and the diols can be produced with high selectivity and high productivity 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) with a high yield.

Means for Solving the Problems

Since the present inventors first disclosed a process for continuously producing the dialkyl carbonates and the diols using the continuous multi-stage distillation column, there have been many

proposals regarding improving this process. However, these have been on a small scale and short operating time laboratory level, and there have been no disclosures whatsoever on a specific process or apparatus enabling mass production on an industrial scale stably for a prolonged period of time based on findings obtained through actual implementation. The present inventors have thus carried out studies aimed at discovering a specific process enabling the dialkyl carbonates and the diols to be produced on an industrial scale of, for example, not less than 2 ton / hr for the dialkyl carbonates and not less than 1.3 ton / hr for the diols stably for a prolonged period of time with a high yield, high selectivity and high productivity. As a result, the present inventors have reached to the present invention.

That is, in the first aspect of the present invention, there are provides:

1. a process for industrially producing a dialkyl carbonate and a diol in which the dialkyl carbonate and the diol are continuously produced through a reactive distillation system of taking a cyclic carbonate and an aliphatic monohydric alcohol as starting materials, comprising the steps of:

- continuously feeding the starting materials into a continuous multi-stage distillation column in which a homogeneous catalyst is present;

- carrying out reaction and distillation simultaneously in said column;

- continuously withdrawing a low boiling point reaction mixture containing the 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, wherein:

said continuous multi-stage distillation column comprises a cylindrical trunk portion having a length L (cm) and an inside diameter D (cm) and having thereinside a tray with number of stages n , and comprises a gas outlet having an inside diameter d_1 (cm) provided 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_2 (cm) provided 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 said gas outlet, and at least one second inlet provided in the middle portion and / or the lower portion of the column above said liquid outlet, wherein:

(1) the length L (cm) satisfies the formula (1);

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

(2) the inside diameter D (cm) of the column satisfies the formula (2);

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

(3) a ratio of the length L (cm) to the inside diameter D (cm) of the column satisfies the formula (3);

$$4 \leq L / D \leq 40 \quad (3),$$

(4) the number of stages n satisfies the formula (4);

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

(5) a ratio of the inside diameter D (cm) of the column to the

inside diameter d_1 (cm) of the gas outlet satisfies the formula (5);

$$3 \leq D / d_1 \leq 20 \quad (5),$$

(6) a ratio of the inside diameter D (cm) of the column to the inside diameter d_2 (cm) of the liquid outlet satisfies the formula (6);

$$5 \leq D / d_2 \leq 30 \quad (6), \text{ and}$$

(7) an aperture ratio of each tray is in a range of from 1.5 to 10%,

2. the process according to item 1, wherein a produced amount of the dialkyl carbonate is not less than 2 ton / hr,

3. the process according to item 1 or 2, wherein a produced amount of the diol is not less than 1.3 ton / hr,

4. the process according to any one of items 1 to 3, wherein said d_1 and said d_2 satisfy the formula (7);

$$1 \leq d_1 / d_2 \leq 5 \quad (7),$$

5. the process according to any one of items 1 to 4, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2300 \leq L \leq 6000$, $200 \leq D \leq 1000$, $5 \leq L / D \leq 30$, $30 \leq n \leq 100$, $4 \leq D / d_1 \leq 15$, and $7 \leq D / d_2 \leq 25$, respectively,

6. the process according to any one of items 1 to 5, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2500 \leq L \leq 5000$, $210 \leq D \leq 800$, $7 \leq L / D \leq 20$, $40 \leq n \leq 90$, $5 \leq D / d_1 \leq 13$, and $9 \leq D / d_2 \leq 20$, respectively,

7. the process according to any one of items 1 to 6, wherein the aperture ratio of each tray is in a range of from 1.7 to 8.0%,

8. the process according to any one of items 1 to 7, wherein the aperture ratio of each tray is in a range of from 1.9 to 6.0%,

9. the process according to any one of items 1 to 8, wherein said tray is

a sieve tray having a sieve portion and a downcomer portion,

10. the process according to item 9, wherein said sieve tray has 100 to 1000 holes / m² in said sieve portion thereof,

11. the process according to item 9 or 10, wherein a cross-sectional area per hole of said sieve tray is in a range of from 0.5 to 5 cm²,

12. the process according to item 10 or 11, wherein a aperture ratio (a ratio of a total cross-sectional area of the holes to a total area of the sieve plate containing the area of the hole portion) of said sieve tray is in a range of from 1.9 to 6.0%.

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

13. a continuous multi-stage distillation column for carrying out transesterification between a cyclic carbonate and an aliphatic monohydric alcohol and distillation, the continuous multi-stage distillation column comprising:

a cylindrical trunk portion having a length L (cm) and an inside diameter D (cm);

a tray having number of stages n provided inside said trunk portion;

a gas outlet having an inside diameter d₁ (cm) provided at a top of said column or in an upper portion of said column near to the top;

a liquid outlet having an inside diameter d₂ (cm) provided at a bottom of said column or in a lower portion of said column near to the bottom;

at least one first inlet provided in the upper portion and / or a middle portion of said column below said gas outlet; and

at least one second inlet provided in the middle portion and / or the lower portion of said column above said liquid outlet; wherein:

(1) the length L (cm) satisfies the formula (1);

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

(2) the inside diameter D (cm) of the column satisfies the formula (2);

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

(3) a ratio of the length L (cm) to the inside diameter D (cm) of the column satisfies the formula (3);

$$4 \leq L / D \leq 40 \quad (3),$$

(4) the number of stages n satisfies the formula (4);

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

(5) a ratio of the inside diameter D (cm) of the column to the inside diameter d_1 (cm) of the gas outlet satisfies the formula (5);

$$3 \leq D / d_1 \leq 20 \quad (5),$$

(6) a ratio of the inside diameter D (cm) of the column to the inside diameter d_2 (cm) of the liquid outlet satisfies the formula (6);

$$5 \leq D / d_2 \leq 30 \quad (6), \text{ and}$$

(7) an aperture ratio of each tray is in a range of from 1.5 to 10%,

14. the continuous multi-stage distillation column according to item 13, wherein said d_1 and said d_2 satisfy the formula (7);

$$1 \leq d_1 / d_2 \leq 5 \quad (7),$$

15. the continuous multi-stage distillation column according to item 13 or 14, wherein L, D, L / D, n, D / d_1 , and D / d_2 for said continuous

multi-stage distillation column satisfy the following formulae; $2300 \leq L \leq 6000$, $200 \leq D \leq 1000$, $5 \leq L / D \leq 30$, $30 \leq n \leq 100$, $4 \leq D / d_1 \leq 15$, and 7

$\leq D / d_2 \leq 25$, respectively,

16. the continuous multi-stage distillation column according to any one of items 13 to 15, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae;
 $2500 \leq L \leq 5000$, $210 \leq D \leq 800$, $7 \leq L / D \leq 20$, $40 \leq n \leq 90$, $5 \leq D / d_1 \leq 13$, and $9 \leq D / d_2 \leq 20$, respectively,

17. the continuous multi-stage distillation column according to any one of items 13 to 16, wherein the aperture ratio of each tray is in a range of from 1.7 to 8.0%,

18. the continuous multi-stage distillation column according to any of items 13 to 17, wherein the aperture ratio of each tray is in a range of from 1.9 to 6.0%,

19. the continuous multi-stage distillation column according to any one of items 13 to 18, wherein said tray is a sieve tray having a sieve portion and a downcomer portion,

20. the continuous multi-stage distillation column according to item 19, wherein said sieve tray has 100 to 1000 holes / m^2 in said sieve portion thereof,

21. the continuous multi-stage distillation column according to item 19 or 20, wherein a cross-sectional area per hole of said sieve tray is in a range of from 0.5 to 5 cm^2 ,

22. the continuous multi-stage distillation column according to item 20 or 21, wherein a aperture ratio (a ratio of a total cross-sectional area of the holes to a total area of the sieve plate containing the area of the hole portion) of said sieve tray is in a range of from 1.9 to 6.0%.

Advantageous Effects of the Invention

It has been discovered that by implementing the present invention, the dialkyl carbonates and the diols can be produced each with a high yield and a high selectivity of not less than 95%, preferably not less than 97%, more preferably not less than 99%, on an industrial scale of not less than 2 ton / hr, preferably not less than 3 ton / hr, more preferably not less than 4 ton / hr, for the dialkyl carbonates, and not less than 1.3 ton / hr, preferably not less than 1.95 ton / hr, more preferably not less than 2.6 ton / hr, for the diols, stably for a prolonged period of time of not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours, from the cyclic carbonates and the aliphatic monohydric alcohols.

Brief Description of the Drawing

FIG. 1 shows an example of schematic drawing of the continuous multi-stage distillation column for carrying out the present invention, the distillation column having a tray with n stages (shown schematically in FIG. 1) provided inside a trunk portion thereof.

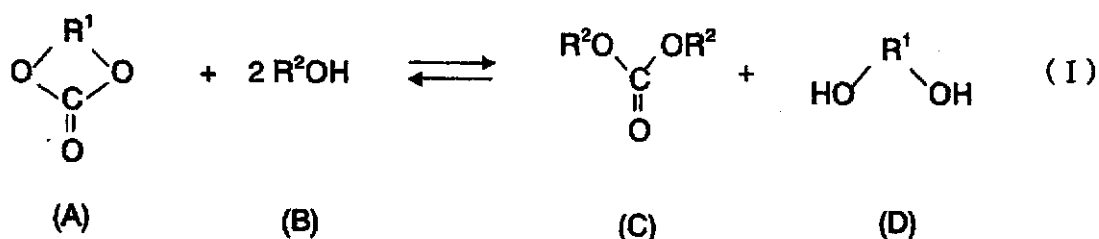
Description of Reference Numerals

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

Best Mode for Carrying Out the Invention

In the following, the present invention is described in detail.

The reaction of the present invention is a reversible equilibrium transesterification reaction represented by following formula (I) in which a dialkyl carbonate and a diol are produced from a cyclic carbonate and an aliphatic monohydric alcohol;



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

The cyclic carbonate used as a starting material in the present invention is a compound represented by (A) in the above formula (I). Preferable examples of the cyclic carbonate include alkylene carbonates such as ethylene carbonate or propylene carbonate; 1,3-dioxacyclohexa-2-one, 1,3-dioxacyclohepta-2-one, or the like, ethylene carbonate or propylene carbonate being more preferable due to ease of procurement and so on, and ethylene carbonate being particularly preferable.

Moreover, the aliphatic monohydric alcohol used as the other

starting material is a compound represented by (B) in the above formula (I). The aliphatic monohydric alcohol having a lower boiling point than that of 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), or 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.

In the process of the present invention, a homogeneous catalyst is made to be present in the reactive distillation column. The method of

making the homogeneous catalyst be present may be any method, but it is preferable to feed the catalyst into the reactive distillation column continuously so as to make the catalyst be present in a liquid phase in the reactive distillation column.

In the case that the 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 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. The catalyst must be present in at least 5 stages, preferably at least 7 stages, more preferably at least 10 stages.

As the catalyst used in the present invention, any of various catalysts known from hitherto can be used. Examples of the catalyst 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, tetraisopropoxytitanium, titanium acetate, and titanium acetylacetonate;

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

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 ;

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 ;

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, 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

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 is as the homogeneous catalyst.

Alternatively, it is also preferable to use, as the homogeneous catalyst, a

mixture obtained by dissolving a compound as above in the starting material of the reaction, the reaction mixture, the reaction by-product or the like, or by reacting to bring about dissolution.

The amount of the catalyst used in the present invention varies depending on the type of the catalyst used, but 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.

There are no particular limitations on the method of continuously feeding the cyclic carbonate and the aliphatic monohydric alcohol into a continuous multi-stage distillation column constituting the reactive distillation column in the present invention; 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 above. 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 materials are fed continuously into the distillation column in a liquid form, in a gaseous form, or as a mixture of a liquid and a 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 the 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 the present invention, the starting materials fed in may contain dialkyl carbonate and / or diol being the products. 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 present reaction 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 a process in which the diaryl carbonate is produced from the dialkyl carbonate and the aromatic monohydroxy compound, the aliphatic monohydric alcohol being by-produced in this process 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 diaryl carbonate and so on. The by-produced aliphatic monohydric alcohol may be used as is as the starting material in the present invention, or may be used as the starting material after amount of contained material having a higher boiling point than that of the aliphatic monohydric alcohol has been reduced through distillation or the like.

The cyclic carbonate preferably used in the present invention is one produced through reaction between, for example, an alkylene oxide such as ethylene oxide, propylene oxide or styrene oxide and carbon dioxide; a cyclic carbonate containing small amounts of these raw material compounds or the like may be used as the starting material in the present invention.

In the present invention, a ratio between the 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 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 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 3 from to 15, yet more preferably from 5 to 12. Furthermore, if much unreacted cyclic 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 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 the present invention, preferably not less than 2 ton / hr of the dialkyl carbonate is continuously produced; the minimum amount of the cyclic carbonate continuously fed in to achieve this is generally 2.2 P ton / hr, preferably 2.1 P ton / hr, more preferably 2.0 P ton/hr, relative to 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 1.9 P ton / hr.

FIG. 1 shows an example of schematic drawing of the continuous multi-stage distillation column for carrying out the production process according to the present invention. Here, the continuous multi-stage distillation column 10 used in the present invention comprises a structure having end plates 5 respectively above and below a cylindrical trunk

portion 7 having a length L (cm) and an inside diameter D (cm), and having thereinside a tray 6 with number of stages n, and has a gas outlet 1 having an inside diameter d_1 (cm) at a top of the column or in an upper portion of the column near to the top, a liquid outlet 2 having an inside diameter d_2 (cm) at a bottom of the column or in a lower portion of the column near to the bottom, at least one first inlet 3 (a, e) provided in the upper portion and / or a middle portion of the column below the gas outlet 1, and at least one second inlet 3 (b, c) and 4 (a, b) provided in the middle portion and / or the lower portion of the column above the liquid outlet 2, and moreover must be made to satisfy various conditions so as to be able to carry out not only distillation but also reaction at the same time so as to be able to produce preferably not less than 2 ton / hr of the dialkyl carbonate and / or preferably not less than 1.3 ton / hr of the diol stably for a prolonged period of time. Note that FIG. 1 merely shows one embodiment of the continuous multi-stage distillation column according to the present invention, and hence the arrangement of the tray 6 is not limited to that shown in FIG. 1.

The continuous multi-stage distillation column according to the present invention satisfies not only conditions from the perspective of the distillation function, but rather these conditions are combined with conditions required so as make the reaction proceed stably with a high conversion and high selectivity, specifically:

(1) the length L (cm) must satisfy the following formula (1);

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

(2) the inside diameter D (cm) of the column must satisfy the following formula (2);

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

(3) a ratio of the length L (cm) to the inside diameter D (cm) of the column must satisfy the following formula (3);

$$4 \leq L / D \leq 40 \quad (3),$$

(4) the number of stages n must satisfy the following formula (4);

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

(5) a ratio of the inside diameter D (cm) of the column to the inside diameter d_1 (cm) of the gas outlet must satisfy the following formula (5);

$$3 \leq D / d_1 \leq 20 \quad (5),$$

(6) a ratio of the inside diameter D (cm) of the column to the inside diameter d_2 (cm) of the liquid outlet must satisfy the following formula (6);

$$5 \leq D / d_2 \leq 30 \quad (6), \text{ and}$$

(7) an aperture ratio of each tray must be in a range of from 1.5 to 10%.

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 0.25L, 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. Here, "L" is defined as above.

It has been discovered that by using the continuous multi-stage distillation column that simultaneously satisfies the above formulae (1) to (6), and for which the aperture ratio of each tray is in a range of from 1.5

to 10% as in the above item (7), the dialkyl carbonate and the diol can be produced on an industrial scale of preferably not less than 2 ton / hr for the dialkyl carbonate and/or preferably not less than 1.3 ton / hr for 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 the process of the present invention is not clear, but this is supposed to be due to a composite effect brought about when the conditions of the above formulae (1) to (6) and the condition on the aperture ratio of each tray are combined.

The term "the aperture ratio of each tray" used in the present invention means, for each tray in the multi-stage distillation column, the ratio between the total area of openings in the tray through which gas and liquid can pass and the area of the tray having these openings therein. Note that for the tray having a downcomer portion, the area of the portion in which bubbling substantially occurs, i.e. excluding the downcomer portion, is taken as the area of the tray.

The present invention relates to the reactive distillation method in which not only distillation is carried out but rather reaction is carried out at the same time, and a high conversion and high selectivity (high yield) are attained; for achieve this, it has been discovered that in addition to the above formulae (1) to (6) being satisfied, it is important to make the aperture ratio be in a specified range as above. It should be noted that

preferable ranges for the respective factors are described below.

If L (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 the conversion enabling the desired production amount to be attained, L must be made to be not more than 8000. A more preferable range for L (cm) is $2300 \leq L \leq 6000$, with $2500 \leq L \leq 5000$ being yet more preferable.

If D (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 must be made to be not more than 2000. A more preferable range for D (cm) is $200 \leq D \leq 1000$, with $210 \leq D \leq 800$ being yet more preferable.

If L / D is less than 4 or greater than 40, then stable operation becomes difficult. In particular, if L / D 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 become liable to occur, bringing about a decrease in the selectivity. A more preferable range for L / D is $5 \leq L / D \leq 30$, with $7 \leq L / D \leq 20$ being yet more preferable.

If n is less than 10, 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, n must be made to be not more than 120. Furthermore, if n 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 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 is $30 \leq n \leq 100$, with $40 \leq n \leq 90$ being yet more preferable.

If D / d_1 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 / d_1 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 / d_1 is $4 \leq D / d_1 \leq 15$, with $5 \leq D / d_1 \leq 13$ being yet more preferable.

If D / d_2 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 / d_2 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 / d_2 is $7 \leq D / d_2 \leq 25$, with $9 \leq D / d_2 \leq 20$ being yet more preferable.

Furthermore, it has been found in the present invention that it is still more preferable if said d_1 and said d_2 satisfy the following formula (7);

$$1 \leq d_1 / d_2 \leq 5 \quad (7).$$

The aperture ratio of each tray must be in a range of from 1.5 to 10%. If the aperture ratio is less than 1.5%, then the apparatus

becomes large relative to the required production amount, and hence the equipment cost becomes high. Moreover, the residence time increases, and hence side reactions (e.g. a reaction between the reaction product diol and unreacted cyclic carbonate) become liable to occur. Moreover, if the aperture ratio is greater than 10%, then the residence time for each tray decreases, and hence the number of stages required for attaining a high conversion increases, and hence the problems described above for when n is large arise. For such reasons, a preferable range for the aperture ratio is 1.7 to 8.0%, with 1.9 to 6.0% being more preferable.

Furthermore, in the present invention, the aperture ratio may be the same for all of the trays, or may differ. In the present invention, it is generally preferable to use the multi-stage distillation column in which the aperture ratio of the tray in the upper portion thereof is greater than the aperture ratio of the tray in the lower portion thereof.

The term "prolonged stable operation" used in the present invention means that the continuous multi-stage distillation column can be operated 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 the high conversion, high selectivity, and high productivity.

A characteristic feature of the present invention is that the dialkyl carbonate and the diol can be produced stably for a prolonged period of time each with high selectivity and preferably with high productivity for the dialkyl carbonate of not less than 2 ton / hr and high productivity for

the diol of not less than 1.3 ton / hr. The dialkyl carbonate and the diol are more preferably produced in an amount of not less than 3 ton / hr and not less than 1.95 ton / hr respectively, yet more preferably not less than 4 ton / hr and not less than 2.6 ton / hr respectively. Moreover, another characteristic feature of the present invention is that in the case that L , D , L / D , n , D / d_1 , and D / d_2 for the continuous multi-stage distillation column satisfy respectively $2300 \leq L \leq 6000$, $200 \leq D \leq 1000$, $5 \leq L / D \leq 30$, $30 \leq n \leq 100$, $4 \leq D / d_1 \leq 15$, and $7 \leq D / d_2 \leq 25$, and the aperture ratio of each tray is in a range of from 1.7 to 8.0%, not less than 2.5 ton / hr, preferably not less than 3 ton / hr, more preferably not less than 3.5 ton / hr of the dialkyl carbonate, and not less than 1.6 ton / hr, preferably not less than 1.95 ton / hr, more preferably not less than 2.2 ton / hr of the diol can be produced. Furthermore, another characteristic feature of the present invention is that in the case that L , D , L / D , n , D / d_1 , and D / d_2 for the continuous multi-stage distillation column satisfy respectively $2500 \leq L \leq 5000$, $210 \leq D \leq 800$, $7 \leq L / D \leq 20$, $40 \leq n \leq 90$, $5 \leq D / d_1 \leq 13$, and $9 \leq D / d_2 \leq 20$, and the aperture ratio of each tray is in a range of from 1.9 to 6.0%, not less than 3 ton / hr, preferably not less than 3.5 ton / hr, more preferably not less than 4 ton / hr of the dialkyl carbonate, and not less than 1.95 ton / hr, preferably not less than 2.2 ton / hr, more preferably not less than 2.6 ton / hr of the diol can be produced.

The term "selectivity" for each of the dialkyl carbonate and the diol in the present invention 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 term "conversion" in the present invention 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 the present invention that the high conversion can be maintained while maintaining high selectivity in this way.

The continuous multi-stage distillation column used in the present invention is a distillation column having n stages of the tray therein. Examples of the tray include a bubble-cap tray, a sieve tray, a ripple tray, a ballast tray, a valve tray, a counterflow tray, an Unifrac tray, a Superfrac tray, a Maxfrac tray, a dual flow tray, a grid plate tray, a turbogrid plate tray, a Kittel tray, or the like. In the case that there are stages in the continuous multi-stage distillation column in which the catalyst is not present and hence reaction substantially does not take place (e.g. stages above the stage at which the catalyst is introduced), a distillation column in which these stages are packed with the packings, i.e. a multi-stage distillation column having both a tray portion and a portion packed with the packings, is also preferable. Examples of the packings include irregular packings such as a Raschig ring, a Lessing ring, a Pall ring, a Berl saddle, a Intalox saddle, a Dixon packing, a McMahon packing or Heli-Pak, or regular packings such as Mellapak, Gempak, Techno-pack, Flexipac, a Sulzer packing, a Goodroll packing or Glitschgrid. Furthermore, the term "number of stages n " used in the present invention means the number of tray in the case of the trays, and

the theoretical number of stages in the case of the packing. The number of stages n in the case of the multi-stage distillation column having both the tray portion and the portion packed with packings is thus the sum of the number of trays and the theoretical number of stages.

For the process of the present invention, it has been discovered that the high conversion, high selectivity, and high productivity can be attained if n stages of any of the above trays are used, but 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 that each sieve tray preferably has 100 to 1000 holes / m^2 in the sieve portion. A more preferable number of holes is 120 to 900 holes / m^2 , yet more preferably 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 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 . The number of holes in the sieve portion may be the same for all of the sieve trays, or may differ.

It has also been discovered that, for such sieve trays, the aperture ratio of each tray is generally in a range of from 1.5 to 10%, preferably from 1.7 to 8.0%, it is particularly preferable if the aperture ratio of each tray in a range of from 1.9 to 6.0%.

It has been shown that by adding the above conditions to the

continuous multi-stage distillation column, the object of the present invention can be attained more easily.

When carrying out the present invention, the dialkyl carbonate and the diol are continuously produced by continuously feeding the cyclic carbonate and 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 the upper portion of the column in a gaseous form, and continuously withdrawing a high boiling point reaction mixture containing the diol from the lower portion of the column in a liquid form.

Moreover, in the present invention, as the continuous feeding of the starting material cyclic carbonate and aliphatic monohydric alcohol into the continuous multi-stage distillation column, 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 the present invention is considered to equate to the average residence time of the reaction liquid in the continuous multi-stage distillation column. 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 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 the present invention 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 material constituting the continuous multi-stage distillation column used in the present invention is generally a metallic material such as carbon steel or stainless steel. In terms of the quality of the dialkyl carbonate and diol to be produced, stainless steel is preferable.

EXAMPLES

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

Example 1:

<Continuous multi-stage distillation column>

A continuous multi-stage distillation column as shown in FIG. 1 having $L = 3300$ cm, $D = 300$ cm, $L / D = 11$, $n = 60$, $D / d_1 = 7.5$, and $D / d_2 = 12$ was used. The trays in the distillation column were sieve trays, each having the cross-sectional area per hole in the sieve portion thereof of approximately 1.3 cm^2 and a number of holes of approximately 180 to $32 / \text{m}^2$. The aperture ratio of each of the trays was in a range of from 2.1 to 4.2%.

<Reactive distillation>

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

continuously introduced into the distillation column from an inlet (3-a) 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 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 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 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 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 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 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 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 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%.

Example 2:

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

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%.

Example 3:

The continuous multi-stage distillation column as shown in FIG. 1 having $L = 3300$ cm, $D = 300$ cm, $L / D = 11$, $n = 60$, $D / d_1 = 7.5$, and $D / d_2 = 12$ was used. The trays in the distillation column were sieve trays, each having the 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 $340 / \text{m}^2$. The aperture ratio of each of the trays was in a range of from 2.5 to 4.5%.

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 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 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%.

Example 4:

The continuous multi-stage distillation column as shown in FIG. 1 having $L = 3300$ cm, $D = 300$ cm, $L / D = 11$, $n = 60$, $D / d_1 = 7.5$, and $D / d_2 = 12$ was used. The trays in the distillation column were sieve trays, each having the cross-sectional area per hole in the sieve portion thereof of approximately 1.3 cm^2 and a number of holes of approximately 240 to 360 / m^2 . The aperture ratio of each of the trays was in a range of from 3.0 to 5.0%.

7.546 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. 7.742 Ton / hr of methanol in a gaseous form (containing 8.95 % by weight of dimethyl carbonate)

and 17.282 ton/hr of methanol in a liquid form (containing 6.66 % 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 = 8.36. 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 top temperature of 65 °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 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 24.641 ton / hr, contained 9.527 ton / hr of dimethyl carbonate, and 15.114 ton / hr of methanol. A liquid continuously withdrawn from the bottom 2 of the column at 7.804 ton / hr contained 5.436 ton / hr of ethylene glycol, 2.34 ton / hr of methanol, and 23 kg / hr of unreacted ethylene carbonate. Excluding the dimethyl carbonate contained in the starting material, the actual produced amount of dimethyl carbonate was 7.708 ton / hr, and excluding the ethylene glycol contained in the catalyst solution, the actual produced amount of ethylene glycol was 5.31 ton / hr. The ethylene carbonate conversion was 99.7%, 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, the actual produced amount per hour was 7.708 ton for dimethyl carbonate, and 5.31 ton for ethylene glycol, the ethylene carbonate conversion was 99.8%, the dimethyl carbonate selectivity was not less than 99.99%, and the ethylene glycol selectivity was not less than 99.99%.

Industrial Applicability

According to the present invention, it has been discovered that the dialkyl carbonate and the diol can be produced each with a high selectivity of not less than 95%, preferably not less than 97%, more preferably not less than 99%, on an industrial scale of not less than 2 ton / hr, preferably not less than 3 ton / hr, more preferably not less than 4 ton / hr, for the dialkyl carbonate, and not less than 1.3 ton / hr, preferably not less than 1.95 ton / hr, more preferably not less than 2.6 ton / hr, for the diol, with a high yield stably for a prolonged period of time of not less than 1000 hours, preferably not less than 3000 hours, more preferably not less than 5000 hours, from a cyclic carbonate and an aliphatic monohydric alcohol.

CLAIMS

We claim:

1. A process for industrially producing a dialkyl carbonate and a diol in which the dialkyl carbonate and the diol are continuously produced through a reactive distillation system of taking a cyclic carbonate and an aliphatic monohydric alcohol as starting materials, comprising the steps of:

continuously feeding the starting materials into a continuous multi-stage distillation column in which a homogeneous catalyst is present;

carrying out reaction and distillation simultaneously in said column;

continuously withdrawing a low boiling point reaction mixture containing the 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, wherein:

said continuous multi-stage distillation column comprises a cylindrical trunk portion having a length L (cm) and an inside diameter D (cm) and having thereinside a tray with number of stages n , and comprises a gas outlet having an inside diameter d_1 (cm) provided 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_2 (cm) provided 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 said gas outlet, and at least one second inlet provided in the middle portion and / or the lower portion of the column above said liquid outlet, wherein:

(1) the length L (cm) satisfies the formula (1);

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

(2) the inside diameter D (cm) of the column satisfies the formula (2);

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

(3) a ratio of the length L (cm) to the inside diameter D (cm) of the column satisfies the formula (3);

$$4 \leq L / D \leq 40 \quad (3),$$

(4) the number of stages n satisfies the formula (4);

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

(5) a ratio of the inside diameter D (cm) of the column to the inside diameter d_1 (cm) of the gas outlet satisfies the formula (5);

$$3 \leq D / d_1 \leq 20 \quad (5),$$

(6) a ratio of the inside diameter D (cm) of the column to the inside diameter d_2 (cm) of the liquid outlet satisfies the formula (6);

$$5 \leq D / d_2 \leq 30 \quad (6), \text{ and}$$

(7) an aperture ratio of each tray is in a range of from 1.5 to 10%.

2. The process according to Claim 1, wherein a produced amount of the dialkyl carbonate is not less than 2 ton / hr.

3. The process according to Claim 1 or 2, wherein a produced amount of the diol is not less than 1.3 ton / hr.

4. The process according to any one of Claims 1 to 3, wherein said d_1 and said d_2 satisfy the formula (7);

$$1 \leq d_1 / d_2 \leq 5 \quad (7).$$

5. The process according to any one of Claims 1 to 4, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2300 \leq L \leq 6000$, $200 \leq D \leq 1000$, $5 \leq L / D \leq 30$, $30 \leq n \leq 100$, $4 \leq D / d_1 \leq 15$, and $7 \leq D / d_2 \leq 25$, respectively.

6. The process according to any one of Claims 1 to 5, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2500 \leq L \leq 5000$, $210 \leq D \leq 800$, $7 \leq L / D \leq 20$, $40 \leq n \leq 90$, $5 \leq D / d_1 \leq 13$, and $9 \leq D / d_2 \leq 20$, respectively.

7. The process according to any one of Claims 1 to 6, wherein the aperture ratio of each tray is in a range of from 1.7 to 8.0%.

8. The process according to any one of Claims 1 to 7, wherein the aperture ratio of each tray is in a range of from 1.9 to 6.0%.

9. The process according to any one of Claims 1 to 8, wherein said tray is a sieve tray having a sieve portion and a downcomer portion.

10. The process according to claim 9, wherein said sieve tray has 100 to 1000 holes / m² in said sieve portion thereof.

11. The process according to Claim 9 or 10, wherein a cross-sectional area per hole of said sieve tray is in a range of from 0.5 to 5 cm².

12. The process according to Claim 10 or 11, wherein a aperture ratio (a ratio of a total cross-sectional area of the holes to a total area of the sieve plate containing the area of the hole portion) of said sieve tray is in a range of from 1.9 to 6.0%.

13. A continuous multi-stage distillation column for carrying out transesterification between a cyclic carbonate and an aliphatic monohydric alcohol and distillation, the continuous multi-stage distillation column comprising:

a cylindrical trunk portion having a length L (cm) and an inside diameter D (cm);

a tray having number of stages n provided inside said trunk portion;

a gas outlet having an inside diameter d₁ (cm) provided at a top of said column or in an upper portion of said column near to the top;

a liquid outlet having an inside diameter d₂ (cm) provided at a bottom of said column or in a lower portion of said column near to the bottom;

at least one first inlet provided in the upper portion and / or a

middle portion of said column below said gas outlet; and

at least one second inlet provided in the middle portion and / or the lower portion of said column above said liquid outlet; wherein:

(1) the length L (cm) satisfies the formula (1);

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

(2) the inside diameter D (cm) of the column satisfies the formula (2);

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

(3) a ratio of the length L (cm) to the inside diameter D (cm) of the column satisfies the formula (3);

$$4 \leq L / D \leq 40 \quad (3),$$

(4) the number of stages n satisfies the formula (4);

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

(5) a ratio of the inside diameter D (cm) of the column to the inside diameter d_1 (cm) of the gas outlet satisfies the formula (5);

$$3 \leq D / d_1 \leq 20 \quad (5),$$

(6) a ratio of the inside diameter D (cm) of the column to the inside diameter d_2 (cm) of the liquid outlet satisfies the formula (6);

$$5 \leq D / d_2 \leq 30 \quad (6), \text{ and}$$

(7) an aperture ratio of each tray is in a range of from 1.5 to 10%.

14. The continuous multi-stage distillation column according to Claim 13, wherein said d_1 and said d_2 satisfy the formula (7);

$$1 \leq d_1 / d_2 \leq 5 \quad (7).$$

15. The continuous multi-stage distillation column according to

Claim 13 or 14, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2300 \leq L \leq 6000$, $200 \leq D \leq 1000$, $5 \leq L / D \leq 30$, $30 \leq n \leq 100$, $4 \leq D / d_1 \leq 15$, and $7 \leq D / d_2 \leq 25$, respectively.

16. The continuous multi-stage distillation column according to any one of Claims 13 to 15, wherein L , D , L / D , n , D / d_1 , and D / d_2 for said continuous multi-stage distillation column satisfy the following formulae; $2500 \leq L \leq 5000$, $210 \leq D \leq 800$, $7 \leq L / D \leq 20$, $40 \leq n \leq 90$, $5 \leq D / d_1 \leq 13$, and $9 \leq D / d_2 \leq 20$, respectively.

17. The continuous multi-stage distillation column according to any one of Claims 13 to 16, wherein the aperture ratio of each tray is in a range of from 1.7 to 8.0%.

18. The continuous multi-stage distillation column according to any of Claims 13 to 17, wherein the aperture ratio of each tray is in a range of from 1.9 to 6.0%.

19. The continuous multi-stage distillation column according to any one of Claims 13 to 18, wherein said tray is a sieve tray having a sieve portion and a downcomer portion.

20. The continuous multi-stage distillation column according to Claim 19, wherein said sieve tray has 100 to 1000 holes / m^2 in said sieve portion thereof.

21. The continuous multi-stage distillation column according to Claim 19 or 20, wherein a cross-sectional area per hole of said sieve tray is in a range of from 0.5 to 5 cm².

22. The continuous multi-stage distillation column according to Claim 20 or 21, wherein a aperture ratio (a ratio of a total cross-sectional area of the holes to a total area of the sieve plate containing the area of the hole portion) of said sieve tray is in a range of from 1.9 to 6.0%.